2020 MARY KAY O'CONNOR PROCESS SAFETY SYMPOSIUM

Beyond Regulatory Compliance: Making Safety Second Nature
In Association with IChemE

2020

PROCEEDINGS

MARY KAY O'CONNOR PROCESS SAFETY CENTER
TEXAS A&M ENGINEERING EXPERIMENT STATION

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WELCOME TO THE 2020 ANNUAL PROCESS SAFETY SYMPOSIUM PROCEEDINGS

I am delighted to welcome you to the 2020 Annual Symposium of the Mary Kay O'Connor Process Safety Symposium Proceedings. The 2020 Symposium is the 23th in the series and is offered virtually for the first time due to the impacts of the ongoing coronavirus pandemic. Our symposium honors the memories of our namesake, Mary Kay O'Connor, and our founding director, Professor M. Sam Mannan. The symposium is an important annual event that focuses on research, education, training and service issues that impact process safety and risk management. Your participation is very much appreciated, particularly in these trying time as we all struggle with the impacts of the coronavirus pandemic on our lives and day to day operations. Your involvement is crucial in making the symposium a success and to advance the cause of process safety technologies and concepts to the end of making industry safer. We believe that proactive improvements in process safety programs are good business and have a positive impact on the industries bottom line particularly in these difficult times.

The objectives for holding this annual symposium are three-fold. First, this annual event provides an independent and unbiased forum for exchange of ideas and discussion where industry, academia, government agencies and other stakeholders come together to discuss critical issues of research and advances in the field of process safety. Secondly, it provides an excellent platform for networking whereby process safety professionals can build peer-to-peer connections for future and also gain knowledge of the various services they can avail from others. Finally, we strongly believe that as we navigate the uncertain waters of COVID-19 today, good, robust research can help solve the complex and intriguing problems faced by the industry today. Identifying these problems and exchanging ideas and opinions with the expertise brought together through discussions at the symposium will provide context to help resolve the issues at hand

In addition, participants in the Symposium can also take this opportunity to benefit from being acquainted with the cutting-edge research done at the Mary Kay O'Connor Process Safety Center.

These proceedings contain the symposium program, the papers presented at the symposium and submitted before the deadline, and other informative items from the center.

We wish you maximum benefit from this symposium and strongly encourage you to participate in the virtual discussions. Please feel free to contact me or other center personnel with your ideas and input regarding the symposium and other activities of the center. We look forward to welcoming everyone to our face to face symposium at Texas A&M University in October 2021. Best wishes for a safe return to normalcy.

Stewart W. Behie, Ph.D., P.Eng.
Interim Director, Mary Kay O'Connor Process Safety Center
Professor of Practice, Artie McFerrin Department of Chemical Engineering
Texas A&M University

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- Process Safety Risk Management Plan
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Process Safety in the 21st Century

As an industry, our inability to learn from past incidents and demonstrate that process safety is improving has led to the project Process Safety in the 21st Century and Beyond. The aim of this project is to envision better process safety by outlining efforts that each stakeholder can take.
How was the project undertaken?

Gaining a global perspective of the key challenges in process safety is the first important step. The challenges were considered across four stakeholders: industry, academia, regulators, and society. To determine the challenges, a series of workshops at international symposia were undertaken, including in the UK (with input from other European countries), North America, Asia, Australia/New Zealand, and the Middle East. Various methods of consultation were used, but the key questions remained consistent. In process safety:

- What are the key industry challenges?
- What are the key academic challenges?
- What are the key regulatory challenges?
- What are the key societal challenges?

These questions were answered by professionals from various levels in industry, academia, and regulatory bodies. Once the challenges were identified, a top five list was drawn up for each stakeholder group.

Our goal with this document is to lay out a series of actions to be undertaken at various levels and across all stakeholders to improve process safety because people have a right to not get hurt. To enable this vision, this roadmap is a call to action to all stakeholders and not just process safety professionals.
We invite you to look at the opportunities and think about how you can influence them and positively impact process safety. Every professional is obliged to improve process safety because engineering and science are essential to us all and it must be sustainable in all senses of the word, including process safety. If we, as engineers, do not develop new strategies for continuous improvement, the engineering profession will become irrelevant to society and the need for process safety will become extinct, thus increasing process safety incidents. A question that needs to be answered is where this roadmap is intended to take us. The simple answer is that the roadmap and the associated journey are focused towards improvements in process safety performance, which will ultimately lead us to our vision of zero incidents.

In Association with **IChemE**

The Institution of Chemical Engineers (IChemE) is the global professional membership organization for chemical, biological and process engineers and other professionals involved in the chemical, process and bioprocess industries. With a membership exceeding 44,000 members in over 120 countries, and offices in Australia, New Zealand, Singapore, Malaysia and the UK; IChemE aims to be the organization of choice for chemical engineers.

We promote competence and a commitment to the best practice, advance the discipline for the benefit of society and support the professional development of our member. We are the only organization licensed to award Chartered Chemical Engineer and Professional Process Safety Engineer status.

IChemE exists because chemical engineering matters.

OUR MISSION

IChemE's four key aims are:

- To build and sustain an active international professional community, united by a commitment to qualifications and standards that foster excellence and the delivery of benefits to society.
- To engage with others to promote development, understanding of chemical engineering and an appreciation of its importance.
- To provide support and services to individuals, employers and others who contribute to improving the practice and application of chemical engineering.
- To enable chemical engineers to communicate effectively with each other and with other disciplines.

To support these aims, we operate as an effective, efficient and responsive organization, providing leadership and demonstrating good practice as well as complying with our obligations as a charitable organization.

IChemE is a registered charity in England & Whales (214379) and a charity registered in Scotland (SC 039661).

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Awards and Scholarships

Trevor Kletz Merit Award

The Merit Award recognizes an individual who has made significant contributions to the advancement of education, research, or service activities related to process safety concepts and/or technologies. The contributions or accomplishments leading to the annual Merit Award need not be associated with the Center but must fit within the central theme of the Center, i.e., Making Safety Second Nature. In establishing the Merit Award, the Steering Committee underscores the importance of promoting and recognizing significant contributions and accomplishments of practitioners and researchers worldwide.

The Harry H. West Memorial Service Award

The Service Award was established by the Steering Committee to honor and recognize individuals who have contributed directly to the success of the Center and have played a significant role in advancing the mission of the Center.

Lamiya Zahn Memorial Safety Scholarship

On July 31, 2004, an explosion and fire occurred in a university apartment on the Texas A&M University campus. Four members of the family of Saquib Ejaz, a chemical engineering graduate student - were critically injured and hospitalized. Saquib's mother and his four-year old daughter, Lamiya Zahin subsequently died from injuries sustained in the fire.

In fond and living memory of Lamiya, the Department of Chemical Engineering and the Mary Kay O'Connor Process Safety Center have established the Lamiya Zahin Memorial Safety Scholarship. Graduate students are encouraged to apply for the $1,000 scholarship by writing a 1000-word essay on "Safety Innovations in Research Projects".

Trevor Kletz Merit Award Recipient

Dr. William (Bill) Rogers, TAMU Safety Engineering Lecturer

Bill is currently a Lecturer at the Department of Chemical Engineering, Texas A&M University and has been associated with various research activities at MKOPSC for over twenty years. He has published and continues to publish numerous peer-reviewed articles on process safety and risk management and has been teaching Risk Analysis and Quantitative Risk Analysis at for the last 10 years. Bill developed several Safety Engineering courses at the Center and taught them at different times and has inspired thousands of undergraduate and graduate students with his enthusiasm and passion for process safety and his unparalleled dedication to teaching. His key contribution to process safety has been disseminating the importance of quantitative risk assessment in engineering problems to his students. Each student passing his class enters the industry knowing the importance of "uncertainty" in risk assessment. There is probably no other educator with a greater number of students in process safety than Bill. In his silent way, Bill continues to pave the way to safer processes by imparting wisdom on the fundamentals of QRAs among the large number of his students.

Harry West Service Award Recipients

Jeff Thomas, MKOPSC Technical Advisory Committee Member

Research Fellow and Volunteer Mentor, Mary Kay O'Connor Process Safety Center

Jeff has been a long-time supporter of the Center and has worked hard in providing direction through the Steering Committee and Technical Advisory Committee and special projects related to process safety. Jeff has also been an active participant in the MKOPSC annual symposium serving on the technical program committee for the past few years as well as serving as track chairs, reviewing presentations, and helping coordinate activities to make this event successful. He engages with the students and is always keen to offer assistance

Lamiya Zahin Memorial Safety Scholarship Recipient

Cassio Brunoro Ahumada
PhD Chemical Engineering Student, Texas A&M University
Graduate Research Assistant, MKOPSC

Cassio Brunoro Ahumada is a doctoral candidate in the Chemical Engineering Department. He holds a Master's in Chemical Engineering from Texas A&M and a Bachelor's in chemical engineering from the Federal University of Espirito Santo, Brazil.

His research investigates how the congestion pattern variation affects the deflagration-to-detonation transition (DDT) mechanism on flammable gaseous mixtures. He is also involved in many safety-related projects, including facility risk assessments, facility siting, and vapor cloud explosion modeling. During his time as a graduate student at TAMU, he interned at Tesla's car manufacturing site and Wood PLC, conducting activities related to process safety management and technical consulting.

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We appreciate all of our member companies and its representatives. Their expertise in our Steering and Technical Advisory Committee is essential to the success of the Center. Email us at mkopsc@tamu.edu if you would like to become a member.

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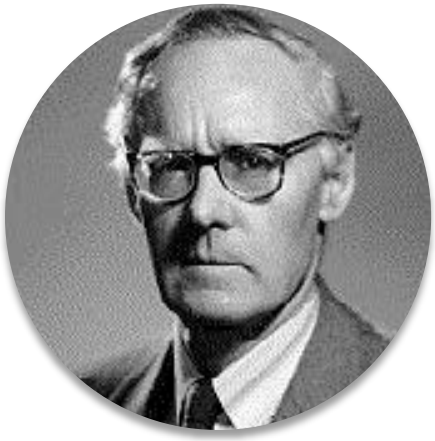
Frank P. Lees Memorial Lecture



Frank P. Lees
1931-1999
Emeritus Professor of
Chemical Engineering
Loughborough University,
United Kingdom

Frank P. Lees is noted for his monumental three-volume work, Loss Prevention in the Process Industries (2nd edition, 1996), an extraordinary accomplishment for one man and an outstanding compendium of our present knowledge of process safety. It is not, however, a scissors and paste job, a mere collection of other people's thoughts; he thoroughly surveyed and evaluated present knowledge and provided his own comments on it. Lees also produced an immense amount of original work, particularly on QRA, HAZOP, consequence analysis and computer applications, and trained a generation of students to follow in his footsteps.

The Mary Kay O'Connor Process Safety Center recognizes the contributions made by Frank Lees to the field of process safety and loss prevention. His teachings and findings will be the guiding light and inspiration for many in this generation and future.

Dr. Katherine Lemos delivered a keynote session as the 2020 Frank P. Lees Memorial Lecture.

Day 1 Keynote Speaker

Keynote Lecture: *Mission-Oriented Leadership*

Katherine A. Lemos, Ph.D.

Chairperson and CEO
U.S. Chemical Safety Board

Katherine A. Lemos, Ph.D. was nominated by the President in June 2019 and confirmed by the Senate in March 2020.

Prior to her confirmation as Chairperson and CEO of the U.S. Chemical Safety Board (CSB) Dr. Lemos served as Director for Northrop Grumman Corporation's Aerospace Sector, driving performance improvements across the product lifecycle with a focus on engagement early in the value stream.

As an expert in accident investigation, human decision-making and safety management, she is known for her innovative and strategic approaches leveraging advances in analytics and autonomy. She has a documented record of turning new technologies into solutions trusted by operators, overseers, and the public they serve.

Before joining Northrop Grumman in 2014, Dr. Lemos worked at the Federal Aviation Administration (FAA) as a technical leader and program manager in Aircraft Certification and Aviation Safety. Prior to this she worked for the National Transportation Safety Board (NTSB) as a Senior Human Performance Investigator in Aviation Safety, and then as Special Assistant to Vice Chairman of the Board.

In academia Dr. Lemos focused her research on decision-making, studying the influence of information and technology on beliefs and behaviors to more reliably yield safe outcomes during risky and uncertain conditions. In aviation, Dr. Lemos conducted applied research to balance the strengths of technology and humans for optimal performance. Dr. Lemos earned a B.B.A. from Belmont University, a M.S. from California Lutheran University, and a Ph.D. from the University of Iowa.

Throughout her career, Dr. Lemos has focused on improving safety and efficiency at the level of the individual and the organization. She has contributed individually as a researcher, professor and technical expert, and also contributed as a leader in managing programs and initiatives, bringing consensus and order to efforts that result in tangible safety and efficiency outcomes.

Day 2 Plenary Panelists

Plenary Panel: *Integrating Pandemic Preparedness and Response Into Business Continuity and Risk Management Planning*

Gerald Parker

Director, Pandemic and Biosecurity Policy Program,

Scowcroft Institute, Bush School of Government and Public Services

Paul Thomas

Vice President, Health, Environment, Safety & Security,

Occidental Chemical Corp.

Malick Diara

Public Health Manager, Workplace Infectious Disease Control Manager,

ExxonMobil

Richard Wells

VP Gulf Coast Operations,

Dow Chemical Corporation

Stewart Behie, Moderator

Interim Director,

MKOPSC

Symposium Coordinators, MKOPSC

Paola Camposeco
paolacamposeco@tamu.edu

Brandy Tuck
brandytuck@tamu.edu

Symposium Student Assistants

Aubrey Bilicek
Aubrey694@tamu.edu

Tanner Thornton
tannerthornton@tamu.ed

Laurel Patterson
laurel0601@tamu.edu

Lauren Guerra
laurenguerra@tamu.edu

Symposium Track Chairs

The Mary Kay O'Connor Process Safety Center would like to recognize and thank the **Track Chairs** who have volunteered their time to assist with the abstract review, selection process, and session coordination. Their input, expertise, and leadership have been essential to the Symposium's success.

Day 1, Track I Chairs: Risk/Consequence Analysis & Design Aspects

Robert, Dow
RJBellair@dow.com

Robin Pitblado, DNV GL
robinpitblado@gmail.com

Day 1, Track II Chairs: Human Factors-People In action

Mindy Bergman, TAMU
Mindybergman@tamu.edu

Ranjana Mehta, TAMU
rmehta@tamu.edu

Farzan Sasangohar, TAMU
sasangohar@tamu.edu

Day 1, Track III Chairs: Managing Operations and Maintenance

Jeff Thomas
jjt1234@aol.com

Trish Kerin, IChemE
TKerin@icheme.org

Day 1 and Day 2 Keynote and Panel Moderator

Stewart Behie, MKOPSC
Stewart_behie@tamu.edu

Day 2, Track I Chairs: Risk/Consequence Analysis & Design Aspects

Brenton Drake, Dow
bdrake1@dow.com

Marisa Pierce, DNV GL
mmarisa.pierce@dnvgl.com

Day 2, Track II Chair: Human Factors-People In action

Camille Peres, TAMU
peres@tamu.edu

Day 2, Track IV Chair: Research and Next Generation

Nick Gonzales, Shell
nick.gonzales@shell.com

Day 2, Track V Chairs: Explosions and Flammability

Delphine Laboureur,
Von Karman Institute
delphinelaboureur@gmail.com

Chris Cloney, Dust Safety Science
chris@dustsafetyscience.com

Symposium Technical Support Team

The Mary Kay O'Connor Process Safety Center would like to recognize and thank the **Technical Support Team** who put in countless of hours to make the virtual symposium successful. They are the ones behind the scenes who took on the challenge of tackling the set up and handle of the virtual sessions. Their virtual session research and coordination was integral to the Symposium's success.

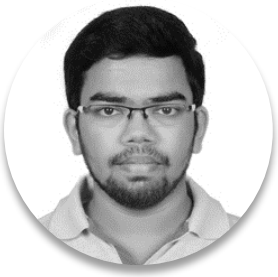
Henry Goyette, MKOPSC
Technical Program Coordinator
hgoyette1@outlook.com

Harold Escobar, MKOPSC
hu.escobar2208@tamu.edu

Jesus Alvarez, MKOPSC
jesusalvarez@tamu.edu

Noemi Contreras, MKOPSC
noemicontreras@tamu.edu

Zhuoran Zhang, MKOPSC
zhuoranzhang@tamu.edu



Bushan Pawar, MKOPSC
bhushan.pawar@tamu.edu

Sankhadeep Sarkar, MKOPSC
sankhadeep.sarkar@tamu.edu

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DAY 1: TUESDAY, OCTOBER 20 | Virtual Symposium

8:00AM	**Welcome & State of the Mary Kay O'Connor Process Safety Center — Dr. Stewart Behie, Interim Director, MKOPSC**		
	Welcome Webinar — Session Room D		
8:30AM	**Break**		
	Track I: Risk/Consequence Analysis & Design Aspects	**Track II: Human Factors—People in Action**	**Track III: Managing Operations and Maintenance**
	Session Room A	**Session Room B**	**Session Room C**
	Risk Assessment I **Session Chair: Rob Bellair**	**Training/Engagement** **Session Chair: Mindy Bergman**	**Modeling and Asset Integrity** **Session Chair: Jeff Thomas**
8:45AM	Importance of Process Safety Time in Design **Shanmuga Prasad Kolappan,** *TechnipFMC*	Session Break	*RBI Study using Advanced Consequence Assessment for Topside Equipment on Offshore Platforms* **Chetan Birajdar,** *Monaco Engineering Solutions*
9:15AM	Limitations of Layers of Protection Analysis (LOPA) in Complicated Process Systems **Abdulaziz Alajlan,** *Saudi Aramco*		Indicators of an Immature Mechanical Integrity Program. **Derek Yelinek,** *Siemens Process & Safety Consulting*
9:45AM	On the Usage of Ontologies for the Automation of HAZOP Studies **Johannes I. Single**, *CSE Center of Safety Excellence*	Virtual Reality Process Safety in Counterfactual Thinking **Kianna Arthur,** *Texas A&M University*	Remember the à la Mode: Lessons Learned from Ammonia Release at Frozen Foods Warehouse **Matthew S. Walters,** *Exponent, Inc.*
10:15AM	**Break**		
	Risk Assessment II **Session Chair: Robert Bellair**	**Human Performance/Decision Making I** **Session Chair: Mindy Bergman**	**Recalling and Learning from Incidents** **Session Chair: Jeff Thomas**
10:30AM	An Efficient and Effective Approach for Performing Cost Benefit Analysis, with Two Case Studies **Henrique (Henry) M. Paula,** *Galvani Risk Consulting, LLC*	Is Attentional Shift the Problem (or something else) with Hazard Statement Compliance? An Experimental Investigation Using Eye-Tracking Technology **S. Camille Peres,** *Texas A&M University*	Process Related Incidents with Fatality—Trends and Patterns **Syeda Zohra Halim,** *MKOPSC*
11:00AM	Does Your Facility Have the Flu? How to Use Bayes Rule to Treat the Problem instead of the Symptom. **Keith Brumbaugh,** *aeSolutions*	Risk management entails decision making: Does decision making in complex situations come down to somebody's gut feeling? **Hans Pasman,** *MKOPSC*	Application of Mind Mapping to Organize and Recall Potential Hazards. **T. Michael O'Connor,** *MKOPSC*
11:30AM	Integrating the PHA and FSS into a Site Risk Assessment Life Cycle. **Colin Armstrong and Sam Aigen,** *AcuTech*	Decision Making using Human Reliability Analysis **Fabio Kazuo Oshiro,** *Monaco Engineering Solutions*	Would a HAZOP, LOPA, or STPA have Prevented Bhopal? **Howard Duhon,** *GATE Energy*
12:00PM	**Lunch Break**		
1:00PM	**Keynote Address: "Mission-Oriented Leadership" by Katherine A. Lemos, Ph.D.** **Chairperson and CEO, U.S. Chemical Safety Board**		
	Keynote Webinar — Session Room D		
2:00PM	**Break**		
	Track I: Risk/Consequence Analysis & Design Aspects	**Track II: Human Factors—People in Action**	**Track III: Managing Operations & Maintenance**
	Session Room A	**Session Room B**	**Session Room C**
	SIS—LOPA **Session Chair: Robin Pitblado**	**Safety Culture and Leadership** **Session Chair: Ranjana Mehta**	**Improving Process Safety with Technological Advances** **Session Chair: Trish Kerin**
2:15PM	A Framework for Automatic SIS Verification in Process Industries using Digital Twin **Nitin Roy,** *California State University, Sacramento*	Improving Industry Process Safety Performance through Responsible Collaboration **Ryan Wong,** ExxonMobil Research and Engineering; and **Shanahan Mondal**, CVR Energy	Predictive Process Safety Analytics and IIoT - PSM Plus: The AI+PSM Analytic Framework **Michael Marshall,** *Tratus Group*
2:45PM	The use of Bayesian Networks in Functional Safety **Paul Gruhn,** *aeSolutions*	How Much Does Safety Culture Change Over Time? **Stephanie C. Payne,** *Texas A&M University*	Guidance to Improve the Effectiveness of Process Safety Management Systems in Operating Facilities **Syeda Zohra Halim,** *MKOPSC*
3:15PM	My Vision of Future Instrumented Protective Systems **Greg Hall,** *Eastman Chemical Company*	Administering a Safety Climate Assessment in a Multicultural Organization: Challenges and Findings **Atif Mohammed Ashraf,** *Texas A&M University*	Unified Wall Panel System (UWPS) - A Value Engineering Solution for Protective Construction in the Petroleum Industry **Scott Hardesty,** *Applied Research Associates*
3:45PM	**Break**		
	Relief Systems **Session Chair: Robin Pitblado**	**Procedures** **Session Chair: Farzan Sasangohar**	**Exploring NaTech Events and Domino Impacts** **Session Chair: Trish Kerin**
4:00PM	Overlooked Reverse Flow Scenarios **Gabriel Martiniano Ribeiro de Andrade, Chris Ng and Derek Wood,** *Siemens Process & Safety Consulting*	A Comparison of Procedure Quality Perceptions, Procedure Utility, Compliance Attitudes, and Deviation Behavior for Digital and Paper Format Procedures **Joseph W. Hendricks,** *Texas A&M University*	Protect Process Plants From Climate Change **Victor Edwards,** *VHE Technical Analysis*
4:30PM	Failure Under Pressure: Proper Use of Pressure Relief Device Failure Rate Data based on Device Type and Service **Todd W. Drennen,** *Baker Engineering and Risk Consultants (BakerRisk)*	Practical Writing Tips To Prevent Human Error When Following Procedures **Monica Philippart,** *Ergonomic Human Factors Solutions*	Process Safety Implications in a Changing Environment **Trish Kerin,** *IChemE Safety Centre*
5:00PM	Additional Engineering and Documentation to Reduce Pressure Relief Mitigation Cost **Gabriel Martiniano Ribeiro de Andrade, and Kartik Maniar,** Siemens	The Impact of Hazard Statement Design in Procedures on Compliance Rates: Some Contradictions to Best (or Common) Practices **Joseph W. Hendricks,** *Texas A&M University*	A Critical Evaluation of Industrial Accidents Involving Domino Effect **Ravi Kumar Sharma,** *Indian Institute of Technology - Roorkee*

DAY 2: WEDNESDAY, OCTOBER 21 | Virtual Symposium

8:00AM	**Welcome & Mary Kay O'Connor Process Safety Center Awards — Dr. Stewart Behie, Interim Director, MKOPSC**		
	Welcome Webinar — Session Room D		
8:15AM	**Break**		
	Track I: Risk/Consequence Analysis & Design Aspects	**Track II: Human Factors—People in Action**	**Track V: Explosions and Flammability**
	Session Room A	**Session Room B**	**Session Room C**
	Risk Assessment III **Session Chair: Brenton Drake**	**Human Performance/Decision Making II** **Session Chair: Camille Peres**	**Explosion Modelling** **Session Chair: Delphine Laboureur**
8:30AM	Applying PHA Methodologies such as HAZOP and Bowtie to Assessing Industrial Cybersecurity Risk **John Cusimano,** *aeSolutions*	Preventing Cognitive-Attributed Errors in Safety Critical Systems: A Path Forward **Tom Shephard,** *Wood*	The Influence of the Velocity Field on the Stretch Factor and on the Characteristic Length of Wrinkling of Turbulent Premixed Flames **Tássia L. S. Quaresma,** *University of Campinas*
9:00AM	Large Hydrocarbon Tank Fires: Modelling of the Geometric and Radiative Characteristics **Ravi Kumar Sharma,** *Indian Institute of Technology - Roorkee*	Two Views of Evaluating Procedural Task Performance: A Transition from Safety-I to Safety-II Approach **Changwon Son,** *Texas A&M University*	Towards a Comprehensive Model Evaluation Protocol for LNG Hazard Analyses **Filippo Gavellia,** *Blue Engineering and Consulting*
9:30AM	Risk assessment of a large chemical complex during the construction phase using Intuitionistic Fuzzy Analytic hierarchy process. **Suresh G,** *Bharat Petroleum Corporation, Kochi Refinery*	Beyond Human Error: Integration of the Interactive Behavior Triad and Toward a Systems Model **Joseph W. Hendricks,** *Texas A&M University*	Beirut: How behaves Ammonium Nitrate Exposed to Fire and How Strong and Damaging is its Explosion? **Charline Fouchier,** *von Karman Institute of Fluid Dynamics*
10:00AM	**Break**		
	Risk Mitigation **Session Chair: Brenton Drake**	**Fatigue and Stress** **Session Chair: Camille Peres**	**Explosion Phenomena I** **Session Chair: Delphine Laboureur**
10:15AM	Development of Resilient LNG Facilities **Onder Akinci,** *Daros Consulting*	Operator Performance Under Stress: A Neurocentric Virtual Reality Training Approach **Ranjana Mehta,** *Texas A&M University*	Flammable Mist Hazards Involving High-Flashpoint Fluids **Simon Gant,** *UK Health and Safety Executive*
10:45AM	Development of Risk Mitigation Programs using a Quantitative-Risk-Based Approach **Rafael Callejas-Tovar,** *BakerRisk*	Towards a Predictive Fatigue Technology for Oil and Gas Drivers **John Kang,** *Texas A&M University*	Measuring Suspended Explosive Dust Concentration from Images **Yumeng Zhao,** *Purdue University*
11:15AM	Incorporating Mitigative Safeguards with LOPA **Edward Marszal,** *Kenexis*	Validation of the Fatigue Risk Assessment and Management in High-Risk Environments (FRAME) Survey **Stefan V. Dumlao,** *Texas A&M University*	The HBT-A Large-Scale Facility for Study of Detonations and Explosions **Elaine S. Oran,** *Texas A&M University*
11:45AM	**Lunch Break**		
	Track I: Risk/Consequence Analysis & Design Aspects	**Track IV: Research and Next Generation**	**Track V: Explosions and Flammability**
	Session Room A	**Session Room B**	**Session Room C**
Sessions	**Consequence Analysis: Gas Release** **Session Chair: Marisa Pierce**	**Next Generation Process Safety I** **Session Chair: Nick Gonzales**	**Explosion Phenomena II** **Session Chair: Chris Cloney**
12:45PM	Hole Size Matters **Jeffrey D. Marx,** *Quest Consultants Inc.*	Identifying contributing factors of pipeline incident from PHMSA database based on NLP and text mining techniques. **Guanyang Liu,** *MKOPSC*	Development of Flammable Dispersion Quantitative Property-Consequence Relationship Models Using Machine Learning. **Zeren Jiao,** *MKOPSC*
1:15PM	How Can I Effectively Place My Gas Detectors **Jesse Brumbaugh,** *aeSolutions*	Causation analysis of pipeline incidents using artificial neural network (ANN) **Pallavi Kumari,** *MKOPSC*	An Unsupervised Model to Predict the Liquid In-cylinder Combustion Risk Ratings of Marine Fuels **Chenxi Ji**, *MKOPSC*
1:45PM	Consequence Assessment Considerations for Toxic Natural Gas Dispersion Modeling **SreeRaj Nair, Noma Ogbeifun,** *Chevron - MCBU*	Development of Hazard Factor for Engineered Particles **Nabila Nazneen,** *MKOPSC*	Fireball and Flame Venting Comparisons **Peter A. Diakow,** *BakerRisk*
2:15PM	**Break**		
2:30PM	**Plenary Panel: Integrating Pandemic Preparedness and Response Into Business Continuity and Risk Management Planning** **Panelists: Gerald Parker,** Director, Pandemic and Biosecurity Policy Program, Scowcroft Institute, Bush School of Government and Public Services, TAMU; **Paul Thomas,** VP Health, Environment, Safety & Security, OxyChem; **Malick Diara,** Public Health Manager, Workplace Infectious Disease Control Manager, ExxonMobil; and **Richard Wells,** VP Gulf Coast Operations, Dow Chemical Corporation. **Moderator:** Stewart Behie, Interim Director, MKOPSC		
	Panel Webinar — Session Room D		
3:45PM	**Break**		
	Session Room A	**Session Room B**	**Session Room C**
	Reactive Chemicals **Session Chair: Marisa Pierce**	**Next Generation Process Safety II** **Session Chair: Nick Gonzales**	**Consequence Analysis: Flammability** **Session Chair: Chris Cloney**
4:00PM	Modelling and Simulation to Predict Energetic Material Properties **Kok Hwa Lim,** *Singapore Institute of Technology*	Can a Virtual Reality Application Better Prepare Millennials and the Z-Generation for Working with Systems in the Process Industry? **Nir Keren,** *Iowa State University*	Numerical Simulation of Methane-Air DDT in Channels Containing Trace Amounts of Impurities **Logan N. Kunka,** *Texas A&M University*
4:30PM	Safety Assessment of Low Temperature Radical Initiators for Proper Storage and Safe Handling Conditions **Cuixian (Trisha) Yang,** *Merck & Co*	Process Safety Risk Index Calculation Based on Historian Data **Prasad Goteti,** *Honeywell Process Solutions*	The Use of Bent Poles as a Detonation Indicator **J. Kelly Thomas,** *BakerRisk*
5:00PM	Analysis of Pressure Behavior during Reaction Runaway and Estimation of Available Depressurization Design. **Yuto Mizuta,** *Mitsubishi Chemical*	A Brief Review of Intrusion Detection in Process Plants and Advancement of Machine Learning in Process Security. **Sinijoy P J,** *Cochin University of Science and Technology*	Machine Learning Based Quantitative Prediction Model for Chemical Mixture Flammability Limits **Zeren Jiao,** *MKOPSC*

DAY 1: Tuesday Oct 20 Summaries

Track I - Risk/Consequence & Design Aspects
Category: Risk Assessment I Session

8:45 AM to 9:15 AM
Importance of Process Safety Time in Design

Speaker: Shanmuga Prasad Kolappan

This work demonstrates the importance of this Process Safety Time and its role in the design of the safety protection system, in particular the factor of SIF response time to achieve potential risk mitigation. Although the Process Safety time is complex predict, its importance is being demonstrated with a simple example.

9:15 AM to 9:45 AM
Limitations of Layers of Protection Analysis (LOPA) in Complicated Process Systems

Speaker: Abdulaziz Alajlan

This presentation presents a case study showing the limitations of applying LOPA in upstream scenarios to develop protection layers requirements for a complicated network of pipelines and processing units with unlimited number of causes contributing to the risk. It compares LOPA with the more sophisticated, more quantified other techniques such as Fault Tree Analysis (FTA). Based on the cases analysis, it is recommended that LOPA can be used to assess simple scenarios with limited number of causes, while more complicated cases are better assessed using FTA. Detailed analysis is presented in the paper to support such recommendation.

9:45 AM to 10:15 AM
On the Usage of Ontologies for Computer-aided HAZOP Studies

Speaker: Johannes. I. Single

The presentation is about the usage of ontologies for the automation of HAZOP studies. Therefore, the questions are answered why computer-aid should be provided and what added value can be provided. After that, the process and plant representation, as well as the representation of expert knowledge via ontologies, is explained. Special attention is paid to the representation of expert knowledge and the automatic drawing of conclusions. The necessary steps are explained by means of a hexane storage tank. Furthermore, the automatic generation of HAZOP worksheets based on an inference algorithm is presented and discussed. The results are summarized at the end of the presentation.

Category: Risk Assessment II Session

10:30 AM to 11:00 AM
An Efficient and Effective Approach for Performing Cost Benefit Analysis, with Two Case Studies

Speaker: Henry M. Paula

Cost benefit analysis is a powerful tool to help managers sort through the recommendations and effectively/efficiently prioritize them. It consists of evaluating the risk reduction and the estimated cost associated with each recommendation, including Capital Expenditures (CAPEX) and Operational Expenditures (OPEX). This presentation provides a simple, efficient, and effective approach for performing cost benefit analysis, illustrated through two case studies. This method is not intended to replace more detailed methodologies. Rather, it is a complementary tool particularly useful for applications with many recommendations.

11:00 AM to 11:30 AM
Does your facility have the flu? Use Bayes rule to treat the problem instead of the symptom

Speaker: Keith Brumbaugh

Stop pretending to meet ultra low targets. Use Bayes theory to identify problems and prioritize the management of independent protection layers (IPLs).

11:30 AM to 12:00 PM
Integrating the PHA and Facility Siting into a Site Risk Assessment Life-Cycle

Speakers: Colin Armstrong and Sam Aigen

PHAs provide an accepted framework in organizations which details scenarios to be evaluated, credible safeguards, and the organization's acceptable risk criteria. Siting studies may consider risk in the same way as PHAs, but organizations typically fail to align the two assessments.

Category: SIS – LOPA Session

2:15 PM to 2:45 PM

A Framework for Automatic SIS Verification in Process Industries using Digital Twin

Speaker: Nitin Roy

Increasing complexity of distributed control systems (DCS) and control logics has made (safety instrumented systems) SIS validation complex and time-consuming. IEC and ISA safety standards recommend comprehensive logic checks of Safety instrumented functions. It can take months to check logic in delivered product. This work introduces automated testing of logic in process plants using Digital Twins. This method makes the process efficient and saves considerable amount of time, manpower and in turn capital. The verification which takes months can be reduced to weeks. It also ensures the verification is comprehensive and accurate making the system safer. In this work we also review the current practices in SIS verification and future improvements.

2:45 PM to 3:15 PM

The use of Bayesian Networks in Functional Safety

Speaker: Paul Gruhn

Functional safety engineers follow the ISA/IEC 61511 standard and perform calculations based on random hardware failures. These result in very low failure probabilities, which are then combined with similarly low failure probabilities for other safety layers, to show that the overall probability of an accident is extremely low (e.g., 1E-5/yr). Unfortunately, such numbers are based on frequentist assumptions and cannot be proven. Yet accidents are not caused by random hardware failures, they are typically the result of steady and slow normalization of deviation (a.k.a. drift). Bayes' theorem can be used to update our prior belief (the initial calculated failure probability) based on observing other evidence (e.g., the effectiveness of the facility's process safety management process). The results can be dramatic.

3:15 PM to 3:45 PM

My Vision of Future Instrumented Protective Systems

Speaker: J. Gregory Hall

I will share my vision of what future Instrumented Protective systems IPS will look like and what is our current objective to achieve that future.

Category: Relief Systems Session

4:00 PM to 4:30 PM

Overlooked Reverse Flow Scenarios

Speakers: Gabriel Martiniano Ribeiro de Andrade, Christopher Ng, Derek Wood

Reverse flow scenarios due to latent check valve failure are critical in the design of relief systems, but often overlooked or incompletely evaluated. This type of scenario is often controlling for relief device sizing, especially for systems involving high differential pressure across pumps or compressors. This presentation reviews current industry best practices to evaluate such scenarios. Specific application examples are then presented to highlight key aspects for the analysis, including identification of pressure sources as well as potential paths for reverse flows, location of and credit for relief devices, initiating events, and limiting basis for system pressure rating. In addition, potential to relieve both forward plus reverse flow simultaneously should be evaluated. Guidance is also provided to determine if vapor, liquid, or two-phase relief should be expected and whether liquid displacement or other non-obvious backflow from utility header should also be considered. Criteria to allow credit for system settle-out pressure, if applicable, and how to evaluate such credit are also provided. As system complexity increases, tips on how to estimate relief loads accurately and efficiently are also provided. Lastly, consideration of other safeguards beyond relief devices for high-risk cases is also discussed.

4:30 PM to 5:00 PM

Failure under Pressure: Proper Use of Pressure Relief Device Failure Rate Data based on Device Type and Service

Speaker: Todd W. Drennen

Component failure rate data is used in a variety of quantitative and semi-quantitative study methods related to process safety and reliability, including Fault Tree Analysis (FTA) and Layers of Protection Analysis (LOPA). In each of these methodologies, failure rate data is used to determine the probability that specific protective components, such as pressure relief devices, will fail to function as designed when called upon to prevent an incident. In the case of pressure relief devices, standardized probabilities of failure on demand are often applied with minimal consideration of the device type or the process service in which the device is employed. This presentation will examine pressure relief device failure rate data from multiple published sources, categorize the data based on device type and service, and then develop guidelines for determining probability of device failure on demand based on the proposed device type and service categories. Additionally, this presentation will provide commentary on the administrative aspects of relief device handling relative to observed relief valve reliability.

5:00 PM to 5:30 PM

Additional Engineering and Documentation to Reduce Pressure Relief Mitigation Cost

Speakers: Gabriel Martiniano Ribeiro de Andrade and Kartik Maniar

Unit revalidation and baseline studies for pressure relief analysis can result in a long list of potential deficiencies, which may result from an increase in unit throughput, changes to industry guidance or standards, changes to company internal guidelines for such studies, conservative assumptions in the absence of required data or based on simplified initial approach, management of change (MOC) at system or unit level, or may be a combination of all these. This presentation addresses what kind of additional engineering tools or processes can be applied on typical systems during revalidation studies, such as reactor loops, columns, turbines and heat exchangers, to ensure a more accurate representation of the relief scenarios to validate the deficiencies. In addition, the paper addresses what improvements in MOC processes can be implemented in order to capture, assess, and reduce the cumulative adverse effect to unit pressure relief analysis due to changes.

Track II - Human Factors – People in Action **Category**: Training/Engagement Session

9:45 AM to 10:15 AM

Virtual Reality Process Safety in Counterfactual Thinking

Speaker: Kianna Arthur

Counterfactual thoughts (i.e., "If only...") are common thoughts of the mental landscape. The current study is interested in further investigating if these thoughts can be used to improve workplace safety and future adherence to procedures.

Category: Human Performance/Decision Making I Session

10:30 AM to 11:00 AM

Is Attentional Shift the Problem (or something else) with Hazard Statement Compliance? An Experimental Investigation Using Eye-Tracking Technology

Speaker: S. Camille Peres

The current study utilized eye-tracking technology to determine whether or not participants are attending to hazard statements based on two different exemplar designs that have yielded the largest gap in hazard statement compliance. In other words, do we observe significant differences in attention to hazard statements based on a few predominant design characteristics (i.e., warning icon, yellow highlighting, numbering, and borders)?

11:00 AM to 11:30 AM

Risk management entails decision making: Does decision making in complex situations come down to somebody's gut feeling?

Speaker: Hans J. Pasman

Every day, we take hundreds of decisions and perhaps, all guided by intuition. However, if it concerns a decision to choose an option from several alternatives that must fulfil a number of requirements, and where a wrong decision can have disastrous consequences, a rational method should be preferred. The presentation presents an overview of available decision aiding methods suitable to apply at risk management.

11:30 AM to 12:00 PM

Decision Making using Human Reliability Analysis

Speaker: Fabio Kazuo Oshiro

In this presentation the human error probability was quantified using standardized methods. The study was based on the evaluation of some methodologies of human reliability and decision making. The method was assessed through a case study of an accident occurred in 2004 at Formosa Plastics Corp. Illiopolis. Initially, an analytical method was developed as Hierarchical Task Analysis (HTA), then by Predictive Human Error Analysis (PHEA) and a qualitative analysis using Systems for Predicting Human Error and Recovery (SPEAR). To complete the study a quantitative assessment using Fault Tree Analysis (FTA) and Human Error Assessment and Reduction Technique (HEART) was developed.

Category: Safety Culture and Leadership Session

2:15 PM to 2:45 PM

Improving Industry Process Safety Performance through Responsible Collaboration

Speakers: Ryan Wong and Shanahan Mondal

Since 2012, the American Fuel & Petrochemical Manufacturers (AFPM) and the American Petroleum Institute (API) have been working together on industry process safety programs under the umbrella of Advancing Process Safety (APS) programs. This presentation will provide learning from experience case studies from participants on how they are using the APS programs, what changes they have implemented at their sites, and what improvements have been achieved by those changes. Examples include practice sharing documents, training, industry safety bulletins, and Walk the Line. Ultimately, I will discuss a variety of practical takeaways that highlight the application and sharing of industry lessons learned.

2:45 PM to 3:15 PM

How Much Does Safety Culture Change Over Time?

Speaker: Stephanie C. Payne

Literature review of longitudinal studies of safety culture over time.

3:15 PM to 3:45 PM

Administering a Safety Climate Assessment in a Multicultural Organization: Challenges and Findings

Speaker: Atif Mohammed Ashraf

This presentation provides a brief description of the administration of a safety climate assessment across four different sites to a 1200 employee organization in the Middle East. The inter-disciplinary approach between psychology and engineering to formulate a science-based safety climate survey is highlighted. Challenges including but not limited to achieving maximum survey participation, overcoming language barriers and the involvement of contractors is discussed. Finally, a review of the strengths and areas in need of improvement concerning safety at the respective sites will be presented.

Category: Procedures Session

4:00 PM to 4:30 PM

A Comparison of Procedure Quality Perceptions, Procedure Utility, Compliance Attitudes, and Deviation Behavior for Digital and Paper Format Procedures

Speaker: Joseph W. Hendricks

There is a dearth of research on digital (hand-held, interactive; not .pdf) procedures in the process safety industries. This study surveyed chemical processing and logistics employees to determine if there are substantial differences in procedure quality perceptions, deviations and attitudes between digital and paper procedure formats. Results suggest that quality perceptions are better for digital procedures. Other results indicated differences in deviations, compliance and utility attitudes with non-trivial effect sizes. Future research and limitations will be discussed.

4:30 PM to 5:00 PM

Practical Writing Tips To Prevent Human Error When Following Procedures

Speaker: Monica Philippart

This presentation begins explaining how people process information and how written information can lead to human error, to help the audience recognize the importance of adhering to the writing tips are subsequently presented. There are many writing standards and guidelines. The Pareto principle was applied to collect a practical selection of tips that procedure writers in high-risk industries can easily implement to reduce the likelihood of error when users follow their procedures. The goal of this presentation is to help improve the safety and effectiveness of operations by reducing human error when following written procedures.

5:00 PM to 5:30 PM

The Impact of Hazard Statement Design in Procedures on Compliance Rates: Some Contradictions to Best (or Common) Practices

Speaker: Joseph W. Hendricks

This study was designed to examine whether or not certain design elements of hazard statements (HS) actually impact compliance rates. We manipulated four HS elements (present vs. absent) – Icon, Number, Fill (Highlight), and Boxed in a virtual environment (2nd Life) – leading to a 16 condition within-subjects design. We observed a range of approximately 20 percentage points in compliance rates across the various conditions. The two most robust findings were that the presence of a warning icon surprisingly hurt performance and numbering was consistently helpful. Banner blindness is considered one possible explanation for the effects and future research will be discussed.

Track III - Managing Operations and Maintenance
Category: Modeling & Asset Integrity Session

8:45 AM to 9:15 AM

RBI Study using Advanced Consequence Assessment for Topside Equipment on Offshore Platforms

Speaker: Chetan Birajdar

An important driver that influences the production profile of an offshore facility is the reliability of production critical equipment. The improvement in production profile can be achieved by minimizing the equipment downtime using reliable components, inclusion of redundant units and effective-efficient inspection maintenance service.

9:15 AM to 9:45 AM

Indicators of an Immature Mechanical Integrity Program

Speaker: Derek Yelinek

Oil, gas, and chemical facilities face many challenges to ensure overall safety, risk management, operational efficiency, and improved reliability. One area of focus will be on the maturity of your Mechanical Integrity program. Organizations frequently benchmark their programs' effectiveness against best-in-class mechanical integrity programs. In this presentation, three common indicators that your Mechanical Integrity program might be limiting your strategic capabilities will be discussed as well as ways to mature your program toward a best-in-class state.

9:45 AM to 10:15 AM

Remember the à la Mode: Lessons Learned from Ammonia Release at Frozen Foods Warehouse

Speaker: Matthew S. Walters

Intentionally opening a line that carries a hazardous substance — a procedure known as a line break — is often necessary for performing maintenance activities on pipes, valves, pumps, compressors, and other process equipment. However, inadequate or improper line break practices may increase the risk for a loss of containment event, complicate troubleshooting efforts if a loss of containment occurs, or inadvertently expose workers to hazardous materials. A case study that examines an incident related to a line break in a frozen foods warehouse will be presented. The loss of containment event described here provides valuable lessons that can aid in developing an effective procedure for safe process operation following a line break, and the impact that improper line break procedures can have on leak identification and system troubleshooting.

Category: Recalling and Learning from Incidents Session

10:30 AM to 11:00 AM

Process Related Incidents with Fatality- Trends and Patterns

Speaker: Syeda Z. Halim

A database of the Occupational Safety and Health Administration (OSHA) captures incident data from investigations for fatal incidents and hospitalizations since 1984. OSHA Region 6 includes 5 states including Texas and Louisiana, where much of the US chemical manufacturing and petroleum refining industry is located. An analysis of process related investigations by OSHA in Region 6 shows that large-scale multi-fatality incidents have been significantly decreased since the implementation of Process Safety Management (PSM) program in 1995. It is noticeable that currently majority of the fatalities occurs in single fatality incidents. Our preliminary analysis suggests that these individual process related fatalities are a result of operating and maintenance activities that are not well addressed by current process safety practices or by personal safety measures. An analysis of such incidents and their circumstances will be conducted proving recommendations for improved performance to reduce the incidents with single fatality.

11:00 AM to 11:30 AM

Application of Mind Mapping to Organize and Recall Potential Hazards

Speaker: T Michael O'Connor

Failure to learn lessons from previous incidents has been a recognized problem for decades. We attempt to improve this situation for certain high hazard tasks. Mind maps are used to categorize hazards from past incidents to aid in their recall and avoidance.

11:30 AM to 12:00 PM

Would a HAZOP, LOPA or STPA have Prevented Bhopal?

Speaker: Howard Duhon

This presentation attempts to answer three questions: 1) Would a HAZOP on the Bhopal MIC design, conducted in the 1960's, have prevented the tragedy? Would a LOPA have prevented it? 3) would and STPA have prevented it?

Category: Improving Process Safety with Technological Advances Session

2:15 PM to 2:45 PM

Predictive Process Safety Analytics and IIoT - PSM Plus: The AI+PSM Analytic Framework

Speaker: Michael Marshall

With an IIoT predictive application environment as the backdrop and an asset integrity and process safety analytic framework as the primary enabler, the paper and presentation discuss methods, metrics, performance analyses, and KPI benchmarking techniques for driving Operational Excellence as it relates to the ultimate concern of any PSM program, i.e., the loss of primary containment (LOPC) and associated impacts to production, profitability and process safety.

2:45 PM to 3:15 PM

Guidance to Improve the Effectiveness of Process Safety Management Systems in Operating Facilities

Speaker: Syeda Zohra Halim

In this presentation we analyze the recent trend in process safety incidents and identify issues behind current incidents. Based on the identified issues we recommend methods to improve the effectiveness of process safety systems.

3:15 PM to 3:45 PM

Unified Wall Panel System (UWPS) - A Value Engineering Solution for Protective Construction in the Petroleum Industry

Speaker: Scott Hardesty

An overview of the novel protection technologies being developed for use in the UWPS, including high- cementitious structural paneling, non-aramid advanced mineral fiber reinforcement and metallic foam energy absorption.

Category: Exploring NaTech Events and Domino Impacts Session

4:00 PM to 4:30 PM

Protect Process Plants From Climate Change

Speaker: Victor H. Edwards

Outlined here is how to conduct a climate risk vulnerability assessment for a process plant.

4:30 PM to 5:00 PM

Process Safety Implications in a Changing Environment

Speaker: Trish Kerin

While much research has been undertaken in natural hazards triggering technological disasters (Natech) it still remains a challenging area. It can be difficult to move past the psychological bias to focus on the possible incident outcome without discounting a seeming incredible cause.

5:00 PM to 5:30 PM

A Critical Evaluation of Industrial Accidents Involving Domino Effect

Speaker: Ravi Kumar Sharma

In this study, an incident analysis of 326 accidents since 1961 involving the domino effect in process, storage plants and the transportation of hazardous materials were analysed. Coding of incidents were done based on data obtained from different sources. The domino incident database analysis includes several categories such as fatalities over time, incidents over time, and incidents with respect to location, materials involved, causes and consequences.

BACK TO THE PROGRAM

DAY 2: Wednesday Oct 22 Summaries

Track I - Risk/Consequence & Design Aspects
Category: Risk Assessment III Session

8:30 AM to 9:00 AM

Applying PHA Methodologies such as HAZOP and Bowtie to Assessing Industrial Cybersecurity Risk

Speakers: John Cusimano, Jacob Morella, Tim Gale

Process hazard assessments (PHA) are a well-established practice in process safety management. These assessments focus on failures (aka deviations) that are typically caused by equipment failures or human error. By design, PHAs do not consider cyber threats to industrial control systems (ICS). However, cyber threats represent additional failure modes that may lead to the same health, safety and environmental consequences identified in the PHA. Functional safety (i.e. ISA 84 / IEC 61511) and industrial cybersecurity standards (i.e. ISA/IEC 62443) recognize this issue and provide guidance on how to integrate these two disciplines to ensure that cyber incidents cannot impact process safety.

9:00 AM to 9:30 AM

Large Hydrocarbon Tank Fires: Modelling of the Geometric and Radiative Characteristics

Speaker: Ravi Kumar Sharma

A proven methodology, called Cyber PHA, based on ISA/IEC 62443-3-2 has been developed and applied to conduct ICS cyber risk assessments throughout the process industries. This presentation will describe the methodology with examples of actual applications to identify, rank and mitigate cyber risk in ICS systems. Furthermore, we will demonstrate how Bowtie Analysis can be used to visualize the results and apply degradation factors and controls related to cyber barrier assurance.

9:30 AM to 10:00 AM

Risk assessment of a large chemical complex during the construction phase using Intuitionistic Fuzzy Analytic hierarchy process

Speaker: Suresh G.

This presentation will discuss the risks that possess in critical operations carried out during construction phase of project were considered. Which are categorized based on a novel method of Interval Valued Intuitionistic Fuzzy Analytical hierarchy process. The different categories are Catastrophic, Critical, Serious, Minor. The Analytical hierarchy process (AHP) possess the unique advantage of comparing parameters that have no units or scale of measurements.

Category: Risk Mitigation Session

10:15 AM to 10:45 AM

Development of Resilient LNG Facilities

Speaker: Onder Akinci

Resilient Design of LNG Facilities is discussed in this study. Problem statement, design fundamentals, case studies and best practice recommendations are presented.

10:45 AM to 11:15 AM

Development of Risk Mitigation Programs using a Quantitative-Risk-Based Approach

Speaker: Rafael Callejas-Tovar

This presentation will utilize example case studies to demonstrate how a quantitative-risk-based approach can be leveraged in a risk mitigation program to optimize risk mitigation solutions such as building reinforcement, building replacement, and/or scenario mitigation. Also, the presentation will present examples of facility siting issues that the processing industries struggle with, such as focusing on implementing solutions to mitigating explosion hazards while neglecting other equal or high risk hazards, or implementing solutions company-wide that might be only effective for some assets, which results in unnecessary costs that do not mitigate the risk effectively.

11:15 AM to 11:45 AM

Incorporating Mitigative Safeguards with LOPA

Speaker: Edward Marszal

Layer of Protection Analysis (LOPA) is a ubiquitous tool for assessing risk in more detail than a purely quantitative HAZOP but not so much as a QRA. It is efficient due to its conservative simplifications. Unfortunately, the simplifications prevent assessing the benefit of mitigative safeguards, such as fire detection systems that trigger water deluges. This presentation presents a methodology for extending to LOPA process to allow assessment of mitigative safeguards. The extension requires the evaluation of the situation where the safeguard fails to operate and also the residual risk that remains even if the mitigative safeguard effectively functions. For preventive safeguards the residual risk is zero, but for mitigative safeguards it cannot be ignored.

Category: Consequence Analysis: Gas Releases Session

12:45 PM to 1:15 PM
Hole Size Matters

Speaker: Jeffrey D. Marx

Consequence analysis is an integral tool for many process safety studies.

1:15 PM to 1:45 PM
How Can I Effectively Place My Gas Detectors

Speaker: Jesse Brumbaugh

The approach taken for selection and placement of gas detectors is found to vary widely between different companies. There is a growing interest in not only the confidence but also the effectiveness of these gas detection systems as a key mitigation barrier. The intention of this presentation is to provide a methodology that is both effective and cost efficient while also presenting the main considerations that design engineers and process safety professionals should address for the gas detection system elements of (1) a comprehensive gas detection philosophy, (2) appropriate detector technology selection, and (3) correct detector placement.

1:45 PM to 2:15 PM
Consequence Assessment Considerations for Toxic Natural Gas Dispersion Modeling

Speakers: SreeRaj Nair and Noma Ogbeifun

Consequence modeling provides information on the potential impact zone and is key for process risk management.

Category: Reactive Chemicals Session

4:00 PM to 4:30 PM
Modelling and Simulation to Predict Energetic Material Properties

Speaker: Kok Hwa Lim

With the rapid development and advancement in computing power, modelling and simulation (M&S) has demonstrated its vast potential in predicting the properties of energetic material and helping to design energetic material. One such application is predicting crystal packing and crystalline structure from first-principle simulation. Such technique has demonstrated the ability to distinguish different polymorphs of the same energetic molecules and accurately predict the crystal structure and density. In addition to the ability to predict detonation pressures and velocities of more established classes of energetic materials based on their thermochemical code or empirical equations, M&S has also demonstrated its ability to screen designed energetic materials for potential application. The application of M&S vastly improves the safety of developing potential energetic materials - the ability to screen potential energetic materials based on M&S-predicted heats of formation and detonation properties means that less hazardous experiments are required to be conducted as well as reducing developmental cost.

4:30 PM to 5:00 PM
Safety Assessment of Low Temperature Radical Initiators for Proper Storage and Safe Handling Conditions

Speaker: Cuixian (Trisha) Yang

Commercially available azo-type low temperature radical initiators provide efficient initiation of many chemical reactions. However, the azo group initiators are energetic compounds that also have thermal stability issues at ambient or even sub-ambient temperatures. These initiators can also generate nitrogen gas during slow decomposition under heat and/or light, which could present a safety challenge for shipping, storage and usage. In order to define safe storage and handling conditions, a variety of calorimetry studies were carried out. Exotherm and pressure data were collected from these studies in an effort to gain a better understanding of the decomposition kinetics. Thermal-kinetics and thermal safety model simulations were then used to obtain the self-accelerating decomposition temperature (SADT) and decomposition activation energy for the azo-type initiator. This methodology for thermal decomposition kinetics data and parameter determination, acquired with 5mg to 1g scale samples, enables safe storage, handling, and scale-up process preparation.

5:00 PM to 5:30 PM
Analysis of Pressure Behavior during Reaction Runaway and Estimation of Available Depressurization Design

Speaker: Yuto Mizuta

Analysis of pressure behavior during reaction runaway and estimation of available depressurization design, dynamic simulation by Aspen, runaway experiment by ARSST, two-phase flow model of ISO.

Track II - Human Factors - People in Action **Category**: Human Performance/Decision Making II Session

8:30 AM to 9:00 AM

Preventing Cognitive-Attributed Errors in Safety Critical Systems: A Path Forward

Speaker: Tom Shephard

The presentation provides background and example models, methods and tools for assessing and eliminating cognitive attributed errors in active human barriers.

9:00 AM to 9:30 AM

Two Views of Evaluating Procedural Task Performance: A Transition from Safety-I to Safety-II Approach

Speaker: Changwon Son

This presentation provides the development of new procedural task performance measures based on Safety-II perspective, an emerging safety paradigm.

9:30 AM to 10:00 AM

Beyond Human Error: Integration of the Interactive Behavior Triad and Toward a Systems Model

Speaker: Joseph W. Hendricks

In an effort to move beyond the "human error" explanation for safety incidents, we surveyed individuals employed in the process safety industry and were primarily from the Oil & Gas and Chemical industries. Results indicated that perceptions of procedure quality was the focal variable in all of the results, including positive relationships with attitudes toward the procedure change process and negative relationships with procedure deviations, and both safety incidents and near-misses. Additionally, we integrated the three elements of the Interactive Behavior Triad—person, task, and context—into Dekker's Model 2 of safety. We found support for two-way interactions using moderator regression analyses. We conclude that these elements are important factors to consider when evaluating and developing procedure systems.

Category: Fatigue & Stress Session

10:15 AM to 10:45 AM

Operator Performance Under Stress: A Neurocentric Virtual Reality Training Approach

Speaker: Ranjana Mehta

Operators in the process industries work under extreme pressures in complex hazardous environments that are associated with critical consequences at the cost of lives. Thus, ensuring operator safety is of utmost importance in this domain, and in particular in stressed contexts. Advances in Virtual Reality (VR) have enabled cost-effective, relatable, and remote trainings that can potentially transform the future of operator training in complex environments.

10:45 AM to 11:15 AM

Towards a Predictive Fatigue Technology for Oil and Gas Drivers

Speaker: John Kang

Towards a Predictive Fatigue Technology for Oil and Gas Drivers.

11:15 AM to 11:45 AM

Validation of the Fatigue Risk Assessment and Management in High-Risk Environments (FRAME) Survey

Speaker: Stefan V. Dumlao

The oil and gas extraction (OGE) industry continues to experience a fatality rate nearly seven times higher than that for all U.S. workers. OGE workers are exposed to intensive shift patterns and long work durations inherent in this environment. This leads to fatigue, thereby increasing risks of accidents and injuries. In the absence of any regulatory guidelines, there is a critical need for the development of comprehensive fatigue assessment practices specific to OGE operations that take into consideration not only the various OGE-specific sources of fatigue, but also the barriers associated with effective and feasible fatigue assessments in OGE work. In response to this need, Shortz, Mehta, Peres, Benden, and Zheng (2019) developed the Fatigue Risk Assessment & Management in high-risk Environments (FRAME) survey. Further, they provided evidence that the FRAME survey content captures fatigue-related information specific to the OGE industry not found in any one other measure of fatigue. The present study expands on these efforts by examining the psychometric properties (i.e., reliability and validity) of the FRAME survey—a critical step before the survey can be recommended for use in practice. A sample of 210 OGE and petrochemical refinery workers were sought to participate in this study. Linkages between the FRAME survey and a number of fatigue-related measures validated for use outside of the OGE industry will be examined. Once data analysis is complete, the FRAME survey will be refined for implementation, and recommendations for implementation will be provided.

Track IV - Research and Next Generation **Category**: Next Generation Process Safety I Session

12:45 PM to 1:15 PM

Identifying contributing factors of pipeline incident from PHMSA database based on NLP and text mining techniques

Speaker: Guanyang Liu

Recently there are a few attempts that develop methods to enable automated content analysis of incident reports by natural language processing (NLP) techniques, but with a manual list of key words still needed, the methods are not intelligent or automated enough to extract information that is outside the pre-defined vocabulary. In this work, advanced NLP techniques for text mining, are employed to identify causal relations from incident reports based on unsupervised learning and co-occurrence network algorithms. The proposed method is capable of extracting latent causal factors of the incident causes described in the reports and indicating the potential of identifying root causes with more comprehensive training text data applied in the future work.

1:15 PM to 1:45 PM

Causation analysis of pipeline incidents using artificial neural network (ANN)

Speaker: Pallavi Kumari

Failure of hazardous liquid (HL) pipelines is a potentially significant hazard to people, property and the environment. One of the main causes of HL pipeline failures is corrosion. To predict cause and consequences of corrosion in HL pipelines, this article presents an artificial neural network (ANN) using incidents data collected by the Pipeline Hazardous Material Safety Administration (PHMSA) of the US Department of Transportation corresponding to the onshore HL transmission pipelines in the US between 2010 and 2019. From this incident database, 70 attributes has been selected for their ability to predict corrosion. Using selected attributes as input to the ANN model, the model is constructed and optimized for its hyper parameters; and it predicts the type of corrosion, total cost of property damage, net material loss and type of incident (rupture/release) with 60-90% accuracy. In order to establish credibility of developed ANN model, the model accuracy obtained using ANN model is compared against another machine learning model.

1:45 PM to 2:15 PM

Development of Hazard Factor for Engineered Particles

Speaker: Nabila Nazneen

Particles from nano to micro ranges, being a comparatively new discovery in the field of science and engineering, poses many risks to the industry. Apart from their health effect, scientists and engineers are concerned about their explosion possibilities. A hazard factor would be able to identify the hazard level of nano to micron range particles and help take proper controls of the risk associated with them. This study creates a database of the different properties of various particles and fashions a hazard factor to formulate a ranking system. The hazard factor is based on properties like particle size, maximum overpressure, maximum rate of pressure rise, minimum ignition energy, minimum explosible concentration, minimum ignition temperature, etc. Based on the data collected, the research uses statistical analysis to check the behavior of the properties and modify the NFPA ranking to formulate the hazard factor.

Finally, the factors will be ranked to precisely identify the hazard level against their respective properties. This hazard factor will be an effective indicator of a potential hazard that the engineered particles may hold and alert the users to take preventive action to moderate the risk of the hazard.

Category: Next Generation Process Safety II Session

4:00 PM to 4:30 PM

Can a Virtual Reality Application Better Prepare Millennials and the Z-Generation for Working with Systems in the Process Industry?

Speaker: Nir Keren

Use of Virtual Reality to Enhance Students' systems mental model.

4:30 PM to 5:00 PM

Process Safety Risk Index Calculation Based on Historian Data

Speaker: Prasad Goteti

This presentation details the Process Safety Risk Index calculation based on Historian data using a real-life example from the process industry.

5:00 PM to 5:30 PM

A Brief review of Intrusion Detection System in Process plants and advancement of Machine Learning in Process Security

Speaker: Sinijoy P J

Rapid technology growth has given rise to new vulnerabilities and threats to the Computer Oriented Process Plants. Production Plant's dependency on Computers, Sensors, IIoT and networks wide opened Intrusion based threats and attacks. Intrusion Detection System [IDS] has become very popular due to its demand in Industry as the protection of information/processes controls are to be secured from the reach of unauthorized personals.

Track V - Explosions and Flammability
Category: Explosion Modelling Session

8:30 AM to 9:00 AM

The Influence of the Velocity Field on the Stretch Factor and on the Characteristic Length of Wrinkling of Turbulent Premixed Flames

Speaker: Tássia L. S. Quaresma

The stretch factor accounts for the effects of strain, due to fluid dynamic effects, and curvature, due to propagation. However, within the BML approach, flame stretching is commonly neglected due to difficulties in modelling the stretch factor, which is assumed as a constant parameter and equals to unity. Recent works based on more sophisticated analyses such as LES and DNS have suggested that this consideration may lead to an inaccurate representation to flames with a non-zero mean curvature. Therefore, we propose a dynamic expression to the stretch factor within the BML approach that is based on the physical understanding of the phenomenon. The study is carried out within an in-house developed RANS code for simulation of turbulent reacting flows in complex geometries. The approach explores the influence of the velocity field on the flame surface and its contribution to flame stretching, which it is not accounted in typical flamelet approaches. We follow the reasoning line that the velocity divergent influences the fluid hydrodynamics when reaching obstacles ahead of the flame front and the stretch factor as well as the characteristic wrinkling length of the flame. The hypothesis is that the velocity divergent contributes to the flame stretching the same way it contributes to the fluid motion. As a consequence, the stretch factor is modelled as a function of the velocity, changing both in time and space.

9:00 AM to 9:30 AM

Towards a Comprehensive Model Evaluation Protocol for LNG Hazard Analyses

Speaker: Filipo Gavellia

Blue Engineering and Consulting and the Gas Technology Institute are collaborating on a DOT-PHMSA sponsored project to develop a new set of Model Evaluation Protocols, that will allow the review of modeling tools for each of the main hazards associated with LNG facilities. The new MEP will include methodologies for evaluating the modeling of hazard phenomena such as release, dispersion, vapor cloud explosion, and pressure vessel failure events. This presentation will describe the framework of the new MEPs and provide an update on the status of the work.

9:30 AM to 10:00 AM

Beirut: How behaves Ammonium Nitrate Exposed to Fire and How Strong and Damaging is its Explosion?

Speaker: Charline Fouchier

The Beirut explosion on the 4th of August is one more accident to be added to the long list of tragedies caused by Ammonium Nitrate. While many investigations have been conducted to understand better the behaviors of the molecule, it is still unclear how the Ammonium Nitrate can detonate in an unconfined environment while heated by fire.

A rapid summary of the state of art on Nitrate Ammonium is given, followed by an analysis of the Beirut accident, with a proposed scenario that could have led to the explosion. Finally, methods to estimate the explosion energy, based on the blast arrival time, the damages on buildings and the crater dimensions are applied on the Beirut accident and compared.

Category: Explosion Phenomena I Session

10:15 AM to 10:45 AM

Flammable Mist Hazards Involving High-Flashpoint Fluids

Speaker: Simon Gant

This presentation on "flammable mist hazards involving high-flashpoint fluids", given by Simon Gant from the UK Health and Safety Executive (HSE), provides an overview of the work led by HSE on flammable mists over the last decade, including preliminary results from an ongoing joint industry project. The work has involved a collaboration between HSE and the Gas Turbine Research Centre (GTRC) at Cardiff University.

10:45 AM to 11:15 AM

Measuring Suspended Explosive Dust Concentration from Images

Speaker: Yumeng Zhao

Dust dispersion during powder handling and processing is of great concern for both workers' health and explosion risk. Dust emission locations in industries can vary during handling and processing, while dust concentration sensing would require the installation of an additional equipment in every location prone to dust generation. A method of using a digital camera or photograph to measure the dust concentration based on two target intensity value has been developed at Purdue University. The method was developed based on the relationship between the suspended dust concentration and extinction coefficient. Calibrated equations have been developed for cornstarch, grain dust, and sawdust. This method does not require any training and can be integrated with security system cameras and/or other independent imaging source.

11:15 AM to 11:45 AM

The HBT-A Large-Scale Facility for Study of Detonations and Explosions

Speaker: Elaine S. Oran

This presentation describes the design and development of the new HBT, the large-scale shock and detonation tube facility for the study of deflagrations, detonations, and the transition processes.

Category: Explosion Phenomena II Session

12:45 PM to 1:15 PM

Development of Flammable Dispersion Quantitative Property-Consequence Relationship Models Using Machine Learning

Speaker: Zeren Jiao

Incidental release of flammable gases and liquids can lead to the formation of flammable vapor clouds. When their concentrations are above the lower flammable limit (LFL), or ½ LFL for conservative evaluation, fires and explosions can result with the presence of an ignition source. The objective of this work is to develop highly efficient consequence models to accurately predict the downwind maximum distance, minimum distance, and maximum vapor cloud width within the flammable limit. In this study, a novel quantitative property-consequence relationship (QPCR) model is proposed and constructed for the first time to accurately predict flammable dispersion consequences in a machine learning and data-driven manner. Flammable dispersion database consists of 450 leak scenarios of 41 flammable chemicals were constructed using PHAST simulations. A state-of-art machine learning regression method, extreme gradient boosting algorithm, was implemented to develop models. The coefficient of determination (R2) and root-mean-square error (RMSE) were calculated for statistical assessment and the developed QPCR models achieved satisfactory predictive capabilities. All the developed models have high accuracy, with the overall RMSE of three models being 0.0811, 0.0741, and 0.0964, respectively. The developed QPCR models can be used to obtain instant flammable dispersion estimations for novel flammable chemicals and mixtures at much lower computational costs.

1:15 PM to 1:45 PM

An Unsupervised Model to Predict the Liquid In-cylinder Combustion Risk Ratings of Marine Fuels

Speaker: Chenxi Ji

A novelty unsupervised machine learning approach to establish an in-cylinder combustion risk criterion for marine fuels.

1:45 PM to 2:15 PM

Fireball and Flame Venting Comparisons

Speaker: Peter A. Diakow

This presentation compares flame jetting distances from vented explosion tests to predictions made using NFPA 68, EN 14994 and the FLACS CFD code.

Category: Consequence Analysis: Flammability Session

4:00 PM to 4:30 PM

Numerical Simulation of Methane-Air DDT in Channels containing Trace Amounts of Impurities

Speaker: Logan N. Kunka

Accidental explosions are not only deadly, but often disastrous to the industry. Understanding these explosions are key to protecting against them. This session explores explosions and deflagration-to-detonation transition (DDT) in natural gas filled channels typically found in sealed sections of coal mines. A detailed investigation is presented on how heavy hydrocarbon impurities, present in realistic natural gas compositions, effect the run-up distance to DDT.

4:30 PM to 5:00 PM

The Use of Bent Poles as a Detonation Indicator

Speaker: J. Kelly Thomas

This presentation covers an analysis of the response of poles to both deflagration and detonation loading with disc-shaped clouds to evaluate the use of bent poles as a detonation indicator.

5:00 PM to 5:30 PM

Machine Learning Based Quantitative Prediction Model for Chemical Mixture Flammability Limits

Speaker: Zeren Jiao

Flammability limits (FL), including lower flammable limit (LFL) and upper flammable limit (UFL), are crucial for fire and explosion hazards assessment and consequence analysis. In this study, by using an extended FL database of chemical mixture, quantitative structure-property relationship (QSPR) models have been established using gradient boosting (GB) machine learning algorithm. Feature importance based descriptor screening method is also implemented for the first time to determine the optimal set of descriptors for model development. The result shows that all developed models have significantly higher accuracy than other published models, with the test set RMSE of LFL and UFL models being 0.058, 0.129, respectively. All the developed QSPR models can be used to obtain reliable chemical mixture FL estimation and provide useful guidance in fire and explosion hazard assessment and consequence analysis.

BACK TO THE PROGRAM

MARY KAY O'CONNOR PROCESS SAFETY CENTER
TEXAS A&M ENGINEERING EXPERIMENT STATION

23rd Annual Process Safety International Symposium
October 20-21, 2020 | College Station, Texas

Importance of Process Safety Time in Design

Shanmuga Prasad K*, Vijay S
TechnipFMC
Process Safety Department
Technip India Ltd, 19, Velachery Main road, Chennai, India- 600 032
* Presenter E-mail: shanmugaprasad.k@technipfmc.com,

Abstract

The importance of time can be well conveyed in the proverb "Time and tide waits for none". The essential actions not taken in time could propagate to Hazardous events especially in the Process industry. Hence these plants should consider this vital criterion called Process Safety Time as part of their design. Although Process Hazard techniques are applied to identify the hazards of the plants, the time factor and dynamical behaviours of the Process are often not considered in these assessments as they are complex. But in recent times dynamic models are available, although the application is limited to be integrated in the design especially the Functional safety i.e. Safety Instrumented System (SIS) and LOPA. So, what is this Process Safety Time? This paper demonstrates the importance of this Process Safety Time, how it can be assessed and its role in the design to minimize the risk of undesirable events leading to hazards.

Keywords: Functional Safety, SIS, Process Safety, Design concept

1. Introduction

The undesirable event in any industry are the accidents. Accidents cause serious impacts that can affect people, environment and asset with in facility or around the facility depending upon its magnitude. The industrial accidents occurs due to various reasons particularly in the oil and gas which handles various hazardous inventories. Although the companies take precautionary measures to avoid such situations. The very important task by any company is to minimize the accidents by identifying the potential hazards anticipated in their plants. Companies generally employs various Process Hazard Analysis techniques such as HAZID, HAZOP, LOPA, What-If Analysis etc. [7], to identify these hazards and to evaluate the various risks evolving from these hazards. Despite these various safety studies being carried out by the companies, accidents do happen due to various other unidentified/ hidden hazards. In particular the factor of time is an important parameter in any process. The essential actions not taken in time could propagate to Hazardous events.

Particularly the response time required to take corrective action is very critical. For Process industry, this response time required depends upon the dynamical behaviours of the Process which are mostly complex to predict as it depends upon various factors such as effective operation, heat and mass transfer, type of design, reaction kinetics etc. So, a good understanding of the Process is essential to determine the response time which will be based on this unique term called the 'Process Safety Time (PST)' dictating the precise response time to be considered. This PST can be conveyed as the time that is available to take action on the process to bring it back it to a safe condition once the process value gets deviated from the normal level.

The estimation of PST is system dependent and hence it relates to the dynamic behavior of the process, equipment design limits and process control system within the context of a unmitigated specific hazard scenario which could result from various initiating events. In the recent times the dynamic models are given good importance to determine how the Process acts upon time. Though the application of these models are very limited due to various reasons such as the cost, project schedule, difficulties in integrating in the design of the plants particularly in the SIS design etc.

This work demonstrates the importance of this Process Safety Time and its role in the design of the Plants, particularly the factor of SIF response time to achieve potential risk mitigation. In a nutshell, the SIF response time shall be lesser than that of the Process Safety Time to bring down the process to a safe state in order to eliminate the occurrence of hazardous event.

2. Basis of Safety Design

Safety design is the process of designing the plants to prevent potential accidents and it often comes with a multidisciplinary nature which means that a very broad array of professionals are actively involved in the aspect of Safety engineering.

Safety engineering also is the key component for eliminating hazards that would otherwise be controlled by either administrative controls such as the SIS, alarms etc. or use of personal protective equipment as a barrier between a hazard and a worker. These engineered safeguards include machine guards, selection of less hazardous equipment, development of maintenance schedules to ensure equipment safety, audit and inspection procedures, selection of safer tools, safety review of new equipment, employee maintenance training, safe design of the flow of

material and people through a facility and risk analysis for both possible man-made and natural incidents.

Layers of Protection Analysis (LOPA) is one of the risk analysis technique which has many applications and it can be applied in the safety design to assess/ design to ensure sufficient protections are available against the potential hazards. Using LOPA, the design the protective devices or safety barriers on the Process plants can be done which acts a 'Layers of Protection' in between the hazard and the receptors. They are designed to act as an armour to protect against the hazards which may harm people, environment and also the commercial interests of the company. These layers of Protections can be identified and assessed using the LOPA for effective control over process upsets. These layers starts from the prevention of hazard by Process control either by automatic or manual even by mitigation of the consequences including the emergency responses.

LOPA is a semi-quantitative analysis, intermediate between a qualitative Process Hazard analysis and the Quantitative Risk Analysis. "LOPA typically uses order of magnitude categories for initiating event frequency, consequence severity and the likelihood of failure of Independent Protection Layers (IPLs) to approximate the risk of scenario"[3].

The following figure portrays a typical Layers of Protection in the process industry.

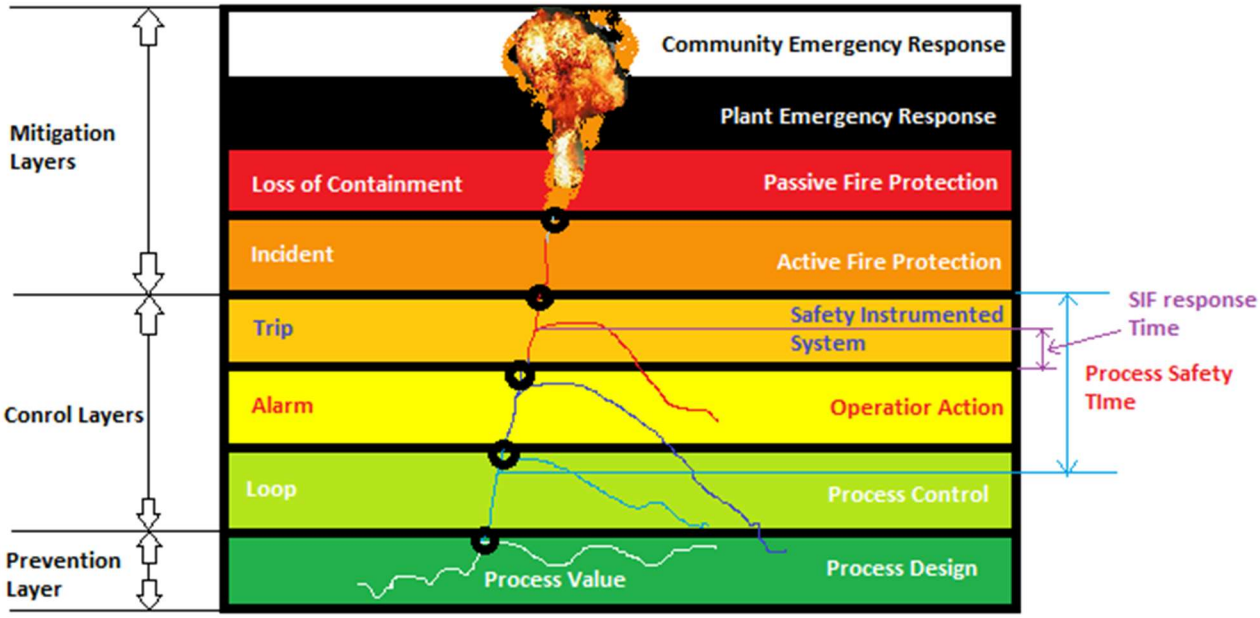

Figure 1: Typical Layers of Protection against a possible accident [4]

The first layer of protection is the Prevention layer in which the Process design itself maintains desired process value not being escalated. The next three layers are 'control layers' designed to prevent a safety related event. The first of the control layer (2nd layer) is the basic process control system which provides safety through proper design of control of the process. The second of the control layer (3rd layer) is the alarm system which provides the appropriate information to process operators, supporting them in the identification of the cause of the unsafe situation and allows them to take actions to restore the plant to normal operation. The third of the control layer (4th layer) is

the Safety Instrumented System (SIS). A typical SIS comprises of several Safety Instrumented Functions (SIFs) with sensors, valves and logic system to make appropriate decisions and take action on the process to bring it back to a safe state. The rest of the layers are termed as Mitigation layers (Active& Passive Fire Protections and the emergency response plans) as they act upon only if the incident occurs and are designed to mitigate the impact of the hazardous event.

Importance is placed on the Operator Action and SIS layers where the alarm and SIF are designed and installed for a specific function of the process, and the design information ensures that the device specified will meet those requirements.

The efficiency of the Operator action part of the control layer depends upon various factors such as proper training, capability, physical ability, response time etc. and it's very much arguable since it is a manual. For the next control layer of SIS comprising of SIF's, the response time is very critical, as it should act upon quickly to bring back the process to a safe state in a time much lesser than of the Process Safety Time. Hence the important task is to estimate this Process Safety Time (PST) although it is quite difficult to assess the exact condition at which the hazard scenario develops, and it is very contrary and hence it is better to evaluate the lower limit of the time at which the hazard might occur under worst case condition.

As said earlier the PST spans from the deviation of the process value until where the incident occurs. And particularly the response time of this layer shall be minimum and should fit well with the PST for a successful protection as if it crosses could result in an incident which is like action taken after the accident happened

3. Hazardous event Timeline

This section briefs about a typical event timeline for a hazardous process scenario as shown in figure-2. The event timeline explains how a process at a normal operating process value could propagate into an hazardous event and in particular how the Safety Instrumented System (SIS) protection layer could bring down the process value to normal operating level.

The normal operating range of a typical process value floats around the value as shown being control by the BPCS i.e. Basic Process Control system. Once the process value gets deviated from the normal operating range due to any reason and it is not being controlled further then it would reach the safe operating limit. Eventually the process value could exceed the design limit of the process/ equipment resulting in potential accident. And if suitable actions are taken then the process value could fall back to the normal operating range. In this case suitable action is considered to be taken by the SIS. The response time taken for this SIS is critical as delayed action would not control the process value exceeding the limits leading to potential hazardous event.

The following sections briefs the various factors in this event timeline including Process Safety Time, Process Lag time, trip delays, Time to Trip and Safety Margin. Further it defines the method for evaluation for better understanding of the response time required for efficient hazard management.

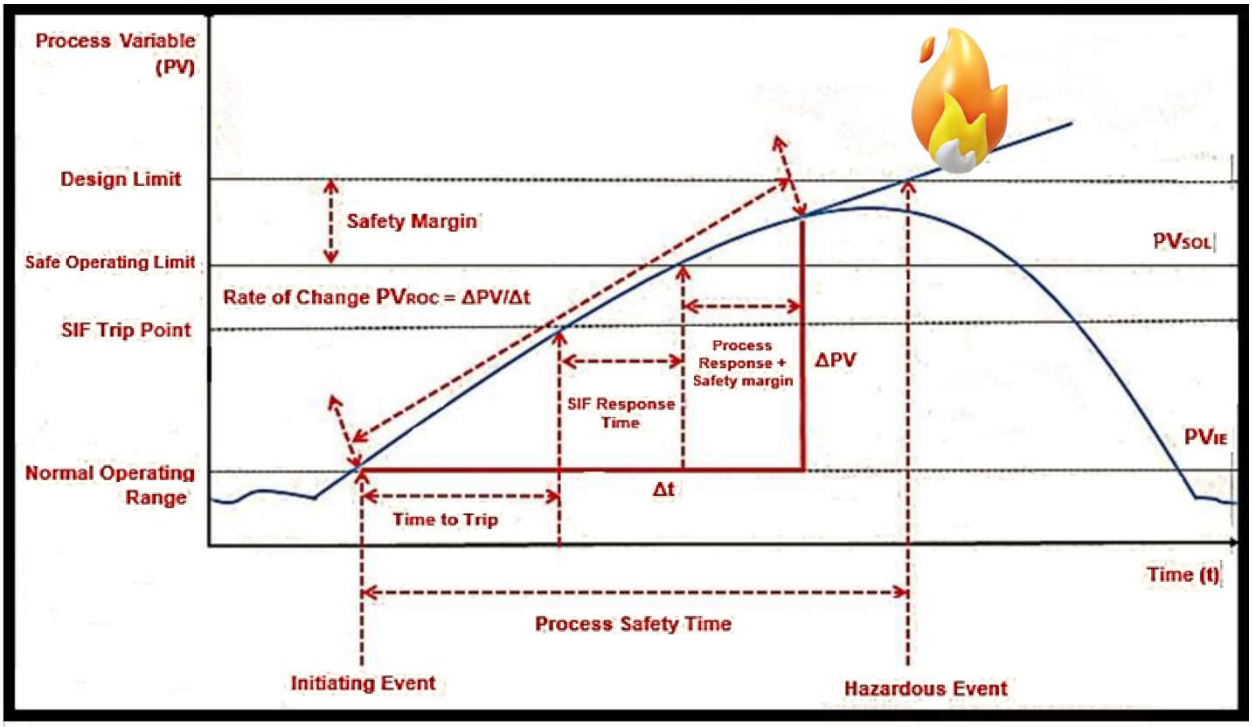

Figure 2: Typical Hazardous Event - Timeline

Process Safety Time (PST)

Process safety time is a function of the behavior of process and process equipment within the context of a specific unmitigated scenario. The term Process safety time (PST) is defined as per different codes and standards are as follows:

IEC 61511:2003 Part 2 defines PST as "the time between a failure occurring in the process or the basic process control system (with the potential to give rise to a hazardous event) and the occurrence of the hazardous event if the safety instrumented function is not performed".

IEC 61508:2010 Part 4 defines PST as "the period between a failure, that has the potential to give rise to a hazardous event, occurring in the EUC [equipment under control] or EUC control system and the time by which action has to be completed in the EUC to prevent the hazardous event occurring".

API 556 second edition, 2011 defines PST as "the interval between the initiating event leading to an unacceptable process deviation and the hazardous event"

The overall concept of the process safety time (PST) in simple words shall be the amount of time that is available to take action on the process to bring it back it to a safe condition after the initiation of hazardous event which may leads to out-of-control condition that can cause severe consequences.

PST is unique to each cause-consequence pair even when multiple initiating events may lead to the same consequences. The potential impact on process caused by the initiating event shall be in different ways due to the different process involved, different process equipment, different operating modes, design conditions and consequences affecting different risk receptors.PST is

quite difficult to measure because the exact conditions under which a hazard scenario may occur is unpredictable. PST should not be considered as a single specific value at which a hazardous event will immediately occur in all circumstances. Instead, PST can be estimated approximately by identifying the lower boundary line at which the hazardous event occurs so that protection layers available with sufficient response time for that credible scenario.

An analytical approach shall be taken into consideration for estimating the PST by considering the theories involved in process by knowing the normal operating and design limits of the equipment. The process variable associated with the occurrence of the hazardous event to be identified first. The point at which the deviations occurs in normal operating conditions shall give the approximated value of the initiating event to occur and the point at which the hazardous event occurs can be estimated approximately by the abnormal conditions of the process equipment i.e. equipment's exceeding its design limit which no longer can be prevented. The time taken between the initiating event and the hazardous event depends on the rate of change of process conditions from the initiating event to the hazardous event.

This can be represented by the equation as follows:

PV_{HE} - The value of the process value of interest at the time the Hazardous Event occurs or can no longer be prevented, may be assumed to be at the design limit of the equipment.

PV_{SOL} - The value of the process value of interest at the time assumed to be at the safe operating limit of the equipment.

PV_{IE} - The initial value of the process value of interest at the time of the Initiating Event, may be assumed to be at the extreme of the normal operating range nearest the hazard.

PV_{ROC} - The Estimated Rate of Change of the process value of interest under worst-case credible conditions in the context of the specific hazard scenario.

Time difference between normal operating range and design limit = PV_{HE}— PV_{IE}

Rate of change of PV of interest under worst case credible conditions PV_{ROC} = dPV/ dt

$$\textbf{Process Safety Time PST} = \frac{PV_{HE} - PV_{IE}}{PV_{ROC}}$$

From the graph it is noted that PST can also be evaluated by the summation of time to trip, SIF response time, process response and safety margin. It can be represented by the following equation

Process Safety Time= Time to Trip+ SIF Response Time + Process Response Time + Safety Margin Time

PST = TTT+ SRT+ PRT+SMT

SIF trip point

SIF trip point is the point at which the safety instrumented functions identifies and trips the system while the deviation occurs. It will bring back the process to normal operating condition to avoid the occurrence of undesirable scenario. The undesirable event may occur when the safety instrumented functions doesn't respond appropriately, and the operating parameters of the

equipment exceeds the design limit. This consequence may create injuries or even fatalities to the personnel's in the plant which leads to loss of reputation, assets and life.

SIF Response Time (SRT)

SIF response time is the length of the time from successful detection onset of an incident until the time at which the final control elements have acted and performed their function. The SIF response time includes the whole SIF loop components including initiator, Logic Solver and the Final element including the delays in the signal transmission. The SIF response time evaluation requires good process engineering knowledge and engineering judgement for its performance.

Time to Trip (TTT)

Time to trip is the length of the time from the initiation of the hazardous event until the time at which the SIF acts upon by successful detection.

Process Response Time (PRT)

Process Response Time is the length of the time once the SIF have completed its action until the time taken for the process value to come under the safe operating limit. This time varies from process to process. For example- for a storage tank scenario with level SIF, the Process Response time will be very shorter as the level stops rising once the SIF acts upon, however for the situations such as for Reactors the process reaction will take considerable amount of time to normalize.

Safety Margin Time (SMT)

Safety Margin Time is the length of time applied over the SIF response and Process Response Time in addition as a Safety Factor and this time factor usually depends upon the company's decision.

Operating Limits

Operating Limits are the values or ranges of values within which the process parameters normally should be maintained when operating. These values are usually associated with preserving product quality or operating the process efficiently. Whereas the Safe operating limits are established for critical process parameters, such as temperature, pressure, level, flow, or concentration, based on a combination of equipment design limits and the dynamics of the process. [6]

4. Example on Process Safety Time and Response Time calculation

This section discusses how the SIF response Time to be evaluated and the factors to be considered while performing the calculation with a simple example.

The following example illustrates a Surge Tank containing hexane liquid with an overflow line. The level control valve which on the line which feeds the tank, maintain the level by controlling the flow from upstream process into the tank, as the outlet flow is being drawn by a Pump to downstream consumers constantly.

Let us consider the maximum allowable level of the tank as 10 m and at 100% i.e. 10.0 m (PV_{HE}) the tank overflows which will be the potential hazard scenario that we focus upon in this illustration. Although there is a presence of dyke containment, it does not have any value in this

calculation as it is only a mitigative protective functional layer to prevent the escalation of the potential consequence of pool fire scenario. This is done by containing and reducing the diameter of the liquid pool to be formed due to the overflow of the tank.

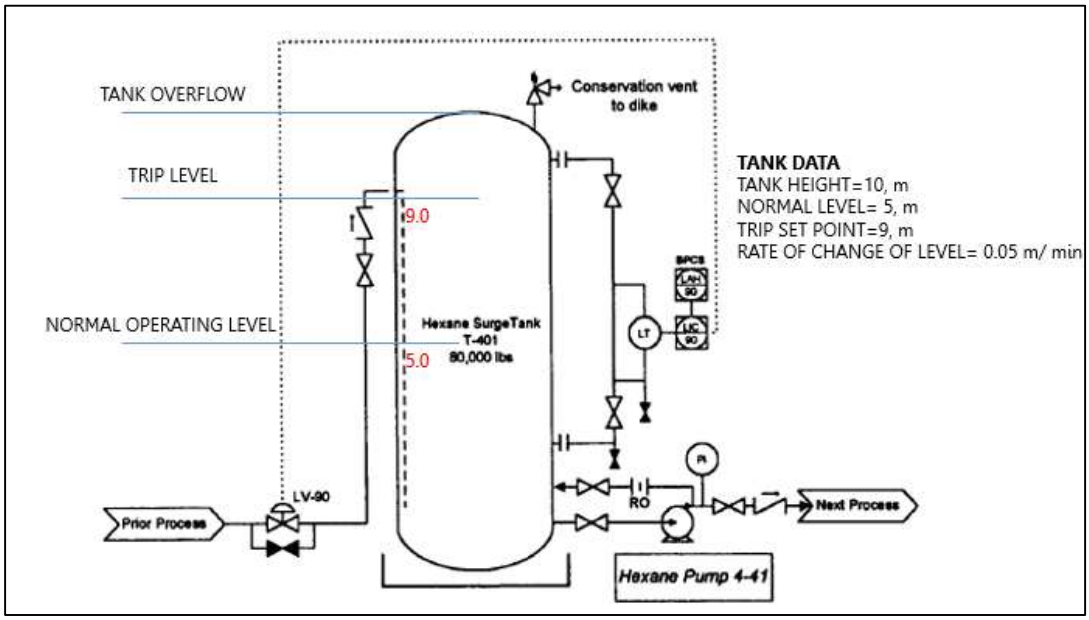

Figure 3: Illustration -Process Surge Tank-Safeguards available[4]

Let us consider there is a SIF in this tank for evaluation, consisting of Level transmitters as an Initiator, DCS as the Logic solver and the final control element being the shutoff valve present at the inlet of the tank next to the control valve as shown in the figure.

The initiating event in this example can be considered as the failure of the tank inlet control valve being fail open, feeding more to the tank which can eventually lead to hazardous scenario of overflow of the tank with potential consequences. The normal operating level in this tank can be considered as 50% i.e. 5.0 m (PV_{IE}), controlled by a Basic Process Control System (BPCS) at level to maintain the level. At the level 80% i.e. 8.0 m being considered High level, alerts the operator by an Alarm to act upon to take corrective action. At the level of 90% i.e. 9.0 m being the considered as the 'High High' level in which the SIF acts upon to safely shutdown the system by closing the inlet shut off valve at the inlet line of the tank. The maximum rate of change of level (PV_{ROC}) in the tank is considered as 0.05 m/ min at maximum incoming flowrate.

So, here the PST ranges from the deviation of normal level at 5m until the 10m at which the incident occurs i.e. the tank getting overflowing and the hexane getting into the dyke. This in turn can lead to hexane accumulation forming a pool inside the dike with potential risk of a pool fire scenario. As explained earlier the Process Safety Time in this example is calculated as below

$$\textbf{Process Safety Time (PST)} = \frac{PV_{HE} - PV_{IE}}{PV_{ROC}}$$

$$PST = \frac{(10.0 - 5.0)}{0.05}$$

Overall PST = 100, minutes

From this we know the overall Process Safety time in this case is 100 minutes. Then the time to trip, which is the time taken for the SIF to get activated on High High level can be calculated as

$$\textbf{Time to Trip (TTT)} = \frac{\textbf{SIF response Trip level (HHH)- } PV_{IE}}{PV_{ROC}}$$

$$TTT = \frac{(9.0-5.0)}{0.05}$$

$$TTT = 80, \text{ minutes}$$

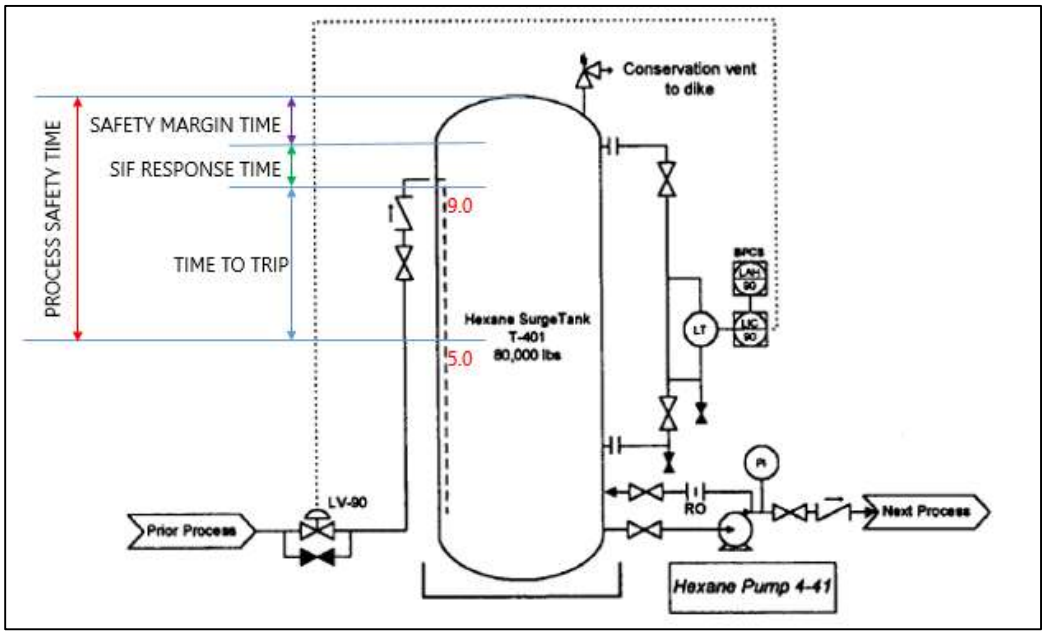

Figure 4: Surge Tank- Hazard Time line

There is no Process Response Time applicable in this example as it is only a tank. The level stops rising immediately once the inlet valve is closed, unlike in Reactors, columns etc. where the stabilization takes considerable time. But a Safety Margin Time (SMT) can be considered based upon the client requirement/ engineering judgement for safety purposes. In this case SMT is considered as 5 minutes.

So, with this information available, we can calculate the required SIF response Time (SRT) as below,

SIF Response Time (SRT) = Process Safety Time- Time to Trip – (Process Response Time+ Safety Margin Time)

$$SRT = PST - TTT - (PRT + SMT)$$

$$SRT = 100 - 80 - (0 + 5)$$

$$SRT = 15 \text{ minutes}$$

Hence from these calculations, it is found the SIF response time required for safeguarding against the overflow hazardous scenario is 15 minutes. So, the entire time taken for the entire SIF to respond starting from the level transmitter sensing successfully, the logic solver takes the decision

as per the interlock designed in time to close the valve including the time loosed in the signal transfers shall be within 15 minutes. This is the actually the maximum response time and SIF are expected to act within this for a successful hazardous scenario mitigation. The following figure represents the same calculation results in graphical form.

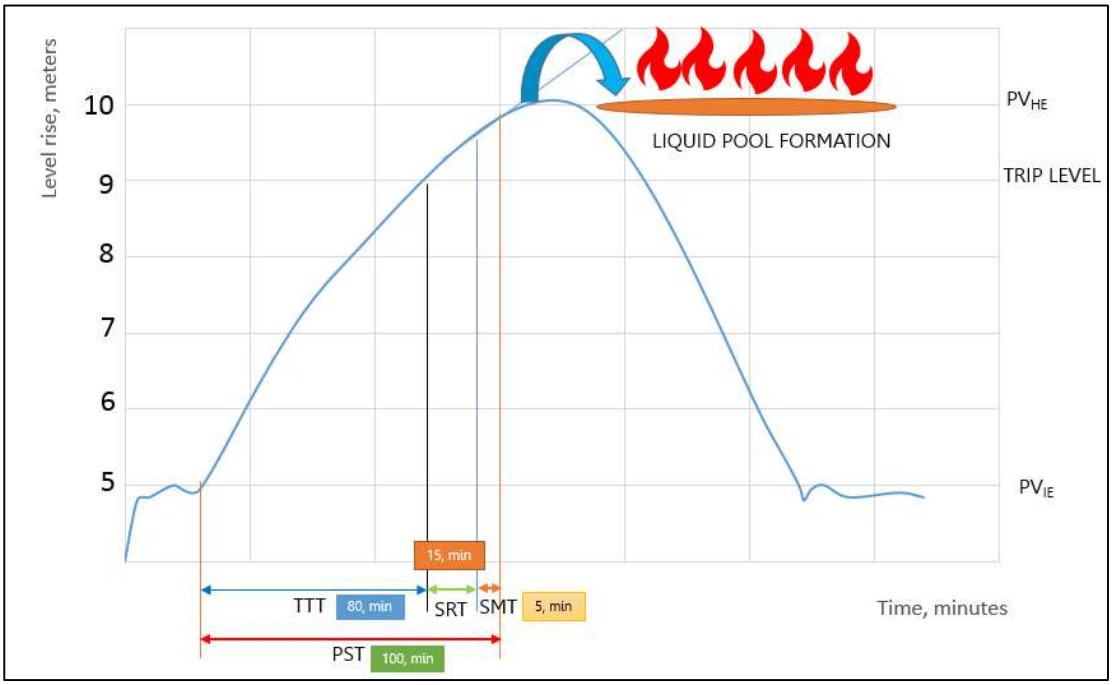

Figure 5: Example on PST-Graphical representation

5. Conclusion

So, this paper highlighted the importance of the Process Safety Time and its influence in the design for a successful safeguarding of the personnel, environment and property from the potential hazardous consequences. The estimation of PST is system dependent which means it varies from process to process. For evaluating a precise PST, it requires deep knowledge in Process and also good engineering judgement skills to preform design very close to the reality.

The Process Safety Time in the example can be easily estimated as the rate of change is linear in nature and the response time can be easily judged. In reality, many Processes have much shorter Process Safety Time for the shutdown functions especially for the SIF loops protecting the rapidly occurring Process. These process might occurs at exponential rates, such as for hazardous scenarios leading to potential Fire, explosions, toxic releases, runaway reactions and other serious consequences where the time taken for the incident to occur is really quick.

Also, if the shutdown protections systems to be designed, it shall consider factors of physical, chemical, kinetic and thermodynamic nature of the Process including the signal delay time lags. There will undoubtedly be uncertainty associated with the prediction of dynamic process where the conditions can be complex. So taking conservative assumptions and consistent safety margins throughout the evaluation is very important for a good protection against potential hazards. We

should also consider the factor of Process lag time on top of it as it again varies from Process to Process.

Typically, the SRT will be one half of the Process Safety Time which is not the case in rapid processes. It is very important that the whole response time of SIF loop should be lesser than the PST and this shall be done selecting appropriate instruments to be in the loop to serve the purpose. Hence care shall be taken for the SIS particularly in the SIL classification, Safety requirement Specification for instruments, SIL verification and validation throughout their lifecycle for effective protection. But to validate the complete response time would be quite difficult, so a good engineering judgement is essential and can be warranted by focusing more on the core attributes of the equipment to be selected and some more detailed analysis.

On conclusion, the importance for the PST shall be considered as one of the critical information in the design of safety systems of the plant. It requires deep knowledge about the process and shall be addressed through a consistent and timely approach close to reality for an effective protection of people, environment and assets against the potential hazards.

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23rd Annual Process Safety International Symposium
October 20-21, 2020 | College Station, Texas

Limitations of Layers of Protection Analysis (LOPA) in Complicated Process Systems

Arafat Aloqaily, Abdulaziz Alajlan
Loss Prevention Department, Saudi Aramco
ARAFAT.ALOQAILY@ARAMCO.COM; AMOQAILY@YAHOO.COM,
ABDULAZIZ.ALAJLAN@ARAMCO.COM

Layer of Protection Analysis (LOPA) has been used as a tool to conduct risk assessments for determining the required level of protection in the oil and gas processes for long time. It is easy to use and can provide a reasonable quantifying approach to determine the required number of independent protection layers in a given process, based on the perceived risk levels and acceptable risk. Since its development, LOPA has gained more popularity over other techniques such as the risk graph or table, mainly because it is more quantitative than the other approaches. However, some literature suggests that LOPA has its limitations including subjectivity as it depends on the judgement of the team conducting the study to determine the potential risk and hazard to protect against. In addition, LOPA is not suitable for complicated multi-causes scenarios.

This paper presents a case study showing the limitations of applying LOPA in upstream scenarios to develop protection layers requirements for a complicated network of pipelines and processing units with unlimited number of causes contributing to the risk. It compares LOPA with the more sophisticated more quantified other techniques such as Fault Tree Analysis (FTA). Based on the cases analysis, it is recommended that LOPA can be used to assess simple scenarios with limited number of causes, while more complicated cases are better assessed using FTA. Detailed analysis is presented in the paper to support such recommendation.

Keywords: LOPA, Fault Tree Analysis, FTA, LOPA Limitations, Complex Process Systems, SIL Assessment, SIL Verification, SIL Assignment

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On the usage of ontologies for the automation of HAZOP studies

Johannes. I. Single, Jürgen Schmidt, Jens Denecke
CSE Center of Safety Excellence (CSE Institut)
Joseph-von-Fraunhofer-Str. 9, 76327 Pfinztal, Germany

Presenter E-mail or Presenters E-mails: johannes.single@cse-institut.de, jens.denecke@cse-institut.de, juergen.schmidt@cse-institut.de

Abstract

Hazard and Operability (HAZOP) studies are conducted to identify and assess potential hazards which originate from processes, equipment, and process plants. These studies are human-centered processes that are time and labor-intensive. Also, extensive expertise and experience in the field of process safety engineering are required. In the past, there have been several attempts by different research groups to (semi-)automate HAZOP studies. Within this research approach, a knowledge-based framework for the automatic generation of HAZOP worksheets was developed. Therefore, ontologies are used as a knowledge representation formalism to represent expert knowledge from the process and plant safety (PPS) domain. Based on that, a reasoning strategy is developed using semantic reasoners to identify hazards based on the developed ontologies in a HAZOP similar manner. The developed methodology is applied within a case study that involves a storage tank containing hexane. The automatically generated HAZOP worksheets are compared to the original worksheets. The results were evaluated and show that an ontology-based reasoning algorithm is well-suited to identify equipment-based hazardous scenarios. Node-based analyses can also be carried out by slightly adapting the method. The presented method can help to support HAZOP study participants and non-experts in conducting HAZOP studies.

Keywords: Computer-aided HAZOP, Ontology, Hazard identification, Ontology-based reasoning, Next Generation HAZOP

1 Introduction

Within Hazard and Operability (HAZOP) studies processes, process plants and equipment are systematically examined to identify and assess potential hazards and operability issues. HAZOP studies are carried out regularly within the life cycle of a process plant, for instance, during planning, commissioning, and revision. The methodology is human-centered and intended as a moderated brainstorming technique that is conducted by an interdisciplinary team of specialists. Hazard identification is usually based on the assessments of experts. It thus depends on their experience, education, and training, but also company policies and soft facts, e.g., group dynamics, safety culture, and sentiment. These circumstances make it time and labor-intensive process.

The automation of HAZOP studies and the provision of a computer-aid for HAZOP studies is a research topic for more than 30 years. Several researchers and research groups investigated the status quo regarding the automation of HAZOP studies and the general improvement of the method [1], [2] and [3]. The key results of this study are briefly described in the following.

Principally, computer systems for the computer-aid of HAZOP studies are concerned with the main tasks (1) creating and using a representation of the process plant and equipment, (2) storing relevant expert knowledge, (3) automatically infer conclusions about plausible hazardous scenarios and identifying safeguards [3]. There are numerous technologies available to perform these tasks. Typically, the representation of the process plant is based on graph-theoretical concepts, such as signed-directed graphs (SDG). There are various knowledge-representation formalisms, e.g., rule-based, frames, ontologies, to represent HAZOP relevant knowledge. Moreover, all previous approaches have an individual and technology-dependent strategy to identify hazardous scenarios.

Single et al. divided existing approaches to the automation of HAZOPs into three generations that are shown in Figure 1 [3]. The first approaches (Generation I) were based on so-called expert systems, which were mainly rule-based. This means relevant general and specific knowledge is represented using IF-THEN rules, e.g., “IF flammable substance THEN avoid source of ignition”. Generation II (see Figure 1) approaches integrated models regarding the behavior of process variables, process plant, or the intention of the equipment with rule-based elements. Approaches of generation III (see Figure 1) combined the usage of a model with case-based reasoning (CBR) (see [4]) to improve their reasoning capability.

Furthermore, a distinction between knowledge-based and data-driven methods can be made. Knowledge-based systems use semantic models to represent domain-specific relationships and system behavior. Based on those conclusions are drawn using reasoners or inference engines. Data-driven methods use raw or pre-processed data to train models to draw conclusions and predict system behavior. Historical data, process data, simulated data, or even records or documents, can be relevant here. Within this research, the focus is on knowledge-based systems and approaches.

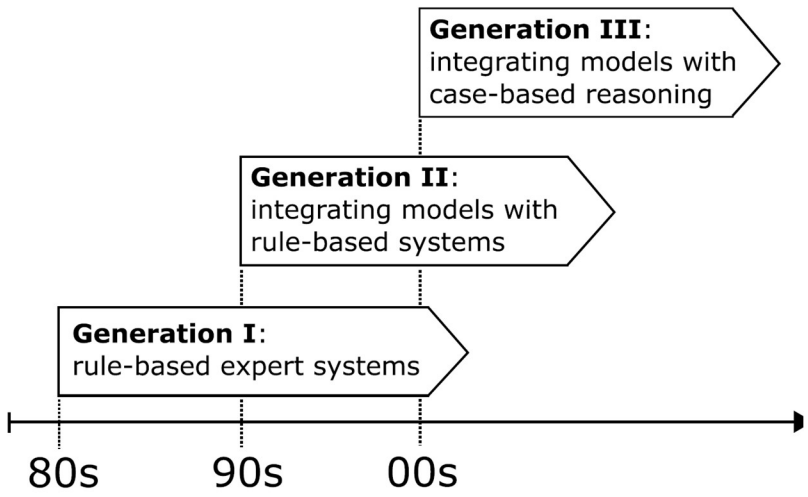

Figure 1: Different generations of automation approaches (adapted from [3])

1.1 Representation of knowledge

Conclusions are drawn within HAZOP studies based on the experience and knowledge of the participants. Therefore, the modeling of expert knowledge and its representation is a crucial stage in achieving the goal of automating HAZOP studies. Within the scope of this research, concepts from the knowledge domains substances, processes, process plants and units, equipment, hazards and malfunctions, causes, and consequences are considered.

There are various knowledge representation formalisms, such as *rules*, *semantic nets*, *frames,* and *ontologies* [5]. Rules can be used to represent logical implications and conditional actions [6]. Semantic nets make use of nodes to represent objects and edges (connections) to represent the relationships between these objects. Lehmann described a semantic net as a "graph of the structure of meaning" [7]. Frames can be used to represent stereotypical situations [8]. Davis et al. state that frames are useful for the definition of terms, objects, and for describing taxonomic relationships and relationships between objects [6]. Ontologies can also be used to represent objects and the semantic and hierarchical relationships between them. Depending on the used ontology language, restriction in the form of axioms can be added to specify the meaning of objects.

Ontologies have already been proposed within the scope of HAZOP studies by other research groups [9], [10]. Zhao et al. [11] used ontologies in their HAZOP expert system. Other research groups used ontologies within decision support systems [12], for the support of FMEA studies [13], or the representation of knowledge that was extracted from chemical accident databases [14].

Semantic nets, frames, and ontologies can be assigned to the group of knowledge graphs. Furthermore, frames are the more expressive successor of semantic nets. Frames and ontologies have similar capabilities, and differences can only be identified by looking at formal frame and ontology languages. Also, there are ontology languages that use rule-based elements [15]. Therefore, ontology and frame languages are briefly discussed in the following.

1.2 Ontology languages in brief

The development of ontology languages goes back to the 70s. One of the first logic-based ontology languages was KL-ONE [16]. Another acknowledged ontology language is the *Resource Description Framework* (RDF) language that was developed in the late 90s. It is intended for describing machine-understandable web resources, compare [17]. It can be used to explain concepts and their relations and forms the basis for other ontology languages. RDF models are formalized using the Extensible Markup Language (XML). RDF Schema (RDFS) is an extension of RDF [18]. DAML+OIL is based on RDF and was additionally provided with formal aspects of description logic (DL) [19]. It is the predecessor of the OWL language. The *Web Ontology Language* (OWL) can be used to unambiguously describe concepts, their relations, and restrict concepts using axioms. There are three sublanguages of OWL with different levels of expressiveness: OWL Lite, OWL DL, and OWL Full. To overcome the limitations of OWL regarding the ability to represent general rules the *Semantic Web Rule Language* (SWRL) was developed. SWRL is an extension of OWL with horn-like rules in the form of the *Rule Markup Language* (RuleML) [15]. OWL2 was developed to overcome drawbacks of OWL1, such as limited expressivity regarding properties and extended the set of built-in datatypes [20].

2 Objectives

This paper aims to present an ontology-based method to generate HAZOP worksheets automatically. Also, the importance of the ontological model and its semantic concepts is presented. Furthermore, a strategy is developed and described to infer logical conclusions from the proposed ontology. In the process, extended concepts such as causes (primary and secondary), chain of consequences, and safeguards are identified. After the ontology-based method is specified, it is applied within a case study to a hexane storage tank. Within the case study, an equipment-based automatic HAZOP is conducted using the described method. The automatically created HAZOP worksheets are compared to the original HAZOP worksheets and assessed.

3 Methodology

A computer system that utilizes a knowledge base to draw conclusions and infer facts is called a knowledge-based system. The first step in development is the conceptualization of relevant knowledge. The conceptualization process is shown in the upper part of Figure 2 (conceptualization process). It includes the design of a structure and the modeling and formalization of knowledge. Furthermore, the process plant, equipment, and substances must be adequately represented. The results of the conceptualization process are ontology-based knowledge representation and an object-oriented process unit model library (see Figure 2, upper part).

The conclusions that can be drawn based on the ontologies depend on the inference strategy that is used to evaluate the ontology (see Figure 2, lower part). The starting point is the selection of the relevant process units, processes, and the involved substances. This is done based on an object-oriented process unit model library. After the selection of the required input data, an inference algorithm infers causes, consequences, safeguards, and related concepts based on the (process) deviations, process unit, and substance information. This is shown in Figure 2.

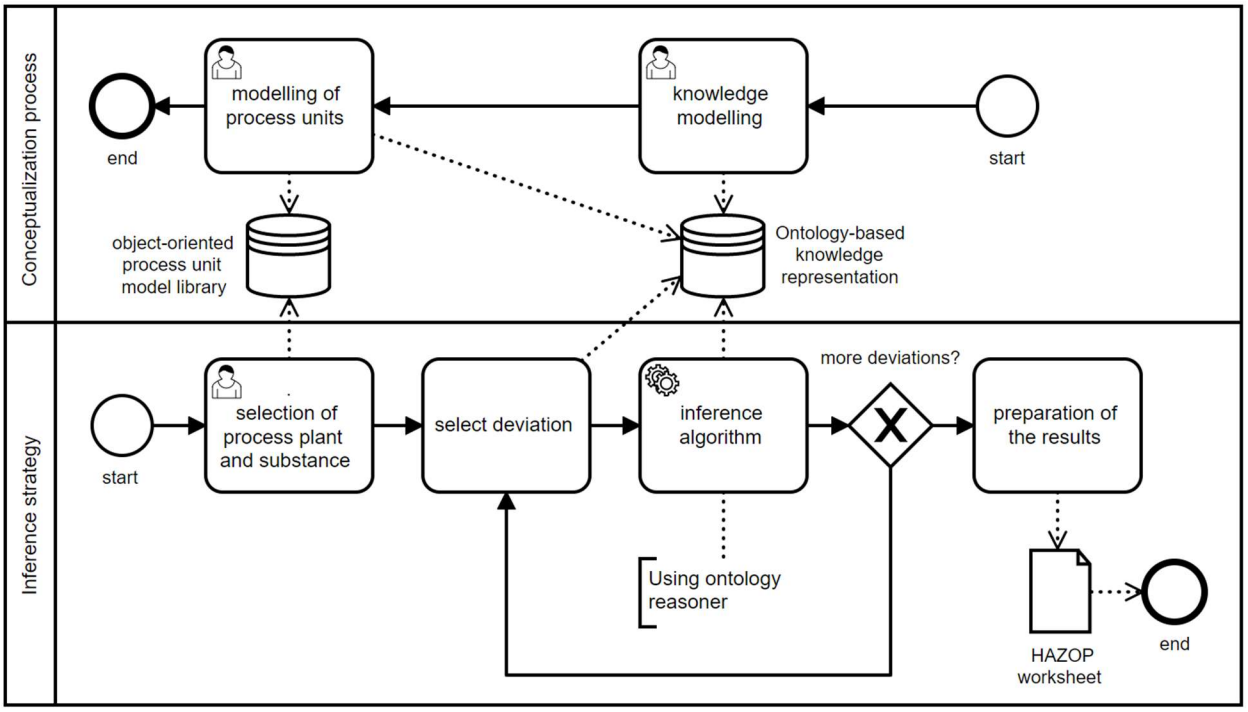

Figure 2: Conceptualization process and application of the evaluation logic

3.1 The ontological model of HAZOP concepts

The design of a knowledge model of the relevant concepts is the first step within the conceptualization phase (see Figure 2, upper part). This means that the concepts and the relationships between them must be identified and modeled carefully. In this context, modeling refers to set-theoretic and semantic modeling. Conclusions can be drawn from this semantic model regarding relevant scenarios.

According to Baybutt, "There are other elements of scenarios that may be important and often are not recorded in HAZOP study worksheets" [21]. The issue of a complete description of causal relationships within scenarios is considered within this research. Human experts can relate situations based on their experience to draw conclusions. Computer systems need well-defined models to infer facts and draw conclusions. For this reason, an extended HAZOP model is used in this thesis to illustrate causal relationships.

In this work, *causes* and *super causes* and *effects,* and *consequences* are distinguished. An ontological model always represents a simplification of reality. It is assumed that the core concepts *cause* and *effects* may be associated with deviations. *Effects* can lead to *consequences,* while *consequences* can have subsequent consequences. Furthermore, *safeguards* are implemented to mitigate *consequences* and *effects* and prevent *causes*. These core concepts are shown in Figure 3 and described in the following:

- **deviations** are composed of guidewords and (process) parameters and describe deviations from the design intent (part of a typical HAZOP worksheet),

- **causes** are a typical feature of HAZOP worksheets, and they describe the causes of the process deviations under consideration (part of a typical HAZOP worksheet),
- **super causes** are higher-level (primary) causes of causes, for instance, the cause *wrong rotating speed* of a pump could have the super cause *malfunction speed control*,
- **effects** arise from process deviations, and within the proposed model representation, for instance, this includes *rupture*, *overfilling* or *abnormal evaporation*,
- **consequences** potentially result from one or more effects; for instance, the effect *rupture* could have the consequence *loss of primary containment*. Also, consequences can have subsequent consequences which form a causal chain, e.g., the *formation of ex-atmosphere* leads to an *explosion* (part of a typical HAZOP worksheet),
- **safeguards** describe preventive, mitigative, primary and secondary safeguards, operational measures, and alarms (part of a typical HAZOP worksheet).

These fundamental relationships between the core concepts (*super cause*, *cause*, *effect, consequence, deviation*) are not detailed enough to be used for the automatic identification of scenarios. Therefore, complementary concepts and relationships are introduced to complete the ontological model. Also, there are relationships between the core and complementary concepts. These concepts include:

- **substance** involves properties of the substance, such as state of aggregation or hazardous attributes (e.g., flammability),
- **process unit** describes units (e.g., atmospheric storage tank) and operation related equipment (e.g., circulation pump, drain valve),
- **process** describes the interaction between substances and units,
- **circumstances** additional requirements that describe conditions such as ignition sources or other environmental conditions**.**

These considerations result in an ontological model that is shown in Figure 3. The core concepts are directly connected to the complementary concepts, such as process, intended function, and substance. Without taking the process unit, substances, or other circumstances into account, no reliable conclusions can be drawn about the HAZOP relevant concepts. Based on concepts deviation, unit, substance, and additional circumstances, potential causes, and effects can be modeled. Plausible causes and effects can only be identified based on an adequate representation of the process unit. In the case of oversimplification, specific causes and effects cannot be identified. The "OperationRelatedEquipment" concept is used for this purpose (compare Subsection 4.1).

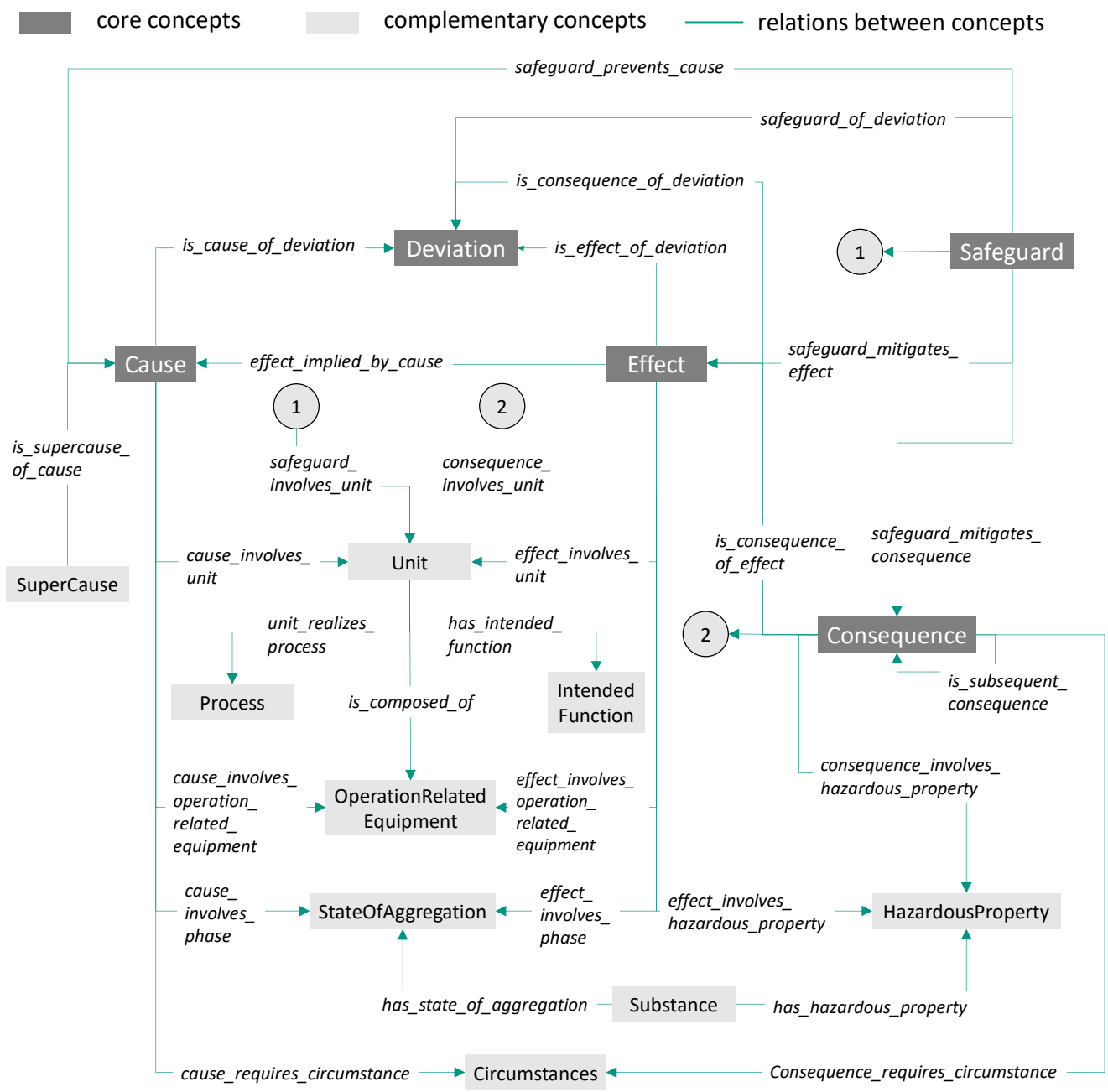

Figure 3: Complete ontological model of relevant concepts

Super(ordinate) causes (or primary causes) are used to describe causes further. For instance, the cause “ExternalLeakage” could have the super cause “DefectiveSeal”. Furthermore, consequences follow effects and depend on the substance properties, process unit, and additional circumstances. For instance, the *hazardous properties* of substances significantly influence potential *consequences*. For instance, the release of a flammable gas could lead to the formation of an ex-atmosphere, while the release of an inert gas could pose the danger of suffocation. Also, the additional circumstance „Confinement“ is required for the identification of the consequence “Explosion”. Proposals for safeguards can be modeled based on causes, effects, and consequences.

“Safeguard” concepts are also connected to the “Unit” concept since they strongly depend on the process unit and equipment.

The designed ontology was formalized using the Web Ontology Language (OWL) that was recommended by the World Wide Web Consortium (W3C) in 2004 [22]. Furthermore, the OWL DL sublanguage was chosen because of its expressiveness and efficient reasoning. Thus, the classes (concepts), properties (relationships), individuals (instances), and axioms are formalized with the OWL ontology language. The Python module Owlready2 was used to create the ontology in an object-oriented manner programmatically [23].

Within this research, the OWL DL ontology is mainly based on classes, object properties to relate these classes and axioms to restrict these classes. Annotations are another component of the OWL ontology language. Annotation properties can be used to model additional information, such as labels, descriptions, or further resources. They can be added to classes, instances, or properties (objects and data). Within this work, they were used to provide explanations regarding the ontological model, explain concepts in detail, and provide sources.

3.2 Inference algorithm

The inferences that can be drawn from the ontologies depend on the evaluation strategy (compare Figure 2, lower part). The scope of this paper is an equipment-based HAZOP. This means the node under consideration consists of a single process unit, including the directly involved equipment, e.g., circulation or transfer pump, cooling jacket, or power supply. In case a node consists of several process units, the inference algorithm must be executed for each unit, and additionally, interactions between the units must be considered. However, the evaluation of the ontologies always follows the same scheme.

The ontology is evaluated with different inputs and objectives within an inference algorithm to generate equipment-based HAZOP worksheets automatically. The call of a reasoner is an integral part of the inference algorithm, which is called multiple times. Reasoners are software packages that are used to infer logical consequences from ontologies. This means reasoners assume classification tasks, such as the computation of subsumption relations (e.g., concept A is a subset of concept B). According to Wang et al., a “classifier tries to build a model that satisfies all the axioms in the ontology” [24]. Furthermore, reasoners check the consistency of an ontology. Within this research, the HermiT reasoner was used, which is implemented in the Owlready2 package [25]. The inference algorithm is implemented using Python.

The developed inference algorithm is schematically shown in Figure 4 and described in the following. Based on the *deviations*, *substances*, and *process units*, *potential causes*, and *effects* are inferred. In this context, substance properties such as phase and hazardous attributes are relevant, but also the equipment, the intended function of the node (equipment) under consideration, and the process are essential. Furthermore, additional circumstances, such as ignition source or environmental conditions (e.g., confinement), are considered.

Super causes (primary causes) are inferred based on the causes and the process unit and associated equipment. Super causes are used to describe higher-level causes and thus the background of the causes in more detail.

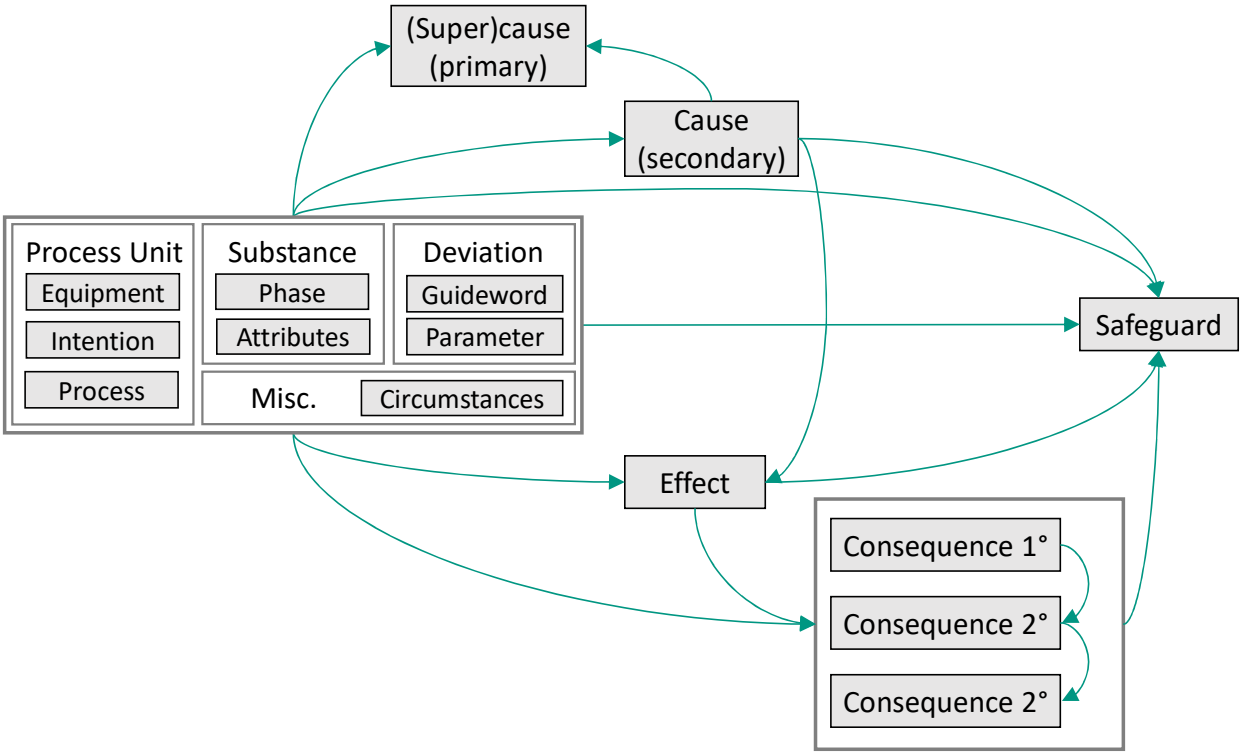

Figure 4: Inference strategy

The developed knowledge-models are based on formal logic. This context shall be illustrated using an example and with the help of description logic (DL):

$$OperatorError \equiv SuperCause \sqcap \exists isSupercauseOfCause.ClosedOutletValve.$$

The object properties (relationships between concepts), such as *isSuperCauseOfCause,* can be found in the ontological model in Figure 3.

Also, effects can be derived directly from causes. For instance, the cause "contamination by water" can imply the effect "drain line fracture" (e.g., under the condition of a cold temperature). Using description logic (DL), this can be expressed as:

$$\begin{aligned} DrainlineFracture \equiv\ & Effect \sqcap \\ & \exists effectImpliedByCause.ContaminationByWater \sqcap \\ & \exists isEffectOfDeviation.LowTemperature \sqcap \\ & \exists effectInvolvesPhase.Liquid \sqcap \\ & \exists effectInvolvesOperationRelatedEquipment.DrainValve. \end{aligned}$$

Effects can lead to consequences that are expressed as causal chains that consist of primary, secondary, and tertiary consequences, e.g., loss of primary containment → formation of the explosive atmosphere → explosion.

The hazardous attributes of substances and additional circumstances, e.g., ignition source, have a direct influence on the inferred consequences. For instance, this can be expressed as:

$$
\begin{aligned}
Fire \equiv\ & Consequence \sqcap \\
& \exists isSubsequentConsequence.LossOfPrimaryContainment \sqcap \\
& \exists consequenceInvolvesHazardousAttribute.Flammable \sqcap \\
& \exists consequenceRequiresCircumstance.IgnitionSource.
\end{aligned}
$$

Safeguards or required actions can either be derived based on potential effects and consequences or directly on probable causes. This means, there are preventive and mitigative safeguards, for instance:

$$
\begin{aligned}
PressureVacuumReliefValve \equiv\ & Safeguard \sqcap \\
& \exists safeguardInvolvesUnit.StorageTankUnit \sqcap \\
& \exists safeguardMitigatesEffect.CollapseOfEnclosure.
\end{aligned}
$$

4 Case Study: Hexane Tank

The proposed method is applied within a case study to evaluate its capabilities and the results. The case-study involves a hexane storage tank [26]. A schematic P&ID of the system is shown in Figure 5. It involves a storage tank (T-301) that stores liquid hexane and is half full in standard operation. Due to the vapor pressure of hexane, the storage tank is pressurized. The tanks' liquid level is controlled by a control loop, including a high-level alarm. The hexane tank is frequently filled from a tank truck via the inflow line and the unloading pump (3-40). The makeup pump (3-41) is supplying a downstream process that is not further specified.

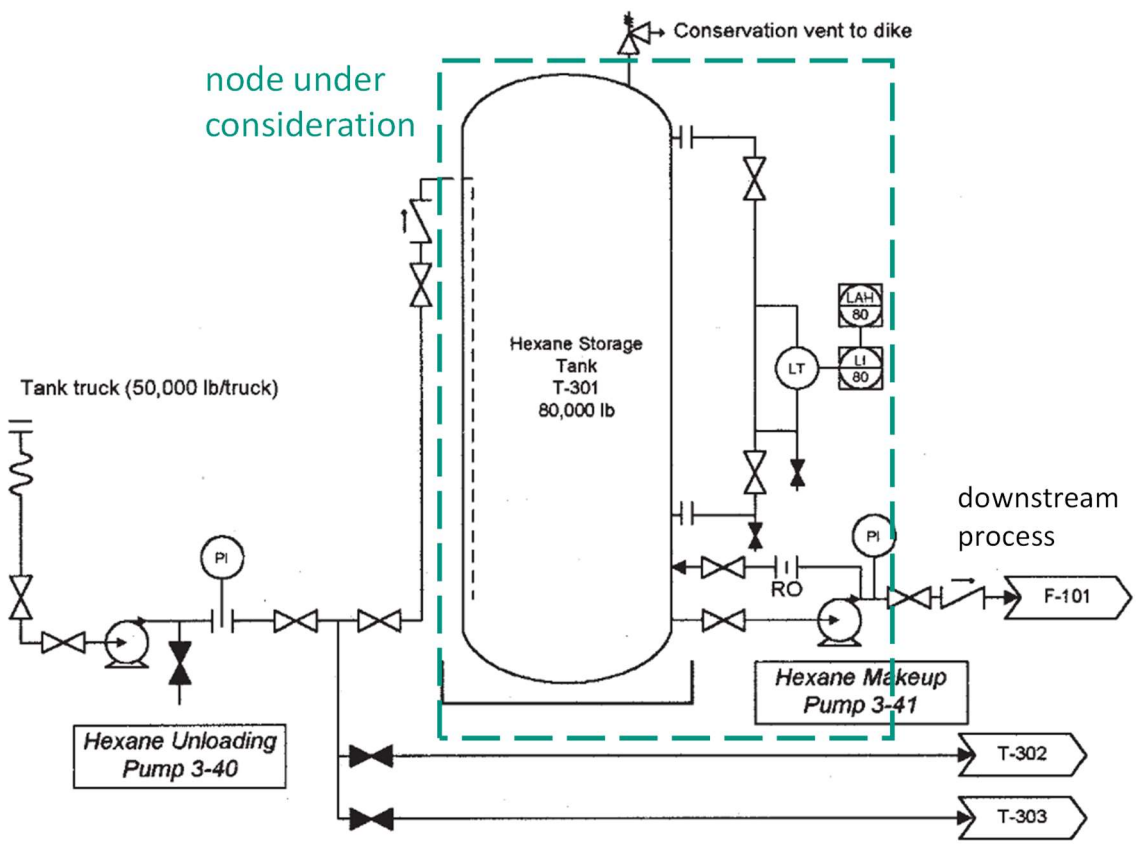

Figure 5: P&ID of the hexane tank, adopted from [26]

4.1 Representation of the process unit and substance

The hexane tank is represented using the following concepts for modeling the process unit:

- Process unit: *storage tank (T-301)*,
- Operation related equipment: *drain line* and *valve*; *inlet* and *outlet valve*; *level transmitter*, *level indicator,* and *controller; recirculation/transfer pump (3-41),*
- Intended function: storage of liquid hexane.

Furthermore, substance attributes, such as the state of aggregation and the hazardous properties of hexane, are also considered:

- Intended state of aggregation: *liquid*,
- Hazardous properties: *flammable*, *harmful*, *health* and *environmental hazard*.

Additionally, circumstances such as potential *ignition sources*, the *presence of ambient air*, *human intervention*, and the *introduction of impurities* during the filling process are assumed. These details regarding the substance and the process unit and involved operation related equipment and additional circumstances are modeled as ontology concepts that serve as an input for the inference algorithm.

4.2 Results of the proposed methodology

The application of the described methodology leads to automatically created HAZOP worksheets that are shown in Table 1 and in the appendix in Table A-1, Table A-2, Table A-3. Overall, 39 potential scenarios with eight different deviations were identified. The selected deviations are the same as in the original HAZOP example [26]. Within the proposed approach, different names were used for the deviations. Instead of the deviation *high concentration of contaminants,* the deviation *other than composition* was used. Also, instead of *loss of containment,* the deviation *elsewhere flow* was used. Based on the inference strategy, one cause or effect per scenario is identified, while several super causes and a chain of consequences are possible. In addition to the typical HAZOP worksheet columns, the considered substances, process unit, super causes, and effect are shown. Each row represents a scenario. In conventional HAZOP worksheets, no distinction is made between causes and super causes and effects and consequences. These are recorded in the same column. Within the automatically generated HAZOP worksheets, details such as super-causes, substances, and equipment are explicitly listed for each scenario to enable plausibility checks. There are also scenarios where no super cause or effect or no other consequence has been identified. Furthermore, the same effects can lead to different chains of consequences. Different chains of consequences form different scenarios with different safeguards and are listed in separate rows. For instance, the effect "Rupture" could lead to the consequence chains "LOPC, FormationOfExAtmosphere, Explosion" or "LOPC, DirectIgnition, Fire", compare Table 1. Within the automatically generated HAZOP worksheets, there are no references to other scenarios. In each series, the scenario is described in full, including the process unit under consideration and the substance. Each scenario (each row) is different but may share identical columns.

Table 1: Automatically created extended HAZOP worksheet for the deviations “high level” and “low level”

Pos.	Unit	Substance	Deviation	Supercause	Cause	Effect	Consequence	Safeguard
0	StorageTankUnit (T-301)	Hexane	HighLevel	FailureControlLoop, OperatorError	IncorrectFilling	Rupture	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PressureVacuumReliefValve
1	StorageTankUnit (T-301)	Hexane	HighLevel	FailureControlLoop, OperatorError	IncorrectFilling	Rupture	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PressureVacuumReliefValve
2	StorageTankUnit (T-301)	Hexane	HighLevel	OperatorError, ValveFailure	ClosedOutletValve	Rupture	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PressureVacuumReliefValve
3	StorageTankUnit (T-301)	Hexane	HighLevel	OperatorError, ValveFailure	ClosedOutletValve	Rupture	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PressureVacuumReliefValve
4	StorageTankUnit (T-301)	Hexane	LowLevel	DepositionOfImpurities	BlockedInflowLine	PumpDryRunning	PotentialProcessInterruption	FlameArrester, LevelControllerLowAlarm
5	StorageTankUnit (T-301)	Hexane	LowLevel	OperatorError	LossOfInflow	PumpDryRunning	PotentialProcessInterruption	FlameArrester, LevelControllerLowAlarm
6	StorageTankUnit (T-301)	Hexane	LowLevel	FailureControlLoop, OperatorError	IncorrectFilling	PumpDryRunning	PotentialProcessInterruption	FlameArrester, LevelControllerLowAlarm
7	StorageTankUnit (T-301)	Hexane	LowLevel	FailureControlLoop, OperatorError	IncorrectFilling	UnfavourableVaporConcentration	FormationOfExAtmosphere, Explosion	DefineExProtectionZone, FlameArrester, Inertization, LevelControllerLowAlarm
8	StorageTankUnit (T-301)	Hexane	LowLevel	FailureControlLoop, OperatorError	IncorrectFilling	UnfavourableVaporConcentration	PotentialProcessInterruption	FlameArrester, LevelControllerLowAlarm
9	StorageTankUnit (T-301)	Hexane	LowLevel	OperatorError, ValveFailure	ClosedInletValve	PumpDryRunning	PotentialProcessInterruption	FlameArrester, LevelControllerLowAlarm
10	StorageTankUnit (T-301)	Hexane	LowLevel	OperatorError, ValveFailure	ClosedInletValve	UnfavourableVaporConcentration	FormationOfExAtmosphere, Explosion	DefineExProtectionZone, FlameArrester, Inertization, LevelControllerLowAlarm
11	StorageTankUnit (T-301)	Hexane	LowLevel	OperatorError, ValveFailure	ClosedInletValve	UnfavourableVaporConcentration	PotentialProcessInterruption	FlameArrester, LevelControllerLowAlarm

4.3 Original HAZOP results

Within the original HAZOP study, the following deviations were considered: *high level*, *low level*, *high temperature*, *low temperature*, *high pressure*, *low pressure*, *high concentration of contaminants,* and *loss of containment*. For practical use, the deviations are directly composed of guidewords and parameters. The results of the original HAZOP worksheets are shown in Table 2 and

Table 3. The direct comparison of the automatically generated with the original results is made in the column "Identified".

Table 2: HAZOP table (part 1) based on [26]

Pos.	Deviation	Causes	Identified	Consequence	Identified	Safeguard/Recommendations	Identified
0	High level	Flow from tank truck not discontinued before tank capacity has been reached Inventory control error – truck arrives before needed	Yes Yes	High pressure	Indirect	Level indication with high-level alarm (audible in the control room) Hexane unloading procedures with a checklist that includes checking field reading of tank before unloading	Yes Other
1	Low level	Inventory control error – truck arrives too late Low flow or no flow – line from the tank truck to hexane storage tank T-301 through hexane unloading pump	Yes Yes	No safety consequences – potential process interruption if not refilled before the downstream feed tank is empty	Yes	-	Other
2	High temperature	No credible causes identified	Other	-	Other	-	Other
3	Low temperature	Low ambient temperature while there is water contamination in the tank	Indirect	Possible freezing of accumulated water in the heel of the tank or the tank's drain line or instrument lines, resulting in fracture of the drain line and loss of containment	Indirect	-	Other
4	High pressure	High level	Other	Release of hexane through the relief valve into the tank's dike Fire hazard affecting a large area if not contained by the dike (consequence category 4 or 5) Loss of containment (if the overpressure	Yes Yes Yes	-	Other

				cause exceeds the tank pressure rating)			

In case the automatically created results match the original results, it is indicated with a "yes" in the "Identified" column. In the case of a similar conclusion within a different scenario, it is indicated with an "Indirect". In case another conclusion is drawn, it is indicated with an "Other". For instance, in Table 3, another safeguard for the "low pressure" deviation was identified, compare to Figure A-2. Also, there can be additional scenarios that have been found. For instance, multiple "high temperature" deviation scenarios have been identified, while no scenarios have been listed in the original worksheets, see Table 2.

Table 3: HAZOP table (part 2) based on [26]

Pos.	Deviation	Causes	Identified	Consequence	Identified	Safeguard/Recommendations	Identified
5	Low pressure	Tank blocked in before cooldown, following steam-out	Yes	Equipment damage resulting in a collapse of the tank under a vacuum	Yes	Standard procedures for steam-out of vessels	Other
6	High concentration of contaminants (Other than composition)	Water not completely drained following a steam-out or washout High concentration of contaminants – line from the tank truck to hexane storage tank T-301 through hexane unloading pump	Yes Yes	Possible freezing of accumulated water in the tank during a period of low ambient temperature	Yes	-	Other
7	Loss of containment (Elsewhere flow)	Corrosion Erosion External fire External impact Gasket, packing or seal failure Improper maintenance Instrument or instrument line failure Material defect Sample station valve leaking Vent or drain valve leaking Low temperature	Partially	Release of hexane Fire hazard affecting a large area particularly if the capacity of the dike is exceeded	Yes Yes	Operation/maintenance response as required, including isolation if needed Capability to manually isolate the tank Periodic non-destructive inspection per API recommended practice and ASME code Relief valve that discharges to the tank's dike Dike sized for 1.5 times the capacity of the tank Emergency response procedures	No No Yes Yes Yes No

	High pressure (if the overpressure cause exceeds the equipment pressure rating)	

4.4 Comparison and discussion of the HAZOP worksheets

Within a conventional HAZOP, potential causes and consequences are usually identified based on a deviation and information concerning the node. For instance, within the original HAZOP worksheet in Table 3 (Pos. 5) based on the deviation "low pressure", the cause "tank blocked in before cooldown" and the consequence "equipment damage resulting in a collapse". Based on these results, appropriate safeguards can be selected. Within the developed extended worksheet, there are multiple causes of the deviation "low pressure" (see Table A-2, Scenario 27-32). For example, the cause "ObstructedVentPath" with the super causes "FaultyInstallation" and "HumanError" have been automatically identified. Furthermore, the effect "CollapseOfEnclosure" with two different consequence chains "LossOfContainment" and "Fire" or "Explosion" have been identified. Based on these findings, multiple safeguards were proposed, e.g., "CollectingBasin", "FlameArrester", "PeriodicalExamination", "PressureVacuumReliefValve" (compare Table A-2, Scenario 27-28).

From a qualitative point of view, a large part of the causes and consequences were identified (compare Table 2 and Table 3). The direct comparison of HAZOP results requires the interpretation of scenarios. Different chains of causes or consequences are shown separately in a new row (see Table 1, Table A-1, Table A-2, Table A-3). Different deviations may lead to similar scenarios. Some scenarios show similarities or were interpreted differently with similar conclusions. For instance, in the original HAZOP, the *consequence* of the *high level* deviation is "high pressure". Within the automatically generated worksheets, the *consequence* is "Rupture". In the original worksheet, this consequence is again listed under the *high pressure* deviation. References to other scenarios have been avoided, and all scenarios have been described in full to improve readability. Furthermore, the *consequence* "freezing and fracture of the drain line" was recognized within this work with the deviation *other than composition*. In the original HAZOP, it was recognized with the deviation *low temperature*. This *consequence* is the result of the two deviations *low temperature* and *other than composition*.

These examples show that different terms can be used while similar conclusions can be drawn. This is also a typical issue within conventional HAZOP studies because experts use different vocabulary, which also depends on company-specific guidelines. Within the automatically created worksheets, *Causes*/*Supercauses*, and *Effects*/*Consequences* are listed separately. In the original HAZOP worksheet, all causes and consequences are listed together. For instance, in the original worksheet, a cause of the deviation "loss of containment" is "Corrosion". Within the proposed approach, "Corrosion" is a super cause while the corresponding cause is "ExternalLeakage". A separate listing of *Causes*/*Supercauses* and *Effects*/*Consequences* improves the transparency of causal relationships of the automatically generated results and helps to identify and resolve inconsistencies. Each scenario to be identified must first be represented within the ontological model. Scenarios must also be abstracted and simplified in such a way that they can be represented within the ontological model. New concepts can only be implemented by taking the other concepts and the ontological structure into account. Otherwise, existing concepts could be dissolved, or the wrong conclusions could be drawn and the wrong scenarios identified accordingly. Since the automatically generated results are based on ontologies, a plausibility check by human experts is required.

Safeguards depend heavily on the hazard potential, risk assessment, industry, and company-specific guidelines. Some of the listed safeguards can also correspond to general recommendations that are not tied to a specific scenario. For example, a flame arrester for a storage tank containing a flammable liquid could be recommended (see position ten in Table 1). Thus, the automatically identified safeguards are proposals that still require expert evaluation.

In general, more scenarios were recorded within the proposed method. On the other hand, the original HAZOP was intended to demonstrate different methods. It is questionable to what extent the completeness of the original HAZOP was the claim of the authors [26]. Nevertheless, this HAZOP example is well suited to compare the quality of the results. The numbers of identified causes and consequences of the HAZOP worksheets are shown in Figure 6 and Figure 7. To determine the number of concepts identified, chains of causes and consequences are counted. For instance, the scenario with the super cause "ExternalFire" and the cause "ThermalExpansion" would count as one in Figure 6. The scenario with the effect "Rupture" and the consequence "LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion" would count as two consequences in Figure 7. This means that intermediate events in the consequence chain, such as "FormationOfExAtmosphere" are not counted separately.

Overall the number of automatically identified causes (own: 34, original: 22) and consequences (own: 25, original: 13) is higher than in the original HAZOP. Within the proposed approach, more causes and consequences have been identified, especially regarding the "high temperature" deviation. In the original worksheet, more causes regarding the "elsewhere flow" deviation have been identified. In both HAZOP approaches, many scenarios consider a loss of containment. This can lead to a fire or even an explosion due to the flammability of hexane. Within the original HAZOP worksheet, the scenario of an explosion was not considered. Therefore, a different number of consequences can be explained.

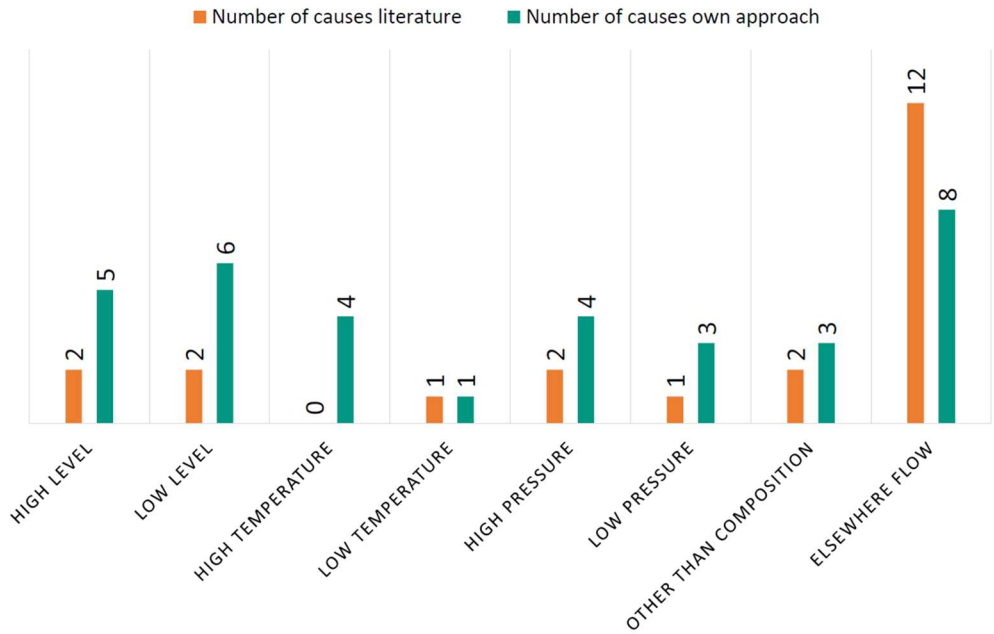

Figure 6: Comparison of the identified causes

The focus of the original HAZOP was not on the identification of safeguards or recommendations. The selection of safeguards depends strongly on the identified causes and consequences. Since the causes and consequences differ, the number of identified safeguards is not directly compared within this paper.

A quantitative evaluation allows metrics to be calculated to measure the extent or focus on specific deviations. For example, the mean value of the identified causes (own approach) is 4.25 per deviation, while the mean value of the consequences is 3.125 per deviation. Concerning Figure 6 and Figure 7, it can be concluded that the number of causes per deviation varies more than the number of consequences. This information can be used, for example, to improve the ontological model. Concerning HAZOP automation systems, HAZOP metrics can be used to track the quantitative effect on the HAZOP results of changes to the ontological model.

Nevertheless, the comparison of the number of identified scenarios does not allow any statement about the completeness or the quality of the results. This can only be done based on qualitative aspects through the analysis and interpretation of scenarios by human experts. Based on these expert assessments, concepts in the ontology can be extended to identify further scenarios in the future. Within this case study, the presented knowledge-based system was able to generate qualitatively equivalent results compared to the original HAZOP.

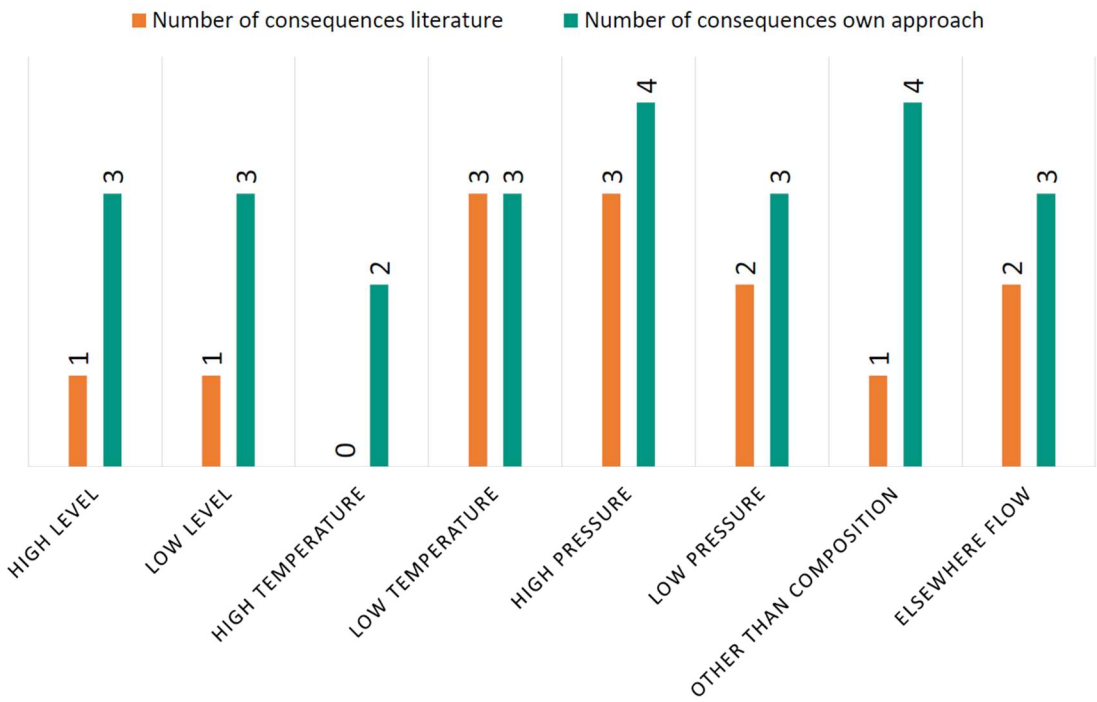

Figure 7: Comparison of the identified potential consequences

5 Conclusions

In this research approach, a knowledge-based methodology is developed to generate HAZOP worksheets automatically. For the representation of knowledge from the HAZOP domain, ontologies are used. Within the knowledge modeling process, it is particularly essential to define relationships between knowledge concepts unambiguously. The completeness of the results is directly dependent on the quality and completeness of the concepts of the ontology. After the design, conceptualization, and formalization of a suitable ontology, an inference strategy was designed to draw conclusions about the ontology. Based on that, HAZOP worksheets have been generated automatically. Conventional HAZOP studies form socio-technical systems where the results depend on the skills of the HAZOP team. In computer systems, the results depend directly on the representation of knowledge and the evaluation algorithms. Human error no longer occurs only within the HAZOP study but also in the modeling process of the ontology. The described methodology is applied within a case study to a hexane storage tank as an equipment-based HAZOP analysis. Afterward, the automatically generated results are evaluated and compared to the original HAZOP results. The presented results show that the developed ontology and the ontology-based reasoning algorithm is well-suited to generate equipment-specific HAZOP worksheets automatically. More research is needed regarding the application of the method within node-based HAZOP studies, regarding the extraction of information on process units from P&IDs, and the propagation of hazards throughout the process plant. At this moment, there is no automatic risk assessment, and it must be carried out by human experts. Furthermore, the identified safeguards are proposals that always require expert judgment. Finally, this research approach contributes to demonstrating the capabilities of knowledge-based systems for the use in HAZOP studies.

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Appendix A

Table A-1: Automatically created extended HAZOP worksheet for the deviations “other than composition” and “elsewhere flow”

Pos.	Unit	Substance	Deviation	Supercause	Cause	Effect	Consequence	Safeguard
12	StorageTankUnit (T-301)	Hexane	OtherThanComposition	CondensationAirHumidity, IntroductionOfRainwater	ContaminationByWater	DrainlineFracture	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	AutomaticWaterDetectionSystem, CollectingBasin, DefineExProtectionZone, FlameArrester, PeriodicalSampleTaking
13	StorageTankUnit (T-301)	Hexane	OtherThanComposition	CondensationAirHumidity, IntroductionOfRainwater	ContaminationByWater	DrainlineFracture	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, AutomaticWaterDetectionSystem, CollectingBasin, FlameArrester, PeriodicalSampleTaking
14	StorageTankUnit (T-301)	Hexane	OtherThanComposition	Corrosion	Contaminants	PoorProductQuality	NoSafetyConsequence	AutomaticWaterDetectionSystem, FlameArrester, PeriodicalSampleTaking
15	StorageTankUnit (T-301)	Hexane	ElsewhereFlow	Corrosion, DefectiveFlangedJoints, DefectiveWelds, Erosion, MaterialDefect, MechanicalDamage, OperatorError	LeakingDrainValve	PoolFormation	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester
16	StorageTankUnit (T-301)	Hexane	ElsewhereFlow	Corrosion, DefectiveFlangedJoints, DefectiveWelds, Erosion, MaterialDefect, MechanicalDamage, OperatorError	LeakingDrainValve	PoolFormation	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester
17	StorageTankUnit (T-301)	Hexane	ElsewhereFlow	Corrosion, DefectiveFlangedJoints, DefectiveWelds, Erosion, MaterialDefect, MechanicalDamage	ExternalLeakage	PoolFormation	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester
18	StorageTankUnit (T-301)	Hexane	ElsewhereFlow	Corrosion, DefectiveFlangedJoints, DefectiveWelds, Erosion, MaterialDefect, MechanicalDamage	ExternalLeakage	PoolFormation	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester

Table A-2: Automatically created extended HAZOP worksheet for the deviations “high pressure” and “low pressure”

Pos.	Unit	Substance	Deviation	Supercause	Cause	Effect	Consequence	Safeguard
19	StorageTankUnit (T-301)	Hexane	HighPressure	DepositionOfImpurities	BlockedOutlet	Rupture	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PressureReliefValve, PressureVacuumReliefValve
20	StorageTankUnit (T-301)	Hexane	HighPressure	DepositionOfImpurities	BlockedOutlet	Rupture	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PressureReliefValve, PressureVacuumReliefValve
21	StorageTankUnit (T-301)	Hexane	HighPressure	DepositionOfImpurities	BlockedOutlet	Crack	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PressureReliefValve, PressureVacuumReliefValve
22	StorageTankUnit (T-301)	Hexane	HighPressure	DepositionOfImpurities	BlockedOutlet	Crack	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PressureReliefValve, PressureVacuumReliefValve
23	StorageTankUnit (T-301)	Hexane	HighPressure	AbnormalHotIntake, ExternalFire, SolarRadiation	ThermalExpansion	Rupture	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PressureReliefValve, PressureVacuumReliefValve
24	StorageTankUnit (T-301)	Hexane	HighPressure	AbnormalHotIntake, ExternalFire, SolarRadiation	ThermalExpansion	Rupture	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PressureReliefValve, PressureVacuumReliefValve
25	StorageTankUnit (T-301)	Hexane	HighPressure	AbnormalHotIntake, ExternalFire, SolarRadiation	ThermalExpansion	Crack	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PressureReliefValve, PressureVacuumReliefValve
26	StorageTankUnit (T-301)	Hexane	HighPressure	AbnormalHotIntake, ExternalFire, SolarRadiation	ThermalExpansion	Crack	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PressureReliefValve, PressureVacuumReliefValve
27	StorageTankUnit (T-301)	Hexane	LowPressure	FaultyInstallation, HumanError	ObstructedVentPath	CollapseOfEnclosure	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PeriodicalExaminationVentOpening, PressureVacuumReliefValve
28	StorageTankUnit (T-301)	Hexane	LowPressure	FaultyInstallation, HumanError	ObstructedVentPath	CollapseOfEnclosure	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PeriodicalExaminationVentOpening, PressureVacuumReliefValve
29	StorageTankUnit (T-301)	Hexane	LowPressure	MalfunctionTransferPump	ExcessiveFluidWithdrawal	CollapseOfEnclosure	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PeriodicalExaminationVentOpening, PressureVacuumReliefValve
30	StorageTankUnit (T-301)	Hexane	LowPressure	MalfunctionTransferPump	ExcessiveFluidWithdrawal	CollapseOfEnclosure	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PeriodicalExaminationVentOpening, PressureVacuumReliefValve
31	StorageTankUnit (T-301)	Hexane	LowPressure	-	AmbientTemperatureChange	CollapseOfEnclosure	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester, PeriodicalExaminationVentOpening, PressureVacuumReliefValve
32	StorageTankUnit (T-301)	Hexane	LowPressure	-	AmbientTemperatureChange	CollapseOfEnclosure	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester, PeriodicalExaminationVentOpening, PressureVacuumReliefValve

Table A-3: Automatically created extended HAZOP worksheet for the deviations “high temperature” and “low temperature”

Pos.	Unit	Substance	Deviation	Supercause	Cause	Effect	Consequence	Safeguard
33	StorageTankUnit (T-301)	Hexane	HighTemperature	MechanicalDamage	HeatInputByRecirculationPump	AbnormalEvaporation	FormationOfExAtmosphere, Explosion	DefineExProtectionZone, FlameArrester
34	StorageTankUnit (T-301)	Hexane	HighTemperature	MechanicalDamage	HeatInputByRecirculationPump	ExceedingFlashPoint	FormationOfExAtmosphere, Explosion	DefineExProtectionZone, FlameArrester
35	StorageTankUnit (T-301)	Hexane	HighTemperature	AbnormalHotIntake, ExternalFire, SolarRadiation	AbnormalHeatInput	AbnormalEvaporation	FormationOfExAtmosphere, Explosion	DefineExProtectionZone, FlameArrester
36	StorageTankUnit (T-301)	Hexane	HighTemperature	AbnormalHotIntake, ExternalFire, SolarRadiation	AbnormalHeatInput	ExceedingFlashPoint	FormationOfExAtmosphere, Explosion	DefineExProtectionZone, FlameArrester
37	StorageTankUnit (T-301)	Hexane	LowTemperature	-	AmbientTemperatureChange	BrittleFracture	LossOfPrimaryContainment, FormationOfExAtmosphere, Explosion	CollectingBasin, DefineExProtectionZone, FlameArrester
38	StorageTankUnit (T-301)	Hexane	LowTemperature	-	AmbientTemperatureChange	BrittleFracture	LossOfPrimaryContainment, DirectIgnition, Fire	AutomaticFireSuppressionSystems, CollectingBasin, FlameArrester

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An Efficient and Effective Approach for Performing Cost Benefit Analysis

Henrique M. Paula, Founder & President, Galvani Risk Consulting, LLC, The Woodlands, TX USA
Donald K. Lorenzo. P.E., Process Safety Instructor, Knoxville, TN USA
Marcelo Costa Jr., Risk Engineer, Orlândia, SP, Brazil

*Presenter E-mail or Presenters E-mails: hmpaula@galvanitx.org

Abstract

Risk management involves the application of one or more of a variety of inter-related techniques (hazard and operability [HAZOP], hazard identification [HAZID], facility risk review [FRR], etc.). Most of these applications result in recommendations or suggestions for risk reduction. In fact, the number of recommendations is often significant (well over 100 in many cases and thousands in a case study discussed in this paper). A large number of recommendations is beneficial because each recommendation provides an opportunity for risk reduction and/or other actions for asset improvement. However, a large number of recommendations can overwhelm the managers responsible for their implementation, making it difficult to decide what to do and/or when to do it. Additionally, there may be overlap or similarities of recommendations from the application of different techniques, sometimes confusing their review and resolution.

Cost benefit analysis is a powerful tool to help managers sort through the recommendations and effectively/efficiently prioritize them. It consists of evaluating the risk reduction and the estimated cost associated with each recommendation, including capital expenditures (CAPEX) and operational expenditures (OPEX). This paper provides a simple, efficient, and effective approach for performing cost benefit analysis. This method is not intended to replace more detailed methodologies. Rather, it is a complementary tool particularly useful for applications with many recommendations.

Keywords: Cost Benefit Analysis, Efficiency, Effectiveness, Risk Reduction, Prioritize

1. SUMMARY OF THE METHODOLOGY

This paper summarizes a cost benefit analysis approach based on: (a) training course manuals on process hazard analysis (PHA) and quantitative risk assessment (QRA) [1], (b) research performed for the U.S. Coast Guard [2], and (c) several studies conducted by the authors for oil production and refining companies [3] [4].

The basis of the approach is that the priority of a recommendation is (a) directly proportional to the risk reduction expected from the implementation of the recommendation and (b) inversely proportional to the cost of implementation:

$$\text{Priority} = \Delta\ \text{Risk} / \text{Cost}$$

That is, the Priority, or ***Benefit to Cost Index*** (***BCI***), is the ratio of the ***risk reduction*** to the cost of implementation of the recommendation.

1.1 Evaluation of Risk Reduction (Δ Risk)

The expected ***risk reduction*** (Δ Risk) is:

- The expected risk associated with continuing to operate under the current situation (i.e., if the recommendation is not implemented)
- Minus the expected risk associated with continuing to operate after the changes are implemented (i.e., if the recommendation is implemented)

If we assume that the risk associated with a scenario is the product of (a) the frequency of occurrence and (b) the consequence(s), then,

$$\Delta Risk = \sum_{n=1}^{D} F_{n,\ \text{before}} \times C_{n,\ \text{before}} - \sum_{n=1}^{D} F_{n,\ \text{after}} \times C_{n,\ \text{after}}$$

Where:

D = number of accident scenarios affected by the recommendation

F_n = frequency of accident scenario n

C_n = consequences of accident scenario n

The consequences of interest may include any combination of a variety of concerns, including worker safety, public safety, environmental, business interruption, reliability, and so forth.

The frequency and consequences <u>before</u> the implementation of a recommendation (current situation) are evaluated during the hazard analysis [5], and this is typically accomplished using risk matrices. The

frequency and consequences after the implementation of each recommendation are evaluated as follows, for each recommendation individually:

- **Identify all risk scenarios that would be affected by the recommendation.** That is, the risk review team identifies the scenarios associated with the risks that the recommendation is trying to reduce. A recommendation may impact a scenario by reducing the frequency of the scenario, by mitigating one or more consequence(s) associated with the scenario, or by doing both.
- **Assess the expected impact that each recommendation has on (a) the frequency and (b) the consequences associated with each affected scenario.** This involves several individual evaluations for each recommendation and is accomplished by selecting Impact Categories from Table 1. Specifically, the team selects the impact category that best applies to the frequency of the scenario and additionally an impact category for each consequence of interest.
- **Evaluate the risk after the implementation of a recommendation.** This is accomplished by multiplying the original assignments of the frequency / consequences for each scenario by the corresponding Risk Reduction Factor from Table 1.

Table 1 — Example Categories for Assessing the Benefits of Implementing Recommendations [2]

Impact Category	Benefits of Implementing Recommendations	Risk Reduction Factor
1	**No Impact** The recommendation does not help reduce the frequency or a specific consequence of a scenario	1.00
2	**Small Impact** The recommendation helps reduce the frequency or a specific consequence of a scenario, but this reduction is relatively small (no more than about 10%)	0.90
3	**Small to Medium Impact** The recommendation definitely helps reduce the frequency or a specific consequence of a scenario (as much as 50%)	0.50
4	**Medium to Major Impact**	0.10

Table 1 — Example Categories for Assessing the Benefits of Implementing Recommendations [2]		
Impact Category	**Benefits of Implementing Recommendations**	**Risk Reduction Factor**
	The recommendation significantly reduces the frequency or a specific consequence of a scenario (as much as 90%)	
5	**Major Impact** The recommendation essentially eliminates the frequency or a specific consequence of a scenario (more than about 99%)	0.01

1.2 "Limitations" of the Evaluation of Risk Reduction (Δ Risk)

The methodology presented in the previous section to evaluate ΔRisk does not address some potential contributors to risk:

- The expected risk associated with making the modifications suggested by the recommendation (or simply the ***modification risk***)
- The possibility that the recommendation will increase risk by creating, for example, new hazards

We added quotation marks to the word "limitations" to recognize that although the methodology does not consider these contributors, these risks are in fact outside the scope of a cost benefit analysis. This is discussed in more detail next.

Regarding the ***modification risk***, suppose the implementation of a recommendation requires construction at the facility. Also, suppose that at least a portion of the process at this facility continues to operate during construction. It is possible that an accident could occur during construction (e.g., a crane accident that damages process equipment and causes a release of hydrocarbons). Accidents may also result from other deficiencies or errors during the implementation of the recommendation, including during the phases of design, engineering, procurement, manufacturing, training of operations/maintenance staffs, and several others.

Regarding the possibility that a recommendation may increase the risk of some scenarios or create ***new hazards***, consider, for example, a recommendation to add a fire sprinkler to reduce the risk of burning a building down. While the proposed sprinkler system should reduce the risk of fire, it may also increase in the risk of water damage (e.g., from inadvertent operation of the sprinkler).

Our methodology does not consider the ***modification risk*** or ***new hazards*** because of the difficulty in evaluating them at the time that we perform the cost benefit analysis. To properly evaluate these risks, the analysts would need detailed information about the change (design documentation, construction plans, updated P&IDs, revised operating procedures, etc.), and this information is unlikely to be available when performing the cost benefit analysis.

In general, experienced safety/risk analysis teams try to account potential detrimental effects of the recommendations. In addition, operating companies have management systems in place to ensure adequate controls of modifications (e.g., a Management of Change [MOC] system), including procedures for all activities associated with the implementation of the recommendations. In the case of adequate controls, the risk of implementing the recommendations should be small compared to the other risks addressed here. At any rate, we have not accounted for this issue in our previous applications of the approach presented in this article.

1.3 Evaluation of the Cost of Implementation of Recommendations

The expected cost is evaluated using Cost Categories and Cost Ranges, as illustrated in Table 2. In selecting a cost category for each of the recommendations, the review team considers the total cost associated with the recommendation, including capital expenditures (CAPEX) and operational expenditures (OPEX) related to design, engineering, procurement, construction, installation, training (e.g., operational and maintenance staffs), etc.

Obviously, this method for cost evaluation is only an approximation based on the experience of the review team. A precise cost can only be estimated after managers review each recommendation and decide the specific action that should be taken to address it. That is, the cost estimate depends on the details of implementation of each recommendation, and this information is generally not available when performing the cost analysis. However, ranges like those in Table 2 are broad enough that it is possible to select a reasonable cost category even without these details.

2. CASE STUDY 1

Over a period of a little over one year, a major oil company conducted a series of safety, hazard, and risk studies for nine production facilities in the Middle East. Most of these facilities were gas oil separation plants (GOSPs) with typical equipment such as manifolds, separators, coalescesr, desalters, gas compressing, oil pumping, control room, electrical substations, chemical injection systems, water treatment facilities, oil storage, pipelines and so forth. The motivation for these studies came from internal company guidelines, insurance requirements and recommendations from incident investigation reports for these are other company operating facilities. There were five consequences of interest in this case study – people

(worker and public), assets, environmental, production (i.e., business interruption), and company's reputation.

Table 2 — Example Cost Categories [3] [4]	
Cost Category	**Cost Range[1]**
5	Up to US $ 10,000
4	From US $10,000 to US $100,000
3	From US $100,000 to US $ 1,000,000
2	From US $1,000,000 to US $ 10,000,000
1	More than US $10,000,000

2.1 Safety and Risk Studies

To help satisfy the company's purpose and specific objectives, the company retained safety and risk consultants to conduct a total of nine studies for each of the nine sites (i.e., a total of 81 studies):

1. Hazard and operability analysis (HAZOP) and Facility risk review (FRR)
2. Hazard identification (HAZID)
3. Quantitative risk assessment (QRA), including event/fault tree, vulnerability analysis, and consequence analyses
4. Safety integrity level (SIL) assessment
5. Hazardous area classification review (HACR) and assessment of electrical/instrumentation equipment
6. Control of substances hazardous to health (COSHH) assessment
7. Permit to work (PTW) review
8. Design review
9. Asset integrity review (AIR)

[1] In using these cost categories in oil & gas applications, the analysts typically include all applicable CAPEX/OPEX costs (engineering, procurement, manufacturing, installation, operation, maintenance, training, etc.) for a period of time (e.g., five years).

The FRR [5] was of particular interest in the cost benefit analysis. As part of the FRR, the analysis team developed a list of risk scenarios for each facility, which included incidents that can lead to the release of hazardous materials with potential for fires, explosions etc. These, in turn, can generate the consequences of interest (impacts on people, assets, environment, production, and company's reputation).

The analysis team then assessed the frequency and consequences of each scenario. Figure 1 illustrates the results of the FRR for one of the nine facilities. The matrix in Figure 1 considers the impact "production" (i.e., business interruption), and there were similar matrixes for the other four consequences of interest.

Impact on Production		Frequency				
		1 (A)	2 (B)	3 (C)	4 (D)	5 (E)
Consequence	5	6, 11, 12, 18, 21	27, 30, 37, 40, 43	15		
	4	5, 20, 46, 49	3, 17, 26, 29, 36, 39	2, 14, 32, 53, 55, 57, 59		10, 31
	3	24, 48, 52, 60, 61	42, 56	13, 25, 28, 34, 35, 38, 45	58	1, 33
	2	8, 9, 51, 62, 63		16, 19, 41	23, 44, 47	
	1			22, 50	4, 7, 54	

Figure 1 – Risk Matrix for Impact on Production

2.2 Consolidation of Recommendations

The nine safety and risk studies generated an average of 260 recommendations for each of the nine plants for a total of over 2,300 recommendations. But it was clear that there were overlaps and similarities among several of the recommendations from each of the nine distinct studies (HAZOP, HAZID, QRA, etc.). Thus, it was convenient to group or consolidate recommendations that addressed similar or related issues. The consolidation provided two benefits: (a) reduced the number of recommendations for the cost benefit analysis and (b) facilitated the work of managers by grouping similar issues for review and resolution.

For example, recommendations 1.43, 1.97, 1.113, 2.18, 2.19, 2.21, 3.3 and 4.9 addressed issues related to the fire protection system. Note that the recommendation number starts with the number of the study and finishes with the unique identifier from that study. For example, Recommendation 1.43 means the 43rd recommendation from the first study, which was a HAZOP study. Recommendation 2.18 is the 18th

recommendation from the HAZID. Because 1.43, 1.97, 1.113, 2.18, 2.19, 2.21, 3.3 and 4.9 all addressed the same issues, it was convenient to group them for the purpose of the cost benefit analysis. For illustration purposes, the combined description of this group of recommendations is as follows:

> The HAZID, HAZOP/FRR, QRA and SIL studies identified potential deficiencies in the firefighting capabilities at the facility. The specific recommendations for the fire water system, pumps, distribution, etc., include:
>
> - ✓ Performing an engineering review of the entire fire water supply and distribution system to assess the adequacy of: (1) fire water pumps regarding their efficiency (i.e., head pressure); (2) pump start-up (i.e., manual vs. automatic); (3) the capacity of fire water tank; (4) fire pump drive-redundancy (i.e., diesel/electric), (5) the deluge systems on the tanks, (6) the design criteria (especially materials of construction) of the rupture disks of the foam pourer systems on the tanks. In this review, consider whether the fire water pumps should be replaced
> - ✓ Adding emergency cooling systems (e.g., fire curtain, sprinklers) for selected equipment and providing long-range fixed monitors at critical locations
> - ✓ Developing / improving a testing program for the entire system and components, including written procedures for testing and test acceptance criteria
> - ✓ Developing / improving a maintenance program for the entire system and components, including written procedures and schedules for performing the different maintenance tasks
> - ✓ Generating a complete set of comprehensive P&IDs for the entire system and components

The consolidation reduced the number of recommendations to a little more than 900, which is about 40% of the original 2,300 or so recommendations.

3. RESULTS FOR CASE STUDY 1

Tables 3 and 4 illustrate the results from the cost benefit analysis. Both tables show the top recommendations only as there were about 100 consolidated recommendations for this facility. The first column in these tables shows the recommendation numbers. The second column shows the impact category (using the definitions from Table 1) applicable to the frequency of each scenario. The next columns show the impact category (again from Table 1) for each of the 5 consequences of interest: people (worker and public), assets, environmental, production, and company's reputation.

The appropriate cost category (using the definitions from Table 2) appears next in Tables 3 and 4. Thus, for each recommendation, it was necessary to make one assessment of impact category on the frequency of the event, five assessments of impact category for the consequences of interest, and one assessment of cost

category, resulting in 7 assessments per recommendation per risk scenario. Assuming an average of 100 recommendations and 63 risk scenarios per each of the 9 facilities, the total number of assessments was 7 x 100 x 63 x 9 = 396,900 (about 400,000). The number of assessments would have exceeded 1 million without the consolidation of the recommendations, thereby demonstrating the usefulness of the consolidation.

The last two columns in Tables 3 and 4 show the ***risk reduction*** (Δ Risk) and the ***BCI***, as defined in Section 2. In these tables, the ***risk reduction*** and ***BCI*** for a group of recommendations represent the combined impact of all recommendations in the group, not the impact of each one of them individually. The evaluation of the risk reduction and BCI was accomplished using Excel in this case study.

One final remark about these results is that the ranking by ***risk reduction*** and ***BCI*** are often different for each recommendation. For example, Recommendation 1.45 is ranked nineth by ***risk reduction*** but first by ***BCI***. The appropriate ranking for the recommendation depends on management objectives and will be discussed in more detail in the conclusion section.

4. CASE STUDY 2, INCLUDING RESULTS

This was a reliability study of a manufacturing process using fault tree analysis. Compared to Case Study 1, this was a more focused (narrower scope) and deeper (more detailed) analysis. The manufacturing process produces a consumer product that reaches millions of customers in the United States of America daily. The objective of the analysis was to evaluate the ***frequency*** that products with a specific type of defect would reach the consumers. The company had established a quantitative goal for this ***frequency*** and believed to be meeting or exceeding the goal. However, the company decided to conduct the study to further verify compliance with the goal as well as to identify opportunities for further improvements.

Figure 2 shows the results of the study regarding the potential implementation of four recommendations (1 through 4). The base case is the current situation, and it is arbitrarily set at 100% in the figure. The implementation of Recommendations 1, 2, 3, and 4 result in the reduction of the frequency of defects to 98%, 98%, 14%, and 34% of the current frequency, respectively. Recommendations 1 and 2 have small impact in reducing the frequency of defects. However, Recommendations 3 and 4 are effective.

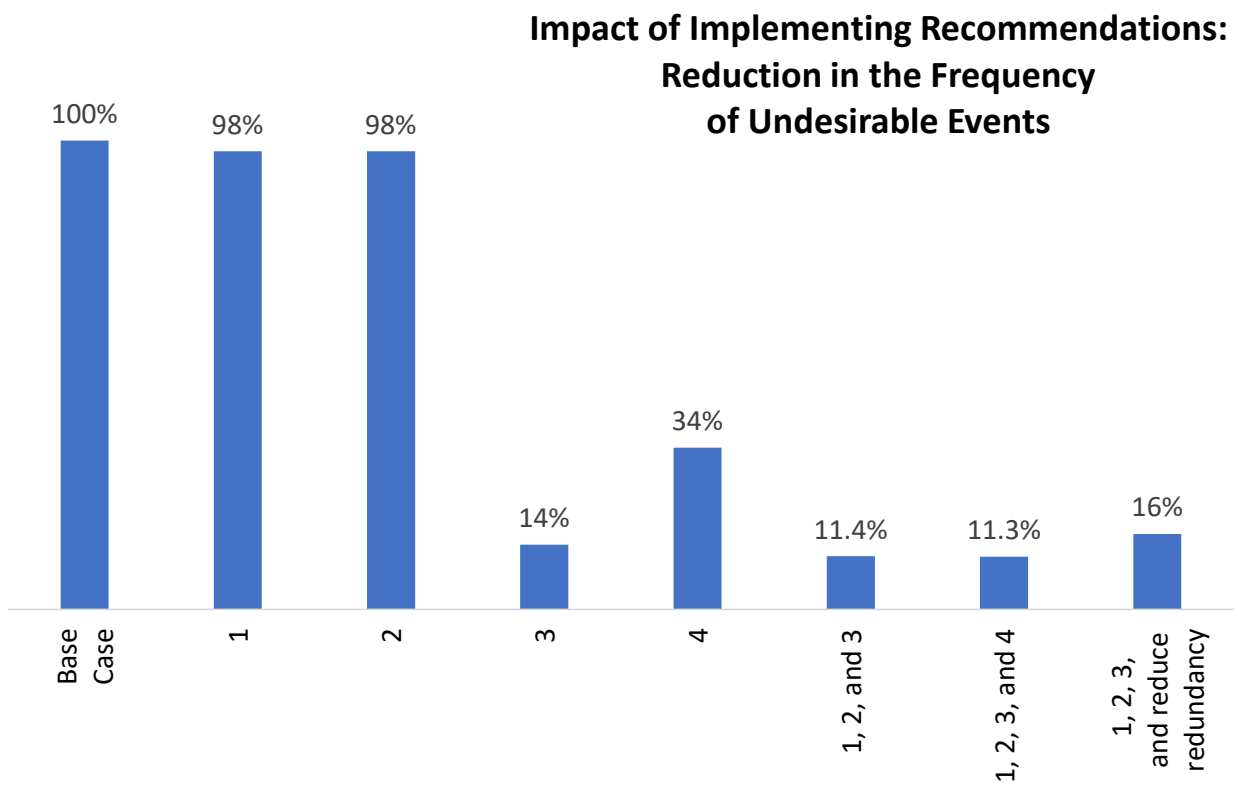

Figure 2 – Results for Case Study 2

Since there were only 4 recommendation, Case Study 2 went beyond Case Study 1 and evaluated the risk reduction for ***combinations*** of implementation of the recommendations. This would be difficult to conduct in Case Study 1 because the number of recommendations is much larger, even after the consolidation. As shown in Figure 2, implementing Recommendations 1, 2, and 3 results in a reduction to 11.4% of the current frequency of the undesirable consequence. This is a small improvement over implementing Recommendation 3 by itself, but it was worthwhile because the cost of Recommendations 1 and 2 was small. Adding Recommendation 4 to this combination brings almost not benefit (i.e., drops to 11.3% instead of 11.4%) even though Recommendation 3 by itself had a significant impact. The reason is that Recommendation 3 and 4 address the same issues in different ways. Thus, Recommendation 4 has almost no beneficial impact if Recommendation 3 is also being implemented.

One final observation regards the last bar column in Figure 2. It also consisted of implementing Recommendations 1, 2, and 3. However, there was redundant equipment in this system that was not providing meaningful protection. The combination of implementing these 3 recommendations ***and removing*** the redundancy resulted in a frequency reduction to 16% of the current value. Since this was strictly an issue of defective products (i.e., no safety considerations) and since the system already exceeded the reliability goal, this alternative was attractive because of the cost reduction. In the end, the company

achieved an overall reduction of the frequency to 16% of the current value, and the savings from the removal of redundancy more than paid for the cost of the study.

5. CONCLUSIONS

Two key results of interest are the rankings of the recommendations by the expected ***risk reduction*** (Table 3) and by the ***benefit to cost index*** (***BCI***) (Table 4). Note that the higher the ***risk reduction***, the higher the motivation to implement the recommendation because it provides greater potential to reduce the overall risk to a lower level. That is, ***risk reduction*** helps identify the most effective recommendations. In general, significant reduction in the risk at the facility can only be achieved by implementing at least a few of the recommendations ranked high by **risk reduction**.

The ***BCI*** is the ratio of the ***risk reduction*** to the cost of implementation of the recommendation. Note that the larger the ***BCI*** for a recommendation, the greater the risk reduction per unit of capital investment. That is, BCI helps identify the most efficient recommendations (i.e., most risk reduction per monetary unit). Therefore, high ***BCI*** often implies "quick wins." However, high ***BCI*** does not necessarily guarantee a significant reduction in the overall risk.

It is crucial that managers understand the definitions and meanings of the two measures provided in Tables 3 and 4 because, as illustrated in Section 4, the rankings provided by ***risk reduction*** and ***BCI*** may be different. That is, a recommendation may receive different priority depending on whether managers want to focus on efficiency or effectiveness. Furthermore, it is an iterative process, because given the risk reduction achieved by implementing the highest priority recommendations, the ***risk reduction*** and ***BCI*** will likely be different (smaller) for subsequent recommendations.

In summary, the cost benefit methodology presented here offers an approach to sort through the recommendations from safety, hazard, and risk evaluations and prioritize them effectively and efficiently. Its simplicity makes it particularly useful for applications with a large number of recommendations.

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Table 3 — Recommendations Ranked by *Risk Reduction*[2]									
Recommendation Number(s)	**Impact Category**						**Cost Category**	***Risk Reduction***	***BCI***
	Frequency	**Consequence**							
		People	**Assets**	**Environmental**	**Production**	**Reputation**			
1.43, 1.97, 1.113, 2.18, 2.19, 2.21, 3.3, 4.9	1	2	3	2	3	3	3	35%	3.5E-01
1.73	3	1	1	3	1	1	3	27%	2.7E-01
1.1a, 1.7	2	2	2	2	2	2	3	19%	1.9E-01
1.1b, 1.7, 1.54, 1.109, 1.116, 9.16	2	2	2	2	2	2	3	19%	1.9E-01
1.3, 1.22, 1.29, 1.55, 1.99, 2.8, 2.9, 9.4, 9.13	2	2	2	2	2	2	3	19%	1.9E-01
1.24, 5.3	2	2	2	2	2	2	3	19%	1.9E-01
1.13, 1.39, 1.47, 2.7, 2.23, 7.C, 9.19	2	2	2	2	2	2	3	18%	1.8E-01
1.74	1	3	3	3	3	2	3	18%	1.8E-01
1.45	4	1	1	1	1	1	4	16%	1.6E+00
1.6	2	1	1	1	1	1	4	9%	9.3E-01
1.41	3	1	1	1	1	1	4	9%	9.2E-01
1.42	3	1	1	1	1	1	4	9%	9.2E-01
1.49	3	1	1	1	1	1	4	9%	9.2E-01
1.100	2	2	2	2	2	2	3	9%	8.8E-02

[2] This table presents all recommendations with ***Risk Reduction*** equal to or greater than 5%.

Table 3 — Recommendations Ranked by *Risk Reduction*[2]									
Recommendation Number(s)	**Impact Category**						**Cost Category**	***Risk Reduction***	***BCI***
	Frequency	**Consequence**							
		People	**Assets**	**Environmental**	**Production**	**Reputation**			
1.111, 2.22, 9.7	2	2	2	2	2	2	3	9%	8.8E-02
1.115	2	2	2	2	2	2	3	9%	8.8E-02
1.46, 9.8	2	2	2	2	2	2	3	9%	8.8E-02
9.3	2	2	2	2	2	2	3	9%	8.8E-02
1.38	5	1	1	1	1	1	4	6%	6.0E-01
2.10, 5.1	1	2	2	2	2	2	3	5%	4.6E-02

Table 4 — Recommendations Ranked by BCI[3]									
Recommendation Number(s)	**Impact Category**						**Cost Category**	***Risk Reduction***	***BCI***
	Frequency	**Consequence**							
		People	**Assets**	**Environmental**	**Production**	**Reputation**			
1.45	4	1	1	1	1	1	4	16%	1.6E+00
1.6	2	1	1	1	1	1	4	9%	9.3E-01
1.41	3	1	1	1	1	1	4	9%	9.2E-01
1.42	3	1	1	1	1	1	4	9%	9.2E-01
1.49	3	1	1	1	1	1	4	9%	9.2E-01
1.38	5	1	1	1	1	1	4	6%	6.0E-01
2.1, 3.1	1	2	2	2	2	2	4	0%	4.6E-01
1.43, 1.97, 1.113, 2.18, 2.19, 2.21, 3.3, 4.9	1	2	3	2	3	3	3	35%	3.5E-01
1.37	2	1	1	1	1	1	4	3%	3.0E-01
1.48, 1.112, 2.4, 3.5	1	2	2	2	2	2	4	3%	3.0E-01
1.73	3	1	1	3	1	1	3	27%	2.7E-01
1.69	2	1	1	1	1	1	5	0%	2.3E-01
1.75	2	1	1	1	1	1	4	2%	2.0E-01
1.1a, 1.7	2	2	2	2	2	2	3	19%	1.9E-01

[3] This table presents all recommendations with ***BCI*** equal to or greater than 10% of the largest BCI in the table.

Table 4 — Recommendations Ranked by BCI[3]									
Recommendation Number(s)	**Impact Category**						**Cost Category**	***Risk Reduction***	***BCI***
	Frequency	**Consequence**							
		People	**Assets**	**Environmental**	**Production**	**Reputation**			
1.1b, 1.7, 1.54, 1.109, 1.116, 9.16	2	2	2	2	2	2	3	19%	1.9E-01
1.3, 1.22, 1.29, 1.55, 1.99, 2.8, 2.9, 9.4, 9.13	2	2	2	2	2	2	3	19%	1.9E-01
1.24, 5.3	2	2	2	2	2	2	3	19%	1.9E-01
1.40	2	1	1	1	1	1	4	2%	1.8E-01
1.13, 1.39, 1.47, 2.7, 2.23, 7.C, 9.19	2	2	2	2	2	2	3	18%	1.8E-01
1.26, 4.3	2	1	1	1	1	1	4	2%	1.8E-01
1.27	2	1	1	1	1	1	4	2%	1.8E-01
1.28	2	1	1	1	1	1	4	2%	1.8E-01
1.74	1	3	3	3	3	2	3	18%	1.8E-01
1.71	3	1	1	1	1	1	4	2%	1.7E-01

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Does your facility have the flu? Use Bayes rule to treat the problem instead of the symptom

Keith Brumbaugh
aeSolutions
Millennium Tower, 10375 Richmond Ave, #800, Houston, TX 77042
Keith.brumbaugh@aesolns.com; keithbru@gmail.com

Introduction

Is our industry addressing the problems facing it today? We idealize infinitesimally small event rates for highly catastrophic hazards, yet are we any safer? Have we solved the world's problems? Layers of protection analysis (LOPA) drives hazardous event rates to 10^{-4} per year or less, yet industry is still experiencing several disastrous events per year.
If one estimates 3,000 operating units worldwide and industry experiences approximately 3 major incidents per year, the true industry accident rate is a staggering 3 / 3,000 per year (i.e. 10^{-3}). All the while our LOPA calculations are assuring us we have achieved an event rate of 10^{-6}. Something is not adding up! Rather than fussing over an unobtainable numbers game; wouldn't it be wiser to address protection layers which are operating below requirements? We are (hopefully) performing audits and assessments on our protection layers and generating findings. Why are we not focusing our efforts on the results of these findings? Instead we demand more bandages (protect layers) for amputated limbs (LOPA scenarios) instead of upgrading those bandages to tourniquets. Perhaps the dilemma is we cannot effectively prioritize our corrective actions based on findings. Likely we have too much information and the real problems are lost in the chaos. What if there was a way to decipher the information overload and visualize the impact of our short comings? Enter Bayes rule to provide a means to visualize findings through a protection layer health meter approach; to prioritize action items and staunch the bleeding.

Keywords:

Bayes, Bayes rule, Bayes theory, LOPA, IPL, SIS, SIF, SIL Calculations, systematic failure, human factors, human reliability, operations, maintenance, IEC 61511, ANSI/ISA 61511, hardware reliability, proven in use, confidence interval, credible range, safety lifecycle, functional safety assessment, FSA stage 4, health meter.

The State of Our Industry

The objectives of this paper are to look at some issues with the contemporary safety lifecycle industry and provide solutions. A major trend across industry has been an overall decrease of the tolerable catastrophic event likelihood (i.e. multiple fatalities) down to one such event every 100,000, or even 1,000,000 years[1]. This lowering of the target has the good intention of making a facility safer. After all, superb targets will make superb facilities, should it not? The downside of extravagant targets is they are harder to achieve. Realistically these targets are impossible to achieve once one considers the real uncertainties of physical systems (systematic error). Note that it is hard-enough meeting the targets already set, how will making the targets even smaller help matters?

Smaller tolerable risk targets will have the result of producing more Independent Protection Layers (IPL) with greater integrity requirements. This leads to a "Forest-for-the-tree syndrome" where a plant is trying to manage more IPLs than it can handle, missing the bigger picture of plant health and safety (e.g. key performance indicators). If every IPL has its own multifaceted maintenance and management requirements, how can a facility ever manage all responsibilities effectively?

Figure 1- Can't see the Forest for the Trees

The solution in this paper is simple, a facility should focus on managing what it is capable of managing. Minuscule targets have good intentions of making everyone safer by providing more protection but throw enough IPLs at a problem and one will soon reach the tipping point, the

[1] One event every 100,000 or 1,000,000 years is a target of 10^{-5} and 10^{-6} respectively.

straw that broke the camel's back. When everything is a problem due to much information nothing will be managed effectively. Furthermore, when there are fewer IPLs to distract a plant management team, the team is free to focus on the real problems. If one can identify problems, then effective management will occur. One of the best ways to identify problems is with a periodic health check of the IPLs. This paper will present one such health check called the "Bayes truth meter." The concept with the Bayes truth meter is to strip away all guess work and generic data, instead showing true IPL health reflecting a facility's own systematic biases.

To back up the claim that targets of 10^{-5} and 10^{-6} are unobtainable, consider figure 2 which is a list of current investigations from the Chemical Safety Board (CSB) over the course of the last year (circa August 2020).

Current Investigations

1/29/20… fatally injured three contractors…

1/24/20…fatally injured two workers …

11/27/19…An explosion and fire ….

10/26/19… death of one worker [&] member of the public...

From <https://www.csb.gov/investigations/>

Figure 2- Chemical Safety Board list of open incident investigations for catastrophic events

There were three major multiple fatality accidents (and one near miss) over the course of one year. This sort of catastrophe would likely earn the maximum hazard mitigation target from any client following OSHA. If targets of 10^{-5} or 10^{-6} were achievable, at least 300,000 to 3,000,000 operating facilities would be needed to average out the three major accidents from year. Unfortunately, the current estimated number of petrochemical facilities in the United States is around 2,300 (keep in mind the CSB only covers US operations).

If the number of facilities is generously rounded up to 3,000 and divided by the three catastrophic accidents, the average industry catastrophic event rate is 10^{-3}, far short of the 10^{-5} or 10^{-6} targets! To put this into a visual form consider figure 3, pretend the 100 boxes represent the industry's 10^{-5} mitigation target.

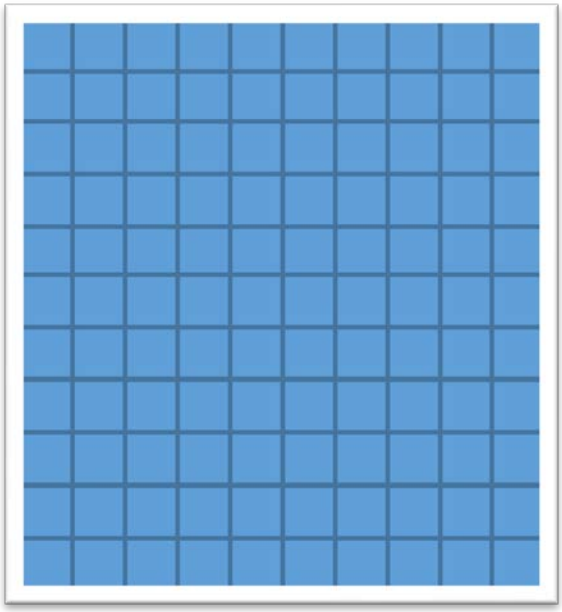

Figure 3- Industry target, 100 boxes represent a hypothetical 10^{-5} (10^{-6} would be 1,000 boxes!)

Figure 4 represents how far off the industry is from the mitigation target (10^{-3}):

Figure 4 - Reality, 1 box represents where the industry is, 10^{-3}! Compare to 10^{-5} above

As the graphics show the industry accident rate is off target by at least 2 orders of magnitude (3 orders if 10^{-6} is considered). This should be a wakeup call that the industry is missing something major. Something interesting to note is a 10^{-3} catastrophic event rate is the same as the lower bound of Human Error Probability (i.e. HEP). HEP is systematic error and is not typically considered in any calculations outside of Human Reliability Analysis (HRA)[2]. Maybe this is the factor the industry is are missing?

The Issues with our Current Approach

Returning to the ideas touched on previously, of increasing risk reduction targets to make everything "safer." There are good intentions behind ultra-low targets, however meeting these obscene targets would require throwing every feasible protection layer that can be mustered at the problem, hoping that something "sticks" (i.e. is effective). This leads to safety barrier overload. The idea is when there are too many things to manage, and everything is important, how can the signal be separated from the noise. If everything is important, what is *drastically*

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[2] For more information on HRA and Human Factors consider the following two papers:
Conducting a Human Reliability Assessment to support PHA and LOPA, Dave Grattan - https://www.aesolns.com/post/conducting-a-human-reliability-assessment-to-support-pha-and-lopa

Can we achieve Safety Integrity Level 3 (SIL 3) without analyzing Human Factors?, Keith Brumbaugh and Dave Grattan - https://www.aesolns.com/post/can-we-achieve-safety-integrity-level-3-sil-3-without-analyzing-human-factors

important? What is behaving fine and doesn't need attention? What is behaving fine right now but might be a problem a few years down the line? What has no chance of working when demanded, putting the facility at risk right now, and is going to drive the system over the edge of a cliff?
When there are too many protection layers to manage due to astronomical LOPA targets, no one will know which warning signs are important and which are just a nuisance. Protection layer overload leads to a "forest-for-the-trees-syndrome." As an example, the most sophisticated protection layers designed by top dollar consulting companies will be all for naught once they fall out of maintenance. How good is the gold plated protection layer when the valve has polymerized stuck due to never being tested? What might lead to maintenance oversights? Perhaps there were too many protection layers with not enough manpower to manage them.

Another problem with astronomical protection layer targets is Safety Instrumented Functions (SIFs) will need to be applied with high safety integrity calculations to meet a target. These calculations will "prove" a protection layer is good enough to close a 1 in 10,000 year gap, yet is that number real? Theoretically a Safety Integrity Level (SIL) calculation is correct if we consider hardware failures alone and the system operates in a vacuum, yet as soon as a human touches the system good luck maintaining that integrity level without highly sophisticated management practices. And the problem only gets worse as more high integrity protection layers are added to the facility.

All of this naysaying may seem to be blasphemous coming from safety system engineer, implying lofty risk reduction targets cannot be met, but has reality been considered? As previously mentioned, the industry catastrophe rate is sitting around the lower bound of Human Error probability (10^{-3}). This is key to understanding what has not been addressed in traditional LOPA math and SIL calculations. It is the authors' opinion that some very major degradation factors are being missed when modelling protection layer integrity. The elephant in the room is systematic error.

This is all not to say that the industry is in a bad state. There are a lot of good people out there doing important work, trying to make everyone safer. Their contributions should not be discounted as they are all based on lessons learned in blood. Yet it feels like the industry has gotten as far as it can with its current practices, floundering between moving forward or backwards depending who is asked. The next step forward in process safety needs to be in the best direction possible.

Do it Better with Bayes

The Bayesian approach allows matching optimistic rare event assumptions and IPLs with real-world observations, turning fantasy into reality. This approach allows one to base plant health metrics on observed evidence. Otherwise the industry is stuck with using generic data which is not specific to the facility's own systematic biases. If typical SIL calculation modelling data is based on industry averages, figure 4 shows how good industry average is.

A Bayes approach will likely show a facility isn't as good as it hoped it was. The problem is inductive reasoning has been used to predict catastrophic rare events. This is like the black swan

theory, historically the world used to think all swans were white since a black swan had never been observed in nature, but then low and behold, they were discovered eventually. Bayes rule would have allowed factoring in the possibility of a black swan occurring. A black swan would always be in the realm of possibilities, as more evidence was gathered such as feathers, third party sighting reports, occurrences in other similar species, etc; the model could have been updated to better predict where reality laid. The traditional model would have said "There has been millions to billions of sightings of white swans, no black swan has ever been seen in nature, therefore there is no black swan."

Back to the process safety industry, one can make a similar comparison between the current industry and a Bayesian approach. The current industry approach is based on frequentist-based statistics. This approach requires enormous amounts of data in order to derive a conclusion, such as the millions to billions of white swans and no black swans. If the analogy is given little thought, all that is known with certainty is there is a good chance of operating in a safe state, but there is no idea of the dangerous state; how bad is it, how it would unfold, and how likely it is. The only way to know the answer to the dangerous state questions is to collect data (which is likely a trick that can only be performed once). A frequentist approach requires enormous amounts of data to definitively state a comparative frequency. It should be obvious that a facility will not have, nor will it ever want enormous amounts of data for a rare catastrophic event. Contrast the frequentist approach with the Bayesian approach. The Bayesian approach allows for the input of subjective data in a logical manner. This method allows *all* relevant evidence to be factored into the model. To again use the black swan analogy; things like feathers, third person accounts, and occurrences in similar species can be directly related to near misses, audit results, and similar process accidents. Bayes allows one to factor in systematic biases and errors. Frequentist methods cannot do this.

Bayes rule can answer the question, "is a catastrophic event rate of 10^{-6} obtainable." Most likely if Bayes rule was embraced, the results would show that the industry is aiming for something that is unachievable, "biting off more than one can chew."

When Bayes shows that 10^{-6} can't be met, a facility will need to step back and ask, "what are we really trying to do here." The answer that makes the most sense is the facility is trying to make the most money while not "going boom." Since the facility's resources (time and money) are limited, then the "not going boom" part needs to be focused on the systems that need the most help while not spending all the surplus resources. Bayesian methods can provide an outlook on how each individual protection layer is behaving. Advanced warnings can be given based on evidence, staunching the bleeding of a bad acting barrier.

How to Apply Bayes Rule to Process Safety?

Implementing Bayes in process safety can be as simple or as difficult as one cares to make it. The author's previous paper[3] went over a simple approach to implement Bayesian Methods into the management of Safety Instrumented Functions. It is suggested to review the referenced paper

[3] *What is Truth – Do our SIL calculations reflect reality*, Keith Brumbaugh 2019, https://www.aesolns.com/post/what-is-truth

for further details as well as a rough “how to” example. The approach does not need to be limited to SIFs; all protection layers are ripe for a Bayesian management approach.

Begin a Bayes model with any protection layer, with any theoretical achieved Probability of Failure on Demand (PFD). Convert the achieved PFD point value, such as 0.01, into a probability distribution (Poisson or example). If the distribution is known this would be preferable, but often times the distribution is not known.

The probability distribution should represent all of the possible PFD the protection layer could hold. The boundary of the distribution should contain all realistic values which the protection layer could ever be. For example, it might be expected a SIF can operate somewhere between a PFD 0.1 to 0.0001. The probability distribution also assigns the likelihood that the protection layer *is* any of the particular PFD values.

The initial probability distribution is known as the prior. The Bayesian prior distribution is then updated over time with new evidence to form a posterior. Below is a conceptualized representation of a SIF which has undergone a Bayesian conversion and updating process.

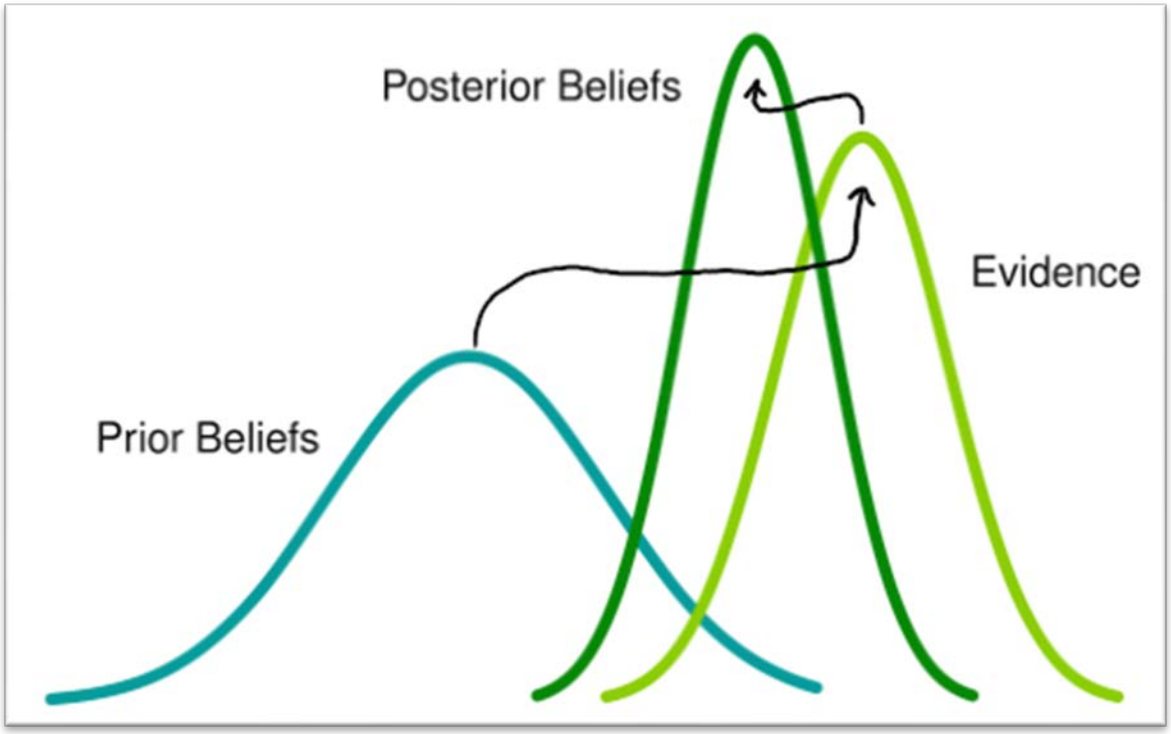

Figure 5- Example of probability distribution with prior, evidence, and posterior

There are two types of evidential data that can be used, the first is quantitative data. Quantitative data can be absolutely proven as true or false (a Bernoulli trial). The protection layer is subject to a test and the result is recorded. Items that fall under this category are proof test results and actual demands on the system (planned or unplanned).

The other type of evidential data is qualitative. Qualitative data is based upon expert opinion. The application of qualitative data may seem subjective, but the precedent of using qualitative data has already been set for process safety. Today it is accepted industry practice to use subjective judgements in LOPA, don't forget that LOPA drives everything (most of the time)! If qualitative judgements are documented and justified, the industry has no problem with them. A qualitative Bayesian update can fall under the same scrutiny. So long as the application of a qualitative Bayesian update is made to be repeatable and predicable there should be no issues. This can be achieved with a repeatable checklist from a common assessment task. Data which fits the qualitative bill are audits, Functional Safety Assessments (FSA), and Human Reliability Analysis. All of this data is aimed at discovering systematic errors by a repeatable and well-established practices.

The probability distribution from figure 5 can also be represented as a cumulative distribution. The data source of a cumulative distribution is the exact same as a probability distribution, the difference is the likelihoods are summed from 0% to 100%.

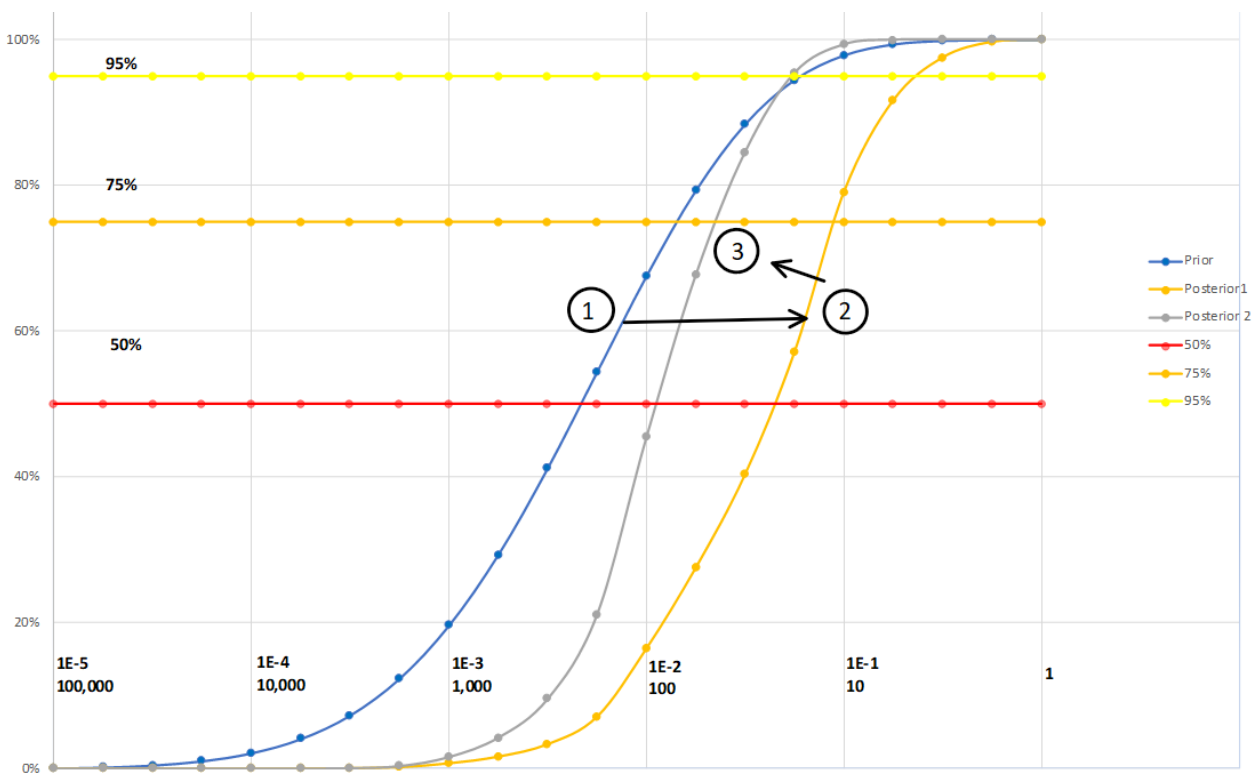

Figure 6- A cumulative distribution of a Prior, and 2 subseqeunt posterior updates

Figure 6 is a cumulative distribution of the probability distribution from figure 5. In figure 6, the X-axis represents an IPLs probability of failure on demand. The X-axis is unitless, representing either PFD on the top, or Risk Reduction Factor (RRF) on the bottom (inverse of PFD). Each vertical line represents one order of magnitude with 4 gradients within (logarithmic PFD). For example, looking at the far-right side of the graph, the four different gradients from right to left

represent 2.5, 5, 7.5, up to 10 RRF. The important take away is the further to the left the point of interest, the better (i.e. smaller PFD).

The Y-axis represents the cumulative likelihood from 0% to 100%. The 3 colored lines at 95%, 75%, and 50% are credibility levels. The point where these credibility level lines intersect the cumulative distribution curve represents the upper credible interval, which says an IPL is a particular PFD value or better. Note that these credibility levels are arbitrary, however they align with examples from the IEC 61511 standard.[4] As an example, the curve labelled "1" intersects the 95% credibility limit around 25 RRF. With this intersection, a statement can be made that it is 95% credible that the IPL represented by distribution "1" is 25 RRF or better.

It might be apparent that as better RRF numbers are targeted, the credibility decreases. For example, on the same "1" curve there is only a 50% credibility the IPL is 125 RRF or better. This is the key concept when trying to determine how "good" an IPL is in a Bayesian system, enabling one to state the credibility an IPL is meeting a certain performance target.

With the basic concepts of credibility and probability addressed, this example can move onto the concept of a Bayesian update. In figure 7, the curves "1," "2," and "3" represent a theoretical Safety Instrumented Function. The system starts with a SIL calculation result converted to a distribution as seen in curve "1." The SIL calculation using traditional methods returned a PFD of 1.3×10^{-2} (77 RRF), this achieved RRF value seeded a Poisson distribution to intersect 77 RRF at the 75% credibility level.

Next, pretend there is a failure during the first proof test. This is a simple Bayesian update with quantitative data represented in curve "2", 1 test, 1 failure. Curve "2" has shifted to the right, a worse result. Now the 75% upper credibility limit is around 12 RRF. Once this warning sign is discovered from the Bayesian update, pretend the management team initiates a root cause analysis, identifies the problem, fixes the problem, then performs a Management of Change Functional Safety Assessment with a positive assessment result. This positive assessment is a qualitative update and is applied as curve "3." Observe curve "3" has shifted back to the left, a better result. As the example ends the system is better than after the failed test of "2," but not as good as it originally started in curve "1."

Making Bayes more Intuitive

The graphs in the previous section were full of details but unfortunately, they are not intuitively obvious. In order to make the concept of a Bayesian update more intuitive it would be beneficial to simplify the information into what is important. After all, management of IPLs, and addressing bad actors is the most valuable application of Bayes. This paper introduces a concept called the "Bayes Truth Meter." Imagine that a corporate criterion will accept an IPL upper credibility limit of at least with 50% credibility, preferably 75%, and over achievement at 95% (see IEC 61511-2018, Part 2, Figure A.7 as basis for the levels). Stripping away all of the distribution "mumbo jumbo," the IPL example of the Prior from Figure 6 (i.e. curve "1") is shown in Figure 7 converted to the Bayes Truth Meter.

[4] IEC 61511-2018 - Part 2, Figure A.7 – Typical probabilistic distribution target results […]

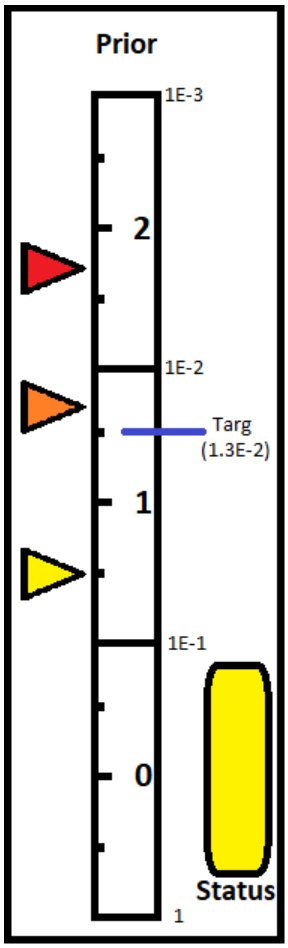

Figure 7- Bayes Truth Meter. Prior "1" from Figure 6.

To quickly describe the meter, the red, orange, and yellow pointers show where the 50%, 75%, and 95% upper credibility limits cross the cumulative distribution curve.

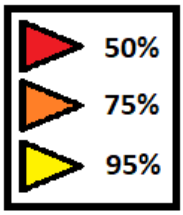

Figure 8 - Credibility markers (key)

The blue bar shows the PFD target (recall the previous example was a SIF with a target of 1.3×10^{-2}, i.e. 77 RRF).

There is a “Status” button in the lower right corner to quickly tell how an IPL is operating in relation to its target. Red is bad, Orange is Ok, Yellow is good, and Green is Great.

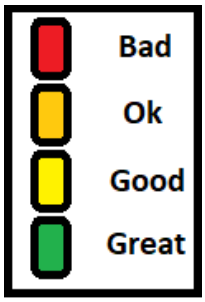

Figure 9 - Status lights (key)

The status light changes based on where an IPLs target (the blue line) lies in relation to its upper credibility levels. For the meter in Figure 7, the 75% credibility marker is better than the target, so Status light is yellow (good).

Following the previous example, the system encounters its first Bayesian update, a failed test.

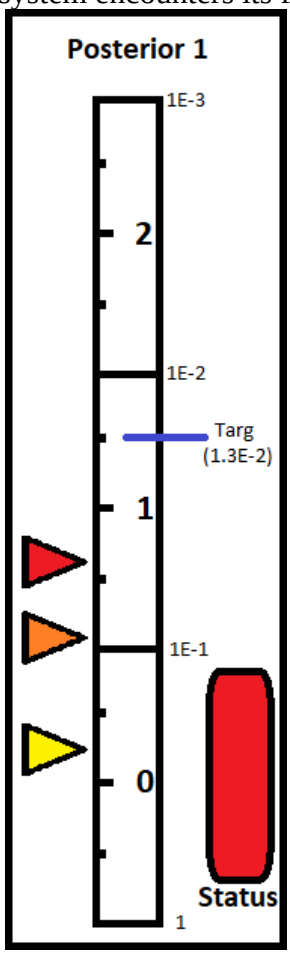

Figure 10 - Bayes Truth Meter. Posterior 1 (i.e. curve "2" from Figure 6).

Figure 10 depicts a failure during the first proof test. The meter shows the target is worse than even the 50% credibility marker. It is not depicted in the meter, but the target is only 20% credible! This poor result has put the status light in the “Bad” zone. Management would know

that it is time to focus attention on this SIF before a problem develops, and at least try to recover to the “OK” status.

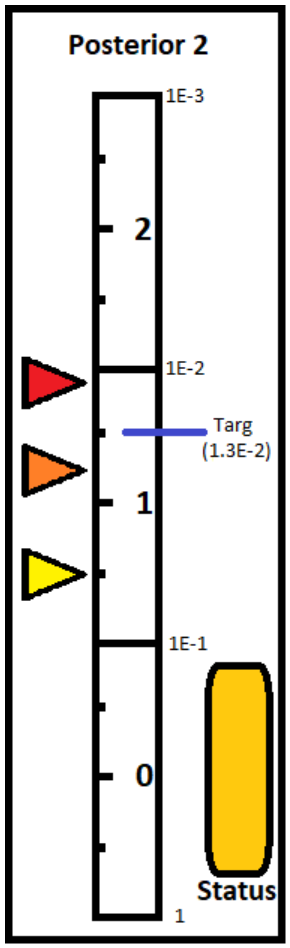

Figure 11 - Bayes Truth Meter. Posterior 2 (i.e. curve "3" from Figure 6).

The final update shown in figure 11 represents the root cause analysis was performed to determine the source of the failure, the cause was fixed, and then an MOC functional safety assessment was run with favorable results. This subjective judgement updates the meter from Figure 10 and shows the system has recovered to the “OK” Status. It is not back to “Good,” but there are likely more important issues demanding attention at this point.

Does Bayes Prove we are Aiming too High?

Returning to a previous point made in this paper, once a Bayes engine has been implemented, the difficulty in achieving the lofty targets set by current industry practice will become apparent (10^{-5} and 10^{-6}). To prove the point, consider the same trial run previously in figures 7 through 11, but with a SIL 2 SIF instead of a SIL 1.

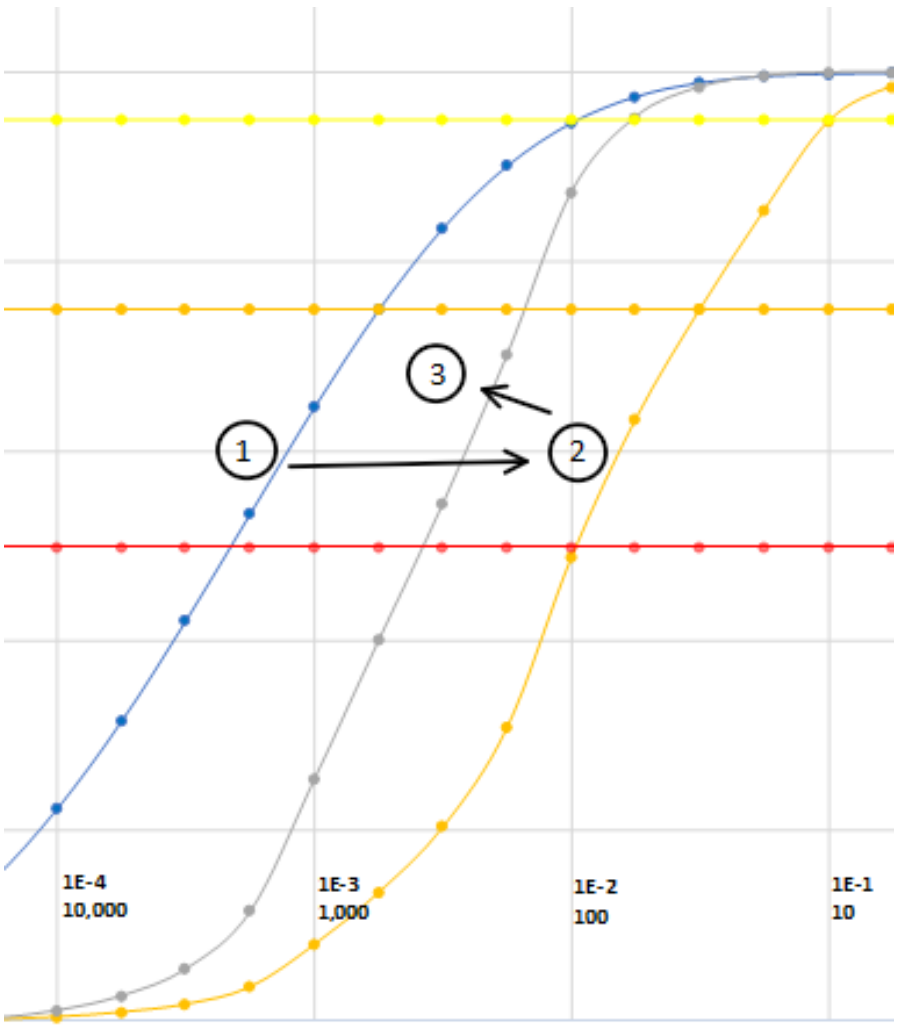

Figure 12 - SIL 2 cumulative distribution with 1 prior and 2 posterior updates

The example in figure 11 is a SIL 2 SIF with near identical parameters as the previous SIL 1 SIF an order of magnitude greater. This example has seeded the prior target of 1.3×10^{-3} at the 75% upper credibility limit. The poor results are shown in this table, but the Bayes truth meter makes it easier to understand.

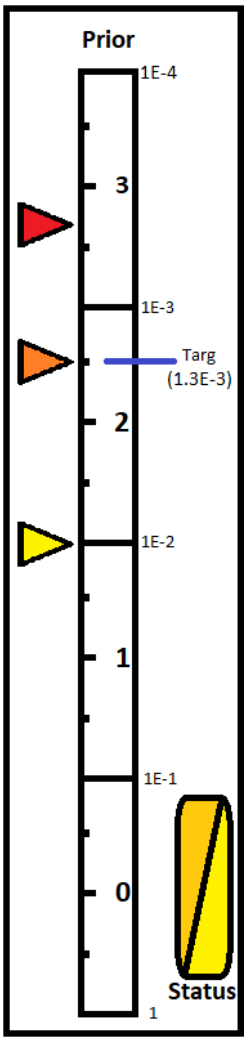

Figure 13 - Bayes Truth Meter. SIL 2 Prior "1" from Figure 12

Figure 13 shows the prior from figure 12 (plot 1) converted to the Bayes truth meter. The status light indicates the system is on the line between “Good” and “Ok” (i.e. the target PFD is at the 75% credibility marker). Next the system is subjected to the same one test, one failure.

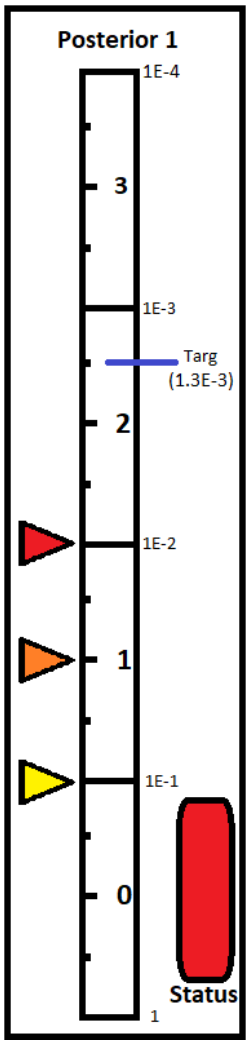

Figure 14 - Bayes Truth Meter. SIL 2 Posterior 1 (i.e. curve "2" from Figure 12).

It can be seen just as in in the previous SIL 1 example (figure 10), the system has dropped significantly. Notice however, this drop is more drastic. Like last time the target has a very low credibility, in fact it is only 13% credible that the SIF is operating at the target. Next a similar root cause analysis with a recovery factor is applied to the SIF.

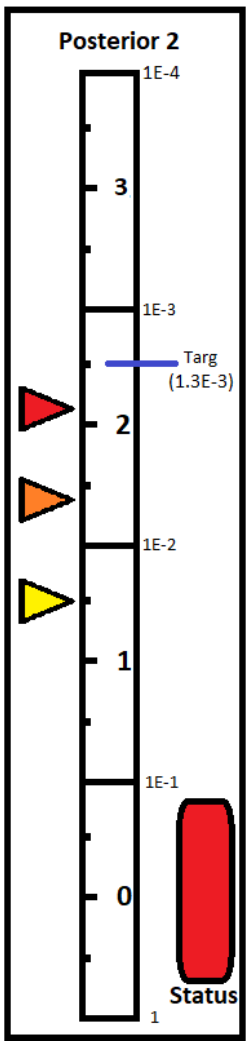

Figure 15- Bayes Truth Meter. SIL 2 Posterior 2 (i.e. curve "3" from Figure 12).

Unfortunately, even after a recovery factor has been applied the SIF is still in the “Bad” zone. It may seem like not enough recovery was applied, however the recovery factor applied was an entire order of magnitude greater than the SIL 1 example (due to the SIL target was an order of magnitude greater). In fact the recovery may have been granted too much weight (recovery factors are also subjective).

This might seem unfair but to put everything into context, consider the target again. The target is around one failure every 1000 years, however in just 5 years there was a failure. How much evidence would it take to convince someone that the function has been fixed and is operating back at the 1 in 1000 level?

Consider hurricane Harvey. That hurricane was a 1 in 1000 level event, yet Houston, Texas experienced this major hurricane a few years back. The city of Houston has implemented new safety measures to help combat any future flooding events, but would anyone living in Houston today claim that there will never be another hurricane Harvey in their lifetime? The answer is

most likely no, and every hurricane season for the next 20+ years the entire city will be on full alert (until the next generation comes along, thinking they know better than their elders).

Conclusion

It is the authors' belief that if the industry started to implement Bayes into its models it will quickly be demonstrated that lofty 10^{-5} and 10^{-6} event frequency targets can never be met. As witnessed in the SIL 2 SIF example, just one failure at any time during a facility's operating history will quickly shatter the illusion that a SIF can reach SIL 2 (not to mention SIL 3). Imagine that a target of 10^{-6} would require at least three IPLs of this same magnitude to mitigate the target. Good luck!

When a facility pretends it can meet 10^{-6} it is ignoring the elephant in the room, systematic errors and common cause failures. These failures are real, but their impacts are largely unknown. A Bayesian model can prove that they are worse than the industry gives credit.

If the industry were to acknowledge that 10^{-6} isn't possible, then what is possible? Back on figure 4 it was shown the industry is operating around a 10^{-3} catastrophic event rate on average, but that doesn't seem like a good target. To compare the process safety industry to the airline industry, the approximate probability of dying in an airplane crash is also 10^{-5} (see footnote [5]). Keep in mind the airline industry has much simpler systems designed for one purpose only, yet still has multiple fatality accidents. It might be best to split the difference between where the process safety industry is, and where the airline industry is. This admits the difficulties due to the complexity of process safety systems, realizing that process safety can never be as simple as an airplane's safety system.

If the industry were able to accept a lower target it would be much easier to close a LOPA gap. Lower targets would equate to few IPLs to manage. Keep in mind also that these IPLs would show a realistic number based on the Bayesian model, updated with real evidence. Less IPLs would lead to more effective management of IPLs and greater ease on maintenance. The Bayes Truth Meter approach allows a plant management to focus on bad actors. Finally, with less "trees" (IPLs) to manage, a facility is free to focus on the "forest" as a whole (overall plant health).

[5] In the year 2019 there were 10 major airline crashes with multiple fatalities. There are approximately 18 million flights per year on average. The odds of dying in a plane crash is around 10^{-5} to 10^{-6}. https://www.1001crash.com/. Airline systems are very sophisticated, and have one goal in mind, technologies are mostly the same, and failure modes are well understood. Compare to the process safety industry.

Figure 16 - Less trees = easier to manage the forest

MARY KAY O'CONNOR PROCESS SAFETY CENTER
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Integrating the PHA and Facility Siting into a Site Risk Assessment Life-Cycle

Sam Aigen and Colin Armstrong
AcuTech
SAigen@acutech-consulting.com
CArmstrong@acutech-consulting.com

Abstract

The PHA process has been implemented in industry for decades, and PHA stakeholders are already fluent in risk communication. PHAs provide an accepted framework in organizations which details scenarios to be evaluated, credible safeguards, and the organization's acceptable risk criteria. Siting studies may consider risk in the same way as PHAs, but organizations typically fail to align the two assessments.

PHAs already contain the hazard scenarios and safeguards and the organization's risk criteria. Aligning the PHA scenarios and safeguards with the siting study can improve the quality of the siting study; generic release scenarios are generally included in a siting study but could be improved by process-specific hazard scenarios from the PHA. PHA recommendations can create an unnecessary cost to the organization if the consequences and risk ranking is not accurate. Conversely, PHA scenarios that fail to identify major risk potential may result in increased risk exposure for personnel and the business. Aligning the qualitative risk criteria from the PHA and the siting study quantitative risk criteria can allow PHA scenarios consequences and level of risk to be accurately identified, result in cost-effective and more accurate risk reduction recommendations, and improve the organization's ability for consistent risk-based decision making.

MARY KAY O'CONNOR PROCESS SAFETY CENTER
TEXAS A&M ENGINEERING EXPERIMENT STATION

23rd Annual Process Safety International Symposium
October 20-21, 2020 | College Station, Texas

A framework for automatic SIS verification in Process Industries using digital twin

Nitin Roy
California State University
Sacramento, CA
nitinroy@icloud.com

Abstract

Increasing complexity of distributed control systems (DCS) and control logics has made (safety instrumented systems) SIS validation complex and time-consuming. IEC and ISA safety standards recommend comprehensive logic checks of Safety instrumented functions. It can take months to check logic in delivered product. This work introduces automated testing of logic in process plants using Digital Twins. This method makes the process efficient and saves considerable amount of time, manpower and in turn capital. The verification which takes months can be reduced to weeks. It also ensures the verification is comprehensive and accurate making the system safer. In this work we also review the current practices in SIS verification and future improvements.

Keywords: automation and control, functional safety, process safety, process simulation

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The use of Bayesian Networks in Functional Safety

Paul Gruhn, P.E., CFSE
aeSolutions
Houston, Texas, USA
paul.gruhn@aesolns.com

Abstract

Functional safety engineers follow the ISA/IEC 61511 standard and perform calculations based on random hardware failures. These result in very low failure probabilities, which are then combined with similarly low failure probabilities for other safety layers, to show that the overall probability of an accident is extremely low (e.g., 1E-5/yr). Unfortunately, such numbers are based on frequentist assumptions and cannot be proven. Looking at actual accidents caused by control and safety system failures shows that accidents are *not* caused by random hardware failures. Accidents are typically the result of steady and slow normalization of deviation (a.k.a. drift). It's up to management to control these factors. However, Bayes' theorem can be used to update our prior belief (the initial calculated failure probability) based on observing other evidence (e.g., the effectiveness of the facility's process safety management process). The results can be dramatic.

Keywords: PSM, Process Safety Management, Bayes' Theorem, SIS, Safety Instrumented System, SIL, Safety Integrity Level, Swiss Cheese Model, Normalization of Deviation, Drift

1. Introduction

Some statistics are easy. For example, what's the probability of a fair 6-sided die rolling a 3? That shouldn't challenge anyone. The answer is based on frequentist principles and can be proven by testing or sampling.

Some seemingly simple statistical examples aren't as simple as they might first appear. For example, imagine there is a one in a thousand chance of having a particular heart disease. There is a test to detect this disease. The test is 100% accurate for people who have the disease, and 95% for those that don't. This means that 5% of people who don't have the disease will be incorrectly diagnosed as having it. If a randomly selected person tests positive, what's the probability that the person actually has the disease?

Some statistical cases are not simple at all. For example, what's the probability of *your* plant having a catastrophic process safety accident within the next year? You and others might have designed and calculated it to be as safe as driving a car (i.e., 1/10,000 per year), but how can you *prove* it? Frequentist based statistics cannot be used to confirm or justify very rare events. Do you believe your plant is safer (or worse) than any another facility you may have visited? Might there be variables, conditions, or precursors that you could *observe* that might *affect* your belief? And if so, might you be able to evaluate and *quantify* their impact on risk?

The answer is 'yes'.

(The answer for the heart disease example above is 2%. See the annex at the end of this paper for the solution if you didn't get the correct result.)

2. Bayes basics – updating prior beliefs

Past performance is *not* an indicator of future performance, especially for rare events. Past performance would *not* have indicated (at least not to those involved at the time) what would happen at Bhopal, Texas City, or any other accident you can think of. How many managers have you heard say, "We've been running this way for 15 years without an accident; we *are* safe!"

What's the probability of dying in a vehicle accident? In the US there are about 35,000 traffic deaths every year. Considering our population, that works out to a probability of about 1/10,000 per year. You're obviously not going to live to be 10,000 years old, so the probability of your dying in a car crash is relatively low. Yet might there be factors that *influence* this number, ones that you might be able to observe and *control*?

Imagine the following: A salesman you know — but have never met — picks you up at your office and drives you both out for lunch. What probability would you assign to being in a fatal accident? On the way to the restaurant you notice him texting while driving, speeding, and being a bit reckless. You're a bit distressed, but you know you don't have far to go, and you keep your mouth shut. At your one-hour lunch you see him consume three alcoholic beverages. Assuming you'd even be willing to get back in the car at that point (there's always Uber), what probability would you assign to being in a fatal accident *now*? (Records have shown that alcohol is involved in 40% of traffic fatalities, speeding 30%, and reckless driving 33%. You are 23 times more likely to crash while texting. Seatbelts reduce the risk of death by 45%.) This is an example of updating a prior belief based on new (even subjective) information. That's Bayes' theorem.

So one *can* observe conditions and make even subjective updates to previous predictions. People do this all the time. Even insurance companies do this when setting premiums (as premiums are not simply based on past performance).

3. A real example: Bhopal

Bhopal was the worst industrial disaster of all time. The facility was designed and build in the 1970's and the accident took place in 1984. While this was a decade before layer of protection analysis (LOPA) was introduced, it's useful to use this technique to evaluate the original design

and compare it to the operation on the day of the event. This is *not* an attempt to explain *why* the event happened, nor should this be considered an example of 20/20 hindsight. This is simply an attempt to show how Bayes' theorem might be used in the process industry.

The facility in Bhopal was patterned after a successful and safe plant in the US. There were inherently safe design principles and multiple independent protection layers to prevent the escalation of an event caused by the possible introduction of water into a storage tank. These are listed in Table 1, along with *sample* probabilities for their failure.

Description	Probability of failure
Stainless steel construction	.01
Nitrogen purge	.1
Refrigeration system	.1
High temperature alarm	.1
Empty reserve tank	.1
Diluting agent	.1
Vent gas scrubber and flare	.1
Rupture disk and relief valve	.1
All safety layers failing at the same time	**1E-9**

Table 1: The possible performance of safety layers at Bhopal

Considering an initiating event frequency of perhaps 0.1/yr (a common number used in LOPA for many initiating events), the risk associated with this event would appear to be much lower than the risk of driving a car. Yet how could this be *proven*? In reality *none* of the layers were effective at Bhopal and the accident happened within the first five years of operation (i.e., within the assumed time period of practically any single initiating event). All the layers at Bhopal didn't magically fail at the same time. Trevor Kletz was well known for saying, "All accidents are due to bad management." Ineffective management allowed all the layers to degrade (and there were common causes between many of them) to the point where *none* of them were available the day the event happened. Normalization of deviation — or drift — was not unique to Bhopal. This is a serious issue that affects *many* facilities even today. *How might we be able to model this?*

4. Functional Safety Engineers and math

Functional safety engineers focus on the ISA/IEC 61511 standard. Following the lifecycle of the standard involves determining a performance requirement for each safety instrumented function (SIF) and evaluating that the intended hardware design meets the performance requirements (and changing the design if it doesn't). This entails performing calculations considering the device configuration, failure rate, failure mode, diagnostic coverage, proof test interval and much more. Yet the calculations only involve *random* hardware failures, and the numbers are often so low that they cannot be proven by frequentist statistics and sampling. The standard does discuss systematic

failures (e.g., human errors of specification, design, operation, etc.), but *not* in a quantitative manner.

What *really* causes accidents involving control and safety systems? Figure 1 is well-known to all functional safety practitioners. (The results were published by the United Kingdom Health and Safety Executive more than 20 years ago, and it's unlikely that any of the values have changed since.) Few, if any, accidents have been due to a random hardware failure, yet that's what everyone is focusing on in their calculations. How might we include the *management* related issues shown in Figure 1 in our overall modelling? And if we were to do so, how much might it change our answer?

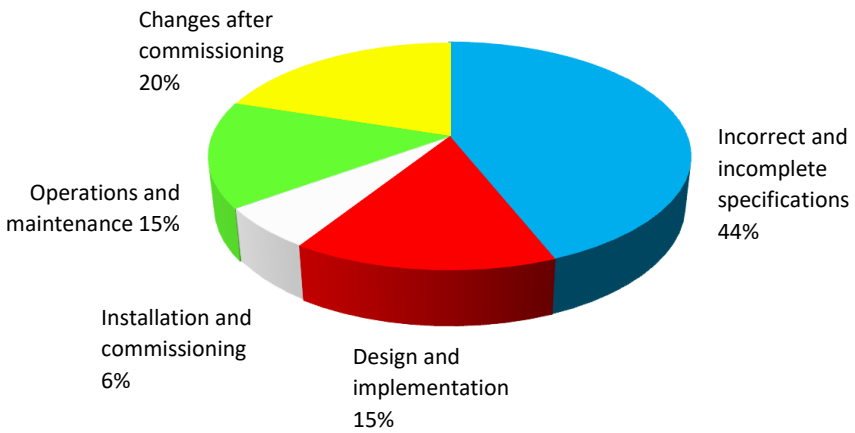


Figure 1: The causes of accidents involving control and safety systems

5. What safe plants and safe drivers have in common

What's the definition of a safe plant? Some have responded, "One that hasn't had an accident." As discussed earlier, such thinking is flawed. Similarly, what's the definition of a safe driver? One that hasn't had an accident? If the salesman driver mentioned earlier tried to reassure you by saying that he drives that way *all the time* and he's *never* had an accident, would you be reassured? It should obvious to everyone that a safe driver is one who follows the rules and laws, doesn't drive under the influence of alcohol or drugs, is not distracted by texting, wears a safety belt, keeps the car in good condition, etc. Yet does doing so *guarantee* there will not be an accident? Obviously not, but it does lower the probability. The same applies to a safe plant. You *don't* define safety by the *absence* of a very rare event; you define it by the *presence* of common and readily observable behaviors. And we *can* model this!

6. What the swiss cheese model, process safety management, and Jenga have in common

James Reason came up with the swiss cheese model in the late 1990's, as shown in Figure 2. It's a graphical representation of protection and mitigation layers. The effectiveness of each layer is represented by the size and number of holes in each layer. The holes are controlled management; the more effective the management, the few and smaller the holes. Accidents happen when the holes line up and a single event can proceed through each layer.

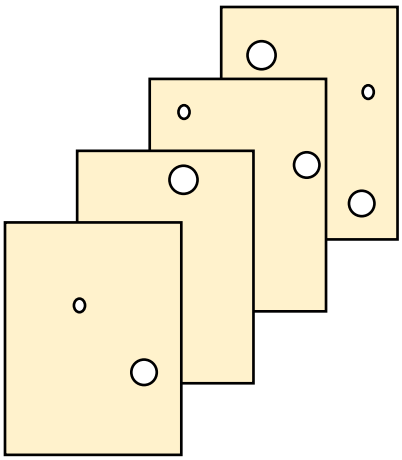
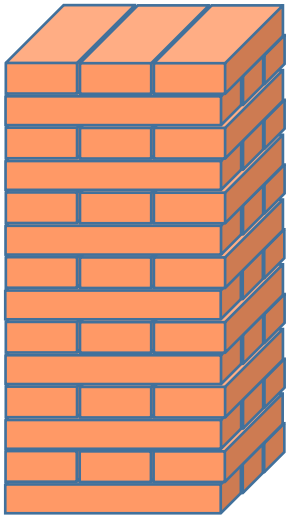


Figure 2: The swiss cheese model

A similar concept can be represented graphically by comparing the 14 elements of the OSHA process safety management (PSM) regulation to a Jenga tower, as shown in Figure 3. Think of the 14 main elements as layers, and the sub-elements as individual pieces within each layer. An effective implementation of all the clauses in the regulation would be similar to a complete Jenga tower, or a swiss cheese model with very few holes, and small ones at that.



1. Trade secrets
2. Compliance audits
3. Emergency planning and response
4. Incident investigation
5. Management of change
6. Hot work permit
7. Mechanical integrity
8. Pre-startup safety review
9. Contractors
10. Training
11. Operating procedures
12. Process hazards analysis
13. Process safety information
14. Employee participation

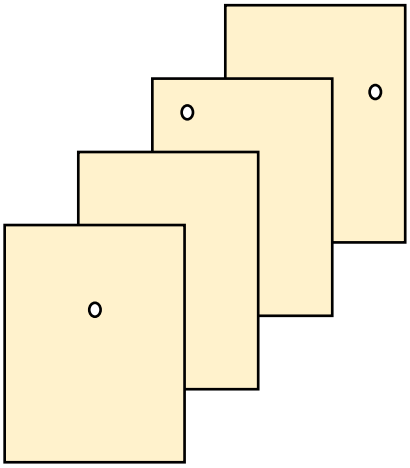


Figure 3: (Effective) Process Safety Management, Jenga, and the swiss cheese model

But how many people working in process plants truly believe their facility has *all* the pieces in place, and that they are *all* 100% effective? Perhaps your facility is more like the tower and swiss cheese model in Figure 4.

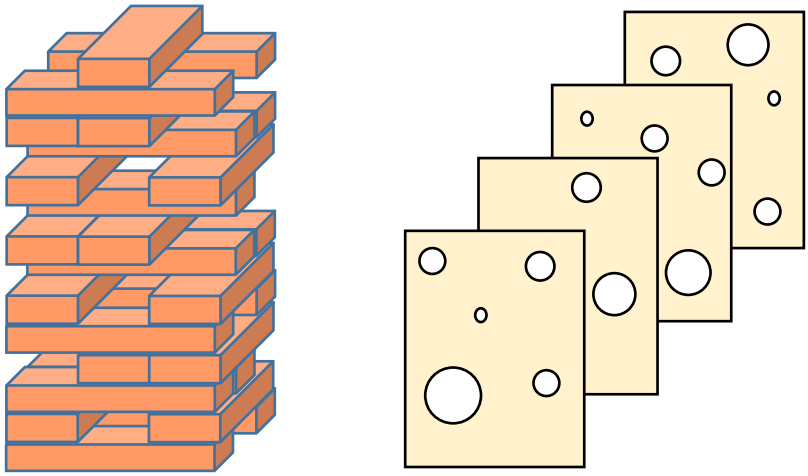


Figure 4: (Ineffective) Process Safety Management, Jenga, and the swiss cheese model

What's deceptive is that the tower in Figure 4 is still standing. Everyone then naturally assumes they must be OK. ("We've been operating this way for 15 years and haven't had an accident yet; we must be safe.") Yet anyone would realize the tower is not as strong or as resilient as the one in Figure 3. Langewiesche said "Murphy's law is wrong. Everything that can go wrong usually goes **right**, and then we draw the **wrong** conclusions." Might we be able to evaluate the completeness of the tower, or the number of holes in the swiss cheese model, and determine the impact on safety? If you knew the various layers were imperfect, might you be able to update your "prior belief" based on newly acquired information, even if that information were *subjective*?

7. Bayesian networks

Functional safety practitioners will be familiar with fault trees and event trees. What might be new to many are Bayesian networks, a simple example of which is shown in Figure 5. Just as with the other modelling techniques, there is math associated with how the network diagram elements interact with each other. There are also commercial programs available to solve them automatically, as diagrams can get large and complex and the math too unwieldy to solve by hand. One interesting aspect of Bayesian networks is that the math and probability tables may be based on *subjective* ranking (e.g., low, medium, high).

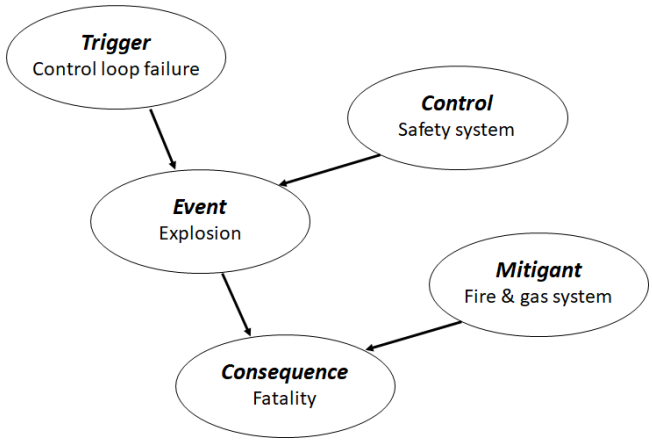


Figure 5: Sample Bayesian network

The case of interest here is to model the impact of the PSM program on the performance of a safety instrumented function (SIF). Imagine a SIF with a target of safety integrity level (SIL) 3. Imagine a fully fault-tolerant system (sensors, logic solver, and final elements) with a calculated probability of failure on demand of 0.0002. The reciprocal of this number is the risk reduction factor (RRF = 5,000), which is in the SIL 3 range as shown in Table 2.

SIL Target	RRF Range
4	10,000 – 100,000
3	1,000 – 10,000
2	100 – 1,000
1	10-100

Table 2: SIL and risk reduction factor

As noted earlier the calculations are based on frequentist statistics and the numbers cannot be proven. But as cited in the examples above, our "prior estimate" could be updated with new information, even if it were subjective. This example can be represented in the simple Bayesian network shown in Figure 6.

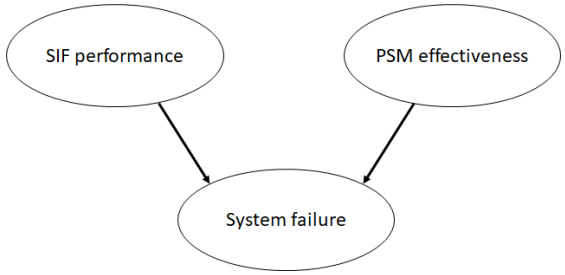


Figure 6: Bayesian network including a subjective factor

It's easiest to understand the solution of a Bayesian network if it can be shown graphically. This simple example can be solved with an event tree. All that's left for us to decide is what numerical values to assign to the ranking scales for the possible effectiveness of the overall PSM program. Admittedly there are many factors that could be evaluated here (e.g., competency, staffing levels, completeness of procedures, effectiveness of management of change, effectiveness of testing, etc.). The example here will simply group all these factors together. Two example ranges are shown in Table 3. Before proceeding, do you think these values are *reasonable*?

Ranked Scale	Optimistic Value	Pessimistic Value
Very high	99.99%	99%
High	99.9%	90%
Medium	99%	80%
Low	90%	60%
Very low	< 90%	< 60%

Table 3: Possible ranking for the effectiveness of a PSM program

The event tree using one value of PSM effectiveness (99%) is shown in Figure 7. Table 4 lists the results for all the possible values.

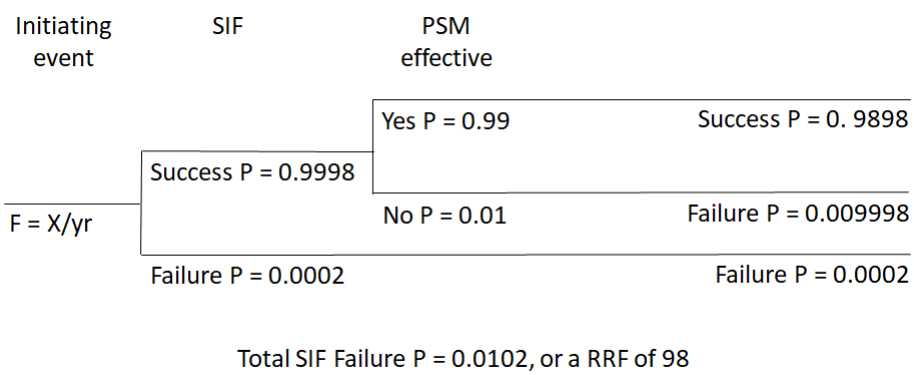


Figure 6: Bayesian network including a subjective factor

Ranked Scale	Optimistic Value	SIF RRF	Pessimistic Value	SIF RRF
Very high	99.99%	3,300	99%	98
High	99.9%	833	90%	10
Medium	99%	98	80%	5
Low	90%	10	60%	3
Very low	< 90%	<10	< 60%	<3

Table 4: SIF performance based on PSM effectiveness

The initial idealistic calculation (our prior belief) showed the risk reduction factor to be 5,000. Including another factor to update our belief results in a *dramatic* change to the number. Simply achieving SIL 2 may end up being very difficult in the real world. Admittedly, assigning numbers to the qualitative rankings of the PSM program will be a point of contention. Before showing these results to your subject matter expert (SME) team members, ask them one simple question; what do *they* think is the overall effectiveness of the PSM program at *their* facility? *Then* show them the results.

8. Conclusion

Being a safe driver is accomplished by following all the rules that are known to help avoid accidents. Similarly, operating a safe plant is accomplished by following all the rules and regulations effectively. Yet it's easy for functional safety engineers to focus instead on math and hardware calculations. The frequentist based statistical calculations result in extremely small numbers that cannot be proven. However, the prior belief probability can be updated with even subjective information. Doing so can change the answer orders of magnitude. The key takeaway is that the focus of functional safety should be on effectively following all the steps in the ISA/IEC 61511 safety lifecycle and the requirements of the OSHA PSM regulation, not the math (or certification of devices). Both documents were essentially written in blood through lessons learned the hard way by many organizations.

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4. "Process safety management of highly hazardous chemicals", 29 CFR 1910.119
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Annex: Solution to the heart disease problem

Only one person out of a thousand has the disease. Yet if 5% of the people test as false positives, that would be 50 people out of a thousand that are *diagnosed*, but do *not* actually have the disease. So the probability of actually *having* the disease based on test results is one out of 51 people (the 50 false positives, plus the one who actually has the disease), which is just under 2%.

Every medical test result in false positives. Don't be mislead by your medical practitioner who may not have a full understanding of the statistics!

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"My Vision of Future Instrumented Protective Systems"

J. Gregory Hall
Principal Electrical Engineer
Eastman Chemical Company
jghall@eastman.com

Abstract

Instrumented Protective Systems (IPS) are essential to reducing the potential risk of possible process hazards in a process facility. IPS are composed of any combination of sensor(s), logic solver(s), and final element(s) used to implement protective functions that detect abnormal or unacceptable operating conditions and take action on the process to achieve or maintain a safe state. IPS are used to reduce the process risk associated with health and safety effects, environmental impacts, loss of property and business interruption costs. What is the future of IPS in respect to design, construction, operation, and maintenance throughout their safety life cycle?

The following will be discussed:

1. Virtual reality HMI – Human Machine Interface
2. Continuous improvement of IPS standards, practices, and procedures
3. Modular engineering safety designs
4. Real time availability of documentation (e.g., drawings, specifications, logic diagrams)
5. Real time identification of potential and actual failures
6. Safety integrity level (SIL) automatically maintained when a failure/s occur
7. IPS informs operations, maintenance, and engineering of failures with possible solutions
8. Manual proof tests that achieve 100% proof test coverage
9. Real time automated proof testing with 100% proof test coverage
10. Reduce (or possibly eliminate) systemic failures and human error

Keywords: IPS - Instrumented Protective Systems

IPS – Instrumented Protective Systems

Instrumented Protective Systems (IPS) are essential to reducing the potential risk of possible process hazards in a process facility. IPS are composed of any combination of sensor(s), logic solver(s), and final element(s) used to implement protective functions that detect abnormal or unacceptable operating conditions and act on the process to achieve or maintain a safe state. IPS are used to reduce the process risk associated with health and safety effects, environmental impacts, loss of property and business interruption costs. IPS are composed of three categories SIS (Safety Instrumented Systems), Safety, and Non-Safety interlocks.

My Vision of the Future IPS

"Star Trek" has shown glimpses of what the future of IPS could look like. Of course, my favorite characters are "Mr. Scott", "Geordi", and "Data". "The Captain" constantly relied upon his "Engineer" to push the Enterprise past the limits of its design specifications without having an incident and breaking into a million pieces. Not a good thing to do in space. Does that sound familiar to your process facility? Management constantly pushing operations to increase production, reduce downtime, reduce capital cost, and lengthen the time between maintenance turnarounds.

In 1966, "Star Trek" the show was so technically advanced with its personal communicators, talking computer, flat screen monitors, and tricorders. In 1987, "Star Trek the Next Generation" brought more advanced computers and a robot named "Data", which presented information, documentation, and real-time data about the operation and health of the Enterprise to Geordi and offer solutions to fix problems in the middle of space, far from any space port. This was done through voice, holographs, and other virtual reality HMI. Geordi's special glasses allowed him to see and communicate with the computer from any location. All these things have become common place in today's world and are being applied to IPS, process control, and safety systems.

I see a future where engineering, operations, and maintenance personnel will have instantaneous access to real-time data and information about the operating status and health of IPS. Thus, allowing solutions to be conceived and implemented before production is reduced, lost, or shutdown. We must improve our engineering documentation, metadata management, process control, and safety systems to ensure data and information flow into and through our computer systems to provide this real-time data and information to whomever needs it at their current location 24/7 and for them to be able to respond accordingly.

Let's look at some topics which will support this objective:

1. <u>Engineering Documentation</u>

It all begins with engineering documentation used to design, construct, and implement IPS. Each interlock is assigned an IPS classification based on the hazard and risk assessment. The interlock designer must meet specific requirements for each IPS classification. Standards, practices, and procedures outline the engineering documents that will become electronic records in the instrumentation database. These records are available as needed through the computer system for engineering, operations, and maintenance during the IPS safety lifecycle. The objective is to ensure accurate and correct information is entered into the record management system. With any database, "garbage in" creates "garbage out".

Once these documents are entered into the instrumentation database, they must be revised and modified to reflect any changes done after engineering during operation and maintenance. If these are not kept current, it will take more time and effort to troubleshoot problems and keep process facilities operating at maximum up-time.

2. Metadata Management

Metadata is the newest term for data management. There is an overload of engineering, operating, and maintenance data available and we continue to want more. The key is to gather and coordinate this data from different computer systems. There are metadata standards to assist with this effort which define standard data fields. There are many programming challenges for IT, process control, and IPS programmers to work together to provide real-time IPS operating and health status in a useable format to whomever needs it at any time and location.

3. Operations / Maintenance

Operations is constantly being pushed to meet production schedules and reduce down-time which directly conflicts with IPS maintenance (e.g., calibration, verification, proof testing, and repair). SIS and Safety interlocks must be tested on regular frequencies which typically corresponds with plant turnarounds to perform full stroke valve testing. SIS interlocks must always maintain the required SIL (safety integrity level) or additional constraints or measures are required to replace any risk gap/s lost due to failures, defeats, or repairs and must be completed within the MTTR (mean time to restore). IPS interlocks can be designed with redundant or additional instrumentation, and the safety logic solver programmed to maintain the SIL when a failure occurs, or defeats are used to allow time for repairs to be completed, proof testing completed, and the interlock restored. If the SIL is maintained, the repair does not have to be completed within the MTTR (mean time to restore).

Identifying actual and potential instrumentation failures is a work in progress. Instrumentation and control system manufacturers are working to improve diagnostics capabilities and provide standard programming / configuration modules in the safety logic solver. SIS initiator voting schemes can be configured to allow one initiator failure or one

initiator defeat at any one time and still maintain the required SIL. This allows on-line calibration, verification, and repair to be accomplished without affecting the integrity of the interlock.

There is lots of potential in developing better on-line and automated proof testing tools and procedures. Imperfect testing has a big effect on SIS proof test intervals and determining the end of life of instrumentation. We must constantly work on methods and procedures to improve testing and move closer to 100% proof test coverage with automated safety shutoff valves being most difficult to implement. On-line and automated proof testing will need to advance to make headway on proving valve operation. There is a fine line between more frequent testing and repairing or replacing valves more frequently. This is an interest subject, because every time a valve is touched with in the field, systematic errors can occur which can be worst than imperfect test coverage.

Robotics, robots, and drones are advancing very quickly and taking over dangerous jobs that were performed by human. These devices are providing real-time data with video to show us equipment located in area where human can't go. They can make repairs and other vital functions. This is only the beginning of what can be done.

4. Virtual Reality

There have been many advancements in Virtual reality in recent years and we are beginning to look like "Star Trek". Operators and maintenance personnel are using remote HMI (Human Machine Interface) in the field to operate, diagnose, maintain, test, and view real-time data. Tablets and headsets are becoming more common.

Engineers, technicians, or others have remote access to the process control / safety systems which can email, text, or call to alert specific personnel about problems and provide possible solutions. If automated on-line testing is available, it can be done within seconds of the alert.

Virtual reality simulation assist engineering with process designs and eliminate potential problems before they are installed in the field. Equipment and processes can be redesigned to be intrinsically safe and reduce the need for IPS.

CONCLUSION

It is a great time to be involved with IPS and the exciting possibilities of making our systems safer. "Star Trek" and other science fiction books, shows, and movies help us see what is possible. We are moving toward interactive systems that will keep us inform of the operating status and health of IPS and process control systems. There will be advancements in proof test coverage, on-line testing, and automated testing. Manufacturers will develop standard software modules to detect failures, allow defeats for repairs, and automate on-line testing to restore to

normal operation while maintaining the integrity level of the SIS or Safety interlock the entire time.

It all begins with improving our documentation, metadata management, process control, and safety systems to easily provide real-time data and information about the operation and health of IPS to engineering, operations, and maintenance to support operating facilities 24/7. This will allow timely solutions to be conceived and implemented before production is reduced, lost, or shutdown.

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Overlooked reverse flow scenarios

Gabriel G Martiniano Ribeiro de Andrade, Christopher Ng, Derek Wood
Siemens Energy
4615 Southwest Freeway Suite 900, Houston, TX, 77027
andrade.gabriel@siemens.com, chris.ng@siemens.com, derek.wood@siemens.com

Abstract

Reverse flow scenarios due to latent check valve failure are critical in the design of relief systems, but often overlooked or incompletely evaluated. This type of scenario is often controlling for relief device sizing, especially for systems involving high differential pressure across pumps or compressors. This paper reviews current industry best practices to evaluate such scenarios. Specific application examples are then presented to highlight key aspects for the analysis, including identification of pressure sources as well as potential paths for reverse flow, location of and credit for relief devices, initiating events, and limiting basis for system pressure rating. In addition, potential to relieve both forward plus reverse flow simultaneously should be evaluated. Guidance is also provided to determine if vapor, liquid, or two-phase relief should be expected and whether liquid displacement or non-obvious backflow from a utility header should also be considered. Criteria to allow credit for system settle-out pressure, if applicable, and how to evaluate such credit are also provided. As system complexity increases, tips on how to estimate relief loads accurately and efficiently are also provided. Lastly, consideration of other safeguards beyond relief devices for high-risk cases is also discussed.

Keywords: safety, pressure relief, reverse flow, back flow, check valve failure

Introduction

Overpressure protection design is an integral requirement to ensure safe operation of process plants. While external fire and blocked outlet are some of the more frequently identified scenarios, less obvious ones such as reverse flow scenarios are sometimes overlooked, but could potentially result in similar or even more severe consequences for plants if not properly accounted for.

In particular, this paper draws upon the collective experience of the authors in overpressure protection design and analysis and is focused mainly on reverse flow scenarios. Methods to

establish system boundary and to identify interfaces between relatively high-pressure (HP) and lower-pressure (LP) systems are discussed, together with methods to identify initiating events, basis for sustaining reverse flow, and available reverse flow paths. Application examples are shown to demonstrate key concepts and provide guidance for estimating relief requirement consistent with current industry best practices such as API STD 521.

Reverse flow scenarios should be considered for an LP system with potential for exposure to an unintended flow from a HP system, such as due to failure of a pump or compressor or other initiating event. By itself, such an event could lead to overpressure if there is no check valve or other safeguard installed to prevent reverse flow. Even with check valve(s) installed, the potential for overpressure remains valid due to the latent and unrevealed failure of check valve(s), i.e. stuck wide-open or leaking, which in turn could provide a reverse flow path during such an event.

Identifying potential reverse flow sources

Low / high pressure system interface

Proper evaluation of the potential for reverse flow across LP / HP system interfaces is critical. The LP/HP system interfaces are usually indicated by the presence of a pump or compressor to facilitate fluid flow from the LP system to the HP system during normal operation (refer to Figure 1 and Figure 2). Other indicators of such interfaces include presence of check valves, significant difference in equipment design pressures across interconnected systems, and change in piping specification. Multiple or switching operating modes or line-ups for individual systems, if applicable, should be evaluated separately.

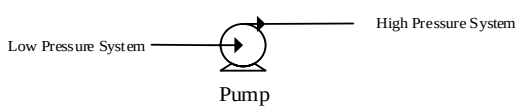

Figure 1. Low / high pressure system interface – Pumps

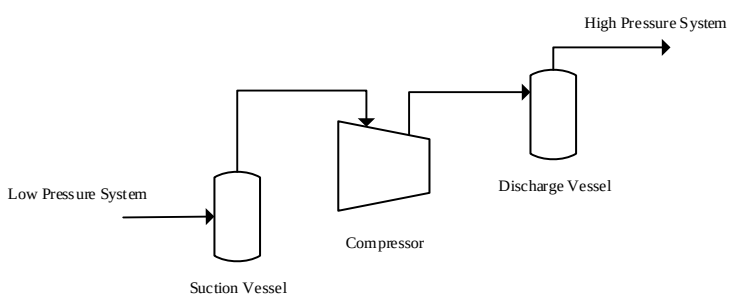

Figure 2. Low / high pressure system interface – Compressors

Initiating event and reverse flow sources

After identifying the LP/HP interface, the initiating event for the reverse flow scenario needs to be identified.

While losing the pump or compressor is typically what could allow reverse flow to occur, the initiating event can be anything from a power failure to an inadvertent closure of the suction valve that causes the pump to trip. Even without a pump/compressor, misdirected flow could occur due to a blocked outlet downstream of where multiple feeds with different maximum operating pressures combine. Identifying all events that may lead to reverse flow is important in understanding what may occur during the reverse flow and where the protected (and isolable) system boundaries are.

The next step would be to identify the potential source(s) for reverse flow once the initiating event happens. Once the pump/compressor at the LP/HP interface is lost, evaluate if the HP system has another source or feed and if its high pressure and reverse flow to the LP system would be sustainable based on:

a) Relatively large vapor inventory in HP equipment, pipeline, or utility header.

b) Vapor generation due to continued heat input

c) Other compressors continuing to operate (e.g. not related to initiating event)

d) Other liquid feed pumps continuing to operate (e.g. not related to initiating event)

In this paper, our main example is related to a hydrotreater feed surge drum as show in Figure 3. In this system, the initiating event for potential reverse flow to the feed surge drum would be the loss of the feed pump to the downstream reactor system. During this event, the reverse flow cannot be driven by the normal liquid once the pump is lost. Therefore, the backflow sources need to be identified based on the downstream system.

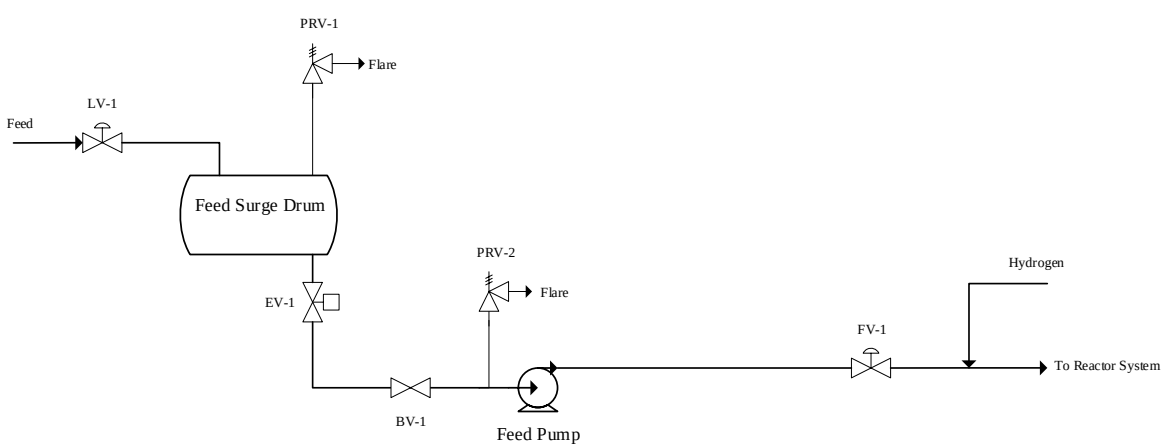

Figure 3. Hydrotreater feed surge drum system introduction

A hydrotreater would typically have some significant HP vapor inventory consisting mainly of hydrogen (in the reactor loop) and a hydrogen make-up compressor (which would not be lost during our initiating event). Thus, the reverse flow might be sustainable for some time, resulting in overpressure of the feed surge drum upon loss of the feed pump. Upon losing the liquid feed to

the reactor system, the consumption of the make-up hydrogen would decrease and could possibly sustain, or even increase, the normal reactor system pressure. For a conservative approach, no credit should be taken for any trip that might shut off all remaining unit feed. The maximum potential operating pressure in the reactor should then be used as the basis for reverse flow.

Whereas the feed surge drum in a hydrotreater is used in the example above, other similar applications are possible such as a feed surge drum feeding a column as shown in Figure 4.

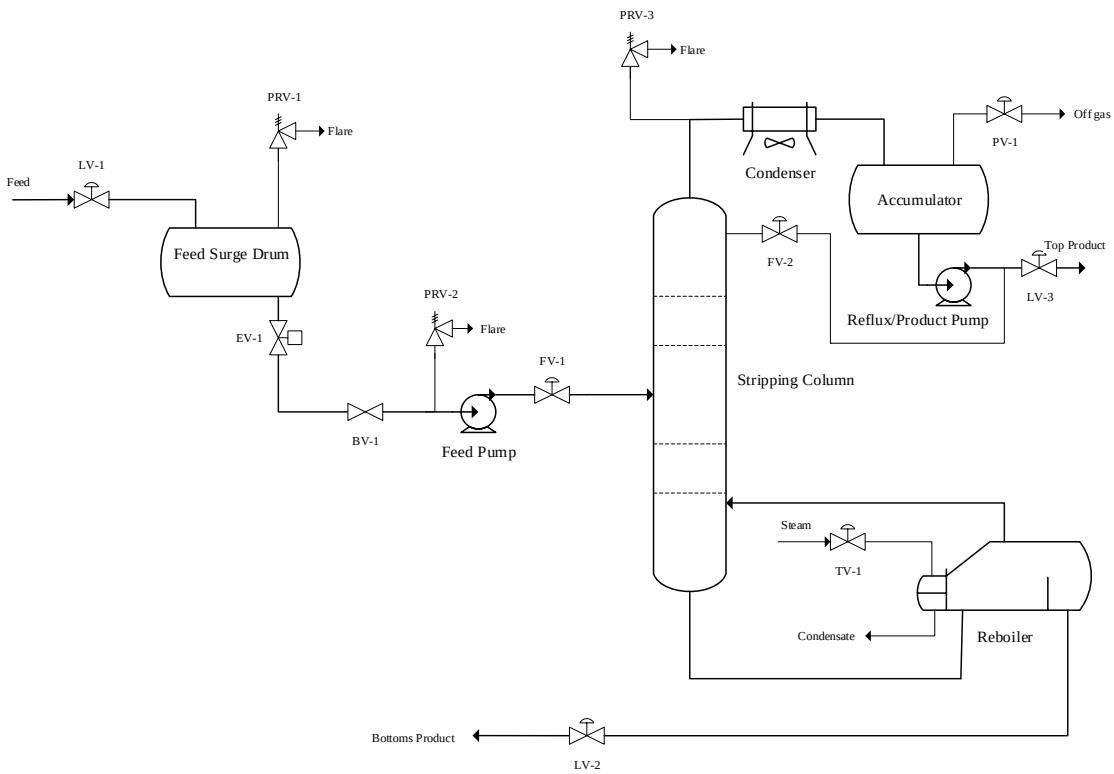

Figure 4. Stripping column feed surge drum system introduction

The column system could also be adversely impacted by the same initiating event as the loss of the feed pump might also result in reduced liquid traffic and less cooling in the column; therefore, pressure at the column could increase above its normal operation as reboiler duty might continue.

Another initiating event might involve a power failure impacting the following equipment: feed pump, condenser (air cooler fans), and reflux pump. The overall impact to the HP system (the column system) would need to be evaluated to understand what maximum pressure should be used for estimating the reverse flow rate.

Per Figure 3 and Figure 4, the LP system is assumed to have a relatively small volume compared to the HP system. If the LP system has a significant volume and the HP system does not have another sustained HP source, the two systems may equalize to a pressure between them based on the available total system volume and the total quantity of vapor in the systems. This is referred to as a settle-out condition. As an example, closed loop compressor systems, such as for refrigeration,

are commonly designed to ensure that the settle-out condition does not result in overpressure of any equipment in such a system.

Identifying paths for reverse flow

The next step is to determine what path the reverse flow might take back to the protected LP system. There are several possible flow paths to examine, depending on the general arrangement of equipment as shown in Figures 5, 6, and 7 below. The total reverse flow will be split among all available paths between the two systems with higher flow rates through the paths of lower resistance.

Check valves are intended to prevent backflow from a higher pressure system, but for the purposes of relief analysis, one should conservatively assume that a check valve can fail and allow reverse flow. This latent failure might not be noticed during normal operation (as there's forward flow through the check valve) and only becomes apparent after the pump/compressor stops and reverse flow begins.

Control valves, including pump minimum flow or recycle control valves, are often present in the reverse flow path and should always be evaluated for their potential to allow reverse flow. Depending on the initiating event and control logic, a normally closed control valve may go wide open and become a significant path for reverse flow.

Centrifugal pumps and compressors should be considered open paths for reverse flow. Positive displacement (PD) pumps and compressors require more analysis as internal components would typically restrict the available path. Reverse flow through PD machines should be considered on a case by case basis to account for the design of the rotating equipment, its operational history, and its maintenance program.

One common mistake in evaluating reverse flow is to assume that reverse flow might only occur through the main process line. Branches from the main process line, such as the minimum flow line depicted in Figure 5, may also provide open paths for reverse flow and should be evaluated. Another common mistake is to evaluate reverse flow paths independently. Because reverse flow is based on pressure differential and hydraulic resistance, each path between the LP and HP systems may experience varying rates of reverse flow simultaneously.

In Figure 5, reverse flow may occur through both the Feed Pump and through the minimum flow valve FV-2. Adding these flows together can result in a significantly higher total required relief rate than for either path alone. In Figure 6, even if flow through FV-2 is considered the path of least resistance due to the presence of dual check valves on the Feed Pump discharge, the potential for flow through the check valves should still be considered. In Figure 7, with the dual check valves downstream of the minimum flow line, the check valves can represent a common limiting element for reverse flow. This shows the importance of proper location of limiting elements within the process.

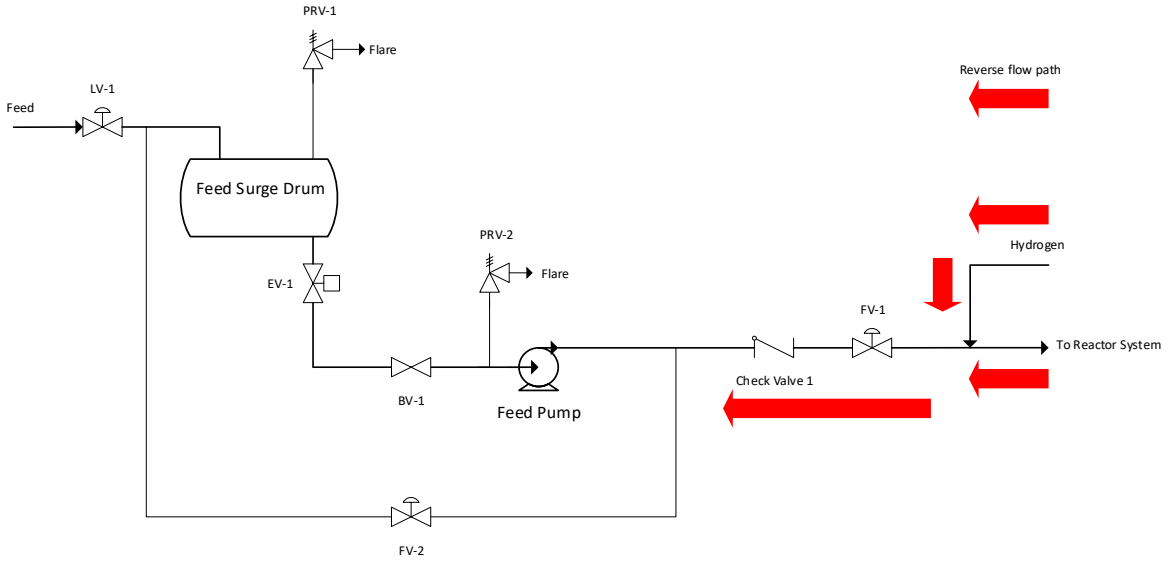

Figure 5. Reverse flow path - single check valve

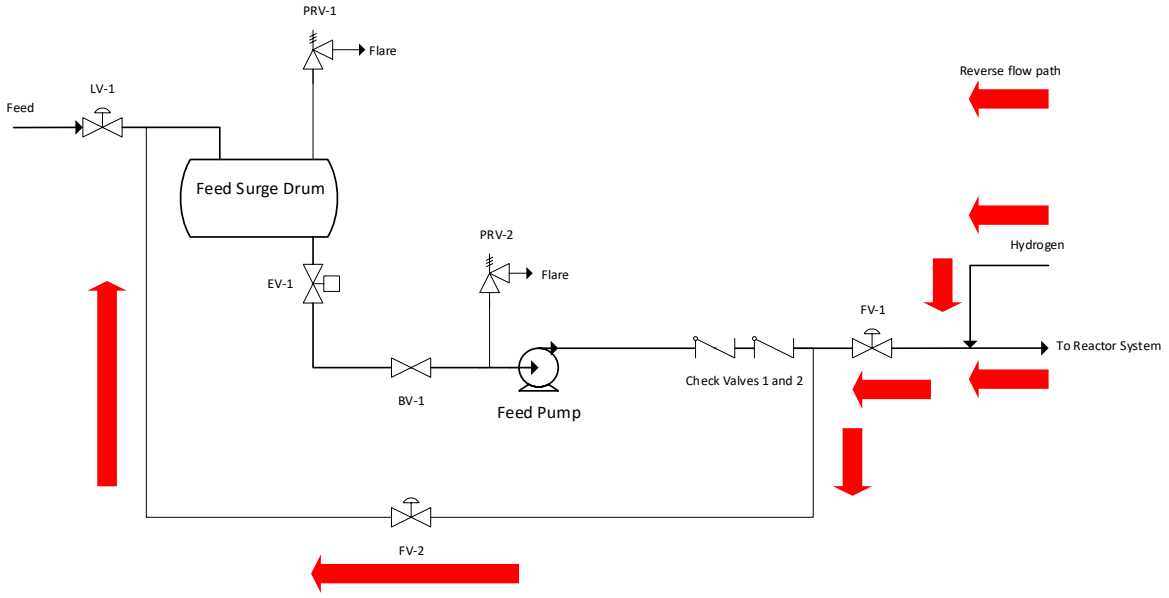

Figure 6. Reverse flow path – minimum flow control valve

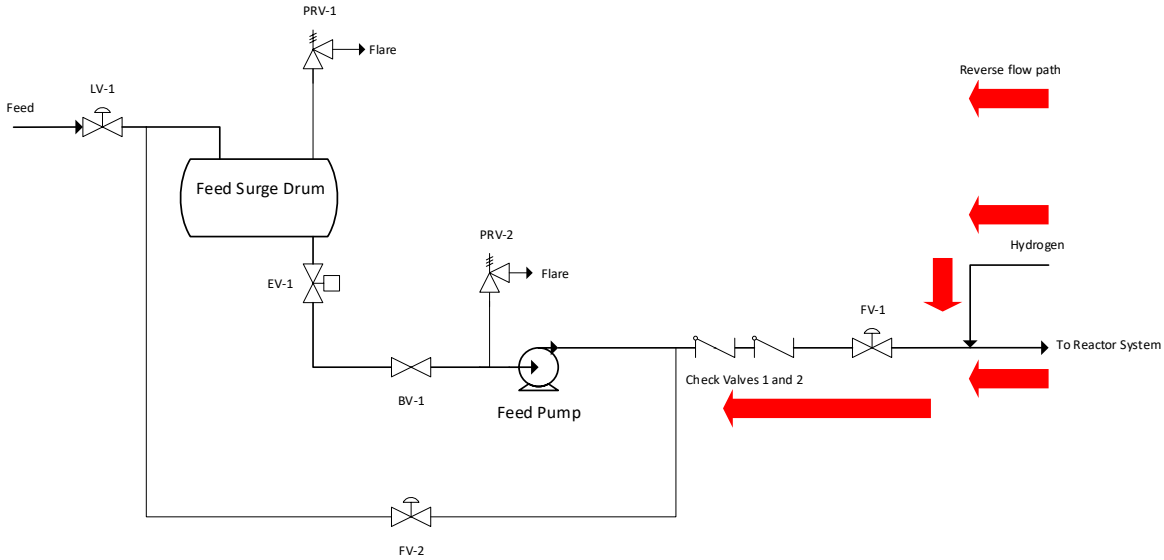

Figure 7. Reverse flow path – dual check valves

Identifying the protected system - impact of relief path location and initiating event

Next, the required location of relief devices, the boundaries of the protected system, and the type of relieving fluid should then be considered as the overpressure protection requirements and available relief path or device would depend on the initiating event.

In Figure 8, the high-pressure reactor system downstream normally contains a 2-phase inventory. What encompasses the protected system, as well as where and what size the relief valves need to be, must take many factors into account.

First, determine the possible initiating events for the scenario and the protected system boundaries. If the focus is only on failure of the Feed Pump (red boundary in Figure 8), there would be an open path back to relief valve PRV-1 on the Feed Surge Drum; thus, it might seem that there is no need for PRV-2 on the pump suction. However, if the initiating event were inadvertent closure of BV-1 (green boundary), it becomes clear that the pump suction valve and piping will become isolated from PRV-1 (taking no credit for opening of minimum flow valve FV-2). The initiating event would also impact what the limiting system pressure would be as the isolated pump suction piping might be rated higher than the upstream Feed Surge Drum.

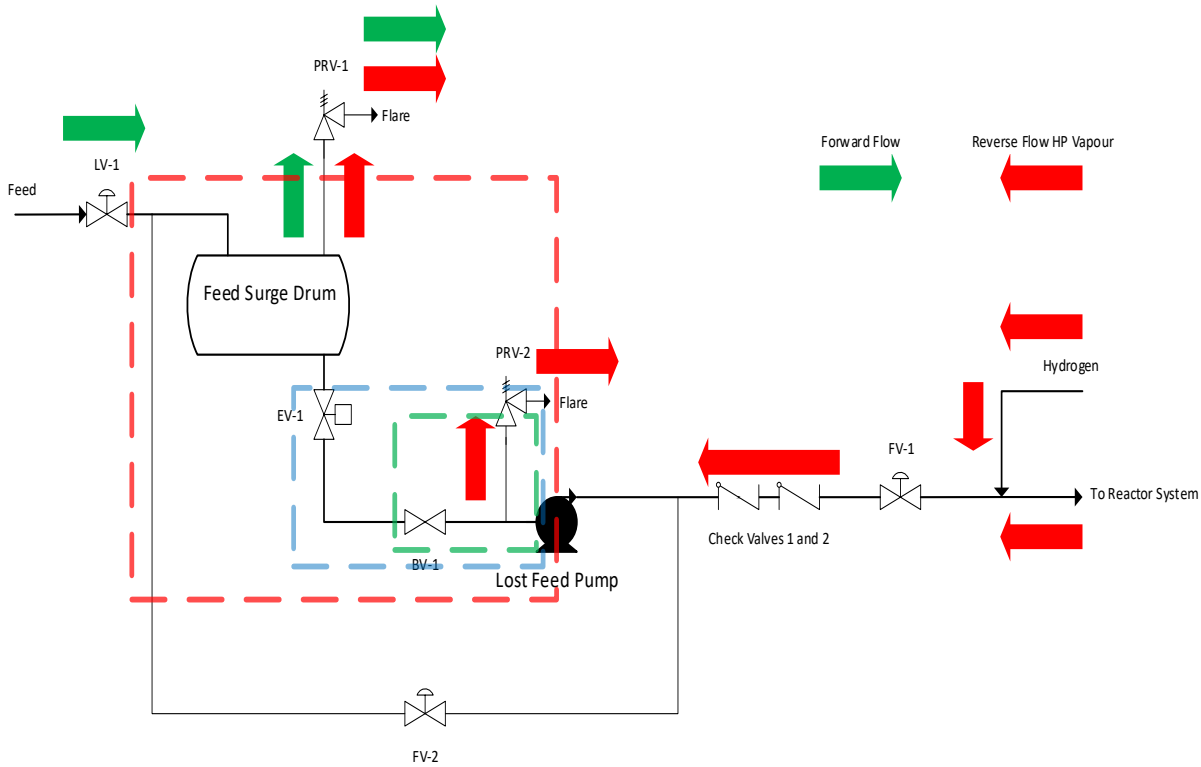

Figure 8. Protected system boundaries by initiating event

If the pump suction piping is designed for the pressure found in the HP system (which is common), perhaps no relief valve is necessary at this location. However, if the higher pressure rating only extends back to BV-1 and if the initiating event were failure of EV-1 (blue boundary), then PRV-1 may still be necessary to protect the piping from EV-1 to BV-1. As can be seen, the boundaries of the protected system and the requirements for overpressure protection can be impacted by the initiating event.

Next, consider what relief loads would be required for the upstream relief valve(s). This might be primarily affected by where the limiting elements are within the process. In Figure 5, with only one check valve downstream of the pump and assuming latent failure of that check valve (i.e. stuck wide-open), both PRV-1 and PRV-2 may see relatively large relief loads. In Figure 6, flow to PRV-2 will be limited by the dual check valves downstream of the pump, whereas flow to PRV-1 will likely be limited by the capacity of FV-2. In this case, PRV-2 may have a relatively small required relief rate while PRV-1 might have a relatively large required relief rate. In Figure 7, with the dual check valves optimally placed after the minimum flow line, both relief valves may see relatively small relief loads. Refer to the next section on how the reverse flow rate through dual check valves might be estimated. Depending on the exact design of the system in Figure 7, PRV-2 may be all that is necessary to protect against reverse flow and PRV-1 may be able to be designed for other scenarios.

Another factor to consider is what phase of fluid will be relieved. If a pump discharge line joins a header with several other pump discharges, the reverse flow may be all liquid. If the downstream system contains a large vapor inventory, the reverse flow may be vapor or two-phase. Looking at

the hydrotreater system above, there is potential to consider multiple fluid phases for relief given the downstream Make-up Gas Compressor (assuming it is unaffected by the initiating event), and the large 2-phase inventory in the reactor system itself.

Moreover, the relief valve on the pump suction (PRV-2) is located on liquid-full piping. The initial fluid through the relief valve will be existing liquid inventory that is being displaced by the reverse flow fluid. In a typical set-up such as in Figure 7, the dual check valves might be located near the pump discharge and the relief valve might be located near the pump suction. As such, the volume of liquid between the two locations tends to be small, and it might be justified to not consider the initial liquid for relief. But in some systems, the check valves may be located a significant distance downstream of the pumps or the relief valve might not be right near the pump suction. There is then a significant amount of liquid that must be displaced before the reverse flow fluid reaches the relief valve. The displacement of the liquid may need to be considered in such cases. In the worst-case scenario, high-pressure downstream vapor could flow across the check valves and become low-pressure, low-density vapor that would displace the liquid back to the relief valve. Given the typical orders of magnitude difference between vapor and liquid densities, this displacement rate can become prohibitively large for relief valve sizing and may need to be mitigated by other methods. Based on the specific details of each scenario/system, good engineering judgement should be applied to determine what fluid phase should be considered for the relief stream.

Estimating the reverse flow relief load

Depending on each application, there could be multiple potential types of reverse flow rate calculations to consider, such as:

a) No resistance for reverse flow (either no check valve present or single check valve stuck wide-open). Refer to Figure 5.

b) Dual check valves designated as safety critical, but severe leakage expected. Refer to Figure 7.

c) Reverse flow through alternative path (e.g. minimum flow recycle or bypass line). Refer to Figure 6.

d) Reverse flow through multiple diverse/parallel paths (e.g. combination of above paths).

The reverse flow rate calculation would be based on the differential pressure between the HP source (driving force for reverse flow) and the LP system, as well as the resistance for flow through the available paths described above.

Along a path for reverse flow, there may be key components besides the check valves that may contribute additional resistance for reverse flow, such as control valves or other piping restrictions. However, credit for positive instrumentation response to process control should not be taken; therefore, control valve positions should be assumed to be fully open (if the expected control response would not be favorable) or to remain at normal position (if the control response would be favorable).

The subsections below are intended to illustrate general techniques to estimate reverse flow rates based on the arrangement of check valves and other flow restricting devices. The approach is based

on collective industry experience as embodied in API STD 521 and is not intended to be prescriptive in nature. The reader is cautioned to adapt accordingly to account for the design, operational, and maintenance constraints for each specific application. For example, a check valve that is designated as safety-critical might be required to demonstrate history of reliable service under specific process conditions and fluid properties. The check valve might also be required to follow enhanced inspection and maintenance program.

Single check valve or no resistance for reverse flow

For cases where there is only a single check valve or no check valve present to prevent reverse flow, calculating the reverse flow rate by assuming the following for the check valve (if present):

a) Failed check valve has no resistance for reverse flow; or

b) Failed check valve has the same resistance for reverse flow as in the forward flow direction (open)

The resistance from all control valves or pipe fittings along the path should be taken into account to estimate the reverse flow rate in general, regardless of how many check valves are installed.

In cases where multiple check valves are present but not designated as safety critical, the reverse flow rate should be estimated the same way as if a single or no check valve were present.

Dual check valves designated as safety critical

For cases where two or more check valves designated as safety critical are present and installed in series, the following can be assumed:

a) The smallest of the check valves completely fails wide-open.

b) The remaining check valve(s) has severe leakage.

API STD 521 presents two acceptable methods or calculating leakage rates:

a) Treat the leaking check valve as an orifice with the bore diameter equal to 10% of the check valve nominal diameter; or

b) Estimate the leaking check valve as an orifice with a bore diameter that would allow for 10% of the normal forward flow through the check valve. The calculation would typically consider maximum operating pressure of the HP system and maximum allowable accumulation of the LP system.

Reverse flow through an alternative path or multiple paths

Where applicable, the distribution of reverse flow across multiple parallel paths should be taken into account. The resistance through each path should be evaluated, and. a network analysis may need to be considered. It is critical to evaluate if there is any common limiting element such as a control valve that could limit the total reverse flow rate to reduce the complexity of the case.

Fluid considerations and physical limitations

In certain cases, the reverse flow rate might be physically limited to a lower flow rate than calculated simply based on differential pressure and flow resistance. For example, the capacity of the pump may be the limiting factor for a HP liquid source. Alternatively, the vapor generation rate of a reboiler or the capacity of a compressor may be the limiting factor for a HP vapor source.

Vapor reverse flow

In the hydrotreater feed surge drum system example presented in this paper, if the initiating event is losing the feed pump, refer to Figure 9, the HP vapor will relieve through surge drum and suction piping relief devices.

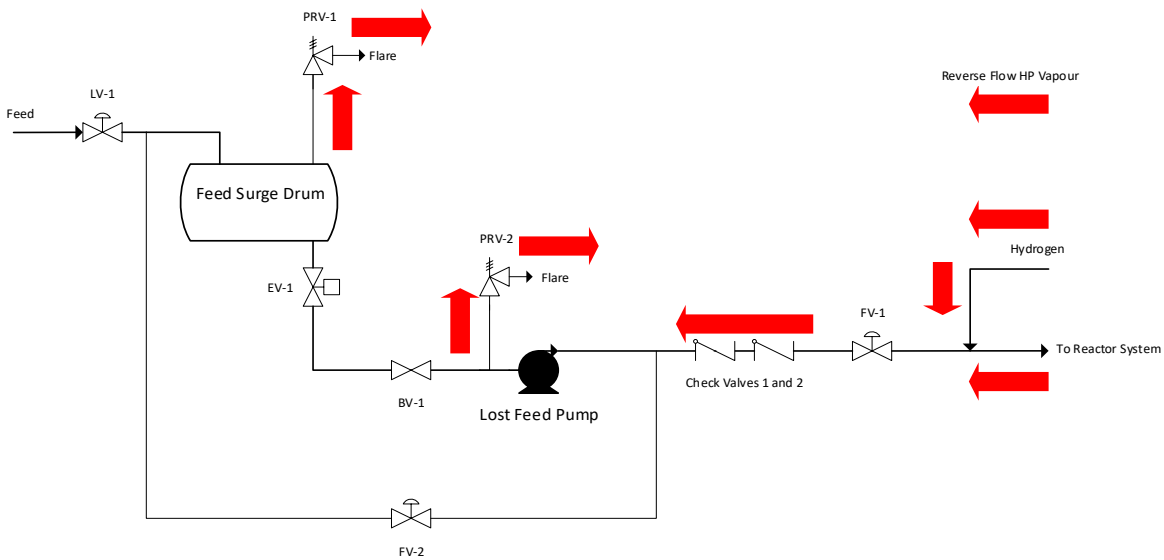

Figure 9. Reverse flow – vapor

Expanding on the hydrotreater feed surge drum case, the reverse flow could also impact the system upstream of the hydrotreater feed surge drum (refer to example on Figure 10). A common event, such as a power failure, could impact multiple systems and the initiating event would have to be evaluated to fully understand the impacts of reverse flow in HP / LP interfaces.

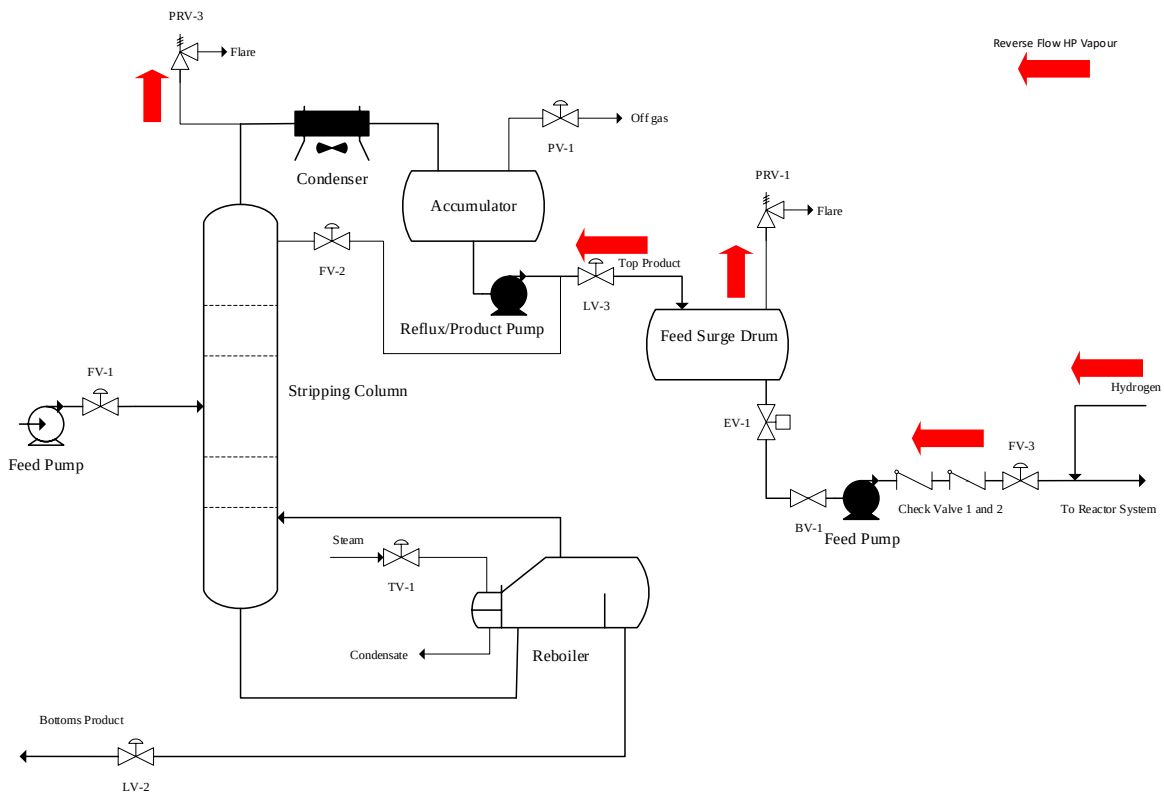

Figure 10. Reverse flow – column system

Liquid Displacement

In the hydrotreater feed surge drum system example presented in this paper, for reverse flow cases where the initiating event would be isolating the feed pump suction valve (either automatically or manually, refer to Figure 11), even though the driving force for reverse flow is a HP vapor, before the relief device can discharge vapor, the liquid trapped in the piping would need to be displaced. In this case, the liquid would be pushed at a rate where the volumetric flow rate of liquid would be equal to the volumetric flow of vapor. This rate can be obtained by multiplying the mass flow rate of vapor by the ratio between the liquid density and vapor density.

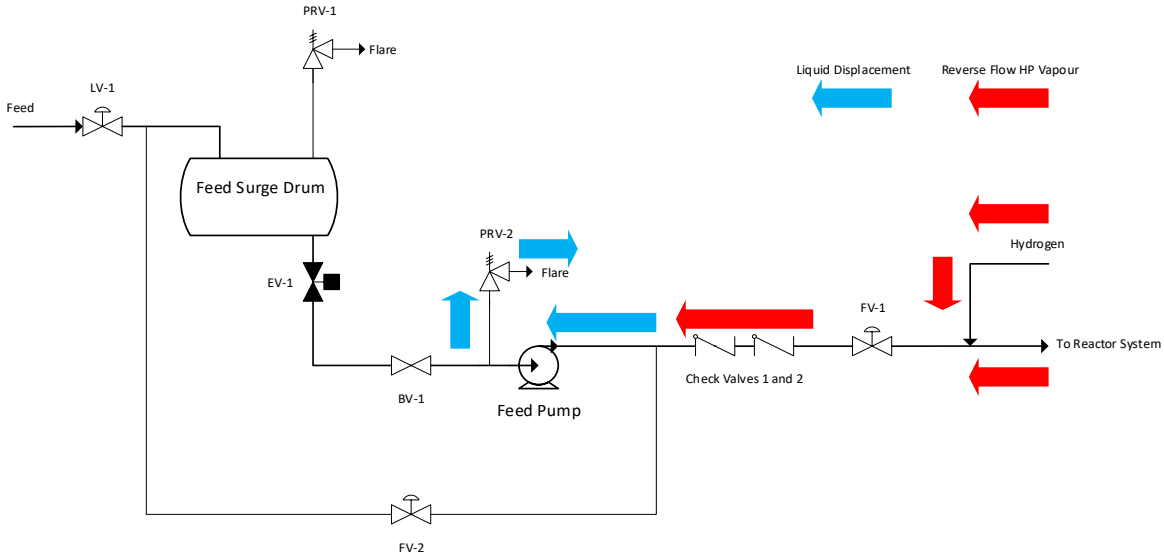

Figure 11. Reverse flow – liquid displacement

Impact of other pressure sources beyond reverse flow (estimating the total relief load)

In the previous section, multiple techniques to estimate the reverse flow rate was discussed. However, the total relief load for a reverse flow scenario may not necessarily be limited to the reverse flow itself. The same initiating event that could cause a reverse flow might also trigger other related impacts, such as blocked outlet.

Reverse flow + forward flow

As discussed, the loss of a feed pump in a hydrotreater may expose the feed surge drum system to reverse flow. If the feed to the surge drum could continue based on available upstream source pressure, the total relief load could be a mixture of the normal feed flow rate plus the reverse flow rate as show in Figure 12.

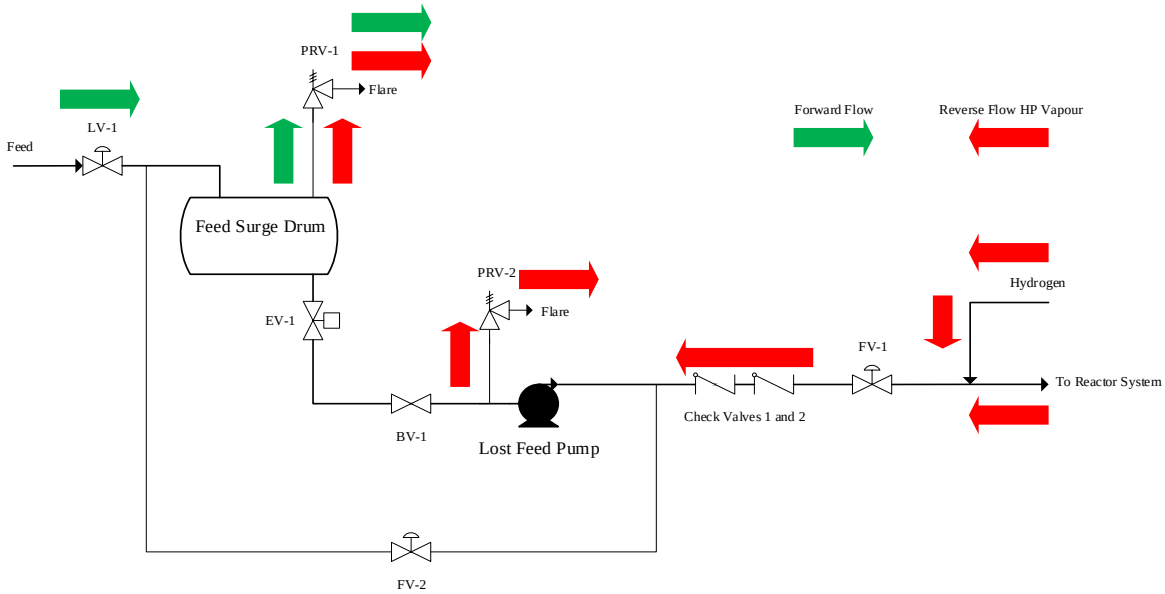

Figure 12. Reverse flow + forward flow

In another example involving a column system, the initiating event of reverse flow may also relate to other potential overpressure scenarios such as power failure that in turn would result in loss of cooling or reflux failure. As shown in Figure 13, a power failure might impact the Condenser fans, Reflux/ Product Pump, and Feed Pump. With respect to the column, loss of overhead cooling and reflux would then occur while column feed and reboiler duty could continue. An additional load to consider might be cascading reverse flow from downstream equipment via the Feed Surge Drum. A broader analysis could determine if the reverse flow might be sustainable, taking account for relative design pressures of interconnected equipment and available pressure sources downstream.

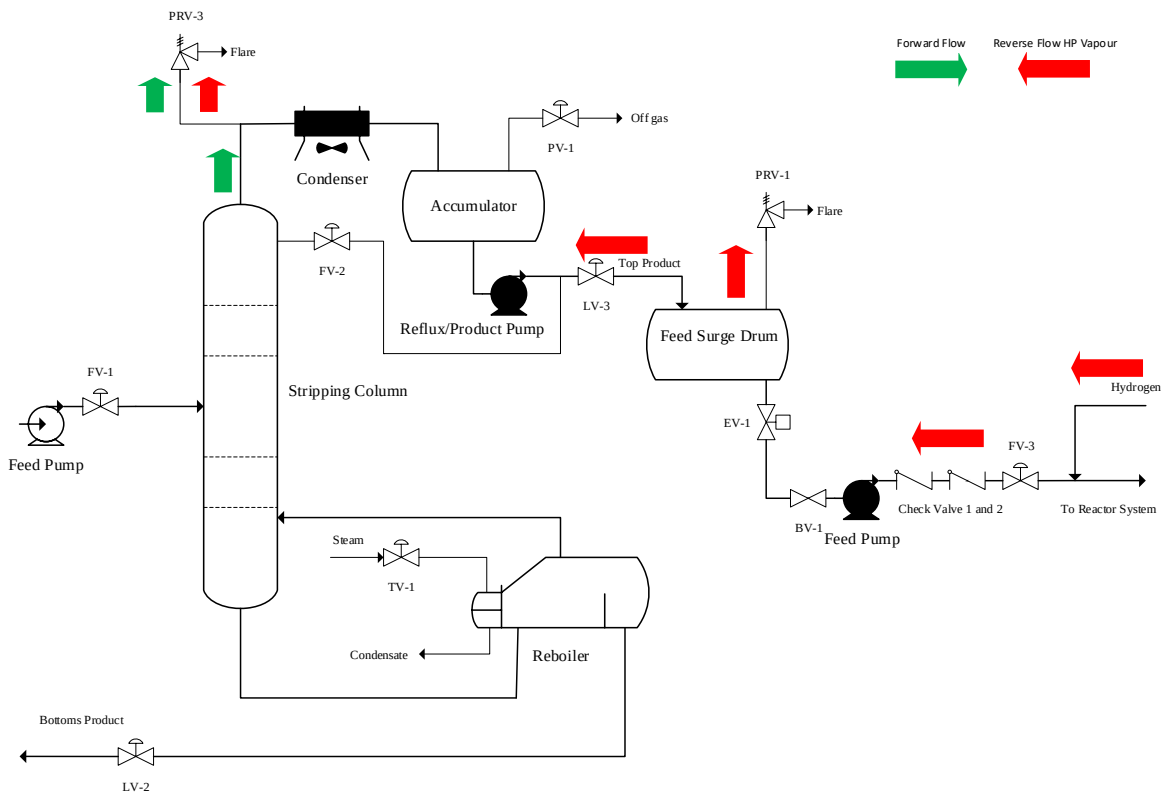

Figure 13. Reverse flow + system excess forward flow

Typical mitigation options and prevention through inherently safer design

Upon evaluating the relief load and capacity, it may be determined that overpressure protection is not sufficient and additional mitigation is required. Mitigation options may take several forms, as discussed below. In general, the design and implementation of any additional safeguard must meet the required risk reduction factor and how it is implemented should be determined on a case by case basis.

Additional relief capacity

For low risk systems with moderately undersized relief valves, the preferred mitigation may be a larger relief valve or adding relief capacity. However, for higher risk systems and/or those in which the relief valves are significantly undersized, the following mitigations could also be considered.

Dual check valves

Oftentimes, a relief valve is found to be undersized for a reverse flow scenario because there is only one check valve after the pump or compressor and there are no other significant limiting elements on the reverse flow path. The addition of a second check valve is a typical mitigation to lower the required relief rate. For more details on the typical requirements of taking credit for dual check valves, please see the section above on “Dual check valves designated as safety critical”.

If possible, the dual check valves should be placed after the branch for any alternate flow paths back to the protected system (such as minimum flow lines) so that the total reverse flow would be limited by the common check valves.

Even with the presence of dual check valves, if the pressure ratio between the high-pressure and low-pressure systems is high enough, it may be required that the check valves have different designs (dual diverse). This could prevent a potential common-mode failure and thereby lower the risk that both valves will fail simultaneously. An implied assumption would be that both check valve designs would be adequate for a specific application. Otherwise, it might be more reliable to provide redundancy by using a common proven design, rather than implementing diverse yet unproven design.

Minimum flow trip

An additional layer of protection that may be applied is a minimum flow trip on the reverse flow path. In the event that forward flow stops, i.e. feed pump fails, isolation valve(s) would be closed to prevent reverse flow.

SIL rated Safety Instrumented Function (SIF)

For higher risk systems, such as those with high pressure ratios between HP and LP systems, a minimum flow trip may not be adequate. A Safety Instrumented Function (SIF) with appropriate Safety Integrity Level (SIL) rating may be necessary to mitigate the potential risks of reverse flow. In cases where it is not practical to install adequate total relief valve capacity, the SIF may be put in place to remove reverse flow. Alternatively, a SIF may be designed to cut prevent continued forward flow from the upstream system during the same event. What SIL rating is required and the exact design of the SIF should be determined on a case by case basis. IEC 61511 provides general guidance on SIF design. Note that even when considering a SIL rated SIF, reverse flow can be a rapid process and valve closure times initiated by the SIF should be quick enough to prevent overpressure.

Inherently safer design

For new installations, the most practical protection against overpressure due to reverse flow may be inherently safe design of the system. This can include, but is not limited to, any of several factors. One might design as much equipment and piping as possible to meet the maximum pressure from a downstream system. In Figures 5-7 above, this could mean designing the piping for high pressure all the way back to EV-1. In some instances, re-rating existing equipment to higher design pressure might be feasible.

Installing dual and, if warranted, diverse check valves on particularly high-pressure ratio systems and making sure those check valves are located downstream of any branches off of the main process line could limit potential for reverse flow. Paying attention to the possible risks of reverse flow during initial design can eliminate the need for costly mitigations in the future.

Conclusion

While the evaluation methods discussed in this paper draw upon general guidance from API STD 521, the reader is cautioned that associated risks for individual systems might vary and depend on

the specific design, operation, and maintenance program for each plant. For example, there's potential for higher leakage rates through check valves that are poorly maintained or normally operating in dirty/ fouling conditions than ones that are properly maintained and in clean service. Beyond relief devices used in the examples, additional or alternate safeguards might be warranted, especially for higher risk applications such as those involving very high differential pressure between the LP and HP systems. The risk management program at a plant should be applied to ensure installed safeguards would be considered adequate. Results from other plant assessments such as Process Hazard Analysis (PHA) or Layer of Protection Analysis (LOPA) should also be considered.

Disclaimer

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Failure Under Pressure: Proper Use of Pressure Relief Device Failure Rate Data Based on Device Type and Service

Todd W. Drennen, P.E.*
Baker Engineering and Risk Consultants, Inc. (BakerRisk©)

Michael D. Moosemiller
Baker Engineering and Risk Consultants, Inc. (BakerRisk©)

*Presenter E-mail: TDrennen@BakerRisk.com

Field Code Changed

Abstract

Component failure rate data are used in a variety of quantitative and semi-quantitative study methods related to process safety and reliability, including Fault Tree Analysis (FTA), Quantitative Risk Assessment (QRA), and Layers of Protection Analysis (LOPA). In each of these methodologies, failure rate data are used to determine the probability that specific protective components, such as pressure relief devices, will fail to function as designed when called upon to prevent an incident. In the case of pressure relief devices, standardized probabilities of failure on demand are often applied with minimal consideration of the device type or the process service in which the device is employed. This paper will examine pressure relief device failure rate data from multiple published sources, categorize the data based on device type and service, and then develop guidelines for determining probability of device failure on demand based on the proposed device type and service categories. Additionally, this paper will provide commentary on the administrative aspects of relief device handling relative to observed relief valve reliability.

Keywords: Pressure Relief Valve, Rupture Disk, Vacuum Breaker, Failure Rate Data, Fault Tree Analysis (FTA), Quantitative Risk Assessment (QRA), Layers of Protection Analysis (LOPA) Probability of Failure on Demand

1 Introduction

The majority of industry facilities employ pressure relief systems. These systems are designed to prevent plant overpressure scenarios, thereby protecting personnel and the public from explosions, fires, and toxic exposures, preventing environmental releases, and preventing damage to equipment, piping, and buildings.

In the consideration of process safety risks by quantitative methods such as Fault Tree Analysis (FTA) and semi-quantitative methods such as Layers of Protection Analysis (LOPA), it is critical to quantify the probability that a relief device will fail to operate when called upon to prevent an overpressure scenario. For a relief device, or any other type of safeguard being analyzed quantitatively, this parameter is commonly referred to as the Probability of Failure on Demand (PFD).

In the Center for Chemical Process Safety (CCPS) Guidelines for Initiating Events and Independent Protection Layers (IPLs) in LOPA, typical PFD values for various types of pressure relief devices are given as shown in Table 1 below [1]. This paper will analyze pressure relief device failure rate data from a variety of published sources and compare the results of that analysis to these PFD values.

Table 1: PFD Values for Pressure Relief Devices per CCPS Guidelines [1]

IPL Classification and Description	PFD*
Spring-operated pressure relief valve	1.00E-02
Dual spring-operated pressure relief valves, no isolating valves present	1.00E-03
Dual spring-operated pressure relief valves, single manual valve can isolate one PRV	1.00E-02
Dual spring-operated pressure relief valves, single manual valve can isolate both PRVs simultaneously	1.00E-01
Pilot-operated pressure relief valve	1.00E-02
Buckling pin relief valve	1.00E-02
Rupture Disk	1.00E-02
Spring-operated pressure relief valve with rupture disk (on inlet, assumes non-fragmenting type disk and monitoring for disk burst between disk and PRV)	1.00E-02
Conservation vacuum and/or pressure relief vent	1.00E-02
Conservation vacuum and/or pressure relief vent	1.00E-02
Vacuum breaker	1.00E-02

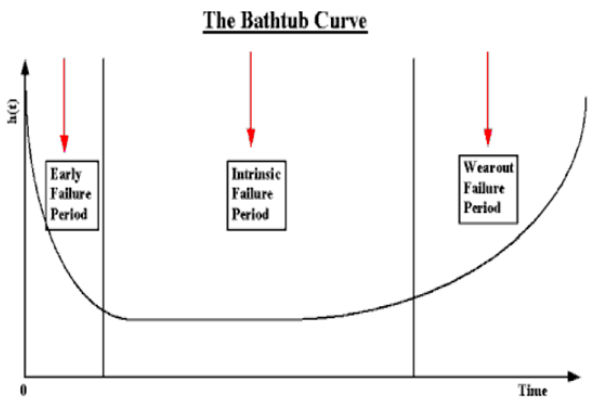
Assumes properly sized device and piping for specific scenario, clean service, and correct metallurgy

2 Discussion of Failure Types

For pressure relief devices, as with any mechanical component, there are multiple potential failure modes; however, not every failure mode necessarily constitutes a failure on demand. Failure of any component or system over its useful life is typically illustrated by a bathtub-shaped curve [2], as shown in Figure 1. This curve shows an initial high instance of failure during what is commonly

called the “early failure” or “infant mortality” period in which failure occurs due to manufacturing defects and/or improper installation. The second period of the curve is relatively flat through the component useful life and is commonly called the “intrinsic failure period,” as it characterizes the inherent component failure rate during its useful life. The third period of the curve slopes upward as the component service time extends beyond the useful life; therefore, it is called the “wear-out” or “breakdown” failure period.

Figure 1: Example Component Bathtub Curve [2]



It should be noted that there are certain types of failures, administrative in nature, which are outside the scope of this paper. For example, the authors are aware of instances in which multiple pressure relief valves were installed at a facility without removing shipping pins designed to protect the valves from damage during shipping, but which prevented the valves from operating as designed. Pressure relief valves can also be damaged through improper transport from the warehouse or shop to the field. However, other types of administrative failures would fall in the intrinsic failure period and therefore be germane to this paper. An example of this type of failure would be improper management of relief device manual isolation valves causing the protected equipment to be isolated from pressure relief protection.

For the purpose of this paper, only the intrinsic failure period will be considered, as relief devices would be expected to be either repaired to “like-new” condition or replaced prior to being operated beyond their useful life. Even when the early and late period failures are removed from consideration, not all intrinsic failure types would be failures on demand.

Mechanical failures of pressure relief devices fall into four primary categories of interest: failure to open as designed, delayed operation, spurious opening, and leakage. Failure to open as designed and delayed operation (i.e., the device opening at a higher pressure than intended) would be considered types of failure on demand, sometimes characterized as “dangerous” failures, as in either case the relief device would fail to open within the acceptable tolerances of its set pressure. Spurious opening and leakage, however, would not be considered failure on demand, as neither would prevent the device from opening when called upon to prevent overpressure. However,

spurious opening and leakage of pressure relief devices could cause releases of hazardous materials, and their risks should be properly evaluated, but discussion of that evaluation would not fall within the scope of this paper.

3 Data Constraints and Assumptions

The data constraints applied to the analysis performed for this paper are as follows:

- Data was only considered for pressure relief valves, as significantly more data was available for them than for other types of pressure relief devices, such as rupture disks and low pressure tank vents.
- Failure rate data was only considered if it was documented on a per unit time basis (i.e. per year or per hour), as opposed to a per demand basis, as pressure relief valve demand rates can vary significantly by industry, by site, and by service, while a time basis provides a more defensible comparison across different locations and circumstances. Relief valve failure mechanisms are also more likely to be time-dependent than demand-dependent.

The assumptions made regarding the data analyzed for this paper are as follows:

- All sources of data use the same criteria for defining their failure modes – for example, that "failure to open on demand" means that the relief valve did not open at the same "X%" overpressure. This failure mode definition issue is pervasive in industry across all equipment types, as noted in the CCPS book on data collection [3].
- As discussed in Section 2, the only failure types considered to be failure on demand were failure to open as designed (a.k.a. valve stuck closed, valve seizes closed, etc.) and delayed operation (a.k.a. failure to open fully at relief pressure, 10% heavy, etc.).
- For data sets for which no test interval was specified, the assumed test intervals were as follows:
    - 12 months for valves located on offshore facilities, based on the requirements established by the U.S. Bureau of Safety and Environmental Enforcement (30CFR 250.880) for pressure relief valves on U.S. offshore facilities [4]
    - 5 years for valves located at onshore facilities, as this was the maximum reported test interval for all data that listed test intervals
- For data that was categorized both by calendar time and by service time, the data sets based on service time were used.
- When calculating PFD, the following equation was used to obtain the average PFD (PFDavg) [5]:

$$\mathbf{PFDavg} \approx \tfrac{1}{2}\,\lambda_{DU}\,T_1$$

Where:
λ_{DU} = Rate of dangerous undetected failure
T_1 = Test interval

Note that in Tables 10, 11 and 12 in Section 5, this equation is simplified to "λt/2" for readability. Also, note that in the case of a pressure relief valve in continuous service, all dangerous failures (i.e. failures to operate on demand) are assumed to be undetected.

4 Data

This section presents the failure rate data analyzed for this paper. The data from each published source is given in table format.

Table 2 below presents PRV failure rate data obtained by Exida [6]. Note that the "useful life" of the PRVs studied (i.e., the period of time in which the failure rate is low and constant in the absence of proof testing) is reported to be 4 to 5 years.

Table 2: Exida PRV Failure Rate Data [6]

Failure to Open*				
Max. Rate	1.00E-07	per hour	8.76E-04	per year
Min. Rate	1.00E-08	per hour	8.76E-05	per year

All rates assume a "useful life" between proof tests of 4 to 5 years.

Table 3 shows PRV "fail dangerous" probabilities from Lees [7]. These probabilities were calculated using the PFDavg equation, shown in Section 3, based on a test interval of one year.

Table 3: Lees PRV "Fail Dangerous" Probability Data [7]

"Fail Dangerous" Probabilities*	
Max. Probability	1.00E-02
Min. Probability	4.00E-03

All probabilities assume annual testing.

Tables 4, 5, and 6 show failure rate data from the Offshore Reliability Data (OREDA) handbooks from 2002 [8], 2009 [9], and 2015 [10], respectively, with the additional summary data shown at the lower left. Table 7 shows failure rate data from Parry [11]. Note that, for Table 7, failure categories a and b are considered failures on demand as discussed in Section 2.

Table 4: OREDA 2002 PRV Failure Rate Data [8]

Item	Pop.	Install-ations	Operational Time (hrs.)	Fail to Open On Demand (Lower, per hour)	Fail to Open On Demand (Mean, per hour)	Fail to Open On Demand (Upper, per hour)	Delayed Operation (Lower, per hour)	Delayed Operation (Mean, per hour)	Delayed Operation (Upper, per hour)	Combined (Lower, per hour)	Combined (Mean, per hour)	Combined (Upper, per hour)
Relief Valves	278	7	7,169,800	1.60E-07	1.68E-06	4.61E-06	1.00E-08	1.98E-06	7.79E-06	1.70E-07	3.66E-06	1.24E-05
PSV - Conventional	170	2	4,884,700	3.30E-07	2.59E-06	6.63E-06				3.30E-07	2.59E-06	6.63E-06
PSV - Conventional 1.1" to 5"	148	2	4,244,500	3.60E-07	2.94E-06	7.61E-06				3.60E-07	2.94E-06	7.61E-06
PSV - Bellows	32	2	428,400	9.20E-07	4.05E-06	8.97E-06	6.80E-07	4.39E-06	1.08E-05	1.60E-06	8.44E-06	1.98E-05
PSV - Bellows 1.1" to 5"	25	2	339,400	1.30E-07	8.73E-06	2.76E-05	3.60E-07	5.84E-06	1.71E-05	4.90E-07	1.46E-05	4.47E-05
PSV - Pilot Operated	34	3	804,100				0	1.13E-06	4.39E-06	0.00E+00	1.13E-06	4.39E-06
Aggregate	**653**	**15**	**17,066,800**	**2.77E-07**	**2.45E-06**	**6.50E-06**	**2.84E-08**	**1.11E-06**	**4.09E-06**	**3.05E-07**	**3.56E-06**	**1.06E-05**
All Conventional	**318**	**4**	**9,129,200**	**3.44E-07**	**2.75E-06**	**7.09E-06**				**3.44E-07**	**2.75E-06**	**7.09E-06**
All Bellows	**57**	**4**	**767,800**	**5.71E-07**	**6.12E-06**	**1.72E-05**	**5.39E-07**	**5.03E-06**	**1.36E-05**	**1.11E-06**	**1.11E-05**	**3.08E-05**
All Pilot	**34**	**3**	**804,100**				**0.00E+00**	**1.13E-06**	**4.39E-06**	**0.00E+00**	**1.13E-06**	**4.39E-06**

Aggregate Lower Failure Rate Per Year	**2.67E-03**
Aggregate Mean Failure Rate Per Year	**3.12E-02**
Aggregate Upper Failure Rate Per Year	**9.28E-02**

Mean Conventional Failure Rate Per Year	**2.41E-02**
Mean Bellows Failure Rate Per Year	**9.77E-02**
Mean Pilot Operated Failure Rate Per Year	**9.90E-03**

Table 5: OREDA 2009 PRV Failure Rate Data [9]

Item	Pop.	Install-ations	Operational Time (hrs.)	Fail to Open On Demand (Lower, per hour)	Fail to Open On Demand (Mean, per hour)	Fail to Open On Demand (Upper, per hour)	Delayed Operation (Lower, per hour)	Delayed Operation (Mean, per hour)	Delayed Operation (Upper, per hour)	Combined (Lower, per hour)	Combined (Mean, per hour)	Combined (Upper, per hour)
Relief Valves	130	8	3,026,200	6.00E-08	5.70E-07	1.53E-06	9.00E-10	4.64E-06	2.07E-05	6.09E-08	5.21E-06	2.22E-05
PSV - Conventional	23	3	659,100	2.00E-08	1.36E-06	4.26E-06				2.00E-08	1.36E-06	4.26E-06
PSV - Conventional 1.1" to 5"	13	1	483,900	1.00E-07	2.07E-06	9.81E-06				1.00E-07	2.07E-06	9.81E-06
PSV - Pilot Operated	74	6	1,266,000	4.00E-10	1.86E-06	8.43E-06				4.00E-10	1.86E-06	8.43E-06
PSV - Pilot Operated 1.1" to 5"	25	2	413,600	2.00E-08	5.28E-06	2.06E-05				2.00E-08	5.28E-06	2.06E-05
Aggregate	**265**	**20**	**5,848,800**	**4.31E-08**	**1.40E-06**	**5.36E-06**	**4.66E-10**	**2.40E-06**	**1.07E-05**	**4.35E-08**	**3.80E-06**	**1.61E-05**
All Conventional	**36**	**4**	**1,143,000**	**5.39E-08**	**1.66E-06**	**6.61E-06**				**5.39E-08**	**1.66E-06**	**6.61E-06**
All Pilot	**99**	**8**	**1,679,600**	**5.23E-09**	**2.70E-06**	**1.14E-05**				**5.23E-09**	**2.70E-06**	**1.14E-05**

Aggregate Lower Failure Rate Per Year	**3.81E-04**
Aggregate Mean Failure Rate Per Year	**3.33E-02**
Aggregate Upper Failure Rate Per Year	**1.00E-01**

Mean Conventional Failure Rate Per Year	**1.45E-02**
Mean Pilot Operated Failure Rate Per Year	**2.37E-02**

Table 6: OREDA 2015 PRV Failure Rate Data [10]

Item	Pop.	Install-ations	Operational Time (hrs.)	Fail to Open On Demand (Lower, per hour)	Fail to Open On Demand (Mean, per hour)	Fail to Open On Demand (Upper, per hour)	Delayed Operation (Lower, per hour)	Delayed Operation (Mean, per hour)	Delayed Operation (Upper, per hour)	Combined (Lower, per hour)	Combined (Mean, per hour)	Combined (Upper, per hour)
Relief Valves	143	9	3,746,700	2.00E-08	4.50E-07	1.35E-06	9.00E-10	4.00E-06	1.77E-05	2.09E-08	4.45E-06	1.91E-05
PSV - Pilot Operated	86	8	1,895,700	0.00E+00	1.28E-06	6.02E-06				0.00E+00	1.28E-06	6.02E-06
Aggregate	**229**	**17**	**5,642,400**	**1.33E-08**	**7.29E-07**	**2.92E-06**	**5.98E-10**	**2.66E-06**	**1.18E-05**	**1.39E-08**	**3.38E-06**	**1.47E-05**

Aggregate Lower Failure Rate Per Year	**1.22E-04**
Aggregate Mean Failure Rate Per Year	**2.97E-02**
Aggregate Upper Failure Rate Per Year	**1.29E-01**

Mean Pilot Operated Failure Rate Per Year	**1.12E-02**

Table 7: Parry PRV Failure Rate Data [11]

Fluid Service	# of Valves	Total Valve-Years	# of Failures	Total Failures Per Year	# of Failures Per Year by Category								Failure Rates Per Year by Category								Dangerous Failure Rate Per Year (Categories a & b)	Dangerous Failure Rate Per Hour (Categories a & b)
					a	b	c	d	e	f	g	h	a	b	c	d	e	f	g	h		
Steam	33	280	39	0.139						39			.000	.000	.000	.000	.000	.139	.000	.000	0.000	0.000E+00
Water	21	210	45	0.214			3	1		41			.000	.000	.014	.005	.000	.195	.000	.000	0.000	0.000E+00
Helium	22	150	34	0.227						31	3		.000	.000	.000	.000	.000	.207	.020	.000	0.000	0.000E+00
Nitrogen	19	190	8	0.042		3		1		4			.000	.016	.000	.005	.000	.021	.000	.000	0.016	1.802E-06
Chemicals	12	57	38	0.667	4	11	8			15			.070	.193	.140	.000	.000	.263	.000	.000	0.263	3.004E-05
Lube/ Hydraulic Oil	29	141	47	0.333	1	19	9			18			.007	.135	.064	.000	.000	.128	.000	.000	0.142	1.619E-05
Natural Gas	73	144	18	0.125		7	7	1		3			.000	.049	.049	.007	.000	.021	.000	.000	0.049	5.549E-06
Ammonia	83	390	90	0.231	6	37	25			22			.015	.095	.064	.000	.000	.056	.000	.000	0.110	1.259E-05
Carbon Dioxide	104	401	123	0.307	4	46	38		1	29			.010	.115	.095	.000	.002	.072	.000	.000	0.125	1.423E-05
Hydrogen Mixtures	57	274	137	0.500	8	45	38		1	45			.029	.164	.139	.000	.004	.164	.000	.000	0.193	2.208E-05
Combined	**453**	**2237**	**579**	**0.259**	**23**	**168**	**128**	**3**	**2**	**247**	**3**	**0**	**.010**	**.075**	**.057**	**.001**	**.001**	**.110**	**.001**	**.000**	**0.085**	**9.747E-06**

The failure definitions are as follows:

(a) Failure to open at the set pressure
(b) Failure to open fully at the relieving pressure
(c) Premature opening below the set pressure
(d) Failure to reseat after opening
(e) Valve chatter, i.e. rapid opening and closing
(f) Leakage through the valve seat
(g) Leakage through the valve body
(h) Rupture of the valve body

Aggregate Dangerous Failure Rates Per Year:

8.54E-02	(All)
7.85E-02	(Clean Vapor Service)
9.77E-02	(Other Service)

Table 8 presents PRV "failure to open" rate data obtained by SINTEF [12]. Note that, for these data, a test interval of four years was reported.

Table 8: SINTEF PRV "Failure to Open" Rate Data [12]

"Failure to Open" Rate				
Reported Rate	1.00E-06	per hour	8.76E-03	per year

Table 9 presents PRV failure rate data obtained by the United Kingdom Atomic Energy Authority (UKAEA) [12]. Note that, for these data, the effective test interval was calculated by dividing the total number of valve-years by the total number of tests for each data set.

Table 9: UKAEA PRV Failure Rate Data [13]

Valve Type	# Valves	# Tests	Valve-Years	Effective Test Interval (yrs.)	# Seize Closed	# 10% Heavy	# Dangerous Failures	Dangerous Failure Rate Per Year	Dangerous Failure Rate Per Hour
Conventional	3906	7459	12651	1.70	130	340	470	3.72E-02	4.24E-06
Bellows	522	1587	2659	1.68	3	35	38	1.43E-02	1.63E-06
Pilot	77	135	188	1.39	2	0	2	1.07E-02	1.22E-06
All	4505	9181	15498	1.69	135	375	510	3.29E-02	3.76E-06
Valve Service	**# Valves**	**# Tests**	**Valve-Years**	**Effective Test Interval (yrs.)**	**# Seize Closed**	**# 10% Heavy**	**# Dangerous Failures**	**Dangerous Failure Rate Per Year**	**Dangerous Failure Rate Per Hour**
Air	52	102	166.6	1.63	3	3	6	3.60E-02	4.11E-06
Ammonia	47	93	142.2	1.53	0	2	2	1.41E-02	1.61E-06
Aromatic	272	795	1256.0	1.58	7	19	26	2.07E-02	2.36E-06
C2/C3	109	201	298.5	1.49	3	4	7	2.34E-02	2.68E-06
Crude Oil	30	59	103.0	1.75	0	6	6	5.82E-02	6.65E-06
Feed	7	23	35.7	1.55	0	1	1	2.80E-02	3.20E-06
Fuel Gas	55	153	239.8	1.57	9	12	21	8.76E-02	1.00E-05
Fuel Oil	52	95	153.8	1.62	1	5	6	3.90E-02	4.45E-06
Hydrogen	40	110	191.5	1.74	0	2	2	1.04E-02	1.19E-06
Light HC	47	106	185.8	1.75	0	1	1	5.38E-03	6.14E-07
LPG	13	43	67.9	1.58	0	0	0	0.00E+00	0.00E+00
Lube Oil	68	171	314.4	1.84	1	8	9	2.86E-02	3.27E-06
Mid. Dist.	63	93	170.8	1.84	2	2	4	2.34E-02	2.67E-06
Nitrogen	30	65	119.5	1.84	0	2	2	1.67E-02	1.91E-06
Organic	81	236	381.0	1.61	7	22	29	7.61E-02	8.69E-06
Process	133	297	569.0	1.92	9	19	28	4.92E-02	5.62E-06
Steam	150	352	610.0	1.73	2	12	14	2.29E-02	2.62E-06
Thermex	23	54	84.2	1.56	1	1	2	2.38E-02	2.71E-06

5 Results

The data presented in Section 4 was analyzed to obtain PFD results from each source based on the reported failure rate data and test intervals, using the equation shown in Section 3 with the exception of the Lees data [6], which was already presented as probability. With these probabilities calculated, they were aggregated into various categories by taking the geometric mean of probabilities across multiple sources. In this manner, the disparate data could be categorized by valve type from the available sources, as shown in Table 10, or by fluid service, as shown in Table 11. However, the same could not be said of valve size or set pressure, as only one of the available sources categorized the valves based on these factors.

Table 10: PRV PFD Data Categorized by Valve Type

Source		Probability of Failure on Demand	Basis
OREDA 2002:	Mean (Conventional)	1.21E-02	Assume annual testing per 30CFR 250.880 (λt/2)
	Mean (Bellows)	4.88E-02	
	Mean (Pilot Operated)	4.95E-03	
OREDA 2009:	Mean (Conventional)	7.27E-03	Assume annual testing per 30CFR 250.880 (λt/2)
	Mean (Pilot Operated)	1.18E-02	
OREDA 2015:	Mean (Pilot Operated)	5.61E-03	Assume annual testing per 30CFR 250.880 (λt/2)
UKAEA:	Conventional	3.16E-02	Calculated effective test interval = 1.70 years (λt/2)
	Bellows	1.20E-02	Calculated effective test interval = 1.68 years (λt/2)
	Pilot	7.40E-03	Calculated effective test interval = 1.39 years (λt/2)
Geometric Mean (All Above Data)		**1.16E-02**	--
Geometric Mean (Conventional)		**1.40E-02**	**Note the limited data set**
Geometric Mean (Bellows)		**2.42E-02**	**Note the limited data set**
Geometric Mean (Pilot)		**7.02E-03**	**Note the limited data set**

While the data could be categorized by valve type, as shown in Table 10, it is important to note that all of the available data was taken from offshore (OREDA) and Nuclear (UKAEA) facilities; each represent specific niches among processing facilities rather than typical chemical process industry examples. Additionally, there is no apparent correlation between PFD and valve type between the OREDA and UKAEA data for any valve type other than pilot-operated. Note that bellows valves have a higher PFD than conventional valves per OREDA, but a lower PFD per UKAEA. Furthermore, statistical hypothesis testing of these data, such as a chi-square test, is of dubious value due to both the small number of valve types evaluated (three) and the small numbers of data sets available for each valve type (three for conventional, two for bellows, and four for pilot-operated). As such, these data should be approached with skepticism.

Table 11: PRV PFD Data Categorized by Valve Service

Source		Probability of Failure on Demand	Basis
Parry:	Ammonia	1.65E-01	3-year inspection interval (λt/2)
	Lube/Hydraulic Oil	2.13E-01	
	Natural Gas/Fuel Gas	7.29E-02	
	Nitrogen	2.37E-01	
	Steam	0.00E+00	
UKAEA:	Ammonia	1.08E-02	Calculated effective test interval = 1.53 years (λt/2)
	Lube/Hydraulic Oil	2.63E-02	Calculated effective test interval = 1.84 years (λt/2)
	Natural Gas/Fuel Gas	6.86E-02	Calculated effective test interval = 1.57 years (λt/2)
	Nitrogen	1.54E-02	Calculated effective test interval = 1.84 years (λt/2)
	Steam	1.99E-02	Calculated effective test interval = 1.73 years (λt/2)
Geometric Mean (All Above Data)*		**5.35E-02**	--
Geometric Mean (Ammonia)		**4.22E-02**	**Note the limited data set**
Geometric Mean (Lube/Hydraulic Oil)		**7.48E-02**	**Note the limited data set**
Geometric Mean (Natural Gas/Fuel Gas)		**7.07E-02**	**Note the limited data set**
Geometric Mean (Nitrogen)		**6.04E-02**	**Note the limited data set**
Geometric Mean (Steam)*		**1.99E-02**	**Note the limited data set**

Probabilities of zero removed from geometric mean calculations.

The data could also be categorized by valve service, as shown in Table 11, but it is unclear whether there was any overlap between the Parry and UKAEA data, either in terms of individual valves evaluated or in terms of the industry from which the data was collected. The UKAEA data came from the nuclear industry in the UK, and the Parry data, collected by a UK-based industry cooperative, may also contain some data obtained from nuclear facilities. While a chi-square test suggests that the differences in failure rates between valves in different services are statistically significant, it is noteworthy that the data above only includes those services that were evaluated by both Parry and the UKAEA. As such, skepticism seems warranted regarding these data, as well as the data categorized by valve type shown in Table 10.

Regarding the aggregation of data from all sources for pressure relief valves of all types and services, though, a broader spectrum of data are available. Table 12 illustrates this analysis, and returns a geometric mean PFD of 6.76×10^{-3} per year. On an order of magnitude scale, this value would approximate to 10^{-2} per year, consistent with the typical PFD assumed for a single pressure relief valve (spring or pilot operated) laid out in the CCPS guidelines [1].

Table 12: Overall PRV PFD Data Summary

Source		Failure Rate (Per Year)	Probability of Failure on Demand	Basis
Exida:	Min.	8.76E-05	1.75E-04	Min. Test Interval = 4 years (λt/2)
	Max.	8.76E-04	2.19E-03	Max. Test Interval = 5 years (λt/2)
Lees:	Min.	8.00E-03	4.00E-03	Defined by Lees as yearly inspection (λt/2)
	Max.	2.00E-02	1.00E-02	
OREDA 2002:	Lower (All)	2.67E-03	1.34E-03	Assume annual testing per 30CFR 250.880 (λt/2)
	Mean (All)	3.12E-02	1.56E-02	
	Upper (All)	9.28E-02	4.64E-02	
OREDA 2009:	Lower (All)	3.81E-04	1.91E-04	Assume annual testing per 30CFR 250.880 (λt/2)
	Mean (All)	3.33E-02	1.66E-02	
	Upper (All)	1.00E-01	5.00E-02	
OREDA 2015:	Lower (All)	1.22E-04	6.08E-05	Assume annual testing per 30CFR 250.880 (λt/2)
	Mean (All)	2.97E-02	1.48E-02	
	Upper (All)	1.29E-01	7.40E-03	
Parry		8.54E-02	1.28E-01	3-year inspection interval (λt/2)
SINTEF		8.76E-03	1.75E-02	4-year inspection interval (λt/2)
UKAEA:	Conventional	3.72E-02	3.16E-02	Calculated effective test interval = 1.70 years (λt/2)
	Bellows	1.43E-02	1.20E-02	Calculated effective test interval = 1.68 years (λt/2)
	Pilot	1.07E-02	7.40E-03	Calculated effective test interval = 1.39 years (λt/2)
Geometric Mean (All)		**9.23E-03**	**6.76E-03**	**--**

One can argue that the comparisons in Tables 10 and 11 are specious, given that the PFDs that are calculated are a function of differing test intervals. To clarify, a relief valve is not more likely to fail after six months in service just because its next test is 3 ½ years away (4-year test interval), compared to the same valve also in service for six months, whose next test is six months away (1-year test interval). In that sense, the best comparisons between data sets are on a 'per unit time' basis. Nonetheless, the comparisons are useful in defending the commonly held values for relief valve reliability as used in LOPA and similar studies.

6 Conclusions

On the macro level, the data analyzed in this paper serves to validate the PFD values for spring operated and pilot operated pressure relief valves presented in the CCPS IPL guidelines [1]. However, no definitive conclusion can be drawn from the PFD data categorized by valve type or by valve service due to the limited amount of available source data. In the cases of valve size and set pressure, the amount of available source data was even smaller. Therefore, it is recommended that more failure data be collected for pressure relief valves and that these data be better categorized by valve type, service, size, and set pressure. This was a primary goal of the CCPS Process Equipment Reliability Database (PERD) effort that created the CCPS data collection and analysis guidelines [3]. To our knowledge, this effort has generated some relief valve data, but little of it has been published. With more robust and detailed data, further analysis could be conducted to determine whether a correlation exists between any of these valve parameters and valve PFD.

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Additional Engineering and Documentation to Reduce Pressure Relief Mitigation Cost

Kartik Shrenik kumar Maniar, Gabriel Gandhi Martiniano Ribeiro de Andrade
Siemens Process and Safety Consultiing
4615 Southwest Freeway, Suite 900
Houston, TX 77027
kartik.maniar@siemens.com, andrade.gabriel@siemens.com

Abstract

Unit revalidation and baseline studies for pressure relief analysis can result in a long list of potential deficiencies, which may result from an increase in unit throughput, changes to industry guidance or standards, changes to company internal guidelines for such studies, conservative assumptions in the absence of required data or based on simplified initial approach, management of change (MOC) at system or unit level, or may be a combination of all these. This paper addresses what kind of additional engineering tools or processes can be applied on typical systems during revalidation studies, such as reactor loops, columns, turbines and heat exchangers, to ensure a more accurate representation of the relief scenarios to validate the deficiencies. In addition, the paper addresses what improvements in MOC processes can be implemented in order to capture, assess and reduce the cumulative adverse effect to unit pressure relief analysis due to changes.

Keywords: Additional Engineering, Pressure Relief, Management of change, management of change, Dynamic Simulation, Mitigation, rate dependent

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Virtual Reality Process Safety in Counterfactual Thinking

Sadia Najneen, Rachel Smallman, Mindy Bergman, Camille Peres, Cassie Lewis, & Joseph W. Hendricks
Environmental and Occupational Health
Psychological and Brain Sciences
Texas A&M University
College Station, Texas 77843-3122, USA
sadianajneen@tamu.edu

Abstract

Counterfactual thinking focuses on how the past might have been different and allows us to mentally manipulate our past behavior and imagine a better (or worse) alternative outcome. These thoughts are usually prompted by negative events that block one's goals. Workers in high-risk jobs often report that they are more likely to attend to potential risks in their work if they have experienced a work-related negative event. This may be due to their engaging in counterfactual thought and applying that to future situations. This pilot study is investigating whether the benefits of counterfactual thinking can be overtly included into training paradigms for workers in high risk industries. Data is being collected using a virtual reality (SecondLife®) warehouse where participants will complete two performance tasks. Between the two tasks, we seek to capitalize on a negative incident (an explosion) by having participants engage in counterfactual thinking by being prompted by a "good" or "poor" counterfactual training prompt, or a control task (not associated with counterfactual thinking; three total conditions). Participants' performance between the first and second task will be compared for the two counterfactual and control conditions. Currently there is limited research investigating the application of counterfactual training to this domain and the current research will address this gap. If successful, this training methodology may be able to minimize the risk of future incidents (and maximize performance/safety) thus it is an important line of research because it may save lives, money, and reduce injuries and incidents overall.

Keywords: Counterfactual Training, Procedure, Safety, Virtual Reality, Human Performance

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Is Attentional Shift the Problem (or something else) with Hazard Statement Compliance? An Experimental Investigation Using Eye-Tracking Technology

Jonathan Walls[a], **S. Camille Peres[b]**, & Joseph W. Hendricks[b]

[a]Psychological and Brain Sciences

[b]Environmental and Occupational Health

Texas A&M University

College Station, Texas 77843-3122, USA

peres@tamu.edu

Abstract

Recent empirical work has demonstrated some counter-intuitive findings regarding hazard statement design when embedded in procedures. Notably, this recent research suggests warning icons have the opposite intended effect of leading to higher compliance rates. The current study utilized eye-tracking technology to determine whether or not participants are attending to hazard statements based on two different exemplar designs that have yielded the largest gap in hazard statement compliance. In other words, do we observe significant differences in attention to hazard statements based on a few predominant design characterisitcs (i.e., warning icon, yellow highlighting, numbering, and borders)? Futhermore, do these attentional shifts lead to different compliance rates? Forty participants were asked to complete four rounds of tasks using the constituent procedures for those tasks. Ongoing repeated measures analyses (ANOVA, HLM) are being used to determine a) do the designs matter? and b) if differences exist, do they impact compliance? Preliminary results suggest there are differences in attention and they reflect what we expected based on previous research – previous low-compliance designs are associated with less attention. Future directions for this line of research will be discussed.

Keywords: Hazard Statement, Virtual Reality, Eye-Tracking

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Has decision making in complex situations to be just by somebody's gut feeling? Many methods exist to do it more rationally and by merging opinions.

Hans J. Pasman, William J. Rogers, and Stewart W. Behie

TEES Mary Kay O'Connor Process Safety Center
Artie McFerrin Department of Chemical Engineering
Texas A & M University, College Station, 77843 Texas, USA
hjpasman@gmail.com

Abstract

After a risk assessment has been completed and feasible risk reduction measures have been reviewed, safeguard selection must be made and/or a decision taken what residual risk would be conditionally accepted. The straightforward way is setting up a binary decision tree and compare for a certain event, e.g., the risk reduction gain versus cost of two alternatives. However, often many contributing factors must be considered, such as: the nature, importance and context of the risk, availability of the measures, procurement and maintenance costs, presence of personnel and particularly the public near the hazardous area, vulnerability of the environment, and determination and weighting of important contributing safety factors.

In such more complex cases the decision problem may take the form of building argumentation for a preferred solution with a team of experts, or making a choice from a number of options and selection criteria using independent experts/stakeholders' opinions. The former is known as the Toulmin model of argumentation, the latter are Multi-Criteria Decision Making (MCDM) methods in which one criterion being weighted as more important than another and expert opinion weighting factors are based on, e.g., education and experience. The outcome of the latter will be a ranking of the alternatives. Where the Toulmin model will squeeze out explicit rational arguments sharpened by rebuttal ones, in MCDM methods expert's gut feeling may dominate, but due to the weighting and mathematical processing best compromise ranking of the options is the result of a synthesis process. Most known is the simple linear model of the Analytic Hierarchical Process (AHP), but a number of more sophisticated methods will be briefly described. In Multi- Attribute Utility Theory (MAUT) utility is a guiding principle, hence economics dominate. A few methods will be selected for working out an example.

Keywords: Risk management decisions; Structured reasoning; Multicriteria decision making, Multi-attribute utility; risk-based decision management

1. Introduction

As follows from the title, this paper is about decision making support methods, and more specifically in the context of risks by hazardous material processing in the widest sense. It is, however, not about the behavior decision theory of Slovic et al. [1] that deals with risk perception of people, nor with types of psychic mechanisms of human decision making, such as naturalistic decision making in highly demanding operational situations [2]. Rather this paper is about decision making, often in a team, whether safety is sufficiently secured, or whether additional safeguards/barriers must be added and which ones will be adequate. These decisions are taken after having identified and characterized the hazards, assessed the risks in a certain situation, and the question arises of weighting risk reducing measures against what cost and effort. Now, most cases may be relatively straightforward and may be taken intuitively, but there are in plant designs and operations also a considerable number of complex cases in which an optimal solution is not so simple. There may be alternatives, uncertainties, immeasurable contributing effects, different views, different criteria, and overruling business objectives, while a decision should be well substantiated, and recorded so that the basis for the decision and the reasoning can be retrieved and perhaps reproduced. So, in those cases rational decision making or at least applying a rational method, rather than only intuitively, is important.

Because there is a well-recognized demand, in particular in business and management decision making under risk, many approaches and methods to prepare and aid in decision making have been developed, and a wealth of literature is available. Of general character is the paper on decision analysis by Keeney [3] and the book on smart choices by Hammond, Keeney, and Raiffa [4]. For engineering purposes Hazelrigg [5] presents a collection of approaches, such as optimization, probabilistic operations, utility, financial, and economic considerations. Predictive risk assessment, though, is afflicted with much uncertainty. This is due to incompleteness of hazard identification and scenario definition, of model deficiency and lack of reliable data. Of course, the expectation is that large scale digitization will provide a source of “big” data, while interpreting analytics will distil from those data useful information, on which more accurate predictions can be based. Although a glut of data will undoubtedly contribute importantly, so-called expert opinion and judgment will remain necessary in the decision-making process with regard to risk. As made clear, e.g., by Baybutt [6], engineering judgment and expert opinion will be relied on despite the heuristics and the many types of cognitive biases that may influence the resulting judgment. Human judgment is simply subjective and imprecise, although training may have an improving effect. Also, personality traits play an important role such as strong focus on a target, risk aversion or appetite, (im-)patience, and impulsivity.

Asking a number of experts to give their opinion, at least initially independently of each other, will to a certain extent compensate for the limitation of the individual person. Based on expert judgment, an array of methods has been developed covering variations on the theme of judging alternatives/options against criteria resulting after mathematical treatment in a ranking of the

alternatives. Most of these methods originated in the 1970s and 1980s, but over the years these methods have been improved and extended.

All of the above methods have applications in many fields, such as economics, engineering, management, and business. With regard to risk assessment it is in generating and ordering data for tools, such as Failure Mode and Effect Analysis (FMEA), combining severity and frequency data into risk for a variety of operations, or judging alternative safety improving solutions against a set of criteria. Their application for these purposes is rather unknown in the safety community, and the objective of this paper is to raise awareness about the use of these techniques for improving risk management and process safety. Although this paper has the character of a review, it is not a real review of the massive body of literature in the field, which would "overshoot' the objective. Our intention is to present just an overview given data and constraints of various possibilities for aiding decision making under uncertainty how best to curb risk. Section 2 contains the overview. A listing of methods covered is presented in Table 1, and in Section 3 brief summaries will be given with, in most cases, a few recent example articles.

2. Overview of decision-making methods

In Table 1 a brief characterization of the various methods is presented with principle, objective or application, and, as far as possible, references at origin. A rough categorization is as follows:

- *Qualitative methods* stressing developing consensus, such as in the Delphi approach as briefly explained before, or developing strong argumentation (Toulmin, see Section 3), or comparing with past cases: Case Based Reasoning (CBR) [7],[8],[9].
- *Quantitative methods*, which form the majority of cases. These methods range from numerically quite simple to more sophisticated ones. The former are the Balanced Scorecard and the Decision Tree. The sophisticated ones depend much on computer support and are classed as Multi-Criteria Decision-Making (MCDM) or Multiple-Criteria Decision Analysis (MCDA) methods, in which sometimes distinction is made between Multi-Attribute Decision-Making (MADM), when there are discrete options and Multi-Objective Decision Making (MODM) when the best options are sought in case of multiple objective variables. The majority of MCDM methods will be given attention in Section 4, and special cases, DEMATEL and ANP, in Section 5. Multi-Attribute Utility Theory (MAUT) or Multi-Attribute Value Theory (MAVT) and pure mathematic approaches, such as Pareto Front or Frontier determination, also called Multi-Objective Optimization Problem (MOOP) will be briefly reviewed in Section 6 and 7. In Section 8 comparisons are discussed.

The question put to the expert can have different forms. In most of the methods the question pertains to a pairwise comparison of items and attributing a weight factor. Based on, for example, education and experience, even the importance of experts itself can be weighted. So, the reasoning behind the choice is not made explicit. There are, however, two methods that focus on the reasoning. This can be according to a questionnaire developed by a team and answered by a second team followed by a second or further rounds till reaching consensus as in the *Delphi method* [10],[11], or by formulating arguments and rebuttals to support a certain decision as in

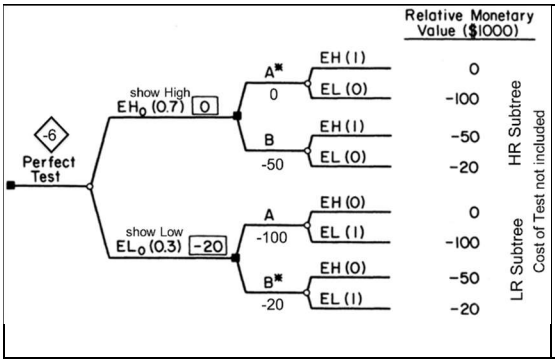
the *Toulmin approach* [12]. Because the latter approach underpins a certain choice in a very systematic way, it will be treated in more detail in Section 3.

Table 1 Overview of Decision Aiding Methods

CATE-GORY	METHOD	ORIGINATOR	Year of origin	PRINCIPLE	OBJECTIVE or APPLICATIONS	Reference
	Qualitative/rational					
Reasoning	Delphi	RAND	1944-1950	Iteration by question rounds and responses feedback by two teams of anonymous experts, until consensus.	Future prediction, policy making	[10],[11]
	Model of Argumentation	Toulmin	1958	Team deliberation to develop arguments for a decision based on reasons why, evidence, logic and rebuttals	Developing well underpinned decision	[12]
	Case Based Reasoning	Schank and Leake	1982-89	AI origin; for decision making memories of similar experiences are important; causal; patterns; algorithm	Decision making; other, e.g. hazard identification	[7],[8]
	Quantitative/estimates					
	Scorecard	Game watchers	?	Evolvement to Likert scale (e.g., 0-4) indication of extent of sufficiency, success, wellness state	Balanced scorecard of KPIs for business strategy	[17],[18]
	Binary Decision Tree	Lee, Bell System	1959	Boolean function represented as directed acyclic graph of branching decision nodes and terminal nodes	Logic circuits design, Decision making under Risk	[22],[23]
MCDM	AHP	Saaty	1977	Experts estimate on a 1-9 scale pairwise importance of criteria and alternatives followed by matrix operations	Outranking, selection of best alternative	[28][38]
	SAW-WSM, and WPM	Various authors	1950s	Criteria weights (AHP), divide by max entry/column, sum rows (SAW, WSM), multiply per row to power of weight (WPM	Outranking, selection of best alternative	[40],[41], [43]
	FAHP	Chang	1996	Triangular membership fuzzy estimates to reflect selection uncertainty; further procedure as AHP	idem	[29]
	BWM	Rezaei	2015	Best-worst version of AHP, simplifies solution procedure	Outranking, selection of best alternative	[48],[49]
	ELECTRE	Benayoun, Roy, Sussman	1966	Pairwise comparisons of alternatives; criteria are weighted; con-/discordance, discrimination; many versions	Smallest set of best alternatives; sorting	[55]
	PROMETHEE	Brans	1982	Pairwise comparison; intensity of preference, i.e. different types of stronger and weaker preference	Outranking, selection of best alternative	[58]
	ORESTE	Roubens	1982	Pairwise comparison; preference, indifference and conflict ordering; Besson ranking; distances calculation	idem	[61]
	TOPSIS	Hwang & Yoon	1980	Chosen alternative should have shortest distance from ideal solution, farthest from negative-ideal solution	Ranked preference order	[65],[66]
	VIKOR	Opricovic	1980	AHP weighted property criteria; distances like TOPSIS; interpretation --> compromise; maximum group utility	Best material or item choice based on performance	[68]
	DEMATEL	Battelle Mem. Institute	1972	In situation of multiple factors experts estimate influence strength of one on another --> cause and effect	Analysis of large variety of complex cases	[30],[77]
	ANP	Saaty	2008	Experts estimate on a 1-9 scale priorities as in AHP but follow then matrix transitivity operations as DEMATEL	Cause-effect network nodes, outranking alternative	[36]
MAUT	Utility risk analysis	Keeney & Wood	1977	List objectives with attributes; select option preferences; chose utility function; aggregate to multi-attribute score	Selection of best alternative given constraints	[94]
	Prospect theory	Kahneman-Tversky	1979	Prospect theory puts expected utility for risky decision in perspective: losses-gains; human biases weighting probability	Major step in knowledge on human risk perception	[31]
	Economic Risk Analysis	Chavas	2004	Utility function of individual preference; expected reward utility; risk premium/aversion; Arrow-Pratt coefficient	Risk analysis in economic/monetary sense	[104]
	TODIM	Gomes & Lima	1992	Gain-loss as Prospect theory; multi-attribute value function; pair comparison criteria; alternative dominance	Rank decision under risk as perceived by human	[32],[33]
MODM	Pareto front(ier)	Many authors	?	Continuous decision space; Pareto optimal set, front, or frontier presents the locations of Pareto optimality	Decision making, choice of optimal solutions	[107],[108], [109]

Case Based Reasoning (CBR) developed in the late 1980s and early 1990s thanks to the rapid growth of computer power. CBR is a product of Artificial Intelligence research and known under the expert systems. The challenge is, of course, to store case information systemically in such way that the typical characteristics of a situation can be recognized and the case retrieved. An obvious application to risk assessment is hazard identification/scenario definition, e.g., [13] applying a similarity mechanism algorithm. This is too the idea behind DyPASI (Dynamic Procedure for Atypical Scenarios Identification) [14]. In a way, the Dutch Storybuilder project of Bellamy et al. [15] comes close to CBR; it made use of a large occupational accident database, which enabled typical case bowties development. Su et al. [16] with the same objective developed a case-based reasoning method for pre-control of worker safety.

The *Balanced Scorecard* is a method for performance measurement and business strategy decisions (Kaplan and Norton [17],[18]). There are, however, many other applications. The scorecard often takes the form of simply selecting a key performance indicator number from a range, e.g., 0 – 4, called a Likert scale, representing very high, high, medium, low, and very low grades. One can also select the linguistic term itself. In case it is about a statement, a range can be: fully agree, agree, neutral, disagree, and fully disagree. Subsequently, different opinions can be aggregated simply by averaging or in case the experts are weighted by multiplying each value with the normalized weight. If experts are uncertain what term to choose, one resorts to applying fuzzy set and logic [19]. The latter can be type-1, often cast as a triangular membership function or if the expert is uncertain about membership as a (interval) type-2 set [20].



Legend:
The test concerns process efficiency. If that turns out H = High, design A is preferred, or B if L=Low. E = expected mean value
The events EH_0 (test shows High) and EL_0 (shows Low) represent the perfect test results, in which the conditional probability of EH_0 becomes = 1 if EH, or becomes = 0 if EL, and similarly for EL_0. Prior probability EH_0 (= 0.7) estimated by experts, favors design A.

$E(A|EH_0) = P(EH_0|EH)(CostA\text{-}EH)+P(EL_0|EH)\ (CostA\text{-}EL) = 1(0)+0(-100) = 0$
$E(B|EH_0) = P(EH_0|EH)(CostB\text{-}EH)+P(EL_0|EH)\ (CostB\text{-}EL) = 1(-50)+0(-20) = -50$
$E(A|EL_0) = P(EH_0|EL)(CostA\text{-}EH) + P(EL_0|EL)\ (CostA\text{-}EL) = 0(0)+1(-100) = -100$
$E(B|EL_0) = P(EH_0|EL)(CostB\text{-}EH) + P(EL_0|EL)\ (CostB\text{-}EL) = 0(-50) + 1(-20) = -20$
Therefore: E(Perfect Test) = 0.7 × 0 + 0.3 × (–20) = – 6. Test costs are not included and should be compared to the gain of choosing the right design.

Figure 1. Decision tree of what value of information of performing a perfect test can contribute to the choice between two process designs, A or B, adapted after Ang and Tang [21]. In contrast to expectation, in the test design B shows best. The gain of choosing the right design without test would be: $-6 - 0.3 \times (-100) = 24$ k$, where the test cost is 10 k$, so it pays.

Another long existing method is that of the *Binary Decision Diagrams* (BDDs) [22] with Boolean functions at the branch nodes: 1 or 0, true or false. BDD trees are directed acyclic graphs [23]. The approach, originally developed for designing electrical circuits, has been exploited for assisting predictive decision making under risk [24] in, e.g., a design process. Here one can take into account the value of acquiring more information e.g., by performing a test to lower uncertainty, versus the risk of taking the decision now. To that end, both the costs of the test, and the cost of the risk of choosing the wrong design must be expressed in dollars. Figure 1 shows the tree of a test to determine the efficiency of a process in order to choose design A, if

the test shows a high efficiency, or B otherwise. Prior to the test, the probability of the test result (high or low) shall be estimated by experts. However, such trees can easily be implemented as well in Bayesian network [25, pp. 398-399].

Multi-Criteria Decision-Making (MCDM) methods are all about ranking alternatives given criteria and other constraints. One of the criteria is often cost. A simple cost-effectiveness approach ignores besides a single effectiveness criterion, other valued characteristics/qualities of the alternative to be chosen. Therefore, a trade-off/compromise solution should be sought that satisfies most of the criteria. This means that a multiplicity of objectives that are non-measurable or comparable should be optimized, e.g., not only the objective of the least possible cost solution, which depends on a number of factors such as investment cost, usability, maintainability, etc., but also as the (conflicting) objective the lowest possible risk. For a case, the latter may depend on different weighted criteria as consequence and probability. This is not a simple problem and it turned out to generate a multiplicity of methods. The search for such compromise started as early as the 1970s, in which Zeleny [26] who was one of the pioneers also giving an interesting view on the developing management science at the time [27]. In MCDM by pairwise comparison, experts express preferences, often using a form of Likert scales, which provide alternatives of a variety of possibilities would fit best a series of different criteria. Both criteria and experts can be weighted. The result after a few matrix operations is a ranking of the alternatives from the best to the worst.

Experts can be asked to produce a relative numerical or linguistical graded term, when judging alternatives with respect to a certain measure, e.g., a property, a key performance indicator, an index or another value. The most popular method is Saaty's Analytic Hierarchy Process (AHP) [28], solving of which is claimed to be improved by the Best-Worst Method (BWM) [29]. AHP, some simpler methods and BWM are described in Section 4. Other methods in this category and their fuzzy versions will be discussed more extensively in Section 5.

In the early 1970s Battelle Memorial Institute developed the Decision-Making Trial and Evaluation Laboratory (DEMATEL) method to analyze cause-effect relationships of the many interacting factors playing a role in complex problems. The method enabled investigating problems, such as those of the world [30], resulting in better decisions. Experts are asked to indicate by pairwise comparison the strength of influence of one factor upon another. After mathematical treatment the *cause-effect* relations between the factors can be identified. DEMATEL results can be input to Bayesian Network (BN) analysis, and for a different application to the Analytic Network Process (ANP). This will be explained in detail in Section 6.

A separate category of approach is formed by the Multi-Attribute Utility Theory (MAUT) in which the discriminating measure is the utility of an alternative given criteria and constraints. If the decision making is about risk, the utility function can be chosen according to human perception of risk proposed by noble prize winners Kahneman and Tversky [31]. This function is embodied in the TODIM method [32],[33]. All of this will be summarized in Section 7. Pareto Front optimization method is also a means to enable improved decision making and will be briefly described in Section 8.

In Section 9 the various categories of methods will be compared, in Section 10 results of applying several methods to a simple process safety example will be compared, and the paper conclusions will be made in Section 8. Given the wide scope of the field a few more, for our purpose less interesting methods, can be found.

3. The Toulmin method

In his book the Uses of Argument [12] philosopher Stephen Toulmin went into great depth analyzing how an argument can be built best. It starts with a *claim* or conclusion based on certain *data,* evidence, or grounds. Between data and claim shall be a *warrant* that given the data it is allowed implicitly to make the claim. The warrant can be logical, if this-then-that, no other possibility, or can be an allowable assumption. Next are a *qualifier* indicating the warrant's strength and the *rebuttal,* which will describe conditions the warrant will not be applicable. Hence, the rebuttal feeds the qualifier. To make a further refinement about the warrant, it will be given a *backing* to ascertain with its authority the warrant is valid in this case.

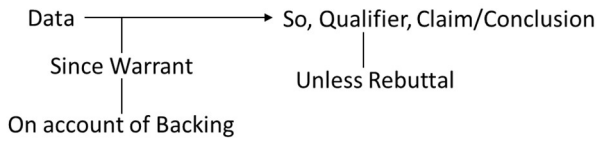


Figure 2 Toulmin's argumentation structure, adapted after [5]

A simple example is the following: Given the reliability information of a valve (datum) and the fact that this information is from OREDA, 2015 [34] (warrant), a source of authority and applicable here (backing). However, the information may not fully cover the actual valve under realistic conditions as these are very different from those covered by OREDA (rebuttal), which introduces additional uncertainty. The reliability datum is presumably (qualifier) suitable for application in our risk assessment (claim), but it needs an uncertainty estimate (conclusion). This example just illustrates Toulmin's layout, as he called it, which is depicted in Figure 2. The value of his argumentation structure is the use in complex situations of a decision to be taken by a team, e.g., about a risky operation with the claim it will still be acceptable. If such a scheme would have been followed and a rebuttal would have been given in the case of the 2010 Deepwater Horizon decision to go ahead despite the negative pressure measurement [35], the event may not have occurred. Of course, following such a scheme takes time, but a team can practice for it, and for complex design decision time may be less of a limitation.

4. AHP, SAW, WSM, WPM, Best-Worst, and fuzzy alternatives

AHP

Given criteria and alternatives an expert team is asked by the decision maker to propose the priority ranking of options by a process of importance weighting. The objective of the Analytic Hierarchy Process (AHP) method is to determine a hierarchy of a top layer of criteria importance grades with below a derived layer of ranked alternative options (candidates, policies, solutions, and many others), all based on input scores of experts. To achieve this, Saaty [28],[36],[37],[38] developed the required matrix operations in 1977. He proposed a scale of importance, ranging

from 1 (equal importance) up to 9 (extreme importance), about which experts in pairwise comparison must agree on, first with respect to n criteria and then m options. As a result, first a square reciprocal matrix, $n \times n$ is composed with preference scores of the first element to all others in the first row $p_{1,j}$ with j = 1,…,n, hence with $p_{1,1} = 1$. In the second row $p_{2,j}$ scores are presented to all others, which leads to $p_{21} = 1/p_{12}$, hence in general $p_{ij} = 1/p_{ji}$ etc. Keeping in mind the cost and other negative criteria, where a higher amount produces a lower preference. The sum of a row yields the priority, which based on the sum of all rows is normalized to a sum of 1 of the priority vector.

If judgments are transitive, that is if $x_i > x_j$ and $x_j > x_k$, then $x_i > x_k$, and the $n \times n$ matrix perfectly reciprocal, the principal eigenvalue of that square matrix is n. However, in case of expert judgment, certainly with a larger number of elements promoting errors, consistency cannot be assured and the principal eigenvalue becomes $\lambda_{\max}$. A consistency index is proposed: $CI = (\lambda_{max} - n)/(n - 1)$. In [28] error influences are analyzed by determining this index for matrices with randomly generated elements. The mean CI-value of those increases with the number of elements while asymptotically leveling off. An actual CI divided by one of a random square matrix with the same number of elements forms the consistency ratio CR. This ratio should be lower than 0.1 for a matrix of 5×5 and decreasing to 0.05 for a 3×3 one [36], otherwise the consistency of judgments should be checked.

$$A = \begin{array}{c} \\ \\ A_1 \\ A_2 \\ \vdots \\ A_m \end{array} \begin{array}{c} \begin{array}{cccc} c_1 & c_2 & \cdots & c_n \\ w_1 & w_2 & \cdots & w_n \end{array} \\ \begin{bmatrix} p_{11} & p_{21} & \cdots & p_{n1} \\ p_{12} & p_{22} & \cdots & p_{n2} \\ \vdots & \vdots & \ddots & \vdots \\ p_{1m} & p_{2m} & \cdots & p_{nm} \end{bmatrix} \end{array}$$

Figure 3 Priority $m \times n$ matrix, A, with a is alternative, c criterion, w weight, p preference.

After the operation is performed for the criteria which yields the criteria weights, it is followed by that of the m alternatives for each of the n criteria. Finally, an aggregating $m \times n$ matrix will be set-up of the criteria c_j and criteria importance weights w_j, with $j = 1, \ldots, n$, and with below each criterion c_j the priority vector column p_{ij} of alternatives A_i, with $i = 1, \ldots, m$, multiplied with the weight w_j of the criterion (Figure 3). Summing the rows and normalizing produces a ranking vector of the alternatives. A simple example on risk reduction options for a batch reactor potentially subject to runaway based on subjective judgments is worked out in the Appendix 1.

In case of many elements, small clusters can be formed and another layer in the hierarchy can be established. Another solution method is logarithmic least squares [39]. AHP became popular, but when digging deeper, problems, such as rank reversal, needed to be solved [36].

SAW, WSM, WPM

The Simple Additive Weighting, also called Weighted Sum Model (WSM), has a longer history than AHP. According to Abdullah and Adawiyah [40] in their review paper, the first time the method was mentioned was 1959 [41]. Preferences should be dimensionless or of the same dimension. In [42] an example is worked out with respect to personnel selection. The method is

less known than AHP, but it is rather similar with respect to the first stage of weighting of the criteria and the consistency check. However, instead of setting up square matrices of the alternatives for each criterion, it directly creates a $m \times n$ matrix, in which experts estimate preferences for each criterion column. As the next step, each column entry is divided by the highest value in that column ($p_j/p_{\max,j}$). After multiplication with the corresponding criterion weight and summing of a row yields a score of each alternative ($A_i = \sum_{j=1}^{n} w_j\, p_j$). For the rank presentation, results are normalized. An example with the same input values as in Appendix 1 is presented in Appendix 2. In this example, the difference in result is minimal, but due to the different ranking procedures of the alternatives with respect to a criterion there will be differences, although it is difficult to pinpoint exactly by what these arise. SAW has often been applied with fuzzy set preference estimates [40].

WPM, Weighted Product Model, differs from WSM by instead of summing, scores are multiplied. WPM is even older than WSM. Hence, after column normalization, preference of alternative A_1 over A_2 is calculated as $P(A_1/A_2) = \prod_{j=1}^{n}(p_{1j}/p_{2j})^{w_j}$, if this ratio > 1 than A_1 is ranked higher than A_2. For finding the entire ranking, this is repeated for all $i = 2, \dots, m$. The ratios are dimensionless, so different types of criteria become comparable. It is only that a higher criterion value should be a benefit. In Appendix 3 with the previous inputs, an example is given.

Triantaphyllou and Mann [43] compared AHP, a revised AHP [44], WSM, and WPM. After having varied the number of criteria, alternatives and weight ranges, authors [43] concluded that no method was fully reliable. For that reason, the title of the article closed with the wording "A Decision-Making Paradox"! In the Appendix from the input matrix of [43], which contains a rather extreme example, using all four methods preferences are calculated and results all differ. However, for the runaway risk reducing example (less extreme input), AHP, WSM, and WPM yielded the same results. Differences in outcomes of methods are due to underlying, not explicitly made assumptions. In developing the mathematics of the expected utility theory (see MAUT below), Fishburn [44]. [45], [46] has specified assumptions, also called axioms, for additive and multiplicative operations on Cartesian product sets (e.g., two sets form a $m \times n$ matrix).

BWM

Just a few years ago, Rezaei [48] developed the Best-Worst Method (BWM) which mitigates some of the drawbacks of AHP: redundant comparisons and inconsistency. BWM runs for a large part parallel to AHP, but it starts after experts examine the criteria to sort out the best and worst ones. All other criteria are pairwise compared with those two as reference. A vector will be built of how much the best criterion is preferred to the others using the 1-9 scale (elements a_{Bj}), followed by the opposite analogue for the worst one (elements a_{jW}). The optimal criteria weights w_j^* would be found if $\left|\frac{w_B}{w_j} - a_{Bj} = 0\right|$ and $\left|\frac{w_j}{w_W} - a_{jW} = 0\right|$. To that end a mathematical treatment is accomplished in which for all j the minimum is determined of the maximum of the above absolute differences subject to the constraints $\sum_j w_j = 1$ and $w_j > 0$. For the best solution it would be required that $\left|\frac{w_B}{w_j} - a_{Bj}\right| \leq \xi$ and $\left|\frac{w_j}{w_W} - a_{jW}\right| \leq \xi$ for all j subject to the above

constraints. After solving, optimal weights and ξ* are found. Ideally, ξ* should be 0. Next, maximum allowable values of ξ are calculated and used as consistency index (CI) as a function of a_{BW}, i.e., the preference of the best criterion over the worst one. The larger the latter, the larger ξ can be allowed. The consistency ratio is derived as ξ*/CI.

Because the solution of the optimization problem with more than three criteria might generate multiple optima, in a follow-on study [49] first a non-linear interval-based solution method on the weights is proposed to rank the criteria in such a case. Then, the problem is re-formulated to min $\max_j\{|w_B - a_{Bj}w_j|, |w_j - a_{jW}w_W|\}$ with $\sum_j w_j = 1$ and $w_j > 0$, for all j, which can be transferred to minimizing ξ^L with $|w_B - a_{Bj}w_j| \leq \xi^L$ and $|w_j - a_{jW}w_W| \leq \xi^L$ under the mentioned constraints. This approach allows linear programming with a unique solution. BWM produced optimum weights near the center of the intervals.

Mi et al. [50] conducted a bibliometric analysis on the method and commented on various aspects of BWM. The conclusion reached was that that the approach was quite successful: in just a few years an overwhelming number of papers applied the method and combinations with most of the methods described below have been proposed. Using this study saves time and effort (??). Several future research topics are open.

Fuzzy AHP (FAHP)

Following an older paper [51], Chang [29] presented in 1996 an evaluation procedure for a fuzzy form of AHP. For this, decision makers do not need to express their judgment in a number but due to uncertainty they can give it a range or base it on a linguistic term determined by a fuzzy set with a symmetric triangular membership function centered at an importance figure. Despite the assertion of Saaty that this was an unnecessary addition to his method [52], it became generally applied. Recently, Chan et al. [53] in a robust mathematically-based analysis accompanied by numerical experiments indicated that with increasing size of matrices the difference in result is due to the difference by the solution methods (eigenvalue for AHP versus logarithmic least squares for the fuzzy version), and it does not pay to apply the fuzzy sophistication. Increasing fuzziness by applying trapezoidal sets, even worsens the result. Only, when a small matrix is highly consistent, FAHP will contribute.

5. Miscellaneous refined MCDM methods

So far, MCDM methods treated criteria, although weighted, as independent and strict. The following more complex methods attempt to make trade-offs leading to improved compromises.

ELECTRE

ELECTRE is of French origin: ELimination Et Choix Traduisant la REalité, translated as Elimination And Choice Expressing Reality. In their review of the many ELECTRE method versions, Govindan and Brandt Jepsen [54] mention that the first description of ELECTRE is from 1966 [54]. This decision aiding method is described in detail in [56]. The objective is ranking a set A of alternative solutions (actions) to a problem. That is realized by determining action consequences (attributes, aspects, etc.) allowing pairwise comparison between actions, which is made in terms of preference (strict, P, or weak, Q), indifference, I, and incomparability,

R. Consequences can be qualitative or quantitative, deterministic or stochastic and may be based on incomplete or imprecise knowledge. Criteria F need not to be constants, but they can range between thresholds following even a function (pseudo-criterion). Actors are the analyst, who derive recommendations for consideration by the decision-makers. Further, by a matrix operation, a concordance index is determined as the strength of criteria (weights) in favor of an action being asserted as outranking, and discordance as the opposite. As a result, actions are recommended or not in different degrees, or with modifications. There are three groups of ELECTRE methods [54], [56] specialized respectively in choosing the best set (ELECTRE I), ranking of best down to worst (ELECTRE II, III -with weights, and IV without), and sorting/assigning to a hierarchy of classes, by the ELECTRE Tri methods that is further detailed in [57]. In [54] the many hundreds of applications are reviewed. In summary, the method has certain characteristics of Toulmin, but also of AHP.

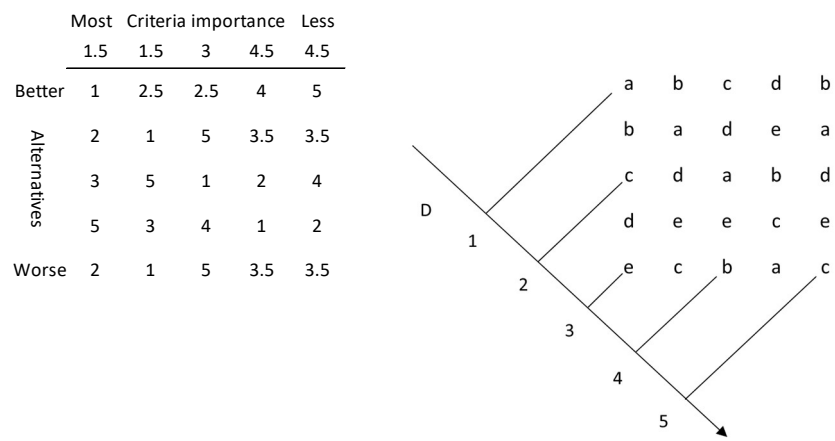
PROMETHEE

The Preference Ranking Organization METHod for Enrichment of Evaluations, PROMETHEE, with its GAIA extension (Geometrical Analysis for Interactive Aid) for visualization of the decision problem, stems from the early 1980s. The first article is by developer Brans in French in 1982, but Brans and Vincke explained it the same year in English [58]. The method is a ranking one, where PROMETHEE I yields a partial ordering of actions/alternatives and version II a full ordering. In a more recent article [59], further versions are mentioned. The ordering can be depicted in a directed dominance graph. PROMETHEE uses fewer variables and avoids some of the vague concepts of ELECTRE that deal with intransitivity and the pseudo-criterion. One distinguishes preference P of a over b, written as aPb, and indifference I as aIb. The latter means that pair values are equal, hence intransitive, but in reality, that is rarely the case. Therefore, the intensity of preference of action a over b is introduced as six functions varying from the strict step criterion with value 1 to a variety of less clear situations in which the difference of criterion values of a and b is modeled in different ways and with different thresholds, even as a Gaussian transition. Together these functions are judged to cover reality adequately. The overall preference is obtained by summing the outcomes of the various functions. Applications of PROMETHEE cover the same type of decision problems as ELECTRE. In [59] references to more than 200 papers are given. Wolters and Mareschal [60] treat sensitivity analysis of PROMETHEE. This can be useful when uncertainty exist about the right value of the criterion importance or that of an alternative, or to find out between what limits the ranking is stable. A method is described how to conduct such analysis for the previous additive-type of MCDM methods.

ORESTE

Roubens developed ORESTE (Organisation, Rangement Et Synthèse de données relationnelles) and published in 1979 in French and in 1982 in English [61]. Pastijn and Leysen [62] a few years later explained the method in a practical way. Just like with PROMETHEE, the incentive to develop the method was to avoid determining ELECTRE's criteria weights. In ORESTE these are replaced by ranks. Again, it is about ordering a set A of alternatives versus a set C of criteria. Also, here two steps: first a complete preordering is achieved (preference P and indifference I), and subsequently in a conflict analysis some parts are removed due to incomparability R. The

preordering is also in steps. Experts give the criteria a weak order in terms of aPb or aIb, (or in a different notation $a \succ b$ and $a \sim b$ resp.) and based on the criteria and the alternatives. Next, the results are written as Besson ranks, which e.g., at an input of [1,2,2,3] due of the 'tie' between 2 and 2 would produce the ranking [1,2.5,2.5,4]). This will yield a matrix of criteria decreasing in importance versus alternatives decreasing in desirability (Example in Fig.3 *left*). The global alternative preference scores will be based on projection of this matrix on its principal diagonal in which left is better. Relative distances D to an arbitrary origin on the better side are determined such that if for criterion j holds aP_jb then $D_j(a) < D_j(b)$; the same is true if a for criterion c_1 is equal to b for c_2 and c_1Pc_2. The projection may be performed linearly (orthogonal or oblique), and non-linearly.



	Most	Criteria importance			Less
	1.5	1.5	3	4.5	4.5
Better	1	2.5	2.5	4	5
Alternatives	2	1	5	3.5	3.5
	3	5	1	2	4
	5	3	4	1	2
Worse	2	1	5	3.5	3.5

Figure 4 *Left:* ORESTE example showing ranks; *Right*: Orthogonal projection on the position matrix of the example adapted after [62]. For further explanation, see text.

In Figure 4 the linear orthogonal projection is shown. In agreement with the above inequalities, for example, makes the 4th position in the 1st column equivalent to the 3rd position in the 2nd column. Hence, for this type of projection, the substitution rate T between criteria ranks and action/alternative ranks is $T = \frac{\partial D_j/\partial rc_j(a)}{\partial D_j/\partial r_{\ j}} = 1$, in which rc_j is the rank of the criterion and $rc_j(a)$ the rank of the alternative. It means that loss of a rank position of an alternative is made up by the increase of importance of a criterion by one rank. For the other types of projection, we refer to [62]. The PIR structure of ORESTE makes it according to Liao et al. [63] more reliable. These latter authors presented references which subsequently made method improvements and introduced a sophisticated type of fuzzy set (hesitant fuzzy linguistic set), where earlier a traditional fuzzy set was applied [64].

TOPSIS

Yoon and Hwang developed TOPSIS (Technique for Order Preference by Similarity to Ideal Solution) at the Kansas State University in 1980 [65]. They followed a concept already mentioned by Zeleny [26]. A selected alternative should have the shortest distance to a positive ideal solution and the longest to a negative-ideal one. It starts with a matrix of columns i of M alternatives versus rows j of N criteria, which have weights attributed to them. Preferences x_{ij} of each alternative with respect to each criterion are expressed in a $m \times n$ matrix. In its simplest

form the evaluation goes as follows: Criteria preference figures should be normalized by computing $r_{ij} = x_{ij} / \sum_{i=1}^{m} (x_{ij}^2)^{1/2}$. In each column the elements r_{ij} are multiplied with the corresponding criteria weights yielding $v_{ij} = w_j r_{ij}$ values. Next, as ideal solution a vector A^+ is composed of the highest of elements ${v_i}^+$ of the j columns representing benefits and the lowest of the conflicting cost column, and as negative ideal one A^- the lowest benefit elements ${v_i}^-$ and the highest cost. Distances to the ideal and negative ideal are derived by calculating for each row ${S_\mathrm{i}}^+ = \{\sum_{j=1}^{N} (v_{ij} - {v_j}^+)^2\}^{1/2}$ $i = 1, 2, \ldots, M$, and ${S_i}^- = \{\sum_{j=1}^{N} (v_{ij} - {v_j}^-)^2\}^{\frac{1}{2}}$ $i = 1, 2, \ldots, M$. Finally, the relative closeness for each alternative to the ideal solution follows as ${S_i}^- / ({S_i}^- + {S_i}^+)$. The largest is the best alternative and the smallest the worst. The results can be shown graphically in 2-dimensional objective space, see Figure 5.

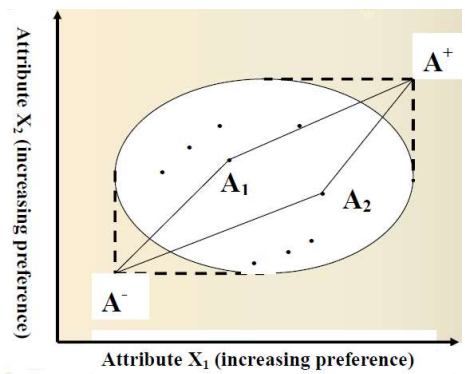


Figure 5 TOPSIS 2-attribute illustration of the selection optimization problem as proposed by Yoon and Hwang [65]. Attributes X_1 and X_2 are criteria, A^+ and A^- the positive and negative ideal solutions, and A_1 and A_2 examples of alternatives that are difficult to distinguish.

Later, in [66] the method was extended to a truly MCDM, in which a multiplicity of objectives that are non-measurable or comparable could be optimized, as explained in Section 2. In that case distance functions to the ideal and its opposite are formulated with variable exponent 1, 2, up to infinite. Then, the lowest distance to ideal and the largest distance to its opposite are chosen. These can only be partially optimal simultaneously. The portions are modeled by triangular fuzzy sets and a max min -solution sought, which represents the optimal satisfactory level.

TOPSIS has also been given a fuzzy version [67].

VIKOR

VIKOR has been developed by Opricovic in Belgrad in 1998. The method also seeks a compromise of feasible solutions, and allows non-commensurable and conflicting criteria. The one closest to the ideal will be the best. The acronym VIKOR stands for the Serbian "VIsekriterijumska optimizacija i KOmpromisno Resenje" (Multicriteria optimization and Compromise Solution). In [68] Opricovic and Tzeng compared VIKOR with TOPSIS. It is again about an $m \times n$ matrix of a columns i of M alternatives and rows j of N criteria with elements

f_{ij}. Also, VIKOR applies the L_p space metric borrowed from Zeleny [26], [69], when Oprovic worked with Duckstein in 1980 [70], and which concept is used in [65] as well: $L_{p,i} = \left\{\sum_{j=1}^{M}[w_j(f_j^+ - f_{ij})/(f_j^+ - f_j^-)]^p\right\}^{1/p}$, $1 \leq p \leq \infty$; $i = 1, 2, \dots, N$, where w is criterion weight (e.g., determined with AHP), and the + and – sign are indicating maximum and minimum. In this equation $(f_j^+ - f_{ij})$ has to be replaced by $(f_{ij} - f_j^-)$ in case, e.g., cost or other non-beneficial conflicting criteria are considered.

So far, the computation is analogue to that of TOPSIS, the difference is in establishing the ranking (see also [71]). For that the extremes of p are substituted: $L_{1,i} = S_i$ and $L_{\infty,i} = R_i$; next, $Q_i$values are derived: $Q_i = \nu((S_i - S_i^-)/(S_i^+ - S_i^-)) + (1 - \nu)((R_i - R_i^-)/(R_i^+ - R_i^-))$, $0 \leq \nu \leq 1$, and ν is called "weight of the strategy of majority of criteria" and has default value 0.5. The alternatives are now ranked in decreasing order in three rows according to the values of Q, S and R. The best choice will be the one with the smallest value, provided that the difference of value of the second best and the best is larger than $1/(M-1)$, while as a second condition is that the best choice Q_i should also be the best of S_i and R_i. Then, for decision making at $\nu = 0.5$ a stable compromise solution (maximum utility of the group represented by min S, and a minimum of individual regret by min R) is obtained and there is consensus on the criteria and the weights. There will be majority expert voting in case a value $\nu > 0.5$ is needed and a veto with $\nu < 0.5$. Also, a sensitivity analysis of the weights can be carried out. If the second condition is not met, the first two alternatives are best, while all Q_i alternatives are indifferent if the difference of the last and the first is smaller than $1/(M-1)$. Chatterjee and Chakraborty [29] dig deeper and describe a number of variants. Oprovic and Tzeng [72] indicated how to perform weights stability interval and trade-off assessment and compared VIKOR with PROMETHEE and ELECTRE and found similar results. A fuzzy version of VIKOR is described in [73].

6. DEMATEL (with BN) and ANP

DEMATEL

The Decision-Making Trial and Evaluation Laboratory (DEMATEL) method [30] is interesting, because it identifies dependent factors, which undoubtedly are frequently present in the MCDM problem. After many years of modest use, there has been over the last decade a strong increase in safety and risk management related articles applying DEMATEL (in 2007, 2 articles; in 2019, 41 [74]). Si et al. [75] reviewed 218 articles on DEMATEL methodology and applications.

The method is as follows: When considering a complex system in which many factors determine an outcome, once having the factors identified $\boldsymbol{F} = (F_1, F_2, \dots, F_n)$ experts can be asked to rate the mutual influence of factors on a Likert scale 0 - 4 from no influence to very strong one, or if desired on a wider more refined scale. Because the weight of experts judging importance is not the same, weighting factors can be derived by applying AHP or even better BWM [74]. The results of each of the K experts are collected in K square $n \times n$ matrices $Z_k = \left[x_{ij}^k\right]_{n \times n}$, with rows i collecting the influences of the factors j on i. Then, elements are aggregated by averaging

to $z_{ij} = \frac{1}{K}\sum_{k=1}^{K} x_{ij}^{k}, \quad i,j = 1,2,\dots,n$, yielding the direct influence matrix Z. The next step is to normalize Z to X by calculating $X = Z/s$, where $s = \max\left(\max_{1\le i\le n}\sum_{j=1}^{n} z_{ij}\,,\ \max_{1\le j\le n}\sum_{i=1}^{n} z_{ij}\right)$.

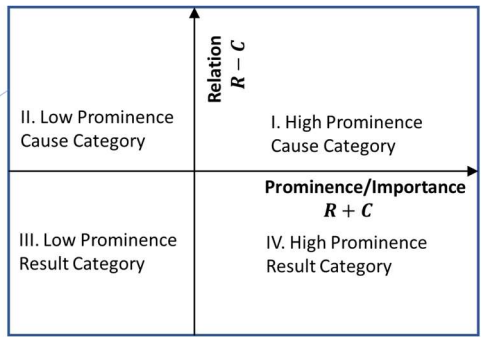


Figure 6 The DEMATEL cause-effect influence relation diagram [76].

To take account of direct but also indirect influences of each factor on any other factor, a comparison is made with the state transition matrix process in a Markov chain multiplying in each step existing states with corresponding transition probabilities (for illustration, see [76]). After each multiplication step influences decrease. Hence, the total influence or T-matrix is computed as $T = X + X^2 + X^2 + \cdots + X^h$, with $h \to \infty$ to the limit value. Analogue with the geometric series formula written as $\sum_{h=1}^{\infty} x^h = x/(1-x)$, the matrix multiplication yields $T = X/(I - X)$, where I denotes the identity matrix. The sums of the rows of T contain the influences of factor F_i on other factors: $R = \sum_{j=1}^{n} t_{ij}$, while the influences F_j receives from all others is contained in the columns: $C = \sum_{j=1}^{n} t_{ij}$. These two quantities can be added $R + C$ representing all influence and called *Prominence*, and $R - C$, the net effect of influence to others, called *Relation.* Results can be plotted in the four quadrants of an influence map shown in Figure 6.

A causality structure of a system can best be described by a Bayesian Network (BN) [75], [76]. So, following [73], [77], or [78] in case of a probabilistic problem the quadrant I (and II) causal factors can in principle be parent nodes in a Bayesian Network (BN) representing factors not influenced directly by others, but influencing child nodes. BNs nodes connect to each other by directional arcs representing a probabilistic cause-effect structure, for safety applications see, e.g., [25, Chap 7] or [79]. However, BNs must be acyclic and are evaluated by determining the joint probability of variable nodes. In case observed data, algorithms are available to sort out the most probable BN structure, e.g., [80]. Here, though, it concerns expert opinions on each other influencing factors and although DEMATEL results indicate causality, the transformation is not fully straightforward. The indirect influences may cause cycles, which based on their insight experts should remove by setting a threshold $(R - C) =$ constant in the influence map of Figure 6. As a first attempt the threshold is set as the average of the T-matrix values. There may also be factors that are correlated by a latent common cause, while an unavoidable cycle may be solved by a creating a Dynamic BN.

After experts have reviewed the BN structure, parametric values for the node variables are derived. The weighted mean of parent factors and the variance in the responses of the experts

enable, represent the resulting of the effect on a child by a ranked node as described by Fenton and Neil [81]. The BN can now be evaluated applying available software (e.g., AgenaRisk [81] or GeNIe [82]) via an underlying truncated normal distribution fit. For a further validity assessment of the expert responses-based model, [83] proposes two types of sensitivity analysis. The first one applies the sensitivity module in the software producing a tornado plot, while comparing the results with matrix values obtained from the T-matrix by summing the effect of factor i on factor j and the other way around in which i has the value of previous j, and j of previous i. The second consists of determining the sensitivity to interventions.

DEMATEL has also been combined with other methods, e.g., [85] with interval type-2 fuzzy set for investigating factor influences in a knowledge management case, while [86] describes an extension with a rather new Pythagorean fuzzy set method [87] applied to probabilistic process safety analysis.

ANP

The Analytic Network Process (ANP) [36], [38] has been derived by Saaty from the earlier AHP. Where AHP is top-down in a linear two-layer hierarchy of criteria and options. ANP consists of a network of each other dependent decision clusters representing individual (organizational) units with mutually influencing decisions, and hence solving for an optimum is non-linear. Through AHP in each cluster, a matrix of criteria versus options can be formed, while the whole is brought together in a supermatrix of matrices containing all the judgments. For achieving the result, Saaty [36], in analogy to DEMATEL, points out that by raising the supermatrix to high powers at the limit, a steady state of priorities is achieved. In case of an investment study, Lee et al. [84] applied DEMATEL in combination with ANP to sort out with the former how factors influence each other within a cluster and with the latter to combine the clusters by means of a supermatrix in which the clusters are weighted.

Khakzad et al. [88] applied ANP for an assessment of security of hazardous facilities.

7. MAUT (or MAVT), TODIM

MAUT

Utility (or Value) is a concept introduced in the middle of the 19th Century by British philosophers Jeremy Bentham and John Stuart Mill. It has become important in economics in view of investment and risk. Based on the work of many scientists the Multi-Attribute Utility Theory, (MAUT) developed a decision tool. Important, in addition, has been the 1947 von Neumann-Morgenstern [89] expected utility concept, which expresses that given two risky options (L and M, called 'lotteries') a rational person will choose the option maximizing the consequence according to the person's utility function. Four conditions (axioms) have been formulated: *Completeness*, i.e. three possibilities exist: either L or M is preferred, or the decision maker is indifferent about the two; *Transitivity*, if L is preferred over M and M over N then L will be preferred over N; *Continuity*, which means that in case of three options there is a probability p at which M is indifferent to pL and $(1-p)$N; and *Independence*, i.e., the preference situation is not changed by a change of an unavailable third option. Another breakthrough has been that in 1954 in the preference of a rational person, Savage [90]

distinguished subjective probability, and hence, subjective utility. This led to his *normative* decision theory. In 1964 Pratt et al. [91] published the foundations of decision under uncertainty in the context of expected utility. A person facing the choice of a 'lottery', is calculating chances from a Bayesian point of view. Pratt et al. [91] added as a condition that the person is consistent and called it a *constructive* approach. In 2016 Shafer [92] reinforced this view arguing that in the people's mind Savage was taking preference too much first and the subjective, hence Bayesian, probability second. She turned it around into the *constructive decision theory*: ahead of decision constructing a belief function on consequences and weighting goals.

According to Fishburn [89], *independence* is the most important condition assumed to hold in MAUT in order to enable linear transformation of elements, which allows to write an addition of utilities as a multiplicative formula. Fishburn [45] has been much engaged in the mathematical aspects of utility and published his foundations of the theory for decision making in 1970. It deals largely with the concept of a person's utility (consequence preference, order, additivity, probability and uncertainty, expected utility, a lottery/gamble, e.g., to win a large amount with probability p or a less amount for sure, i.e., the certainty equivalent, utility dependence on present wealth) and it explains many of the terms concerning preference relations shown later in the mathematics of MCDM.

Applications followed, e.g., Keeney and Wood [94] selecting one of five possible system alternatives/options for a water resource problem dealing with 12 attributes (or terms in other references: selection criteria, fundamental objectives, or aspects) X, covering many aspects from cost to flexibility, their estimated range of best to worst expressed in the attribute unit or in a subjective interval of 0-100. Next, the *preferential* independence of the attributes is investigated by judging indifference at some value trade-off of a pair of attributes at any fixed level of other attributes. Then, *utility* independence of attributes is verified by judging which value within the range of each attribute is indifferent to either 50% chance of the best, or 50% of the worst of the range and verifying this "certainty equivalent" (expected value or mean) will not change with the level of the other attributes. This "lottery" is repeated for this latter certainty equivalent value and the best value of the attribute. If still no change, utility independence is ascertained. Finally, two intuitively indifferent attributes are selected (here, 'water shortage' and 'flood protection'), and a 50-50 "lottery" on two options performed, either both attributes in best or both in worst state, or the other option one in best and one in worst. Because the latter was judged to be the preferred option. It implied *additive* independence, so that the sum of the 12 attribute utilities can be written as an additive equation, which by a transformation can be converted in a multiplicative one, Keeney [95]: $1 + ku(x_1, x_2, \dots, x_{12}) = \prod_{i=1}^{12}[1 + k\, k_i u_i(x_i)]$, where utilities $u(x)$ are scaled between 0 and 1, the scaling constants k_i are positive but less than 1 and constant $k > -1$. Then the largest k_i is determined by setting all attributes on worst level and finding out, which one would be preferred when setting to best. This is the largest k_i and appears to be 'water shortage'. Following the procedure to setting one to best level and finding indifference in consequences with another attribute enables ordering k_i. By selecting indifferent pairs of attributes and applying on each of 11 pairs the multiplicative utility equation, ratios of their respective k_i values are found. One further equation is needed to find the real scale k, and this is accomplished by again selecting 'water shortage' and 'flood protection' and by

performing a lottery. According to the decision maker's subjective judgment, the state of first best and the second worst is indifferent with a probability p (here, 0.6) to the situation of the two states best versus $1 - p$ of the two states worst. Then, by applying the above multiplicative equation the value of k is found and the order of option utilities be derived (in terms of ref. [120] this is the impact model). The definition of utility functions of each attribute is by estimating a few points over the attribute interval span and fitting those with a linear or exponential function. The utility value of each option can now be calculated. This whole procedure can be followed by a sensitivity analysis in which trade-offs, and hence k_i values are varied. The method hinges on quite a few judgments, while the independence conditions cause constraints. Improvements of the method would be to sharpen objectives, include uncertainty margins, better utility definitions, and taking account of changing conditions (dynamics).

More applications followed, e.g., on engineering [96] and [97] for some recent ones and [98] on inherent safety design decisions, but matters also became more complex. The 1977 engineering application [99] mentioned that depending on the character of the decision maker there can be risk neutrality, risk proneness, and constant or decreasing risk aversion. In 1979 psychologists Kahneman[†] and Tversky [31] published their renowned paper on Prospect Theory, which contained a strong critique on the expected utility theory and with which they were awarded the Nobel Prize. A few years later this paper was followed by [100] in which they claim that people's notion of probability of a risky event is not what the abstract mathematical utility theory assumes. If there is a chance to gain a certain amount say $1000 against a certain prize, one is easier prepared to go for it if one possesses only $4000, than in case of $40,000. Also, people are more sensitive to a certain amount of loss than to the same amount of gain. This leads to the schematic value function of Figure 7. Therefore, in case of a risky decision, people will be rather risk averse, i.e, they are not prepared to pay proportionally more for a larger return. On the other hand, they are easier willing to pay proportionally more to reduce their loss, hence there is certain risk appetite.

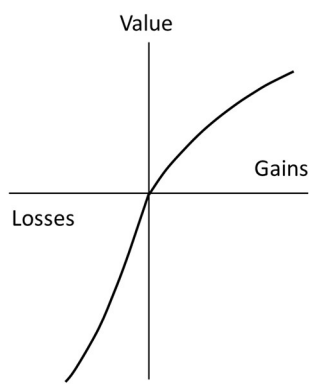


Figure 7 Utility/value function reproduced according to Kahneman and Tversky [31]: concave with risk aversion behavior in the positive quadrant; convex and steeper with risk propensity in the negative one.

[†] For this work Daniel Kahneman received the Nobel prize in Economics in 2002, 6 years after the death of Amos Tversky (the prize is not awarded posthumously)

Wang et al. [101] applied the prospect theory and MCDM on failure modes and effect analysis (FMEA) for the evaluation of risk factors of which values are aggregated using fuzzy measure and the Choquet integral. It all serves to determine FMEA's RPNs (risk priority numbers) and prioritize failure modes of. Grabisch [102] reviewed in the context of MAUT various aggregation approaches including the Choquet integral for fuzzy measures such as membership functions μ. If there exist functions $f : X \to [0,1]$ with respect to μ, then the Choquet integral $C_\mu(f(x_1), \dots, f(x_n)) := \sum_{i=1}^{n}(f(x_{(i)}) - f(x_{(i-1)}))(\mu(A_{(i)})$, where $f(x_{(0)}) = 0$ and $\mu(A_{(i)})$ represents the importance weight of the set of criteria $A_{(i)}$.

Huang et al. [103] note that decision makers do not need to decide groupwise on preferences but can make their choices individually, while the results can be aggregated by means of the SAW method. So, it is even possible to perform a MCDM applying internet communication means.

In 2004, Chavas [104] treated decision making under risk comprehensively, expanded the expected utility hypothesis with some further assumptions, and described the mathematics for various cases of risk appetite, neutrality, constant decreasing, and increasing risk aversion.

TODIM

As mentioned by Gomes and Rangel [32], Gomes and Lima developed the MCDM method TODIM (Interactive and Multicriteria Decision Making in a Portuguese acronym) in 1992. The method applies the value function of the prospect theory [31] and the additive utility function [95] requiring verification of independence of attribute options. Further, the method uses pairwise comparison of criteria, enables elimination of inconsistencies, allows linguistic grades, fuzzy inputs and interdependencies of options, but no trade-offs. The method has been characterized as having on the one-hand American MAUT features and on the other, French ones of ELECTRE or PROMETHEE.

Again, there are m alternatives and n criteria[‡]. Experts weight the criteria c, after which weights w_c are normalized, the highest value criterion indicated as the reference one, and all weights divided by the reference, w_{rc}. Next, the dominance δ of alternative A_i with performance P_{ic} over alternative A_j with P_{jc} is determined as: $\delta(A_i, A_j) = \sum_{c=1}^{n} \Phi_c(A_i, A_j)\ \forall\ (i,j)$. Applying the value curve of Prospect Theory on the terms $\Phi_c(A_i, A_j) = 0$, if $P_{ic} - P_{jc} = 0$, or $\Phi_c(A_i, A_j) = [w_{rc}(P_{ic} - P_{jc}) / \sum_{c=1}^{n} w_{rc}]^{1/2}$, if $P_{ic} - P_{jc} > 0$, and $(-\frac{1}{\theta})[(\sum_{c=1}^{n} w_{rc})(P_{jc} - P_{ic)}/w_{rc}]^{1/2}$, if $P_{ic} - P_{jc} < 0$, where $\theta$is an attenuation factor controlling the shape of the loss curve.

After calculation for each criterion the square matrix A_{ij}, the overall matrix shall be calculated by summing and normalizing by $\xi_i = \frac{\sum_{j=1}^{m} \delta(A_i,A_j) - \min \sum_{j=1}^{m} \delta(A_i,A_j)}{\max \sum_{j=1}^{m} \delta(A_i,A_j) - \min \sum_{j=1}^{m} \delta(A_i,A_j)}$. The last step is ordering to rank. A sensitivity analysis is recommended with respect to the reference criterion, weights, the attenuation choice, and the alternative performance ones.

TODIM knows fuzzy applications too, for examples [105], [106].

[‡] In contrast to [40] we follow the conventional matrix notation of $m \times n$, and not $n \times m$.

8. Pareto Front Optimization

To complete the array of decision analysis and aiding techniques, the Pareto Frontier method is mentioned in this section and an overview is given on how it works. In its simplest form the frontier forms the envelope around optimal choices of alternatives in case of two objective variables, plotted on Cartesian coordinates, as shown in Figure 8 [25, p. 395]. Hence, it is a Multi-Objective Optimization, often added in the name with Programming to MOOP or MOP, because in particular with multi-dimensional cases the optimizing will become rather intricate.

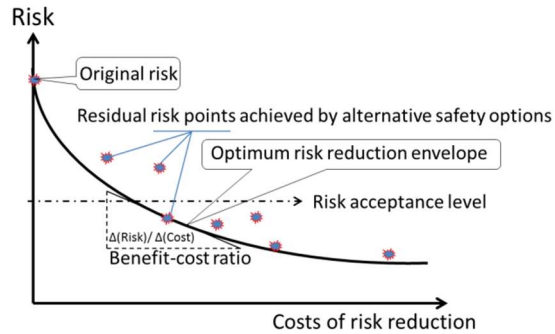


Figure 8 Risk versus costs to reduce it. The envelope is an illustration of a Pareto front, from [25].

Note that the Optimum risk reduction envelope must be an interval or range that includes an estimate of the aggregated *epistemic uncertainty of the data* used to construct the envelope. Otherwise, the Costs of risk reduction below the Risk acceptance level will be underestimated.

An explanation of Pareto Front and a number of two-dimensional examples is provided by Ščap et al. [107]; solutions at the front dominate the ones away from it. Many algorithms to perform optimization have already been proposed. Giagkiozis and Fleming [108] focus on multi-objective evolutionary algorithms. This includes metamodeling and the use of surrogate models to speed up the computational process, Pareto estimation via either dominance-based algorithms or decomposition ones, and Radial Basis Function Neural Network technique to fit and map results. The latter enable optimization, e.g., in 3-dimensional space with three objective variables. An example of optimizing a design of a bridge has been worked out by Pouraminian and Pourbakhshian [109]. These authors applied ANSYS for structural analysis, determined a Pareto front by optimizing using the swarm particle technique and subsequently used VIKOR to determine the optimal point at the Pareto front.

9. Comparison, Discussion and Software

In 2013 Velasquez and Hester [110] presented an overview of MCDM methods, which to some extent emphasized MAUT but in the light of current knowledge was incomplete. A few years later, Saaty and Ergu [111] counted 35 MCDM methods, of which only the most applied ones have been summarized in this paper. Although these authors did not review all methods, which was judged too much of a task, they concluded that different methods applied to the same problem resulted in different answers. No “super-method” existed. They also formulated 16

evaluation criteria, of which the first few are “Simplicity of execution”, and “Comprehensive structure: breadth and depth”. Comprehension is in particular important when factors, such as benefits, opportunities, costs, and risks are influenced from political, social, and technological angles, to name a few.

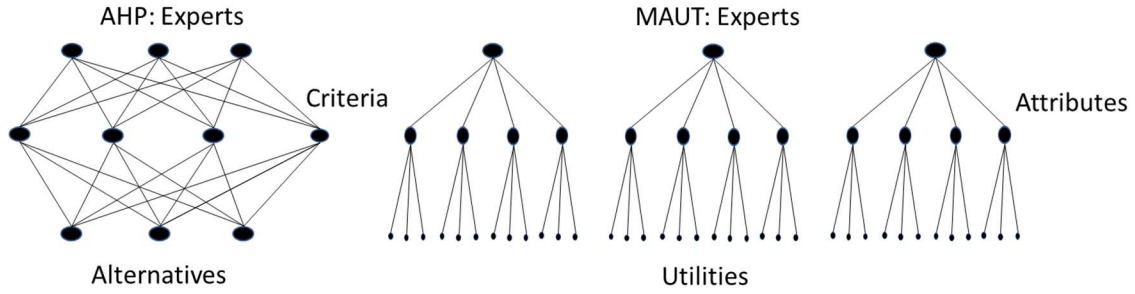


Figure 9 Hierarchical structures of AHP and MAUT respectively, adapted after [112].

Several papers present results of comparisons, a few have already been mentioned above, such as [68], [72], not showing dramatic differences between method outcomes. An interesting comparison between two radically different methods: AHP and MAUT, was made by Belton [112] who showed the fundamental difference between the two approaches. AHP treated importance weights of criteria and scores of the alternatives given a criterion the same way. In contrast, to achieve a certain goal MAUT derives first utility functions of the attributes based on judgment inputs and then determines the utility values of the alternative solutions (options). Belton made it clear by a schematic drawing showing influences of three “experts” on AHP criteria and from those to each alternative, while in case of MAUT utility values determined by the three “experts” yield a series of outcomes. The difference is schematically presented in Figure 9. Outcomes of comparable inputs to the two methods for an example with rather extremely chosen figures showed relatively large dispersion and a certain bias. De Leeneer and Pastijn [113] compared ORESTE with PROMETHEE and a SAW-like method, finding slight, but no serious differences, although of course the additive method is not suited for incomparability.

The possible advantage of the use of fuzzy set theory in decision analysis remains a point of discussion, Dubois [114].

Sensitivity analysis would help to identify major influencing parameters. However, because of its complexity with a method such as ELECTRE it is difficult to perform a sensitivity analysis, see [94].

On the web one can find “decision radar” [115] offering to calculate online for free a problem with TOPSIS, ELECTRE, SAW, Linear Assignment, or AHP. The program uses the term *indicator* for criterion and *choice* for alternative/option. In case of AHP only the consistency is checked and the priority matrix calculated. Linear Assignment is a cost minimization for a number of tasks to be assigned to the same number of agents (balanced) or a different number (unbalanced), where each agent has a different cost for the task assigned. So, it can be applied as a decision-making method. The system can be solved by linear programming. There are also

software toolkits with a license for sale for ELECTRE I and III [116] and ELECTRE III and IV [117]. PROMETHEE software can be freely downloaded [118], while VIKOR, TOPSIS and DEMATEL, including their fuzzy versions, can be solved online [119]. The International Society on MCDM [120] offers an even much broader variety of software, among other online MAUT Decision Navigation ("Entscheidungsnavi"), an MCDA package for R, a MATLAB solver for MOP and more. Finally, there is the Creative Decisions Foundation [121] established by Thomas L. Saaty in 1996 providing for free the educational software Super Decisions that assists step-by-step in applying the AHP and ANP methods.

10. Results of Calculations

A simple example has been evaluated, inspired by a classical process safety problem of selecting the best option to protect against the effects of a thermal runaway of a closed batch reactor: a pressure relief valve (*PRV*), a *bursting disk* or a *dump* tank. The selection criteria are *availability*, *clean-up* effort after a runaway, effectiveness of *risk reduction* and *capital costs* of the investment. Suppose a panel of experts comes to the conclusion that the ratio of importance of these criteria is: 1 : 0.333 : 10 : 2, while their preferences of the options with respect to each of the criteria are collected in Table 2. (These input judgments are debatable and, of course, depend on local conditions and opinions).

Table 2 Expert input of option preference for each criterion relative to a PRV

	Availability	Clean-up	Risk red.	Cap. Cost	Rank
PRV	1	1	1	1	3
Burst.Disk	3	0.25	2	0.1	2
Dump	1.2	0.666	3	3	1

Calculations have been performed according to AHP process, SAW-WSM and WSP using MS Excel, of which details are provided in the Supplementary material, while for all other MCDM methods use was made of the software mentioned in Section 9. Results are presented in Table 3.

Table 3 Ranking results calculated according to various decision aiding methods

Method	AHP	SAW/WSM	WPS	PROMETHEE	ELECTRE	TOPSIS	VIKOR	MAUT	Rank
PRV	0.19	0.19	0.20	-0.8	D > P	0.13	0.14	0.18	3
Burst.Disk	0.30	0.31	0.35	-0.1	D > B	0.37	0.31	0.31	2
Dump	0.51	0.50	0.44	0.9		0.50	0.54	0.51	1

Note the agreement in ranking of the alternatives despite quantitative differences. ELECTRE only provides relative preferences. The quantitative output of TOPSIS (closeness vector) and VIKOR (S group utility) is high when rank is low; the figures were converted by taking reciprocal values and normalizing. MAUT output is through the inflection of the utility curve versus the defined range of the attribute quite sensitive to the extent of willing to take risk, which was assumed here as: inflection up = risk aversion –> availability; straight line = neutral –> risk reduction and clean-up; down = risk appetite –> capital cost.

A recent example of a practical application of AHP and TOPSIS is a Multi-Attribute HAZOP [122] based on a wellhead P&ID. Experts identified 50 hazards and produced fuzzy hazard scenario risk factor inputs of severity, frequency, undectability, sensitivity to maintenance effectiveness, and sensitivity to failure of safety measures. Risk factor weights were determined with the aid of AHP, while in a second step the hazards are ranked with TOPSIS.

11. Conclusions

There is an abundance of methods available to support decisions under uncertainty, which one way or the other can be applied in risk management and risk governance. The methods have been described rather superficially and only a fraction of the literature on these methods has been referenced. The methods have very different footing and are each suited for different situations. They vary from qualitative methods focused on rational reasoning, such as Toulmin's approach, or recalling past similar cases, to quantitative methods. The latter can be probabilistic, such as decision tree, or based on MCDM preference ranking. The importance preference scores are for the greater part based on expert estimates, but where possible measured properties, test outcomes, or calculation results can be source too. Preference data can be processed in matrix operations varying from simple to complex ones; the latter trying to obtain the best of compromises with graphical output that even can be multi-objective. Most methods produce a ranking of options. Some are technologically oriented to obtain an optimum choice; others have an economical/financial objective. Some specialized methods, such as ELECTRE Tri, are dedicated to categorization/classification of options. DEMATEL exposes influences among factors that play a role in a problem field. MAUT is guided by the value persons attribute to their individual utility function. No method constitutes a panacea [43], [111]. In many cases base methods are combined with other techniques, such as fuzzy inputs, aggregation of output, e.g., with SAW, or in the case of DEMATEL further processing of output with ANP, or even better Bayesian Network.

Some methods are worked out on a simple example case. Results are presented for a rather simple case of the best protection option for a reactor, which ma runaway. Appendices in Supplementary material contain details. For most methods described software is available at the Internet.

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Decision Making Using Human Reliability Analysis

Fabio Kazuo Oshiro
Monaco Engineering Solutions Limited
United Kingdom
fabio.oshiro@mes-international.com

Abstract

Brazil Regulatory Agencies have a risk criteria which have been defined based on quantitative risk approach. The most well-known regulations are defined for São Paulo and Rio de Janeiro cities which have a strong relationship with IBAMA (Brazilian Environmental Agency). Their standards establish the activities to be developed defining the use of human reliability techniques to calculate the human error probability. Human errors are the main factors of the industrial accidents and their effects are not being assessed systematically with the level of details that are used during a typical human reliability assessment.

The objective of this paper is to analyse some methodologies of human reliability considering the external (observable) and internal (cognitive) human factors. The calculation of the human failure frequency of QRA studies are being performed by consulting firms in a subjective and conservative manner when compared to failure of equipment analysis.

In this paper, the human error probability was quantified using standardised methods. The study was based on the evaluation of some methodologies of human reliability and decision making. The method was assessed through a case study of an accident occurred in 2004 at Formosa Plastics Corp. Illiopolis. Initially, an analytical method was developed as Hierarchical Task Analysis (HTA), then by Predictive Human Error Analysis (PHEA) and a qualitative analysis using Systems for Predicting Human Error and Recovery (SPEAR). To complete the study a quantitative assessment using Fault Tree Analysis (FTA) and Human Error Assessment and Reduction Technique (HEART) was developed. The recommendations were assessed in two different categories: First, using the Weighed Score Method based on the management point of view and second, through the HEART and FTA methods representing the point of view of the operators.

The paper concludes that results based on operational focus were more objective and transparent when compared to results from management techniques. This is because the operational indicators were easier to interpret and less subjective with less financial concerns involved. Also, the methodologies used provided a thorough understanding of the events in each phase of the accident.

Keywords: decision making, human reliability, accident in Formosa-IL, human error probability, performance factors influence. A maximum of eight keywords should be included.

1 INTRODUCTION

There are numerous studies related to human behavior and each one possesses specific characteristics. Basically, they are differentiated in external (observable) and internal (cognitive) factors and the selection of analytical method depends on the availability of information and the viability of cognitive analysis. Nowadays, in Brazil, the frequency of human error is evaluated in a subjective and conservative way when compared to equipment failure and its quantification, which can be developed using methods that represents the risk closer to reality. This study was based on the evaluation of human reliability and decision making methodologies, followed by a practical application of human reliability assessment of an accident which occurred in 2004 at Formosa Plastics Corp. Illiopolis.

2 METHOLODOGY

The description of the accident which occurred at the Formosa-IL plant was extracted and summarized from the investigation report (Chemical Safety and Hazard Investigation Board, 2007). The plant layout of Formosa-IL is presented in Figure 2.1.

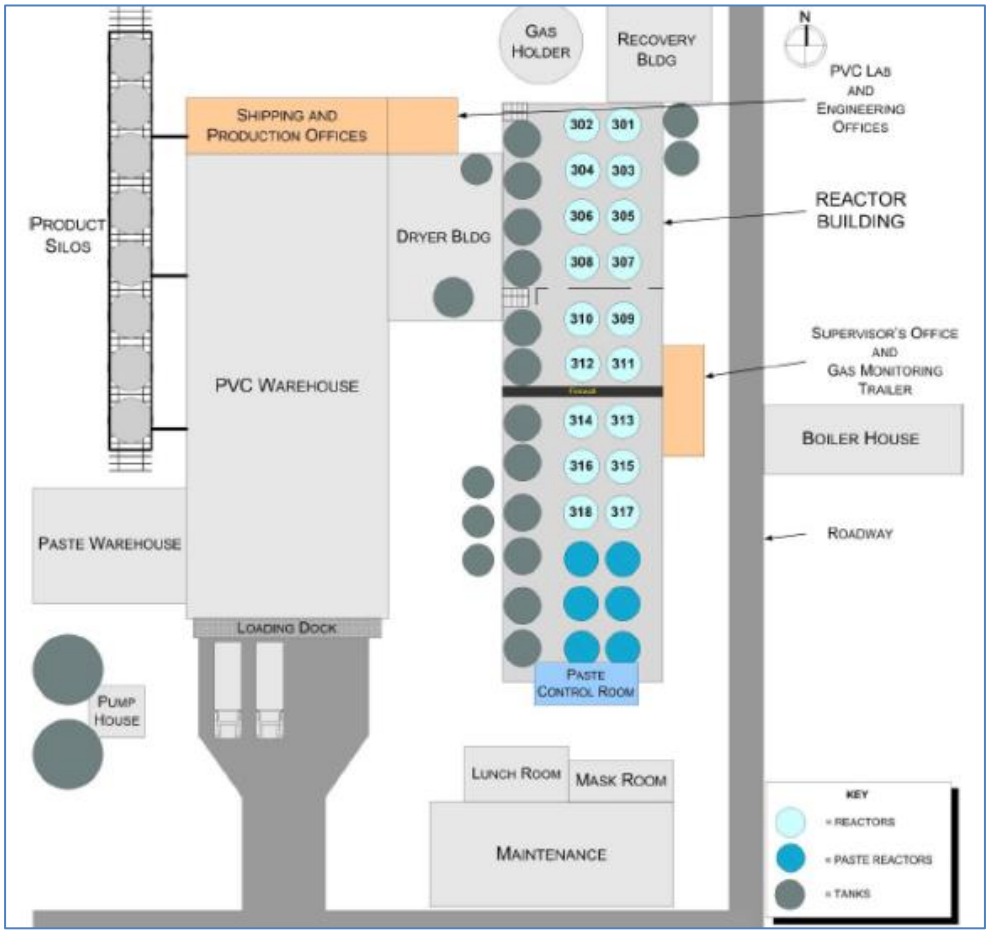

Figure 2.1: Layout of the plant of Formosa-IL (Chemical Safety and Hazard Investigation Board, 2007)

The method for human reliability assessment used in this study is presented in Figure 2.2.

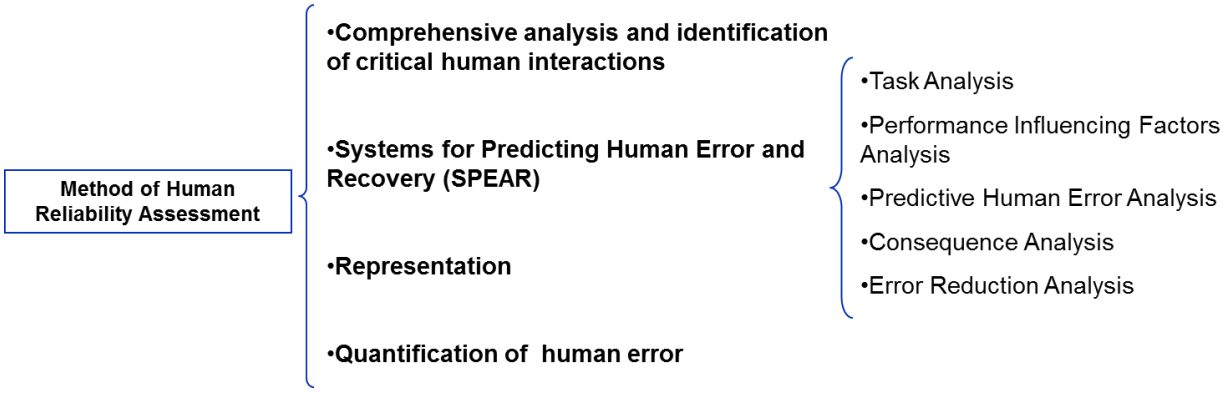

Figure 2.2: Method of human reliability assessment (AICHE/CCPS, n.d.)

The first step of human reliability evaluation consists of general analysis and identification of human interactions. The cleaning of reactors was identified as the critical activity by the Chemical Safety and Hazard Investigation Board (CSB) at the Formosa-IL plant. Normally, before an installation, it is necessary to get information about the most critical operational and maintenance activities directly from the operational team through meetings to stimulate transparent communication about work activities.

The second step consists of the SPEAR methodology application. Initially, the action oriented technique HTA was used in chart and tabular format to represent the activity i.e. reactor cleaning. Following the completion of the task analysis, the Performance Influencing Factors (PIF) analysis was developed in accordance to AICHE/CCPS classification. The last three steps of SPEAR were completed using the Predictive Human Error Analysis (PHEA) where consequences and error reduction analyses were developed. The results were obtained in tabular form preserving the logic between the type of human errors, its consequences and the measures for risk reduction.

The third step of the human reliability assessment, defined as Representation, was developed using Fault Tree Analysis (FTA) and Influence Diagram Analysis (IDA) to represent the accident at Formosa-IL.

In the last step which the human error was quantified, the Human Error Assessment and Reduction Technique (HEART) was used to estimate the probability of human error to quantify the FTA and the IDA developed in the previous step.

3 RESULTS AND DISCUSSION

The human reliability assessment results for the accident at the Formosa-IL PVC plant is presented in this section. The first step of the human reliability assessment consisted of the comprehensive analysis and identification of human interactions, previously identified by the CSB. The second step consisted of the application of the SPEAR and the results are presented in the following section.

3.1 SPEAR (SYSTEMS FOR PREDICTING HUMAN ERROR AND RECOVERY)

3.1.1 Hierarchical task analysis (HTA) of the Formosa-IL accident

The simplified HTA presented in Figure 3.1 was developed to verify the reactor cleaning activity. It does not consider all of the steps of reactor cleaning in detail. A detailed analysis, which requires significant time and effort, should be prepared involving operators and engineers in charge of the

area. The considerations in Plan 0 were extracted from the CSB report while activities considered in Plan 1 were assumed to be more realistic.

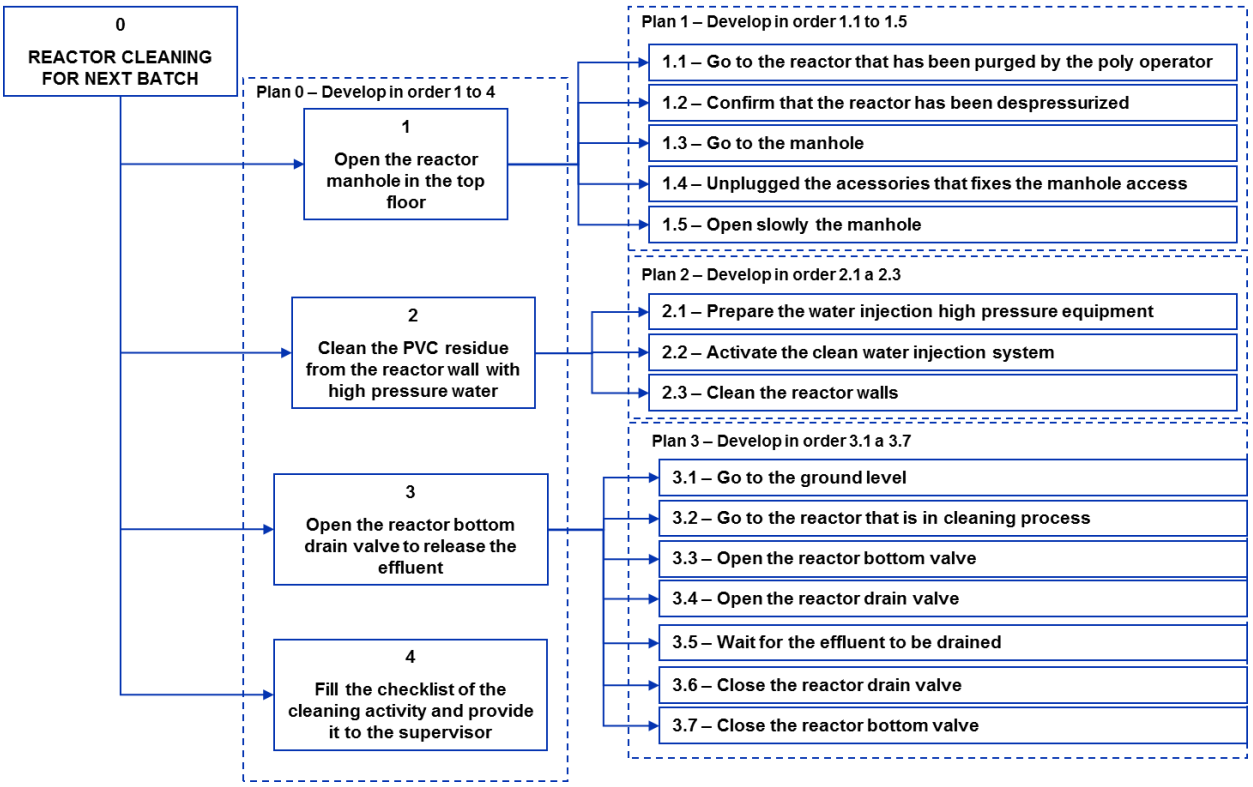

Figure 3.1: Hierarchical Task Analysis of Reactor Cleaning

It is reported in the accident report that the blaster operator went to the wrong group of reactors, this activity corresponds to step 3.2 i.e. go to the reactor that is in cleaning process. The HTA shown in Table 3.1 was developed to detail this step. The CSB report presented the communication system and the location of the reactor only. Therefore, any missing information was considered to be close to reality.

Table 3.1: HTA Table for reactor cleaning

Task Step	Input (registers)	Output (action	Communication	Time and Task dependency	Second function, distraction	Comments
3.2 – Go to the reactor that is in cleaning progress	Identification of reactor tag on reactor bottom and control panel	Operator must check that the reactor tag is in accordance with reactor cleaning progress	By voice, operator on ground level shall go to the other operator to communicate. There is no intercom and radios are not part of routine operation.	Delays in start-up of next batch. Cleaning progress is sometimes not appropriate and should be re-done	Other functions in parallel with cleaning progress	Residual VCM can be released if cleaning process is inappropriate Hazards: operator injuries Operators must use appropriate PPE

The purpose of this exercise was to present some tools to identify where faults were occurring. Ideally, these analyses should be developed for all existing critical activities in an industrial plant.

It is observed in Table 3.1 that improper opening of a reactor during cleaning, whilst it is in operation is not considered as part of the task. The HTA developed follows the concept of the method, it addresses with precision and detail the activities to be performed and not the possible deviations. These should be analysed using alternative tools. The development of HTA allows procedures to be more appropriately defined and training to be more efficient. However, it does not show possible errors that may occur.

3.1.2 Analysis of PIFs

After the task analysis, it is important to evaluate the PIFs of the operators during reactor cleaning. The scale used for assessment of PIFs is shown in Table 3.2.

Table 3.2: Scales used for assessment of PIFs of reactor cleaning

Rating Scale PIF	Procedure	Physical Work Environment
Worse – 1	• There are no written procedures or standards for implementation of activities. • Not integrated with training.	• High level of sound • Poor lighting • High or low temperatures, high humidity or high winds
Average – 5	• Written procedures available, but not always used. • Standardized methods to perform the task.	• Moderate levels of noise • Temperature and humidity variables
Better – 9	• Detailed procedures and checklists available. • Procedures developed using analysis task. • Integrated with training.	• Noise levels at optimal levels • Lighting based on analysis of the task requirements • Temperature and humidity at optimum levels

The list of standard PIFs was used to identify factors that could influence the reactor cleaning activity. The list is not a formal definition of PIFs, and depending on the activity, this list should be developed and reviewed by the plant analysts. This evaluation was based on the descriptions of the accident presented by the CSB. Numerous deficiencies were commented and considered in the evaluation. Factors with value of 5 were considered relevant to the study although no information was found in the accident report (Chemical Safety and Hazard Investigation Board, 2007).

- Operating environment
  - Weather: 5
  - Illumination: 5
  - Working hours and breaks: 5
- Work details
  - Place/access: 3
  - Identification:2
  - Displays and controls identification: 3
  - View of critical information and alarms: 3
  - Clear the instructions: 1
  - Quality of controls and warnings: 1
  - Grade of support of diagnosing fault: 1
  - Conflicts between safety and production requirements: 2
  - Training for emergencies: 1
- Characteristics of the operator
  - Skills: 5
  - Risk assumption: 5
- Social and organization factors:
  - Clarity of responsibilities: 3
  - Communications: 1
  - Authority and leadership: 2
  - Commitment of management: 2
  - Overconfidence in technical safety methods: 2
  - Organizational learning: 1

The assessments were conducted after the investigation of the accident once the faults had already been analysed. If this assessment were performed before the accident, judgement would most likely be different and higher notes would be obtained.

The results demonstrate that there were mainly deficiencies in group task characteristics and organizational and social factors. Within the group task characteristics, specific categories such as clarity of instructions, quality of checks and warnings, degree of support on fault diagnosis presented the worst reviews. These deficiencies could have occurred due to the absence of a supervisor allowing for a hierarchical distance and lack of communication between operations and management. Emergency procedures training were considered the most critical as the effects of the accident would be very different if the operators were adequately trained in evacuation procedures. In the group organizational and social factors, specific categories communications and organizational learning presented the worst results, although authority and leadership, commitment of management, overconfidence in technical safety methods were also considered critical.

These reviews can be justified mainly because there was already evidence of criticality of the bypass procedure of the safety interlock and no effective modification was performed. Also, there was no routine of communication such as radios and intercoms, nor adequate availability of supervisor, and such evidence were not considered by management.

3.1.3 Predictive Human Error Analysis (PHEA), Consequences and Error Reduction

Table 3.3 presents the PHEA methodology that analyses human error and cognitive perspective developed to assess the step of task 3.2. The information was extracted from the CSB accident report, but the logic was developed using the methodology. It is observed that the analysis of consequences of SPEAR is associated with each type of human error defined by PHEA. This way, the consequences are associated with a cause that was identified during the study. A strategy to reduce the error should be developed based on the consequence, that depending on the criticality of the consequence should be mandatory or not.

Table 3.3: Human Error Analysis (PHEA) of the reactor cleaning activity (step 3.2)

Task Step	Type of task	Type of error	Description	Consequences	Recovery	Strategy to reduce the error[1]
3.2 – Go to the reactor that is in cleaning progress	Action	Action in the wrong direction	Move in the wrong direction of the right reactors	Operator will be in the wrong group of reactors	Reactor identification at the bottom of reactor and control panel	Optimize layout of the reactors in order to facilitate identification
	Action	Right action on wrong object	Operator performs bypass of interlock system and drains the reactor in operation	Large release of vinyl chloride monometer (VCM) followed by explosion and fire	None	- Evacuation System - Study of protection layers - Historical analysis - Improve procedures and training
	Action	No action	Absence	Delay in drainage	None	
	Action	Omitted action	Absence	Delay in drainage	None	
	Checking	Omission of checks	Operator does not check the reactor identification that should be drained	Impossibility to drain reactor due to interlock activation	Indication of interlock activity in the control panel	Include in checklist the activity verification of reactor to be drained
	Checking	Right check in the incorrect object	Blaster operator confirms that the reactor is in cleaning process, but is on the wrong reactor	Impossibility to drain reactor due to interlock activation	Indication of interlock activity in the control panel	Include in checklist the activity verification of reactor to be drained
	Checking	Wrong check in the correct object	Blaster operator is in the correct reactor but confirms that another reactor is in cleaning process	Operator goes to another reactor and will not drain it due interlock activation	Operator of the upper level will fix the blaster reactor	Improving procedures and training
	Checking	Wrong check in the wrong object	Blaster operator is in the wrong reactor and confirms that another reactor is in cleaning process	Operator goes to another reactor and will not drain it due interlock activation	Operator of the upper level will fix the blaster reactor	Improving procedures and training
	Recovery	No information	Blaster operator has no confirmation about which reactor is in cleaning process	Operator will be in the wrong group of reactors	Operator will go to the upper level and verify which reactor is in cleaning process	

[1] – Strategies to reduce the error should be related mainly to changes in procedures, training, equipment and design.

The results of PHEA allow the main PIFs contributing to the risk to be analysed. Table 3.4 shows the PIFs related to types of errors evaluated in PHEA.

Table 3.4: Identification of the most critical PIFs during cleaning reactor activity

Type of error	Performance Influencing Factors (PIFs)
Action in the wrong direction	Distraction, practices with unfamiliar situations or poor identification
Right action in the wrong object	Distraction, poor identification, poor lighting, identification of displays and controls or poor communication
No action	Practices with unfamiliar situations or working hours and breaks
Omitted action	Practices with unfamiliar situations, working hours and breaks or distraction
Omission of checks	Distraction or poor communication
Right check in the wrong object	Distraction, poor identification, poor lighting, identification of displays and controls or poor communication
Wrong check in the right object	Distraction, poor identification, poor lighting, identification of displays and controls or poor communication
Wrong check in the wrong object	Distraction, poor identification, poor lighting, identification of displays and controls or poor communication
No information	Poor communication or poor authority and leadership

The interlock pressure system theoretically prevents an undue drainage on the various operator errors on by-pass activation. It is the last protective barrier of the preventive system. However, its actual efficiency should be further evaluated through studies of LOPA.

The list of possible error types with PIFs demonstrates that factors such as distraction and working hours and breaks contribute directly to errors related to the operator's physical state. Identification of displays and controls, poor identification and poor lighting are related to visual factors that influence the decisions of the operator. Poor authority, poor leadership and poor communication refer to organizational policy and practices whilst unfamiliar situations refer to operator experience.

3.2 REPRESENTATION

3.2.1 Fault Tree Analysis

The fault tree analysis representing the development of the accident was developed and is shown in Figure 3.2.

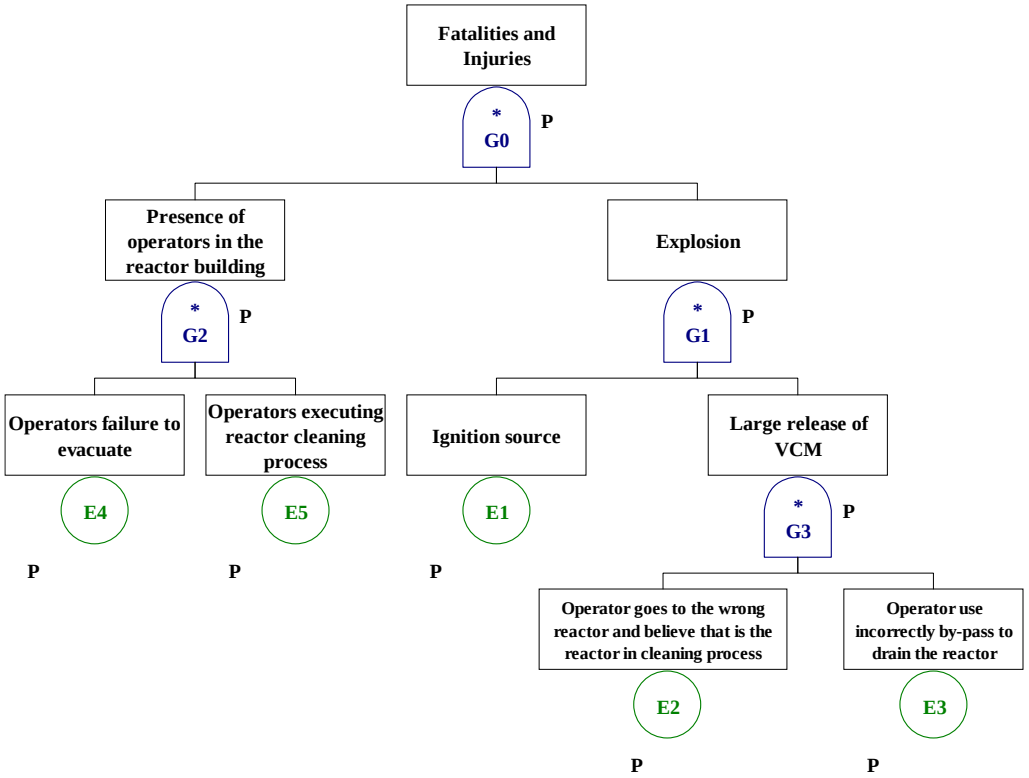

Figure 3.2: Fault tree representation of a large release of VCM scenario followed by explosion and fire causing fatalities

The basic events are directly influenced by the root causes that contribute to the occurrence of the top event (accident). Below is the list of root causes related to each basic event (Chemical Safety and Hazard Investigation Board, 2007).

Basic Event E2 - Operator believes he went to the reactor which required cleaning, when in fact he went to the reactor in operation

- There is no status indicator in the reactor
- Symmetrical layout of reactors
- Similarity of reactors
- Overload of blaster operator

Basic Event E3 - Operator uses the bypass valve to open the bottom valve of reactor in operation

- Bottom valve of the reactor does not open (interlock system - pressure above 10 psi)
- Existing system bypass
- No physical control of air injection hoses of emergency
- No bypass procedure during normal operation
- Supervisor unavailable

Basic Event E4 - Employees fail to evacuate the area

- Ambiguous procedures about how to control large releases of VCM
- Insufficient evacuation training
- No routine drills

3.2.2 IDA (Influence Diagram Analysis)

IDA allows a simplified and detailed view of the factors that influence the event (see Figure 3.3). The main elements that affect the scenario are represented by the ellipse, while the white square represents the uncertainty that led to the accident. The hexagons correspond to the investment possibilities that need to be performed. These investments are shown in blue. The IDA provides a quick and practical decision model and the great value of the diagram is its power of communication since it is easily understood and allows a large amount of information to be considered.

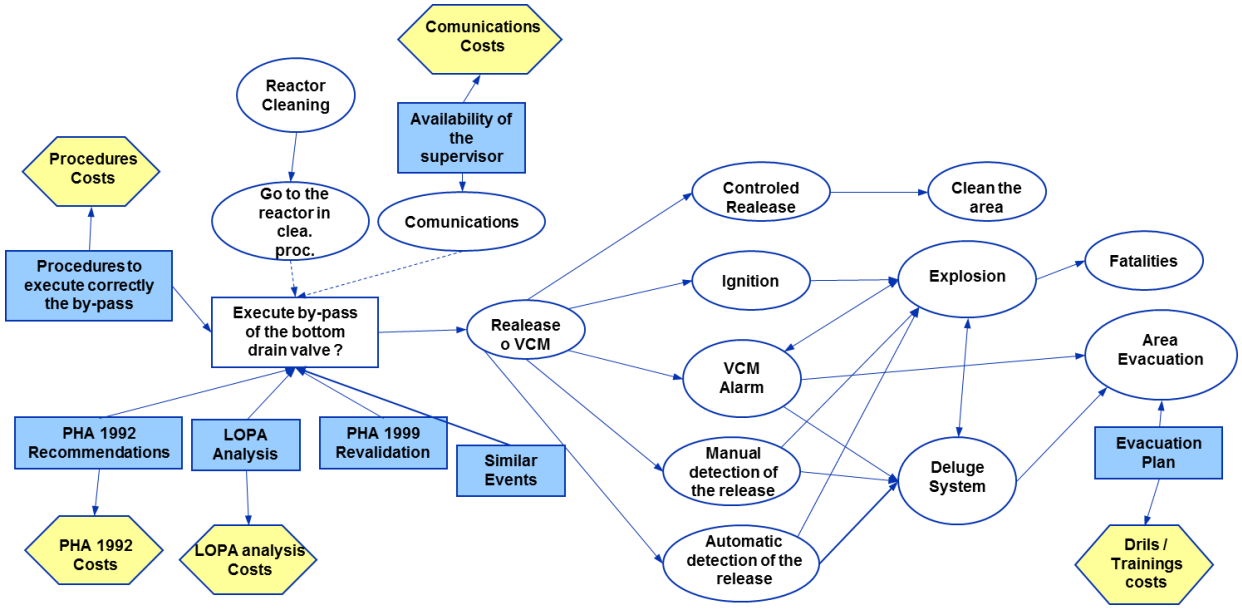

Figure 3.3: Diagram of influences of the Formosa-IL accident

The diagram identifies possible investments, but not the best option. Detailed quantitative studies could show which ones would be the priority investments, although much time, effort and knowledge are required for their development.

The problem of prioritization of the best investments can be treated in two ways. The first one from a management point of view using more general information of the organization, obtained through the analysis of PIFs. The weighted score method for decision making was used due to the following characteristics concerning prioritization of the recommendations:

- Limited financial and resource time
- Investments are independent of one another
- There are factors that may not have been considered, but the diagram logic is consistent
- Evaluation can be developed by a group, but the approval is directed to a single person (the manager)
- Quantification of the whole scenario is not as precise
- Some significant and consistent records need to be presented to justify the selection

The other way is through an operational focus based on the estimated probability of human error through the technique of reduction, evaluation of human errors HEART and quantification of the accidental scenario using the FTA.

The recommendations that were suggested to Formosa-IL are summarized below. Assessments were prepared to prioritize the recommendations.

- Recommendation A - Increase the supervisor availability
- Recommendation B – Implementation of Layer of Protection Analysis (LOPA) studies
- Recommendation C - Implementation of Recommendations from Process Hazard Analysis PHA1992
- Recommendation D - Procedures for use of bypass during normal operation

3.3 QUANTIFICATION OF HUMAN ERROR

3.3.1 Quantification of Basic Events for the FTA

Quantification is an important step in defining the impacts of possible improvements in the reactor design. For quantification of the fault tree, it is necessary to estimate the probability of fault of the basic events. The human error probabilities (HEPS) are estimated using the HEART Methodology.

- Basic Events E2 to E4 have the associated probability of fault presented in Table 3.5:

Table 3.5: Probability of fault of Basic Event E2 to E4

Basic Event E2	**Central**	**B5**	**B95**
Generic Task E	0.02	0.007	0.045
Errors Producing Conditions (EPC)		**Proportion**	**Calculation**
Pathway capacity overload particularly caused by simultaneous presence of non-redundant information (x 6)		0.2	(6-1) x 0.2 +1 = 2.0
No direct scheduled and clear confirmation of an intentional action (x 4)		0.3	(4-1) x 0.3 +1 = 1.9
Assessed probability of fault	0.02 x 2 x 1.9 = 0.076		
Basic Event E3	**Central**	**B5**	**B95**
Generic Task B	0.26	0.14	0.42
Errors Producing Conditions (EPC)		**Proportion**	**Calculation**
No direct scheduled and intentional action of a clear confirmation. (x4)		0.1	(4-1) x 0.1 +1 = 1.3
Little or non-independent checking or testing of output (x 3)		0.2	(3-1) x 0.2 +1 = 1.4
Assessed probability of fault	0.26 x 1.3 x 1.4 = 0.47		
Basic Event E4	**Central**	**B5**	**B95**
Generic Task I	0.03	0.008	0.11
Errors Producing Conditions (EPC)		**Proportion**	**Calculation**
Lack of familiarity with the situation that is potentially important, but that occurs infrequently, or which is unprecedented. (x 17)		0.5	(17-1) x 0.5 +1 = 9
Assessed probability of fault	0.03 x 9 = 0.27		

- Basic Event E1 (Source of ignition) has the ignition probability of 30% (Uijt de Haag, 1999).
- Basic Event E5 (Operators present for cleaning process of reactor) has a probability of 4/24 representing 4hr of day (day and night) on the lower level.

- **Critical analysis of the probabilities of the basic events**

Table 3.6 summarizes the probabilities of occurrence of the basic events.

Table 3.6: Probability of occurrence of the basic events

ID	Description of Basic Event	Details	Probability
1	Igniting source	The probability of ignition of a flammable fluid depends on parameters such as fluid molecular weight, discharge rate of leakage, temperature of self-ignition, energy and presence of an igniting source. It varies depending on the fluid and operational storage conditions which influence its rate of release. The calculation of ignition could be determined using advanced software, but the value of 30% (Uijt de Haag, 1999) is consistent for the purpose of this study.	30%
2	Operator incorrectly goes to reactor in operation and believes to be in reactor in cleaning process	The displacement of the operator to a reactor for cleaning process is considered part of the routine and occurs in a daily basis. The reactors have indicators at the bottom and on the control panel. The probability of 7.6% relatively low compared to the others can be accepted, since the only deficiency evaluated is the identical arrangement of the reactors.	7.6%
3	Operator uses bypass to open bottom valve of reactor in operation	The probability of use of the by-pass valve to open the bottom of the reactor corresponds to 47% which is a high value for use of bypass security systems. Normal safety standards do not allow security systems to be shut down even during maintenance. Since this procedure of bypass of this safety valve was common in company of Formosa – IL, the value is quite representative.	47%
4	Employees fail to evacuate the area	Normally the fault of operators during evacuation in major accidents should correspond to very low values; the calculated value of 27% that corresponds to almost 1 fault every 3 times is very representative.	27%
5	Operators present for the reactor cleaning process	It is considered that there are operators in the surrounding areas of the reactor during the cleaning process for approximately 4 hours of the day.	16.7%

- **Quantification of the scenario of the Formosa-IL accident (FTA)**

The probabilities calculated using the HEART method can be used to quantify the fault tree of the accident in Formosa-IL, as shown in Figure 3.4. The purpose of this calculation is to identify the impact of each change in project.

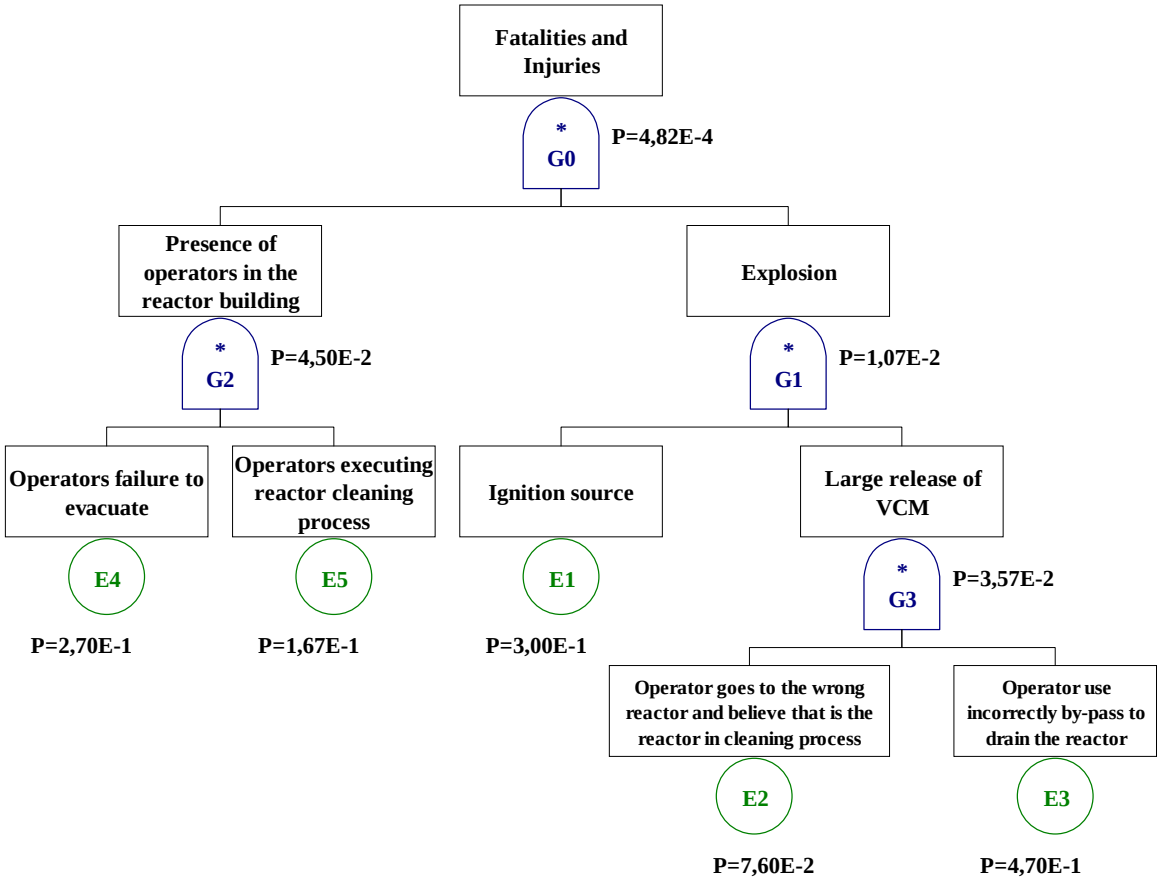

Figure 3.4: Representation and quantification of fault tree of a large release of VCM scenario followed by explosion and fire causing fatalities

3.3.2 Quantification of the IDA (MANAGEMENT FOCUS)

The management has no detailed information of operation; therefore, the decision making process is based on general techniques that do not require specific information of the activity in question. The general view allows an evaluation of the system as a whole, ensuring that the interactions of various sectors occur in the best possible way.

Each recommendation was evaluated through a score. This technique can be performed by different managers from different sectors through an individual assessment of the various stakeholders, yielding a final average. Table 3.7 shows the weight of each recommendation considered to quantify the IDA.

Table 3.7: Weight of evidence

Weight of evidence	Effective	Ineffective
What is the weight of evidence of procedures for the use of by-pass in normal operation to ensure bypass of the interlock with safety	0.3	0.7
What is the weight of evidence of the implementation of the recommendations of the PHA 1992 to ensure bypass of the interlock with safety	0.6	0.4
What is the weight of the evidence of implementing LOPA studies to ensure bypass of the interlock with safety	0.8	0.2
What is the weight of evidence for increasing the availability of the supervisor to ensure bypass of the interlock with safety	0.2	0.8

Table 3.8 shows the results of IDA quantification.

Table 3.8: Weight of evidence to conduct by-pass of the bottom valve of the reactor with safety

If	And	And	And	Success	Fault	Total Weight	Weighted success	Weighted fault
D The procedures for using the by-pass in normal operation	**C Implementation Recommendations for PHA 1992**	**B Implementing LOPA Studies**	**A Increase the availability of the supervisor**					
Effective	Effective	Effective	Effective	0.95	0.05	0.0288	2.7%	0.1%
Effective	Effective	Effective	Ineffective	0.90	0.10	0.1152	10.4%	1.2%
Ineffective	Effective	Effective	Effective	0.90	0.10	0.0672	6.0%	0.7%
Ineffective	Ineffective	Effective	Effective	0.90	0.10	0.0448	4.0%	0.4%
Ineffective	Effective	Effective	Ineffective	0.85	0.15	0.269	22.8%	4.0%
Effective	Ineffective	Effective	Effective	0.80	0.20	0.0192	1.5%	0.4%
Effective	Ineffective	Effective	Ineffective	0.70	0.30	0.0768	5.4%	2.3%
Ineffective	Effective	Ineffective	Effective	0.60	0.40	0.0168	1.0%	0.7%
Effective	Effective	Ineffective	Effective	0.60	0.40	0.0072	0.4%	0.3%
Effective	Effective	Ineffective	Ineffective	0.50	0.50	0.0288	1.4%	1.4%
Ineffective	Ineffective	Effective	Ineffective	0.50	0.50	0.1792	9.0%	9.0%
Ineffective	Effective	Ineffective	Ineffective	0.40	0.60	0.0672	2.7%	4.0%
Effective	Ineffective	Ineffective	Effective	0.40	0.60	0.0048	0.2%	0.3%
Ineffective	Ineffective	Ineffective	Effective	0.30	0.70	0.0112	0.3%	0.8%
Effective	Ineffective	Ineffective	Ineffective	0.10	0.90	0.0192	0.2%	1.7%
Ineffective	Ineffective	Ineffective	Ineffective	0.01	0.99	0.0448	0.0%	4.4%
							68.2%	31.8%

The Weighted Score Method determines the possible combinations between the recommendations and presents a successful weighted acceptance. Combinations that have the higher weighted success should have their cost of implementation verified. The implementation of the recommendations B and C correspond to the combination that attracts most managers and presents a probability of weighted success of 22.8%. The implementation of recommendation B only is very effective, but the weighted success of the activity is only 9%, being the third favourite. The second preferred combination corresponds to recommendations B, C and D with 10.4% probability of success. The implementation of all recommendations, obtaining the highest probability of success is in the eighth position. Recommendation A was considered of low efficiency (weighted success 0.3%) and consequently its implementation makes no significant contribution to the existing combinations. This analysis is based on the subjective judgment of management group members and the values used in this study were estimated.

3.4 RECOMMENDATION IMPACT USING FTA (OPERATIONAL FOCUS)

To each proposed recommendation, the EPC is re-assessed considering the reduction fraction in its value and quantifying the fault tree of the top event once more. This way it is possible to observe how each recommendation can contribute to reducing the probability of occurrence of the top event.

Table 3.9 shows the probability of accident occurrence and their respective relative reduction considering the implementation of each recommendation. From the operational point of the view, recommendation B has the largest 92% reduction in the probability followed by a 50% reduction of recommendation A. The third largest reduction of 34% is related to recommendation D.

Table 3.9: Impact of implementation of recommendations

ID	Recommendation	E2		E3		E4		FTA	
0	Without recommendations	0.076	Reduction	0.47	Reduction	0.27	Reduction	4.82E-04	Reduction
B	Implement studies of LOPA	0.076	0%	0.34	29%	0.03	89%	3.88E-05	92%
A	Increase the availability of supervisor	0.04	47%	0.45	6%	0.27	0%	2.43E-04	50%
D	Procedures for use of by-pass in normal operation	0.076	0%	0.31	35%	0.27	0%	3.18E-04	34%
C	Implementation of Recommendations PHA1992	0.076	0%	0.35	26%	0.27	0%	3.59E-04	26%
	A+B+C+D	0.04	47%	0.27	42%	0.03	89%	1.65E-05	97%

3.5 COMPARISON BETWEEN MANAGEMENT AND OPERATIONAL FOCUS

The results of the two focuses are similar showing that if implemented, recommendation B has higher potential for reduction in the prevention of an accident. Although recommendation A is not well qualified in management focus, it is the second best option according to the operational focus. This difference probably derives from the management group's choice to disregard this recommendation. Recommendation C was most prominent in terms of management than operation. Recommendation D presented similar classification in both focus.

4 CONCLUSIONS

There are numerous studies related to human behavior and each one possesses specific characteristics. Basically, they are differentiated in external focus (observable) and internal (cognitive). The method to be selected for analysis depends on the availability of information and the viability of cognitive analysis.

The human error probability was calculated based on both observable and cognitive focus following the structure of the SPEAR method. The observables factors were obtained from the HTA and the cognitive factors were analyzed with the application of PHEA. The most important step that ensured that both factors were considered in the calculation of the probability of human error is the development of the FTA based on the causes and consequences evidenced in PHEA.

The development of IDA is also based on the results of the task analysis and the analysis of human errors, which allows a visualization of variables and uncertainties of the decision process that, must be performed by managers. The results of the management focus can be less transparent than the operational focus, as it is more subjective and may be related to the interests of the decision makers.

The results of the operational focus take more objective factors into consideration with more precise indicators as its assessment is based on mental models of the plant process, which facilitates

the evaluation. These different results demonstrate the need to consider the operating environment in decision making and that they are essential for the calculation of the probabilities of human errors. This study shows that cognitive studies are not simple and are not always feasible. The efforts to calculate the probability of human error should be evaluated. Although the objective of this study was to assess the probability of human error, the results of this cognitive study provide information and possible recommendations that may contribute to reducing risks at the industrial plant.

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MARY KAY O'CONNOR PROCESS SAFETY CENTER
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Improving Industry Process Safety Performance through Responsible Collaboration

Mawusi Bridges and Alyse Keller
American Fuel & Petrochemical Manufacturers
1800 M Street, NW Washington, DC 20036
akeller@afpm.org and mbridges@afpm.org

Shanahan Mondal
CVR Energy Inc.
2277 Plaza Dr #500
Sugar Land, TX 77479
smondal@cvrenergy.com

Ryan Wong
ExxonMobil Research and Engineering
22777 Springwoods Village Parkway
Spring, TX 77389-1425
ryan.l.wong@exxonmobil.com

Abstract

Since 2012, the American Fuel & Petrochemical Manufacturers (AFPM) and the American Petroleum Institute (API) have been working together with industry members to improve the industry's process safety programs under the umbrella of Advancing Process Safety (APS) programs. This paper presents "learning from experience" illustrating the ways that industry participants are using the APS programs, what changes they have implemented, and what improvements have been achieved by those changes. Examples include practice sharing documents, training, industry safety bulletins, and Walk the Line. Most importantly, this paper provides a variety of practical takeaways that highlight the sharing an application of industry lessons learned.

Keywords: Process Safety Culture, Learn from Experience, Leadership

1 Advancing Process Safety (APS) Overview

The American Fuel & Petrochemical Manufacturers (AFPM) and American Petroleum Institute (API) Advancing Process Safety (APS) program is a comprehensive suite of programs and resources to assist industry in improving their process safety performance and raise awareness of possible process safety hazards. The APS program was first conceptualized in 2010 as a process developed by industry for industry. This paper outlines how the program was developed, how it continues to improve, and highlights testimonials from operating companies that have utilized the resources to improve safety.

1.1 Why Advancing Process Safety?

Prior to 2010, there were several high-profile events involving the petroleum refining and petrochemical industries. Safety has always been a top priority, but it was obvious that industry needed to evolve if it was going to maintain its license to operate. In 2010, several Senior Executives from AFPM and API member companies saw an opportunity to share safety lessons learned and help all of industry. That was the first meeting of what is now called the Process Safety Advisory Group (PSAG). The PSAG members empowered their Process Safety Directors to develop tools to help industry improve. The Process Safety Directors created the Process Safety Workgroup (PSW), which oversees the APS resource development program.

Advancing Process Safety Guiding Principles:

- Focus on improving industry's process safety performance.
- Prioritize mitigating higher-consequence risks – supported by data and industry experience.
- Avoid being prescriptive and provide a range of tools to achieve desired outcomes.
- Protect companies' liability, intellectual property, and antitrust exposure.

The philosophy of the APS program is that it is in every company's best interest to improve process safety across the industry. The PSAG challenged the legal subgroup to develop a way for the programs to exist and thrive, while navigating the obstacles and risk landscape. After 18 months of development and the commitment from industry and senior leaders, AFPM and API launched the APS program in the Spring of 2012.

Obstacles to Success

- History of competitive silos
- Trade Group Logistics
- Legal concerns
- Antitrust
- Documenting areas for opportunity
- Inadvertently setting industry standards and expanding RAGAGEP
- Sharing Safety Practices

These voluntary industry programs attract participants by providing high-value resources that are relevant to petroleum refining and petrochemical facilities of all sizes. The APS programs leverage the unique strengths of both AFPM and API member companies to ensure program growth and success.

1.2 Advancing Process Safety Programs in 2020

The APS programs are industry-led initiatives to continuously improve process safety performance through enhanced information sharing, communication, and responsible collaboration. Under the oversight of the PSAG and PSW there are seven formal subgroups that

develop and provide industry with tools to help improve process safety. The subgroups are described below:

Hazard Identification and Practice Sharing (HIPS)
The HIPS subgroup provides an avenue for identifying hazards and sharing industry hazards and safety practices. Hazard Identification documents and associated shared practices are available to all member companies.

Human Reliability Subgroup
The Human Reliability Subgroup brings together industry experts and practitioners to focus on identifying safety improvement from a human perspective. This subgroup develops tools that may reduce the likelihood and/or consequence of human error related events.

Industry Learnings & Outreach (ILO)
The ILO subgroup provides data analysis, identifies trends, and communicates opportunities to the other subgroups. The data analysis includes the annually submitted API RP 754 Process Safety Event data, an event sharing database, and trends from the API Process Safety Site Assessment program.

Mechanical Integrity (MI) Subgroup
The MI Subgroup brings together industry experts and site-level practitioners to identify opportunities to reduce the process safety risks associated with the operation, inspection, and maintenance of fixed equipment in refineries and petrochemical facilities. Work products focus on education and awareness and include an educational brochure on existing API mechanical integrity Standards and Recommended Practices [1]. To aid in industry awareness efforts, the related API standards and codes committees have a standing agenda item reserved for the MI Subgroup to provide updates.

Process Safety Regional Networks
The Process Safety Regional Networks enable knowledge sharing between site-level process safety practitioners from petroleum refinery and petrochemical facilities. The networks serve as a conduit for the flow of information between the subgroups with the goal of improving overall safety performance. A map of the networks can be found in Appendix 6.2.

Process Safety Site Assessment Program (PSSAP)
The Process Safety Site Assessment Program (PSSAP) uses highly experienced, independent, third-party assessors to evaluate the health of site-level process safety programs using a standard set of protocols. The assessors, who average over 40 years of industry experience, spend a week at a site evaluating the quality of written programs and the effectiveness of field implementation. PSSAP provides a benchmark report to assist each site understand their performance relative to their peers and to highlight the opportunities for improvement. A list of the assessment protocols can be found in Appendix 6.3.

Walk the Line (WTL)
Walk the Line addresses the human reliability challenges of day-to-day operations. Specifically, it focuses on three fundamentals: (1) changing the culture by setting the expectation for energy control, (2) operational continuity through the use of operating discipline tools, and (3) operational readiness. These three fundamentals help Operators understand the importance of operating line-ups in a plant and potentially reducing related process safety events.

Each subgroup develops valuable resources based on their areas of focus. By working to reduce process safety events from a variety of perspectives, a wide audience of industry practitioners are reached. These resources are disseminated through the AFPM Safety Portal, monthly webinars, newsletters, and the process safety regional networks.

Figure 1 illustrates the continuous improvement of the APS program over the years.

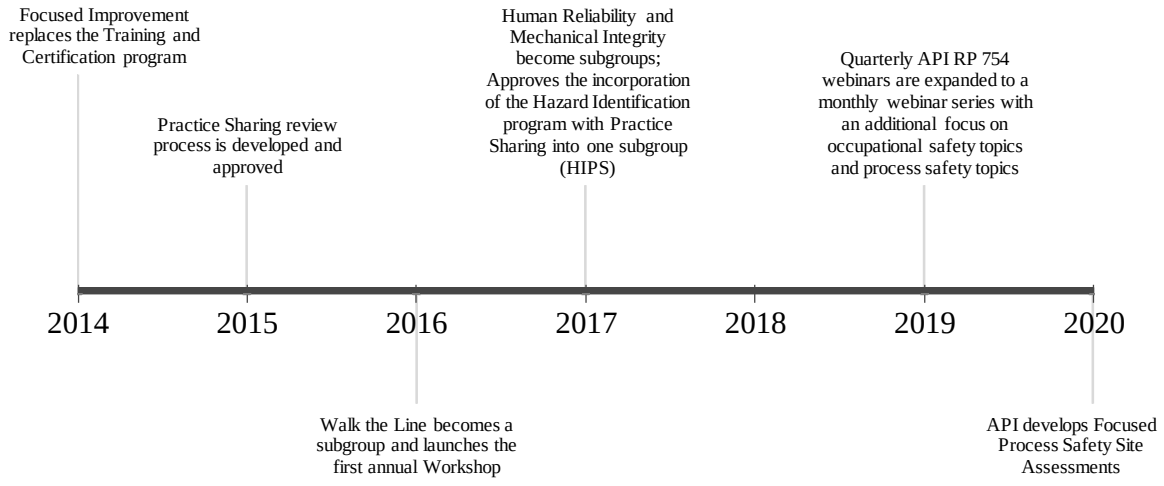


Figure 1: APS Continuous Improvement Milestones Since Program Origination.

With a foundation of data-driven programs, APS has evolved since 2010 and will continually improve.

2 Industry Performance

2.1 API RP 754 Process Safety Indicators

In 2010, the first edition of the *ANSI/API RP 754 Process Safety Indicators for the Refining and Petrochemical Industries* was published; the second edition was published in 2016 [2]. This recommended practice (RP) created a single common definition for a process safety event

thereby allowing industry to collect, analyze, and benchmark process safety performance. The ILO subgroup leverages this information to analyze and share learnings to aid in process safety improvements.

Figure 2 shows the quantitative improvements in PSE rates between 2011 - 2018 (2018 is the most recent data available). The combined Tier 1 and Tier 2 PSE rate per 200,000 employee hours for both petroleum refineries and petrochemical facilities has decreased since 2011. Based on 2018 data, the 3-year rolling average PSE rate per 200,000 workforce hours is for each facility type is noted below:

- Petroleum Refinery
  - Tier 1: 0.0649
  - Tier 2: 0.1772
- Petrochemical Facility
  - Tier 1: 0.0714
  - Tier 2: 0.1614

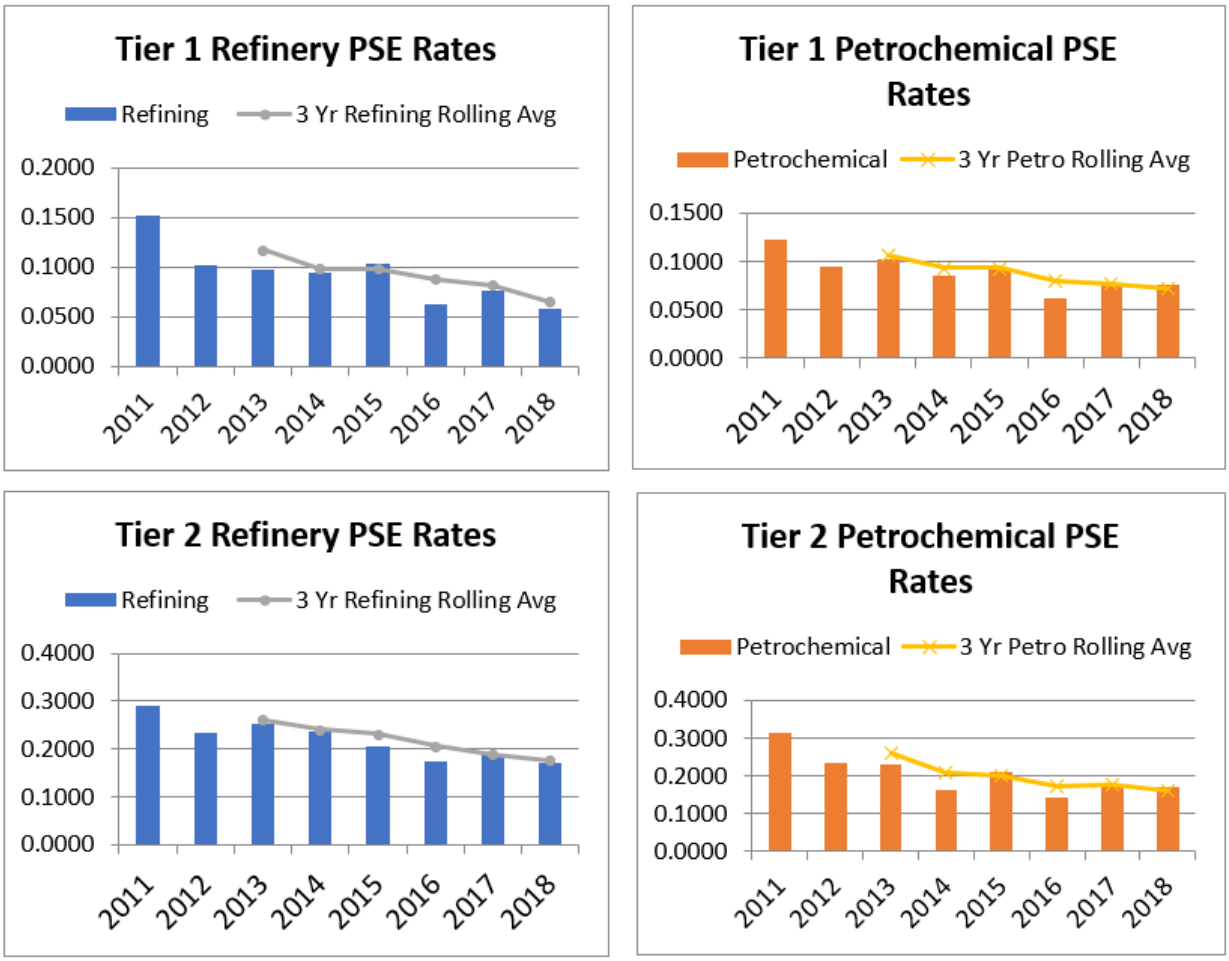


Figure 2: Process Safety Event Performance for Petroleum Refining and Petrochemical Facilities.

2.2 Process Safety Site Assessments

Industry improvements are also reflected in the protocol performance from the API Process Safety Site Assessments. Figure 3 illustrates how assessments performed in the 2016 to 2019 timeframe (green squares) are scoring better than those done earlier in the program (2012 to 2015 – blue dots), with the largest improvement in the Mechanical Integrity-Fixed Equipment scores. Facility Siting is the exception as additional questions were added after RP 756 was published. The belief is that sites that had an assessment performed in the later years have benefitted from the programs and learnings of the APS effort. For example, Hazard IDs, learnings from site assessments, API RP 754 metrics, discussions at Regional Networks, and the efforts of the MI Subgroup, among others have assisted these later sites to strengthen their process safety programs. Some sites are beginning to be assessed for the second time. This will provide specific data points on site improvements.

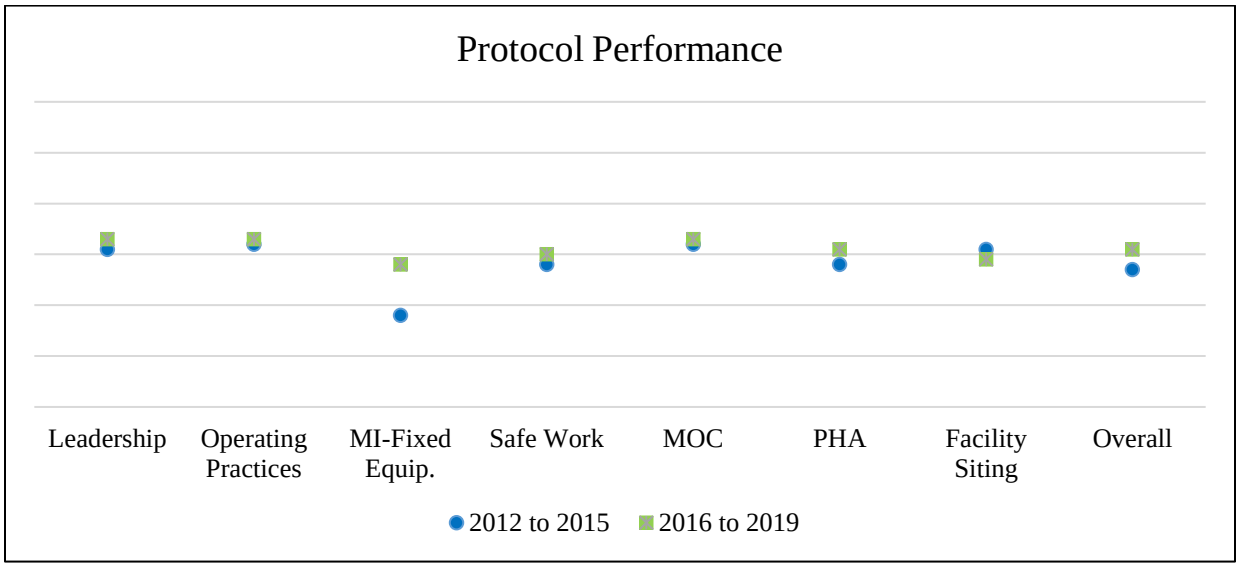


Figure 3: Protocol Performance from API Process Safety Site Assessments.

Beginning in 2020, API will offer PSSAP Focused, which aims to conduct more assessments at small refining, petrochemical, and chemical facilities in order to improve process safety performance across the industry. This new addition to the Process Safety Site Assessment Program is s tailored design to cover the seven original protocols and address key process safety activities in a shorter time frame. Specifically, PSSAP Focused utilizes smaller assessment teams and fewer protocol questions.

2.3 Walk the Line

The Walk the Line (WTL) program, developed by AFPM in coordination with Jerry Forest of Celanese to share with industry, seeks to address Loss of Primary Containment (LOPC) causes related to: open ended lines/ valve left open, operational readiness, and line-up error [3-7]. Annual analysis on the API RP 754 Tier 1 and Tier 2 PSE data indicates a downward trend in WTL related events, as shown in Figure 4. LOPC causes related to line-up errors has seen a ~90% reduction, whereas causes related to operational readiness and open-ended lines/valve left

open have seen a ~9% and ~20% reduction, respectively. Based on the number of events, the greatest opportunity for improvement is with open ended lines/valve left open.

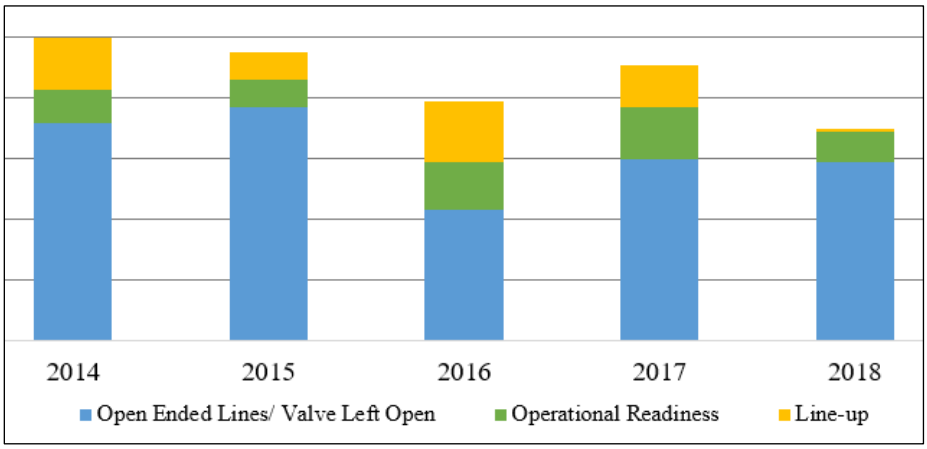


Figure 4: LOPC Reduction in Walk the Line Related Events [8].

2.4 Available APS Resources

There are over 1,100 tangible documents and resources on the AFPM Safety Portal available for use by AFPM member companies. Through data analysis and benchmarking reports, industry can effectively learn, share, and improve. Appendix 6.1 includes a complete list of documents available on the AFPM Safety Portal.

Available Resources

- 20 Hazard Identification Documents
- 57 Practice Sharing Documents
- 19 Safety Bulletins and Learning Reports
- 25 Webinars
- 747 Event Sharings
- 277 Conference Papers/Presentations

The remainder of the paper discusses examples and use cases from operating companies to illustrate the use of APS to drive safety improvement.

2.5 Event Sharing Database Industry Use Examples

The process safety event sharing database contains a collection of high-level, blinded, industry events, submitted voluntarily by operating companies. Users can query the database, searching on criteria that includes process type, API 754 consequences, mode of operation, equipment type, cause, and keywords. Examples of how event sharing has been used include:

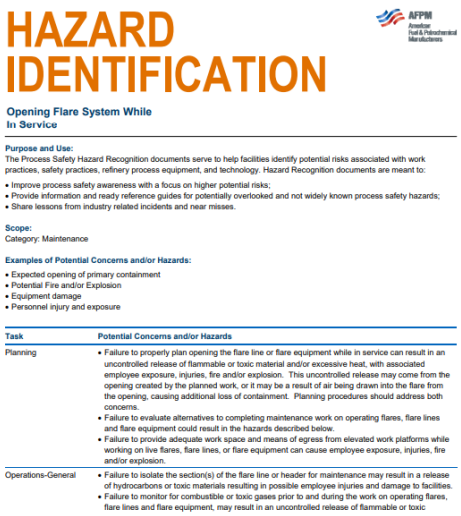
- Gathering related events to discuss in a Process Hazard Analysis (PHA)
- Searching for topics for safety meetings and toolbox talks
- Reviewing Crude Unit events to support the preliminary hazard review for a new Crude Unit design
- Discussing related events during design reviews for new or replacement equipment
- Searching related events and the associated corrective actions following PSE investigations, which may be included in the reports to potentially prevent a future event that has occurred in industry

3 Operating Company Testimonials on How the APS Products Have Helped Improve Safety

The following testimonials are from operating companies on how the APS tools led to tangible improvements.

3.1 Large Company Testimonial A: US Petroleum Refining Company

This company participates in every aspect of the APS suite of programs and due to their involvement has made enhancements to corporate and site procedures.



HAZARD IDENTIFICATION

AFPM American Fuel & Petrochemical Manufacturers

Opening Flare System While In Service

Purpose and Use:
The Process Safety Hazard Recognition documents serve to help facilities identify potential risks associated with work practices, safety practices, refinery process equipment, and technology. Hazard Recognition documents are meant to:

- Improve process safety awareness with a focus on higher potential risks;
- Provide information and ready reference guides for potentially overlooked and not widely known process safety hazards;
- Share lessons from industry related incidents and near misses.

Scope:
Category: Maintenance

Examples of Potential Concerns and/or Hazards:

- Expected opening of primary containment
- Potential Fire and/or Explosion
- Equipment damage
- Personnel injury and exposure

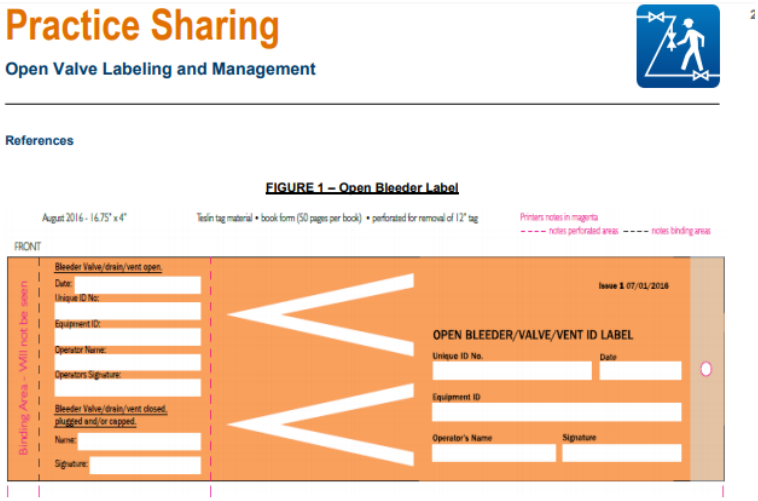
Task	Potential Concerns and/or Hazards
Planning	• Failure to properly plan opening the flare line or flare equipment while in service can result in an uncontrolled release of flammable or toxic material and/or excessive heat, with associated employee exposure, injuries, fire and/or explosion. This uncontrolled release may come from the opening created by the planned work, or it may be a result of air being drawn into the flare from the opening, causing additional loss of containment. Planning procedures should address both concerns. • Failure to evaluate alternatives to completing maintenance work on operating flares, flare lines and flare equipment could result in the hazards described below. • Failure to provide adequate work space and means of egress from elevated work platforms while working on live flares, flare lines, or flare equipment can cause employee exposure, injuries, fire and/or explosion.
Operations-General	• Failure to isolate the section(s) of the flare line or header for maintenance may result in a release of hydrocarbons or toxic materials resulting in possible employee injuries and damage to facilities. • Failure to monitor for combustible or toxic gases prior to and during the work on operating flares, flare lines and flare equipment, may result in an uncontrolled release of flammable or toxic

Figure 5: Hazard Identification Document: Opening Flare System While in Service [9].

The Hazard Identification document, *Opening Flare System While in Service*, Figure 5, addresses potential hazards present throughout the job – from planning through execution. This document was instrumental in this company creating a new corporate-wide standard regarding live flare work.

Based on petroleum refining and petrochemical Tier 1 and Tier 2 PSE data analysis conducted in 2016, approximately 43% of the events occurred in piping systems. Approximately 40% of the piping system events were related to small bore, <2" diameter, pipes. These industry metrics helped drive new corporate initiatives and programs including small-bore piping program, Corrosion Under Insulation (CUI) program, and piping vibration initiative.

Analysis of the 2018 Tier 1 and Tier 2 PSE data showed that approximately 33% of the PSE events have a human error cause and about 25% of these events are related to WTL [8]. This information helped drive internal programs including verification of returning equipment from maintenance, pre-startup piping and instrumentation (P&ID) walk downs, verification of energy isolation, and open valve labeling and management as seen in Figure 6.



Practice Sharing

Open Valve Labeling and Management

References

FIGURE 1 – Open Bleeder Label

August 2016 - 16.75" x 4"

Tesin tag material • book form (50 pages per book) • perforated for removal of 12" tag

Printers notes in magenta
---- notes perforated areas ---- notes binding areas

FRONT

Binding Area - Will not be seen

Bleeder Valve/drain/vent open.
Date:
Unique ID No:
Equipment ID:
Operator Name:
Operators Signature:

Bleeder Valve/drain/vent closed, plugged and/or capped.
Name:
Signature:

Issue 1 07/01/2016

OPEN BLEEDER/VALVE/VENT ID LABEL

Unique ID No.
Date
Equipment ID
Operator's Name
Signature

Figure 6: Practice Sharing Document: Open Valve Labeling and Management [10].

This company conducted process safety site assessments at each of their sites. Opportunities identified in the site assessments led the company to take the following actions:

- Updated a corporate specification to reduce intervals on process safety valves (PSV) testing when they leak during the offline inspection pop test.
- Updated corporate practices to develop a standard format of Corrosion and Material Diagrams.
- Developed a new site level soil-to-air interface inspection procedure.
- One or more sites updated their piping inspection procedure to include visual inspection of piping supports.
- One or more sites reviewed and revised their warehouse Alloy Verification procedure.
- One or more sites developed a local procedure for inspection under fireproofing.
- One or more sites developed a new procedure for inspecting process unit and pipe rack structures.

3.2 Large Company Testimonial B: Multinational Petroleum Refining and Petrochemical Company

Based on recommendations from a Process Safety Site Assessment, this company was able to incorporate the feedback into their LOPC reduction plan and make improvements to their Lock Out/Tag Out process. This input helped streamline their process and make roll out and site adoption easier.

Through involvement with the WTL program, this company has incorporated the philosophy and concepts corporate wide as a core pillar of process safety.

Process Safety Site Assessment Facility Testimonial

"Our company has benefited from several of the APS programs, with the most benefit from the Process Safety Site Assessments. We found that the Assessors are seasoned veterans with 25, 35, even more than 40 years of experience - truly experts in their field. They come with an attitude of wanting to help and are not judgmental. By talking to them and learning from their experiences, and through the use of the industry assessment protocols, our Subject Matter Experts (SMEs) get a clearer picture of the industry good practices."

This company has incorporated the WTL philosophy and concepts corporate wide as a core pillar of process safety. By leveraging WTL practices and Human Reliability principles, they have been able to maximize their program effects on the front line. Examples include:

- Through translating the WTL terms into their company "language," they have found a greater rate of acceptance and adoption.
- A presentation at the 2019 WTL Workshop discussed the concept and implementation of a valve line up board. The program focuses on operator ownership, having a clear visual to tag open bleeds, and having a simple system to audit. This board has been incorporated into one site's daily operating meetings. The presentation is available on the AFPM Safety portal and a related practice share is currently in development. This is a great example of Operations knowledge sharing and improvement facilitated at the WTL Workshop.

3.3 Medium Company Testimonial: US Petroleum Refining Company

Process Safety Site Assessment Facility Testimonial

"… One of the most important takeaways of the process was the opportunity for our staff to learn through engagement with the assessors …The assessors had a wealth of plant experience that allowed them to dive into the key aspects related to the health of our overall process safety management …"

Through involvement with the WTL program, this company is better able to speak a common process safety language among their petroleum refining, pipeline, and terminal business units. The practical approach to address specific areas for improvement are applicable to all facility types. Since implementing WTL, this company has seen noticeable improvements around culture and operator ownership.

As a mandatory step in their workflow, teams updating safe work practices review the APS resources. By reviewing these materials, including event lessons learned, Hazard Identification and Practice Sharing Documents, and Safety Bulletins, they benefit from industry knowledge to help make their company safer. The useful and relevant topics are then incorporated into their company practices. This has been an effective way to help make their company safer and familiarize a wider audience with the APS resources.

3.4 Small Company Testimonial A: US Petroleum Refining Company

Many petroleum refineries are in geographically remote areas. This can present a challenge for companies to benchmark, learn, and improve. This small company utilizes the Process Safety Regional Network meetings to develop a larger network of process safety peers and help close any gaps, by leveraging lessons learned from network companies.

The Regional Networks provide an opportunity for talent development and training, as many process safety professionals have an Engineering or Operations background. The meetings provide practical examples from industry peers that help process safety professionals supplement their technical training and help to further understand how to set up and roll out effective programs. The peer-to-peer discussions also help with the cultural adoption of programs by leveraging peer experiences.

Small companies may not have internal Subject Matter Experts (SMEs). The process safety regional networks provide an opportunity for invited topical experts (e.g. Fired Heaters, Facility Siting, Safety Instrumented Systems, Mechanical Integrity, Alarms, etc.) to interact with the members. The subject matter covered at these meetings is both practical and challenging, giving the participants a deeper understanding and better questions to ask their colleagues.

3.5 Small Company Testimonial B: US Petroleum Refining Company

The annual API RP 754 PSE analysis and report allows companies and sites to benchmark their process safety performance against industry.

- This company also uses the industry data during internal investigation team training, highlighting the importance of having good investigation recommendations and implementing mitigations to minimize risks.

- They also use the PSE metrics as a company and site key performance indicator (KPI). This helps set goals and keeps the importance of minimizing PSEs at the forefront for all employees, from the front line to senior leadership.

A focused learnings report was developed, based on 2018 PSE Tier 1 and Tier 2 Petroleum Refining and Petrochemical data analysis (Figure 7). This two-page document provides a clear and concise communication, identifying data trends and available resources to help. This report highlighted industry focus areas and helped this company support budget requests for related projects, such as high-high alarms on tanks to potentially help prevent overfill events.

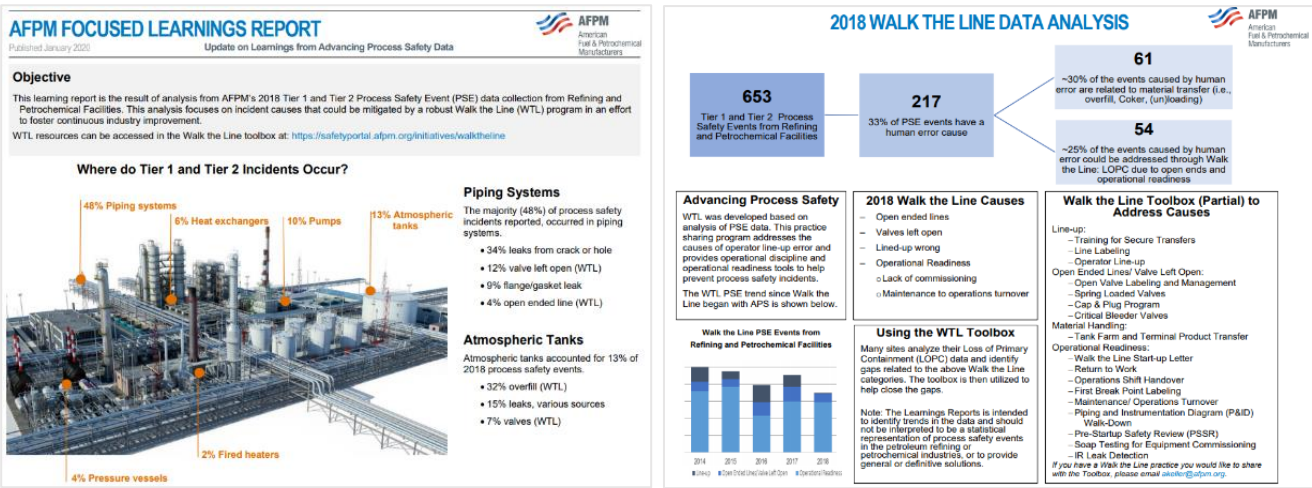


Figure 7: 2018 AFPM Focused Learnings Report [8].

This company developed and conducted an event lessons-learned training exercise at a process safety regional network meeting. Participants were given a picture of the area after the event, high-level details (facility type, equipment, type of LOPC, and Tier classification), and the cause category descriptions. See Figure 8 below.

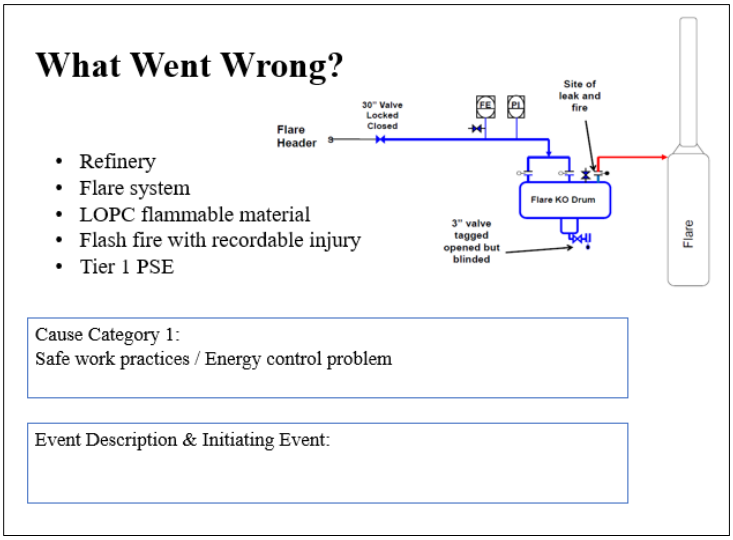


Figure 8: Example of "What Went Wrong?" Participant Worksheet [11].

Participants then broke out into small groups to determine "what went wrong," potentially causing the event. This type of discussion allowed participants to talk through the event, drawing on their own experiences to determine the event cause. At the end of the breakout, the group reassembled and shared interpretations of the event. The facilitator then showed the "correct answers" based on the investigation findings. The exercise template was shared with network members and has been used at different companies. One company has utilized this during a site Safety Day to review their own events and lessons learned. Following this training, the front-line employees expressed this was an engaging and helpful exercise to discuss events, with a focus on hazard recognition.

3.6 Small Company Testimonial C: US Petroleum Refining Company

This small company has leveraged the APS programs to maximize their learnings and growth opportunities through interactions with larger companies. While improvements have been numerous, they have seen the greatest step changes in the following three areas:

- A Process Safety Site Assessment at one facility revealed a deficiency in their MI approach/ program. Based on this learning, the company has created a corporate Director of MI (position filled and in place for 2.5 years) to provide a centralized approach to MI across all sites. They are currently implementing a formal Fixed Equipment Mechanical Integrity (FEMI) program.
- Participation in WTL has resulted in the creation of "WTL Process Safety Committees" at the sites. Based on the 2019 WTL Workshop presentation, the sites have incorporated the use of a committee to not overwhelm one person and keep the initiative fresh and moving forward. These committees are in addition to their Health & Safety Committee. Their Operators are interested in this program and are excited about the opportunity to participate. The presentation is available on the AFPM Safety Portal and a related practice sharing document is in development.
- A popular topic of focus at regional network meetings has been around Human Factors and Learning Teams. The examples and lessons learned shared by larger companies has greatly benefited this smaller company. While the implementation of Learning Teams is in its infancy at this company, all sites have conducted one Learning Team following an event. The front-line employees are energized by this process and are learning focused.

Process Safety Site Assessment PSSAP Assessor Testimonial

"I have taught over 100 process safety courses. I think the API PSSAP has a much bigger impact and provides much better training than any of the courses that I have taught. The hands in the field activities with industry experts for 40+ hours with protocols of best practices creates a phenomenal learning environment and opportunity."

4 Conclusion

Through the development and implementation of a collaborative safety program, the petroleum refining and petrochemical industry has moved the needle in process safety performance. Company executive leadership has influenced, encouraged, and supported a culture of commitment by their employees to improve process safety. Individuals have utilized the

available resources such as event sharing, PSE data, Hazard Identification and Practice Sharing documents to learn from industry events to help drive improvement at their sites.

The responsible collaboration developed through APS has evolved beyond process safety. The Process Safety Regional Networks inspired the creation of additional regional networks for topics like Occupational Safety, Mechanical Integrity, and Hydrofluoric (HF) Acid Alkylation Operations. These new regional networks engage with existing industry groups, such as the Mechanical Integrity RP standards committees and the HF Alky Safe Operations Forum. These responsible collaborations further reinforce the importance of industry helping industry improve safety. AFPM and API have also shared the APS model with other industry trade associations, as improving safety is in everyone's interest.

The knowledge sharing relationships extended beyond the AFPM and API member companies. Beginning in 2018, industry Senior Executives have participated in "fly ins" with regulators in Washington, DC. These fly ins provide an opportunity to develop relationships and share the industry focus on safety improvements through the APS program. Proactive efforts such as these illustrate that industry takes its commitment to safety and environmental stewardship seriously and works to maintain its license to operate.

A successful program needs to continually improve and look forward. As the workforce demographics change, so may the available training tools. To proactively address this need, the AFPM Immersive Learning subcommittee was formed to build tools to help improve training. There is also a drive to understand more about Process Safety Events, with a focus on potential leading indicators. APS and the newest focus areas will continue to evolve to address the needs of the industry.

These programs could not have developed into the impactful tools they are without widespread industry support and collaboration. Advancing Process Safety has a strong foundation and will continue to move the needle over the next ten years. Thank you to all the companies who help advance process safety and have shared your testimonials. For more information and to get involved, please email safetyportal@afpm.org.

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6 Appendix

6.1 List of APS Documents Available

Hazard Identification Documents

- Atmospheric Tank Preparation for Out of Service Maintenance
- Atmospheric Storage Tank Operation
- Critical Crane Lifts *(updated 2019)*
- Fired Heaters
- Flare Operations
- Hot Taps *(updated 2019)*
- Injection Point and Process Mixing Point Hazards
- Liquid Petroleum Gas *(updated 2019)*
- Maintenance/Operations Turnover and Verification After Maintenance
- Opening Flare System While in Service
- Process Sampling for QAQC *(updated 2019)*
- Shift Handover *(updated 2018)*
- Vacuum Trucks *(updated 2018)*
- Winterization

Practice Sharing Documents

- Control of Defeat of Process Safety Critical Devices
- Energy Isolation Procedure
- Evaluation of H2S Release Potential Consequences
- Falling Objects Preventative Measures
- Flange Bolting and Gasketing
- HAZOP Vulnerabilities List
- Higher Risk Work Procedure
- MOC/PSSR Responsible Person Certification
- Non-occupied Building Audit Forms
- Preventing Back Overs by Mobile Equipment
- Radio Communications Protocol
- Review of Industry Incidents
- Safe Work Permit Audit Form
- Threaded Lanyard Plug
- Training Curriculum for Process Positions

Safety Bulletins

- AFPM Annual Learnings Report for 2016
- AFPM Annual Learnings Report for 2018
- Exothermic Reaction Between Bleach and FRC
- Flammability Hazards of H2S Accumulation in Sulfur Tanks
- Flange Assembly Incidents
- Glass and HF Acid in Alkylation Units
- Hazards of Flooding Tank Floating Roofs
- Hazards of Plumber's Plug and Isolation Tool Use
- Hazards of Purged Tanks
- Injection/Mixing Point Hazards
- Insulation of Refractory Lined Equipment
- Reflection on the Texas City Isom Tragedy after 10 years
- Residual Elements in HF Alkylation Piping Caused Failure

Human Reliability

Practice Sharing documents:

- Human Performance Considerations Related to Permit to Work Overview Letter
- Part 1: Establishing Scope of Activities Managed Outside of a Permit to Work Document
- Part 2A: Enhancing Work Permit Execution: Energy Isolation – Lock Out Tag Out
- Part 2B: Enhancing Work Permit Execution: Joint Job Walks
- Part 2C: Enhancing Work Permit Execution: Safeguard Verification and Stop Work Triggers
- Part 3: Improving the Permit-Related Auditing Process
- Part 4: Enhancing Permit Issuer/ Receiver Roles (Competence, Training, and Execution)
- Fired Equipment Fuel Rich (Flooding) Warning Practice
- Fired Equipment - Assessment and Troubleshooting
- Fired Equipment Light-off Checklist
- Guidance for Fired Equipment Re-start

Safety Bulletin:

- Human Performance Related to Fired

Equipment Safety Bulletin

Mechanical Integrity

Hazard Identification documents:

- Critical Check Valves
- Deadlegs
- Hoses
- Small Bore Piping
- Temporary Repair of Piping and Piping Components

Practice Sharing documents:

- Fixed Equipment Key Process Indicators (KPIs)
- Integrity Operating Window Alarm Management
- Temporary Leak Repair Checklist and Form

Safety Bulletins:

- Hazards of Installing Temporary Repairs for Piping Systems - Clamps
- Hazards of Piping Vibration
- Hazards of Corrosion Under Insulation (CUI)

Walk the Line

Hazard Identification Documents

- Maintenance / Operations Turnover and Verification After Maintenance
- Operator Line-up
- Operations Shift Handover

Practice Sharing documents:

- Accountability Checks
- Cap and Plug Program
- Critical Bleeder Valves
- Checklist Standard Operating Procedures (SOP)
- Continuous Improvement
- First Break Point Labeling
- Open Valve Labeling and Management
- Informal Unit Walk-through
  - Operator Evaluation Rounds
  - Line Labeling
  - Management of Change (MOC) Tagging
  - Measurements & Metrics
  - Operating Instructions
- Operator Shift Notes
- Piping & Instrumentation Diagram (P&ID) Walk Down
- Pre-Startup Safety Review (PSSR)
- Process Safety Officer
- Refinery Tank Farm and Terminal Product Transfer
- Return to Work
- Safe Operating Limits
- Shift Change Communications
- Shift Change Meeting
- Shift Handover Process Safety Field Audit
- Soap Testing for Equipment Commissioning
- Spring Loaded Valves
- Use of Infra-Red Camera to Detect Leaks during Start-up
- Walk the Line Start-Up Letter

Newsletters:

- 2016 Summer Workshop – What's Next?
- 2018 Walk the Line Workshop Overview
- 2019 Walk the Line Workshop Overview
- Conduct of Operation Tools
- Conduct of Operations - Engineering Discipline Tools
- Operational Discipline – Design Tools
- Operational Readiness
- Walk the Line Training
- Walk the Line with Confidence
- What is Walk the Line?

Toolbox Topics: Front Line Supervisor "What if" scenarios

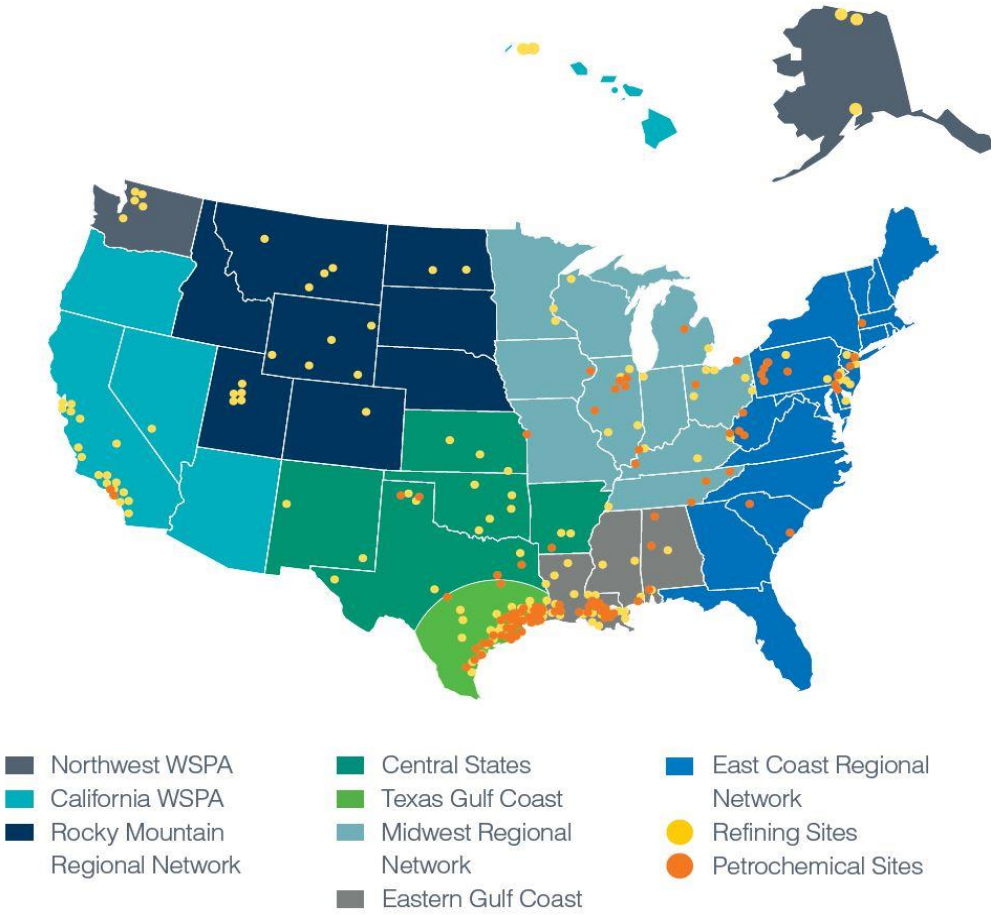
- Toolbox Topic 1: Bleeders Toolbox
- Toolbox Topic 2: Coker Switch
  - Toolbox Topic 3: Instrument Lineup
  - Toolbox Topic 4: Barge Loading
  - Toolbox Topic 5: Wrong Valve Operated
  - Toolbox Topic 6: LOTO error
  - Toolbox Topic 7: Maintenance
  - Toolbox Topic 8: Similar Valve Operation
  - Toolbox Topic 9: More Bleeders

- Toolbox Topic 10: Line Open
- Tracking Walk the Line Incidents

Training Templates:

- Training for Front Line Supervisors
- Training for Secure Transfers
- Abbreviated Training

6.2 Map of Process Safety Regional Networks



6.3 List Process Safety Site Assessment Protocols

Sites can select the following areas for assessment:

- Process Safety Leadership
- Operating Practices
- Mechanical Integrity
- Safe Work Practices
- Management of Change (MOC)
- Process Hazard Analysis (PHA)
- Facility Siting
- Product Storage & Transfer
- Incident Learning
- Hydrofluoric Acid (HF) Alkylation/API RP-751 (satisfies the 3-year audit requirement)

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23rd Annual Process Safety International Symposium
October 20-21, 2020 | College Station, Texas

How much does safety culture change over time?

Stephanie C. Payne* and Stefan V. Dumlao

Texas A&M University

scp@tamu.edu

Abstract

The Bureau of Safety and Environmental Enforcement (BSEE) defines safety culture as "the core values and behaviors of all members of an organization that reflect a commitment to conduct business in a manner that protects people and the environment" (BSEE, 2013, p. 1). The Committee on Offshore Oil and Gas Industry Safety Culture noted "Operators and contractors should assess their safety cultures *regularly* as part of a safety management system" (Recommendation 5.1, emphasis added). Interestingly, "regularly" is not defined; thus, it is unclear how frequently safety culture should be assessed. In all fairness, despite 30 years of research on safety culture (Zohar, 2010), the science of safety culture change is quite limited. In an effort to advise oil and gas companies on how frequently they should assess safety culture, we review the empirical literature and identify all of the longitudinal studies that have examined safety culture over time (i.e., minimum of two assessments). In our review, we track the industry/jobs for each sample, the average time period over which safety culture has been examined, as well the full range of time lags tested. We summarize this literature and the conclusions to date concerning the extent to which safety culture changes over time and identify the extent to which further research is warranted.

Keywords: safety culture

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Administering a safety climate assessment in a multicultural organization: Challenges and findings

Atif Mohammed Ashraf , Dr. Luc Vechot, Dr. Stephanie Payne, Dr. Tomasz Olewski
Mary Kay O' Connor Process Safety Center ---Qatar, Department of Psychology – Texas A&M University
Texas A&M University at Qatar, Education City, PO 23874, Doha, Qatar ,[b] 4235 TAMU, College Station, TX 77843-4235
: atif.mohammad_ashraf@qatar.tamu.edu

Abstract

Substantial improvements have been made in the realm of safety performance by applying inherently safer designs concepts and implementing effective HSE management systems. However, one of the biggest challenges faced by multinational corporations today is strategically managing their workforce – particularly a multicultural workforce. In addition to many advantages, a multicultural workforce faces language barriers and cultural differences in perceptions of safety-related phenomena including risks, hazards, and personal and process safety which can inhibit critical safety communications, as well as the overall health of the organization's safety culture.

This presentation will provide a brief description of the administration of a safety climate assessment across four different sites to a 1200 employee organization in the Middle East. The inter-disciplinary approach between psychology and engineering to formulate a science-based safety climate survey will also be highlighted. Challenges including but not limited to achieving maximum survey participation, overcoming language barriers and the involvement of contractors will be discussed. Finally, a review of the strengths and areas in need of improvement concerning safety at the respective sites will be presented. Apart from this, the relationship between national culture values and safety-related psychological constructs will also be examined. This research provides an initial peek at the influence of organizational safety culture over and above national culture on safety knowledge, motivation, and behaviours, which has important theoretical and practical implications for workplace safety.

Keywords: Safety culture, safety climate, safety psychology, Human factors

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A Comparison of Procedure Quality Perceptions, Procedure Utility, Compliance Attitudes, and Deviation Behavior for Digital and Paper Format Procedures

Joseph W. Hendricks & **S. Camille Peres**
Environmental and Occupational Health

Texas A&M University

College Station, Texas 77843-3122, USA

peres@tamu.edu

Abstract

There is a dearth of research on digital (hand-held, interactive; not .pdf) procedures in the process safety industries. This study surveyed employees ($N = 32$) at a large, chemical processing company in both chemical processing and logistics divisions. The goal was to determine if there are substantial differences in procedure quality perceptions between digital and paper procedure formats in addition to differences in procedure deviation behavior. Preliminary results indicate that for those using both paper and digital formats, quality perceptions are significantly higher for digital formats, with digital being perceived as higher quality. Further, although not statistically significant, preliminary analyses indicate workers deviate more frequently when using paper procedures vs. digital. Ongoing analyses include a between-subjects analysis examing these variables for workers that use only paper procedures vs. those that use both. We will compare paper vs. paper and paper vs. digital for these analyses. Additionally, we will present data on attitudes towards procedures regarding both utility (how useful they are) and compliance. Future directions for procedure format will be discussed.

Keywords: Process Safety, Procedures, Human Performance

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Practical Writing Tips To Prevent Human Error When Following Procedures

Monica Philippart, PhD
Ergonomic Human Factors Solutions
Austin, TX 78733
monica@ergoHFsolutions.com

Abstract

Operating procedures are used daily in high-risk industries. They are part of the interface between users and other system components, and contribute to the user's understanding of those other components. Ideally, operating procedures help users perform their tasks effectively, efficiently and safely. Users differ from each other in many ways and, unfortunately, those who write procedures are not usually trained to account for these differences.

Incident investigations show that a large part of adverse events in processing operations is related to procedures. Specifically, systemic problems in written procedures negatively impact human reliability and reduce process safety performance across organizations. There are a few simple recommendations that, if followed by writers of operating procedures, can significantly improve user performance, ultimately contributing to the safety and effectiveness of the operational system as a whole.

This paper explains how people process information, describes how written information can lead to human error, and offers practical tips that those writing procedures in the petroleum industry can easily implement to help their readers avoid perception, interpretation and decision-making errors. The selection of tips is based on the author's experience in teaching the subject, conducting incident investigations, and reviewing operating procedures in the aerospace and petroleum industries for more than two decades.

Keywords: Human Error, Human Factors, Operating Procedures, Human Reliability, Process Safety, System Safety.

Introduction

Procedures are used for many purposes and in many situations. Depending on the organization, procedures can be used, for example, for:

- Construction, manufacturing and design.
- Testing, validation, verification and quality assurance.
- Normal operations.
- Emergency, abnormal and off-nominal operations or conditions.
- Interactions with contractors.
- Scheduling a conference room or fixing the photocopy machine.
- Daily living activities in conditions that are out of the ordinary such as now, with the Coronavirus pandemic.

A Standard Operating Procedure (SOP) can be defined as a documented step-by-step set of instructions that guide operators in carrying out a specific task with a certain degree of uniformity. By contributing to the user's understanding of the hardware, software, and human interactions with other components of the system, it serves as a job aid and, though not a substitute, can serve as a supplement to training. Its primary purposes are to ensure consistent operating practices, to inform users of hazards associated with tasks, and to provide relevant control measures.

Industrial organizations are required to use SOPs by statutory safety and health regulations and by guidelines such as the U.S. Occupational Safety and Health Administration (OSHA) Process Safety Management (PSM) and Guidelines for Risk Based Process Safety (RBPS). Users of operating procedures differ from each other in countless ways including level of expertise, time constraints, work styles, reading abilities, personal circumstances, experience and more. Writers can make operating procedures easier to use by accounting for user differences. However, those who write procedures are typically engineers who may not be qualified for technical writing, and could be either very familiar with the job or writing instructions for a completely new job.

Problems with SOPs are regularly pointed out as one of the major causes of incidents in chemical process industries. For example, in approximately 70% of over 60 incident investigations reviewed by the U.S. Chemical Safety Board (CSB), SOPs were not properly developed, incomplete, or not followed as instructed during the course of the incidents.

Because SOPs are so widely used, and because they continue to contribute to incidents, this paper focuses on practical tips specifically selected for writing effective SOPs. Although these tips were chosen with SOPs and the typical petroleum industry SOP writer in mind, they can also be used for other types of user documentation including user guides and manuals, user handbooks and general technical instructions, job performance aids, quick reference guides, and instruction placards.

There are many writing standards and guidelines, some specific to particular industries such as aviation or nuclear. This paper leverages those resources, applying the Pareto principle to offer a practical selection of tips that the intended audience can most easily implement while significantly improving the usability of their SOPs to reduce their organization's susceptibility to human error. The selection is limited to guidance specific to writing the SOP text and was made based on the writing shortfalls most commonly observed by the author in the petroleum industry. In addition to the selection excluding other existing writing standards and guidelines, there are other aspects relevant to the development of first-rate SOPs that fall outside of the scope of this paper but are equally important including, but not limited to:

- Purpose of the document and procedure
- User population
- Document organization
- Document and page design
- Writing for electronic procedures

In order to recognize the importance of the writing tips that will be provided, it is helpful to understand how people process information and how written information, in particular, can easily lead to human error.

How People Process Information

Information processing begins with input from the sensory organs. Once sensory stimuli (touch, heat, sound waves, photons of light, etc.) are received by the person, they need to be interpreted for their meaning to be understood. Countless factors affect this interpretation, resulting in possibly different conclusions from the same stimuli depending on the individual and the circumstances under which the stimuli are received. After a conclusion has been made on the meaning of the stimuli, a person can then make a decision as to how to respond, if at all. Depending on the stimuli, responses could also be involuntary (such as when a hand moves away from a burning handle); but, for the purpose of this paper, the scope of the stimuli is limited to information presented through written procedures, which will be addressed as SOPs.

Figure 1 illustrates how people interact with a system. In the case of reading an SOP, the reader must first be able to see the information on the document (step 1 in the figure). Next, he must be able to decode what he sees (words, numbers, pictures, graphs or anything else) and understand what it means (2a). The decision on how to proceed (2b) is based on a combination of his understanding of this information and many other factors (explained in the next section). The reader will then act or not (3) based not only on the decision he made, but again on numerous other factors including the capabilities of the individual and the object of the action.

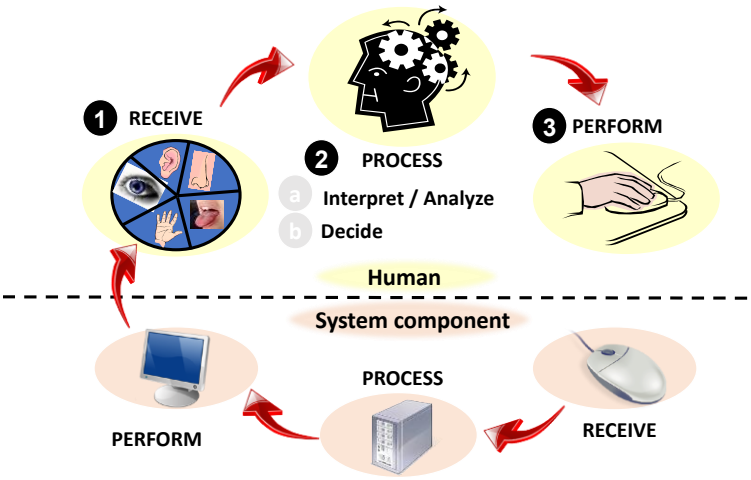

Figure 1. Human-System Interaction

The bottom half of Figure 1 illustrates the same steps on the part of the system component on which the person performed the action. That system component could be a thing or another person. The resulting action, or lack thereof, provides feedback to the first person in the form of another sensory stimuli, and the cycle continues.

This is, of course, a simplified illustration. In reality, people are continuously receiving sensory stimuli and, therefore, constantly interpreting, making decisions and responding to information.

How Written Information Can Lead To Human Error

Myriad factors affect each of the information processing steps just described. In the case of processing information from SOPs, those factors could negatively affect an individual's ability to see the information provided, causing errors of perception. Some of the factors leading to errors of perception could be internal to the individual, such as the person's vision (maybe he needs glasses and doesn't have them, or can't differentiate the colors in a color-coded graph) or preoccupation (causing him to be distracted and skip reading a step on the procedure). Some of the factors leading to errors of perception could be external to the individual, such as the document being unreadable after being exposed to the elements or due to poor lighting.

Similarly, factors internal to the individual could negatively affect how information provided in SOPs is interpreted after being adequately perceived, leading to errors of interpretation. Cultural background, knowledge, intelligence, memory, attention, ability to focus and ability to divide attention are only some examples. Examples of external factors that could lead to errors of interpretation are the amount of information provided in the procedure (both, insufficient information and too much information, could be detrimental), the clarity of the information, the quality of the information, the individual's workload, and consistency in many respects, such as the terminology used throughout the procedure.

Decision-making, the last step of information processing, is also highly susceptible to error due to internal and external factors. Many of the internal and external factors that affect interpretation also affect decision-making. Experience, knowledge, motivation, attitude, risk propensity, fatigue and stress are internal factors that could negatively influence a person's decision after having adequately interpreted information. From his knowledge, a person could correctly interpret the red traffic light that signals drivers to stop at an intersection; but that person could decide to not stop because he also knows (same factor – knowledge) that there are rarely other vehicles and never law enforcement at this intersection.

Examples of external factors that could affect decision-making are the clarity of the information, the amount of information, the timeliness of the information, time constraints, having conflicting goals, and workload. But also factors such as the existence of rules and their clarity, and the enforcement of the rules strongly affect decision-making. The driver in the example would probably stop at the red light if there was a police officer present.

Although most internal factors are difficult and some are not possible to control, they can, to an extent, be predicted, taken into account and offset with external factors. Employers can hire people adequately skilled and with sufficient experience (internal factors), and can also provide training, additional experience, and instructions (external factors).

SOPs are external factors that organizations can manage and, when properly accomplished, can prevent perception errors, interpretation errors and decision-making errors.

Writing Tips

Regardless of the intent of the writer, other elements such as how a procedure is written, its design, the context in which it is used, and the physical and mental abilities of the person at the time of use are ultimately going to dictate the extent to which the procedure is followed. This section focuses on improving the content of SOPs specifically by helping the writer present information more effectively.

Words matter. They are the most basic building blocks of SOPs and must be chosen carefully. The following tips are a selection of standards and guidelines that help convey written information, and meet the following criteria:

- Reputable source
- Prevent writer errors commonly found in petroleum industry procedures
- Reduce likelihood of perception, interpretation or decision-making errors
- Improve the usability of procedures
- Easy to understand without training
- Easy to implement without training
- Easy to remember with a simple checklist

For simplicity and brevity, tips are followed only by examples if no elaboration is necessary. Self-explanatory tips stand alone. Tips have been grouped into categories only to make reading easier; so the titles of the groups are not precise.

Conciseness

- Use short sentences. Long sentences tend to be more complex and therefore difficult to understand. Question whether you need every word; especially modifiers like absolutely, actually, really and very.
- Express only one idea in each sentence. Break complex ideas into parts and make each one the subject of its own sentence.
- Use short paragraphs.
- Include only one topic in each paragraph.

Clarity

- Avoid ambiguous terms like "approximately," "properly" and "as required."
- Convert double negatives into positives when possible. For example: "Turn off the alarm when everyone has evacuated." instead of "Do not turn off the alarm if not everyone has evacuated."
- Be precise. For example, specify the target value and allowable range: "Adjust pressure to 45 psig (range 40 to 50 psig)."
- Avoid mental calculations. For example: "Adjust pressure to 45 psig (range 40 to 50 psig)" instead of: "Adjust pressure to 45 +/- 5 psig".
- Use imperative mood for commands. For example: "Close the valve." or "Verify the valve is closed." instead of "The valve must be closed."
- Use pronouns that clearly refer to a specific noun. If a pronoun could refer to more than one person or object, repeat the name of the person or object, or rewrite the sentence. For example: "After the Engineer appoints a Consultant, the Consultant must…" instead of: "After the Engineer appoints a Consultant, he or she must…"
- Use the same units of measure and the same unit increments in the procedure as those on the instrument. Don't ask to measure 47 lb if the scale reads in 5 lb increments or in kg.
- Use "must" instead of "shall." According to the Federal Plain Language Guidelines (Rev 1, May 2011), "shall" is obsolete and imprecise, and the US Courts have been eliminating the term in favor of "must" in their Rules of Procedure.
    - Use "must" for an obligation.
    - Use "must not" for a prohibition.
    - Use "may" for a discretionary action.
    - Use "should" for a recommendation.

Ease of reading

- Use sub-titles to help with place keeping.
- Use familiar rather than original words.
- Use short rather than long words.
- Don't use slashes between words. "And/or" is a common example. Slashes are ambiguous. Write out what you mean; for example, "either X, or Y, or both" if all three options are, in fact, available.
- Use terms and expressions consistently within and across procedures. For example, standardize words commonly used such as "Review," "Verify," "Record" and "Perform." Consider standardizing complete sentences too. Consistency provides the following benefits:
    - Makes reading (perception) easier and faster.
    - Makes understanding (interpretation) easier and faster.
    - Helps anything different stand out (perception).
    - Helps avoid confusion (interpretation) – if you refer to the same thing with different terms, readers may wonder if they are different things.

Highlighting

- **Use bold for emphasis.** ALTHOUGH CAPITAL LETTERS DRAW THE READER'S ATTENTION, IT IS NOT A GOOD EMPHASIS TECHNIQUE BECAUSE IT MAKES IT HARDER TO READ. <u>Underlining will also draw the reader's attention, but is harder to read and appears to be a link when read in electronic form.</u>
- Limit emphasis to important information; otherwise, you'll dilute its impact.
- Use Warnings to alert the reader of potential hazards to personnel if instructions are not obeyed.
- Use Cautions to alert the reader of potential damage to equipment if instructions are not obeyed.
- Use Notes to call attention to important supplemental information.
- Make all Warnings, Cautions and Notes distinct from the rest of the text and distinct from each other. Example:

WARNING
Not making a Warning distinct from the rest of the text could cause the reader to inadvertently miss vital information, potentially resulting in equipment damage or personnel injury.

- Place Warnings, Cautions and Notes immediately before or after the step to which they apply. If they apply to a complete section, place at the beginning of that section.

- Include the hazard and consequence of failure to obey instructions in Warnings and Cautions.
- Include in Warnings the necessary PPE for the task, if different from what is already in use.
- Do not allow Warnings, Cautions or Notes to be separated from the step to which they apply due to page breaks.
- Include only one topic in each Warning, Caution and Note.

Conditions

- Start conditional sentences with the "if" provision, followed by the "then" provision.
- If there are several "if" provisions, use a different sentence for every "if."
- If there are multiple "if" as well as "then" provisions, use an "if-then" table. Tables organize the material by a situation (if something is the case in one column) and the consequence (then something else happens in a parallel column).
- Use tables or bullet lists to present multiple exceptions or conditions. Examples:

  Turn off the faucet if:
  - The water has reached the Max mark.
  - The water coming out is no longer clear.
  - The technician requests to stop.
  - Water stops coming out.

If the water…	**You must…**
Has reached the Max mark	Turn off the faucet
Is no longer coming out clear	Turn off the faucet, clear the filter, then continue to run water
Stops running	Turn off the faucet and contact the technician

Lists

- Use numbers or letters to designate items in a list if future reference or sequence is important. Otherwise, use bullets (solid round or square preferably).
- Make sure each of the bullets in the list can make a complete sentence if combined with the lead-in sentence. The last bullet in the following example does not complete a sentence with the lead-in sentence.

Vertical lists (lead-in sentence):

- Highlight levels of importance.
- Help the user understand the order in which things happen.
- Make it easy for the user to identify all necessary steps in a process. Add blank space for easy reading.
- Are an ideal way to present items, conditions, and exceptions.
- Use bullet lists to help your user focus on important material (This does not make a complete sentence if combined with the lead-in sentence.)

Conclusion

Operating procedures are used daily as part of the interface between users and other system components in high-risk industries. Investigations continue to point to procedures as important contributors to incidents. This paper begun describing how people process information, and explained how and why information processing is so vulnerable to human error. The purpose of this preliminary information was to help readers recognize the importance of adhering to the writing tips presented.

The purpose of the paper was to provide practical tips that SOP writers in the petroleum industry can easily implement to help users avoid errors when following procedures. The paper can be used as a training aid and the tips as a checklist for SOP writers. Because most of the tips apply to other types of technical documents in addition to SOPs, the list can be used for other purposes as well, eventually creating systemic changes that will not only reduce the likelihood of human error, but also improve human reliability and performance across organizations.

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The Impact of Hazard Statement Design in Procedures on Compliance Rates: Some Contradictions to Best (or Common) Practices

Joseph W. Hendricks[a], S. Camille Peres[a], Timothy J. Neville[a], & Cara A. Armstrong[a]
[a]Ergonomics Center, Department of Environmental and Occupational Health
Texas A&M University System
College Station, TX 77843-8371, USA
jwh@tamu.edu

Abstract

This study was designed to examine whether or not certain design elements of hazard statements (HS) actually impact compliance rates. Participants (N = 52) were trained on how to carry out eight tasks (each with an associated procedure) in a virtual 2nd Life® warehouse over the course of 32 trials. We manipulated four HS elements (present vs. absent) – Icon, Number, Fill (Highlight), and Boxed – leading to a 16 condition within-subjects design. We observed a range of approximately 20 percentage points in compliance rates across the various conditions. Some of the conditions on the lower end of compliance rates included elements such as a warning icon and boxing elements. One interpretation proposed for this observed effect is banner blindness, with the implication being individuals may be habituated to these types of elements and therefore ignore them. Future research directions are discussed.

Keywords

Procedures, safety compliance, hazard statements

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Use of Consequence Modelling for Meticulous Risk Based Inspection Calculations of Offshore Topsides Static Mechanical Equipment

Bahram Fada, Chetan Birajdar
Monaco Engineering Solutions Limited
United Kingdom
Chetan.birajdar@mes-international.com

Abstract

An important driver that influences the production profile of an offshore facility is the reliability of production critical equipment. The improvement in production profile can be achieved by minimizing the equipment downtime using reliable components, inclusion of redundant units and effective-efficient inspection maintenance service. One technique of achieving an effective-efficient inspection maintenance service is to employ the Risk-Based Inspection (RBI). RBI is a decision-making method for inspection planning based on risk – comprising the Consequence of Failure (CoF) and Probability of Failure (PoF) and its effective implementation depends on thorough understanding of both components, CoF & PoF. The PoF is related to the extent of, and uncertainty in, the degradation related to the component's resistance to its loading. The calculations methods for PoF are detailed in several industry accepted recommended practices like API, DNV, etc. CoF is defined as the outcome of a leak given that the leak occurs, and its extent depends on the inventory properties, storage and processing conditions and the surrounding area around the loss of containment. This paper focusses on application of effective CoF estimation techniques for Topside Static Mechanical Equipment on offshore installations.

API 581 is widely used industry guidelines for RBI technique and provides calculations for PoF and CoF. The mechanical static equipment on a typical offshore installation will be located on various decks (levels) and within an open or enclosed module of each deck, while the mechanical static equipment on a typical onshore installation will be located at ground level. Hence, the consequences of loss of containment on offshore installation can be observed at various heights/decks (depending on the release locations); while that on onshore installation will be predominantly at ground level. The API 581 calculation methods do not differentiate between the type of installation. While the CoF calculations are extensive for application on onshore facilities, their application on the offshore facilities might provide misleading results.

Many of the offshore installations have detailed consequence analysis performed as a part of Quantitative Risk Analysis (QRA), Fire and Explosion Risk Analysis (FERA), etc. This paper suggests a method to use results of the detailed consequence analysis as an input to the CoF calculations in RBI Assessment. The CoF results calculated using this methodology have been compared to those using the API 581 methodology for several offshore installations. The results obtained by employing this technique suggests a considerable improvement in the CoF results. Since the results of existing consequence assessment are used and no new study or calculation is required to be performed to employ this RBI technique, it is not considered as cost intensive technique to implement.

This method suggests using detailed consequence assessment results as an input to RBI technique can aid in developing more cost-effective inspection program for Mechanical Static equipment on offshore installations at no additional effort.

Keywords: Risk Based Inspection, Consequence Modelling, Static Mechanical Equipment, Reliability, Offshore

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Indicators of an Immature Mechanical Integrity Program

Derek Yelinek
Siemens Process & Safety Consulting
4615 Southwest Freeway, 900, Houston, TX 77027
derek.yelinek@siemens.com

Abstract

In the oil, gas and chemical businesses it has been considered the nature of the business to delegate responsibility and decisions for safety, maintenance and inspections to plant personnel. Historically, this made sense.

However, in today's changing environment, factors around economics, investments, operations, and safety make this approach a risky and expensive way of doing business. Executives are ultimately accountable for revenue, profitability, operational efficiency, overall safety, and business risk. A staggering array of factors impact these outcomes. Multiple facilities, processes, equipment types and personnel expertise must be overseen and coordinated.

Unfortunately, the degree of executive accountability is not always matched by the quality of information, visibility and decision tools generally available. The good news is that mature mechanical integrity programs and information systems can significantly improve this state.

We'll look at three indicators that your Mechanical Integrity (MI) program might be limiting your strategic capabilities:

1. Mechanical integrity is not a standardized program that is universally applied across the organization
2. Traditional information practices are inadequate to meet new requirements
3. Arbitrary time-based inspection practices are still used

When each facility is left to establish and follow their own set of mechanical integrity practices, organizations risk or experience adverse impact in:

- Lower Revenue due to interruptions, downtime, inefficiencies
- Higher operational costs due to reactive maintenance, ineffective inspections, mismanaged equipment and personnel
- Higher organizational risks of incidents, compliance infractions, insurability issues

• Higher management risk due to lack of information and visibility needed for effective management; commensurate with the level of accountability

Keywords: asset integrity

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Remember the à la Mode: Lessons Learned from Ammonia Release at Frozen Foods Warehouse

Matthew S. Walters*, Sean J. Dee, Russell A. Ogle
Exponent, Inc.
4580 Weaver Parkway Suite 100, Warrenville, IL 60555 USA
*mwalters@exponent.com

Abstract

Intentionally opening a line that carries a hazardous substance — a procedure known as a line break — is often necessary for performing maintenance activities on pipes, valves, pumps, compressors, and other process equipment. However, inadequate or improper line break practices may increase the risk for a loss of containment event, complicate troubleshooting efforts if a loss of containment occurs, or inadvertently expose workers to hazardous materials. In this paper, best practices and considerations for line breaks into hazardous piping systems and responding to suspected leaks of hazardous materials will be reviewed. Mitigation and prevention strategies including mechanical integrity and leak detection programs will also be discussed. Then, a case study that examines an incident related to a line break in a frozen foods warehouse will be presented. The incident resulted in a reportable release of anhydrous ammonia from a bank of parallel compressors. Just hours prior to the release, contractors conducted a line break to remove an ammonia compressor from service. Gas detector alarms activated during post-shift hours in the unmanned process building as anhydrous ammonia escaped the refrigeration system. Operators discovered the release upon reporting to work the following morning and initiated an emergency shutdown. After reviewing the work performed during the line break, personnel believed they had identified the cause of the leak and restarted the system. However, the leak persisted after the restart before being properly diagnosed during a second shutdown. The loss of containment event described here provides valuable lessons that can aid in developing an effective procedure for safe process operation following a line break, and the impact that improper line break procedures can have on leak identification and system troubleshooting.

Keywords: line break, mechanical integrity, Process Safety Management

Remember the à la Mode:
Lessons Learned from Ammonia Release at Frozen Foods Warehouse

1 Introduction

The U.S. Occupational Safety and Health Administration (OSHA) and U.S. Environmental Protection Agency (EPA) both require employers[1] to develop and implement specific safe work practices to control hazards at facilities covered by OSHA's Process Safety Management (PSM) program and EPA's Risk Management Plan (RMP) rule [1,2]:

> *The [employer] shall develop and implement safe work practices to provide for the control of hazards during operations such as **lockout/tagout**; **confined space entry**; **opening process equipment or piping**; and **control over entrance into a facility** by maintenance, contractor, laboratory, or other support personnel. These safe work practices shall apply to employees and contractor employees. [**emphasis added**]*

Recognized and generally accepted good engineering practices (RAGAGEP) are well-established for developing lockout/tagout (LOTO) [3,4] and confined space entry [5,6] procedures. Additionally, resources that aid the development of administrative controls for access to hazardous locations are readily available [7,8]. RAGAGEP for opening process equipment or piping containing hazardous chemicals, however, are less established and are often situationally dependent [9]. This paper explores practices and considerations for intentionally opening piping that carries a hazardous substance, commonly referred to as line breaking. A loss of containment incident from an ammonia refrigeration system will also be reviewed, demonstrating potential consequences of line breaking and emphasizing the importance of developing and properly implementing a line breaking procedure at PSM/RMP covered facilities.

2 Line Breaking Overview

Line breaking is often necessary for performing maintenance or troubleshooting activities related to pipes, valves, pumps, compressors, and other process equipment. However, a line break by definition compromises the mechanical integrity of a system and increases the risk for loss of containment. A three step approach to mitigate hazards associated with line and equipment opening was presented in a document produced under OSHA's grant program, as summarized in Figure 1 [10]. Aspects of these steps are discussed in the following sections. Note that Figure 1 is meant to provide a general overview and may not provide a comprehensive list of all considerations related to line breaking. (I saw

[1] OSHA regulations refer to "employer," while RMP regulations refer to "owner or operator."

Figure 1. General steps for safe and successful line and equipment breaking [10].

2.1 Review and Plan for the Hazard

Understanding the hazards associated with a line break is critical to developing a safe work plan. The Review step described in Figure 1 is an opportunity to perform hazard identification and risk analysis (HIRA) to help ensure adequate controls are in place prior to opening piping or equipment. A hierarchy of hazard controls is typically followed, with personal protective equipment (PPE) often specified as the last line of defense [11]. Examples of items that may be reviewed prior to line breaking include, but are not necessarily limited to:

- Safety data sheets (SDSs) to inform the necessary level of PPE.
- Piping and instrumentation diagrams (P&IDs) and other relevant documentation to develop a procedure for isolating, draining, purging, and/or flushing a specific piece of equipment.
- Facility layout to determine means of egress and access to first aid stations.
- Other activity occurring at the facility so that adequate barricades can be put in place to prevent workers not involved with the line break from potential exposure.

It is recommended that a subject matter expert (SME) participate in the review of potentially hazardous tasks such as line breaking [9]. Consider an incident where contract employees removing scaffolding 80 feet above the ground were exposed to phosgene gas after an employee from a separate contractor working underneath them performed a line break [12]. This incident highlights the advantages of involving knowledgeable personnel including a SME in the HIRA to develop a holistic line breaking plan that encompasses more than just the physical act of opening equipment.

2.2 Isolation Considerations

The first action listed under the Do step in Figure 1 is “isolate.” Isolation is the separation of a worker from hazardous energy sources, including electrical, mechanical, hydraulic, pneumatic, chemical, and thermal. In line breaking, an important consideration is the hazard presented by the process fluid. In the event of a failure, different isolation methods present the potential for different leakage rates. Greater process fluid hazards imply lower tolerable leakage rates. The hazard level of an unintended release of a process fluid into the ambient surroundings is characterized by a threshold atmospheric concentration [13]. For example, toxicity hazards are often characterized by a threshold concentration called the Immediately Dangerous to Life and Health (IDLH) concentration. The threshold concentration for a flammability hazard is the lower flammability limit (LFL). For a process fluid that is both toxic and flammable, one typically finds that the IDLH concentration is much lower than the LFL concentration. Hence, if a process fluid that is both toxic and flammable, safety consideration will often dictate a smaller tolerable leakage rate due to the toxicity threshold criteria compared to a process fluid that is only flammable. As another example, a process fluid capable of undergoing flashing during its release (a superheated liquid) will pose a greater hazard level than a fluid that does not flash (a subcooled fluid), because flashing fluids discharge vapor into the atmosphere at a faster rate than simple evaporation.

Common methods of isolation for line breaking, in order of increasing effectiveness in preventing unintended leakage, include (i) valved isolation, (ii) double block and bleed, and (iii) positive isolation (blinds and blanks) [14,15]. These three common isolation methods are shown in Figure 2 and detailed below. Other specialized methods of isolation, such as pipe plugs or inflatable bags, are not described in this paper and should be considered on a case-by-case basis. As previously discussed, the method of isolation should be reviewed during the HIRA, prior to any work being performed. In addition to the nature and severity of the process fluid’s hazards, additional considerations may influence the selection of the method of isolation. These other factors may include, but not necessarily be limited to feasibility, accessibility, and urgency.

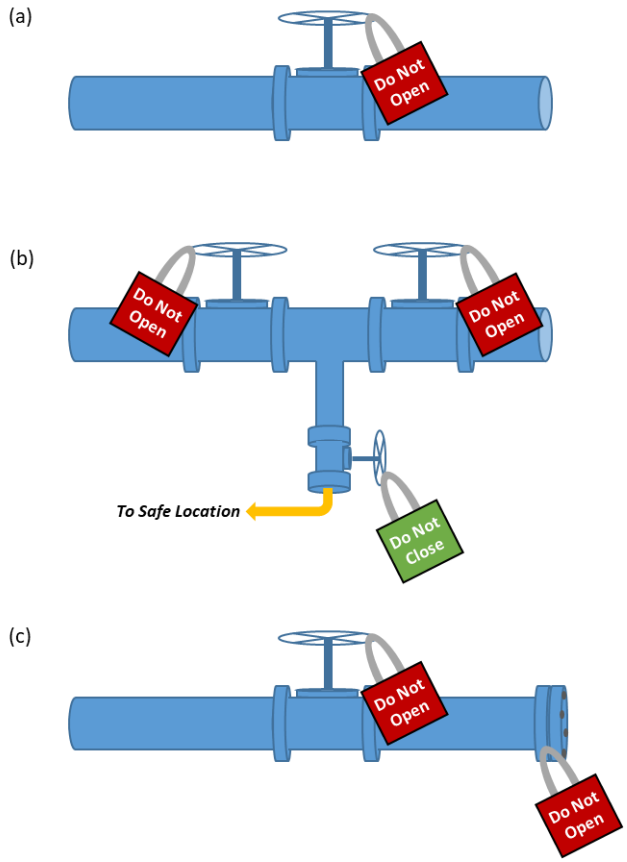

Figure 2. Three common isolation strategies during line breaking are (a) valved isolation, (b) double block and bleed, and (c) positive isolation using a blind or blank (blank flange shown).

2.2.1 Valved Isolation

A closed and locked isolation valve (see Figure 2a) is the simplest isolation method of the three reviewed here, but requires periodic inspection and verification to be effective. Isolation valves can become corroded, jammed, or otherwise mechanically compromised, and it is often not possible to confirm the effectiveness of isolation valve closure prior to breaking into a system [14,15]. Even a configuration with two isolation valves in series is typically not considered the most effective isolation strategy. Isolation valves are also susceptible to human error, such as misalignment or inadvertent opening. For example, a misaligned handle on a natural gas supply plug valve caused pipefitters working at a Minnesota school to believe that the gas piping they were relocating was isolated [16]. In fact, the plug valve was in the open position, and the subsequent gas leak and explosion resulted in nine injuries and two fatalities.

An isolation strategy that relies solely on an isolation valve as protection from hazardous substances should be given careful consideration during the review process. In situations where valved isolation is the only practical or feasible option, particularly for older equipment, additional levels of monitoring and protection may be necessary. The Case Study presented below demonstrates how a loss of containment event occurred through a closed isolation valve that was intended to contain the process fluid (ammonia) during a line break. Additional monitoring and alternative protections that could have been implemented to mitigate the consequences due to the leaky isolation valve are discussed in the Lessons Learned section.

2.2.2 Double Block and Bleed

The double block and bleed method of isolation consists of an open bleed valve located between two closed isolation (block) valves (see Figure 2b). The objective of the bleed valve is to vent hazardous materials to a safe location (e.g., a vent header that is routed to environmental controls), so there will be no pressure accumulation between the two isolation valves even if the upstream isolation valve leaks [17]. However, a build-up of pressure may occur if the bleed line is too small, too long, or discharges into a vent system that is susceptible to backpressure [14]. In one near-miss event, a compressor was severely overpressurized with flammable gas by backflow through a double block and bleed configuration [18]. Operators were unaware that the bleed valve, which had the intended purpose of reliving pressure by routing material to a flare header, was still open as they began to de-isolate the system after maintenance work was completed. Pressure increased in the flare header during the de-isolation process, resulting in significant backpressure occurring between the two isolation valves and ultimately leading to backflow of flammable gas.

An additional concern for the double block and bleed isolation strategy is the ability to adequately handle the potential hazardous material at the safe bleed location. In ammonia refrigeration systems like the one discussed in the Case Study, emergency relief vents typically emit directly to the atmosphere. This may be an acceptable safe location for small leaks, but consequences of large or sustained leaks through the upstream isolation valve should be considered for double block and bleed configurations.

2.2.3 Positive Isolation (Blinds and Blanks)

Positive isolation through blinding or blanking (see Figure 2c) is recognized as the most effective method of isolation during line breaking [14,15]. A *blind* refers to a solid plate inserted between flanges (such as a slip blind or spectacle blind), while a *blank* refers to a solid plate used to cap a disconnected line. Blinds and blanks require temporary isolation for insertion and removal; in other words, the isolation procedure is itself a line break. Therefore, PPE selection is an important consideration for positive isolation strategies. An appropriately sized blind or blank that is compatible with the service fluid and rated to the system maximum allowable working pressure is expected to provide continuous isolation when properly installed. The presence (or lack thereof) of blinds and blanks is also readily apparent to operations and maintenance staff.

2.3 LOTO during Line Breaking

Following isolation and any necessary draining, purging, or flushing, the “Do” step in Figure 1 specifies that LOTO procedures be performed. LOTO consists of at least two administrative controls that are part of an effective isolation, de-energization, and verification procedure [3,4,19]:

- Lockout: The placement of a lockout device (such as a lock) on an energy isolating device to ensure the energy isolating device and the equipment being controlled cannot be operated.
- Tagout: The placement of a prominent warning (such as a tag) on an energy isolating device to indicate that the energy isolating device and the equipment being controlled may not be operated until the warning is removed.

While the purpose of both a LOTO procedure and a line breaking procedure is to control hazardous energy, this paper seeks to emphasize and reinforce the distinctions between LOTO and line breaking. At processing facilities with hazardous chemicals, a primary purpose of line breaking procedures is to

identify an appropriate hierarchy of controls that protect workers [9]. The hierarchy encompasses engineering controls (e.g., energy isolation device), administrative controls (e.g., barricade to restrict access), and PPE. LOTO is clearly an important administrative control during line breaking that helps ensure hazardous materials remain isolated, but it is just one of many controls that should be considered. It is important to recognize that LOTO is more than simply placing a lock and tag on a piece of equipment or valve. Workers must first correctly identify the source(s) of the hazardous energy, so they can be confident that their isolation strategy is appropriate. Next, after LOTO is performed, it is common to attempt to verify (Try Out) the LOTO was successful. Incorporating these additional considerations of LOTO into a line break hazard evaluation can further protect workers from the hazardous energy source. However, as described in the Case Study, a properly implemented LOTO program can be compromised if other inadequacies exist in the execution of a line break.

2.4 Line Breaking Procedures and Permits

Line breaking procedures for routine tasks (e.g., regular sampling for product quality) are usually included in standard operating procedures and do not require a permit [9]. However, nonroutine line breaks performed for the purpose of maintenance or troubleshooting warrant particular attention because they often involve much greater risk than routine operations. There is an expectation that PSM/RMP covered facilities develop nonroutine line breaking procedures and implement a permitting system, but the information that should be included in these procedures and permits is not clearly defined [20]. The following guidance is provided for *Nonroutine Work Authorizations* in the nonmandatory appendix of OSHA's PSM regulation [21]:

> *…A work authorization notice or permit must have a procedure that describes the steps the maintenance supervisor, contractor representative or other person needs to follow to obtain the necessary clearance to get the job started. The work authorization procedures need to reference and coordinate, as applicable, lockout/tagout procedures, **line breaking procedures**, confined space entry procedures and hot work authorizations… [**emphasis added**]*

Operating procedures are generally written in sufficient detail such that a qualified worker can consistently and successfully perform the task [22]. But as the term "nonroutine" suggests, it is not possible to develop a one-size-fits-all procedure to cover the potential hazards for every line break scenario. The uncommon nature of nonroutine tasks also adds to the challenge of reliable execution, since workers will likely only have limited or infrequent experience performing the job. Typically, line breaking procedures require a review of potential hazards, such as those discussed in Section 2.1 above, with the idea that a more detailed plan will be incorporated into the line breaking permit following the review. This review also allows workers with limited experience opening a particular piece of equipment or piping to become familiar with the potential hazards. Considerations for effective line breaking procedures and permits are presented in the Lessons Learned from the Case Study.

2.5 Response to Suspected Leaks

Even with appropriate work practices and controls, there is still risk associated with opening piping and equipment that contains hazardous material. Industry guidance documents can be used to develop a response plan if leaks occur during line breaking activity [23,24]. For example, *API RP 574* recommends that a safety review meeting should be conducted to establish a hot zone around the leak site with specific PPE requirements to perform work inside this zone. The review should consider the

temperature, pressure, and remaining inventory of process fluid within the piping, as well as potential damage mechanisms and related factors. *API RP 574* also cautions the incident response team in making assumptions about the leak's cause, noting that incidents have occurred where investigative personnel assumed they knew the cause of a small leak on an operating line and were caught unprepared when the leak suddenly became quite large. The Case Study describes a situation where operators believed they understood and isolated the source of an ammonia leak that occurred following a line break, only to initiate a second emergency shutdown after the uncontrolled ammonia release persisted.

3 Case Study: Release from Ammonia Refrigeration System

This Case Study describes a release of anhydrous ammonia from a refrigeration system at a frozen foods warehouse. Ammonia has been used as a refrigerant dating back to the 1850s and remains popular in industrial refrigeration systems [25]. While generally considered to be environmentally benign, ammonia is potentially hazardous to human health with an IDLH of 300 ppm. Ammonia is also flammable, with an LFL of 16% in air [26]. Additionally, facilities with more than 10,000 pounds of anhydrous ammonia are covered under PSM/RMP [27]. The ammonia release discussed here occurred following a line break performed during compressor maintenance on a PSM/RMP covered process. Lessons Learned from the incident are discussed in the following section.

3.1 Ammonia Refrigeration System Overview

A simplified process flow diagram (PFD) for the subject ammonia refrigeration system is shown in Figure 3. Three parallel screw compressors discharge hot, pressurized ammonia gas to an evaporative condenser. The condensed ammonia flows through three cascaded Receivers to a Surge Vessel, with letdown valves between each stage. The total liquid ammonia holdup in the Receivers and Surge Vessel is greater than 10,000 pounds. Liquid ammonia is extracted at various pressures to provide refrigeration for a cooler, freezer, and ice cream storage area located within the distribution warehouse. The ammonia is partially vaporized as heat is transferred to the refrigerant. Ammonia vapor collected from the Medium Temperature Receiver is recycled to the suction side of the parallel screw compressors.

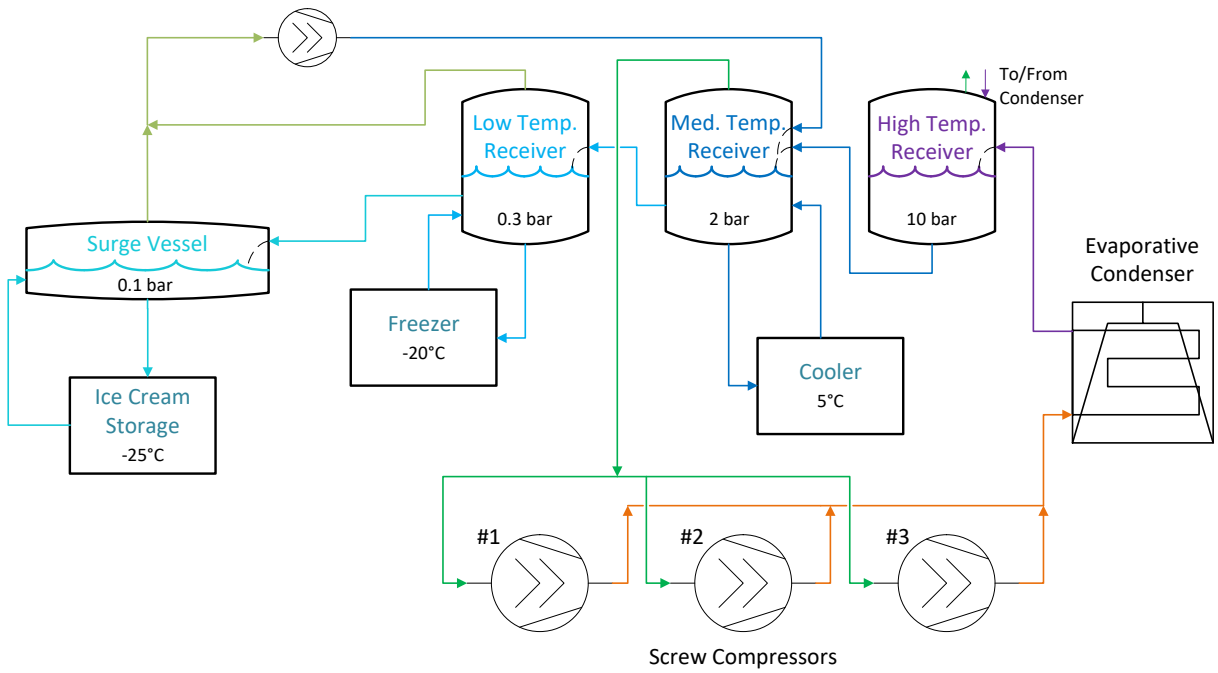

Figure 3. Simplified PFD for PSM/RMP covered anhydrous ammonia refrigeration system discussed in this Case Study.

3.2 Line Break during Compressor Maintenance

The screw compressors at the facility were scheduled to be rebuilt as part of a capital improvement project to increase efficiency and enhance system reliability. A contractor experienced with the subject ammonia refrigeration system was responsible for performing the maintenance and was working to remove Compressor #2 during the day of the incident. Facility personnel recognized that removing Compressor #2 constituted a line break and filled out a Line Breaking Permit using the company's standard form. The Permit specified the following requirements for removing Compressor #2:

- Notification of affected personnel
- Use of PPE (respirator and gloves)
- Depressurization of piping prior to opening
- LOTO of electrical and mechanical equipment
- Positive forced ventilation active during work

The Permit did not include any additional details on how to accomplish the requirements listed above. Notably, a procedure was not developed for effective isolation of the system. Once the Permit was signed, contractors began work to remove Compressor #2. As shown in Figure 4, the suction and discharge lines for Compressor #2 were isolated by closing and locking Gate Valves #1 and #2 on the suction side and Y-pattern Globe Valve #1 on the discharge side. The disconnections occurred at Compressor #2's flanges, which were located downstream from Gate Valve #2 and upstream from Y-pattern Globe Valve #1. Ammonia detectors in the compressor building briefly alarmed as contractors removed Compressor #2, but all workers in the area were wearing appropriate PPE with the positive forced ventilation system activated.

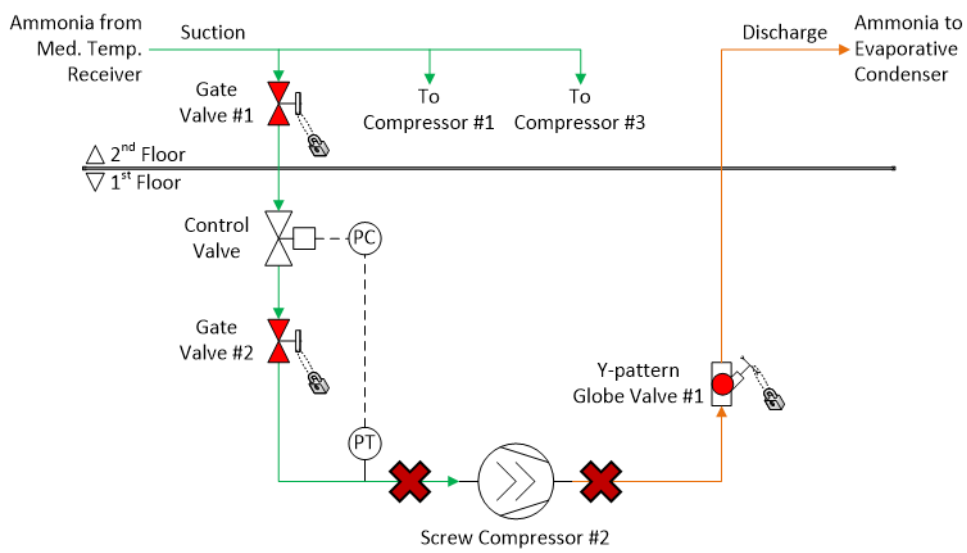

Figure 4. Isolation strategy used during the removal of Compressor #2.

3.3 Ammonia Leak and Incident Response

Following the removal of Compressor #2, Gate Valves #1 and #2 and Y-pattern Globe Valve #1 remained closed and locked. Facility personnel believed that the LOTO procedures they had executed (see Figure 4) were sufficient to control the hazardous energy, and no positive isolation strategies were implemented. Compressors #1 and #3 were running and the refrigeration system appeared to be operating normally at the end of the day shift.

The distributed control system (DCS) for the subject refrigeration process was monitored remotely by the system vendor, with no operations staff onsite during the overnight hours. Shortly after workers left the facility on the day Compressor #2 was removed, ammonia gas detectors began to continuously alarm in the compressor area. This caused the positive forced ventilation system to activate automatically. However, the gas alarms did not automatically trigger an emergency shutdown because the system shutdown procedure required operators to manually close certain valves. Process data show that the level in the High Temperature Receiver started to slowly drop following the initial detection of ammonia. The gas alarms at the facility are recorded in the DCS, but the system did not notify local operators of the alarm. Gas detectors continued to alarm throughout the night, at times reaching the instrument's upper detection limit. Level measurements in the Medium and Low Temperature Receivers also began to drop.

A refrigeration operator reported to work the following morning and discovered an ammonia release in progress upon opening the door to the mechanical building that housed the screw compressors. The operator immediately evacuated the area and initiated an emergency shutdown of the refrigeration system. The contractors who had performed the maintenance on Compressor #2 were called back to the facility and informed that the refrigeration operator had observed a cloud of ammonia in the area of Gate Valve #2. Based solely on the operator's visual observation, it was determined that a small ammonia leak had occurred through the closed and locked Gate Valve #2 on the suction side of Compressor #2. Contractors removed and replaced Gate Valve #2, which had previously been scheduled for replacement as part of the facility's mechanical integrity program, and installed a blank on the downstream side of the new valve. No additional leak investigation was conducted.

The system was restarted following the replacement of Gate Valve #2. No modifications were made to the discharge side isolation strategy. After two hours of operation, ammonia detectors again alarmed in the compressor area at the same time operators smelled ammonia and visually observed liquid droplets at the Y-pattern Globe Valve #1. A second emergency shutdown of the refrigeration system was initiated. Facility personnel determined that the source of the second ammonia leak was Y-pattern Globe Valve #1, depicted in Figure 7. Contractors used a re-seating tool on the globe valve to ensure uniform contact between the valve plug and seat ring. Inspection of the valve bonnet also revealed wear on the packing. The facility did not have replacement packing in stock, so the bonnet of Y-pattern Globe Valve #1 was capped to contain any possible leaks through the packing. The system was restarted for a second time and no subsequent ammonia releases were detected.

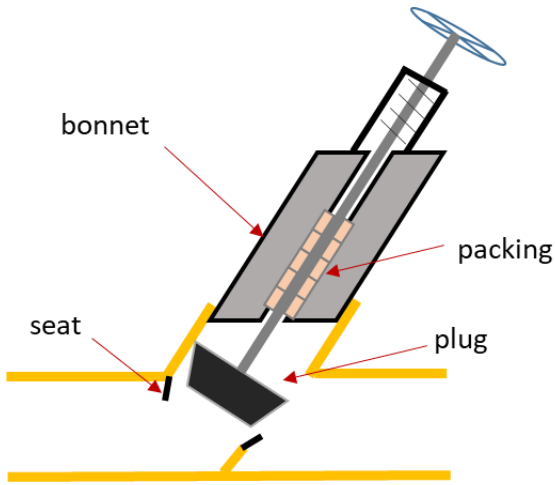

Figure 5. Schematic of Y-pattern Globe Valve #1 installed on the discharge side of Compressor #2.

3.4 Incident Summary

Key events leading up to and during the ammonia release are summarized in Figure 6. Although there were no known onsite or offsite injuries from the incident described here, a reportable quantity of ammonia was released under the EPA's Emergency Planning and Community Right-to-Know Act (EPCRA) [28]. Lessons Learned from the incident are discussed in the following section.

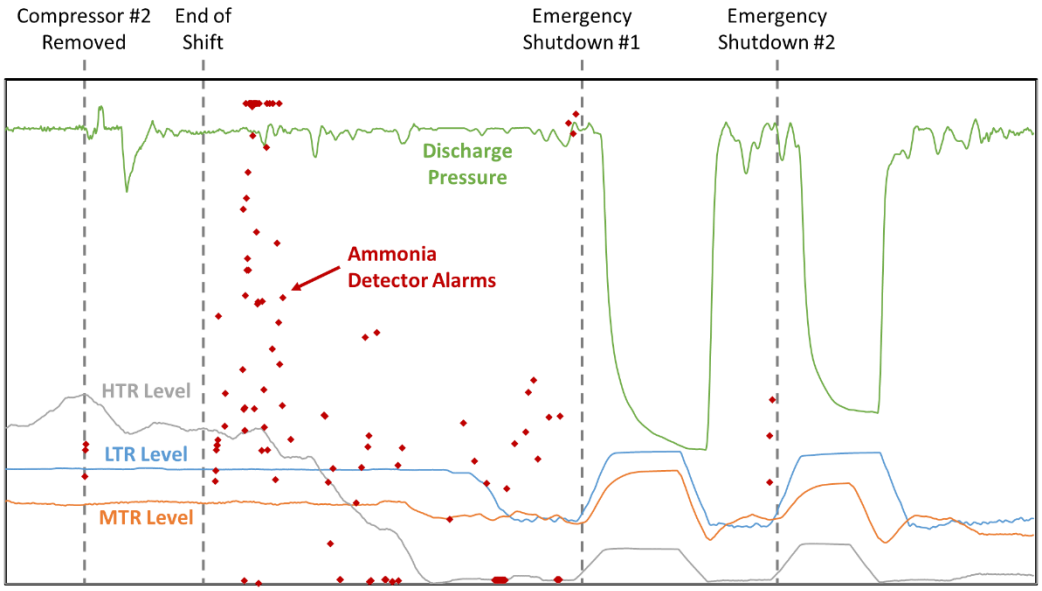

Figure 6. Key events, process data trends, and alarms related to the ammonia release described in the Case Study.

4 Lessons Learned

4.1 Line Breaking and LOTO Procedures Are Related yet Distinct

A dedicated line breaking procedure that is distinct from LOTO procedures can help ensure consistent safe work practices when opening piping that may contain hazardous materials. While a line breaking procedure existed at the facility described in the Case Study, operators largely viewed it as an extension of LOTO. The operators believed that properly implementing LOTO on the valves previously shown in Figure 4 would provide continuous, reliable isolation, and therefore the only isolation requirement included in the Line Breaking Permit was LOTO. A more thorough HIRA may have exposed limitations in this isolation strategy.

Figure 7 provides an example of a line breaking procedure that was presented in a document produced under OSHA's grant program. While this example procedure is similar to the line breaking procedure that was in place for the subject ammonia refrigeration system, the "Hazard Review" step in the sample document was not listed in the subject facility's line break procedure. Facilities should consider emphasizing the importance of conducting a hazard review in their line breaking procedures, with the goal of highlighting that effective isolation strategies commonly involve many other considerations in addition to implementing a LOTO procedure. Additionally, incorporating more specific guidance on possible isolation strategies into line breaking procedures (e.g., additional guidance if a simple valved isolation is implemented) may assist facility personnel in making more informed decisions. Finally, a requirement within the line breaking procedure to involve a SME in the review step may increase the likelihood that an effective isolation strategy is developed.

PROCEDURE OVERVIEW

1. SCOPE

WESTVACO POLICY:
No person may break into, open, or disassemble any equipment which might contain hazardous material, pressure or temperature until a L.E.O. permit has been completed.

2. LINE and EQUIPMENT OPENING PERMIT

WESTVACO has established a permit system to be followed during line breaking and process equipment opening which might contain hazardous material, pressure or temperature.

3. HAZARD REVIEW
*Knowledge needed of:
1. Hazardous material in system
2. Potential physical hazards
3. Barricading as required
4. First aid if exposed
5. Location of safety equipment

4. PREPARATION and ISOLATION
*Prior to initiating line and equipment opening operations/maintenance shall insure the following is complete.
1. Isolated
2. Drained
3. Purged
4. Flushed
5. Lock out/Tag out
6. Heat tracing de-energized
7. Pipe supports as needed
8. Other (special instructions supplied by supervision)

5. PROTECTIVE EQUIPMENT PACKAGE REQUIRED
*hard hat/safety glasses/safety shoes required in all areas.

PACKAGE A:	PACKAGE B:	PACKAGE C:
Rubber Gloves	Rubber Gloves	Rubber Gloves
Protective Boots	Protective Boots	
Chemical Goggles	Chemical Goggles	Chemical Goggles
Face Shield	Face Shield	Face Shield
Chemical Suit	Chemical Suit	
Respirator		Respirator

PACKAGE D:	PACKAGE E:	Other:
	Rubber Gloves	To Be Specified
Chemical Goggles	Chemical Goggles	
Face Shield	Face Shield	

6.PROCEDURE
Follow above steps for LINE AND EQUIPMENT OPENING.

7.START UP:
Inspect work area to verify completion of work before start-up.

Figure 7. Example Line and Equipment Opening (L.E.O.) procedure presented in a document produced under OSHA's grant program that requires a hazard review and the completion of a permit before a line break may be performed [10].

4.2 Permit Systems for Line Breaks Can Help Promote Safe Work Practices

In a 2019 investigation report, the U.S. Chemical Safety and Hazard Investigation Board (CSB) concluded that "consistently using safe work practices, such as line-breaking permits, ensures that each time a piece of process equipment is put into a nonroutine situation, personnel use effective safeguards" [29]. The CSB also noted that it is important to train personnel on different situations that would require a line break permit to ensure the permit process is followed, particularly under nonroutine conditions. The operators at the subject ammonia refrigeration system did in fact recognize that a line breaking permit was required to remove Compressor #2. However, the completed permit only contained a generic checklist that lacked detail on how to provide adequate safeguards. Facilities should consider a

requirement to attach the procedure developed during the HIRA to line breaking permits, which may help promote a more effective permitting system. A permit that consists of only a simple checklist used for all line breaks may not provide facility personnel with sufficient detail to enact appropriate controls in every scenario.

4.3 A Robust Mechanical Integrity Program Can Help Inform Isolation Strategies for Line Breaking

The intent of a mechanical integrity program is to ensure that equipment is properly designed, installed, and remains fit for use until it is retired. A successful mechanical integrity program typically includes the following components [30]:

- A mechanism to ensure that equipment is maintained in a manner appropriate for its intended application,
- A preventative maintenance program that reduces the need for unplanned maintenance, and
- An inspection and testing program that helps recognize when equipment deficiencies occur and reduces the likelihood that deficiencies lead to serious accidents.

Maintenance records for Compressor #2 indicated that the suction and discharge valves were included in the annual maintenance inspection program and were determined to be in satisfactory condition a year prior to the incident. Both the suction and discharge valves, however, were nearing the end of their expected useful lives and were scheduled to be replaced as part of the Compressor #2 maintenance activity. Had the facility leveraged their mechanical integrity program records as part of a line break HIRA, an alternative isolation strategy may have been selected that did not rely on isolation valves that were scheduled for replacement. Implementing a robust mechanical integrity program, and using it when reviewing hazards for a line break, can help manage risks when opening piping and equipment.

4.4 Effective Monitoring During Line Breaking Can Help Mitigate Consequences of Leaks by Improving Timeliness of the Leak Response

Failure to monitor and act upon the ammonia detector alarms allowed the ammonia release described in the Case Study to persist unchecked during the overnight hours at the unmanned frozen foods warehouse. Monitoring for hazardous gas releases is particularly important during nonroutine operations such as line breaking. Following the incident, an automatic notification system was implemented so that the on-call operator will receive an alert to their phone in the event of an ammonia alarm. This new system allows for a more timely response to ammonia releases that occur outside of working hours. It also provides operators with advance warning so that they may take appropriate action, such as donning a respirator or acquiring a handheld gas detector, prior to entering a building where a leak may be occurring.

4.5 Leak Investigations Should be Considered When an Unanticipated Release Occurs

As Figure 6 shows, gas detectors briefly alarmed at the same time Compressor #2 was disconnected from the piping system. This small ammonia release associated with residual low-pressure gas within the removed equipment was anticipated — the line breaking permit required respiratory protection, forced ventilation, and restricted access to prevent worker exposure to residual ammonia gas. Therefore, it is

unlikely that the facility would initiate a leak investigation associated with the gas detector alarms that occurred during the equipment removal.

The release that was discovered the following morning, however, was unexpected. Had this release triggered a leak investigation, facility personnel would likely have reached two significant conclusions: (1) the leak was much larger than initially thought (due to the reduction in level in the HTR, MTR, and LTR) and (2) the discharge isolation valve was also a possible source of the leak. Despite the operator's visual observation of an ammonia cloud near the suction isolation valve, any leak from the suction side must have passed through two closed gate valves and a closed control valve. Additionally, the suction side of the compressor operates at much lower pressure compared to the discharge side. While it certainly is possible for a leak to occur on the suction side, the magnitude of the release should have placed suspicion on the single discharge isolation valve that was responsible for containing the higher-pressure side of the gas systems. Had operators fully appreciated the size of the leak, it is likely they would have taken action to improve the isolation strategy on *both* the suction and discharge side isolation valves before restarting the system.

5 Conclusion

Unfortunately, the Case Study described here is not an isolated occurrence — numerous unintended ammonia releases have been documented that were purportedly associated with line breaking activities [31,32]. More broadly, line breaking can be a nonroutine yet ubiquitous task at process facilities that is inherently hazardous. Despite U.S. regulations requiring PSM/RMP facilities to control hazards during line breaking, the regulations do not provide prescriptive requirements and RAGEGEP for line breaking is not fully developed. This paper has presented several Lessons Learned from an ammonia release that occurred following a line break on a refrigeration system. These learnings represent good engineering practices that can help reduce risk associated with opening piping or equipment that potentially contains hazardous material.

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Process Related Incidents with Fatality and the Effectiveness of the Process Safety Management Program

Syeda Z. Halim*, T. Michael O'Connor, Noor Quddus, and Stewart W. Behie

Mary Kay O'Connor Process Safety Center, Texas A&M University System

College Station, TX 77843, United States

zohrahalim@gmail.com

Abstract

A database of the Occupational Safety and Health Administration (OSHA) captures incident data from investigations for fatal incidents and hospitalizations since 1984. OSHA Region 6 includes 5 states including Texas and Louisiana, where much of the US chemical manufacturing and petroleum refining industry is located. An analysis of process related investigations by OSHA in Region 6 shows that large-scale multi-fatality incidents have been significantly decreased since the implementation of Process Safety Management (PSM) program in 1995. It is noticeable that currently majority of the fatalities occurs in single fatality incidents. Our preliminary analysis suggests that these individual process related fatalities are a result of operating and maintenance activities that are not well addressed by current process safety practices or by personal safety measures. An analysis of such incidents and their circumstances will be conducted proving recommendations for improved performance to reduce the incidents with single fatality.

Keywords: lessons learned, work control, procedures,

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Application of Mind Mapping to Classify and Recall Potential Hazards

T Michael O'Connor and Ingry Ruiz

Mary Kay O'Connor Process Safety Center

mikeocon@gmail.com

Abstract

Applying mind-mapping to planning and executing workplace activities could provide a means of reducing errors in common tasks such as welding/cutting, filling/emptying tanks and confined space entry. Trevor Kletz, one of the "founding fathers" of process safety engineering (Vechot et al, 2014), emphasizes the inability of organizations and individuals to learn from past mistakes in his final book, "*What Went Wrong?: Case Histories of Process Plant Disasters and How They Could Have Been Avoided, 5th ed. (2009)"*. Due to the complexity of both human and organizational behavior, finding a solution to previous mistakes has been an immense challenge for many companies. As a result, more effective methods to increase individual and organizational learning are needed. Mind-mapping appears to offer a means to organize and and assist in the recall of hazards that have resulted in previous incidents.

This paper describes a methodology for examining previous incidents pertaining to certain common tasks. The emphasis in this effort is to help identify hazards that are hidden from view, and/or not normally encountered and to organize them in a way that is easy to recall or identify. Applying this method can improve procedures and work permits, as well as training and as a check by those directly involved before undertaking a hazardous task.

Keywords: Lessons Learned, Hazard Recognition, Hot Work, Permitting, Confined Space, Tank Filling, Hazardous Substances, Work Control

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Would a HAZOP, LOPA or STPA Have Prevented Bhopal

Author: Howard Duhon, P.E.
Co-Author: Pranav Kalantri
GATE Energy
13501 Katy Freeway, Houston, TX 77079

hduhon@gateinc.com

Abstract

In the early morning hours of Dec 3, 1984, a large toxic gas release from a Union Carbide India Limited (UCIL) pesticide plant in Bhopal, India swept over a large, densely populated area south of the plant. About 500,000 people downwind were exposed to the gas cloud. Thousands of people died in the immediate aftermath (we don't know how many) and tens of thousands were severely injured.

Bhopal has changed the way we think about safety culture. PHAs such as HAZOPs have become popular largely because of Bhopal. In this paper we ask and attempt to answer three questions:

- Would a HAZOP have prevented Bhopal?
- Would a LOPA have prevented Bhopal?
- Would an STPA have prevented Bhopal?

To be clear, the question is not whether these technologies, applied today, would prevent a Bhopal-like tragedy in the future. The question is, would any of these technologies, applied to the Bhopal design in the 1960's, have prevented the accident or mitigated the impact.

Keywords: Bhopal, HAZOP, STPA, LOPA, PHA

1 Background

1.1 Bhopal – The City

Bhopal is a bustling metropolis of 2 million people. The city and surrounding area are home to a large open- air zoo, a fascinating museum of Indian tribal life, a collection of historical palaces and temples and a cave with stone-age paintings that is a UNESCO World Heritage site.

It is located in the geographic center of India. Excellent rail connections make it an attractive location for manufacturing. That is one reason why Union Carbide India Limited (UCIL) chose to locate an agricultural chemicals plant there in the 1960's.

1.2 The Accident

In the early morning hours of Dec 3, 1984, a large toxic gas release from that plant swept over the city.

Thousands of people died in the immediate aftermath. We don't know how many. A commonly accepted estimate is 2000 (2007, D'Silva), but it may be as high as 8000 (2004, Amnesty International). Tens of thousands were severely injured. Thousands of those died prematurely of their injuries in the months and years following the release.

And the accident caused significant social and economic problems. An already poor area was made much poorer. Young women exposed to the gas cloud carry a social stigma and have had great difficulty finding husbands.

1.3 Personal Connection and Site Visit

I worked for Union Carbide Corporation (UCC), the parent company of UCIL, at the time of the accident, though not in the agricultural products area of the business.

While traveling in India in 2013 I took the opportunity to visit the site.

The plant has been idle since the accident, rusting away, overgrown with trees and shrubs; caught in a struggle between people who want it torn down and others who want it to be preserved as a UNESCO World Heritage Site.

Thirty-five years on the accident is still alive in the neighborhood around the plant. Billboards and graffiti demand restitution. Hospitals and rehabilitation centers, there to treat the injured, abound.

1.4 Purpose of this Paper

The Bhopal accident and other deadly accidents in that period have had a marked effect on safety culture in the processing industries. The widespread development and adoption of Process Hazard Analysis (PHA) methods is one result.

In this paper I present a thought exercise to investigate whether earlier adoption of PHAs would have prevented Bhopal.

- Would a HAZOP have prevented Bhopal?
- Would a LOPA have prevented Bhopal?
- Would an STPA have prevented Bhopal?

1.5 The Root Cause?

Questions remain about what happened. It is very difficult to study an accident of this magnitude because so many people had/have vested interests in crafting the story. I'll present my best estimates, my opinions, of what happened.

Trevor Kletz has argued that there is no such thing as a root cause – only a point at which we stop asking questions. In this case, D'Silva (2007) has argued that it is appropriate to take the inquiry back to the days of the Raj, the colonial occupation of India, because the residue of colonialism impacted the psyche the people and the political and legal systems of the country in ways that contributed to the tragedy.

Nancy Leveson has made a more definitive argument against the idea of root causes with her STPA technology. Charles Perrow (1984) made similar arguments in Normal Accidents. Accidents are best understood as the results of loss of control of complex systems.

I agree that there is no such thing as the root cause, but it is a useful fiction for the analysis that follows.

1.6 The Legal, Political, Social Environment

General poverty in the country and abject poverty in the neighbourhood around the plant.

Recent independence after centuries of colonial rule.

Pro-India and anti-foreign sentiment.

All technology, once imported into India was, by law, to become Indian technology.

"Indian hands" must have control over decision-making

Government approval for foreign workers required - replacement of foreign workers ASAP.

Very difficult to get profits out of the country – unable to export cash, Union Carbide funded a shrimping fleet and exported shrimp!

Lack of safety culture.

Increasingly restrictive laws led to the exodus of many established foreign companies including IBM and Coca Cola. But UCC stayed.

UCIL the local subsidiary of UCC owned property around the plant that was supposed to be a buffer zone. A shanty-town developed on that property. Some homes even used the plant's concrete fence as one wall of their home. UCIL was powerless to evict the squatters.

1.7 Losing Money

The Indian market for pesticides was much smaller than 'anticipated'. The over-estimate was partly due to wishful thinking. It was partly due to a drought in the country that impoverished farmers who then could not afford to buy pesticides. It was also partially due to culture – the slower than expected adoption of the idea of chemical pesticides among subsistence farmers.

Actually, UCC knew that the plant was dramatically oversized, but had to accept the Indian governments projections to get approvals. The plant never operated at more than 50% of capacity and operated at 25% capacity in 1984.

Problems with the plant design made it much more expensive to produce pesticide in India than it would have been to import it.

UCIL's competitors were the low-cost producers for two reasons. Some local providers sold unproven products. Indian government provided incentives for local producers.

A decision had been made to close the MIC facility, dismantle it, move it out of the country. The MIC that was vented was the last batch of MIC planned to be produced.

At the time of the accident the MIC plant was shut down and was being decommissioned. A large quantity of MIC was stored, due to be reacted into pesticide in the next few days.

Many of the employees were about to be unemployed.

1.8 Safety Culture

UCIL management was unable to instil an appropriate safety culture. This may have been difficult in any case, but was exacerbated by the fact that the company was losing money. In any part of the world, safety is one of the first things to suffer in the name of economy.

An antagonistic relationship existed between management and workers that made it very difficult to study accidents and near misses. Workers lied and covered up their actions to avoid blame.

1.9 The Assassination of Indira Gandhi

The political situation at the time was particularly tense. The last batch of MIC was produced in late October,1984. It would have been converted into pesticide soon after that, but the assassination of Indira Gandhi resulted in riots and the plant was shut down for the month of November.

2 Process/Plant Description

MIC is an intermediate in the development of a family of pesticides called carbamyls of which Sevin was the principle product of the Bhopal pesticide plant.

Phosgene Plant:	CO + Cl ➔ Phosgene	
MIC Plant:	Phosgene + Methylamine ➔ MIC	[Continuous Plant]
1-Napthol Plant	(new technology, worked at pilot scale, never worked at full scale)	
Sevin Plant:	MIC + 1-Naphthol ➔ SEVIN pesticide	[Batch Plant]

Figure 1 illustrates the important features of the pesticide production facilities focusing on MIC storage.

- MIC was produced onsite (MIC Production Plant). This was a continuous process.
- MIC was consumed onsite as a raw material in the Pesticide Plant (MIC Consumer). This was a batch process.
- MIC storage was required between the continuous and batch plants. MIC storage was supposed to be kept to a 'minimum'.
- A Caustic Scrubber was provided to neutralize MIC vented from the storage tanks.
- A flare was provided to burn MIC vented from the storage tanks.
- A refrigeration system was provided to keep the stored MIC cold. Keeping the MIC cold decreases the reaction rate of MIC with water and other contaminants.

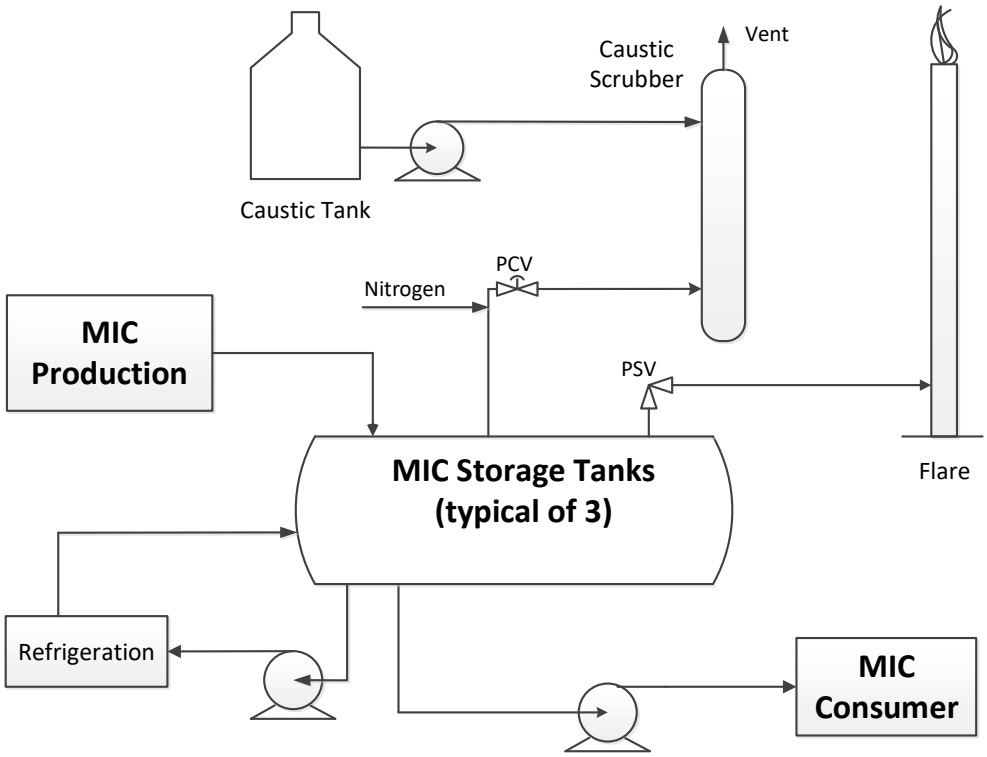

Figure 1 – Plant Process Schematic

3 The Accident

3.1 Initiating Event

People who believe in root causes also believe in initiating events or triggering events. The triggering event for this accident was the introduction of a large amount of water into one of the MIC storage tanks E-610. The MIC-water reaction is highly exothermic resulting in a massive gas release.

There is dispute about how the water got into the tank. Early speculation centered on a filter washing operation, without proper isolation, in another part of the plant. But I think it is now well established that the triggering event was sabotage – water intentionally introduced into the tank by a disgruntled employee.

The source of the water is not important for our current discussion. Water introduction into the tank should not have created a catastrophe, regardless of the source of the water.

3.2 Safeguards Bypassed or Broken

Several safeguards were designed into the plant to prevent an MIC release or at least to minimize its impact. The safeguards were probably adequate for handling any typical initiating event. They may not have been adequate to handle the quantity of water injected into the tank that day. But we will never know that, because all of the safeguards were bypassed, out-of-service, or otherwise rendered ineffective.

3.2.1 No Means of Adding Water to the Tank

It is common in industrial facilities to have vent and drain valves on piping systems to make it easy to vent and drain systems and to inject water, steam, nitrogen, air, etc. for purging or cleaning the system. To avoid accidental water injection the Bhopal facility was installed without drain valves or vent valves.

In fact, this was hardly an impediment. It appears that the saboteur simply removed a pressure gauge and installed a hose connection in its place (1988, Kalelker).

3.2.2 Minimizing MIC Stored Volume

There were three storage tanks, E-610, E-611, E-619. The tanks were very large – 8 ft in diameter x 40 ft long with a capacity of 15000 gal.

E-619 was dedicated to storage of off-spec MIC. It was empty.

E-610 and E-611 were provided for MIC storage. The plant safety manual noted that each tank was larger than necessary and specified that one tank should always be empty and the other less than 50% full.

In fact, both E-610 and E-611 were 70% full.

3.2.3 Refrigeration System Out of Service

The rate of an exothermic reaction is decreased by decreasing the temperature. A refrigeration system was provided to keep the MIC at about 0 ^{o}C, 32 ^{o}F. Had the tank been operated at that temperature the initial reaction rate would have been much lower and the event may have been less catastrophic.

Ironically, the refrigeration system was turned off long before the accident at least partly as a safety measure. The pump provided to circulate MIC through the refrigeration unit was subject to seal leaks. After one catastrophic seal failure the refrigeration system was shutdown permanently.

Two pumps are shown on Figure 1. The pump provided to transfer MIC to the pesticide plant was also prone to seal leaks and had also been shut down. Instead of pumping MIC to the Sevin Plant, the MIC tank was pressurized with nitrogen to transfer MIC.

3.2.4 Caustic Scrubber Failure

The vented MIC escaped through the Vent Gas Scrubber (Caustic Scrubber). The scrubber was provided to neutralize vented MIC with caustic (sodium hydroxide).

There are conflicting reports on the operation of the scrubber; that the scrubber was out of service for maintenance, that the caustic tank was empty as a cost cutting move, that the system was operating, but that the flowmeter was not working, hence we have no direct evidence that caustic was pumped to the scrubber.

All of that is irrelevant. The scrubber was designed for venting rates much lower than was experienced that day. The gas flowrate that night was 4 to 5 times the scrubber design rate. At that flowrate the caustic would likely have been blown out the top of the column with little contact with the gas.

3.2.5 THE FLARE WAS OUT OF SERVICE!

A section of pipe in the flare header was corroded and the flare was taken out of service.

This is almost unfathomable! Why was this allowed to happen?

Perhaps, since this was the last batch of MIC to be processed, the risk of operating for a few days without the flare was deemed an acceptable risk. Maybe there were plans to repair the pipe prior to starting up the pesticide plant, though I haven't seen that reported. It is likely that plant management was simply unaware of the compromised flare system.

3.2.6 Shanty Town in the Plant Buffer Area

UCIL bought a significant amount of property around the plant that was to be a buffer area.

Many of the people who died lived in that buffer zone.

The poorest of the poor set up a shanty town on UCIL property along the plant fence, some literally using the plant's concrete fence as one wall of their house. UCIL tried multiple times to have the shanty town removed, but was unsuccessful; the shanty town residents were voters and local politicians supported them.

Indeed, the squatters were eventually given property deeds to the land that they occupied.

3.2.7 Ineffective Emergency Response

No on-duty UCIL employees were killed in the event. This is largely due to relevant knowledge – the plant operators knew what was happening and which direction the wind was blowing, so they knew which way to evacuate.

An effective emergency response would undoubtedly have saved many in the community. UCIL issued no meaningful alarm to the community and provided no information to civil authorities until about 2 hours after the initial release.

The impacted citizens ran – of course. That was probably the worst thing they could have done. A wet towel over their heads would probably have been far more effective. But they probably didn't know that and certainly didn't know what was happening.

3.2.8 Ineffective Treatment of the Injured

A final safeguard would have been effective treatment of the injured. In the immediate aftermath the doctors didn't know the cause of the incident. And even if they had known, they would not have known how to treat the injured and would not have had the necessary supplies.

3.2.9 Internal Communication Failures

This is a remarkable series of defeated safeguards and it seems incredible that a plant would be operated that way. As I read the various accident reports I got a sense that these decisions were made by different people at different times. It is possible that no single person knew that all of these safeguards were out of service.

It is a fundamental weakness of defence in depth that an individual can bypass a single safeguard, secure in the knowledge that the other safeguards will provide protection.

Normalization of deviance, a term popularized by Diane Vaughn in The Challenger Launch Decision, probably contributed. If you bypass safety systems long enough and continually get away with it, it begins to look normal.

4 Would a HAZOP, LOPA or STPA have Prevented it?

I'll now turn my attention to the main purpose of this article – speculation as to whether a HAZOP, LOPA or STPA would have prevented Bhopal.

To be clear, there is little doubt that any of these methodologies applied today, given our current safety culture and 20/20 hindsight would prevent a Bhopal type tragedy in the future. The question on the table is whether any of them would have made a difference if applied 50 years ago.

Prior to the analysis, we need to be humble about our ability to predict the past in this way. We humans suffer from many cognitive biases that limit our ability to effectively perform such a study.

4.1 The Dunning-Kruger Effect

Perhaps the most appropriate cognitive bias to consider at this point is the Dunning-Kruger effect. Dunning and Kruger noted that we overestimate our ability in areas where we are not competent. They noted "If you're incompetent, you can't know you're incompetent... The skills you need to produce a right answer are exactly the skills you need to recognize what a right answer is."

I was certainly incompetent at assessing the Bhopal design at the time of the event. When I first saw the Bhopal P&IDs a few weeks after the accident, I considered it to be an appropriate design. And all of my colleagues thought the same. The Dunning-Kruger effect in action! These were opinions that we were clearly not competent to make at the time, but which we held

very confidently. The safeguarding provided was typical; the kind of design we expected see. I think that we believed that two or three levels of protection were enough, no matter what the underlying inherent risk.

4.2 Performing the Bhopal HAZOP

Using the Risk Matrix in Figure 2, the MIC HAZOP for water injection would likely have resulted in the following:

NODE:	MIC Storage Tank
Deviation-Guideword:	Different Flow
Cause:	Scenario 1: Large amount of water enters tank via human error
Safeguards:	No hose connections
	Procedures to isolate for maintenance
	High Pressure Alarm
	High Temperature Alarm
	PSV, Flare
	PCV, Scrubber
	Control storage volume (only 1 tank used and not more than 50% full)
	Diluent
Risk Ranking:	Frequency Judgement: A or B
Consequence:	5 - Multiple Fatalities
	Yellow (see Figure x)

NODE:	MIC Storage Tank
Deviation-Guideword:	Different Flow
Cause:	Scenario 2: Large amount of water enters tank via sabotage
Safeguards:	High Pressure Alarm
	High Temperature Alarm
	PSV, Flare
	PCV, Scrubber
	Control storage volume (only 1 tank used and not more than 50% full)
	Diluent
Risk Ranking:	Frequency Judgement: A
Consequence:	5 - Multiple Fatalities
	Yellow

My sense, humbly proffered, is that a HAZOP done on the Bhopal MIC Storage Tanks in the 1970s would have returned a yellow risk ranking and, at best, would have result in minor design modifications. The recommendations might well have referred to operating procedures rather than physical changes.

It is conceivable, maybe likely, that a HAZOP team would have seriously questioned the size of the MIC storage tanks, but in fact, the size of the tanks had already been seriously questioned.

There is an important change that a HAZOP might well have recommended. The refrigeration system design featured a pump that circulated MIC through the refrigeration system. A HAZOP may have flagged pump seal leaks as a significant hazard and resulted in design improvements in that system.

5 - CATASTROPHIC multiple fatalities	Yellow	Yellow	RED	RED	RED
4 - MAJOR single fatality	Green	Yellow	Yellow	RED	RED
3 - SEVERE severe lost time injury	Green	Green	Yellow	Yellow	RED
2 - MINOR reportable injury	Green	Green	Green	Yellow	Yellow
1 - SLIGHT first aid	Green	Green	Green	Green	Yellow
	Never Heard of in the Industry	Happens in the Industry	Happens in our Company	Likely to Happen at this Plantt	Likely to Happen Multiple Times at this Plant
	A	B	C	D	E

Figure 2 – HAZOP Risk Matrix

5 Would a LOPA have Prevented it?

What follows is a simplified LOPA conducted via a Required Risk Reduction matrix (Duhon, Cronin, 2015). The matrix, Figure 3, has the following properties:

1. Both the horizontal axis (frequency) and vertical axis (severity) have one order of magnitude difference from one row/column to the next. Note that I have extended the vertical axis to show 10, 100 and 1000 fatalities rather than the single 'Multiple Fatalities' severity used in the HAZOP matrix above.
2. Rather than red, yellow, green, the cells have numbers. Each number represent the required order of magnitude risk reduction in the given design.

This form of the risk matrix allows explicit identification of the unmitigated risk and explicit accounting for the impact of safeguards.

8 1000 Public Fatalities	4	5	6	7	8
7 1000 Fatalities/100 Public Fatalities	3	4	5	6	7
6 100 Fatalities/10 Public Fatalities	2	3	4	5	6
5 - CATASTROPHIC 10 Fatalities	1	2	3	4	5
4 - MAJOR single fatality	0	1	2	3	4
3 - SEVERE severe lost time injury		0	1	2	3
2 - MINOR reportable injury			0	1	2
1 - SLIGHT first aid				0	1
	1/10000 Years	1/1000 Years	1/100 Years	1/10 Years	1/1 Year
	A	B	C	D	E

Figure 3: Required Risk Reduction Matrix

Performing a LOPA via the RRR

Let's do a LOPA on the two scenarios we considered for the HAZOP,

Scenario 1: Large amount of water accidently introduced into the tank via human error.

Likelihood: 1/100 years, 1/1000 years???

Severity: 100 fatalities, 1000 fatalities???

Conclusion: Somewhere in the B6 to C8 range with an RRR of 3 to 5.
Let's assume that the LOPA team settled on an RRR = 4

Applying IPLs:

PSV/Flare:	2
PCV/Scrubber:	0 (the scrubber was sized for minor releases)
Pressure Alarm:	0 -1 (Probably 0. Would have depended on identification of available responses such as addition of diluent.)
Temperature Alarm:	0
Refrigeration	0 or 1 (Probably 0. Refrigeration would have slowed the reaction initially, but eventually the run-away reaction would have overwhelmed the refrigeration system.)
Buffer zone around plant:	1?? (Would LOPA team have considered the buffer zone?)
No vent or drain connections	1??
Diluent	0 (See pressure alarm - LOPA team would have sought information of the nature of the diluent, its effect on the reaction rate, the ability of operators to implement it, etc.)
Sum of IPLs:	3 to 4

Scenario 2: Large amount of water accidently introduced into the tank via sabotage.

This discussion would have proceeded much as in scenario 1 except that one safeguard would not apply to this scenario – the lack of drain and vent connections which made human error less likely, but did not make sabotage any harder or less likely.

Would a LOPA have prevented Bhopal? The most obvious impact of a LOPA would have been to discount the scrubber as a safeguard/IPL for a major release. It was sized for minor venting. It seems likely that a LOPA would not have found adequate safeguards to balance the threat. Ergo, a LOPA would likely have driven process modifications of some sort.

6 Would an STPA have Prevented it?

STPA is a systems-theory-based PHA methodology developed by Nancy Leveson (Leveson, 2011). It is currently mainly used in highly complex, highly hazardous industries such as military, space, aeronautics, medicine and autonomous vehicles.

The guiding principle of STPA is that accidents happen when we lose control. The methodology, in-a-nutshell is:

1. Define the system and scope of study
2. Develop the control structures
3. For each control action identified in the control structure, identify how we could lose control via:
    a. Safe action not applied
    b. Unsafe action applied
    c. Action applied too early, too late or out of order
    d. Action applied too long or not long enough
4. For each unsafe control action identified, identify causal factors
5. Make recommendations to prevent the unsafe control actions

STPA uses control structures to identify all control actions (process related, economic, regulatory, social or otherwise), that directly or indirectly affect the control objective under study. Such control structures help us think about control loops other than process control loops, which is perhaps the most important impact of STPA.

Given the hazardous nature of MIC, the obvious control objectives that would have been analysed by the STPA team were controlling water ingress and human exposure in case of a leak.

Figures 4 and 5 show the control structure (at two levels of detail) to represent these objectives.

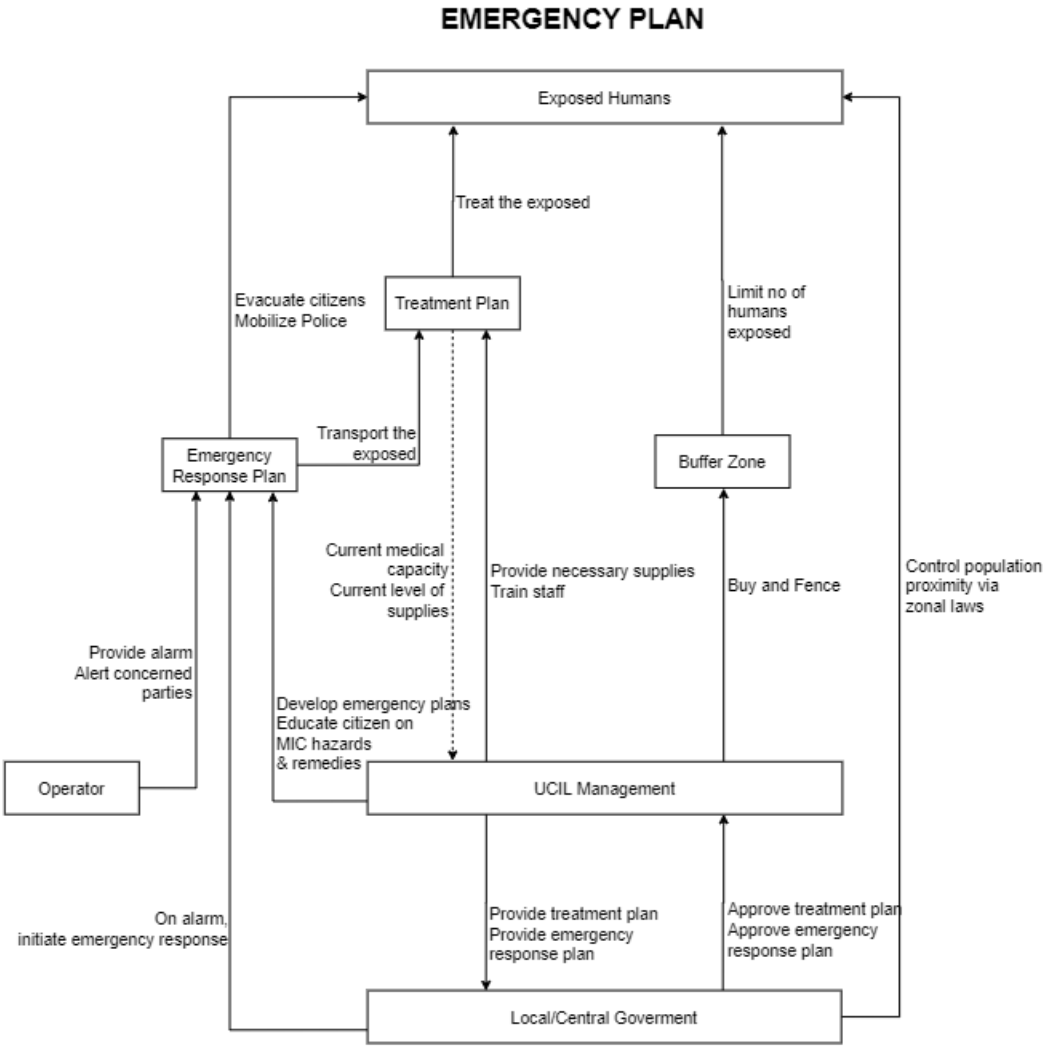

Figure 4 – Bhopal MIC Release Control Structure (High Level) Plan

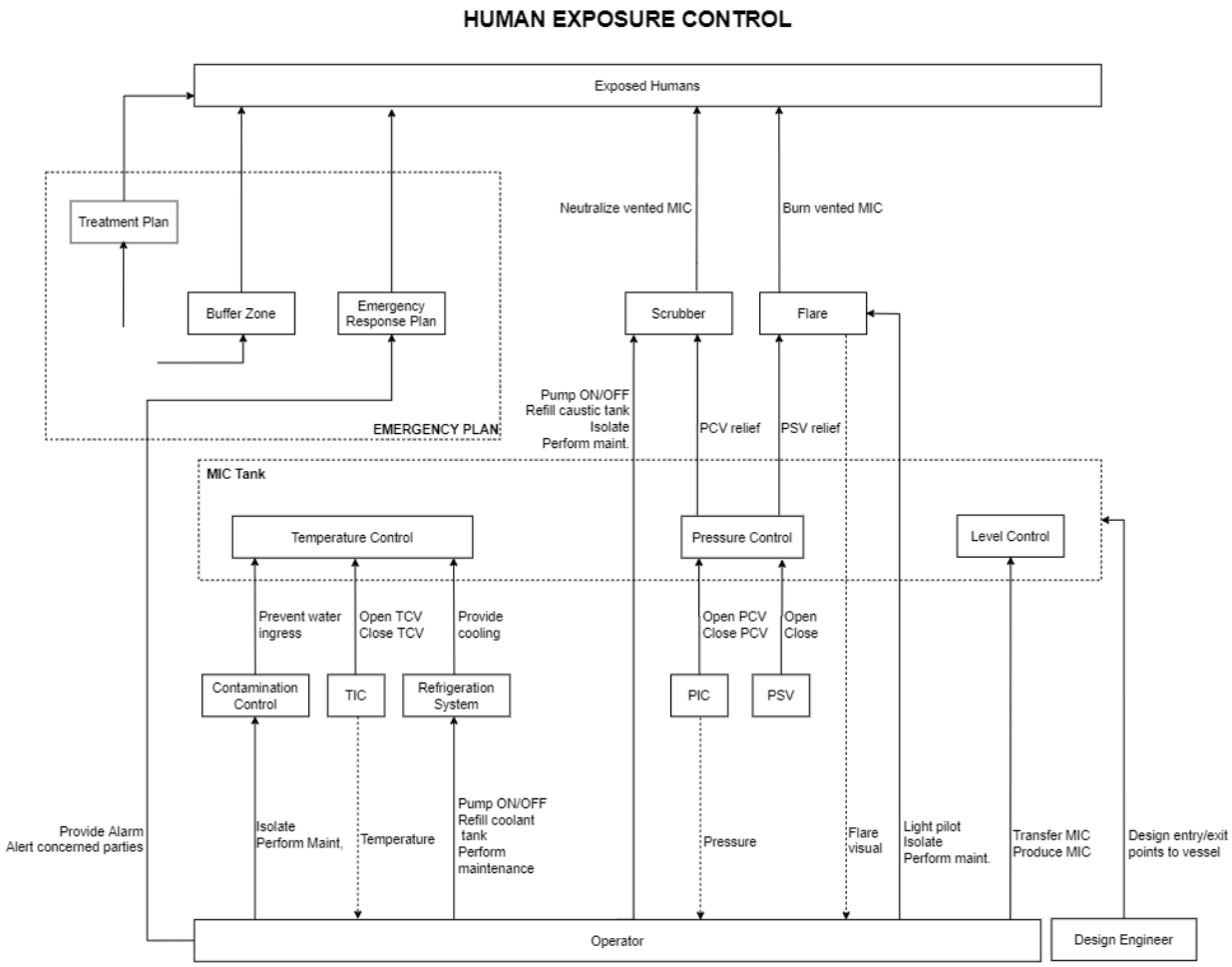

Figure 5 – More Detailed MIC Release Control Structure for Emergency

In an STPA, all control actions identified in control structures are rigorously analysed to identify any potential unsafe control actions (UCAs) and make recommendations to prevent them. Tables 1-6 show how UCAs are identified. The explicit emphasis on control would likely have led the STPA team to evaluate several control functions more extensively than HAZOP or LOPA teams would have using UCA tables as follows:

Table 1: Identifying Unsafe Control Actions (1/5)

Objective: Control Water Ingress
Controller: Operator
Action: Water Wash Filter

Not providing causes Hazard	Providing Causes Hazard	Providing Too early/ Too late	Stopped too soon/Applied too long
Washing of filters causes hazard if the MIC Storage Tank is not isolated	Washing of filters causes hazard if the water system pressure is high enough for water to reach the MIC storage tank	-	-
-	Washing of filters causes hazard if the MIC Storage Tank isolation is not leak proof	-	-

Table 2: Identifying Unsafe Control Actions (2/5)

Objective: Control Water Ingress
Controller: Design Engineer
Action: No drain or vent valves installed to prevent accidental water hose attachment

Not providing causes Hazard	Providing Causes Hazard	Providing Too early/ Too late	Stopped too soon/Applied too long
-	Providing an entry point for MIC causes hazard if the MIC inlet pipeline is susceptible to accidental injection	-	-
-	Providing an access point for PSV causes hazard if it is easily removable and accessible for other injection	-	-

Context: Unlike traditional methods of hazard identification, that lead to human error as the cause of failure, STPA considers human error as a symptom of failure. It requires the team to identify system states that make such 'errors' possible. Ergo, an STPA is useful in realizing design flaws that are vulnerable to human error.

Water entered the tank either through a hose attached following removal of a pressure gauge (sabotage) or via improper isolation for a maintenance operation. In either case, the STPA team would likely have questioned the level of control provided for keeping water out of the tank. It shouldn't have been as easy as removing the pressure gauge! It should not have been as 'easy' as forgetting to install slip blinds!

Table 3: Identifying Unsafe Control Actions (3/5)

Objective: Control Human Exposure
Controller: Government Agencies and UCC/UCIL
Action: Provide and Maintain Buffer Zone Around the Plant (Note: UCC/UCIL bought a great deal of land around the plant, but only fenced a few acres of the plant proper.)

Not providing causes Hazard	**Providing Causes Hazard**	**Providing Too early/ Too late**	**Stopped too soon/Applied too long**
Not providing a buffer zone (failure to purchase the land) causes hazards if there are settlements in the unsafe area		Buffer zone is not provided if people are already living on the land and cannot be evicted	Buffer zone may be effective early in plant life, but stopping the policing of property rights in the buffer area by the government later in plant life will lead to hazards if illegal settlements are established.
Not controlling settlement into the buffer zone			

Context: By explicitly identifying the Buffer Zone as a 'control' that can fail, rather than a safeguard to be taken credit for, an STPA would have likely prompted the management to enclose the entire Buffer Zone within the plant fence rather than depending on local and state governments to police property rights.

Table 4: Identifying Unsafe Control Actions (4/5)

Objective: Control Human Exposure

Controller: Government Agencies and UCC/UCIL

Action: Developing and providing an emergency response plan

Not providing causes Hazard	Providing Causes Hazard	Providing Too early/ Too late	Stopped too soon/Applied too long
Not developing a response plan causes hazards if the Medical professionals are not trained to treat MIC exposure			Hazard develops if training to medical professionals is provided early, but not updated when required
Not developing a response plan causes hazards if the Medical professionals do not have the supplies to treat MIC exposure.			Hazard develops if supplies are provided early, but not maintained
Not developing a response plan causes hazards if the First responders are not trained for MIC release			Hazard develops if first responder training is provided early, but not updated when required
Not developing a response plan causes hazards if the People are not trained on how to respond to a release			Hazard develops if people are trained early, but training is not maintained

Table 5: Identifying Unsafe Control Actions (5/5)
Objective: Control Human Exposure
Controller: Operators
Action: Sound Alarm

Not providing causes Hazard	Providing Causes Hazard	Providing Too early/ Too late	Stopped too soon/Applied too long
Not sounding an alarm causes hazard as People will not be aware of the leak	Sounding an alarm causes hazard if Alarm is sounded too often and is hence ignored.	Sounding an alarm causes hazard if Alarm is sounded too late for effective response	Sounding an alarm causes hazard if Alarm is stopped too soon.
-	Providing an alarm causes hazard if the people are not aware of how the leak related alarm sounds	-	-

Context: There should have been a unique alarm in place that the common people recognized. There should have been a plan in place to mitigate the effects, even if a leak occurred. Citizens were reported to be running on the streets, unaware of the consequences, helpless and uninitiated. The simplest remedies like a wet towel on the face could have mitigated the effects. Why were citizens not aware? Why were the most basic remedies for MIC exposure unknown? Why couldn't the police and the military mobilize the affected on time? An STPA would have not only identified the need to develop an Emergency Response Plan, but also analysed the developed plan to realize any potential gaps in emergency readiness.

Table 6: Identifying Unsafe Control Actions (6/6)
Objective: Control Human Exposure
Controller: Operators
Action: Provide incident reports

Not providing causes Hazard	Providing Causes Hazard	Providing Too early/ Too late	Stopped too soon/Applied too long
Not providing timely incident reports to the management causes hazard	Providing incorrect incident reports causes hazard	-	-

-	Providing incomplete incident reports causes hazard	-	-

Context: Perhaps the most distinctive feature of an STPA is its ability to address socio-political aspects of a system. To this day, even after thorough investigation and multiple interviews with the operators, we still do not have a clear explanation of what exactly went down in Bhopal. Operators offered opposing explanations, some were hesitant to speak due to fear and some others were perhaps just disgruntled by the state of affairs in the company. There was a clear lack of communication and trust between the UCIL management and the plant operators. Did they provide the management with previous incident reports? (minor leaks had occurred before the major accident). Did they sound the alarm on time? Were they hesitant to tell the truth? All of this boils down to the communication established between the operators and management. If the communication gaps would have been addressed earlier, perhaps the operators might have spoken up about their disgruntle instead of acting out or covering up. By recognizing the communication between operators and management as a control action, an STPA would have, at the very least, initiated a discussion about such communication gaps that were prevalent.

7 Conclusion

The process industries began routinely doing HAZOPs following, and at least partially as a result of, Bhopal and other significant accidents in that time period. I (Duhon) began to wonder if a HAZOP performed in the 1960s on the Bhopal MIC Plant design would have prevented the accident.

I found this question difficult to answer.

When we started doing LOPAs I asked the question anew. LOPA methodology corrects a common blind spot. A HAZOP team will 'take credit' for almost anything that looks like a safeguard. LOPA technology identifies and quantifies the effectiveness of each possible safeguard. A LOPA would almost certainly not have taken credit for the MIC caustic scrubber for example.

And yet, it still wasn't clear to me that a LOPA team in the 1960s would have made the necessary changes to prevent the accident or adequately mitigate the results.

And now that we are at least contemplating the use of STPAs I've asked myself the question once more. Would an STPA, performed in the 1960s, have prevented Bhopal? This time it is easier to have some confidence. The basic premise of the STPA methodology is that accidents occur when we lose control. This starting point leads us to generate a much larger and richer set of things that might go wrong.

It is much more likely that an STPA would have prevented Bhopal, and by extension, that it will prevent the next 'unimaginable' tragedy.

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PSM Plus: Predictive Process Safety Analytics and IIoT

Michael Marshall, PE*
Tratus Group
Findlay, Ohio
michael@tratusgroup.com

Abstract

EHS and APM software platforms fail to effectively couple asset integrity with a process safety complement in the form of risk reduction metrics, KPI benchmarking, scorecards, dashboards, reports, alerts, and other data displays and notifications.

"Data Rich & Insights Poor" is a characteristic observation in organizations not fully deploying digital analytics and transformation tools. The ineffectiveness of incident reduction processes, tools and software applications in use today can generally be characterized as follows:

- A lot of data is being generated, but is underutilized for data-driven analytics and systemic root cause solutioning to enable whole classes of defects to be resolved
- Lots of emphasis on compliance, but too little focus on process safety risk reduction
- Other than API 754 PSE Tier 1 and 2 KPI comparisons, there is little evidence of competitive KPI/indices benchmarking of the much more numerous near miss and unsafe conditions data of Tier 3 and 4 PSEs
- Programs lack business perspective regarding the impact of asset integrity on process safety and incident reduction as a function of mechanical availability and associated lost production costs (a huge driver considering that every 1% gain in mechanical availability is worth about $8 million of additional margin capture per year in a typical 200,000 bpd refinery)
- Not utilizing a predictive approach involving algorithmic correlation relative to causation
- Incapacity to link condition monitoring and failure analysis for predictive analytics, advanced pattern recognition, machine learning and artificial intelligence

So, with an IIoT predictive application environment as the backdrop and an asset integrity and process safety analytic framework as the primary enabler, this paper discusses methods, metrics,

performance analyses, and KPI benchmarking techniques for driving Operational Excellence as it relates to the ultimate concern of any PSM program, i.e., the loss of primary containment (LOPC) and associated impacts to production, profitability and process safety.

Keywords: asset integrity, process safety, software, incident management, metrics, KPI benchmarking

If I were designing process safety software with asset integrity in mind, (and vice versa) it would look something like this…

Introduction

Two high-level indicators used to evaluate manufacturing cost effectiveness are mechanical availability and maintenance costs as a percent of replacement asset value (RAV). It is widely accepted by Oil & Gas companies that world class manufacturing performance means operating at or above ***97% mechanical availability*** as well as spending less than 2% on maintenance as a percent of replacement asset value (RAV).

In order to achieve such "best in class" targets, tools must be used to ***analyze and trend performance relative to those measures***. Deep-dive methods must surface indicators which drive toward systemic root causes of inadequate performance and reveal both asset integrity and process safety "AIPSM" incidents ***as a function of economic impact*** (lost production plus direct losses). As such, lost profit opportunity ($LPO) becomes a measure of loss of primary containment (LOPC) incidents and near misses characterized by equipment anomalies and upset/malfunction operating conditions.

By way of example, five years after deploying an earlier scaled-down spreadsheet version of just such an incident-focused asset integrity and process safety analytic framework at a multi-refinery company, the following results were realized across the enterprise:

- **Safety**: 27% reduction in Process Safety Management (PSM) and environmental incidents
  - 24% reduction in fires and explosions
  - 39% reduction in spills and releases
  - 48% reduction in near misses

- **Mechanical Availability**: Improved equipment reliability and asset utilization
  - Increased mechanical availability by 1.5% overall (up to 5% at some sites)
  - Reduced production losses by 47%
  - Reduced unplanned outages by 25%

  } Savings of $20 million per year (four refineries)

- **Costs**: Reduced maintenance costs by 18%
  - Achieved maintenance costs of 2% RAV at most sites (< 3% at others)

In spite of the initial reliance on spreadsheet facilitated data acquisition and aggregation, the rigor and structure of this unique asset integrity approach was successful in bringing into focus many common cause systemic failures, bad actors and associated process safety issues highlighted by the economic impact to the organization.

Additionally, a calibrated/weighted asset risk ranking tool and methodology facilitated the proper allocation of tools and resources for identifying performance optimization opportunities and driving Operational Excellence (OE) initiatives. Emphasizing the value of this asset integrity approach drove the proper prioritization of opportunities and virtually guaranteed the successful outcome of the exercise.

If 97% mechanical availability is now considered world-class asset integrity, could sustainable 98% or 99% availability be achievable by coupling the incident investigation and reporting analytic framework being introduced, which let's call *PSM Plus* for purposes of this paper, with condition monitoring Industrial IoT (IIoT) technologies like predictive analytics, Advanced Pattern Recognition (APR) and machine learning? Considering that ***every 1% gain in mechanical availability is now worth about $8 million of additional margin capture per year in a typical 200,000 bpd refinery,*** the low-cost, high impact potential of a systemic RCFA approach like *PSM Plus* is a logical next step for IIoT predictive analytics.

So, with IIoT predictive analytics as the backdrop and the asset integrity and process safety analytic framework of *PSM Plus* as the primary enabler, this paper discusses methods, metrics, performance analyses, key performance indicators (KPIs) and benchmarking techniques for driving OE as it relates to the ultimate concern of any PSM program, i.e., the loss of primary containment (LOPC) and associated impacts to production, profitability and process safety.

This paper reveals how *PSM Plus* ties together seemingly connected but functionally disparate asset integrity and process safety fundamentals into a collaborative analytic framework involving:

- Knowledge Management Systems Platforms
  - Asset Integrity Management (AIM) Systems (like Metegrity Visions™)
  - EHS Management Systems (like Enablon, Gensuite, Visium KMS)
- Problem Solving Methodologies
  - Continuous improvement programs (A3 problem-solving, 8Ds, TapRoot®, DMAIC, Six Sigma, etc.)
- Predictive Maintenance with APR and trending technologies
  - DCS systems, operational data and data historians
- Benchmarking Tools
  - Gap analysis and regulatory reporting
  - "Solomon Associates style" industry-wide process safety benchmarking

PSM Plus is at the core joining these systems and tools into a cohesive and synergistic set of technologies for managing asset integrity and process safety. It includes several business methods for evaluating ***people, processes and tools (and technology)*** and focuses on the three high value OE business drivers of ***risk management, cost reduction, and productivity improvement***.

"If you don't measure it, you can't manage it!" [1, 2]

The PSM rulemaking was exceptional in its vision some thirty years ago but could have been made much better by the inclusion of metrics and KPI benchmarking. As is often said "what gets measured gets done" is likely a leading reason why so many asset integrity and process safety programs have failed to grow and continuously improve relative to industry best practices and OSHA expectations.

Even though API 754 (published 2010 and significantly revised in 2016) (**Figure 1**) makes the case for establishing common metrics across industry with its public reporting of higher consequence process safety events (PSEs), it is of limited usefulness as a meaningful benchmarking tool without the public sharing and comparison of the much more numerous near miss and unsafe conditions data of the Tier 3 and 4 PSEs.

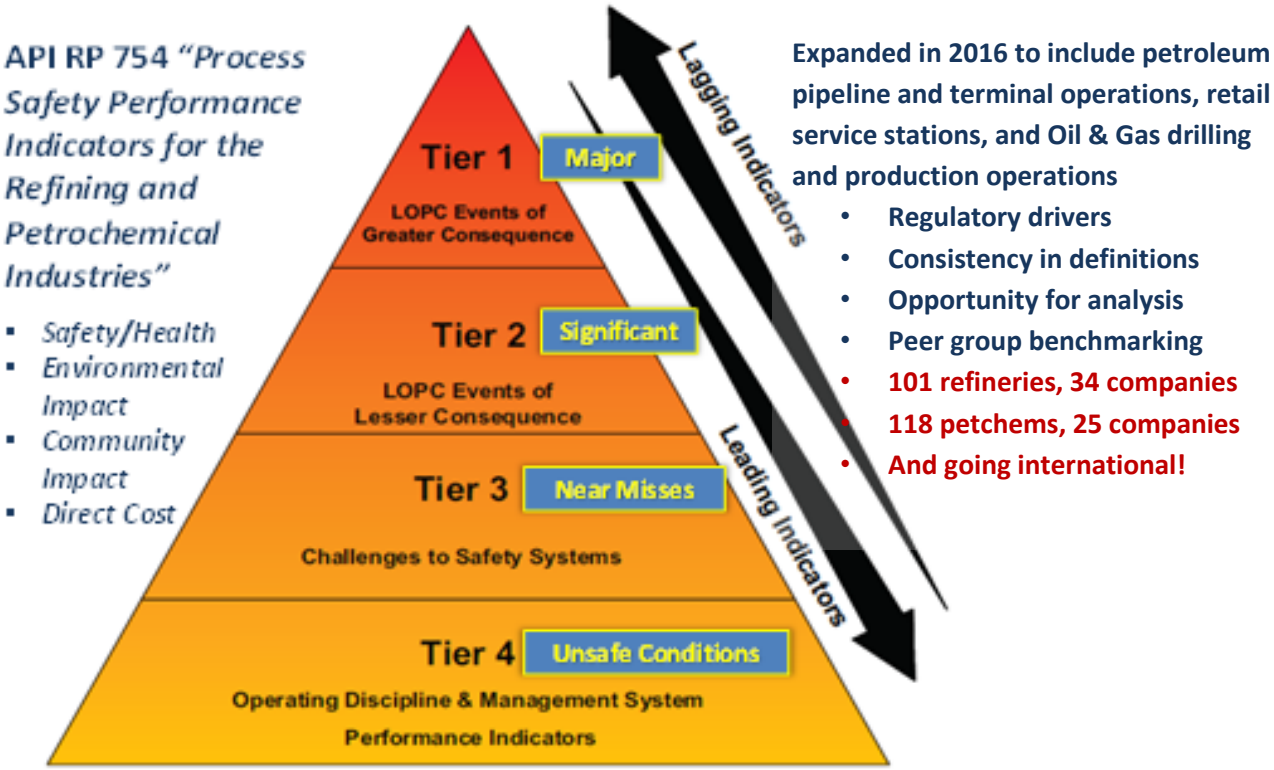


Figure 1. API RP 754 "Process Safety Performance Indicators of the Refining and Petrochemical Industries"

Although a good benchmarking program overall, the US Chemical Safety Board (CSB) characterizes the shortcomings of API 754 (Gomez, 2012) [3] as follows:

1. The statistical power of the few higher severity Tier 1 and 2 events is insufficient to detect effect
2. The Tier 1 and 2 numbers are lagging indicators and thus of limited usefulness as performance indicators

3. The lower consequence near miss and management system failure Tier 3 and 4 events occur in larger number and are thereby more reflective of process failures, and yet are not publicly reported for industry trend analysis, KPI benchmarking, and continuous improvement

Undoubtedly, analyzing for systemic root cause and publicly benchmarking the much more numerous "free lessons" of Tier 3 and Tier 4 PSE findings according to company size, type, peer group and other deeper dive comparators, etc., specific performance improvement opportunities could be better assessed and thereby further utilized to enhance best practice and regulatory conformance.

Of the fourteen PSM elements, incident investigation is the one which provides the best window on asset integrity, plant reliability and process safety risk management, and which gets the most attention from regulators, especially the CSB. Incident analyses almost always show that loss of primary containment (LOPC) is preventable, with mechanical failure far exceeding the next highest categories of operator error, other/unknown and upset/malfunction which all together constitute the leading process safety risk opportunities for improved performance in the process industry today. Utilizing a data-driven predictive analytics/decision support framework like *PSM Plus* for systemic root cause analysis drives incident and risk reduction by enabling whole classes of defects to be resolved across an enterprise and throughout facilities company-wide.

Incident Mitigation and Management - Where is your organization?

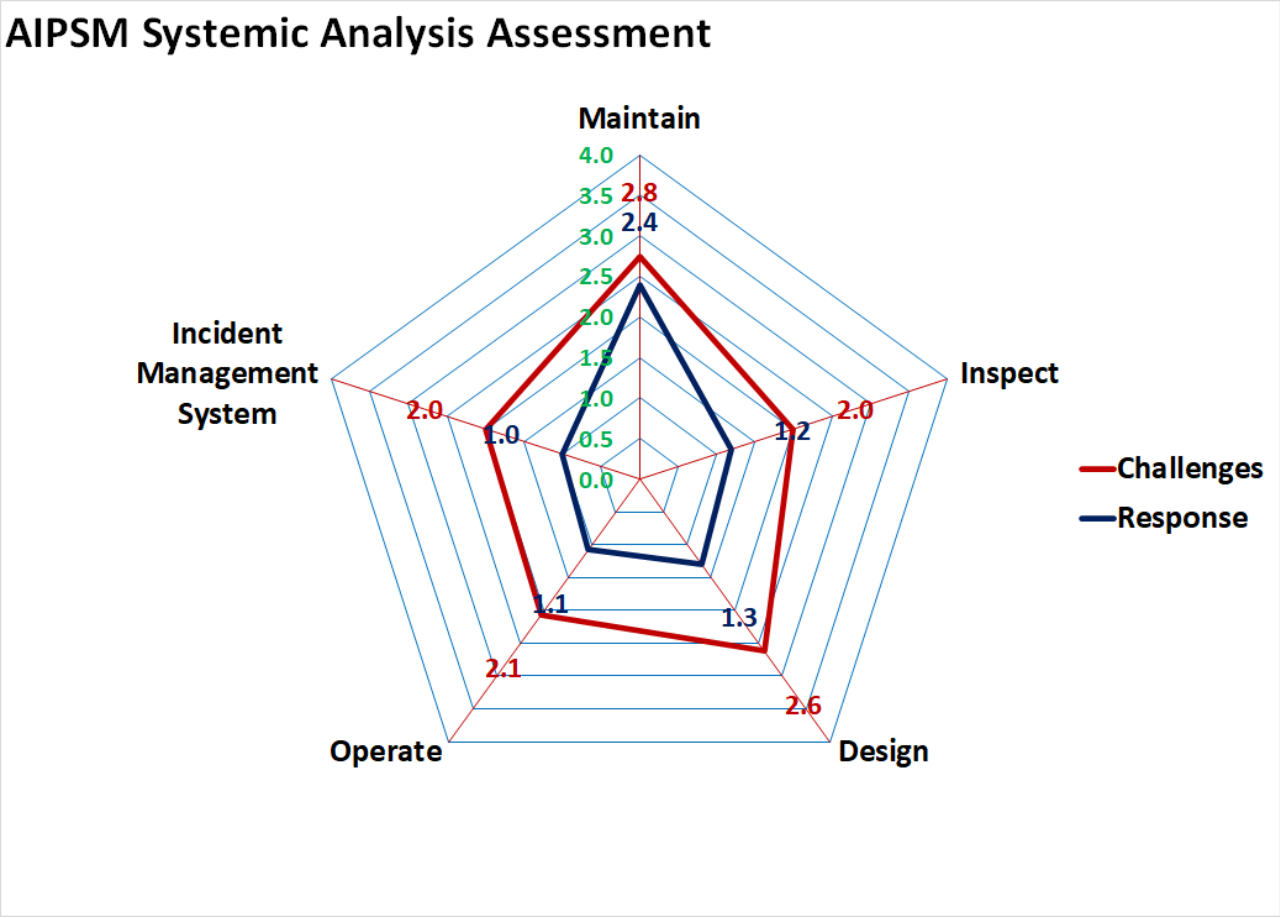
Major process safety events often arise from the unforeseen interactions of human and organizational factors, leading to what is called an "organizational incident." The "people, processes and tools" management system framework of the PSM rule focuses heavily on work processes and program reviews, yet fails to adequately address the organizational "people" aspects of leadership and management commitment, management reviews, authorities and accountability, human factors, organizational culture as well as continuous improvement by way of metrics/KPIs for assessing program effectiveness and maturity.

With respect to minimizing organizational incidents, any number of the following factors may contribute to the ineffectiveness of incident management system (IMS) processes and tools (software and other technologies) in use today, especially with Tier 3 and 4 PSE analysis:

- Inadequate root cause failure analysis (RCFA) quality - Lack of critical thinking in process safety event (PSE) causal analysis
  - RCFA methods like the "Five-Whys" are utilized superficially and with minimal causal factors analysis
  - Search for a single root cause promotes a flawed reductionist view of incident causation given that multiple root causal factors most often contribute to a PSE
  - Human factors and equipment failure are not analyzed for systemic root cause(s) and systems-focused solutions, thereby not enabling whole classes of defects to be resolved
  - Associations in the data are not sought out for high level systemic analysis and planning
  - EHS platforms lack a process safety complement and are not configured to incorporate and display metrics, KPIs, scorecards, dashboards, reports, portals, alerts, and analyses
  - Programs lack business perspective regarding the impact of asset integrity on process safety and incident reduction
- Management systems are not properly designed for optimization planning and process improvement
  - Lots of emphasis on compliance, but too little focus on process safety risk reduction
  - A more robust data collection, analysis and reporting Plan-Do-Check-Act (PDCA) management system structure is necessary to satisfy regulatory expectations for the demonstration of program effectiveness and maturity as well as for leadership and management commitment
  - Other than API 754 PSE Tier 1 and 2 KPI comparisons, there is little evidence of competitive KPI/indices benchmarking of the much more numerous near miss and unsafe conditions data of Tier 3 and 4 PSEs across the enterprise and company-wide
  - Limited evidence of employee (frontline) involvement and "closest to the work" mentality in incident reduction policies, practices and programs
  - Insufficient evaluation of asset integrity and process safety performance relative to economic impact (lost production plus direct costs)
  - Lack of mechanisms for management system consistency and sustainability
- A lot of data is being generated, but is underutilized for data-driven analytics and solutioning - "Data Rich & Insights Poor" is a characteristic observation in organizations not fully or properly deploying digital analytics and transformation tools
  - Not utilizing a predictive approach involving algorithmic correlation relative to causation

- Not leveraging data infrastructure and information systems like OSISoft PI and CMMS platforms
- Limitations exist due to inflexible IT architectures which are not conducive to robust analytics, especially predictive analytics and IIoT digital transformation processes and technologies
- Limited use of mobile digital graphical interfaces to facilitate end user experience by way of dashboards, scorecards and reports with data views, charts, maps, tables, KPI's and alerts built on a predictive analytics platform
- Inability to link condition monitoring and failure analysis for predictive analytics

The *PSM Plus* AI+PSM (and IIoT-enabled) coupling of process safety (incident management) with asset integrity (= maintain + inspect + design + operate) utilizes a management system framework (people, processes, tools/technology) to drive risk mitigation and management in conformance with industry RAGAGEP (recognized and generally accepted good engineering practice). A strategically designed 20-element assessment protocol based on ***AIChE CCPS book "Risk Based Process Safety"*** as well as other industry best practices is used to analyze and identify implementation opportunities for performance improvement, process optimization and ***systemic risk reduction***.



Investing more organizational effort into the design and implementation of incident management system (IMS) programs is key to assuring that the right information is gathered from failure events and is applied to ***managing risk systemically***. In so doing, "lessons learned" becomes more than a clichéd phrase but instead an integral part of the organizational incident reduction process. *PSM Plus* was developed to address this overall industry-wide need.

A Closer Look at *PSM Plus* and its AIPSM coupling

PSM Plus (**Figure 2**) and its AIPSM approach can be deployed as a web-based analytic framework (methodology) for incident investigation and reporting, with the following capabilities at its core:

- Looks at process safety through the lens of asset integrity and lost production impacts
- Establishes a common framework for assessing asset integrity and process safety management effectiveness and maturity for KPI benchmarking between plants
- Categorically risk ranks issues and guides problem solving teams to high value deep-dive investigation of systemic problems, prioritized by safety as well as economic impact
- Provides an analytic framework/filtering tool for root cause analysis teams now inundated with API 754 PSE investigations
- Makes the system point to root causes with data sufficient enough to facilitate systemic analysis, thereby eliminating the need to review large numbers of individual detailed reports
- Gives specific attention to equipment failure analysis and human factors fundamentals
- Utilizes FMEA algorithms to assist the user in root cause determination and solutioning
- Limits the number of distinctive event cause categories with a coding structure that connects industry-specific causal factors with a concise set of basic and root cause categories
- Prioritizes by API 754 PSE Tier definitions and measures by economic impact to the corporation, providing a clear understanding of lost profit opportunities associated with asset integrity and process safety incidents
- Utilizes a data-driven predictive analytics/decision support framework for systemic root cause analysis in order to enable whole classes of defects to be resolved across the enterprise and throughout facilities company-wide
- Provides early warning of precursor/incipient failure stages of equipment degradation and associated impacts on mechanical availability, process safety and profitability
- Provides for an added capability of linking condition monitoring and failure analysis to advanced pattern recognition for predictive analytics

PSM Plus AIPSM (i.e., the spreadsheet version) is a field-tested and proven system for the characterization, classification and categorization of asset integrity and process safety incidents risk-ranked and prioritized by API 754 PSE potential as well as economic impact (production plus direct losses). Sophisticated machine learning techniques scour historical incidents to find meaningful patterns in the *PSM Plus* data to prioritize and guide investigative teams to high value problem-solving exercises.

Most importantly, *PSM Plus* AIPSM prioritizes systemic operational problems and guides engineers and process safety specialists to focus on high value investigation exercises as measured by economic impact to the organization. It captures and structures ***the 20% of data that 80% of operators, engineers, managers and corporate executives want to see*** by tapping into the data rich potential of an enterprise asset management (EAM) system and surfacing that 20% of key information as KPIs.

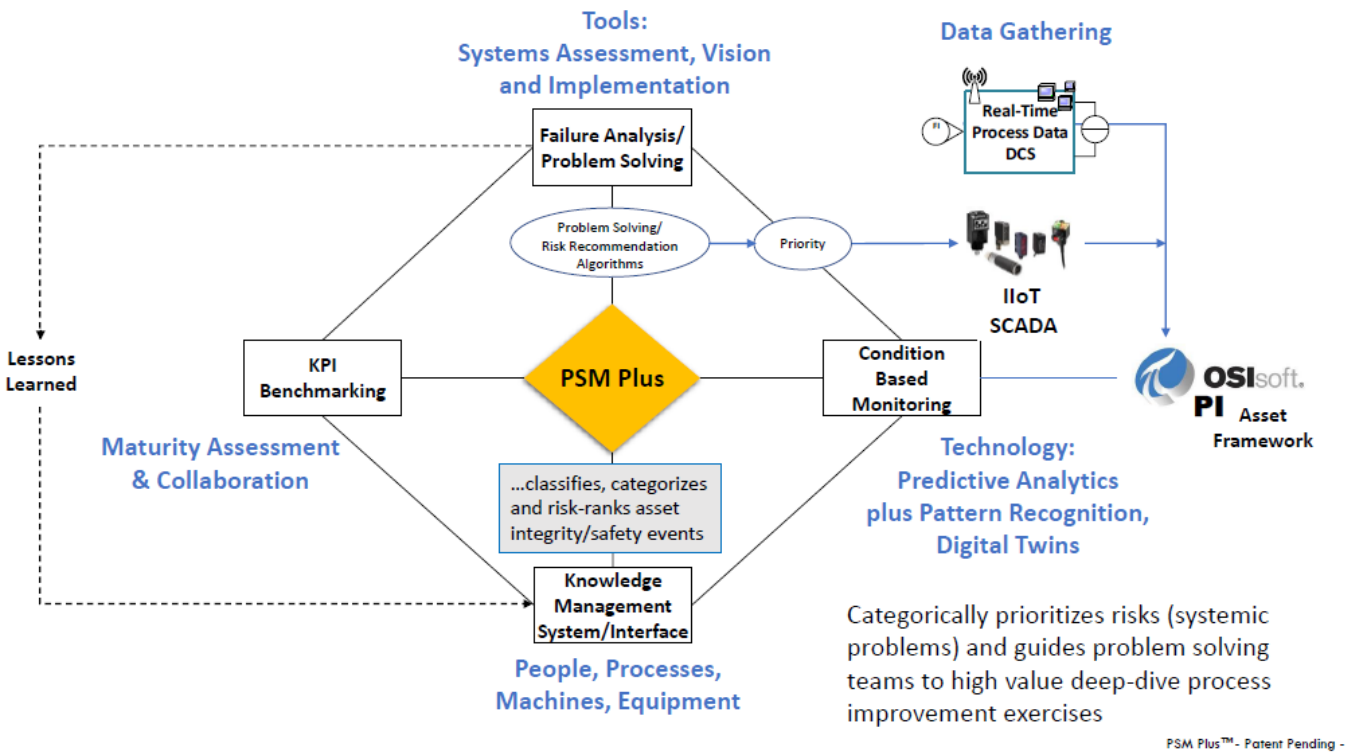


Figure 2. Business Model for *PSM Plus* and Predictive Maintenance with APR Analytics Design, Implementation and Partner Collaborations

Sophisticated machine learning technologies can be used to find meaningful patterns from the *PSM Plus* data to prioritize issues and guide problem solving teams to high value investigation exercises. Solutioning algorithms and methodologies such as A3, Eight Disciplines (8Ds), TapRoot® or ABS RCA Handbook problem solving are used to conduct extensive investigations, invoking data from the APR tools to conduct RCFA and establish patterns and trends to apply proactive measures to predict and prevent future events from occurring.

This incident investigation and reporting analytic framework has the added capability of linking condition monitoring and incident investigation to ***Advanced Pattern Recognition*** (APR) tools by automatically establishing event frames that identify asset integrity and ***process safety events and lost profit opportunity*** ($LPO). The APR tools use this information to mark the time series data in a way that connects the anomaly or incident to the operational parameters leading up to the event.

Applying a rigorous structure and discipline to the investigation process, *PSM Plus* goes well beyond analysis of the high severity Tier 1 and 2 PSEs to draw even more insight from the much

more numerous Tier 3 and 4 events. Every year in a typical refinery, Tier 3 and 4 numbers far exceed Tier 1 and 2 events by many hundreds, and yet are often not adequately analyzed (or reviewed at all) for root cause due to a lack of proper tools and resources. Obviously, the higher incidence of these lower severity near miss and unsafe conditions is statistically more relevant than the relatively few higher severity PSEs and are thereby more reflective of common cause failures.

The characterization, categorization, risk-ranking and prioritization of that abundance (better seen as near-miss "free lessons") of data (**Figure 3**) is especially critical for identifying systemic problems and converting into leading indicators of more serious PSE potential. In order to drive continuous improvement with mechanical availability and process safety, ALL incident data must be analyzed for systemic effect in order to enable whole classes of defects to be resolved across an enterprise and maximize the knowledge base necessary to reduce the risk of LOPC occurrence as well as minimize lost profit opportunity ($LPO).

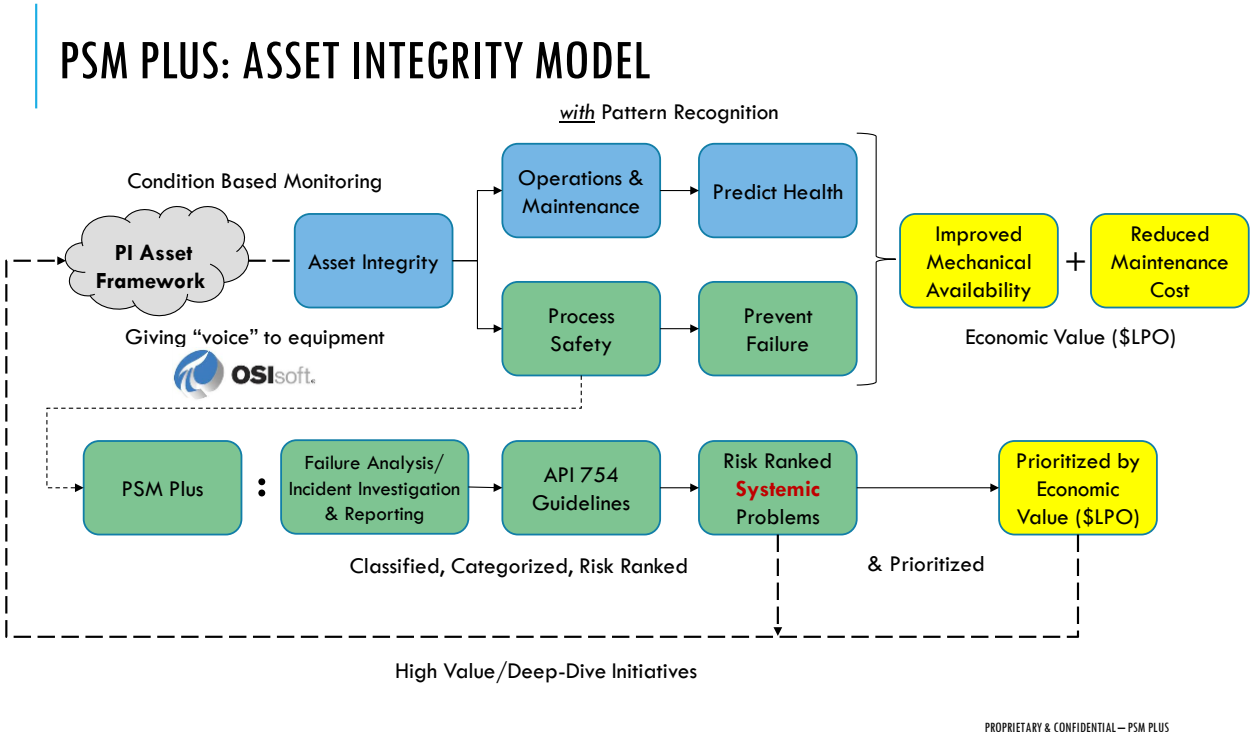


Figure 3. IIoT and the *PSM Plus* Asset Integrity Model

The *PSM Plus* analytic framework is the ideal complement for ***benchmarking process safety management program effectiveness and maturity*** as well as the establishment of a potential industry PSM accreditation model. Such a model might entail conformance with elements of the ***AIChE CCPS book "Risk Based Process Safety"*** and also include a more robust capture, analysis, and benchmarking of API 754 Tier 3 and Tier 4 PSEs relative to incident precursors, data patterns, IOW excursions as well as other leading indicators.

Furthermore, with the proliferation of low-cost sensors and robust wireless communications technologies, plant systems will soon be flooded with thousands more data points, alarms and alert notifications. As such, knowing what data to capture, manage, display and control is essential to proper metrics development and analysis. By not understanding the importance of data collection and quality, that endeavor will just be an exercise in "garbage in, garbage out."

Without quality data and problem-solving analytics, you will be on the wrong path with your IIoT application and its expected return on investment (ROI). The role of *PSM Plus* in the IIoT digital transformation journey is as an analytic framework (methodology) for incident investigation and reporting which classifies and categorizes asset integrity incidents, and then risk-ranks and prioritizes according to API 754 PSE potential as well as economic impact to the corporation. Knowing ***"why it happened, and how to prevent it from happening again"*** is central to what *PSM Plus* is, and what it does better, faster and smarter than any software analytics offering on the market today (**Figure 4**).

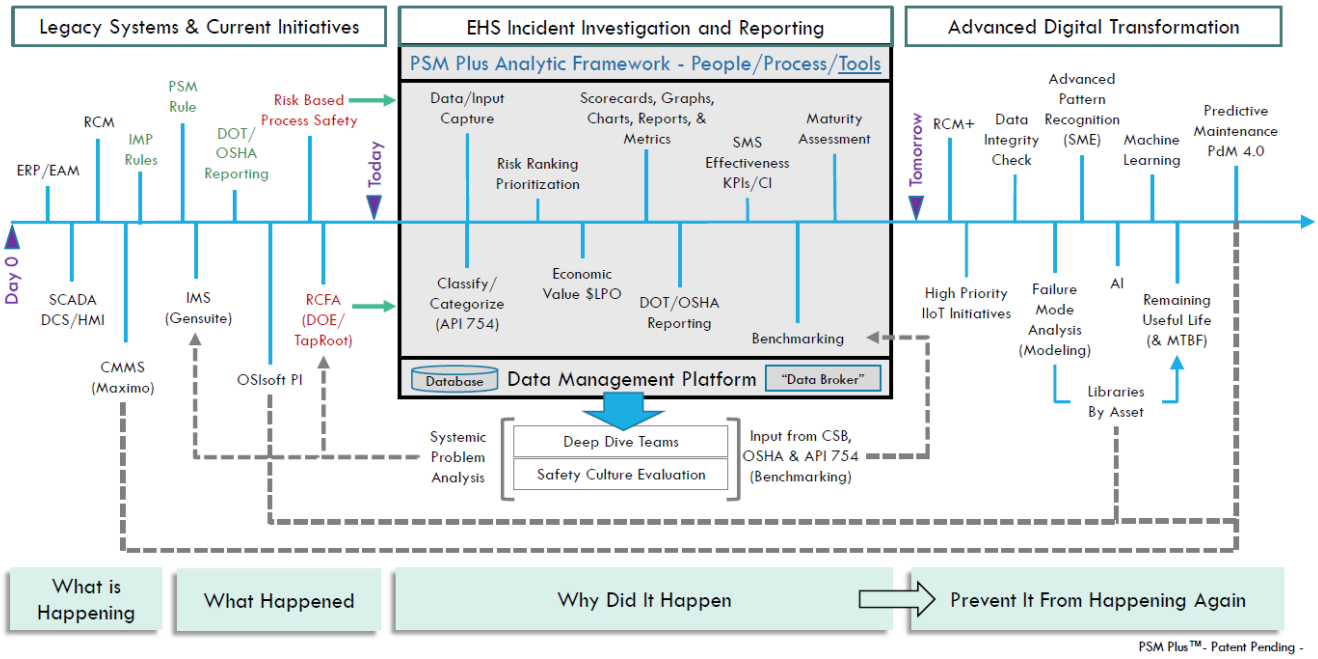


Figure 4. The role of *PSM Plus* in the digital transformation journey… ***"Why did it happen, and how to prevent it from happening again"***

For companies already using user-configurable (in theory) software packages and the API 754 incident management/classification system, incorporating the *PSM Plus* predictive analytics framework only involves reconfiguring fields and output to incorporate and display metrics, KPIs, scorecards, dashboards, reports, portals, alerts, and analyses.

The user-configurable software capability greatly facilitates this initiative, and results in a low cost, high impact opportunity for significantly enhancing risk/incident reduction efforts not only at the site level, but also corporate-wide. Such an initiative greatly complements ongoing safety culture improvement programs by further leveraging the often overlooked "people" soft risk/skills applications and improvements among stakeholders at all levels.

Achieving Operational Excellence – A Technological Evolution [4]

The convergence of information technology (IT) and operations technology (OT) data has been greatly facilitated by the proliferation of low-cost sensors and internet-protocol-enabled devices. Connecting people, processes, machines and equipment via the internet is the next wave of the industrial revolution which is being called the Industrial IoT (IIoT), or Industry 4.0. It is here, and it will be the early adopters who gain the competitive advantage in this new frontier of the Oil & Gas industry.

Chevron is one of those trailblazing pioneers, and has recently partnered with IoT services from Microsoft to enable thousands of pieces of refining and oil field equipment with wireless sensors by 2024 to predict exactly when equipment will need to be serviced. This is just part of Microsoft's $5 billion, four-year investment in the Oil & Gas sector to drive best-in-class improvements in mechanical availability, profitability and ultimately process safety and risk management. ***However, an analytic framework for <u>extending</u> Mean Time Between Failure (MTBF) and thereby driving continuous improvement in mechanical availability must also be part of the solution!***

Process plants are built around a myriad of machine and equipment assets like pumps, compressors, heat exchangers, piping, vessels, control valves and instrumentation, with the integrity of those assets being key to managing plant reliability and process safety risk. Given the demands of PSM program elements like mechanical integrity, hazard assessment, procedures, change management, incident investigation, and information management, it can be difficult for plant personnel to keep up. IIoT coupled with analytic tools holds the promise to help.

IIoT is converging IT with OT by connecting and enabling information exchange between sensors and analytics utilizing the data. Pre-built process safety analytic software or "apps" running on premise or in the cloud can quickly analyze data from performance and condition monitoring sensors and automatically find trends to generate alerts and reports, thus freeing up engineers from performing those same tasks manually.

As with Chevron and their Microsoft IIoT joint venture, many Oil & Gas companies are making significant investments in systems and processes to better manage and enhance the performance of their assets and ensure the safety of their work environments. These include investments in:

- Enterprise Asset Management (EAM) systems that capture a wide variety of data related to design, construction, commissioning, operations and maintenance of plant, equipment and facilities
- Digital Control Systems (DCS) that provide the automation necessary to control and manage production…generating enormous volumes of data that is stored in data historians
- Knowledge Management Systems that record incidents and act as repositories for capturing lessons learned, developing best practices, and reporting for regulatory compliance
- Universal Data Connector technologies like Eramosa eRIS™ acting as a "data broker" to unlock data from proprietary platforms, thus enabling the extraction of data from any database, software product or EAM system

- Problem solving and RCFA methodologies that invoke a disciplined approach to resolving issues and eliminating defects

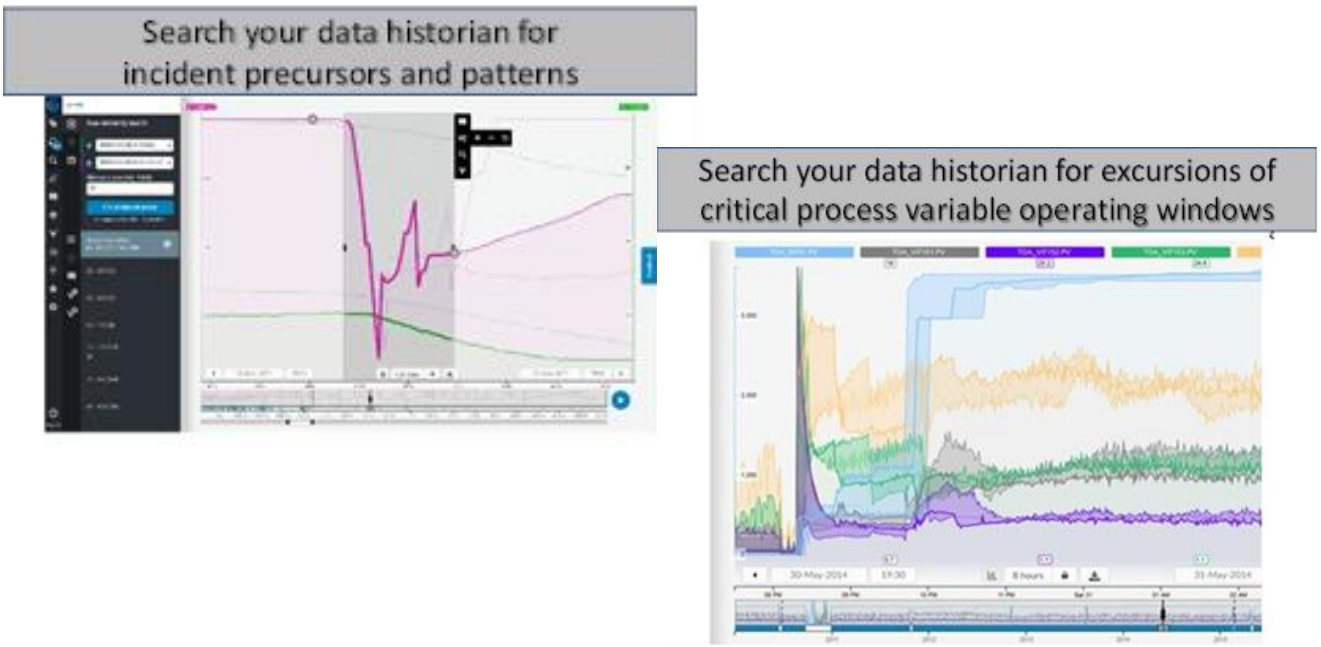


Figure 5. TrendMiner "Incident Precursors and Patterns"

Additionally, new investments are being made in emerging technologies where the convergence of low-cost sensors with robust wireless communications technologies make it economically feasible to outfit more equipment for condition-based monitoring. This proliferation of instrumentation feeds directly into the creation of more data which can be used in new predictive maintenance with Advanced Pattern Recognition (APR) systems. These systems, fueled by recent strides in machine learning and artificial intelligence, hold incredible promise for unlocking meaningful insights from historical data and recognizing trends within integrity operating envelopes (API 584 IOWs) that could predict undesirable outcomes from occurring (**Figure 5**).

Undoubtedly, the promise of these technologies will change the way process and design engineers interact with data. Having said that, it is equally important to establish confidence that the data being stored for analysis accurately reflects the characteristics of the original signal. For this reason, tools like Pattern Discovery Technologies Compression*Insight*™ product should be employed to monitor for data historian configuration issues and data point behavior anomalies.

Unfortunately, the majority of these systems or processes exist as islands. Incremental value can only be realized when connected in a meaningful way that addresses asset integrity and process safety risk relative to economic impacts. *PSM Plus* creates this value.

Summary

IIoT is still in its infancy and remains a mystery to most process plant managers who have either not heard of it or do not understand its potential. Nevertheless, IIoT is clearly here and happening now, and ***convincing people that this rapidly emerging technology is not just another pioneering effort but instead "what good looks like" will be the challenge.***

As such, and with an IIoT application environment in mind, *PSM Plus* is a predictive process safety analytic framework which drives conformance to the safety management systems guidance of OSHA 1910.119 (Process Safety Management), API 1173 (Pipeline Safety Management) and API 754 (Process Safety Performance Indicators) by classifying and categorizing safety incidents to uncover systemic problems that can be risk ranked and prioritized by loss production impacts to the organization. This guides deep-dive process and design engineering teams to high value, high impact returns on problem solving exercises. In its use of enterprise-wide benchmarking KPIs, the methodology has proven to dramatically reduce process safety and environmental incidents, improve equipment reliability and reduce maintenance costs…saving millions of dollars in annual production losses (***every 1% gain in mechanical availability is worth about $8 million of additional margin capture per year in a typical 200,000 bpd refinery***).

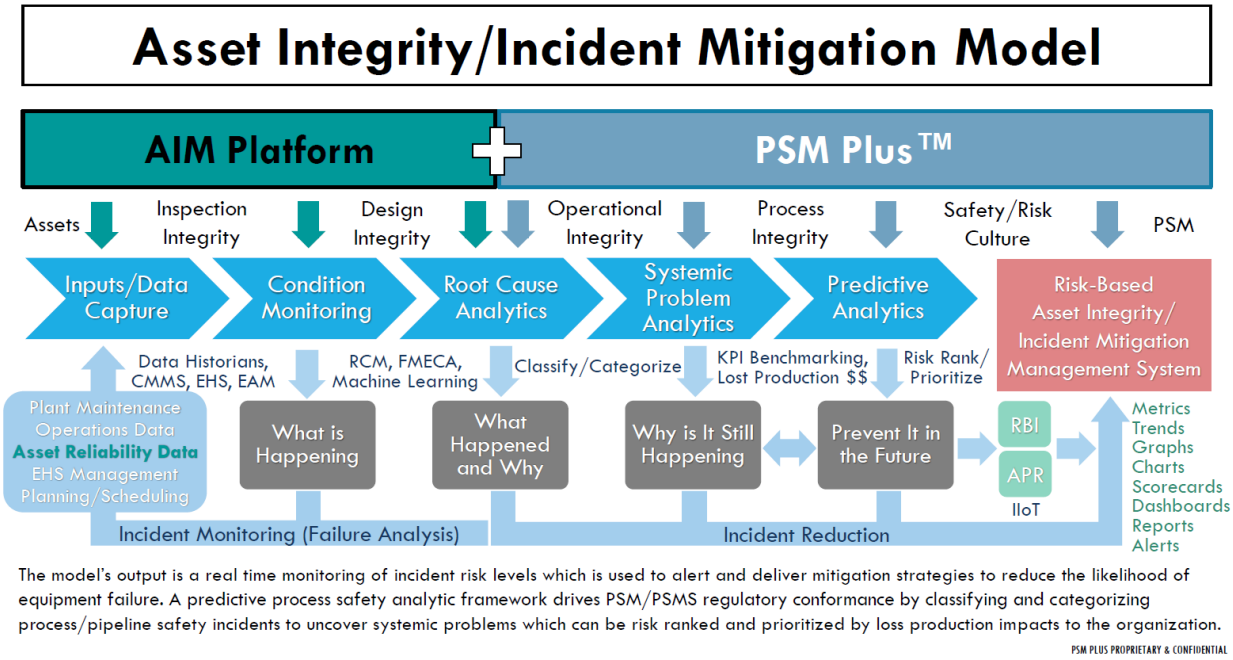


Figure 6. Asset Integrity Platform with *PSM Plus* Tools and Analytics

Rather than just another pioneering effort, *PSM Plus* is instead a proven management systems methodology (of people, processes, tools/technology) which can either be configured into an existing knowledge management system (like Metegrity Visions™, Enablon, Gensuite, Visium KMS, Sphera, etc.) or "bolt-on" as a standalone module deployed as a powerful complement to APR and machine learning products (like TrendMiner, Seeq, Falkonry, ECG, GE SmartSignal/Predix, etc.,), the powerful combination of which is further enabled by the cost-effective proliferation of wireless sensor technologies (from Emerson, Honeywell, Siemens, ABB,

Flowserve, Endress+Hauser, etc.) as a powerful complement to specific and unique technology offerings.

Many market-leading AIM and EHS platforms as well as countless other disparate IIoT software companies offer an array of predictive tools and analytics integrated with asset monitoring, but none incorporate a "real world" analytic framework like that of *PSM Plus* which utilizes field-tested and proven metrics, asset classification and categorization, risk ranking and prioritization, KPI benchmarking and associated economic impacts (production plus direct losses) (**Figure 6**).

Rather than first understanding the application environment and then focusing on how software solutions can help, many software designers start with a "solution" and then search for problems to solve. To the contrary, and as based on decades of SME experience, *PSM Plus* was developed with a hands-on, Reliability Centered Maintenance (RCM) "closest to the work" mentality as well as a first-hand appreciation of management rank dynamics from field supervisor to department manager to plant manager to corporate VP.

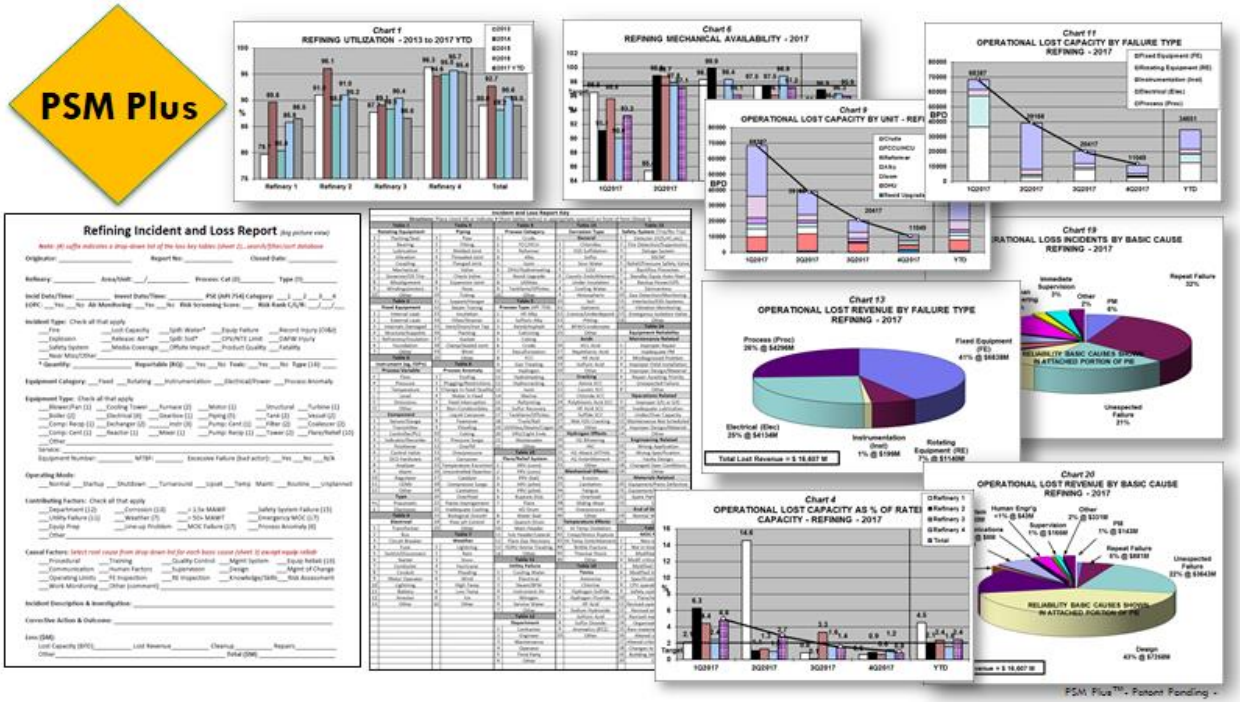


Figure 7. *PSM Plus* Incident Management System RCFA, Data Views, Reports

With that reporting hierarchy in mind, the primary goal of *PSM Plus* is to analyze and trend cost minimization, drive asset optimization and conformance to process safety RAGAGEP (recognized and generally accepted good engineering practice) not for just any one facility, but across all facilities as well as enterprise-wide, and ultimately throughout industry (via API 754 adaptation). ***"Following the leader" in a range of best-in-class to next-to-last is what RAGAGEP conformance is all about,*** and in this highly regulated industry, there is strength as well as comfort in numbers.

There are some 650 major refineries globally and many hundreds more petrochemical plants (not to mention pipelines and midstream assets). Given the widespread application (now at nearly all US facilities and growing internationally) of the relatively new Tier 754 incident benchmarking standard and the abundance of data being collected, the opportunities for an IIoT asset monitoring application coupled with the incident investigation and reporting analytic framework of *PSM Plus* are numerous and especially ripe for the early adopters **(Figure 7)**.

Besides enhancements to process safety, just considering that ***every 1% gain in mechanical availability is now worth about $8 million of additional margin capture per year in a typical 200,000 bpd refinery,*** the low-cost, high impact potential of *PSM Plus* is well worth exploring as a complement to any predictive maintenance/analytics platform.

Conclusion

The regulatory climate changed considerably following the highly publicized incidents at BP Texas City in 2005, Tesoro Anacortes in 2010, Chevron Richmond in 2012, and ExxonMobil Torrance in 2015. Each happened not due to a failure of equipment, instrumentation, facility siting, operator, procedure, communication, supervision, or training, but rather a failure of all those things together, i.e., ***a management system failure***. In addition to tens of millions of dollars in enforcement actions, legal consequences are now getting personal as was the case for plant management in the aftermath of the 2011 Chevron Pembroke incident. Systemic failures were cited in a May 2019 sentencing hearing at Swansea Crown Court with the plant declared "fundamentally unsafe" due to a series of errors and failings that contributed to a multi-fatality incident.

"These systemic failures are manifest failings which paint a picture of a workplace which had become, over time, fundamentally unsafe" …Chevron Pembroke refinery explosion, Swansea Crown Court, May 2019 sentencing

"Should have been foreseen and acted upon"

So, could a Chevron Pembroke type incident be averted by your process safety programs as they exist today? From an asset integrity perspective, do you have formalized programs and tools at both the site and corporate levels for effectively identifying and communicating systemic failures? Are you keeping up with industry pace-setters regarding asset integrity and process safety RAGAGEP? And, how might regulators answer these questions for you should a catastrophic incident occur? *PSM Plus* was designed to address these concerns.

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About the Author:

Michael Marshall, PE is an Oil & Gas industry consultant with Michael Marshall LLC (email: michael@tratusgroup.com), and has 39 years' experience working in the downstream, midstream and petrochemical industries. While working first with Chevron (10 years) and then Marathon Petroleum Company (23 years - retired), he progressed through various in-plant and corporate refining facility and project engineering, operations, maintenance, and equipment inspection/reliability supervisory and managerial positions. It was Mike's many years of hands-on experience while serving in frontline engineering, operations, maintenance and inspection roles which instilled in him the importance of properly designed risk minimization and management systems, performance metrics and KPIs. He has unique insight and expertise in areas of risk-based design relative to loss of containment (LOPC) damage mechanisms, safety integrity systems and overpressure protection.

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The Insanity we call Process Safety

Syeda Zohra Halim, Noor Quddus, and Stewart Behie
Mary Kay O'Connor Process Safety Center, Texas A&M University System
College Station, TX 77843, United States
zohrahalim@tamu.edu

ABSTRACT

The overall number of incidents in petrochemical industry in USA has varied over the years, but the coverage gained by a number of incidents in 2019 alone has been significant. A large number of these significant incidents have been in Texas, especially in and around the oil and gas hub near Houston. These incidents have involved fatalities and injuries, large fires, and some with explosions or long plume hovering over the city of Houston. In one occasion, the emergency response was extended for several days. As one incident made it to the news, the next incident had raised more questions and concerns among regulatory authorities as well as the public. Analysis of the recurrence of such major incidents in the recent years may indicate a deficiency in the underlying process safety measures across the industry.

Process safety encompasses the safety triad: prevention, mitigation and response. Understanding of what causes incidents and taking proactive measures is important to prevent them in the first place. Many companies have good safety programs in place to prevent incidents, however the incidents keep on happening. It is essential to identify how to build a stronger safety triad and take proactive measures against the issues to reduce the incidents. The current paper looks at the factors at play that significantly contribute to the failure behind the incidents and proposes measures to address these factors. From the analysis of the factors, it is evident that both short term and long-term planning and implementation is required by companies in collaboration with regulatory agencies and academic institutions.

Keywords: Process safety excellence

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Unified Wall Panel System (UWPS) - A Value Engineering Solution for Protective Construction in the Petroleum Industry

Scott Hardesty, A. Mangold, and K. White

Applied Research Associates / Protective Technologies Group / ATFP Associates

shardesty@ara.com, amangold@ptgus.com, ken.white@atfpassociates.com

Abstract

Employee and operator safety at all petroleum facilities is a constant that must be maintained at the highest level, regardless of production status or industry financial climate. Value engineering offers protective construction solutions at reduced cost, decreased installation time and long operational life expectancy. The Unified Wall Panel System (UWPS) is a value engineered methodology that offers full-performance protective qualities in the industrial safety regimes of blast overpressure, fire/thermal loading, seismic events, high wind, ballistic and fragmentation resistance. The UWPS is a composite structure wall configuration developed through extensive R&D, full-scale testing and real-world applications by industry leaders in hazard analysis, threat mitigation and facility construction. New and innovative applications of technologies include novel employment of high-strength cementitious structural paneling, utilization of non-aramid advanced mineral fiber reinforcement, and metallic foam energy absorption. These cutting-edge technologies provide key advantages in the modular UWPS concept currently being considered for installation across a wide spectrum of applications, including military facilities, public utilities, academic institutions, medical campuses and a variety of publicly or privately accessible locations. The UWPS approach provides for the option of a shelter-in-place response for operational continuity during a catastrophic event driven by manmade and natural incidents. Petroleum industry safety in operational zones, control rooms, and administrative areas can be enhanced through utilization of the UWPS and value engineering methodology.

Keywords: UWPS, Value Engineering, Fire, Blasts, Cementitious Panel, Mineral Fiber, Metallic Foam

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Protect Process Plants from Climate Change

Victor H. Edwards
VHE Technical Analysis
P. O. Box 940849
Houston, TX 77094-7849
E-mail: m2vhe@att.net

Abstract

Climate change is occurring and some of the changes will impact the safety of process plants. This paper summarizes potential adverse impacts on process plants and then outlines how to conduct a Climate Risk Vulnerability Assessment (CRVA) for a process plant. One product of the CRVA is a plan of action to eliminate or reduce adverse climate effects on the process plant.

Keywords: Climate change, Process plant, Climate Risk Vulnerability Assessment, Effects on process plants, Hazards from climate change.

Climate change is happening

"The vast majority of actively publishing climate scientists – 97 percent – agree that humans are causing global warming and climate change." (NASA, 2020)

These changes will impact the biosphere and human civilization.

What is climate change? It is a change in global or regional climate patterns.

In particular climate change is a change apparent from the mid to late 20th century onwards and attributed largely to the increased levels of atmospheric carbon dioxide produced by the use of fossil fuels.

Table 1 lists some of the effects of climate change.

Some of these changes are already occurring.

Table 2 outlines some of the potential effects of climate change on process plants (Edwards, 2019; Kulinowski, 2019).

Stronger hurricanes are just one possible future hazard. Figures 1, 2, and 3 are illustrations of some of the effects of Hurricane Rita in 2005 in Southwestern Louisiana on process plants (Courtesy of Roy Sanders).

Table 1 – Potential Effects of Climate Change

- **Higher temperatures**
- **More frequent and stronger droughts; more frequent and stronger wildfires**
- **Stronger hurricanes, tropical cyclones, tornados, and other storms**
- **Higher rainfall events leading to flooding**
- **Rising sea levels**
- **Adverse effects on the biosphere, such as extinction of some species**
- **Impacts on human activities such as human displacement, disruption of supply chains, and disruptive climate effects**

Table 2 – Potential Impacts on Process Plants

- **Flooding with equipment damage**
- **Wind damage to process plants from tornados, tropical storms, and straight-line winds**
- **Water shortages of process water, cooling water, and/or potable water**
- **Power outages**
- **Disruption of supply chains**
- **Market disruptions**
- **Process upsets**
- **Loss of containment of process fluids, potentially leading to fire, explosion, toxic gas release and/or environmental pollution**
-

Figure 1 – Cane covering pipe rack in plant in Lake Charles, LA, 2005

Figure 2 – Cooling tower fan housing, Lake Charles, LA, Hurricane Rita, 2005

Figure 3 – Tank Battery in Cameron, LA, after Hurricane Rita 2005.

Hurricane Rita came ashore on September 24, 2005 as a Category 3 hurricane with winds of 115 mph. It caused $18.5 billion (2005) in damage. Figures 1 through 3 show examples of damage.

Hurricane Laura came ashore at Cameron, Louisiana on Thursday, August 27, 2020 as a Category 4 storm. Based on windspeed, Laura was the fifth strongest on record in the US history. Laura's 150 mph winds made it the strongest hurricane to pass over the state of Louisiana, matching the 1856 Last Island Hurricane in intensity. Process plants in the area experienced damage from Laura, but details are not available at this time of writing.

Protect your plant from climate change

Your plant may already be at risk from extreme weather events and the risk may increase with time due to effects of climate change. One approach to identify hazards and protective measures to reduce risk is to conduct a Climate Risk Vulnerability Assessment (CRVA). Numerous CRVA's have been conducted in the last two decades to evaluate climate risk scenarios and to plan risk reduction measures for people and for a number of types of infrastructure, such as coastal communities, river systems, crops, bridges and urban water and wastewater systems. To date, I have been unable to find any published CRVA's for process plants.

What changes should occur during the siting, design, and construction of process plants?

•

• First, conduct a Climate Risk Vulnerability Assessment (CRVA). This will be illustrated by adaptation of the U S Climate Resilience Toolkit.(https://toolkit.climate.gov/ Accessed 12 September 2020) for potential plant sites in a five step process See Table 3.

Table 3 –Five Steps to Resilience (Climate Risk Vulnerability Assessment-

---

1. **Explore hazards**
2. **Assess vulnerabilities and risk**
3. **Investigate options**
4. **Prioritize and plan**
5. **Take action**

Step 1. Explore Hazards

1.1 - Gather a CRVA team of qualified people who want to protect local assets.
Table 4 (below) is an example. Process Hazards Review experience with this plant a plus because climate change effects may increase process hazards.
It may also be useful to include local leaders in the discussions of potential local effects.

Table 4 - CRVA - Example Team Members

Facilitator (often a chemical engineer)

Project Engineer (part-time)

Process Engineer or Technologist

Mechanical Engineer (part-time)

Civil Engineer

Electrical/Instrument Engineer (part-time)

Production representative

Maintenance representative

Climatologist/Meteorologist (part-time)

1.2 - Check past weather events and future climate trends.

Climate Explorer is just one free software package available from the US Climate Resilience Toolkit.- Explore climate projections for any county or parish in the contiguous United States or borough in Alaska with this software. Compare maps of past and projected conditions for coming decades. Consult historical records to see when weather has veered outside of normal climate.

Other software packages available from U S Climate Resilience Toolkit – For example, river flooding prediction and storm surge prediction

For an example of advanced quantitative modeling of hazardous weather events see Kytomaa, et al., 2019.

1.3 List in a CRVA spreadsheet the things you value that could be damaged.

In the spreadsheet, list the plant assets with one line for each asset. Table 5 is an example of some of the assets of interest in a typical process plant. The facilitator of the CRVA normally prepares the draft CRVA spreadsheet prior to the group review by the CRVA Team.

Table 5 – Examples of Plant Assets Potentially at Risk
Plant administration building
Cooling tower
Feed storage tank
Electrical substation
Product storage tank
Feed pump and motor
Heat exchanger
Reactor
Structure of Reaction Building
Distillation column
Biotreatment system

Table 6 is a list of the column headings of a CRVA spreadsheet. This spreadsheet is adapted from the U. S. Climate Resiliency Toolkit. Table 7 is the upper left portion of a typical CRVA spreadsheet. Each line is for a specific asset. When an asset is at risk from more than one hazard, additional lines are added for each hazard. For example, the plant administration building would be at risk from both flooding (first line) and hurricane (second line).

1.4 Assess the plant's current physical status

- For existing plants, inspect facility to detect and document any deficiencies in assets, such as mechanical damage, corroded structures, vessels, or equipment and adequacy of drainage features; Record deficiencies in brief report and highlight in column two of CRVA spreadsheet. Also indicate year asset was placed in service.
- For existing plants, review design basis of structures, vessels, and equipment to see if existing facilities meet current building codes and equipment standards; identify necessary modification to meet load demands, Record deficiencies in brief report and highlight in column two of CRVA spreadsheet.
- These two activities are best conducted by appropriate members of the CRVA team to build team knowledge prior to team CRVA risk review meetings.

Table 6 – Column Headings of a CRVA Spreadsheet

- **Key Assets or Resources**
- **Physical Condition**
- **Weather or Climate Hazard**
- **Potential or Historical Consequences**
- **Climate stressors and trend**
- **Non-climate stressors and trend**
- **Potential tipping point and probability (probable, increasingly probable, decreasingly probable, improbable)**
- **Sensitivity (high, medium, low)**
- **Adaptive Capacity (high, medium, low)**
- **Vulnerability (high, medium, low)**

Table 7 – Upper Left Corner of a CRVA Spreadsheet

Key Assets or Resources	When Installed & Condition	Weather or Climate Hazard	Potential or Historical Consequences	Climate stressors and trend	Non-climate stressors and trend
Plant administration building	1990/Fair	Record rainfall	Flooding causing Structural damage; water damage	Rainstorms of increasing frenquency and intensity; increasing trend	Upstream development of watershed expanding flood plain; increasing trend
Cooling tower	2005/Good	Hurricane	Collapse of cooling tower; extended loss of cooling causing plant shutdown	Hurricanes of increasing frequency and intensity; increasing trend	Corrosion; increasing trend
Product storage tank	1990/Fair	Hurricane	Collapse of storage tank; loss of containment of product; potential fire; environmental damage	Hurricanes of increasing frequency and intensity; increasing trend	Storage tank near end of life; corrosion; increasing trend

Power substation	1989/Poor	Record rainfall	Loss of electrical power; damage to electrical equipment, extended piant shutdown	Heavy rain events, consecutive days of rain, increasing trend	Electrical equipment near end of life; increasing trend

2 Assess vulnerabilities and risk with CRVA team review

Conduct team CRVA risk review. Determine which of your assets are subject to harm

Assess each asset's vulnerability

Estimate the risk to each asset

Document risk review in CRVA spreadsheet. After this exploration, you'll discover if weather and climate represent a hazard to things you value.

3. Investigate options

Consider solutions to mitigate highest risk.

Check how others have responded to similar risks

Reduce your list to reasonable actions

4. Prioritize and plan

Evaluate costs, benefits, and your team's ability to accomplish each option

Rank the expected value of each option

Integrate the highest value actions into a stepwise plan

5. Take action

Move forward with the stakeholders who accept responsibility and bring resources to take action.

Check to see if your actions are increasing your resilience.

As you move forward, you'll monitor, review, and report on your project.

Summary and Conclusions

Climate change is happening and some of its effects present risks to process plants. Reported here is a method to conduct a Climate Risk Vulnerability Analysis (CRVA) for a plant at risk. The purpose of the CRVA is to identify those climate hazards and to plan modifications to eliminate or reduce those risks by adapting and using the U. S. Climate Risk Resiliency Toolkit

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Process safety implications in a changing environment

Trish Kerin
IChemE Safety Centre
Level 7, 455 Bourke St Melbourne, VIC Australia
tkerin@icheme.org

Abstract

While much research has been undertaken in natural hazards triggering technological disasters (Natech) it still remains a challenging area. It can be difficult to move past the psychological bias to focus on the possible incident outcome without discounting a seeming incredible cause. We have seen some notable instances of Natech incidents, including the Fukushima Daiichi Nuclear Power Plant meltdown following an earthquake and subsequent tsunami, as well as the impacts of Hurricane Harvey on the Houston industrial facilities. More recently, there is a natural disaster of significant proportions taking place in Australia, with a prolonged and intense bushfire season. As at January 8, 2020 over 10,700,000 hectares have burnt across 7 states and territories (only one territory remains unburnt) and the fires are not yet all under control. This has burnt a significant range of environments, even razing whole towns. Twenty-eight people have lost their lives during this fire season to date. These towns contain gas storage and water treatment facilities, which can have ongoing process safety implications. Major cities are clogged by smoke, creating ongoing health issues for the public. This paper will discuss the impact of the 2019-2020 Australia bushfires and the assumptions made on how to prepare for a natural disaster.

Keywords: Hazard, Case Histories, Risk assessment, Natech

Australian Bushfires

Bushfires, sometimes known as forest fires in other countries, are a natural part of the Australian landscape. The season for the states in Southern Australia usually commences in summer (December to February) and continues on to autumn or fall (March to May). In Queensland and

New South Wales, the season starts earlier in spring (September to November) or early summer. The intensity of the fires is impacted by several factors, including fuel load, fuel moisture, wind speed, ambient temperature and relative humidity largely. While some fires are the result of arson, but the most common sources of ignition is natural, being a dry lightning strike [1]. As Australia is generally a hot, drought prone land mass, it is at risk of fire each year.

Indigenous Australians have used fire as a land management and hunting tool for tens of thousands of years. The National Museum of Australia states "The first bushfires in the colony were reported in 1797. As Aboriginal people were driven off their land, their regime of low-intensity fire management went with them, and bushfires became more prevalent. From then, the government sought to limit Aboriginal and settler use of fire as an agricultural tool." [2]

When the fires are particularly intense, a fire storm situation can develop, where the fire creates its own weather system. Colloquially known as "pyrocumulus" a fire can create a "flammagenitus" formation [3]. See Figure 1 for their formation details. Where the weather system is strong enough, fire tornadoes can also form [4].

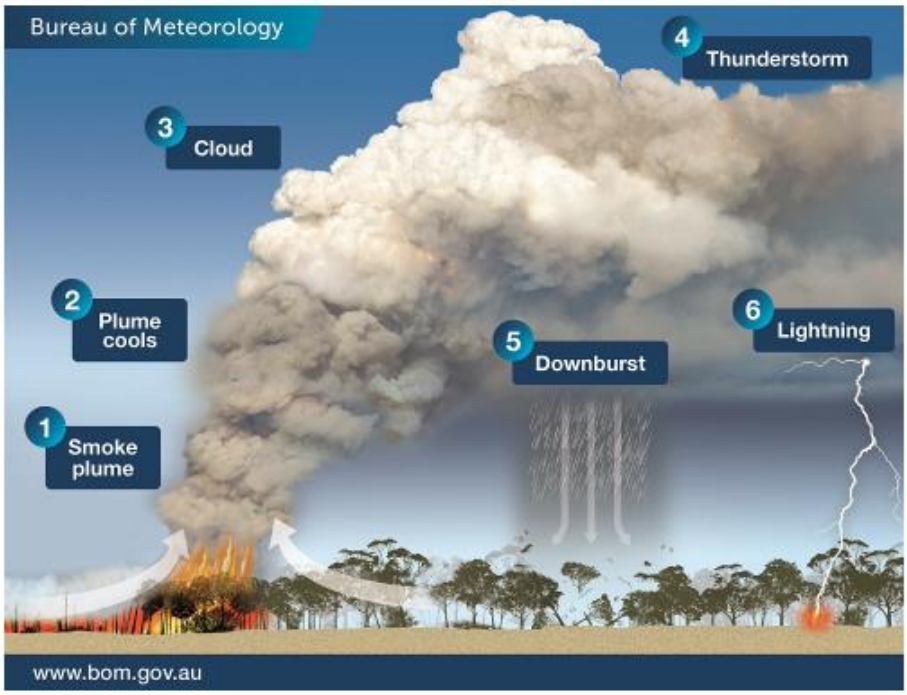

Image: Pyrocumulonimbus cloud development

1. A plume of hot, turbulent air and smoke rises.
2. Turbulence mixes cooler air into the plume, causing it to broaden and cool as it rises.
3. When the plume rises high enough, low atmospheric pressure causes its air to cool and cloud to form.
4. In an unstable atmosphere a thunderstorm can develop: pyrocumulonimbus cloud
5. Rain in the cloud evaporates and cools when it comes into contact with dry air, producing a downburst.
6. Lightning may be produced and can ignite new fires.

Figure 1. How pyrocumulus develop [3]

Figure 2. Photo of fire tornado. Proto by Brett Hemmings ©2019 Getty Images

In a high wind situation, the fires can ignite well in advance of the fire front, via wind borne embers, effectively resulting in the fire advancing at very fast speeds [5].

These conditions can lead to catastrophic bushfire seasons in Australia. The five deadliest bush fires on record in Australian are shown in Table 1. It should be noted, this list does not include the fires of the 2019-2020 season. The total area burnt in the 5 most deadly bushfires as noted in Table 1 is 2,955,300 hectares.

Date	Name	Fatalities	Other details
7-8 February 2009	Black Saturday (Victoria)	173	3500 buildings destroyed including >2000 homes, >450,000 hectares burnt [2]
16-18 February 1983	Ash Wednesday (Victoria and South Australia)	75	Over 2,800 buildings destroyed >486,030 hectares [6]
13-20 January 1939	Black Friday (Victoria)	71	Over 650 houses destroyed and 1,355,000 hectares [7]
7 February 1967	Black Tuesday (Tasmania)	62	1300 homes razed, 264,270 hectares [8]
1 February to 10 March 1926	Gippsland fires and Black Sunday (Victoria)	60	Estimated 400,000 hectares [9]

Table 1. Five deadliest bushfires on record in Australia [10]

The 2019-2020 Season

The Australian Bushfire season in 2019 commenced in September 2019, which is quite early, and continued into March 2020. Significant fires raged through November and December as well as into January, through March. By comparison the fire in Table 1 all occurred between mid January to March. This suggest that the season was a long and unprecedented one.

In the 2019-2020 fire season in excess of 17,000,000 hectares burned across New South Wales, Victoria, Queensland, Australian Capital Territory, South Australia and Western Australia. Including more than 80% of the World Heritage listed Greater Blue Mountains and 54% of the NSW components of the World Heritage listed Gondwana Rainforests [11]. This area burnt is almost 6 times the land area of the 5 deadliest fires combined (Table 1), or by comparison approximately one quarter the size of Texas. However these fires raged across the country, covering the equivalent distance from New York City to Orlando and across from Jacksonville to Phoenix. It is not known how many animals perished in the fires, but is estimate to be in excess of one billion [11].

Thousands of people fled to beaches as the fires bore down on them, some stuck there for several days. Many were eventually evacuated by a Navy rescue mission from the sea [12]. Figure 3 shows people huddled on the Mallacoota Pier during daylight as the fire threatened the town.

The scenes on Mallacoota pier as the fire bore down on the township. *(Instagram: @Travelling_aus_family)*

Figure 3. Impact of the fire on visibility during daylight, showing Mallacoota Pier

As a result of the fires, smoke haze impacted major cities along the South East of Australia. Figure 4 shows the smoke over Sydney Airport on the afternoon of 5 December 2019. The smoke posed

not only a health hazard due to airborne particulates but also triggered building and facility smoke alarms in several cases [13].

Figure 4. Photo of sun from Sydney Airport at approximately 5pm © T Kerin

Process Safety Implications

There are a number of process safety implications to a natural event such as this bushfire season. The areas impacted are mostly small regional towns, however this does not mean they do not have industry that can be impacted. Table 2 lists the types of impacts that are possible. These are in addition to the fires causing disruption to emergency response equipment and facilities. Fire fighters are focused on the bush fire impact, and unable to respond to other types of incidents. Road closures occur, impacting escape routes and other forms of emergency response.

Equipment/facility type	Process safety implications
Liquid Petroleum Gas (LPG) and other fuel storage	Potential fire impact on fuel sources Potential heat damage to storage tanks and ancillary equipment Potential triggering of fire detectors
Waste disposal facilities	Potential for fire to spread to waste resulting in unknown substances being burnt Potential inability to contain waste fires

Timber mills	Potential for fire to spread to timber mills and woodchip piles Potential inability to contain wood chip fires Potential impact on treatment chemicals used
Water treatment facilities	Potential impact on treatment chemicals used
Refrigerated warehouses	Potential impact on ammonia or other refrigerant gas storage
Dairies	Potential impact on ammonia or other refrigerant gas storage Potential impact on cleaning chemicals used
Wineries, breweries and distilleries	Potential impact on cleaning chemicals used Potential impact on alcohol storage tanks
Electricity generation	Potential impact on fuel sources (gas, coal, distillate) Generation disrupted and impacting customers safety
Unconventional gas extraction	Potential fire impingement on extraction and production facilities
Aviation	Inability to fly through smoke, both low visibility and engine damage risk [14]

Table 2. Process safety impacts due to bushfires

Process Safety Learnings [15]

A key to responding to bushfire scenarios is adequate response planning and robust business continuity planning. This is a core principle of Natech management. Natech triggers include events such as earthquakes, hurricanes, floods, lightening, volcanic eruptions and bush fires to name a few. Natech events require specific risk assessment and planning. There are also several assumptions that need to be made when considering Natech. These include considering loss of utilities and communication systems, roads becoming unusable, the inability to evacuate people and additional demands placed on emergency responders. It is also common to see domino affects during Natech events.

Consideration should be given to the vulnerability of safety systems and recognize they may be unavailable then needed. For example, items such as sensors and alarms may not be available due to power outage or fire impingement.

Emergency plans should be assessed taking into consideration possible Natech events. While evacuation may be a credible response, it should be assumed roads may not be available. It should also be assumed that hazardous material leaks could occur, impacting the wider population and rescue efforts. From a Natech perspective it should be assumed that consequences could eventuate, even then they seem unlikely. These consequences should be considered irrespective of the cause.

Conclusion

Adequate risk assessment, planning and emergency response activities are critical to manage the impacts of a natural disaster. We need to think broadly about the consequences that could impact our facilities, regardless of the cause and ensure that we have some type of response planned. It may not always be economically feasible to design protections for these types of scenarios, but we do need to have credible response plans available to address them if they eventuate.

From a bushfire perspective, there are some strategies which can be implemented to minimize the impact of a bushfire season. These include controlled fuel reduction burns to reduce the available fuel loading in the warmer months, as well as land clearing away from properties that need protecting. These go a long way to reducing the likelihood of fire striking a particular area. From a mitigation perspective the use of fire retardant materials and coatings can minimize the spread once the is present, and properly designed fire bunkers can save lives if shelter in place is needed.

We cannot eliminate the occurrence of natural hazards, we must learn to live with them and protect our people and facilities from them. But to do this we need to better understand the phenomena and adequately plan our response.

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A critical evaluation of industrial accidents involving domino effect

Ravi Kumar Sharma*[1], Nirupama Gopalaswami[2]
[1]Centre for Excellence in Disaster Mitigation and Management,
Indian Institute of Technology Roorkee, 247667, India
[2]Honeywell Process Solutions, Houston, TX 77042, United States
ravikrsharma16@gmail.com

Abstract

The domino effect has been responsible for several catastrophic accidents that have occurred in petro-chemical processes and the storage industry. In this study, 326 accidents since 1961-2017 involving the domino effect in process, storage plants and the transportation of hazardous materials were analysed. Coding of incidents was done based on data obtained from different sources. The domino incident database analysis includes several categories such as fatalities over time, incidents over time, and incidents with respect to location, materials involved, causes and consequences. The analysis has shown that explosions are the most frequent cause of the domino effect followed by fires. The accidents involving domino effects show that process plants and transportation sector are the most probable industry for a domino accident (33%) followed by storage terminals (20%). The domino effect sequences were analysed using relative probability event trees, which may be useful in further work to understand the domino effect and reduce the probability of its occurrence in future. In the present study, we have proposed a three-stage methodology for the assessment of risk due to the domino effect. The results show that quantitative risk assessment of escalation hazard is fundamentally important to address preventive measures.

Keywords: Domino effect; Accident database; Risk analysis; Major accident hazard; Fires; Explosions

1. Introduction

The domino effect has been responsible for several catastrophic accidents that have occurred in petro-chemical processes and the storage industry. The consequences of these accidents are at various levels and may affect not only the industrial sites but also people, economy and the environment. The destructive potential of such accidents is widely recognized but very less attention has been paid to this subject in the scientific and technical literature [1-4]. In the area of risk assessment, the domino effect has been documented in the technical literature since 1947 [4-5]. However, no well-assessed procedures have been developed for the quantitative assessment of risk caused by the domino effect. Therefore, the assessment of domino accidental events remains an unresolved problem. Moreover, there is widespread uncertainty in the escalation criteria and in the identification of the escalation sequences that should be considered in the analysis of domino scenarios, either in the framework of quantitative risk assessment or in land-use planning. The probability of domino effects is relatively high due to the development of industrial plants, proximity of such facilities to other installations, their inventories and the transportation of hazardous substances [5].

The severity of domino accidents has caused concern in the legislation and in technical standards aimed at the assessment and prevention of accident escalation. Therefore, the European legislation widely recognized the assessment of domino hazards since the first "Seveso" Directive (Directive 82/501/EEC), which was adopted in 1982. Currently, these requirements have been extended to the assessment of possible ''domino'' scenarios both on-site and off-site. Such requirements are compulsory for industrial sites falling under the obligations of the ''Seveso-II'' Directive (Council Directive 96/82/EC), as amended by Directive 2003/105/EC [6-7]. Therefore, the domino effect is a significant concern in risk analysis. A good understanding of the main hazards and features of this phenomenon can help identify additional safety measures, facility siting studies such as minimum safe distances between certain types of equipment.

In spite of the relevant attention dedicated in the legislation, there is no well-accepted approach to date for the analysis of domino related hazards. Several authors have analysed the categories involved in domino accidents. Bagster & Pitblado [8] and Khan and Abassi [9] analysed the probability of occurrence and adverse impacts of such 'domino' or 'cascading' effects. Cozzani and Salzano [10-11] studied the contribution of a blast wave as a primary event and assessed the overpressure threshold values for damage to equipment caused by blast waves originating from primary accidental scenarios. Reniers [12] analysed the efficiency of current risk analysis tools for preventing external domino accidents. They proposed a meta-technical framework for optimizing the prevention of external domino accidents, emphasized the importance of combining inherent safety criteria with conventional active and passive protection [12-13]. Antonioni *et al.* [14] developed a methodology for quantitatively assessing the contribution of domino effects to overall risk in an extended industrial area. Subsequently, several technical standards have introduced preventive measures, such as safety distances, thermal insulation or emergency water deluges, etc. to control and reduce the probability of domino events. However, a relevant uncertainty exists in the threshold values assumed in such assessments [2,11].

Similar studies on pipeline domino effects were conducted by Ramírez-Camacho *et al.* [15]. The authors studied the possibility of domino effect in parallel pipelines with different configurations. They had concluded that the main risks associated with the domino effect are erosion by fluid-

sand jets and the thermal action of jet fires in pipelines. Ramírez-Camacho *et al.* [16] analyzed about 1063 pipeline incidents to illustrate the risk associated with Pipeline and Hazardous Materials Safety Administration (PHMSA) maintains a public database of pipeline incidents. A 20-year trend analysis by PHMSA on significant incidents reveal losses in pipeline industry amounting to 7 Billion dollars in USA alone [17]. Similar major losses due to pipeline incidents have been reported regularly by European Gas Pipeline Incident Data Group (EGIG) in Europe, the Transportation Safety Board (TSB) and National Energy Board (NEB) of Canada [18-19].

In view of above, the domino effect is an important aspect of risk assessment because the understanding of main hazards and features of the phenomenon can be used to introduce additional safety measures. The past accident analysis in chemical process industries bestows great importance on identifying their triggers, sequences, and consequences. Retrospection can provide pointers for developing accident prevention strategies.

Due to its importance, the compilation and analysis of 326 past accidents involving domino effects have been carried out in this paper to study their behaviour. The analysis reveals that explosions were responsible for domino effects in almost 57% of the cases, followed by fires (43%). Explosions and fires can cause subsequent accidents, and their physical effects can trigger a domino sequence [20]. The severity of the ensuing scenario can considerably increase the influence of a domino effect. A historical analysis of domino effects carried out by Darbra *et al.* [3] show that 59.5% of accidents in seaport areas were due to fires, 34.5% were explosions and 6% were toxic clouds. The assessment carried out in the present study shows that 34% of domino accidents have occurred in process plants and transportation section whereas 20% in storage terminals of hazardous materials. Storage areas, which usually contain large amounts of hazardous materials, are also common settings for domino effect scenarios. This is evident with the recent incident in Tiajin where a series of explosions killed 173 people and injured hundreds of others at a container storage station [21].

The domino effect sequences were analysed using relative probability event trees. The most frequent sequences were i) explosion→ fire (26%), ii) fire→ explosion (20.3%), and iii) fire→ fire (12%). In the last decade, three major petroleum storage area accidents occurred in Buncefield, UK (2005), Puerto Rico, USA (2009), and Jaipur, India (2009) [22]. In addition to this, Amuay refinery accident occurred in Venezuela on the 25 August 2012 [23]. A similar study on 226 domino incidents mainly focused on developing countries and concluded that the most frequent domino accident sequence was explosion → fire (24.8%) followed by explosion → fire → explosion amount to about 8% [24].

Few authors have analysed historical surveys of the domino effect. For example, Abdolhamidzadeh *et al.* [1] have presented an inventory of 224 major process industry accidents involving 'domino effect'. Darbra *et al.* [3] examined 225 accidents involving the domino effect, which occurred from 1961 to 2007. The aspects analysed included the accident scenario, the type of accident, the materials involved, the causes and consequences, and the most common accident sequences. Kourniotis *et al.* [25] examined a set of 207 major chemical accidents that occurred between 1960 and 1998, 114 of which involved a domino effect according to their criteria. Ronza *et al.* [26] performed a survey of 828 accidents in port areas and constructed relative probability event trees to analyse the sequence of the 108 accident scenarios in which a domino effect was observed. This paper addresses the development of revised criteria to assess the possibility of

escalation of accidental scenarios, resulting in domino accidental events. The main purpose of the study was to obtain a better understanding of the causes of hazardous event escalation and mitigation measures that prevent transforming minor accidents into disasters.

2. Definitions and event sequence of domino effects

There is no universally accepted definition of the term 'domino effect' in the context of accidents in the chemical process industries to date. Domino effects have also been widely used interchangeably with cascading events. Most of the scientists define the situations wherein a loss of containment accident in a process unit becomes the trigger for one or more loss of containment accidents in the same or adjacent process units.

Lee [27] defines the domino effect as "*a factor to take into account of the hazards that can occur if leakage of a hazardous material can lead to the escalation of the incident"*.

Delvosalle [28] considers all of the aspects and define domino accidents as "*a cascade of events in which the consequences of a previous accident are increased both spatially and temporally by the following ones, thus leading to a major accident"*.

The AIChE-CCPS (American Institute of Chemical Engineers - Centre for Chemical Process Safety) [29] defines a domino effect as "*an incident that starts in one item and may affect nearby items by thermal, blast or fragment impact, causing an increase in consequence severity or in failure frequencies"*.

A recent definition provided by Cozzani & Salzano [10] is: "*a domino accidental event will be considered as an accident in which a primary event propagates to nearby equipment, triggering one or more secondary event resulting in overall consequences more severe than those of the primary event"*.

For pipelines the domino effect can be defined as event where in two or more pipes installed in the same hallway: a jet released from one of the pipes can seriously damage another one by abrasion (underground pipes) or thermal action (jet fire) [15-16].

Therefore, these definitions are used as a framework for the selection of accidents. Based on these definitions one can say that a relatively minor accident can initiate a sequence of events that causes damage over a larger area and lead to several severe consequences, which is typically referred as a *domino effect*.

According to Reniers [12], domino effects are classified into two categories: single-company (internal) domino effects and multi-company (external) domino effects. Internal domino effects signify an escalation accident occurring inside the boundaries of one chemical plant. In external domino effects, one or more secondary accidents occur outside the boundaries of the plant where the primary event occurs. Although external domino effects often have more severe consequences than internal domino effects, this phenomenon has received less attention from prevention managers in existing chemical clusters. The reason for this relatively extraordinary observation is threefold [12]. First, they are less frequent; second, their modelling is highly complex; and third,

they are difficult to investigate because several companies are involved. The analysis of technical literature and case histories concerning past accidents shows that all of the accidental sequences, where a relevant domino effect took place, have three common features namely event occurrence, their propagations and escalation vectors as shown in Fig. 1 and discussed below [30].

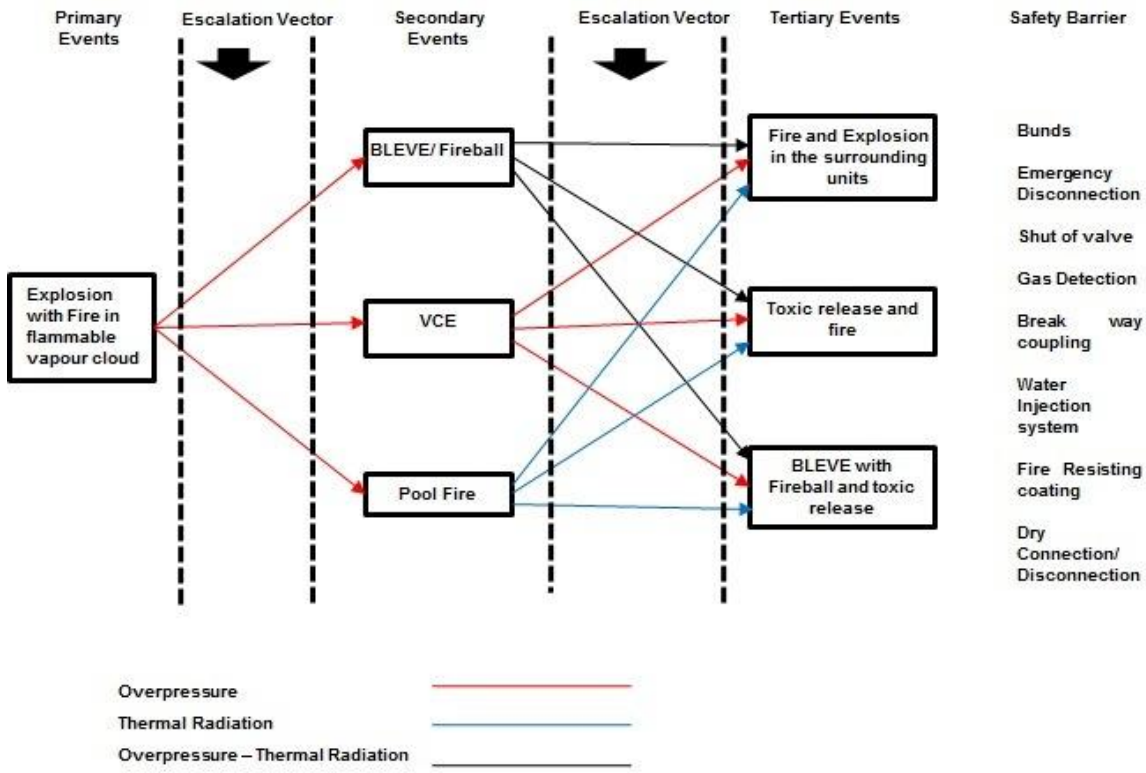

Fig. 1. Domino sequences that may be triggered by an initiating event explosion.

- A "primary" event (fire, explosion), which triggers the domino effect.
- The propagation of the accident enhanced by the effect of escalating vectors by which "secondary" accidents are triggered as a result of the primary event. (Domino effect enablers)
- An "escalation" vector that enhances consequence effects and propagations with secondary accidents having higher severity than the primary one.

Thus, it is important to understand that the propagation sequence is relevant only if it results in an 'escalation' of the primary event, i.e., triggered by an 'escalation vector' originating from the primary scenario. By these definitions, all knock-on accidents including the accidents that occur within a single process unit would fall under the umbrella term 'domino effect' [1].

2.1 Coding of domino events

To study the historical analysis of domino accidents, one of the critical tasks is to establish the criteria for differentiating domino accidents from non-domino accidents. The hurdles involved in interpreting records of past accidents create difficulties in conducting past accident analysis (PAA)

for stand-alone accidents and an increased level of difficulty for the PAA of domino events. Therefore, it is important to develop the most appropriate definition for the domino effect. There are certain well-known difficulties associated with the task of obtaining records of past accidents [1, 31-32] as listed below:

a) a well-established mechanism was not developed for reporting and maintaining records of domino accidents that occurred in many countries, particularly in the previous century;

b) industries and governments intended to under-report the accidents to reduce liability;

c) integral inaccuracy of several available records; for example, explosion and fire accidents were often recorded in a generic sense, and in several situations it was difficult to identify the specific event type;

d) contradictory descriptions of the incident and incompetence of assessment leading to an inability to resolve the uncertainty due to lack of unassailable evidence;

e) indistinct documentation of sequence of accidents in an incident

f) Lack of proper methodologies to address these scenarios in current risk assessment methodologies.

In developing countries, the lack of proper documentation and inventory of accidents obscure the involvement of the domino effect. Hence, there is no method to confirm whether the accident had involved a 'domino effect'. To classify a series of accidents as a domino event, it is necessary to establish that the event conforms to the definition of a domino effect. Usually incomplete and imprecise records of the past accidents create complications to determine whether more than one process unit was involved in an incident involving multiple accidents. We conducted this study recognising these limitations and surveyed the records of the following sources:

- The COMAH (Control of Major Accident Hazards) database
- MARS (Major Accident Reporting System)
- MHIDAS (Major Hazardous Incident Data Service)
- Lee's Loss Prevention in Process Industries
- FABIG- Fire and Blast Information Group FABIG
- The U.S Chemical Safety Board Reports
- NTB -U.S National Transportation Board Accident reports
- U.S Department of Transportation – Pipeline and Hazardous Materials Safety Administration (PHMSA) pipeline incident database
- TSB - The Transportation Safety Board of Canada
- EGIG - European Gas Pipeline Incident Data Group Reports

In present literature review, only accidents that occurred over the past 50 years were considered. Accidents that occurred prior to 1961 were excluded, as they happened in a technological environment in which safety measures and risk planning were not comparable with those currently

in place. This, although, has reduced the number of accidents studied but increased the quality and significance of the sample.

2.2 Major chemical accidents worldwide

Major chemical accidents are consistent with trends in industries for developing untapped and technologically challenging sources of hydrocarbons. Further, even with developments in engineering practice and hazard awareness, a decade-wise analysis demonstrates that large losses continue to occur. There are a number of good examples where the industry does appear to have learnt from incidents and made improvements on a global basis. Corporations involved in the hydrocarbon processing industry are becoming additionally sophisticated in their risk assessment and risk management approaches and practices. There may, however, be several cultural barriers to learning from these major incidents. These barriers are identified as time, litigation, fear of adverse publicity, internal procedure, disclosure of confidential information, and commitment to safety (of both companies and individuals). Such barriers are specific to the developing world [33]. This study increases the awareness of domino accident losses across the industry and provides a resource from which lessons can be learnt. Table. 1 summarizes the major catastrophic domino accidents worldwide with the largest losses. In this study, the large property damage losses have been grouped by type of facility into five categories: refineries, petrochemical plants, gas processing plants, storage terminals/distribution, and upstream.

Table 1. Major chemical accident events in the last 25 years *

Date	**Plant Type**	**Event type**	**Incident and Incident Location**	**Property Loss (US $ million)**	**Injuries / Fatalities**
7 Jul 1988	Upstream	Fire/ Explosion	Piper Alpha Incident in North Sea, UK	1600	165/--
19 Mar 1989	Upstream	Fire/Explosion	Gulf of Mexico, United States	750	7/--
23 Oct 1989	Petrochemi cals	Explosion/Fire	Phillips Petroleum Chemical Plant in Pasadena, USA	Over 1000	100/25
29 Sep 1998	Natural gas plant	Explosion/ Fire	Esso Longford gas explosion in Longford, Victoria, Australia	13000	--/2
25 Dec 1997	Gas Processing Refinery	Fire/ Explosion	Sarawak, Malaysia	430	--/--
13 May 2000		Explosion/ Fire	Enschede The Netherlands	-	1000/22

21 Sep 2001	Petrochem	Explosion/ Fire	Toulouse, France	610	3000/ 30
19 Jan 2004	Gas Processing	Fire /Explosion	Skikda, Algeria	580	74/ 27
23 Mar 2005	Refinery	Fire/Explosion	Texas, United States	1500	170/15
11 Dec 2005	Petroleum	Fire /Explosion	Hertfordshire, England	1443	43/0
23 Oct 2009	Refinery	Fire /Explosion	Bayamon, Puerto Rico	6.4	--/--
29 Oct 2009	Petroleum	Explosion/Fire	Jaipur, India	32	150/11
02 Apr 2010	Refinery	Fire/Explosion	Washington, United States	-	4/--
21 Apr 2010	Upstream	Fire/Explosion	Gulf of Mexico, USA	590	--/--
6 Jan 2011	Refinery	Fire/Explosion	Fort McKay, Albert, Canada	600	--/--
25 Aug 2012	Refinery	Explosion/Fire	Amuay, Venezuela	1000	150/48
18 Sep 2012	Refinery	Explosion/Fire	Tamaulipas, Maxico	-	46/26
18 Apr 2013	Fertilizer Plant	Explosion/Fire	Texas, USA	-	100/15
23 Aug 2013	Refinery	Explosion/Fire	Visakhapatnam, India	-	14/37
22 Nov 2013	Pipeline	Explosion/Fire	Qingdao, China	-	62/-

*Sources:[1,23,34-35]

3. Statistical analysis of domino accidents

3.1 Distribution of accidents over the last five decades

As shown in Table 2, there has been a significant increase in the number of accidents over the years from the 1960s up to 2010 with the exception of the 1991-2000 decade. The implementation of Clean Air Act in 1991 and Process Safety Management programs could have a reason for less incidents during the 1991-2000 decade. During the implementation of Clean Air Act, there were also significant changes in incident reporting which could showing a decreasing trend in the 1991-2000 decade. This increase in rest of the decades can be attributed to two main reasons. First, the chemical industry has undergone continuous expansion: more and larger process plants and storage areas have been created that are more prone to fire and explosion hazards. Second, access to information about accidents has improved gradually over the time. A considerable number of accidents that occurred during the 1960s and before were not recorded and the information were lost.

The number of fatalities has also been increasing every decade with the exception of the 1991-2000 decade (see Fig.2). The decade from 1981-1990 showed an exceptionally high number of fatalities due to two of the biggest accidents. The Mexico City accident in 1984 led to 650 deaths, which is higher than 60% of the total fatalities in that decade. This decade had the worst industrial accident - Bhopal gas tragedy in 1984, but it is not factored in the computations because it was not a domino incident [36]. Even during the years 2001–2010, two major accidents in Neyshabur and Zahedan (Iran) had higher than 45% of the total fatalities in that decade. Thus, one or two major accidents in a decade have typically led to the increased number of fatalities. Major accidents, which are found to be domino in nature, should be controlled at the initial stages to minimize the fatalities. Despite the number of accidents, the domino effect has received much less attention than other aspects of risk assessment. The number of fatalities has also been high in the 2001-2010 decade. Since 2011, the number of accidents has decreased every year. This decreasing trend could be due to increasing automation of industries, new strict regulations and prompt action in the case of emergencies.

Table 2. Domino accidents and fatalities per decades (1961-2017)

Decades	**Number of Accidents (Total = 326)**	**Percentage**	**Number of fatalities (Total = 3505)**	**Percentage**	**Per accident death rate**
1961-1970	36	11%	131	4%	4
1971-1980	41	13%	357	10%	9
1981-1990	60	18%	1058	30%	18
1991-2000	49	15%	572	16%	12
2001-2010	101	31%	943	27%	9
2011-2017	39	12%	444	13%	11

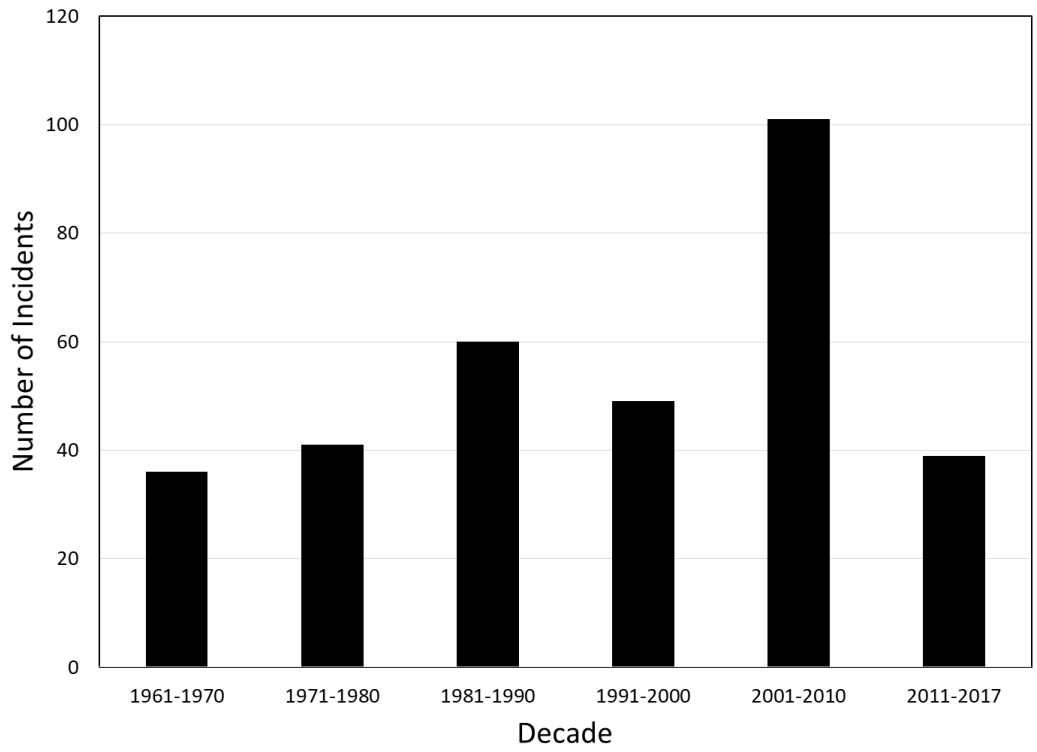

Fig. 2. Number of domino accidents in the chemical industry in each decade

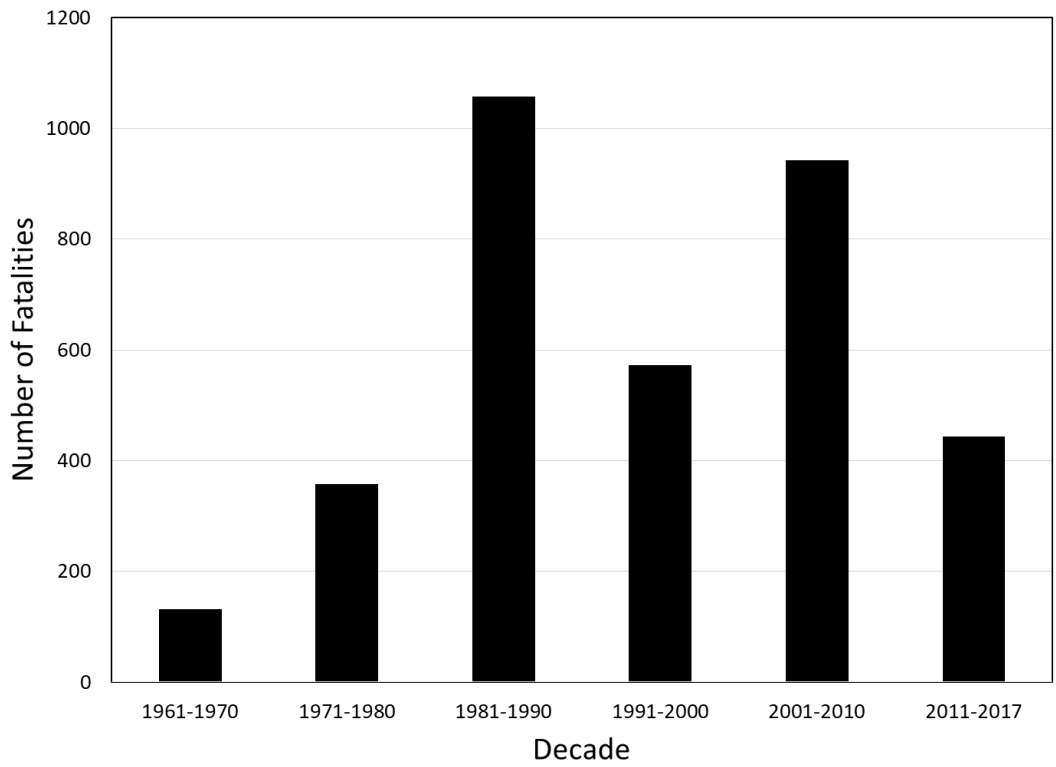

Fig. 3. Number of fatalities in major domino accidents in each decade

3.2 Location-specific accidents

Accidents were divided into the following categories according to the country in which they occurred:

- the European Union (~ 16%, France, Italy and Germany have a large number of domino accidents than the rest of EU Nations);
- other developed countries (~ 58%, United States, Canada, Australia);
- Developing countries (~ 26%, Asia and North African countries).

A certain degree of bias may exist because preference was given to information on accidents that occurred in Europe and the United States. This is because most of the institutions that manage the databases used in the study are based in these countries and the information on them is generally more exhaustive.

It has been observed that about 74% of accidents involving domino effects are recorded from developed countries as illustrated in Fig 4. The large scale of process plants and associated storage and transportation facilities in developed countries could be the major contributor to the high percentage, while some loss of data in the rest of the world must be considered. Data on the number of accidents are mostly obtained from the organizations of developed countries and could be the reason for high percentages. The possible loss of data in developing countries could contribute to the lower percentage. Data from developing countries may also be incomplete. However, the available data seems to be enough to show the overall trend and a representative sample is shown in Fig.4.

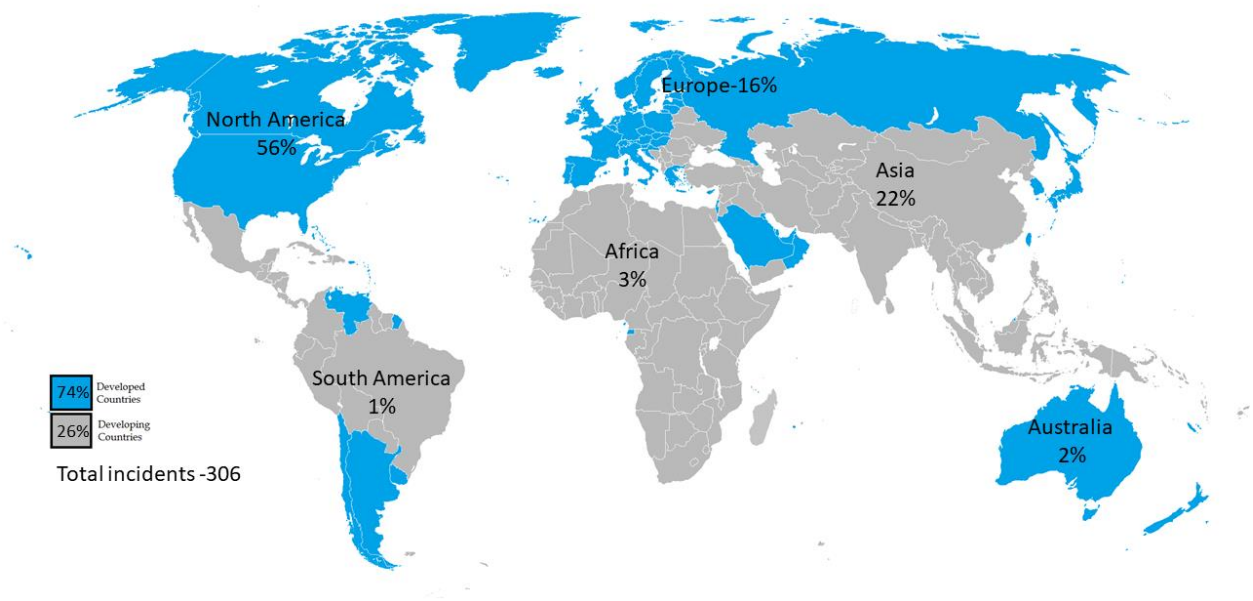

Fig.4. Domino accidents in various parts of the world

3.3 Substance Type

Most of the domino accidents involved more than one substance. Although the number of substances involved in accidents is higher, only the substance involved in the primary accident is categorically mentioned. In domino accidents, the substance in the primary event may involve other substances in secondary or further events leading to the involvement of a large number of substances in some of the worst accidents. A relatively small number of accidents involved only one substance. Flammable substances were involved in most of the accidents (89%) and were the substances most frequently found in domino accidents [3].

In most of the domino accidents, flammable substances were involved. The analysis of 166 domino accidents as shown in Table 3 illustrates that Crude oil is by far most frequently involved (43 cases, 26%) followed by natural gas (19%), propane (11%), LPG (10%), gasoline was found in 10% cases and diesel oil involved 4%. LNG and vinyl chloride incidents contribute to about 4% of the incidents. Ethylene, chlorine, hydrogen and methanol were involved in the same number of accidents (2 % each one).

Table 3. Materials involved in major domino accidents

Substance	No of accidents (Total = 166)	Percentage
Crude Oil	43	26%
Natural Gas	31	19%
Propane	19	11%
LPG	17	10%
Gasoline	16	10%
Vinyl chloride	7	4%
Diesel oil	6	4%
LNG	6	4%
Ethylene oxide	5	3%
Ethylene	4	2%
Chlorine	4	2%
Hydrogen	4	2%
Methanol	4	2%

3.4 Industries Type

All chemical accidents have been categorized according to the types of industries, such as process plants, storage terminals and transportation. It has been observed that most of the accidents (33%) are from process plants and transportation followed by 20% in storage terminals. The percentage-based distribution is shown in Fig. 5. Process plants could have the highest percentage of accidents due to the elevated operating conditions of reactants and the complex nature of reactions involved. The probability of domino accidents in process plants could also be high due to the congestion of

equipment and their connectivity. The pipeline industry contributes to the major incidents in the transportation sector. The pipeline industry has had losses amounting to about 7 billion dollars in the USA [6]. Storage terminals are at risk due to large amount of chemicals retained at one place. Recent incident like Tianjin, China (2015) explosion and west, Texas, United States (2013) explosion highlight the impact that storage terminals can pose.

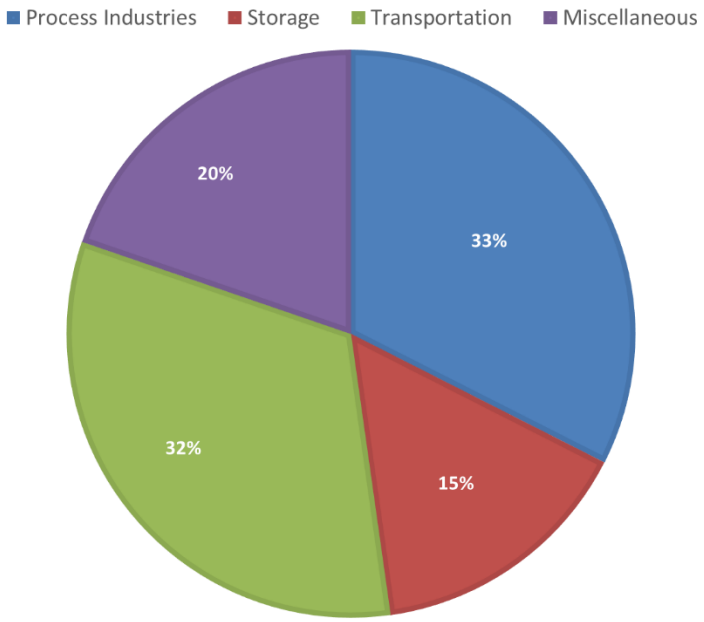

Fig. 5. Percentage of major domino accidents in various types of industries

3.5 Causes

The cause of the primary accident is a significant aspect of the analysis of domino effect accidents. The Major Hazard Incident Data Service reported several generic causes: external events, human error, impact failure, mechanical failure, instrument failure, violent reaction (runaway reaction), upset process conditions and services failure [37]. Although some of the generic causes for accidents are self-explanatory (for example, violent reaction), the accidents due to human error have greater complexity because other causes, such as violent reaction or mechanical failure could also be a result of human error.

Of all the external causes, accidents (fire and explosion) in other plants were the most frequent types. When the primary event was an explosion, it was typically impossible to ascertain from the information available whether it was the blast wave or the fragment projection that caused the secondary accident. When human error was the generic cause of the accident, general operations, general maintenance, overfilling and procedural failures were the main specific causes. The specific causes are shown in Fig. 6.

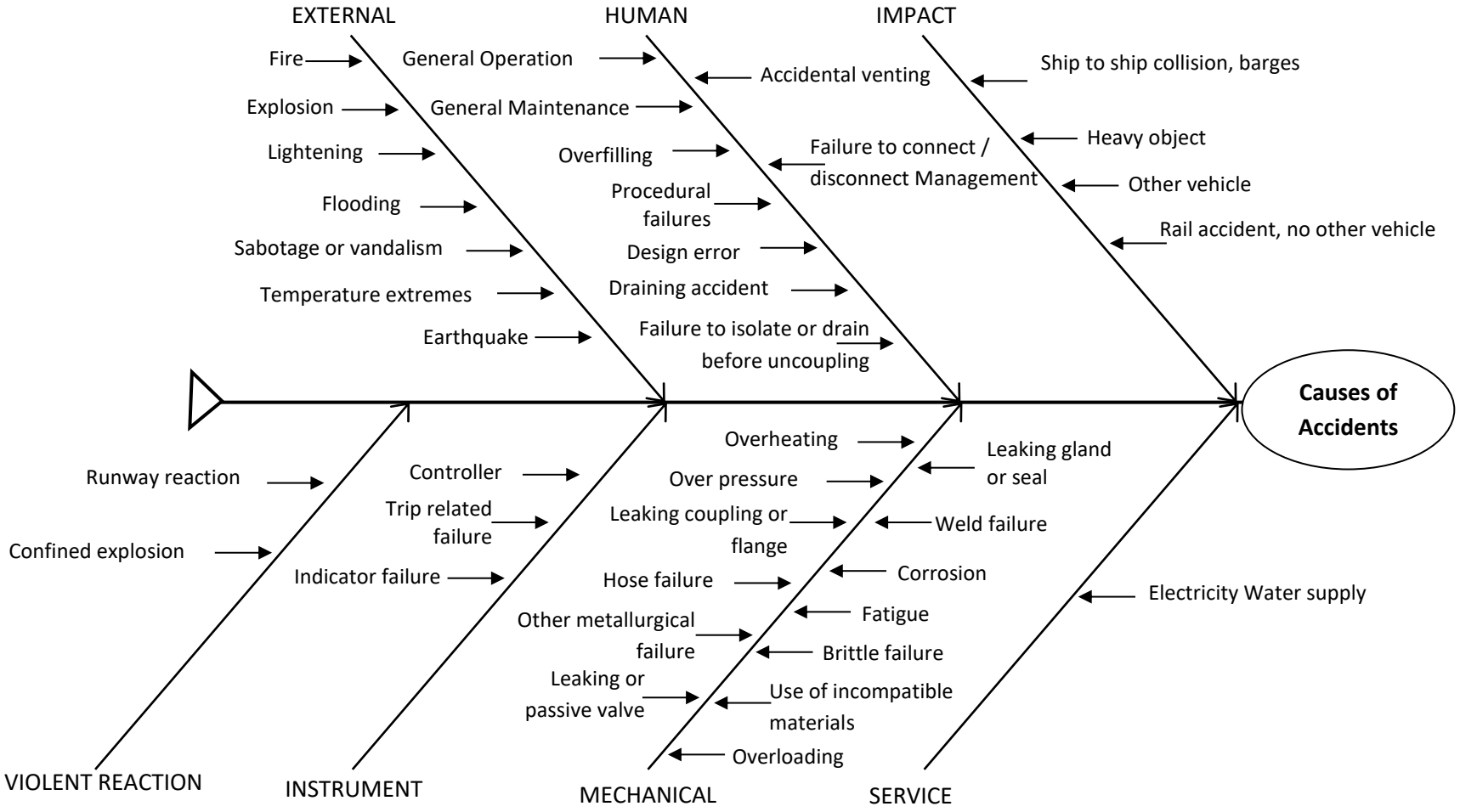

Fig. 6. Fish bone diagram of domino accident causes

4. The domino effect methodology

The methodology proposed here for evaluating the potential for a domino effect involves a three-stage procedure as illustrated in Fig. 7. This staged approach has an increasing degree of complexity. In any hazard analysis, it is initially prudent to evaluate whether it is possible to demonstrate acceptability on the basis of the "consequences" being tolerable or non-hazardous (i.e. acceptable) followed by a second stage that considers whether the "probability or frequency" is tolerable. The third stage involving risk assessment is only necessary if it is not possible to show that the site separation was acceptable from a consequence and a frequency viewpoint. If it is still not possible to demonstrate risk tolerability at the third stage, then it will be necessary to investigate and include risk mitigation measures. This approach is reflected in the proposed methodology for domino assessment.

Stage 1 includes an assessment of the maximum hazard ranges for sites A and B and an evaluation of whether these hazard zones extended to susceptible critical plants on their site,

Stage 2 includes an assessment of whether the frequencies of all incidents affecting critical plants exceed some notional threshold value, and

Stage 3 includes a combined Quantified Risk Assessment for both sites.

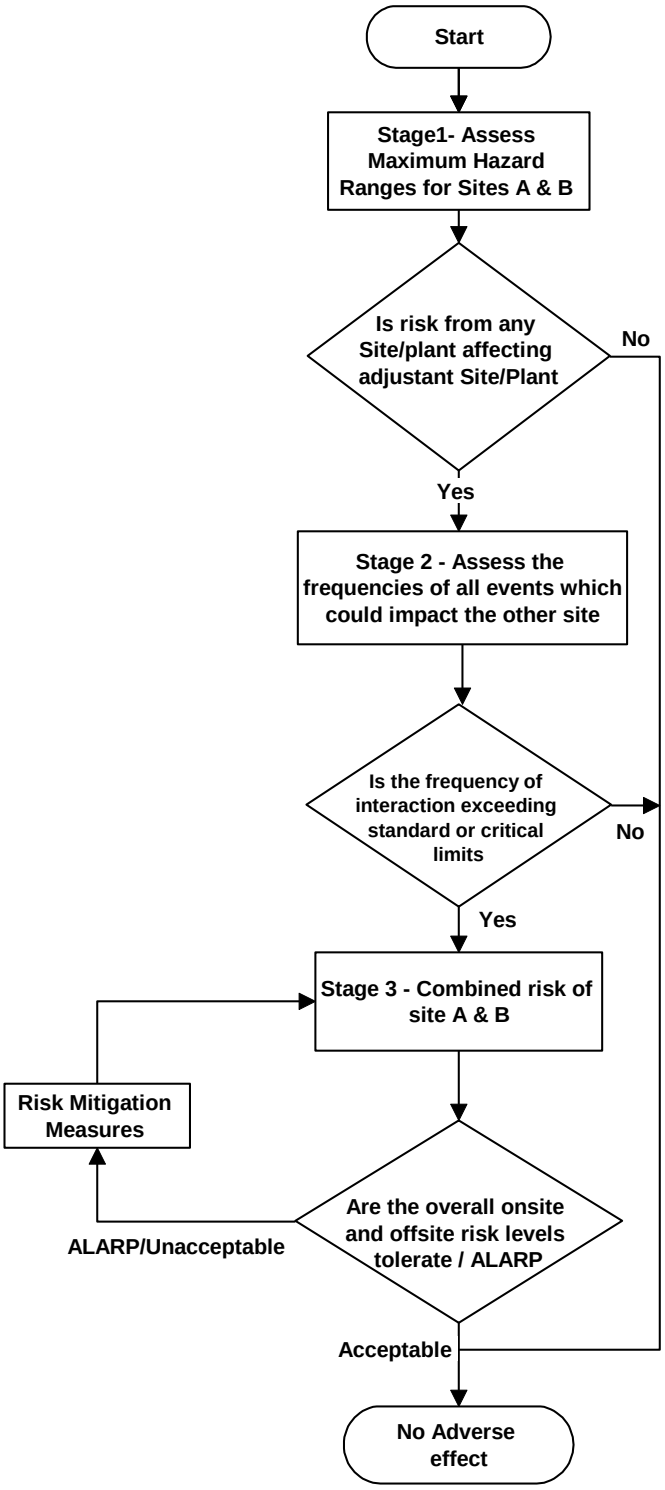

Fig.7 Overall domino assessment methodology

4.1 Input data for analysis

Accurate data are necessary to evaluate the correct effects of any accident. If additional and accurate data are available, the prediction of impacts will have higher reliability. The information and data required for the domino assessment methodology are listed below.

- A layout of the site
- Locations of units in the layout which may generate primary events of concern
- Operating conditions of all processes
- In-plant and surrounding population data
- Probable ignition sources in the plant
- Weather data
- Consequence analysis
- Analysis of the primary events with respect to failure frequencies
- Effects of primary events at adjacent facilities due to overpressures, heat radiations and toxic releases
- Assessment of secondary events
- Analysis of factors which led to a domino effect

4.2 Identification of escalation vectors

After the identification of primary accidental events, the escalation vectors associated with each scenario should be defined (step 2 in Fig. 1). The propagation of the primary event due to the escalation vectors and its effects typically generate at least one secondary target. Thus, the physical effect due to the primary event that caused damage to the exposed individuals is often different from that responsible for escalation. Therefore, it is crucial to understand that each accidental scenario should be associated with a "vulnerability vector" (used to estimate the damage to the exposed individuals) and to one or more 'escalation vectors'.

Any scenario can generate the three following escalation vectors. Escalation vectors and criteria for the primary and probable secondary scenarios shown in Table 4 are [30]:

- heat radiation and/or fire impingement,
- overpressure, and
- toxic release.

Table 4. Escalation vectors and criteria for the primary and probable secondary scenarios

Primary scenario	Escalation vector	Escalation criterion	Expected secondary scenarios
Mechanical explosion	Fragments, over pressure	16 kPa	Pool fire, Jet fire, BLEVE, toxic release
Confined explosion	Overpressure	Fragment impact	All
BLEVE	Fragments, over pressure	Fragment impact	All
VCE	Overpressure, fire impingement	16 kPa	All
Pool fire	Radiation, fire impingement	15 kW/m^2	All
Jet fire	Radiation, fire impingement	15 kW/m^2	All
Flash fire	Fire impingement	LFL	Tank fire
Fireball	Radiation, fire impingement	Engulfment	Tank fire
Toxic release	-	-	-

Therefore, the selection of credible escalation scenarios based on reliable models for equipment damage is a central issue to allow the assessment and the control of risk due to domino accidents.

5. Conclusion

Domino effect accidents have been noticed in the tragic history of many past accidents and a more realistic way of addressing intrinsic risks of chemical and petrochemical plants. Assessment of domino accidents is a difficult task due to the involvement of more than one flammable substance and secondary or tertiary events. Due to the complex nature of such accidents, very few studies on its analysis have been published. The present review and analysis of published literature has led to records of 326 major domino events spread over five decades. The domino accidents have been summarized indicating the number of fatalities, locations, involvement of substances and domino sequences.

The number of accidents over five decades has been increasing due to the expansion of chemical industries and improvements in accessing information on accidents. In addition, we observed that one or two major accidents in a decade led to a higher number of fatalities in last five decades. More than 75% of domino accidents occurred in developed countries, which seems rational due to a large number of industries located there. There is also the possibility of loss of data regarding accidents in developing countries, thereby leading to a lower percentage. However, the review and analysis has also shown that domino accidents in underdeveloped countries have higher severity compared with countries that are technologically more advanced.

The analysed accidents have indicated that fires and explosions are the primary domino effect events. Thus, precautionary measures should be adopted while handling flammable materials, which are the most common substances in domino accidents. The most frequent sequences are explosion→fire, fire→ explosion and fire→fire. The results show that the quantitative assessment of escalation hazard is a key tool to understand the credible and critical domino scenarios in

complex industrial sites. Therefore, significant efforts should be devoted to improving safety in such operations, especially in storage facilities where most transfer operations are performed.

Past accident analysis enables an understanding of how accidents occur and provides useful inputs for the development of loss prevention strategies, and therefore, it is an important component of loss prevention Research and Development (R&D) activities. Consequence analysis in the case of domino accidents is a complex task as no clear guidelines for identifying it are available. The escalation criteria described in this study may represent an initiating point in quantifying risk of domino effects.

The domino effect is an important aspect in risk analysis, as knowledge of the main hazards and features of this phenomenon can be used to identify additional safety measures. However, risk assessment techniques have intrinsic limitations due to the complexities introduced by event interactions and multi component or multiphase systems encountered in real situations. Thus, it is imperative to study the risk assessment of major past accidents and thereby to take appropriate measures to prevent major accidents in the future.

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Applying PHA Methodologies such as HAZOP and Bowtie to Assessing Industrial Cybersecurity Risk

John Cusimano*, Jacob Morella, Tim Gale
aeSolutions, Industrial Cybersecurity Division
Greenville, SC
john.cusimano@aesolns.com

Abstract

Process hazard assessments (PHA) are a well-established practice in process safety management. These assessments focus on failures (aka deviations) that are typically caused by equipment failures or human error. By design, PHAs do not consider cyber threats to industrial control systems (ICS). However, cyber threats represent additional failure modes that may lead to the same health, safety and environmental consequences identified in the PHA. Functional safety (i.e. ISA 84 / IEC 61511) and industrial cybersecurity standards (i.e. ISA/IEC 62443) recognize this issue and provide guidance on how to integrate these two disciplines to ensure that cyber incidents cannot impact process safety.

A proven methodology, called Cyber PHA, based on ISA/IEC 62443-3-2 has been developed and applied to conduct ICS cyber risk assessments throughout the process industries. This paper will describe the methodology with examples of actual applications to identify, rank and mitigate cyber risk in ICS systems. Furthermore, we will demonstrate how Bowtie Analysis can be used to visualize the results and apply degradation factors and controls related to cyber barrier assurance.

Keywords: industrial cybersecurity, ics cybersecurity, cyber pha, cyber bowtie, isa/iec 62443, cyber-risk, cyber-security

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Large hydrocarbon tank fires: Modelling of the geometric and radiative characteristics

Ravi Kumar Sharma*
Centre for Excellence in Disaster Mitigation and Management,
Indian Institute of Technology Roorkee, 247667, India
*Presenter E-mail: ravikrsharma16@gmail.com

Abstract

Storage tank farms are essential industrial facilities to accumulate oil, petrochemicals and gaseous products. Since tank farms contain huge mass of fuel and hazardous materials, they are always targets of serious accidents such as fire, explosion, spill and toxic release which may cause severe impacts on human health, environmental and properties.

This article presents an IOCL Jaipur, 2009 incident investigation of the gasoline tank fires flame characteristics. Although having a safe layout cannot prevent initiating accidents, it effectively controls and reduces such accidents' adverse impact. Therefore, appropriate modeling studies are needed to simulate the potential threat of large-scale pool/ tank fires in fuel storage depots for quantitative consequence analysis and effective preventive measures. The estimated flame height lies between 0.9 and 1.5, which is within the observed range. In contrast, the estimated surface emissive power lies between 27 and 123 kW/m2 respect to time, as determined by adopting various models for large-scale tank fires. The results show turbulent flame with constant loss burning rate per area, different flame height, and different heat release rate. The irradiances (E_r) are assessed with a point source model and are validated by the DNV Norway-based risk assessment PHAST software with a maximum percentage error of 25%.

This paper highlights the scale of threat posed to people, assets and the environment by hydrocarbon storage tank fires, and discusses tank fire hazards and key factors that influence the prevention and suppression of oil storage tank fires.

Keywords: Hydrocarbon pool fire; Flame height; Thermal radiation; Surface emissive power; Fire safety

1. Introduction

With the rapid development of the petrochemical industry and the completion of petroleum oil's strategic reserve, the safety of large oil depots and large storage tanks has become more and more critical. However, in recent years, fire and explosion accidents occur frequently in large-scale storage tanks, and these accidents often cause severe losses. In the last decade, there have been three major large-scale storage tank fire accidents that have exhibited striking similarities. The Buncefield oil storage depot accident in the UK on 11 December 2005, the Caribbean Petroleum Corporation fuel depot accident in Puerto Rico, the USA on 23 October 2009, and the Indian Oil Corporation Ltd (IOCL) accident in Jaipur, India on 29 November 2009 [1]. The Amuay refinery accident occurred in Venezuela on 25 August 2012 [2]. A wide range of similarities has been observed among these accidents, most notably the vapour cloud explosion (VCE) followed by the multiple tank fires. These accidents demonstrate the large-scale destruction of the surroundings and serious environmental implications and underline the necessity of appropriate measures to prevent such devastating accidents [3]. Hence, learning from past accidents is essential for the future safe operations of storage tanks.

Many of the past accident reports state that tank fires are the common disaster forms in petroleum industries and result in more intense radiation, and higher flame, which can cause serious impact on the surrounding personnel and equipment and can also lead the boiling liquid to a vapor explosion or vapor cloud explosion [4].

The different types of fire that may occur at a petroleum industries jet fire, flash fire, fireball, and pool fire - great importance is attached to pool fire as it is the most frequent, constituting more than 60% of all the fire incidents [5]. Pool fire often triggers explosions as also newer pool fires result from explosions. Pool fires can be vast and persistent, challenging to douse [6]. A buoyancy-driven, turbulent non-premixed flame is formed above the pool. The resulting fire is distinguished from other fires by a very low initial momentum and the propensity to be strongly influenced by buoyancy effects. Therefore, the research about hazard analysis and damaged area of the pool fire in tank fire is of great significance for its prevention.

Compared to other types of fires, full-surface fires formed after a fire in a petroleum oil storage tank have the characteristics of high burning rate, high flame temperature, intense radiant heat, and high challenge in the safety assessment of the tank after the fire [7] . Moreover, the wind's direction and magnitude often play a decisive role in the progress and spread of the fire. Once the fire becomes challenging to control, firefighters will encounter many difficulties in the process of firefighting and rescue [8].

The hazard assessment of pool fire incidents comprises an estimation of the mass burning rate, flame geometry, flame temperature, and, more prominently, evaluate emitted radiation by the flame [9]. The thermal radiation evaluation plays a significant role in assessing the resistance of equipment in the proximity of fire and verifying the possibility of domino effects [10]. To avoid too conservative results, imposing anti-economic geometric constraints, such as spacing or barrier and other potential active and passive preventive measures, a realistic incident assessment is significant.

Additionally, fire risk evaluation involves the application of developed risk criteria to assess the risk level. Fire risk treatment improves existing risk controlling measures, develops new risk control measures, and implements these measures to reduce fire risk [11]. Therefore, fire risk analysis is the only part of the fire risk management process. It serves as the foundation of regulatory decision-making on implementing risk-reducing actions or choosing appropriate risk treatment measures or not [12-13]. Research related to fire risk analysis, therefore, critical and essential.

2. Aims and objectives

The main objective behind tank fire of pool fire research is the necessity of understanding large-scale tank fires' behavior. In most situations, tank fires could lead to a domino effect [14]. However, most studies have focused on small-scale pool fires, which differ significantly from the large-scale pool and tank fires [15]. To avoid too conservative results, imposing anti-economic geometric constraints, such as spacing, a realistic scenario evaluation is needed. Predicting large flames' relative characteristics is still subject to considerable uncertainty because several parameters associated with sizeable turbulent diffusion flames cannot be determined accurately [16]. This paper aims to evaluate flame characteristics, flame surface emissive power, and irradiance at various distances by using various models to recommend proper preventive measures.

One of the motivations of this study was to expand the fundamental understanding of fuel fire dynamics to improve the ability to predict hazards from a fire in the given accident scenario, establish the utility of forensic tools, and validate empirically-based correlations used to model fire scenarios. In order to analyse the given scenario, one must first determine, by modeling or by direct observation, how massive the fire is and how intensely it is burning, requiring knowledge of such characteristics as flame geometry, flame temperature, and heat release rate. Once the fire's size and intensity are established, heat transfer models can be used to predict hazard levels to the fire surroundings [17-21]. The ability to simulate real-life accident scenarios has been limited. However, because tests with small-scale fires do not fully simulate the physics of large-scale fires, and outdoor tests with more massive fires are subject to poorly controlled ambient conditions [19].

In most cases, photographic or video images were used to characterize the fire's shape, yet visual methods have limited effectiveness in large and sooty fires. Differences in the methods used to define and measure flame geometry have also contributed to significant scatter in the results. Through characterization of the temperature field in the fire plume, the overall geometry of the fire could be described in greater detail than by looking at video images alone.

3. Description of site and the incident initiation

3.1. Tank Storage Facilities during the Fires

The affected plant, shown in Fig. 1, covered 485,625 m2 and contained 11 tanks with a total capacity of about 110,370 m3 to store gasoline, diesel, and kerosene. Besides, there were five underground tanks, each of 70 m3 capacity, to store gasoline and anhydrous alcohol [22]. Seven buildings, a lubricating oil warehouse, and a truck loading facility also came within the affected

area's ambit. The 3 m height compound wall enclosed the entire region, confined by trees and buildings.

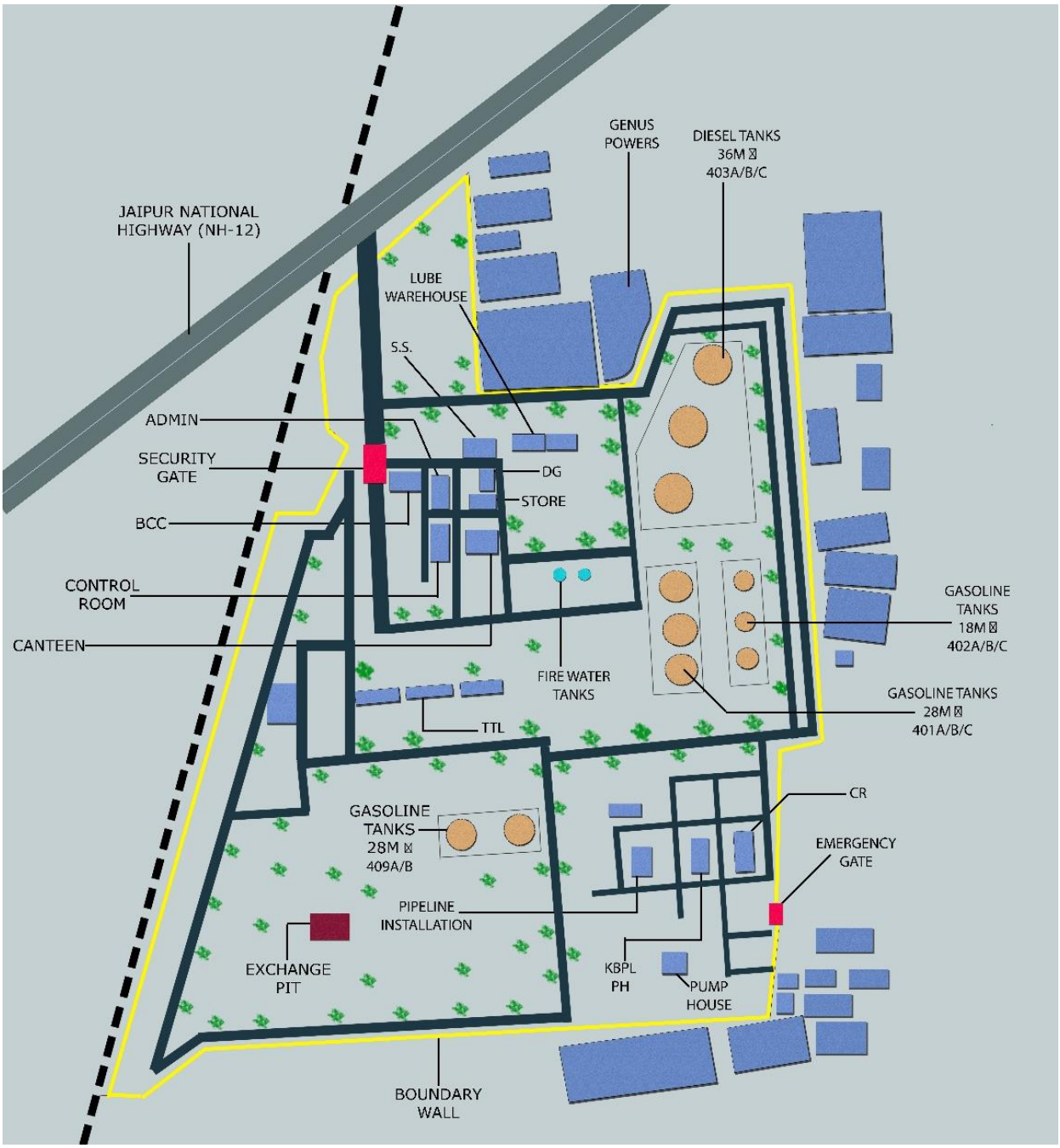


Fig. 1. Pre-accident view of IOCL Jaipur depot tank storage site and the immediate neighbourhood

3.2. Jaipur IOCL Incident

On, October 29, 2009, at 19:30 hrs (IST), fires that engulfed 11 large fuel storage tanks - a significantly high proportion of the IOCL Jaipur oil storage and transfer depot at Sitapura industrial area, India, followed a series of explosions in rapid succession. The powerful explosion occurred that followed by a great fireball with a temperature of 1727°C, which covered the entire installation [23]. The consequent fires involved a number of fuel storage tanks on the site. The first explosion was followed by further explosions leading to multiple tank fire, which involved 9 tanks containing different materials immediate after the explosion. This was followed by additional 2 tank explosions due to effect of highly intense heat radiation caused by fire in 9 tanks [22]. Thus, 11 large storage tanks of various sizes exploded and caught fire that resulted in the complete destruction of the facilities and buildings within the premises of the terminal as shown in Fig 2.The expansion of the fire to the neighbouring 2 tanks happened without explosions due to the high thermal radiation. For large, black, smoky hydrocarbon fires, the estimation of the critical thermal separation distance is not only dependent on the total fire but also on the height of a hot and clear burning zone. Additionally, for multiple tank fires, as occurred in Jaipur, there is a considerable increase in the mass burning rate, the flame height, the surface emissive power, as well as the thermal separation distance [24].

Subsequently, the depot was completely destroyed and widespread damage was caused to the neighbouring properties. This devastating accident resulted in 11 fatalities and injuries to more than 150 people. About 5000 people had to be evacuated from their homes in the adjacent area. The fire burned for 11 days, destroying most of the site and emitting a large plume of smoke. The cost of the incident in terms of damage to property and loss of business is estimated to be approximately USD 60 million. In the aftermath of the incident, the Ministry of Petroleum and Natural Gas Committee (MoPNG) was formed to oversee the investigation.

3.2.1. Response Assessment

The management of IOCL took the decision to allow the petroleum products to burn out in order to avoid further aggravation of the accident in the interest of public safety [22]. IOCL personnel and local firefighters were trained only for a worst-case scenario involving one tank on fire, rather than 11 tank fires at the same time caused by a vapor cloud explosion. Without sufficient equipment or training, local responders attempted to fight the multiple tank fires but failed as the fire encompassed more tanks.

a) *Insufficient equipment*: Tank terminals like IOCL, Jaipur were not required to conduct a risk analysis where they consider the potential of a vapor cloud explosion and multiple tank fires. Neither IOCL, Jaipur deport nor the fire department had the requisite amount of foam and adequate equipment to effectively fight and control a fire involving multiple tanks.

b) *Insufficient preplanning with local fire departments or firefighter training at the site level*: IOCL, Jaipur did not preplan with local emergency responders, set up mutual aid with other hazardous materials sites, or adequately train facility personnel to address a tank farm fire involving multiple tanks. In fact, training for IOCL, Jaipur terminal personnel was limited only to fighting a fire involving one tank, not an incident involving multiple tanks.

c) *Limited emergency preparedness*: Local fire departments did not have sufficient training or resources to respond to industrial fires and explosions, which resulted in firefighting delays from insufficient foam and equipment. The limited training and resources of the local fire departments resulted in an inefficient firefighting operation.

(a)

(b)

Fig. 2. IOCL Jaipur Depot on the (a) first night and (b) second day of the fire that followed the explosion. Multiple tank fires are clearly visible.

3.3. Selection of Typical Damage Images

To understand and illustrate the postulated scenarios, a selection of typical damage images from the IOCL Jaipur incident is shown in Figs. 3-5.

(a) (b)

Fig. 3. Damaged tanks after the incident (a, b)

(a) (b)

Fig. 4. Warehouse and crushed drums on the loading bay

Fig. 5. Damaged depot pumping station

4. Analysis of tank oil fire

4.1. Mechanism of tank oil combustion

Wang & Joulain [25-26] reported that the oil tank's burning area could usually be divided into three layers, i.e., fire layer, mixture layer of air, and flammable vapor and oil liquid layer. In oil combustion, negative pressure occurs in certain areas because of oxygen consumption. It inhales surrounding clod air into this area, which mixes with flammable vapor and forms a mixture layer. The mixed layer's existence makes the flame wave, and its combustion rate gradually tends to a constant value. The mixed layer is influenced by the diffusion velocity of surface vapor; in other words, that the diffusion velocity of surface vapor determines the process of combustion.

In the process of combustion, part of the heat diffuses to the outside through the thermal radiation and heat convection when others feedback to the surface of oil through heat conduction of tank wall, heat convection of hot smoke, and thermal radiation of flame. It keeps flammable vapor producing steadily from the surface and propels combustion. Combustion of vapor, heat feedback and evaporation are three related and circulatory links and only intervene in this loop and slow its process, and the combustion could be stopped.

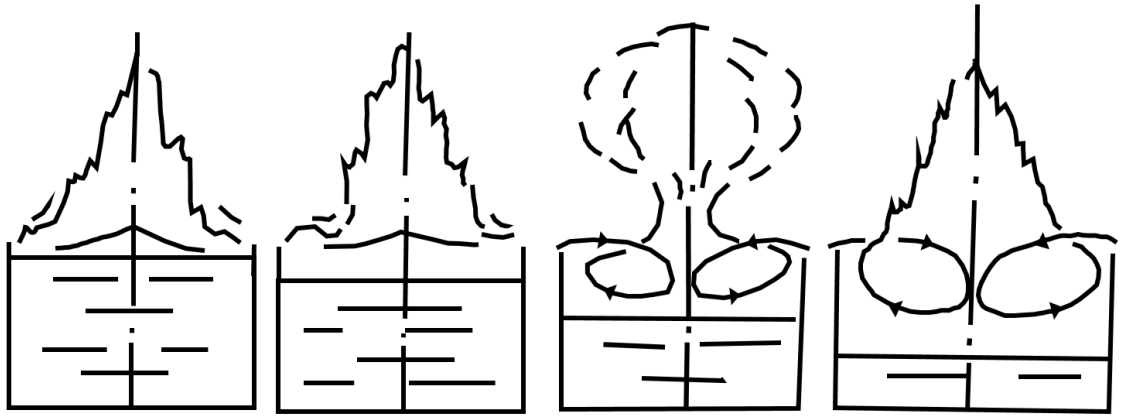
4.2. Flame Characteristics

The strong airflow occurs in oil tank fire. The flame's characteristics are closely related to the tank's diameter, the liquid's nature, and the liquid level.

A high level (the distance between the fuel surface and the tank top is D/10 to D/5) tank fire is likely to be a pool fire. Its flame is typically a turbulent buoyant diffusion flame—the base of the flame-like a cone. The negative pressure created about 0.5 bar, at the tank center near the fuel surface. The negative pressure sucks the surrounding air towards the tank's axis, but as the fuel level high, air cannot be sucked into the tank, and the air, as shown in Figs. 6a and 6b forms only an inverted circular valley [25].

When the fuel level was decreased to the range D/5 to D/3 (middle-level tank fires), the air sucked into the tank, and ullage is being formed, but at the same time, the burning gas rises, and the cold air and hot gases form an irregular interlocking pattern. There are many "fireballs" on the top of the tank. The flame has a neck and also has apparent pulsation with mushroom clouds, as shown in Fig. 6c [25].

As soon as the fuel level is lower than D/2, cold air enters into the tank but has difficulty penetrating the flame. As the concentration of oxygen declines in the ullage, combustion, and the tank's negative pressure develops progressively weak—a large quantity of the fuel-rich combustion products and smoke escape from the tank along the wall. The flame is shorter, and the neck disappears. There are some non-burning fuel-rich "black holes" near the perimeter of the tank, as shown in Fig. 6d [25].



(a) (b) (c) (d)

Fig. 6. Typical tank fire flames: a. high level, b. some fuel bum out, c. middle level, d. low level [25]

5. Combustion calculation model of the pool fire in fire tank

The study of pool fire in the fire tank mainly includes the geometry formation of the liquid pool, the shape, height, temperature of the flame, heat radiation, and harm degree of flame. In accordance with the liquid pool's geometry changes over time, the liquid pool's geometry of pool fire in the fire barrier can be divided into two types of a constant geometry of the pool and geometry change of the pool over time. Among the combustion parameters which determine the overall structure of a pool fire, the most important is flame height and radiation intensity.

5.1. Mass burning rate

The burning rate is a fundamental parameter for describing the pool fire; it is affected by the heat radiation, the flame location and shape, the fuel container's thermal conductivity, and other factors [27]. Usually, the pool fire development process can be divided into three stages. The first stage is the combustion, with the burning rate accelerated in the course of flame temperature rising. After a while second stage (steady combustion), the heat obtained by the fuel from the flame becomes comparable with the heat transmitted from the fuel to the ambient medium. The burning rate reaches a stable value (Fig. 7). In the third stage, the burning rate falls owing to the inadequate supply of fuel.

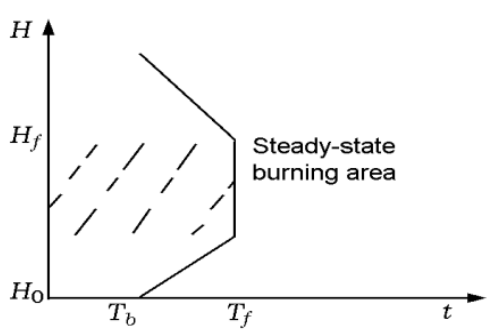


Fig. 7. Pool fire development process.

The time averaged mass burning rate $\bar{m}_f^{''}$ can be calculated by multiplying the time averaged burning velocity and the liquid density of fuel. The equation for the mass burning rate is given below:

$$\bar{m}_{\mathrm{f}}^{''} = 10^{-3}\,\Delta H_c/\Delta H_{\mathrm{v}} \tag{1a}$$

This equation is valid for a wide range of gaseous and liquid fuels [28].

For the maximum mass burning rate, the following equations are given [28].

$$\bar{m}_{\mathrm{f}}^{''}(D) = \bar{m}_{\mathrm{f,max}}^{''}\left(1 - e^{-k\beta D}\right) = \bar{m}_{\mathrm{f,max}}^{''}\varepsilon_{\mathrm{F}} \tag{1b}$$

$$\bar{m}_{\mathrm{f}}^{''}(D) = \rho_{\mathrm{f}}\bar{\mathrm{V}}_{\mathrm{f,\ max}} \approx 1.27 \times 10^{-6}\,\Delta H_c/\Delta H_{\mathrm{v}}\left(1 - e^{-k\beta D}\right)\rho_{\mathrm{f}} \tag{1c}$$

The estimated $\bar{m}_{\mathrm{f}}^{''}(D)$for large tank gasoline fire is 0.055 kg/m^{2}s. The maximum mass burning rate of gasoline in a large tank on fire reported by various authors' ranges from 0.055 to 0.083 kg/m^{2}s [16,29].

5.2. Flame Geometry

The geometry of a flame depends mostly on flame pool diameter, flame length, mass burning rate, temperature and the flame radiative properties. These properties are characteristically taken as averaged in time. The measurements derived from different assessments for the influence factors and the geometry of large flames is described in the following section.

5.2.1. Relative Flame Height

As another critical characteristic parameter, the flame height is directly related to the heat transfer process of pool fire and the flame's influence around the environment.

The flame height is a dynamic parameter, and the flame tip is often taken to be the point of 50% intermittency [28], because the visual effect of the flame is not entirely representative of its height. The best known and most widely adopted correlation for calculating the ratio between the flame height and the diameter of a circular pool is described by Bubbico *et al.*[30]and Chen and Wei [31].

The flame height is generally taken as the maximum visible height or the time-averaged visible height [32]. The time-averaged relative $(\bar{H}/D)$ and maximum relative $((\bar{H}/D)_{\mathrm{max}})$ visible flame height are dependent on the Froude number (Fr_f) and the dimensional wind velocity $(\bar{u}_W^*)$ and can be estimated by the following correlations [33].

$$\bar{H}/_D = a\,Fr_f^b\,\bar{u}_W^{*c} \tag{2a}$$

and

$$\left({}^{H}/_{D}\right)_{max} = a\, Fr_f^b \bar{u}_W^{*c} \tag{2b}$$

There are more correlations with many empirical parameters, including a, b, and c, which are experimental parameters and are given in Table1 [28].

Table. 1. Parameters for the determination of the dimensionless visible flame heights used in Eqs. (2a, b) [34]

Correlation	a	b	C	Comment
Munoz 1	8.44	0.298	-0.126	Measured on gasoline and diesel pool fires: $(H/D)_{max}$
Munoz 2	7.74	0.375	-0.096	Measured on gasoline and diesel pool fires: $(\bar{H}/D)$

The height of the visible flame is a function of the pool diameter and the burning velocity [35]. For the IOCL Jaipur incident, an assessment of the maximum, visible and relative flame heights of gasoline tank fires was conducted assuming that the 'c' parameter in Eq. (2b) was zero because there was no wind effect. The modified equation can therefore be written as:

$$(H/D)_{max} \approx a\mathrm{Fr}_f^b = \mathrm{a}\left(\frac{\overline{m''}_f}{\rho_a\sqrt{gD}}\right)^b \tag{3a}$$

Thus, the estimated $(H/D)_{\max}$ ratio for the gasoline tank (D = *24m*) fire is 1.5. For a large hydrocarbon pool fire where $D \geq 9\ m$, the time-averaged relative flame height ($\bar{H}/D$) is calculated using Eq. (2a) [33] and Table. 1 and approximates to

$$(\bar{H}/D)_{\mathrm{calc}} \approx a\mathrm{Fr}_f^b = 7.74\left(\frac{\overline{m''}_f}{\rho_a\sqrt{g\,D}}\right)^{0.375} = 0.9 \tag{3b}$$

With $\overline{\mathrm{m}''}_{\mathrm{f,max}}$ (D = 24 m) ≈ 0.055 kg/ (m^2s) for a gasoline pool fire, ρ_a = 1.29 kg/m^3, and the parameters a and b from Table 1, the calculation using Eqs. (3a, b) results in

$$0.9 \leq (H/D)_{max,calc} \leq 1.5 \tag{3c}$$

An empirical relationship was observed between the maximum and average flame height. Thus, a single correlation could be used to estimate both dimensions [16]:

$$(H/D)_{\max} \approx 1.6\,\bar{H}/D \tag{4}$$

The empirical relationship in Eq. (4) was also considered valid for the IOCL Jaipur tank fires.

5.2.2. Height of the Clear Burning Zone by MSFM

In the Modified Solid Flame Model (*MSFM*), $\overline{SEP}^{ma}_{\mathrm{MSFM}}(D,\eta) \equiv \overline{SEP}^{ma}_{\mathrm{cl}}, \overline{H} \equiv \overline{H}_{cl}$ and $\bar{\eta}_{rad} \equiv \overline{\eta}_{rad,\mathrm{cl}}$. Thus, the relative height of the hot clear burning zone (yellow luminous), $\overline{H}_{\mathrm{cl}}/D$, which is not covered with a black smoky layer, can be calculated by Eq. 5a [33]:

$$\overline{H}_{cl}\,(D)/D = \frac{\overline{\eta}_{rad,\mathrm{cl}}(D)\overline{m''}_{\mathrm{f,max}}(-\triangle H_c)}{4\overline{SEP}^{ma}_{cl}} \tag{5a}$$

Eq. (5a) is valid only for gasoline and kerosene fires. Within the extent of the MSFM exponential correlation between $\bar{\eta}^{\mathrm{exp}}_{\mathrm{rad}}$ and the pool diameter [36], the following is valid:

$$\bar{\eta}_{\mathrm{rad,cl}}(D) = \bar{\eta}^{\mathrm{exp}}_{\mathrm{rad}}(D) \;=\; 0.35\ \mathrm{e}^{-0.05\mathrm{D}}, \overline{SEP}^{\mathrm{ma}}_{\mathrm{cl}} \approx 100\ \mathrm{kW/m^2} \tag{5b,c}$$

Eq. (5a) with Eq. (5b) results in the numerical value equation:

$$\overline{H}^{\mathrm{MSFM}}_{\mathrm{cl}}/D = 2.5\ \times 10^{-3}\bar{\eta}^{\mathrm{exp}}_{\mathrm{rad}}(D)\overline{m}''_{\mathrm{f,max}}\,(-\Delta H_c) \tag{5d}$$

For a large tank gasoline fire, $\overline{m}''_{\mathrm{f,max}}\,(-\Delta H_c) \approx 3630\ \mathrm{kW/m^2}$ [16].

$$\overline{H}^{\mathrm{MSFM}}_{\mathrm{cl}}/D \;\approx 2.1\ \times \mathrm{e}^{-0.05\mathrm{D}} = 0.6 \tag{5e}$$

The $\overline{SEP}^{ma}_{cl}$ defines the relative height $\overline{H}_{cl}/D$ of the clearburning zone, as well as $\bar{\eta}_{red,cl}\,(D)$, the effect of the thermal radiation of a large tank fire on the neighbourhood, e.g., the contents of a tank (burning or not burning).

5.2.3. Height of the Clear Burning Zone by considering the (C/H) ratio

The modelling of the clear flame length has been proposed by Pritchard and Binding [37] and Ditali [38]. It was reported that the height of the clear flame varied by approximately 30% of the maximum flame length for fires up to 25 m in diameter to 0% for fire diameters of 5 m or more [39]. The hydrocarbon fuel has a major role in the production of smoke within the fire affecting the height of the clear flame. The (*C/H*) ratio is used to illustrate the saturation of a hydrocarbon fuel and the tendency to generate soot [38].

5.2.3.1. Pritchard and Binding Correlation:

Pritchard and Binding [37] used the *C/H* ratio to characterise the effect of the fuel type in the correlation of the clear flame height by Eq (6a).

$$H_{Cl}/D = 11.404\ (m^*)^{1.13}(U^*_9)^{0.179}(C/H)^{-2.49} \approx 0.3 \tag{6a}$$

where the C/Hratio for gasoline is 0.43.

$m^*$and U^* can be calculated by Eqs. 6b and 6c, respectively

$$m^* = m''/\rho_a\,(gD)^{1/2} = 6\times 10^{-3} \tag{6b}$$

and

$$U^* = U\Big/\left(\frac{g}{m''\,D/\rho_a}\right)^{1/3} \approx 0.5 \tag{6c}$$

5.2.3.2. Ditali Correlation:

Ditali et al., [38] produced a similar correlation (Eq. 7), based on a separate set of experiments, with a lower dependency on the (C/H) ratio.

$$H_{\mathrm{cl}}/D = 12.4\,\left(m''\right)^{0.61} D^{-0.6}(C/H)^{-0.15} \approx 0.4 \tag{7}$$

Comparison of the above two correlations with clear flame data shows that the Pritchard and Binding correlation provides a better prediction than the Ditali correlation [32]. Hence, the Pritchard and Binding correlation represents the best available method for predicting clear flame height.

5.3. Flame Temperature

The flame temperature is a function of time and height, as described by Planas and Casal [39]. Eq. 8 gives the correlation used for the flame temperature.

$$\mathrm{T}_f(t,h) = \frac{10^4\,.t}{(34+210\times \mathrm{H}+8.51\times \mathrm{t})} + 298 \tag{8}$$

In the IOCL Jaipur incident, the estimated flame temperature of the gasoline tank (*D = 24 m*) was approximately 957ºC (1230 K), which lies within the range of 1100 K-1240 K as reported by various researchers for large-scale gasoline pool fires [17,40-43].

5.4. Surface Emission Power (SEP)

The heat radiation flux on the flame's surface is usually associated with the fuel properties, the burning extent, the geometry, size of the flame and the flame surface location, and the flame shape and temperature. We should select different mathematical models to calculate the heat radiation flux on the flame's surface, depending on the pool diameter. Generally, we assume that the energy uniformly radiates from the cylindrical flame's top and side face to the surroundings. The surface thermal radiation flux on the flame's surface should be calculated according to the Mudan model. The following formula calculates the mass heat release rate

A key parameter for the estimation of the thermal radiation of tank or pool fires is the Surface Emissive Power (*SEP*) [34,44, 46]. It is usually defined as the heat flux due to thermal radiation at the surface of the flame in kW/m^2 [47]. The flame surface area (A_F) should be considered in the

calculations of *SEP* because it depends on the geometry of the flame. The thermal radiation, *SEP*, of a tank or pool fire can be calculated using the radiation models, such as the Solid Flame Model (*SFM*), the Modified Solid Flame Model (*MSFM*), the Two-zone Radiation Model (*TRM*) and Thermal Radiation for Single and Multiple Tank Fires Model (*TRSMFM*). These models consider the effect of heat feedback enhancement on *SEP*.

5.4.1. Solid Flame Radiation Model (SFM)

SFM is used to compute the maximum surface emissive power of a specific pool or tank fire [47-48]. In this model, the flame is considered as a cylinder with the circular base having a homogeneous temperature around the flame as shown in Fig. 8. This model can also be considered as a single-zone radiation model, with no black soot portion, having a high emissivity of $\bar{\varepsilon}_{\mathrm{F}} \approx 0.95$, just like optically thick flames. The time-averaged maximum surface emissive power $\overline{SEP}_{SFM}^{ma}$ is calculated using the Eq. (5.9a) as given by [47].

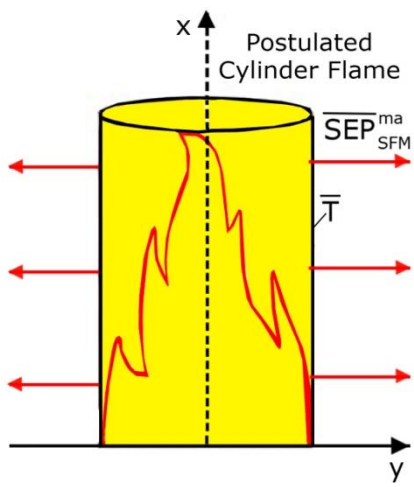


Fig. 8. Solid flame radiation model: The flame is an equally radiant cylinder [33]

$$\overline{SEP}_{\mathrm{SFM}}^{\mathrm{ma}} = \bar{\varepsilon}_{\mathrm{F}}\, \sigma\, (\mathrm{T}_f^4 - \overline{\mathrm{T}}_{\mathrm{a}}^4) \neq f(Df) \tag{9a}$$

With the calculated flame temperature of 1230K from Eq. (8), the constant, surface emissive power is estimated as 123 kW/m^2.

5.4.2. Modified Solid Flame Model (MSFM)

In this model, the flame is divided into two parts: a luminous part where the flame can be clearly seen with high emissive power and an upper part where dark smoke covers the flame with sudden bursts of luminous flames, as shown in Fig. 9. The moving border between these two parts depends on the fuel, pool diameter, and oxygen content of the burning zone [44]. Especially for large pool diameters, an alternative equation for the time-averaged maximum surface emissive power $\overline{SEP}_{MSFM}^{ma}(D, \eta)$ is proposed by [34] (Eq. 10a).

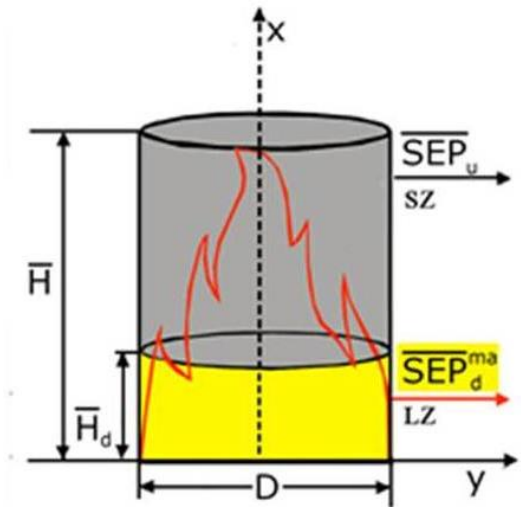


Fig. 9. MSFM: the flame is divided into a clear luminous zone with a high radiation (LZ) and a non-radiating soot zone (SZ). [33]

$$\overline{SPM}^{ma}_{\mathrm{MSFM}}\ (D,\eta) = \frac{\bar{\eta}_{\mathrm{rad}}(D,\eta)\overline{m''}_{\mathrm{f}}\ \ (-\triangle H_c)}{4\bar{H}\ (D)/D} \tag{10a}$$

Mc Grattan *et al.*, [48] found an exponential relationship between η_{rad} and the pool diameter:

$$\eta_{rad} = 0.35\ e^{-0.05D} \tag{10b}$$

MSFM is a two-zone radiation model where the *SEP* of the lower clear burning zone (*LZ*) is denoted by $\overline{SEP}^{ma}_{\mathrm{cl}}$ (Eq. 5.10c), whereas the *SEP* for upper black soot zone (*SZ*) is denoted by $\overline{SEP}_{\mathrm{u}}$. The *SEP* of two zones, depending on the area fraction of the smoke zone ($\bar{a}_{\mathrm{SZ}}$), can be calculated according to Eq. (10d) [32].

$$\overline{SEP}^{\mathrm{ma}}_{\mathrm{cl}}(D) = \overline{SEP}_{\mathrm{max}}\ (1 -\ e^{-kD}) \tag{10c}$$

$$\overline{SEP}_{\mathrm{act}} = (1 -\ \bar{a}_{\mathrm{SZ}})\overline{SEP}^{\mathrm{ma}}_{\mathrm{cl}} + \bar{a}_{\mathrm{SZ}}\overline{SEP}_{\mathrm{SZ}} \tag{10d}$$

From a hazard prediction point of view, the summation of thermal radiation from black soot and radiation from the luminous spots on an equivalent area basis is used to reach an average emissive power for the fire. If we consider two assumptions of 35% and 65% for the surface area covered with black smoke and the remaining part with luminous spots, the time average emissive power is given by the following expressions (Eqs.10e - 10f).

$$\overline{SEP}_{\mathrm{act}} =\ 0.65\ [140] +\ 0.35\ [20] = 98\ \mathrm{kW/m^2} \tag{10e}$$

$$\overline{SEP}_{\mathrm{act}} =\ 0.35\ [120] +\ 0.65\ [20] = 55\ \mathrm{kW/m^2} \tag{10f}$$

Where gasoline-pool fires show 1) $\bar{a}_{\mathrm{SZ}}$ = 0.35 and $\overline{SPM}^{\mathrm{ma}}_{\mathrm{MSFM}}$ = 140 kW/m^2 for Eqs. (10e) and 2) $\bar{a}_{\mathrm{SZ}}$ = 0.65 and $\overline{SPM}^{\mathrm{ma}}_{\mathrm{MSFM}}$ = 120 kW/m^2 for Eq. (10 f), and the $\overline{SEP}_{\mathrm{SZ}}$ = 20 kW/m^2 for Eqs. (10e, 10f) with k ≈ 2.0.

5.4.3. Two-zone Radiation model (TZM)

As illustrated in Fig. 10, most hydrocarbon fuel fires become optically thick when the diameter is approximately 3 m or larger. Although the thermal radiation from black soot is low, the hot spots appearing on the flame surface due to turbulent mixing have a higher emissive power. Corresponding to the empirical radiation model according to Mudan [17] for sooty pool fires and the time-averaged surface emissive power, the following Eq. (11a) is to be used.

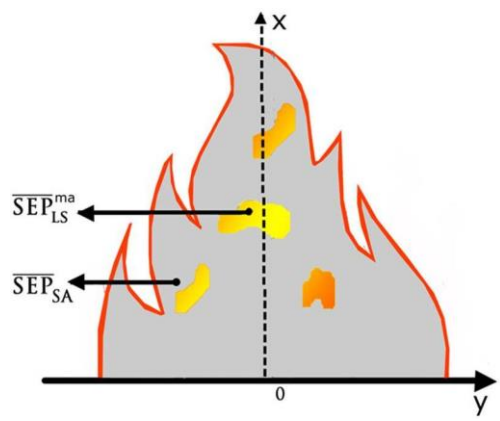


Fig. 10. Two-zone radiation model [17]

$$\overline{SEP}_{act}(D) = \overline{SEP}^{ma}_{LS}\bar{a}_{LS}(D) + \overline{SEP}_{SA}\,(1 - \bar{a}_{LS}\,(D)) \tag{11a}$$

where, the area fractions are estimated by Eq. (11b).

$$\bar{a}_{\mathrm{LS}}(D) = \bar{A}_{\mathrm{LS}}/\bar{A}_{\mathrm{F}} = 1 - \bar{a}_{\mathrm{SA}} = e^{-sD} = \mathrm{e}^{-0.12D} \tag{11b}$$

According to Mudan and Croce [49], a uniform surface emissive power of flames for smoky hydrocarbon fuels can be determined by Eq. (11c). Although the thermal radiation from black soot is low, the hot spots appearing on the flame surface due to turbulent mixing have a higher emissive power.

$$\overline{\mathrm{SEP}}_{\mathrm{act}}(D) = \overline{\mathrm{SEP}}^{\mathrm{ma}}_{\mathrm{LS}}\mathrm{e}^{-sD} + \overline{\mathrm{SEP}}_{\mathrm{SA}}\,(1 - \mathrm{e}^{-sD}) \tag{11c}$$

$$\overline{\mathrm{SEP}}_{\mathrm{act}}(D) = 140\ \mathrm{e}^{-0.12D} + 20\ (1 - \mathrm{e}^{-0.12D}) = 27\ \mathrm{kW/m^2} \tag{11d}$$

Mudan and Croce [49] proposed an actual $\overline{SEP}_{act}(D)$ averaged over the flame surface based on the means $\overline{SEP}^{ma}_{LS}$ = 140 kW/m^2 ≠ $\eta(D, \eta)$ and $\overline{SEP}_{SA}$ = 20 kW/m^2 ≠ η (D, η). For larger pool fires with $D \geq 20$ m, $\overline{SEP}_{act}$ $(D) \approx 20\ (1 - e^{-0.12D})$ is also valid so that for larger pool and tank fires, the hot and luminous spots [right side first in Eqs. (11a, c) are eliminated.

5.4.4. Thermal Radiation for Single and Multiple Tank Fires (TRSMF)

For multiple tank fires, as occurred in the IOCL Jaipur incident, the interaction of neighbouring tank fires has a considerable effect on the *SEP* of the individual tank fires due to heat feedback enhancement. To determine the surface emissive power of a flame, the flame surface area A_F has to be calculated.The thermal radiation, that is, the maximum surface emissive power SEP^{ma} (without blocking by black soot), of a tank fire can be calculated with [28, 32-33,44-45,49].

$$SEP^{ma} = \eta_{rad}(D)\,\bar{Q}_c/\bar{A}_F = 114 \text{ kW/m}^2 \tag{12a}$$

With

$$\bar{Q}_c = \bar{m}_f''(-\Delta H_c)A_P \tag{12b}$$

For the cylinder, the flame area is given by

$$\bar{A}_F = \pi D\bar{H}(D) + \pi D^2/4 \tag{12c}$$

The time-averaged $\bar{A}_F$ is determined from the instantaneous area A_F, which is influenced by the flame fluctuations. According to Eq. (12a), doubling of m_f'' as a result of the interaction brings about a doubling of the thermal radiation of the hot spots. These types of effects were investigated theoretically and experimentally by Gawlowski *et al.*, [46].

6. Irradiance

The received thermal flux, i.e., the irradiance at any point, is calculated by a point source model, which assumes that heat radiation of the flame is irradiated from a point that equally disperses in a radial direction from the emission point as a sphere, as shown in Fig. 11. [47].

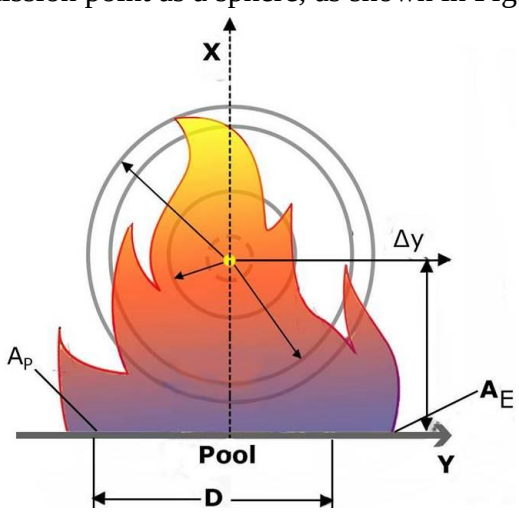


Fig. 11. Point source radiation model [44]

The point source radiation model (PSM) calculates the mean irradiance (received thermal radiation flux) from the following relationships [49]

$$E_r = \tau_a \bar{\eta}_{rad} m_f \Delta H_c A_F F_P \quad (13a)$$

The view factor F_P is calculated according to the fundamental relation of view factor with respect to distance (Eq. 13b).

$$F_P = 1/4\pi x^2 \quad (13b)$$

The *PSM*, however, has only a very limited range of validity. In particular, in the near field, great uncertainties exist.

In the IOCL Jaipur incident, the mean irradiance E_r versus distance was calculated for the gasoline tank (*D = 24*) fires with the point source (*PS*) radiation model and was validated with the DNV Norway-based risk assessment PHAST Software estimation, as shown in Fig. 12. The percentage error between the estimated and calculated irradiance is 17% at a 100 m distance from centre of flame.

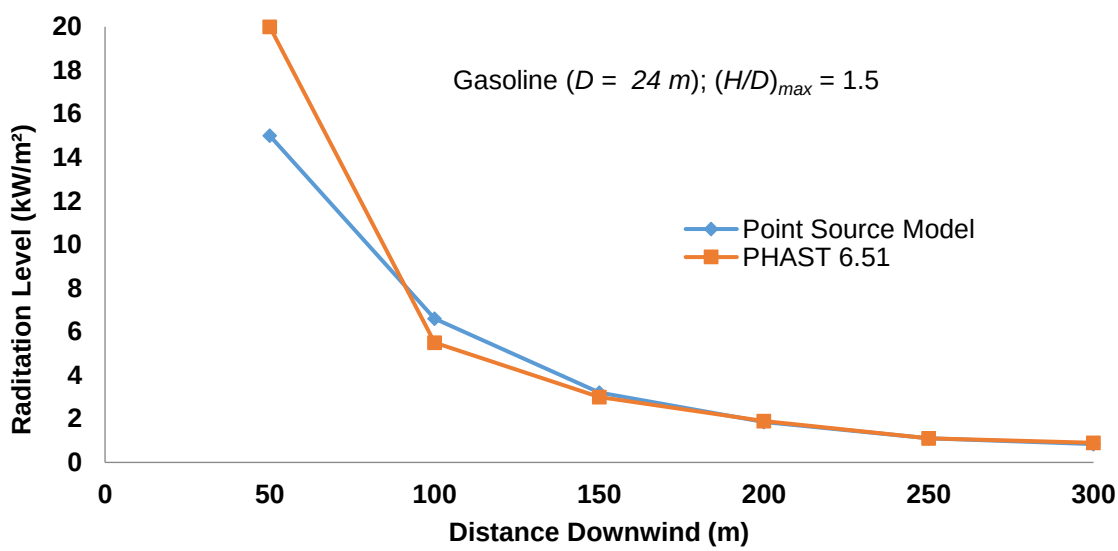


Fig. 12. Estimation of the received thermal flux from a gasoline (D = 24) tank fire with the point source model and PHAST software

7. Results and discussion

7.1. Flame Shape

The $(H/D)_{max}$ ratio computed by the Munoz correlation is 1.5, whereas the observed value lies in between 1.0 and 1.7. Thus, the calculated value is within the observed value. The average value of $(\bar{H}/D)_{\text{calc}}$ is 0.9. The clear burning zone heights (H_{Cl}/D) were obtained by various models such as *MSFM*, Pritchard and Binding, and Ditali models. The Pritchard and Binding and Ditali models use (*C/H*) ratios to indicate the saturation of the hydrocarbon fuel. The *MSFM* model gives a

maximum value of 0.6, whereas the Pritchard and Binding and Ditali models predict 0.3 and 0.4, respectively. This trend shows that the flame height in the Jaipur incident case was unusually large.

The estimated maximum, relative, and clear burning zone flame heights using the above correlations for the IOCL Jaipur incident are shown in Table 2.

Table. 2. Flame heights for a gasoline tank (D = 24 m) on fire in the IOCL Jaipur incident

(H/D)	Ratio	Flame Height (m)	Models
$(H/D)_{max\ observed}$ *	1.0 – 1.7	24-40	*Observed Range*
$(H/D)_{max}$	1.5	36	*Thomas relation*
$(\bar{H}/D)_{calc}$	0.9	22	*Thomas relation*
$\bar{H}_{cl}^{MSFM}/D$	0.6	15	*MSFM*
H_{Cl}/D	0.3	7	*Pritchard and Binding Correlation*
H_{Cl}/D	0.4	10	*Ditali Correlation*

**Observed Flame Height at the time of incident of IOCL, Jaipur*

7.2. Thermal radiation intensity

The analysis of the *SEP* in the Jaipur accident, estimated with various models, is given in Table. 3, which indicates that a higher *SEP* ($\overline{SEP}_{\mathrm{SFM}}^{\mathrm{ma}}$ ≈123 kW/m^2) value was reached by *SFM*, where the flame is considered as a single luminous zone. In the case of multiple tank fires, the interaction of neighbouring tank fires, as in case of Jaipur incident, has a considerable effect on the *SEP* (SEP^{ma} ≈ 114 kW/m^2) of the individual tank fires. In the later stage, the radiation from inside the flames is blocked by absorption of dense soot parcels. Subsequently, the effect of the sooty zone is calculated by *MSFM* with considerable assumptions ($\overline{SPM}_{MSFM}^{ma}$ ($\bar{a}_{SZ}$ = 0.35) = 98 kW/m^2 and $\overline{SPM}_{MSFM}^{ma}$ ($\bar{a}_{SZ}$ = 0.65) = 55 kW/m^2). In addition, the turbulent mixing phenomena also influences the *SEP* ($\overline{SEP}_{act}$ ≈ 27 kW/m^2), which can be estimated by *TZM*. Babrauskas [40] reported that as the pool or tank diameter increases, the fire regime changes from laminar to turbulent.

Table. 3. Surface Emission Power (*SEP*) of a gasoline tank on fire in the IOCL Jaipur Incident

Models	(SEP)	kW/m^2
SFM	$\overline{SEP}_{\mathrm{SFM}}^{\mathrm{ma}}$	123
MSFM	$\overline{SPM}_{\mathrm{MSFM}}^{ma}$ $\bar{a}_{\mathrm{SZ}}$= 0.35	98
MSFM	$\overline{SPM}_{\mathrm{MSFM}}^{ma}$ $\bar{a}_{\mathrm{SZ}}$ = 0.65)	55
TZM	$\overline{\mathrm{SEP}}_{\mathrm{act}}$	27
TRSMFM	SEP^{ma}	114

7.3. Thermal heat flux

Piont source model and PHAST software were used to estimate of Thermal heat flux with respect to distance. The estimated thermal heat flux was at 50m distance approximately 15kW/m^2 by point source model while about 20kW/m^2 by PHAST. While at 100m distance about 5kW/m^2 thermal

heat flux by Point model and 6.5kW/m^2 by PHAST. There is 20% error at nearer distances, whereas the effective values at higher distances are almost equal.

8. Preventive Measures in Oil Depots

Decisions regarding investments into fire safety generally have to be made under uncertainty [50-51]. This stems both from the inherent randomness of large fire events and from the fact that we are not able to fully understand and model the underlying phenomena. The overall goal of quantitative fire risk assessment is to support decisions on risk reduction measures by estimating their impact on the expected consequences (e.g. financial losses or human fatalities) of all possible fire scenarios [50,52-53]. The basic requirement for the fire risk models to be used for decision-making is to assess a risk as a function of the safety measures installed by incorporating decision variables. Another important requirement is to model risk-relevant characteristics of the fuel storage terminals and installations. Finally, the model should assess the risk as accurately as possible. The available fire risk reduction measures are grouped into the following main categories, as shown in Fig. 13.

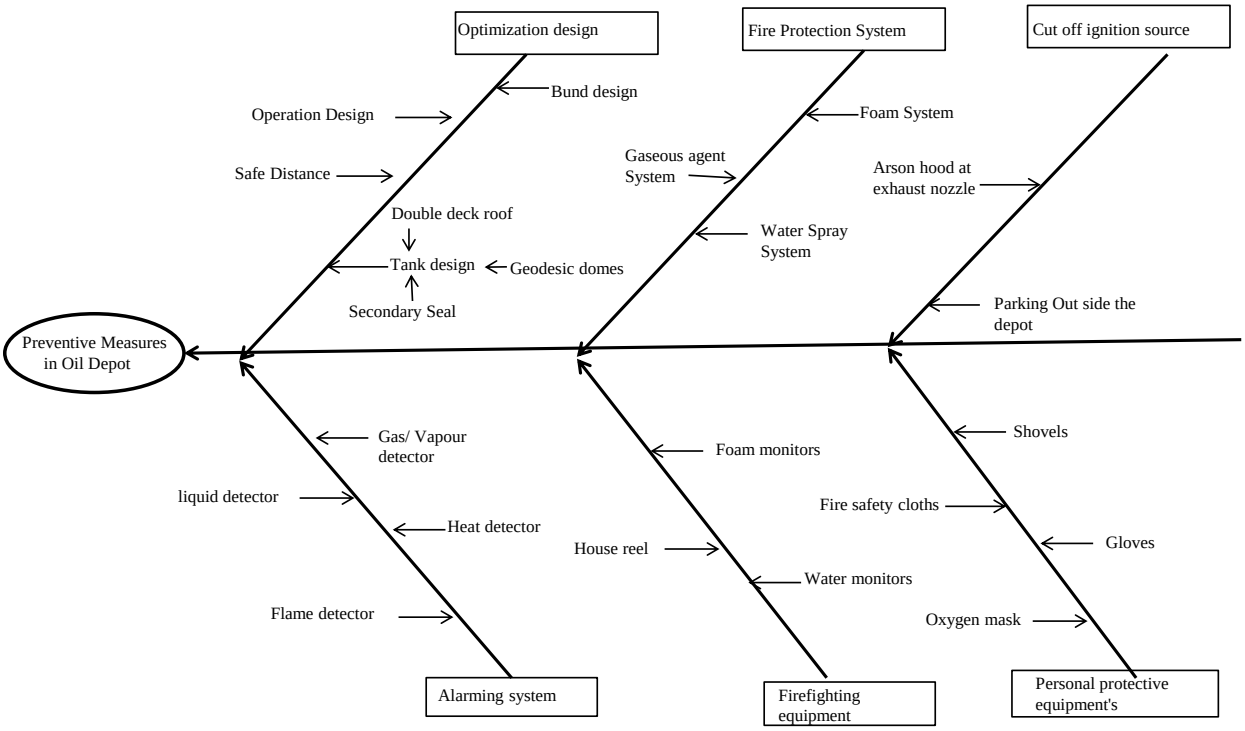


Fig. 13. Fishbone diagram of pool or tank fire prevention

9. Conclusion

In this paper, fire risk analysis and implementing a quantitative analysis models are presented. Large-scale pool fires of liquid hydrocarbons show fundamentally different characteristics such as generally much higher mass burning rates, large flame heights and high irradiances. To measure, calculate and study the fire characteristics, simulations and modelling of hydrocarbon large-scale

pool fires were performed using various models like SFM, MSFM and TZM etc. The present simulations are based on the assumption of a complete combustion without wind influence.

The computed mass burning rate and flame temperature, based on the data collected, are within the ranges reported in the peer-reviewed literatures. The modelling analysis of the Jaipur accident revealed that the height to diameter ratio $(H/D)_{max}$ lies between 0.9 and 1.5. This ratio is approximately 1.6 times the average value and lies well within the observed values. For the computation of the clear burning zone, the Pritchard and Binding correlation results in better values than the Ditali and *MSFM* models.

The surface emissive power from a tank fire was calculated by changing the percentage of black smoke and luminous spots covering the flame. Four models namely *SFM, MSFM, TRM, TRSMFM* were used to predict the values of *SEPs* that showed a most likely fire scenarios occurred during the accident. The irradiance at various distances based on the point source model were compared with the DNV Norway-based risk assessment PHAST 6.51 Software and it is found that the result was correct with a maximum error of 20%.

A physical explanation of the Jaipur accident with regard to the relative flame heights *H/D* and thermal radiation (*SEP*, *E*) is in principal possible, in particular when considering the effective consequence models including the observations regarding multiple tank fires. However, it is necessary to overcome the lack of field data, especially with regard to the *H/D*, *SEP*, and E_r for larger individual as well as multiple tank and pool fires for their more realistic characterization. Furthermore, continuous efforts are required to improve the large tank fire modelling to simulate real life scenarios coping with changes in technology and management of petrochemical storage terminals. Accordingly, future research can be carried out on the verification methods of key parameters, the measurement method of the impacts of fire on business community, heritage and the environment, and the appropriate risk management strategy to reduce fire risks in storage terminals.

Nomenclature

a_{sz}	[-]	area fraction of smoke zone
$\bar{A}_F$	[m^2]	time averaged flame area
$\bar{A}_{\mathrm{LS}}$	[m^2]	time averaged flame area of yellow luminous spots
A_P	[m^2]	pool area [m^2]
BPCL		Bharat Petroleum Corporation Limited
C/H		carbon to hydrogen atomic ratio in fuel
D	[m]	pool or tank diameter
E_r	[kW/m^2]	irradiance at the horizontal distance Δy from flame surface
F_P	[m^{-2}]	point source view factor
$Fr_{\mathrm{f}} \equiv \frac{\bar{m}_f''}{\rho_{a\sqrt{gD}}}$	[-]	fuel Froude number
g	[m^2/s]	gravitational acceleration
H	[m]	height of the combustion zone
H_{cl}	[m]	height of the hot, clear burning zone not obscured by black smoke
$\bar{H}/D$	[m]	time-averaged relative height or height of the visible flame [m]
$\left(H/D\right)_{max}$	[m]	maximum relative height of the visible flame
ΔH_c	[J/kg]	Heat of Combustion
H_{cl}	[m]	height of the hot, clear burning zone not obscured by black smoke
$\triangle H_c$	[J/kg]	(specific) heat of combustion
ΔH_{v}	[J/kg]	Latent heat of vaporization
HPCL		Hindustan Petroleum Corporation Limited
m^*	[-]	dimensionless mass burning rate of fuel
$\bar{m}_f''$	[kg s^{-1}m^{-2}]	mass burning rate
$\bar{m}_{\mathrm{f,max}}''$	[kg s^{-1}m^{-2}]	maximum mass burning rate
Q	[kW]	heat release rate of the fire
$\bar{Q}_{\mathrm{c}}$	[kW]	heat of combustion
SEP	[kW/m^2]	Surface Emission Power of the flame
t	[s]	time
T_a	[K]	ambient temperature
T_f	[K]	average surface emission temperature of the flame (K)
t	[s]	time
u_w^*	[m/s]	wind velocity
U_9^*		dimensionless wind speed measured at a height of 9m
u^*		Dimensionless wind speed
$\bar{\mathrm{V}}_{\mathrm{a,max}}$	[m/s]	maximum burning velocity
x	[m]	distance from the point source to the targe

Greek symbols

η_{rad}	radiative fraction of the fire
β	correction factor referring to the optical path height
kβ	mean beam length corrector-flame attenuation coefficient product
τ_a	atmospheric transmissivity
$\bar{\varepsilon}_{\mathrm{F}}$	effective emissivity of the (gray) flame
ε_{F}	flame emissivity

σ	[$kWm^{-2}K^{-4}$]	Stefan-Boltzmann-constant
ρ_a	[kg/m^3]	ambient density
ρ_f	[kg/m^3]	density of fuel

Subscripts

a	ambient conditions
act	actual quantity (i.e., the luminous flame is partly obscured by black smoke)
calc	calculated or predicted quantity
cl	hot, clear burning zone of the height or height H_{cl}
exp	experimental quantity
hs	hot spots
kβ	mean beam length corrector extinction coefficient product, m^{-1}
la	laminar
LS	yellow luminous spots
ma	maximum SEP, i.e. the flame is not obscured with black smoke
max	maximum value of a quantity
P	pool
SA	area of black smoke
sp	parcel of black smoke
SZ	zone of black smoke

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Risk assessment of a large chemical complex during the construction phase using Intuitionistic Fuzzy Analytic hierarchy process

Dr. A B Bhasi, Dr. V R Renjth, Suresh G*
School of Engineering, Cochin University of Science and Technology
Kerala, India
*Presenter E-mail: sureshkankhal@cusat.ac.in

Abstract

The safety during construction phase of large chemical complex is very critical in terms injuries or property damages. In this paper the risks that possess in critical operations carried out during construction phase of project were considered. Which are categorised based on a novel method of Intuitionistic Fuzzy Analytical hierarchy process. The different categories are Critical, Serious, Minor. The Analytical hierarchy process (AHP) possess the unique advantage of comparing parameters that have no units or scale of measurements. The Intuitionistic Fuzzy Analytical hierarchy process (IFAHP) further improves the AHP in terms of vagueness, uncertainty and handles the imprecise data.

Keywords: Chemical complex, Construction project, Risk assessment, IFAHP

1.Introduction

The construction activities mostly labour intensive and is the largest economic activity after the agriculture sector. The construction sector mainly divided into building, infrastructure, and industrial. In the industrial construction sector. We consider the chemical industry, where piping, reactors, vessels and other are present. These construction activities are mainly carried out through unorganised sectors. In unorganised sectors, the countries like India the labours are mainly migrant

speaks different languages illiterate. So, ensuring and practicing safety in this sector is a difficult task. The evaluation and data obtained through this study cannot be generalized. What should be reviewed for preparing the future general safety plan in regional level.

This study utilizes multi criteria decision making techniques like intuitionistic fuzzy analytical hierarchy process (IFAHP) in Safety Engineering. This is a natural way to quantify a parameter, that cannot be measured or difficult to measure by using any fundamental scale or derived scales like kilogram, metre, kilowatt etc. But we can quantify the parameter by comparing the intensities or changes in it. The AHP is a technique, that utilizes pairwise comparison to prioritise a component from a group of components. Which are tangled together like Pasta strings, in which strands are separate. In pairwise comparison, the elements in pairs are compared against a given criteria. The intuitionistic fuzzy analytical hierarchy process improves AHP in terms of vagueness, uncertainty, ambiguity and handles imprecise data.

2. Literature Review

The IFAHP is utilized in various domains such as, supply chain management, engineering, e-commerce, banking, risk assessment. The some of the applications of IF-AHP in the various disciplines are described below

2.1 IFAHP in prioritizing alternatives

Sadiq [16] applied to select best drilling fluid in environmental decision-making process. It utilized a similarity measure to reduce the number of alternatives by grouping different alternatives into a single class or cluster. Also, for ranking generalized mean and standard deviations are followed.

Jian Wu [19] applied interval valued IFAHP in e-commerce domain, here a score judgement matrices and obtained its associated interval multiplicative matrix to calculate the priority vector.

Cengiz Kahraman [7] applied Intuitionistic Fuzzy originated Interval type -2 FAHP in the application of dam less hydroelectric power plants. The triangular IF linguistic evaluation scale is utilized in pairwise comparison, which is transferred into Triangular Type-2 Fuzzy (TT2F) pairwise comparison matrix. Applied classical AHP to find best alternatives and finally TT2F is defuzzied.

Nirmala [11] proposed Triangular IFAHP with location index number and fuzziness index function for represent the TIFN in the application to select best computer

Zeshui Xu [20] proposed an algorithm to repair inconsistencies in his article for global supplier selection.

Yao Yu [21] described to rank the risk factors in transnational public- private partnership projects based on IFAHP. The consistency is checked by the distances between given intuitionistic

preference relation and its perfect multiplicative consistent intuitionistic preference relations were used. An algorithm is proposed for repair the inconsistent relations.

2.2 IFAHP in risk estimation

Hoang Nguyen [10] Introduced IFAHP in ship system risk estimation. the priority vector of consequences is found out and a membership knowledge measure of IFV's is utilized for final ranking.

Selcuk Cebi [2] is proposed IVIFAHP for warehouse risk estimation, score judgement matrix and possibility degree matrix were utilized for prioritization.

2.3 Construction safety

Safety and health in construction by ILO [5] and Health and safety in construction [17] describes an overall picture of safety at construction sites. Every year many construction site workers are killed or injured as a result of their work, others suffer ill health, such as musculoskeletal disorders, dermatitis or asbestosis. The hazards are not, however, restricted to those working on sites. Children and other members of the public are also killed or injured because construction activities have not been adequately controlled. So, the construction sector should be analysed for improvements in the sense of health and safety.

3. Basic concepts

The basic concepts of intuitionistic fuzzy AHP, which are adopted in this study is described here.

3.1 Intuitionistic Fuzzy sets

The concept of Fuzzy sets given by Zadeh is generalised by Atannasov [1] by introducing membership function and non-membership function for that elements of the universe of discourse. An Intuitionistic Fuzzy Set A in X is defined as an object of the following form

$$A = \{[x, \mu A (x), \upsilon A (x)] / x \in X\}$$

Where the function $\mu A: X \rightarrow [0,1]$
$\upsilon A: X \rightarrow [0,1]$

$\mu A (x)$ – degree of membership

$\upsilon A (x)$ – degree of non-membership

such that $0 \leq \mu A (x) + \upsilon A (x) \leq 1$

Obviously, each ordinary fuzzy set may be written as

$$\{[x, \mu A (x), (1-\mu A (x))] / x \in X\}$$

The value of $\pi A(x) = 1 - \mu A(x) - \upsilon A(x)$ is called the degree of non- determinacy or uncertainty or hesitancy.

3.2 IF Operations[20]

Let $a_{pq} = (\mu_{pq}, \upsilon_{pq})$; $a_{rs} = (\mu_{rs}, \upsilon_{rs})$

1. $a_{pq} + a_{rs} = (\mu_{pq} + \mu_{rs} - \mu_{pq}\mu_{rs}, \upsilon_{pq}\upsilon_{rs})$
2. $a_{pq} \times a_{rs} = (\mu_{pq}\mu_{rs}, \upsilon_{pq} + \upsilon_{rs} - \upsilon_{pq}\upsilon_{rs})$
3. $\lambda a_{pq} = (1-(1-\mu_{pq})^{\lambda}, \upsilon_{pq}^{\lambda})$, $\lambda > 0$
4. $a_{pq}^{\lambda} = (\mu_{pq}^{\lambda}, (1-(1-\upsilon_{pq})^{\lambda})$, $\lambda > 0$

4. Prioritization methodologies for Intuitionistic Fuzzy AHP

4.1 Structuring hierarchy

Humans have the ability to perceive things and ideas, to identify them and to communicate what they observe. Our mind structures complex reality into its constituent parts. These in turn into their parts and so on hierarchically. The numbers of parts usually range between five, and nine.

By breaking down the reality into homogeneous clusters and subdividing these clusters into smaller ones. We can integrate large amount of information to structure of a problem and form a more complete picture of the whole system.

4.2 Setting priorities

Humans also have the ability to perceive relationship among the things they observe, to compare in pairs of similar things against certain criteria and discriminate between both members of a pair by judging the intensity of their preference for one over the other. The relationship between the elements of each level of hierarchy by comparing the elements in pairs. This relationship represents the relative impact of the elements of a given level on each element of the next higher level. The latter element serves as a criterion is called a property.

The result of the discrimination process is a vector of priority of relative importance of the elements with respect to each property. This pair wise comparison is repeated for all the elements in each level. The final step is to come down the hierarchy by weighing each vector by the priority of its property. This synthesis results in a set of net priority weights for the bottom level. The elements with the highest weight in the one that merits the most serious consideration for action although the others are not rule out entirely.

4.3 Establish priorities

In a decision problem the first step in establishing the priorities [14] of elements is to make pair wise comparisons. That is, to compare elements in pairs against a given criteria. For pairwise comparison, a matrix is the preferred form. The matrix is simple, well-established tool that offer a framework for testing consistency, obtaining additional information through making all possible comparisons and the analysing the sensitivity of overall priorities to change in judgements.

To begin with the pairwise comparison process, start at the top of the hierarchy to select the criterion 'C'. That will be is used for making the first comparison. Then from the level immediately below, take the elements to be compared A_1, A_2, …. An

In the matrix compare the element A_1 in the column on the left with element A_1, A_2, …. An in the row on the top with respect to the property 'C' in the upper left-hand corner. Then repeat the column elements A_2 and so on. To compare elements, ask how much more strongly does these elements or activity possess or contribute to, dominate, influence, satisfy or benefit, the property then does the element with which it is being compared.

To fill in the matrix for pairwise comparison, we use numbers to represent the relative importance of one element over another with respect to the property.

The fundamental scale for pairwise comparisons is described below

Table 1. Fundamental scale for pairwise comparisons

Intensity	Definition
1	Equal importance
3	Moderate importance
5	Strong importance
7	Very strong importance
9	Extreme importance
2,4,6,8	For comparisons between the above values
Reciprocals	If activity 'i' has one of the above non zero numbers assigned to it, when compared with activity 'j', then 'j' has the reciprocal value when compared with 'i'
1.1-1.9	When elements are close and nearly indistinguishable
1.3	Moderate importance
1.9	Extreme importance

Experience was confirmed that a scale of nine units [15] is reasonable and reflect the degree to which one can discriminate the intensity of relationship between elements.

Using the scale in a social, psychological, or political context, express the verbal judgement first and then translate them to the numerical values. The numerically translated judgements are approximations, and their validity can be evaluated by a test of consistency.

4.4 Intuitionistic preference relations

The pairwise comparisons were represented by intuitionistic fuzzy values.

An intuitionistic fuzzy set A on the universe of discourse X = {x_1, x_2…., x_n}is represented as

A = {[x, μ A(x), υ A(x)] / x ϵ E}

The Intuitionistic preference relations A_{pq} is conveniently represented as

A_{pq} = (μ_{pq}, υ_{pq}), where μ_{pq} represents the degree to which the object x_p is preferred over the object x_q , υ_{pq} represents the degree to which the object x_p is not preferred over the object x_q and the π = 1-μ_{pq}- υ_{pq} is represented as a hesitancy function with the conditions

μ_{pq} , υ_{pq} ϵ [0,1], $\mu_{pq} + \upsilon_{pq} \leq 1$, $\mu_{pq} = \upsilon_{qp}$, $\mu_{qp} = \upsilon_{qp}$,

$\mu_{pq} = \upsilon_{qp} = 0.5$ for all p, q = 1, 2…..., n

4.5 Consistency checking `

Consistency means the ability to establish relationship among objects [8] or ideas in such a way that they are coherent, or they relate well to each other. It is uncertain that our judgements in the pairwise comparison matrix is perfect reciprocals of the transpose position, because we integrate new experiences into our consciousness so that the relationships may change, and some consistency may lose. The newest ideas that affect our lives tend to maintain coherence among the objects of our experiences, that causes us to rearrange some of our previous commitments. If we decided to never change our minds, we would be afraid to accept new ideas. All knowledge must be admitted into our narrow corridor between tolerable inconsistency and perfect consistency.

In this work we used the algorithm by Xia and Xu [9] to construct a perfect multiplicative consistent intuitionistic preference relation

$\bar{A} = (\bar{a}_{pq})_{nxn}$

For q= p+1, let $\bar{a}_{pq} = (\bar{\mu}_{pq}, \bar{\upsilon}_{pq})$, where a

$$\bar{\mu}_{pq} = \frac{\sqrt[q-p-1]{\prod_{r=p+1}^{q-1} \mu_{pr}\mu_{rq}}}{\sqrt[q-p-1]{\prod_{r=p+1}^{q-1} \mu_{pr}\mu_{rq}} + \sqrt[q-p-1]{\prod_{r=p+1}^{q-1} (1-\mu_{pr})(1-\mu_{rq})}}, \quad q > p+1$$

$$\bar{\upsilon}_{pq} = \frac{\sqrt[q-p-1]{\prod_{r=p+1}^{q-1} \upsilon_{pr}\upsilon_{rq}}}{\sqrt[q-p-1]{\prod_{r=p+1}^{q-1} \upsilon_{pr}\upsilon_{rq}} + \sqrt[q-p-1]{\prod_{r=p+1}^{q-1} (1-\upsilon_{pr})(1-\upsilon_{rq})}}, \; q > p + 1$$

$\bar{a}_{pq} = a_{pq}$, q = p + 1;

$\bar{a}_{pq} = (\bar{\upsilon}_{pq}, \bar{\mu}_{pq})$, q < p

By applying these equations, we can only update less than half of the elements in the original intuitionistic preference relations to construct the perfect multiplicative consistent Intuitionistic preference relations

$\bar{A} = (\bar{a}_{pq})_{nxn}$ for A

The intuitionistic preference relations A is an acceptable multiplicative consistent intuitionistic preference relation if d (A, $\bar{A}$) < τ

where d (A, $\bar{A}$) is distance measure between the intuitionistic preference relations A and its corresponding perfect multiplicative consistent intuitionistic preference relations $\bar{A}$. Which can be calculated by

$$d(\bar{A},A) = \frac{1}{2(n-1)(n-2)} \sum_{p=1}^{n} \sum_{q=1}^{n} \left(\left|\tilde{\mu}_{pq} - \mu_{pq}\right| + \left|\tilde{\upsilon}_{pq} - \upsilon_{pq}\right| + \left|\tilde{\pi}_{pq} - \pi_{pq}\right| \right)$$

"τ" is the consistency threshold.

The inconsistent intuitionistic preference relations $A = (a_{pq})_{nxn}$ can be transformed in to corresponding perfect multiplicative consistent intuitionistic preference relations $\bar{A} = (\bar{a}_{pq})_{nxn}$.

The distance d ($\bar{A}$,A) between intuitionistic preference relations and the transformed one can be calculated. If distance d ($\bar{A}$,A) is too larger then the transformed intuitionistic preference relations $\bar{A}$ can not be represent the initial preference of the decision-maker. It is desirable that the modified intuitionistic preference relations should not only have acceptable multiplicative consistent but maintaining the original preference information of the decision-maker as much as possible.

Hence it is beneficial to combine initial intuitionistic preference relations A and the corresponding perfect multiplicative intuitionistic preference relations $\bar{A}$ into a joined intuitionistic preference relation

$\tilde{A} = (\tilde{a}_{pq})_{nxn}$, where each element is defined as

$$\tilde{\mu}_{pq}^{(t)} = \frac{(\mu_{pq}^{(t)})^{1-\sigma} \quad (\tilde{\mu}_{pq}^{(t)})^{\sigma}}{(\mu_{pq}^{(t)})^{1-\sigma} \quad (\tilde{\mu}_{pq}^{(t)})^{\sigma} + (1-(\mu_{pq}^{(t)})^{1-\sigma}(1- \quad (\tilde{\mu}_{pq}^{(t)})^{\sigma}}$$

$$\tilde{\upsilon}_{pq}{}^{(t)} = \frac{(\upsilon_{pq}{}^{(t)})^{1-\sigma} \quad (\tilde{\upsilon}_{pq}{}^{(t)})^{\sigma}}{(\upsilon_{pq}{}^{(t)})^{1-\sigma} \quad (\tilde{\upsilon}_{pq}{}^{(t)})^{\sigma} + (1-(\upsilon_{pq}{}^{(t)})^{1-\sigma}(1- \quad (\tilde{\upsilon}_{pq}{}^{(t)})^{\sigma}}$$

Where t is the number of iterations, σ is a controlling parameter, that is determined by the decision-maker the smaller the value of σ, Ã =A; if σ = 0, Ã =A; if σ = 1, Ã = Ā also Ã is an intuitionistic preference relation.

The generally the combined intuitionistic preference relations Ã contains not only the original preference information of the intuitionistic preference relations but the preference information of its corresponding perfect multiplicative consistent preference relations Ã. The controlling parameters σ also represents the preference of the decision-maker to some extent. Based on the above-mentioned analyses the equation to repair inconsistent intuitionistic preference relations described

Through this equation we can improve the consistency level of any intuitionistic preference relation without losing much original information. This iterative method we can save a lot of time of the decision maker.

4.6 Composite priorities and ranking

The intuitionistic preference relations do not directly give the priorities. According to Saaty's concepts, n- dimentional vectors ω = (ω_1, ω_2……. ω_n) is estimated from the multiplicative preference relations. The ω_p is the weight which accurately represents relative dominance of the alternative A_p among the alternatives in A.

The intuitionistic preference relations A= $(a_{pq})_{nxn}$, where $a_{pq} = (\mu_{pq}, \upsilon_{pq})$, since μ_{pq} , υ_{pq} ϵ [0,1] , $\mu_{pq} + \upsilon_{pq} \leq 1$, then $\mu_{pq} = 1 - \upsilon_{pq}$. So, the intuitionistic preference relation's membership and non-membership pairs. Which can be transformed into (μ_{pq},1- υ_{pq}) the interval valued preference relations. So that, the intuitionistic preference relations A= $(a_{pq})_{nxn}$ is transformed into an interval valued preference relation $A' = (a'_{pq})_{nxn} = (\mu_{pq}, 1-\upsilon_{pq})_{nxn}$ based on the operational laws of intervals the priority weights are estimated using the formulae

$$\omega_p = \left(\frac{\sum_{q=1}^{n} \mu_{pq}}{\sum_{p=1}^{n}\sum_{q=1}^{n}(1-\upsilon_{pq})}, \; 1 - \frac{\sum_{q=1}^{n}(1-\upsilon_{pq})}{\sum_{p=1}^{n}\sum_{q=1}^{n} \mu_{pq}} \right)$$

For an intuitionistic fuzzy set π A (x) = 1- μ A(x) + υ A(x), the degree of uncertainty of the membership of element x ϵ X to the set A. In ordinary fuzzy set π A (x) = 0 for every x ϵ X.

π A (x) = [0,1] for all x ϵ X , Let α = (μ_α, υ_α, π_α) be an intuitionistic fuzzy value

Szmidt and Kaeprzyk [4] proposed a relation to rank the intuitionistic fuzzy value

ρ(α) = 0.5 (1+ π_α)(1- μ_α)

The smaller the value of ρ(α), the greater the intuitionistic fuzzy value α in the sense the amount of positive information included and reliability of information.

5. Procedure for intuitionistic fuzzy analytic hierarchy process

The procedure for intuitionistic fuzzy analytic hierarchy process is described below

1. Identify the objective, safety factors, sub factors and construct the hierarchy of the problem
2. Evaluate intuitionistic preference relation (IPR) by pairwise comparison between each factor, sub factors against the respective Criterion of severity, occurrence, detectability
3. Evaluate perfect multiplicative consist preference relation from IPR and measure the distance to check the consistency.
4. If it is inconsistent ($\tau > 0.1$) repair the inconsistent IPR by applying the auto correction formula in an iterative manner until $\tau < 0.1$.
5. Calculate the priority vector for each consistent IPR
6. Find out the overall weights from lowest level to the highest level by intuitionistic fuzzy (IF) operation
7. Evaluate the overall weights by multiplication of severity, occurrence, detectability weights to find risk priority number by IF operations.
8. Classify the safety factors in this case according to the boundary values decided as major, critical, minor and finally rank the safety factors using the equations.

6. Illustrations

Ensuring safety of construction site is a complex activity consisting of various factors and criteria. The multi criteria decision making techniques can be applied to analyse the problem in this work intuitionistic fuzzy analytical hierarchy process (IFAHP) is utilized for the risk assessment of a large chemical complex during construction phase.

The construction activities mainly include building construction, infrastructure development and industrial construction. Here in this problem we are conducted studies of a large chemical complex construction. The construction activities considered here includes only civil works like excavation, road construction, buildings. Fabrication and erection of structures, platforms, piping, vessels, other equipment, etc. Inspection and pre commissioning activities like, water flushing, air blasting, leak testing, storage, and handling of chemicals.

The methods adopted in this study consisting mainly the analysis of the past incident happened at different construction site and expert opinion [13] is recorded. Around twenty-one risk factors were identified and can be explained in detail. The risk factors identified are clustered in five groups as the primary risk factors and then to its respective sub factors.

Table 2. Risk factors

	Working at Height
1	Scaffolding, ladders
2	Falling from height, slips
3	Dropping Objects from Height
4	Safety belt & other protective gears
	Lifting & moving equipment
5	Manual lifting/moving
6	Lifting equipment
7	Vehicle and other moving equipment
	Confined space
8	Lighting & ventilations
9	Standby / watch
10	Simultaneous operations inside
11	Administrative control
	Electrical
12	Dragging of live cables
13	Energization of equipment / loto
14	Circuit breaker/earthing/extension
15	Short circuit/insulation failure
16	Striking underground/overhead lines
	General site management
17	Work permits
18	Usage of proper ppes
19	Barricading/ sign boards
20	House keeping
21	Handling of Chemicals

The various issues noticed under the sub factors [6] are

Scaffolding and Ladders: - Unhealthy scaffolding, substandard scaffolding (without bottom plate, toe guards, handrails). Not providing ladders to climb in the scaffolding.

Falling from height and slips, dropping of object from height, Safety belt and other protective gears [3] are self-explanatory

Manual lifting and moving [12] : - It is one of the prime areas of concern and lifting postures of manual lifting is very important to avoid injuries, sprains etc. Often peoples carry load and moving alone, or as a group will also lead into incidents due to slips for imperfect perceptions.
Lifting equipment: - The lifting tackles, chain blocks, pulleys, pads etc. should be take the loads successfully. But due to wear and tear or other damages. It should not take the certified loads. This should be checked before starting of the work. Lifting equipment should be statutorily carried out the load test within the stipulated time. Cranes mostly have interlocks and other safety integrated features, bypassing of these features are alarming.

Vehicle and other moving equipment: - Vehicles and other moving equipment possess a great threat in the construction site due to poor visibility, constraints in passages, reverse moments, signalling problem, communication and perception problems.

Ventilation and lightings: - In the confined space work the ventilation is most important factor. In construction site due to scarcity of utilities the eductors or other artificial air supplies is mostly not provided. The poor lighting also causes incidents inside the confined space like falling into deep inside the vessel.
Standby / watch: - In many cases, it is found that the standby person assigned with many other duties. It is most important to provide a full time dedicated standby person and watch.
Simultaneous operation inside confined space: - It has been seen that contradicting SIMOPS where carried out in confined space example, welding and flushing welding and painting.
Administrative control: - These are like entry registers, gas checking etc.

Dragging of live electrical cable: - This is found that a common activity in construction sites to drag the live cables from one location to another. Often the joints or insulation damaged areas where found sparking while dragging, especially welding cables.
Energisation of equipment / LOTO [18] : - It has been seen that the partial energization to check the equipment like pumps, compressors, fans, etc. and at this stage the Lotto system is not implemented. So, identification of energised equipment is become difficult. Also, in each equipment, three Types of supply comes. The main power supply, control supply and ancillary supplies to ancillary equipment like lube pumps, for space heaters, etc. Often it is difficult to identify the equipment supply is live or not.

Circuit breakers /earthing /extensions: - In many occasions the circuit breakers are found unhealthy and could not serve the function when it is activated. Earthing and bonding is also found improper. Taking multiple and series extension is found without circuit breakers.

Short circuiting and insulation failure: - Short circuiting is mainly due to wetting of electrical equipment due to water leaks, rains, flooding, etc. Installation failure due to dragging or crushing of lines.

Striking overhead and underground lines: - This happens mainly while moving material from one location to another. During excavation, crane moments, etc.

Work permit system: - It has been seen that construction site work permit system practice is very lenient.
Usage of proper personal protective equipment (PPEs): - It is found that usage of all mandatory PPEs is liberal. But the basic PPEs like helmet shoes etc. found strictly followed.
Barricading / Sign boards: - It is found that barricading is provided impractically, people try to select shortcuts due to the practical difficulties in barricading.
Housekeeping: - Generally it is weaker area in construction sites.

Handling of chemicals: - Health hazards, flammability, corrosion toxicity was considered. chemicals used at sites for cleaning, passivation, cutting also as raw materials, catalyst etc.

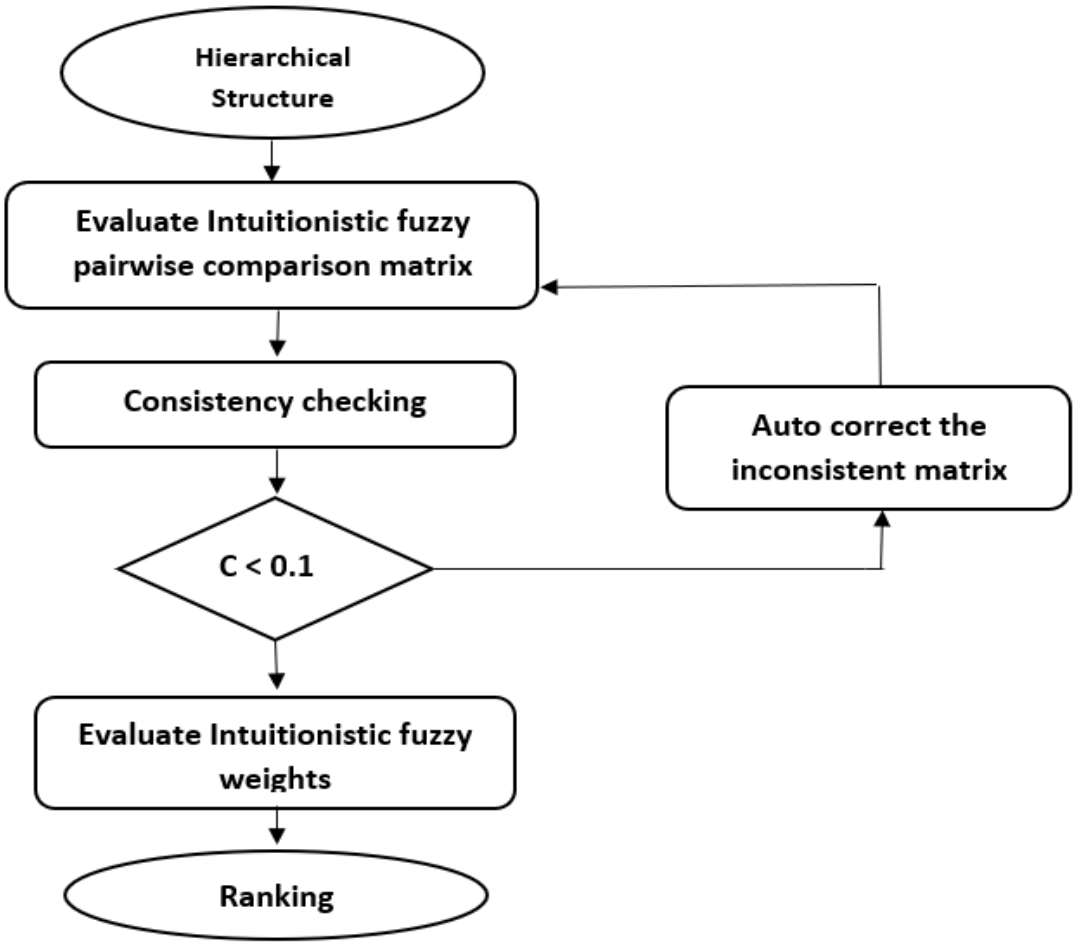

Fig 1. Schematic diagram of IFAHP

The pairwise comparison of factors under each criterion of severity, occurrences and detectability is evaluated as per the procedure explained.

Here the criterion of severity is illustrated.

Linguistic variables used in this work

Table 3. Linguistic scale for evaluation

Linguistic term	Symbol	IF value
Very high	VH	(0.7,0.1)
High	H	(0.55.0.25)
Equal	E	(0.4,0.4)
Low	L	(0.25,0.55)
Very low	VL	(0.1,0.7)

The intuitionistic preference relation under the criterion of severity for the primary factors are tabulated as below

Table 4. Intuitionistic preference relation under the criterion of severity for the primary factors

	Working at Height		Lifting & Moving		Confined Space		Electrical		Site Management	
Working at Height	0.4	0.4	0.25	0.55	0.55	0.25	0.55	0.25	0.7	0.1
Lifting & Moving Equip	0.55	0.25	0.4	0.4	0.55	0.25	0.55	0.25	0.7	0.1
Confined Space	0.25	0.55	0.25	0.55	0.4	0.4	0.4	0.4	0.55	0.25
Electrical	0.25	0.55	0.25	0.55	0.4	0.4	0.4	0.4	0.55	0.25
General Site Manage	0.1	0.7	0.1	0.7	0.25	0.55	0.25	0.55	0.4	0.4

Evaluation of perfect multiplicative consistent intuitionistic preference relation

$$\bar{\mu}_{pq} = \frac{\sqrt[q-p-1]{\prod_{r=p+1}^{q-1} \mu_{pr}\mu_{rq}}}{\sqrt[q-p-1]{\prod_{r=p+1}^{q-1} \mu_{pr}\mu_{rq}} + \sqrt[q-p-1]{\prod_{r=p+1}^{q-1} (1-\mu_{pr})(1-\mu_{rq})}}, \quad q > p+1$$

Let us consider $\bar{a}_{15}$

$\bar{\mu}_{15}$ = 0.545814

$\bar{\upsilon}_{pq}$ = 0.146296

Table 5. Perfect multiplicative consistent intuitionistic preference relation primary factors under severity

0.400	0.400	0.250	0.550	0.289	0.289	0.366	0.231	**0.546**	**0.146**
0.550	0.250	0.400	0.400	0.550	0.250	0.449	0.182	0.599	0.100
0.289	0.289	0.250	0.550	0.400	0.400	0.400	0.400	0.449	0.182
0.231	0.366	0.182	0.449	0.400	0.400	0.400	0.400	0.550	0.250
0.146	0.546	0.100	0.599	0.182	0.449	0.250	0.550	0.400	0.400

d (Ā,A) = $\frac{1}{2(n-1)(n-2)}\sum_{p=1}^{n}\sum_{q=1}^{n}\left(\left|\tilde{\mu}_{pq}-\mu_{pq}\right|+\left|\tilde{\upsilon}_{pq}-\upsilon_{pq}\right|+\left|\tilde{\pi}_{pq}-\pi_{pq}\right|\right)$

d (Ā, A) = 0.23 >0.1 so apply auto correction formulae

$$\tilde{\mu}_{pq}{}^{(t)} = \frac{(\mu_{pq}{}^{(t)})^{1-\sigma} \quad (\tilde{\mu}_{pq}{}^{(t)})^{\sigma}}{(\mu_{pq}{}^{(t)})^{1-\sigma} \quad (\tilde{\mu}_{pq}{}^{(t)})^{\sigma} + (1-(\mu_{pq}{}^{(t)})^{1-\sigma}(1- \quad (\tilde{\mu}_{pq}{}^{(t)})^{\sigma}}$$

Auto correction first iteration

Table 6. Perfect multiplicative consistent intuitionistic preference relation primary factors under severity

0.4	0.4	0.25	0.55	0.337	0.281	0.401	0.235	0.578	0.136
0.55	0.25	0.4	0.4	0.55	0.25	0.469	0.194	0.62	0.1
0.281	0.337	0.25	0.55	0.4	0.4	0.4	0.4	0.469	0.194
0.235	0.401	0.194	0.469	0.4	0.4	0.4	0.4	0.55	0.25
0.136	0.578	0.1	0.62	0.194	0.469	0.25	0.55	0.4	0.4

d (Ā, A) = 0.108254 >0.1

Auto correction second iteration

Table 7. Perfect multiplicative consistent intuitionistic preference relation primary factors under severity

0.400	0.400	0.250	0.550	0.327	0.283	0.394	0.234	0.572	0.138
0.550	0.250	0.400	0.400	0.550	0.250	0.465	0.192	0.616	0.100
0.283	0.327	0.250	0.550	0.400	0.400	0.400	0.400	0.465	0.192
0.234	0.394	0.192	0.465	0.400	0.400	0.400	0.400	0.550	0.250
0.138	0.572	0.100	0.616	0.192	0.465	0.250	0.550	0.400	0.400

d ($\bar{A}$, A) = 0.083895 < 0.1

Weight vectors for primary factors under severity calculated by

$$\omega_p = \left(\frac{\sum_{q=1}^{n} \mu_{pq}}{\sum_{p=1}^{n} \sum_{q=1}^{n} (1-\upsilon_{pq})}, \ 1 - \frac{\sum_{q=1}^{n} (1-\upsilon_{pq})}{\sum_{p=1}^{n} \sum_{q=1}^{n} \mu_{pq}} \right)$$

Table 8. Weight vectors for primary factors under severity

Primary factors under severity (sp)		μ	υ
Working at height	ω1	0.1228	0.6301
Lifting & moving equipment	ω2	0.1631	0.5850
Confined space	ω3	0.1136	0.6588
Electrical	ω4	0.1122	0.6632
General site management	ω5	0.0682	0.7388

Overall priority of dropping object from height under severity =

weight of working at height [sp (ω1)] x weight of dropping object from height [sh(ω3)]

= (0.1228,0.6301) x (0.1696,0.5682) = (0.0208,0.8403)

Overall priority dropping object from height under severity = (0.0208,0.8403)

Overall priority dropping object from height under occurrences = (0.255, 0.832)

Overall priority dropping object from height under detectability= (0.0139,0.873)

Risk priority number of dropping object from height = S x O x D

= (0.0208,0.8403) x (0.255, 0.832) x (0.0139,0.873) = (0.000007,0.99659)

Rank the intuitionistic fuzzy value= $\rho(\alpha) = 0.5\ (1+ \pi_{\alpha})(1- \mu_{\alpha})= 0.50169564$

The products of the risk priority numbers are greater than 0.5015 is considered as critical

The products of the risk priority numbers are in between 0.501 & 0.5015 is considered as major

The products of the risk priority numbers are less than 0.501 is considered as minor.

Table 9. Pair wise comparison under severity

	Working at Height		Lifting & Moving		Confined Space		Electrical		Site Management	
Working at Height	0.4	0.4	0.25	0.55	0.55	0.25	0.55	0.25	0.7	0.1
Lifting & Moving Equipment	0.55	0.25	0.4	0.4	0.55	0.25	0.55	0.25	0.7	0.1
Confined Space	0.25	0.55	0.25	0.55	0.4	0.4	0.4	0.4	0.55	0.25
Electrical	0.25	0.55	0.25	0.55	0.4	0.4	0.4	0.4	0.55	0.25
General Site Management	0.1	0.7	0.1	0.7	0.25	0.55	0.25	0.55	0.4	0.4

WORKING AT HEIGHT	scaffolding, ladders		falling from height		dropping objects		belt & protective gears	
scaffolding, ladders	0.4	0.4	0.4	0.4	0.25	0.55	0.55	0.25
falling from height, slips	0.4	0.4	0.4	0.4	0.25	0.55	0.7	0.1
dropping objects from height	0.55	0.25	0.55	0.25	0.4	0.4	0.7	0.1
safety belt & other protective gears	0.25	0.55	0.1	0.7	0.1	0.7	0.4	0.4

Lifting & Moving Equipment	lifting/moving		lifting equipment		moving equipment	
manual lifting/moving	0.4	0.4	0.25	0.55	0.25	0.55
lifting equipment	0.55	0.25	0.4	0.4	0.25	0.55
vehicle and other moving equipment	0.55	0.25	0.55	0.25	0.4	0.4

CONFINED SPACE	lighting & ventilations		standby / watch		simultaneous operations		administrative control	
lighting & ventilations	0.4	0.4	0.55	0.25	0.55	0.25	0.55	0.25
standby / watch	0.25	0.55	0.4	0.4	0.25	0.55	0.25	0.55
simultaneous operations inside	0.25	0.55	0.55	0.25	0.4	0.4	0.55	0.25
administrative control	0.25	0.55	0.55	0.25	0.25	0.55	0.4	0.4

ELECTRICAL	dragging of live cables		energization/LOTO		breaker/earthing		short circuit/insulation		structing lines	
dragging of live cables	0.4	0.4	0.25	0.55	0.25	0.55	0.25	0.55	0.1	0.7
partial energization of equip/LOTO	0.55	0.25	0.4	0.4	0.4	0.4	0.4	0.4	0.55	0.25
circuit breaker/earthing/extension	0.55	0.25	0.4	0.4	0.4	0.4	0.4	0.4	0.55	0.25
short circuit/insulation failure	0.55	0.25	0.4	0.4	0.4	0.4	0.4	0.4	0.55	0.25
striking Underground/Overhead lines	0.7	0.1	0.25	0.55	0.25	0.55	0.25	0.55	0.4	0.4

GENERAL SITE MANAGEMENT	work permits		usage of proper PPEs		Barricade/sign boards		House keeping		handling of chemicals	
work permits	0.4	0.4	0.25	0.55	0.25	0.55	0.25	0.55	0.55	0.25
usage of proper PPEs	0.55	0.25	0.4	0.4	0.55	0.25	0.55	0.25	0.55	0.25
Barricading/ sign boards	0.55	0.25	0.25	0.55	0.4	0.4	0.55	0.25	0.55	0.25
House keeping	0.55	0.25	0.25	0.55	0.25	0.55	0.4	0.4	0.55	0.25
handling of chemicals	0.25	0.55	0.25	0.55	0.25	0.55	0.25	0.55	0.4	0.4

Table 10. Pair wise comparison - occurrence

	Working at Height		Lifting & Moving		Confined Space		Electrical		Site Management	
WORKING AT HEIGHT	0.4	0.4	0.55	0.25	0.7	0.1	0.55	0.25	0.7	0.1
LIFTING & MOVING EQUIPMENT	0.25	0.55	0.4	0.4	0.7	0.1	0.7	0.1	0.7	0.1
CONFINED SPACE	0.1	0.7	0.1	0.7	0.4	0.4	0.55	0.25	0.25	0.55
ELECTRICAL	0.25	0.55	0.1	0.7	0.25	0.55	0.4	0.4	0.25	0.55
GENERAL SITE MANAGEMENT	0.1	0.7	0.1	0.7	0.55	0.25	0.55	0.25	0.4	0.4

WORKING AT HEIGHT	scaffolding, ladders		falling from height		dropping objects		belt & protective gears	
scaffolding, ladders	0.4	0.4	0.4	0.4	0.1	0.7	0.55	0.25
falling from height, slips	0.4	0.4	0.4	0.4	0.25	0.55	0.7	0.1
dropping objects from height	0.7	0.1	0.55	0.25	0.4	0.4	0.7	0.1
safety belt & other protective gears	0.25	0.55	0.1	0.7	0.1	0.7	0.4	0.4

LIFTING & MOVING EQUIPMENT	lifting/moving		lifting equipment		moving equipment	
manual lifting/moving	0.4	0.4	0.7	0.1	0.55	0.25
lifting equipment	0.1	0.7	0.4	0.4	0.25	0.55
vehicle and other moving equipment	0.25	0.55	0.55	0.25	0.4	0.4

CONFINED SPACE	lighting & ventilations		standby / watch		simultaneous operations		administrative control	
lighting & ventilations	0.4	0.4	0.55	0.25	0.55	0.25	0.55	0.25
standby / watch	0.25	0.55	0.4	0.4	0.4	0.4	0.25	0.55
simultaneous operations inside	0.25	0.55	0.4	0.4	0.4	0.4	0.55	0.25
administrative control	0.25	0.55	0.55	0.25	0.25	0.55	0.4	0.4

ELECTRICAL	dragging of live cables		energization/LOTO		breaker/earthing		short circuit/insulation		striking lines	
dragging of live cables	0.4	0.4	0.55	0.25	0.25	0.55	0.25	0.55	0.7	0.1
partial energization of equipment/LOTO	0.25	0.55	0.4	0.4	0.25	0.55	0.4	0.4	0.55	0.25
circuit breaker/earthing/extension	0.55	0.25	0.55	0.25	0.4	0.4	0.4	0.4	0.55	0.25
short circuit/insulation failure	0.55	0.25	0.4	0.4	0.4	0.4	0.4	0.4	0.55	0.25
striking Underground/Overhead lines	0.1	0.7	0.25	0.55	0.25	0.55	0.25	0.55	0.4	0.4

GENERAL SITE MANAGEMENT	work permits		usage of proper PPEs		Barricade/sign boards		House keeping		handling of chemicals	
work permits	0.4	0.4	0.25	0.55	0.25	0.55	0.25	0.55	0.55	0.25
usage of proper PPEs	0.55	0.25	0.4	0.4	0.4	0.4	0.55	0.25	0.55	0.25
Barricading/ sign boards	0.55	0.25	0.4	0.4	0.4	0.4	0.55	0.25	0.55	0.25
House keeping	0.55	0.25	0.25	0.55	0.25	0.55	0.4	0.4	0.55	0.25
handling of chemicals	0.25	0.55	0.25	0.55	0.25	0.55	0.25	0.55	0.4	0.4

Table 11. Pair wise comparison - detection

	Working at Height		Lifting & Moving		Confined Space		Electrical		Site Management	
WORKING AT HEIGHT	0.4	0.4	0.4	0.4	0.25	0.55	0.4	0.4	0.4	0.4
LIFTING & MOVING EQUIPMENT	0.4	0.4	0.4	0.4	0.25	0.55	0.4	0.4	0.55	0.25
CONFINED SPACE	0.55	0.25	0.55	0.25	0.4	0.4	0.25	0.55	0.55	0.25
ELECTRICAL	0.4	0.4	0.4	0.4	0.55	0.25	0.4	0.4	0.4	0.4
GENERAL SITE MANAGEMENT	0.4	0.4	0.1	0.7	0.25	0.55	0.4	0.4	0.4	0.4

WORKING AT HEIGHT	scaffolding, ladders		falling from height		dropping objects		belt & protective gears	
scaffolding, ladders	0.4	0.4	0.25	0.55	0.25	0.55	0.4	0.4
falling from height, slips	0.55	0.25	0.4	0.4	0.4	0.4	0.55	0.25
dropping objects from height	0.55	0.25	0.4	0.4	0.4	0.4	0.55	0.25
safety belt & other protective gears	0.25	0.55	0.25	0.55	0.25	0.55	0.4	0.4

LIFTING & MOVING EQUIPMENT	lifting/moving		lifting equipment		moving equipment	
manual lifting/moving	0.4	0.4	0.55	0.25	0.55	0.25
lifting equipment	0.25	0.55	0.4	0.4	0.4	0.4
vehicle and other moving equipment	0.25	0.55	0.4	0.4	0.4	0.4

CONFINED SPACE	lighting & ventilations		standby / watch		simultaneous operations		administrative control	
lighting & ventilations	0.4	0.4	0.55	0.25	0.55	0.25	0.55	0.25
standby / watch	0.25	0.55	0.4	0.4	0.25	0.55	0.55	0.25
simultaneous operations inside	0.25	0.55	0.55	0.25	0.4	0.4	0.55	0.25
administrative control	0.25	0.55	0.25	0.55	0.25	0.55	0.4	0.4

ELECTRICAL	dragging of live cables		energization/LOTO		breaker/earthing		short circuit/insulation		striking lines	
dragging of live cables	0.4	0.4	0.25	0.55	0.25	0.55	0.25	0.55	0.4	0.4
energization of equipment/LOTO	0.55	0.25	0.4	0.4	0.55	0.25	0.55	0.25	0.55	0.25
circuit breaker/earthing/extension	0.55	0.25	0.25	0.55	0.4	0.4	0.4	0.4	0.55	0.25
short circuit/insulation failure	0.55	0.25	0.25	0.55	0.4	0.4	0.4	0.4	0.55	0.25
striking Underground/Overhead lines	0.4	0.4	0.25	0.55	0.25	0.55	0.25	0.55	0.4	0.4

GENERAL SITE MANAGEMENT	work permits		usage of proper PPEs		Barricade/sign boards		House keeping		handling of chemicals	
work permits	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.25	0.55
usage of proper PPEs	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Barricading/ sign boards	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.25	0.55
House keeping	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.25	0.55
handling of chemicals	0.55	0.25	0.4	0.4	0.55	0.25	0.55	0.25	0.4	0.4

Table 12. Ranks of risk factors

	RANKING OF RISK FACTORS		severity			occurrence			detectability						RPN RANK		Category
	WORKING AT HEIGHT		μ	υ		μ	υ		μ	υ		μ	υ		ρ		
1	scaffolding, ladders		0.01428	0.86294		0.01674	0.85816		0.00866	0.89508		0.00000207	0.99796		0.50101781	6	Major
2	falling from height, slips		0.01486	0.86649		0.01796	0.86323		0.01357	0.87443		0.00000362	0.99771		0.50114294	3	Major
3	dropping objects from height		0.02082	0.84027		0.02553	0.83263		0.01393	0.87260		0.00000741	0.99659		0.50169564	1	Critical
4	safety belt & other protective gears		0.00874	0.89983		0.02271	0.84004		0.00941	0.89761		0.00000187	0.99836		0.50081845	9	Minor
	LIFTING & MOVING EQUIPMENT																
5	manual lifting/moving		0.01275	0.88115		0.02599	0.82185		0.01355	0.87492		0.00000449	0.99735		0.50131971	2	Major
6	lifting equipment		0.01734	0.86489		0.01266	0.88819		0.01017	0.89851		0.00000223	0.99847		0.50076436	10	Minor
7	vehicle and other moving equipment		0.02331	0.84120		0.01873	0.85974		0.01029	0.89828		0.00000449	0.99773		0.50112829	4	Major
	CONFINED SPACE																
8	lighting & ventilations		0.01559	0.86425		0.01284	0.88738		0.01737	0.85478		0.00000348	0.99778		0.50110664	5	Major
9	stand by / watch		0.00998	0.89817		0.00958	0.90816		0.01166	0.88795		0.00000112	0.99895		0.50052284	18	Minor
10	simultaneous operations inside		0.01595	0.86717		0.01101	0.90119		0.01666	0.86302		0.00000292	0.99820		0.50089596	7	Minor
11	administrative control		0.01121	0.89149		0.00812	0.91614		0.01090	0.89479		0.00000099	0.99904		0.5004777	20	Minor
	ELECTRICAL																
12	dragging of live cables		0.00906	0.89932		0.00780	0.91569		0.01377	0.86780		0.00000097	0.99888		0.5005601	16	Minor
13	energization of equipment / LOTO		0.01510	0.87300		0.00598	0.93136		0.02390	0.83422		0.00000216	0.99855		0.50072041	14	Minor
14	circuit breaker/earthing/extension		0.01505	0.87374		0.00852	0.91555		0.01722	0.86115		0.00000221	0.99852		0.50073805	13	Minor
15	short circuit/insulation failure		0.01510	0.87300		0.00846	0.91616		0.01810	0.85786		0.00000231	0.99849		0.5007544	11	Minor
16	strickling Under ground/Over head lines		0.01009	0.89390		0.00515	0.93307		0.01182	0.88248		0.00000061	0.99917		0.50041668	21	Minor
	GENERAL SITE MANAGEMENT																
17	work permits		0.00664	0.91118		0.00889	0.89444		0.01181	0.87644		0.00000070	0.99884		0.50057854	15	Minor
18	usage of proper PPEs		0.01155	0.88970		0.01428	0.87170		0.01227	0.88023		0.00000202	0.99831		0.50084545	8	Minor
19	Barricading/ sign boards		0.00912	0.90299		0.01464	0.87053		0.01214	0.88118		0.00000162	0.99851		0.50074456	12	Minor
20	House keeping		0.00708	0.91652		0.01052	0.89136		0.01200	0.88144		0.00000089	0.99892		0.50053673	17	Minor
21	handling of chemicals		0.00502	0.92752		0.00734	0.90757		0.01615	0.85310		0.00000060	0.99902		0.50049147	19	Minor

7. Conclusions

Construction sector are the largest economic activity in the current scenario. Ensuring safety in this sector contribute smooth functioning of the projects sites. Since it is more human intensive and carried out through unorganised sectors and it is influenced by regional cultures. It inherently has uncertainties and vagueness. The intuitionistic fuzzy analytic hierarchy process (IF-AHP) has been proposed for the risk assessment and ranking of safety factors during construction phase of a chemical complex. The IF-AHP structure is very easy to understand and the pair-wise comparison between the factor scales the differences. It deals successfully the vagueness and uncertainties.

In this study critical activities in the construction site is that “dropping of objects from heights” the activities have major consequences are “manual lifting and moving”, “falling from height and slips “,” vehicle and other moving equipment”,” lighting and ventilations in the confined spaces”, all others factors are found to be minor.

7.1 Limitations

1. It is influenced by regional standards, practices, and organizational cultures.
2. Results cannot be generalized.
3. It is a perception-based assessment.

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Development of Resilient LNG Facilities

Onder Akinci*[1], Andrew Staszak[2], Hyun-Su Kim[1], Connor Rivard, Michael Stahl[2]
[1]Daros Consulting
[2]SciRisq Inc.
onder.akinci@darosconsulting.com

Abstract

Demand for natural gas and increased production in the US enabled development of new LNG export facilities. In parallel to development of new facilities, federal and local codes in the US have evolved significantly in the last decade. Thus, today's standards require design and construction of robust facilities. Depending on the location of LNG plant, effects of hurricanes, earthquakes, hydrocarbon accidents or floods are typically considered in the design. This study examines evolution of resilient design requirements and best practices for LNG export terminals. Case studies are presented to demonstrate impact of resilient design requirements on design of a facility.

LNG export terminals are considered to be some of the safest hydrocarbon processing facilities in the industry. Global experience indicates that LNG facilities can be successfully constructed and safely operated in some of the harshest environments close to the arctic circle, near the flood zones, on an offshore vessel or in high seismicity regions. Robust design of plant structures as per current standards and mature liquefaction technologies ensure that LNG terminals are capable of withstanding major accidents or natural events. Additionally, Facility Siting and Consequence Analysis for LNG plants require design and process safety measures against fires and explosions.

When these onerous design cases are compared with those for typical upstream and downstream facilities or civil infrastructure, it can be further inferred that risks to public are generally smaller for LNG plants. Plants designed according to current standards using state of the art tools are expected to be resilient against natural hazards and accident events.

Keywords: LNG, Resilience, Inherent Safety, Facility Siting, Blast, Fire, Cryogenic, PFP, Facility Siting, Safety Critical Elements

1. Introduction

Resilience for Oil&Gas facilities can be defined as the capability to recover from or adjust easily to extreme events or changes beyond typical design conditions. Considering the importance of Oil and Gas infrastructure and assets, it is imperative to have a better understanding of resilience in the context of process safety and operations. Resilience expectations and requirements for designated critical infrastructure sectors including the energy sector were also captured in Presidential Policy Directive (PPD-21) Critical Infrastructure Security and Resilience [1]. Roles and responsibilities of owners and the government agencies were listed in PPD-21 to ensure advancement of a national unity of effort to strengthen and maintain secure, functioning, and resilient critical infrastructure.

Design codes and standards developed by industry organizations including ASME, API, NFPA and ASCE provide provisions for conventional and extreme event design of equipment and structures at the facilities. These codes (or recommended practices) also cover design for extreme events such as high magnitude earthquakes, floods, fires, explosions, line break and overpressurization scenarios. These code provisions are typically intended for protection of life and asset. However, the plant may not be fully functional after a major event and in some cases, it may take months or years to return to service. There are not widely accepted guidelines in the industry for resilient design of LNG plants. Use of best practices, provision of extra design margins and lessons learned from previous projects are regarded as mitigation measures to resist extreme events. However, these types of measures do not provide the assurance required to confirm resiliency of a facility.

To design a resilient facility, the applicable risks and response limits should be established. Additionally, the level of risk tolerance and design objectives shall be laid out at an early stage. It is understood that expecting no or minimal damage after any applicable catastrophic event may not be feasible. Guidelines and recommendations are presented in this study for development of LNG facilities in a practical way. The following steps are discussed for resilient design:

- Identify major risks and consequences
- Design and implement barriers to mitigate risks
- Evaluate benefits of barriers and test against scenarios beyond the design basis to understand failure mechanisms
- Quantify the limits states and return periods for governing cases

The steps can be integrated with or added as a design iteration after designing the facility for conventional loads.

2. Process Safety Risks

The assessment and quantification of process safety risks throughout the energy industry is considered a well understood topic. Extensive reviews, papers, analyses, regulations, and guidance

document development have been conducted over the years and has covered most aspects of the topic. The focus on LNG specific risk has only recently become a more significant focus, likely spurred by the significant increase of LNG demand, which in turn has spurred an increase in LNG project developments.

In general, the basic processes of identifying hazards, quantifying their extents, and developing frequencies and probability of events are no different for LNG facilities. There are numerous papers and resources which cover these topics [2 to 5]. A major difference is due to the need to address cryogenic hazards facility wide. There are other significant differences of the elements which should be considered as standard practice or at a minimum addressed qualitatively in the assessment of process safety risks for resilient LNG design.

Before we proceed, the topic of criteria must be addressed. As noted in the introduction, general risk guidelines and approaches focus on the safety of personnel but allow significant leniency in protection of the facility. This is not unintentional nor incorrect, but resilient LNG facility design requires a much stronger focus on the protection of assets. This focus leads to a number of distinguishing elements when addressing process safety risk and will be the focus of this section.

In an effort to review these critical differences and outline elements for consideration, we will summarize risk simply as the product of *Likelihood* and *Consequence*. Students of risk analysis acknowledge that risk calculation is much more complex, but the focus of this paper is on a higher level than nuanced and detailed calculations. Accepting our simplified model, we can examine representations of *Likelihood* and *Consequence* separately and discuss those elements that are relevant to the design of resilient LNG facilities.

2.1 Elements of Consequence

When evaluating consequences, a number of critical topics can be identified which should be addressed during evaluation of process safety risk. The following six elements were chosen for discussion as they have been found, in practice, to contribute more significant influence over the results and ultimately to design decision making. Other differences and elements may be critical to specific LNG facilities, but the below list can be applied to the majority of LNG facility types.

2.1.1 Multi-phase Physics

In any robust risk assessment, the multi-phase potential of releases would be addressed and that is no different in LNG facilities. The difference lies in the types of hazards needed to be addressed. Typically, jet/spray fires along with rain out and liquid spill resulting in pool fires should be addressed, but additional consideration should be given to cryogenic extents. Within pool spills this can be somewhat trivial but within 2-phase sprays determining the extents of impact areas can be more challenging. In order to provide assessments of impacts to facility assets and determine protection levels, these consequences need to be determined. Thus, more robust modelling platforms or calculation approaches should be considered which can predict or estimate liquid and vapor contents within spray plumes as well as temperature differences. Figure *1* provides an example comparison between conventional modelling versus consideration for multi-phase releases in modelling. The difference results in a significant shift in hazard exposure extents and has a direct effect on risk results.

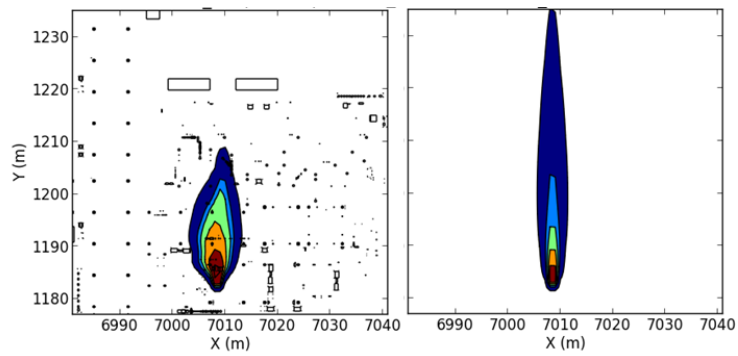

Figure 1. Example of cold vapor/spray modelling differences (right- expanded consideration for material phase, left- simple approach).

2.1.2 Transient Conditions

For many facility types an assessment of consequences based on steady state conditions is acceptable. This is particularly true in facilities where the focus is solely on personnel safety, as ignoring transient release conditions produces conservative results. As noted, ensuring design evaluation of assets requires more detailed information. When designing equipment (valves, co-located vessels, structures) the determination of load duration is critical. Whether that load is thermal radiation or cryogenic exposure, a steady state (or indefinite) duration is unacceptable. Thus, when designing resilient LNG facilities, consideration must be given to the duration of events. This can be addressed conservatively based on initial release rates and available inventories, but for larger and more complex facilities it is recommend that this be addressed through the development of release rate curves. Release rate curves can be developed by considering isolation timings, blowdown and relief systems, and connected inventory, providing variations in release rate over time. Consequences then can be determined at various critical times. These critical times would depend on facility specific protection approaches and goals. Figure *2* [6], shows the direct influence of transient considerations on jet fire size. Neglecting these changes in a hazard generally results in a conservative assessment but has also been found to lead to incorrectly influenced cost-benefit design decisions.

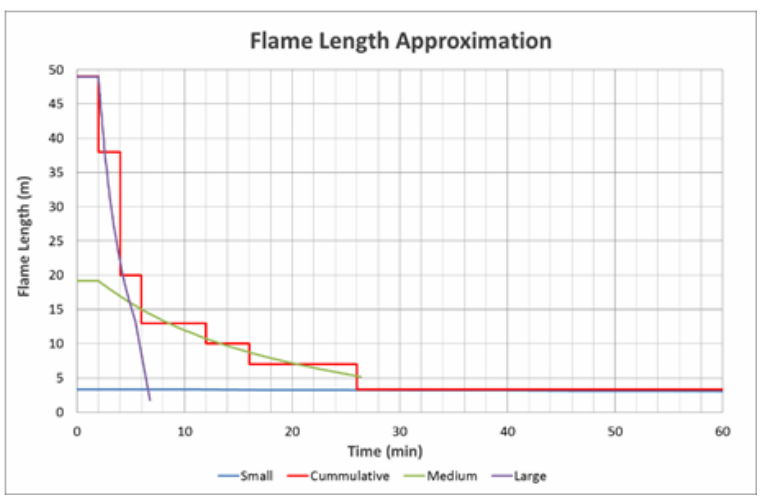

Figure 2. Example of transient release modelling and the effect on fire size [6].

2.1.3 Physical Barrier Influences

Essentially all of the LNG specific elements discussed here are connected, thus highlighting their criticality in assessment of a resilient facility. Physical barrier influences are an explicit example of that, as they can play a major role in the transient effects of consequences. In general, for process safety risk assessments, a majority of the consequence assessment will be based on phenomenological modelling. These are usually open-field calculations, where adjustments can be made to account for phenomena such as dispersion limitations for indoor releases or pool spread due to bunding. In continuing with the theme of an analysis that can assess or fully influence the design of assets, risks should consider more elements of physical barrier influences. Examples of these would be the use of extensive curbing and drain-ways in LNG facilities and their effect on pool formation and duration, the use of flange guards to limit cryogenic spray and similar events.

2.1.4 Active Mitigation

A resilient facility will utilize active mitigation systems both to limit process hazards from occurring (e.g. process controls, blowdown, redundancies) and limit consequences resulting from process failures (e.g. isolation, foam, water spray). In order to design these systems, they must be assessed and/or their effect accounted for within process safety risk. There are an extensive number of approaches which can be used to address these types of elements, which will not be focused on here, but the approach should be dependent on the whole approach of the risk assessment (introduced error and level of assumptions and simplifications).

2.1.5 Performance and Optimization

As discussed above, active mitigation and control is critical to the design of a resilient facility and are typical of LNG facilities. Asset and general risk can be greatly influenced (reduced) through the use of optimized process safety controls. This encompasses systems such as isolation, shutdown, blowdown, detection, and other active mitigations. Assessment approach must consider or provide means to evaluate their effectiveness. Through these evaluations both performance verification and optimization of the system design can be conducted. It is important to note, that in many cases (facility and overall design philosophy dependent) these elements may be addressed through additional studies and analysis. Where the supporting data is taken from the process safety risk assessment.

2.1.6 Receptor and Equipment Response

Specific to asset protection as it relates to process safety, elements of receptors and equipment should include a more detailed or robust examination. Examples of these are critical valve survivability and structural integrity of piperacks and process module decks. If process controls are being used for a resilient design, the risk assessment should not only account for their operation in the calculations but should also provide an assessment (and design criteria as necessary) to ensure that those elements will survive and can perform their function. Additionally, this type of analysis is required to assess the risks of escalation events. Consequences must be determined in a way that allows for escalation prediction. Figure *3* shows an example of geometric effects on LNG vapor cloud explosion events. When evaluating and determining the design limit or performance requirements of these critical receptors, those conducting the risk assessment should be able to identify when more complex modelling is required. In cases such as that shown below, critical receptors

located in “U” shaped portion of the structure below, could be under-evaluated with simple modelling approaches. This would result in systems, unable to perform their intended functions following an initiating event.

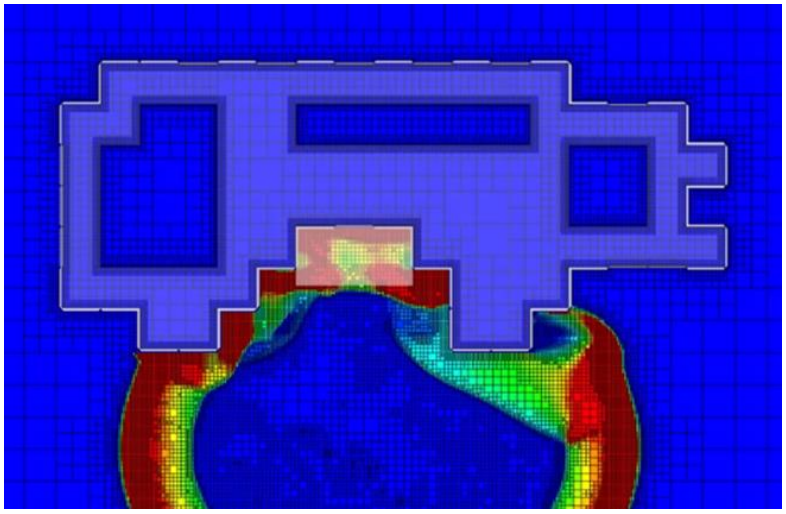

Figure 3. Example of LNG vapor cloud explosion effects including complex geometric effects.

2.2 Elements of Likelihood

In a similar approach to *Consequence* evaluation, a number of critical elements specific to *Frequency* evaluation of process safety risk and resilient LNG facility design can be identified. Six critical elements have been highlighted for discussion. It is important to note, the interconnectivity between these and the *Consequence* elements discussed above. Generally, for any expansion or complexity added to consequence quantification in a risk assessment an equal expansion or complexity must be added in the frequency quantification to address those (or the reverse).

2.2.1 Mechanical Integrity

Mechanical integrity (MI) programs in general are growing in use and application across the energy industry. It has been proven as a statistical driver in reducing process failures and thus reducing process safety risk. Resilient LNG facility operations, particularly those projects conducted in the last half decade have been leaders in this area. Thus, it is important to include the effects of mechanical integrity in the assessment of risks. This can be accomplished through modifiers in release frequencies, increased probabilities of safety system success, etc. Depending on the MI system and details available, these modifications to the assessment may only be applied to specific areas of the facility.

2.2.2 Cryogenic Design Standards

The basis of the frequency side of the risk equation is built on historical release data. This data is an amalgamation of equipment specifications and types from all over the world and from all kinds of processes (though separated into equipment types). Equipment specific to LNG processes can and do have higher design standards and lower tolerances. This should not be neglected in assessing process safety risks. This can be addressed in assessment through the use of specialized leak frequency databases (although statistical

relevance of samples can be questionable) or most commonly through the use of modifiers. In sections of equipment with known high design standards, modifiers can be applied to existing databases or leak frequency results. Development and justification of these LNG specific modifiers is an area of needed academic evaluation, but the use of leak frequency modifiers for other processes has been commonly used in risk assessment. Thus, employing good engineering practices would be considered an acceptable approach.

2.2.3 Survivability and Impairment

As discussed in the *Consequence* section, a resilient LNG facility design and assessment of risks will encompass numerous survivability and impairment metrics for structures, critical receptors, and safety systems. These must be addressed on the Frequency side of the equation. Thus, consideration should be given to the probability of success for active systems, probability of impairment at given consequences, probability of escalation of equipment given that impairment. It is recommended that these elements be grouped and addressed in broad terms if singling out of elements is not required. Introduction of too many differentiating probabilities greatly increase the computational requirements and can make identification of risk drivers more complicated. An example of this is LNG tank fires. In recent years, regulatory agencies have stressed that facilities must fully address the risk related to these events. In these cases, advanced modelling can be conducted to inform the consequences, but also to aid in verifying performance and selection of timings, probabilities of impairment for use in the *Frequency* side of the equation. Figure *4*, provides an example of modelling changes in an LNG tank fire over time and which are then used to predict the timing of impairment.

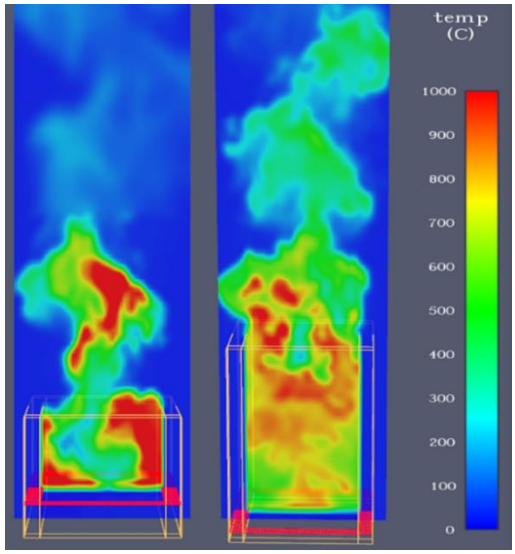

Figure 4. Example of computational modelling conducted to provide performance verification and timing of impairment.

2.2.4 Event Probabilities and Deviations

Similarly, as noted above for survivability and impairment, event probabilities should address the specific consequences developed for LNG facilities. This includes but is not limited to cryogenic hazard endpoints and transient changes or durations of consequences. In most cases, this requires the development and use of event trees specific to LNG

operations and processes. The look and complexity of these event trees will be entirely dependent on the module, system, design features, etc. In some cases, a single set of LNG process event trees may be acceptable, but in other cases they may need to vary by process area or module.

2.2.5 Ignition

Traditional process safety risk can address ignition through a number of approaches. There are complex source grading and point based systems, but generally release rate models are employed. There are a number of release rate models, some of which are applied as a singular function across all process areas and those that use numerous functions for varying equipment types, process module locations, and size. The latter is generally recommended for the design and assessment of resilient LNG facilities, but with additional considerations. Modifiers may be applied or used for areas with integrated process safety controls and active response systems. Additionally, modifiers may be applied for cryogenic releases which require more explicit (open flame) or higher energy ignition sources. It should be noted that cryogenic releases can undergo a number of temperature changes, thus caution should be used in applying these modifiers.

2.2.6 Safety System Intervention

As noted in both the discussion of *Consequence* with regards to Active Mitigation and Performance and Optimization, resilient LNG facility designs utilize progressive safety system interventions. These encompass process safety controls to limit the origination of events and active systems to limit the effects of events. An example of two LNG specific systems are water spray and foam deployment for vapor dispersion and fire mitigation. These elements are critical to design and operation and must be addressed in the assessment of process safety risks. Developing appropriate probabilities of success and failure of these systems should be included in any process safety risk assessment. In many cases, the performance of these systems can be difficult to identify with analysis. For example, fire and gas detections systems can be the main drivers of intervention. Using performance verification assessments, such as shown in Figure 5, can provide the quantitative basis needed to calculate the *Frequency* side of the risk equation.

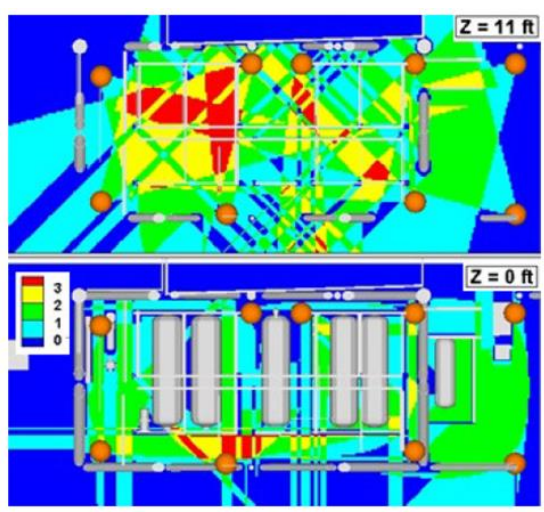

Figure 5. Example of fire and gas mapping to determine performance probabilities.

3. Facility Siting

Hazards associated with location of process plant permanent and temporary buildings are evaluated according to API 752 [7] and 753 [8] for onshore facilities in the US. These standards were mainly intended for petrochemical facilities. Some LNG plants follow API 752 to locate the occupied buildings outside the fire and blast zones. However, applicability of these facility siting standards to LNG facilities can raise questions. Current framework for facility siting of LNG plants account for risk to public outside the plant boundaries. However, risks to plant buildings and personnel are critical elements of resilient design. It is recommended to perform facility siting studies for LNG plants. Location and blast (and fire) rating of buildings can then be specified accordingly. This can also be integrated with a comprehensive risk-based approach (QRA) and checked against the applicable risk criteria as discussed in NFPA 59A-2019 [9].

Facility siting standards and recommended practices target improving safety of public and plant personnel. For the case of unoccupied buildings such as equipment rooms or process modules, facility siting guidelines offer neither mitigation measures nor require protection against extreme loads. This critical aspect of resilient design is left to the plant owner and the approach can vary significantly based on the operator's experience and risk criteria. As noted, a robust QRA addressing all aspects of risk (process safety, environmental, siting, etc.) is the most effective form of evaluation. That analysis must address the specifics of LNG facilities and processes, such as those discussed in this paper, otherwise the result would unlikely meet the requirements of a "resilient LNG facility design."

4. Environmental Risks

Resilience of infrastructure against environmental loads has been studied extensively, but not all the industries have participated in these efforts. Flooding of Energy and Oil Facilities during the storms of last decade allude to areas of improvement in terms of resilient design against the environmental conditions. Design of a major petrochemical plant beyond the requirements of conventional design code allowables depend on owner's design criteria. Also, changes in design codes and criteria over time can be a critical limitation for aging plants. Failures at major LNG facilities due to natural events are very rare and not documented well, but implementation of lessons learned from previous events at other types of facilities provide valuable information about potential vulnerabilities [10].

Conventional design load cases for LNG facilities may include earthquake, wind, flood and tsunami depending on the location of facility as discussed in NFPA 59A [9] and ASCE 7 [11]. These codes provide the applicable load cases and refer to design codes from ACI, API, AISC and other industry organizations. For some cases, government agencies or owners specify additional load cases such as wind-borne debris and blast projectile impact [12]. The main objective of design against environmental loads is protecting safety of the public and staff on site. This is achieved by

checking capacity of structural systems and barriers against specified loads. Some examples include Operating Basis and Safe Shutdown earthquakes or rare hurricanes.

The natural hazards have varying levels of return periods. For less frequent events the allowable design limits might be higher whereas for low return period events more stringent design criteria applies. It is believed that use of lower allowable stresses limits the damage due to a more frequent event and results in a resilient design. However, the inherent safety factors from design codes does not always ensure resilience. It is recommended to assess and quantify the level of damage to critical systems due to credible events so that plant can return to service after a short outage period.

If production at a facility depends on external or an independent power generation unit, design basis and reliability of those systems need to match the plant's availability targets. For a hypothetical case, some elements of the power system could be designed for earthquake or wind loads corresponding to a relatively lower return period. The plant may then lose power if the design basis of power supply or generation system is exceeded, and it can take a considerable time to restore production operations.

Combining the actual environmental risks with process safety risks can give a more refined risk profile for the plant's safety and availability. Thus, the original design or upgrades can account for resiliency in a more structured and quantified way. This approach can help owners, investors, buyers, and insurers better understand the plant's characteristics.

5. LNG Transportation Risks

In many cases produced LNG is required to be transported to other areas where the consumption and demand is high. Transportation plays a critical role in successful LNG operations. More efficient transportation methods for LNG has been long sought since the liquefied gas required special facilities for safe delivery and protection against losses due to regasification. It is well established to transport LNG via a large ship with LNG cargo containment system, LNG carrier. Recently, floating units with LNG production, storage, and offloading (LNG FPSO) or with LNG storage and regasification facilities (FSRU) have been newly constructed or converted from existing gas carriers, then installed at nearshore or offshore sites. Several international and national codes and standards apply transportation of LNG to ensure safety of the vessel, terminals, and the communities nearby the shipping channels. Additionally, these provisions ensure high operability rates and rapid recovery from adverse effects.

For design and construction of LNG carrying ships, it is mandatory to conform to International Liquefied Gas Carriers Code (IGC Code) under the SOLAS convention in 1974 [13]. LNG FPSO and FSRU are also recommended to be designed following the IGC Code. Classification societies, i.e., ABS, BV, DNVGL, provide specific design guides in their steel ship design rules for new construction of LNG carriers while prescribing classification rules for design, construction, and inspection of offshore LNG FPSO. For LNG carriers that are operated in North America, ship management procedures and emergency plans are required to be submitted to and approved by the

US Coast Guard. ISO 28460 relates to marine operations during LNG carrier's port transit and the cargo transfer at the ship-shore interface [14].

International Maritime Organization (IMO) classifies LNG carriers together with their cargo containment system (CCS), which consists of insulated cargo tanks inside the inner hull. The LNG carrier types by CCS are presented in Figure 6 at a high level. It is worth noting that in the last decade, the membrane type (No96 and Mark III) for LNG carrier's CCS became more popular to maximize the cargo volume. For all types, the liquefied gas carrying ship is constructed to adopt double hull structures that provide hold spaces such as the ballast tank or cofferdam between the inner hull and outer shell. The hold spaces between the inner hull and outer shell are typically required to have 4 to 6 ft widths for protection of the cargo tanks in case of ship side-on collision and grounding accidents.

To design resilient floating LNG (LNG FPSO, FSRU, etc.) and LNG carriers several technical challenges have taken to be considered. These include sloshing impact to cargo containment system, structural integrity under accidental collision events, cryogenic spill induced material brittle failure, fire risk escalated by flammable gas formation due to spillage of cryogenic liquid as the temperature rises, and explosion risk due to rapid expansion of the liquid-to-gas volume. In the past couple of decades, many studies have been carried out on sloshing response and collision behaviour of the gas carrying or storing facilities. The main defence mechanism for the accidental events should be eliminating the risk and the safeguards should be implemented to mitigate damage to safety critical systems. This in turn would ensure continuation of service or limited damage after an event.

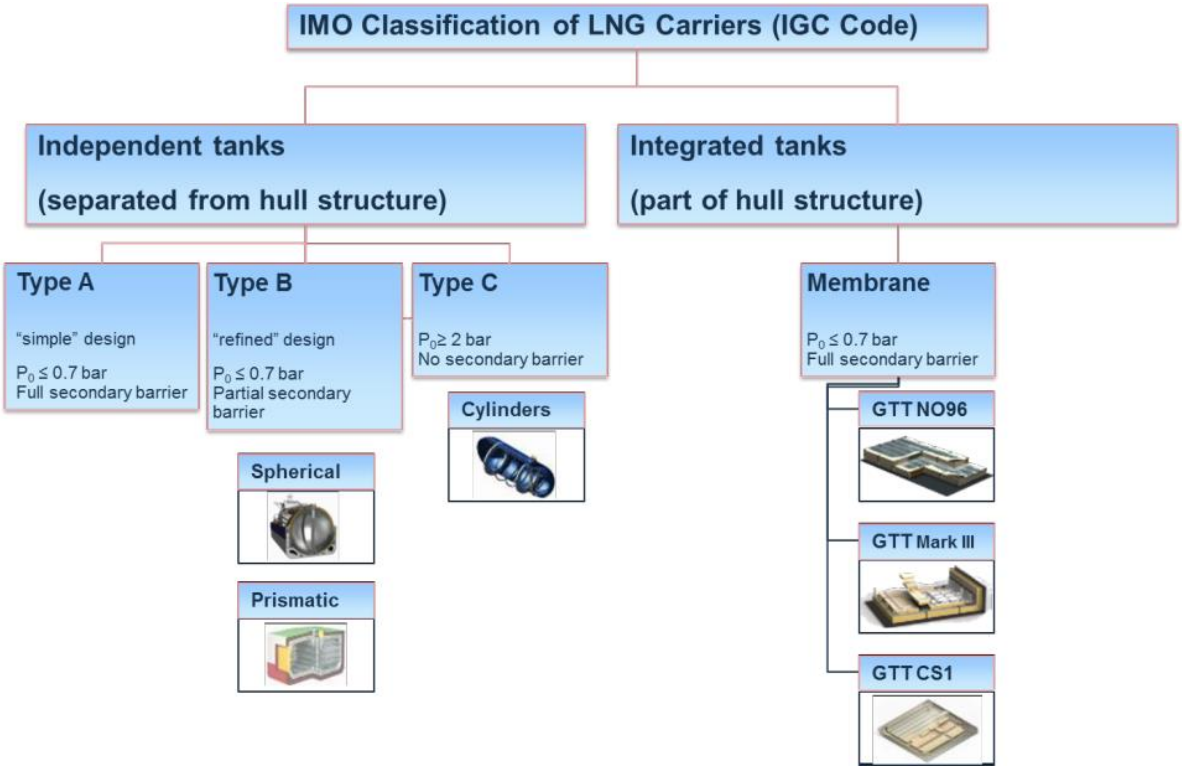

Figure 6. LNG carrier types by cargo containment system [15].

A previous study investigated application of two row LNG membrane-type CCS configuration during development of a Floating LNG facility. This study showed advantages of central cofferdam structure to withstand topside module loads and higher liquid filling level to reduce sloshing impact [16]. A numerical method for evaluation of the ultimate impact strength of membrane-type LNG containment system under a sloshing impact load was proposed by H. Lee et al. [17]. Additionally, J.M. Sohn et al. [18] investigated a Mark III (membrane) type LNG CCS subjected to sloshing impact pressure by numerical dynamic analysis using nonlinear finite element method and concluded that the membrane-type LNG cargo container can withstand 15 bar or greater sloshing impact loads without loss of structural integrity. Sloshing assessment guidance for membrane-type LNG cargo containment system is provided by major classification societies including Bureau Veritas [19], DNV GL [20], and Lloyds Register [21].

A range of maximum credible failure cases for LNG terminals due to accidental events was summarized by R.M. Pitblado et al. [22]. Also, there are studies suggesting that a spherical (Moss) type gas carrier can withstand a side-on collision, without breaching the LNG cargo area, due to the same size LNG carrier at 6.6 knots speed or fully loaded VLCC at 1.7 knots speed. H. Bogaert and B. Boon [23] introduced the Maritime Collision Model, MARCOL, for assessing the penetration probability of LNG tanks at LNG terminals. S.R. Cho et al. [24] implemented case studies for dynamic collision analyses with soft bow design options, as shown in Figure 7, to minimize damage to LNG cargo compartment of membrane type carrier.

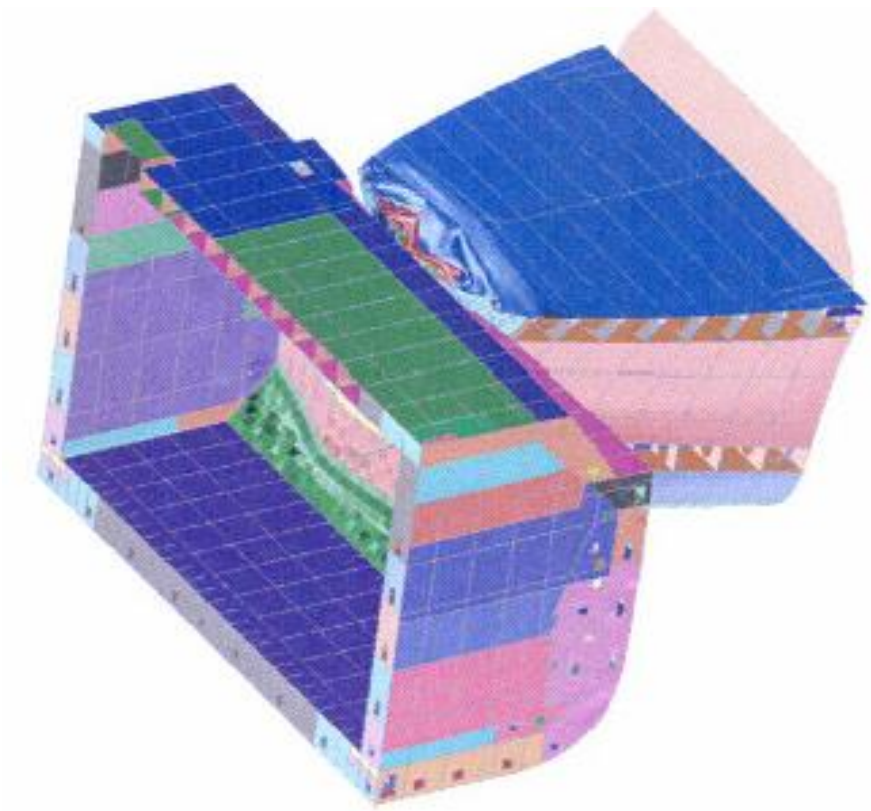

Figure 7. Deformed shape of bow striking ship and membrane type LNG carrier side [24].

As discussed in this section, there has been significant improvements in design of LNG carriers. The rulesets required by the class societies and owners result in a robust hull design that can withstand damage or recover from accidental conditions with minor repairs. Experience gained in the ship building industry demonstrates how resilient design practices can be beneficial to overall supply chain by increasing reliability even under adverse conditions.

6. Design and Assessment of Barriers and SCEs

Barriers at LNG plants protect the staff, public and the asset from accidental events. These can be grouped into several categories depending on the threat that is considered. Common applications include blast resistant components, fire protection and cryogenic spill protection. Elimination of threat should be the first line of defence, but it would not be practical for many scenarios. Therefore, blast and fire resistant design plays a critical role in safety and resilience of plants. For onshore and offshore LNG facilities different design criteria applies. When resilience is considered, onshore plants can implement some of the best practices from offshore applications.

For the plant to survive an accident with minimal damage, it is imperative to protect people. In addition to this, protecting the asset and mainly functionality of the Safety Critical Elements (SCEs) should be considered. The scope of this effort can include process modules, equipment, piping systems, support structures, tanks, and buildings. An initial screening analysis can be performed to group the systems into categories. Then, critical systems can be designed to withstand extreme events for a predefined failure frequency based on plant owner's risk acceptance criteria.

Some examples of this comprehensive approach are listed below for illustration purposes.

- Fire Hazards and Protection: Design for credible events including jet fire as applicable. Typically, API 2218 [25] and UL 1709 [26] are followed for passive fire protection. However, it is well known that gas plants are also susceptible to jet fire risks. API RP 2FB [27] guidelines can be followed at a high level to mitigate jet fire risks at onshore plants. Also, jet fire rated PFP per ISO 22899-1 [28] should be specified for modules and equipment in areas with jet fire risks. Use of risk based PFP optimization methods as discussed in a previous study by Akinci et al. [29] can avoid overconservatism and enable placement of Passive Fire Protection (PFP) to where it is really needed for safety and resilience.
- Cryogenic Spill Protection: Determine areas with cryogenic spill risk and specify Cryogenic Spill Protection (CSP) accordingly. The selected CSP product should be tested and certified per ISO 20088-1 [30] rather than assuming a certain product can resist cryogenic spill. Also, effect of water ingress or presence of moisture (if applicable) in the CSP system should be considered as rapid cooling and freezing expansion can have an adverse effect on the integrity. Additionally, the cryogenic spill might be followed by a fire and dual systems (CFP + PFP) are required for some applications.
- Piping Systems: Design critical piping systems for blast and fire loads and protect structural supports (e.g. piperacks) against applicable fire scenarios including jet fire. This does not only protect the asset, but also minimizes the risk of escalation. FABIG guidelines [31]

recommend use of different loads and acceptance criteria based on the return periods of events. Use of linear static analysis methods can be acceptable for events with 100 or 200 year return periods, but transient non-linear analysis methods are recommended for extreme events to avoid conservatism [29].

- Active Fire Protection: Design deluge systems for blast loads to survive the initial event. Protecting only certain parts of the systems limits availability after an event.
- Selection of PFP: Selected PFP products can have a dual purpose and resist both fire and cryogenic spill events. Spray applied systems or removable systems can be specified. Use of removable systems (e.g. flexible jackets) can be advantageous in some cases. These systems enable inspection of protected structural members, equipment, and piping. Also, these systems can be replaced in case of damage after an accident event instead of replacing the protection and protected member depending on the intensity of fire.
- Safety Critical Elements: SCEs at a plant can consist of mechanical, electrical, and structural systems. Control of process systems is critical to safe shutdown. Most plants are designed to be fail safe. This approach successfully mitigates escalation risks, but plant may still suffer from major damage. Protection of SCEs can reduce the damage in a plant due to an extreme event. This in turn increases resilience of critical systems. Emergency Systems Survivability Analysis (ESSA) can provide guidance for identification, assessment and qualification of critical systems to perform the design functions.
- Design of Equipment Buildings: Buildings (occupied and unoccupied) at Oil and Gas facilities have been historically designed using ASCE blast guidelines [32] when they are in blast zones. The intent of this design guide and the focus are on protection of building occupants. Therefore, functionality and survivability of equipment buildings or modules need to be addressed separately. A simplified sketch of a typical equipment room is provided in Figure 8. For most plants, the focus of fire and blast design has been protection of the building enclosure. However, for equipment buildings, contents of the building can be equally critical. The functionality of these buildings can be maintained or restored in a short period of time with proper design measures. Exterior of the building can protect ingress of blast overpressures or heat, but inertial loads experienced by the internal components can be detrimental. The following recommendations are expected to increase resilience of elevated equipment buildings in fire and blast zones:
    - Design building enclosure for low damage level [32] to minimize permanent damage to structural members.
    - Provide adequate spacing between building walls and equipment to allow blast deformation, and do not attach critical equipment on the walls.
    - Check spacing between equipment and the potential interaction effects due to inertial effects.
    - Add blast and fire resistant skirts around the base of building to prevent gas accumulation underneath, and to protect the cables and cable trays from blast (drag) and fire loads.
    - Provide fire and blast dampers at the penetrations to avoid ingress of blast pressures and flame.
    - Check accelerations of elevated buildings due to blast loads and design columns and bracing systems to limit internal damage.

- Procure seismically qualified type equipment per IEEE [33] or similar standards, and design equipment anchors for inertial loads due to blast deformations. This will increase survivability of critical equipment.
- Check dynamic response of cable trays and supports such as hangers due to inertial loads (e.g. rapid deformation of building roof induces inertial loads on the cable trays).
- Qualify elevated floor and cable penetration details for blast deformations and inertial loads (cables might be damaged if there isn't sufficient slack).
- Provide backup power systems to maintain functionality of equipment if power supply might be interrupted due to an accidental event.
- Check internal temperature of building due to an external fire event and required survivability times under full load considering heat generated by the equipment. Consider qualifying the HVAC and power supply for blast and fire cases if required to maintain functionality.
- Perform a comprehensive ductility level analysis using detailed models to capture interaction of structural, mechanical and electrical systems [29].

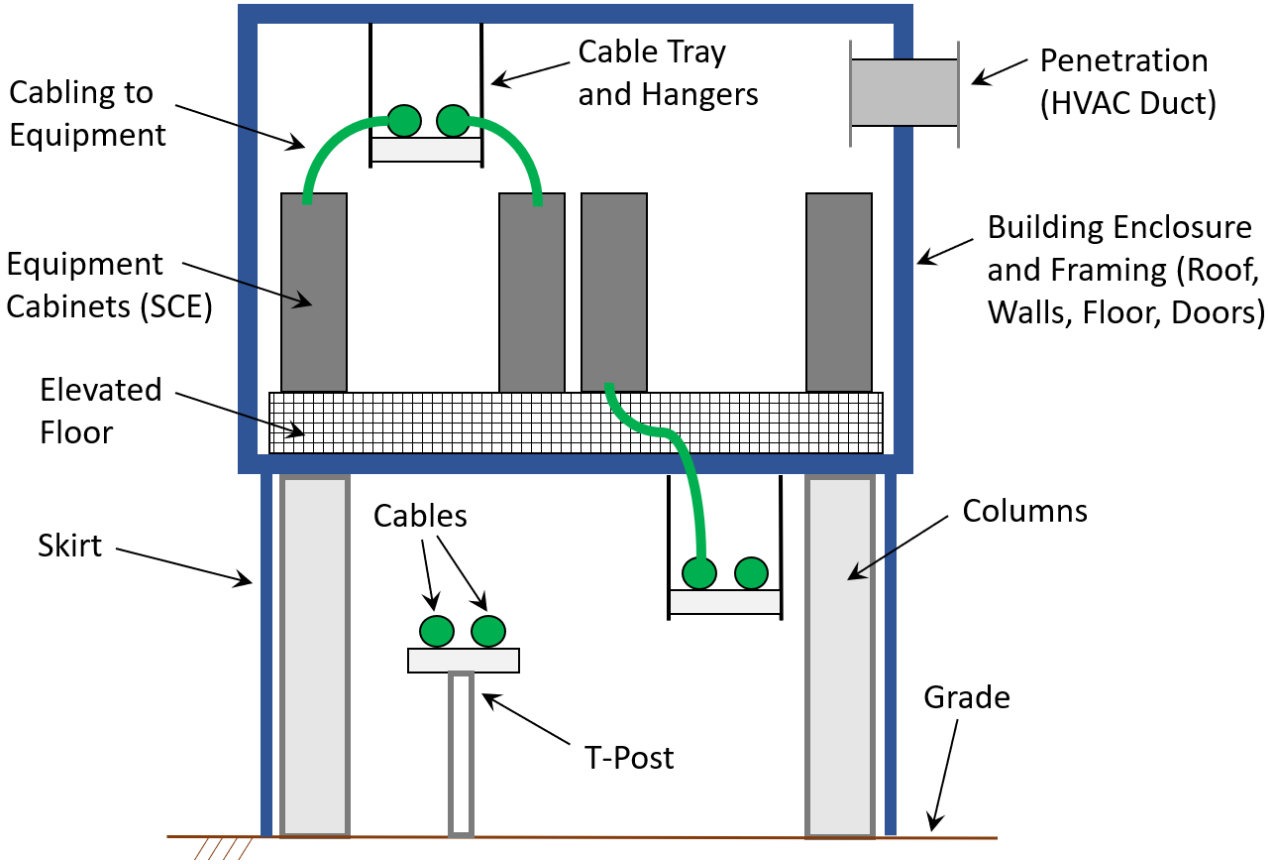

Figure 8. Schematic of a typical equipment building.

7. Best Practices

Key to designing resilient LNG facilities is a thorough assessment of the range of conditions the facility can experience. Process hazard analyses (PHA) are critical studies to be performed that systematically identify scenarios which may exceed the current design limits, highlight what preventative and mitigative barriers are in place, and identify gaps in the design that may warrant the implementation of additional barriers. Hazard and Operability Studies (HAZOP), Layers of Protection Analysis (LOPA), and other hazard analyses are powerful tools to identify hazards with the goal of reducing the risk from these hazards to as low as reasonably practicable (ALARP). The Center for Chemical Process Safety (CCPS) has published guidance on a variety of hazard analysis techniques [34] and simplified risk assessment methods [35].

To conduct robust hazard analyses, it is essential that the teams performing these analyses are multidisciplinary. Representation from engineering, control systems, operations, commissioning, technology licensors, equipment specialists, etc. should be considered based on the process and systems being assessed. The hazardous scenarios that are identified often form the basis for which scenarios are considered for detailed consequence modelling, which ultimately are used to establish design requirements in the event of fires, explosions, or cryogenic spills. Design and operations procedures accounting for these risks are expected to result in more robust and resilient plants. Use of appropriate methods and numerical tools (CFD, FEA etc.) in calculation of hazards and analysis of safeguards can help to optimize the design without overconservatism.

8. Summary and Conclusions

Several process safety risk assessment and design aspects have been discussed in this study for resilient design. While separated as *Consequence* and *Frequency* contributors to the risk equation, it is clear that these are interconnected. Any element included in one aspect of the risk equation must be appropriately addressed in the other half. For example, an assessment or design can only gain insight and value from the inclusion of safety system intervention into the effect of release rates over time, if these elements are addressed on the probabilistic side of the risk equation.

It is important to note, that any robust design is not just about the evaluation or assessment of process safety risks, but should be used in sensitivities, ALARP determination, cost benefit analysis and the like. It is only through these mechanisms that a resilient design can truly be developed. In most cases this means that there will be some design cycles, or iteration during the design phase, which ultimately will lead to optimization and proof of performance (i.e. verification of a resilient design and operation).

Resilient design can be achieved practically if proper analysis methods and tools are used. Quantification of risks and effective mitigation systems are key to success in this context. Owners can make informed decisions and set the design criteria based on their acceptable risk tolerance. Use of advanced analysis methods (CFD, FEA etc.) and risk-based design (with QRA) reduce the

conservatism and have been proven to be an effective approach for Oil and Gas facilities. The entire supply chain from production to regasification can be analyzed and made more resilient. Resilient design guidelines and methods discussed in this study can be implemented for other process plants as well.

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Development of Risk Mitigation Programs using a Quantitative-Risk-Based Approach

Rafael Callejas-Tovar†, Karen Vilas, John Dyer, Joshua Bruce-Black, William A. Mather
Baker Engineering and Risk Consultants
11011 Richmond Avenue, Suite 700, Houston, TX
† Presenter E-mail: rcallejas@bakerrisk.com

Abstract

In order to address onsite hazards and risk per American Petroleum Institute Recommended Practices (API RP) 752 and 753, most United States-based companies and sites are conducting detailed facility siting studies using either a consequence-based or risk-based approach. These detailed analyses can give companies valuable feedback concerning the overall risk profile of their facilities with respect to corporate and industry best practice risk tolerance criteria. However, many companies are left wondering, "What next?" In other words, once the hazard and risk profiles have been determined, owners/operators are struggling with implementing a prioritized action item list to systematically drive down the site risk profile to As Low As Reasonably Practicable (ALARP).

In order to reduce risk to ALARP, companies are gravitating towards the implementation of risk mitigation programs. Such programs can involve multi-year programs and require significant investment across a number of company facilities. If a quantitative risk assessment (QRA) is available, it can be used as a powerful tool to develop cost-effective risk mitigation programs. A QRA provides useful information about the dominant hazards (explosion, fire, toxic, etc.) and highest risk receptors, and allows a company to prioritize investment across all of its assets, or at individual facilities as needed.

This paper will utilize example case studies to demonstrate how a quantitative-risk-based approach can be leveraged in a risk mitigation program to optimize risk mitigation solutions such as building reinforcement, building replacement, and/or scenario mitigation. Also, the paper will present examples of facility siting issues that the processing industies struggle with, such as focusing on implementing solutions to mitigating explosion hazards while neglecting other equal or high risk

hazards, or implementing solutions company-wide that might be only effective for some assets, which results in unnecessary costs that do not mitigate the risk effectively.

Keywords: Risk mitigation, quantitative risk analysis, facility siting, risk-based building design.

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Incorporating Mitigative Safeguards with LOPA

Austin Bryan, Ed Marszal
Kenexis
3366 Riverside Drive, Columbus, OH 43221 USA
Edward.Marszal@Kenexis.com, Austin.Bryan@Kenexis.com

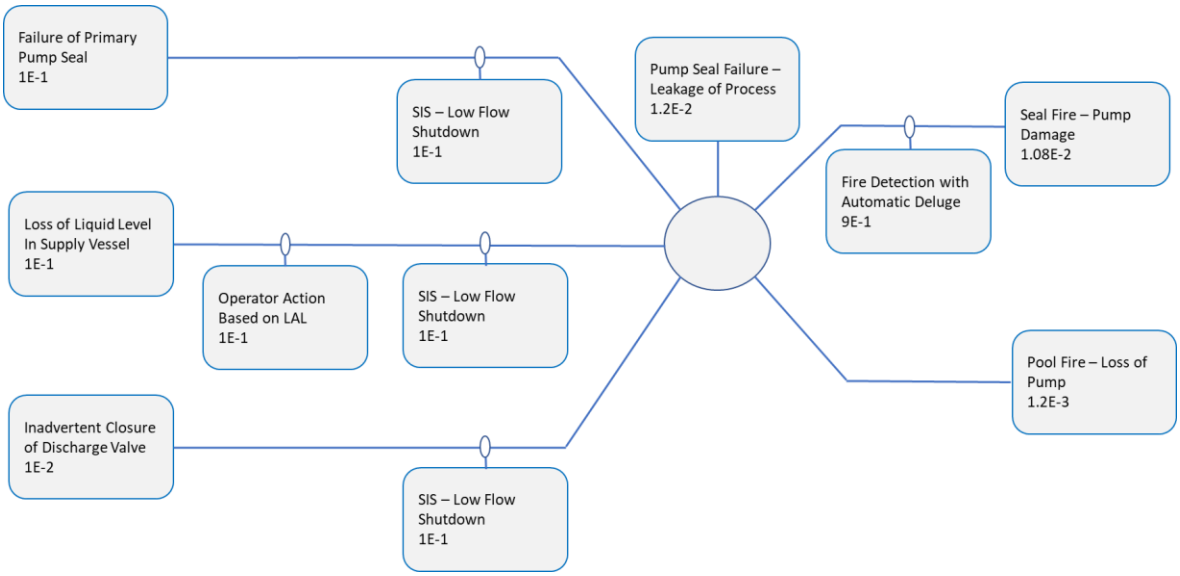
Incorporating Mitigative Safeguards with LOPA

Layers of Protection Analysis (LOPA) has become ubiquitous in the process industries as a risk assessment and management tool. Designed to bridge the gap between fast but fully qualitative PHA methodologies like Hazard and Operability (HAZOP) studies and more refined but cumbersome quantitative risk assessments (QRA), LOPA provides an economical means of quickly analyzing the risk of a system in a manner that is both reproducible and defensible. However, LOPA is not a panacea. While many, even most, risk analysis scenarios can be adequately covered by LOPA, due to the assumptions inherent in LOPA, many safeguards and hazard scenarios cannot be adequately represented in a traditional LOPA. Worse, when people or organizations dictate that LOPA shall be used as the risk analysis tool of choice, these limitations and lack of understanding by those implementing LOPA can result in flawed analysis and possible exposure to risk above tolerable limits.

Many of the assumptions used in LOPA are around how Independent Protection Layers (IPLs) are identified and how risk reduction is allocated for them. IPLs should be independent, specific, effective/reliable, and auditable. Does this mean that safeguards which do not meet one or more of these requirements do not provide any benefit with respect to safety? Does a shared component between a control loop which initiates a scenario and an interlock designed to stop the scenario, such as a valve or transmitter, prevent the interlock from stopping the developing scenario if a non-shared component is the cause of the failure? If fire detection initiates water deluge, is there no value in the deluge if it only reduces the severity of the scenario and does not eliminate it? Obviously not, but the LOPA rules for IPLs assure that assumptions that are required to allow for the simplifications that are built in to LOPA are maintained, and that the results obtained will not over-estimate the risk reduction provided by those IPLs.
One of the largest requirements is that IPLs be effective/reliable. This is generally interpreted as that an IPL should provide a risk reduction of at least 10. This is in keeping with LOPA's general

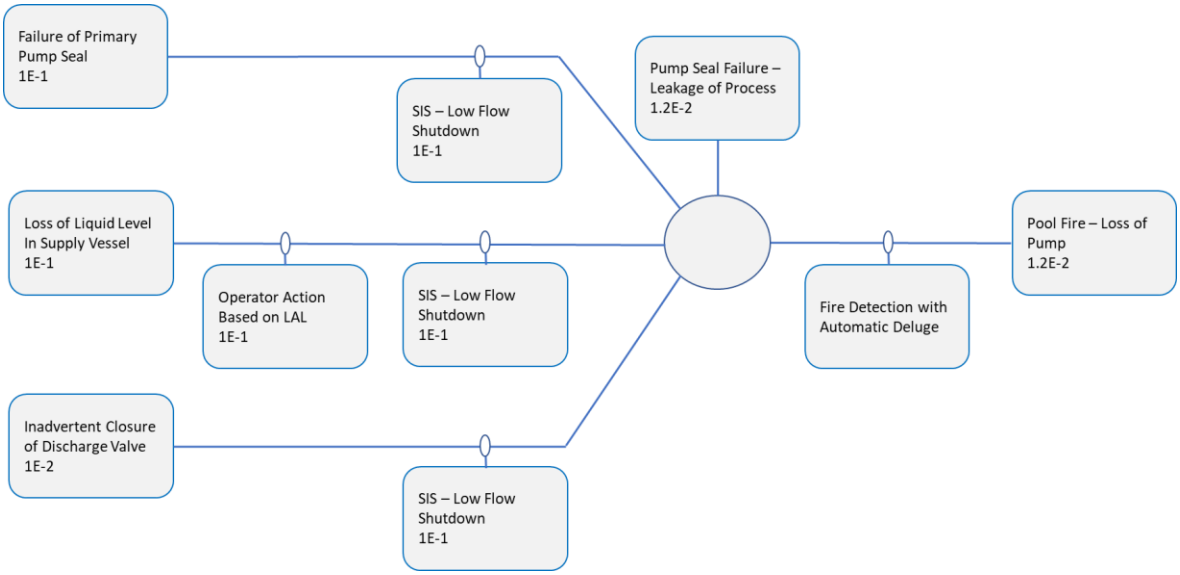
order of magnitude level of analysis and is typically stated as an IPL must have a PFD of less than 0.1. This is true for preventative protective layers, for which successful activation prevents the consequence from occurring. Taking a simple example of failure of a pump seal, we can represent the scenario using a bow-tie diagram as shown in Figure 1:

Figure 1: Pump Seal Failure Bowtie Diagram



The left side of the bow tie diagram, essentially a fault tree, includes initiating events and preventative protective layers. To the right of the primary loss of containment event, what is essentially an event tree shows the impact protective measures have on the ultimate consequence, with multiple consequences being possible. With LOPA's assumptions, the event tree is simplified, with results being reduced to either the worst case consequence occurs, or no consequence being the only options. Mitigative protective layers are effectively removed.

Figure 2: Pump Seal Failure Bowtie Diagram



Only the preventative protection layers, located on the fault tree side of the bowtie diagram, are typically considered during a LOPA study. While some conditional modifiers (occupancy, probability of ignition) may occasionally be included, this is only done when the effect of the mitigation is to effectively eliminate the consequences. From the example, we can see that LOPA gives us a more conservative estimation of risk.

Crediting the mitigative layers is problematic. Typically, in order to be credited, a mitigative protection layer must effectively completely prevent the consequences. However, with a slight adjustment, which can be easily incorporated into software designed for LOPA, mitigative layers which reduce but do not eliminate the consequence can be analyzed. In order to do so, we need to understand the mitigative protective layer, and take into consideration not only what the probability of failure of the protective layer should be (which will now include not only failure of the hardware to initiate an action, but the probability that the action, even when properly performed, does not adequately reduce the consequence) but also the expected consequence when the protective layer successfully mitigates the consequences. Effectively, the scenario is split into a mitigated and unmitigated results, with both requiring evaluation, allowing for a more nuanced analysis without performing a full QRA. The net effect of this is to reduce the frequency at which the worst case scenario, referred to as the residual risk, is expected to occur, but with the added onus of requiring a check that the risk associated with the mitigated scenario is tolerable.

Figure 3: Risk Matrix for Unmitigated Risk

Consequence						
	VH	1	2	3	4	5
	H	1	1	2	3	4
	M	1	1	1	2 Unmitigated	3
	L	1	1	1	1	2
	VL	1	1	1	1	1
		1E-5/yr	1E-4/yr	1E-3/yr	1E-2/yr	1E-1/yr.
		Likelihood				

Figure 4: Risk Matrix with Mitigated Risk

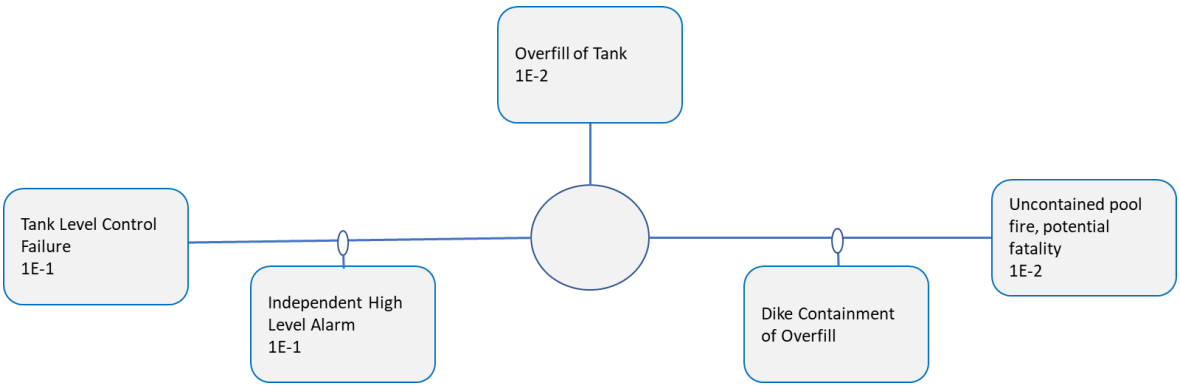
Consequence						
	VH	1	2	3	4	5
	H	1	1	2	3	4
	M	1	1	1 Residual	2 Unmitigated	3
	L	1	1	1	1 Mitigated	2
	VL	1	1	1	1	1
		1E-5/yr	1E-4/yr	1E-3/yr	1E-2/yr	1E-1/yr.
		Likelihood				

Preventative protection layers are still accounted for as usual, applying to both the mitigated and unmitigated scenario outcomes. If both the mitigated and unmitigated risk are tolerable, then no further action is needed. If either (or both) risk ranks are intolerable, then additional risk reduction measures can be recommended to address the risk gap.

There are a number of advantages to performing this analysis. The first is that we can more accurately assess risk for scenarios which rely more heavily on mitigation. This can reduce the potential for costly, over-engineered safeguards while, unlike a full QRA, being easily integrated

and applied during the LOPA. Consider a tank with a berm or dike around it. There are a number of ways it can be modelled in LOPA using simplifications. Some companies, viewing it as a mitigation and not preventative, do not credit dikes in their LOPA, especially in cases where even when contained ignition of the liquid could result in a potential injury. Others, viewing the dike as a reliable safeguard against a large spill that could impact other units, may assume the dike always works and assume that the dike contains the spill as part of the consequence. These are both simplifications used to examine the scenario within LOPA, but both can result in inaccurate analysis. A typical LOPA scenario for dealing with overfill and spillage of liquid hydrocarbon from the tank, in which the dike is not credited for risk reduction, is represented in Figure 5.

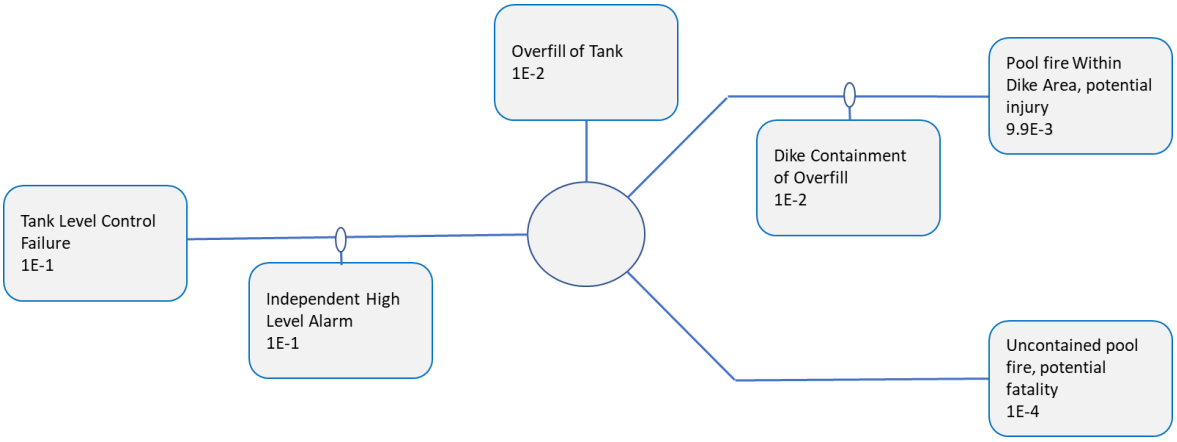
Figure 5: Bowtie Diagram of Tank Overfill



Assuming a target risk of 1E-4, an analysis of the scenario including preventative safeguards results in a risk gap of 2 orders of magnitude. Typically, this would require implementation of two additional orders of magnitudes of risk reduction, such as a SIL 2 high level interlock, to prevent overfill of the tank. If we take into account the benefits of the dike, which the team determines would significantly reduce the consequence of the release, with ignition expected to result in an injury instead of a fatality. Typical guidance for LOPA, such as that presented in *Guidelines for Initiating Events and Independent Protection Layers in Layers of Protection Analysis* published by CCPS, would not credit such a dike as it does not fully eliminate the consequence. Assuming that the dike meets the other requirements (properly maintained / inspected / drained, sufficient containment volume, and wall height to prevent slosh over the walls from hydraulic waves), it is not unreasonable to assign some level of risk reduction to the containment.

Splitting the mitigation impacts into a mitigated and unmitigated risk ranks, assuming that we can use the recommended PFD value typically assigned for dikes which could be considered preventative of 0.01, we can see that the frequency of a potential fatality has been reduced from 1E-2 to 1E-4, but that we also have an injury potential of 9.9e-3.

Figure 6: Bowtie Diagram of Tank Overfill with Dike



Note that the fatality event is now at a tolerable risk level, while the injury level likelihood is above tolerable by roughly an order of magnitude. Looking at a risk matrix, we can see the same:

Figure 7: Risk Matrix of Tank Overfill with Dike

Consequence						
	VH	1	2	3	4	5
	H	1	1 Residual	2	3 Unmitigated	4
	M	1	1	1	2 Mitigated	3
	L	1	1	1	1	2
	VL	1	1	1	1	1
		1E-5/yr	1E-4/yr	1E-3/yr	1E-2/yr	1E-1/yr.
		Likelihood				

We can again see that the fatality risk is tolerable while the injury risk in above tolerable. In this analysis, we would still need to make a recommendation, but it would be for only a single order of magnitude of risk reduction instead of the two order of magnitude as we previously recommended when the dike was not considered. From a LOPA worksheet standpoint, the scenario is split into two outcomes. The higher severity outcome with the TMEL of 1E-4, credits the dike and provides the residual risk, while the lower severity shows the consequences that exist when the dike functions as intended, and provides out mitigated risk. The LOPA worksheets for this are shown in Figure 8.

Figure 8: LOPA Worksheets for Tank Overfill with Dike

Deviation	Cause	Frequency	Consequence	CAT	S	TMEL	IPL	PFD	MEL	RRF
1.1. High Level	1.1.1. Failure of Level Control LT-101	0.1	1.1.1.1. Potential overfill of tank. Potential release of hydrocarbon liquids and subsequent ignition. Potential for significant fire and possible fatality.	Safety	H	1E-4	1. Independent high level alarm LT-102	0.1	1E-4	1
							3. Dike around tank would contain liquid and limit fire to dike area	.01		
			1.1.1.2. Potential overfill of tank. Potential release of hydrocarbon liquids to dike. Potential ignition and pool fire within the dike, resulting in possible injury to personnel.	Safety	M	1E-3	1. Independent high level alarm LT-102	0.1	1E-2	10

The second, and perhaps more important, reason to incorporate mitigative functions into LOPA is when the LOPA is specifically analyzing mitigative safety instrumented functions (SIF). While uncommon, there are SIF which are purely mitigative in nature. This can lead to confusion when companies and LOPA practitioners attempt to define a SIL level for these functions, as required by IEC / ISA 61511 – Functional safety – Safety instrumented systems for the process sector. Consider a low pressure shutdown on a pipeline. The purpose of such an interlock is to act as an isolation in the event of a significant breach and loss of containment, limiting the losses and reducing the potential size of such a release (the definition of a mitigation safeguard). Assuming a required risk reduction factor of 100, it is easy and tempting to just assign a SIL 2 target to the SIF.

The problem is that, much like with the dike, even if we were to apportion a SIL 2 PFD value of 0.01 to the low pressure SIF, it may not actually address the risk if the consequences of the pipeline failure when the SIF takes its safety action is sufficiently high. This exact scenario has been presented to the author, and the company felt that they were following the correct course of action by assigning a SIL 2. However, analysis of the LOPA revealed that the potential outcome for the unmitigated release was a multiple fatality event, as it would likely result in the formation of a large gas cloud which would impact a number of units if ignited, and thus likely catch a number of operators within the potentially fatal blast zone. A mitigated release would significantly reduce this, but the team could not rule out the potential for an operator to be in the area of the release, so while the low pressure SIF would prevent the large VCE, it would not prevent a smaller VCE/flash fire in the unit where the pipeline came into the facility, and would likely result in a fatality.

Looking at the risk matrix in *Figure 9*, we can see the risk associated with the multiple fatality event has been addressed by implementing a SIL 2, however, we still have unacceptable risk in the form of a single fatality event. Note that further reducing the PFD of the SIF does not impact this risk, as a SIL 3 function would still result in unacceptable risk, as shown in *Figure 10*.

Figure 9: Risk Matrix for SIL 2 Mitigative Function

Consequence						
	VH	1 Residual	2	3 Unmitigated	4	5
	H	1	1	2 Mitigated	3	4
	M	1	1	1	2	3
	L	1	1	1	1	2
	VL	1	1	1	1	1
		1E-5/yr	1E-4/yr	1E-3/yr	1E-2/yr	1E-1/yr.
		Likelihood				

Figure 10: Risk Matrix for SIL 3 Mitigative Function

Consequence						
	VH	1 Residual	2	3 Unmitigated	4	5
	H	1	1	2 Mitigated	3	4
	M	1	1	1	2	3
	L	1	1	1	1	2
	VL	1	1	1	1	1
		1E-5/yr	1E-4/yr	1E-3/yr	1E-2/yr	1E-1/yr.
		Likelihood				

Because the initial release will result in a potential fatality, mitigative functions cannot sufficiently reduce the risk, and assigning that risk to a mitigative SIF results in a false sense of security. Appropriately evaluating the SIF as a mitigative function shows that the scenario will require preventative measures to reduce the likelihood of the loss of containment in order to adequately reduce risk.

Implementing the use of mitigative protective layers in an IPL analysis will take some effort. Software and templates for LOPA will need to be updated to allow for easy incorporation of mitigative safeguards into scenarios with the ability to separate consequences before and after mitigation (and simultaneously evaluate both). Guidance on when and what mitigative scenarios should be considered will need to be developed and facilitators trained. Once this has been done though, accounting for the risk reduction of important mitigation safeguards will fall easily within the LOPA workflow and allow for a more accurate view of risk and further reduce the need to take scenarios to expensive and time consuming quantitative risk analysis.

Keywords: LOPA, Layer of Protection Analysis, Mitigation, Consequence Mitigation

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Hole Size Matters

Jeffrey D. Marx* and Benjamin R. Ishii
Quest Consultants Inc.
Norman, Oklahoma
jdm@questconsult.com

Abstract

In the world of consequence and risk analysis, there are many variables and parameters that we study, modify, and debate. A significant variable that is often under-appreciated is release hole size. Most work in consequence and risk analysis is focused on loss of containment (LOC) events. For these evaluations, the hole size used to represent that LOC, when combined with the released material's thermophysical properties, is critical in determining a mass release rate, which then directly affects the magnitude of the flammable or toxic hazards that are calculated. Accordingly, matters regarding hole size have a significant effect on the predictions produced by consequence analysis. The selected hole size for a single release scenario directly effects the impacts of a hazard; and the hole size associated with the maximum hazard extent may vary based on the hazard being evaluated. When an evaluation turns to a quantitative risk analysis (QRA), the importance of hole sizes is compounded due to the added effects of frequencies that are assigned to the various hole sizes. This issue of hole size is explored with an evaluation of some common practices and assumptions, and some of the associated pitfalls. The effect on predicted consequences for various hazards, and how a range of sizes selected in a QRA could affect the risk predictions, are also explored. It will be clear that hole size matters.

Keywords: Consequence modeling, Consequence analysis, Risk analysis , QRA

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How Can I Effectively Place My Gas Detectors

Jesse Brumbaugh*, Vinny Simoes and Chris Hickling
aeSolutions, Inc.
Houston, TX 77084
jesse.brumbaugh@aesolns.com

Abstract

Several Recognized and Generally Accepted Good Engineering Practices (RAGAGEPs) exist to help someone make their selection and placement of gas detectors (e.g. ISA-TR84.00.07, NFPA 72, UL-2075). However, there are no real consistent approaches widely used by companies. Historically, gas detection has been selected based on rules of thumb and largely dependent on experience. Over the last several years there has been a growing interest in determining not only the confidence but also the effectiveness of those gas detection systems. In fact, incorrect detector placement far outweighs the probability of failure on demand (of the individual system components) in limiting the effectiveness of the gas detection system.

An effective gas detection system has three elements:

1. A comprehensive Gas Detection Philosophy
2. Appropriate Detector Technology Selection
3. Correct Detector Placement

The Gas Detection Philosophy clearly specifies the chemicals of concern and the intended purposes, i.e. detection of toxic or combustible levels, voting requirements, alarm rationalization, and control actions.

Appropriate Detector Technology Selection includes consideration of the target gas and the required detection concentration levels.

The primary approaches for Detector Placement are geographic and scenario-based coverage. Geographic coverage places detectors on a uniform grid, and sometimes areas risk ranked to reduce the number of detectors required. Scenario-based coverage has a range of leak models and places gas detectors based on the dispersion modeling results.

All three elements for effective gas detection (philosophy, technology, and placement) are interdependent but understanding their relationships is of paramount importance to design an effective gas detection system.

The intention of this paper is to present the main considerations that design engineers and process safety professionals should address for each gas detection system element in order to obtain the best return on your investment when placing your gas detectors.

Keywords: Instrumentation, Reduction of Risk, Risk Assessment, Protection, Detection System, Alarms and Operator Interventions, Detector, Gas Detection/Dispersion Prediction

Introduction

Many plants make use of a gas detection system (GDS) to protect both onsite and offsite personnel, and many plants deal with leaks frequently. Most leaks that occur are small. Being able to distinguish between a nuisance condition and a more serious leak that may cause harm to personnel is an important consideration to make when designing a GDS. On one hand, many plants have had experience with nuisance alarms from either inadequate detector technologies, poor gas detector placements, or in some cases, a design based on a philosophy that does not make appropriate leak distinctions. On the other, many plants also make use of outdated or legacy technology that may have been designed and positioned based on Recognized and Generally Accepted Good Engineering Practices (RAGAGEPs) at the time, but are deficient and unable to detect a significant number of serious leaks, or may have missed an opportunity for updating after a significant process modification or expansion.

While there are standards that exist to support selection and placement of gas detectors (e.g. ISA-TR84.00.07, NFPA 72, UL-2075), the reality is the optimal solution can vary from plant to plant (e.g. a congested offshore platform, a batch chemical plant with multiple recipes, a refinery, a sour well) and inconsistent approaches are oftentimes found. Our understanding of gas dispersion and the ability to model and predict the release behavior has grown significantly. Using this knowledge and expertise, as well as practical plant experience from start-up and testing, much more educated and informed decisions can be made to increase confidence in a system. Ignoring nuisance alarms can degrade the overall effectiveness of the GDS. For example, incorrect detector placement alone can be more detrimental to the overall GDS performance over individual component probability of failure on demand to the point where it cannot be credited as an effective independent protection layer (IPL).

Gas Detection Philosophy

The gas detection philosophy drives the design basis of the entire GDS, and without a good structure, sub-optimal decisions can be made at later stages such as when selecting the right technology or placing detectors. The philosophy should provide general direction and intent on how to implement the GDS. Several philosophy decisions may differ from plant to plant, so it is important to consult a subject matter expert (SME) as well as those experienced and knowledgeable with the operations and design of the plant.

For plants handling flammable and toxic gases, continuous monitoring can be an effective risk management strategy. Some plants may opt for fixed detectors while others may choose personal gas detection. A combination of the two may be best, based on an understanding of the hazardous gases present and the amount of time a person may have until a major injury or life-threatening condition arises from the gas. Likewise, detection of a serious leak should be effectively communicated to alert personnel to the presence of the hazard and to shut down and isolate the section of the process that is having the leak to prevent further escalation. Communication to personnel is often in the form of audible alarms and strobes to signify areas with active hazards and direct personnel to safe locations. Personnel in control rooms or other remote locations should have a good human-machine interface (HMI) that is intuitive and helps the operator to accurately identify the area of the plant were the hazard is present and provide a quick response, such as shutting down specific building air intakes and exhaust fans to not draw toxic or combustible fumes into different rooms if needed.

The philosophy should define specific performance goals for the GDS and effective strategies to achieve these goals, including the selection and location of gas detectors. Among those goals should be a focus on best personnel protection, including both potential toxic and flammable effects. For onsite personnel, monitored toxic effects with established end points should include those with atmospheres that are Immediately Dangerous to Life and Health (IDLH), Emergency Response Planning Guidelines (ERPGs), and probability of fatality where life-threatening conditions could develop after exposure to a toxic atmosphere for over a specific duration. For flammable atmospheres, exceeding the lower flammable limit (LFL) provides the opportunity for combustion should a source of ignition be contacted. In order to not exceed IDLH, ERPG-3, or LFL concentrations, a fraction of these concentrations should be monitored. The specific fraction depends on multiple factors, including the available detector technology, manufacturer recommendations, voting, numbers of installed detectors, an indoor or outdoor location, and detector reliability while avoiding spurious activations or nuisance alarms. For some highly toxic chemicals, continuous monitoring may not be practicable at these levels under certain conditions and alternative RAGAGEPs can be used that will still protect personnel from a potential leak. These action levels can lead to such things as requiring a self-contained breathing apparatus (SCBA) or continuously supplied breathing air in areas where a highly toxic atmosphere could become present.

Each highly hazardous chemical should be thoroughly documented, including:

- Different toxic or combustible levels;
- National Fire Protection Association (NFPA) 704 ratings;
- Basic physical properties;
- Applicable American National Standards Institute (ANSI), National Institute of Occupational Safety & Health (NIOSH), and National Institute of Standards and Technology (NIST) codes and standards;
- Locations of hazards;
- Consequences of deviation; and
- Detection criteria.

With hazard identification, include a cause and effect diagram for each chemical and a list of specific areas that may pose a hazard such as trenches or other areas where hazardous gases that are heavier than air are more likely to accumulate. These areas can be identified through dispersion modeling. Aside from highly hazardous chemicals, there is also the potential for inert gases that are not toxic to personnel to cause an oxygen deficient atmosphere and potential asphyxiation. Some inert gases may have additional hazards that need to be factored in, such as carbon dioxide (CO_2) resulting in sleepiness that can diminish the egress of personnel.

The philosophy should describe in detail the method for detection, alarming, and actions to be taken. Detection should be configured appropriately to focus on hazardous leaks that may harm personnel or the plant instead of small leaks. Occupational health requirements can require toxic concentrations to be reported using time-weighted averages. Detector voting and methods for early detection should be detailed in the philosophy. The philosophy should list the requirements for audible and visual alarms along with strobes and other forms of visual notification including alarm rationalization. The philosophy document should describe what hazardous conditions are being alarmed and ensure each location where hazards and an operator may be present provide alarms that can be seen or heard. Having both visual and audible alarms combined is a best

practice. For example, a room with a detected leak should strobe both exterior and interior to the room to show field personnel where the hazard is and provide a safe means of egress. Restricting notifications to where the hazard may be present can avoid unnecessary evacuations from other areas of the plant, which is especially useful for indoor releases. Outdoor releases should consider a worst-case release. Zoning and alarms should help personnel identify the source of the leak so that a safe egress route can be taken to the safest muster point location combined with available windsocks. A windsock should be visible at all outdoor locations with personnel present during operator rounds or from occupied buildings where a toxic or combustible hazard could become present. When covering mitigation, list all methods for response including the instrumented control actions on alarm, room ventilation requirements, and mitigation systems such as fogging or water deluge to ensure the response will mitigate the specific chemical hazard.

Appropriate Detector Technology Selection

Many different gas detection technologies exist, including infra-red (IR), electrochemical, catalytic bead, semiconductor, and laser. Each of these technologies has different strengths and weaknesses. Some technologies allow for lower detection limits. Target gases for different detectors can have cross sensitivities with different chemicals that one should be aware of when making the selection of which detector technology will be selected. Different technologies have specific maintenance requirements or costs to be aware of.

Detectors can either be of the fixed point or open path gas detector (OPGD) variety. OPGDs have the advantage of covering a very wide area, which is especially useful outdoors when it can replace multiple point detectors and reduce overall costs. OPGDs will have limitations on the overall length of the beam. OPGD technology is typically limited to the infra-red and laser types. It should be remembered that an OPGD measures the quantity of the target gas in its beam and it does not measure the concentration. The OPGD reports concentrations by the linear meter or foot, such as in LEL-meters for combustible clouds. 1 LEL-m and action at 2 LEL-m is a common setting for a combustible OPGD. For a toxic OPGD, a major issue is you cannot quantify what an operator who happens to be in that area has been exposed to. In order to better estimate the ppm value of toxic clouds (typically for reporting purposes), sometimes an extra point detector or two are included in the area at higher risk leak locations.

The gas detector technology for the target gas of interest should be stated in the gas detection philosophy. Catalytic bead detectors can pick up any combustible gas. However, they suffer from unrevealed failure modes, need regular maintenance, and require the presence of oxygen to operate. They are commonly the choice for hydrogen since IR cannot detect hydrogen. IR point detectors are good general-purpose detectors able to find any hydrocarbon gas which absorbs at their wavelength, and are able to detect at ppm or LEL% levels. IR OPGDs use the same technology as point IR gas detectors though typically operate at a different wavelength. Electrochemical and semiconductor detectors are a common choice for toxic gases and are for point detection only. Laser based OPGDs are used for toxic gases.

Correct Detector Placement

Industry standards followed for gas detection include ANSI/ISA-TR12.13.02 & 03, ANSI/ISA 84.00.01 & 07, API RP 14C, API RP 500, and NFPA 72. Basic rules of thumb applied for placement include locating detectors at breathing height for toxic gases, 1-2 feet above ground for heavy gases such as propane, and for gases that are lighter than air either above the leak source or

as high as possible if those gases may accumulate in specific areas such as hydrogen in a battery room. Additional considerations should also be made for conditions that may cause the gas to behave differently, such as cryogenic conditions, as both liquefied natural gas (LNG) and liquefied ammonia are known to disperse low to the ground while the vapors are cold. Other rules of thumb that are important are to place detectors near air ductwork intakes or room outlets, in areas accessible for maintenance, and away from locations that can be damaged by general maintenance or frequently traveled and not in areas where flooding can disable or damage the equipment.

Two RAGAGEPs for placing gas detectors that are mentioned in these standards include scenario and geographic-based methods. Older plants that have not used either of these methods are oftentimes found to have large gaps in detector coverage.

A commonly employed strategy is to place gas detectors to ensure detection of a 5-meter cloud for combustible hazards. The geographic method is simply applying this strategy globally to where all possible areas that handle a specific hazardous material have a hypothetical leak with a 5-meter diameter cloud covered, resulting in a uniform 5-meter spacing of detectors. Basic rules of thumb discussed earlier are then used. The development of a gas detector placement drawing from this method is very simple and cheap.

While geographic methods are successful at leak detection, it can result in more detectors than are necessary, which in turn leads to higher installation and plant operating expenses. This can especially lead to excessive gas detectors for toxic hazards, as the 5-meter cloud methodology was intended for combustible clouds. As a result, many companies prefer to use scenario-based coverage over geographic methods. Scenario-based methods use dispersion modeling to guide detector placement decisions. Scenario model selection involves identifying a variety of leak points, hole sizes, and leak directions. After performing dispersion modeling, detectors are subsequently put in the optimal locations based on leak point and critical receptors of concern locations to detect the hazardous gas.

Tools used for gas dispersion prediction include Gaussian plume, empirical models, and computational fluid dynamics (CFD) models. Gaussian plume is the most simplistic of them all and utilizes basic equations and constants, empirical models add more resolution by making predictions from experimental observations and include software such as the DNV-GL PHAST unified dispersion model (UDM) in a 2--dimensional field, and CFD models use a full suite of transport equations while maintaining conservation of momentum, mass, and energy in a 3--dimensional field that includes geometric and topographic interactions.

Outdoor leaks are largely found to follow the active wind direction. The reason is that the mass and momentum of the air is far greater than the leak itself and will eventually carry it out of the plant in that direction. However, there are exceptions to this typical behavior. For example, it has been found from dispersion modeling that leaks which are especially large and at high pressures have a very large initial momentum and can travel a good distance on their own, especially if along the ground. Sometimes this can result in toxic lethality and flammable thresholds being primarily dictated by leak direction rather than wind direction. Also, topography plays a very large role in dispersion especially for cryogenic plumes, or heavier than air gases. Trenches and diking as well as buildings and large pieces of equipment in plants very frequently dictate gas dispersion behaviors at ground level. OPGDs are often used around the perimeter of secondary containment

of hazardous chemicals due to the ability of the geometry to contain and channel gases. Also, wind currents can interact with large buildings and tanks in such a way that turbulence eddies and recirculation zones influence the dispersion in different ways depending on wind direction. Depending on gas density, outdoor vertical leaks in the air may not be detectable from ground-based detectors and an elevated open path detector beam, as well as point detectors at the air intakes of potentially affected buildings, may be needed. All these behaviors can be difficult to predict from simply looking at the geometry, and as such, dispersion modeling that incorporates CFD may be critical to understanding these behaviors and making better decisions in detector placements on a case by case basis.

Indoor leaks are largely found to follow the initial leak direction and are eventually drawn out by room out-take ducts. Room ventilation rates are many times smaller than outdoor areas, but still, influence the dispersion of leaks from the room airflow patterns. When there are many rooms to evaluate, one of the more useful tools for predicting dispersion patterns in a room is a basic airflow model which can be quickly performed using CFD models. Observing these patterns as well as the locations of all equipment handling highly hazardous chemicals, one can make reasonable predictions as to where the vapors of a release will eventually travel as the room exhaust ventilates the air. Incorporating different leak scenarios into these models helps support and confirm these decisions, as well as finding common pathways that multiple vapor clouds may take. Modeling can also expose counter-intuitive behaviors. For example, while typical detector placement practices for lighter than air gas releases is to place the detectors as high as possible, an air supply coming from the ceiling with exhaust ducts along the ground has shown ammonia gas clouds being pushed away from the ceiling and spreading to other areas along the ground. Indoor models can also help in revealing regions with still air (dead zones) where gas clouds may not migrate while dispersing. Finally, heavier than air releases are often found to have the highest concentrations and largest footprint at grade level, suggesting that detectors as low to the ground as possible should typically be used for heavy gases.

Some detector technologies can have additional constraints to be mindful of. For example, electrochemical based detectors contain an electrolytic solution that is directly exposed to the air. Electrochemical detectors must be sheltered from the rain to avoid degradation. If the air that this solution is exposed to has a very high velocity, it can lead to rapid evaporation and depletion of the solution, which can lead to problems such as increased maintenance and inaccurate readings. On the other hand, placement in a dead zone can lead to non-detection of hazardous gases. For these situations, gas dispersion models using CFD are quite useful in finding the right balance with gas detector placement decisions. Considerations for the leak models are numerous, and it can be easy to take different paths. Underspecifying the models can lead to a suboptimal solution with gaps in coverage, which is especially problematic if the GDS is being treated as an IPL and it requires a high amount of coverage. On the other hand, over specifying the models can result in more analysis and computational time than is required. As such, a balanced approach is recommended, and a series of steps can be followed to come up with an effective solution to detector coverage that is not as time intensive, including with CFD based methods. The flow chart in **Figure 1** demonstrates this proposed scenario-based work flow.

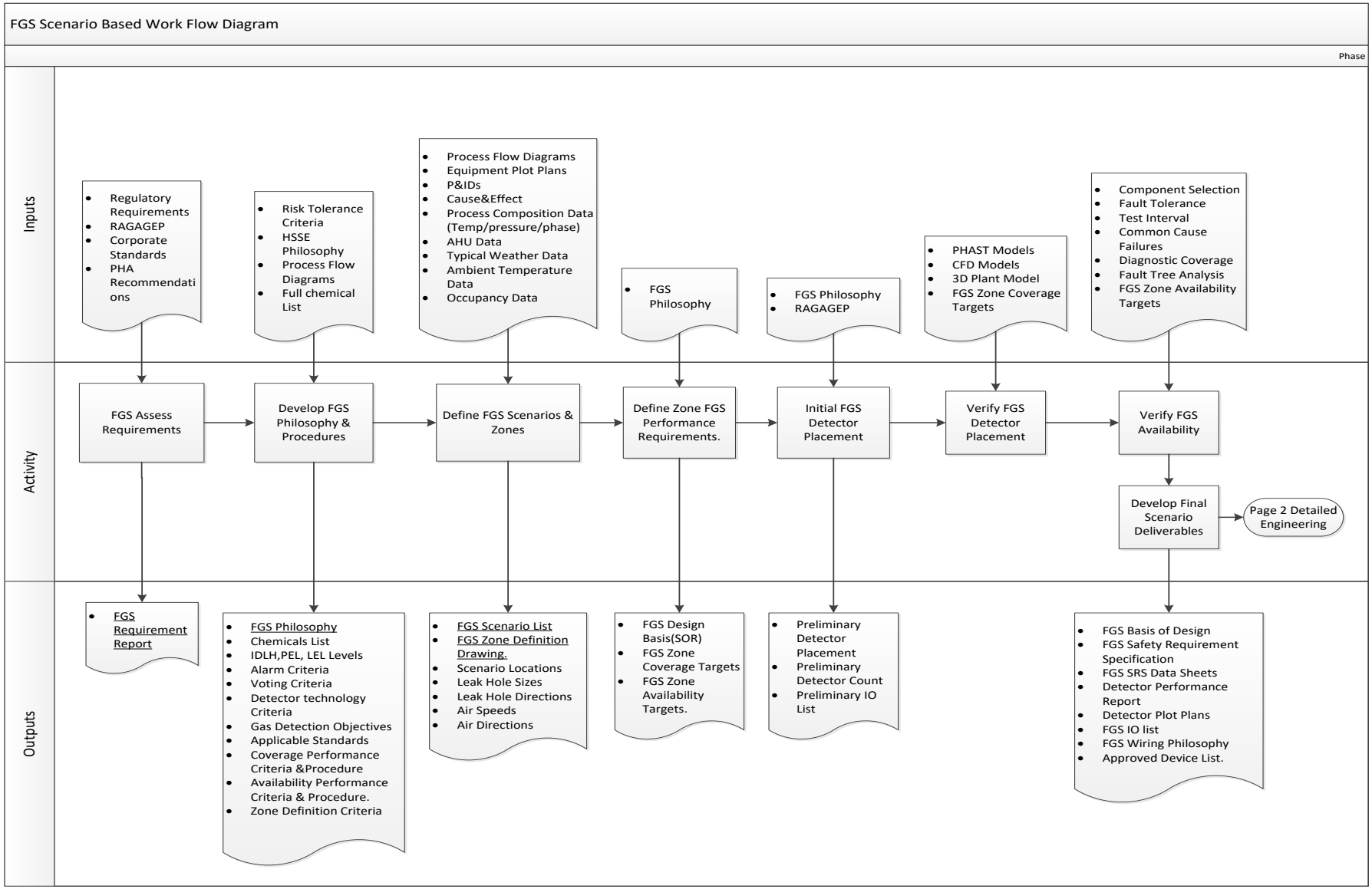
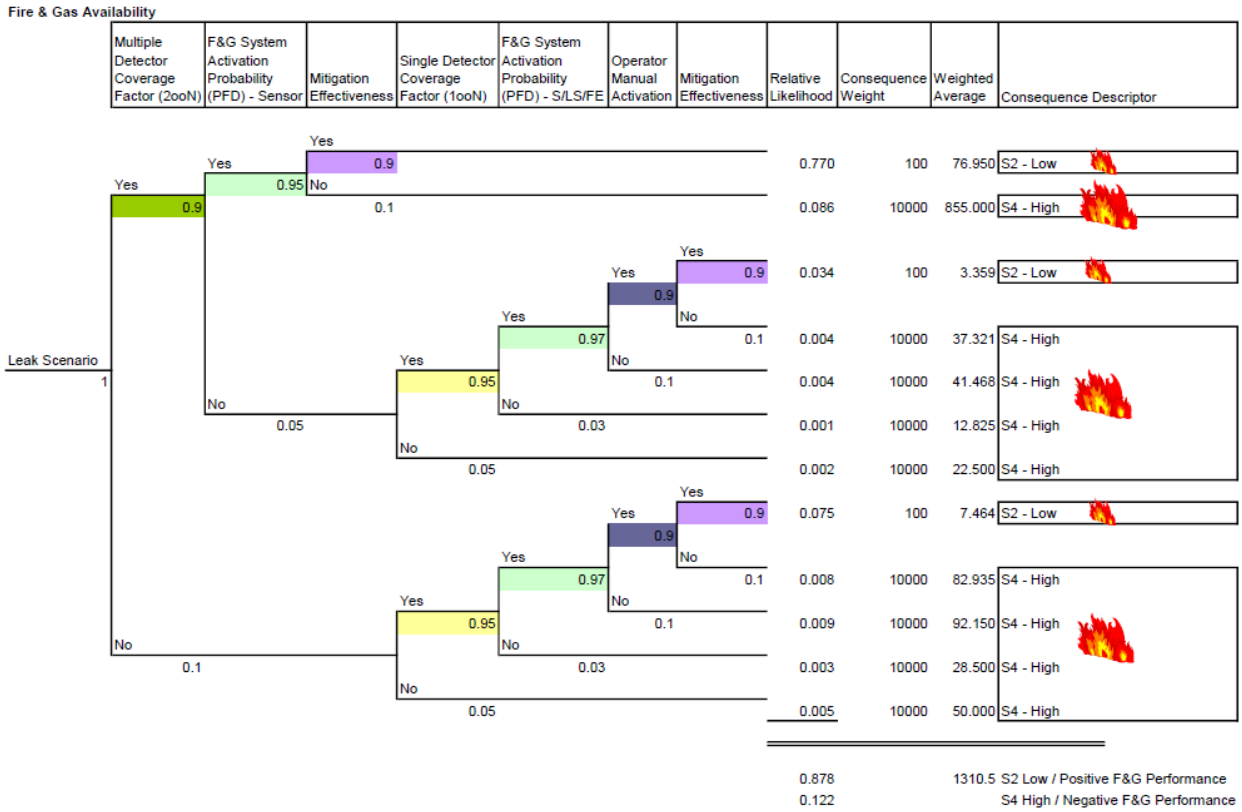


***Figure 1**: FGS Scenario-Based Workflow*

After developing the gas detection philosophy and gathering all process safety information (PSI), a key step to more quickly optimize detector placement is coming up with an initial detector coverage map using RAGAGEP methods. After having the initial design coverages, PHAST or CFD models can be used to assess the initial coverage performance. Some locations that are nearly identical can utilize the same models across numerous areas. For example, multiple trains, multiple similar rooms, etc. This will reduce the number of models that need to be run. Time to alarm can be included in an assessment of maximum available response time (MART). Safety availability performance provides confidence levels in the GDS based on voting requirements, technology reliability, and mean time to failure (MTTF) of different components (e.g. system hardware and mitigation measures). Based on these metrics, the coverages are adjusted based on predicted behaviors from the models until an optimal design is achieved. Then, fire & gas safety (FGS) requirements are specified, and then the GDS can move to detailed engineering.

It is important to note that beyond having a highly effective detector coverage, there are limitations of what a GDS can reasonably do. For example, consider following ISA-TR84.00.07 and employing voting to improve reliability. After the percent coverage is determined, a fault tree may be used to estimate the overall mitigation effectiveness of the GDS system.

***Figure 2**: Fault Tree Analysis on Fire & Gas System Performance*



To achieve an overall risk reduction of SIL-2 or higher, one would need well over 99% coverage and for all components to be SIL-2 capable. Trying to achieve SIL-2 with a GDS would be over specifying performance. Rather, ISA-TR84.00.07 advises that a GDS not be considered as an IPL

if the overall effectiveness is less than 90%. One can see in the fault tree example in **Figure 2** how even 90% coverage with 2ooN voting and 95% coverage with 1ooN voting will result in an overall effectiveness under 90% if the entire system is limited by a mitigation effectiveness of 90%. Consequently, both the detector coverage and the mitigation effectiveness should be above 90% to be credited as an IPL, which can necessitate modeling of the mitigation system as well to prove a higher level of effectiveness.

Conclusions

All three elements required for effective gas detection (philosophy, technology, and placement) are interdependent but understanding their relationships is of paramount importance in designing an effective gas detection system. In addition, modeling provides additional details that support other important plant safety decisions. Model driven sensor locations enable informed emergency planning. For example, ground level detectors located between the source of the release and the critical receptors can provide early warning to building occupants to take the specified emergency action. It can also reveal gases that may be drawing into a building air intake that can be a considerable distance from the leak source. 3D modeling that incorporates wake effects from buildings can show the plume reaching areas that may not be immediately intuitive such as air handlers on the back side of a building. Room ventilation patterns may also result in several non-intuitive behaviors. It is therefore essential to understand hazardous chemical properties, potential release sources, release directions, and the ventilation patterns in the room for proper gas detector placement. For some clients, the plant fence line is an important consideration for potential impact to public receptors, and dispersion models provide valuable information on time for emergency response and likely concentration impacts.

Considering the recommended approach of scenario-based coverage through dispersion modeling, although it may increase the initial project cost, over the long term, it typically has a lower overall cost due to reduced detector quantities, reduced maintenance, etc. It also gives confidence and an engineering basis behind detector placement decisions to reduce life cycle costs, overall risk to onsite plant personnel, and overall risk to offsite public receptors.

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Consequence assessment considerations for toxic natural gas dispersion modeling

SreeRaj R Nair, Noma Ogbeifun
Chevron Corporation
Midland, Texas, USA
snair@chevron.com, Noma.Ogbeifun@chevron.com

Abstract

The numerical simulation of gas dispersion and estimation of consequence impact is of importance in Oil and Gas industry's process safety management. For natural gas fields with toxic components like Hydrogen sulfide, the toxicity impact zone drives business decisions related to equipment design, facility siting, layout, land use planning, and emergency response measures. Proprietary tools or empirical models which are calibrated using experiment database are often used for carrying out consequence modeling.

The selection of a tool and a suitable dispersion model, based on the cloud behavior, at the source of dispersion is critical for the impact zone estimation. It is observed that, the fluid phase and the cloud density are key for determining the appropriate dispersion model. Incorrect parameter selection could lead to an inaccurate consequence impact zone estimation which could result in disproportionate risk management efforts.

This paper summarizes the methodology and results from extensive set of consequence modeling studies done for potential release events associated with Gas exploration and production. The analysis focuses on the significance of composition, temperature, pressure and hole sizes in release source term determination as well as the atmospheric parameters that could impact the dispersion and estimation of impact zones. The output of the study and analysis provides (i) input and guidance on selection of dispersion model to represent appropriate cloud behaviors (e.g. buoyant or dense gas dispersion) and (ii) critical parameters that should be included in the sensitivity assessment to determine the consequence impact zone.

Keywords: Consequence Modeling, Facility Siting, Toxicity, Hydrogen sulfide, Impact zone, Parameter sensitivity, Natural gas composition

1. Introduction and motivation

Understanding of the process related risks is key in natural gas exploration and transportation process safety management. Several major toxic natural gas release incidents (see Table 1) have happened in the recent past resulting in human fatality, environmental damage and asset loss (BSEE 2014, Jianwen 2011). Predictive risk assessments are carried out to determine the extent of hazardous level distances (impact zone) and how frequently the event occurs (Nair & Wen 2019).

Estimation of the potential impact zone from different accident scenarios through scenario-based consequence modeling forms integral part of process risk assessment (US DoT 2018). An important contribution to the calculation of the impact zone comes from the modeling of atmospheric dispersion following the accidental release of toxic fluids. Impact zone estimation by consequence modeling is typically carried out using proprietary tools or empirical models (Hanna et.al. 1982, Nair & Wen 2019). These models and tools have a range of applicability and are validated using experiments (Hanna 1982, Pandya 2012). Preventive and mitigation measures for risk management are prioritized using the study outputs. Variation in model inputs impact results and different parameters have different influences on the results (US DoT 2018). Incorrect selection of the approach, tool and uncertainty in the input could lead to an inaccurate impact zone estimation which could result in disproportionate risk management efforts. This challenge can be addressed by better understanding of the cloud behaviour following release and sensitivity analysis of the modeling inputs and parameters.

Table 1: Toxic natural gas Incidents

Incident	Consequence and description
1950, Poza Rica, Mexico low altitude temperature inversion	Twenty-two persons died and 320 were hospitalized as a result of exposure to hydrogen sulfide for 20-minute period.
1974-1991, Sour gas gathering line releases, USA (EPA records)	11 incidents, Multiple fatalities, Unspecified number of wildlife died
1992, Gezi, The Zhao 48# well; H_2S gas well blowout	6 fatalities and 24 poisoning; under pit operation corporation, Petroleum administration, Bureau of North China
2003, Kaixian blowout (Chongqing "12.23" incident), high sulphur gas	240+ fatalities, 2000+ hospitalization, 65000 evacuated; direct economic loss of $900 million
2006, Sichuan (The Luo 2# well)	About 10000 people evacuated
2010-2014, Southeast Saskatchewan, Canada	43 sour gas leaking facilities (with average H_2S concentrations at 30,000 ppm)
2013, Kashghan field, Kazakhstan	200 km of leaking pipeline, $3.6 billion to replace

2. Methodology and tools

Impact zones from potential outcomes like explosion overpressure, hazardous levels of toxic concentration, flammable cloud extent or thermal radiation from fires are estimated by (i) experiments / field trials, (ii) using index, spacing tables, (iii) predictive (consequence) modeling. This study follows the scenario-based model approach for gas transmission integrity management risk assessment as per ASME B31.8S-2004 (US DOT 2018) to determine the key parameters that should be subjected to sensitivity analysis. The analysis focuses on the significance of composition, release source terms and environmental parameters that could impact the dispersion and estimation of impact zones. The key stages of consequence modeling are given in Figure 1.

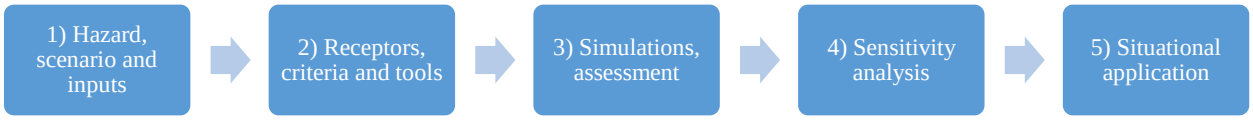


Figure 1: Methodology - Consequence modeling and sensitivity analysis

2.1 Hazard, scenario and input

Natural gas is a clean and naturally occurring hydrocarbon gas mixture which is an efficient source of energy. Natural gas in the event of unplanned release followed by ignition can result in flash fire, jet fire or explosion leading to thermal radiation or overpressure impact to personnel, environment or asset. Natural gas consists primarily of methane (CH_4) and rest of the composition depends on the reservoir (gas field) location. One-fifth to one-third of all natural gas resources in the world could fall under the sour gas classification (Kelly et.al. 2011). Natural gas is usually considered 'sour' if there are more than 5.7 milligrams of H_2S per cubic meter of natural gas, which is approximately equivalent to 4 ppm by volume under standard temperature and pressure (Speight, 2007). H_2S is highly toxic (fatal effects at low concentration), extremely flammable and corrosive. The molecular mass of CH_4 is 16 g/mol (lighter than air) and that of H_2S is 34.1 g/mol, slightly heavier than air (29 g/mol) at standard conditions. Pipeline accidents accounted for 70% of the accidents involving natural gas and the most frequent causes were mechanical failure of the pipelines or due to significant changes to the surrounding environment (Bariha et.al, 2016). 90% of sour natural gas releases could result in toxic cloud dispersion with potential impacts (Muhlbauer, 2004).

Accidental releases of toxic natural gas from transfer pipeline (gathering line from wells to processing facilities) and atmospheric dispersion was considered for this evaluation. A ground level release from pipeline (6-inch diameter) transferring toxic natural gas and downwind dispersion of the cloud was considered as the base case event scenario. Input and parameters for consequence modeling are given in Table 2 and an illustration of the release and dispersion scenario is given in Figure 2.

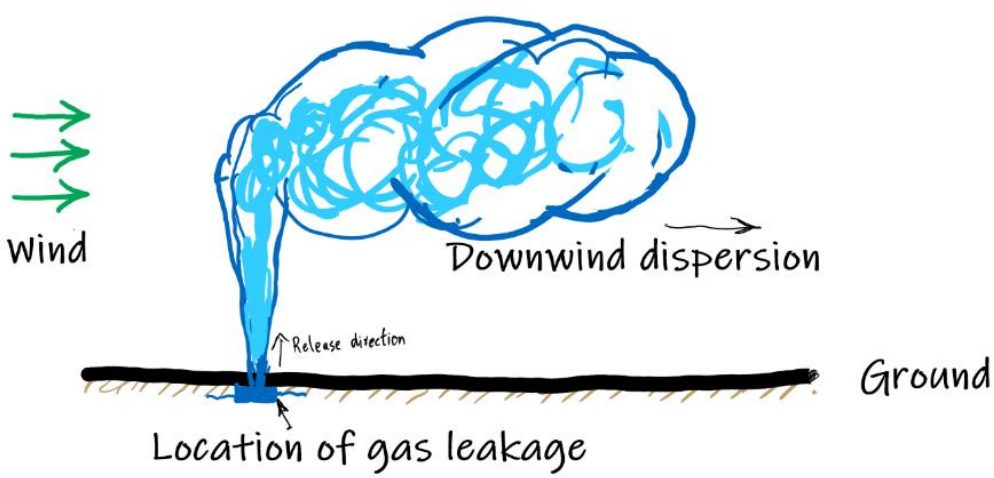


Figure 2: Scenario – toxic natural gas release from a pipeline and dispersion

Table 2: Input, parameters and sensitivity values

Input / Parameter	Base value	Sensitivity values
Release source term		
Hole size	2 in	1 in, 3 in, 4 in
Temperature	Medium (77 ºF)	Low (20 ºF), High (120 ºF)
Pressure	Medium (115 psia)	Low (50 psia), High (500 psia)
Orientation	Horizontal (1º from grade)	Upwards (45º from horizontal)
Environmental parameter		
Atmospheric stability	Stable (F)	Neutral (C, D)
Wind speed	Low (3.4 mph)	Med (13 mph), High (20 mph)
Humidity	Medium 50%	Low (20%), High (80%)
Terrain (surface roughness)	Med – Level country / Cut grass (0.2 in)	Low - Mud flats, Snow (0.0004 in) High – Wooded / urban area (3.9 in)

Toxic natural gas from eight reservoirs across different geographic regions were analysed. The natural gases considered (represented as S1 to S8) include H_2S composition ranging from 2% to 28% are shown in Table 3. The gas densities at 700 psia are shown at the gathering system supply pressure for a typical reservoir.

Table 3: Toxic natural gas composition (mol %)

Natural gas composition	S1	S2	S3	S4	S5	S6	S7	S8
Water	1.7%	3.0%	1.0%	0.7%	0.3%	0.0%	0.0%	1.0%
Nitrogen	0.1%	0.4%	0.5%	0.9%	1.2%	1.0%	1.1%	0.3%
H_2S	**2.6%**	**7.2%**	**9.6%**	**13.8%**	**14.1%**	**15.7%**	**17.0%**	**28.0%**
CO_2	5.4%	8.1%	11.2%	2.2%	3.3%	3.1%	8.3%	3.2%
C1	14.1%	49.6%	50.3%	78.1%	63.9%	57.4%	18.3%	35.0%
C2	28.0%	10.1%	9.5%	0.7%	10.1%	10.5%	24.0%	14.7%
C3	32.3%	9.7%	8.3%	0.9%	4.1%	5.8%	19.6%	10.7%
i-C4	3.3%	2.2%	2.0%	0.3%	0.6%	1.3%	0.0%	1.6%
n-C4	9.8%	4.4%	3.4%	0.7%	1.2%	3.0%	7.9%	3.1%
C5s	2.7%	3.4%	2.7%	0.7%	0.6%	1.8%	2.4%	1.6%
C6+	0.3%	2.0%	1.6%	0.9%	0.4%	0.5%	1.4%	0.8%
Molar mass (g/mol)	38.3	30.2	29.9	21.7	24.2	26.7	36.8	31.7
Gas density @ 700 psi	6.03	3.42	3.82	2.54	3.22	3.82	5.61	4.18

2.2 Criteria and Tools

Tools: The study utilized commercially available and validated tools (i) Aspen HYSYS for phase equilibrium estimations, and (ii) Canary by Quest for consequence (release and dispersion) modelling. Aspen's HYSYS is an Industry's leading process simulation software and Canary is an US EPA approved model atmospheric dispersion of natural gas.

Criteria: Toxic and flammable cloud dispersion to concentrations of personnel impact are analysed through modeling. The hazardous level (distance to the end points of interest) used for this study is given in

Table 4.

Table 4: Hazardous levels of pipeline release of natural gas

Component	Accidental consequence	Level-3	Level-2	Level-1
Natural gas (see values in Table 5)	Flash fire (flammable vapor cloud distance)	Upper Flammability Limit Methane (CH_4) 16% Propane (C_3H_8) 9.5%	Lower Flammability Limit (LFL) CH_4 4% C_3H_8 2%	50% LFL CH_4 2% C_3H_8 1%
Hydrogen sulfide (H_2S)	Toxic concentrations of exposure results in health effects or death	500 ppm potential for respiratory arrest, loss of consciousness	100 ppm Immediately dangerous to life and health (IDLH), coughing, dizziness	75 ppm Acute Exposure Guideline Level #3; loss of sense of smell in minutes

2.3 Simulations and results

Phase-equilibrium estimation (Aspen HYSYS): For this analysis, the HYSYS process simulator was used to perform flash and property calculations to better understand fluid phase behaviour under process and release conditions. The Peng-Robinson (PR) equation of state (EOS) was used because it provides better phase and equilibrium estimations close to/at the critical point as well provide better liquid densities estimations for gas and condensate systems when compared to Soave-Relich-Kwong EOS and Non-Random Two-Liquid EOS (Guerra 2006, Aspentech 2013). Furthermore, Aspentech, the licensor for the HYSYS software, has made several enhancements to the original PR EOS model to extend its range of applicability (Temperature, Pressure, and binary interaction parameters) to improve predictions of non-ideal systems (Aspentech 2013). Phase envelopes for each sample was generated using the phase envelope tool in HYSYS and used to verify canary input phase equilibrium calculations.

A phase envelope (isopleths or contour plot) is the Pressure-Temperate (P-T) projection of the phase diagram of a multicomponent system of fixed composition (see Figure 3). It is made up of two parts: (1) the bubble curve and (2) the dew curve. The bubble point curve is a curve that denotes the formation of the first bubble of vapor at a given pressure and temperature. Similarly, the dew point is a curve that denotes the formation of the first drop of condenses vapor at a given pressure and temperature. These two curves join at a critical point where the latent of vaporization is assumed to be zero denoting the co-existence of vapor, and liquid, phases. Transition from a liquid to a two phase and to vapor phase occurs in moving across from left to right.

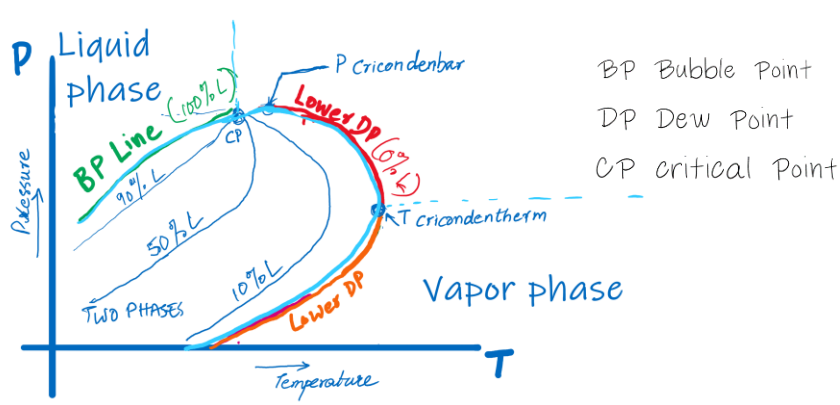


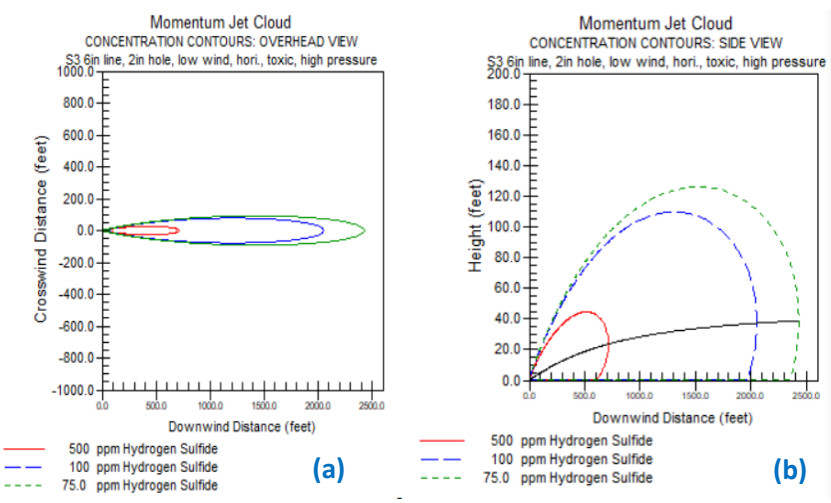
Figure 3: Multi-component Phase Diagrams

Distance to downwind dispersion estimation (Canary by Quest): Consequence modeling (release and dispersion) from 2-inch hole in horizontal direction for stable wind condition and low wind speed was carried out and considered as the base case for this study. The flow through pipeline is fixed at 50lb/s and release from a hole is assumed to be continuous (60 minutes) and disperses in an open field (no impingement). The number of components in the toxic natural gas compositions was optimized (given Table 5) for more accurate phase representation within the Canary multi-component model. Using the phase envelope generated from HYSYS, the input compositions were modified to ensure similar sample molar mass and H_2S composition which is acceptable for this comparative study. The analysis helped to understand the influence of user-adjustable parameters on model outputs.

Table 5: Multi-component compositions for release and dispersion modelling (mol %)

Natural gas composition	S1	S2	S3	S4	S5	S6	S7	S8
H_2S	2.6%	8.1%	10.4%	14.3%	14.6%	16.5%	17.8%	29.6%
CO_2	5.5%	9.1%	12.1%	2.2%	3.4%	3.3%	8.8%	3.4%
CH_4	14.3%	55.7%	54.5%	81.0%	66.1%	60.1%	19.2%	37.0%
C_2H_6	28.5%	11.3%	10.3%	0.8%	10.4%	11.0%	25.3%	15.5%
C_3H_8	32.8%	10.9%	9.0%	0.9%	4.3%	6.1%	20.6%	11.3%
C_4H_{10}	13.3%	4.9%	3.7%	0.8%	1.3%	3.1%	8.3%	3.3%
Molar mass (g/mol)	38.7	27.7	26.8	20.0	22.8	24.5	34.5	29.1
UFL	11.3%	15%	16%	16.6%	16.1%	15.8%	14.3%	16.6%
LFL	2.6%	4.0%	4.2%	4.8%	4.3%	4.1%	3.2%	3.7%

The results of dispersion were recorded for the maximum concentration along downwind central line concentration for an averaging time of 60 seconds. A typical output from Canary is given in Figure 4.



(a) Overhead (plan) view: illustrates the toxic natural gas footprint of three H_2S cloud concentrations in the downwind dispersion along the central line with the cloud width

(b) Side view: illustrates the elevation cross section of the cloud dispersion with the cloud height (black line indicates cloud central line)

Figure 4: Hydrogen sulfide momentum jet cloud - dispersion isopleths (a) Overhead view (b) Side view

Is should be noted that for risk assessment application, the width of the cloud and the averaging time plays a significant role (US EPA 2017).

2.4 Uncertainty and sensitivity analysis

For developing confidence in understanding a model, evaluate how variations in a model's outputs can be apportioned to variations in the inputs, which often referenced as sensitivity analysis. Sensitivity analysis approach by varying one input parameter at a time which holds other parameters at central values. The sensitivity outcomes are dependent on these central values. Each of the eight toxic gas compositions, were subjected to the sensitivity to the range of values for input and parameters. The results are presented using histograms or quantitative measures to compare the sensitivity of the uncertain input and parameter.

3. Results and discussion

This section reports the results of the simulations and discuss the sensitivity to the input and parameters. The aim is to identify the most important parameters from amongst a large number that affect model outputs. This will help in optimizing the time and resource usage for consequence modelling in risk assessment. The analysis is carried out on two sets:

- Material and release conditions (Source term): fluid composition, hole size, temperature, pressure, release orientation
- Environmental conditions: atmospheric stability, wind speed, humidity, terrain

3.1 Sensitivity: Fluid composition

The compositions analyzed include toxic gases with molar mass lower, similar and higher than that of air (28.9 g/mol). A comparison of the molar mass and H_2S composition used in this study is given in Figure 5.

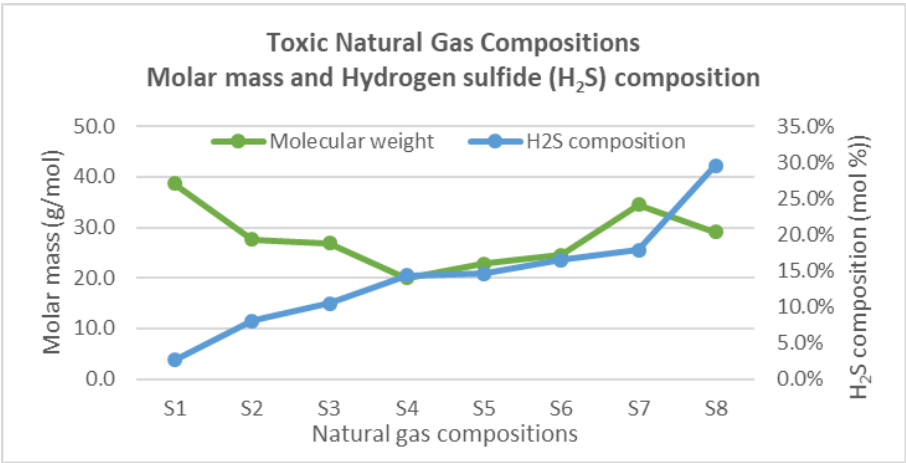


Figure 5: Composition comparison – molar mass

Over the years, certain heuristics have been used as source term input parameters for modeling multiphase releases and ensuing dispersion. Examples of these heuristics include choosing a pure component of the same molecular weight in place of the mixture, distilling mixture composition to a handful of components, choosing to model natural gas as a pure methane, etc. Although convenient, these modeling assumptions can result in hazard estimations that diverge from reality

with the biggest problem being the inability to accurately account for thermodynamic effects like phase splits and composition changes during release conditions (Johnson and Marx, 2003).

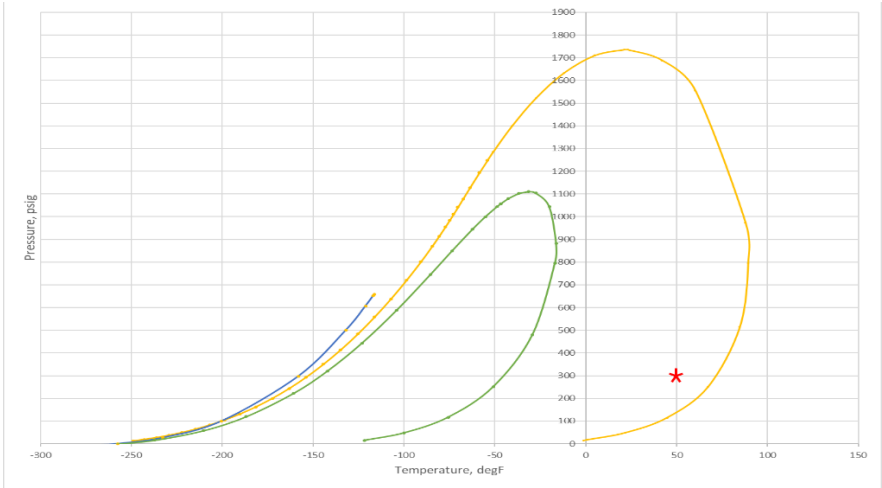
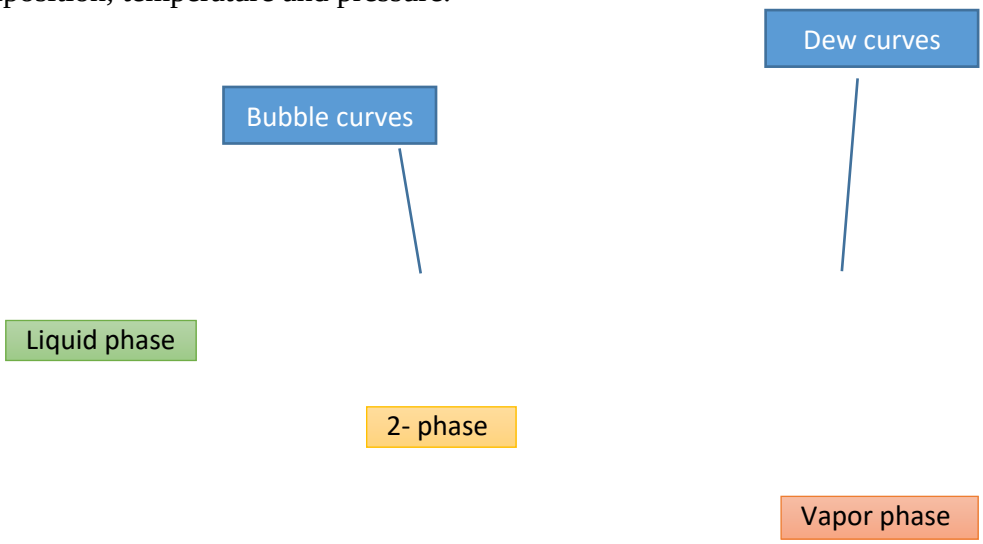


Figure 6: Phase equilibrium curves for methane (blue), Methane-ethane-Hydrogen sulphide (green), and S4 sample (yellow)

Consider the case in Figure 7 of a natural gas pipeline operating at 50 degF and 300 psia with S4 composition in Table 5. Going by the popular heuristic of modeling natural gas as 100% methane (blue line) it was observed that at pipeline operating conditions, the release is purely vapor with buoyant properties. Similarly, if the natural gas mixture (simplified C4) with three (78% CH_4, 8% C_2H_6 and 14% H_2S) components (green line), at pipeline vapor is mostly vapor too. However, a detailed composition of the mixture (S4, Table 5) reveals the release contains vapor, aerosol, and liquid phases which were missed earlier highlighting the importance of composition in dispersion modeling and the need to perform sensitivity analysis.

The phase envelope of eight toxic natural gas compositions given in Figure 7 illustrates that the phase of a multicomponent toxic natural gas could vary (liquid, 2-phase or vapor) with a change in the composition, temperature and pressure.



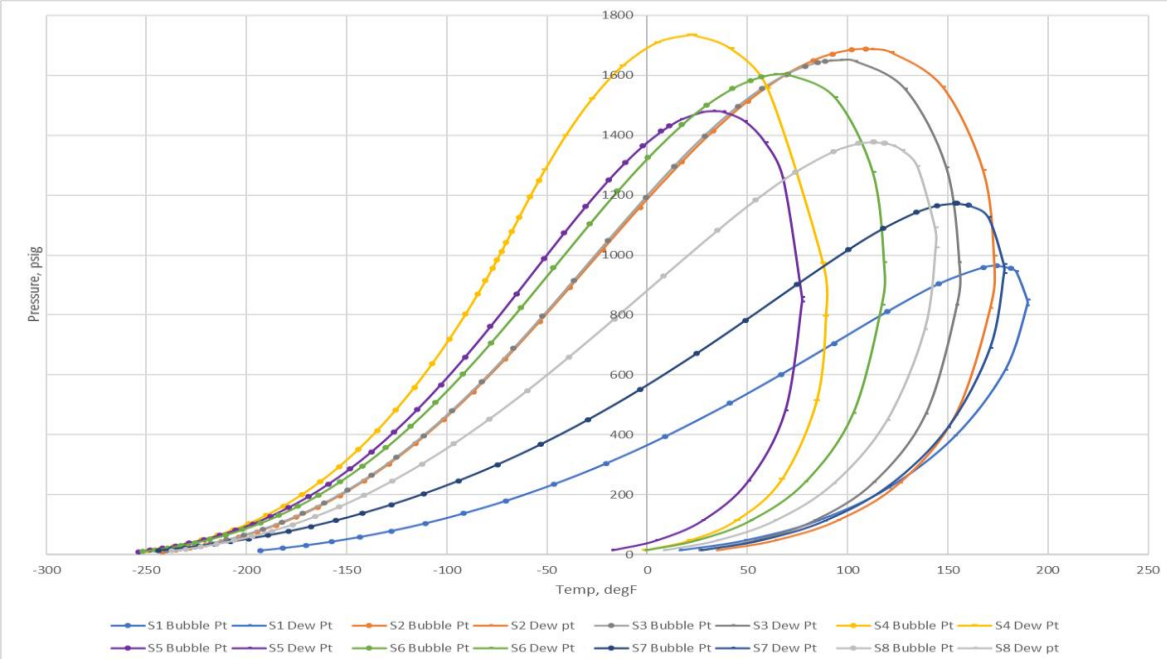


Figure 7: Phase equilibrium curve – toxic natural gas compositions

Density of fluid and related buoyancy (positive, neutral, negative) plays a major role is selecting the dispersion modelling approach (passive, dense etc) for estimating downwind distances (Nair & Wen, 2019). Released fluid density is driven by fluid's molar mass molar mass, release pressure and temperature. The Bubble curve and the Dew curves shift towards to right with an increase in molar mass (S1, S7, S8). This is due to the higher molar mass from higher composition of C4+ hydrocarbons and hydrogen sulphide contribution. The phase of the released material is critical since it determines the release and dispersion model used (e.g. heavy gas vs gaussian); an inappropriate selection can lead to erroneous results. For example, the fluid phase of S5 (MW 24.2) and S6 (MW 26.7) with similar molecular mass could yield different results for a given pipeline operating pressure and temperature (at 800 psig and 100 ^{o}F, S5 will be vapor, whereas S6 will be 2-Phase. Sensitivity to the changes in temperature and operating pressure was further analysed and given in section 3.3.

The downwind dispersion distances for the eight toxic natural gas compositions to toxic and flammable hazard levels are given in Figure 8. The downwind distances for LFL ranges from 27ft (S4, S5) to 60ft (S1) and H_2S 100ppm cloud ranges from 820ft (S2) to 1775ft (S8). The following observations inferred from the results:

i. Distance to H_2S toxic hazard level is significantly larger than flammability hazard levels. For example, results of toxic gas composition S2, toxicity downwind distance to 500ppm = 261ft and 100ppm = 820ft whereas the flammable cloud downwind distance UFL = 8ft and LFL = 30ft).
ii. Downwind distance of toxic dispersion is maximum for those release with higher compositions of H_2S (S7, S8) and with higher molar mass (S8, S7, S1).
iii. Downwind distance of toxic cloud dispersion is higher for toxic gas with higher H_2S composition (S8) while the downwind distance of flammable cloud dispersion is higher for composition with higher molar mass (S1, S7).

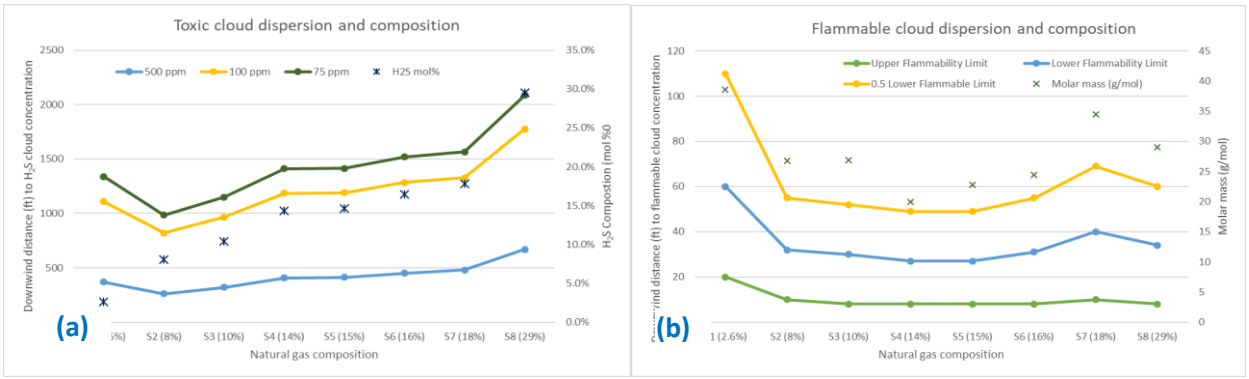


Figure 8: Sensitivity composition: Downwind distance to (a) H_2S concentration, (b) Flammable cloud

3.2 Impact of water vapor in natural gas

Well fluids may become saturated in the presence of produced water during production and transmission. As part of the analysis in this paper, the impact of water saturation on natural gas (with H_2S) dispersion in the event of a release was studied. Using the water saturate tool in Aspen HYSYS, S6 natural gas sample was saturated at 115 psig and 77 degF to estimate the new composition given in Table 6.

Table 6: Wet (saturated) and Dry base – natural gas (S6) compositions (mol%)

Component	Dry basis	Wet basis
H_2O	0.00%	**0.15%**
Nitrogen	1.00%	1.00%
H_2S	15.68%	15.66%
CO_2	3.10%	3.09%
Methane	57.34%	57.25%
Ethane	10.49%	10.47%
Propane	5.79%	5.79%
i-Butane	1.30%	1.30%
n-Butane	3.00%	2.99%
n-Pentane	1.80%	1.80%
n-Hexane	0.50%	0.50%

Consequence modeling was performed using Canary to assess the impact of water saturation on downwind dispersion to H_2S hazard level dispersion distance (see Figure 9).

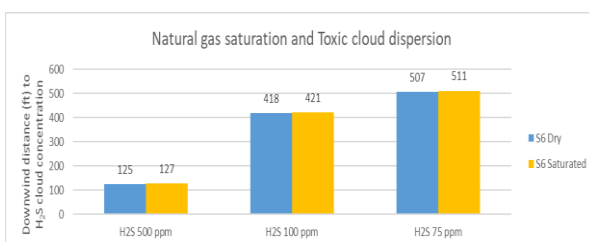


Figure 9: Sensitivity – natural gas saturation: Downwind distance to H_2S concentration

The results suggest that water saturation of natural gas was not a significant parameter in downwind dispersion to H_2S hazard levels.

3.3 Release source terms and sensitivity

3.3.1 Sensitivity – Release hole size: A representative hole size is assumed to represent a release resulting from loss of fixed equipment (pipeline) integrity (e.g. corrosion, erosion) or from operational upsets (e.g. blocked outlet). Three representative hole sizes (small, medium and large) was considered for the study. Release rates from three-hole sizes (1, 2 and 3 inch) and downwind dispersion for the eight toxic gas compositions were estimated, see Figure 10. The following observations were inferred from the results:

i. Release rates grow significantly with increase in hole size irrespective of the composition. For S1 composition, the release rates varied from 2.9 lb/s to 21.5 lb/s.
ii. Release rates were higher for compositions with larger molar masses (S1, S7) and the difference is significant for larger hole sizes.
iii. Similar release rates (e.g. 11.5 to 12.7 lb/s, 3-inch hole) for compositions with molar mass less than 27 g/mol (S2, S3, S4, S5, S6) for all hole size. However, significantly higher release rate (21.5 lb/s) for S1 with molar mass 39 g/mol.
iv. Downwind dispersion distance to 500ppm H_2S concentration from 1-inch hole releases was proportional to the H_2S composition.
v. Longest downwind dispersion reported (3inch releases), for 500ppm H_2S concentration was for S8 composition (28% H_2S, molar mass = 29 g/mol), while 100ppm was for S7 (18% H_2S, molar mass = 34 g/mol). Downwind dispersion following release from larger hole sizes are influenced by H_2S concentration and molar mass.

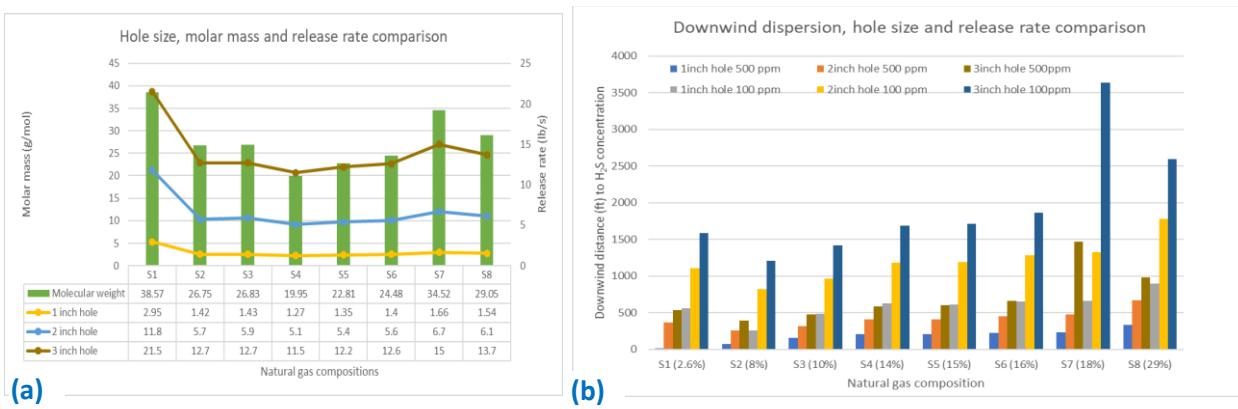


Figure 10: Sensitivity – Hole size: (a) molar mass and release rate; (b) downwind distance to H2S concentration

Downwind dispersion of toxic cloud is dependent on hole size, release rate and composition. The failure mechanism and related hole size (small, medium or large) need to appropriately be determined. For larger hole releases, the composition and molar mass is significant, whereas for smaller hole releases, the difference in composition did not have a significant impact to the downwind dispersion distances.

3.3.2 Sensitivity – Pressure: Release and dispersion at three pressure conditions (low = 50 psia, medium = 117 psia, high = 500 psia) at release for the eight natural gas compositions were compared. The release rates for S8 composition varied from 2.6 lb/s (low pressure) to 50 lb/s (high pressure) and the downwind distances for 500ppm varied from 435 ft (low pressure) to 4050 ft (high pressure). Following observations are inferred from the results given in Figure 11:

i. The release rates (2 to 3 lb/s) for all compositions were similar at low pressure;

ii. The release rates (4.5 to 5.5 lb/s) were comparable for compositions S2 to S8 at medium pressure, but higher (11.8 lb/s) for S1 with highest molar mass.
iii. For high pressure, the compositions (S1, S7, S8) with higher molar mass (>29 g/mol) have significant higher release rates (>38 lb/s) compared to the compositions (S2, S3, S4, S5, S6) with lower molar mass (<29 g/mol).
iv. For high H_2S compositions (S7, S8), the dispersion distances were significantly longer for high pressure releases (500 ppm exceeds 2750 ft compared to less than 1000 ft for natural gas with less than 18% H_2S).

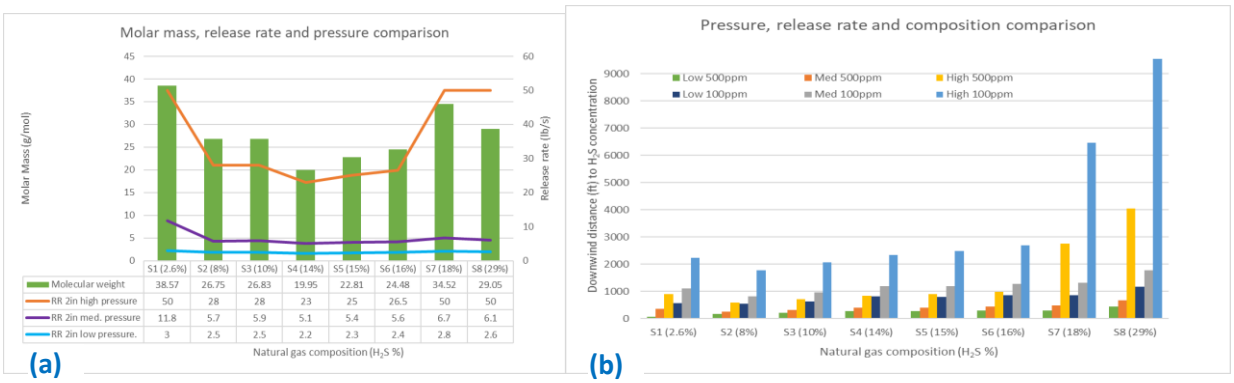


Figure 11: Sensitivity – Pressure: (a) molar mass and release rate; (b) H_2S downwind distances

Release rates are dependent on operating pressure and have higher release rates for higher pressure for all compositions considered. Downwind dispersion for high pressure releases are sensitive for compositions with greater than 18% H_2S content. For such cases with significantly higher impact zone, further analysis should be carried out before implementing risk reduction measures.

3.3.3 Sensitivity – Temperature: Release and dispersion at three temperatures (low = 20°F, medium = 77°F, high = 120°F) for different natural gas compositions were compared. The release rates varied from 2.2 lb/s (S4 low pressure) to 50 lb/s (maximum) and the downwind distances for 500ppm varied from 68ft (S1 low pressure) to 4050ft (S8 high pressure). For the study, the atmospheric temperature and surface temperature also has been assumed as the same temperature as that of the fluid temperature. Following observations are inferred from the results given in Figure 12:

i. For medium and high temperature conditions, the release rates are similar irrespective of the compositions.
ii. Similar release rates (~5 lb/s) were estimated for compositions with molar mass 29 g/mol and less for the range of temperatures evaluated.
iii. Significantly higher release rates were estimated for compositions with molar mass greater than 30 g/mol under low temperature conditions.
iv. For all three temperature conditions, downwind dispersion distances similar for all compositions with molar mass less than 30g/mol.
v. Downwind dispersion distances for composition with greater than 30g/mol similar for medium and high temperature, whereas significantly higher for low temperature releases.

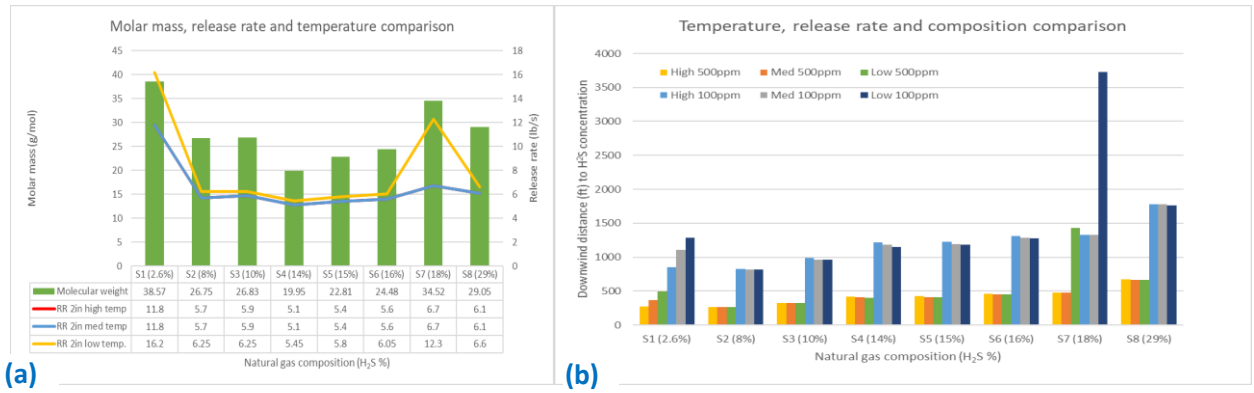


Figure 12: Sensitivity – Temperature: (a) molar mass and release rate; (b) H_2S downwind distances

Release rates and downwind dispersion are sensitive to low temperature for those compositions with >30 g/mol. For such cases with significantly higher impact zone, further analysis should be carried out before implementing risk reduction measures.

Sensitivity – Release orientation: Release and dispersion from two release orientations, horizontal and upwards (at 45deg from horizontal) for the eight natural gas compositions and from 2-inch hole at 77oF and 115psia were compared. Release rate for each composition will be the same for both orientations. The orientation options were limited to two as the scenario considered was at the ground level. 500ppm downwind distance for horizontal orientation ranges from 260ft (S2) to 668ft (S8) whereas for upwards ranges from 20ft (S1) to 335ft (S8). Following observations were inferred from the results given in Figure 13:

i. Downwind dispersion distances are higher for horizontal orientation compared to upwards orientation for all compositions.
ii. For both orientation, downwind dispersion distances for 500ppm and 100ppm were similar for compositions with H_2S concentrations 14% to 18% (S4, S5, S6, S7), but significantly lower for compositions with low (<10%) H_2S concentrations (S1, S2) and significantly higher for compositions with high (>20%) H_2S concentrations (S8).
iii. For dispersion from upwards releases, the downwind dispersion distance increases with the increase in H_2S concentration.
iv. For dispersion from horizontal release, S1 with 2.6% H_2S (highest molar mass and release rate) dispersion distances are higher than S2 (8% H_2S) and S3 (10% H_2S).

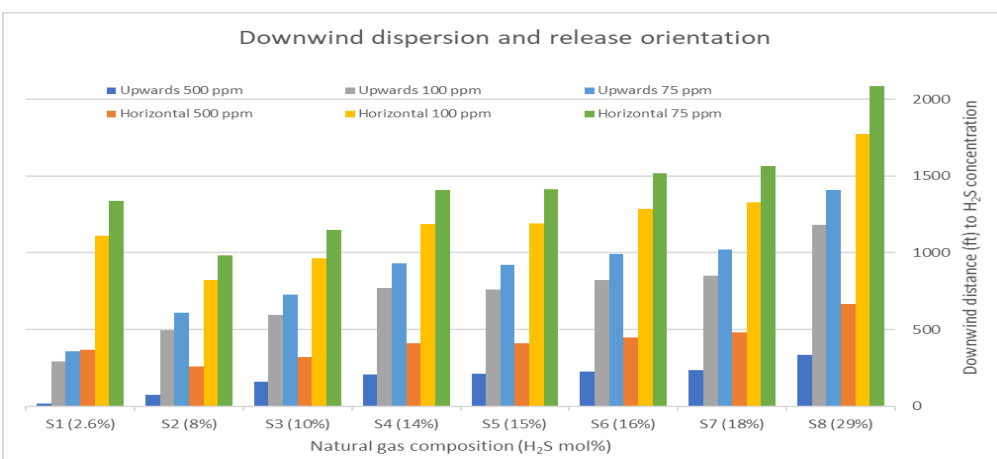


Figure 13: Sensitivity – Release orientation: H_2S downwind distances

The analysis implies that downwind dispersion is sensitive to the orientation of release. Hence, appropriate orientation based on the failure mode and expected location (elevation) of the receptors of concern should be used for consequence modeling.

3.4 Impact zone: Environmental parameters

3.4.1 Sensitivity – Atmospheric stability and wind speed: Dispersion for different natural gas compositions and from 2-inch hole at 77oF and 115psia under three atmospheric stability conditions and wind speeds (3.4F: stable and low wind speed, 13D: Neutral and medium wind, 20C: slightly unstable and high wind) were compared. Following observations, were inferred from the results given in Figure 14:

i. The longest downwind dispersion irrespective of the composition was recorded for stable conditions and low wind speed.
ii. For dense gas (negatively buoyant) compositions (S1, S7) with higher molar mass (>29 g/mol), the downwind dispersion for Neutral and Medium wind (13D) was significantly higher.
iii. For lightly unstable and high wind speed (20C) conditions, the downwind distances for 100 ppm was less than 200 ft for all compositions whereas for stable and low wind speed (3.4F) conditions, the distances exceeded 800 ft.
iv. For compositions with molar mass <29 g/mol (positively buoyant), the downwind distance for 20C conditions are higher than 13D conditions. Under these conditions, the cloud is behaving more as heavy gas and closer to ground level, whereby higher concentration cloud travels further downwind.

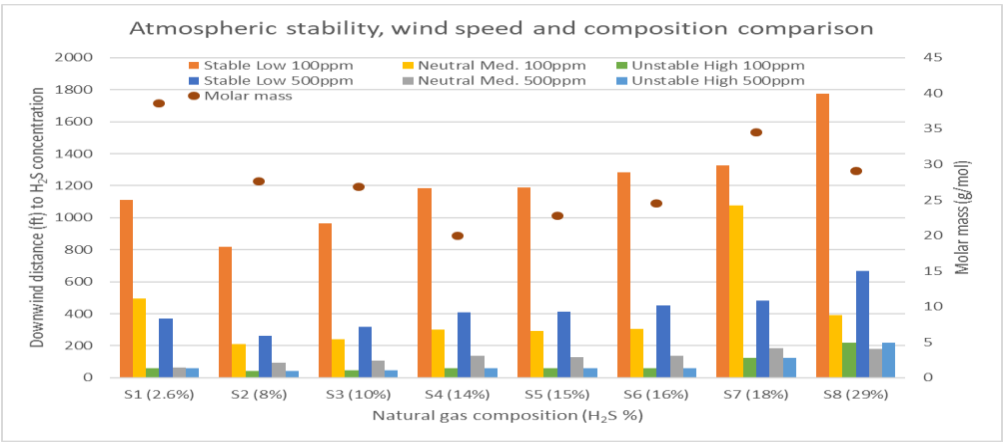


Figure 14: Sensitivity – Atmospheric stability and wind speed: H_2S downwind distances

Note: Canary tool couples (transition) from jet dispersion to heavy gas dispersion when the central line touches ground level. This modeling factor is reflected in results for S7 under 13D conditions and for all compositions under 20C conditions.

For S8 composition with higher H_2S concentration, the downwind distance to 100 ppm extends to 1775 ft at low wind and stable conditions (3.4F) compared to 390 ft and 220 ft for neutral stability and higher wind speeds. For a location with predominant neutral stability and medium wind speed (like 13.4D), if the risk management bases the impact zone distance worst-case stability and wind (1775ft) which is about 5 times typical (390ft), then the risk management (e.g. emergency

planning) incur significantly higher cost and effort. This comparison highlights the importance of determining wind speed and stability class appropriate for the location. It is however, advisable to have a range of stability and wind speed to represent the variations during 24 hours and through the year.

3.4.2 Sensitivity – Terrain: Dispersion for different toxic gas compositions and from 2-inch hole at 77°F and 115 psia over three different terrains (mud flat, level country or cut grass, urban area) were compared. The terrains were considered flat (without obstructions) and the turbulence from terrains were addressed by surface roughness parameter as given in Table 2. Following observations are inferred from the results given in Figure 15:

(i) With increase in surface roughness, the downwind dispersion decreases. Downwind dispersion distances for Urban area was significantly lower than (1/3rd) for all compositions except S1.

(ii) Downwind dispersion distances for Mud flat and Cut grass is similar for all compositions except S1 with the highest molar mass. This implies that dispersion of toxic gas with less than 35 g/mol molar mass is not sensitivity to surface roughness <0.2 inch.

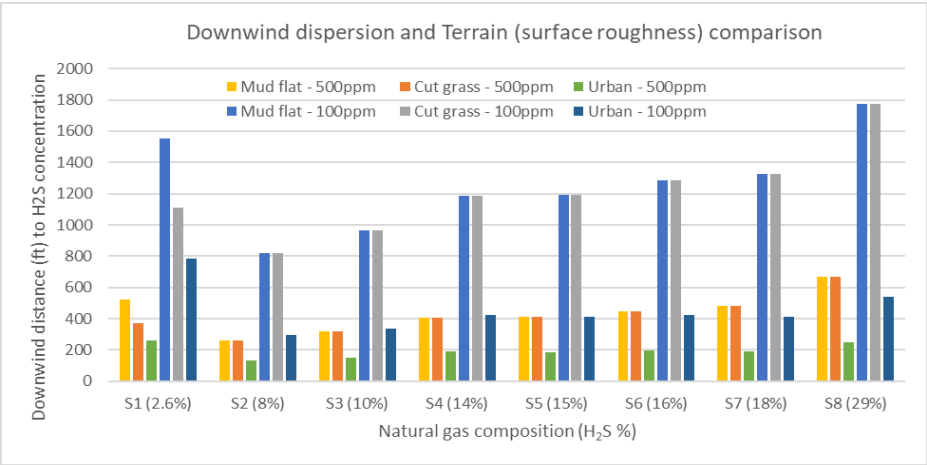


Figure 15: Sensitivity – Terrain: H_2S downwind distances terrain

3.4.3 Sensitivity – Humidity: Dispersion for different toxic gas compositions and from 2-inch hole at 77°F and 115 psia at three humidity conditions (low =20%, medium = 50%, high =80%) were compared. Results given in Figure 16 implies that humidity has no significant impact on the downwind dispersion of toxic natural gas.

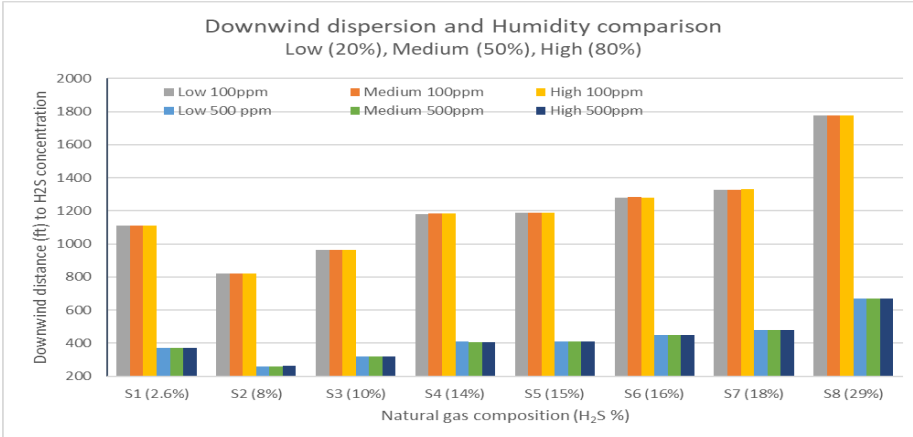


Figure 16: Sensitivity – Humidity: H_2S downwind distances

3.5 Discussion

The statement is often made that natural gas is lighter than air and the properties of a mixture is determined by the mathematical average of the properties of the individual constituents. Such mathematical bravado and inconsistency of thought is detrimental to safety and must be qualified (Speight, 2011). During expansion from elevated pressure, released toxic gas could be colder and heavier than air close to the release source with the potential to accumulate in low-lying areas (Nair & Wen 2019).

3.5.1 Findings: From the range of simulations (using HYSYS) and consequence modelling (using Canary), it was concluded that for a similar type of release event, the toxic hazard impact zone could be orders of magnitude different. Comparative study was carried out for eight different toxic natural gas compositions with H_2S concentration ranging from 2.6% to 29%. It was observed that the downwind distances to hazardous levels ranges from less than 50 ft to more than 5000 ft for a loss of containment from toxic natural gas pipeline transfer line. The range of results were obtained by varying input on the release (source term) conditions and certain environment conditions. From the parametric sensitivity analysis for a release event from a natural gas transfer pipeline at ground level using eight different compositions, the following observations and recommendations were made:

1. Downwind dispersion distances to concentrations of interest (impact zone) is dependent of the natural gas composition and release rates. Detailed assessment should be carried out taking account of the phase and component characteristics of the fluid in question.
2. Selection of appropriate evaluation criteria (hazardous levels – toxic, flammable) is critical in determining the impact zone from accidental releases. For natural gas with toxic (H_2S) component, the potential impact zone of concern will be dominated by toxicity.
    a. Flammable cloud dispersion distance is dependent on the molar mass of natural gas composition; significant longer distances for natural gas with molar mass greater than air (>29 g/mol).
    b. Toxicity impact zone is dependent on the H_2S composition along with the molar mass. Significantly longer downwind distance of toxic cloud dispersion for natural gas with higher (>18 mol%) H_2S concentration.
3. Phase equilibrium properties of the release should be considered in determining the release phase as low temperature and high-pressure releases can have longer impact zone

distances. Detailed review (prior to implementing risk mitigation) should be carried out for high pressure releases of compositions with >18 mol% H_2S & molar mass >29 g/mol and for low temperature releases of compositions with molar mass >30 g/mol.

4. Selection of representative release hole size and orientation of the release have significance in the impact zone estimation. The cause and mode of failure need to be evaluated to determine representative source term.
5. Natural gas cloud dispersion is sensitive to the turbulence related parameters, i.e. stability class, wind speed and surface roughness. Following environmental parameter selection and sensitivity evaluations are suggested:
    a. Terrain effects for dense (molar mass > 30g/mol) toxic natural gas with molar mass, surface roughness selection to evaluate terrain for low (cut grass) and medium (process plant / urban).
    b. Site specific set of atmospheric stability and wind speed to be selected to represent the predominant conditions. Sensitivity to be evaluated for higher and lower wind speeds and corresponding atmospheric stabilities.

3.5.2 Application: Role of consequence modeling results in the risk management efforts is analysed with the study findings. The results from the parameter sensitivity analysis for natural gas composition S4 transposed to geographical location. The potential impact to public (personnel) corresponding to each impact zone radius was estimated for comparing the levels of risk. A comparison with composition S7 and possible risk management considerations are also discussed. The downwind distances to 100 ppm H_2S cloud is summarized in Table 7.

Table 7: Natural gas (S6) compositions (mol%) and downwind distance to 100ppm H_2S

Case sensitivity (distance in ft)	S4	S7
Molecular weight	20	35
Base case (2in, 3.4F, 77oF, 115psia)	1185	1327
Sensitivity: Temperature – Low (20oF)	1150	3725
Sensitivity: Pressure – High (500psia)	2330	6450
Sensitivity: Wind & Stability – Medium, Neutral	300	1075
Sensitivity: Surface roughness – High (0.1m)	420	410

Impact zones for selected few cases are illustrated in

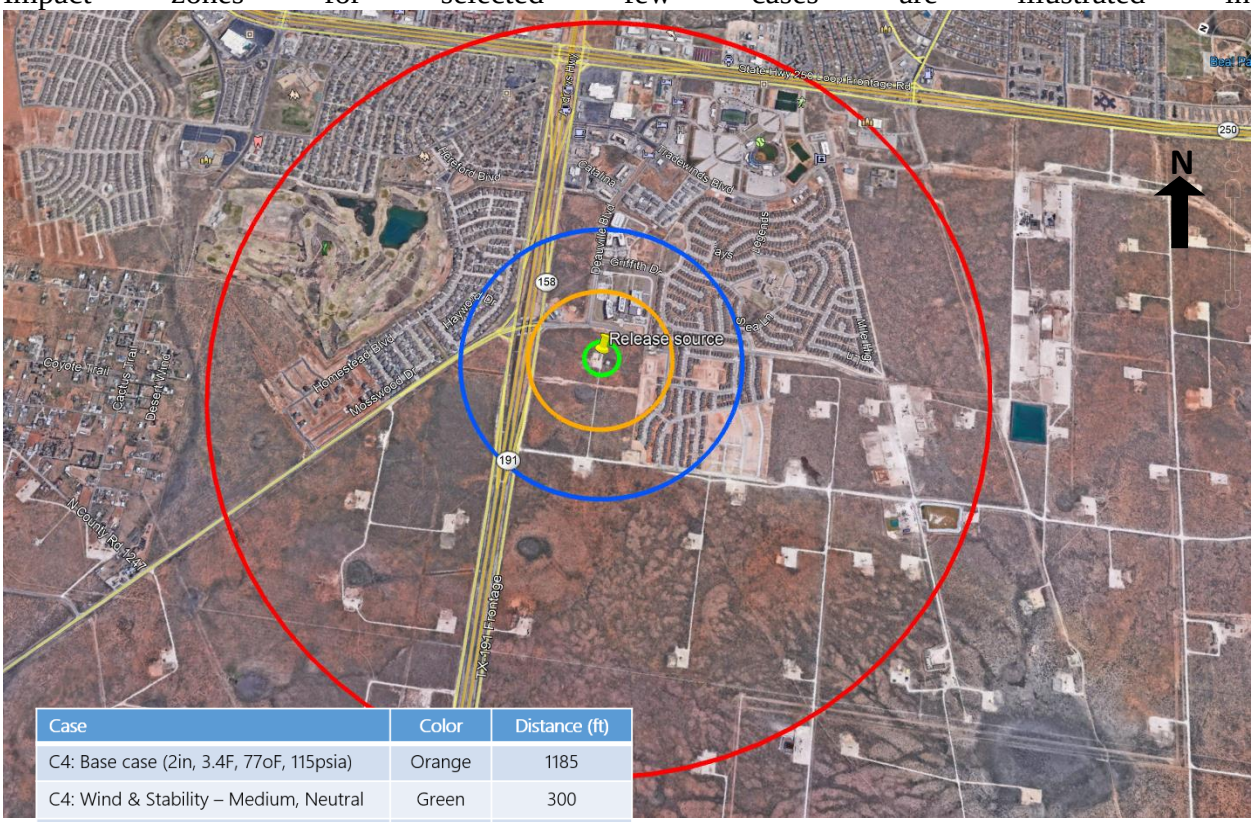


Fig wnwind distances and potential impacts, the yellow pir ed circles represents the impact zone for different set of input and parameters are given in Table 7. The impact area for a release event will be a section of the circle with orientation dependent on wind direction.

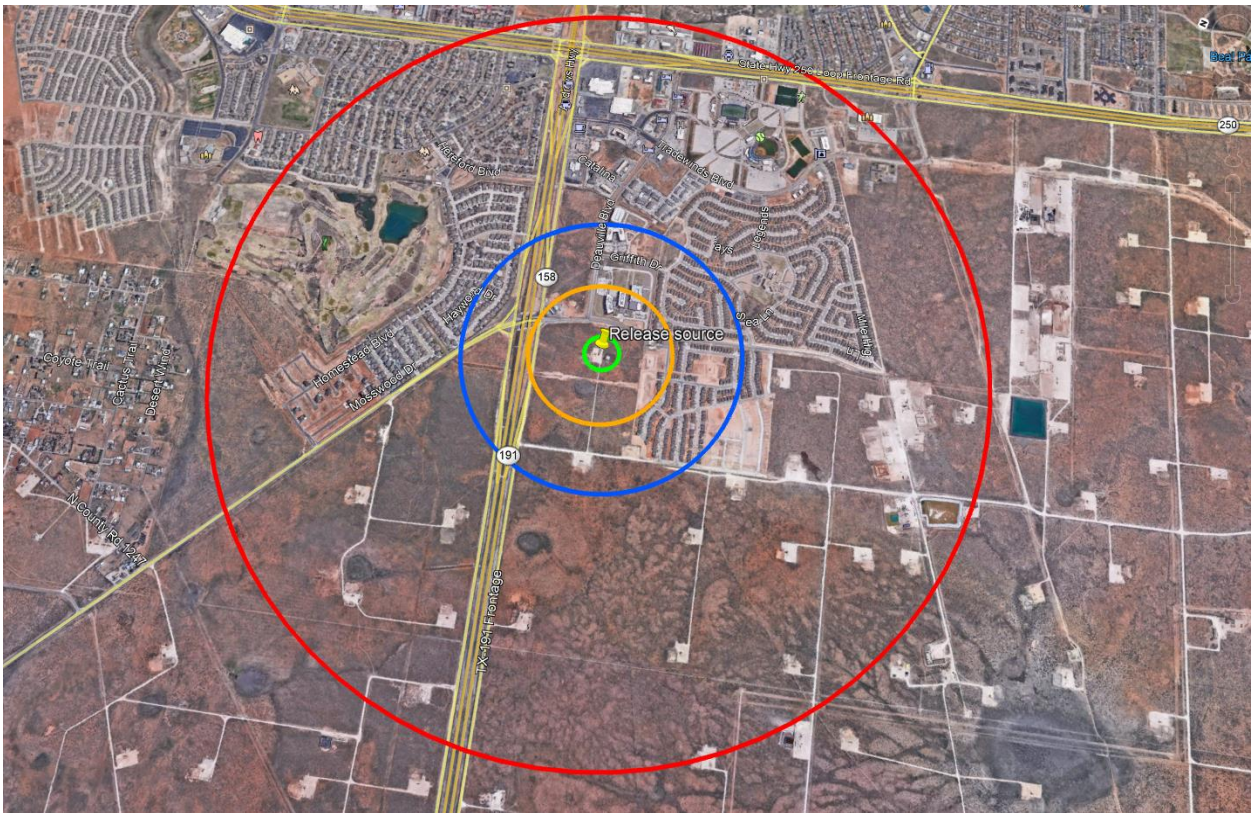


Figure 17: Parameter sensitivity summary - H_2S downwind distances and potential impacts

The representative set of cases with impact zones, corresponding potential consequence and risk management considerations are given in Table 8. The base case impact zone (Orange color and radius 1185 ft), the 100 ppm H_2S cloud (IDLH – concentration level) could reach an office building or residential area. This implies that in the event of a release under the given base case conditions and Southerly (towards North) wind, more than 500 personnel could be exposed to natural gas cloud with 100 ppm or more for a period until the release is isolated and such an exposure could result in coughing and dizziness. Risk reduction measure considerations should be to reduce the impact zone radius including reducing the pipeline diameter or restricting the horizontal release orientation (e.g. laying pipeline underground). However, modeling using the site-specific representative wind speed and atmospheric stability (13D - medium and neutral) instead of worst-case conditions (3.4F – stable and low wind conditions), the impact zone estimated was much smaller (300 ft, Green color). The impact zone was limited to the facility surroundings (without personnel exposure) and whereby the risk management limits were limited to maintaining the exclusion zone (restricting personnel access / habitats). Similarly, for the impact zone and potential consequences for operating under higher pressure or for S7 composition is given in Table 8.

Table 8: Natural gas impact zone – parameter sensitivity and risk management considerations

Case sensitivity	Color	Consequence / concern	Risk management considerations
C4: Base case (2in, 3.4F, 77oF, 115psia)	Orange	500+ (1 x Office, 30 houses)	Perform site specific assessment Risk reduction through buried lines, smaller diameter pipeline
C4: Wind & Stability – Medium, Neutral	Green	Environmental impact	Manageable risk, maintain exclusion zone

C4: Pressure – High (500psia)	Blue	2000+ (2 x office, 100+ houses)	Operational controls (e.g. at lower pressure)
C7: Pressure – High (500psia)	Red	25,000+ (Ball park, Supermarket, neighbourhoods)	Elevated risk, consider alternate route

A worst-case consequence modelling estimate may not be the best for risk management, instead a 'credible' worst-case scenario need to be determined and subjected to consequence modeling. The credibility of a set of modeling input should be determined considering the site specific operating conditions, fluid characteristics, type of failure and likelihood of environmental conditions. Once the risk levels are evaluated, sensitivity analysis on modeling input can be used further to determine the risk management efforts.

4. Concluding remarks

Numerical simulation of release and dispersion of natural gas provides an enhanced information on the potential impact zone which forms an essential part for risk-based decision making, especially in engineering projects and emergency planning. For toxic natural gas, with components like Hydrogen Sulfide (H_2S), the toxicity impact zone drives business decisions related to equipment design, facility siting, layout, land use planning and emergency response measures. The study focused on potential accidental release from pipeline at ground level transferring toxic natural gas. Eight natural gas compositions were subjected to a range of release source terms and environmental parameter sensitivity analysis. The multi-component phase diagram was developed using HYSYS and release followed by and dispersion were estimated using Canary. Analysis was carried out for by changing one parameter at a time for release and environmental conditions.

The analysis concludes that the release and dispersion of toxic natural gas is significantly impacted by (i) fluid composition, molar mass and fluid phase, (ii) release hole size and orientation, (iii) low temperature and high pressure, and (iv) surface roughness, wind speed and stability. The study findings highlights the significance of the use of a multicomponent consequence model when the potential for the formation of a two-phase system exists. Sensitivity modelling for the key parameters is the suggested approach to overcome this challenge. Incorrect selection of the modeling approach, input and environmental parameters could lead to an inaccurate consequence impact zone estimation which could result in disproportionate risk management efforts. this challenge can be addressed by better understanding of the cloud behaviour following release and sensitivity analysis of the modeling inputs and parameters.

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Modelling and Simulation to Predict Energetic Material

Hongwei Xi and Kok Hwa Lim*
Singapore Institute of Technology
10 Dover Drive, Singapore 138683

*Presenter E-mail: kokhwa.lim@singaporetech.edu.sg

Abstract

With the rapid development and advancement in computing power, modelling and simulation (M&S) has demonstrated its vast potential in predicting the properties of energetic material and helping to design energetic material. One such application is predicting crystal packing and crystalline structure from first-principle simulation. Such technique has demonstrated the ability to distinguish different polymorphs of the same energetic molecules and accurately predict the crystal structure and density. In addition to the ability to predict detonation pressures and velocities of more established classes of energetic materials based on their thermochemical code or empirical equations, M&S has also demonstrated its ability to screen designed energetic materials for potential application. The application of M&S vastly improves the safety of developing potential energetic materials - the ability to screen potential energetic materials based on M&S-predicted heats of formation and detonation properties means that less hazardous experiments are required to be conducted as well as reducing developmental cost.

Keywords: energetic material, safety, modelling and simulation, density, heat of formation, detonation property

1. Introduction

Energetic materials, such as propellants, pyrotechnics, and explosives, belong to a special class of materials, which have many industrial and civic applications, but pose dangers and hazards at the same time. In recent decades, active research in the design and synthesis of novel high energy density materials (HEDMs) led to the development of many novel HEDMs, for example, hexanitrohexazaisowurtzetane (HNIW or CL-20) [1], 1,3,3-trinitroazetidine (TNAZ) [2], octanitrocubane (ONC) [3] and 1,1-diamino-2,2-dinitroethene (FOX-7) [4], were designed and synthesized, but the experimental exploration of HEDM can be a rather hazardous process.

With the rapid development of simulation technique and computing power, modelling and simulation (M&S) has become an important auxiliary tool for the design and development of HEDMs. Theoretical M&S for HEDMs can evaluate all their material properties more safely, efficiently and economically, which include crystal form and density, heat of formation and detonation pressures and velocities. Thus, M&S is not only a precise predictor of material properties of classical and new-synthesized energetic material, but also a screening tool for potential high energy density molecules (HEDCs) from plenty of theoretical target for the future experimental synthesis. This M&S screening tool for HEDMs, help to reduce the number of hazardous, time-consuming and cost in the rapid advancement and development of HEDM.

2. Determining crystal form and density

The crystal density is a very important parameter in HEDM and new energetic materials must first achieve the density of more than 1.90 g/cm^3 at the present stage. To date, several theoretical methodologies have been developed to predict the crystal densities of HEDM based on their geometric structures or electronic structure.

In the past, a frequently used method is the group additivity method (GAM) [5]. GAM divided an energy density molecule into several appropriate functional groups. Then, the volume of this energy density molecule is evaluated by summing up all the volumes of these fragments. In recent years, GAM method has been updated to consider intermolecular interactions of molecules in the solid state. However, it does not include the effect of crystal form on crystal

volume. Moreover, GAM method relies heavily on the experimental data: when novel promising HEDC possesses the special group beyond the GAM database, GAM method often finds itself limited in application.

Based on the calculation of quantum mechanics (QM) and the analysis of the electron cloud of HEDC, an initial theoretical method predicting the density of HEDC was advised [6-7]. The density accuracy of the QM method is derived entirely from the appropriate theoretical method and high calculational level, not a response to the experimental data. Theoretical molecular density of each target HEDC needs to exactly evaluate the volume of 0.001 electrons $Bohr^{-3}$ electron density envelope, which is computed by Monte Carlo integration. In order to ensure the accuracy of Monte Carlo integration, QM volume of HEDC is gained from the arithmetic average value of more single-point molar volumes, for example, more than 100 times. QM density of HEDC is highly consistent with the experimental data (see Table 1) and thus it has been widely adopted at the exploration stage of promising HEDMs.

Table 1. Experimental[a] and theoretical, calculated at the B3LYP level, density (at g /cm) of TNT, RDX, HMX.

	TNT		RDX		HMX	
	Expt.[a]	Theory	Expt.[a]	Theory	Expt.[a]	Theory
Density	1.65	1.63	1.81	1.79	1.90	1.88

[a] see reference [8].

Since the predicted density from QM calculations does not take into account intermolecular interactions and crystal packing, and QM method is also unable to distinguish between different polymorphs of the same HEDC, to increase the accuracy of HEDMs prediction, we advised a new technique to evaluate the crystalline density of HEDM, which includes two consecutive steps: crystal packing of molecules and first-principle simulation of crystalline HEDMs [9].

Our calculations have shown that the crystal form of the same HEDC molecule is an important factor to determine the crystal density of HEDM [9]. It is even more remarkable that the new method can discover the promising crystal forms of synthesized energetic materials, which are more thermodynamically stable, possess higher density, and yet experimentally evasive

polymorphs. [10]

3. Evaluating of heat of formation

Heats of formation (HOF) is a key parameter predicting explosive performances of HEDMs. Since experimental measures of HOF for HEDMs is difficult and hazardous, theoretical measures using computer codes have become a popular subject in the exploration of HOF of energetic material.

In general, HOF of HEDMs is always calculated at the level of Density Functional Theory (DFT) [11-12], which can accurately estimate HOF for HEDMs and avoid the shortcomings of other theoretical methods. Since DFT method includes the electronic correlation and its calculations is not expensive, DFT method, especially B3LYP [13-15] methods, can be employed to estimate HOF for most HEDMs. However, the theoretical method needs specifically-designed isodesmic reactions, during which the target molecule of energetic material need be broken down into several small molecules containing the same component bonds, thus, different isodesmic reactions can generate different HOF data for the same molecule at the same computational level. In addition, there must be accurate experimental HOF values for small molecules generated in the isodesmic reactions. Thus, there remains some disadvantages in the method of isodesmic reactions.

At present, a theoretical method with sufficient accuracy [16], the atomization scheme together with high-level calculational models, which include the Gaussian-n (G2, G2(MP2), and G3) [17-19] and complete basis set (CBS-4M, CBS-Q, and CBS-QB3) [20-22] models have been well developed to evaluate HOF of energetic material. The method can accurately estimate HOF with mean absolute deviations of less than 4.0 kJ/mol from experimental data (see Table 2).

Table 2. Experimental and theoretical, calculated at the G2 and CBS-Q level, heat of formation (at kJ/mol) of oxazole, pyrazole and1*H*-tetrazole.

	oxazole		pyrazole		1*H*-tetrazole	
	Expt.[a]	Theory	Expt.[a]	Theory	Expt.[b]	Theory
HOF	-15.50	-17.03[c] -16.02[d]	179.40	183.01[c] 180.37[d]	333.46	334.59[c] 332.79[d]

[a] See references [23] and [24].

[b] Experimental HOF of 1*H*-tetrazole come from the sum of HOF (237.23 kJ/mol, at 298 K and the solid phase) and the enthalpy of sublimation (96.23 kJ/mol).

[c] Theoretical HOF calculated at the G2 level, see reference 16.

[d] Theoretical HOF calculated at the CBS-Q level, see reference 16.

4. Predicting of detonation perform

Detonation velocity (*D*) and detonation pressure (*P*) are two essential parameters, which are used to characterize detonating performance of HEDMs. Detonation velocity represents how fast the detonation wave propagates and detonation pressure is the value of the Chapman-Jouguet pressure from shock wave impedance measurements.

Since experimental data is lacking new HEDMs, detonation velocity and detonation pressure of organic CHNO explosives are traditionally predicted by applying the empirically derived Kamlet-Jacobs equations: [25]

$$D = 1.01\ (NM^{1/2}Q^{1/2})^{1/2}(1+1.30\rho_0) \quad (1)$$

$$P = 1.558\rho_0^2 NM^{1/2}Q^{1/2} \quad (2)$$

where *N* is moles of detonation gases per gram of explosive, *M* is average molecular weight of gases, *Q* is chemical energy of detonation, $\rho 0$ is the density of explosive.

Although the Kamlet–Jacobs equations were derived decades ago, they are still being applied to predict the detonation properties for many CHNO HEDMs, especially for new HEDM and theoretically designed potential HEDM. For example, the Kamlet–Jacobs value of detonation

velocity and detonation pressure for RDX are theoretically estimated as 9.03 km/s and 35,2 GPa [26] respectively, which are close to the experimental values of 8.754 km/s and 33.8 GPa [27].

In recent years, it has become increasingly popular to predicted detonation performance using existing thermodynamic computer codes. At the present stage, detonation parameters of energetic materials can be more effectively calculated using the program, EXPLO5™ [28]. Using both the density and heat of formation for the crystal of HEDMs, EXPLO5™ enables the calculation of detonation parameters such as detonation velocity, detonation pressure and detonation energy based on the Chapman-Jouguet detonation theory. The detonation velocity and detonation pressure can be derived with an error range of less than 10%. Furthermore, the ability of EXPLO5™ goes beyond the system of CHNO explosives, whose database includes 38 elements (C, H, N, O, Al, Cl, Si, F, B, Ba, Ca, Na, P, Li, K, S, Mg, Mn, Zr, Mo, Cu, Fe, Ni, Pb, Sb, Hg, Be, Ti, I, Xe, U, W, Sr, Cr, Br, Co, Ag, and Zn) such that EXPLO5™ can predict detonation performance of a wide variety of HEDMs.

Table 3. Detonation velocity (*D*, at km/s) and detonation pressure (*P*, at GPa) of TNT, RDX, HMX from experiment[a] and EXPLO5™ calculation[a].

	TNT		RDX		HMX	
	Expt.[a]	EXPLO5	Expt.[a]	EXPLO5	Expt.[a]	EXPLO5
D	6,930	6,809	8,700	8,793	9,110	9,179
P	19.0	18.70	33.8	33.68	38.7	37.82

[a] see reference [27].

6. Concluding Remarks

Although experimental data of energetic material is preferred over the modelling and simulation data, reliable experimental data of HEDM is not often available as its experiments is always hazardous. Researches on new HEDM are always short of experimental data, especially in the exploration of novel HEDM materials.

With the rapid development and advancement of simulation technique and computing power, M&S has shown its vast potential in predicting the properties of energetic material and screening the promising HEDCs from a plenty of theoretical targets. As the M&S results collaborate with the experimental data, it has turned into an indispensable tool to predict the material properties with accuracy and check the reliability of theoretical target.

Since M&S initially evaluated the material properties and predicted detonation perform only based on the calculation of quantum mechanics, beyond the experiment, it can effectively avoid the potential hazards of HEDM experiments. At the same time, M&S can reduce the cost of experimental design and synthesis of potential HEDMs.

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MARY KAY O'CONNOR PROCESS SAFETY CENTER
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23rd Annual Process Safety International Symposium
October 20-21, 2020 | College Station, Texas

Safety Assessment of Low Temperature Radical Initiator for Proper Storage and Safe Handling Conditions

Cuixian Yang*, Ralph Zhao, Amanda Peterman, Analisse Rosario, Josh Bader, Tom Vickery, Megan Roth, Adam Fine

Environmental & Process Safety Engineering (EPSE), Chemical Engineering R&D
Merck & Co., Inc., 126 E. Lincoln Ave, Rahway, NJ 07065, USA
Cuixian.yang@merck.com

Abstract

Commercially available azo-type low temperature radical initiators provide efficient initiation of many chemical reactions. However, the azo group initiators are energetic compounds that also have thermal stability issues at ambient or even sub-ambient temperatures. These initiators can also generate nitrogen gas during slow decomposition under heat and/or light, which could present a safety challenge for shipping, storage and usage. In order to define safe storage and handling conditions, a variety of calorimetry studies were carried out. Exotherm and pressure data were collected from these studies in an effort to gain a better understanding of the decomposition kinetics. Thermal-kinetics and thermal safety model simulations were then used to obtain the self-accelerating decomposition temperature (SADT) and decomposition activation energy for the azo-type initiator. This methodology for thermal decomposition kinetics data and parameter determination, acquired with 5mg to 1g scale samples, enables safe storage, handling, and scale-up process preparation.

Keywords: Azo Radical Initiator, Self-accelerating Decomposition Temperature (SADT), Storage and Handling, Thermal-Kinetics Simulation, Thermal Instability

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Analysis of pressure behavior during reaction runaway and estimation of available depressurization design

Yuto Mizuta[a,b], Motohiko Sumino[a], Youichi Kunitou[a], Atsumi Miyake[b]
[a] Safety Engineering Center, Mitsubishi Chemical Corporation,
1000, Kamoshida-cho, Aoba-ku, Yokohama-shi, Kanagawa, Japan
[b] Yokohama National University,
79-1, Tokiwadai, Hodogaya-ku, Yokohama-shi, Kanagawa, Japan
*Presenter E-mail: mizuta.yuto.ma@m-chemical.co.jp

Abstract

There are possibilities of pressure increasing in chemical equipment in case of some upset during operation of reaction, pressurization or transfer. Therefore, safety valves are installed on equipment to prevent rupture of equipment. Pressure increasing by reaction runaway is difficult to estimate, because two-phase flow might occur. Therefore, it would be hard to say all equipment have enough size as safety valves. Vent sizing method for two-phase flow by reaction runaway has been developed by DIERS established in 1987, and ISO 4126-10 was published in 2010 and it got the global standard. However, the ISO model often overdesigns the vent size due to the assumption that the liquid level in the reactor does not change during the runaway reaction. In this study, analysis of reaction rate of reaction runaway was carried out by ARSST, and then process simulation of pressure increasing and depressurization phenomenon was carried out by Aspen. Considering liquid level decrease in the reactor during reaction runaway would give more accurate design of vent size.

Keywords: Reaction Runaway, Vent Sizing, Two Phase Flow, ISO 4126-10, Aspen, Dynamic Simulation, Liquid Decrease

1. Introduction

In chemical plants, safety valves are installed to prevent rupture of equipment by undesired pressure-rise due to reaction and failure during pressurization or transport operations. Typical initiating abnormal events leading to pressure-rise are malfunction of control valves, cooling system failure of reflux or internal coil of reactor, tube failure of heat exchanger and fire case etc. Though there are many considerations for those cases, pressure-rise caused by reaction runaway is difficult to predict. Especially, about the case of two-phase flow occurred by reaction runaway, it is not easy to estimate adequate vent size. Vent sizing

method for two-phase flow was developed by DIERS under the auspices of AIChE in 1987, and ISO 4126-10 [1], Safety devices for protection against excessive pressure -Part10: Sizing of safety valves for gas/liquid two-phase flow, was published in 2010. After that, JIS B 8227 [2] which is translation of ISO 4126-10 was published in 2013 in Japan. However, the organization which able to implement vent sizing with JIS method would be few because of hard to understand each analysis procedure. On the other hand, recently, explosion accidents due to reaction runaway occur successively, and it is considered important to review protection layers. Therefore, it is necessary to review safety protection layers for such equipment, especially the safety valve. If the calculation of vent sizing by ISO model can be carried out, sometimes bigger diameters than diameters of reactor is obtained. It is impossible to implement to install safety valve to the reactor. The assumption of ISO method is that equipment are confined and inventories do not change during runaway reactions. However, actual equipment have some gas lines such as exhaust lines or reflux lines. Vaporization of solvents and decrease of inventory would temper reaction runaway phenomenon by vaporization and prevention of two-phase flow, and it could lead to make vent size much smaller. Before finding out the effects of gas lines, the construction of detailed simulation considering with mass balance, reaction rate, temperature and pressure changes during runaway reaction is carried out in this paper.

2. Model equipment and analysis method

The construction of detailed model is carried out with model substances and equipment in this study. There are many severe accidents at chemical plants dealing with MEKPO (Methyl Ethyl Ketone Peroxide) in Japan, Korea and Taiwan. MEKPO is well known as unstable substance and thermal runaway occur by high temperature condition or some contamination. There are many studies about MEKPO in east Asia, and many knowledges are obtained, but there are few studies about vent sizing.

2.1 Model equipment

The model equipment is assumed in this study as shown in Figure 1. The diameter of reactor is 1.5m, height is 3m, operation conditions are atmospheric temperature and pressure, and MEKPO is produced by reaction of MEK and hydrogen peroxide. MEKPO solution is 28 wt%, and the solvent is the mixture of dimethyl phthalate and toluene. The set pressure of safety valve is assumed 300 kPag. As the Assumption of abnormal scenario is cooling system failure of the reactor, the temperature rise leads to reaction runaway and pressure rise due to generation of non-condensable gas. Though it is necessary to decide the set pressure and the diameter of the safety valve for adequate depressurization, the diameter of safety valve is focused in this study.

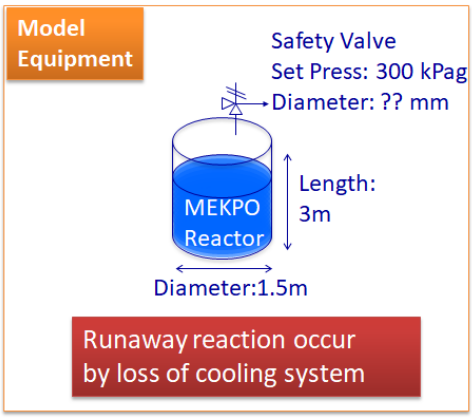

Figure 1. Specification of model equipment of MEKPO reactor

2.2 Analysis method for detailed simulation

The detailed simulation is constructed by Aspen Plus, Aspen Dynamics and Aspen Custom Modeler in this study. Aspen is simulation software specialized for chemical processes, and it includes equipment model with characteristic of unit operation and many chemical properties. The procedure of the construction of estimation model for the diameter of safety valve as shown in Figure 2. It is necessary to analyze the behavior of temperature and pressure rise during runaway reaction. Dynamic simulation is carried out with Aspen Dynamics by export of static model from Aspen plus. For the construction of Aspen Model, the definition of reaction formula, reaction rate and heat of formation of reactants and products are necessary. Furthermore, the behavior of temperature and pressure rise in the reactor is calculated by the combination among heat of reaction, the amount of generated gas and vapor pressure of solvent. And then, validation should be checked which the

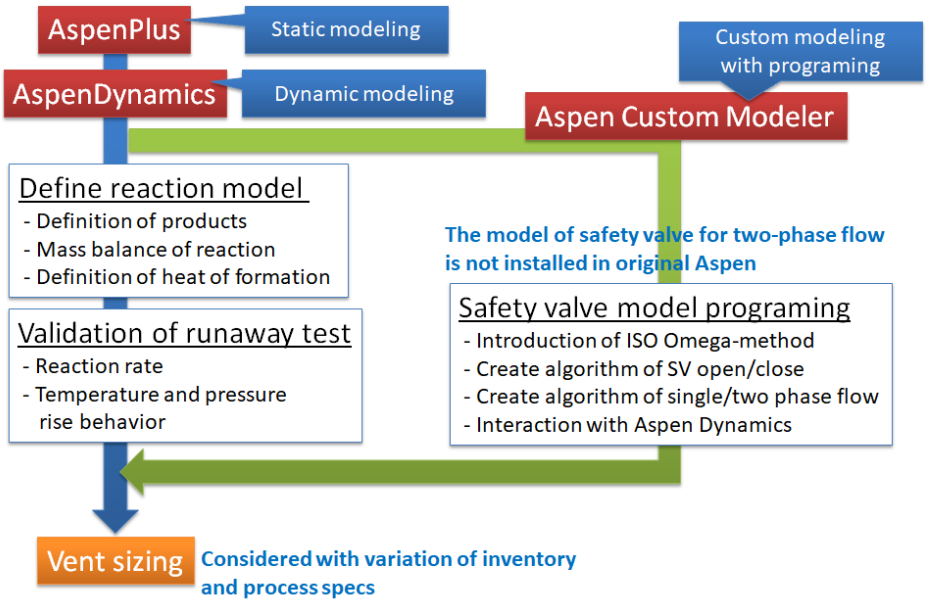

Figure 2. The procedure of the construction of estimation model for the diameter of safety valve in Aspen

simulation result matches the experimental one. On the other hand, safety valve model is not included in Aspen Dynamics and it is necessary to make a model in Aspen Custom Modeler.

3. Initiatives for each action assignment

The action assignments of construction of model is introduced in each section as shown in Table 1. At first, Aspen Dynamics model is studied as the definition of reaction model and validation of runaway experiment, and they are described in section 3.1 and 3.2. And then, the model of safety valve is programmed in Aspen Custom Modeler, and it described in section 3.3. Finally, the calculation results are compared with ISO model and benchmark tests as checking accuracy of detailed model in section 3.4 and 3.5.

Table1. The list of action assignments

Chapter No.	Action assignments
3.1	Definition of reaction model
3.1.1	Mass balance of reaction
3.1.2	Estimation of reaction rate
3.1.3	Estimation of heat of reaction
3.2	Validation of runaway experiment
3.3	Making the model with Aspen Custom Modeler
3.3.1	Installation of ISO Omega-method
3.3.2	Generation of two-phase flow
3.3.3	Solution for transient phenomenon
3.4	Comparison of results with ISO model
3.5	Detailed simulation with considering liquid decrease

3.1 Definition of reaction model

ARSST [3, 4] test was carried out to define reaction model such as reaction formula and reaction rate by experimental data of temperature and pressure. In order to define reaction formula, it is necessary to define the kinds and numbers of products on the right side of reaction formula. And it is also necessary to save mass balance on both sides of reaction formula. After then, the estimation of reaction rate was carried out by the experimental data of ARSST. Finally, heat of formation of each substance was estimated to calculate heat of reaction. The reaction model was created by defining a set of parameters.

3.1.1 Mass balance of reaction

In order to create a reaction model, it is necessary to take the mass balance among C, H and O before and after decomposition of MEKPO as shown in Eq.1. The pressure vessel was installed in the ARSST as shown in Fig. 3 to collect and analyze generated gas by the gas chromatography, and also residual pressure in the ARSST is measured to determined number of moles. On the other hands, It is difficult to identify the composition of products in liquid

phase, RESIDUE is defined as representative substances in liquid phase. The RESIDUE is so-called fitting parameter in the reaction formula, and the construction of molecular x, y and z and numbers of molecular n7 is fitted to take mass balance.

$$MEKPO \rightarrow n_1H_2 + n_2O_2 + n_3CO + n_4CO_2 + n_5CH_4 + n_6C_2H_6 + n_7RESIDUE \quad \text{(Eq. 1)}$$

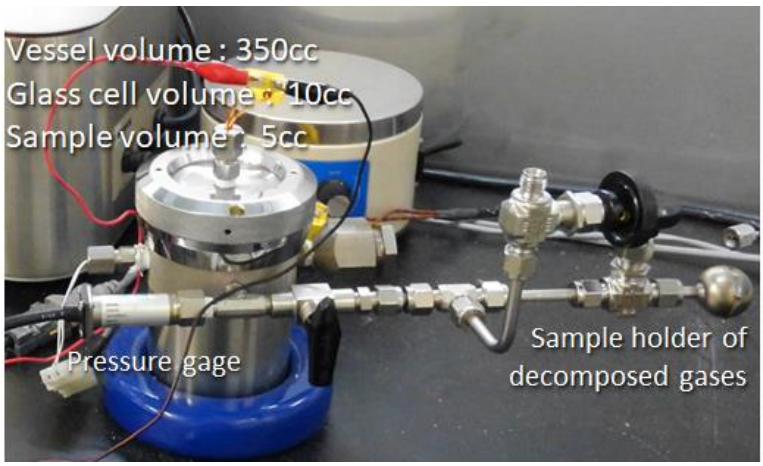

Figure 3. Gas collection of generated gas by reaction runaway

3.1.2 Estimation of reaction rate

The reaction rate is estimated by the temperature data obtained by the ARSST test. The reaction rate as shown in Eq.2 is assumed that the reaction mechanism does not change during runaway reaction, and it is fitted by the Arrhenius type reaction rate.

$$r = \frac{d}{dt}\left\{n_0|_{MEKPO}\left(1 - \frac{T - T_0}{\Delta T_{max}}\right)\right\} = A_{pre}exp\left(\frac{-E_a}{RT}\right) \quad \text{(Eq. 2)}$$

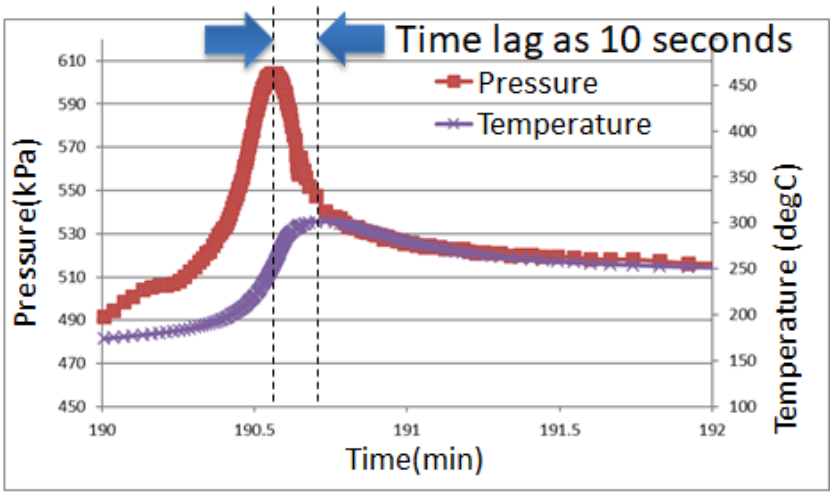

Figure 4. Experimental data of temperature and pressure

According to compare temperature and pressure data, it was observed which peak value of them are different for 10 seconds as shown in Fig. 4. It means that estimation of reaction rate with only temperature data, it would underestimate behaviour of runaway reaction, and the peak time obtained with pressure data would be close to real peak time of runaway reaction. It is considered that until the peak time temperature rose by reaction rate, and then, after the peak time temperature decreased by cooling curve as shown in Fig. 5.

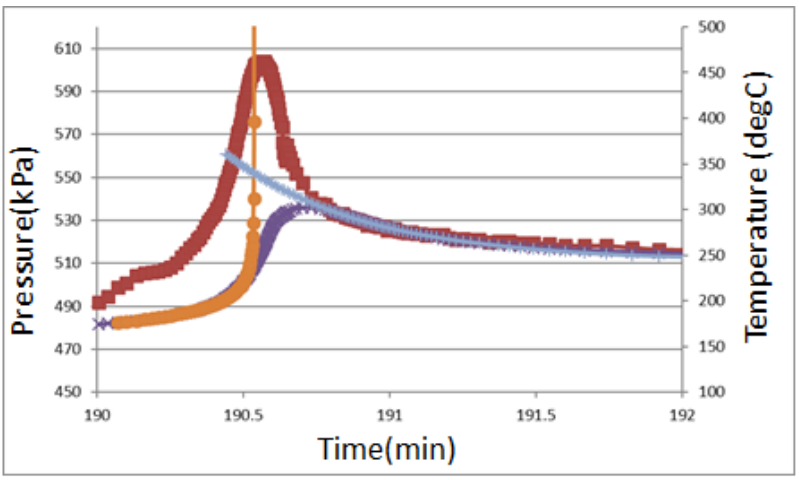

Figure 5. Corrected temperature data estimated by reaction rate and cooling curve

The blue line is Arrhenius line estimated with corrected temperature data and the red line is original line in Fig. 6. In case that the reaction rate is estimated with only the experimental data, the slope of dT/dt was small in the region where the violent runaway occurred, and the runaway behaviour would be underestimated.

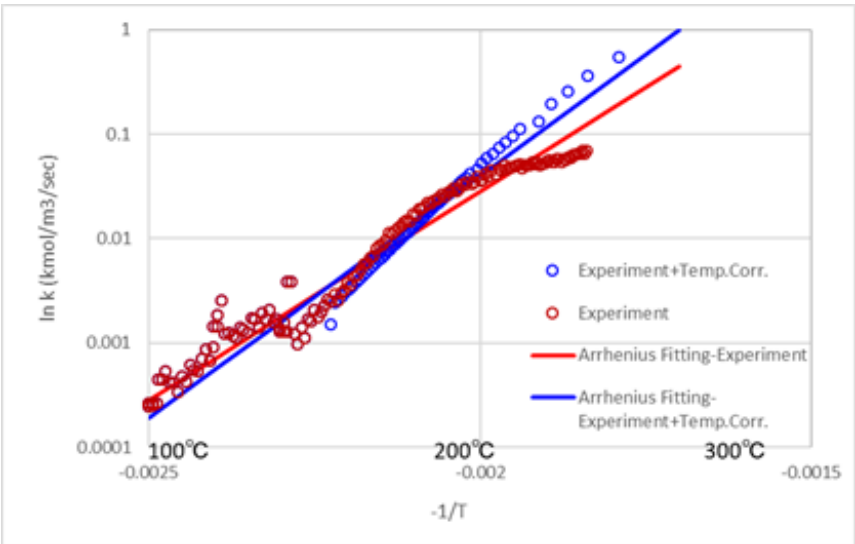

Figure 6. Estimated reaction parameter with and without corrected temperature data

3.1.1 Estimation of heat of reaction

The heat of reaction is calculated by the difference of total heat of formation before and after the reaction as shown in Equations 3 and 4. Regarding MEKPO and RESIDUE, these substances are not in the Aspen physical property library, so it is necessary to estimate their heat of formation. The heat of formation of MEKPO was calculated by PM3 model by SPARTAN, the molecular orbital calculation software. The heat of formation of RESIDUE was calculated from the relationship between the heat of formation of other substances, the composition, the specific heat, and the adiabatic temperature rise obtained from the ARSST test as shown in Eq.3, Eq.4 and Fig. 7.

$$H_{f\_MEKPO} = \sum_{i=products} (n_i H_{f\_i}) + Q_{reac} \qquad \text{(Eq. 3)}$$

$$\Delta T_{max} = \frac{Q_{reac}}{C_p M_{test}} \qquad \text{(Eq. 4)}$$

Products	n: Moles	H_f
	–	kcal/mol
H2	0.05	0
O2	0.31	0
CO	0.25	−26.4
CO2	0.75	−94
CH4	0.12	−17.8
C2H6	0.19	−20
RESIDUE	0.82	Unknown

Figure7. The heat of formation of products from runaway reaction

3.2 Validation of runaway reaction

After defining the mass balance, reaction rate, and heat of reaction, the calculations were converged in the low temperature region where almost no reaction occurred, and a steady model was created in Aspen Plus. The data was exported to Aspen Dynamics and the runaway behaviour was analysed corresponding to the experimental conditions of ARSST. As a method to make MEKPO runaway in a model converged at low temperature, thermal energy was applied to the reactor to raise the temperature to near the experimental temperature of ARSST by task function. When the temperature is raised up to temperature range around 100°C, self-decomposition gradually starts and eventually reaction runaway occurs. The calculated results were compared with the ARSST test result as shown in Fig. 8, and the experimental result was reproduced in terms of the time to runaway reaction and the temperature rise range, so the validity of the model was verified.

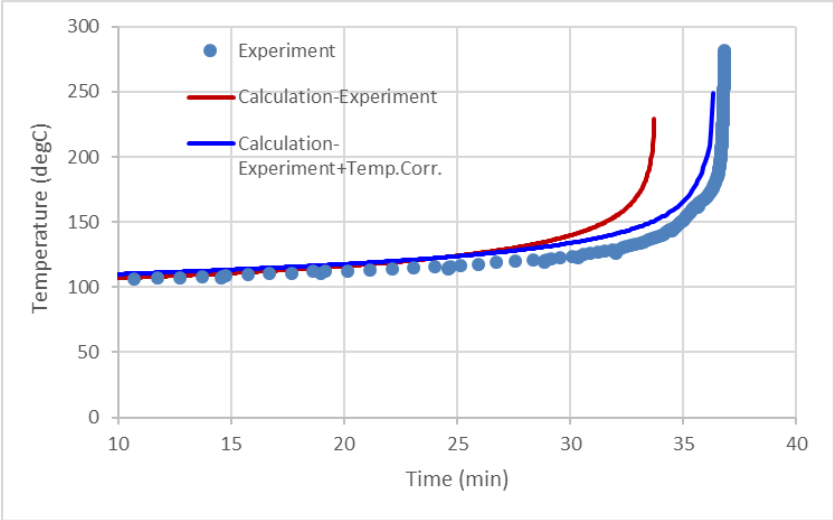

Figure 8. Validation of Aspen calculation result by comparison with ARSST test

3.3 Making the model with Aspen Custom Modeler

As the safety valve model is not installed in Aspen Dynamics, it is necessary to make the model based on ISO method with Aspen Custom Modeler. As shown in Fig.9, temperature and pressure behaviour are calculated with Aspen Dynamics, and then data are transported to Aspen Custom Modeler. According to intensity of reaction runaway, two-phase flow or gas phase flow are selected and the flow rate from reactor is calculated, and then the flow rate is returen to Aspen Dynamics to converge a series of process phenomenon.

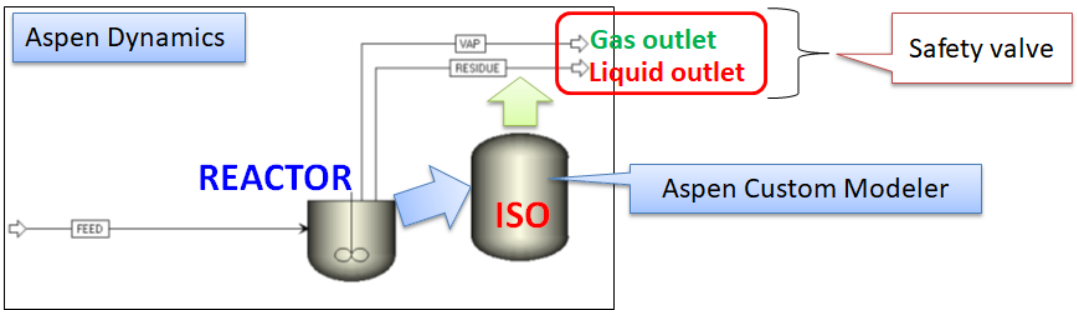

Figure 9. The interaction between Aspen Dynamics and Aspen Custom Modeler

3.3.1 Installation of ISO Omega-method

The calculation procedure of vent sizing based on ISO omega-method is that mass flow rate of generated gas is estimated by the temperature at the set pressure of safety valve and it is equal to mass flow rate through safety valve and then the diameter is calculated. On the other hand, in case of Aspen model, defined reaction formula and reaction rate give change over time of temperature and pressure, and pressure behaviour in the reactor is calculated with relationship between mass flow rate of generated gas and mass flow rate through safety valve. The calculation procedure of ISO model and Aspen model is shown in Fig.10 and 11. Blue

rectangles mean input parameters and green ones are intermediate parameters and red ones are output parameters. The input and output parameters of Aspen model are not same as ISO model, and modifying equation are necessary to calculate for vent size.

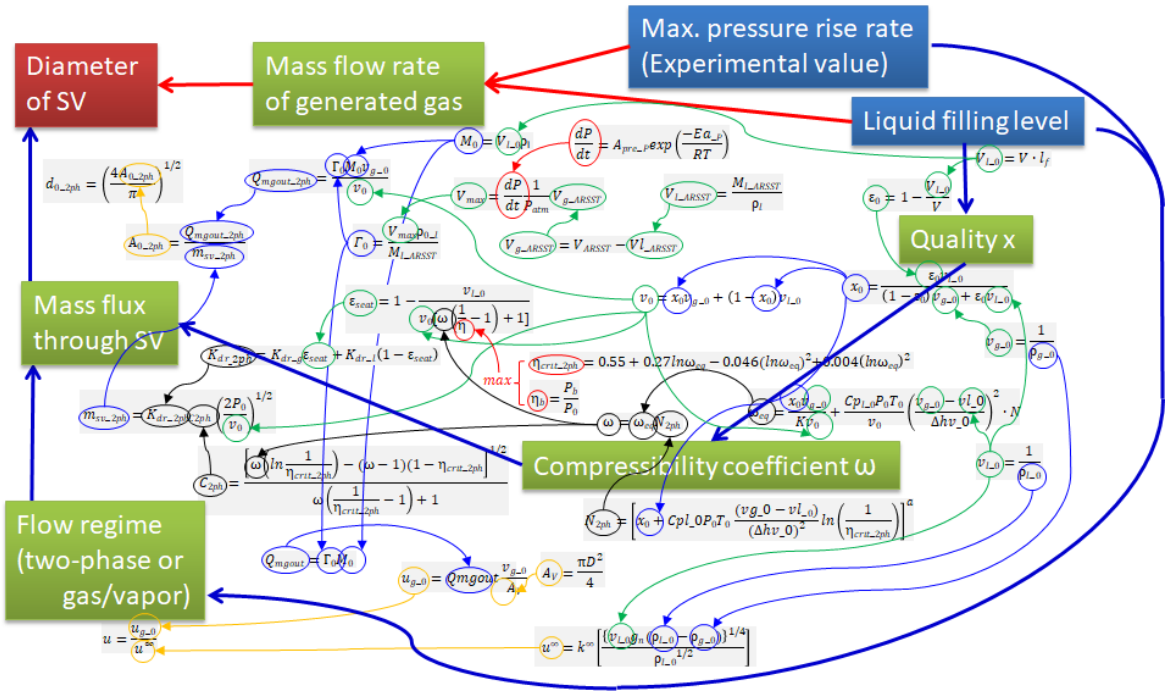

Figure 10. The calculation procedure of vent sizing in ISO model

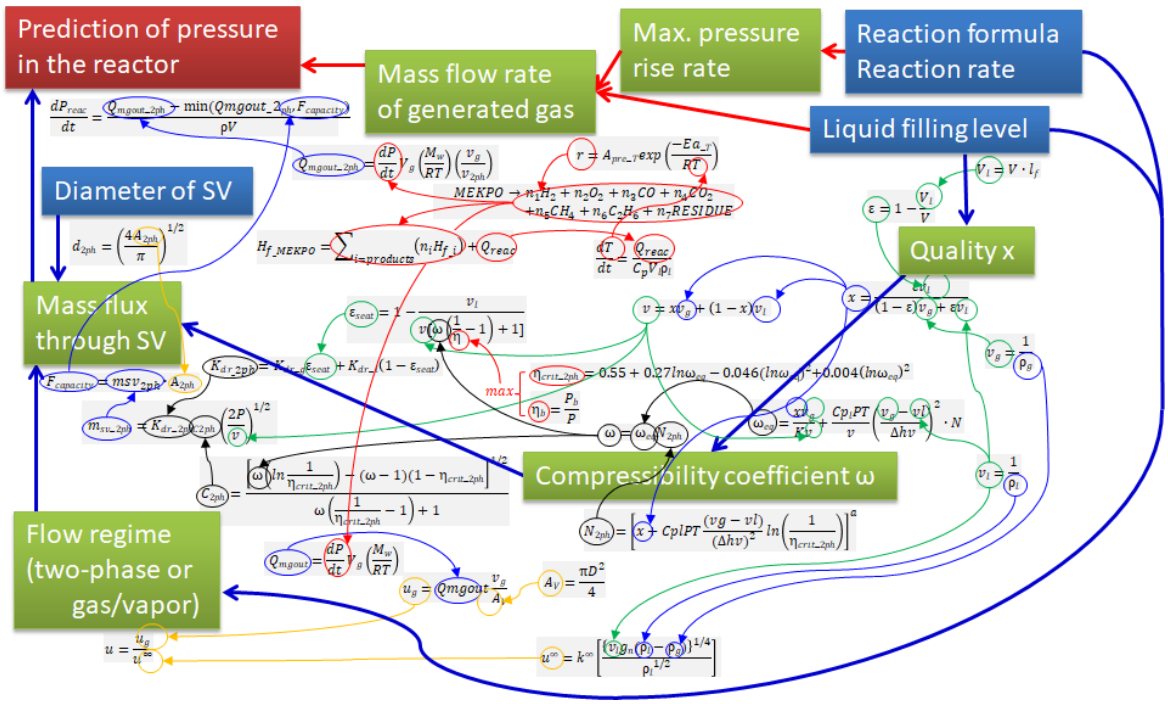

Figure 11. The calculation procedure of vent sizing in Aspen model

3.3.2 Generation of two-phase flow

It is necessary to determine the flow condition, that is two-phase flow or gas/vapor single phase flow, for the calculation of mass flow rate through safety valve. Two-phase vapor-liquid disengagement is shown in Fig. 12 which is the diagram based on DIERS test. The diagram is organized on two axis, average void fraction and dimensionless superficial vapor velocity. Dimensionless superficial vapor velocity of Churn-turbulent fluid is shown in Eq.5 and of Bubbly fluid is shown in Eq.6. According to knowledge of DIERS test, the turbulent condition in violent runaway reaction would be Churn-turbulent fluid, the flow condition is determined with Eq.5.

$$\Psi = \frac{2\varepsilon_d}{1 - C_0 \cdot \varepsilon_d} \qquad \text{(Eq. 5)}$$

$$\Psi = \frac{\varepsilon_d \cdot (1 - \varepsilon_d)^2}{(1 - {\varepsilon_d}^3)(1 - C_0 \cdot \varepsilon_d)} \qquad \text{(Eq. 6)}$$

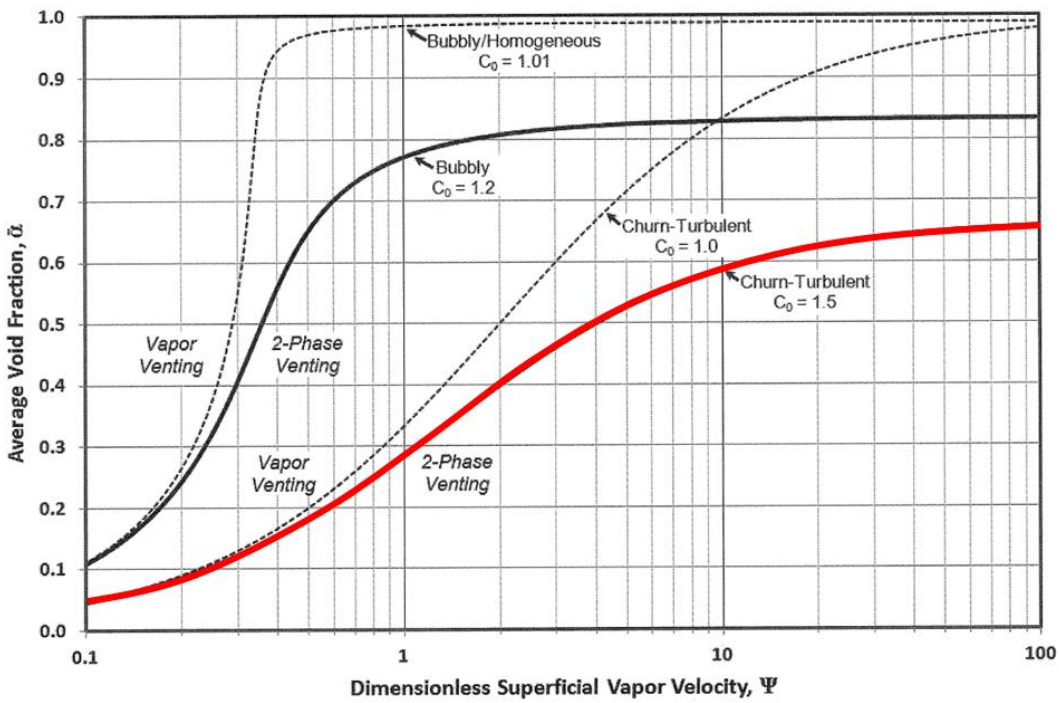

Figure 12. Two-phase vapor-liquid disengagement

3.3.3 Solution for transient phenomenon

When runaway reaction progress, dimensionless superficial vapor velocity increase gradually, and if pass through the borderline of two-phase flow, that is red line in Fig. 12, it is necessary to exchange the equation from single-phase to two-phase. If in order to exchange equation by “if” sentence, mass flow rate through safety valve is discontinuity before and after occurrence of two-phase flow, and calculation cannot continue in Aspen Custom

Modeler. Therefore, mass flow rate through safety valve is calculated with interpolation formula to keep continuity by hyperbolic function as shown in Eq.7.

$$tanh\_void = tanh(void\_crit - void)/2 + 0.5 \qquad \text{(Eq. 7)}$$

The value of vertical axis *tanh_void* is calculated by the relationship between *void_crit* and *void*, and then mass flow rate through safety valve *F* is calculated with Eq.8.

$$F = \tanh\_void \times F_{2ph} + (1 - \tanh_{void}) \times F_{1ph} \qquad \text{(Eq. 8)}$$

Default type of hyperbolic function vary from -1 to 1 as value, but in order to exchange equation it is necessary to adopt range between 0 to 1 and tanh is multiplied by 2 and is added 0.5 as in Eq.7. In the same way, when safety valve open, transition phenomena occur as mass flow rate from no flow with safety valve close to flow with safety valve open. However, in this case safety valve open before pressure in the reactor reach at set pressure of safety valve as shown in Fig.13, and the phenomena that pressure cannot rise to set pressure is obtained. Therefore, hyperbolic function is redefined the equation as in Fig.9 and Fig.14, that is to take value which is 0 or hyperbolic value in max function like activation curve used in machine learning.

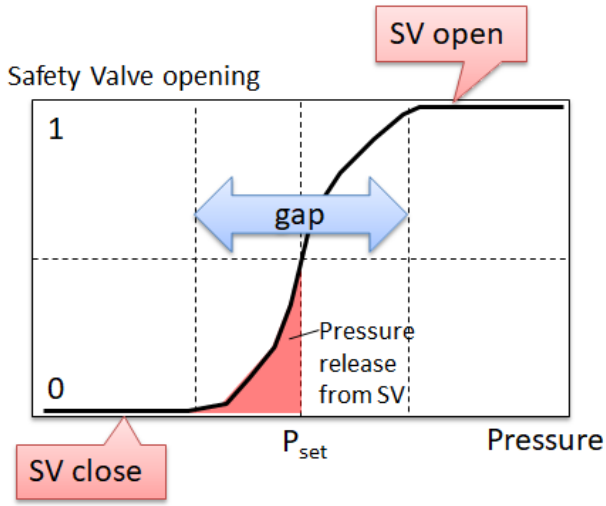

Figure 13. Exchanging equation of flow condition with hyperbolic function

$$tanh\_Pset = max(0, tanh((Preac - Pset) * tan\_factor\_Pset)) \qquad \text{(Eq. 9)}$$

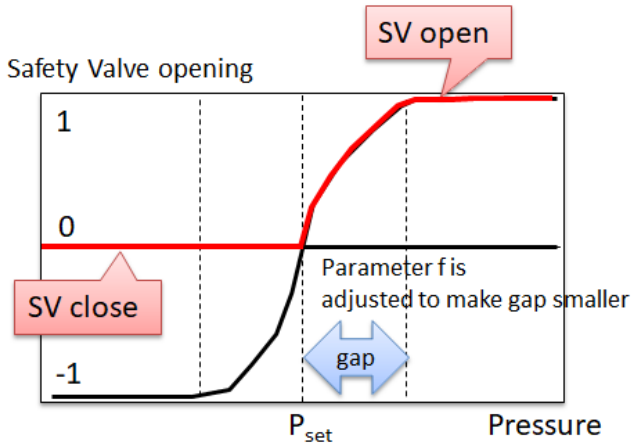

Figure 14. Exchanging equation of flow condition with re-defined hyperbolic function

When pressure in the reactor under set pressure of safety valve, safety valve is closed because of max function. And then if pressure in the reactor rise over set pressure, safety valve opens immediately because *tanh_Pset* goes to 1, and it makes possible to solve the problem that pressure leaks before set pressure.

$$F_{SV} = \tanh\_Pset \times F \qquad \text{(Eq. 10)}$$

3.4 Comparison of results with ISO model

The result of detailed simulation described from chapter 3.1 to 3.3 is compared to ISO method. Originally, this model is the model that variation of liquid level during runaway reaction is considered, but, as case 1 for comparison to ISO method, the part of model is changed not to change liquid level. The calculation results of detailed simulation case 1 and ISO method are plotted on Fig.15. Fig.15 is the graph described in ISO regulation, and the meaning is same as Fig.12, though vertical axis is not same as average void fraction and critical filling threshold. The relationship between average void fraction and critical filling threshold is like Eq.11.

$$\varphi_{limit} = (1 - \varepsilon_d) \times 100 \qquad \text{(Eq. 11)}$$

Regarding ISO method, the plot is only one at set pressure. And regarding detailed simulation case 1, there are a series of plots while runaway reaction. When reaction rate is not so fast plots are on left side in the graph, and as reaction rate increase plots move to right direction. About slight increasing liquid level in case 1 is caused by liquid expansion with temperature rise. The destination of case 1 is almost same position of ISO method. As equation in detailed simulation is same as ISO method, correspondence of these results would be one of validation for the accuracy of constitution of equation, physical properties

and experimental parameters. And estimated diameter of safety valve of detailed simulation case 1 and ISO method are 2.2m and 2.3m on each.

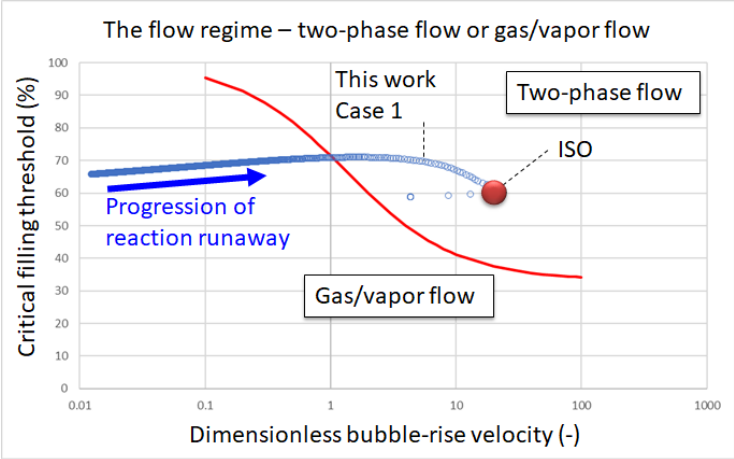

Figure 15. Comparison of calculation results between detailed simulation case 1 and ISO method

3.5 Detailed simulation with considering liquid decrease

The calculated results of detailed simulation case 2 is plotted on the graph as shown in Fig.16. Reaction runaway starts from left side and plots move on the trace of case 1 because variation of liquid level is not significant in single phase discharge region. When the plots move into two-phase region, liquid is discharged from safety valve as two-phase flow, and liquid level decrease to 45%. As the liquid level decreases, the amount of gas generated by decomposition of the liquid phase also decreases and liquid fraction in two-phase flow is decreased, so the estimated result of the diameter of safety valve is 0.7 m.

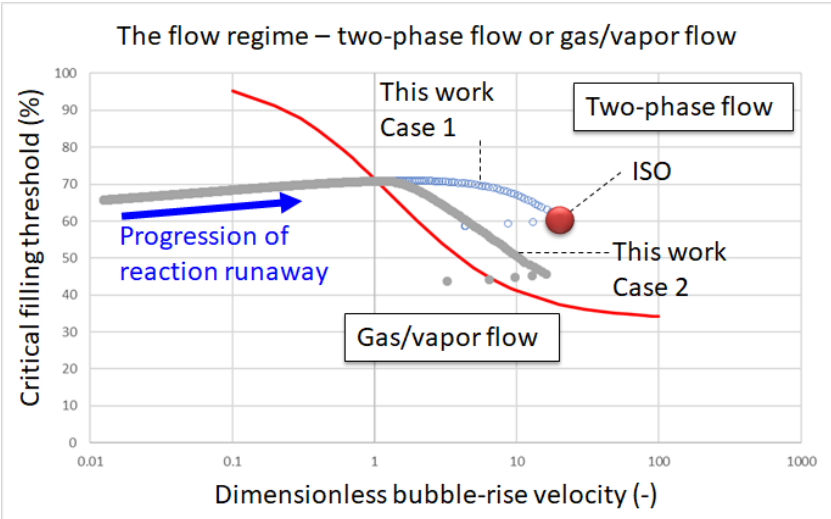

Figure 16. Comparison of calculation results between detailed simulation case 1, 2 and ISO method

Though accuracy of detailed simulation will be carried out verification test in future work, and here, accuracy is checked by the benchmark test data in literature [5] as shown in Fig.17. The critical conditions for vent sizing are plotted as benchmark test results, and the black line in the graph is critical border line for vent sizing. Actual rupture incidents data are also plotted on the graph under the critical borderline, and it means that diameters of safety valves of rupture incidents are underestimated. The calculation results of detailed simulation case 1 and 2 are also plotted on the graph. The data of case 2 is plotted on the borderline and case 1 is above the borderline. It means that case 2 would give accurate vent size and case 1, this is almost same as ISO method, would give bigger vent size.

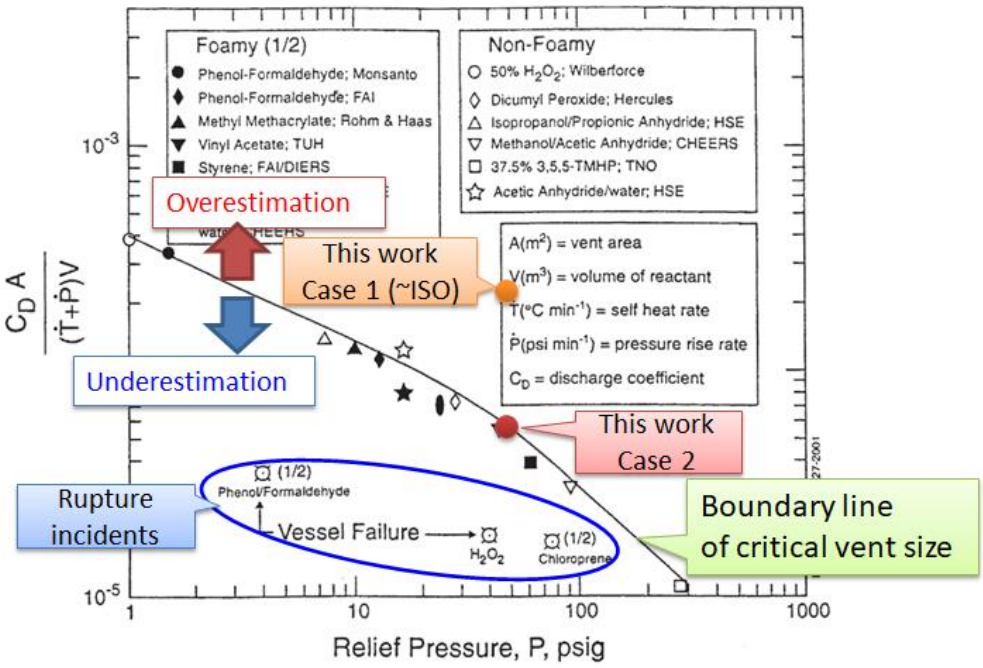

Figure 17. Comparison of vent size among benchmark test and the calculation results of case 1 and 2

4. Conclusion and future work

The detailed simulation model is constructed with Aspen software, ISO Omega-method and ARSST test. The model is considered variation of liquid level during runaway reaction, and it is necessary to solve the problem of numerical discontinuity of valve opening and occurrence of two-phase flow. And for increasing accuracy of the model, reaction rate during runaway reaction is assumed correctly, and accurate setting of reaction formula and heat of formation is also necessary. The calculation result of case 1 gives almost same result as ISO method, and according to Fig.16 estimated diameter of safety valve would be overestimated. On the other hand, the calculation result of case 2 plot on the borderline of benchmark test, and estimated diameter of safety valve would be accurate. In the future work, verification test will be carried out to check accuracy of the detailed simulation model, and then the test and simulation considered exhaust gas line will be carried out. The current model is assumed confined reactor, but Some lines such as exhaust gas line or reflux line are often installed on the actual reactor. Analysing more

realistic equipment would give more reasonable and accurate design for depressurization against runaway reaction.

References

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Preventing Cognitive Attributed Errors in Safety Critical Designs: A Path Forward

Tom Shephard
Safety and Automation Consultant
Houston, Texas

*Presenter E-mail: tsheptx@gmail.com

Abstract

For the first time in human history, it is now possible to take a comprehensive, cognitive-focused, first principles approach to safety critical design, the topic of this presentation. Why is this important? Experts assessing the root cause of catastrophic accidents commonly site human error as a causal factor. These same experts often assert that a latent error in the design is the causal source of that error. An example is a mismatch between the cognitive demands placed on the user and the user's actual cognitive capability.

Recent catastrophic accidents provide ample evidence that existing methods failed to reliably mitigate cognitive-attributed errors in safety critical designs. The most widely employed are rooted in methods developed in the last century. In response, global industry leaders and organizations issued article and white paper 'calls for change' in these areas. In addition, new technology available in 1990's triggered an exponential growth in the base knowledge that articulates the fundamental nature, attributes, capabilities and executional functioning of the very different automatic (aka, subconscious) and conscious processes. Little of this new information is widely known or currently applied. Subsequently, industry tribal knowledge is broadly incomplete and often erroneous, a hidden bias that contributes to latent design errors.

So, what is the path forward? What form does a new design process take? Any new method should be purposely designed to explicitly identify and mitigate cognitive-attributed design errors at the task phase (detect, decide, act) under all plausible situations. It should utilize the latest available, peer-reviewed information on human cognition and apply equally well to the design of an active human barrier or a safety critical task. From that frame, the presentation provides an overview of one possible solution. Presented examples of automatic and conscious processes should aid in understanding what this new information looks like and the tools and expertise needed to apply it.

Keywords: human factors, barriers, cognitive ergonomics, situation awareness, tasks, process safety, design process, emergency response

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1 Introduction

For the first time in human history, it is now possible to take a comprehensive, science-driven, cognitive-focused approach to the design of active human barriers and safety critical tasks, the topic of this manuscript. Why is this important? Experts assessing the root cause of catastrophic accidents commonly site human error as a causal factor. Many of these experts also assert that a latent error in the design is the causal source of that error. This occurs when there is a mismatch between the cognitive demands of the design and the cognitive capabilities of the human.

The Deepwater Horizon (DWH) accident triggered a global issuance of calls-for-change whitepapers and articles in cognitive ergonomics (CIEHF 2016, Johnsen et al. 2017, OESI 2016, IOGP 2012, and SPE 2014). Common themes include improvements to situation awareness and reducing the unrealistic cognitive demands in the design and the effects of cognitive behaviours and biases, e.g., confirmation bias. Recent accidents continue to provide evidence that the existing design methods are inadequate. Some of the most widely employed are rooted in methods developed in the past century. New technology, first available in the mid-1990's, triggered an exponential growth in neuroscience research that provides an expanded and new understanding of the nature, attributes, and executive functioning of the profoundly different conscious (aka, System 2) and automatic processes (aka, System 1, unconscious, etc.). Little of this new information is widely known or currently applied in the O&G and petrochemical sectors. Subsequently, industry tribal knowledge is broadly incomplete and often erroneous, a hidden bias that contributes to latent design errors.

So, what is the path forward? What form does a new design process take? To start, a new methodology should be designed to prevent cognitive mismatches in situation-specific activities that take place within the barrier detect, decide, and act phases. It should consider the recommendations in the above-referenced articles and white papers. It should use and apply the best available science and practices currently used in other industries, and information available from peer-reviewed and widely recognized sources. From that frame, this manuscript presents a set of prototype processes and tools guided by these statements, i.e., a white paper approach. The intent is to show: example processes and tools and the information they generate, a new methodology that identifies and mitigates cognitive mismatches at the situation-based activity level, and how these tools can be used in a capital project environment. Given the state of the industry (and the many roadblocks that appear to be stifling progress), the author believes it may be helpful to provide an improved starting point, i.e., one that addresses the primary deficiencies in current practice. Additional supporting information is provided. A novel table summarizes and contrasts the fundamentally different functioning, capabilities, limitations, and behaviours of the automatic and conscious processes. A summary of the cognitive issues that are likely to occur in all safety critical designs is also included. Finally, the manuscript highlights the skills and knowledge needed to apply the more advanced cognitive-focused processes.

1.1 Industry Insights and Recommendations

The follow are representative statements and recommendations from the above referenced articles and white papers. Excerpts are inserted throughout this manuscript to indicate how the statements and recommendations in these documents are addressed in the presented processes and tools.

Human Factors in Barrier Management – White Paper (CIEHF 2016)

"Characteristics of good human barrier elements…..Expectations about the human performance needed are realistic: a) identify the situation that needs action; b) knowing or being able to work out what needs to be done; c) being able to do it in the time available, and under the likely conditions; d) having some means of knowing that the action had the intended effect"

"The intentions and expectations of human performance that are implicit in the decision to rely on people as part of the barrier system are rarely made explicit."

"All three levels of SA involve significant cognitive complexity and rely heavily on what psychologists often refer to as the operator's "mental model". A mental model captures the operators understanding of how a system operates and how it behaves…."

"When the cognitive dimension of incidents is properly investigated, there is often a significant discrepancy between what the operator thought was the state of the world, what was happening, or how an equipment or a process would have behaved and what the actual state of the world was, or how the system did behave."

Missing Focus on Human Factors – Organizational and Cognitive Ergonomics – in the Safety Management for the Petroleum Industry. (Johnsen et al 2017)

"Expertise on organizational ergonomics and cognitive ergonomics are missing from companies and safety authorities and are poorly prioritized during development."

"Incident investigations have revealed that cognitive and organizational ergonomics seldom are mentioned and explored."

"Latent errors are related to designers, high-level decision makers and managers, where the adverse consequences may lie dormant within the system for a long time and only becoming evident when combining with other factors to breach the system's defenses."

"The missing focus on cognitive and organizational ergonomics…may create weaknesses and holes in defenses/barriers….These issues have not been properly addressed in new versions of the Norsok S-002 to be published in 2017."

SPE Technical Report, The Human Factor: Process Safety and Culture (SPE 2014)

"Incident investigations often identify deficiencies in the design or implementation of the interface between people and technology as contributing to the loss of reliable human performance. This is sometimes referred to as "design-induced human error".

"The extensive research into the psychology of how irrationality and cognitive bias lead to poor risk assessment and decision-making and the practice solutions to counter these biases should be used more extensively to improve the training of people involved in safety-critical operations in the E&P industry."

Human Factors and Ergonomics in Offshore Drilling and Production: The Implications for Drilling Safety (OESI 2016)

"In the offshore O&G environment, many of the current interfaces for daily and emergency tasks have not been specifically designed to facilitate and support human performance."

One incident…"revealed the cause of the incident was not related to poor decision-making, but rather the absence of SA and poor mental models" "Studies have reported finding the loss of SA can result from something as simple as inattention and is also a function of experience and training"

"In a study of 332 offshore incidents…"more than 40% of drilling activity incidents were associated with inadequate SA. A majority of those errors (67%) occurred at Level 1 SA…"

"Confusion occurs when a worker misinterprets the observed behavior of the operating system in light of their mental model of the system."

1.2 Terms and Definitions

Active Barrier - 'Active' denotes a barrier type that performs the required safety function only upon detection of a pre-defined condition or state. A preventive or mitigation barrier may be designed as an active barrier type.

Active Human Barrier – A 'barrier' that relies on a human to perform one or more barrier phases, i.e., the detect, decide or act phase. This barrier type includes the 'Active Hardware + Human Barrier' defined in CCPS (2018, p. xv). Emergency response barriers are active human barriers.

Barrier – "A control measure or grouping of control elements that on its own can prevent a threat developing into a top event (prevention barrier) or can mitigate the consequences of a top event once it has occurred (mitigation barrier)….." (CCPS 2018, p.vx)

Barrier Element – The components that comprise the barrier system, i.e., *Physical, Human* and *Organizational* elements.

Barrier Function – The safe state achieved when all barrier tasks and phases are completed and executed as intended.

Barrier Phase – The uniquely different *Detect, Decide* and *Act activity* stages that comprise the barrier safety function. Each is referred to as a 'phase'.

Barrier System - The "…. system that has been designed and implemented to perform one or more barrier functions. A barrier system describes how a barrier function is realized or executed." (Sklet 2006) The barrier function is achieved by the collective functioning of its Physical, Human and Organizational elements.

Cognitive-attributed design error – a mismatch between a cognitive activity that is inherent to a design and the cognitive capabilities of the human assigned to perform the activity.

Task Response Time – The target response time assigned to a task. The execution of all tasks that comprise a barrier must be correctly completed within the specified process safety time (PST) for that barrier.

Abbreviations:

ALARP – As Low as Reasonably Practicable

DDA – Detect, Decide, Act

IPL – Independent Protection Layer. An active human barrier is an IPL if it is designated as such in a LOPA or equivalent risk assessment process

LOPA – Layer of Protection Analysis

HAZOP – Hazard and Operability Study

HE – Human Element

HMI – Human Machine Interface

OE – Organizational Element

PE – Physical Element

PPE – Personal Protective Equipment

PST – Process Safety Time

RA – Risk assessment

2 Barrier Constructs and Models

2.1 Define Barriers as Tasks

From the IOGP Report 460: Cognitive Issues Associated with Process Safety and Environmental Incidents, (IOGP 2012):

"A "safety-critical human task" is an activity that has to be performed by one or more people and that is relied on to develop, implement or maintain a safety barrier."

"Although some OGP members require safety-critical human tasks to be specifically identified and managed, the safety-critical nature of operator activities is not always recognized. It seems that the required performance standard, or the consequences of individuals not performing tasks to the required standard, is often poorly understood. There also seems to be an insufficient understanding of the demands that safety-critical tasks can make on human performance, what is needed to support the required level of performance, and the ways in which human performance could fail in undertaking the tasks, or the inherent unreliability associated with the task."

"…members should work towards adopting being able to satisfy themselves that safety-critical human barriers will actually work and the risk of human unreliability in performing them is effectively managed and reduced….members should work towards adopting practices to identify and understand safety-critical human tasks. "

From HSE (Principle 8) Assessment Principles for Offshore Safety Cases (APOSC), 2006:

"Safety critical tasks should be analysed to demonstrate that task performance could be delivered to the specified performance when required. This demonstration should draw upon recognised good practice in human factors."

"Human performance problems should be systematically evaluated. This should involve evaluating the feasibility of tasks, identifying control measures and providing an input to the design of procedures and personnel training, and of the interfaces between personnel and plant. The depth of analysis should be appropriate to the severity of the consequences of failure of the task."

Active human barriers can be defined as one or more tasks. Depending on the approach taken in a task analysis process, some barriers may have one person performing two or more tasks. Refer to Figure 1. Other barriers (e.g., emergency response barriers) require several persons who must perform these tasks in a manner that realizes and achieves the barrier safety function within the PST.

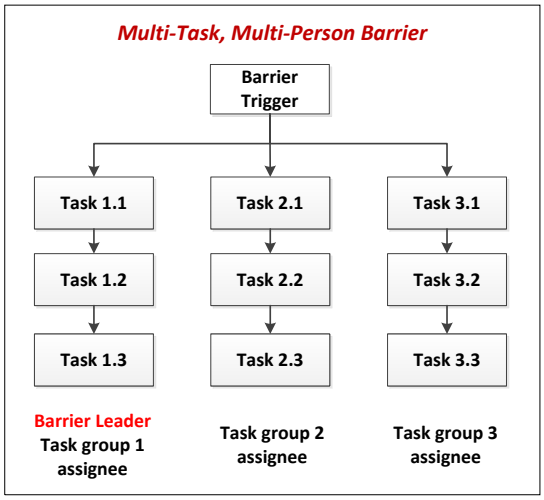


Figure 1 – A Multi-Person, Multi-Task Barrier

Emergency Response barriers common to offshore O&G production and drilling facilities typically rely on several persons, i.e., the ER command team, who perform the command and control tasks in all ER barriers. (Flin et al., 1996)

Multi-person barriers introduce a new set of design requirements, e.g., identify a qualified barrier leader to coordinate the actions of others assigned to execute their barrier tasks.

Active human barriers and SCTs are similar in that they both perform safety critical functions and rely on one or more humans to achieve the safety function. Indicated in Table 1, there are some important differences. One difference is the activation of an active human barrier is unplanned and therefore may be a surprise, as compared to a planned and scheduled SCT.

Active Human Barrier				**Safety Critical Task**
Barrier Type / Attributes	***Preventive***	***Control/Recovery (Mitigation)***	***Emergency Response (Mitigation)***	***Preventive***
Occurrence	Unplanned	Unplanned	Unplanned	Planned
How many Human Elements *(HE)*	1 (typical)	1 * (typical)	2 or more (typical)	Varies
Tasks needed to achieve the barrier function	1 (typical)	1 * (typical)	Two or more (typical) (One or more tasks assigned to HE)	1 or more / assigned HE
Active Barrier?	Yes	Yes	Yes	** No
Workload Demand	Assumed manageable within stated response time	Assumed manageable within stated response time	Situation and peak workloads may exceed HE capacity for periods of time.	Assumed manageable
Specified response time (Process Safety Time or PST)	Yes	Yes	Varies May establish target response times: • Medical or recovery response • Incident timeline, e.g., ship avoidance (collision) barrier PST may be affected by a time constraint attributed to an external system or barrier: • Barrier is dependent on a passive barrier with a specified endurance time (e.g., a firewall) • One or more barrier elements depend on an external, time/capacity limited support system, e.g., battery-backed power systems	Generally, no

* The number of unique tasks from an HTA depends on the how the task analysis team frames the task.
** Active only when scheduled.

Table 1 - Compare Active Human Barrier and Safety Critical Tasks

Preventive barriers have a Process Safety Time (PST) that is the time available to complete the preventive response before the top event occurs. With mitigation barriers, the PST is the time available to achieve the protective function after the top event occurs. In both cases, the PST reflects the dynamic aspects that are unique to the hazardous event. Each barrier is assigned a PST, to which the assignees must complete all barrier functions and achieve the specified safe state within this period. Time pressure is a common contributor to stress and, in many cases, to human error; human error can lead to barrier failure or degraded performance. A primary objective of an effective active human barrier and SCT design process is to eliminate or mitigate barrier design attributes that contribute to human error. Active human barriers are the primary focus of this manuscript.

2.2 Frame Tasks in the Form of a Detect, Decide, Act (DDA) Model

From CIEHF (2016, p 20) "Active Barriers must have detect-decide-act functionality – i.e., they must have one or more elements that allow them to:

> *...Detect the condition that is expected to initiate performance of the barrier function...*
>
> *...Decide what action needs to be taken, and;*
>
> *...Take the necessary action."*

"Detect – decide – act functionality can be inherent in a single barrier element, or can involve a combination of barrier elements working together (such as a sensor raising an alarm, a human understanding the meaning of the alarm and knowing what action to take, and then the human using a technical system to effect the action)."

Tasks are a compilation of cognitive and physical activities. The ability to define and assess these activities requires that the task be expanded into a form that supports the definition and assessment process. The model adopted in this manuscript is the Detect, Decide, Act (DDA) model from the Center for Chemical Process Safety (CCPS), indicated in Figure 2.

Figure 2 - Activity Phases in Active Barriers (CCPS 2018)

For reference, each activity in Figure 2 is referred to as a '*phase'*. The *detect* phase includes the alarm or event that activates the barrier/task.

2.3 Integrate Situation Awareness (SA) into the DDA Model

SA… "must be understood and applied at an adequate level of technical depth. SA also needs to be understood and managed at both the individual and team levels." (IOGP 2012)

"SA has been acknowledged as the basis for good decision making within complex systems, including the O&G industry where poor performance can lead to devastating results." (OESI 2016)

Sharp end knowledge-based mistakes may be…."caused by bad HMI design, for example, operator's poor problem solving due to lack of sufficient support via HMI to sustain excellent situation awareness." (Johnsen et al 2017)

The referenced call-to-action documents affirm the need to improve the understanding and application of situation awareness in operational and design practices. Suggestions on how this should be achieved were limited. Situation awareness (models and application) have been

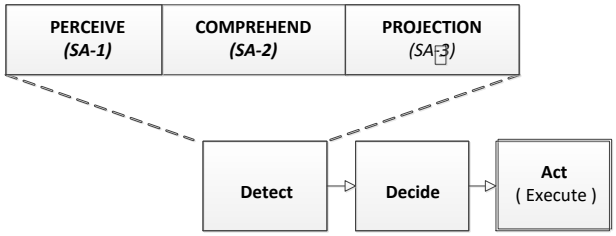
successfully used in other high risk, high consequences industries for several decades. The dominant and the most widely recognized and referenced model is Dr. Mica Endsley's three-stage SA model (Endsley 1995). Endsley's model is adopted for this manuscript. It proposes three stages of situation awareness:

- **Perception** (SA-1) refers to the acquisition of information that is perceivable and available to our senses.
- **Comprehension (SA-2)** is the product of combining the SA-1 information with one's stored knowledge and experience to develop an understanding (mental picture) of what the information means.
- **Projection (SA-3**) is the product of using one's expertise and understanding of how (and how quickly) the current situation (SA-2) is changing over time, to predict or anticipate how conditions may change in the future, near term.

Because time is the singular resource that often places the greatest demand on humans assigned to perform barriers/tasks, it is important to recognize that time is an essential aspect of SA.

- "The rate at which information changes is that part of SA…. that allows for the projection of future situations." (Endsley and Jones 2012, p. 19)
- "A critical part of SA is often understanding how much time is available until some event occurs or some action must be taken." (Endsley and Jones 2012, p. 19)

Figure 3 below shows the adopted approach for integrating Endsley's model it into the DDA model and thus, into the barrier design process.



Adapted from Figure 2.1, Endsley, 2012 p15.
DDA is the Active Human Barrier model in CCPS 2001, 2018.

Figure 3 - Integration of SA Model into DDA Model

This construct is similar to those used in other publications and guidance documents, listed in Table 2 below.

Reference	Model Phases
SINTEFF 2011 (CRIOP), Figure 5.1	Observation/Identification, Interpretation, Planning/Choice, Action/Execution
IFE/SINTEF (2015) (Petro- HRA Guideline)	Detect, Diagnose, Decide, Act
NUREG 2011 (p. 27)	Detection and Noticing, Understanding and Sensemaking, Decision Making, Action
Energy Institute 2020, p 28 (Implied)	Detect, Diagnose, Decide, Activate

Table 2 - Comparison of Task Models

2.4 Barrier Elements: Physical, Human and Organizational

The final construct needed to support the design process is to name and frame the barrier elements that, together, achieve the barrier safety function or task goal

From CIEHF (2016):

"Barriers should be considered as barrier systems: i.e., in nearly all cases, for the barriers to perform as expected, a combination of elements need to perform their individual functions in a coordinated manner."

"Failure of any barrier or barrier element to perform its function, or to be identified as being unlikely or incapable of performing its function when demanded, should therefore be treated as a significant event."

Currently there is no *globally accepted* guidance document that defines, names and frames the various barrier elements. As such, the terms adopted in this manuscript are *Physical*, *Human* and *Organizational*, indicated in Figure 4.

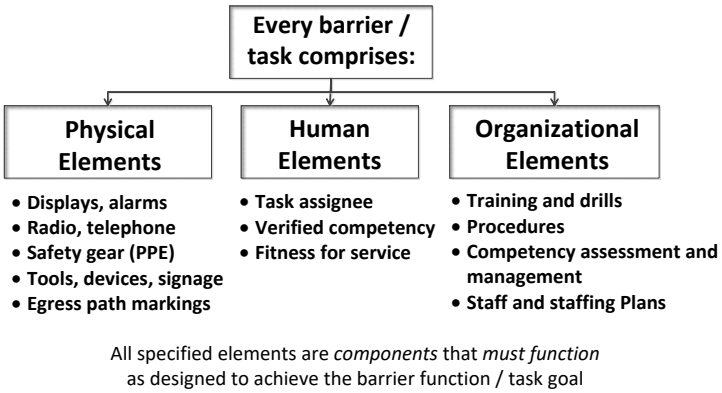


Figure 4 - Barrier / Task Elements

Physical Element (PE): Examples include, but are not limited to, technical systems, HMI displays and interfaces, instrumentation, notification devices, signage, paint markings that delineate egress routes, PPE, lifeboats, safe havens, etc.

Human element (HE): This is the person(s) assigned to perform one or more barrier task/phases. The HE is verified to be available, fit-for-service, and meet all specified competency requirements.

Organization Element (OE): Examples include, but are not limited to, staff and staffing plans, organizational charts, procedures, training, competency assessments, competency management, etc. (OE includes all components that are not human or physical in nature.)

Discussion

"….an early – and often problematic – focus only on safety critical elements (rather than on tasks and activities) in the UK offshore safety regime meant that wider critical aspects of the human element were often missed. A proper focus is needed on the totality of what people do, not just on the performance of the technical systems." (CIEHF 2016)

"...technical standards often lack design features necessary to incorporate human factors and ergonomics considerations. "(SPE 2014)

Comment: Some country and regulatory standards (e.g., PSA 2013) and industry guidance documents (CCPS 2018, CIEHF 2016, SINTEFF 2016) use the terms 'Technical, Organization and Operational' for barrier and task elements. For this manuscript, the terms 'Physical, Human and Organizational' are used. They may be easier to remember, and the expected-versus-actually word meanings may be less confusing. This may not be true for the similar sounding terms 'Organizational' and 'Operational'. IEC 61511, first published in 2004, used the similar sounding terms 'verification' and 'validation'. Not surprising, they are often misunderstood or swapped. Perhaps the more compelling reason for adopting the term 'human' is to offset the historical hardware-focused paradigm that remains prevalent in many industry documents. Given this perspective, and because the human element is the most challenging element to address in the barrier design process, it does not seem helpful to exclude 'human' as one of the elements.

3 Risk Assessment and Barrier Identification

3.1 Introduction

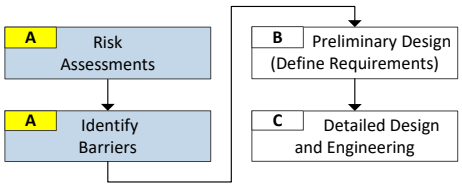


Figure 5 - Process Overview

From this section forward, the manuscript presents the prototype design processes indicated in Figure 5. These processes are amenable to the typical, stage-gated project execution model. They apply to active human barriers, though could also be applied to the design of safety critical tasks. In the following sections, the design activities and supporting information are discussed.

Information is also provided on execution considerations, i.e., the suggested participants to include in a task analysis workshop.

3.2 Proposed Methodology and Processes

From Figure 6, this section begins by identifying some of the existing risk assessment and proscriptive approaches currently used to identify a required active human barrier or SCT. As noted in the figure, an important product of these processes is the information listed in Figure 6. (This information is required input to the next project stage, i.e., preliminary design.)

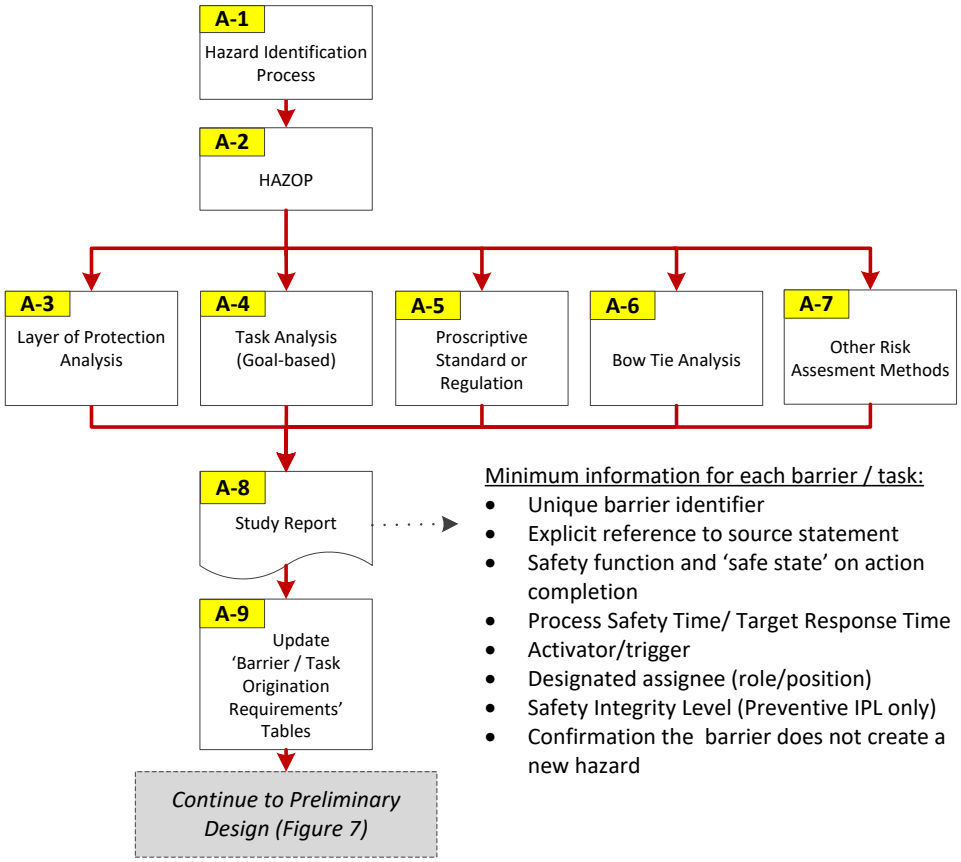


Figure 6 - Risk Assessments and Barrier Identification

If the barrier is defined in a LOPA, it may also be assigned a Safety Integrity Level (i.e., a minimum risk reduction factor) that must be achieved by the barrier.

Step A-4, the task analysis, is the process that may be appropriate for barriers that require two or more tasks to achieve the safety function. This is commonly the case for emergency response barriers. For example processes used to perform a hierarchical-type task analysis, see Shepherd (2001), Endsley and Jones (2012, p. 63-78) or IEF/SINTEFF (2015, p 35-39).

"A particular risk is that the inputs to a barrier analysis are not realistic and properly informed about operational realities." "The operator needs to be enabled to contribute fully to the

process through training and preparation. It is also essential that any analysis session has an adequate task analysis as an input…." (CIEHF 2016)

"So while preventive barriers (those on the left-hand side) typically operate over a timescale that can be measured in weeks, days and hours, mitigation barriers (those on the right-hand side) typically have to operate in a timescale of hours and minutes. This can create pressure for people to perform to extremely high standards in situations of both stress and time pressure." (CIEHF 2016)

As good- practice, the information listed in Figure 6 should be provided by the processes indicated as A-3 to A-7.

With Step A-8, it may be helpful to provide the base barrier information for single task barriers in a format similar to Table 3. If this information is not the product of the risk assessment then this introduces the issue of who provides it (e.g., qualified, authorized, accountable) and when. From the author's experience, errors in this information can be highly consequential, e.g., it may later be found that the PST cannot be me or the barrier creates a new hazard. Both are conditions that make the barrier infeasible in its current form.

Barrier Origination Requirements: Single Task Barriers								
Barrier Identifier	**Source Reference**	**Barrier Activator**	**Safety Function**	**Safe State on Completion**	**Process Safety Time**	**Human Role/ Assignee**	**Safety Integrity Level (SIL)**	**Creates a New Hazard?**
Barrier A	Risk assessment, etc.	Activator alarm or event	Define safety function	Define safe state at barrier completion	Time (minutes)	ID Role / Person	LOPA designated barriers only	Y/N
Barrier B	"	"	"	"	"	"	"	?

Table 3 - Originating Barrier Requirements: Single Task Barriers

If the barrier requires two or more tasks, the task is defined in terms of a task goal. The goal of each task contributes directly to achieving the barrier safety function. The base requirements of each task could be captured in a table similar to Table 4. Together this collection of tasks provides the design basis information needed to progress the barrier design.

Barrier / Task Origination Requirements: Multi-Task Barrier						
Barrier Name: **Barrier ID:** **Barrier Response Time (PST):** **Barrier Activator:** **Source Reference:** Risk assessment, LOPA,						
Task ID	**Task Goal**	**Task Activator**	**Human Role / Assignee**	**Task Safe State**	**Task Target Response Time**	**Safety Integrity Level (SIL)**
Task A	Define task goal	Activator alarm or event	Role / Person	Define safe state at task completion	Time (minutes)	LOPA designated barriers only
Task B	"	"	"	"	"	"

Table 4 - Originating Barrier Requirements: Multi-Task Barriers

3.3 Execution Considerations

"There is often a lack of awareness of the difference between "work-as-imagined" and the "work-as-done" (Hollnagel, 2014). "Work-as-imagined" reflects an idealised, office-based view of how task and processes are to be performed without recognizing the many situational – established work practices, practical difficulties, uncertainties, completing goals and stresses – that exists in reality at the front line. "Work-as done" captures the reality of how work is actually done…." (CIEHF 2016)

From an execution perspective, the team that participates in the RA and task analysis should include one or more knowledgeable operations specialist that has direct experience with the barrier type. A gap between the Work-as-Imagined (WAI) and the Work-as-Done (WAD) can be contributor to barrier failure. Therefore, it is essential that knowledgeable and experienced persons participate in these processes to prevent early mistakes in barrier identification and definition.

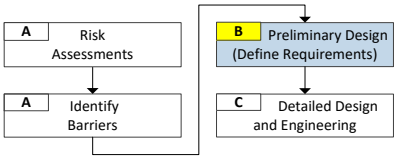
Task Analysis Participants	
Review facilitator and scribe	Facilitator: Plan, prepare and facilitate the task analysis Scribe: Record decisions, action items, etc.
Operations	Senior operations specialist(s)
Technical disciplines	Facilities engineer (layout, mechanical systems, safety critical systems, etc.) Others as needed
Process safety engineer	Knowledge of the facility's process safety design basis, risk and safety design studies Responsible for tracking the review action items

Table 5 - Suggested Task Analysis Team

4 Barrier Definition (Preliminary Design)

4.1 Introduction

This section describes the prototype barrier definition process, also referred to as the preliminary design phase. All information listed in the previous risk assessment and identification process should now be available. Figure 7 below identifies the steps needed to evaluate and define the design-basis barrier/task requirements. If using a typical stage-gated project methodology, the suggestion is to perform this work in the Front-End Engineering Design Phase of the project.



4.2 Proposed Methodology and Process

"*The performance needed to deliver the required functionality should be capable of being described clearly: i) what state or events would imitate the performance, ii) what task(s) are*

involved in carrying out the function, and iii) when the function has been achieved." (CIEHF 2016)

From Figure 7, the design progresses backwards from the response action(s) needed to achieve the barrier/task safety function and safe state.

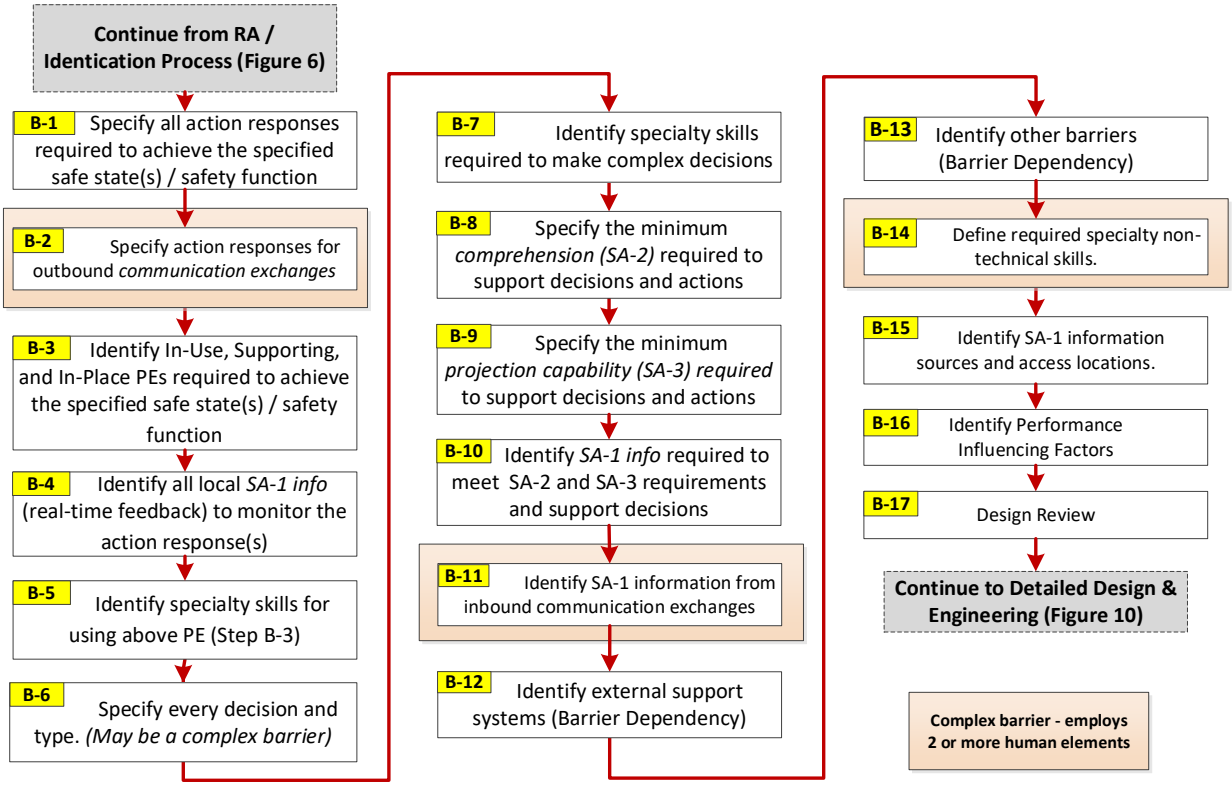


Figure 7 - Overview of Preliminary Design Process

Barriers with Multiple Tasks

Complex barriers have two or more tasks. Each barrier/task is developed using the above process. Recall Figure 1 in Section 2.1. A barrier may have two or more *human elements*. This introduces additional requirements that define how these persons interact and coordinate their efforts in a manner that reliably achieves the stated safety function and safe state. The term *non-technical skills* is often used to identify these requirements. (NTS are further discussed in Step B-14.)

Team Situation Awareness (TSA) applies to the barrier design process for multi-person barriers. For brevity, it is not included in this manuscript. For background, see Endsley and Jones (2012) and Salmon et al. (2009).

Step B-1 Specify Action Responses to Achieve the Safety Function

Define (specify) every *action response* required to achieve the barrier / task safety function and safe state. A barrier or task may have several action responses or response steps. Actions may range from

- Pressing an Emergency Shutdown pushbutton in the field or at a control room console,
- Controlling a fire hose or foam monitor,
- Updating an Incident Command Board by hand marking and updating the information on the status board.

Note: The last item may seem simple but can be complex from a cognitive perspective. Performance of the step requires receiving information from different sources at scheduled or unscheduled times. The information may be provided over a phone or radio communication (a sustain vigilance task). With each conveyance, the barrier-critical information must be correctly understood, captured and accurately recorded. Recording hand-written information on the incident command board must be legible and include the required SA-1 information. The information should also be recorded in the expected locations and use pre-determined terms. For further insight, see Taber (2010).

Step B-2 Specify Required Actions that are Outbound Communications

Barriers that employ two or more human elements require that they communicate with each other. The barrier leader conveys status information and instructions to coordinate actions. Others convey status feedback information or requests. This step defines all required response action(s) that convey outbound information in a communication exchange. Each communication should be uniquely identified, and the following information defined or specified:

- Sender and intended receiver(s)
- Message goal and purpose: convey instruction, coordinate actions, etc.
- Message type: real-time (two-way) communication, email, voice communication using a public address system (one-way), etc.
- Sender location and environment, e.g., noisy environment, proximity to danger, etc.
- Estimated message frequency, timing, and duration
- Transmission form: voice, visual, conference call, text, etc.
- Transmission systems (PE): phone, public address, video conferencing, etc.

(The receiver side of the exchange is addressed in Step B-11.)

Observation: The sender must fully attend to (focus on) the exchange to prevent an incorrect, incomplete, or inappropriate conveyance. During the exchange, the sender and receiver(s) are not available for other activities, i.e., this is a sustained vigilance activity. As such, the exchange duration should be limited to the most efficient and effective conveyance in the least amount of time. Overly lengthy exchange durations or a message conveyance error can contribute to a degraded or failed barrier / task.

Step B-3 Identify Physical Elements that Enable or Support Actions

For each action response defined in *Steps B-1 and B-2,* identify all physical elements required to perform and support the action and achieve the specified safe state. PEs to consider and identify in this step include:

- **Direct-Use PE**: Physical equipment and devices that are *directly* used to perform the action response, e.g., a fire hose, hand-held radio, stretcher, HMI data entry display, or Emergency Shutdown pushbutton.
- **Support PE**: Physical equipment and devices required to protect or support the person performing the action response, e.g., personal protective equipment, Scott air pack, portable gas detector, hand-held radio, flashlight, etc.
- **In-Place PE**: Physical features or space that must be in place to achieve the specified safe state, e.g., a protected rally/muster area.

Step B-4 Identify Required Feedback (SA-1 Information) from PE Identified in Step B-3

Identify the real-time SA-1 feedback required (if any) that may be required *by the person performing the action response.* The feedback is from the PE identified in Step B-3. Examples include:

- Direct Use PE: Information that provides real-time feedback on the effectiveness, performance, or success of the response action.
- Support PE: Information that provides real-time feedback on the operational state of support PE, e.g., the remaining air in a Scott air pack.
- In-Place PE: Information (if required) that identifies the operational state of the required in-place PE, e.g., the status of a protected muster area or rally point.

Step B-5 Define Specialized Skills Required to Use PE Defined in Step B-2

Identify the specialty skills (if any) needed to correctly use, apply or monitor the PE identified in Step B-3. These 'skills' include specialized training and knowledge that should be verified by a competency assessment process.

A deficiency in an essential skill can contribute to a degraded barrier/task performance or failure, or an injury to the action responder or others.

Step B-6 Define Decisions

Step B-6 defines every *decision* required to guide each barrier/task response action. Decision-making should be performed in a manner that is reliable, correct, appropriate, and early enough to consistently complete the action response and achieve the safe state within specified the response time (PST).

Observation: Once defined, decisions provide the basis for identifying the necessary SA-1 information, and comprehension (SA-2) and projection capability (SA-3) requirements, i.e., essential inputs to the decision-making process.

Discussion

"OGP's Human factors Sub-Committee believes that improved understanding and management of the cognitive issues that underpin the assessment of risk and safety-critical decision-making could make a significant contribution to further reduction the potential for the occurrence of incidents." (IOGP 2012)

"The situations people find themselves in can also influence the quality of their decision-making. Time pressure, poor information presentation, ambiguity of information and conflicting goals can lead to poor decisions." (SPE 2014)

Barrier decisions are often a primary contributor to barrier failure, and the most cognitively demanding. Cognitive demand (workload) and the time needed to complete the decision-making process can increase if:

- The SA-1 input information changes rapidly
- The barrier requires numerous and complex decisions
- Goal conflicts exists (Sträter 2005, p. 51, Woods et al 2010, p. 88).

A late decision may result in a failure to achieve the safe state within the specified barrier / task response time.

The time needed to make decisions increases exponentially as the number of decisions increase. (See ***Hicks Law*:** Response time = $b*\text{Log}_2 (n+1)$ where 'n' is the number of decisions. This applies to binary type decisions. (Hicks 1952))

Observation: In current practice, barrier / task decisions are often not fully identified, understood, or addressed in the barrier design process. Consider the following. With many active human barriers, the operating company chooses to insert a human into the barrier to perform a function that, as perceived, cannot be reliably performed by a fully automated barrier that resides in a safety instrumented system. A common expectation, the human has knowledge and judgement that will reduce the number of unintended barrier activations, i.e., nuisance trips. This expectation is often encompassed within an organizational practice or denoted by the terms 'Good Process Practice' or 'Well- Control.' In these cases, the operator is expected to make production versus safety judgements. Expectations of this type are implied requirements that are often not documented or integrated into the barrier design process. With all such requirements, this creates a potential entry point for hidden 'design' errors that place the barrier/task at risk.

With the possible exception of emergency response barriers, active human barriers provide a degree of flexibility, i.e., the operator can choose when to complete the actions response, provided it can be completed within the specified barrier / task response time. As will be shown, this offers flexibility but can also contribute to reduced barrier/task reliability or failure.

The following are plausible decisions that occur with active human barriers:

1) What is the required action response for this barrier?
2) Is the barrier / task activator signal valid?
3) Do I have enough information to act?
4) Do I initiate the barrier response now or wait?

Item 2 could be viewed as a *production versus safety* type decision, i.e., the approach the operator uses to make this determination might not be specified or included (using clear language) in operating procedures and training programs.

Item 4 introduces a special set of cognitive issues that may be unknown to barrier designers. Consider the situation - the barrier activator alarm is detected and the operator choses to delay the action response for one of many possible reasons. This decision sets up the following possible scenarios. (The following conversation is an early venture into the cognitive information included in Appendix D.)

1) The need to remember a future action relies on Prospective Memory, a known human weakness. The future action is stored in working memory (WM), which has a limited store capacity. In addition, the information in WM may fade (is forgotten) if not periodically refreshed. This can occur in as little as 20-30 seconds. (Endsley and Jones 2012, p. 33). A second concern pertains to time. Humans do not possess an internal clock that accurately tracks time from the perspective of 'clock time'. This creates the need to 'watch the clock' to not lose track of time. Both concerns increase the likelihood that the deferred action is forgotten or executed late, i.e., the barrier/task fails.

   Consider: Perhaps this is a preventive type active human barrier that was identified in a LOPA with a risk reduction (RR) factor of 10 taken. To increase the likelihood that this RR can be approached or, perhaps realized, consider adding features to improve reliability. Example, provide a timer that starts timing when the barrier activator occurs. The timer setpoint is set to the PST. Have the timer alarm when it is within x minutes from the PST and the action response is not started. This alarm alerts the operator of an incomplete faction, an attempt to refocus the operator on the pending action. If the response action does not occur as it nears timeout, consider automatically initiating the action response.

2) During the deferred action period, what happens if additional demands occur? Perhaps other high (or higher) priority alarms activate, or a shutdown pre-alarm alerts the operator of a pending process shutdown that will occur if prompt action is not taken. These new events introduce one or more increasingly complex decisions, i.e., which issue to attend to first? This places the original barrier at risk because attention/WM is a limited resource and the time-pressured situation increases the likelihood that non-rational biases and behaviours may influence or drive these decisions.

3) Another consequence of deferring the action in step 1, the operator may believe this act 'frees up' cognitive capability to address issues that may seem to be more immediate. Factually, the deferred action continues to consume WM, i.e., holding the pending action in working memory requires some level of attention to remind oneself of the future action. Perhaps the same is done for a second demand. The load effect is additive, and the WM capacity is limited. Another factor to consider, WM capacity may be reduced in

response to stress, excessive workload, lack of sleep, fear, etc. As the cognitive load begins to exceed one's capability, the operator may revert to a behaviour referred to as *attention tunnelling*, i.e., the person chooses to focus on one item and ignore others. The worked item is not necessarily the highest priority. Humans are often driven by undetected and non-rational biases and behaviours that may cause the person to work on a lower priority task instead of a pending higher priority task that may be more complex.

What are additional design questions to consider that can affect barrier/task reliability?

- What are acceptable reasons to defer a barrier response action once the activator has triggered?
- What is an acceptable practice on how long the operator can wait to perform the response action(s)? Is it OK to do so at the last minute as a matter of practice?
- If multiple demands occur at the same time, what guidance and training is provided on which priorities to address first? Does the training include drills under time pressure?

From CIEHF Table 1 (2016):

"Characteristics of good barrier human elements…. Does not require operator to make real-time judgements that involve safety/performance trade-offs."

"Characteristics of poor human barrier elements…. "Relies on complex judgement or decision-making, especially when there is conflict between safety and performance"

Comment: The above and similar requirements are stated in many industry practice and guidance documents and standards. Given the earlier discussion, compliance to this requirement seems questionable in many cases. Decisions unique to emergency response and well control barriers are inherently complex. A continuation of this type of mismatch appears to be a significant disconnect between standards and achievable practice, and a lost opportunity to provide the realistic guidance needed to improve the reliability of these barrier types.

Step B-7 Identify Specialty Skills (if any) Required to Make Each Decision

"Developing analytical and non-analytical reasoning skill has been shown to improve the quality of decision-making, as has the use of experiential training methods." (SPE 2014)

Decision types are:

- **Skill based:** The decision type eventually becomes habituated, i.e., is achieved using automatic processes (The 'skill' is reliable *if* the situation does not change and it occurs frequently enough to become habituated.)
- **Rule based:** The decision is based on the requirements in a procedure. The cognitive challenge is to recall the correct procedure. This invokes a conscious process to evaluate the validity of recalled procedures. (The necessity to rely on memory may be offset by a support application that automatically and reliably presents the correct procedure. The operator may use it if sufficiently trusted.)

- **Knowledge based:** A procedure is not available, or the situation is novel. Conscious processes are required to make this type of decision, an approach that is often perceived to be the least reliable. (Reason 1990)
- **Recognition-Primed Decision Making (RPD)** – RPD requires a high level of experience and expertise. This type of decision-making is commonly used by those charged with making high-risk, high-consequence decisions, especially when it occurs under time pressure. The approach relies on having sufficiently accurate mental models and the confidence and ability to mentally simulate the possible outcomes.

The barrier leader assigned to an emergency response barrier or other complex barrier types, should demonstrate the competency for RPD decision making. (For information, see Flin et al., 2008, Salas and Klein 2001) If knowledge-based decision is a possibility, this warrants an assessment to determine if and why this type of decision is needed, i.e., the barrier/task may be infeasible.

Skill and rule-based decisions are not 'special' from the context of this step.

Step B-8 Specify Comprehension (SA-2) Requirements

Specify the minimum understanding and comprehension required to correctly guide each decision and action response. This begins by examining each decision and action response to understand what must be comprehended and understood to make an informed decision and guide actions.

Comprehension is achieved by correctly understanding the meaning of the SA-1 information and its relationship to decisions and actions. The breadth and depth of a comprehension requirement depends on the process and hazard type, the barrier safety function and safe state, and the nature and source of the SA-1 information. The following examples may provide insight into the thought-processes needed to reveal and confirm comprehension requirements.

Example 1: **Facility:** Process Plant. **Barrier safety function:** Activate (press) the process safety shutdown pushbutton on activation of the High High (HH) level alarm on tank A. The barrier is defined as a single task assigned to and performed by the Control Room Operator.

- Understand the correct response action when the HH level alarm activates (i.e., the barrier trigger alarm), and barrier response time (PST).
- How long ago did the alarm activate, e.g., how much time remains to complete the action response within the PST?
- What is the priority of this barrier relative to other barriers and safety critical alarms?

Example 2: **Facility:** Offshore production platform. **Barrier safety function**: When the general alarm activates, use the designated egress routes to safely and promptly transit to the assigned emergency response stations (applies to assigned emergency responders) or to a primary or

secondary muster area or one of the designated alternates if the primary and secondary areas cannot be reached (applies to non-essential personnel). **The task goal for non-essential personnel:** When the general alarm activates, use the designated egress routes to safely and promptly transit to the primary or secondary muster areas or one of the designated alternates if the primary and secondary areas cannot be reached.

- What is my response when a general alarm activates?
- Where are the designated primary and secondary muster stations?
- Given my current location, how long will it take to reach the primary or secondary muster station?
- What are the alternate routes and escape/evacuation options if the routes to the primary and secondary muster stations are blocked? What are the other evacuation options and the hazards for using them?
- Is the selected route passable and safe, e.g., understanding the threats from debris, a route that is visually obstructed by smoke, or a nearby toxic gas warning beacon is active (lit).

*Case Study: In the DWH accident, the muster alarm (muster barrier) was activated **after** the following occurred: well blowout in progress, several explosions, widespread fires, injuries and fatalities, the loss of all emergency lighting, and the destruction of building sections and escape routes. For insight into the mental state and the many challenges presented to mustering personnel, see Skogdalen et al. (2011).*

Example 3 – **Facility:** An offshore production platform. **Situation:** A non-recoverable catastrophic event places personnel at acute risk; the facility will soon be uninhabitable. **Barrier safety function:** abandon the facility using the primary evacuation method (if feasible) or one of the alternate escape methods (lifeboat, ladder to sea, etc.) and promptly move away to a safe distance. **The barrier task goal** assigned to the Offshore Installation Manager, the barrier leader**:** Assess conditions to guide a decision to activate the abandon barrier (sound the abandon alarm).

- Do current conditions warrant abandoning the facility?
- How much time do I have to make this decision?
- How much time is needed to achieve the safe state, e.g., personnel evacuated and moved far enough away from the facility?
- How many injured do we need to move?
- Where are personnel currently located? Is everyone accounted for?
- What is the status of the evacuation vessels?
- Who is nearby that can assist with rescue operations? How far away?
- Is everyone wearing the required PPE and gear?
- What is the weather and sea state, and how does that affect the evacuation?
- Once evacuated, how long before rescue?
- What are the hazards attributed to the abandon process, e.g., potential for injury or fatality during and after the process?

- What notifications are needed?

Discussion

Comprehension relies on having the requisite mental models (MM) that encompasses the SA-1 information and the circumstances that are unique to barrier. The required MM is the product of having the right experience (e.g. depth, duration, and applicability to the barrier) and knowledge (i.e., procedural, technical, and executional). The product of combining the MM with the SA-1 information should provide the understanding and comprehension needed.

Mental Model (MM) refers to long-term memory structures and content. MM are "mechanisms whereby humans are able to generate descriptions of system purpose and form, explanations of system functioning and observed system states, and prediction of future states." (Rouse 1985) They include prototype representations (schemata) and associated action sequences (scripts). (Endsley 2012 p. 21-23)

Step B-9 Specify the Projection /Anticipation (SA-3) Requirements

Similar to Step B-8, specify the required capabilities (if any) to anticipate / project what may happen in the near future, given how the SA-1 information is changing. Because this capability tends to be limited to those with increased expertise, this may affect personnel selection and staffing. (This requirement is in additional to the knowledge and experience requirements identified in Step B-8.)

Example SA-3 requirements may include:

- A capability to project how and how quickly conditions may escalate, e.g., a toxic or flammable gas leak.
- A capability to anticipate knock-on effects that results when a barrier response action is taken.
- Anticipate workload spikes that can occur during emergency operations.

Step B-10 Define SA-1 Information Required to Support each SA-2, SA-3 and Decisions

Define all SA-1 information needed to guide decisions and achieve the specified SA-2 and SA-3 requirements.

Step B-11 Identify Required SA-1 Information: Inbound Communications

Identify all required SA-1 information received via a communication exchange.

Identify why the information is needed e.g., input to support a SA-2 requirement, to coordinate actions with others, etc. As applicable, identify the communication exchange using the same ID assigned in Step B-2.

Define information: the receiving location and environment, e.g., noisy, proximity danger, etc.

If the source of the message is not identified by Step B-2 e.g., the information comes from an external source, identify and record the 'sender' information listed in Step B-2.

See Step B-2 for additional information.

Step B-12 Identify Supporting System (Barrier Dependency)

Identify every external 'support system' required to maintain the operation state or performance of barrier elements. This step also defines the SA-1 information needed to monitor the performance and operational status of the support system.

Discussion

Active human barriers often depend on one or more external systems to maintain the operational state or performance of one or more barrier physical elements. Common support systems may include:

- Emergency power and distribution systems
- Battery-backed power systems (UPS)
- Instrument air and distribution systems
- Communication networks
- Emergency lighting
- HVAC systems

The barrier reliability is now affected by the reliability of the support system. As such, these systems should be monitored and maintained at a level that equals or exceeds the level applied to the barrier system.

Step B-13 Identify External Barriers (Barrier Dependency)

Identify every external barrier that is required to achieve and/or maintain the defined barrier/task safe state. This step also defines the SA-1 information needed to monitor the performance and operational status of the external barrier (if any).

Discussion

The barrier/ task may rely on other barriers (e.g., an external passive barrier) to achieve or maintain its specified safe state. Passive barriers are often designed to provide the safety function for a defined duration, i.e., the specified endurance time. Consequently, the safe state of the protected area (e.g., a muster/rally area) is maintained only as long as the passive barrier provides its protective function. Refer to Figure 8. A firewall and fireproofing protect the muster area (muster barrier) from a fire event for a period defined by their respective endurance times.

The endurance time of the passive barriers constrain the time available to complete control/recovery barriers safety functions.

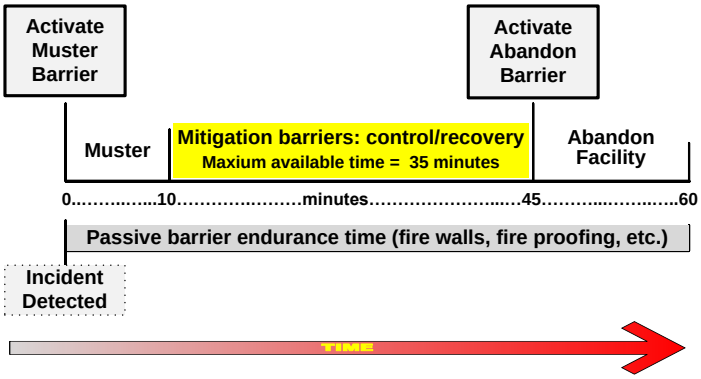


Figure 8 - Time Relationship of Passive and Active Mitigation Barriers (Offshore Example)

The barrier reliability is now affected by the reliability of the external barrier on which it depends. As such, the external barrier should be monitored and maintained at a level that equals or exceeds the level applied to the barrier system.

Step B- 14 Identify Non-Technical Skills (NTS)

Step B-14 defines the non-technical skills (NTS) required to achieve the barrier function within the process safety time. This step applies to multi-person barriers only.

NTS include the follow skill areas. Define the task specific NTS requirements for each barrier task.

- Communication (See note below)
- Teamwork
- Leadership
- Managing stress
- Coping with fatigue

Note: For *communications*, the information conveyance aspects are defined in various steps in Figures 7 and 11. Situation awareness and decision-making are often identified as NTS. Their requirements are defined in the various steps in Figures 7 and 11.

"There is lack of non-technical skills such as communications and decision-making. "(Johnsen et al. 2017)

"Non-technical skills are the cognitive and social skills that compliment workers' technical skills…. the cognitive, personal and social resource skills that compliment technical skills, and contribute to safe and efficient task performance." (Flin et al. 2008, p. 1)

"If an incident occurs, the first minutes of the response are critical to escalation prevention and to the successful conclusion of the event….Before personnel can go forward for formal assessment in emergency management, they first require training in handling emergencies at the scene and an appraisal of their capabilities under duress. Emergency management also requires specific qualities and skills, which are essentially different from those demanded by daily activities." (OPITO 2014)

"Developing proper technical and nontechnical competencies are a critical part of assuring operational safety. Both are necessary, but neither alone is sufficient… In the operational safety context, key nontechnical competencies typically include situational awareness, leadership, teamwork, communication, decision-making, risk awareness, etc." (SPE 2014)

For additional information on NTS, see: *Safety at the Sharp End*, (Flin et al. 2008) and *Introducing behavioural markers of non-technical skills in oil and gas operations.* (IOGP 2018)

Step B-15 Identify the Source and Access Locations for SA-1 Information

Step B-15 examines all SA-1 information requirements (Steps B-4, 10, 11, 12, and 13) and specifies:

- The source of each information item, e.g., technical device, signage, paint marking, communicated information, etc. (This is essential input design/procurement information for these devices and systems.)
- The location(s) where the information is accessed. This requirement can affect the type, number and location of displays or signalling devices. The physical locations may add new technical requirements, e.g., adds a new device or adds a requirement that a device must be certified for use in a hazardous area.

Step B-16 Identify Performance Influencing Factors

This step identifies the Performance Influencing Factors that can affect each barrier element. Refer to Appendix B for further information.

Step B-17 Populate Task Specifications (p/o Performance Standard)

This step compiles and records the information in the form of a Task Specification.

Figure 9 provides a high-level overview of the information developed during the preliminary design process. For single task barriers, one task specification is developed. With multi- task barriers, a task specification is developed for each task. The term *Task Specification* is used here to differentiate this information from the additional information defined in a later design process, i.e., Performance Standards. See Appendix A for a more complete discussion on Task Specifications and Performance Standards.

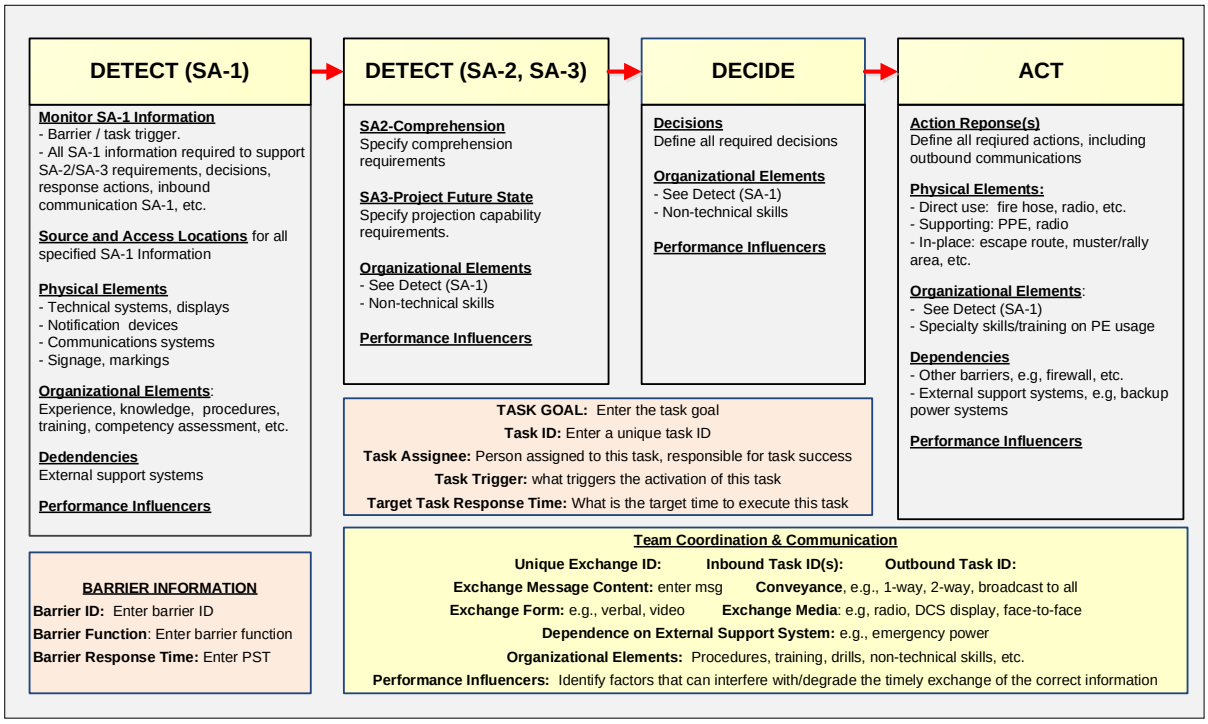


Figure 9 – Information Defined in the Preliminary Design Phase (Task Specification)

Observation: Consider the nature and scope of the information captured by the preliminary design process and ask the following questions. How does this information compare to the information developed using your current processes? What are the potential consequences if any part of this information is missing or not defined?

Observation: As a product of the preliminary design phase, compare the information captured in this phase to the information included in a Safety Instrumented Function (SIF) governed by the globally adopted standard IEC 61511. It requires a Safety Requirements Specification (SRS) for each SIF. The SIF is an active barrier that does not rely on a human to perform any of the DDA phase activities. The nature of the information captured in the Task Specification is analogous to the information captured in the SRS. Within the IEC 61511, the SRS is the foundational document that supports and enables the ability to take a full life-cycle approach to the SIFs developed by this standard. Currently, no equivalent standard exists to address active human barriers.

Step B-18 Design Review

This step is the final design review of the Task Specifications developed for each barrier. For a stage-gated project, the product of this step (and the updates that may result from this review) provides the essential information needed to begin the detailed design and engineering process.

Step B-17 marks the final design review and approval process. Reviewed items are accepted, changed, rejected, or deferred. As stated in Section 3.3, the contribution from the operation specialist is essential to preventing a gap between the Work-as-Imagined (WAI) and the Work-as-Done (WAD). A WAI-WAD mismatch can lead to barrier degradation or failure. Table 6 provides the suggested makeup of the design review team (See Figure 7, Step B-17).

Design Review Participants	
* Review facilitator/scribe	Facilitator: Plan, prepare and facilitate the design review Scribe: Record the review decisions, tabled actions, etc.
Operations	Senior operations specialist Owner representative from training/competency management department
Human factors specialist	Cognitive ergonomics specialist who performed the cognitive assessment and mitigation process.
Other technical disciplines	Industrial controls and HMI Display Systems Instrumentation Facilities engineer (layout, mechanical systems, etc.)
Process safety engineer	Knowledge of the facility's process safety design basis, risk and safety design studies. Responsible for tracking the review action items

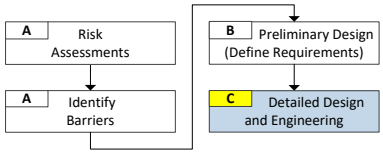
** Consider use of a facilitator and scribe on larger projects that have many barriers / tasks*

Table 6 - Suggested Design Review Team – Preliminary Design Phase

5 Barrier Detailed Design and Engineering

5.1 Introduction

This section identifies the detailed engineering and design processes proposed to fully design and engineer the barriers, tasks and elements. Figure 10 below identifies are the broad processes suggested for this design stage. For brevity, only C4 is presented. This process and the referenced processes in Appendix C provide insight into the suggested design methods and tools to identify and prevent cognitive-attributed errors at the design stage.



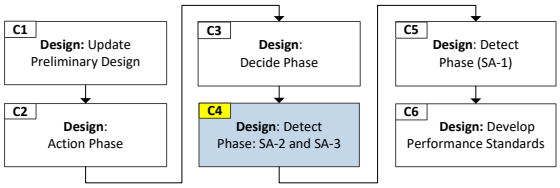


Figure 10 - Overview of Detailed Engineering and Design Processes

5.2 Detect Phase: SA-2 and SA-3 Design (Process C4 in Figure 10)

The process presented in Figure 11 below confirms the SA-2 (comprehension) and SA-3 (projection) requirements are achievable, specifies requirements for increased competency requirements and HMI support displays (if any), and captures new SA-1 requirements (if any) from these steps. Step C4-6 evaluates and mitigates the effects of Performance Influencing Factors. Step C4-7 identifies, eliminates, or mitigates situation-based cognitive mismatches

using a new tool and processes in Appendix C. (These two steps also apply to Processes C2, C3 and C5, indicated in Figure 10.)

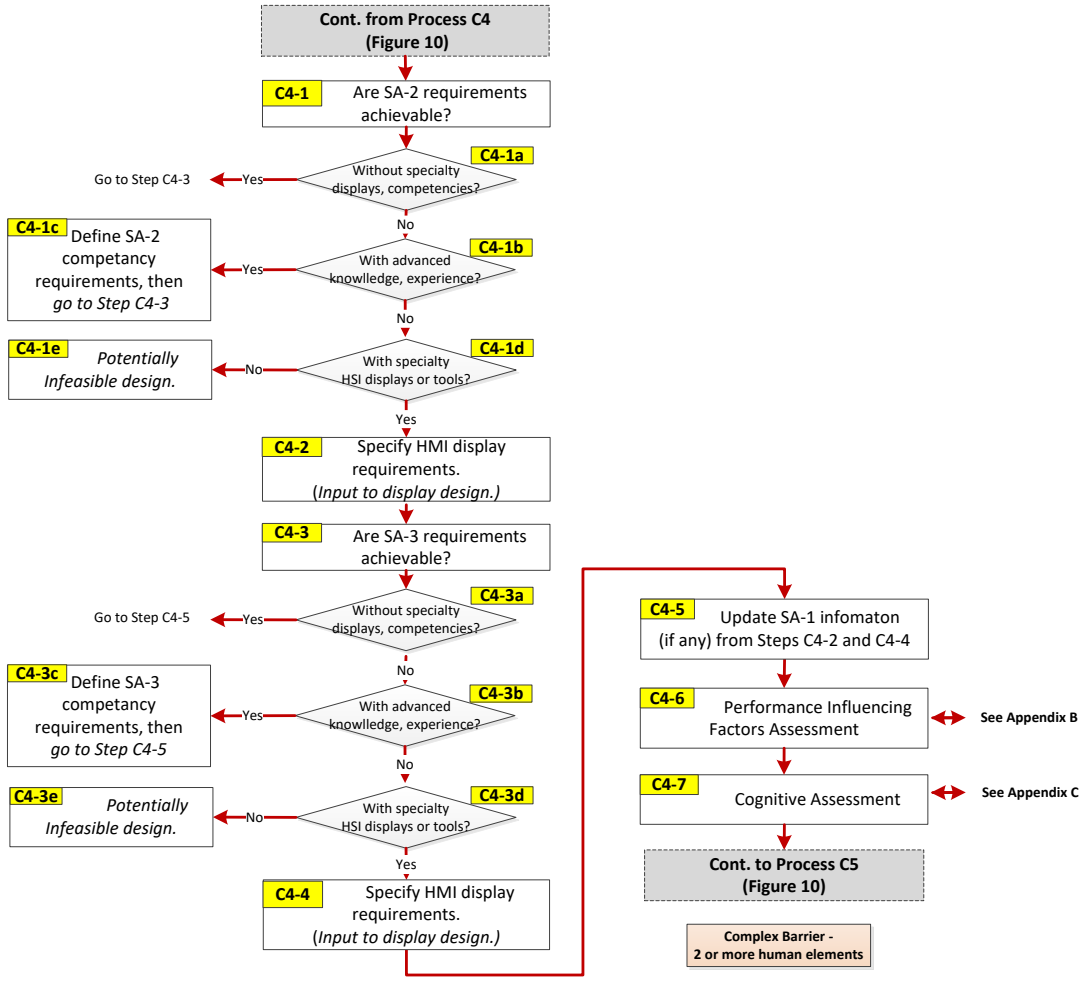


Figure 11- Process C4: SA-2, SA-3 (Detailed Design and Engineering)

Step C4-1 Confirm Feasibility of SA-2 (Comprehension) Requirements

Steps C4-1 and C4-1a seek to examine and verify that each SA-2 requirement is achievable as specified, i.e., it does not require additional knowledge and experience or a new aid-type HMI display.

If not, Step C4-1b seeks to verify if the requirement can be met by increasing the competency requirements (e.g., increased knowledge and/or experience).

Note: *An unrealistically high competency requirement may be misaligned with an operational objective to minimize changes to existing programs, e.g., staffing and staff selection. If this is the case, determine if the SA-2 requirement can be met by adding a new aid-type HMI display that translates and presents the information in a form that requires less technical knowledge or experience. (This may be achieved by a representational or ecological display.)*

Step C4-1d assesses if the requirement can be met by adding a display aid. If so, specify the requirements for the display, e.g., display purpose, function, and performance requirements. If not, the SA-2 requirement is infeasible as currently designed (Step C4-2e). This requires a return to the preliminary design phase or perhaps to the RA/barrier identification phase to seek a different solution.

Step C4-2 If Applicable, Specify and Design an HMI Display to Support a SA-2 Requirement

If Step C4-1d identifies that an aid type HMI display is required, this step defines the display requirements to a level required as input to the display designer/implementer. This may include developing a prototype display for an Operations review.

Step C4-3 Confirm Feasibility of SA-3 (Projection) Requirements

Steps C4-3, C4-3a to e and Step C4-2 are similar to the activities noted above for Steps C4-1, C4-1a, with the difference being the focus on SA-3 requirements. Example differences:

- The SA-3 requirement may be met with increased expertise (technical, process and/or execution),
- A support HMI display, if required, would differ in that its purpose is to provide guidance on a potential future states or condition based on the changing nature of the SA-1 information.

The note in Step C4-1 above also applies to this step.

Step C4-4 If Applicable, Specify and Design an HMI Display to Support a SA-3 Requirement

The activity is the same as Step C4-2, though the objective here is to design a display that supports the SA-3 requirements. The form of the display may change given its different purpose, i.e., show possible changes (near term) that may occur.

Step C4-5 Update the SA-1 Information from New HMI Displays (if any)

Update the design information to add new SA-1 information (if any) that results from adding a new support SA-2 or SA-3 HMI display.

Step C4-6 Assess for Performance Influencing Factors (PIFs)

Step C4-6 evaluates the positive and negative effects of Performance Influencing Factors (PIFs). See Appendix B for further information.

Step C4-7 Cognitive Assessment and Mitigation

See Appendix C for the processes and new tools that apply to this step. To understand the type of technical knowledge needed to employ these processes, see Appendix D.

Appendix A – Task Specification and Performance Standards

Some industry documents use the term 'performance standard' to include all of the information that defines and specifies the requirements and performance of a barrier/task. From that context, a performance standard produced by the presented processes is comprised of 1) the information in the task specifications and 2) other information developed after that information becomes available.

From CIEHF (2016, abridged):

"A human performance standard for barriers, or barrier elements, should have six characteristics:

a) *The human performance the barrier will deliver should be specific to the threat and the situation in when the barrier function is needed…*
b) *It should be clear who is expected to be involved in the delivering the required performance…*
c) *It should identify the level of competence of each of the individuals involved.*
d) *The expected timing of the performance of the function - both the initiation of the performance and its time for completion – should be appropriate to the timescale of the threat.*
e) *The standard for successful performance of the barrier should be defined…*
f) *It should document any expectations by those who approve the barrier about how operations around the barrier will be conducted that are especially critical to performing its function."*

Much of the above-identified information is defined in the preliminary and detailed design activities (Figures 7 and 10, above).

<u>Performance Standards</u>

"…performance means the properties which a barrier element must possess in order to ensure that the individual barriers and its function will be effective. It can include such aspects as capacity, reliability, availability, effectiveness, ability to withstand loads, integrity, and robustness and mobilization time." (PSA, 2013)

"Performance standards "…rarely (if ever) specify the level of human performance that needs to be achieved for the barrier function." (CIEHF 2016)

The PSA quote suggests the scope and type of the additional information to include in a barrier/task performance standard. Process 6 (Figure 10) is the intended process to develop this information. For brevity, this process is not included in this in this manuscript.

Appendix B - Performance Influencing Factors

Step B-16 (Figure 7) and Step C4-6 (Figure 11) refers to this Appendix.

This appendix provides information and guidance on how Performance Influencing Factors (PIFs) and Performance Shaping Factors (PSFs) are addressed in the presented processes.

Note: On terminology, Performance Influencing Factors and Performance Shaping Factors are often used. Depending on the document, they may or may not be interchangeable. In this manuscript, only the term PIF is used, and the term is interchangeable with PSFs.

Tables 7 to 10 list the PIFs and PSFs included the human reliability analysis and task design guidance standards. A few observations can be made when the lists are compared:

- They indicate areas of overlap, but many more areas of divergence
- Given the wide variations, the results produced by these documents may vary considerably.

PSFs	
Competency and training	"These factors should be considered when they appear of relevance to the questions at hand. The performance shaping factors have been selected to represent common root causes found in incidents and accidents across various industries." (SINTEFF 2011, para 5.2.6)
Procedures	
Human-system interface	
Teamwork	
Goal conflicts	
Time of day	
Time available	
Work environment	
Emergency response	
Interventions	

Table 7 – Performance Shaping Factors (SINTEFF (CRIOP) 2011 para 5.2.6)

Areas of safety improvement	PIFs
Control/display design	Procedures
Equipment/tool design	Communications
Memory Aids	Clarity of signs
Training	Competence
Work design	Staffing levels
Procedures	System/equipment interface
Supervision	
Reducing distractions	
Environment	
Communications	
Decision aids	
Behaviour safety	

Table 8 – Performance Influencing Factors (EI 2020, Tables 5, 7, 8, 9)

PSFs
Time
Threat stress
Task complexity
Experience and training
Procedures and supporting documentation
Human-Machine Interface
Adequacy of Organization
Teamwork
Physical working environment

Table 9 – Performance Shaping Factors (IEF/SINTEFF 2015 (Petro-HRA) Table 2.2 pp. 29-30)

Category	PIF
Organization-Based	Training program
	Corrective action program
	Other programs
	Safety culture
	Management activities - Staffing: number, qualifications, team composition, Scheduling: prioritization, frequency
	Workplace adequacy
	Resources: procedures, tools, necessary Information
Team-based	Communication
	Direct supervision: leadership, team member
	Team coordination
	Team cohesion
	Role awareness
Person-based	Attention: to task, to surroundings
	Physical and Phsyc Abilities - Alertness, fatigue, impairment, sensor limits, physical attributes, other
	Knowledge / experience
	Skills
	Familiarity with situation
	Bias
	Morale, motivation, attitude
Situation / Stressor-based	External environment
	Conditioning events
	Task load
	Time load
	Other loads: non-task, passive information
	Task complexity: cognitive, task execution
	Stress
	Perceived situation: severity, urgency
	Perceived decision: responsibility, impact (personal, plant, society)
Machine-based	HMI: input, output
	System responses

Table 10 – Performance Influencing Factors (NUREG 2012, Table 2-2)

From the context of this manuscript, the PIFs/PSFs listed in these tables appear to fall into the following categories:

1. Cognitive, e.g., attention, bias, task load, etc.
2. HMI Displays (a physical element)
3. Organizational, e.g., procedures, experience, skills, staffing, team coordination, etc.
4. Others, e.g., working environment, safety culture, etc.

Item 1 PIFs/PSFs are directly and thoroughly addressed by the cognitive-focused processes in Appendix C. For Item 2, the design requirements for specialty HMI displays are defined by the presented processes. All HMI displays are then assessed using the processes in Appendix C.

In Item 3, the presented processes specify requirements for the applicable organizational elements. To realize these requirements, they may be implemented and delivered under an overarching set of organizational (e.g., corporate) polices and standards. The full realization of the barrier/task organizational requirements may depend on the efficacy and effectiveness of those organizational policies and standards. It may be appropriate to examine them to confirm they can deliver on the stated requirements.

The presented processes do not include a process to address the PIFs/PSFs listed in Item 4, e.g., working environment. A process is required to address these factors as they can affect the performance and reliability of a barrier/task. One possible approach is the checklist approach employed in SINTEF (2011), section 5.4.

Appendix C – Cognitive Assessment Process

A. Introduction

This section provides a new tool and processes that are used to systematically identify cognitive attributed design errors and then identify solutions to eliminate or mitigate each error.

Current industry design standards and practices do not assess a safety critical function at the situation-based activities that occurs within each task phase. In practice, each phase presents the user with a different range of cognitive challenges. These challenges can further vary with changes in external conditions or the immediate state and capabilities of the individual performing the activity. These unique and seemingly transient circumstances create design-human mismatches of the type that can cause human error and, subsequently, barrier/task degradation or failure.

"…most operator errors arise from a mismatch between the properties of the system as a whole and the characteristic of human information processing. System designers have unwittingly created a work situation in which many of the normally adaptive characteristics of human cognition (its natural heuristics and biases) are transformed into dangerous liabilities." (Reasons 1990, p. 238)

"There is often a lack of understanding of the nature or complexity of the tasks – and especially the cognitive elements of those tasks – that need to be carried out for barriers to function as intended." (CIEHF 2016)

"During the discussion about cognitive and organizational ergonomics, it was found that the operator companies had little expertise and lack of relevant knowledge." (Johnsen et al. 2017)

"…most people in the industry lack awareness of the realities and limitations of human cognition and the "tricks' the brain uses to be able to function in the complex modern world." (SPE 2014)

B. Process Overview

For those charged with designing active human barriers and SCTs, some may be using a Human Reliability Assessment (HRA) process as a design tool.

"Safety suffers from the variety of methods and models used to assess human performance. For example, operations is concerned about human error while design is aligning the system to workload or situational awareness. This gap decouples safety assessment from design. As a result, system design creates constraints for the Human working at the sharp end, which eventually leads to errors. Accidents and incidents throughout all industries demonstrate the safety relevance of this gap." (Sträter 2005)

There is a clear need to develop processes and tools that can methodically and effectively identify and prevent cognitive attributed errors in active human barrier and SCTs. What is

needed to develop and implement a process that can achieve this end? What form would it take? What are the minimum essential features and functions? From the author's perspective, the process has at least three components.

1. As was done for physical ergonomics, the process begins with developing a baseline of vetted information that articulates the known human cognitive capabilities and limitations, biases and behaviours that can positively or negatively affect barrier/task reliability, effectiveness and performance.

2. Identify processes and tools that can assess barrier/task elements and components at the situation-based activities in the detect, decide and act phases. Using the information from item 1, the process would assess and identify the cognitive mismatches within each activity, and its most likely cause and context.

3. As a final step, identify the most appropriate and proven approaches that eliminate or mitigate each cognitive mismatch. The solutions should be verified to be effective and consistent with overarching design standards (e.g., company HMI display guidance standards and conventions) and organizational policies (e.g., where possible avoid competency requirements that limit the pool of personnel who could meet those requirements).

The new tool and processes described in this appendix appear to align with the above framing. Refer to Figures 12 and 13 below. These figures present the processes that appear to align with Item 2 above. Table 11 (below) is the proposed tool to address Items 1 and 3.

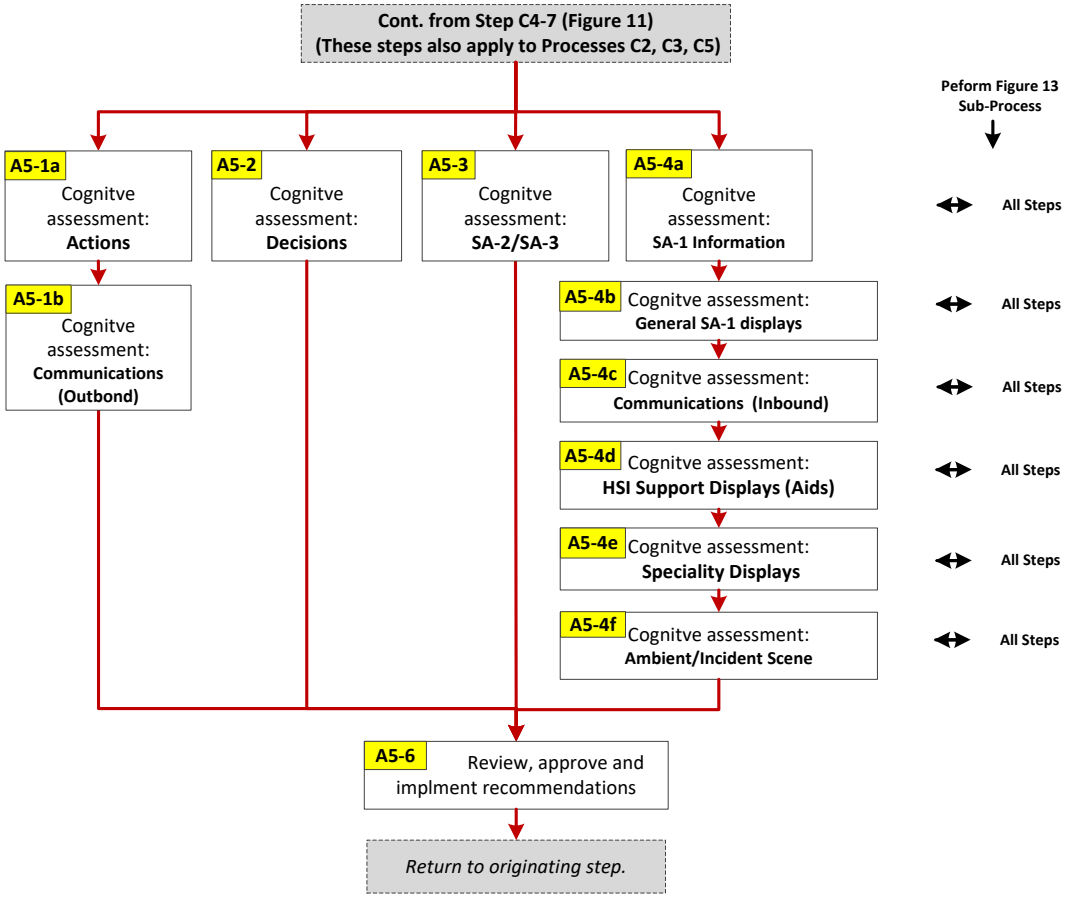


Figure 12 - Cognitive Assessment Process

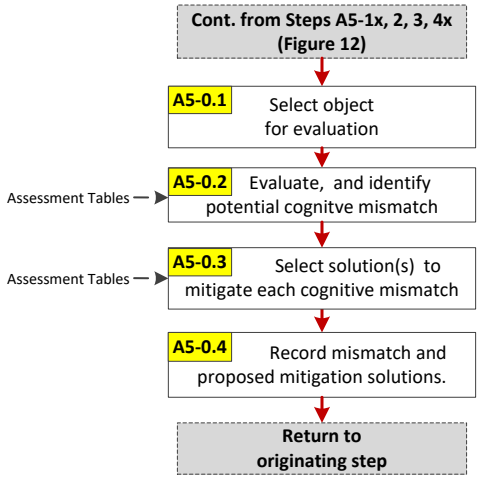


Figure 13 – Sub-Process Applies to Steps in Figure 12

Cognitive Assessment and Mitigation: Long Term Memories (LTM)					
Mismatch ID	**Cognitive Issue**	**Potential cognitive mismatch**	**Applies to Phase:**	**Potential consequence**	**Possible corrective changes to: Physical, Human, Organizational Elements**
LTM -1	Incomplete or inaccurate Mental Model (**MM**) (Endsley and Jones, 2012, pp 21-29)	**Gap or error in MM:** Procedural knowledge: Required decision, actions and permitted response time. Response priority to this barrier relative to other barriers, safety events, production issues Execution Knowledge: Typical time needed to complete decisions and response actions Potential for exposed personnel in area. Technical knowledge: Chemical / process hazards Piping systems Control system functions Barrier dependencies Limitations in displayed information, Potential escalation pathways / rate	**Detect** SA-2 Comprehension SA-3 Projection **Decide** *NOTE: See 'Attention / Attention Management' Table for MM effect on SA-1 detection and cue scanning.*	**Incorrect SA-2 comprehension or SA-2 projection contribute to decision errors that can lead to:** - No action response - Wrong action response - Late action response	**OE1** – For emergency response and exceptionally high workload environment, employ training drills under significant time pressure. (OPITO 2014) **OE2** – Employ formal training programs to develop the SA-2 comprehension (procedural and technical knowledge) and SA-3 projection/anticipation capabilities. **OE3** – Employ testing and drills to verify compliance to SA-2 and SA-3 requirements. **OE4** – Employ dynamic training simulators to develop/demonstrate RPD or SA-3 capabilities. **OE5** – Increase expertise (competency) requirements to improve SA-3 capability. **OE6** – Expand training and drill scope to provide the widest possible range of situations and scenarios to counter the *bounded rationality/keyhole* effect (Reason 2009, p 38, 57, 167/9). Include incidents from other facilities, at least one 'implausible' scenario (Weick 2007, p 28/9). **OE7** – Explicitly educate and train on the relative response priorities upon simultaneous occurrence of multiply barriers, safety situations and production issues. Consider adding OE1 or OE4 to confirm automatic responses. **HE1** – For complex environments (rapidly changing SA-1 information and threat scenarios), select personnel with the aptitude and cognitive capabilities needed to track and accurately comprehend changing conditions, timelines, and consequences. **PE1** – Provide additional SA-1 information when needed to meet the SA-2 comprehension requirements. (See *OE6* above, keyhole effect). Example for flammable or toxic gas detectors: add to the gas detector location display - wind direction and magnitude, aids to understand the potential release point and rate, presence of personnel in the affected area, etc. **PE2**– To support SA-2 requirements, provide advanced displays (e.g., representational or ecologic) that integrate/translate sensor information into a form that directly represents the potential hazard, i.e., the potential consequential/ risk from ignition, explosion and overpressure. **PE3** – To support SA-3 requirements, provide support displays that capture and translate information to show change (absolute, rate, degradation status, etc.) **PE4**- Provide a representational display that instantly shows the operational state (current/trending) of dependent support systems and barriers, e.g., a dynamic bow time display.
LTM -2	MM Drift (Le Coze, 2020, Ch. 8)	Previously verified-competent person's MM changes in ways that deviate from requirements	**Detect:** SA-2, SA-3 **Decide**	See LTM-1	**OE1 –** Competency management program require periodic/update assessments of SA-2 and SA-3 competencies **OE2 -** Competency verification expires, requiring re-verification.

Table 11- Example Cognitive Assessment Table: Mental Models

Table 11 is one of several possible pages pertaining to long- term memories. This tool presents the possible cognitive issues that can contribute (positively or negatively) to a barrier/task enhancement or failure. Using the table (and others that address different issues), the selected object (Step A5-0.1 in Figure 13) is evaluated against cognitive issues in the applicable tables. The process seeks to identify a cognitive mismatch (Step A5-0.2) and select solutions to mitigate the mismatch (Step A5.03). The last column provides possible solutions to eliminate or mitigate each mismatch. Examining the contents of Table 11 in greater detail:

- **Mismatch ID** – This is the unique identifier for the identified cognitive mismatch.
- **Cognitive issue** – Identifies the name of the issue and source reference
- **Potential Cognitive Mismatch** – Describes the nature of the possible mismatches attributed to this issue.
- **Applies to Phase** – Identifies the barrier/task phase where this issue can occur.
- **Potential consequence** – Identifies potential consequences if the mismatch is not corrected
- **Possible Corrective Changes (PCC)** – Identifies a range of possible changes to physical, human and organizational elements that may be appropriate to eliminate, minimize or mitigate the mismatch. (The list is not exhaustive so other options may also be possible. This is considered in the presented processes.) To support design transparency and traceability, the PCC ID is combined with the issue ID to create a unique ID for the selected change.

The suggestion is to create a series of tables (similar to Table 11) that address the most common and likely cognitive-attributed issues to evaluate and mitigate in the design process. Table 12 is one possible listing of the recommended topics, each with its own set of tables.

1. Working Memory (WM), WM span / short term memory (STM)
2. Attention and Attention Management
3. Long Term Memories, e.g., Mental Models, memory call-up heuristics, issues with changing memories (e.g., potential to drift towards unsafe behaviours) (Table 11)
4. Cue / SA-1 information sensing and detection
5. Decisions and decision-making
6. Time pressure / temporal monitoring and tracking
7. Non-rational behaviours, e.g., priming, task switch errors, ease, biases, e.g., confirmation bias
8. Team skills, e.g., role awareness, supervision, communication, team cohesion, team coordination

Table 12 - List of Example Cognitive Assessment Tables

Given the criticality of the tool, persons having the requisite experience and knowledge should vet it. An organizational representative may provide guidance on solution options that more closely aligned with existing organizational standards and practices, e.g., standards on training, personnel selection and staffing, etc.

Table 11 is similar to the tables in NUREG 2012, Appendix A, with some important differences that reflect their different purpose and use:

- The column "Relevant PIF(s)" contains organizational elements that are separately addressed in Figure 7 (Step B-16) and Figure 11 (Step C4-6). See Attachment B for further information on how PIFs are addressed in the presented process.
- The NUREG table is missing the column 'Possible Corrective Changes'.

NUREG 2012 appears to be a comprehensive collection of cognitive issues to consider in the design. The material is extensively sourced and referenced. As such, it appears to be a useful starting point to develop the remaining tables of the type indicated in Table 11. IEF/SINTEFF 2015 (Petro-HRA) may be another useful resource.

Another challenge is the need to seek, assess and select proven solutions to address each mismatch. The information appears to be available but tends to be scattered among published books, peer-reviewed journals and other sources that are seldom accessed by industry practitioners. (Flin 2020, Ch. 15) Possible resources to develop this part of the table includes NUREG 2012, Reason (1990, pp. 239-248), Stanton et al (2010), Sheridan (2002), CCPS 2007, Smith and Hoffman (2018), McLeod (2015), Endsley and Jones (2012), and published research and findings in peer-reviewed journals and academic publications.

A new design process should align with other likely expectations, e.g.

- Project-viable,
- Consistent with the organizations technical and organizational standards and practices,
- Transparent, auditable, and traceable, and
- Achieves an acceptable degree of consistency in the results.

The proposed processes and tools appear capable of meeting these objectives.

C. Proposed Methodology and Processes

As a general process, each step in Figure 12 (excluding Step A5-6) performs the sub-process activities indicated in Figure 13. Each row (unique mismatch) identifies the task phase to which it applies. The following summarizes how the processes in Figures 12 and 13 are intended to function.

Sub-process Step A5-0.1: Select Object for Evaluation

The process progresses in a stepwise approach that begins with selecting an object for evaluation. The following are the suggested objects.

- Step A5-1a – Action Phase: Assess each action response
- Step A5-1b – Action Phase: Assess each communication exchange (sender perspective)

- Step A5-2 – Decision Phase: Assess each decision
- Step A5-3 – Detect Phase: Assess each SA-2 and SA-3 requirement
- Step A5-4a - Detect Phase: Assess each SA-1 item (Refers to items that are *not* assessed in Steps A5-4b through A5-4f).
- Step A5-4b - Detect Phase: Assess each general HMI display while considering its intended purpose, function, presentation/layout, and the user.
- Step A5-4c – Detect Phase: Assess each communication exchange (receiver perspective)
- Step A5-4d - Detect Phase: Assess each support HMI display while considering its intended purpose, function, presentation/layout, and the user.
- Step A5-4e - Detect Phase: Assess each specialty display while considering its intended purpose, function, presentation/layout, and user. (This includes Emergency Response / Incident Command Status boards and similar displays.)
- Step A5-4f - Detect Phase: Assess SA-1 Information from the ambient environment or incident scene, e.g., visual information available from an egress route, accident scene, or an injured person

Sub-process Step A5-0.2: Evaluate Object and Identify Cognitive-Mismatches

Step A5-0.2 evaluates the object for possible cognitive mismatches using the Cognitive Assessment Tables. Additional information needed to perform the assessment (examples): task specifications/performance standards, prototypes or sketches of the proposed support and specialty displays, technical information for physical elements, etc. The assessment considers the situation-based context and nature of the cognitive activity and task. From this frame, the object is assessed against each applicable issue in the Cognitive Assessment Tables. When a mismatch is found, record the following: the assessed object (uniquely identified), the mismatch ID and the assessor's notes (if any) to support the review process in Step A5-6.

The following are examples of what to consider in the assessment:

- A review of *actions* may assess a task requirement to track sequential steps, the effectiveness of the response action, or the need for real-time monitoring of supporting and in-place physical elements.
- A review of *decisions* may assess the decision complexity and type, the time available to make decisions and if the decision is achievable given the presented design and supporting elements.

- A review of the *SA-2 and SA-3 requirements* may assess for the more complex aspects attributed to the LTM and MM or the likely effectiveness of a proposed support display.
- A review of the *communication exchanges* may assess the exchange duration and timing, the effectiveness of the exchanged information, the potential to miss or misunderstand conveyed information, and the information's effectiveness in supporting coordination and cohesion.
- A review of *SA-1 information and associated displays* may assess the salience of the presented information, working memory capacity and store duration demands, or the time and effort needed to use and access the display and the information presented in the display.

Sub-process Step A5-0.3: Recommend Corrective Solution

Step A5-0.3 reviews the mismatch from Step A5-0.2 and selects the most appropriate solution(s) from the cognitive assessment table that eliminates or mitigates the mismatch. If necessary, propose an alternate solution (i.e., not in the tables) if there is a more suitable option. Record the selected changes. For traceability, record the proposed change using a unique identifier that is a combination of the error mismatch ID and the PCC ID.

Sub-process Step A5-0.4: Record Results of Steps A5-01, 2 and 3

In Step A5-04, capture and compile the records from Steps A5-0.1, 0.2 and 0.3 in preparation for the review and approval step, Step A5-6.

Process Step A5-6 – Review and Implement Approved Cognitive Assessment Recommendations

This process, indicated in Figure 12, is common to Steps A5-1a/b, 2, 3 and 4a-f. Upon completing this process, the recommendations from the assessments are evaluated and selected, rejected, or modified. The approved recommendations are then implemented into the requisite detailed design and engineering documents and designs (i.e., physical elements), and organizational elements. The suggested activities to perform in this step:

- Review each recommendation to confirm acceptance, i.e., effective, consistent with project standards, cost and schedule impact, ALARP assessment, etc.
- When multiple recommendations apply to a common phase or element, review for mutual compatibility.
- Confirm the recommendation does not create a new hazard or cognitive challenge
- If a recommendation is not accepted, identify an acceptable alternative.
- Record the review decisions, action items, etc. to maintain design traceability.

D. Execution Considerations

Expertise Requirements

The assessment activities indicated in Figure 13 (assessment sub-process) require one or more persons having the expertise, experience and execution skills to correctly perform the activities. The person should have expert knowledge of the information in Appendix D, and have the knowledge and experience needed to correctly use the assessment tables (e.g. Table 11). This may be a challenge for organizations that have not used personnel with this skillset on capital projects. (This skillset may be found within the cognitive systems engineering discipline and some human factors programs.)

Cognitive Assessment / Recommendations Review Team

Table 13 provides the suggested team makeup to support the common review process indicated in Figure 11, Step A5-6. Use of a facilitator and scribe may be justified on larger projects that have many barriers/tasks to review.

Cognitive Assessment / Recommendations Review Team	
* Review team facilitator and scribe	Facilitator: Plan, prepare and facilitate the design review Scribe: Record the review decisions, action items, etc.
Operations	Senior operations specialist Owner representative from training/competency management program department
Cognitive ergonomics specialist	Human factors/cognitive ergonomics specialist(s) who performed the assessment
Other technical disciplines	Industrial controls and HMI Display Systems Instrumentation Facilities engineer (layout, mechanical systems, safety critical systems, etc.)
Process safety engineer	Knowledge of the facility's process safety design basis, risk and safety design studies Responsible for tracking the review action items

* Consider use of a facilitator and scribe on larger projects that have many barriers / tasks.

Table 13 – Cognitive Assessment and Recommendations Review Team

Execution Considerations to Address Project Disruption

If applied to a major capital project, the execution challenges to implement these processes may be significant in the following terms:

1. A significant reliance on senior operations personnel
2. Requirements for new expertise, e.g., cognitive systems engineer
3. Budget increase to address the new work
4. Schedule challenges to integrate these new activities into the often compresses project schedule

Comment: The scope and nature of the presented processes have many broad similarities to the global standard IEC 61511. IEC 61511 is a complex standard that governs the design and life-cycle management of safety instrumented

functions. Some of the implementation challenges included new software tools (e.g., used to calculate SIF reliability), new documents, (e.g., SIF Safety Requirements Specifications), new expertise requirements (e.g., functional safety experts), many new project steps and activities (e.g., verification and assessment processes) and knock-on/disruptive effects on other disciplines and the design/procurement processes. Over time, success did happen, and new execution models were developed. Though it took many years, today these activities are now fully integrated into the capital project design environment. From the author's experience developing new corporate and project level execution models on this scale it does appear possible to integrate the presented processes into a major capital project. Similar to IEC 61511, the roll out would likely occur over a period of years. Early efforts to do so should be attempted on small and less complex projects to gain experience and benefit from lessons learned and changes that make the processes more efficient or project amenable.

Appendix D - Human Cognition – Example Baseline Information

An Overview and Comparison of Automatic and Conscious Processes

Human cognition is the collective product of interdependent and *very different* subsystems:

- Sensory receptors and pre-processing
- Autonomic (e.g., amygdala) Ex: freeze, fight, flight (FFF)
- *Automatic* processes aka, System 1, subconscious, unconscious process
- Conscious processes aka, System 2, attention/working memory

The relative response time of these very different cognitive processes, based on a single perception cycle and indicative times:

- Autonomic process: ***Very Fast*** *e.g.*, 20 milliseconds (ms) (Edwards 2005)
- Automatic process: ***Fast*** e.g., 70-150 ms (Sträter 2005, p85,128, Carter 2014, p. 121)
- Conscious process: ***Slow*** e.g., 285 ms (Carter 2014, p. 121)

Figure 14 may help to understand what drives a potential human response when startled. The automatic and conscious processes produce very different responses to a given situation. The physical world response depends on which of these very processes controls the physical response at that moment. The startle response is the product of human evolution. It provides the quickest response that may be achieved by temporarily suppressing conscious processing, a process that is relatively slow. As such, the most likely response is one that is automatic, e.g., a habituate response. This may be beneficial in the natural world but is not necessarily the most appropriate response in the technological world. (The automatic response depends on what resides in long-term memory, i.e., the habituated response or skill.)

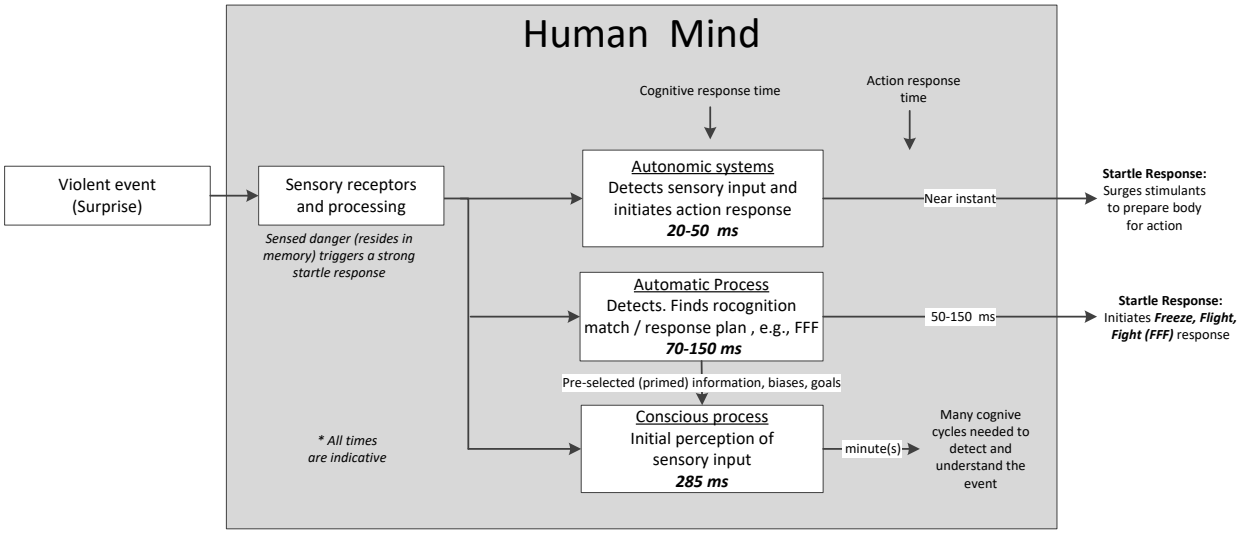


Figure 14 - Overview- Cognitive Processes and Relative Response Times

From this figure, note the pre-priming information based to the conscious processes. It serves up what the automatic processes expected, i.e., information that lies in long term memories and mental models. Conscious processes begin with this pre-selected sensory and expectation information, a source of bias driven by one's mental models.

Humans have two cognitive processes that control our daily perceptions and actions, i.e., the automatic and conscious processes. Summarized in Table 14 below, each has profoundly different capabilities, limitations, behaviours, biases and quirks. (The term 'automatic' is commonly used in the research and academic world; others may know it by the terms subconscious, unconscious or System 1.) Most incorrectly believe that our decisions and actions are realized by a conscious process. This is often not the case. Instead, humans place a high reliance on these automatic processes. Our understanding of the automatic processes increased exponentially starting in the mid 1990's, a time when functional magnetic resonance imaging systems provided a real-time look at the human brain under dynamic conditions. This added to our understanding and confirmed how automatic processes function and interacts with conscious processes.

Function	Automatic Processes	Conscious Processes
Span of Control	***Always active:*** *responsible for* 95% of daily cognitive activities (Mlodinow 2012, p34, Kahneman 2011)	***Active only when called****: On average, actively engaged to affect* ~ 5% of daily cognitive activities (Mlodinow 2012, p34, Kahneman 2011)
Normal Operation	**-** *Automatic,* continuous, and effortless. (Kahneman 2011, p20, Reason 1990, p 98) - Open loop, positive feedback only. (Sträter 2005, p118)	- Highly effortful, ***Lazy tendencies,*** *seeks 'cognitive ease'.* (Kahneman 2011, p21, Ch3) - Closed loop, negative feedback. (Sträter 2005, p118) - Runs concurrent to subconscious processes. (Reason 1990, p132-4)
Executive Mode	A **recognition engine** that continuously compares input stimulus to one's MM seeking a match. *If a match,* automatically selects associated schema/action response. *If not,* calls a conscious process to resolve. (Endsley 2012 p22-3, Kahneman 2011 p11, 24)	***Linear, sequential processing cycles.*** (Reason 2008, p12) Realized by Working memory (WM), essential to all conscious processes. (Reason 1990, p12) WM comprises an executive workspace, short term memory store (WM span), and indirect access to MM and pre-processed sensory data. (Mlodinow 2012, p64, Carter 2014 p157)
Response Time	***Recognition Time:*** **Fast** e.g., 70-200 milliseconds. (Sträter 2005, pp119/126-7, Carter 2014 p. 121) ***Recognition and prepare to initiate a learned response:*** **Fast** e.g., 200 milliseconds. (Sträter 2005, p119/126-7, Carter 2014 p121) *Skill examples:* driving or recognize a face.	***Slow:*** *Se*conds to minutes. (Carter 2014 p121)
Attention	***Attention/WM:*** *To support a habituated skill or action (e.g., driving) captures for* brief periods that go unnoticed by conscious processes. An overly focused conscious process may prevent this brief access (a dangerous condition) which also goes unnoticed.	***Attention activated by:*** - Conscious call (Reason 1990, p132, Kahneman 2011, p105) - Automatic process call, e.g., no mental model match found, etc. (Kahneman 2011, pp 24, 35)

Function	Automatic Processes	Conscious Processes
Observability	***Mostly hidden.*** *Recognition products may be experienced as an intuition or gut feel.* *Note: fMRI imaging technology, first available in the mid- 1990s, resulted in an exponential increase in research and a greater understanding of automatic processes and functions.*	***Partially.*** General visibility into the object of one's directed attention, decisions, results, some conscious processes. (Reason 2008, p12) ***Hidden*** automatic process controlled/influenced activities, e.g., memory call criteria, effects of emotions, priming, goals and beliefs, etc. (Kahneman 2011 p103, Reason 1990 pp11/2)
Decision and Analytical Capabilities	***None (See 'Executive Mode')*** ***Analytical: Limited.*** Some intuitive ability to guestimate averages, but not sums. No statistical capability. (Kahneman 2011 pp92-3)	**Yes.** Powerful analytical and decision capability. Max throughput of 10 bits/second (binary decision) (Reason 2008, p12) ***Caveat****:* Subject to hidden biases, potentially inappropriate short-cuts, memory/execution induced errors, etc. (Kahneman 2011)
Ability to Detect Danger	***Fast, continuous, automatic.*** (Sylvestre 2017 pp66-71) Detected danger limited to learned experiences only.	**Limited, if activated and tasked.** (Sylvestre 2017 pp66-71) Freeze, fight, flight (FFF) response may temporarily suppress conscious process activation.
Ability to Detect Risk	***None*** (Sylvestre 2017 pp 66-71)	**Yes,** but only if activated and tasked. (Sylvestre 2017 pp66-71)
Memory	Manages, selects and serves up **L**ong **T**erm **M**emories (LTM), e.g., mental model. LTMs can change over time with every recall, or with a single, highly emotional event.	***Memories:*** Manages short term memory store within working memory. Store limited to 7 +/- chunks, which can degrade with stress, fear, high attentional demands, etc. Memory may fade in 20-30 seconds if not refreshed. (Endsley and Jones 2012, p33)
LTM Memory Access	***Memory call criteria:*** Initially seeks a similarity match: like-with-like. If no clear solution, seeks the most frequently used (frequency gambling). (Sträter 2005, p110, Reason 2008, pp12 -25, Reason 1990 pp98, 130-147)	**Default: Automatic process controls access to LTM.** With effort and focus, able to: - Accept or reject provided LTM memory. (Reason 1990, p131 2008, p12) - Modify LTM memory call criteria to seek a different LTM memory (Reason 1990, p131, 2008, p12)
Validates Response Before Acting	***No.*** Impulsive behaviour. (Sträter 2005, pp118/9, Kahneman 2011 p85/6)	**Yes, but only with focused effort.** Otherwise: - Does not automatically check input data /decision validity *or* if essential info is missing (*Kahneman 2011 pp46, 84, 86, 99, 105)* - Tends to limit validity checks to confirming information only, i.e., c*onfirmation bias (Kahneman 2011 pp 80-82, 105*
Ability to Self-Monitor, Self-Correct	***None*** (Kahneman 2011, 41-2, 105)	**Yes, but only activated and tasked**. Provides the only means to monitor and modify one's performance, decisions, emotional state and behaviour. With cognitive overload, behaviour becomes more representative of automatic processes, e.g., less tempered / controlled. (Kahneman 2011, p24, 41, Sträter 2005, p119)
Sensory Data	***Sensory Data (~13 million bits/second)*** ***Access to sensory data: Yes***, all senses. High bandwidth. Some latency in accessed data. (Carter 2014, p79) (Temporary data store rate within sensory system up to 2 seconds, depending on sensory type.)	***Access to sensory data:*** **Yes**, *all senses* ***Max input data rate***: 10-50 bits / second stored in WM short term memory. Data latency may be several times longer than automatic process. Automatic process pre-selects sensory data based on what is expected or looked for (MM driven).

Function	Automatic Processes	Conscious Processes
Skills and Habituated Actions	***Habituated Actions and skills:*** Once learned, has full automatic control of all habituated skills, actions and routines. (Sträter 2005, p118) ***Note:*** Skills and habits become automatic with 2-6 months of continued repetition. Prior to that, control is a sliding mix of conscious/automatic processes.	**Initially a skill or habit begins as a consciously controlled activity.** (Reason 2008 pp13-14, Kahneman 2011 p35)
Time	***Temporal Capabilities: Optimal for a 10-20 second time horizon*** ***Track event sequence: Yes*** ***Track clock time: No***	**Event sequence:** Yes, subject to *WM* limitations * **Clock time:** Yes, subject to *attention* limitations, i.e., can accurately track time for periods < 30 seconds, then progressively less reliable. **Time tracking* consumes attention resources, e.g., attention capture or divided attention degrades clock-time tracking.

Table 14 - Comparative Overview of Automatic and Conscious Processes

Observation: The information in Table 14 provides clear indication that 'human error' is systematic, i.e., not random. Example, each person's unique LTMs and MMs contribute to many cognitive outcomes.

Observation: The human mind is not equipped to reliably and accurately track 'clock time', information that is essential to active human barrier barrier/task performance. ***Humans are not reliable clocks.***

Examples of Cognitive Limitations and Issues to Address in Design

The following are examples of the cognitive traits and issues that affect how human respond in different situations that trigger different cognitive challenges.

Attention

Attention is the resource that allows us to direct our attention to an object and bring it into the mind's eye. *This, combined with working memory, can only be directed to one object at a time.* This is a fundamental design constraint. Attention is directed by a conscious process or may also be directed by an automatic process. Attentional focus may degrade with conditions that also degrade working memory. The ability to maintain one's attention focused on an intended object may be jeopardized by the following:

- Distractions and interruptions
- All conditions that negatively affect Working Memory.
- Workload exceeds capacity (increased errors, error types)
- Poor attention management (Misdirected attention)
- Attention capture (more on this later)
- A continuing effort to maintain a physical or mental work-pace that is above one's 'normal' pace
- A continuing effort to maintain one's emotional state or behaviour in the presence of internal or external stress-inducing condition (Kahneman 2011, pp. 39-42)

- Any need to continuously monitor any object, information source, etc., i.e., a sustained vigilance task.

There are two types of attention capture to consider, internal preoccupation and external distractions.

Internal preoccupation (Reason 2008, p. 33-45)

- Excessive workload-induced tunnel vision (ignores information)
- Intended intense focus (lose awareness of surroundings)
- Problems at home (Misdirected attention)
- Fear, FFF activation (Re-directed attention, loss of focus)

External distractions (Reason 2008, p. 33, 45)

- *Interruptions*: 2-way radio call, ambient conversations
- *Sudden distractions*: explosion, panicky voices, smell toxic gas

Attention capture can lead to ……

- Unintended blocking out of other SA-1 inputs
- Automatic withdrawal of attention from a current task
- Execution errors, e.g., place losing, forget or misremember information in WM (Reason 2008, p. 32-3)
- Change blindness: failure to see what is not looked for, i.e., tunnel vision/attention tunnelling (Kahneman 2011, p. 23)
- *Strong habit intrusion:* Automatically perform a familiar task sequence that is <u>not</u> appropriate to the current (though similar) task; 40% of all absent-minded slips. *Reason 2008, p.42)*

<u>Working Memory</u>

From Table 14, conscious processes are slow, stepwise and sequential. Memory errors can occur in the *Detect, Decide*, or *Act* phase. *This can lead to a wide range of human error and therefore barrier/task failure scenarios*. WM error types include:

- Data from memory is forgotten or misremembered
- Forget to remember a pending future task, i.e., prospective memory (Reason 1990, p. 107)
- Lose track of time / poor time management.
- Place losing - What step am I in? (Reason 2008, p. 33)
- Lose track of task priorities and safety-critical objectives

Task Switch Error

Task switch error is a seldom discussed failure mechanism, i.e., the human fails to switch to a higher priority task when it is appropriate and warranted. Common reasons include task priority ambiguity, response priority not included in training and procedures, priorities that reflect plant culture, etc. (Wickens et al. 2015)

A timely switch to a higher priority task may fail to occur or be delayed 30% of the time. (Wickens et al. 2015) The typical cognitive time needed to switch is 300 milliseconds. Switch error examples:

- A general tendency to resist a switch to a different task, even if it is a higher priority (Sträter 2005, p. 50)
- *Plan continuation error* - a strong resistance to change tasks when nearing completion on an existing task. (Sträter 2005, p50)
- "When deciding to perform a task with drastic effects...the human is usually reluctant to undertake the task." (Sträter 2005, p. 50)
- Under high mental load, a switch may fail to occur due to attention capture and cognitive tunnelling. (Wickens et al. 2015)
- If attempting to progress two tasks simultaneously (a high mental load) the more cognitively demanding task may be dropped even though it may be higher priority. (Sträter 2005, p. 51-52, Wickens et al. 2015)

Change Blindness

Change blindness is a potential source of human error that is rooted in attention management, working memory and mental models. Attentional focus allows one to *select* specific sensory information to perceive, while other critical information may go unnoticed. (Kahneman 2011, p 23-4) Sensory perception may also be inhibited if one's attention is fully focused on a difficult mental task. (Kahneman 2011, p. 23-4) In both cases, we can easily miss important sensory information, i.e., *we look but don't see.* (Kahneman 2011, p 23-4) Tracking a change in a value or state requires remembering the initial-state information and, at some time in the future, comparing it to future-state information, a non-trivial reliance on WM/STM.

Cognitive Ease and Confirmation Bias

From the earlier discussion on working memory, this is a limited resource. Conscious processes consume a great deal of energy (e.g., sugars). Unconsciously, humans tend to seek a 'least effort' approach that unloads one's attention resources, a behaviour termed '*cognitive ease*'. (Kahneman, 2011 p. 59-78) The following are example causes and consequences of this behaviour. (Kahneman 2011, Figure 5 p. 60)

Causes of cognitive ease:

- Repeated experience
- Clear display
- Primed idea
- Good mood

Consequences of achieving cognitive ease:

- Feels familiar
- Feels true
- Feels good
- Feels effortless

Clearly, these do not reflect the proactive, rational behaviour expected from those charged with performing safety critical functions. The design challenge is to seek methods that mitigate or minimize this type of behaviour. Example biases and behaviours attributable to cognitive ease:

- **Confirmation bias**: the tendency to validate one's own understanding by seeking *confirming information, but not contrary information*. (Kahneman, 2011 p. 80-1)
- No effort is made to assess if essential information is missing.
- If a maintenance action caused an earlier spurious alarm, one tends to automatically assume the next alarm, under similar circumstances, has the same cause. (Kahneman, 2011 p. 74-5)

Confirmation bias was often cited in the white papers discussed in the introduction, i.e., one of the biases that should be addressed in the safety design process.

Bounded Rationality / Keyhole Effect

The following are two different forms of bounded rationality / key-hole effect.

The first form occurs because SA-1 information is missing, i.e., the information needed to fully understand what is happening is not available or is not in the form that adequately supports understanding (SA-2) or the ability to anticipate what might happen next. This information gap may contribute to a false or incorrect understanding. Example: alarms and indications from flammable gas detectors are often misunderstood. Not often appreciated, this approach does not

directly sense the unsafe condition, i.e., the consequence of gas ignition. The detection indication is an inferred measurement. Further, the scale of the unsafe condition depends on the gas amount (leak size, duration, concentration), where it is (enclosed or congested space, non-classified area), and the effects of the ambient conditions (wind rate and direction). As a consequence, active human barriers that rely on flammable gas detection are inherently problematic, though seldom recognized as such.

Case Study: An incomplete understanding of gas detection alarms contributed to the DWH accident. Critical links in the causal chain that enabled the second devastating explosion near engine room #3 were the failure of two active human barriers that had flammable gas alarm activators, i.e., on confirmation of flammable gas manually trip HVAC systems (closes inlet air to non-classified areas) and manually trip the gas turbine-driven generators located in the enclosed engine rooms (prevents generator over-speed caused by uncontrolled gas ingress into the turbine's combustion air supply).

The second form pertains to person's initial framing of a decision or problem space that fails to consider the actual scenario because their training and experience (e.g., one's mental models) did not include that possibility. We tend to limit our assessment to only those things that we '*know*' are possible, i.e., *it resides in our MM*. (Reason 1990 p. 38). "If I do not know about it and no one told me about it, why would I consider that as a possibility?"

"Mental models are probably one of the single most important concepts in cognitive engineering." (CIEHF 2016)

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Two views of evaluating procedural task performance: A transition from Safety-I to Safety-II approach

Changwon Son*, Farzan Sasangohar, and S. Camille Peres
Texas A&M University, College Station, TX USA

cson@tamu.edu

Abstract

Standard operating procedures (SOPs) play a critical role in achieving safety and productivity of daily operations in process industries. Incident investigations indicate that a majority of adverse events during routine and non-routine operations are attributed to issues associated with SOPs. For example, a recent investigation of fatalities points out the absence of formal procedure required to remove plugging from a waste gas piping system and inadequate emergency procedures for hazard notification and evacuation as major causes of the incident [1]. To ensure the safety of a complex system such as chemical plants, there exist two viewpoints towards human operators' use of SOPs: *Safety-I* and *Safety-II*. First, Safety-I perspective defines safety as the absence of undesired events [2]. Thus, it looks for things that went wrong and seeks to minimize the deviations from prescribed tasks, which is assumed to be a cause of adverse incidents . Second, Safety-II views safety as the presence of desirable and successful outcomes. Therefore, this standpoint highlights things that went right (e.g., adaptive behavior, workaround), and considers performance variability of human operators from SOPs to be inevitable and even necessary to maintain safe operations of the system [3]. In favor of Safety-I approach, conventional measures of SOP performance have long been established and utilized to indicate the degree of safety in relations to procedural systems. Nonetheless, no measures of human operators' SOP performance that adopt or support Safety-II framework exist to date. To address such a gap, this paper identifies limitations of traditional and dominant measures of SOP performance and then proposes a novel idea that harmonizes the two views towards safety. The new measurement concept embraces not only the conventional measures regarding the implementation of SOPs (i.e., following or not following a procedure), but also incorporates human adaptive behaviors. To instantiate the new measures of SOP performance, case studies using real-world examples in the chemical industries are presented. Following this, implications for the proposed measure of SOP performance based on Safety-II viewpoint and future research proposals to support the benefits of the measure are provided.

Keywords: Human performance, operating procedure, adaptation, safety measure

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Beyond Human Error: Integration of the Interactive Behavior Triad and Toward a Systems Model

Joseph W. Hendricks[a], S. Camille Peres[a], Pranav Kannan[b], & Lubna Ahmed[b]

[a]Ergonomics Center, Department of Environmental and Occupational Health
Texas A&M University System
College Station, TX 77843-8371, USA
jwh@tamu.edu
[b]Mary Kay O'Connor Process Safety Center, Artie McFerrin Department of Chemical Engineering,
Texas A&M University
College Station, TX-77843-3122, USA

jwh@tamu.edu

Abstract

In an effort to move beyond the "human error" explanation for safety incidents, we explored issues with procedures from the worker's point of view by using an anonymous, holistic survey developed from interviews with then currently employed operators. The current sample (N = 174; survey) included individuals employed in the process safety industry and were primarily from the Oil & Gas and Chemical industry. The survey was deployed over the course of approximately four weeks. Twelve distinct constructs emerged from the survey (e.g., perceptions of procedure quality, procedure deviation, attitudes toward the procedure change process, etc.). Results indicated that perceptions of procedure quality was the focal variable in all analyses including positive relationships with attitudes toward the procedure change process and negative relationships with procedure deviations, and both safety incidents and near-misses. Additionally, we integrated the three elements of the Interactive Behavior Triad—person, task, and context—into Dekker's Model 2 of safety. We found support for both two and three-way interactions using moderator regression analyses. These results further support a systems view and model of procedure design, implementation, and change processes. We conclude that these elements are important factors to consider when evaluating and developing procedure systems and provide additional information beyond more simplistic "human error" explanations for safety incidents.

Keywords: Procedures, process safety, survey, human error

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Operator Performance Under Stress: A Neurocentric Virtual Reality Training Approach

Yibo Zhu[1] Ranjana Mehta*[1]
[1]Texas A&M University
College Station, TX 77843, USA
rmehta@tamu.edu

Abstract

Operators in the process industries work under extreme pressures in complex hazardous environments that are associated with critical consequences at the cost of lives. Thus, ensuring operator safety is of utmost importance in this domain, and in particular in stressed contexts. Advances in Virtual Reality (VR) have enabled cost-effective, relatable, and remote trainings that can potentially transform the future of operator training in complex environments. However, consideration of operator states still remain a critical gap in ensuring that trainings are effective for ensuring performances in real world emergency response operations. The objective of the present study was to develop and evaluate the effectiveness of a stress-inducing multi-sensory operator emergency response training scenario (e.g., firefighting). Fourteen adults were trained under no-stress and stress-inducing VR training scenarios to assist in fire extinguishment. We monitored participant gaze trajectories, physiological responses, body motions, and functional brain connectivity during the training scenarios and during a post training assessment mission. Preliminary findings indicate that the stressful training scenarios, confirmed using heart rate variability metrics, resulted in reliably different neural patterns than the non-stress training scenarios. We also found that post training assessment mission after the stress training demonstrated greater neural efficiency. These initial findings suggest that training under pressure are associated with development of neurocognitive networks that may be resilient to stress often experienced by operators in real world scenarios.

Keywords: Virtual reality, emergencies, brain, human factors

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Towards a Predictive Fatigue Technology for Oil and Gas Drivers

John Kang, Abedallah Al Kader, Ran Wei, Ranjana Mehta, and Anthony McDonald
Texas A&M University
College Station, TX 77843, USA
rmehta@tamu.edu

Abstract

Driver fatigue is a critical safety risk factor that contributes to tens of thousands of motor vehicle accidents each year, resulting in injuries and deaths that cost society $109 billion annually. The issue is particularly pervasive within shift workers, who are 6 times more likely to be involved in a drowsy driving crash as compared to the general. Shift workers in the oil and gas extraction (OGE) industry may be at a greater risk of fatigue-related motor vehicle crashes because of their exposure to long hours awake with no breaks, monotonous road environments and all-night work shifts. The goals in this study are to develop and validate a predictive fatigue technology and to identify translational strategies through which this technology can be feasibly and effectively employed by the OGE industry to reduce the number of fatigued workers driving on the road, which will ultimately reduce fatigue-related motor vehicle crashes. Driving, physiological, and performance based data were obtained from twenty OGE drivers, 12 day shift and 8 night shift drivers, over the course of 3 work days (12 hour shifts) using vehicle kinematics, chest-based heart rate monitor, and a tablet-based psychomotor vigilance test. In general, night shift drivers exhibited greater physiological load and greater decrements in vigilance and alertness, however these trends were not consistent over the three days. Driving kinematics indicated that there were no significant difference of breaking behavior between day shift and night shift drivers. Only 1 severe breaking event was detected from forward/breaking acceleration data, and 4 close to severe breaking events were detected from generic longitudinal acceleration data. These events were not associated with markedly different physiological responses. Additional data analysis is underway that will further explore driver fatigue levels over multiple days to driving performance. These findings will guide future efforts on developing fatigue prediction algorithms to identify at-risk drivers such that effective fatigue management strategies (e.g., scheduling, rest guidelines) can be implemented.

Keywords: Shiftwork, fatigue, heart rate variability, alertness

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Validation of the Fatigue Risk Assessment and Management in High-Risk Environments (FRAME) Survey

Abedallah Al Kader, Stefan V. Dumlao*, Ranjana K. Mehta, & S. Camille Peres
Texas A&M University
sdumlao@tamu.edu

Abstract

The oil and gas extraction (OGE) industry continues to experience a fatality rate nearly seven times higher than that for all U.S. workers. OGE workers are exposed to intensive shift patterns and long work durations inherent in this environment. This leads to fatigue, thereby increasing risks of accidents and injuries. In the absence of any regulatory guidelines, there is a critical need for the development of *comprehensive* fatigue assessment practices specific to OGE operations that take into consideration not only the various OGE-specific sources of fatigue, but also the barriers associated with effective and feasible fatigue assessments in OGE work. In response to this need, Shortz, Mehta, Peres, Benden, and Zheng (2019) developed the Fatigue Risk Assessment & Management in high-risk Environments (FRAME) survey. Further, they provided evidence that the FRAME survey content captures fatigue-related information specific to the OGE industry not found in any one other measure of fatigue.

The present study expands on these efforts by examining the psychometric properties (i.e., reliability and validity) of the FRAME survey—a critical step before the survey can be recommended for use in practice. A sample of 200 OGE and petrochemical refinery workers were sought to participate in this study. Linkages between the FRAME survey and a number of fatigue-related measures validated for use outside of the OGE industry will be examined. Once data analysis is complete, the FRAME survey will be refined for implementation, and recommendations for implementation will be provided.

Keywords: fatigue, risk management, human factors, validation

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Identifying contributing factors of pipeline incident from PHMSA database based on NLP and text mining techniques

Guanayng Liu

Texas A&M University

College Station, Tx 77840

guanyangapply@tamu.edu

Abstract

Lessons learned from past incidents are essential to enhancing process safety of chemical industry and should be considered as knowledge legacy that evolves over time for corporate and government. Although a wealth of empirical knowledge has been accumulated from public incident databases and incident investigation reports, learnings are still limited due to the high expense of manual content analysis and lack of methodology to gain insights from past incidents.

Recently there are a few attempts that develop methods to enable automated content analysis of incident reports by natural language processing (NLP) techniques, but with a manual list of key words still needed, the methods are not intelligent or automated enough to extract information that is outside the pre-defined vocabulary. In this work, advanced NLP techniques for text mining, are employed to identify causal relations from incident reports based on unsupervised learning and co-occurrence network algorithms. The proposed method is capable of extracting latent causal factors of the incident causes described in the reports and indicating the potential of identifying root causes with more comprehensive training text data applied in the future work.

Keywords: Consequence Modeling, Facility Siting, Toxicity, Hydrogen sulfide, Impact zone, Parameter sensitivity, Natural gas composition

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Causation analysis of pipeline incidents using Artificial Neural Network

Pallavi Kumari, Mengxi Yu, Syeda Zohra Halim, and Noor Quddus
Mary Kay O'Connor Process Safety Center
Texas A&M University, College Station, TX 77840, USA
nooralquddus@tamu.edu

Abstract

Failure of hazardous liquid (HL) pipelines is a potentially significant hazard to people, property and the environment. One of the main causes of HL pipeline failures is corrosion. To predict cause and consequences of corrosion in HL pipelines, this article presents an artificial neural network (ANN) using incidents data collected by the Pipeline Hazardous Material Safety Administration (PHMSA) of the US Department of Transportation corresponding to the onshore HL transmission pipelines in the US between 2010 and 2019. From this incident database, 70 attributes has been selected for their ability to predict corrosion. Using selected attributes as input to the ANN model, the model is constructed and optimized for its hyperparameters; and it predicts the type of corrosion, total cost of property damage, net material loss and type of incident (rupture/release) with 60-90% accuracy. In order to establish credibility of developed ANN model, the model accuracy obtained using ANN model is compared against another machine learning model.

Keywords: Pipeline incidents report; Natural language processing; Artificial intelligence

1 Introduction

Pipelines are one of the safest modes to transport bulk energy and have failure rates much lower than the railroads or highway transportation[1]. However, failures do occur, and sometimes with catastrophic consequences [2]. Although pipelines failures can never be completely avoided, an appropriate and accurate risk analysis of pipeline incidents can result in reasonable and effective risk management measures to reduce the overall risk of failure.

Based on causes, pipeline incidents can be classified in five categories: corrosion,

mechanical, natural, operational error and third party [3, 4]. Corrosion can be further categorized in internal corrosion caused by the material being transported and external corrosion related to the pipeline coating and cathodic protection. Incidents due to mechanical causes consists of cracks and fractures unable to withstand the pipeline flow, and those due to natural causes are caused by events such as floods, earthquake, frost, etc. Incidents due to operational error are caused by fluctuations in operating conditions (e.g. Pressure), and third party incidents represents a damage caused by an operation not carried out by the pipeline management e.g excavation. Among all the causes, corrosion failure is ranked as one of the most frequent failure sources in oil and gas pipelines and difficult to detect [5, 3, 6]. Hence, this article primarily focuses on pipeline incidents related to corrosion.

The area of pipeline incidents analysis can be broadly divided into two categories: Data analysis and causation analysis. Data analysis analyzes pipeline failure data to derive rate of injury, fatality and failure rate. Using several data bases, such as PHMSA, CONCAWE [5, 7]. However Only data analysis does not provide a clear insight into pipeline incidents. In order to do so, causation analysis methods are present, which utilizes methods like neural network, regression technique and Bayesian methods[8, 9, 10]. In the field of using machine learning based methods, there had been a good amount of development to predict the cause of pipeline incidents. Most recently, Shaik et al.[11] has used parameters such as metal loss and weld anomalies, wall thickness and pressure flow to predict repair requirement of a pipeline.

However, these methods only uses a few attributes to predict pipeline incidents, in spite of presence of hundreds of attributes. Hence, to overcome the limitations of both of these approaches, this work proposes to first perform data analysis to select significant attributes from the rich pipeline incident database and, then, to perform causation analysis to predict cause and consequences of corrosion using a machine learning method, artificial neural network (ANN).

This article is organized as follows. Firstly, details of data used for analysis and its preprocessing method are provided. Then the proposed methodology that consists of ANN model development and model testing is illustrated in detail. The proposed methodology is demonstrated on corrosion incidents. Finally, the major findings and conclusions of this study are summarized.

2 Data processing

In North America, the spearhead oil and gas pipeline (OGP) incident database is managed by the Pipeline and Hazardous Materials Safety Administration (PHMSA) [12]. In this region, pipeline operators are required by law to report to the PHMSA every event that involves an undesired release to the environment, which meets any of the following criteria [13]:

1. The incident involves a death or personal injury necessitating in-patient hospitalization
2. Estimated property damage including cost of substance lost is $50,000 or more

In this work, the data has been collected from PHMSA database corresponding to the onshore hazardous liquid transmission pipelines in the US between 2010 and 2019. The collected data has 3592 pipeline incidents, and each pipeline incident has 606 attributes. One of the most frequent cause in OGP failures in the last 10 years has been corrosion causing 721 incidents in the collected data, for which a prediction model is developed in this work.

In order to develop a prediction model for corrosion, at first 70 out of 606 attributes are

selected based on reasoning about their relevance to pipeline failure. There are two types of attributes among selected ones: a) Generic attributes relevant to failure (e.g. time, location and area of incident), and b) Attributes specific to corrosion (e.g. presence of corrosion inhibitors and lining). Since some of the attributes are only populated for few number of incidents, selected attributes have been combined to increase information density of attributes. For example, age of pipe and age of tank has been combined to only account for age of the item involved. In this way, the number of selected attributes has been reduced to 24 from 70.

To further process the data, numerical operation has been conducted on the attributes. For example, age of the item involved in the incident is calculated by the difference of the year of manufacture of the equipment from the year of incident. Additionally, numerical attributes have been categorized in bins. For examples, age of the item involved in the incident ranges from 10 to 120 years, hence, it is categorized into 12 equal bins: 10, 20, 30, 40, 50, 60, 70, 80, 90, 100, 110, 120. Further, only most informative part of some attributes has been utilized for the analysis. For example, local time of incident has been extracted as day and night and used for analysis. Additionally, since most of these attributes (e.g., Operator location, type of commodity releases) are categorical. Hence, to maintain consistency in the dataset, the numerical inputs (i.e., age and diameter of pipe) have been categorized by putting a fuzzy value range is allocated for each linguistic variable.

The selected attributes for input, their count of categories and the categories are listed in Table 0. Among the selected attributes for input, the attributes such as the type of commodity released, area of incident, depth of cover, subpart of system involved and item involved are selected to infer information of the system that is highly likely to undergo an incident. Further, the equipment specification such as coating type, diameter and wall thickness of pipe is selected to give specific information about the pipeline. Here, Pipeline function specifies it is either transportation the commodity from the production site/well to refinery or similar facilities (gathering) or from refinery to final use or port (trunkline/transmission). It also indicates if the pipeline is operating below 20 percent specified minimum yield strength (<=20% SMYS) or above 20 percent specified minimum yield strength (>20% SMYS). Among equipment specifications, age of the item involved in the incident is calculated by the difference of the year of manufacture of the equipment from the year of incident.

Next, inspection related attributes are selected to infer information about the condition of the pipeline. Specifically, internal inspection tool indicator represents the pipeline configuration to accommodate internal inspection tools, and operation complications indicator represents presence of operational factors which significantly complicate the execution of an internal inspection tool run. Here, SCADA in place indicator and CPM in place indicator represent presence of Supervisory control and data acquisition (SCADA)-based system and Computational Pipeline Monitoring (CPM) leak detection system in place on the pipeline or facility involved in the incident, respectively.

As condition monitoring attributes, prior damage is selected which represents observable damage to the coating or paint in the vicinity of the corrosion. Attributes such as corrosion inhibitors, corrosion lining and cleaning dewatering is selected to represent commodity treatment with corrosion inhibitors or biocides presence of interior coating or lining with protective coating, and routine utilization of cleaning/dewatering pigs (or other operations).

Table 1: Input attributes

Attributes	Count	Categories
Operator location	18	TX, GA, CA, WY, PA, OK, IL, KS, AK, CO, OH, MD, UT, HI, NJ, NY, MT, NH
Local time of incident	2	Day, Night
Type of commodity released	4	Crude oil, Refined and/or petroleum product (non-HVL), HVL or other flammable or toxic fluid, Carbon dioxide/biofuel/alternative fuel
Area of incident	3	Underground, Aboveground, Tank including attached appurtenances/transitional area
Depth of cover (in)	4	50, 100, 150, >150
System subpart involved	5	Pipeline including valve sites, Terminal/tank farm equipment and piping, Pump/meter station equipment and piping, Breakout tank/storage vessel including attached appurtenances, Equipment and piping associated with belowground storage
Item involved	12	Pipe, Auxiliary piping (e.g. Drain lines), Tank/Vessel, Weld including heat affected zone, Valve, Relief line, Tubing, Meter/Prover, Flange, Scraper/pig trap/Sump/separator, Pump, Other
Part of pipe involved	3	Pipe body, Pipe seam, Others
Diameter of pipe (in)	5	5, 10, 15, 20, >20
Pipe wall thickness (in)	5	0.1, 0.2, 0.3, 0.4, >0.4
Pipeline function	4	> 20% SYMS regulated trunkline/transmission, <= 20% SYMS regulated trunkline/transmission, > 20% SYMS regulated gathering, <= 20% SYMS regulated gathering
Pipe coating type	11	Coal tar, Fusion bonded epoxy, Cold applied tape, Paint, Asphalt, Extruded polyethylene, Field applied epoxy, Polyolefin, Composite, Others, None
Age of item involved (years)	9	10, 20, 30, 40, 50, 60, 70, 80, >80
Material involved	2	Carbon steel, Others
Internal inspection tools indicator	3	Yes, No, Null
Operation complications indicator	3	Yes, No, Null
SCADA in place indicator	3	Yes, No, Null
CPM in place indicator	3	Yes, No, Null
Age of cathodic protection (years)	5	0, 10, 30, 50, 70
Prior damage	3	Yes, No, Null
Corrosion inhibitors	3	Yes, No, Null
Corrosion lining	3	Yes, No, Null
Cleaning dewatering	4	Yes, No, N/A- Not mainline pipe, Null

Age of corrosion inspection (years)	7	0, 1, 2, 3, 4, 5, >5
Age of hydrotest (years)	6	0, 10, 20, 30, 40, >40
Direct inspection type	4	Yes and an investigative dig was conducted at the point of the incident', Yes but the point of the incident was not identified as a dig site, No, Null

As output of the analysis, four attributes are selected. First, this model differentiates between cause of incidence which is internal and external corrosion. Additionally, to specific the type of incident, the incident is identified as realease, rupture and others. On the other hand, to predict the consequence of the incident, the model output is taken as cost of property damage (in dollars) and net loss of commodity released (in barrels). To increase the computational efficiency of the model, consequences has been categorized in bins of powers of 10.

Table 2: Output attributes

Attributes	**Count**	**Categories**
Cause of incident	2	Internal corrosion, External corrosion
Cost of property damage (dollars)	5	$10^5, 10^6, 10^7, 10^8, > 10^8$
Type of incident	3	Release, Rupture, Others
Net loss of commodity (barrels)	5	$10^1, 10^2, 10^3, 10^4, > 10^4$

3 Methodology

The model developed in this work is an input-output ANN model to capture the causal dependencies and the contribution of the input attributes in the pipeline failures. This model understands the synergy among underlying input attributes and their collective ability to affect the integrity of the pipeline utilizing a wealth of empirical knowledge has been accumulated from public incident databases. The methodology followed to develop the ANN model is described as a flowsheet in Figure 1.

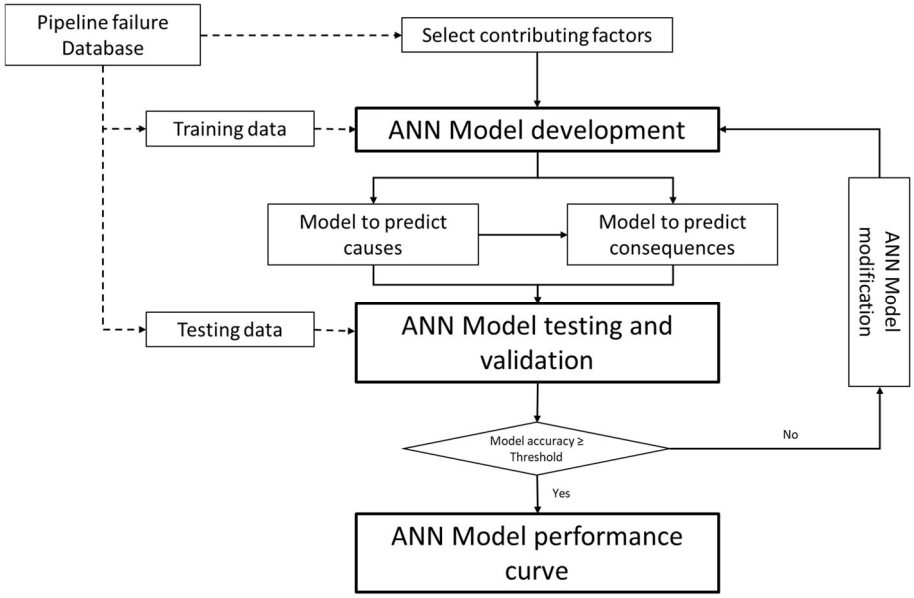

Figure 1: Model development methodology

To develop the ANN model, firstly, the contributing factors or attributes are selected from the PHMSA pipeline failure database as described in data processing section. Then, the entire data has been divided into ratio of 2:1 as training and testing data. Training data is utilized to obtain the parameters of ANN model . The structure of the ANN model developed in this work is presented in Figure 2.

3.1 ANN model development

The network structure is designed to have an input layer, two hidden layers and an output layer. Here, the inputs of the model are the attributes listed in Table 0, which is connected to the hidden layers. The first hidden layer is designed to have with twenty nodes, and the second layer is designed to have twenty five nodes. The second hidden layer is connected to the output layer, i.e., the attributes listed in Table 2. For each attribute listed in Table 2, an ANN model is developed.

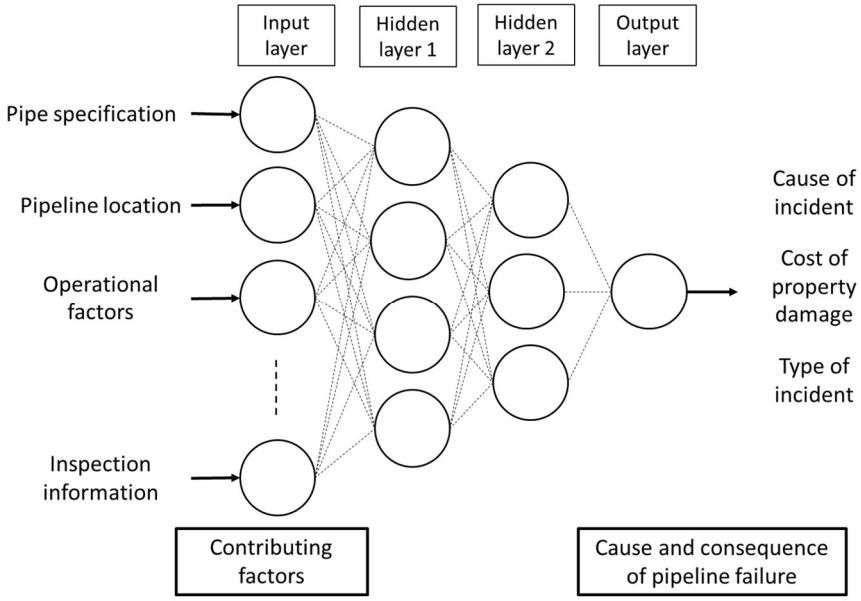

Figure 2: Structure of ANN model

In an ANN model, each node is a processing elements (also known as neuron) which is connected to other processing elements. Typically the neurons are arranged in a layer, with the output of one layer serving as the input to the next layer. A neuron may be connected to all or a subset of the neurons in the subsequent layer, with these connections modeling the causation structure of the pipeline failure. Weighted data signals entering a neuron simulate, and consequently, transfer the information within the network. The input values to a processing element are multiplied by a connection weight that simulates the strengthening of neural pathways in the causation structure of failure. It is through the adjustment of the connection strengths, i.e., weights, that learning is emulated in ANNs. This connection is shown in Figure 3, and the adjustment of connection strength is explained below:

Step 1: For each input, the input value x_i is multiplied with weights w_i; and all the multiplied values is summed to account for contribution from all the nodes in the input layer. Also, bias b is added to the summation of multiplied values.

$$z = \sum_{i=1}^{n} x_i . w_i + b \tag{1}$$

where n is the total number of inputs.

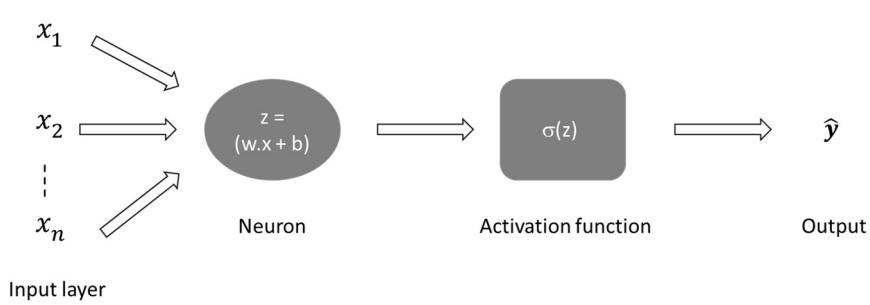

Figure 3: Structure of ANN model

Step 2: The value of z is passed to a non-linear activation function. Activation functions are used to introduce non-linearity into the output of the neurons. Moreover, they have a significant impact on the learning speed of the neural network. Here, a sigmoid activation function, given by Eq. 2, is used for the first hidden layer due to its better gradient propagation and efficient computation.

$$\sigma(z_i) = \frac{1}{1+e^{-z_i}} \tag{2}$$

Here, $i \in \{1, \ldots, n\}$ where n is the total number of inputs.

The output of first layers acts as the input for the second hidden layer. Since, the outputs of the second layers, i.e, attributes from Table 2, are categorical in nature, a softmax activation function, given by Eq. 3, is used due to its suitability for the second hidden layer.

$$\sigma(z)_j = \frac{e^{z_j}}{\sum_{j=1}^{C} e^{z_j}} \tag{3}$$

Here, $j \in \{1, \ldots, C\}$, where C is the total number of categories. In other words, softmax function gives a probability score to each category.

Step 3: The learning algorithm which consist of two parts, backpropagation and optimization is employed. In learning algorithm, loss function i.e., mean square error between actual output (y) and predicted value ($\hat{y}$), is mimimized to get optimized value of w_i and b. Here, a categorical cross-entropy loss function, given by Eq. 4, is used for suitability.

$$Loss\ \ function = -\sum_{i=1}^{C} t_i log(\sigma(z)_i) \tag{4}$$

Here, t_i is number of training points in each category. The loss function is minimized to get optimized weights and biases for each neurons.

3.2 ANN model testing

Using the trained model, output for each point in the testing data, i.e., 1/3rd of the total data is predicted. The model accuracy is calculated using Eq. 5 and reported in the results section.

$$Model\ \ accuracy = 100 * \sum_{i=1}^{C} \frac{Ture\ \ positive_i}{Total\ \ number\ \ of\ \ training\ \ data} \tag{5}$$

In order to establish credibility of developed ANN model, the model accuracy obtained using ANN model is compared against another machine learning model, support vector machine (SVM).

3.3 Support vector machine (SVM)

SVM is a supervised classification model which differentiates between two categories by building a separating plane between them and classifies points by assigning them to one of two disjoint half-spaces [14].

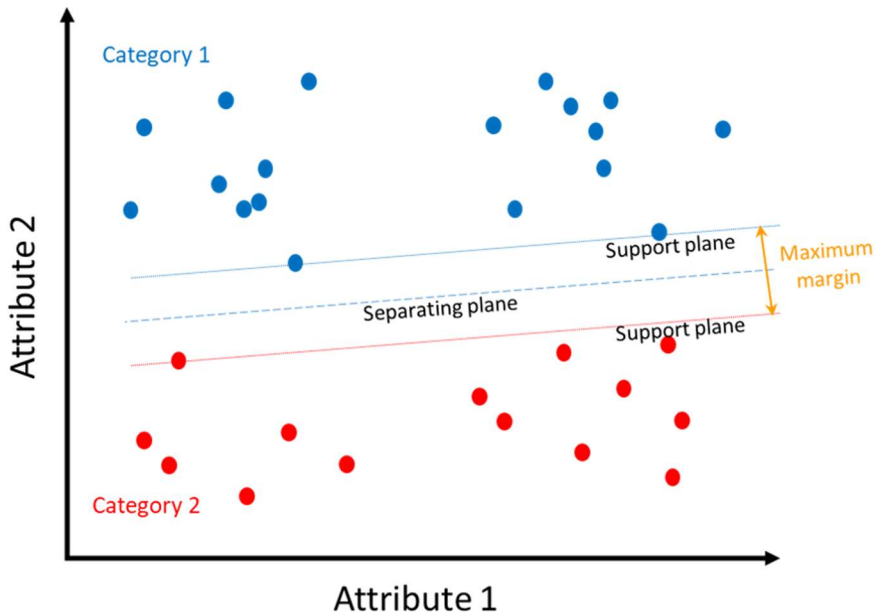

Figure 4: Support vector machine

A separating plane between two categories is characterized by weights and bias w_i, b, which is given by

$$\min_{w_i,b} \sum_{i=1}^{k} w_i y_i - b \tag{6}$$

where k is the number of attributes. To differentiate between multiple classes, i.e., for multiclass classification, SVM divides the classification problem in multiple binary classification problems.

The separating planes are built using training data. The testing data is categorized into different output categories based on which side of separating plane they lie on. SVM Model accuracy is calculated using Eq. 5.

4 Results and discussion

Using the four outputs once at a time, four ANN models are built. Each model has 26 inputs. These four models are validated and tested for their prediction accuracy. Their performances have also been compared against each other.

4.1 ANN Model training

ANN model training performances of the four models developed are compared using learning rate parameter which determines the rate to move toward a minimum of a loss function at each iteration. It can be observed in Figure 5 that learning rate of models with outputs as cause and release type are good; while learning rate of models with outputs as net loss and total cost are low, since they have higher number of categories. Since a lower learning rate implies lower model accuracy, the model accuracy of models, as shown in FIgure 6, with outputs as net loss

and total cost are lower than that of models with outputs as cause and release type.

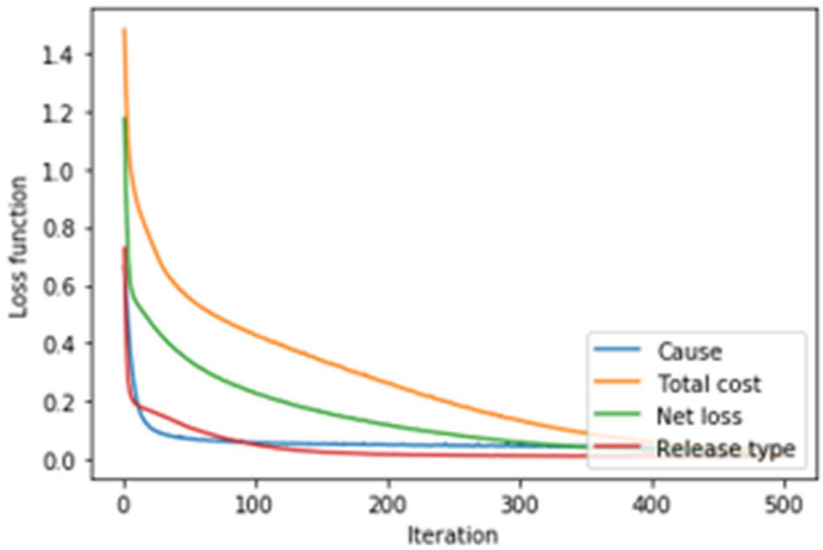

Figure 5: Loss function vs iteration for training data

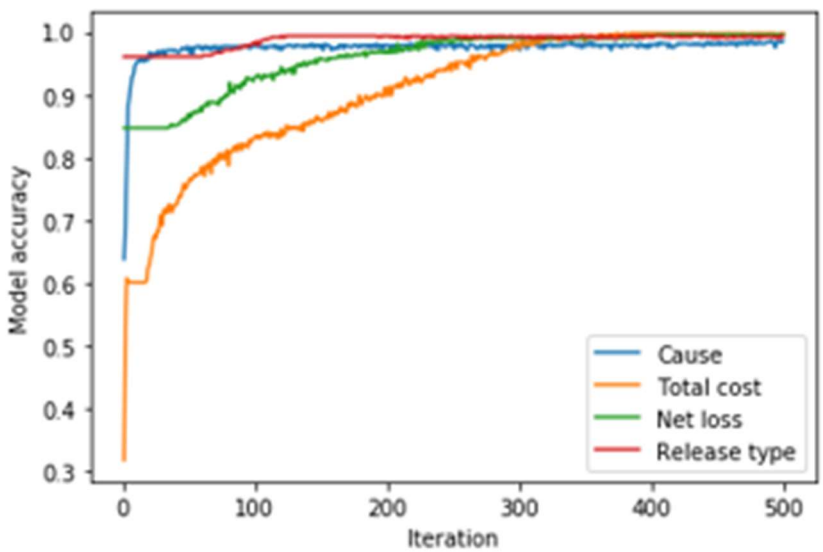

Figure 6: Model accuracy vs iteration for training data

4.2 ANN Model validation and testing

The ANN model validation performances of the four models developed are listed in column 3 of Table 3. They all are validated on a portion of training data with 90-95% accuracy.

The trained and validated model is tested using 1/3rd of the data, and the testing model accuracy is listed in column 4 of Table 3. The number of categories of each output is listed in 2nd column. It can be seen that model accuracy decreases with increase in number of categories.

4.3 ANN and SVM Model comparison

ANN model accuracy for each output is compared against the respective SVM model. As shown in Table 4, the model performance of ANN is higher for cause, net loss and release type model. However, model accuracy for total cost is poor for both the ANN and SVM model. This is due to fact that total cost of an incident is affected by other factors such as population and natural resources near the pipeline, presence of ignition source etc., which are not be present in the database.

Table 3: ANN model accuracy: Validation and testing

		ANN Accuracy	
Output	**No. of categories**	**Validation**	**Testing**
Cause (Internal/External corrosion)	2	97.40	94.54
Total cost ($< 10^6, 10^6, 10^7, 10^8, > 10^8$ in dollars)	5	89.80	60.50
Net loss ($< 10^1, 10^2, 10^3, 10^4, > 10^4$, in bbls)	5	95.76	74.79
Release type (Release, rupture, others)	3	94.53	98.80

Table 4: ANN and SVR model testing accuracy comparison

Output	**No. of categories**	**ANN**	**SVM**
Cause	2	94.54	82.80
Total cost	5	60.50	65.50
Net loss	5	74.79	81.09
Release type	3	98.80	96.60

5 Conclusion

This article presents a new framework for causation analysis of hazardous liquid pipeline incidents focused on corrosion. The proposed technique first collects and processes incident data from PHMSA database. Specifically, it eliminates the redundant attributes and selects 70 attributes resulting in higher information content. The number of attributes are reduced to 30 using process knowledge resulting in higher information density. A reasonably accurate prediction model is developed to predict the cause and consequence of corrosion, which utilizes 70 attributes resulting in higher information content. The attributes are reduced to 30 resulting in higher information density.

The proposed ANN model is applied on the preprocessed incident data and validated with 90-95% accuracy. The model performance is tested on another set of data which results in 95% model accuracy for predictive model of cause and release type and 75% model accuracy for predictive model of net loss. This article shows the strength of ANN method to predict cause and consequences of pipeline incidents and can further be extended to pipeline incidents caused bu other causes such as excavation, natural forces, etc.

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Development of Hazard Index for Engineered Nanoparticles

Nabila Nazneen[§], *Qingsheng Wang*[*]

Mary Kay O'Connor Process Safety Center
Artie McFerrin Department of Chemical Engineering
Texas A&M University
College Station, TX 77843-3122
nabila.nazneen@tamu.edu
qwang@tamu.edu

Abstract

Nanotechnology, being a comparatively new discovery in the field of science and engineering, poses many risks to the industry. Apart from its health effect, scientists and engineers are concerned about its explosion possibilities. A hazard index would be able to identify the hazard level of nanoparticles and help take proper controls of the risk associated with them. This study creates a database of the different properties of various nanoparticles and creates a hazard factor to formulate the index. The hazard factor is based on properties like explosion parameters, size, shape, dispersibility, humidity, toxicity, flammability, reactivity, etc. The study also aims to consider certain other characteristics like the level of available scientific knowledge that may impact the index given that this relatively new field of technology has more risk than the already experimented ones. Based on the hazard factor, the research will use statistical analysis to check the validity of the method and later compare the result with other existing indexes. Finally, the indexes will be ranked to precisely identify the hazard level against their respective properties. This index will be an effective indicator of a potential hazard that the engineered nanoparticles may hold and alert the users to take preventive action to moderate the risk of the hazard.

Key Words: Nanoparticles, hazard, index

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Can a virtual reality application better prepare Millennials and the Z-Generation for working with systems in the process industry?

Nir Keren, Mathew E. Harvey
Iowa State University
Ames, IA
nir@iastate.edu, meharvey@iastate.edu

Abstract

Systems in process facilities are complex. The almost 'endless' variety of components, the interaction among these components, the physical arrangement of the components in the facility, and anticipating system behaviours can overwhelm employees joining the industry. However, these employees need to evolve a sense of what industrial systems are in order to be able to grasp the various system functionalities mentioned above. Johan de Kleer[1] and his colleagues termed this sense as '*Mechanistic Mental Models*' and described it as "the common intuition of '*simulating the machines in the mind's eye'*." One thing observed for millennials and the z-generation joining the workforce is the ease at which they interact with items in the digital realm.

To address the concern above with understanding systems and take advantage of millennials and z-generation tendencies when it comes to functioning in digital environments, the authors and their associates developed a full-scale, 3D, highly interactive virtual reality application titled DesignVR. DesignVR can be used to interact with systems, as well as design and build systems, for desired industrial applications. Experiments were conducted with students to accomplsh just that.

To examine the effectiveness of DesignVR in creating proper mental models for systems, the authors asessed mental models on the following four dimentions:

(1) **System topology**: the structure of the system
(2) **Envisioning**: the inference functionality of systems components
(3) **Casual model**: the ability to describe system and components functionality

[1] de Kleer, J. & Brown, J. S. (1983). Assumptions and Ambiguities in Mechanistic Mental Models. In D. Gentner & A. L. Stevens (Eds.), *Mental models* (pp. 155-190). Hilsdale, NJ: Lawrence Erlbaum Associates Pubs.

(4) **Simulation**: the ability to conduct mental simulation for behaviour.

The utility of DesignVR as a platform for preparing the younger workforce for working with systems in procress industries is documented, discussed, and demonstrated[i].

Keyword: Systems in process industries; virtual reality; mental models of systems.

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[i] DesignVR and the virtual reality hardware it operates with will be available for attendee demonstrations and interactions.

MARY KAY O'CONNOR PROCESS SAFETY CENTER
TEXAS A&M ENGINEERING EXPERIMENT STATION

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PROCESS SAFETY RISK INDEX CALCULATION BASED ON HISTORIAN DATA

Prasad Goteti

P.Eng, TUV (Rheinland) FS Expert, CFSE
Safety Engineering Consultant
Honeywell Process Solution
Houston, TX, USA
prasad.goteti@honeywell.com

1. INTRODUCTION

The Plant Manager wants to know how Safe the Process plant is running in real time. It would be nice if the Manager had one number by which he knew what the current Process Risk is in the plant. Introducing the concept of Process Risk Index.

The intent of this paper is to introduce and explain the concept of Process Risk Index

An example using a Safety Instrumented Function (SIF) in a Process application will be used to explain this concept. The paper will detail how relevant Historian data is collected and analyzed and used to compare "Healthiness" of protection layers with design data like Layers Of Protection Analysis (LOPA) and Safety Integrity Level (SIL) calculation of the SIF to calculate the Process Risk Index.

1.1 ABBREVIATIONS AND KEYWORDS

American Petroleum Industry (API); Computer Maintenance Management System (CMMS); Key Performance Indicators (KPI); Loss Of Primary Containment (LOPC); International Electrotechnical Commissions (IEC); International Society of Automation (ISA); Independent Protection Layer (IPL); Layers Of Protection Analysis (LOPA); Long Sample Time (LST); Occupational Safety and Health Administration (OSHA); Probability of Failure on Demand (PFD); Process Risk Index (PRI); Process Safety Event (PSE); Process Safety Indicators (PSI); Risk Reduction Factor (RRF); Safety Integrity Level (SIL); Safety Instrumented Systems (SIS); Safety Life Cycle (SLC); Short Sample Time (SST); Target Mitigated Event Likelihood (TMEL)

2. INTRODUCTION TO API RP 754

API RP 754 is titled "Process Safety Performance Indicators for the Refining and Petrochemical Industries", the second edition of which came out in April 2016.

The purpose of the Recommended Practice (RP) is to identify leading and lagging indicators in the refinery and petrochemical industries whether for public reporting or for use at individual facilities including methods for the development of Key Performance Indicators (KPI). As a framework for measuring activity, status or performance, the RP classifies Process Safety Indicators (PSI) into four tiers of leading and lagging indicators. Tiers 1 and 2 are suitable for public reporting while Tier 3 and 4 are meant for internal use at individual sites.

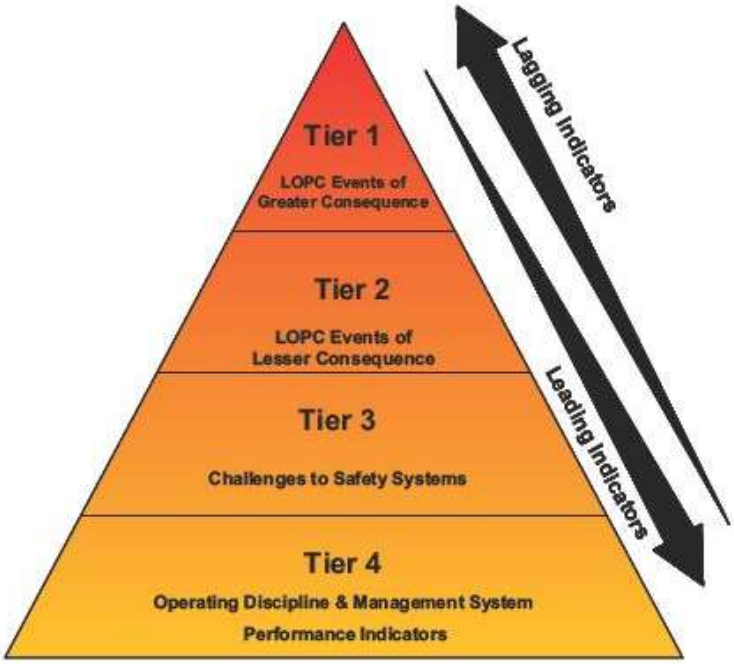


Figure-1 (Ref 8)

A Process Safety Event (PSE) is defined in this RP as an unplanned or uncontrolled Loss Of Primary Containment (LOPC) of any material including non-toxic and non-flammable material (ex. Steam or compressed air) from a process, or an undesired event or condition that, under slightly different circumstances could have resulted in a LOPC of a material.

Leading indicators inform of a potential hazardous event in advance while lagging indicators are based on facts after the hazardous event. Tier 1 and 2 generally would have more lagging than leading indicators, while tier 3 and 4 would have more leading than lagging indicators.

Identifying key leading and lagging indicators for each tier and monitoring them on a continuous basis could give an indication of Process Safety performance of a site. As an example, a major gas leak above the tolerable limits set by the local jurisdiction would classify as a Tier 1, lagging indicator. While an audit finding indicating that a proper PHA was not conducted would classify as a tier 4, leading indicator.

KEY PERFORMANCE (KPI) INDICATORS FOR PROCESS SAFETY

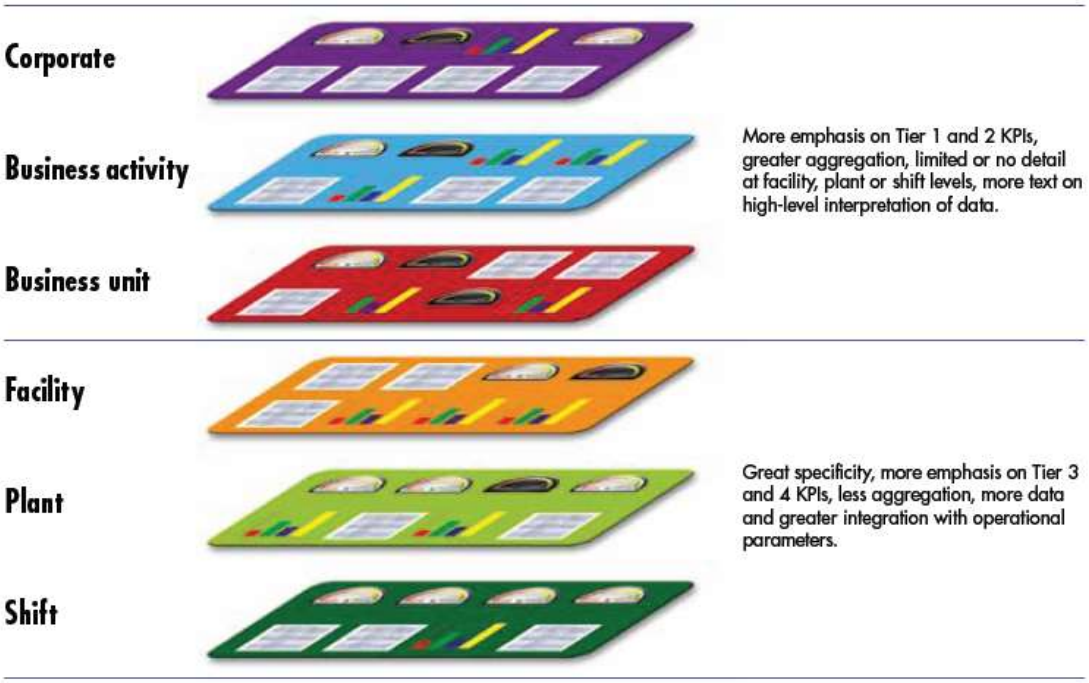


Figure-2 (Ref -9)

3. INTRODUCTION TO IEC/ISA 61511

The IEC/ISA 61511 standard defines the functional safety requirements for Safety Instrumented Systems (SIS). It focuses attention on one type of instrumented safety system used within the process sector, the Safety Instrumented System .

The intent of IEC/ISA 61511 is the management of functional safety. IEC/ISA 61511 details the activities to be performed to meet the functional safety requirement in the form of a Safety Life Cycle (SLC). The SLC covers the Analysis, Implementation and Operation phases to define, design, implement, operate, and maintain an SIS. The end user can develop his own SLC based on the guidance in IEC/ISA 61511 which is then documented in a Safety Plan.

3.1 THE ANALYSIS PHASE COVERS THE FOLLOWING STEPS

1. **Hazard and Risk Assessment** – To determine the hazards and hazardous events in the process, the initiating events leading to the hazardous event, the associated process risk, the Risk Reduction Factor (RRF) required to reduce the risk below acceptable levels, and to identify Independent Protection Layers (IPL) to achieve the necessary risk reduction based on the corporate's acceptable risk criteria.
2. **Non-SIS Solutions Applied** – Inherently Safer Design (ISD) would be the ideal choice for any process plant to achieve process safety. However, this is not always possible due to the hazardous material and chemical processes used in the process industry. Key considerations for ISD, as described by Trevor Kletz [ref. 7], are to minimize, substitute, moderate, and simplify. Also the use of non-SIS layers such as Pressure Relief Valves, Rupture Discs, etc. (refer to Figure 1) are options that a project team considers to meet identified process safety risks. SIS layers would usually be the last option to be considered.
3. **Allocation of safety functions to protection layers** – To identify Safety Instrumented Functions (SIF) as one of the IPLs in step 1 and determine their Safety Integrity Levels (SIL) based on the extent of Risk Reduction Factor (RRF) taken credit for (refer to "Necessary Risk Reduction " in Figure 3). Initial SIL verification calculations for each SIF are sometimes generated as part of this step.
4. **Safety Requirement Specification (SRS)** – Generate a document or set of documents which define the Functional and Integrity requirements of each identified SIF in the SIS.

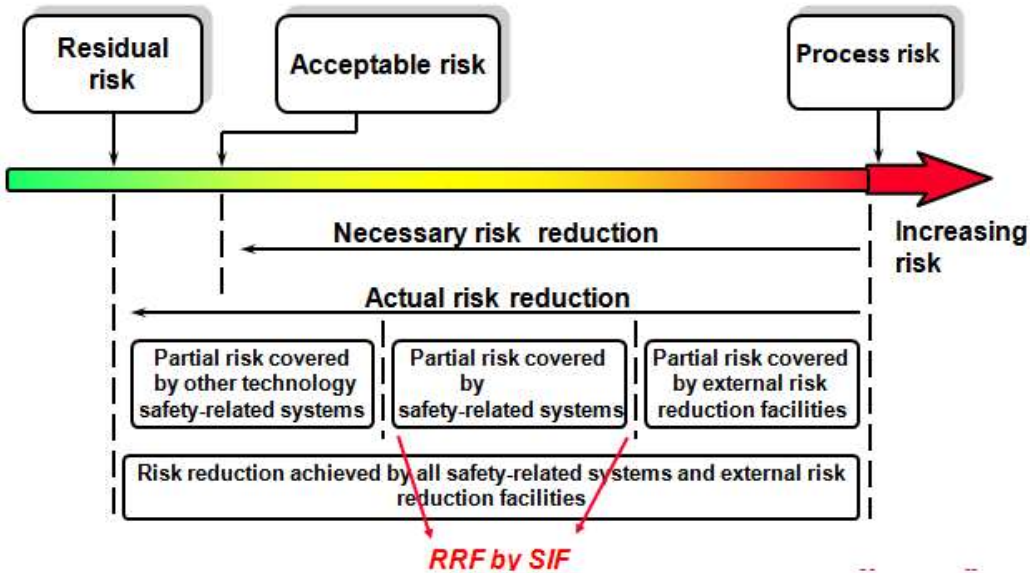


Figure 3 – Risk Reduction by Independent Layers of Protection

3.2 THE IMPLEMENTATION PHASE COVERS THE FOLLOWING STEPS

1. **SIS Design and Engineering** – Design and Engineering of the SIS to meet the Functional and Integrity requirements in the SRS. SIL verification calculations are generated based on the instrumentation selected for each SIF. The RRF calculated for each safety instrumented function (SIF) needs to be equal to or greater than the RRF values determined during the Analysis phase. The RRF of the SIF represents a portion of the total risk reduction as indicated in figure 3. SIL verification calculations are based on various parameters, like failure rates of the instruments, Proof Test Intervals (PTI), and higher diagnostic on valves (Partial Valve Stroke Testing or PVST).
2. **SIS Installation, Commissioning and Validation** – To install, commission, and validate that the SIS meets the Functional and Integrity requirements in the SRS

3.3 THE OPERATION PHASE COVERS THE FOLLOWING STEPS

1. **SIS Operation and Maintenance** – Operate and Maintain the system based on the requirements in the SRS. Look for Key Performance Indicators (KPI) which will inform the Operator of any SIF failures.
2. **SIS Modification** – Use an approved Management Of Change (MOC) procedure to manage any changes to the SIS after installation and commissioning. The MOC process usually begins with an "impact analysis" based on the proposed changes to ensure no negative impact to the original design requirements of the SIS. Proposed changes need to be validated, reviewed, approved, and communicated before the changes are incorporated. Requests for modifications can come from either an operational change in the process or during the once every 5 year OSHA PSM specified HAZOP revalidation.
3. **SIS Decommissioning** – De-commission the SIS when all process hazards no longer require a safety function.

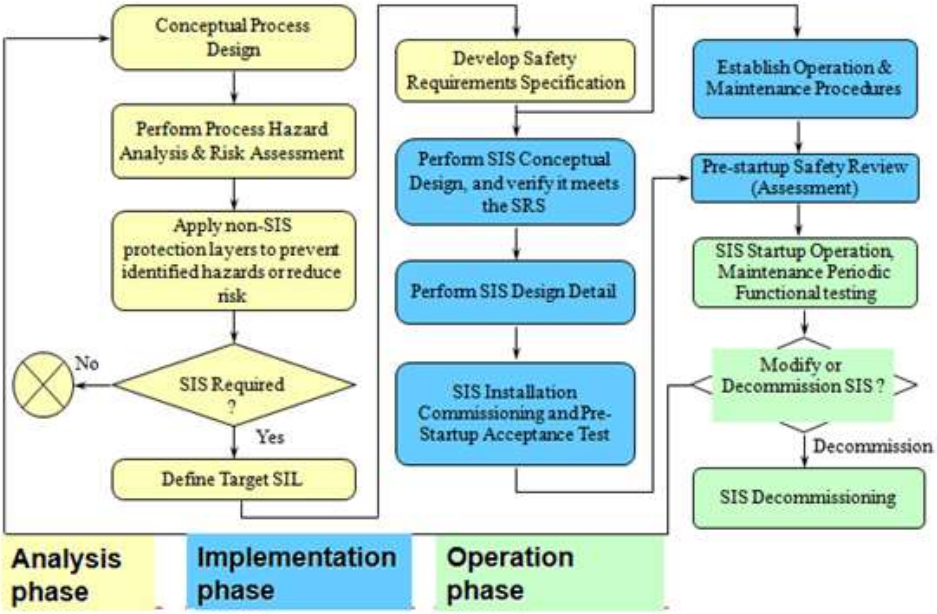


Figure 4 – Simplified representation of the Safety Life Cycle

4. PROCESS RISK INDEX CALCULATION

4.1 CONSIDER THE FOLLOWING PROCESS

Hydrocarbon feed to a pressure vessel (V-1). The Upstream pressure to Vessel V1 is greater than 5 Atmospheres and the Maximum Allowable Working Pressure (MAWP) of vessel V-1 is 5 Atmospheres. The pressure in V-1 is controlled at 3 Atmospheres by PIC-1 through the Basic Process Control System (BPCS). When the pressure crosses the alarm limit due to failure of PIC-1, PZT-4275 will sense and send the signal to a Safety Instrumented System (SIS) logic solver, the interlock PSHH-1 (set at 3.75 Atmospheres) will initiate shutdown of XZV-4275, which is a De-energized To Trip (DTT), Fail Close valve, ie Open when the pressure is normal. The Pressure Safety Valve (PSV-1) pops up in the event the pressure in V-1 reaches 4 Atmospheres.

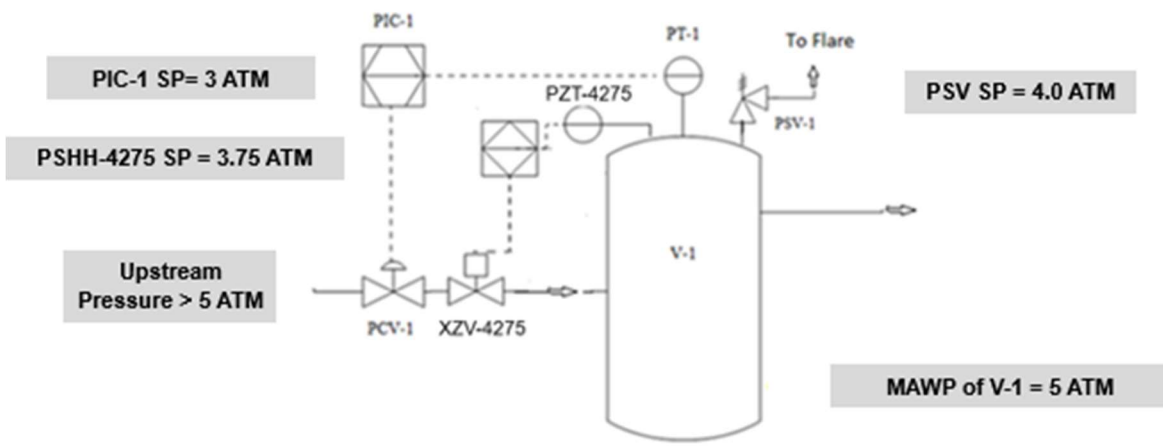


Figure-5

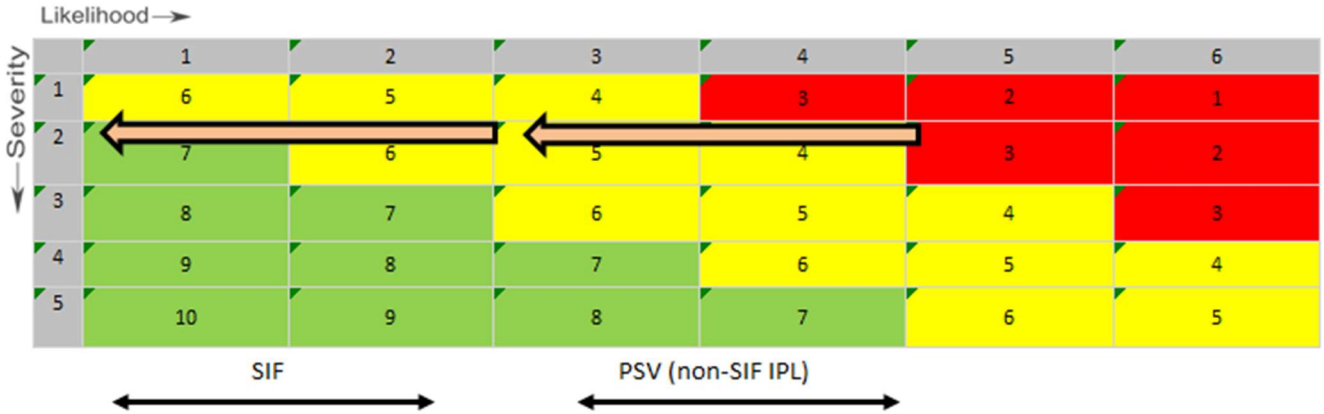
4.2 PART OF ANALYSIS PHASE OF SLC – HAZOP AND SIL DETERMINATION

In this example, it is assumed the PHA study indicated that uncontrolled High pressure in V-1 can lead to vessel rupture and release of hydrocarbons in the atmosphere leading to potential explosion. Refer figure -6, which is the PHA Risk matrix that was used. The various scenarios based on the Risk matrix are detailed in Table 1.

Scenario	Likelihood	Severity	Risk
Without any safeguards	5	2	3 (Unacceptable)
With PSV-1 only as safeguard	3	2	5 (Tolerable)
With PSV-1 and PSHH-1 as safeguards	1	2	7 (Acceptable)

Table-1

The Target Mitigated Event Likelihood (TMEL) for Severity 2 considered is 1E-05 per year.



Where :

	Likelihood
1	Once in 10000 years
2	Once in 1000 years
3	Once in 100 years
4	Once in Ten Years
5	Once a year
6	Multiple times per year

	Severity
1	Multiple fatilities
2	Single Fatility
3	Serious injury
4	First Aid
5	First Aid

Figure-6

4.3 USE OF KPI'S DURING OPERATION PHASE OF THE SAFETY LIFE CYCLE

After the IPLs are designed and implemented, it would be good to know how the IPLs are functioning. Based on API RP 754, the KPI's that can be identified in our example would be (refer figure 1 and 5):

1. **Release of PSV-1 to Flare (Tier 1 or 2 KPI)** –This would mean both the BPCS and SIS loop had failed to maintain the pressure in the vessel below dangerous levels. This would be a lagging indicator and could be classified as Tier 1 or 2 by the individual site based on amount of gas released to flare.
2. **SIF-1 exercised (Tier 3)** –This would mean that the Pressure in the vessel was not controlled by the BPCS loop and reached a limit where SIF-1 had to shut the Hydrocarbon inlet line. This would be a Tier 3, leading indicator as far as LOPC is concerned but a lagging indicator in terms of Process Availability.
3. **Audit findings (Tier 4)** - If an Audit finding indicates that the SIF-1 field instruments are not being Proof Tested as was considered during the SIL verification calculations, this will be informed to the individual site management as a Tier 4 leading indicator.
4. **SIF component, detected failure (Tier 4)** – If the input transmitter of SIF-1 fails and is detected or bypassed, the SIF is now running in a degraded mode. The component needs to be fixed and restored so that SIF-1 can contribute to the risk reduction it was designed for. This would be a Tier 4 leading indicator.

4.4 OPERATION PHASE OF SLC – PROCESS RISK INDEX (TIER 3 OR 4)

Process Risk Index (PRI) is one number which indicates the Process Risk profile of a Process unit in real time (Short term) or over a period of time (Long term). PRI is based on Hazardous event scenarios which are High Severity (Safety , Commercial or Environmental) , ie which have a Base "Unacceptable" risk without any safeguards.

If PRI=0%, it means the Process Risk is within the "Acceptable" criteria of the Operating company
If PRI=100%, it means the Process Risk is in the "Un-Acceptable" criteria of the Operating company

4.5 TYPES OF PROCESS RISK INDEX

Short Term (ST) PRI is for a period of one shift or One day. This is for the Plant operations and maintenance manager to get an idea how their Process plant is doing

Long Term (LT) PRI is for a period of a few months and above. This is for the Senior management and Plant managers to know how the Process plant has been doing in the long term.

4.5.1 Assumptions for ST PRI equations

1. "Safety" is the driver for this hazardous event (not Commercial and Environment). So from now on we will refer to it as Safety Risk Index.
2. PFDactual of SIF and non-SIF IPL is the same as PFD per design
3. The SIF input has 1oo1 input voting
4. All other IPLs are working per design

4.5.2 Variable which effects Short Term (ST) Safety Risk Index

1. SIF "Time in Bypass" over the Short term period. This data is available in the Plant Historian (Figure 7)

4.5.3 Equations for Short Term (ST) Safety Risk Index for ONE scenario

1. **Designed ST Safety Risk** = TMEL (for safety) x Safety Severity
   (the assumption here is that with the designed IPLs, the TMEL has been met)

2. **Actual ST Safety Risk** = IEF x [(PFD of non-SIF IPL x SIF PFD) x (Time SIF NOT in Bypass/SST) + (PFD of non-SIF IPL) x (Time SIF in Bypass/SST)] x Safety Severity

   where :

   IEF = Initiating Event Frequency
   SST = Short Sample Time

3. **ST Safety Risk Index = [Log of (Designed Safety Risk/Actual Safety Risk) / Log of Designed Safety Risk)]*100**

4.5.4 Example for Short Term (ST) Safety Risk Index for ONE scenario

In our example, if SIF-1 input (PZT-4275) is bypassed for 8 hours in a period of 24 Hours, the ST Safety Risk Index calculation is per Table-2 :

Parameter	Source	Value
TMEL	Risk and TMEL definition	1E-05
Safety Severity of "2"	Risk Matrix (Figure 6)	Single fatality
IEF	BPCS failure , IEC 61511	0.1 per year
PFD of PSV-1	To fill the Risk Gap per Table 1	0.01
PFD of SIF-1	To fill the Risk Gap per Table 1	0.01
SST	Assumption	24 Hours
SIF-1 Input Bypassed	Assumption (from Historian data)	8 Hours
Designed Safety Risk	Per Equation 1: TMEL (for safety) (1E-05) x Safety Severity (1)	1E-05 fatalities/year
Actual ST Safety Risk	Per Equation 2: IEF(0.1) x [(PFD of non-SIF IPL(0.01) x SIF PFD(0.01)) x (Time SIF NOT in Bypass (16)/SST(24)) + (PFD of non-SIF IPL(0.01) x (Time SIF in Bypass(8)/SST(24))] x Safety Severity (1)	3.4E-04 fatalities/year
Designed Safety Risk/Actual ST Safety Risk)	(1E-05) / (3.4E-04)	0.0294
Log of (Designed ST Safety Risk)	Log (1E-05)	-5
Log of (Designed Safety Risk/Actual ST Safety Risk)	Log (0.0294)	-1.53148
ST Safety Risk Index (%)	**Per Equation 3, (-1.53148/-5)*100**	**30.629%**

Table-2

4.5.5 Equations for Short Term (ST) Safety Risk Index for MULTIPLE scenarios (example – One Plant)

1. **Designed ST Safety Risk (Multiple)** =
   ∑ (TMEL (for safety) x Safety Severity)

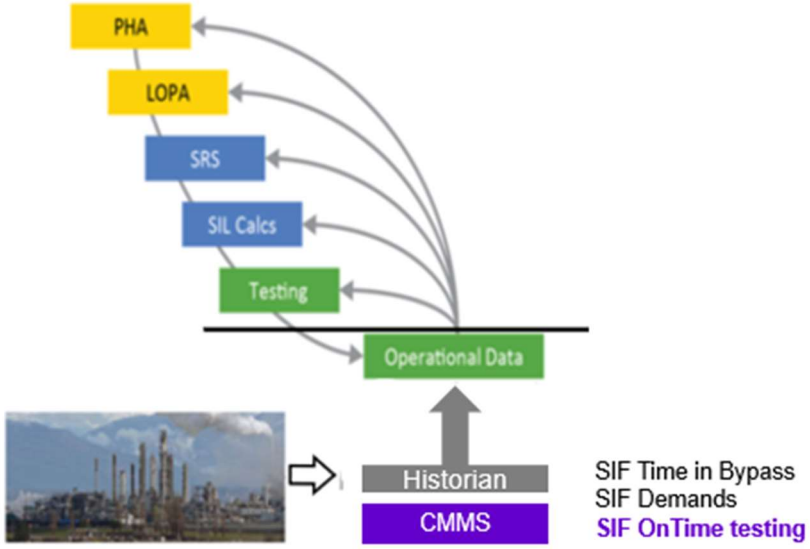
   (the assumption here is that with the designed IPLs, the TMEL has been met for all scenarios)

2. **Actual ST Safety Risk (Multiple)** =
   ∑ (IEF x [(PFD of non-SIF IPL x SIF PFD) x (Time SIF NOT in Bypass/SST) + (PFD of non-SIF IPL) x (Time SIF in Bypass/SST)] x Safety Severity)

   where :
   IEF = Initiating Event Frequency
   SST = Short Sample Time

3. **ST Safety Risk Index (Multiple) =** [Log of (Designed Safety Risk (Multiple)/Actual Safety Risk(Multiple)) / Log of Designed Safety Risk(Multiple))]*100

4. **Worst actor of ST Safety Risk Index =** Highest ST Safety Risk Index (ONE scenario)



For Short Term (ST) Safety Risk Index – Only "SIF Time in Bypass" used

For Long Term (LT) Safety Risk Index – All three parameters are used

Figure-7

4.5.6 Assumptions for Long Term (LT) PRI equations

1. For One scenario which has an "Unacceptable" Risk criteria without any safeguards
2. "Safety" is the driver for this hazardous event (not Commercial and Environment). So from now on we will refer to it as Safety Risk Index.
3. PFDactual of SIF and non-SIF IPL may not be the same as PFD per design
4. The SIF IPL input has 1oo1 input voting

4.5.7 Variables which effect Long Term (LT) Safety Risk Index

1. SIF demand rate. If this is greater than the assumed IEF, then SIF demand rate will be considered in the "Actual LT Safety Risk" equation
2. SIF "Time in Bypass" over the Long Term period
3. IPLs On time testing. If this is different than what was considered during design, then this will effect the PFDactual of the IPLs.
4. The above data is available in the Plant Historian and Computer Maintenance Management System , CMMS. (Figure 7)

4.5.8 Equations for Long Term (LT) Safety Risk Index for ONE scenario

1. **Designed Long Term Safety Risk** = TMEL (for safety) x Safety Severity
   (the assumption here is that with the designed safeguards, the TMEL has been met)

2. **Actual LT Safety Risk** = SIF demands x [(PFDactual of non-SIF IPL x SIF PFDactual) x (Time SIF NOT in Bypass/LST) + (PFD of non-SIF IPL) x (Time SIF in Bypass/LST)] x Safety Severity

   where :
   SIF demands considered as Initiating Event Frequency if SIF demands > IEF
   LST = Large Sample Time
   PFDactual (for SIF and IPL) varies based on "Real test intervals" vs "Design Test intervals"

3. **LT Safety Risk Index = [Log of (Designed Safety Risk/Actual Safety Risk) / Log of Designed Safety Risk)]*100**

4.5.9 Example for Long Term (LT) Safety Risk Index for ONE scenario

In our example, if SIF-1 input (PZT-4275) is bypassed for say a Total of 2 months in a period of One year, the LT Safety Risk Index calculation is per Table-3 :

Parameter	Source	Value
TMEL	Risk and TMEL definition	1E-05
Safety Severity of "2"	Risk Matrix (Figure 6)	Single fatality
SIF demands per year	Assumption (from Historian data), 1 per year > 1 in 10 years (IEF)	1 per year
PFD of PSV-1	Designed to fill the Risk Gap per Table 1	0.01
PFD of SIF-1	Designed to fill the Risk Gap per Table 1	0.01
PFDactual of PSV-1	Based on On-Time testing data	0.01
PFDactual of SIF-1	Based on On-Time testing data	0.01
LST	Assumption	1 year (12 months)
SIF-1 Input Bypassed	Assumption (from Historian data)	2 months (Total)
Designed Safety Risk	Per Equation 1: TMEL (for safety) (1E-05) x Safety Severity (1)	1E-05 fatalities/year
Actual LT Safety Risk	Per Equation 2: SIF demands(1) x [(PFDactual of non-SIF IPL(0.01) x SIF PFDactual(0.01)) x (Time SIF NOT in Bypass (10/12)/LST(1)) + (PFD of non-SIF IPL(0.01)) x (Time SIF in Bypass (2/12)/LST(1))] x Safety Severity (1)	2.65E-04 fatalities/year
Designed Safety Risk/ LT Actual Safety Risk	(1E-05) / (2.65E-04)	0.0377
Log of (LT Designed Safety Risk)	Log (1E-05)	-5
Log of (Designed Safety Risk/LT Actual Safety Risk)	Log (0.0377)	-1.423
LT Safety Risk Index (%)	**Per Equation 3, (-1.423/-5)*100**	**28.464%**

Table-3

4.5.10 Equations for Long Term (ST) Safety Risk Index for MULTIPLE scenarios (example – One Plant)

1. **Designed LT Safety Risk (Multiple)** =
   ∑ (TMEL (for safety) x Safety Severity)

   (the assumption here is that with the designed IPLs, the TMEL has been met for all scenarios)

2. **Actual LT Safety Ris**k **(Multiple)** =
   ∑ (SIF demands x [(PFDactual of non-SIF IPL x SIF PFDactual) x (Time SIF NOT in Bypass/LST) + (PFDactual of non-SIF IPL) x (Time SIF in Bypass/LST)] x Safety Severity)

   Where:
   SIF demands considered as Initiating Event Frequency if SIF demands > IEF
   LST = Large Sample Time
   PFDactual (for SIF and IPL) varies based on "Real test intervals" vs "Design Test intervals"

3. **LT Safety Risk Index (Multiple) =** [Log of (Designed Safety Risk (Multiple)/Actual Safety Risk(Multiple)) / Log of Designed Safety Risk(Multiple))]*100

4. **Worst actor of LT Safety Risk Index =** Highest LT Safety Risk Index (ONE scenario)

5. PROCESS SAFETY RISK INDEX

Depending on who the KPI is for, the Process Safety Risk Index is understood and used in a different manner.

5.1 AT THE CORPORATE LEVEL AND PLANT LEVEL

1. **Process Plant Safety Risk Index (Long Term)** = LT Safety Risk Index (Multiple)
2. **Worst actor for Process Plant Safety Risk Index (Long Term)** = Scenario with Highest LT Safety Risk Index

This will give Senior management at the corporate an insight on how the plant has been running based on the Long Term safety track record

The Long Term Safety Risk Index will help the Plant / Operations Manager to reanalyze risk and take appropriate action based on some of the worst actors which are driving the Safety Risk index up.

5.2 AT THE PLANT LEVEL ONLY

1. **Process Plant Safety Risk Index (Short Term)** = ST Safety Risk Index (Multiple)
2. **Worst actor for Process Plant Safety Risk Index (Short Term)** = Scenario with Highest ST Safety Risk Index

The Short Term Safety Risk Index will help the Plant / Operations Manager to decide on maintenance priorities on a shift or day basis

6. CURRENT LIMITATIONS AND PATH FORWARD

6.1 THE CURRENT PROCESS RISK INDEX CALCULATOR CONSIDERS

1. Considers only Safety Risk Index as it assumes Safety as the Risk driver
2. For a SIF only 1oo1 input voting has been considered
3. Only SIF status (bypassed, Ontime test, Demands)
4. All SIF demands are considered as “Real” demands and not spurious

6.2 PATH FORWARD

1. Process Risk Index will be considered for scenarios which are driven by Commercial or Environmental during the HAZOP stage
2. Non-SIF IPL status will also be included (assuming their status is digitally available)
3. SIFs with MooN input voting, where M>=N and N>1, will also be considered
4. Real failure rates of SIF instruments based on application and collected data

7. CONCLUSION

Process Risk Index, both Short and Long term can provide valuable information to both the Corporate and Plant Managers and Engineers. Based on these Risk Indices, Corporate and plant teams can monitor and improve the Process Safety solutions currently being used in a plant. This reduces the probability of process incidents and increases plant reliability

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MARY KAY O'CONNOR PROCESS SAFETY CENTER
TEXAS A&M ENGINEERING EXPERIMENT STATION

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A Brief Review of Intrusion Detection System in Process Plants and Advancement of Machine Learning in Process Security.

Sinijoy P J[*],Dr.M.Bhasi[**],Dr.V.R.Renjith[***]

[*]*Research Scholar, School of Engineering, Cochin University of Science and Technology, Kerala, India.*

[**] *Professor, School of Management Studies, Cochin University of Science and Technology, Kerala, India*

[***]*Professor, School of Engineering, Cochin University of Science and Technology, Kerala, India*

sinijoy@cusat.ac.in

Abstract

Industrial Cyber Security threats are continuously evolving in complexity and it is a fact that Cyber security is at the top of global risks for decades. Most of the firms are not fully prepared for system intrusion and attacks. Intrusion Detection System (IDS) is to protect critical infrastructure from attackers. A robust IDS can protect process industries from Cyber-attacks. Evasion techniques by attackers make detection a difficult one. The complex interconnected systems demand robust Cyber security techniques. The learning-from-experience strategy using case-based reasoning methodologies and utilization of machine learning are investigated. Detection methods like Anomaly based IDS(AIDS),Signature based IDS(SIDS),Host Based IDS and Network based IDS are discussed.

Paper discusses Cyber Physical Systems and reviews various recent works on IDS comprehensively. Paper also propose to examine evasion techniques used by the attackers along with the advantages and disadvantages of existing systems. Conventional Intrusion Detection System drawbacks are overthrown with the advent of Real Time and Artificial Intelligence based Intelligent Monitoring Systems.

It's vital to develop effective real-time attack monitoring and threat mitigation mechanisms. This paper also propose future research idea on intrusion detection technology with extensive application of Machine Learning.

Keywords: CPS,IDS,

1.Introduction

Dependency on technology and associated cyber security threats are increasing in multitude. A chemical industry is an industrial process plant that processes large scale chemicals. Attackers target process based industries like chemical industry since the consequences are very high in these sectors.

First part of the paper describes Cyber Physical Systems [CPS] which is the process of combining hardware and software components. CPS is a concept that focus on bridging Cyber and Physical world. The biggest threat to CPS is from the targeted controller and various processes controlled by it. An intruder hacking into such systems and changing the value of any of the critical parameters like temperature, pressure in the operational unit can cause heavy damage.

Here in Section 2,it describes about cyber physical system security and intrusion detection systems and Section 3 introduces related work in CPS. Conclusions of the study and references are presented in last Sections.

2.Cyber Physical System Security and Intrusion Detection Systems

2.1 Cyber Physical System Risk Equation

Risk = Threat x Vulnerability x Consequence

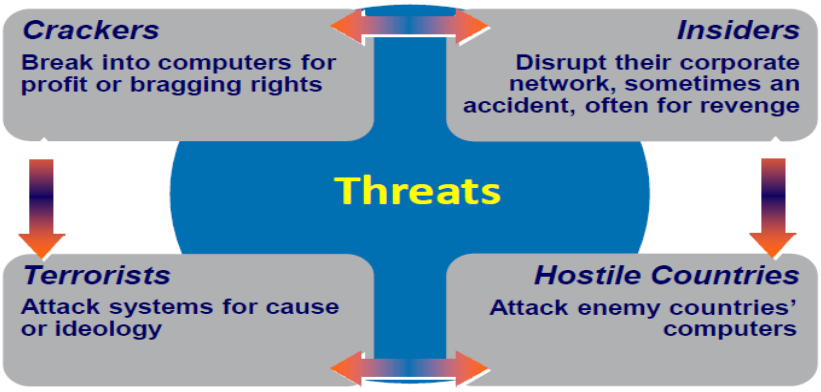

Figure 1:Four types of Threats.

2.2 Cyber Physical System Attack Types

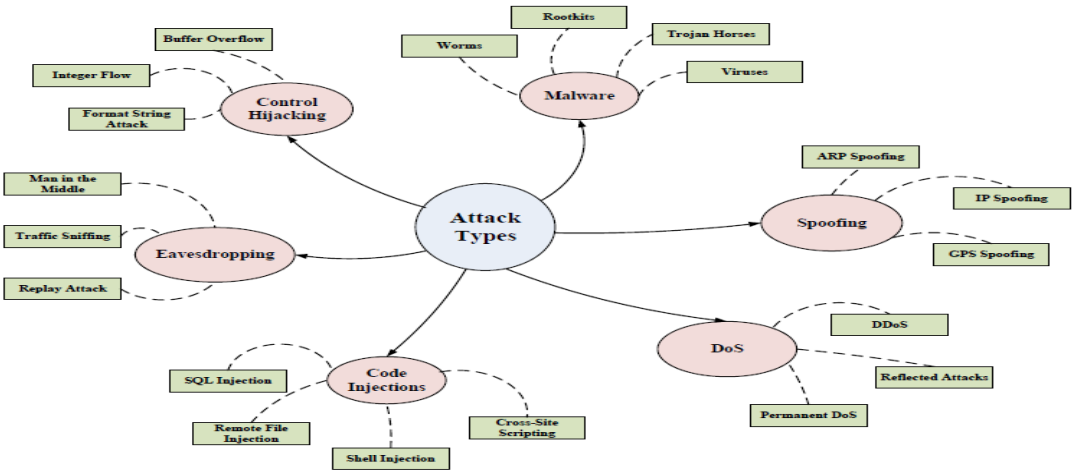

Figure 2:Cyber Physical system Attack types

The above Fig 2 describes common attacks and their sub types in CPS.

A) The intruder take the complete control of the system by control hijacking attack .B)Malware attacks affect the normal functioning of the system, C)Code injection exploits the vulnerabilities of the system by systematic injection of code that changes the complete execution of the program. D) Denial of Service attacks disables the normal services provided by the system. i) Permanent DoS is a type of attack when intruder tries to exploit unpatched vulnerabilities in order to install modified firmware to damage a system. ii) Distributed Dos attack is based on a model where several systems send request to the targeted system and occupies the resources thus making the targeted system unable to serve the purposes. E) Man in the Middle is an active type of attack and occurs when intruder intervenes between communicating entities trying to intercept the packets. F) Spoofing i)IP Spoofing aimed at using another IP Address to pass the security system .ii)GPS Spoofing is based on broadcasting incorrect signal of higher strength than received from satellite in order to deceive the victim.

2.3 Two major types of Threat Mitigation Schemes.

A. Intrusion Detection System [IDS]

B. Rule Based and Machine Learning Algorithm based Framework for Threat Detection.

2.3.1 Intrusion Detection System [IDS]

An Intrusion Detection System (IDS) is a device or software application that monitors a network or Process systems for malicious activity or policy violations. IDS can be classified by where detection takes place [network or host] or the detection method that is employed (signature or anomaly-based).

In terms of the way of detecting the threat, modern IDS can be subsequently divided into three subgroups:

- Anomaly-based assumes detection of the behavioural patterns which are different from the patterns of normal system‘s functioning;
- Signature-based requires a storage with a set of threats models being kept up-to-date, used to identify threats;
- Specification-based, in this mode specifications of the system as whole, as well as of components and interfaces are utilized for detection of suspicious activities.

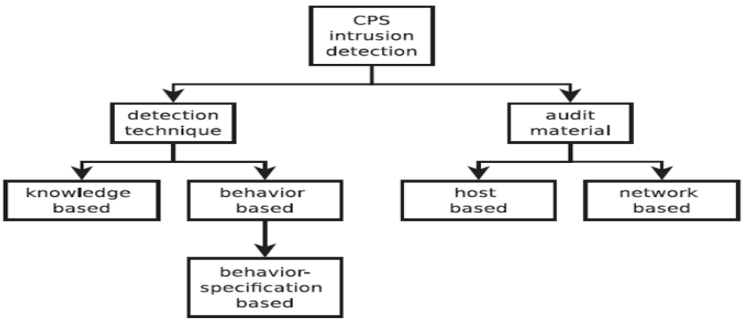

Figure 3

The Fig 3 shows CPS Intrusion Detection classification based on two major things

1. Detection technique: this criteria defines what misbehaviours of a physical component of the IDS considers to detect intrusion
2. Audit Material: This criteria defines how the IDS collects data for data analysis

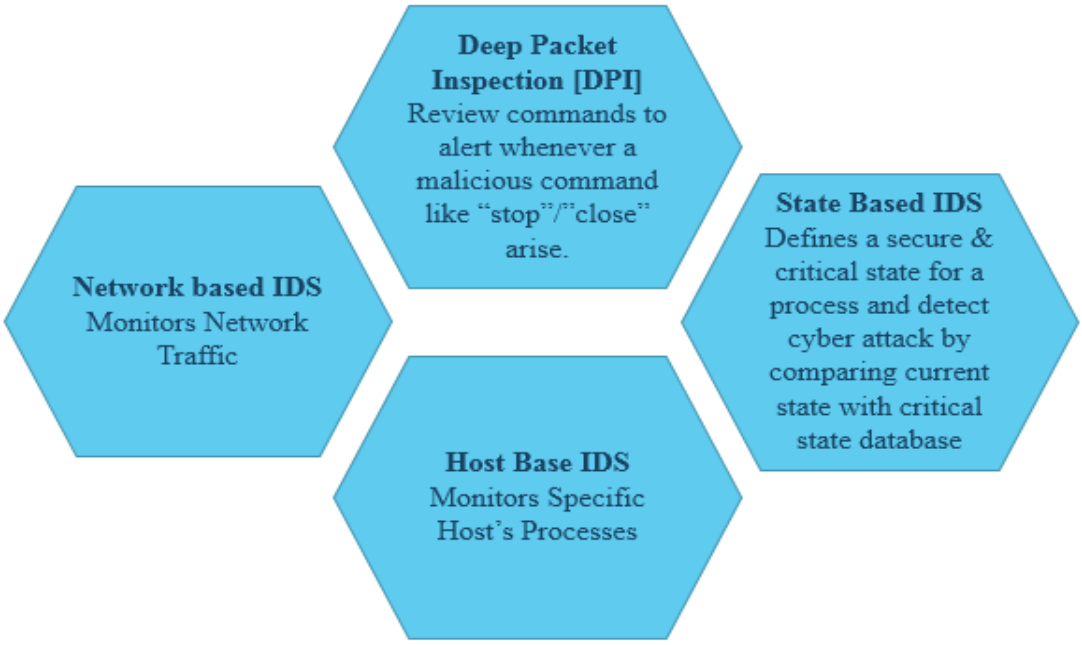

Figure 4:Types of IDS.

2.3.2 Rule Based and Machine Learning Algorithm based Framework for Threat Detection.

Threat detection based on rules from previous knowledge and new rules are created based on Machine Learning Algorithm. The raw input (packets, log files)are captured and forwarded to the feature extraction, where the features relevant to the threat detection are defined and based on this

the features are extracted, in the third step rule based mechanisms are applied to the key features and potential dangerous entities are discarded and the packets that are not identified malicious are checked by ML module(Rule Inference module)

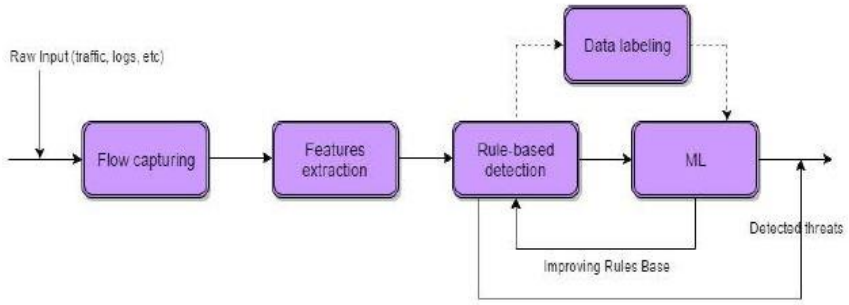

Figure 5: Rule Inference Module.

3. Review of Literature.

3.1 Multi-Layer Data Driven Cyber Attack Detection System for Industrial Control Systems Based on Network, System and process data [1]

Increasing number of attacks on cyber physical systems (CPSs) raises the alarm for Cyber security of industrial control systems (ICSs). Existing efforts of ICS cyber security are mainly based on firewalls, data diodes and other methods of intrusion prevention, which are not sufficient for growing cyber threats from motivated attackers. To enhance the cyber security of ICS, a cyber-attack detection system built on the concept of defence-in-depth is developed utilizing network traffic data, host system data, and measured process parameters. This attack detection system provides multiple-layer defence in order to gain the defenders precious time before unrecoverable consequences occur in the physical system.

A real-time ICS test bed data is used for demonstrating the proposed detection system. Five attacks, including man in the middle (MITM), denial of service (DoS), data exfiltration, data tampering, and false data injection, are carried out to simulate the consequences of cyber-attack and generate data for building data-driven detection models. Four classical classification models based on network data and host system data are studied, including k-nearest neighbour (KNN), decision tree, bootstrap aggregating (Bagging), and random forest, to provide a secondary line of defence of cyber-attack detection in the event that the intrusion prevention layer fails. Intrusion detection results suggest that KNN, Bagging, and random forest have low missed alarm and false alarm rates for MITM and DoS attacks, providing accurate and reliable detection of these cyber-attacks. Cyber-attacks that may not be detectable by monitoring network and host system data, such as command tampering and false data injection attacks by an insider, are monitored for by traditional process monitoring protocols.

Cyber-attack detection system utilizing a defense-in depth concept improves overall cyber-security by combining signature-based and anomaly based analysis of network, host, and process data. Attack scenarios-reconnaissance, DoS attacks, and a data tampering attack, have been conducted to demonstrate the possibility of cyber-attacks and to generate data for studying IDS development.

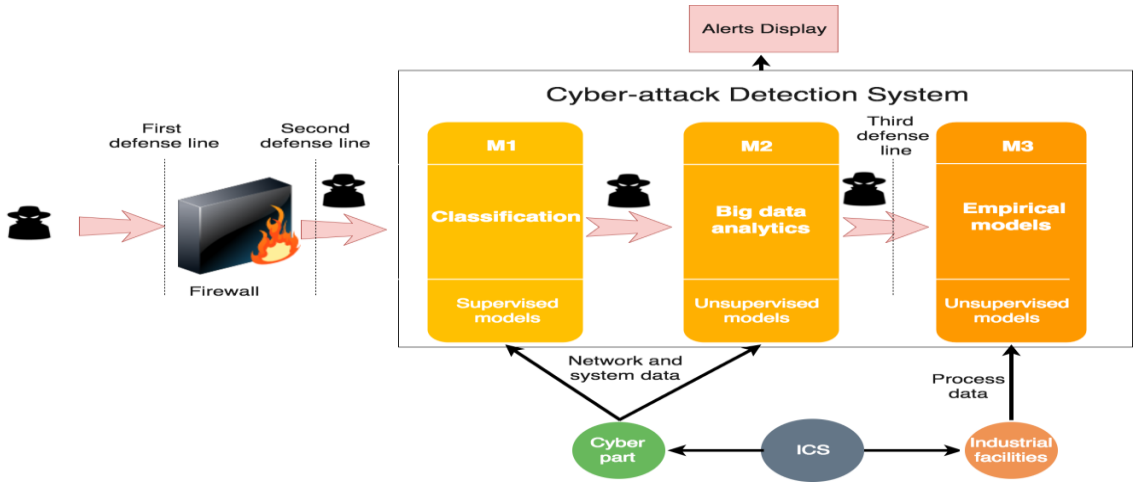

Figure 6: Cyber Attack Detection System

Figure 6 shows the structure of the proposed cyber-attack detection system with a defence-in-depth concept. The first defence layer is the traditional intrusion prevention layer, including firewalls, data diodes, and gateways, which are already widely applied in the industry. However, there are Situations that the attackers could bypass this defence line. The second defence layer consists of data-driven models for cyber-attack detection based on network traffic and system data, including the classification model indicated by M1 and big data analytics models indicated by M2. The classification models are based on supervised learning techniques, which can only detect attacks with behaviours similar to known attacks. Unsupervised big data analytics-based models will provide additional flexibility for intrusion detection; this is an area of ongoing research. M1 and M2 provide early detection of attackers when the attacks cause behaviour deviation from normal Operation. If the secondary layer fails to detect malicious activities, the last defence line monitors process data and uses empirical models indicated by M3 to detect abnormal operation, potentially due to cyber-attack. Model with residual thresholding detection was implemented in the M3 defence layer. The detection results of M1 and M3 using data generated from the physical test bed show that the proposed cyber-attack detection system has a high detection accuracy and a wide attack coverage. This multi-layer detection system improves the robustness of overall intrusion detection and is sensitive to both known and zero-day exploits.

A multi-layer, data-driven cyber-attack system was developed to enhance ICS cyber security by providing wider attack detection coverage by applying the defence-in-depth concept. In order to detect unknown attacks using network and host system data, the unsupervised big data analytics models in M2 will be studied to further enhance the second defence line.

3.2 Deep Learning based Efficient Anomaly Detection for Securing Process Control Systems against Injection Attacks[2]

Modern Industrial Control Systems (ICS) represent a wide variety of networked infrastructure connected to physical world. Depending on the application, these control systems are termed as Process Control Systems (PCS),Supervisory Control and Data Acquisition (SCADA) systems, Distributed Control Systems (DCS) or Cyber Physical Systems (CPS). ICS are designed for reliability; but security especially against cyber threats, is also a critical need. In particular, an

intruder can inject false data to disrupt the system operation.

Anomaly-based detection approaches are used to detect attacks that features the injection of spurious measurement data and proven to be efficient. In this paper, injection attack detection system that uses deep learning algorithms such as stacked auto encoders and deep belief networks that are tailored to identify different types of injection attacks are explained.

A model plant is used to obtain different data such as sensor and actuator measurements and specific attacks were injected into the data. The injected attacks vary in behaviour for training and testing of the proposed schema.

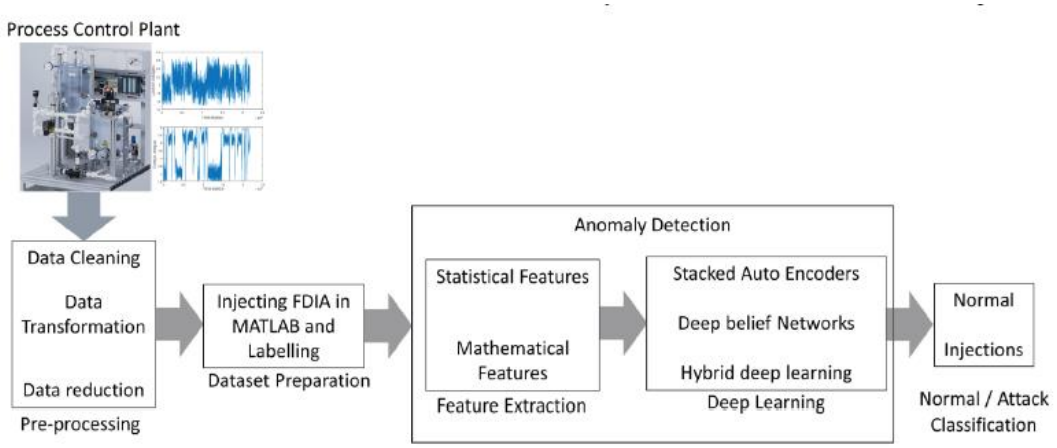

Figure 7 Deep Learning Based Anomaly Detection Architecture.

Main purpose is to detect abnormal behaviours caused by FDIA in plant data. As shown in Fig7 from the process control plant, different analog signals are acquired. The signals include sensor and actuator signals. The signals from the plant are initially pre-processed. Pre-processing of data is a crucial step for various applications. The effort on pre-processing depends on the characteristics of acquired data. It involves transforming raw data into an understandable format. The data obtained from real-world is often incomplete, inconsistent, lack certain common behaviour or trend and is likely to contain errors and corrupted values. Pre-processing techniques such as data cleaning, data transformation and data reduction are proven techniques to resolve the above-mentioned issues.

Data cleaning detects and corrects corrupt or inaccurate records in the collected data. Data transformation includes tasks such as smoothing, normalization, aggregation and generalization of acquired data. Normalization is the key task in data transformation which scales the data to a specified range. Min-Max normalization and z-score normalization techniques are most commonly used normalization techniques. Data reduction is usually done when acquired data is too big to handle or work with. Feature extraction is a key step in deep learning applications such as pattern recognition and image processing. The derived features out of raw data intends to be more informative and non-redundant facilitating the subsequent learning and generalization steps and, in some cases, leading to better human interpretations. Sometimes feature extraction is also considered as a data reduction mechanism discussed in pre-processing.

Statistical features are those which are defined and calculated through statistical analysis. Statistical analysis theory is the frequently used method of data feature extraction in the time domain. Mathematical methods are applied on the raw or pre-processed data to obtain the meaningful information. Mathematical features are also the most commonly extracted features on both time series and time independent transformations. Several mathematical functions from transform theory can be used to translate the signal into a different domain. List of mathematical features include derivate, probability and stochastic process, estimation theory, numerical methods etc. Deep learning is a machine learning technique combining both supervised and unsupervised techniques inspired from human brain. Some common deep learning architectures include Stacked Auto-Encoders (SAE), Deep Belief Networks (DBN), Convolution Neural networks [CNN] are used in image processing applications and requires huge dataset and training time.

In this paper, Two different deep learning techniques, SAE and DBN were used for complex feature extraction and later the classification was done with SVM and SMR. Different techniques have different detection accuracies for different injection attacks. In order to achieve the best detection accuracy hierarchical architecture with ranking approach can be used. The detection accuracies are dependent of type of dataset extracted features and the network architecture along with configuration parameters. Stacked Auto-Encoders (SAE), is the concept of stacking multiple auto encoders together. Deep Belief Networks (DBN) are formed by stacking Restricted Boltzmann Machines (RBM). RBM is a generative stochastic network which can learn a probability distribution over its set of inputs. An expertise in these parameters is necessary to identify which configuration suits well for the individual application.

3.3 Online Monitoring of a Cyber Physical System against Control Aware Cyber Attacks[3]

There have been an increasing number of malware attacks on the industrial control systems like Stuxnet in 2010 , Maroochy Shire Sewage attack in 2000, water filtering plant of Pennsylvania in 2006 and Davis-Besse power plant in Oak Harbor, Ohio in 2003. Increasing vulnerabilities in the cyber physical system have made information security an immediate concern and need for detecting and controlling the spread of such malware. Information security methods like authentication and integrity are inadequate in securing these control systems. Attacks on control system can result in tremendous costs to an organization in rebuild and recovery activities.

Cyber-Physical System (CPS) is integrations of computation with the physical processes. Control systems automate the tasks once performed by the humans by sensing the environmental conditions, executing the programmed logic and then actuating physical equipment to perform a desire task. Control systems are made up of sensors along with computational and communication capabilities.

Data received by the actuator causes necessary action(s) on the physical system. Sensors measure the physical system states and transmit it to the distributed controllers. A control action is a reactive process and failure of any non-redundant sensor or actuator can cause irreparable damage to the system under control.

Statistical techniques like SPRT are useful in the malware detection .In a cyber-physical system like SCADA, data is collected in the form of bug reports and system status logs. These data can

provide the vital historic information for understanding system behaviour and its trends. However, these files are huge in size and difficult to inspect manually.

The paper focuses on using computational geometric techniques for understanding the controller profile to detect anomalous behaviour.

3.4 Detection of Cyber-Attacks with Zone Dividing and PCA[4]

In 2010, an epoch making malware, Stuxnet, was discovered. It was a virus targeting centrifuge controllers in the Iran nuclear fuel factory. After its discovery subspecies have been developed. Although Stuxnet had a specific target, indiscriminate attacks can be committed by them. When control systems are intruded, not only their dysfunction but also serious accidents such as explosion or spill of dangerous substances might occur. Industrial control systems (ICS) require highly reliable security and safety services with urgent priority.

In information networks security measures are frequently taken. Databases of anti-virus software are updated every day. Various security patches are sent from product developers almost every day. However, in control networks anti-virus software is not utilized or security patches are not applied. Because they increase computation load and change link libraries, they might make controllers stop or be in ill conditions.

Therefore, vulnerability of control networks is much less than one of information networks.

Even in information networks, successes of cyber-attacks are reported frequently. The relationship of cyber-attacks and security measures is a cat-and-mouse game. In order to assure the safety of ICS against cyber-attacks, the relationships between safety and cyber-security must be considered and the characteristics of the plant must be taken advantages to develop security measures. PCA can be applied to any kinds of plants if normal operation data are available. Many abnormality detection systems can be constructed for real industrial plants. It is still difficult to distinguish the causes of the abnormal situation as cyber-attack. The detection is very important especially because concealment is included cyber-attack procedures. The combination of zone division and automatic abnormal detection using PCA can be an effective security measure.

In this paper, a design method of control network configuration to improve security and safety is proposed. The network is divided into plural zones. If the security of each zone is set independently, the possibility of the intrusion of the whole area becomes low. How to divide the network and how to detect the abnormality are discussed. Examples of application of zone division and PCA were illustrated. It was shown that the system could detect the relationship changes caused by concealment.

4. Conclusion

The growing number of security incidents in ICS facilities is mainly due to a combination of technological and organizational weaknesses. In the past, ICS facilities were separated from public networks, used proprietary software architectures and communication protocols. Built on the “security by obscurity” paradigm, the systems were less vulnerable to attacks leveraging ICT. Although keeping a segment of communication proprietary, ICS vendors nowadays increasingly use IP-based communication protocols and commercial off-the-shelf software. Also, it is standard to deploy remote connection mechanisms to ease the management during off-duty hours, and

achieve nearly-unmanned operation. The stakeholders seldom enforce strong security policies. User credentials are often shared among users to ease day-to-day operations, seldom updated (and not always revoked), resulting in a lack of accountability. Due to these reasons, ICS facilities have become increasingly vulnerable to internal and external cyber-attacks. Although companies reluctantly disclose incidents, there are several published cases where safety and security of ICS were seriously endangered.

Many machining monitoring systems based on artificial intelligence (AI) process models may be used for optimising, predicting or controlling machining processes.AI has significances when compared to traditional mathematical modelling and statistical analysis.

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The influence of the velocity field on the stretch factor and on the characteristic length of wrinkling of turbulent premixed flames

Tássia L. S. Quaresma*, Tatiele D. Ferreira, and Sávio S. V. Vianna
University of Campinas
School of Chemical Engineering
Albert Einstein Avenue, 500, 13083-852 Campinas, São Paulo, Brazil
t261147@dac.unicamp.br

Abstract

We investigate the effect of the turbulent velocity field on the reaction rate of turbulent premixed flames within the laminar flamelet regime. In the Bray-Moss-Libby (BML) combustion modeling, two parameters account for the effects of turbulence on the flame, namely the stretch factor and the characteristic length of wrinkling. However, difficulties in modeling flame stretch suggests the stretch factor to be assumed constant and equals to unity, which may lead to an inaccurate representation of the flame behavior. Also, the length scale of wrinkling is often calculated as a function of the fluctuating velocity via empirical correlations that depend on adjustable constants. As a first investigation line, we propose an expression for calculating the stretch factor dynamically, based on the divergence of velocity. In a second line we propose a hybrid reaction rate model that incorporates the well-known fractal approach into the BML model, by means of the characteristic length of wrinkling. The initial quasi-laminar regime is considered in the early stages of flame propagation, and the transition from laminar to turbulent is based on the turbulent Reynolds number of the flow. The modeling is carried out within an in-house developed a 3D Navier-Stokes compressible solver for premixed methane-air and propane-air flames in three partially obstructed geometries.

Keywords: Stretch factor, flame wrinkling, turbulent premixed flames, velocity field, fractal approach, BML model

1 Introduction

Numerical modeling of explosions has a number of relevant applications in the fields of physics and engineering. Investigations on phenomena such as sparkle-ignition engines, vapor cloud explosions, and supernovae are all ultimately related to the comprehension of the turbulence-flame interaction. In process safety studies, determining the turbulent flame speed via sophisticated

computational fluid dynamics (CFD) analysis represent a useful technique for predicting damaging effects of accidental gas explosions with a higher degree of accuracy [1].

In turbulent premixed combustion modeling, the reaction rate term is of fundamental importance for predicting flame propagation speed. Within the flamelet regime, the formulation proposed by Bray, Moss and Libby (BML) for the mean turbulent reaction rate incorporates the effects of the turbulent velocity field in both the stretch factor and the characteristic length scale of wrinkling [2].

Flame stretch is said to be responsible for changing the flame surface area by two different mechanisms: straining and curvature. Whereas the effects of curvature are typically neglected for Lewis number equals to unity, stretch due to straining is related to the effects of the divergent local flow field [3]. However, difficulties in modeling such phenomena, often resort to the consideration of an unstretched flame surface, by assigning the stretch factor equals to unity. Recent direct numerical simulation (DNS) studies have suggested that this approximation may lead to an unrealistic representation of the flame behavior, and the stretch factor should not be considered a constant parameter [4].

The characteristic length scale of wrinkling, also referred to as the integral length scale of wrinkling, is said to be responsible for controlling the turbulent flame surface area [3]. The length scale of wrinkling has been classically modelled as a function of the integral length scale of turbulence and the fluctuations of velocity. This relation is commonly obtained from empirical correlations that introduce additional adjustable constants that require tuning.

In this context, we propose two separate investigation lines for the aforementioned limitations of the BML reaction rate model. In the first one, the stretch factor is calculated dynamically based on a normalization of the divergence of the velocity. The second is focused on coupling the well-known fractal model into the length scale of wrinkling eliminating the need for an empirical correlation to be used.

The initial propagation stages of the flame are considered to be quasi-laminar. Transition from laminar to turbulent regimes of propagation is calculated with a blending function that is based on threshold turbulent Reynolds number. The laminar flame surface area is obtained on a geometric basis of spherical flames, as well as the volume of the computational cell [1].

The study is carried out within an in-house developed Reynolds-Averaged Navier Stokes (RANS) code for simulation of turbulent reacting flows in complex geometries, called STOKES. Turbulence is closed by the Boussinesq hypothesis and the k-e model [5], [6]. The two investigation lines have been tested in three different partially obstructed geometries for premixed methane-air and propane-air flames. Results of flame position, flame speed and flame contour inside the combustion chambers are presented and compared with literature data.

2 Methodology

The BML formulation for the mean reaction burning rate in the flamelet regime is considered [2]

$$\overline{w}_c = \rho_R u_L^0 I_0 \frac{g\bar{c}(1-\bar{c})}{|\hat{\sigma}_y|\hat{L}_y}$$

where the two first terms are the reactants density and the unstretched laminar flamelet speed, respectively. The stretch factor is represented by I_o, c is the Reynolds-averaged reaction progress variable, g and $|\hat{\sigma}_y|$ are model constants, and $\hat{L}_y$ is the characteristic length of wrinkling, given as

$$\hat{L}_y = c_L l_L f\left(\frac{u}{u_L^0}\right)$$

where c_L is a model constant, l_L is the laminar flamelet thickness and the function f relates the velocity fluctuations u and the unstretched laminar flamelet speed u_L^0 via the empirical correlation [7]

$$f\left(\frac{u}{u_L^0}\right) = \left[\left(1 + \frac{c_{w1}}{u/u_L^0}\right)^{-1}\left(1 - exp\left[-\frac{1}{1 + c_{w2}u/u_L^0}\right]\right)\right]$$

where two additional constants that require tuning are introduced.

In the initial quasi-laminar flame propagation stage, the laminar flame surface area is calculated as

$$A_{flame} = c_{LAM}V^{\frac{2}{3}}$$

where V is the volume of the hexahedrical uniform computational cell and c_{LAM} is a constant. The blending function from laminar burning to turbulent burning if given by [1]

$$f_u = \max\left(0; 1 - \exp\left(-0.008(Re_{Turb} - Re_{Threshhold})\right)\right)$$

2.1 The dynamic stretch factor I_o

We propose an expression for the stretch factor that is no longer constant, but instead it is calculated based on the influence of the velocity field on the flame surface. We follow the reasoning line that the divergence of velocity contributes to flame stretching the same way the divergence affects the flow, by representing points of both outward and inward fluxes on the surface of the flame.

We call it a dynamic stretch factor, and it is given by a simple algebraic expression based on a normalization between the local divergence of velocity to its maximum value in the previous time step on the surface of the flame

$$I_o = \frac{|\nabla . v|}{max|\nabla . v|}$$

Therefore, in this first investigation pathway, the mean reaction rate is calculated by the following expression

$$\bar{w}_c = \rho_R u_L^0 \left(\frac{|\nabla . v|}{max|\nabla . v|}\right)\left(\frac{g\bar{c}(1-\bar{c})}{|\hat{\sigma}_y| c_L l_L f\left(\frac{u}{u_L}\right)^n}\right)$$

where the function *f* considers an empirical correlation [7]. With the exception of the c_L constant, the values assigned for all model constants are presented in Table 1.

Table 1: Model constants considered in the dynamic stretch factor proposed formulation.

I_o	c_{LAM}	$\mathrm{Re}_{\mathrm{threshold}}$	g	$\lvert\hat{\sigma}_y\rvert$	c_{w1}	c_{w2}
dynamic	4/9	500	1.5	0.5	1.5	4.0

In this formulation, the constants presented in Table 1 are not changed and the assigned values are typical values found in literature [1], [7], [8].

2.1 The hybrid BML model

Following the works of [9] and [10], we calculate the length scale of wrinkling by replacing the function *f* for the classical fractal concept, that describes the flame surface as a fractal [11].

$$f\left(\frac{u}{u_L}\right) = \left(\frac{L_{Turb}}{l_G}\right)^{Df-2}$$

The flame surface is wrinkled by rotating turbulent eddies of different length scales, ranging from the inner cut-off to the outer cut-off, powered by the fractal dimension Df. This study assumes the outer cut-off to be equal to the integral length scale of wrinkling L_{turb} and the inner cut-off to be equal to the Gibson length scale, which is considered to be the smallest scale having a turnover velocity sufficient to wrinkle the flame front [12]

$$L_{Turb} = \frac{k^{3/2}}{\epsilon} \quad l_G = \frac{u_L^{o3}}{\epsilon}$$

where *k* is the turbulent kinetic energy, *e* is the turbulent kinetic energy dissipation rate.

The proposed expression for the mean reaction rate in this approach is given by

$$\overline{w}_c = \rho_R u_L^o \left(\frac{g\tilde{c}(1-\tilde{c})}{\lvert\tilde{\sigma}_y\rvert c_L l_L \left(\frac{L_{Turb}}{l_G}\right)^{Df-2}}\right)$$

where the stretch factor is omitted for its consideration to be constant and equals to unity.

The constants applied in this approach can be seen in Table 2. The constant c_{LAM} was tuned to the value of 0.09 and the turbulent threshold Reynolds is taken as the Reynolds number for internal turbulent flows.

Table 2: Model constants applied in the proposed BML hybrid formulation.

I_o	c_{LAM}	$\mathrm{Re}_{\mathrm{threshold}}$	g	$\lvert\hat{\sigma}_y\rvert$	Df
1	0.09	2500	1.5	0.5	7/3

2.3 Case studies

In the present study, we investigated two approaches for the mean reaction rate in three different geometries that consist of combustion chambers partially obstructed by solids (Figure 1). Chambers (a) and (b) initially contains a stoichiometric mixture of propane and air, whereas chamber (c) is filled with methane and air at stoichiometric proportions prior to ignition.

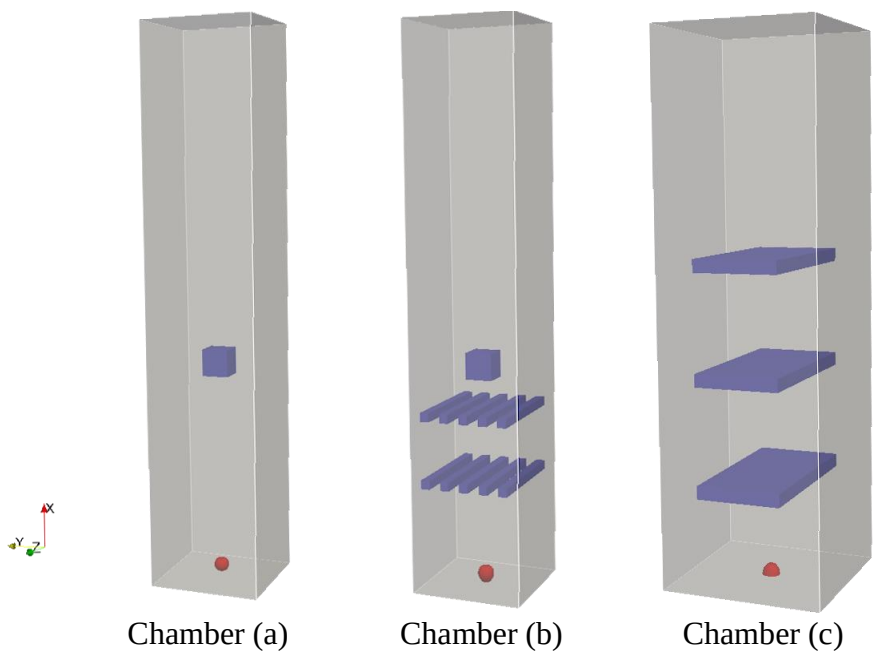


Figure 1: Combustion chambers partially obstructed by solid obstacles.

The internal dimensions of chambers (a) and (b) are 50 x 50 x 250 mm, whereas chamber (c) has 150 x 150 x 500 of internal dimensions. The cubic obstacles in chambers (a) and (b) has a cross section of 12 x 12 mm and are positioned at 100 mm away from the ignition point (in red). Chamber (b) contains two baffle stations located at 50 and 80 mm from the ignition point. Each baffle has 3 x 4 x 50 mm and are positioned 5 mm apart. Chamber (c) contains three rectangular obstacles (75 x 10 x 150 mm) at 100 mm apart from each other and the ignition point [13], [14], [15].

Assigned values for c_L can be checked on Table 3.

Table 3: Model constant c_L applied in the proposed BML hybrid formulation.

	c_L		
	BML	Dynamic I_o	BML hybrid
Combustion chamber (a)	2.0	2.0	3.5
Combustion chamber (b)	1.0	1.0	9.0
Combustion chamber (c)	2.5	2.5	10.0

3 Results

Simulation results applying the BML formulation with a constant stretch factor are presented in Figure 2. These results also consider the empirical correlation for calculating the length scale of wrinkling. They were used as a benchmark for assessing any other results obtained in the BML modification that are presented further.

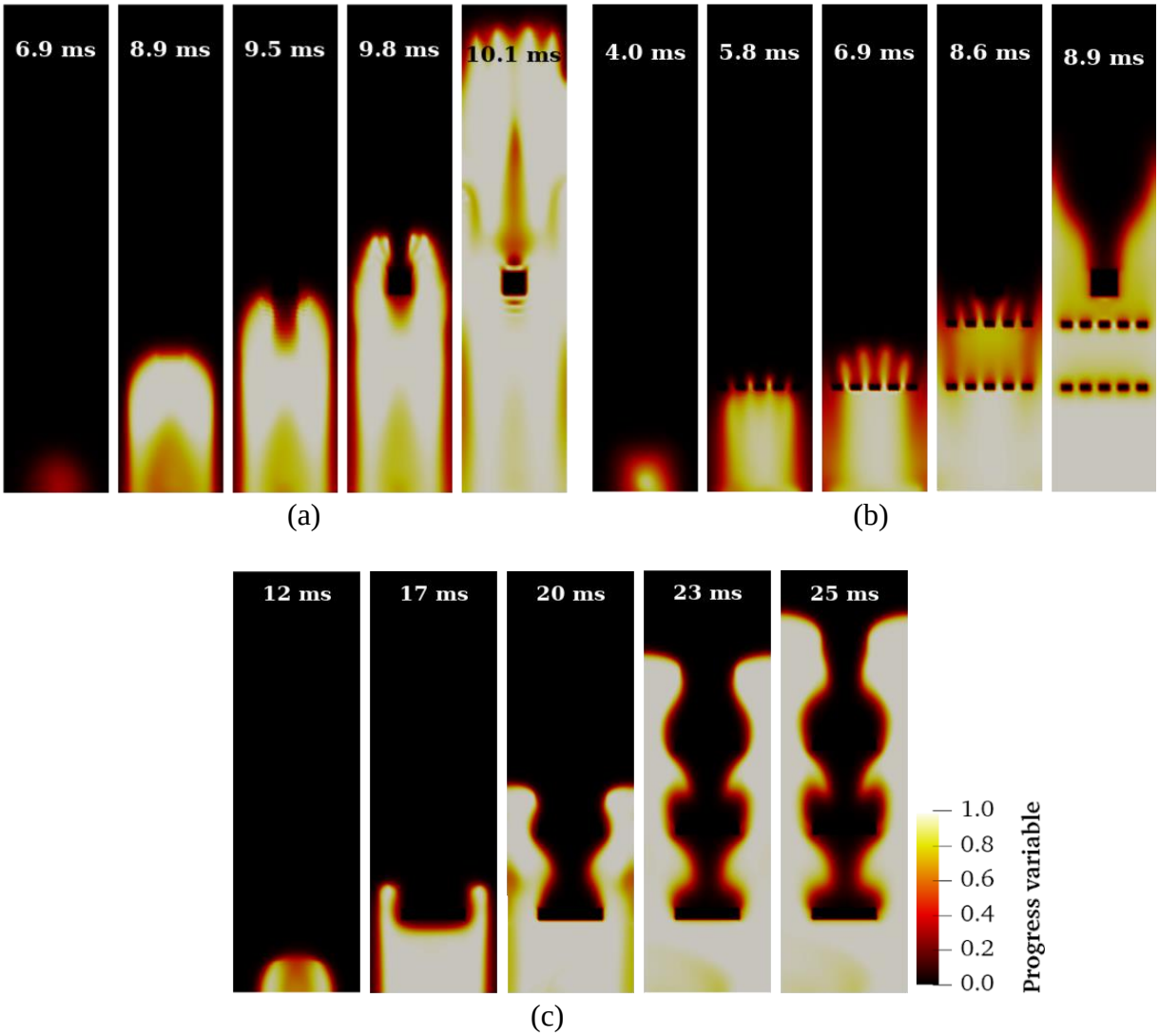


Figure 2: Flame position at different time steps with the classical BML formulation, applying constant I_o and an empirical correlation [1] for calculating the length scale of wrinkling. Flame position (a) in chamber Fig.1a; (b) in chamber Fig.1b; and (c) in chamber Fig.1c.

Flame position at different time steps into the chambers applying the two proposed approaches for BML modification are presented on Figure 3. Simulation results considering the proposed BML hybrid approach can be observed in Figures 3a, 3c, and 3e, whereas flame positions obtained by the proposed dynamic stretch factor are shown in Figures 3b, 3d, and 3f. It can be noted that in all three geometries, the BML hybrid simulations show a well-defined flame contour

and a clear separation between reactants (progress variable equals to 0) and products (progress variable equals to 0). On the other hand, in the dynamic stretch factor simulations results, conversion into products is not fully complete, especially in areas where there is more resistance to the flow Figures 3d and 3f. This behavior can be related to a higher divergence field that contributes to higher rates of flame stretching.

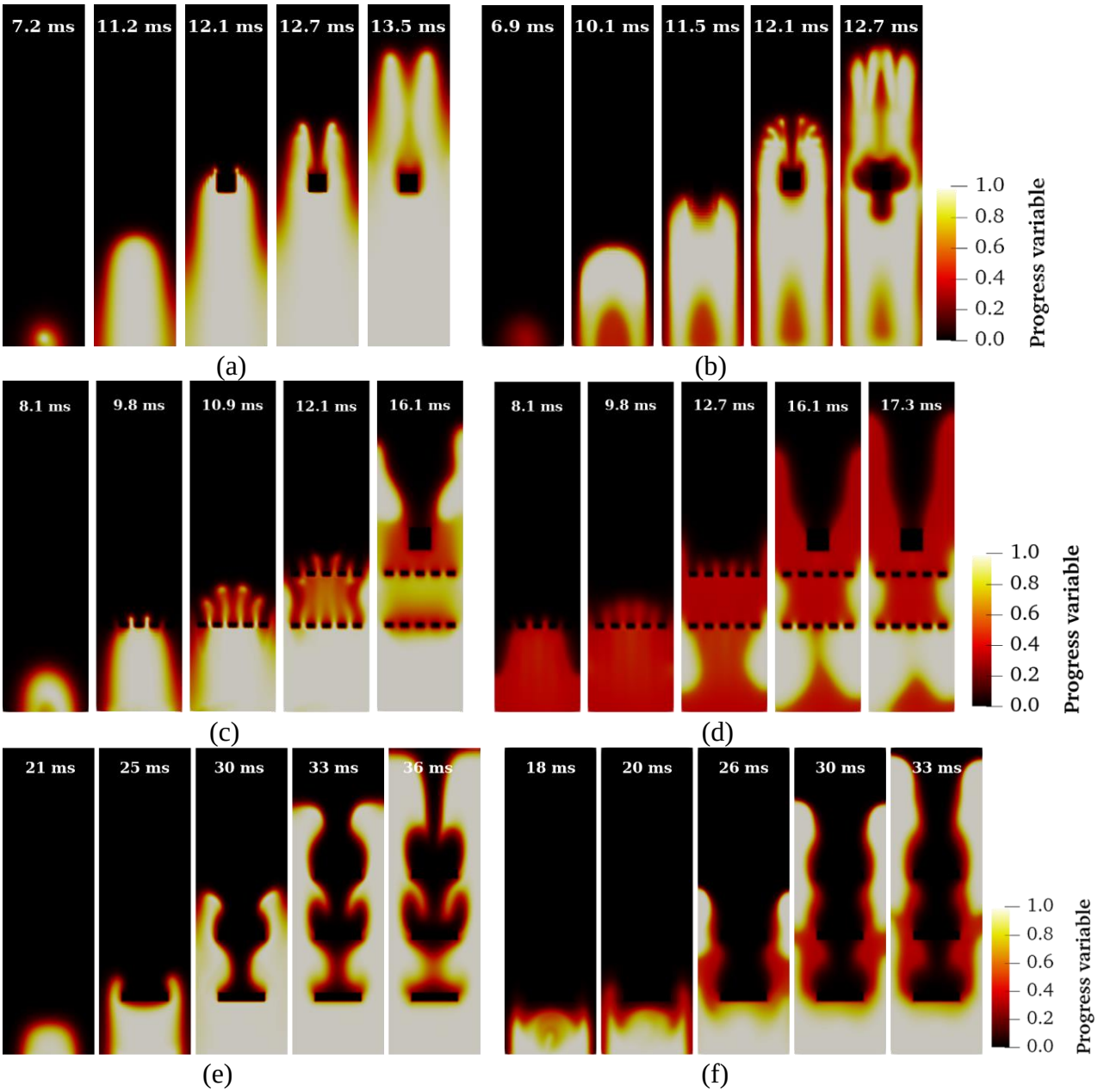


Figure 3: Flame position at different time steps in the partially obstructed combustion chambers. (a), (c), (e) BML hybrid simulations; (b), (d), (f) dynamic stretch factor (I_o) simulations.

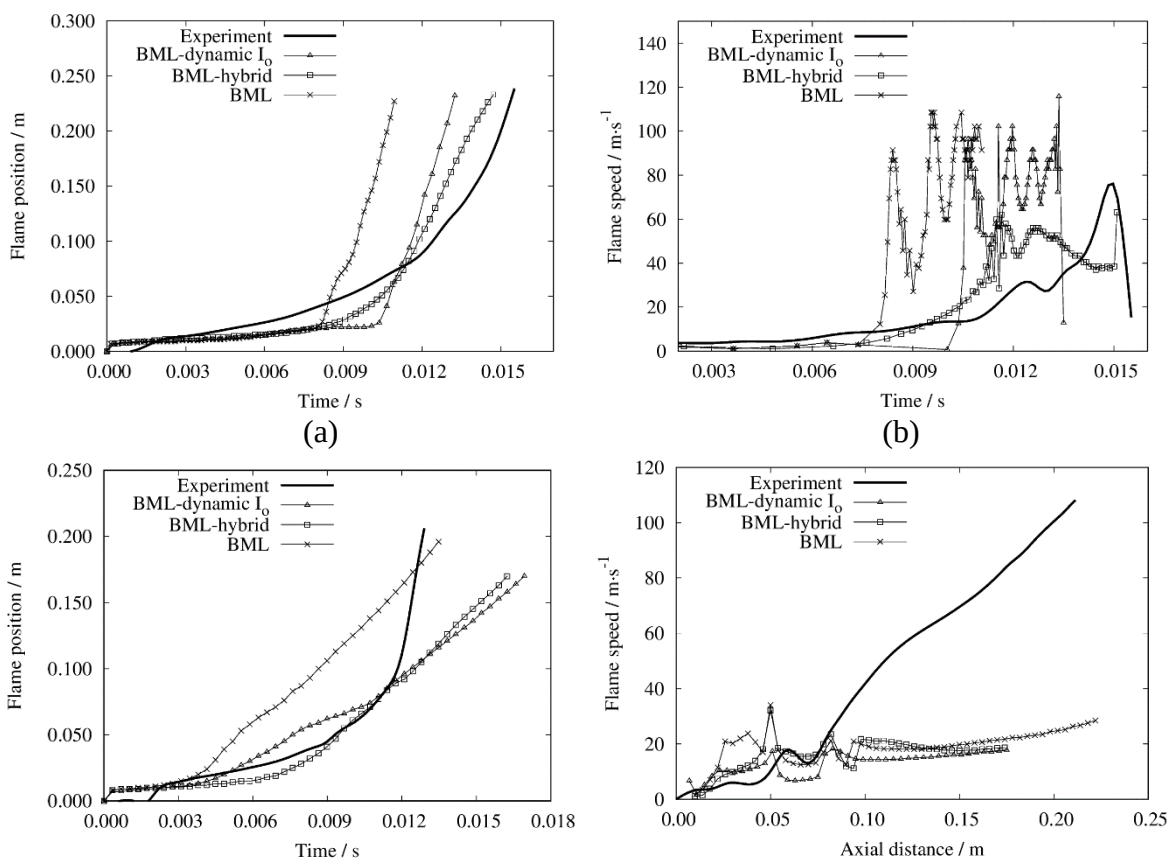
Comparison of flame position in chambers (a), (b) and (c) throughout time can be observed in Figures 4a, 4c and 4e, respectively. Flame speed in chamber (a) is plotted against time in Figure

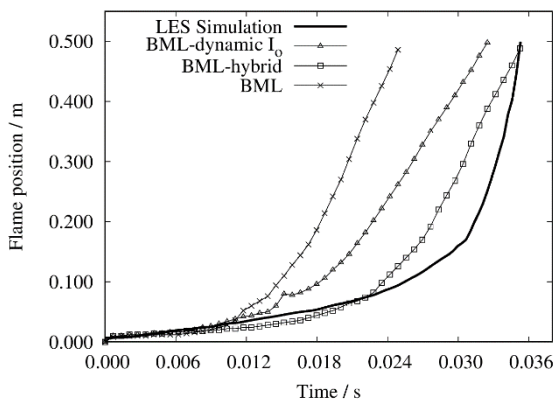
4b, whereas the flame speed in chambers (b) and (c) is plotted against the axial distance from the ignition point, and are shown in Figures 4d and 4f, respectively. The results are compared either with experimental data or LES simulation from literature [13], [14], [15].

In the graphs, curves identified by "BML" refer to the model considering a constant stretch factor and the function f is calculated by the aforementioned empirical correlation [7]. It can be observed that the insertion of the dynamic stretch factor for calculating the mean reaction rate, without changing any other model parameter, acted to decrease flame propagation via flame stretching, which contributed to a slight improvement in agreement with literature data. However, flame position and speed profiles were barely changed and some discrepancy from literature benchmarks still remain.

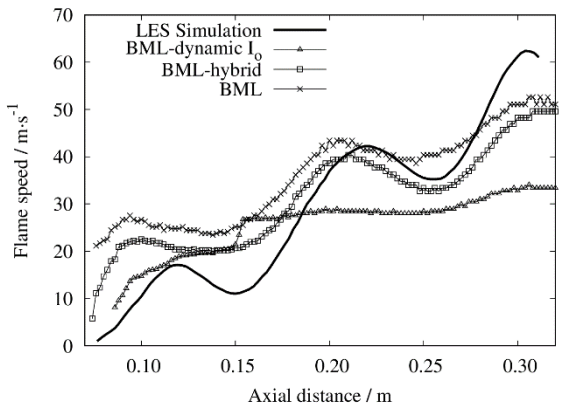
Such discrepancy is diminished with the BML hybrid approach. It can be noted a significant improvement in both flame position and flame speed profiles in Figures 4a, 4b (chamber a), and 4e, 4f (chamber c). Also, it is important to bear in mind that this approach considered the laminar-to-turbulent transition Reynolds as in internal turbulent flows.

None of the BML formulations were able to predict the final flame acceleration in chamber (b), as it can be seen in Figures (c) and (d). These results, when compared to the images in Figures 3c and 3d, may the related to a pronounced re-laminarization effect (reduction in speed and turbulence levels) decreasing flame speed [13].





(e)



(f)

Figure 4: Flame position at different time steps in the partially obstructed combustion chambers. (a), (c), (e) BML hybrid simulations; (b), (d), (f) dynamic stretch factor (I_o) simulations.

4 Conclusion and future work

We have introduced two alternative approaches for modification of the BML reaction rate model. The first one considers a dynamic stretch factor that is calculated taking into account the effect of the divergence of velocity to flame stretching. The second formulation calculates the characteristic length of wrinkling as a function of the fractal concept in which the surface of the flame is wrinkled by the length scales of turbulence, ranging from the Gibson length scale to the integral length scale. Within this approach, the laminar-to-turbulent transition turbulent Reynolds is taken as the transition Re for internal turbulent flows. This formulation showed a significant improvement in predicting flame position and flame speed in two out of three geometries tested. Future work will focus on combining the two proposed approaches for calculating the BML mean reaction rate as well as running simulations for large scale geometries.

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Towards a comprehensive model evaluation protocol for LNG hazard analyses

Filippo Gavelli[a], Bryant Hendrickson[a], Jide Kolade[b], Rich Kooy[b]
[a]: Blue Engineering and Consulting
[b]: Gas Technology Institute

fgavelli@blueeandc.com

Abstract

Every LNG project that is proposed to be built, expanded or significantly modified needs to meet the siting requirements of the applicable regulations. While the requirements and methodology can vary among different countries – for example, U.S. regulations follow a prescriptive approach, while the European standard requires a risk assessment to be performed – a common siting requirement is the safety of the public outside the project's fenceline in the event of accidents within the LNG plant.

In order to quantify the hazard footprints for potential accident scenarios (such as flammable vapor dispersion, pool and jet fires, vapor cloud explosions, etc.), computational tools must be used. Given the multiplicity of tools available, ranging from empirical models to 3D computational fluid dynamics packages, the agency reviewing the project application may not have the methodologies or protocols necessary to determine the suitability of a given model to a certain scenario, its accuracy and any other setup requirements. For this reason, a Model Evaluation Protocol (MEP) was developed in 2007 to allow computational tools for vapor dispersion modeling to be reviewed. This protocol was then successfully applied to two software packages (Phast and FLACS), which were found acceptable for LNG vapor dispersion modeling under US federal regulations.

The 2007 MEP, however, is very limited in scope: in fact, it only addresses vapor dispersion modeling and, more specifically, only from atmospheric releases (e.g., vapors from a liquid spill onto the ground). This means that there are currently no established protocols to evaluate models to simulate hazards such as the flammable or toxic dispersion of a vapor cloud from a pressurized release (e.g., a pipe breach), the overpressures generated by a vapor cloud explosion, etc.

Blue Engineering and Consulting and the Gas Technology Institute are collaborating on a DOT-PHMSA sponsored research project to develop a new set of Model Evaluation Protocols, that will allow the review of modeling tools for each of the above-referenced hazards. The new MEPs will greatly increase the confidence of authorities as well as the public, by defining which models may be used and under which limitations, and what validation factors need to be applied depending on the type of hazards being evaluated. This paper describes the framework of the new MEPs.

Keywords: LNG, model evaluation, hazard analysis, siting

1 Background

The safe siting of LNG facilities requires the quantification of the consequences to people and property from a loss of containment and release of hazardous materials (e.g., flammable and/or toxic materials). In order to quantify the hazard footprints for potential accidental release scenarios (such as flammable vapor dispersion, pool and jet fires, vapor cloud explosions, etc.), computational tools must be used.

Current U.S. federal regulations by the U.S. Department of Transportation, Pipeline and Hazardous Materials Safety Administration (PHMSA) contained in 49 CFR 193 include a list of models required to perform consequence modeling for LNG facility siting: these models include DEGADIS and FEM3A for flammable vapor dispersion distances and LNGFIRE3 for thermal radiation distances from each LNG container and LNG transfer system. Unfortunately, these models have significant limitations that restrict their applicability to only a fraction of flammable dispersion and pool fire scenarios typically involved in a siting study. Furthermore, these models are unable to model other types of hazards (e.g., overpressures) and hazard scenarios (e.g., flashing and jetting releases, jet fires) currently required by PHMSA's guidance (https://www.phmsa.dot.gov/pipeline/liquified-natural-gas/lng-plant-requirements-frequently-asked-questions/h1). Therefore, "new" models (that is, models not explicitly listed in 49 CFR 193) need to be used to perform these calculations.

Given the multiplicity of tools available for hazard modeling, ranging from empirical models to 3D computational fluid dynamics packages, concerns may arise regarding the suitability of a given model to a certain scenario, its accuracy and any other setup requirements; furthermore, the regulatory body reviewing the work may not have the expertise necessary to make such determination. These concerns led to the 2007 development of a model evaluation protocol (MEP) for LNG vapor dispersion [1]; two different dispersion models (Phast v.6.6-6.7 and FLACS v.9.1) were subsequently reviewed according to the 2007 MEP and approved by PHMSA in 2011 [2], [3] for use in vapor dispersion modeling.

However, the scope of the 2007 MEP is quite limited relative to the current regulatory requirements for LNG facility siting: in fact, it only applies to vapor dispersion hazards; additionally, the only data sets included in the validation database represent LNG spills onto water or low-momentum, ground-level gas releases. Therefore, in 2019 PHMSA sponsored a research project to develop a set of model evaluation protocols that would allow the evaluation of models for the calculation of the different types of hazards and hazard scenarios associated with the operation of LNG facilities.

This research project is conducted by Blue Engineering and Consulting Company (BLUE), in collaboration with the Gas Technology Institute (GTI).

2 Scope

The flow chart shown in Figure 2-1 describes the potential outcomes following a loss of containment from a pipe or vessel, and can be used to clarify the scope of the current project:

1. If the released material is flammable, one of three outcomes are possible:
    a. If the release is ignited in proximity of its source, a fire will occur. This can be a jet fire if the release is a pressurized jet (gas or flashing liquid), or a pool fire if the release results in the formation of a liquid pool on the ground. The fire will be "anchored" to the fuel supply (i.e., the source of the jet or the liquid pool, respectively) and will continue until the fuel supply is depleted. If the thermal radiation from the fire impinges onto a pressurized vessel, the possibility of escalation of the event to a Boiling Liquid Expanding Vapor Explosion (BLEVE) may also need to be considered.
    b. If the release is not ignited in proximity of its source, a gas/vapor cloud will form and disperse away from the release location, driven by the wind and the source momentum, and interacting with terrain and obstacles in its path. As the cloud disperses, it mixes with ambient air and is progressively diluted. If the cloud encounters a viable ignition source while still in the flammable range (i.e., the gas concentration is between the lower and upper flammable limit, LFL and UFL, respectively) then one of two scenarios may occur:
        i. if the ignition happens in an open and uncongested area, a flash fire occurs. A flash fire is characterized by the relatively slow-speed propagation of the flame front from the ignition location throughout the flammable cloud. Given the low flame speed, no significant overpressures are produced; however, it is typically assumed that any person caught within the flammable cloud at ignition will be a fatality from the flash fire. The flash fire may also burn back to the release location or the pool location, if still active, which may then start a jet or pool fire.
        ii. If the flammable cloud is ignited in a confined and/or congested area, a vapor cloud explosion (VCE) occurs. A VCE is characterized by significant (i.e., potentially damaging) overpressures, which may be due to the confinement of the combustion products (e.g., when a flammable cloud migrates inside a building and is ignited) or to the acceleration of the flame front as it interacts with multiple obstacles (e.g., when a flammable cloud is ignited within a process area).
    c. If the flammable cloud disperses below the LFL without finding a viable ignition source, no hazardous conditions are generated due to the flammable nature of the release.
2. If the released material is toxic, the dispersion of the toxic cloud creates hazardous conditions, which extend to the area covered by the cloud at concentrations equal to or greater than the toxic threshold of concern (typically the Acute Exposure Guideline Level 2).

3. If the released material is neither flammable nor toxic, other hazards are still possible, including:
    a. Cryogenic exposure due to contact with cryogenic liquid (e.g., spray from a flashing jet) or vapors from the release.
    b. Asphyxiation hazards, due to the displacement of oxygen by the gas/vapor cloud.

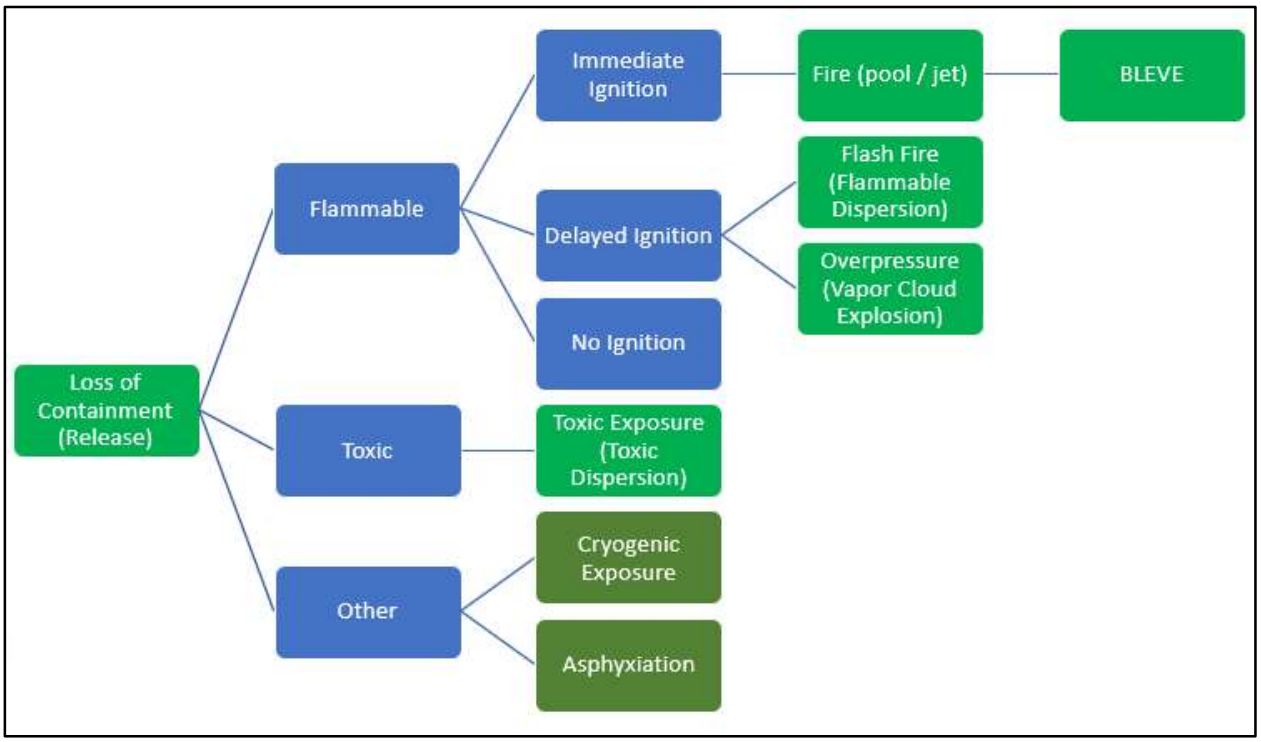

Figure 2-1. Event tree for the consequences of a loss of containment.

The flow chart shown above does not consider every possible hazard scenario – for example, the same release may create both flammable and toxic hazards. However, it highlights the main hazard categories that need to be included in an LNG facility siting analysis. The items shown in green identify steps of the analysis which require modeling. The scope of this project is to develop a set of MEPs so that models used to perform each of these types of calculations may be evaluated, with two exceptions:

- A guidance document on the evaluation of models for LNG fires was released by Sandia National Laboratories [4] and is considered by PHMSA as an MEP for pool and jet fires;
- Cryogenic and asphyxiation hazards are typically associated with flammable and/or toxic releases and modeling of these hazards is usually performed by the same models used to perform flammable or toxic dispersion calculations. Therefore, MEPs for cryogenic or asphyxiation hazard modeling are not required.

Note that the initiating event (loss of containment) does not represent a hazard per se; however, defining the source term for any scenario requires modeling appropriate to the nature of the release; therefore, a model evaluation protocol is necessary for that step as well. In summary, the scope of the current project is to develop MEPs to evaluate models for:

1. Source term.
2. Flammable dispersion (including non-LNG materials and flashing and jetting releases).
3. Toxic dispersion.
4. Overpressures from vapor cloud explosions.
5. Fireballs and overpressures from BLEVEs.

3 Structure of the MEP

The broad scope of this project allowed the development of a general model evaluation methodology, based on an extensive review of the published literature including previous model validation and evaluation efforts. The common methodology allows for an easier development of the individual MEPs and facilitates both the application and evaluation processes, particularly for models capable of calculating different hazard types. Additionally, a consistent methodology is expected to make the entire model evaluation process easier for stakeholders to understand. Of course, the methodology recognizes that certain tasks are inherently hazard-specific and can be adjusted as needed to account for these differences.

For each hazard type described above, the overall model evaluation protocol consists of the following:

1. Statement of scope
2. Evaluator qualifications
3. Model description questionnaire
4. Scientific assessment
5. User-oriented assessment
6. Model Verification
7. Model Validation, which includes:
    a. Model Validation Database (MVD), with:
        i. Experimental data necessary for modeling
        ii. Key variables and physical comparison parameters
        iii. Statistical performance measures
        iv. Acceptability criteria
    b. Sensitivity cases
    c. Uncertainty quantification
    d. Qualitative performance assessment
    e. Quantitative performance assessment

When a model is submitted for review and the model evaluation tasks are completed, a Model Evaluation Report (MER) is prepared, which includes:

1. Evaluation summary, including:
    a. Capabilities and limitations
    b. Validation factors
2. Best practice guidance
3. Recommendations for improvement

In the MEP methodology developed for this project, the MVD is developed as a separate document from the MEP. The reason for separating the two is that the MVD and all its subtasks are clearly hazard dependent, whereas most of the other tasks in the MEP have broader commonalities across hazards; therefore, this approach allows better consistency across MEPs. Additionally, this separation allows for updates of the database and, perhaps more importantly, of the statistical performance measures and acceptance criteria without requiring updates of the protocol itself.

A brief discussion of each subtask is provided below, as guidance for the development of hazard specific documents, which will be the purpose of the remainder of this project.

3.1 Scope of the evaluation

The MEP needs to clearly state the scope of the evaluation for which it is providing guidance. This includes:

- The hazard being evaluated: It is important for the specific hazard to be described, particularly because many modern models can calculate several different types of hazards; each hazard should be evaluated separately, and the model proponent may choose to seek approval only for some modeling capabilities.
- The types of models that may be evaluated: The MEPs developed during the current project are intended to be applicable to any type of model.
- The objectives of the evaluation: In general, a model evaluation can serve multiple purposes, including regulatory approval, performance ranking, model improvement, etc. The main objective of these MEPs is to provide (or deny) regulatory approval and to establish conditions (if any) for such approval.

3.2 Evaluator qualifications

Most of the existing protocols specify that the reviewer should be a third party, in order for the review to be objective and independent. The downside of this requirement is that a truly independent reviewer is unlikely to have the same knowledge and expertise as the model developer, and this may affect the model evaluation. Nonetheless, it is believed that allowing the model developer to perform the entire evaluation would inevitably raise objectivity concerns. Therefore, the following requirements apply to the individual (or group) performing a model evaluation under this protocol:

- The evaluator may not be associated with the model developer or with any of the model distributors. Prior association is acceptable, provided that there is no current or foreseeable collaboration that may raise concerns of objectivity;
- The evaluator must have recognized expertise with the physics of the hazard being considered;
- The evaluator must have recognized expertise with consequence modeling, and specifically with the same type of model (e.g., box, Gaussian, semi-empirical, CFD, etc.) as the one being evaluated. Direct experience with the model being evaluated is not required, however, would be beneficial.

A summary of the evaluator's credentials will be included in the model evaluation report for transparency.

3.3 Model description questionnaire

The first step in the evaluation of a specific model is to familiarize with the model itself. Therefore, the evaluation will begin with a Model Description, which should include information such as:

- Model name, version and release date
- Model type, application areas, theoretical background, solution methods
- Application areas for which approval is being sought
- Hardware requirements and operating system(s)
- User manual, technical references, publications

In order to facilitate the submittal and to ensure consistency of information across models, a questionnaire will be included with each MEP; the questionnaire follows the general structure of the one provided with the existing MEP for LNG dispersion [5], modified as needed based on the specific hazards being evaluated. The model description should be prepared by the model developer or a third-party with deep knowledge of the model.

3.4 Scientific assessment

The purpose of the scientific assessment is to ensure that if the model is able to correctly predict a given scenario, it does so for the "right" reasons. The scientific assessment therefore includes several tasks, such as [6], [7]:

- Review and assess the scientific basis of a model (that is, the governing equations being used to replicate the physical phenomena)
- Describe the model's capabilities and limitations, and any special features, relative to the physical phenomena for which approval is being sought
- Identify potential areas for improvement.

The information necessary to perform the scientific assessment should be obtained from the questionnaire and from other relevant documentation, including technical references provided by the model developer as well as peer-reviewed literature. The model evaluator will need to have in-depth understanding of both physical phenomena and modeling principles involved, in order to provide an independent assessment.

3.5 User-oriented assessment

The purpose of the user-oriented assessment is to evaluate the usability of the model. Therefore, it addresses the following issues:

- User-oriented documentation (e.g., Installation, User Guide), online help, training
- Pre- and post-processing interfaces
- Available model options (material databases, solver options, etc.)

- Simulation set-up (user-friendliness and guidance)
- Level of expertise required to run the model
- Warning and error messages
- Output data (availability and format)

The information necessary to perform the user-oriented assessment should be from the questionnaire and other relevant documentation. Direct experience with the model by the evaluator could be valuable, as it would allow to include 'user' feedback.

3.6 Verification

Verification is the process of ensuring that the implementation of a model is consistent with its theoretical basis. The purpose of verification is to demonstrate that the coding of equations, algorithms and databases in the model is correct. Potential procedures for code verification include:

- Checks for internal consistency (e.g., mass or energy balances)
- Modeling of simple scenarios for which analytical solutions are available
- Examine code behavior for limiting conditions
- Method of manufactured solutions
- Comparison with results from ensemble of other models

Verification can be complicated by confidentiality issues for proprietary models. As such, most of the existing model evaluation protocols assign the responsibility for model verification to model developers; the same approach is followed in this project: the model evaluator should rely upon evidence of verification provided by the model developer and review it to perform the assessment.

3.7 Validation

Validation is the process of comparing model predictions to experimental data for scenarios that test the physics that the model is intended to predict. As discussed in existing protocols [8], "validation" in the true sense of the word cannot be accomplished by simply comparing a model against a finite number of scenarios; what can be accomplished is an "evaluation" which establishes enough confidence in a model to expect that it will perform in an acceptable manner when applied to similar scenarios. However, for consistency with the majority of published literature, the comparison of a model with experimental data in these MEPs will be called "validation" and will be understood to implicitly refer to the data sets included in the validation database, and not to a theoretical, unachievable absolute validation.

3.7.1 Model Validation Database

The purpose of a model validation database is to provide a set of scenarios to be simulated with a computational model, in such a matter that qualitative and/or quantitative comparisons may be made between model predictions and actual observations. In general, the scenarios in the database will consist of experimental data sets; however, real-world scenarios (e.g., accidents) can also be

included provided that there is sufficient information to set up a simulation and evaluate the modeling results.

The following criteria for the selection of relevant data were provided by Karaca [7]:

- Test cases should represent as close to realistic scenarios as possible. The definition of "realistic" is difficult and depends on the type of hazard: in general, the scale of the experiment (strength of the source term, flammable cloud volume, dimensions of the test area, etc.) relative to expected "real life" scenarios should be considered in this assessment.
- Test description should be sufficiently detailed. The goal is for modelers to be able to set up initial and boundary conditions in their simulations as close as possible to the experiment.
- Sufficient meteorological data should be available and obtained from sensors in/near the area of interest. Location and height of sensors should be provided. Time resolution of wind and temperature data should be enough to estimate turbulence and stability parameters. Humidity and precipitation should be provided.
- Measurements should be adequate to reliably describe the hazard. Data needs to be on a sufficiently fine grid and sufficiently high time resolution. Data should be provided at different averaging times to allow comparison with different models.
- Measurements should be of a sufficient quantity to be statistically representative.
- Data processing applied to raw data must be documented.
- Uncertainty of all measured and derived quantities must be provided, together with a description of the method used to define such uncertainty.

Recent work by Skjold et al. [9] introduced an interesting approach for selecting data sets to be included in the MVD: each potential experimental data set is reviewed and scored according to several parameters (e.g., relevance to industry practice; experimental scale; repeatability; quality of measurements; availability of experimental data; etc.). An average score is then calculated for each data set. Only sets with an average score above a predetermined threshold are included in the validation database. This approach therefore provides a clear explanation for the selection of certain data sets over others, instead of "cherry picking" scenarios that may favor one model type over another; given the importance of transparency in a regulatory environment, this approach is followed in the development of model validation databases for the current project.

3.7.2 Key Variables and Physical comparison parameters

The key variables for model validation depend on the hazard being considered. For example, for flammable and toxic dispersion, gas concentration in air is certainly an important variable, however other variables such as temperature may also be used; for vapor cloud explosions, pressure is likely the variable of most interest, however, other variables such as temperature or gas velocity may provide useful information; and so on. Ultimately, the data available from the experimental data sets determines which measured or calculated variables can be used.

Once the key variables have been identified, the physical comparison parameters (PCPs) can be determined. Once again, the choice depends on the type of variable being considered as well as

the available data. For example, the physical comparison parameters previously defined for dense gas dispersion scenarios are as follows [10]:

1. Point-wise concentration
2. Maximum arc-wise concentration
3. Cloud width
4. Predicted distance to the measured maximum arc-wise concentration
5. Distance to the LFL concentration
6. Predicted concentration at the measured distance to the LFL

The selection of PCPs for flammable or toxic dispersion model validation can certainly start from the list above; however, the same PCPs would not be relevant for overpressures or other hazards.

3.7.3 Statistical performance measures and acceptability criteria

Model performance relative to experimental observation and evaluation database can be performed on a qualitative and/or quantitative basis. Qualitative model evaluation consists of comparing predicted and experimental plots of the relevant variables. A qualitative evaluation can be a useful first step in model evaluation as it provides a general indication of the ability of a model to predict a particular scenario.

A quantitative evaluation is necessary for a more detailed and objective assessment of a model's performance. This is typically done by defining a set of statistical performance measures (SPM) that compare predicted and observed physical comparison parameters, as each measure has its advantages and disadvantages [11]. The selection of SPMs should also be consistent with previous work in order to gain experience with which values of SPMs represent a well performing model.

Each SPM should be associated with a range of values which indicate acceptable model performance. Defining acceptable ranges for the SPMs is quite difficult, because there are no theoretical "targets" that can be used for guidance. Instead, the definition of "acceptable" relies to a certain extent on previous experience with model evaluations in a particular area. For example, what could be considered an acceptable bias for a model simulating a complex phenomenon, such as deflagration to detonation transition (DDT), may be considered unacceptable for a model simulating a better understood phenomenon, such as the dispersion of an unobstructed jet release.

As discussed before, there has been vast experience accumulated in the field of dispersion modelling, therefore, acceptability criteria for statistical performance measures for dispersion models can be considered fairly well-established. However, model evaluation experience for other hazards is very limited, therefore, establishing acceptability criteria in those cases will require careful consideration. Additionally, any acceptability range should be periodically re-evaluated as additional experience is gathered and newer models are evaluated.

It is important to remember that model validation is only one of the tasks involved in the model evaluation, even though it is often the most "recognizable". Therefore, meeting all (or failing to meet some) of the SPM acceptability criteria should not automatically qualify (or disqualify) a model for use.

3.7.4 Sensitivity analysis

The purpose of a sensitivity analysis is to evaluate the changes in model predictions due to variations to the input parameters. Therefore, a model should be run multiple times for the same scenario, every time changing one input parameter over a specified range, and the change in modeling results reported – for example, as a tabulated percent of the base case, or as error-bars on a scatter plot.

The guidance provided in existing model evaluation protocols is rather generic, given the many factors that can affect model's predictions, and how factors are likely to be different for each type of model and hazard being considered. An example of requirements for sensitivity analysis is included in PHMSA's Advisory Bulletin ADB-10-07 [12] for the approval of vapor dispersion models. The parameters to be varied and the variability ranges should be consistent with experimental uncertainties in the respective validation datasets, and will be clearly indicated in the MVD.

3.7.5 Uncertainty quantification

Uncertainty quantification (UQ) characterizes sources of uncertainties in a model and propagates their effects to computed quantities of interest. The application of uncertainty quantification in this protocol is to establish credible bounds of predictability on computed quantities of interest and to assess model sufficiency based on computed variances. The protocol presents a well-defined workflow that may be used to assess uncertainties in simulation results for LNG hazard modeling. The workflow builds on well-established, peer-reviewed, guidelines such as those outlined in the American Society of Mechanical Engineers (ASME) Standard for Verification and Validation in Computational Fluid Dynamics and Heat Transfer [13] and the NASA monograph on simulation credibility and uncertainty quantification [14].

The steps in the workflow include:

1. Define UQ objective(s) and define computed quantities of interest.
2. Identify sources of uncertainties: Identify uncertainty sources that materially impact the quantities of interest for the stated UQ objective(s). Uncertainty sources are categorized as:
    a. Parametric uncertainties
    b. Initial condition/boundary condition uncertainties
    c. Source term or forcing function uncertainties
    d. Numerical uncertainties
    e. Model form uncertainties.

    These uncertainties can either be aleatory, epistemic, or mixed form uncertainties.
3. Characterize uncertainties: by assigning a mathematical form (e.g. Gaussian random variable) to the uncertainty source and numerical values (e.g. mean and standard deviation) to the form parameters.
4. Propagate uncertainties: propagate characterized input uncertainties to outputs using problem-appropriate UQ method (e.g. Latin hypercube sampling or stochastic expansion).

5. Summarize UQ results: in a format best suited for the decision process. Possible formats include statistical metrics (e.g. standard deviations), confidence intervals, system response cumulative distribution functions, or probability boxes.

4 Model Evaluation Report

The Model Evaluation Report (MER) represents the final product of a model evaluation, and the only public part of the evaluation. The objectives of the MER are to:

- Provide regulators with the information they need to determine whether to approve or reject a model, and to set any conditions or limitations on its use
- Convince all stakeholders that the model review was conducted in an objective and independent manner

Therefore, the MER needs to include sufficient information to allow the reader (for example, staff from a regulatory agency, model developer and users, or any other stakeholder) to understand the review process and the results.

4.1 Evaluation summary

The main section of the MER will consist of a summary of the model evaluation, which will address each section of the MEP, as described previously. In general, the MER will include the following for each evaluation task:

- A brief description of the purpose of the task
- A list of the information provided by the developer, including the questionnaire, memos/emails and published references
- A discussion whether the model meets or falls short of the requirements of the task. This will include an evaluation against each qualitative and quantitative assessment criterion applicable to that task.

Note that each evaluation task typically includes several subtasks, each with specific requirements. For example, the scientific assessment evaluates the physical models, governing equations and the numerical methods included in the model; the physical models are then broken down into several submodels (dense gas dispersion, atmospheric boundary layer profiles, etc.). The model's performance should therefore be evaluated at the "submodel" level (e.g., comparing different turbulence closure models, if available), to provide detailed information on the model's capabilities and limitations.

4.2 Best practice guidance

As discussed previously concerning sensitivity analysis, a model's output depends on a number of input parameters and options. Therefore, the approval of the model is contingent upon its use in a manner that is consistent with how the validation simulations were performed. For example, if a CFD model was evaluated for vapor dispersion using grid dimensions between 0.5 and 1.0 m and

found to perform well, it may not be reasonable to assume that it would also perform well when using grid dimensions of 2.5 m or larger.

In order to ensure that approved models are used in a manner expected to result in accurate predictions, the MER will include a section highlighting "best practices". These may include:

- Grid sizes
- Time steps
- Sub-models and property databases (e.g., turbulence closure, combustion, etc.)
- Boundary conditions (e.g., boundary layer profiles)
- Initial conditions or source terms

Deviations from best practices guidance when submitting simulations for regulatory approval, for example, may "void" the model approval and require the user to provide additional evidence that the model's predictions should be considered acceptable.

4.3 Recommendations for improvement

Even though model improvement is not a stated purpose of the model evaluation protocols developed during the current project, any observations made during the model evaluation, which could lead to an improvement in the capabilities of a model to accurately predict the consequences of a hazard scenario, should be identified and explained in the MER. However, the model developer will not be required to act on these recommendations as a condition for the approval of future versions of the model.

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MARY KAY O'CONNOR PROCESS SAFETY CENTER
TEXAS A&M ENGINEERING EXPERIMENT STATION

23rd Annual Process Safety International Symposium
October 20-21, 2020 | College Station, Texas

Beirut: How behaves Ammonium Nitrate Exposed to Fire and How Strong and Damaging is its Explosion?

Charline Fouchier, Hans Pasman, Sunhwa Park, Noor Quddus, Delphine Laboureur

Von Karman Institute for Fluid Dynamics, Sint Genesius Rode

TEES Mary Kay O'Connor Process Safety Center, Texas A & M University System

The Beirut explosion on the 4th of August is one more accident to be added to the long list of tragedies caused by Ammonium Nitrate. While many investigations have been conducted to understand better the behaviors of the molecule, it is still unclear how the Ammonium Nitrate can detonate in an unconfined environment while heated by fire.

A rapid summary of the state of art on Nitrate Ammonium is given, followed by an analysis of the Beirut accident, with a proposed scenario that could have led to the explosion.

Finally, methods to estimate the explosion energy, based on the blast arrival time, the damages on buildings and the crater dimensions are applied on the Beirut accident and compared.

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Flammable mist hazards involving high-flashpoint fluids

Simon Gant[1]*, Anthony Giles[2], Richard Bettis[1], Philip Bowen[2]
[1] Health and Safety Executive (HSE), Harpur Hill, Buxton, UK
[2] Gas Turbine Research Centre (GTRC), Cardiff University, Wales, UK
simon.gant@hse.gov.uk

Abstract

In 2009, the UK Health and Safety Executive (HSE) published a review of serious incidents involving ignition of flammable mists of high-flashpoint fluids, i.e. fluids whose vapours cannot be ignited and sustain a flame at normal room temperature (e.g. kerosene, diesel, lubrication oils and hydraulic oils). The review identified 37 incidents which together were responsible for 29 fatalities. In response to the findings, HSE and a consortium of other regulatory and industrial sponsors funded a Joint Industry Project (JIP) on the subject, which ran from 2011 to 2015. The work included a detailed literature review and a series of experiments at Cardiff University on a mist release configuration consisting of a downwards-pointing spray from a 1 mm diameter circular orifice. Test pressures ranged from 1.7 bar to 130 bar and three fluids were tested: Jet A1 (kerosene), a light fuel oil and a hydraulic oil. Computational Fluid Dynamics (CFD) simulations were also performed, and results were compared to existing hazardous area classification guidelines. The work was used to devise a preliminary classification scheme for mist flammability, based on a fluid's flashpoint and ease-of-atomization.

Several important questions remained unanswered following the first JIP relating to the effect of the orifice shape, size and release configuration, and the ignition characteristics of other common fluids, notably diesel. In 2018, HSE launched a follow-on JIP (currently ongoing) which aims to address these issues. The work started with an updated review of flammable mist incidents, published in 2019. Experiments on diesel have started in 2020 at Cardiff University and further, larger-scale experiments are planned for 2020-2021 at the HSE Science and Research Centre, Buxton.

This paper and presentation at the MKOPSC International Symposium 2020 provides an overview of the work led by HSE on flammable mists over the last decade, and a summary of the preliminary results from the ongoing experiments.

Keywords: flammable mist, high-flashpoint fluid, spray, ignition

Introduction

Combustible liquids are typically classified by their flashpoint temperature and are often regarded as being relatively non-hazardous if they are handled at temperatures well below their flashpoint (EI, 2015). However, if a high flashpoint liquid is atomised to produce a mist of fine droplets, it can be ignited below its flashpoint and produce a fire or explosion. These flammable mists hazards are mainly a concern for leaks from pressurized systems (e.g. pumps, pipework, valves) as a result of corrosion, mechanical damage, cracks, seal failures or loosening of screwed fittings (Eckhoff, 1995). Flammable mists can also be produced by condensation of vapour, such as that produced by an overheated bearing in a marine diesel engine (Freeston *et al*., 1956).

In Europe, there are regulations controlling flammable atmospheres, namely the ATEX 'Workplace' Directive (1999/92/EC)[1] across the EU and the Dangerous Substances and Explosive Atmospheres Regulations (DSEAR)[2] in Great Britain. DSEAR was introduced to implement the ATEX Directive and has been retained within the UK following its departure from the EU. These regulations require the identification of any zones where a flammable atmosphere could form, either as part of normal operations or in the event of a reasonably-foreseeable equipment failure. Within such zones, all ignition sources must be controlled by using appropriate equipment rated for use within hazardous areas. Both ATEX and DSEAR cover flammable atmospheres produced by gases, dusts and/or mists.

There is established guidance available on the extent of hazardous areas for flammable gases (e.g. BSI, 2016; EI, 2015), but relatively few guidance documents or standards are available to assess equivalent hazardous areas for flammable mists, or to help select safe equipment for use in flammable mist atmospheres. The relevant British and European standard, BS EN 60079-10-1 (BSI, 2016) contains limited guidance in Annex G, but this is qualitative rather than quantitative. The Energy Institute model code of safe practice EI15 (EI, 2015) provides guidance on area classification for installations handling flammable fluids, which includes tabulated hazard distances for higher flashpoint fluids leaking at different pressures through various specified hole sizes. However, the document acknowledges that "there is little knowledge on the formation of flammable mists and the appropriate extents of associated hazardous areas". The release conditions given in EI15 are also tailored towards relatively large-scale equipment, with hole sizes ranging from 1 mm to 10 mm.

Many areas of concern exist in plant rooms and other enclosed areas. Here, the lack of definitive guidance often leads to the whole plant room being considered as a hazardous area. Many assessments assume that leaks do not form a mist at lower pressures (perhaps below 5 or 10 bar gauge), and it is not clear that all potential mist hazards are fully recognised.

Is there a need for oil mist zoning?

Common items of industrial equipment may have the potential to produce oil mists. For example, hydraulic equipment, lubricating oil systems and delivery lines for high-flashpoint fuels (diesel,

[1] https://eur-lex.europa.eu/legal-content/EN/TXT/?uri=CELEX:31999L0092 (accessed 10 September 2020).
[2] https://www.hse.gov.uk/fireandexplosion/dsear.htm (accessed 10 September 2020).

kerosene etc.) could create aerosols if they failed under pressurised conditions. Such equipment is in widespread use, but the creation and ignition of flammable mists does not seem to be a frequent occurrence. The apparent lack of mist explosion events suggests that there are often mitigating factors preventing flammable mists being created or ignited. Understanding these factors could allow more accurate assessment of when control measures are unnecessary and when they are essential.

It should be noted that the perception of oil mists seems quite different on board ships. Following several incidents involving loss of life, the Safety of Life At Sea (SOLAS) regulations require unattended engine crankcases to be fitted with oil mist detectors and automatic shutdown (IMO, 1974). The International Maritime Organisation (IMO) notes that the majority of engine room fires are the result of oil mist formation and has guidance on the fitting of oil mist detectors in these more open areas.

In 2009, the UK Government's safety regulator, the Health and Safety Executive (HSE), published a review of serious incidents involving ignition of flammable mists of high-flashpoint fluids (Santon, 2009). HSE also recently worked in collaboration with the French national laboratory INERIS[3] and the Université de Lorraine in France (Lees *et al*., 2019) to produce a systematic study of three European national incident databases: the UK's Offshore Hydrocarbon Release Database, the French ARIA database and the German ZEMA databases. These 2009 and 2019 incident reviews both showed that while oil mist explosions were relatively infrequent events, they happen sufficiently often that the possibility of one occurring should not be ignored. For example, the latter study showed that over a 30 month period from 2016 to 2018, there was approximately one incident per month that involved fluid mists or sprays on offshore oil and gas installations operating on the UK continental shelf. Oil mist explosions have led to deaths, injuries and significant property loss.

MISTS – a Joint Industry Project

Following the review of mist incidents in 2009, HSE set up a joint industry project to help improve our understanding of flammable mists. The four-year project started in December 2011 and was jointly sponsored by 16 industry and regulatory partners (HSE, ONR, RIVM, GE, Siemens, EDF/British Energy, RWE, Maersk Oil, Statoil, BP, ConocoPhilips, Nexen, Syngenta, Aero Engine Controls, Atkins, Frazer Nash and the Energy Institute).

Literature review

The first stage of the project was an extensive literature review that examined three fundamental issues: mist flammability, mist generation and mitigation measures (Gant *et al.*, 2012; Gant, 2013). Data on mist flammability were reviewed that included measurements of the Lower Explosive Limit (LEL), Minimum Ignition Energy (MIE), Minimum Igniting Current (MIC), Maximum Experimental Safe Gap (MESG) and Minimum Host Surface Ignition Temperature (MHSIT). One of the significant findings was that the LEL for mists could fall to as low as 10% of the LEL for the vapour of the same substance (i.e. to around 5 g/m^3). Mists were found generally to be more

[3] Institut National de l'Environnement Industriel et des Risques, http://www.ineris.fr (accessed 10 September 2020).

difficult to ignite than the equivalent vapour, due to the energy needed to vaporise droplets prior to ignition.

Regarding releases from pressurized systems, the literature review noted that one of the challenges in conducting tests with mists (as compared to gases and dusts) is the complexity introduced by droplet breakup and agglomeration, impact with surfaces and evaporation. Correlations for primary atomisation, secondary droplet breakup, and droplet impingement were reviewed. Much of the historical work on sprays was found to be motivated by the development of internal combustion engines and gas turbines. The study noted that there was uncertainty in applying correlations developed for those applications to the very different scales and operating pressures typical of industrial equipment requiring hazardous area classification (e.g. pumps and pressurized pipework).

Release classification

Following the literature review, a method to classify releases was developed to group together similar fluids and release scenarios (Burrell and Gant, 2017). This was deemed necessary because of the large number of high-flashpoint fluids in use across a range of industries, which would make detailed case-by-case assessments impractical. The classification method was based on two factors that were considered to be significant for flammable mist formation from leaks of fluids under pressure, namely the flashpoint of the fluid and the propensity for releases to atomize into droplets. The chosen atomization criteria were calculated based on the Ohnesorge number, Oh, which is a dimensionless parameter that depends on the fluid viscosity (μ), density (ρ) and surface tension (σ) and the equivalent diameter of the leak (D):

$$Oh = \frac{\mu}{\sqrt{\rho D \sigma}} \tag{1}$$

Ohnesorge (1936) provided an empirical correlation for atomization breakup that depends on the Reynolds number of the fluid released through the orifice, Re:

$$Oh_c = 745 Re^{-1.22} \tag{2}$$

To characterise the propensity of a given release to atomize, it was proposed to use the ratio of Oh to Oh_c. The greater the value of this ratio (Oh/Oh_c) was taken to indicate a greater propensity for the release to atomize into small, more easily ignitable droplets.

Figure 1 shows the flashpoint plotted against this ratio for a range of different fluids released through a 1 mm diameter hole at 10 bar gauge pressure. These values were chosen to represent a credible accident scenario and it is one of the conditions considered in EI15. It is also readily achievable in experimental studies. The fluids assessed were chosen because they were of particular interest to the organisations sponsoring the JIP. The vertical and horizontal error bars in Figure 1 show the range in flashpoint and (Oh/Oh_c) values resulting from the range in fluid properties, which were taken from various data sources in the literature (see Burrell and Gant, 2017 for details). For biodiesel, the effect of changing the release temperature (which alters the fluid properties) is also shown in the Figure.

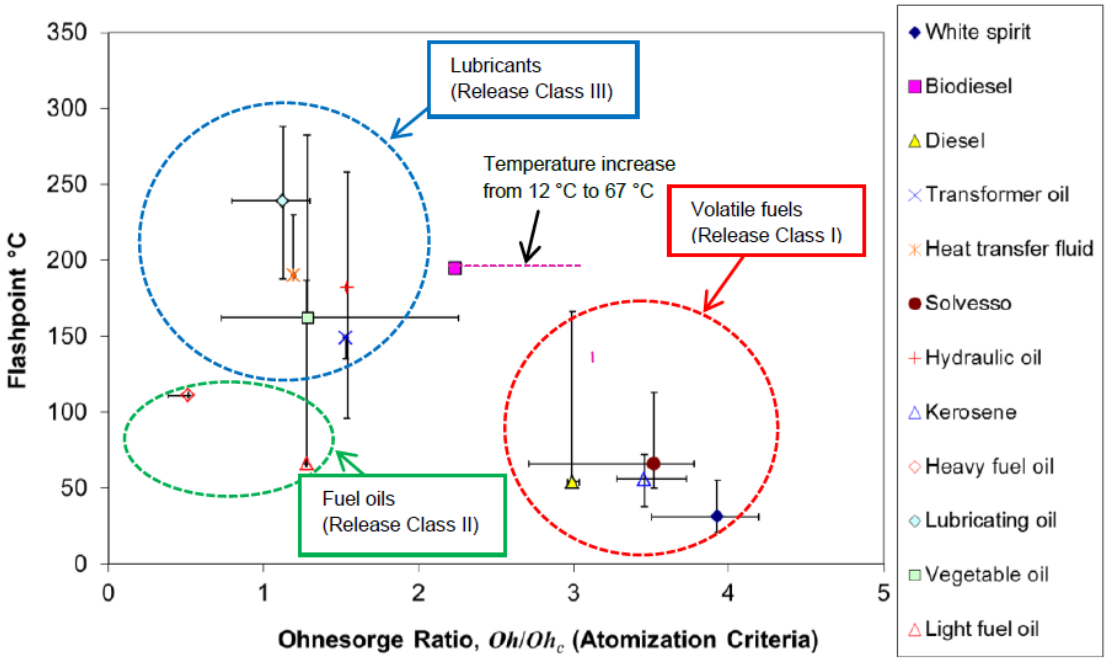

Figure 1. Mist classification by flashpoint and atomisation criteria for a 10 bar release through a 1 mm hole

The fluids appeared to fall naturally into three groups in Figure 1. To generalise this classification system to all fluids, the figure was split into quadrants defining four "Release Classes" (see Figure 2), which can be summarised as:

- **Release Class I**: More volatile fluids that are more prone to atomisation, such as many commercial fuels.
- **Release Class II**: More volatile fluids that are less prone to atomisation, such as viscous fuel oils at ambient temperatures.
- **Release Class III**: Less volatile fluids that are also less prone to atomisation, such as many lubricants and hydraulic fluids at cool (near ambient) temperatures.
- **Release Class IV**: Less volatile fluids that are more prone to atomisation, such as many lubricants and hydraulic fluids at high temperatures that may arise during use.

The specific values used to bound the four Release Classes (i.e. flashpoint of 125 ºC and Ohnesorge ratio of 2) were selected based on the best judgement at the time. As and when new evidence becomes available it is possible (and even likely) that these bounds may need to be revised.

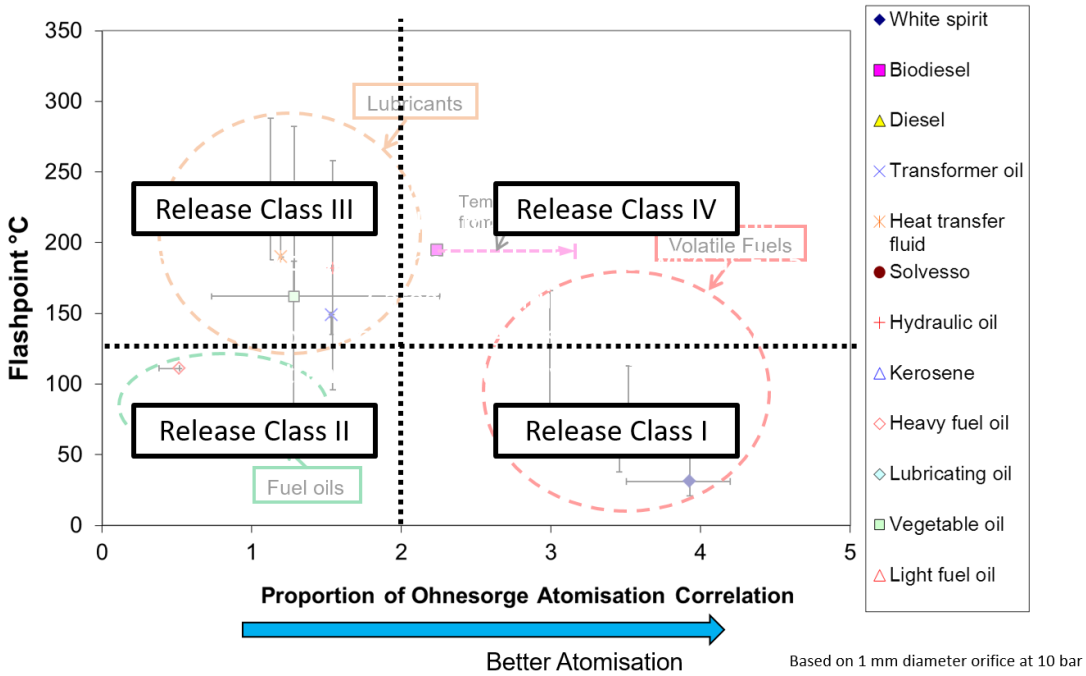

Figure 2. Mist classification diagram overlaid with the four release classes

Experimental studies

To investigate the flammability of mists, experiments were conducted at Cardiff University's Gas Turbine Research Centre (GTRC) using three exemplar fluids: one each from the Release Classes I, II and III (Mouzakitis *et al*., 2017a; 2017b). These experiments were designed to produce a mist from a pressurised leak through a small hole. The aim was to determine whether the mists produced by the three fluids could be ignited and, if so, to use a spark igniter to map the extent of the flammable cloud. Following this, the aim was to measure droplet size distributions and concentrations at the edge of the flammable cloud, to investigate the LEL.

The tests were all conducted using a 1 mm diameter, smooth-bore, cylindrical plain orifice with length-to-diameter ratio of 2. The releases were all directed downwards to prevent asymmetric effects, and the experiments were conducted within a 1.2 m square, 2.5 m tall test chamber (shown in Figure 3). This configuration provided a good starting point using a simple arrangement that is well defined and readily repeatable. The 1 mm hole size is the smallest hole tabulated in the EI15 model code of safe practice and it therefore allowed for direct comparison to the existing guidance.

The ignition tests all used a 1 Joule electric spark (Chentronic's Smartspark[4]). Prior to the start of experimental work, there was considerable discussion within the project's Steering Committee regarding the ignition source. The intention was for the source to represent a credible upper limit for most situations where area classification would be considered. While most commonly occurring electrical sparks are significantly lower in energy, the consensus view was that 1 J represented a reasonable upper limit. Situations with the potential for higher-energy ignition

[4] https://www.chentronics.com/products/smart-spark (accessed 10 September 2020).

sources (or even naked flames) may exist in a few cases, but these were considered to be sufficiently unusual that they would be outside the scope of normal guidance.

During a series of releases, the igniter was placed at a set axial distance from the release point and then tracked across a radius of the jet to locate points just inside and outside of the ignitable envelope. Similar radial tracks were tested at several different distances along the axis of the expanding jet.

Droplet sizes and concentrations were measured using a non-intrusive laser Phase Doppler Anemometer (PDA) system (Dantec Dynamics coherent Innova 70-5 Series argon-ion laser with BSA P60 flow and particle processor[5]). PDA measurements were made following the ignition tests at locations inside and outside of the ignitable envelope at the same locations as those tested for ignition.

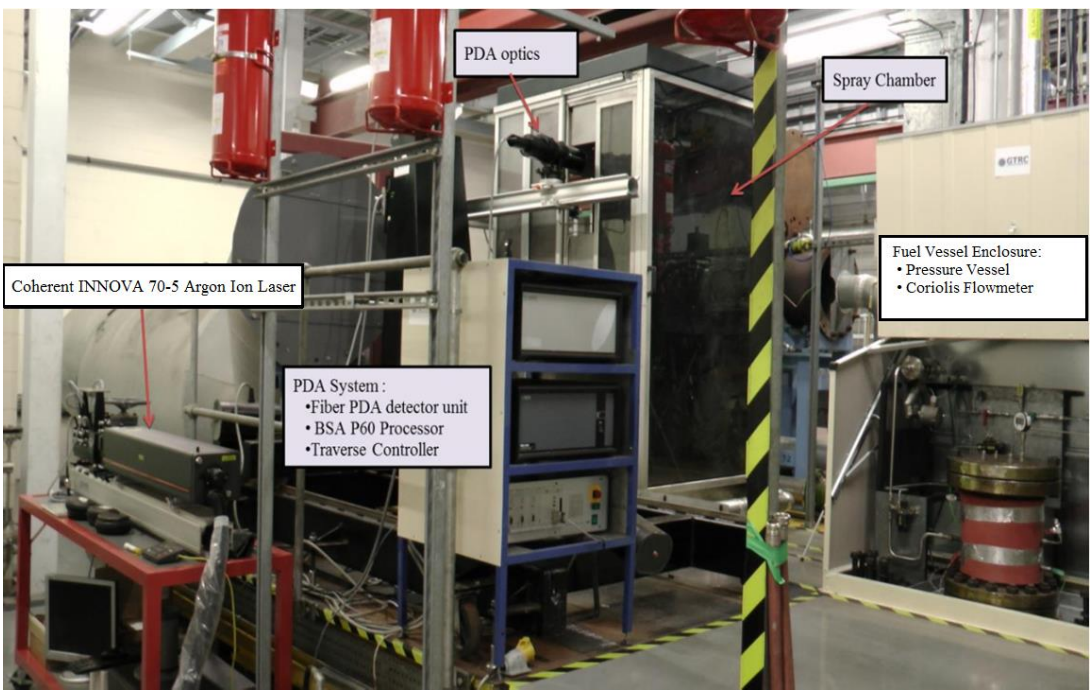

Figure 3. GTRC test apparatus configured for PDA measurements

The three fluids tested were:

- **For Release Class I**: Kerosene (Jet A1), low viscosity, flashpoint = 38 °C)
- **For Release Class II**: Light Fuel Oil (LFO), higher viscosity, flashpoint = 81 °C.
- **For Release Class III**: Hydraulic oil, higher viscosity, flashpoint = 223 °C.

[5] https://www.dantecdynamics.com/ (accessed 10 September 2020).

Tests were also carried out with the same LFO heated to 70 °C. This temperature was still below the flashpoint, but the increase in temperature changed the physical properties sufficiently that the heated LFO was moved from Release Class II into Release Class I.

For each fluid, tests were first carried out at release pressures of 5, 10, 15 and 20 bar gauge. Following these test pressures, further tests were carried out for kerosene and the hydraulic oil at lower and higher pressures (respectively) to determine the limiting pressures were the mist could be ignited.

The tests described above were all carried out for a "free jet" configuration, with the unobstructed spray directed downwards from the top of the enclosure. Following these tests, a further set of results were obtained for LFO and hydraulic oil with a flat mild steel impingement plate located at either 150 mm or 400 mm below the orifice. In both cases, the igniter was located 25 mm above the plate. The aim of these impingement tests was to see whether mists that could not be ignited in the free jet configuration could be ignited after they had impinged at high-velocity onto a solid surface and broken up to produce a finer mist.

Experimental results

Figure 4 shows some example photos of the ignition tests. The tests showed significant differences between the three fluids from the different Release Classes (see Table 1). In the free jet tests, kerosene was found to ignite at all of the test pressures from 5 to 20 bar. The pressure was then reduced in steps of 1 bar to the lowest test pressure possible on the apparatus (1.7 bar gauge) and the kerosene mist ignited in each case. The hydraulic oil showed the opposite behaviour and could not be ignited at any of the pressures between 5 and 20 bar. The pressure was then increased in stages up to 130 bar and the hydraulic oil mist still did not ignite fully, although there were occasional localised flashes near the spark igniter but the flame did not propagate through the mist. LFO did not ignite at ambient temperature across the range of pressures from 5 to 20 bar, but when it was heated to 70 °C it ignited at all of the pressures.

In the impingement tests, the hydraulic oil again did not ignite at any of the pressures. The LFO at ambient temperature did not ignite at a pressure of 15 bar but did ignite at the higher pressure of 20 bar. When the LFO was heated to 70 °C, it behaved in the impingement tests as it had done previously in the free jet tests and ignited at all of the pressures from 5 to 20 bar.

In addition to these ignition test results, it was noted that there were significant differences in the visible appearance of releases with the different fluids. At one extreme, the hydraulic oil released at lower pressures remained largely concentrated in a dense, almost unbroken core of liquid with very few small droplets being formed. At the other extreme, a significant proportion of kerosene was well atomised even at very low pressures.

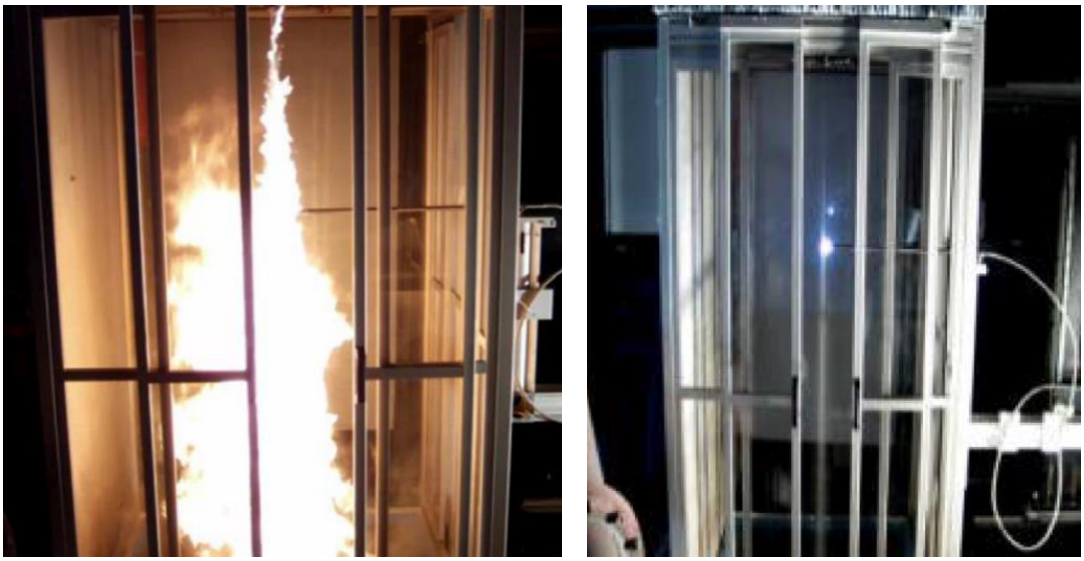

Figure 4. Ignition tests with kerosene (left) and hydraulic oil (right)

Table 1. Ignition Test results

Spray Geometry	Fluid (Flashpoint)	Temperature	Release Pressure (barg)	Ignition
Free spray	Kerosene (FP = 38 ºC)	Ambient	1.7, 2, 3, 4, 5, 10, 15, 20	At all pressures
Free spray	Hydraulic oil (FP = 223 ºC)	Ambient	5, 10, 15, 20, 30, 70, 110, 130	No, but some "flashes" at highest pressures
Free spray	Light fuel oil (FP = 81 ºC)	Ambient	5, 10, 15, 20	No
Free spray	Light fuel oil (FP = 81 ºC)	70 °C	5, 10, 15, 20	At all pressures
Impinging	Hydraulic oil (FP = 223 ºC)	Ambient	5, 10, 15, 20	No
Impinging	Light fuel oil (FP = 81 ºC)	Ambient	15, 20	At 20 barg only
Impinging	Light fuel oil (FP = 81 ºC)	70°C	5, 10, 15, 20	At all pressures

Figure 5 shows an example map of the ignition test results through the kerosene free jet release at 5 bar showing both ignition locations and positions where PDA droplet size and concentration measurements were made. The PDA system worked from direct measurements of individual droplets, collecting many thousands of measurements to obtain a statistical analysis of the aerosol within a small measuring volume. The lack of small droplets in the releases with hydraulic oil and LFO meant that the measurements were of low quality for those fluids. Some good PDA measurements were obtained, but those only corresponded to areas where ignitions were certain.

There was little or no good quality data from locations outside the ignition envelope or even on the borderline.

In the kerosene tests, the minimum calculated concentration from the PDA measurements for a successful ignition point was 3 g/m^3. However, the average concentration of the ignition positions near the edge of the flammable envelope was 69 g/m^3. The average concentration of the unsuccessful ignition points from the PDA measurements was 21 g/m^3. In comparison, the LEL for kerosene vapour is approximately 48 g/m^3 (Zabetakis, 1965).

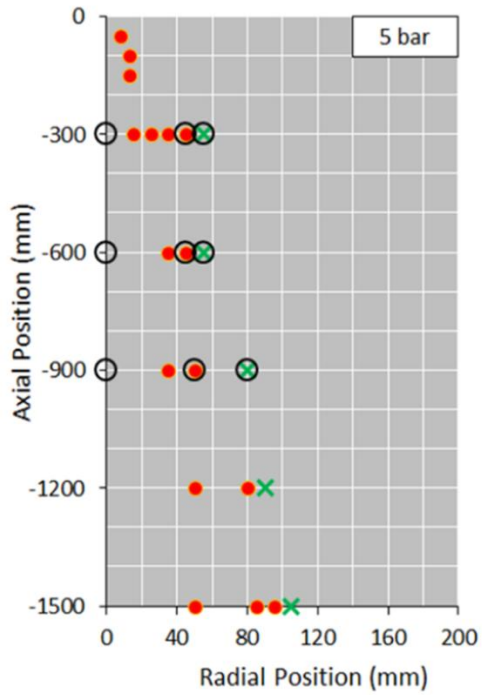

Figure 5. Cross-section through the 5 bar kerosene mist showing the igniter positions as either red dots (where ignition was successful) or green crosses (where ignition was attempted but the mist did not ignite). The locations where PDA droplet size and concentration measurements were made are shown as black circles. The release point is at zero on the axial and radial axes.

CFD modelling

Alongside the experimental studies, a set of CFD simulations were performed using the ANSYS CFX-15 software[6] (Coldrick and Gant, 2017). The model used an Eulerian-Lagrangian approach in which the GTRC spray booth was represented by a fixed computational mesh and the spray was represented by individual computational particles (representing a statistical sample of droplets) that were tracked through the flow. Computational particles were released at the orifice location and droplets were allowed to break apart under aerodynamic forces. The model accounted for the transfer of mass, momentum and energy between the droplets and the surrounding air. Tests were performed to ensure that the results were insensitive to the grid cell size and particle count. For most of the simulations, a grid of 1.3 million nodes was used with 10,000 particles.

[6] http://www.ansys.com (accessed 10 September 2020)

At the orifice, the primary breakup of the liquid into droplets was defined in the model by specifying the initial spray cone angle and initial droplet size. Seven different cone angle models and nine different droplet size models were tested. Two secondary breakup models were also tested to account for aerodynamic forces on droplets. Details of these models are given in the report by Coldrick and Gant (2017). Sensitivity tests with different combinations of models were undertaken and results compared to the data from the GTRC experiments on kerosene at 20 bar. The best performing combination of models was then used to model all of the other tests (i.e. the kerosene tests at different pressures and the hydraulic oil and LFO free jets for pressures between 5 and 20 bar). The best performing primary breakup model was found to be the DNV Phase III JIP Rosin-Rammler correlation (DNV, 2006), which gave predictions within a factor of 2 for the measured droplet concentration and droplet diameter for the kerosene releases. Predictions for the hydraulic oil and LFO were in worse agreement with the measurement data. The main issue there was that the CFD model assumed that the release atomized whereas in the experiments only a small fraction of the liquid was actually atomised. The model therefore predicted flammable concentrations to occur when in practice the mist could not be ignited. Examples of the CFD results are given in Figure 6.

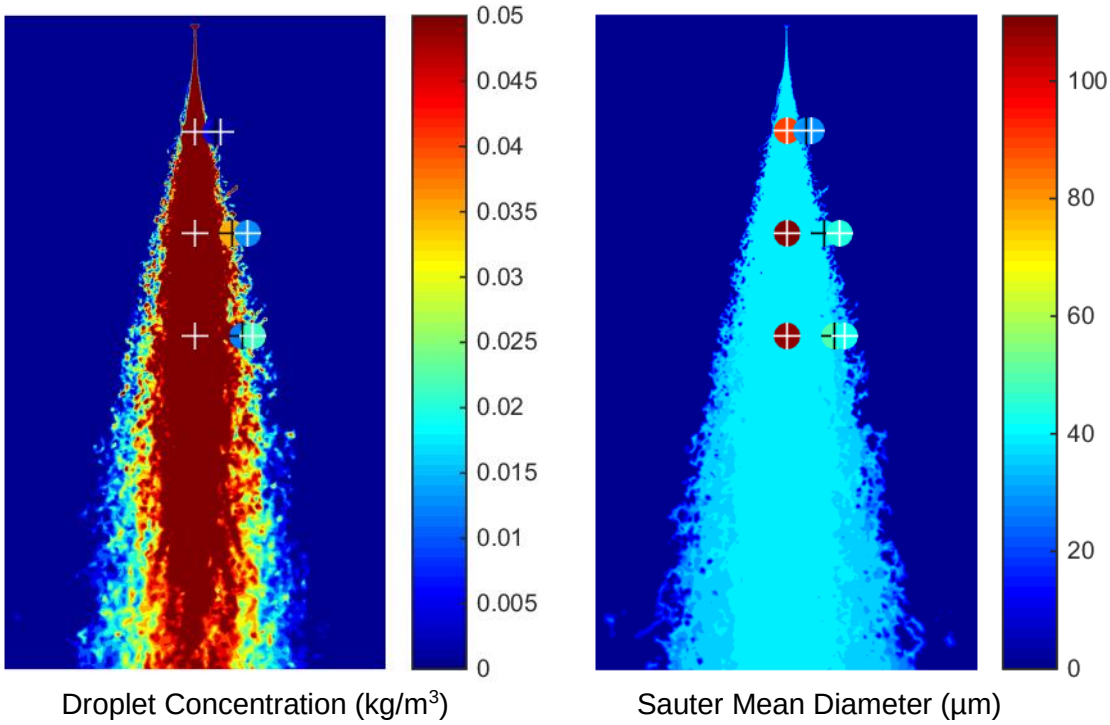

Figure 6. CFD predictions of droplet concentration and size for the kerosene release at 20 bar. Coloured circles show measured values with the same colour scale as the contours. Crosses indicate where ignition occurred (a black cross) or did not occur (a white cross). Ignitions were not attempted on the centreline and only droplet diameter and concentration were measured there (a white cross on the centreline is used to identify solely the measurement location). Note that the scales chosen are not the maximum levels: concentrations in excess of 50 g/m^3 are shown in the left-hand figure as red.

The kerosene CFD model was subsequently used to predict the extent of the flammable mist cloud for Category C fluids in Table C4 of the EI15 code of safe practice (EI, 2015). These EI15 values were originally determined using the consequence modelling software DNV-GL Phast[7] for spray releases directed horizontally in a 2 m/s wind, where the wind was blowing in the same direction as the release (i.e. co-flowing). The hazard range was defined in EI15 as the distance to the LEL, which was assumed to be a droplet concentration of 43 g/m^3. EI15 presents results for four different hole sizes of 1 mm, 2 mm, 5 mm and 10 mm and four pressures of 5 bar, 10 bar, 50 bar and 100 bar. The same set of conditions was modelled using CFD, although the pressures were modelled as gauge pressure whereas the EI15 values are for absolute pressure, i.e. the CFD results were for a 1 bar higher pressure in each case. The configuration of the CFD model was the same as that described earlier in the model validation study, with a vertical downwards spray in nil wind, using the DNV Phase III JIP primary droplet breakup model.

The results comparison (Table 2 and Figure 7) showed that the CFD model gave somewhat larger hazard distances than those given in EI15, particularly for lower pressure releases. The EI15 distances all increase with pressure, but the CFD results exhibit more complex behaviour. This was likely due to the EI15 hazard distance assuming a horizontal release, whereas the CFD value was for a vertically-downwards release.

Table 2. Predicted hazard distances from CFD model compared to EI15 values for Category C fluids

Release Pressure, bar	***Hazard Distance (m) for Release Hole Diameter of:***							
	1 mm		***2 mm***		***5 mm***		***10 mm***	
	EI15	***CFD***	***EI15***	***CFD***	***EI15***	***CFD***	***EI15***	***CFD***
5	2	4.3	4	7.1	8	16	14	28
10	2.5	3.4	4.5	5.7	9	13	17	23
50	2.5	2.8	5	5.4	11	13	21	27
100	2.5	3.0	5	6.0	12	13	22	27

The literature review of mists by Gant (2013) showed that the LEL in quiescent mists could be lower than the 43 g/m^3 value assumed by EI15, by as much as a factor of 10 (i.e. approximately 5 g/m^3). These lower concentration ignitions were observed in experiments with a strong ignition source at the base of a quiescent mist cloud. Given this finding and the longer hazard distances produced by the CFD model for downwards-directed releases, the results suggested that hazardous distances could extend over a spherical volume with a radius around the release point similar to that given in EI15, but with the hazardous zone extending over a greater distance downwards in a cylindrical region below the release point (potentially, to the floor).

[7] http://www.dnvgl.com/phast-and-safeti, accessed 10 September 2020.

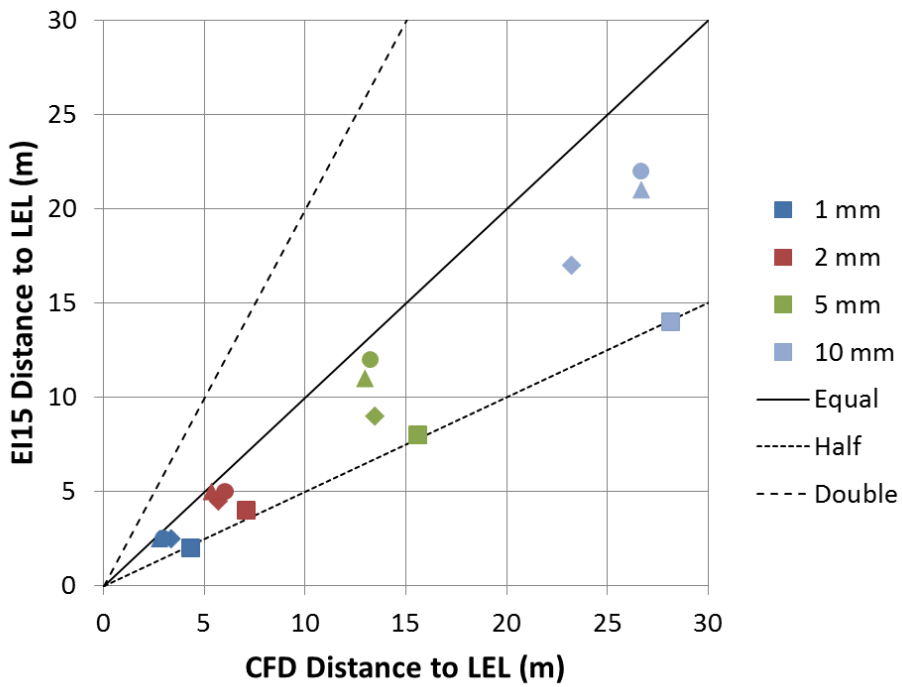

Figure 7. Comparison of CFD predictions to the guidance on hazard distances for mists produced by Category C fluids from Table C4 of EI15 (EI, 2015). Symbols are coloured according to the orifice diameter as shown in the key. Symbol shapes indicate the release pressure as follows: ■ 5 bar, ◆ 10 bar, ▲ 50 bar, ● 100 bar.

Additions to guidance

Based on the findings of the MISTS project, some tentative new guidance was developed (see Figure 8 and Bettis *et al*., 2017). Whilst the MISTS experimental and modelling results confirmed that the EI15 guidance was broadly appropriate, the new results identified differences between fluids that fell within the broad class of EI15 Category C fluids. Where the MISTS experimental findings clearly showed that particular releases did not produce ignitable mists, the new guidance reflected the absence of a flammable zone. In the case of Release Class I, the ignition of kerosene at lower pressures than the lowest pressure of 5 bar in the relevant EI15 table was also highlighted.

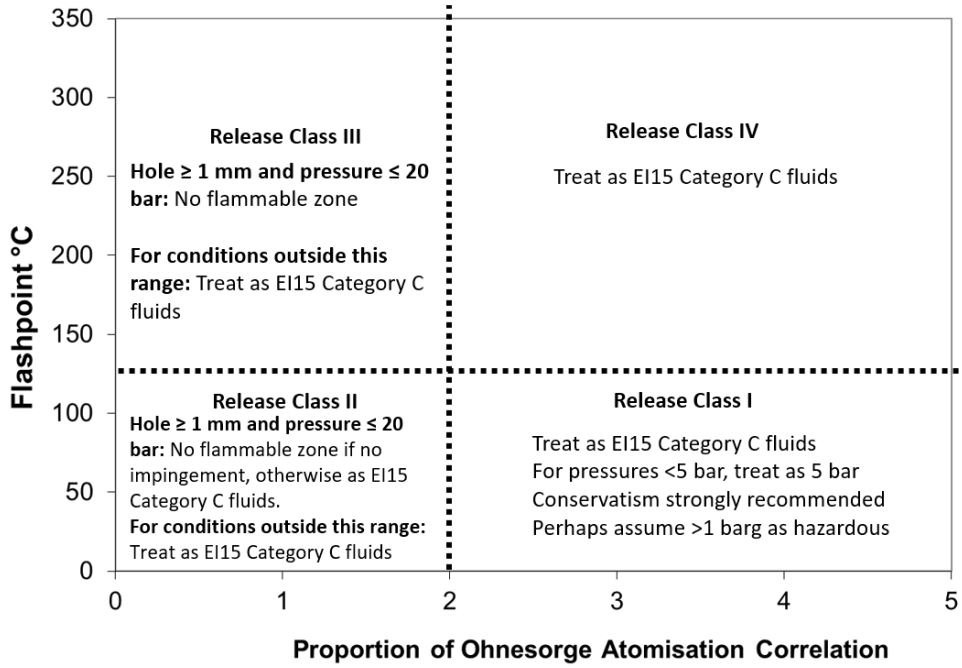

Figure 8. Tentative mists hazardous area classification guidance produced by the MISTS project

MISTS2 project

Following the end of the MISTS project, HSE led a workshop involving other regulatory agencies, industry groups and consultancies to discuss the findings and possible future work. Based on information gathered in that consultation exercise, HSE proposed a second project under its "Shared Research" programme and invited other organisations to contribute time and funding to increase the amount of work that could be undertaken. The new MISTS2 project began in 2018 and it was planned to finish at the end of 2020. However, due to the COVID-19 pandemic, the project is now likely to extend into 2021. In addition to HSE, the project is being supported by Shell, Électricité de France (EDF), the Office for Nuclear Regulation (ONR), the Energy Institute and INERIS. The scope of work for this ongoing MISTS2 project are described below.

Diesel fuel

The somewhat unexpected ignitions of kerosene at very low pressures in the MISTS project raised questions about diesel. It is very widely used and has similar fluid properties to kerosene with a flashpoint around 20 °C higher. Understanding the potential for diesel to create flammable mists, particularly at low operating pressures, is a priority task in the MISTS2 project.

The diesel tests are using the same experimental test procedures as those used in the previous MISTS programme, to allow like-for-like comparison of results. Two different diesel fuels are being tested: the first is an 'ultra-low sulphur' diesel that is typical of the UK vehicle 'pump' diesel (available from petrol stations, or US gas stations), which is largely composed of mineral-oil derived fuel, and the second fuel is a 100% biodiesel. The biodiesel has a flashpoint of 145 °C, significantly higher than the 58 °C flashpoint of the standard 'pump' diesel blend.

At the time of writing (September 2020), the GTRC test rig had been redesigned and rebuilt to allow the duplicate testing in a more robust and safe test environment (see Figure 9) and the ignition tests have been completed. The 'pump' diesel was found to ignite at all the pre-defined pressures of 5, 10, 15 and 20 bar gauge. A test at a lower pressure of 3 bar gauge did not ignite. The biodiesel could be ignited at a release pressure of 20 bar gauge but did not ignite at the lower test pressures of 5 to 15 bar gauge. Work is currently underway to visualise the spray and measure the droplet sizes and concentrations.

Figure 9. GTRC test rig rebuilt for MISTS2 studies

Hole shape

All of the experimental work to date has used a 1 mm diameter drilled circular orifice with a length-to-diameter ratio of two. In practice, accidental releases of fluids will involve a variety of situations where the leak path has a more complex geometry. Examples might include:

- Holes created by corrosion, where leaks are likely to have very short path lengths through thinned material, with rough edges;
- Loosened screwed fittings, where the leak is along the threads;
- Cracked pipes or fittings, where the leak is through a relatively long and narrow opening;
- Damaged or missing seals and gaskets, where the leak is through an arc of the fitting.

To better understand whether the range of possible release paths will alter the likelihood of a flammable mist being created, a series of tests will be carried out with more complex orifices. Additive manufacturing (i.e. 3D printing) will be used to create orifices with different geometries. For each geometry, a range of small size variations will be produced and tested to select ones that

closely match the discharge rates of the circular nozzle. In this way, differences in the mists will only be due to changes in the leak shape rather than flow rate.

Ignitable extent

The MISTS experiments were conducted in a relatively small-scale test chamber that did not provide data on the maximum extent of the flammable cloud on the flow centreline. In the kerosene tests, the mist could be ignited all the way to the floor of the chamber. Since the maximum extent of the flammable cloud is such an important parameter for hazardous area classification, it is proposed in the MISTS2 project to duplicate the GTRC releases in a much larger indoor facility at the HSE Science and Research Centre in Buxton, England (see Figure 10).

The pressurised releases in the HSE Burn Hall will be directed vertically downwards from a boom offset from a 10 m high scaffold. To minimise differences from the MISTS releases and the MISTS2 trials, the same GTRC orifices will be used. The ignition trials will also use the same spark igniter, which GTRC have agreed to loan to HSE for these tests. It is currently proposed to use diesel for these experiments. The HSE test rig will allow the igniter to be placed on the centreline, or slightly offset from it if there is a dense liquid stream in the centre. Ignition locations will extend out to axial distances (below the orifice) in excess of 8 metres, which is well beyond the flammable cloud extent predicted by current models. It is anticipated that these tests will provide evidence to support current guidance and future predictive modelling.

Figure 10. The indoor Burn Hall at HSE Science and Research Centre

Summary

In 2009, the UK Health and Safety Executive (HSE) published a review of serious incidents involving the ignition of flammable mists of high-flashpoint fluids, which identified 37 incidents which together were responsible for 29 fatalities. In response to the findings, HSE and a consortium of other regulatory and industrial sponsors funded the MISTS Joint Industry Project, which ran from 2011 to 2015. The project involved a detailed literature review and a series of experiments at Cardiff University on a mist release configuration consisting of a downwards-pointing spray from a 1 mm diameter circular orifice. Test pressures ranged from 1.7 bar to 130 bar and three fluids were tested: kerosene, a light fuel oil and a hydraulic oil. CFD simulations were also performed, and results were compared to existing hazardous area classification guidelines. One of the notable results from the experimental work was that mists of kerosene (with a flashpoint of 38 ºC) could be ignited with release pressures as low as 1.7 bar. The findings from the MISTS project were used to develop a tentative classification scheme for mist flammability, based on the fluid's flashpoint and ease-of-atomization.

Several important questions remained unanswered following the MISTS project, relating to the effect of the orifice shape, size and release configuration, and the ignition characteristics of other common fluids, notably diesel. In 2018, HSE launched a follow-on project, MISTS2, which is currently ongoing. This new project is conducting tests on diesel, on different orifice shapes and taking measurements of the maximum extent of the flammable mist. Preliminary results have shown that standard 'pump' diesel can be ignited at pressures from 5 to 20 bar gauge, but not at 3 bar gauge. Tests with a higher flashpoint 100% bio-diesel found that it can be ignited at 20 bar gauge but not at lower pressures. Further work is ongoing at GTRC Cardiff University and at the HSE Science and Research Centre in Buxton.

The flammability of mists is a complex subject and there are many unknowns that need to be addressed to develop proportionate, reliable and scientifically-robust hazardous area classification guidance. Compared to the decades of research on flammable gases, the work on mists is still at an early stage. HSE is keen to collaborate with other organisations that share an interest in this topic going forward.

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MARY KAY O'CONNOR PROCESS SAFETY CENTER
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23rd Annual Process Safety International Symposium
October 20-21, 2020 | College Station, Texas

Sensing Dispersed Dust Concentration using Photograph

Kingsly Ambrose*, Y. Zhao and Z. Niu
Department of Agricultural and Biological Engineering
Purdue University
West Lafayette, Indiana 47907, USA
rambrose@purdue.edu

Abstract

Dust dispersion during powder handling and processing is of great concern for both workers' health and explosion risk. Dust emission locations in industries can vary during handling and processing, while dust concentration sensing would require the installation of an additional equipment in every location prone to dust generation. A method of using a digital camera or photograph to measure the dust concentration based on two target intensity value has been developed at Purdue University. The method was developed based on the relationship between the suspended dust concentration and extinction coefficient. Calibrated equations have been developed for cornstarch, grain dust, and sawdust. This method does not require any training and can be integrated with security system cameras and/or other independent imaging source.

Keywords: Dust dispersion, Dust sensing, MEC

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Creation of the *HBT*, a Large-Scale Facility at TAMU for Study of Detonations and Explosions

Elaine S. Oran*
Department of Aerospace Engineering
Texas A&M University
College Station, TX 77843
eoran@tamu.edu

Abstract

Unwanted explosions, expecially those that evolve into deflagrations or detonations, often have devastating consequences. They take lives, destroy homes and places of work, and many leave behind severe and persisting health and economic problems. The creation of the new shock and detonation tube facility at TAMU, *HBT*, is based on the principal that the more we understand about the the dynamics of explosion events and the controlling physics and chemistry of the fuels and other energetic materials, the better we can develop ways to avoid the event or at least to mitigate the damage. The HBT facility will consist of a large cylindrical channel (200 m x 20 m) equipped with an evolving suite of accompanying diagnostics. The channel will be made of thick steel so that it can withstand the strongest explosions, deflagrations, and detonations. The facility will be used to examine explosion properties of materials ranging from hydrogen or natural gas through to heavier hydrocarbons typical of petrochemicals, all in a range of initial conditions. It will also be used to study multiphase effects, ranging from dispersed small reactive or inert particles to larger-scale rubble. Another possible use for this facility is the study of ignition and flame acceleration properties of materials typically present in woodland fires and even materials processing by shock and detonation waves. This presentation will review the most recent state of development of HBT and present a plan for dealing with safety and noise issues.

Keywords: Explosions, vapor clouds, industrial explosions, detonations, deflagrations, DDT.

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Development of Flammable Dispersion Quantitative Property-Consequence Relationship (QPCR) Models Using Machine Learning

Zeren Jiao*, Yue Sun, Yizhi Hong, Pingfan Hu, Qingsheng Wang
Mary Kay O'Connor Process Safety Center
jiaozeren@tamu.edu

Abstract

Incidental release of flammable gases and liquids can lead to the formation of flammable vapor clouds. When their concentrations are above the lower flammable limit (LFL), or ½ LFL for conservative evaluation, fires and explosions can result with the presence of an ignition source. The objective of this work is to develop highly efficient consequence models to accurately predict the downwind maximum distance, minimum distance, and maximum vapor cloud width within the flammable limit. In this study, a novel quantitative property-consequence relationship (QPCR) model is proposed and constructed for the first time to accurately predict flammable dispersion consequences in a machine learning and data-driven manner. Flammable dispersion database consists of 450 leak scenarios of 41 flammable chemicals were constructed using PHAST simulations. A state-of-art machine learning regression method, extreme gradient boosting algorithm, was implemented to develop models. The coefficient of determination (R2) and root-mean-square error (RMSE) were calculated for statistical assessment and the developed QPCR models achieved satisfactory predictive capabilities. All the developed models have high accuracy, with the overall RMSE of three models being 0.0811, 0.0741, and 0.0964, respectively. The developed QPCR models can be used to obtain instant flammable dispersion estimations for novel flammable chemicals and mixtures at much lower computational costs.

Keywords: Consequence Modeling, Extreme Gradient Boosting, Flammable Dispersion, Machine Learning

1. Introduction

Incidental release of flammable materials may result in the formation of flammable vapor clouds. For areas in which flammable gas concentrations are above the lower flammability limit (LFL), fires and explosions will take place when encountering an ignition source, which can be highly hazardous to the process plant and nearby communities. The deadly hydrocarbon vapor cloud explosion happened in BP Texas City fifteen years ago caused fifteen fatalities and 180 others injuries, which truly showed the disastrous consequence of flammable dispersion and explosion (Holmstrom et al., 2006). Flammable dispersion consequence analysis plays a major role in the prevention and mitigation of fire and explosion incidents. In emergency response planning, it is also necessary to conduct consequence analysis for large-scale flammable chemical leaks. When assessing the consequences of flammable dispersion, the areas under the ½ LFL and LFL are critical criterion for determining the safe areas that have been identified in various flammable dispersion research works (Birch et al., 1989). Webber (2002) investigated the possibility of reducing the ½ LFL threshold, finding that the criterion should not be reduced and that the ½ LFL criterion should apply to instantaneous concentrations.

The consequences of flammable dispersion can be predicted using empirical, computational fluid dynamics (CFD) or integrated models. Typical empirical models for gas dispersion include the Pasquill-Gifford model and Britter-McQuaid model, which allow rapid predictions with acceptable accuracy (McQuaid, 1982). However, the Pasquill-Gifford or Gaussian dispersion models apply only to neutrally buoyant dispersions of gases. Furthermore, the Britter-McQuaid model of dense gas dispersion is unable to account for the effects of parameters such as release height, ground roughness, and wind speed profiles (Crowl and Louvar, 2019). CFD models, such as ANSYS Fluent, CFX, and FLACS, are capable of capturing the influence of surface roughness, but are time-consuming and come at significant computational costs (Li et al., 2020; Middha et al., 2010). This makes them particularly cumbersome in emergencies in which instant estimation should be available. Integral models such as HEGADIS, NCAR, and DRIFT, which take advantage of both empirical and CFD methods, have been widely used as a result of their higher prediction accuracies and lower computational costs (Gant et al., 2018). However, these models are limited to free-field dispersion with no obstructions and are generally not applicable to situations involving complex geometries (Dasgotra et al., 2018).

The consequence modeling package PHAST (Process Hazard Assessment Software Tool) is a popular consequence analysis and risk assessment tool that integrates dispersion models to examine the progress of potential incidents from initial release to far-field dispersion, including the modeling of rainout and subsequent vaporization (Witlox et al., 2014). PHAST dispersion simulation results have been widely validated against various experimental results, including both buyout and heavy gases, showing satisfactory agreement between simulated and experimental results (Witlox et al., 2014; Witlox et al., 2018).

Machine learning regression methods have shown significant capabilities for use in data mining and big data analysis in recent years (Jiao et al., 2020a; Shen et al., 2020). There have been increasing applications of machine learning algorithms for hazardous material ratings (Yuan et al., 2020; Jiao et al., 2020b), fire and explosion-related property prediction (Cao et al. 2018; Jiao et al., 2019a; Yuan et al., 2019), and consequence analysis (Sun et al., 2019; Jiao et al., 2020c). For current machine learning algorithm implementation in gas dispersion modeling, Wang et al. (2015) also used PHAST simulation results to validate a neural network-based real-time estimation of chlorine dispersion using gas detector data, illustrating the practicality of PHAST in validation of

proposed dispersion models. Gwak and Rho (2019) compared three different machine learning techniques in predicting CO2 dispersion in a lab environment. However, these works only examined the techniques using one or two chemicals (chlorine, carbon dioxide, or sulfur dioxide) under limited leaking conditions, which is not sufficient to construct a wide spectrum applicable model for real-world applications.

One of the major challenges to be addressed involves the development of a rapid, universal applicable prediction model that is based on leaking conditions and specific properties of flammable chemicals with the dispersion distance data. However, the relationship is deemed highly non-linear and the interaction mechanism remains unknown. Machine learning-based quantitative structure-property relationship (QSPR) analysis is used widely for fire and explosion related property predictions, including flammability limits, autoignition temperature, and flashpoint, which shows higher accuracy and reliability compared with other prediction methods. QSPR can reveal mathematical relationships between the structural attributes and the property of interest at a quantum chemistry level, which can serve to bridge the gap between micro quantum structure descriptors and relatively macro properties. Furthermore, machine learning algorithms can also overcome the high non-linearity between input features and output variables, which make machine learning-based quantitative prediction models suitable for forming linkages from leaking conditions and chemical properties to dispersion consequences.

In order to develop a robust predictive tool for fast flammable dispersion consequence analysis for a wide range of flammable chemicals and leak conditions, machine learning-based quantitative property-consequence relationship (QPCR) models should be developed to better assist in consequence analysis, risk assessment, and emergency response planning. In this study, PHAST flammable dispersion simulations of 450 different leak scenarios were conducted involving 41 flammable chemicals commonly present in the chemical, oil and gas industries. The three key flammable dispersion parameters, which are maximum downwind distance, minimum downwind distance, and maximum vapor cloud width within LFL and ½ LFL criteria, were obtained from the simulation to construct a comprehensive database with nearly 60,000 data points. State-of-art machine learning technique, gradient boosting regression (GBR), was implemented using the Xgboost package in R to correlate the property descriptor with designated dispersion distances to construct the flammable dispersion QPCR models.

2. Methodology

2.1 Database

Database compilation is the first step for big data analysis. In this study, flammable dispersion consequence database was generated using PHAST simulation. The leak condition parameters consist of several components: source condition (release material, location, quantity, etc.), weather condition (wind speed, atmospheric temperature, humidity, etc.), and leak condition (leak size). Source conditions are determined based on specific petroleum process operating conditions. 41 flammable chemicals with 450 leak scenarios with both under ½ LFL and LFL criteria were simulated to construct a database with a total of 19,579 valid flammable dispersion scenarios since some scenarios did not result in the generation of the flammable cloud. Each simulation result contains three key parameters of flammable dispersion: maximum downwind distance, minimum downwind distance, and maximum vapor cloud width. Among 19,579 scenarios that were employed in the QPCR model development, 75% of data points (14,684 scenarios) were randomly

selected as the training set and the remaining 25% (4,895 scenarios) are grouped into the test set to validate the accuracy of the model. Using a single data source for model development can ensure the model's prediction consistency and accuracy.

2.2 Property Descriptors

To construct a quantitative relationship model, finding suitable descriptors as input variables is a crucial step to ensure the QPCR model's accuracy and practicality. In this study, seven different property descriptors were used as prediction features to develop QPCR models by considering different aspects that can influence flammable dispersion. Property descriptors are categorized into three groups: source property, criteria property, and physical property. Source and criteria properties are the same input variables used in PHAST simulation. Beyond that, vapor density is chosen along with lower flammability limits as physical properties since it the most influential factor in flammable dispersion (Crowl and Louvar, 2019). By keeping the descriptor list as short as possible, the developed QPCR model can avoid overfitting caused by unnecessary input variables.

2.3 Gradient Boosting Regression

Gradient boosting regression (GBR) is a prediction model in the form of an ensemble of weak prediction models, typically decision trees. It builds the model in a stage-wise fashion as other boosting methods do, and it generalizes them by allowing optimization of an arbitrary differentiable loss function (Nielsen, 2016). The idea of gradient boosting originated from the observation by Leo Breiman that boosting can be interpreted as an optimization algorithm on a suitable cost function (Breiman, 1997). Explicit regression gradient boosting algorithms were subsequently developed by Jerome H. Friedman (Friedman, 2002). Since then, the developed gradient boosting (also known as boosting tree) algorithm has been widely used in chemical engineering and safety research which shows superior performance compared with other machine learning regression methods such as random forest, support vector machine (SVM), etc.

3. Results and Discussion

3.1 Maximum Distance

In order to visualize the performance of the developed QPCR models, two types of performance evaluation plots are shown in Fig. 1. The plot of QPCR prediction values vs. actual values is shown in Fig. 1a and the prediction residual plot is shown in Fig. 1b with the test set statistical values as R^2=0.9838 and *RMSE*=0.1556. All data points are evenly distributed along the diagonal baseline of Fig. 1a which indicates that the predicted distance is very close to the actual distance. Furthermore, the majority of data residuals shown in Fig. 1b are between ±0.25 and very close to the zero baseline, which proves that the developed QPCR method provides a good estimate of flammable dispersion maximum downwind distance.

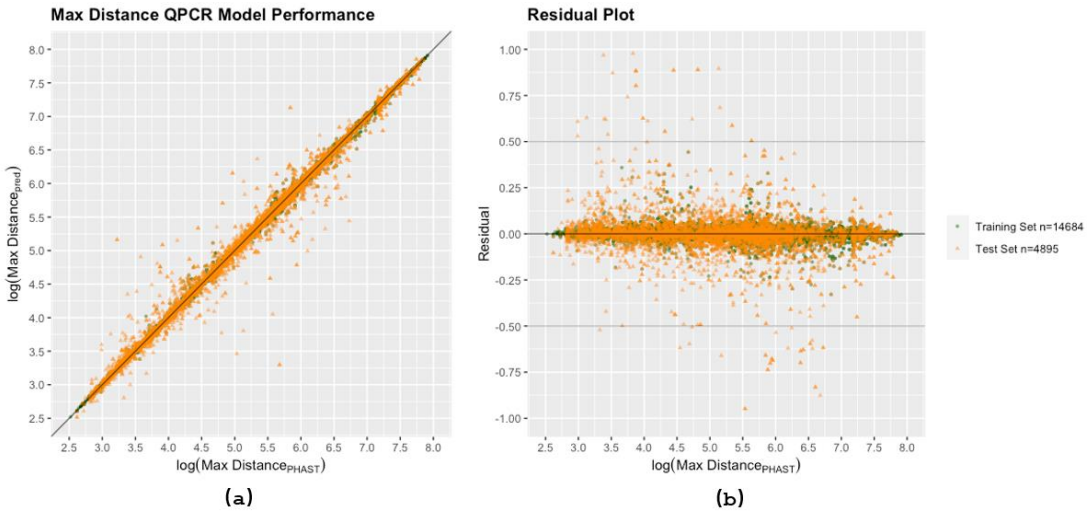

Fig. 1. Maximum Distance QPCR Model Performance Plot: (a) Predicted Value vs. PHAST Simulation Value, (b) Residual Plot

3.2 Minimum Distance

The plot of QPCR prediction values vs. actual values is shown in Fig. 2a, with all data points evenly distributed along the diagonal baseline. Since the minimum distance values are relatively lower than those of the maximum distance, the prediction residual plot shown in Fig. 2b shows that the minimum distance prediction errors are more concentrated around the zero baseline compared with the maximum distance model. The test set statistical values of R^2=0.9837 and *RMSE*=0.1466 also indicates that the minimum distance model performance is slightly better than the maximum distance model.

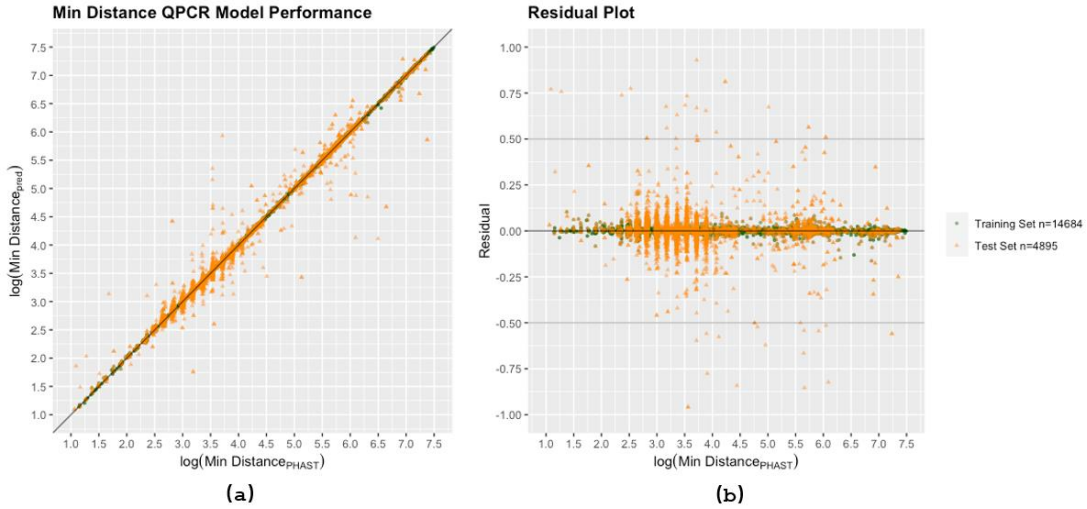

Fig. 9. Minimum Distance QPCR Model Performance Plot: (a) Predicted Value vs. PHAST Simulation Value, (b) Residual Plot

3.3 Maximum Width

The plot of QPCR prediction values vs. actual values is shown in Fig. 3a, since the maximum width data is the least normally distributed data according to the histogram. The test set data points are sparsely distributed along the diagonal baseline compared to the maximum and minimum distance models. The prediction residual plot in Fig. 3b shows that more points are located between 0.25 to 0.50 and -0.25 to -0.50 compared to the previous two models. However, the test set statistical values of R^2=0.9869 and *RMSE*=0.1973 demonstrate that the prediction of the maximum width QPCR model still has satisfactory accuracy.

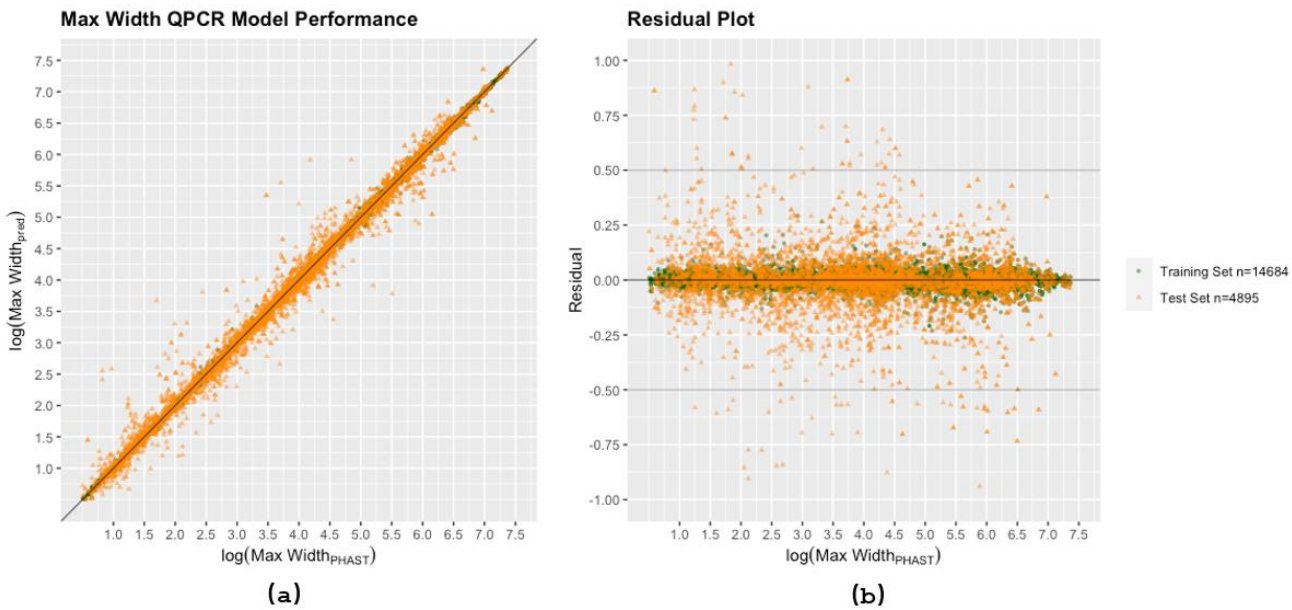

Fig. 3. Maximum Width QPCR Model Performance Plot: (a) Predicted Value vs. PHAST Simulation Value, (b) Residual Plot

3.2.3 Statistical Evaluation

The statistical assessment value of the training set, test set, and overall dataset of three QPCR models is summarized in Table 1 and the model predicted result can be found in Supplementary Table associated with this paper. All three models have very high accuracy in predicting flammable dispersion downwind distance, with the dataset R^2 being higher than 0.995. The fact that the training set R^2 for each of these three models is higher than 0.999 illustrates the power of gradient boosting in detecting small details within the dataset training machine learning prediction models. The independent test set validation also proved the model's superior performance of the flammable dispersion QPCR model, with the test set RMSE lower than 0.2.

Table 1

Statistical Assessment Values of Developed Flammable Dispersion QPCR Models

Method	Training Set	Test Set	Overall

	R^2	$RMSE$	R^2	$RMSE$	R^2	$RMSE$
Maximum Distance	0.9995	0.0267	0.9838	0.1556	0.9956	0.0811
Minimum Distance	0.9999	0.0124	0.9837	0.1466	0.9958	0.0741
Maximum Width	0.9997	0.0265	0.9869	0.1873	0.9966	0.0964

4. Conclusions

In this study, a database was constructed with nearly 20,000 dispersion simulation of 41 flammable chemicals using PHAST, and the consequence analysis results were used to construct and validate the QPCR models. The GBR method was employed to provide a reliable prediction for flammable dispersion downwind distances. The GBR-based QPCR models showed significantly high accuracy for prediction of dispersion downwind distances, with the test set RMSE for maximum distance, minimum distance, and maximum width having values of 0.1556, 0.1466, and 0.1873, respectively. These prediction models illustrate the power of QPCR and machine learning for assisting with consequence analysis and emergency response planning with much higher efficiency and accuracy.

The GBR-based QPCR models do not have specific equations for intuitive applications, such as empirical methods, which is one of the shortcomings of QPCR method. However, an increasing trend in the availability and implementation of digital equipment in consequence analysis and emergency response planning has occurred. Thus, the problem can be overcome by developing a built-in software package that provides instant predictions of hazardous areas with very high accuracy and reliability. The models also showed potential for joint applications with QSPR models of LFL and vapor density so these models can be expanded to other novel chemicals without measured properties. Furthermore, the database must be further expanded to include the detailed influences of weather and environmental conditions including weather categories, wind speeds, and ground temperatures. Additionally, the method to quantify the influences of obstacles must be included so as to allow for more universal applications of the QPCR models.

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A novel liquid in-cylinder combustion risk criterion based on unsupervised clustering algorithms

Chenxi Ji[1,2,3,4], Zeren Jiao[1,2], Shuai Yuan[1,2], Mahmoud M. El-Halwagi [1,2,4*]
1The Mary Kay O'Connor Process Safety Center
2Artie McFerrin Department of Chemical Engineering, Texas A & M University
3College of Marine Navigation, Dalian Maritime University
4 The Gas & Fuels Research Center, Texas A & M University
chenxiji@tamu.edu

Abstract

When shipping fuel selection problem is on the table, the safety factors should always be top prioritized. Currently, the liquid flammability classification mainly rely on flash point and the risk criteria are largely dependent on the two-dimensional matrix of consequence and probability. However, the liquefied marine fuel combustion has its own uniqueness, leading to a less consistent with the common classification standard.

This paper is aiming at providing a more reasonable criterion to classify flammable liquids in the compression ignition engines for further application on safety evaluation of promising marine fuel options. Besides the widely recognized liquid flammability characteristics, this study identifies contributors for in-cylinder flame propagation and the liquid aerosol formulation as well. Then two unsupervised machine learning clustering algorithms, k-means and spectral clustering, are employed to find the specific patterns of the three safety features for the collected liquid organic compounds database. To consider both cluster cohesion and separation, the global mean silhouette value is presented to find the optimal number of clusters and to evaluate the clustering performance of the proposed models. The results agree that the spectral clustering outperforms k-means clustering algorithm on classifying the risk ratings of liquid flammability, flame propagation and aerosol formulation. Moreover, the principal component analysis and the star coordinate diagrams are presented to visualize high dimensional data to two dimensional graphs. Finally, the overall liquid safety performance is evaluated by a novel rating system, liquid in-cylinder combustion risk index (LICRI) via the weight values determined by the information entropy approach.

Keywords: Marine fuel safety; In-cylinder flame propagation; Liquid aerosolization; Spectral clustering algorithm; High dimensional data visualization

1. Introduction

Safety is the top priority for the promising marine fuel selection. Flammability and explosive hazards, which have been well studied, are the major concerns of safety aspects of the tank to propeller (TTP) process aboard ships. The inherent flammable properties of liquids may involve flash point, auto ignition point, upper/lower flammability limit and boiling point. Currently, liquid combustion level is commonly determined by flash point. NFPA 704 (National Fire Protection Association, 2017), the widely recognized liquid flammability classification standard, categorizes liquids into five classes, and Figure 1 shows the NFPA 704 standard and flammability ratings of the promising marine fuels by adopting the NFPA fire diamond.

(a)

Rating	Criteria
0	Materials will not burn in air when exposed to a temperature of 1500°F for a period of 5 minutes
1	Flash point at or above 200°F
2	Flash point between 100 and 200°F
3	Flash point between 73 and 100°F
4	Flash point below 73°F

(b)

Properties	Fuels							
	HFO/MGO	Biodiesel	LPG	LNG	MeOH	EtOH	LH_2	Ammonia
US NFPA 704	2 1 0	1 1 0	4 2 0 SA	4 3 0 SA	3 1 0	3 2 0	4 0 0 SA	1 3 0
Health	1	1	2	3	1	2	0	3
Flammability	2	1	4	4	3	3	4	1
Instability	0	0	0	0	0	0	0	0
Special			SA: asphyxant gas	SA: asphyxant gas			SA: asphyxant gas	

Figure 1. NFPA 704 liquid flammability rating standard (1a); NFPA diamond-based flammability classification for promising marine fuels (1b)

Can we conclude that LNG, LPG and liquefied hydrogen share the most hazardous flammability feature? Maybe not. As many concerns were risen on the hazards of high flash point liquid because of the aerosolization or atomization phenomenon, the study published by Bowen and Shirvill (Bowen and Shirvill, 1994) highlighted the liquid aerosolization hazard could be minimized by the adopting the minimum practicable pressure for operating systems. However, the aerosolization effect of fuel oils was still underestimated, leading it to be the root cause of many incidents in shipping industry (Kohlbrand, 1991; Santon, 2009). Therefore, the flash point driven liquid flammability standard may be too simple to classify the safety level of marine fuel options, especially when considering the common combustion scenario of marine fuels.

Most ships employ 2-stroke diesel compression-ignition (CI) engine as their main power-driven source (Klett et al., 2017). Unlike Otto cycle, the diesel internal combustion engine uses a higher compression ratio, 15 to 20, to ignite the marine fuel (Sivaganesan and Chandrasekaran, 2016). The flame in the cylinder of the CI engine is initially propagated as laminar and later it becomes turbulent. Besides, there is a common operation for the combustion of heavy fuel oil, which needs to be heated to bring the viscosity below 20cst

for achieving proper aerosolization (MAN Diesel & Turbo, 2014). Date back to 1955, Eichhorn (Eichhorn, 1955) firstly presented the aerosols was able to lead an explosion and pointed out as well that the liquid aerosol flammability with fuzzy boundaries was completely different from vapor flammability limits. Moreover, we experimentally confirmed that the n-dodecane in the aerosol state can be ignited lower than the flash point. As shown in Fig. 2, there is a pressure rising when the equivalence ratio decreases to 0.1, but neither n-octane nor n-dodecane is supposed to be ignited since the LFL for both n-octane and n-dodecane vapor are 0.57% and 0.54% respectively, , illustrating the liquid aerosol has a wider flammability range than the bulk liquid (Yuan et al., 2019).

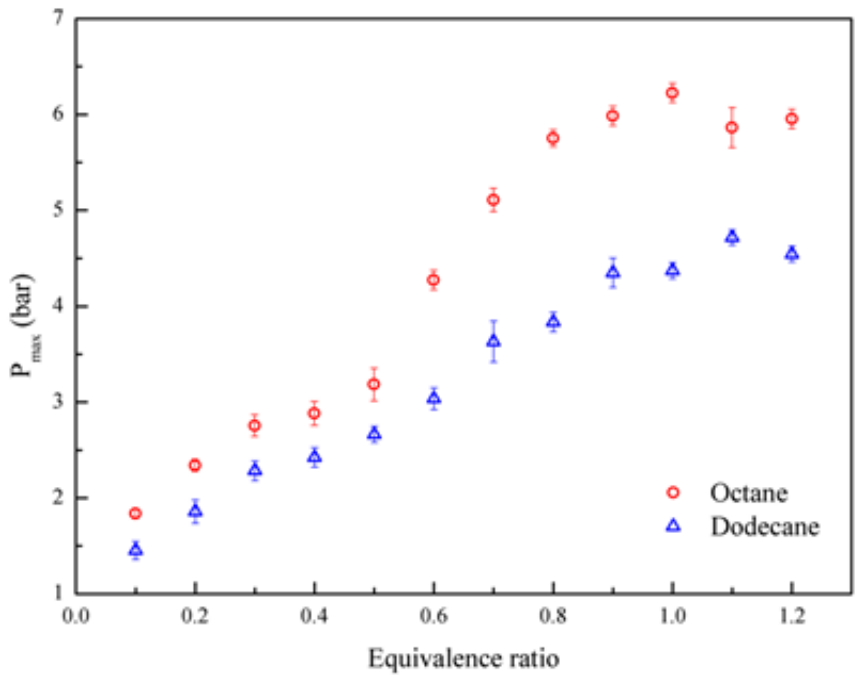


Figure. 2 P_{max} of n-octane aerosol and n-dodecane aerosol explosions for different equivalence ratios (Yuan et al., 2019)

However, the liquid aerosolization and flame propagation, making bulk liquids more hazardous on combustion and explosion in the cylinder of CI engine, have not widely recognized in industry or academia, thus, it is necessary to take the liquid flammability, flame propagation and liquid aerosolization effects into consideration to tell the safety extent of the liquefied fuel options in the CI engine. To fill in gaps for categorizing promising shipping fuels from the perspective of chemical safety and process system engineering, there are two steps to carry on the study of TTP process safety: the first step is to find optimal models for the identified contributors of liquid aerosol formulation and the next step is to adopt the clustering and classification approaches via unsupervised machine learning (ML) algorithms to classify the safety level of promising marine fuels.

In this study, the major inherent properties of the liquid flammability, flame propagation and aerosol formulation will be identified firstly, then two ML clustering algorithms will be executed to classify the collected database to different groups, and a new flammability rating called liquid in-cylinder combustion risk index (LICRI) will be calculated to show the overall liquid safety preference, which can be applied as a reasonable reference when considering the marine fuel selection issue in the TTP process.

2. Identification of liquid in-cylinder combustion contributors

The field of fluid flammability characteristics have been well studied while few works has focused on the inherent property identification for liquid in-cylinder flame propagation and liquid aerosolization. Thus, it is critical to identify the leading factors for liquid fuel aerosolization and flame propagation effects so that a reasonable liquified fuel safety criterion in CI engines may be established accordingly.

This work adopts AIT, FP and flammability range (FR), the range between lower flammability and upper flammability, as the contributors for liquid flammability matrix. Since the liquid in-cylinder flame has a combination feature of both premixed laminar and turbulent, the theoretical models of these two flames are analyzed to identify the significant parameters.

The well known "Two zones" model proposed by Mallard & Le Chatelier (Mallard and Le Chatelier, 1883) is:

$$S_l = \sqrt{\alpha \dot{\omega} (\frac{T_b - T_i}{T_i - T_u})} \tag{1}$$

$$\alpha = \frac{k}{\rho C_p} \tag{2}$$

α is the thermal diffusivity, $\dot{\omega}$ is the reaction rate, k is the thermal conductivity, C_p is the specific heat capacity and ρ is the density. The relationship, shown in Equation 8, between laminar flame and turbulent flame speed presented by Peters (Peters, 2000) has widely accepted and shown a good performance.

$$S_T = S_l + \mu' \left\{ -\frac{a_4 b_3^2}{2b_1} Da + [\left(\frac{a_4 b_3^2}{2b_1} Da\right)^2 + a_4 b_3^2 Da]^{1/2} \right\}, \text{ where } Da = \frac{S_L l}{\mu' \delta_L} \tag{3}$$

As the above equation shows, μ' is the turbulence intensity; b_1, b_4 and a_4 are the turbulence modeling constants with value of 2.0, 1.0 and 0.78; Da is the Damkohler number and l is the turbumence integral length scale, δ_L that denotes the flame thickness is a function of heat capacity, heat conductivity, density and laminar flame speed. Hence, both of the laminar and turbulent flame equations pointed out the dependent varibles of flame propagation for CI engines are heat capacity (HC), liquid density (LD) and

liquid thermal conductivity (LTC), and these three variables construct our liquid in-cylinder flame propagation matrix.

Many literatures (Ballal and Lefebvre, 1979; Danis et al., 1988; Kiran Krishna et al., 2003; Polymeropoulos, 1984; Yuan et al., 2020, 2019) have pointed out the key parameter to determine liquid aerosolization is the droplet size. Among all the theoretical mean diameters of aerosols, the Sauter Mean Diameter (SMD) is the most common one to apply for heat transfer, combustion and dispersion modelling (K. Krishna et al., 2003). In addition, the diesel engine fuel injector can be deemed as an electro spray type of aerosol generator. Most studies conducted on pressure atomizers have focused on the type of injector used in compression ignition engines (Lefebvre and McDonell, 2017). Two SMD formulae proposed by Harmon (Harmon, 1955) and Elkotb (Elkotb, 1982) for plain-orifice type pressure atomizers are listed as below:

$$\mathrm{SMD} = 3300 d_o^{0.3} \sigma^{-0.15} \rho_L^{-0.648} \mu_L^{0.07} U_L^{-0.55} \rho_G^{-0.052} \mu_G^{0.78} \tag{4}$$

where d_o is the discharge orifice diameter, μ_L is the liquid kinematic viscosity, σ is the surface tension, U_L is the liquid flow rate and ρ_L is the liquid density.

$$D_{3,2} = 3.08(\sigma\rho_L)^{0.737} v_L^{0.385} \Delta P_L^{-0.54} \rho_A^{0.06} \tag{5}$$

From the above equation, σ is the surface tension, v_L is the liquid dynamic viscosity, ρ_L is the liquid density, ΔP_L is the liquid pressure differential and ρ_A is the air density.

Liquid dynamic viscosity (LDV) and surface tension (ST), as the inherent properties of fuels, are the determinant parameters for the droplet size of liquid aerosols based on the above two equations. For most practical fuels, any change in dynamic viscosity is always accompanied by a change in volatility, and Ballal and Lefebvre (Ballal and Lefebvre, 1979) indicated as well the quenching distance was dependent on fuel volatility. Besides dynamic viscosity, liquid vapor pressure (LVP) is an evidential index to tell the volatility level of liquids. Therefore, the identified contributors for liquid aerosolization are surface tension, liquid dynamic viscosity and liquid vapor pressure.

3. Liquid in-cylinder combustion safety clustering approaches

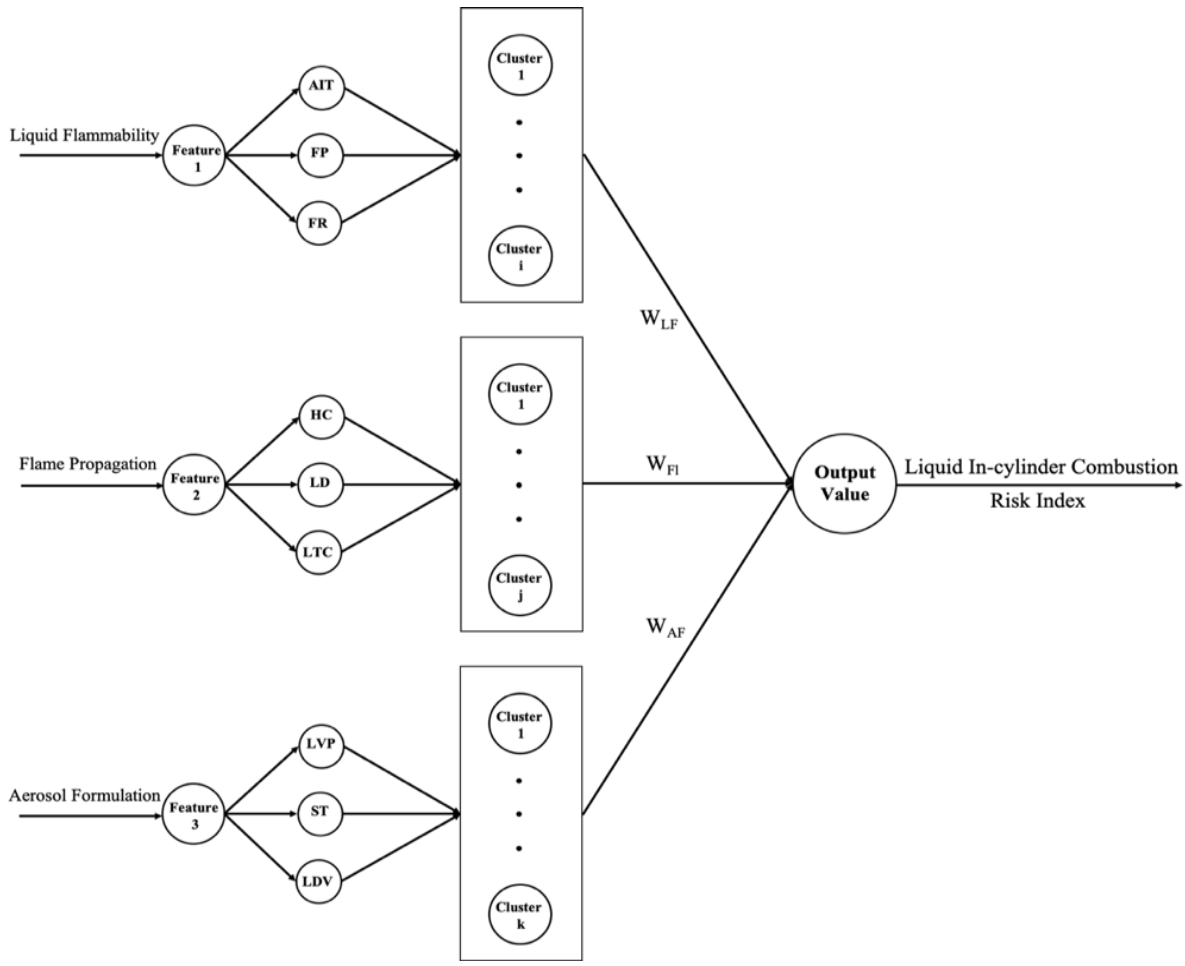


Figure 3. Network of liquid in-cylinder combustion risk index

By integrating the liquid aerosol formulation contributors with liquid flammability and flame propagation, the database of our liquid in-cylinder combustion criterion is built with three evaluation matrix and nine contributors. This work employs two unsupervised clustering algorithms, the network of which is shown in the above figure to categorize the risk rating of liquid flammability, flame propagation and aerosol formulation, then a liquid in-cylinder combustion risk index (LICRI) is presented to tell the overall liquid combustion safety ratings.

$$\text{LICRI} = \sum_{i=1,j\in\{1,2,\ldots,n\}}^{3} W_i^{C_{i,j}} \tag{6}$$

As shown in the above equation, the weight value W_i for the three safety matrices should be determined and normalized before implementing the ML clustering algorithms. The range of LICRI is between 0 to 1, as the values of cluster numbers $C_{i,j}$ are increasing, the less value of LICRI value for one substance would be, illustrating its high risk for liquid in-cylinder combustion. In this study, the DIPPR 801 database ("DIPPR Project 801 - Full Version - Physical Constants - Knovel," n.d.) are preprocessed to collect 703

effective organic compounds in the liquid state under specific temperatures with values on the nine dimensional data, *i.e.*, AIT, FP, FR, HC, LD, LTC, LVP, ST and LDV.

3.1 Information entropy approach

Statistically, information entropy can be referred as the expectation of the amount of information contained in an event. Thus, the entropy value is a useful tool to tell the degree of dispersion. The smaller the entropy value, the greater the degree of dispersion of the indicator, and the greater the influence, *i.e.*, weight vector, of the indicator on the comprehensive evaluation. Compared with other weight value determination approaches, the information entropy method, as an objective approach, has an outstanding capability to distinguish indicators, and it always brings high credibility and precision to avoid subject weight determination (Li et al., 2011). The typical information entropy procedures are summarized as:

- Step 1 Determination of the evaluation matrix

Herein, three parameters within total nine indicators construct the evaluation matrix.

- Step 2 Normalization of the evaluation matrix

This study employs the critical value approach to normalize the evaluation matrix, and the normalized indicators are calculated by:

$$x'_{ij} = \frac{x_{ij} - \min x_j}{\max x_j - \min x_j} \ , \ x'_{ij} = \frac{\max x_j - x_{ij}}{\max x_j - \min x_j} \tag{7}$$

- Step 3 Calculation of information entropy

$$e_j = \frac{-\sum_{i=1}^{m} \frac{x'_{ij}}{\sum_{i=1}^{m} x'_{ij}} \cdot \ln\left(\frac{x'_{ij}}{\sum_{i=1}^{m} x'_{ij}}\right)}{\ln m} \tag{8}$$

- Step 4 Calculation of weight vectors

$$W_j = \frac{1 - e_j}{\sum_j (1 - e_j)} \tag{9}$$

Therefore, the LICRI can be updated after the weight vectors determined, please check supporting information for calculation details.

$$\text{LICRI} = 0.480^{Cluster_i} + 0.323^{Cluster_j} + 0.197^{Cluster_k} \tag{10}$$

3.2 K-means clustering

K-means clustering is based on the distance between objects and centroids with the actual observations as the input. The identified cluster shape is assumed as spheroidal with equal diagonal covariance. As a traditional clustering algorithm, the core idea of k-means clustering approach is to minimize the total

with-in cluster variation, as shown in Equation 31. Firstly, initial cluster assignment is proceeded by randomly assigning a number from 1 to K; then the cluster centroid for each datum of the K clusters will be calculated and each data point will be distributed to the closest Euclidean distance (Likas et al., 2003). The centroids calculation for each cluster will be repeated until the clustering assignments complete, and the objective function of K-means algorithm is shown below:

$$\min C_1, \dots, C_k \left\{ \sum_{k=1}^{K} \frac{1}{|C_k|} \sum_{i,i' \in C_k} \sum_{j=1}^{p} (x_{ij} - x'_{ij})^2 \right\} \tag{11}$$

Where $C_1, \dots, C_k$ denote cluster 1 to k, $|C_k|$ is the number of samples in the kth cluster, p is the number of predictors, and $\sum_{j=1}^{p}(x_{ij} - x'_{ij})^2$ represents the Euclidean distance between two observations in the kth cluster. The study applies Python package Scikit-learn (Pedregosa et al., 2011) to process the K-means cluster algorithm, and the silhouette analysis (scikit-learn, 2017) is adopted to determine the number of clusters.

3.3 Spectral clustering

In contrast to the traditional K-means algorithm, spectral clustering is more adaptable to data distribution with excellent clustering effect and less computational cost. Spectral clustering is an algorithm that evolved from graph theory. The main idea is to treat all data as points in space, and these points can be connected by edges. The edge weight value between two points farther away is lower, and the weight value between two points closer is higher. With the eigenvectors of matrices as the input algorithm, spectral clustering adopts graph distance geometry and arbitrarily identified cluster shape. By cutting the graph composed of all data points, the difference after cutting the sum of the edge weights between the subgraphs is low, while the sum of the edge weights in the subgraphs is quite high, so the purpose of clustering can be achieved (Bürk, 2012; Luxburg, 2006). Referred from Ng's work (Ng et al., 2002), the normalized spectral clustering algorithm is formulated as below:

- Step 1 Split the LICRI database to three data sets and normalize each data set;
- Step 2 Construct similarity graph with adjacency matrix by normal k-nearest neighbor approach by setting number of neighbors of 15 and Sigma value of 1;
- Step 3 Compute the normalized graph Laplacian L and its first eigenvectors $v_1, \dots, v_k$;
- Step 4 Set $V \in \mathbb{R}^{n \times k}$ as the matrix containing the vectors $v_1, \dots, v_k$ and formulate the matrix $U \in \mathbb{R}^{n \times k}$ by normalizing the matrix V;

$$u_{ij} = \frac{v_{ij}}{\sqrt{\sum_k v_{ik}^2}} \tag{12}$$

- Step 5 Let $y_i \in \mathbb{R}^k, i \in \{1,2,\dots,n\}$ be the vector corresponding the i-th row of matrix U.
- Step 6 Cluster the points $(y_i)_{i=1,\dots,n}$ with the k-means algorithm into clusters $C_1,\dots,C_k$, as shown in previous chapter.

The spectral clustering algorithm is implemented with the help of *Matlab* statistics and machine learning toolbox and the fast and efficient spectral clustering package (Ingo, 2020). In contrast to convex data set shape of k-means algorithm, spectral clustering tends to be useful for hard non-convex problems (Hocking et al., 2011).

3.4 Clustered data visualization

Data visualization is another obstacle to show the model performance since the LICRI database have a three-dimensional feature. One available technique as previously discussed is to utilize principal component analysis (PCA) to reduce the dimension of the data sets, and this work employs PCA to automatically consider weight values of principal components and to visualize k-means cluster models; while another method is to build star coordinates (SC), converting high-dimensional database to 2 dimensional coordinate.

Basically, the SC system is a curvilinear coordinate system. By defining the origin as a 2d point $O_n(x,y) = (o_x, o_y)$ and a series of n 2d vectors $A_n = \langle \overrightarrow{a_1}, \overrightarrow{a_2}, \dots, \overrightarrow{a_i}, \dots, \overrightarrow{a_n} \rangle$, the axes can be established and mapped to the Cartesian Coordinates (Kandogan, 2000). The data points D_j from a high dimensional dataset D are converted to data points D_j' of the established 2d Cartesian Coordinates by the sum of all unit vectors $\overrightarrow{u_i} = (u_{x_i}, u_{y_i})$ on each coordinate, and the relationship between the original and converted data points are shown below:

$$D_j'(x,y) = [o_x + \sum_{i=1}^{n} u_{x_i} \cdot (d_{ji} - min_i),\ o_y + \sum_{i=1}^{n} u_{y_i} \cdot (d_{ji} - min_i)\ ] \qquad (13)$$

where $D_j = (d_{j0},\ d_{j1}, \dots,\ d_{ji}, \dots,\ d_{jn}),\ |\overrightarrow{u_i}| = \frac{|\overrightarrow{a_i}|}{max_i - min_i}$

$$max_i = \max\{d_{ji}, 0 \le j < |D|\}, min_i = \min\{d_{ji}, 0 \le j < |D|\}$$

Moreover, the cluster projection diagram of any two response variables can be applied to find the optimal clustering model as well. The star coordinates and cluster projection diagram are integrated with the spectral clustering algorithm to visualize the clustered data of the LICRI database. The sihouette plot, which has been widely applied to show the optimal number of clusters for unsupervised algorithms, is employed to find the better clustering models among three safety features between k-means and spectral clustering algorithms.

3.5 Cluster validation criterion

In order to evaluate which algorithm has a better clustering performance, cluster validation criteria are introduced in the work. Focusing on measuring the fit of a clustering structure itself, the study employs the internal validation indices to consider both cluster cohesion and cluster separation. Three common internal measures of cluster validation have been surveyed, including the Dunn index (Dunn, 1974), Davies-Bouldin index (Davies and Bouldin, 1979) and Silhouette index (Rousseeuw, 1987). All of the three indices have presented as robust strategies to predict the optimal clustering partitions. This work utilizes the Silhouette index as the cluster validation criterion as a result of its interpretation and validation of consistency within clusters of the liquid in-cylinder combustion database. The silhouette validation criterion applies a concise graphical representation, the silhouette plot, to display how well each data points have been clustered. Similar with the Dunn index, the higher the silhouette index is, the better the clustering performance would be. For the data point j in the cluster C_j, the mean distance between I and other data point within the same cluster is defined as:

$$a(j) = \frac{1}{|C_j|-1} \sum_{k \in C_j, j \neq k} d(j,k) \tag{14}$$

where $d(j,k)$ is the distance between two data points j and k within the cluster C_j. The reason of adding the item $\frac{1}{|C_j|-1}$ is because the distance $d(j,j)$ is excluded in the sum.

Then the distance of j to the points in some cluster C_l ($C_l \neq C_j$) other than C_j is defined as:

$$b(j) = \min_{l \neq j} \frac{1}{|C_l|} \sum_{k \in C_l} d(j,k) \tag{15}$$

Next, the silhouette value of one data point j can be expressed as:

$$s(j) = \begin{bmatrix} 1 - a(j)/b(j), & if\ a(j) < b(j) \\ 0, & if\ a(j) = b(j) \\ b(j)/a(j) - 1 & if\ a(j) > b(j) \end{bmatrix} \tag{16}$$

From the above expression, one can find the range of $s(j)$ is between -1 and 1: a vale close to 1 indicates the data point is clustered to the correct cluster whereas the value -1 tells the data point is affected to the wrong cluster. This study applies the mean silhouette value to show the performance for a given cluster C_l, which is denoted as $\overline{s_l}$:

$$\overline{s_l} = \frac{1}{|C_l|} \sum_{j \in C_l} s(j) \tag{17}$$

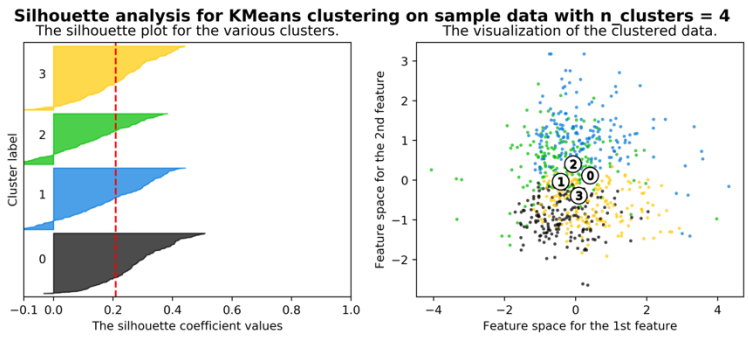
Finally, the overall performance of one specific model is able to be evaluate by the global silhouette index, the mean of the average silhouette values through all the clusters with the cluster number L:

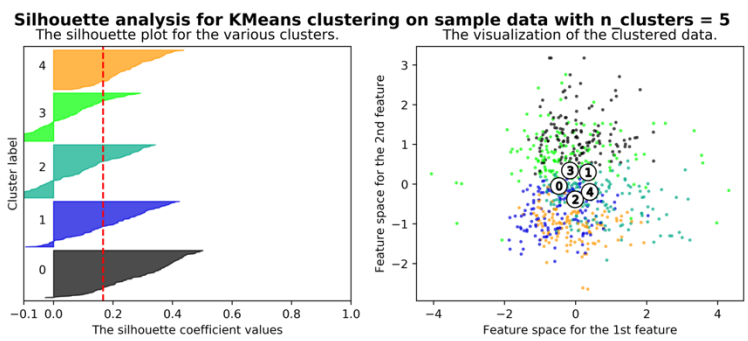
$$\bar{S} = \frac{1}{L}\sum_{L=1}^{L} \bar{s_l} \tag{18}$$

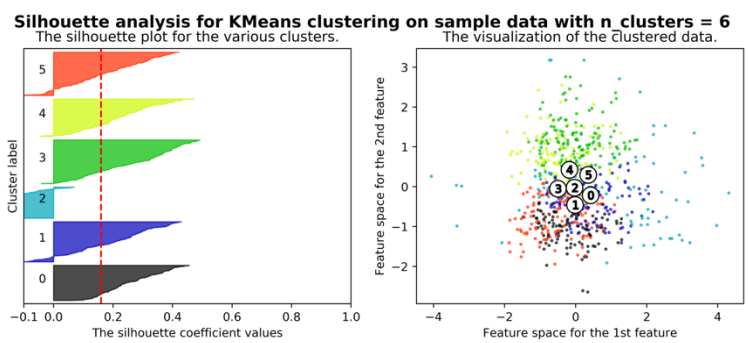
4. Results and discussion

This study employs silhouette analysis to study the separation distance between the final clusters of k-means and spectral clustering algorithms. Also, the silhouette value is adopted to determine the optimal numbers of clusters for the LICRI database. The performance of the unsupervised clustering models is evaluated with the visualization of the clustered data and the sihouette plot.

4.1 K-means clustering







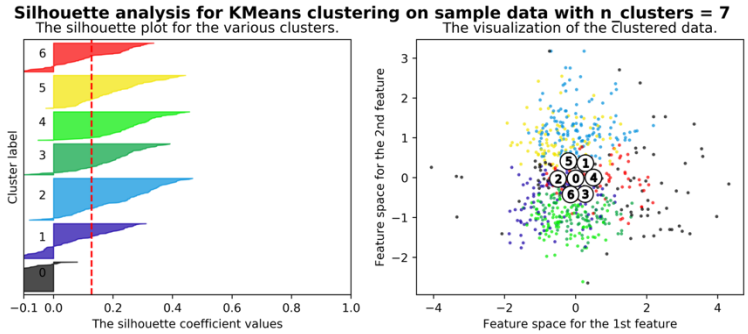


Figure 4. Silhouette plots and the n cluster labelled scatter plots for liquid flammability indicators by integrating PCA ($n \in \{4, 5, 6, 7\}$)

As shown in Figure 4, the silhouette plots of the k-means clustering show the silhouette coefficient values of n clusters for the three indicators of liquid flammability. The n clustering value of 6 and 7 are bad picks because each of them has a negative dominated cluster, indicating those samples might have been assigned to the wrong cluster. The right figure is the n cluster labelled scatter plots, and its horizontal axis is PCA 1 and the vertical axis is PCA 2. The 4 cluster model seems to have less overlap points than others, and this statement is suppoted by the highest average silhouette value, 0.209, locating in 4 cluster model and its narrow fluctuations in the size of the silhouette plots. These bring to the conclusion that the 4 cluster model is the optimal clustering model. However, there are many data points locating in the unclear clusters from the 4 cluster labelled scatter plot.

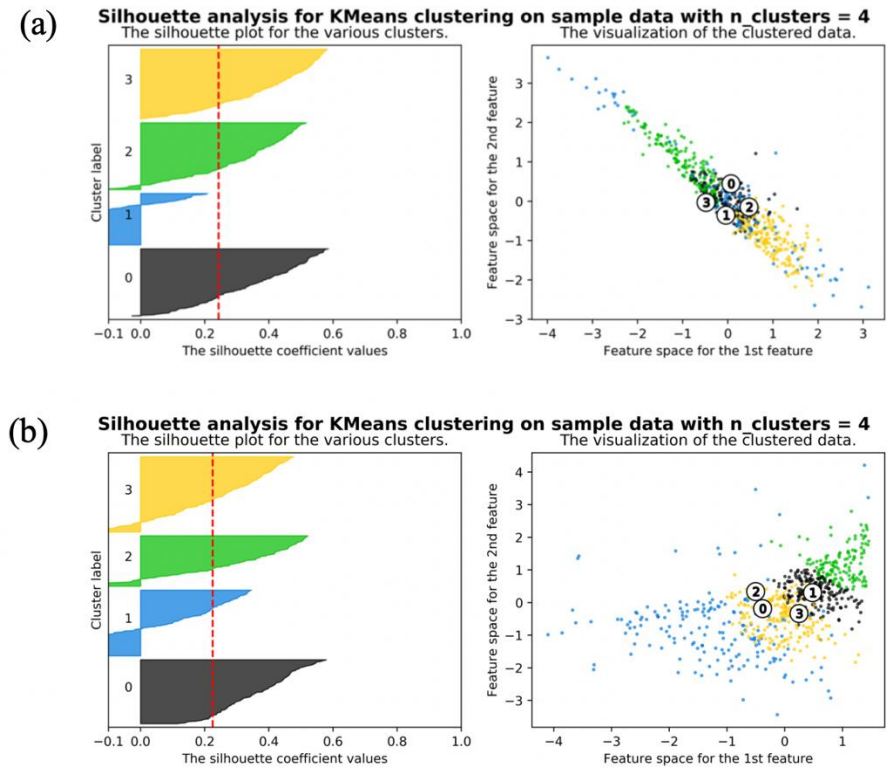


Figure 5. Silhouette plots and the n cluster labelled scatter plots for optimal flame propagation model (5a) and optimal liquid aerosolization model (5b)

Similarily, the study analyzed the silhouette plots and the n cluster labelled scatter plots for flame propagation and liquid aerosolization indicators by integrating PCA, and the optimal models are shown in Figure 5. Surprisingly, even the optimal clustering models have less acceptable cluster labelled scatter plots and wide fluctuation in the values of silhouette coefficient. Therefore, a more accurate clustering algorithm is needed to estalish a reasonable liquid combustion safety criterion.

4.2 Spectral clustering

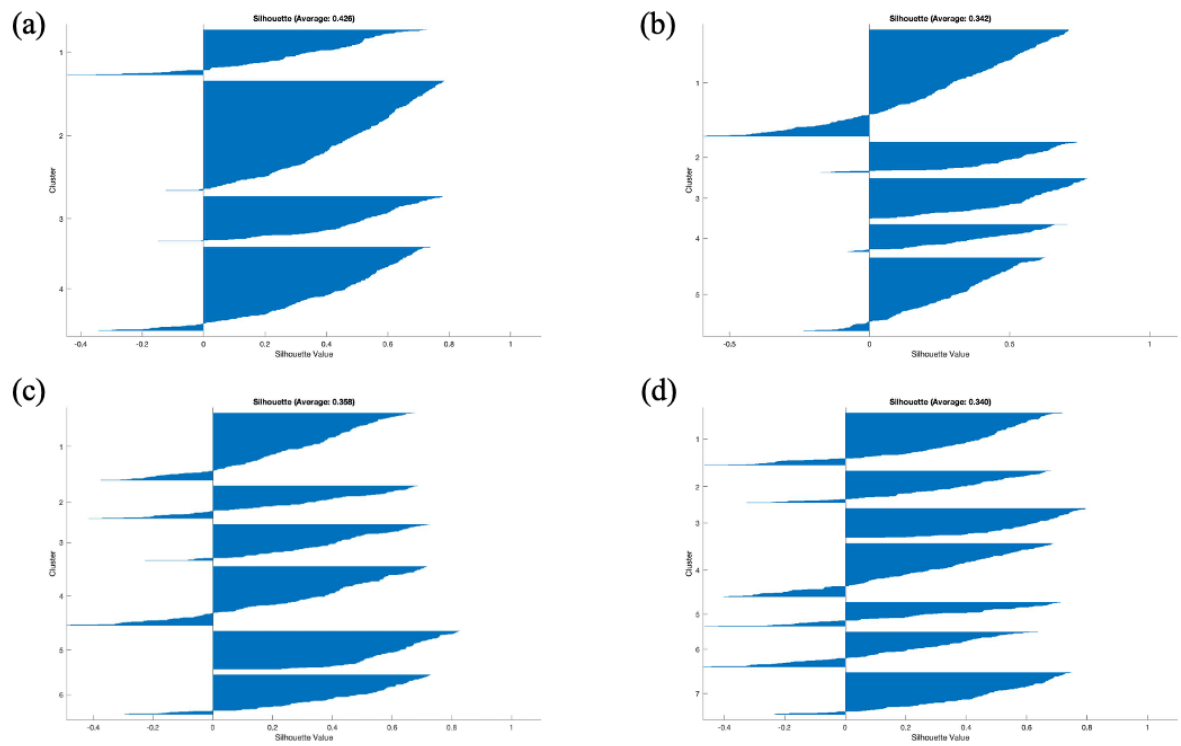


Figure 6. n cluster silhouette plots for liquid flammability indicators ($n \in \{4, 5, 6, 7\}$)

Figure 6 shows the n cluster silhouette plots for liquid flammability indicators. By considering the average silhouette value, the negative silhouette values and the fluctuation of the size of each cluster, the four cluster model is determined as the optimal spectral clustering model. Then the 3-dimentional labelled scatter plot and its star coordinate plot are visualized to testify the performance of the proposed spectral clustering model, see Figure 7.

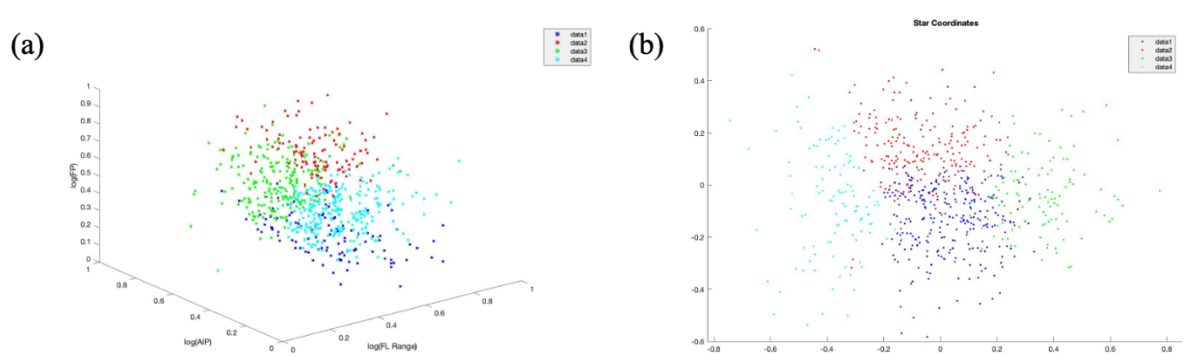


Figure 7. Three-dimensional labelled scatter plot (7a) and the corresponding star coordinate plot (7b)

As shown in Figure 7a, little information can be obtained since the cluster data distribution in space cannot be reflected by the 3-dimentional labelled scatter plot. The 3-dimentional labelled scatter plot is transformed by the theory of star coordinates (Kandogan, 2000), illustrated as Figure 7b. Clearly, this model performs great on clustering liquid flammability indicators, *i.e.*, FT, AIT and FR, only a limited number misclassified points locate in the cluster 1 and cluster 2, while cluster 3 and 4 have excellent clustering feature. Compared with the k-means clustering model with 2 principal components, the spectral clustering model increases the average silhouette value from 0.209 to 0.426, and the clustering effect on the labelled scatter plot improves a lot as well.

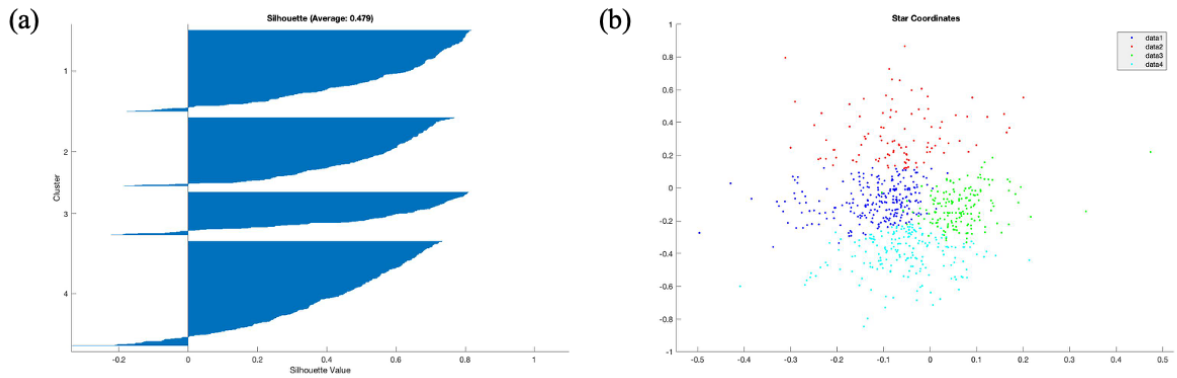


Figure 8. Optimal silhouette plot (8a) and optimal cluster star coordinate plot (8b) for flame propagation

Illustrated in Figure 8, the silhouette plot and the optimal cluster star coordinates plot show a fairly good results on clustering HC, LD and LTC, while the silhouette coefficient value is 0.479 and few outliers in the star coordinate plot has crossed the boundary of each cluster.

To summarize, the spectral clustering models outperform k-means clustering models with two princinpal components for each liquid combustion safety matrix, see Table 1. Also, the silhouette plots of the spectral clustering models present better performance than those of the k-means clustering models, consistent with the values of global average silhouette numbers. The optimal cluster models can be determined by the highest value of silhouette coefficient value, and the 4 cluster models are selected as the optimal clustering models for the three liquid in-cylinder combustion safety features: liquid flammability, flame propagation and aerosol formulation.

Table 1. Average silhouette coefficient value of three liquid combustion safety matrices for two clustering models

	Liquid Flammability				Flame Propagation				Aerosol Formulation			
	4 cluster model	5 cluster model	6 cluster model	7 cluster model	4 cluster model	5 cluster model	6 cluster model	7 cluster model	4 cluster model	5 cluster model	6 cluster model	7 cluster model
K-means clustering models with 2 PCAs	0.209	0.167	0.160	0.128	0.243	0.181	0.153	0.142	0.225	0.219	0.156	0.145
3-dimensional spectral clustering models	0.426	0.342	0.358	0.340	0.479	0.432	0.455	0.453	0.436	0.379	0.291	0.301

By employing the optimal clustering models, the risk ratings of each collected liquid flammability, flame propagation and aerosol formulation are generated by the *Matlab* codes with the help of the calculated information entropy values. The whole clustered data and the weight value calculation can be found in supporting information. The following table shows the example liquids with NFPA 704 flammability level 3 and 4 (National Fire Protection Association, 2017) but different ratings based on the our proposed clustering models.

Table 2. NFPA flammable and highly flammable liquids with different liquid in-cylinder combustion ratings

	NFPA Flammability	Liquid Flammability	Flame Propagation	Liquid Aerosolization	LICRI Value
Methanol	3	2	4	2	0.280
Ethanol	3	4	2	3	0.165
Methoxy acetone	3	4	4	2	0.103
o-Ethyl aniline	4	4	1	4	0.378
Di-(2-Chloroethoxy) Methane	4	3	2	3	0.223

As shown in the Table 2, methanol and ethanol share the same level in the NFPA 704 standard, but the proposed model points out that methanol is a less risky marine fuel on overall liquid in-cylinder risk combustion value than that of ethanol, although methanol has a high-risk rating on flame propagation.

Based on the LICRI values of the extracted substances in Table 2, the safety preferences can be ranked as o-Ethyl aniline, methanol, di-(2-Chloroethoxy) methane, ethanol and methoxy acetone. Following the same way, more promising fuels can be evaluated from the LICRI value of the spectral clustering model.

5. Conclusion

In this study, one novel liquid combustion safety criterion for compression ignition engines is carried out successfully with acceptable clustering outputs to fill in gaps for categorizing promising marine fuels from the perspective of chemical safety and life cycle assessment. This work confirms that the graph theory based spectral clustering performs better than k-means clustering algorithm in the non-convex liquid in-cylinder combustion database. The four cluster models are finalized as the optimal ones for liquid flammability, flame propagation and liquid aerosolization. The liquid organic compound database, comprising 703 substances, are clustered into four groups for the three safety matrices, the low overall rating presents the high-level hazard. The k-means algorithm integrates PCA to automatically optimize weight values of each principal component; while the spectral clustering algorithm employs the star coordinate plots to only reduce the high dimensional data into two dimensional data in the visualization stage. The star coordinate plots provide a great way to visualize high dimensional data sets, and it can solve the most obvious disadvantage of PCA, lack of interpretability.

Compared with the flash point dominated NFPA flammability standard, this criterion gives more information on marine fuel combustion in the CI engines. Also, the global mean silhouette value is reliable to find the optimal number of clusters in this work and it can be served as a robust reference to quantitatively evaluate the goodness of the clustering algorithms. Although the clustered results do show good performance on the cluster 1 and cluster 4, the results still needs to be improved on boundary determination of the cluster 2 and cluster 3. Nevertheless, the unsupervised clustering models with information entropy determined weight values give a more objective way to evaluate risk associated with the liquid in-cylinder combustion since it completely avoids human judgement to build the safety matrices. The future of this work may either improve the clustering results by adopting other clustering techniques such as hierarchical clustering and density-based spatial clustering of algorithms with noise or expand this work to the sustainability study of the promising marine fuel options so that the more greener and safer marine fuel solutions can be found to meet the long term strategy of the International Maritime Organization.

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Fireball and Flame Venting Comparisons: Test Data, CFD Simulations and Industry Standard Prediction

Peter A. Diakow,* J. Kelly Thomas, and Emiliano Vivanco
Baker Engineering and Risk Consultants, Inc.
3330 Oakwell Court, Suite 100
San Antonio, TX 78218-3024 (210) 824-5960
**Presenter E-mail:* PDiakow@BakerRisk.com

BakerRisk has performed vented deflagration testing of congested enclosures over a range of configurations, congestion levels and fuels. This paper provides a comparison of the measured flame jetting distances to predictions made using standard methods commonly used to calculate the associated hazard zone. These methods include the National Fire Protection Association Standard on Explosion Protection by Deflagration Venting (NFPA 68, [2]), the British Standard's Gas Explosion Venting Protective Systems (EN 14994, [3]) and computational fluid dynamics (CFD) analysis.

Nine test series were carried out using BakerRisk's Deflagration Load Generator (DLG) test rig. The DLG is 48-feet wide × 24-feet deep × 12-feet tall, yielding a total volume of 13,800 ft^3 (391 m^3), and is enclosed by three solid steel walls, a roof, and floor. The rig vents through one of the long walls (i.e., 48-foot × 12-foot). The venting face was sealed with a 6-mil (0.15 mm) thick plastic vapor barrier for these tests to allow for the formation of the desired fuel air-mixture throughout the rig. Both slightly hyper-stoichiometric propane and lean hydrogen mixtures have been tested in the DLG. Congestion was provided by an array of vertical cylinders. A range of congestion levels and fill fractions were tested. DLG testing was performed with and without vent panels present.

Flame jetting distances from the venting face of the DLG were measured using high-speed video. Flame jetting distances were predicted using the Fireball Dimensions calculation from NFPA 68 and the Flame Effects calculation from EN 14994. Blind (i.e., pre-test) simulations were also performed using the FLACS CFD code [1]. The flame jetting distance in the CFD simulation was taken as the distance from the DLG vent to where the gas temperature dropped below a specified value; the predicted distance for the fuel concentration to drop below half the lower flammability limit (LFL) was also evaluated as a check on the predicted jetting distance.

Keywords: Fireball, Vented Deflagration, Testing, CFD, NFPA 68, EN 14994, Blast Effects

Introduction

The tests described in this paper were performed in three separate tests programs [4, 5 and 6], referred to herein as test programs 1, 2 and 3 (T1, T2, T3). Nine separate test series were performed, as summarized in Table 1. Each test series in T1 (e.g., T1-A) consisted of three tests and each series in T2 and T3 consisted of two tests, except for T3-B which consisted of one test. The primary objectives of T1 and T2 were to (1) characterize vented deflagration blast loads and (2) compare these loads to those based on standard prediction methods. The primary objective of T3 was to gather vented deflagration data for validation of numerical models [7]. A comparison of the measured flame jetting distances was made with predictions based on the correlations provided in NFPA 68 and EN 14994, and to values calculated using the FLACS CFD code.

Table 1. Combined Test Matrix

Program	Test Series	Fuel	Fuel Concentration	Flammable Volume	Congested Volume	Obstacle to Enclosure Surface Area Ratio (A_r)	Vent Parameters
1	A	Propane	4.33%	100%	100%	0.32	6 mil plastic
	B						20-gauge steel panels
	C			50%			
	D			100%	25%	0.10	6 mil plastic
2	A				50%	0.39	
	B					0.27	
	C					0.15	
3	A				100%	0.38	
	B01	Hydrogen	20%				

Background

A vapor cloud explosion (VCE) is classified as a deflagration if the flame propagates through the unburned fuel-air mixture at a burning velocity less than the speed of sound. The confinement and congestion associated with the volume encompassed by a flammable cloud affect the flame speed, which governs the resulting blast load. Confinement refers to solid surfaces that prevent free expansion of the expanding gas in one or more dimensions (e.g., solid walls, roof, etc.). Congestion refers to obstacles in the path of the flame that generate turbulence (e.g., the vertical cylinders in the rig used in these tests). Turbulence increases both the combustion rate per unit surface area as well as the flame surface area.

During a deflagration, the unburned portion of the flammable cloud is pushed ahead of the flame as the product gases expand. Cloud expansion is a function of the flame speed and the initial flammable cloud volume. Increasing flame speed decreases the amount of expansion prior to consuming the flammable cloud. Increasing the initial flammable cloud volume proportionally increases the expanded volume. Expansion occurs in all directions for an unconfined deflagration, whereas confinement limits free expansion to the unconfined direction(s). During a vented deflagration, flame jetting occurs until the fireball consumes the flammable mixture that has expanded through the vent and/or the vented volume is depressurized.

Test Rig Configuration

The DLG test rig is an enclosure with three solid walls, a roof, and floor, measuring 48-feet wide, 24-feet deep, and 12-feet tall [5]. Venting was allowed through one of the long walls (i.e., 48-foot x 12-foot). The venting face of the rig was sealed with a 6-mil thick plastic vapor barrier, which released (i.e., tore open) at approximately 0.1 psig; the vapor barrier allowed for the formation of a fuel-air mixture inside the test rig. For T1-C, the plastic vapor barrier was installed halfway between the rear of the test rig and the venting surface. For all other tests, the vapor barrier was installed on the rig face, such that the flammable cloud filled the entire enclosure volume. For T1-B and T1-C, steel vent panels (20 gauge, 2 lb_m/ft^2) were installed over the plastic vapor barrier using Fabco® Vent-All explosion relief fasteners to provide a 0.3 psig vent release pressure; vent panel restraint devices were not utilized in these tests. Figure 1 shows photos of the three venting configurations utilized: (1) vapor barrier installed on the external venting face of the rig, (2) steel panels installed on the external face of the rig, and (3) vapor barrier installed at the halfway point in the rig.

The target propane concentration for these tests corresponds to the peak of the laminar burning velocity (LBV) curve for propane-air mixtures (4.33% propane). The peak of the LBV curve constitutes a worst-case mixture, but also corresponds to the region with least change in LBV versus concentration. Minimum and maximum concentration thresholds were established based on a 1% variation in LBV from the peak value. The target hydrogen concentration was 20% since a worst-case hydrogen-air mixture would be expected to result in a deflagration-to-detonation transition (DDT) inside the DLG; a DDT was observed at a higher concentration [6]. Ten sample points were distributed throughout the DLG test rig to allow the uniformity of the fuel-air mixture to be monitored. Each sample point indicated a fuel-air concentration within the tolerance thresholds prior to ignition. The flammable mixture was ignited at the center of the rear wall, opposite the venting surface.

Congestion inside the rig was provided by an array of vertical cylinders (2.375-inch and 2-inch outer diameter) that occupied the internal volume of the rig. For T1 and T3, the 2.375-inch outer diameter cylinders were located at the front of the rig (first two rows) to minimize plastic deformation of the cylinders due to repeated loading; T2 only utilized the 2-inch outer diameter cylinders. The congestion patterns used for T1,T2 and T3 are shown in shown in Figure 2, Figure 3 and Figure 4, respectively. The resulting obstacle-to-enclosure surface area ratios (A_r), which provide a relative congestion measure, are provided in Table 1.

High speed (HS) cameras recording at 1000 frames per second (fps) and high definition (HD) cameras recording at 30 fps were located approximately 200 feet away from the DLG, perpendicular to the venting face. The HS camera recordings were used to evaluated flame jetting distance. However, for test T3-B, is should be noted that hydrogen-air flames are nearly impossible to see since they burn with a very pale blue flame and do not produce soot particles, so the resulting flame length data for this test is approximate. It is recognized that a thermocouple array or radiative heat flux gauges could have provided a more quantitative measure, but flame jetting measurements were not the focus of these test programs. Pressure transducers were also fielded internal and external to the test rig.

Figure 1. Test Rig Venting Configurations

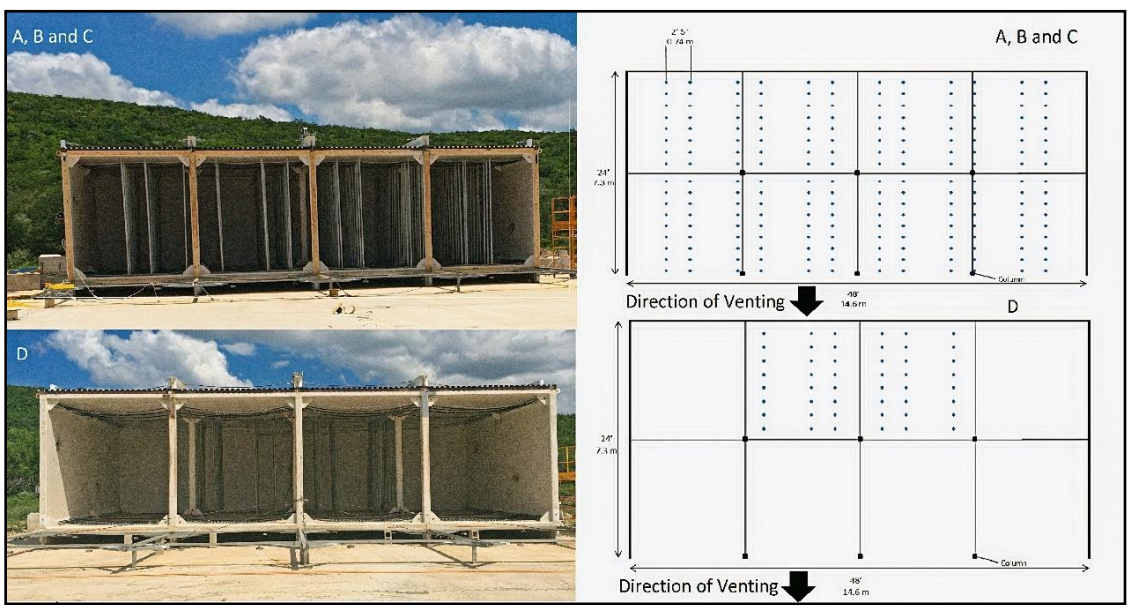

Figure 2. Test Rig Congestion Configurations for T1

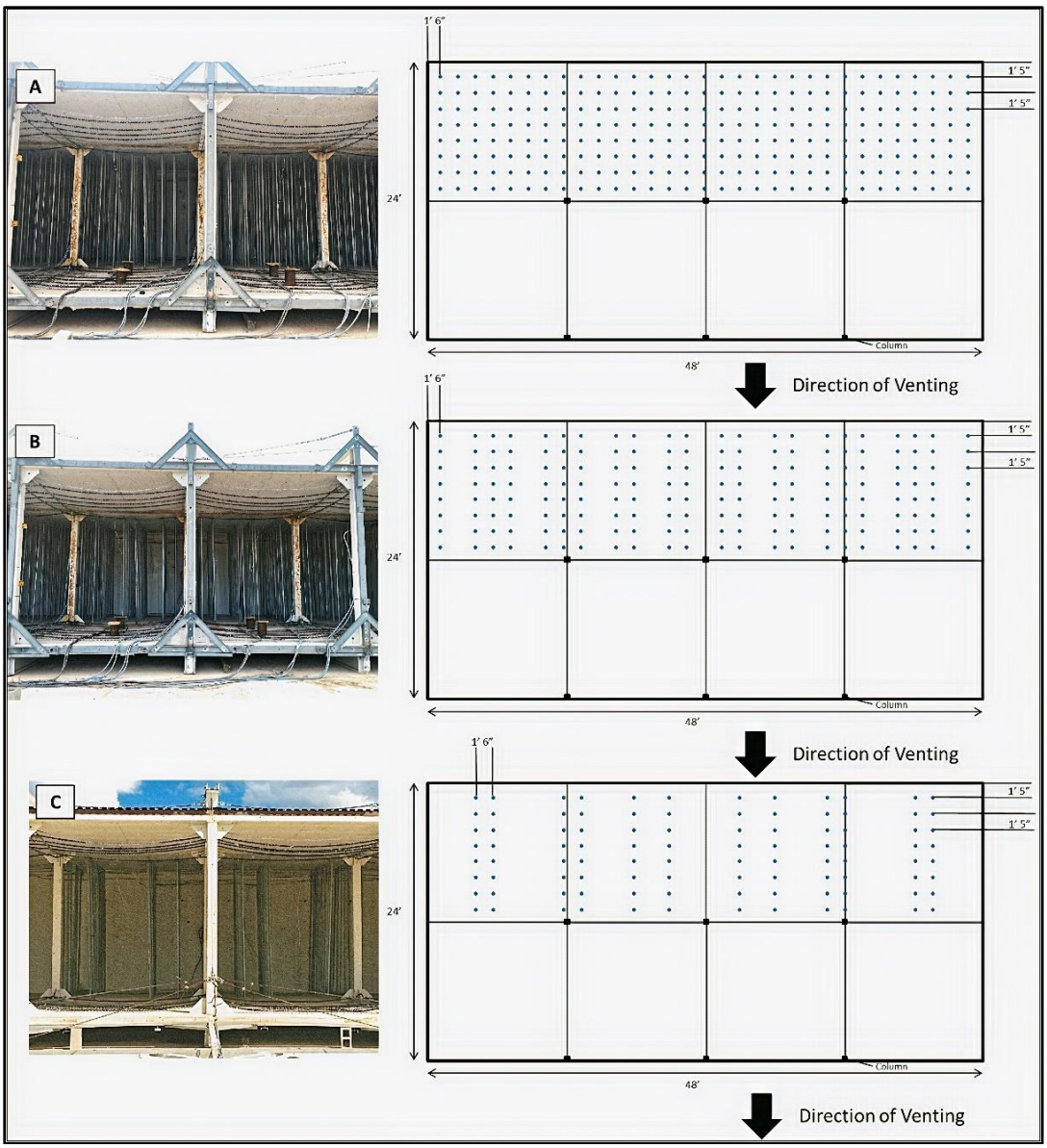

Figure 3. Test Rig Congestion Configurations for T2

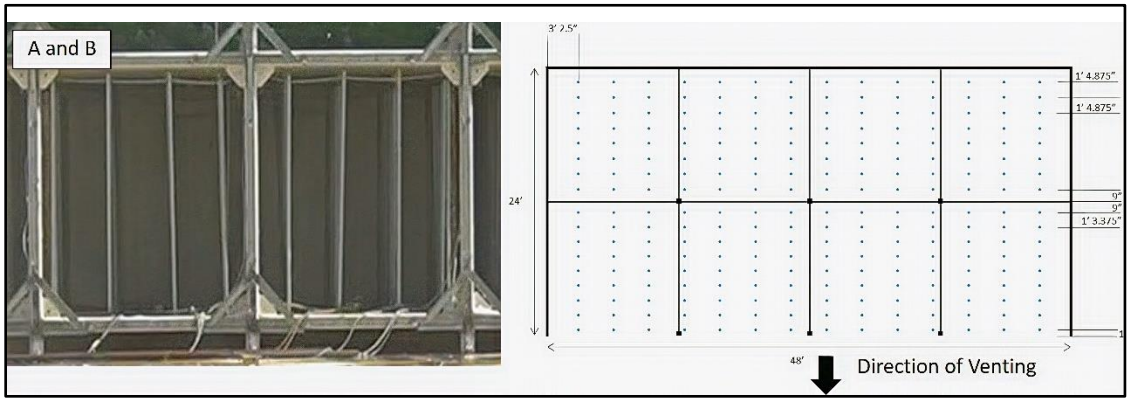

Figure 4. Test Rig Congestion Configurations for T3

NFPA 68 and EN 14994

NFPA 68 provides an equation (section 7.6 *Fireball Dimensions*, eqn. 7.6.1) for the calculation of the hazard zone due to the fireball ejected from a vented gas deflagration. The NFPA 68 hazard zone calculation is based on the vented enclosure volume and number of evenly distributed vents, and is restricted in application to enclosure volumes of 0.1 m^3 to 10,000 m^3 (3.5 ft^3 to 353,000 ft^3). EN 14994 also provides an equation (section 6.4.2 *Flame Effects*, eqn. 13) for the calculation of the flame length ejected from a vent opening. The EN 14994 flame length calculation is based solely on the vented enclosure volume, and is restricted in application to enclosure volumes of 0.1 m^3 to 50 m^3 (3.5 ft^3 to 1,800 ft^3). Note the DLG test rig volume (391 m^3) exceeds the valid range for the EN 14994 predictions by a factor of 8.

Figure 5 shows the predicted flame jetting distance for a range of enclosure volumes using NFPA 68 and EN 14994. The venting face of the DLG rig was taken to be a single vent in the NFPA 68 calculation. The volumes corresponding to 50% and 100% of the DLG test rig are also shown in Figure 5. For the 50% rig volume case, which reflects Test T1-C, the comparison in Figure 5 implies that the effective enclosure volume would be taken to be half the rig volume, while these correlations actually specify the total enclosure volume (i.e., allowance is not provided for a cloud which fills only a portion of the enclosure). The DLG rig volume is greater than the upper limit of the EN 14494 correlation, and the values shown above the upper limit of this correlation are an extrapolation. However, the NFPA 68 and EN 14994 predictions are very similar at 50% or 100% test rig volumes.

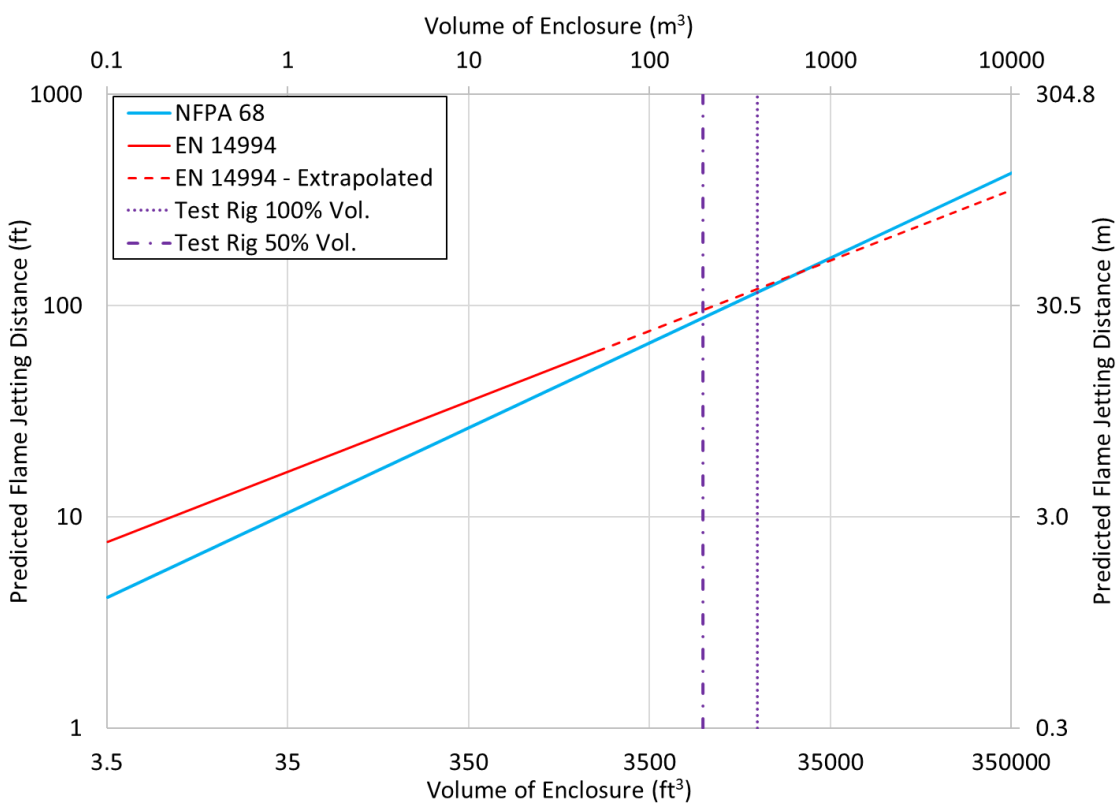

Figure 5. Predicted Flame Jetting Distance as a Function of Volume

FLACS Analysis

Blind (i.e., pre-test) CFD simulations were performed using the FLACS (Flame Acceleration Simulator) code [5, 7] since this code is commonly used in industry for VCE simulations. FLACS solves conservation equations for mass, momentum, enthalpy, turbulence, and species/combustion on a 3D Cartesian grid.

The proper representation of geometry (obstacles and structures) is a key aspect of the development of a FLACS model. Obstacles such as structures and pipes are represented as area porosities (the opposite of blockages) on control volume (CV) faces and are represented as volume porosities in the CV interior. CV surfaces and volumes are each either fully open, fully blocked, or partly blocked. For the partly blocked surfaces or volumes, the porosity is defined as the fraction of the area/volume that is available for fluid flow. The resulting porosity model is used to calculate flow resistance and turbulence source terms from small objects, as well as the flame speed enhancement arising from flame folding in the sub-grid wake. The FLACS geometry created for the T2-A simulations is shown in Figure 6. The flammable volume, flammable cloud composition, and ignition location used for the simulations were the same as for the field tests.

The computational mesh was created following the guidelines in the FLACS user's manual [1], which states that for confined rooms filled with gas from wall to wall, the combustion region must be resolved by a minimum of 5-6 grid cells in the smallest direction of flame acceleration. The smallest DLG rig dimension is 12-feet (height), resulting in a grid cell size of 2-feet. The grid cells are cubical in shape, resulting in an internal mesh 6 cells tall, 12 cells deep, and 24 cells wide. The same cell size was used for the mesh external to the rig.

The simulation region was -328 feet to 328 feet (-100 m to 100 m) in the X direction; -246 feet to 394 feet (-75 m to 120 m) in the Y direction; and 0 feet to 164 feet (0 m to 50 m) in the Z direction. The X direction corresponds to the DLG width (48 feet), the Y direction corresponds to the DLG depth (24 feet), and the Z direction corresponds to the DLG height (12 feet). The origin (0 feet, 0 feet, 0 feet) corresponded to a bottom interior corner of the DLG. A plane wave boundary condition was used for the ±X, ±Y, and +Z directions.

The flame jetting distance in the FLACS CFD simulation was taken as the distance from the DLG vent to the location where the gas temperature dropped below 1000 °K (1340 °F), below the height of the test rig. This choice of temperature is in line with similar comparisons conducted using FLACS [8]. Figure 7 shows exemplar FLACS temperature contour plots for test T2-B; the predicted flame jetting distance for this case was determined to be 70 feet (21.3 m) based on the temperature criteria. As a comparison, the flame jetting distance was also evaluated by identifying the location where the fuel concentration dropped below half the LFL; this should underpredict the flame jetting distance as the fireball can move outward beyond the point where combustion terminates due to the continued expansion and momentum of the product gas.

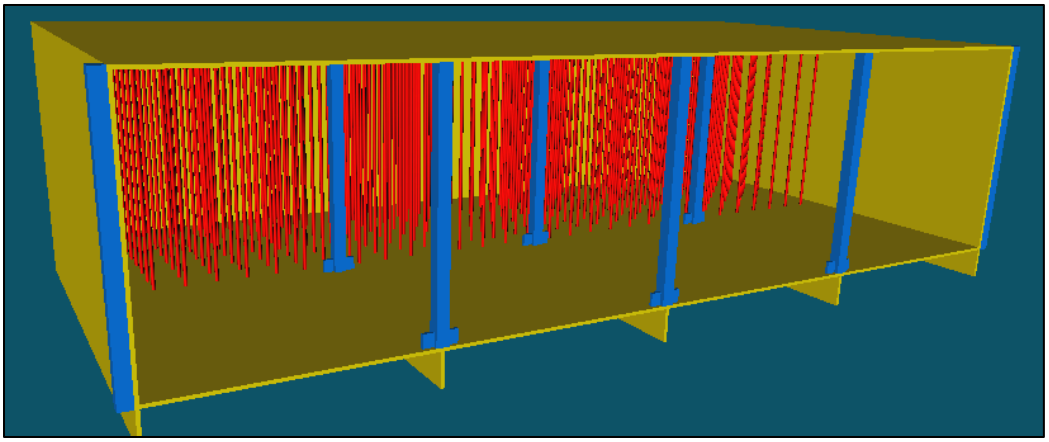

Figure 6. DLG Rig Model Created for FLACS Analysis, T2-A

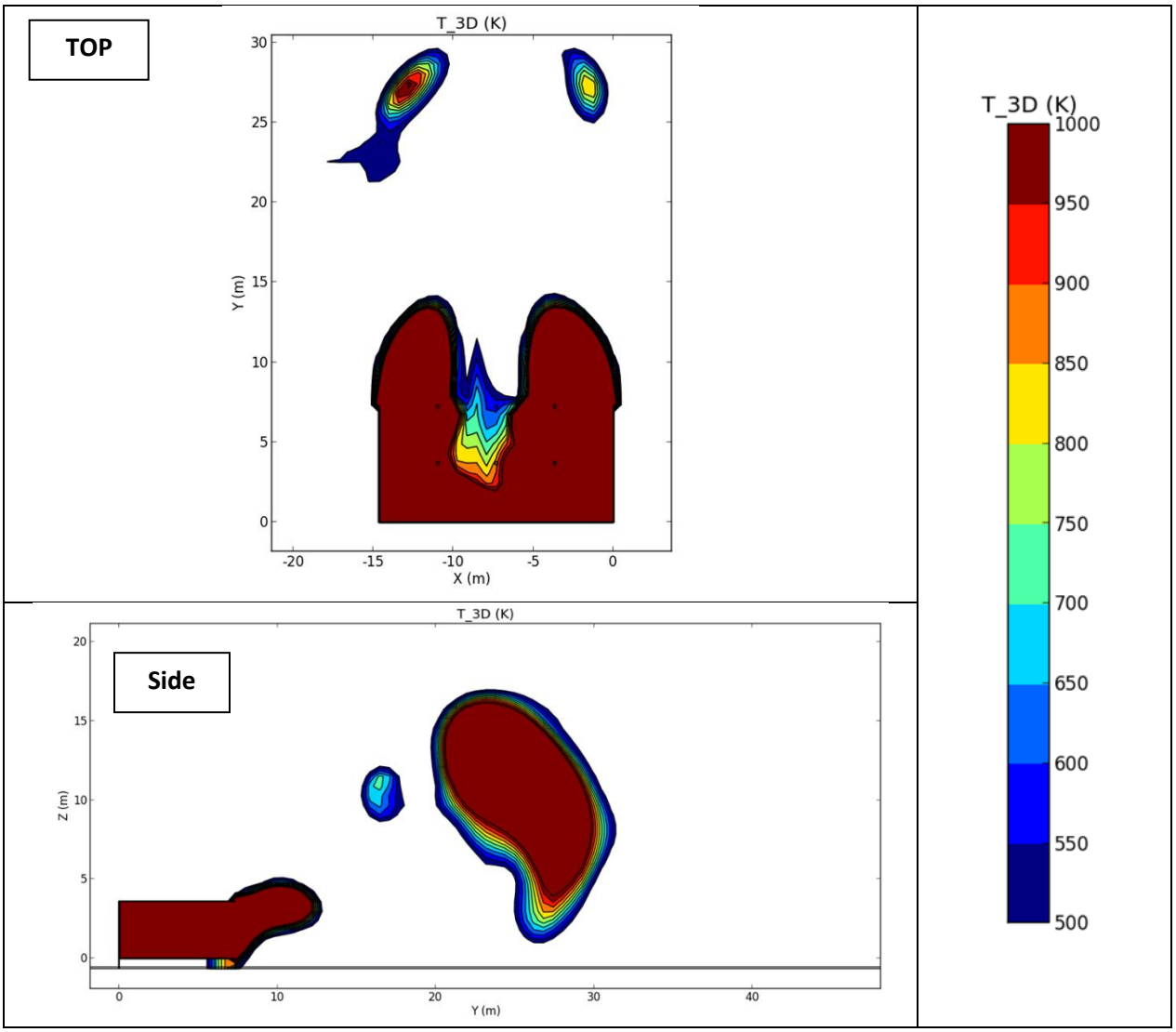

Figure 7. FLACS Analysis Temperature Plots (500 °K - 1000 °K) for T1-D

Results and Discussion

The maximum flame jetting distance from the venting face of the rig was determined for each test using HS video. Figure 8 shows still frames from the T1-D HS video at four different times during flame jetting. For test T1-C, where only the back half of the rig contained a flammable mixture, so that 12 feet (i.e., half the rig width) was added to the observed flame jetting distance.

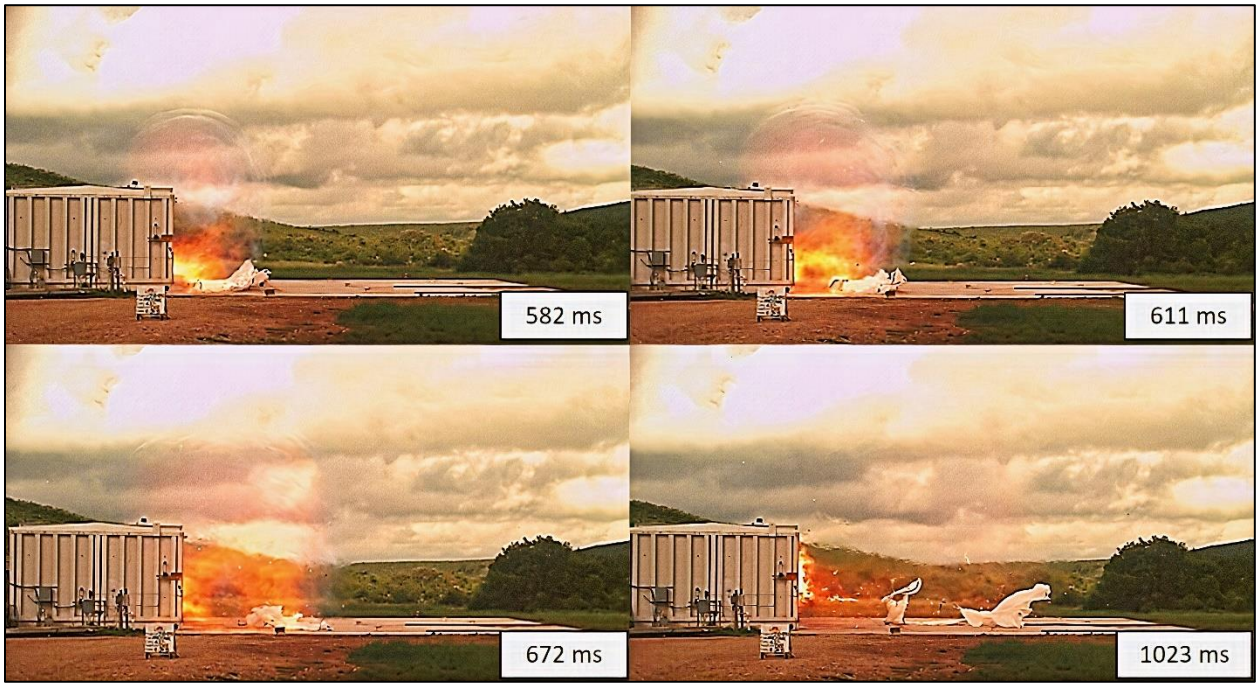

Figure 8. Still Frames from HS Camera Footage, T1-D

Table 2 and Figure 9 provide the average flame jetting distance for each test series along with the predicted values (i.e., from NFPA 68, EN 14994 and based on the FLACS analysis). Table 3 shows the ratio of the predicted to measured flame lengths for each prediction method. As noted previously:

- The EN 14994 predictions are an extrapolation, since the test rig volume (391 m^3) exceeds the valid range for this correlation by a factor of 8,
- The predictions based on the distance at which the flammable gas concentration drops to LFL/2 in the FLACS analysis should underpredict flame length, and
- The test data for the hydrogen test (T3-B) has significant uncertainty.

The NFPA 68 hazard distance predictions are significantly longer than the measured flame jetting distance, with the ratio between the predicted and measured values averaging 2.3 and reaching 2.9 for one test. This is not unexpected, since the "hazard distance" to personnel, the quantity predicted by the NFPA 68 correlation, should extend beyond the actual flame jetting distance (i.e., there is a hazard due to thermal radiation which extends beyond the actual flame). In addition, the NFPA 68 correlation is intended to be conservative. The predictions made using the EN 14994 correlation are slightly more conservative.

Table 2. Flame Jetting Distances for Tests and Predictive Methods

Test ID	Flame Jetting Distance (ft)				
	Test Data	NFPA 68	EN 14994	FLACS (Temperature)	FLACS (Fuel Conc.)
T1-A	52	112	120	171	28
T1-B	42	112	120	170	24
T1-C	36	85	95	153	16
T1-D	43	112	120	67	28
T2-A	51	112	120	232	32
T2-B	48	112	120	142	32
T2-C	51	112	120	116	28
T3-A	67	112	120		
T3-B	38	112	120		

Table 3. Predictive Method Flame Jetting Distance Ratios

Test ID	Flame Jetting Distance (Ratio to Test Data)			
	NFPA 68	EN 14994	FLACS (Temperature)	FLACS (Fuel Conc.)
T1-A	2.2	2.3	3.3	0.5
T1-B	2.7	2.9	4.1	0.6
T1-C	2.4	2.6	4.3	0.4
T1-D	2.6	2.8	1.6	0.7
T2-A	2.2	2.4	4.5	0.6
T2-B	2.3	2.5	3.0	0.7
T2-C	2.2	2.4	2.3	0.6
T3-A	1.7	1.8		
T3-B	2.9	3.2		

The predicted flame travel distances from FLACS using a 1000 °K temperature criterion were conservative compared to the observed flame jetting distances, ajusting the temperature defintion by +/- 100 °K resulted in an approximate change of +/-10 feet (+/- 20%) to the predicted flame length for most cases, but changes of up to 27 feet (52%) were observed in one case. The FLACS simulations show hot product gases traveling upwards and outwards away from the the combustion region. In many cases product gases with a temperature >1000 °K propagated beyond the reported distance above the height of the test rig, where they would not pose a hazard to personel on the ground.

As expected, the flame jetting distance predicted based on the FLACS analysis using a fuel concentration criterion (i.e., LFL/2) significantly underpredicted the measured values in all cases.

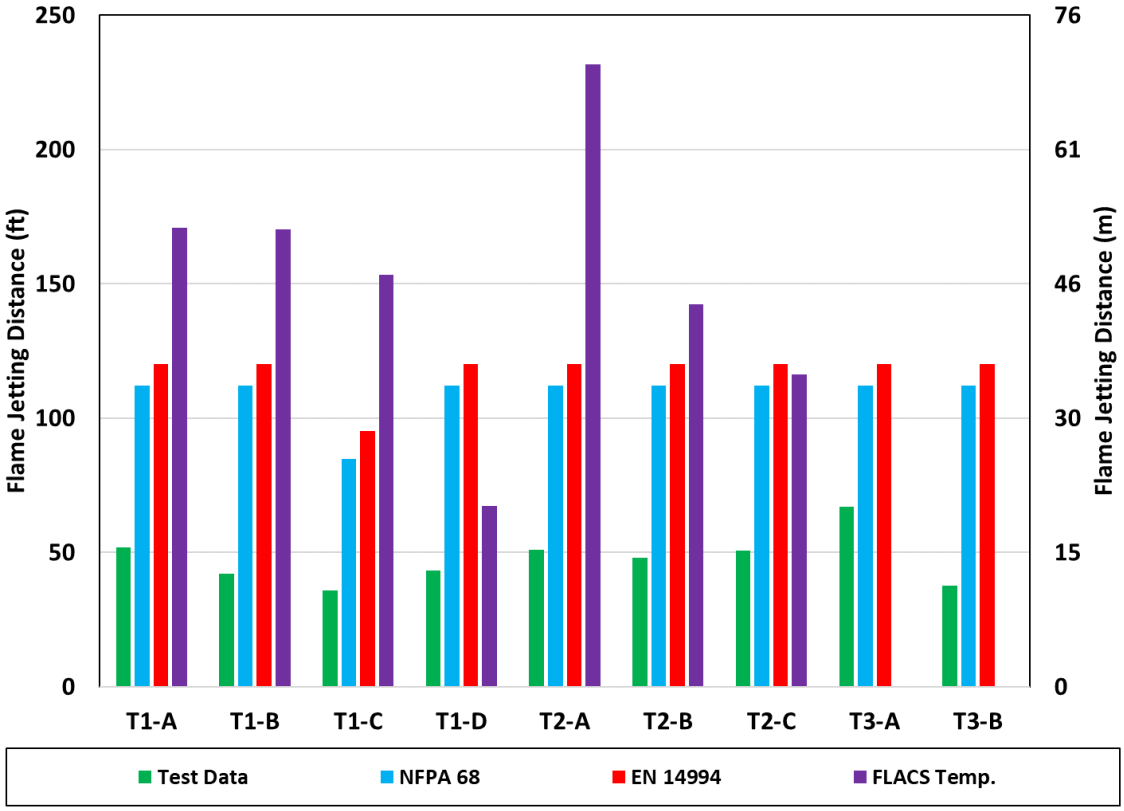

Figure 9. Flame Jetting Distance for Tests and Predictive Methods

<u>Flame Jetting Distance vs. Vent Panel Configuration</u>

The T1 test series without vent panels (T1-A) had a longer average flame jetting distance (52 feet) compared to the test series with panels (T1-B, 42 feet). This likely is because the panels act as a physical barrier for the flame as they move away from the rig, as can be seen in Figure 10. The panels also relieve at a higher pressure (0.3 psig) than the vapor barrier (0.1 psig), which decreases the amount of unburned fuel available for expansion outside of the rig at the time the vent activates, but this effect is small for the range of vent release pressures examined.

FLACS models pressure relief panels using a porosity parameter specified as completely closed (i.e., porosity = 0) until the vent release pressure is reached, and completely open thereafter (i.e., porosity = 1). FLACS does not model the physical barrier to flame expansion provided by the vent panels as they are displaced away from the venting surface (i.e., it is not modelling the fluid-structure interaction). The FLACS predictions based on a temperature criterion for tests T1-A and T1-B relatively unchanged; the FLACS predictions were 3.3-4.1 time greater than the observed values.

The NFPA 68 and EN 14494 correlations do not distinguish between a vent panel and a vapor barrier, nor for the effect of vent release pressure.

Figure 10. HS Camera Still Frames, T1-A (without panels) vs T1-B (with panels)

Flame Jetting Distance vs. Flammable Volume

The T1 test series that utilized a flammable cloud filling 100% of the rig volume (T1-B) had a longer average flame jetting distance (42 feet) compared to the test series where the cloud only filled 50% of the rig volume (T1-C, 36 feet). That is, flame jetting distance was observed to increase with flammable cloud volume, as expected.

The FLACS temperature criterion predictions slightly changed from the T1-B 100% cloud volume case (170 feet) to the T1-C 50% volume case (i.e., 170 and 153 feet, respectively), both of which are an over-prediction of the observed value.

As noted previously, for the NFPA 68 and EN 14494 correlation predictions, the enclosure volume was set equal to the flammable cloud volume for Test T1-C (i.e., 50%), albeit these correlations actually specify the total enclosure volume (i.e., allowance is not provided for a cloud which fills only a portion of the enclosure). The NFPA 68 and EN 14494 predictions, with this caveat, did follow the observed trend, while being conservative by a factor of 2 to 3.

Flame Jetting Distance vs. Congestion Level and Volume

There were several test series (T1-A, T1-D, T2-A, T2-B, T2-C and T3-A) where the only parameter changed was the congestion array within the rig. Flame jetting distance generally increased with congested volume. The observed flame jetting distance was longer for a congestion array filling 100% of the rig (i.e., T1-A and T3-A), and shortest with a congestion array which filled only 25% of the rig (i.e., T1-D); however, it is recognized that the flame jetting distance for

Test T1-A is only slightly larger than for Test T1-D test series. It is expected that flame deceleration in the uncongested portion of the rig limits flame jetting distance, as a very slow flame would have very limited flame jetting. The average flame jetting distances in these tests showed little relationship with the congestion level, with all three of the T2 test series giving roughly the same flame jetting distance despite a wide range of congestion level (see Figure 3).

The FLACS temperature criterion predictions did not follow the trend of the test data. A congestion array filling 50% of the rig (i.e., the T2-A test series) gave the longest predicted flame jetting distance. The FLACS temperature criterion predictions did not show a clear dependence on congestion level, which is agreement with the test data. The NFPA 68 and EN 14494 correlations are not dependent on the congestion parameters.

Flame Jetting Distance vs. Fuel

The T3-A test series with propane had a longer average flame jetting distance (67 feet), compared to the T3-B test series with hydrogen (38 feet). This may be due to high flame speed associated with the hydrogen mixture resulting in little unburned mixture being pushed out of the rig. However, as noted previously, hydrogen-air flames are very difficult to see and hence the test

The NFPA 68 and EN 14494 correlations are not dependent on fuel type.

Conclusions

The NFPA 68 hazard distance predictions are between approximately 2 and 3 times longer than the measured flame jetting distance. However, the "hazard distance" to personnel predicted by the NFPA 68 correlation should extend beyond the actual flame jetting distance, since the hazard extends beyond this distance, and since the NFPA 68 correlation is intended to be conservative. The predictions made using the EN 14994 correlation are slightly more conservative, but represent an extrapolation beyond the upper limit for volume for this correlation.

The flame jetting distances predicted based on the FLACS analysis using a temperature criterion were greater than the measured values for all tests, and generally conservative compatered to the NFPA 68 and EN 14994 predictions. The temperature criterion defined for this work does not account for thermal dosage, future work could look at thermal dose and associated vulnerability to define the hazard zone.

The following observations were made based on test data:

- The presence of unrestrained vent panels (vs. only a vapor barrier) reduced flame jetting distance since the panels act as a physical barrier to the flame.
- Flame jetting distance increases with flammable cloud volume (i.e., the fraction of the enclosure filled with a flammable gas mixture).
- Flame jetting distance generally increases with congested volume (i.e., the fraction of the enclosure filled with congestion). This is likely due to flame deceleration in the uncongested portion of the enclosure.

Acknowledgments

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Numerical simulation of methane-air DDT in channels containing trace amounts of impurities

Logan N. Kunka[*], Carolyn R. Kaplan[+], Elaine S. Oran[*]
[*] Texas A&M University [+] University of Maryland
logankunka@tamu.edu

Abstract

Accidental explosions in industrial settings often cause devastating losses to personnel and industry. Many instances, such as in coal mines, are attributed to the accumulation of dangerous amounts of methane with trace amounts of heavier hydrocarbon gases, which mix with air and create conditions for flame ignition and subsequent deflagration-to-detonation transition (DDT). This paper discusses the conditions under which flames in channel geometries can accelerate to detonation and the effects of trace amounts of impurities. DDT was investigated for the addition of ethane and propane into a methane-air mixture at various geometry scales with constant blockage ratio and channel configuration. Results of small-scale simulations of DDT in channels containing methane and air were compared with existing experimental data. We found that the location where DDT occurred, L_{DDT}, decreased slightly as the percentage of impurity changed. The variation was, in fact, on the order as the stochasticity (uncertainly due to turbulence and turbulence interactions) in the simulation. The detonation cell size, however, decreased with increased amount of impurities, thus resulting in a more robust detonation wave.

Keywords: Explosions, deflagration-to-detonation transition, numerical simulation, reactive flows

Introduction

Accidental industrial explosions are low-probability, high-impact events that cause devastating losses. These accidents are a safety concert to various industries including oil and gas, mining, and fuel refining and transportation [1–4]. For the mining industry in particular, the conditions for such explosions are created in confined regions of underground mines such as abandoned and sealed sections of the mine. Natural gas can accumulate in these sealed sections mixing with air to create an explosive gas mixture.

When detonation occurs in accidental explosions in coal mines or fuel storage facilities, the destructive potential of the explosion increases enormously. If DDT occurs in the system, the

energy release rate drastically increases and the resulting detonation wave can travel at several kilometers per second. The explosion frequently has the same mechanism: ignition, flame propagation and acceleration, then transition to detonation. The route to detonation is the response of a reactive gas to a smaller explosions created in deflagration and from the formation of hotspots [5,6]. Goodwin et. al [7] and Xiao et. al [8] showed that DDT can also occur through shock focusing on the flame front or the unburnt mixture ahead of the flame. It is important to know the distance it takes for such DDT process to occur so that protective seals can be designed accordingly [9]. A worst-case scenario would occur if DDT happened at the seal, when pressure reaches its maximum.

Most work focusing on such explosions has modeled the explosive mixture as pure methane-air [10–12]. While the primary hydrocarbon in natural gas is methane, often trace amounts of impurities such as propane and ethane are included. These heavy hydrocarbons are often 0-20% of natural gas by volume. This work aims to understand the influence of these impurities on the DDT process and ultimately how it effects the run-up distance L_{DDT}.

Numerical Model

The numerical simulations solve the two-dimensional (2D) fully-compressible reactive Navier-Stokes equations for conservation of mass, momentum, energy and species. The reactants are perfectly mixed and are assumed to behave as an ideal gas.

$$\frac{\partial \rho}{\partial t} + \nabla \cdot (\rho \vec{u}) = 0$$

$$\frac{\partial \rho \vec{u}}{\partial t} + \nabla \cdot (\rho \vec{u}\vec{u}) + \nabla p = \nabla \cdot \tilde{\tau}$$

$$\frac{\partial (\rho E)}{\partial t} + \nabla \cdot ((\rho E + p)\vec{u}) = \nabla \cdot (\vec{u} \cdot \tilde{\tau}) + \nabla \cdot (K \nabla T) - \rho q \dot{\omega}$$

$$\frac{\partial (\rho Y)}{\partial t} + \nabla \cdot (\rho T \vec{u}) = \nabla \cdot (\rho D \nabla Y) + \rho \dot{\omega}$$

$$p = \rho R T / M$$

$$\tilde{\tau} = \rho \nu ((\nabla \vec{u}) - (\nabla \vec{u})^T - \frac{2}{3}(\nabla \cdot \vec{u}) I)$$

$$E = \frac{P}{(\gamma - 1)\rho} + \frac{1}{2}(\vec{u} \cdot \vec{u})$$

where ρ is the density, t is the time, p is the pressure, $\vec{u}$ is the vector velocity, T is the temperature, E_a is the specific total energy, Y is the mass fraction, q is the chemical energy release, $\dot{\omega}$ is the chemical reaction rate, κ is the thermal conductivity, D is the mass diffusivity, R is the universal gas constant, M is the molecular weight, ν is the kinematic viscosity, $\tilde{\tau}$ is the viscous stress tensor, I is the unit tensor, and γ is the specific heat ratio.

Combustion and the conversion of fuel to product is modeled using a calibrated single-step chemical-diffusion model (CDM) where the reaction rate ($\dot{\omega}$) is defined as

$$\dot{\omega} = \frac{dY}{dt} = -A\rho Y exp(-\frac{E_a}{RT})$$

where Y is the fuel mass fraction, t is time, A is the pre-exponential factor, ρ is the fluid density, E_a is the activation energy, R is the universal gas constant, and T is the fluid temperature.

The diffusion properties of the mixture are temperature depended and defined as

$$\kappa = \kappa_0 T^{0.7}/\rho,\ D = D_0 T^{0.7}/\rho,\ \mu = \mu_0 T^{0.7}/\rho$$

Chemical-Diffusion Model

The Arrhenius Equation in the previous section describes the conversion of fuel into product as part of the chemical-diffusion model (CDM). To generate these CDMs a genetic algorithm and optimization approach is used to find the optimal value for model parameters (γ, A, E_a, q, κ_0, M_w) such that calculated flame and detonation parameters (T_b, $T_c v$, S_l, x_{ft}, D_{CJ}, x_d) match their specified values [13]. This model has extensively tested in laminar and turbulent flames, detonations, and DDT [10,14,15].

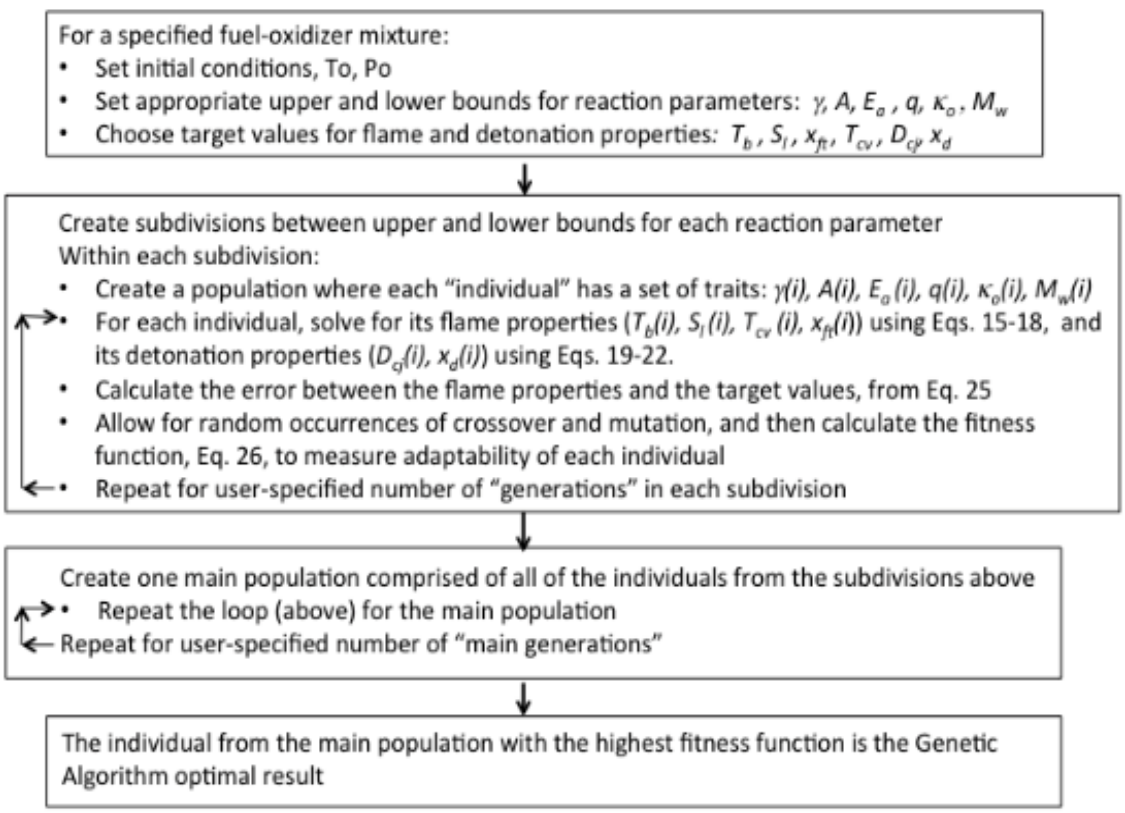


Figure 1: Genetic algorithm for creating Chemical-Diffusion Models [13]

To create a CDM for mixtures with impurities, a percentage of the methane by volume was replaced with the equal amount of heavy hydrocarbon. The mixture parameters were then computed with the genetic algorithm approach. This process was repeated to develop CDMs for

0%, 1%, 2%, 4%, 6%, and 8 % propane and ethane. The CDM for methane-air containing 4% ethane is summarized in in Table 1.

Parameter	Optimal Value	Definition
γ	1.19	Ratio of specific heats
E_a/RT	81.59	Activation energy
A (cm^3/g-s)	1.32×10^{14}	Pre-exponential factor
qM_w/RT	40.50	Normalized heat release
κ_o (g/s-cm-K$^{0.7}$)	6.90×10^{-6}	thermal conductivity
M_w (g/mole)	27.42	Molecular weight

Parameter	Calculated Result	Target	Definition
D_{CJ} (m/s)	1809	1811	Chapman-Jouguet detonation velocity
S_l (cm/s)	38.57	38.53	Laminar flame speed
x_{ft} (cm)	0.0434	0.0435	Laminar flame thickness
T_b (K)	2233	2235	Adiabatic flame temoperature
x_d (cm)	0.960	0.960	Detonation half thickness
T_{cv} (K)	2602	2600	Final temperature at constant volume

Table 1: CDM for Stoichiometric Methane-Air with 4% Ethane

Problem Setup

The computational domain is a long channel with regularly spaced obstacles ($L = d$) with a blockage ratio $br = 0.3$. This configuration is consistent with DDT experiments [16–18] and is typical for various industrial setting and mines. Previous work has shown effects of blockage ratio, obstacle type and obstacle placement on L_{DDT} [14,19,20]. The configurations investigated were $d = 17.4cm$, $d = 52cm$, and $d = 1m$. All geometry ratios were kept constant for each gas mixture.

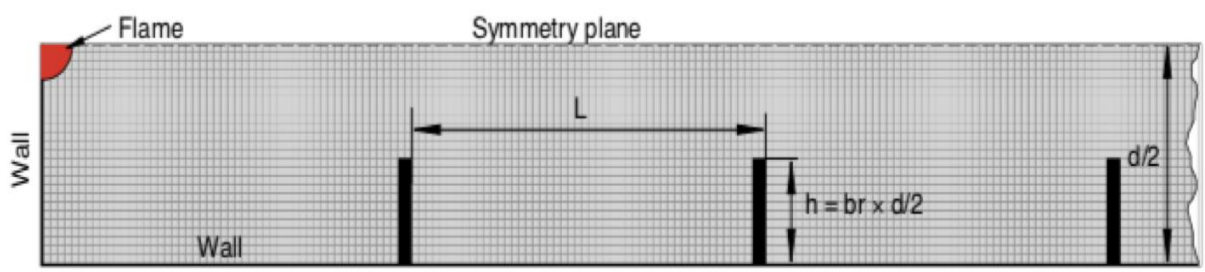


Figure 2: Computational Domain

The numerical model is solved through the Flame Acceleration Simulation Tool (FAST). FAST solves the governing equations using a fifth-order-accurate spatial-reconstruction method that

includes features from the nonlinear error-controlled WENO scheme [21] and HLLC approximate Riemann solver [22]. The time integration uses a third-order explicit Runge-Kutta scheme. The code includes an immersed boundary algorithm that allows flow computations in and around complicated geometries. Adaptive mesh refinement (AMR) is used to dynamically refine the mesh in areas of interest though the use of BoxLib [11]. AMR helps resolve important flow features such as flames, pressure waves, shocks, and boundary layers. The grid spacing was $dx_{max} = 0.27cm$ and $dx_{min} = 0.03cm$.

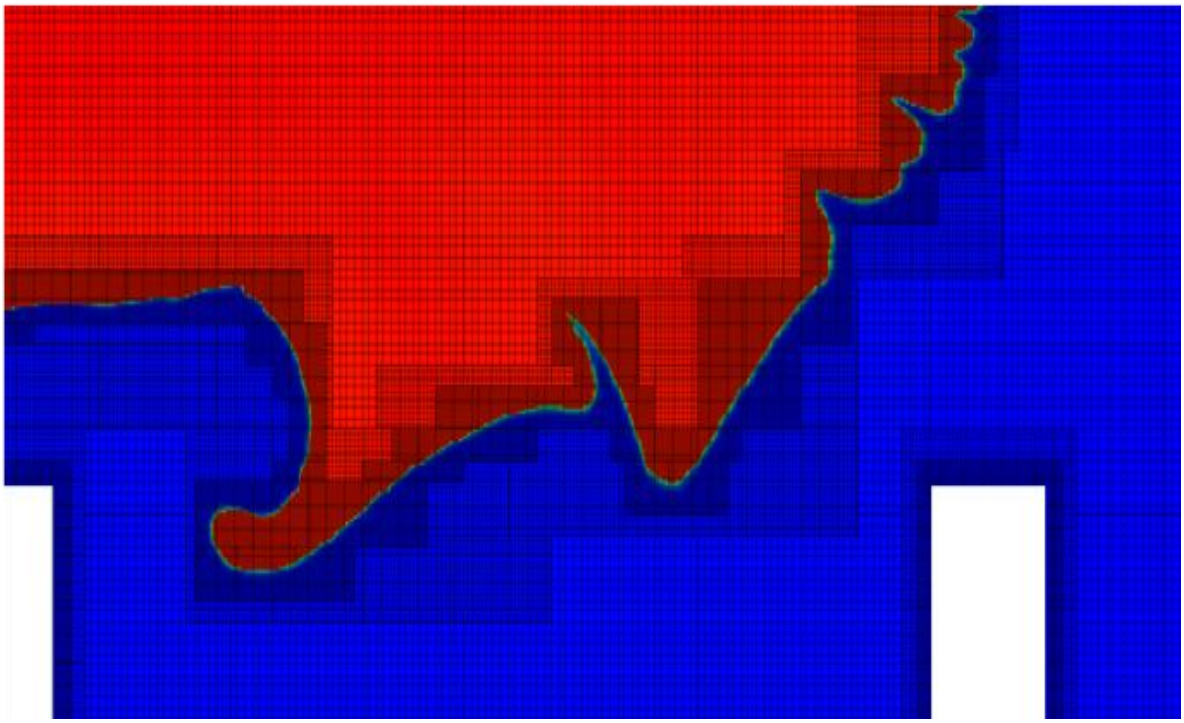


Figure 3: Adaptive Mesh Refinement (AMR)

DDT Process

The deflagration-to-detonation process which occurred in the simulations is seen in the figure below and has been previously described in [10,12,23]. The process begins as the initial flame propagates at the laminar flame speed S_l though the unburned mixture. The hot products increase the background flow increasing the apparent flame speed. As the flame propagates over the obstacles the flame surface becomes distorted increasing the flame surface area. This increase in flame area is the main cause of the increased burning rate which further accelerates the flame. The fast flame will then begin to generate acoustic waves which will eventually coalesce into a shock. As these shocks reflect from the obstacles, the shock-flame interaction (Rayleigh-Taylor instability) causes more distortion in the flame front increasing its velocity. This shock-flame complex will propagate at a velocity below the D_{CJ} in a decoupled state. This shock will continue to gain strength until its reflection from an obstacle increasing the temperature to the ignition temperature, thus forming a hotspot from which DDT will occur. This resulting detonation wave will propagate over the obstacles occasionally extinguishing due to rarefaction over an obstacle, only to subsequently detonate again.

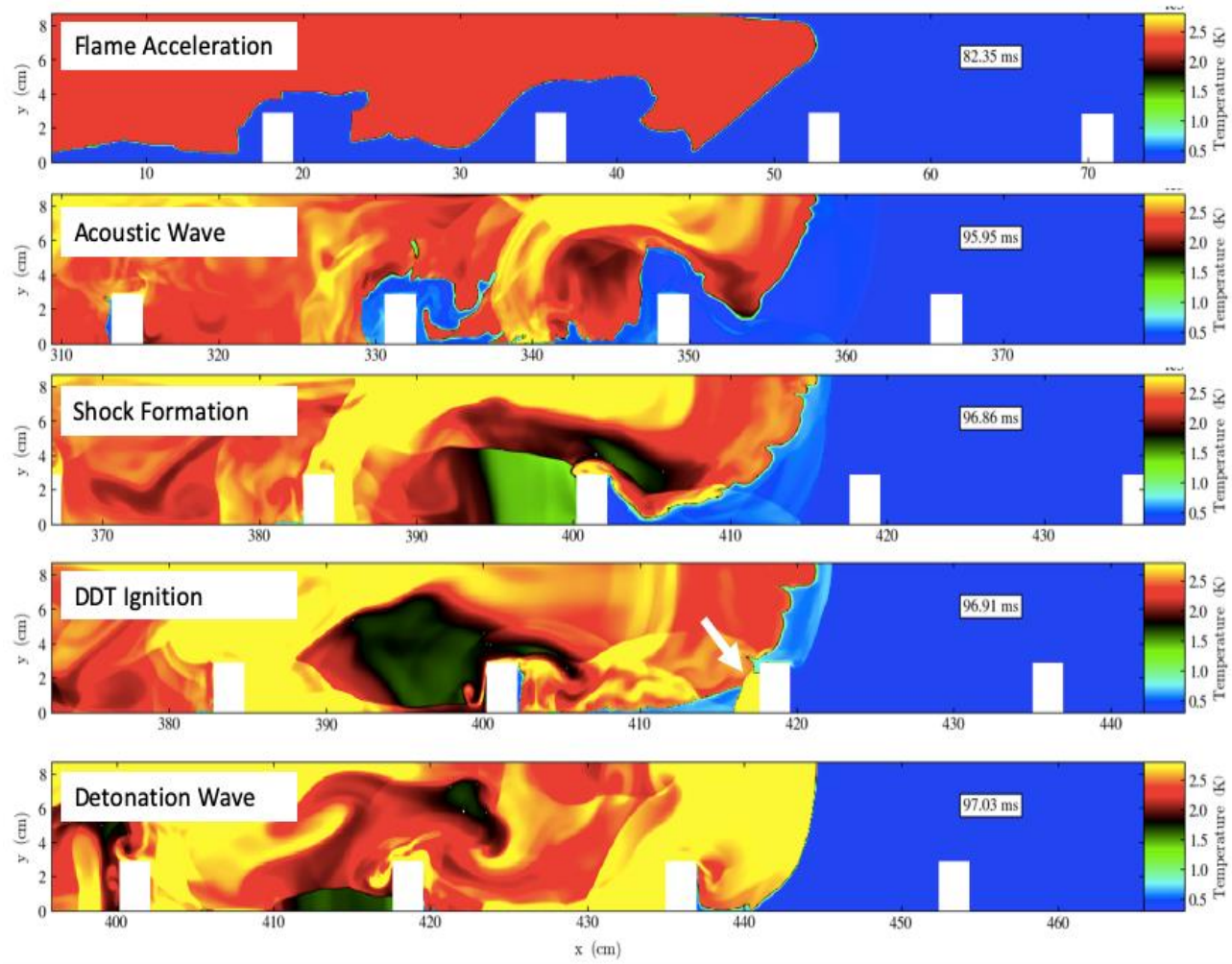


Figure 4: DDT Process

Results and Discussion

We repeated the simulations for each methane-air mixture containing various amounts (0%, 1%, 2%, 4%, 6%, 8 %) of heavy hydrocarbon impurity at each channel size ($d = 17.4cm$, $d = 52cm$, $d = 1m$). Flame speed is tracked as a function of position and time as it propagates down the channel. Detonation initiation starts when the flame velocity reaches the Chapman-Jouguet velocity D_{CJ}. The detonation wave will overshoot this steady-state velocity before settling at the D_{CJ} velocity. Flame velocities for each case are tracked in Figure 5.

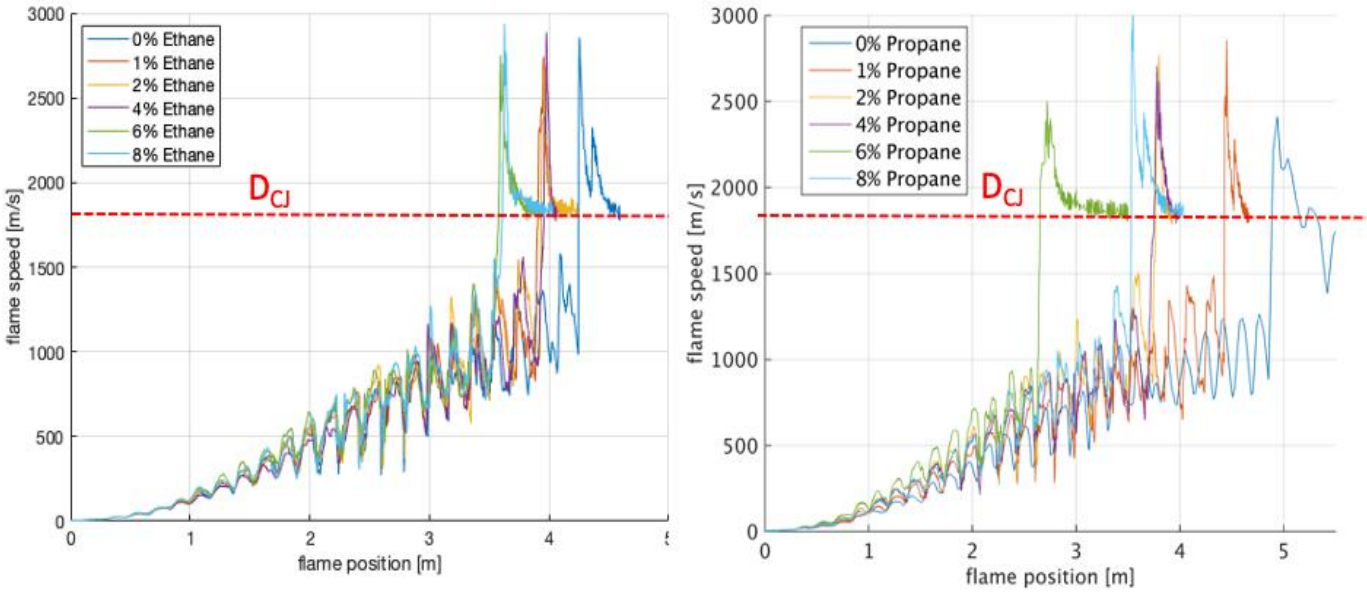
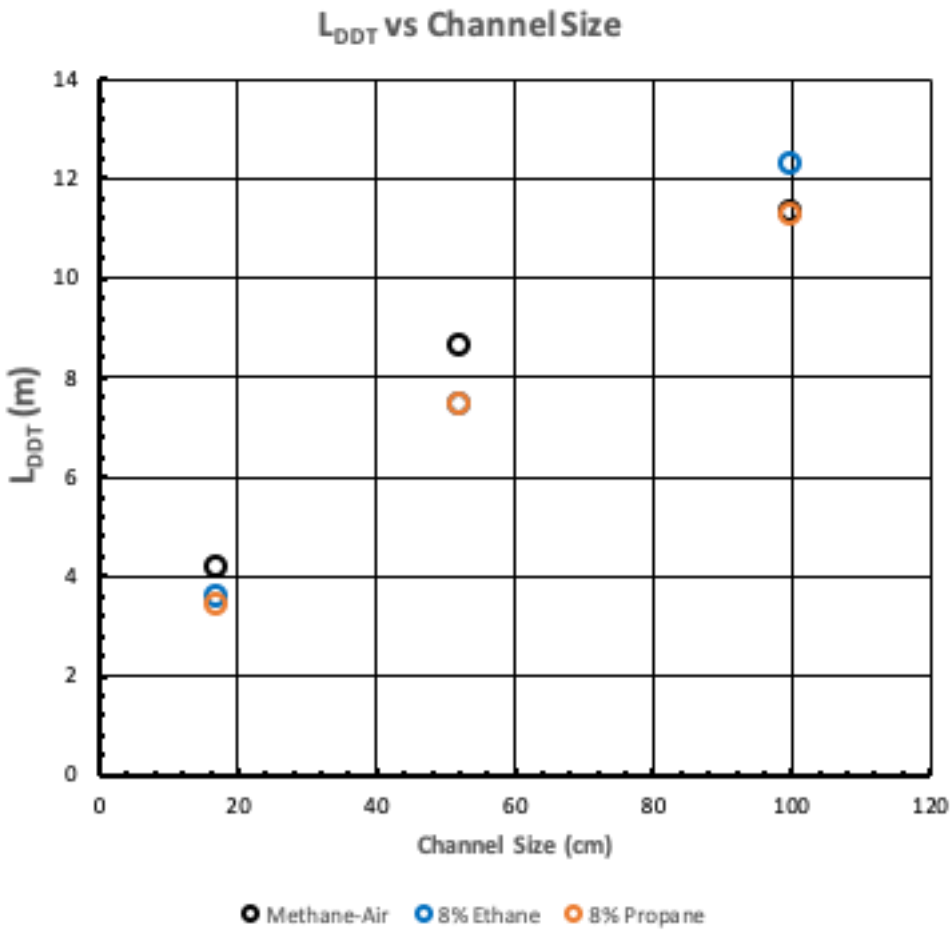
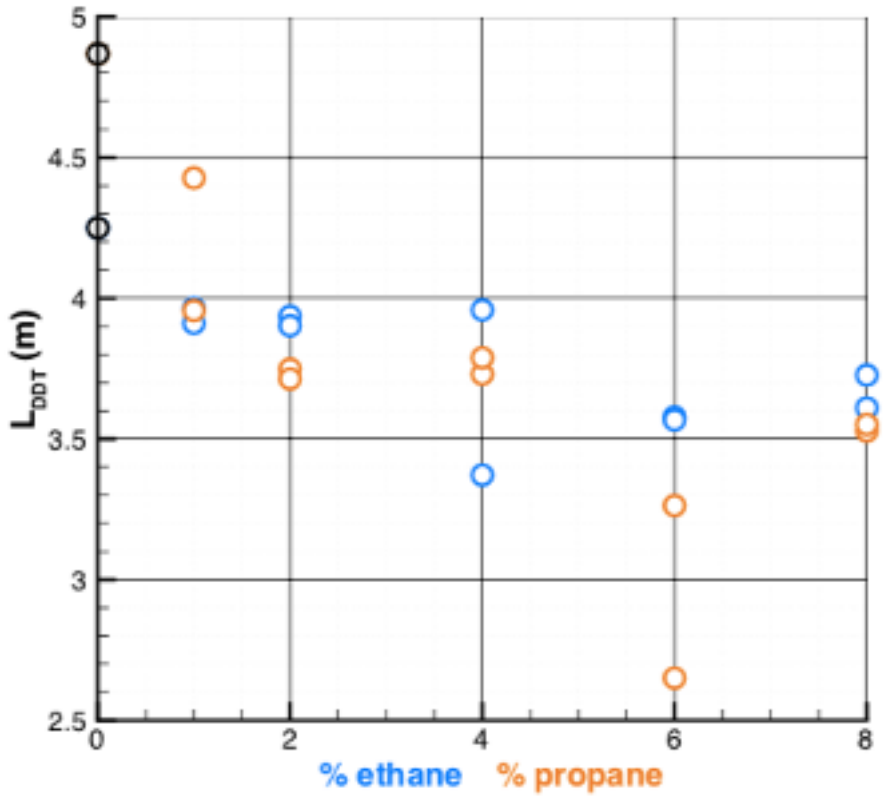


Figure 5: Flame speed as a function of channel length

The numerical simulations show L_{DDT} for blended mixtures contained in channels with periodic obstacles, and found that the effect of ethane or propane concentration on L_{DDT} is relatively weak and difficult to estimate due to stochastic variations of L_{DDT}. For the 17.4cm channel, both ethane and propane seem to reduce L_{DDT} by about the same 15% at concentration 2%, and by 20% at concentration 8%. For larger channels, the scatter increases, and the data do not show a clear trend. Though it is clear L_{DDT} scales nearly linearly with increased channel diameter. As the channel diameter increases, the it takes a longer time for the flame to accelerate to conditions where DDT can occur.


L_{DDT} vs Channel Size
L_{DDT} (m)
Channel Size (cm)
0
2
4
6
8
10
12
14
0
20
40
60
80
100
120
Methane-Air
8% Ethane
8% Propane



This is consistent with the fact that small ethane or propane concentrations have little effect on the laminar flame speed and flame temperature in the resulting mixture. For example, increasing the propane concentration from 0 to 8% increases the laminar flame speed S_l of the stoichiometric methane-propane-air mixture by 6.5% while the flame temperature T_b varies by about 1%. Since S_l and T_b control the flame evolution for the most part of L_{DDT} in obstructed channels, the resulting effect of propane concentration on L_{DDT} is small. On the other hand, the same increase in propane concentration from 0 to 8% decreases the length of the reaction zone in a detonation wave x_d by a factor of 2.8. This should facilitate the shock-induced ignition that is involved at the last stages of DDT, but these last stages are very short compared to the preceding flame evolution in obstructed channels, and thus have little effect on L_{DDT}. A large reduction in x_d suggest a reduction in the detonation cell size. A smaller cell size indicates a more robust detonation wave, thus harder to extinguish.

Propane					
Impurity	$S_L(cm/s)$	$D_{CJ}(m/s)$	$T_b(K)$	$x_{ft}(cm)$	$x_d(cm)$
0%	37.7	1800	2229	0.0448	1.67
1%	38.63	1813	2225	0.0434	1.276
2%	39.03	1807	2234	0.0435	1.056
4%	39.81	1822	2234	0.0442	0.817
6%	39.98	1813	2225	0.0417	0.678
8%	40.17	1828	2206	0.0409	0.591

Table 2: Propane Characteristics

We thus conclude that up to 8% concentrations of ethane or propane in blended methane-air mixtures have no practical effect in reducing the distances to DDT in obstructed channels. The estimated L_{DDT} for blended mixtures are basically the same as for pure methane-air.

Conclusions

Numerical simulation of DDT in methane filled channels with trace percentage (0%-8%) propane or ethane showed little effect on reduce the run-up distance to detonation (L_{DDT}). The the variance in L_{DDT} is on the order of the stochasticity of the simulations (variance do to turbulence and turbulence-shock interactions). Hydrodynamics scaled linearly with larger channels, but the chemical models did not scale with larger channel diameter. Increased heavy hydrocarbon content slightly increased the laminar flame speed (S_l) and adiabatic burning temperature (T_b) which are the primary driver in the DDT process. Increased heavy hydrocarbon did increase the half-reaction thickness (x_d) which suggest a smaller detonation cell size and thus a more robust detonation.

Acknowledgments

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23rd Annual Process Safety International Symposium
October 20-21, 2020 | College Station, Texas

The Use of Bent Poles as a Detonation Indicator

J. Kelly Thomas, W.B. Lowry and B.L. Bingham
Baker Engineering and Risk Consultants
3330 Oakwell Court, Suite 100
San Antonio, TX 78218-3024
(210) 824-5960
KThomas@BakerRisk.com

The response of "slender columnar objects" (e.g., scaffold poles, fence posts, lamp posts, etc.) to the loads imposed by a vapor cloud explosion (VCE) has been recently identified as an indicator of whether a VCE was a deflagration or a detonation. It has been suggested that a slender columnar object deformed such that it has "continuous curvature" rather than a "hinge" is indicative of a detonation. The purpose of the work described in this paper was to examine the response of simple poles to the blast and drag loading from VCEs involving disc-shaped clouds in order to determine the validity of pole response as a detonation indictor.

The blast and drag loading from VCEs involving disc-shaped clouds was evaluated using BakerRisk's Blast Wave Target Interaction (BWTI™) computational fluid dynamics (CFD) code. Cloud diameter to height (D/H) ratios of 50, 100 and 200 were evaluated; cloud dimensions were selected to preserve cloud volume. A hemispherical cloud was also evaluated (D/H = 2). The largest D/H ratio is representative of that involved in the Buncefield and Caribbean Petroleum accidental VCEs. CFD analyses were performed for flame speeds of Mach 0.7 (fast deflagration), 1.0 (very fast deflagration), and 5.2 (detonation). Blast and drag loads were determined for selected locations both within and external to the flammable cloud. As part of this work, dimensionless positive phase peak drag pressure and drag impulse curves were developed for each cloud geometry and flame speed assessed.

The LS-DYNA finite element analysis (FEA) code was used to evaluate the response of a simple pole geometry to the predicted blast pressure and drag load (i.e., density and velocity) histories. A vertical pole fixed into the ground (i.e., a cantilevered beam) was considered in this analysis. A pole outer diameter of 0.5 inches (1.3 cm) was employed in all cases, with the pole wall thickness selected to ensure a reasonable level of response to the predicted blast and drag loading (i.e., the thickness evaluated varied with location, cloud geometry and flame speed). The pole height evaluated was fixed based on the tallest disc-shaped cloud evaluated (i.e., D/H = 50). Both the deformation of the pole during the blast and drag loading and the relative velocity of the pole and

gas were considered in the analysis. A simplified treatment of the reflection of the blast wave from the pole face was used.

The results of this work indicate that the presence of continuous curvature in a simple pole is not necessarily indicative of a detonation. Continuous curvature can be obtained with a deflagration, and may or may not be present with a detonation, depending on cloud geometry and pole location. In fact, the results of this work indicate that the presence of continuous curvature in a simple pole is more likely in a deflagration than in a detonation, although it may occur with either.

Keywords: VCE, CFD, FEA, detonation, blast, drag, detonation indicator, pole damage

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Gradient Boosting Based Quantitative Prediction Models for Mixture Flammability Limits

Zeren Jiao*, Chenxi Ji, Shuai Yuan, Zhuoran Zhang, Qingsheng Wang
Mary Kay O'Connor Process Safety Center
: jiaozeren@tamu.edu

Abstract

Flammability limits (FL), including lower flammable limit (LFL) and upper flammable limit (UFL), are crucial for fire and explosion hazards assessment and consequence analysis. In this study, by using an extended FL database of chemical mixture, quantitative structure-property relationship (QSPR) models have been established using gradient boosting (GB) machine learning algorithm. Feature importance based descriptor screening method is also implemented for the first time to determine the optimal set of descriptors for model development. The result shows that all developed models have significantly higher accuracy than other published models, with the test set RMSE of LFL and UFL models being 0.058, 0.129, respectively. All the developed QSPR models can be used to obtain reliable chemical mixture FL estimation and provide useful guidance in fire and explosion hazard assessment and consequence analysis.

Keywords: Machine Learning, Flammability Limit, Quantitative Structure-Property Relationship, Gradient Boosting

1. Introduction

Flammability properties of chemicals are extremely important in fire and explosion protection. Lower flammability limit (LFL) and upper flammability limit (UFL) is the minimum/maximum concentration of the chemicals that can cause fire or explosion with the presence of the ignition source such as flames, static electricity sparks etc.[1-2] Experimental measurement is the most commonly used method to determine the chemical FL value.[3] However, the experimental setup of FL measurement is expensive, and most of the chemicals are highly flammable and potentially explosive, which is extremely dangerous to conduct the flammability test.[4-6] Besides, chemical mixtures have a various combination which is highly time-consuming and almost impossible to measure all mixture combinations.[7-8] Although Le Chatelier's law can provide a reasonable estimation of flammability limits of fuel mixtures. It is only accurate with certain mixture LFL and mixture UFL with low pure chemical UFL.[3,9-10]

Many research has been done in fire and explosion related property evaluations.[11-12] Prediction of physical properties using quantitative structure-property relationship (QSPR) models has been intensively studied recently due to its high prediction accuracy and reliability.[13-15] However, a QSPR model with high accuracy primarily needs an extensive database that is broad-spectrum applicable to various kinds of novel chemicals. DIPPR 801 database is the most used property database that has more than 1,000 chemicals' properties readily accessible.[16-17] However, it only includes the property of pure chemicals. It is particularly hard to construct a database that is large enough for a valid model development which is the reason why only limited QSPR studies focused on mixture property. Most mixture property research does not have a database larger than 100 chemical mixture.[18-20]

In QSPR study, molecular structure optimization and descriptor calculation is the key to determining whether the suitable subset can be found for model development. Different software are utilized in molecular descriptor calculation and statistical model construction, Dragon is the most popular and powerful descriptor calculator which is capable of calculating more than 5,000 different descriptors.[21] However, it is a commercial software which is not free for use and the descriptor library will not update since it has already been discontinued years ago. These disadvantages will impede the expansion and optimization of the current model. However, programming language Python has grown significantly in recent years with many packages available for chemical informatics.[22-24] Its calculation results have also been proven to be consistent with mainstream descriptor calculation software like Dragon and Gaussian for QSPR studies.[21] The generic algorithm is the commonly used descriptor screening method in safety-related QSPR research, however, the detailed selection mechanism is still unclear and the method is not able to rank the calculated descriptors in a manner of importance in model development for intuitive descriptor screening.

For the regression method, multiple linear regression (MLR) is the standard method for constructing the QSPR model which shows satisfactory predictability and reliability.[25] But the relationship between descriptors and designated properties is nonlinear and the interaction mechanism remains unknown. Forcing a linear regression to the dataset will oversimplify the relationship and lead to inaccurate results. Implementing machine learning into prediction model development is one of the most popular fields in recent years, and many machine learning methods show promising potential for applications in QSPR research.[26-27] Support vector machine (SVM) is the most commonly used machine learning regression method applied to QSPR studies since it has an easy-to-use LIBSVM package.[28] Other than the SVM method, many machine learning

methods can be applied to QSPR study which shown good predictability in other engineering and chemistry related research.[29-32] Yuan et al. applied three machine learning methods to develop QSPR models for the upper flammability limit (UFL) of pure organic chemicals and compared the performance of these models to the traditional MLR model. The result shows that the random forest based model has the best performance.[27] However, the gradient boost algorithm, which is showing a promising result in regression analysis and superior performance than SVM and random forest, has never been applied in safety-related QSPR analysis.[33]

Some researches have been done about LFL mixture prediction, Wang et al. developed a QSPR model based on 86 chemical mixture data and Pan et al. further expanded the database to include more chemical mixtures that showed better performance than Wang et al.'s work.[4,6] Jiao et al. compare different machine learning regression method in LFL mixture model development and they found that the random forest based model tend to outperform MLR and other machine learning method.[8] However, all of their work has limited data available for model development and the database need to be further expanded to include more available experimental data. For mixture UFL prediction models, Wang et al. used support vector machine (SVM) to develop QSPR model of 78 fuel mixture and Shen et al. further improved the model by expanded the database to 86 chemicals, however, both of their developed models are based on the same dataset and the models developed has high test set RMSE, which mean that the UFL prediction model needs to be improved for reliable application.[5,7]

The standard procedure of QSPR model development is shown in Figure 1. In this study, an expanded database consisting of 271 mixture LFL data and 138 mixture UFL data was constructed to develop the QSPR model. The chemical molecular structures were first optimized via Python Psi4 and Rdkit package, and descriptors are then calculated by Mordred package.[21-24] A novel gradient boosting descriptor screening method is proposed to select the optimal subset of molecular descriptors calculated and develop the final QSPR prediction models.[34] Statistical assessment of developed QSPR models was also conducted to validate and test their accuracy, internal robustness, and applicability domain.

2. Methodology

2.1 Database

In this study, multiple experiment results are obtained from literature to construct the mixture FL database. The extended data set contains 271 different mixture LFL data of 15 chemicals and 138 mixture UFL data of 12 chemicals, which is the largest database among all published mixture FL QSPR papers.[35-37] The chemicals in this database contain various chemicals including saturated hydrocarbons (methane, ethane, propane), unsaturated hydrocarbons (acetylene, ethylene, propylene, 1-butane, butadiene), hydrocarbon isomers (n-butane and isobutane), ether (dimethyl ether), ester (methyl formate), halogenated compound (1,1-difluoroethane) and inorganic compounds (ammonia and carbon monoxide), which can ensure its broad-spectrum applicability. The detailed composition and experimental data can be found in the Supporting Information. Subsequently, 75% of data points (203 in LFL database, 103 in UFL database) in the FL database were randomly selected to the training set and the remaining 25% of data points (68 in LFL database, 35 in UFL database) will be grouped in the test set to validate the accuracy of the models.

2.2 Molecular Descriptors Calculation and Determination

In this study, Mordred is employed as the molecular descriptor calculator. Mordred is an open-source Python package which is at least twice as fast as the well-known PaDEL-Descriptor and it can calculate descriptors for large molecules, which cannot be accomplished by other software.[21] Since the molecular structure can significantly affect the calculation result of molecular descriptors, the chemical structures should be optimized before calculation. The 3D structure of chemicals are obtained from the PubChem database, the obtained structures are then optimized using MMFF force field by Python package Rdkit, which is proved to be more accurate than semi-empirical AM1 force field in QSPR study.[24,38] The MMFF optimized structures are then optimized by the B3LYP density functional method with 6-31G(d) basis set in Python package Psi4.[24] The optimized molecules are finally uploaded to Python package Mordred for descriptor calculation. As a result, 541 descriptors were calculated for each chemical in the database for the descriptor screening process.

For the mixture descriptor determination, there are multiple methods including weighted average, normalized weighted average, square-root mole fraction, etc. Among those methods, assuming molecular descriptors vary linearly with mole fraction is the most widely used method in other mixture QSPR study which is also been proved to be the best and most efficient way to determine mixture descriptors. Besides, Le Chatelier's method, which is the weighted average of pure compound FL, is used to be the method to determine the mixture FL. Therefore, in this study, the weighted average method is applied to determine mixture molecular descriptors.[39] The descriptors of the mixture are the sum of each component of descriptor value times the mole fraction of that component:

$$D_{mix} = w_1 D_1 + w_2 D_2 + \cdots + w_n D_n$$

where D_{mix} is the mixture descriptor value, w_i is the mole fraction of chemical i, D_i is the molecular descriptor value of chemical i.

2.3 Gradient Boosting Based Descriptor Screening and Model Development

Gradient boosting (GB) is a machine learning algorithm in the form of an ensemble of weak prediction models, typically decision trees. It builds the model in a stage-wise fashion as other boosting methods do, and it generalizes them by allowing optimization of an arbitrary differentiable loss function.[33] The idea of gradient boosting originated from the observation by Leo Breiman that boosting can be interpreted as an optimization algorithm on a suitable cost function.[40] Explicit regression gradient boosting algorithms were subsequently developed by Jerome H. Friedman.[41] A simplified gradient boosting diagram is shown in Figure 2. Instead of fitting the data hard, the boosting tree method learns slowly which accomplished by only fit the predictors to the updated residuals from the previous tree. The main algorithms of gradient boosting regression are shown below:[42]

2.4 Applicability Domain

According to the Organisation for Economic Co-operation and Development (OECD) principle 3, an applicability domain should be defined for developed QSPR models.[43] The purpose of the applicability domain is to determine whether the model's assumptions are met and for which the QSPR model can be reliably applied. In this study, the applicability domains of mixture FL QSPR models are obtained and evaluated through Williams plot, which is the data set cross-

validation standardized residuals vs. leverage values (h_i). If the leverage value of certain data point is larger than the standard leverage value (h^*), or the standardized residual of a certain data point is larger than 3, the data point should be considered out of the applicability domain.[44] The leverage value and standard leverage value can be calculated by the following equation:

$$h_i = X_i(X^T X)^{-1} X_i^T$$

$$h^* = \frac{3(k+1)}{m}$$

where, X_i is the row-vector of mixture descriptors, X is the training set mixture descriptor matrix, k is the descriptor selected for model development, m is the number of mixture data in the training set.

3. Results and Discussion

3.1 LFL Mixture

The descriptor importance plot of the mixture LFL QSPR model is shown in Figure 4. The descriptor importance matrix and description of selected descriptors are also shown in Table 1. The gain of a descriptor is the average training loss reduction gained when using a feature for splitting. It can be seen from both plot and table that the *AMID_C* is the most important descriptor in gradient boosting based mixture LFL QSPR model development. The total descriptor amount used in final descriptor development is usually 4-6 in order to ensure accuracy without overfitting.[4-8] In this study, the screening threshold of the LFL model is set to 0.01 which results in *AMID_C*, *Mor06p*, *ATS0p*, *TASA*, and *ATS0v* selected as the subset for final model development.

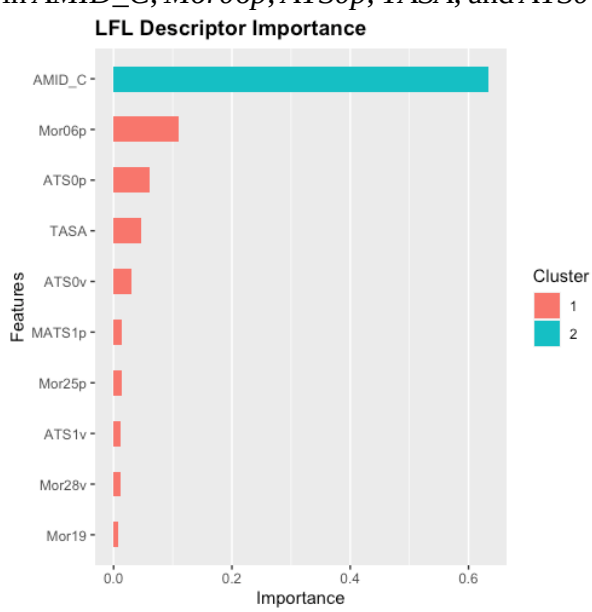

Figure 4. Mixture LFL model descriptor importance plot

Table 1. Mixture LFL descriptor importance matrix and description of selected descriptors

Descriptor	Gain	Description
AMID_C	0.63266410	Averaged molecular ID on C atoms

Mor06p	0.11097867	3D-MoRSE weighted by polarizability (distance = 6)
ATS0p	0.06008590	Moreau-broto autocorrelation of lag 0 weighted by polarizability
TASA	0.04620432	Total hydrophobic surface area
ATS0v	0.03076566	Moreau-broto autocorrelation of lag 0 weighted by vdw volume

3.1.2 Model Performance

Based on the parameter setting that can reach the lowest test RMSE, the *eta*, *max_depth*, and *nround* of the QSPR model are set to 0.2, 10, and 400, respectively.

In order to visualize the performance of the developed QSPR models, two types of performance evaluation plots are shown in Figure 6. The plot of QSPR prediction values vs. experimental values is shown in Figure 6a and the prediction residual plot is shown in Figure 6b with the test set statistical values as R^2=0.986 and *RMSE*=0.058. All data points are evenly distributed along the diagonal baseline of Fig. 6a which indicates that the predicted mixture LFL is very close to the experimental value. Furthermore, all of the data residuals shown in Fig. 6b are less than 0.2 and also close to the zero baseline, which proves that the developed QSPR method provides a good estimate of mixture LFL.

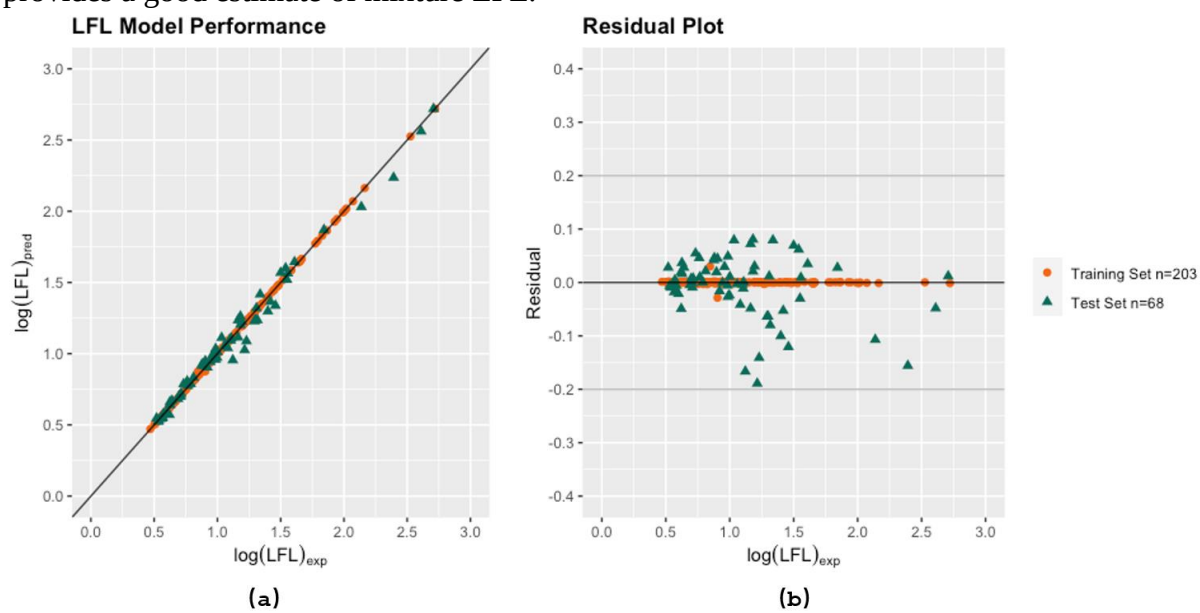

Figure 6. Mixture LFL model performance plot: (a) Predicted value vs. experimental value, (b) Residual plot

Williams plot the mixture LFL model is shown in Figure 7. The applicability domain is the area within ±3 standardized deviations and standard leverage h^* of 0.103. We can see that all data points are located inside the applicability domain. Therefore, the developed model is confirmed to have the capability in predicting chemical mixtures LFL within the corresponding applicability domain.

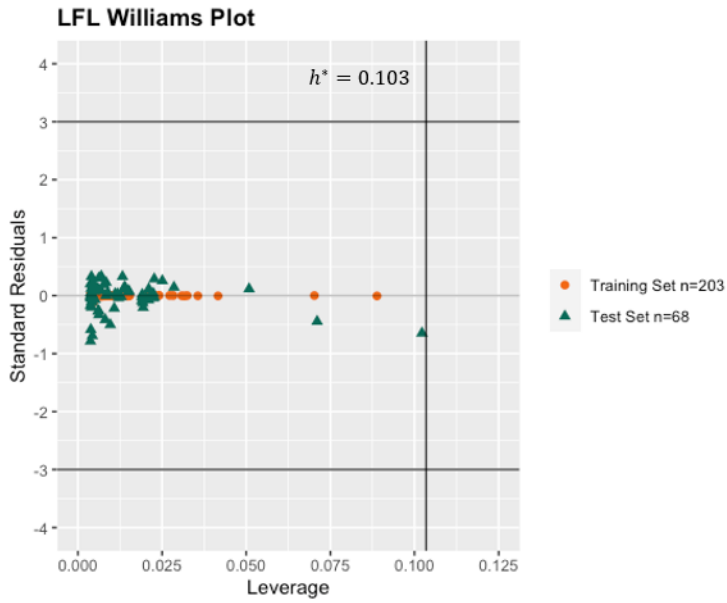

Figure 7. Mixture LFL model Williams plot

3.2 UFL Mixture

The descriptor importance plot of the mixture UFL model is shown in Figure 9. The descriptor importance matrix and description of selected descriptors are also shown in Table 1. It can be seen from the plot that the ten most important descriptors are clustered into three groups. The first two groups contain most of the regression gain among all calculated descriptors. And the *Mor21p* is the most important descriptor in developing UFL QSPR model development. In this study, the descriptors of the first two clusters are selected as the subset for final model development.

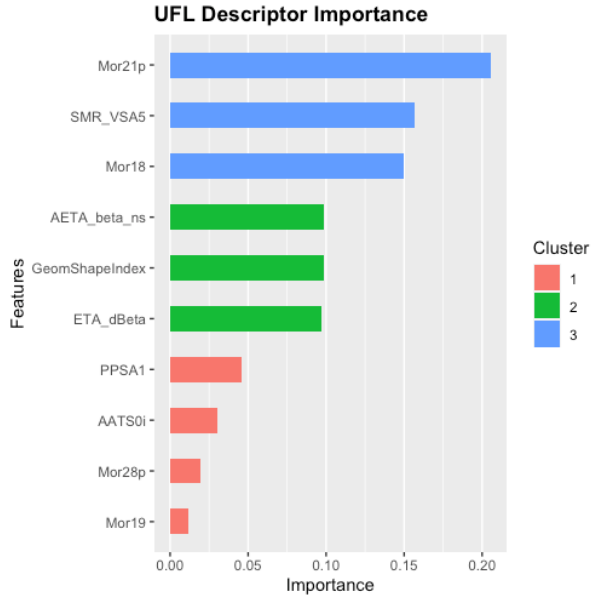

Figure 9. (a) Mixture UFL model descriptor importance plot

Table 3. Mixture LFL descriptor importance matrix and description of selected descriptors

Descriptor	Gain	Description
Mor21p	0.20539972	3D-MoRSE weighted by polarizability (distance = 21)
SMR_VSA5	0.15635616	MOE MR VSA Descriptor 5 (2.45 <= x < 2.75)
Mor18	0.14959226	3D-MoRSE (distance = 18)
AETA_beta_ns	0.09862583	Averaged nonsigma contribution to valence electron mobile count
GeomShapeIndex	0.09827003	Geometrical shape index
ETA_dBeta	0.09674492	ETA delta beta

3.2.2 Model Performance

Based on the parameter setting that can reach the lowest test RMSE, the *eta*, *max_depth*, and *nround* of the QSPR model are set to 0.4, 10, and 200, respectively.

The plot of predicted UFL values vs. experimental UFL values is shown in Figure 11a. Since the UFL data naturally have a wider spread than LFL data, the test set data points are more sparsely distributed along the diagonal baseline compared to the mixture LFL models. The prediction residual plot in Figure 11b also shown the same trend. However, the test set statistical values of R^2=0.935 and *RMSE*=0.129 demonstrate that the UFL QSPR model still has satisfying accuracy.

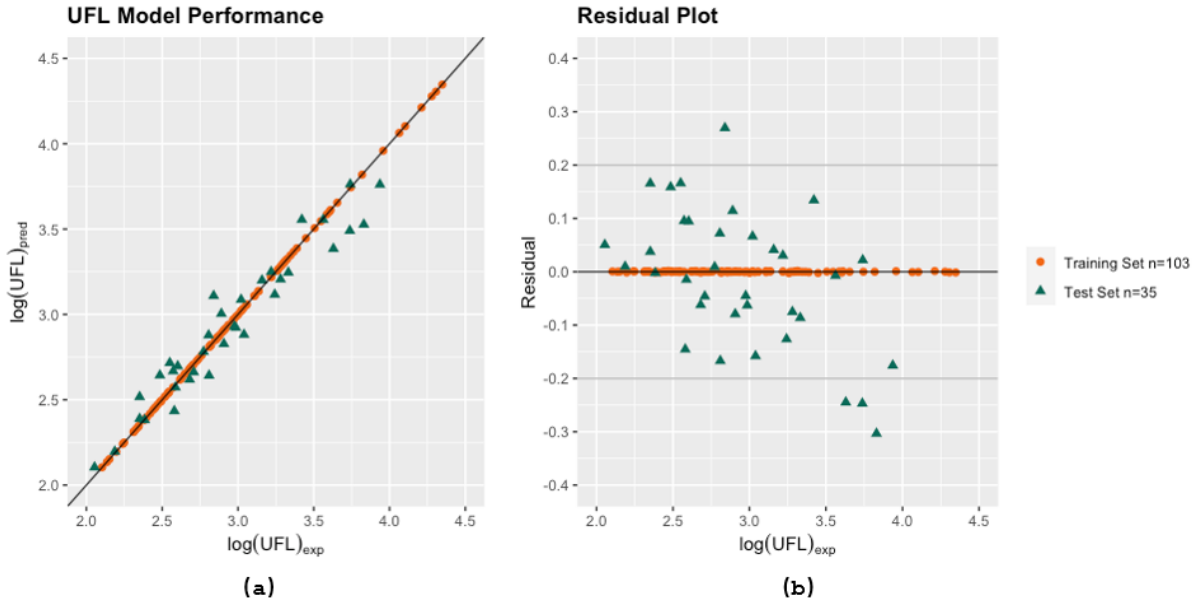

Figure 11. Mixture UFL model performance plot: (a) Predicted value vs. experimental value, (b) Residual plot

Williams plot the mixture UFL model is shown in Figure 12. The applicability domain defined as the area within ±3 standardized deviations with standard leverage h^* of 0.175. It can be seen from the figure that all data points are well-contained under the applicability domain, which indicates that the developed QSPR model's reliability in predicting chemical mixtures UFL within the corresponding applicability domain.

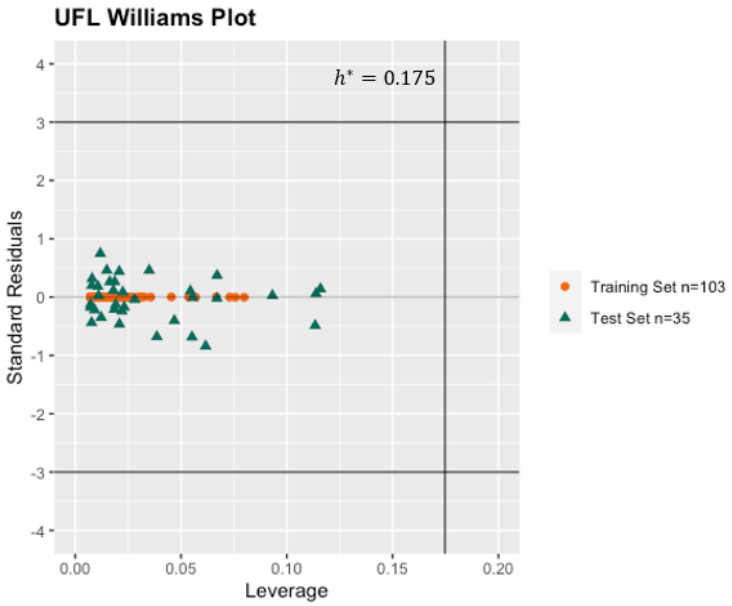

Figure 12. Mixture UFL model Williams plot

4. Conclusions

In this study, the largest database of mixture FL is constructed for the development and validation of the QSPR prediction model. The gradient boosting machine learning method is employed for the first time in the QSPR study to provide a novel descriptor screening and regression method to improve the prediction accuracy of QSPR analysis. The GB-based mixture FL QSPR models shown significantly high accuracy and reliability with the test set RMSE for mixture LFL, mixture UFL of 0.058, 0.129, respectively. The performance of developed QSPR models illustrated the power of gradient boosting descriptor selection and regression in assisting hazardous property prediction. The statistical assessment result also proved that the models developed in this study can reliably apply to flammable chemical mixture hazard assessment and novel chemical design with much higher efficiency and accuracy.

The machine learning based QSPR models do not have a specific equation for an intuitive application like MLR based QSPR model which is one of the shortcomings of GB-based QSPR models. However, the problem can be overcome by developing a built-in software package with interactive user interface, since the whole development process is conducted through open source programming, the free distribution with high expandability can be easily reached in the future. Besides, the database used in this study still needs to be further expanded to include as much chemical mixture FL data as possible to further improve its applicability. And the descriptor calculation capability also needs to be improved to expand available descriptors for screening. In future QSPR model development, it is recommended that the GB-based methods should primary considered for an effective and accurate prediction.

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Speaker Bios

Day 1 Track 1
Risk/Consequence Analysis & Design Aspects
Risk Assessment I

Shanmuga Prasad Kolappan
TechnipFMC

Importance of Process Safety Time in Design Concept

Shanmuga Prasad Kolappan is a driven Process Safety Engineer with considerable rich experience in the field of Safety, Risk and Loss prevention engineering. He has 9 years of experience including Process and commissioning, risk consulting, and loss prevention. He has also had exposure to safety studies such as Hazop, SIL, QRA, and active/ passive fire protection, as well as to safety software such as PHA Pro, PHA works, PHAST, SAFETI, exSILentia, Pipenet, Detect3D and BowTieXP. Shanmuga is continuously learning and updating the recent trends in the field of Safety.

Abdulaziz Alajlan
Saudi Aramco

Limitations of Layers of Protection Analysis (LOPA) in Complicated Process Systems

An experienced Process Safety and Risk Engineer with 20 years of experience in different regions and countries. Author of multiple scientific papers and chairman of safety conferences.

Johannes I. Single
CSE Center of Safety Excellence

On the Usage of Ontologies for the Automation of HAZOP Studies

Johannes I. Single is a researcher at the CSE Center of Safety Excellence in Pfinztal, Germany and he is currently pursuing his PhD at the University of Kaiserslautern. Previously, Johannes worked as a research engineer in the field of software engineering for CSE and interned at BASF and GEA Group. He graduated from Karlsruhe University of Applied Sciences and holds a master's degree in mechanical engineering specializing in process engineering.

Risk Assessment II

Henrique (Henry) M. Paula
Galvani Risk Consulting, LLC

An Efficient and Effective Approach for Performing Cost Benefit Analysis, with Two Case Studies

Dr. Henrique Martini Paula has 40+ years of engineering experience with expertise in integrity management, risk management/PSM activities, risk and reliability analyses, and project quality management. He participated in more than 100 projects in a variety of industries from offshore oil & gas to nuclear to the petrochemical industry. Dr. Paula has provided consulting and training services in over 30 countries, including instructing for the American Petroleum Institute (API), the American Institute of Chemical Engineers (AIChE), and the Process Safety Institute (PSI). He has authored/co-authored well over 100 documents, including journal articles, conference papers, technical reports, and sections of "Guidelines for Chemical Process Quantitative Risk Assessment" and "Guidelines for Developing Quantitative Safety Risk Criteria" (both by AIChE). He served as guest editor for a special issue of the journal Reliability Engineering and System Safety.

Keith Brumbaugh
aeSolutions

Does Your Facility Have the Flu? How to Use Bayes Rule to Treat the Problem instead of the Symptom

Keith Brumbaugh is currently serving as a Discipline Lead in aeSolutions' Safety Instrumented Systems Engineering department, with over fifteen years of experience in instrumentation and safety systems engineering.
Keith is a licensed Professional Engineer (Control Systems - Texas), and holds a Certified Functional Safety Expert (CFSE) certificate. Keith went to Texas Tech University and holds a B.S. in Electrical Engineering, and Minor in Computer Science.

Sam Aigen
AcuTech

Integrating the PHA and Facility Siting into a Site Risk Assessment Life-Cycle

Sam Aigen, CCPSC is a Senior Engineer at AcuTech. Mr. Aigen previously worked in various capacities for ExxonMobil, including the Research and Engineering Company's Central Engineering Office in Virginia and the Refining and Supply Company, based out of the Beaumont Refinery in Texas. He has extensive experience in both process engineering and process design, and is skilled in Process Hazard Analysis, consequence modeling, facility siting, and Quantitative Risk Assessment.

Colin Armstrong
AcuTech

Integrating the PHA and Facility Siting into a Site Risk Assessment Life-Cycle

Colin Armstrong is a technical Lead for numerous FSS and QRA projects in oil, gas, LNG, and specialty chemical industries worldwide. He has experience in all aspects of QRA, consequence modeling, frequency assessment, scenario analysis (FTA, FMEA, event tree, LOPA, etc.) Mr. Armstrong is an instructor of QRA and Consequence Modeling for operating companies and students at University of Maryland. He is also an investigator and expert witness in response to incidents and OSHA citations.

Layers of Protection: Relief Systems

Nitin Roy
California State University

A Framework for Automatic SIS Verification in Process Industries using Digital Twin

Nitin Roy is an Assistant Professor (Safety) at California State University, Sacramento.

Paul Gruhn
aeSolutions

The use of Bayesian Networks in Functional Safety

Paul Gruhn is a Global Functional Safety Consultant with aeSolutions in Houston, Texas. Paul is an ISA (International Society of Automation) Life Fellow, a 30 year member and co-chair of the ISA 84 standard committee (on safety instrumented systems), the developer and instructor of ISA courses on safety systems, the author of two ISA textbooks, and the developer of the first commercial safety system modeling software. Paul has a B.S. degree in Mechanical Engineering from Illinois Institute of Technology, is a licensed Professional Engineer (PE) in Texas, a member of the Control Systems Engineering PE exam committee, and both a Certified Functional Safety Expert (CFSE) and an ISA 84 Safety Instrumented Systems Expert. Paul was the 2019 ISA President.

Greg Hall
Eastman Chemical Company

My Vision of Future Instrumental Protective Systems

Greg Hall is a Principal Electrical Engineer with Eastman Chemical Company with 39 years experience at Texas Operations in Longview, Texas. Greg is the IPS (Instrument Protective Systems) Design engineer, chairman of the Texas Operations IPS Committee, member of the Eastman Corporate IPS Governance Council, and received an Electrical Engineering degree from the University of Texas at Austin.

Relief Systems

Gabriel Martiniano Ribeiro de Andrade
Siemens Process & Safety Consulting

Overlooked Reverse Flow Scenarios

Gabriel Andrade is a lead process engineer for Siemens Energy and has worked in process safety since 2011. Gabriel started his career at Chemtech, a Siemens engineering company in Brazil, after obtaining his chemical engineering Bachelor's degree from the Universidade Federal do Rio de Janeiro in 2010. Passion and dedication brought him to the Siemens process safety group in Houston, where he has happily lived with his wife since 2014.

Christopher Ng
Siemens Process & Safety Consulting

Overlooked Reverse Flow Scenarios

Christopher Ng is a Technical Advisor at Siemens Energy Inc. with over 22 years of process engineering experience in the upstream and downstream industry. He has expertise in process design and modeling, process safety, hazard and risk analysis. He has a bachelor's degree in Chemical Engineering from the University of Western Ontario (Canada) and is a licensed professional engineer in Texas.

Derek Wood
Siemens Process & Safety Consulting

Overlooked Reverse Flow Scenarios

Derek Wood is a lead process engineer for Siemens Energy with 11 years of experience in process engineering in the energy industry, with expertise in process safety, simulation, and design. He has a master's degree in Petroleum Engineering from the University of Texas at Austin.

Todd W Drennen
Baker Risk

Failure Under Pressure: Proper Use of Pressure Relief Device Failure Rate Data based on Device Type and Service

Todd W. Drennen, P.E., is a senior engineer for Baker Engineering and Risk Consultants, Inc. (BakerRisk). He has more than 15 years of experience in process safety, including pressure-relief system design and analysis, simulation of complex process upset scenarios, process hazard analysis (PHA), layers of protection analysis (LOPA), fault tree analysis (FTA), and process safety management (PSM) compliance auditing. He has a BS in chemical engineering from Drexel University and is a licensed professional engineer in Illinois and Delaware.

Gabriel Martiniano Ribeiro de Andrade
Siemens Process & Safety Consulting

Additional Engineering and Documentation to Reduce Pressure Relief Mitigation Cost

Gabriel Andrade is a lead process engineer for Siemens Energy and has worked in process safety since 2011. Gabriel started his career at Chemtech, a Siemens engineering company in Brazil, after obtaining his chemical engineering Bachelor's degree from the Universidade Federal do Rio de Janeiro in 2010. Passion and dedication brought him to the Siemens process safety group in Houston, where he has happily lived with his wife since 2014.

Kartik Maniar
Siemens Process & Safety Consulting

Additional Engineering and Documentation to Reduce Pressure Relief Mitigation Cost

Kartik Maniar is a principal process engineer for Siemens Energy and has worked in the field of process safety since 2006. He has lead and completed various refinery wide pressure relief and flare studies. Recent work has included relief studies based on dynamic simulation and also flare load minimization using non-normal devices and instrumentation credit.

Day 1 Track 2
Human Factors-People In Action
Training/ Engagement

Kianna Arthur
Texas A&M University

Virtual Reality Process Safety in Counterfactual Thinking

Kianna Arthur is a second-year PhD student in the Social-Personality Psychology program. She works with Dr. Rachel Smallman in examining the functionality of counterfactual thoughts (i.e., "If only...") in health-related contexts. This includes both the generation and subsequent consequences of counterfactuals (motivation, behavioral intentions, risk perception, etc.). Kianna also works with Next Generation Advanced Procedures and RIHM Lab with Dr. Camille Peres.

Human Performance/Decision Making I

S. Camille Peres
Texas A&M University

Is Attentional Shift the Problem (or something else) with Hazard Statement Compliance? An Experimental Investigation Using Eye-Tracking Technology

Dr. Camille Peres is an Associate Professor with Environmental and Occupational Health at Texas A&M University as well as the assistant director of Human Systems Engineering with the Mary Kay O'Connor Process Safety Center. Her expertise is Human Factors and she does research regarding: procedures; Human Robotic Interaction in disasters; and team performance in Emergency Operations.

Hans J. Pasman
Mary Kay O'Connor process Safety Center

Risk Management entails decision making: Does design decision making in complex situations come down to somebody's gut feeling?

Dr. Hans J. Pasman studied chemical technology at Delft University of Technology, the Netherlands. Ph.D. in 1964. He worked for Shell before moving to the research organization TNO. He has investigated numerous process industry accidents, worked on a variety of topics and managed units of TNO Defense research. 1980-90s Chairman NATO group on Explosives, OECD group on Unstable Substances, Chairman European Working Group on Risk Analysis, Chairman European Working Party on Loss Prevention. Dr. Pasman is a Co-founder of the European Process Safety Centre and he coordinated late 90s industrial safety research TNO. He was a Professor of Chemical Risk Management at Delft University for nearly 10 years. He is also member of former Dutch Hazardous Substances Council and since 2008 Research Professor at Mary Kay O'Connor Process Safety Center in the Chemical Engineering department of Texas A&M University.

Fabio Kazuo Oshiro
Monaco Engineering Solutions

Decision Making using Human Reliability Analysis

Fabio Oshiro is a Principal Risk, Safety & Reliability Engineer with over 15 years of experience in the Oil & Gas industry. He has vast experience executing more than 100 risk and reliability studies performed on behalf of clients based throughout the globe for the successful completion of upstream, midstream and downstream projects located mainly in Middle East, Brazil and Africa. In the past 7 years Fabio has been working as Reliability Expert based in Surrey, UK.

Safety Culture and Leadership

Ryan Wong
AFPM

Improving Industry Process Safety Performance through Responsible Collaboration

Ryan Wong is currently a Process Safety Analyst with ExxonMobil with about 7 years of experience in the field of process safety. Prior to moving into the field of Process Safety, Ryan spent approximately 5 years at the ExxonMobil Baton Rouge Refinery in process design, process support, and risk management roles. Ryan graduated from the University of Michigan with a Bachelors in Chemical Engineering.

Shanahan Mondal
AFPM

Improving Industry Process Safety Performance through Responsible Collaboration

Shanahan Mondal is Corporate Process Safety lead for CVR Energy providing oversight across the company's refineries and fertilizer facilities. Shanahan has held several positions in site leadership in disciplines including Operations, Process Engineering, Process Control, and Process Safety. Mr. Mondal holds a B.S. degree in Chemical Engineering from the Massachusetts Institute of Technology and has 23 years of experience in the refining and petrochemical industries.

Stephanie C. Payne
Texas A&M University

How Much Does Safety Culture Change Over Time?

Dr. Stephanie C. Payne is a Professor of Psychology and Faculty Fellow of the Mary Kay O'Connor Process Safety Center at Texas A&M University, College Station, TX. Her program of research on workplace and laboratory safety focuses on the measurement of safety climate, antecedents and consequences of climate, and moderators of these relationships. Dr. Payne's safety research has been published in various safety and psychology journals including Journal of Safety Research and Safety Science. Her safety research has been funded by various agencies including the National Institute of Occupational Safety and Health and the National Academies of Science, Engineering, and Medicine. She has collaborated with multiple chemical and oil and gas companies.

Atif Mohammed Ashraf
Mary Kay O'Conner Process Safety Center

Administering a Safety Climate Assessment in a Multicultural Organization: Challenges and Findings

Atif M. Ashraf is a Research Associate at the Mary Kay O'Connor Process Safety Center – Qatar. He obtained an MSc in Chemical Engineering from Texas A&M University in 2016. Atif's research areas include runaway reactions, dust explosion characterizations, toxic gas dispersion modelling, evacuation modelling, risk communication, and human factors. Since 2017, Atif's research, predominantly in the Middle East, has been focused on understanding and assessing safety climate and culture through the application of psychology and engineering principles.

Procedures

Joseph W. Hendricks
Texas A&M University

A Comparison of Procedure Quality Perceptions, Procedure Utility, Compliance Attitudes, and Deviation Behavior for Digital and Paper Format Procedures

Dr. Hendricks earned a PhD in Industrial and Organizational Psychology from Texas A&M University. He is currently a research associate – senior investigator with the Next Generation Advanced Procedures consortium at Texas A&M University.

Monica Philippart
Ergonomic Human Factors Solution

Practical Writing Tips To Prevent Human Error When Following Procedures

Dr. Philippart specializes in managing operational risks associated with human performance. Her career begun at NASA's Kennedy Space Center, where she applied her mechanical and industrial engineering degrees to develop and improve spaceflight equipment and processes. Since 2006, she has dedicated primarily to enhancing deep-water drilling process safety and risk management in the petroleum industry. Dr. Philippart has also developed and imparted courses for NASA and Embry-Riddle Aeronautical University, and has enjoyed working for The Walt Disney Company.

Joseph W. Hendricks
Texas A&M University

The Impact of Hazard Statement Design in Procedures on Compliance. Rates: Some contradictions to Best (or Common) Practices

Dr. Hendricks earned a PhD in Industrial and Organizational Psychology from Texas A&M University. He is currently a research associate – senior investigator with the Next Generation Advanced Procedures consortium at Texas A&M University.

Day 1 Track 3
Managing Operations and Maintenance Modeling and Asset Integrity

Chetan Birajdar
Monaco Engineering Solutions

RBI Study using Advanced Consequence Assessment for Topside Equipment on Offshore Platforms

Chetan has more than 10 years of global experience in the Oil and Gas Industry working for Engineering Contracting Companies, major international operators and specialist consultancies for various onshore and offshore projects in Asia, Middle East, UK & Europe and Eurasia. He has been instrumental in developing and modifying Hazard Management tools used in Risk Calculations (Consequence Modelling, Risk Calculations, SIL Calculations, etc.) and has extensive experience of using various software packages.

Derek Yelinek
Siemens Process & Safety Consulting

Indicators of an Immature Mechanical Integrity Program

Mr. Derek Yelinek is the Risk Based Inspection Lead for Siemens Process & Safety Consulting business located in Houston, TX. Mr. Yelinek has over 10 years of experience in the development, implementation, and management of Mechanical Integrity programs with a focus on Inspection Data Management Systems (IDMS), Risk-Based Inspection (RBI), and procedure and work-process development. His experience ranges in consultant/services as well as the user/owner side, across the oil & gas, chemical, and mining industries. Derek is API 570, 571, and 580 certified and holds a B.S. in Chemical Engineering from Western Michigan University.

Matthew S. Walters
Exponent, Inc

Remember the à la Mode: Lessons Learned from Ammonia Release at Frozen Foods Warehouse

Dr. Walters is a Senior Engineer in the Thermal Sciences Practice at Exponent. Dr. Walters' background is in chemical engineering with specific expertise in the areas of separations, process modeling, and process control. He has applied his expertise to a variety of pollution mitigation systems, including post-combustion carbon capture, gas processing, combustion gas clean-up, and liquid-liquid extraction of contaminants from silicone polymers. Dr. Walters leverages his knowledge of process systems to provide incident investigation and process safety consulting services, as well as proactive technical evaluations of novel systems and processes.

Dr. Walters is a member of the American Institute of Chemical Engineers, where he participates in the Center for Chemical Process Safety committee for Abnormal Situation Management. He is also a licensed professional engineer in Illinois and a Certified Fire and

Explosion Investigator. Dr. Walters received M.S. and Ph.D. degrees in Chemical Engineering from the University of Texas at Austin. He also earned B.S. degrees in both Chemistry and Chemical Engineering from Purdue University.

Recalling and Learning from Incidents

Syeda Zohra Halim
Mary Kay O'Conner Process Safety Center

Process Related Incidents with Fatality-Trends and Patterns

Syeda Zohra Halim is currently employed as a Postdoctoral Research Associate at the Mary Kay O'Connor Process Safety Center (MKOPSC) and as a Lecturer of Chemical Engineering at the Texas A&M University. She overlooks several industry and federal funded projects ongoing at the MKOPSC, generates proposals for new ones and mentors graduate students in process safety-related dissertation projects.
Zohra completed her PhD in Chemical Engineering in Spring 2019 with Mary Kay O'Connor Process Safety Center at Texas A&M University. Her research focused on developing a model for assessing cumulative risk arising from impaired barriers in offshore oil and gas facilities.

T. Michael O'Connor
Mary Kay O'Conner Process Safety Center

Application of Mind Mapping to Classify and Recall Potential Hazards

T. Michael O'Connor is the President of O'Connor Ventures, Inc. in Houston, Texas. O'Connor established the Mary Kay O'Connor Process Safety Center at Texas A&M University, College Station, TX, where he currently serves as a Research Associate, as well as on the Steering Committee and Technical Advisory Committee. He is a member of the Engineering Advisory Council and the Industrial Advisory Board of the School of Public Health at Texas A&M University. Since June 2005, he has been a member of the Board of Directors at StarRotor Corporation, College Station. Formerly he was Vice-Chairman, Heat Transfer Research, Inc His primary interests include process safety, heat transfer in high temperature heat exchangers and furnaces, and metallurgy associated with these applications. He has a BS, Chemical Engineering, University of Missouri – Rolla.

Howard Duhon
GATE Energy

Would a HAZOP, LOPA, or STPA have Prevented Bhopal?

Howard Duhon, P.E. is a founder and a principal of Gibson Applied Technology and Engineering, Inc. (GATE) in Houston, TX. He has 46 years of experience in the Petrochemical and Oil and Gas industries mainly in process design and project engineering roles. For the past 15 years that work has mainly involved deep-water developments and has been focused on managing the interfaces between topsides and other disciplines. Duhon has a chemical engineering degree from the University of Louisiana at Lafayette. Throughout his

career he has had a particular interest in the study of decision theory and in the application of that knowledge to improve project execution. From 2013 to 2016 he served a term on the International Board of Directors of the Society of Petroleum Engineers (SPE) as the Projects, Facilities and Construction Technical Director.

Improving Process Safety With Technological Advances

Michael Marshall
Tratus group

Predictive Process Safety Analytics and IIoT-PSM Plus: The AI+PSM Analytical Framework

Missing Biography

Syeda Zohra Halim
Mary Kay O'Conner Process Safety Center

Guidance to Improve the Effectiveness of Process Safety Management Systems in Operating Facilities

Syeda Zohra Halim is currently employed as a Postdoctoral Research Associate at the Mary Kay O'Connor Process Safety Center (MKOPSC) and as a Lecturer of Chemical Engineering at the Texas A&M University. She overlooks several industry and federal funded projects ongoing at the MKOPSC, generates proposals for new ones and mentors graduate students in process safety-related dissertation projects.
Zohra completed her PhD in Chemical Engineering in Spring 2019 with Mary Kay O'Connor Process Safety Center at Texas A&M University. Her research focused on developing a model for assessing cumulative risk arising from impaired barriers in offshore oil and gas facilities.

Scott Hardesty
Applied Research Associates

Unified Wall Panel System (UWPS) - A Value Engineering Solution for Protective Construction in the Petroleum Industry

Scott Hardesty is a Senior Engineer, Program Manager and Employee Owner at Applied Research Associates / Rocky Mountain Division in Littleton, CO. Joining ARA in 2002, Mr. Hardesty has been focused on the assessment of protection technologies and analysis of existing or emerging threats across a wide spectrum of energetic scenarios. These include ballistic, blast, fragmentation and focused explosive energy. He has executed a wide range of materials testing programs for governmental and commercial clients, including utilization of both conventional and improvised explosives with charge sizes up to 10,000 lb. He is a

subject matter expert in developing customized instrumentation plans for complex test environments. He is the lead facilities coordinator for ARA large scale explosives operations at multiple facilities, and regulatory compliance coordinator responsible for interface with the ATF, DHS, FAA, DCMA, DSS, OSHA or other federal/state/local agencies. Mr. Hardesty served on Active Duty as a Captain (O3) in the United States Army Ordnance Corps after graduating from the Colorado School of Mines with a BS in Engineering (Mechanical Specialty) in 1998.

Exploring NaTech Events and Domino Impacts

Victor Edwards
VHE Technical Analysis

Protect Process Plants From Climate Change

Dr. Edwards retired as Director of Process Safety for IHI Engineering and Construction International Corporation in 2013. Since retiring, Vic has been actively consulting, editing, and writing. He specializes in process safety management and in process and environmental engineering. He has over 70 publications. Recent books include "Careers in Chemical and Biomolecular Engineering" with Suzanne Shelley and Section 10 of the 9th Edition of Perry's Chemical Engineers' Handbook. Vic chaired the 9th Global Congress on Process Safety in 2013 and the first Process Plant Safety Symposium in 1992. In 2015, Edwards received the Walton/Miller Award from the Safety and Health Division of the American Institute of Chemical Engineers. Vic is an AIChE Fellow and a member of AAAS, ACS, NFPA, NSPE, and NYAS.

Trish Kerin
IChemE Safety Center

Process Safety Implications in a Changing Environment

Trish Kerin is a director of the Institution of Chemical Engineers Safety Centre (ISC). After graduating with honors in mechanical engineering, Trish spent several years working in project management, operational and safety roles for the oil, gas and chemical industries. Trish has represented industry on many government committees related to process safety, and sits on the board of the Australian National Offshore Petroleum Safety and Environmental Management Authority and the Mary Kay O'Connor Process Safety Center steering committee. Trish is a Chartered Engineer, registered Professional Process Safety Engineer, Fellow of IChemE and Fellow of Engineers Australia. Trish holds a diploma in OHS and is a Graduate of the Australian Institute of Company Directors.

Ravi Kumar Sharma
Indian Institution of Technology

A Critical Evaluation of Industrial Accidents Involving Domino Effect

Dr Ravi Sharma did his Master of Technology (M.Tech) in Environmental Engineering in 2010 from Indian Institute of Technology (IIT), Roorkee and received the PhD degree, 2014 in the area of Quantitative Risk Assessment from Indian Institute of Technology (IIT), Roorkee. Presently he is working as a Senior Research Fellow in Indian Institute of Technology (IIT), Roorkee. His research interests are Quantitative Risk Assessment (QRA), Fires and Explosions Modelling, Fire Protection Engineering, Hazard Assessment, Inherent Process Safety, Loss Prevention, and Emergency Response Planning.

Day 2 Track 1
Risk/Consequence Analysis & Design Aspects
Risk Assessment III

John Cusimano
aeSolutions

Applying PHA Methodologies such as HAZOP and Bowtie to Assessing Industrial Cybersecurity Risk

John Cusimano is an industrial control system (ICS) / OT cybersecurity expert with a strong background in process control and functional safety engineering. Since 2009, John has started up and successfully led ICS/OT cybersecurity consulting practices at two consulting/engineering firms. John has personally performed countless ICS cybersecurity vulnerability and risk assessments in wide range of industries per NIST, ISA/IEC 62443 and NERC CIP standards. He developed the CyberPHA methodology through a combination of his work on standards committees and by working with key clients who shared his interest in applying process safety engineering discipline to ICS cybersecurity. He was a leader in the development of the ISASecure™ certification scheme. He led the development of 3 ICS cybersecurity courses for ISA (ISA IC33, IC34 and IC34) as well as the accompanying certificate programs. He also led the development of the "ICS Cybersecurity for Manager's" course hosted by SANS. John served as Chairman of the ISA 99 subcommittee that authored the recently approved ISA/IEC 62443-3-2 standard, "IACS Security Risk Assessment & Design".

Ravi Kumar Sharma
Indian Institution of Technology

Large Hydrocarbon Tank Fires: Modeling of the Geometric and Radiative Characteristics

Dr Ravi Sharma did his Master of Technology (M.Tech) in Environmental Engineering in 2010 from Indian Institute of Technology (IIT), Roorkee and received the PhD degree, 2014 in the area of Quantitative Risk Assessment from Indian Institute of Technology (IIT), Roorkee. Presently he is working as a Senior Research Fellow in Indian Institute of Technology (IIT), Roorkee. His research interests are Quantitative Risk Assessment (QRA), Fires and Explosions Modelling, Fire Protection Engineering, Hazard Assessment, Inherent Process Safety, Loss Prevention, and Emergency Response Planning.

Suresh G
Bharat Petroleum Corporation

Risk assessment of a large chemical complex during the construction phase using Intuitionistic Fuzzy Analytic hierarchy process

Suresh G. is a Research Scholar in Safety Engineering. After M Tech in Chemical Engineering, he is working in a large petroleum refinery for the last 21 years.

Risk Mitigation

Onder Akinci
Daros Consulting

Development of Resilient LNG Facilities

Onder Akinci has a PhD degree in Civil Engineering and more than 20 years of R&D, Civil/Structural/Architectural Engineering and Project Management experience. He had leadership roles with major EPC, consulting and LNG project development companies previously. He is a registered Professional Engineer in the state of Maryland. Dr. Akinci has extensive non-linear analysis, structural design, PFP optimization and facility upgrade experience. His areas of expertise include design of structures for fire and blast, earthquakes, hurricanes, dropped object and impact loads. He worked on several onshore and offshore Oil&Gas projects, and supported all phases from concept development to construction."

Rafael Callejas-Tovar
BakerRisk

Development of Risk Mitigation Programs using a Quantitative-Risk-Based Approach

Dr. Rafael Callejas-Tovar is a Senior Engineer working in the BakerRisk® Houston office as part of the Process Safety Group. His work is focused on quantitative risk analysis, consequence modeling, and computational fluid dynamics simulations. He received his PhD degree in Chemical Engineering from Texas A&M University. Rafael has over 8 years of industry and consulting experience in the U.S. with a focus on consequence and quantitative risk analysis for chemical plants, refineries, transportation of hazardous materials, and offshore oil & gas facilities.

Edward Marszal
Kenexis

Incorporating Mitigation safeguards with LOPA

Ed Marszal is President and CEO of Kenexis. He has over 25 years of experience in risk analysis and technical safety engineering of process industry plants, including design of Safety Instrumented Systems and Fire and Gas Systems. Ed is an ISA Fellow and former Director of the ISA Safety Division and 20 year veteran of the ISA 84 standards committee for safety instrumented systems. He is also the author of the "Safety Integrity Level Selection" and "Security PHA Review" textbooks from ISA.

Consequence Analysis: Gas Release

Jeffrey D. Marx
Quest Consulting Inc

Hole Size Matters

Jeff Marx is a Senior Engineer with Quest Consultants in Norman, Oklahoma, USA, and a registered professional engineer in the state of Oklahoma. He earned his Bachelor's degree in mechanical engineering from the University of Oklahoma and a Master's degree in Mechanical Engineering from Georgia Tech. In his 27 years at Quest, Jeff's primary responsibilities have been in consequence and risk analysis studies for the petrochemical industry. This work includes facility siting, building siting studies per API RP 752/753, and quantitative risk analysis (QRA) studies for various corporate and regulatory entities. His work has been involved all aspects of the petrochemical system, including pipelines, gas plants, refineries, LPG terminals, and chemical plants. Much of this work has been involved in the LNG industry, including siting for LNG plants (using 49 CFR 193, NFPA 59A, CSA Z276, EN 1473, and other standards), and as a member of the Canadian Standards Association's Z276 committee, the LNG standard for Canada.Jeff is also responsible for several portions of the CANARY by Quest consequence analysis software, and has helped to develop, maintain, and apply the CANARY+ risk analysis toolset used at Quest.

Jesse Brumbaugh
aeSolutions

How Can I Effectively Place My Gas Detectors

Mr. Brumbaugh is a process safety engineer with a 13 year background in process modeling, holding degrees in chemical engineering and computer science from Texas Tech University. He has worked in the process safety industry for over 7 years, performing models of gas dispersion, vapor cloud explosions, pool and jet fires, and other hazards in a wide range of software packages including computational fluid dynamics (CFD). He has also participated in numerous PHA studies; conducted process simulations in VMG, HYSYS, and CFD; and also numerical methods based models for various types of projects.

SreeRaj Nair
Chevron

Consequence Assessment Considerations for Toxic Natural Gas Dispersion Modeling

Nair is a Process Safety leader with expertise in Technical Safety engineering and safety management. A Chartered Engineer with global experience in stewarding process safety performance and governance in hazardous industries and industry peer groups. Nair, in his current role as Senior Process Risk Engineer, leads technological risk management at Permian operations and projects for Chevron Corporation. Nair, a MSc (Eng.) in Process Safety and Loss Prevention (the University of Sheffield, UK, 2004) is pursuing his PhD in dispersion modeling at the University of Warwick.

Noma Ogbeifun
Chevron

Consequence Assessment Considerations for Toxic Natural Gas Dispersion Modeling

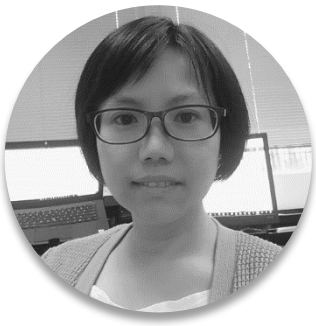
Noma Ogbeifun is a Process Risk Engineer with Chevron Corporation in the Permian Basin. Currently, he performs qualitative and quantitative risk assessment studies like HAZOP, HAZOP-LOPAs, Inherently safer design reviews, consequence modeling, etc. on new and existing facilities. In a prior role as an LNG process engineer, he supported major projects including Wheatstone LNG, Gorgon LNG, Wafra Steam Flood, and Angola LNG. He obtained his Bachelor of Science degree in Chemical Engineering from Purdue University in 2013. Noma has co-authored papers covering subjects within LNG, renewable energy, and process safety published in journals and presented at conferences. In his spare time, Noma is pursuing his Master's in Business Administration at the University of Texas at Permian Basin.

Reactive Chemicals

Kok Hwa Lim
SIT

Modelling and Simulation to Predict Energetic Material Properties

Kok Hwa Lim is currently the Director (Designate), Professional Officer Division and Programme Director, Pharmaceutical Engineering Programme in Singapore Institute of Technology. Mr. Lim is a registered Professional Engineer and Chartered Engineer (Chemical & Process Engineering). Mr. Lim has been elected as Council Member and is currently the Vice President of The Institution of Engineers Singapore. He was a member of the Working Group for Workplace Health Safety in Higher Education and Research Sector. Mr. Lim is frequently invited as speaker/ panelist in local and regional seminars/ conferences on Process Safety. He helped to organize various local seminars/ conferences with industries and professional societies. Mr. Lim has been teaching Process Safety related courses to undergraduate students as well as process safety professional since 2008.



Cuixian (Trisha) Yang
Merck & Co

Safety Assessment of Low Temperature Radical Initiators for Proper Storage and Safe Handling Conditions

Cuixian Yang studied at Xiamen University in China to study Chemical Engineering. In 2007, she was admitted to Tufts University to pursue her PhD in Chemical Engineering, focusing on the development and application of novel nanomaterials based on Tabacco Mosaic Virus as biotemplate. Later on, she joined Professor Klavs Jensen's group in Chemical Engineering Department of MIT as a postdoctoral associate. Her postdoc research involved continuous catalytic hydrogenation in micro packed bed reactor. In 2015, she started her industry career as a senior scientist in Chemical Engineering R&D department of Merck&Co in New Jersey. She participated in many different projects, and gained experience from early-stage process development, scale-up to late-stage process validation and tech transfer to manufacturing scale. Currently, she is working in Environmental and Process Safety Engineering group, to explore new process safety-related techniques and approaches.

Yuto Mizuta
Mitsubishi Chemical

Analysis of Pressure Behavior during Reaction Runaway and Estimation of Available Depressurization Design

Yuto Mizuta graduated from the safety engineering course of Yokohama National University in Japan. Mr. Mizuta has worked for Mitsubishi Chemical for 15 years as a safety engineer. He has experience in evaluating and consulting chemical plants in Japan.
The main activities are risk analysis of chemical processes, consequence analysis, process simulation and consulting for explosion and thermal reactivity.

Day 2 Track II
Human Factors: People in Action
Human Performance/Design Making II

Tom Shephard
Wood (Retired)

Preventing Cognitive-Attributed Errors in Safety Critical Systems: A Path Forward

Tom Shepard has over 40 years working for operating and engineering companies, delivered process safety, control and safety systems, and capital projects to the O&G, refining, petrochemical, pipeline and fuels terminal industries. Projects included some of the world's largest and most complex O&G offshore facilities and world-class refinery projects. Example roles included subject matter expert, project management, discipline lead and discipline department management. A unique skillset is the ability to develop solutions to highly complex, multi-discipline and multi-organizational problems. That skillset led to many successes developing and implementing corporate and project level technical and execution standards, tools, methods, work processes and execution models. Current

focus: Develop, promote and implement new standards, tools and methods for designing safety critical tasks, active human barriers and emergency response systems. The specific focus is on approaches that can reliably and systematically prevent or mitigate cognitive-specific errors in safety critical designs.

Changwon Son
Texas A&M University

Two Views of Evaluating Procedural Task Performance: A Transition from Safety-I to Safety-II Approach

Changwon Son is a Ph.D. candidate in Applied Cognitive Ergonomics (ACE) Lab in the Department of Industrial and Systems Engineering at Texas A&M University. He has obtained his master's degree in Safety Engineering from Mary Kay O'Connor Process Safety Center at Texas A&M University. Son's research is focused on resilience engineering, an emerging paradigm for safety, for complex socio-technical systems such as disaster response, oil and gas processing, and healthcare. After receiving his bachelor's degree in Hanyang University, Seoul, Korea, he worked for Hyundai Heavy Industries, the world's largest shipbuilding and offshore company as safety, health, and environmental manager for over six years

Joseph W. Hendricks
Texas A&M University

Beyond Human Error: Integration of the Interactive Behavior Triad and Toward a Systems Model

Dr. Hendricks earned a PhD in Industrial and Organizational Psychology from Texas A&M University. He is currently a research associate – senior investigator with the Next Generation Advanced Procedures consortium at Texas A&M University.

Fatigue and Stress

Ranjana Mehta
Texas A&M University

Operator Performance Under Stress: A Neurocentric Virtual Reality Training Approach

Ranjana Mehta is Associate Professor in the Department of Industrial and Systems Engineering at Texas A&M University. She is also a graduate faculty with the Texas A&M Institute for Neuroscience at Texas A&M University, director of the NeuroErgonomics Laboratory, co-director of the Texas A&M Ergonomics Center, and a faculty fellow with the Center for Remote Health Technologies and Systems, the Center for Population Health and Aging, and Mary Kay O'Connor Process Safety Center. The NeuroErgonomics Lab examines the mind-motor-machine nexus to understand, quantify, and predict human performance when interacting with emerging technologies (unmanned, collaborative, and wearable systems) in safety-critical extreme environments (e.g., emergency response, space exploration, oil and gas).

John Kang
Texas A&M University

Towards a Predictive Fatigue Technology for Oil and Gas Drivers

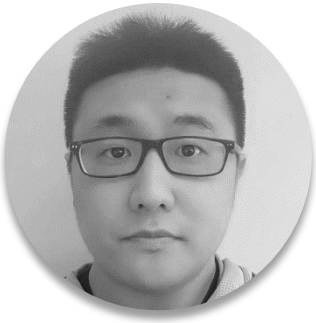
John Kang is a Ph.D. student in Industrial & Systems Engineering at Texas A&M University and has received a BS in Industrial & Systems Engineering from Georga Tech. His research interests are physiological wearables to quantify fatigue and understanding decision making under fatigue or stress in a high-risk environment.

Stefan V. Dumlao
Texas A&M University

Validation of the Fatigue Risk Assessment and Management in High-Risk Environments (FRAME) Survey

Stefan V. Dumlao is a doctoral student studying industrial-organizational psychology at Texas A&M University, College Station. His primary research interests are employee reactions to wearable monitors and occupational safety.

Day 2 Track IV
Research and Next Generation
Next Generation Process Safety I



Guanyang Liu
MKOPSC

Identifying Contributing Factors of Pipeline Incident from PHMSA Database on NLP and Text Mining Techniques

Guanyang Liu, a PhD student in Chemical Engineering with research interest of reaction engineering, process safety, AI applications in process industry

Pallavi Kumari
MKOPSC

Causation Analysis of Pipeline Incidents using Artificial Neural Network (ANN)

Pallavi Kumari is a fourth-year Ph.D. student working with Dr. Joseph Kwon. Her research focuses on root cause and consequence analysis of rare events in chemical process industries using statistical data analysis methods, process modelling and process control techniques. She received her Bachelor's from IIT Kanpur and worked in Reliance Industries Limited, India.

Nabila Nazneen
MKOPSC

Development of Hazard Factor for Engineered Particles

Nabila Nazneen is currently pursuing her Master's in Safety Engineering at the Mary Kay O'Connor Process Safety Center. Her Background is in Chemical Engineering. She attained an MBA degree right after her bachelor degree and worked in the ready-made garments sector for two years before coming to the Process Safety Center. Her research interest is on nanoparticle hazards.

Next Generation Process Safety II

Nir Keren
Iowa State University

Can a Virtual Reality Application Better Prepare Millennials and the Z-Generation for Working with Systems in the Process Industry?

Nir Keren is an associate professor of occupational safety and a graduate faculty member at the Virtual Reality Application Center at Iowa State University. Keren is also the director of the Occupational Safety Program of the NIOSH Heartland Education and Research Center and the Director of the VirtuTrace Laboratory for Applied Decision Making Research in Virtual Reality.

Prasad Goteti
Honeywell Process Solutions

Process Safety Risk Index Calculation Based on Historian Data

Prasad Goteti is a Principal Project Engineer at Honeywell Process Solutions (HPS), Houston Texas USA, in the Safety Engineering Center of Excellence. He is responsible for providing process safety solutions to customers, working on Safety Engineering at the proposal and detailed engineering stage for Safety Instrumented System (SIS) projects, which includes Emergency Shutdown Systems (ESD), Burner Management Systems (BMS) and Fire and Gas Systems (FGS). He is also an Instructor for the TUV Rheinland Germany, approved Functional Safety Training course conducted by HPS Automation College. Prasad holds a degree in Instrumentation from Birla Institute of Technology and Science (BITS), Pilani , India, is a Professional Engineer (P.Eng) with the Association of Professional Engineers and Geoscientists of Alberta (APEGA), Canada, a Certified Functional Safety Expert (CFSE), a TUV Functional Safety Expert (TUV Rheinland, Germany), An Advisory Board member of Purdue Process Safety and Assurance Center (P2SAC) at Purdue Universtiy, a member of WG 7 of ISA TR 84.00.07 and WG 9 of ISA TR 84.00.09 committees

Sinijoy P J
Cochin University of Science and Technology

A Brief review of Intrusion Detection System in Process plants and advancement of Machine Learning in Process Security

Missing Biography

Day 2 Track V
Explosions
Explosion Modeling

Tássia L. S. Quaresma
University of Campinas

The Influence of the Velocity Field on the Stretch Factor and on the Characteristic Length of Wrinkling of Turbulent Premixed Flames

Tássia Lins da Silva Quaresma is a chemical engineer with experience in Risk Analysis and numerical combustion modeling with computational fluid dynamics. Tássia has contributed to several risk analysis methodologies of oil&gas and mining industrial plants. Currently, she has been studying combustion models for turbulent premixed flames focusing on the turbulent-flame interaction in order to predict flame speed and its consequences.

Filippo Gavellia
Blue Engineering and Consulting

Towards a Comprehensive Model Evaluation Protocol for LNG Hazard Analyses

Dr. Filippo Gavelli is a mechanical engineer who specializes in the analysis of heat transfer and fluid flow phenomena, including multiphase flows and cryogenic fluids. He has 18 years of engineering consulting experience and over 25 years of experience in computational fluid dynamics (CFD) modeling, using several research and commercial codes. He applies his expertise to modeling the consequences of hazardous releases and performing risk assessments for Liquefied Natural Gas (LNG) facilities. Dr. Gavelli has over 16 years of experience with modeling hazard scenarios including vapor cloud dispersion, pool and jet fires and vapor cloud explosions; his experience includes more than 50 LNG installations worldwide, including onshore, offshore and floating facilities for LNG import, export, peakshaving, truck loading and bunkering. He has been a member of the NFPA 59A committee for over 13 years and a frequent contributor to LNG safety related conferences and expert panels.

Charline Fouchier
Von Karman Institution of Fluid Dynamics

Beirut: How behaves Ammonium Nitrate Exposed to Fire and How Strong and Damaging is its Explosion

Charline Fouchier is a postdoctoral researcher at the von Karman Institute in Belgium. She completed her Ph.D. degree this year on the *Investigation of the Pollutant Dispersion Driven by a Condensed-Phase Explosion in a Complex Environment*. She has a master's degree in industrial safety, from Ecole national supérieure des Mines d'Alès (France), a master's degree in Environments and Urban Risks from Ecole Nationale supérieure des Mines de Saint-Etienne (France) and a Post-graduate Research Master in Fluid Dynamics from the von Karman Institute (Belgium), during which she won the Excellence in Experimental Research Award for her work on the blast propagation in an urban environment

Explosion Phenomena I

Simon Gant
UK Health and Safety Executive

Flammable Mist Hazards Involving High-Flashpoint Fluids

Simon Gant is a Principal Scientist in the Fluid Dynamics Team at HSE's Science and Research Centre in Buxton, UK, where he undertakes work on incident investigations, research, development of guidance and standards, model reviews and consultancy. He obtained a master's degree in mechanical engineering from Leeds University in 1997 and a PhD in computational fluid dynamics from Manchester University in 2002. His current work is mainly focused on the Jack Rabbit II chlorine release trials, hydrogen energy demonstration projects, carbon capture and storage, and flammable mists

Yumeng Zhao
Purdue University

Measuring Suspended Explosive Dust Concentration from Images

Missing Biography

Elaine S. Oran
Texas A&M University

The HBT-A Large-Scale Facility for Study of Detonations and Explosion

Elaine S. Oran is TEES Eminent Professor in the Department of Aerospace Engineering at Texas A&M University. Previously she was the A. James Clark Distinguished Professor and the Glenn L. Martin Institute Professor at the University of Maryland. For many years before that, she was the Senior Scientist for Reactive Flow Physics at the US Naval Research Laboratory in Washington, DC. She received an A.B. in chemistry and physics from Bryn Mawr College and both a M.Ph. in Physics and a Ph.D. in Engineering and Applied Science from Yale University. She is a Member of the National Academy of Engineering, an Honorary Fellow of the AIAA, and a Fellow of the American Academy of Arts and Sciences. Her recent research interests include chemically reactive flows, turbulence, numerical analysis, high-performance computing, shocks and shock interactions, and rarefied gases, with applications to combustion, propulsion, and all sorts of explosions.

Explosions Phenomena II

Zeren Jiao
MKOPSC

Development of Flammable Dispersion Quantitative Property-Consequence Relationship Models Using Machine Learning

Zeren Jiao joined Mary Kay O'Connor Process Safety Center in September 2016 and he obtained his M.S. degree of chemical engineering in 2018 under the supervision of Dr. Sam Mannan. He is currently a Ph.D. student in the MKOPSC. Zeren Jiao's research focus is on implementing machine learning and big data analysis in process safety.

Chenxi Ji
MKOPSC

An Unsupervised Model to Predict the Liquid In-Cylinder Combustion Risk Ratings of Marine Fuels

Chenxi Ji is currently a research assistant at Mary Kay O'Connor Process Safety Center and Gas & Fuel Research Center of Texas A&M University. He is motivated to apply his process systems engineering and chemical process safety on the shipping and oil & gas industry, seeking to make the oil & gas industry faster, greener, safer and more cost-effective.

Peter A Diakow
BakerRisk

Fireball and Flame Venting Comparisons

Peter Diakow is a Senior Consultant with the Blast Effects group at BakerRisk, with a master's degree in Mechanical Engineering from Queen's University in Canada. Mr. Diakow has over a decade of experience in experimental testing and research with a focus on vapor cloud explosions, vented deflagrations, and deflagration to detonation transition (DDT). At BakerRisk Mr. Diakow is also involved with Facility Siting Studies (FSS), Quantitative Risk Assessments (QRA), Incident Investigations and Dust Hazard Analyses (DHA).

Consequence Analysis: Flammability

Logan N. Kunka
Texas A&M University

Numerical Simulation of Methane-Air DDT in Channels Containing Trace Amounts of Impurities

Logan Kunka is a DOE Computational Science Graduate Fellow and currently a graduate student in Aerospace Engineering at Texas A&M University. He holds a B.S. in Aerospace and Mechanical Engineering from Oklahoma State University. His interests include combustion, computational physics, and high-performance computing. His current research involves the simulation of reactive flows including deflagrations, detonations, and DDT in gases.

J. Kelly Thomas
BakerRisk

The Use of Bent Poles as a Detonation Indicator

Missing Biography

Zeren Jiao
MKOPSC

Machine Learning Based Quantitative Prediction Models for Chemical Mixture Flammability Limits

Zeren Jiao joined Mary Kay O'Connor Process Safety Center in September 2016 and he obtained his M.S. degree of chemical engineering in 2018 under the supervision of Dr. Sam Mannan. He is currently a Ph.D. student in the MKOPSC. Zeren Jiao's research focus is on implementing machine learning and big data analysis in process safety.

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