AGENDA INFORMATION

☒ Regular Meeting Date: May 15, 2023
☐ Other: Date:

Dept. Manager	GM/ Director	CAO

The District of North Vancouver
REPORT TO COUNCIL

May 12, 2023
Case: 08.3060.20/014.23
File: 08.3060.20/014.23

AUTHOR: Dan Milburn, General Manager, Planning, Properties & Permits

SUBJECT: 100 Forester Street, Proposed Hydrogen Liquefaction Facility – Additional Information

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RECOMMENDATION

THAT the May 12, 2023 report of the General Manager, Planning, Properties & Permits entitled 100 Forester Street, Proposed Hydrogen Liquefaction Facility – Additional Information is received for information.

REASON FOR REPORT

Further to the report submitted for the May 15th Regular Council Agenda, this additional information is being provided to Council for consideration.

Respectfully submitted,

Dan Milburn
General Manager, Planning, Properties & Permits

Attachment 1: HTEC DNV Liquid Hydrogen Project Risk Assessment Study

REVIEWED WITH:

❑ Community Planning	____	❑ Clerk's Office	____	External Agencies:	
❑ Development Planning	____	❑ Communications	____	❑ Library Board	____
❑ Development Engineering	____	❑ Finance	____	❑ NS Health	____
❑ Utilities	____	❑ Fire Services	____	❑ RCMP	____
❑ Engineering Operations	____	❑ ITS	____	❑ NVRC	____
❑ Parks	____	❑ Solicitor	____	❑ Museum & Arch.	____
❑ Environment	____	❑ GIS	____	❑ Other:	____
❑ Facilities	____	❑ Real Estate	____		
❑ Human Resources	____	❑ Bylaw Services	____		
❑ Review and Compliance	____	❑ Planning	____		

"HTEC DNV Liquid Hydrogen Project Risk Assessment Study"

For

HTEC

Revised Final Report

May 10, 2023

Prepared By:

Doug McCutcheon and Associates, Consulting
A Division of "Human Factors Impact Ltd."

May 10, 2023

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EXECUTIVE SUMMARY

HTEC plans to construct a hydrogen liquefaction facility in North Vancouver located at 100 Forester Street in the heavy industrial zoned area between the Chemtrade and ERCO facilities and south of the GFL Environmental facility. Hydrogen from the existing ERCO operations will be received and converted to liquid hydrogen for sale to customers. The process will consist of a refrigeration system using liquid nitrogen, two liquid hydrogen storage vessels (90,000 US gallons each) and ability for truck loading of the liquid hydrogen for transfer to customers.

This is a preliminary report, as the facility is not yet fully designed. However, the risk of a fatality associated with the proposed industrial use (*as described above)*, and more particularly the off-site risks associated with release of liquid hydrogen from the proposed use, and the off-site risks associated with any explosions or fires caused by any such release, meet the threshold set by the risk criteria developed by the Canadian Society for Chemical Engineering – Process Safety Management Division, as described in Appendix “4” of this report.

The risk criteria as developed through the Canadian Society for Chemical Engineering – Process Safety Management division (Major Industrial Accidents Council of Canada) *(CSChE-PSM (MIACC))* work. The CSChE-PSM (MIACC) methodology is considered Canada’s best practice for conducting industrial risk assessments. This study leads to a conservative estimate of risk and uses globally accepted methodology and peer reviewed data.

Risk assessments are a requirement of the District of North Vancouver (DNV). A risk assessment acceptable to the District is a preliminary requirement of this proposed rezoning, and, more generally, it is always a requirement where any applicant is seeking DNV discretionary land use permissions in relation to any proposed activity or land use where known hazards are or will be present. This has been evident in the existing neighbouring industrial facility projects.

The risk assessment methodology begins with the identification of hazards that could have an unwanted impact beyond the company property line. For the liquid hydrogen process, the main hazard is a release of liquid hydrogen from a storage tank, from the loading of tank trucks or from a high-pressure pipe leak. The released liquid hydrogen will evaporate, resulting in a vapour cloud, which could explode creating a shock wave and igniting resulting in a radiant heat impact from the ensuing fire. Finally, the use of liquefied nitrogen in the process can also have an impact so this is looked into as well.

The outcome of the risk assessment is to determine any impact off site that could cause fatalities to the public. It is noted that all scenarios were developed based on worst-case situations with no credit taken for mitigating factors. The resulting hazardous scenarios identified for analysis are listed below. This does not necessarily mean every scenario results in an off-site consequence; just that they have been identified as a scenario for further analysis. The scenarios are:

1. A release from a tank connection creating a hydrogen vapour cloud explosion from a liquid hydrogen storage tank release.

2. The failure of a tank truck loading hose leading to a hydrogen vapour cloud explosion due to a liquid hydrogen tank truck release.
3. A fire impinging on one of the storage tanks or a tank truck long enough to create a Boiling Liquid Expanding Vapour Explosion (BLEVE)
4. The ignition of a release of liquid hydrogen leading to a jet fire.
5. A liquid nitrogen release that stays close to ground level as a liquid or vapour.

Using recognized methodologies to characterize the release consequences finds from the scenario analyses:

- A BLEVE analysis uses a full liquid hydrogen storage tank as the worst case and concludes an overpressure enough to cause structural damage to buildings, which, if serious enough, could result in a fatality (1.0psi). That worst case could be felt as far as 713m from the explosion with potential radiant heat impact of 88m. Note the radiant heat from the fireball lasts for a short time of 14 seconds, which is not considered long enough to create enough heat to expose a person offsite to a fatality of site. Note; that despite the severe consequences, the risk is low due to the low probability, this is discussed later.
- Should there be a tank truck BLEVE, the overpressure impact will be out to 113m from the release.
- A hydrogen pipe release that ignites will create a jet fire, which can cause fatalities out to a distance of 20.1m from the flame.
- A spill of liquid nitrogen will form a pool and the vapour will lower the oxygen content in the air below 18% for a distance of 8.1m. An asphyxiation potential for nearby people.

Risk defined by the distance (consequence) to a fatality for an individual exposed and the probability it can happen on an annual basis. Probabilities need to be from reliable sources, such as the work conducted by the UK Health and Safety Executive (HSE). Their work is a collection of outcomes from other research studies and is peer reviewed by the HSE before publishing them. Other sources are available which corroborate the HSE work.

Shown in Appendix "4", the overall risk of a worst-case event for the HTEC operations, is well within the acceptable criteria as defined by the CSChE-PSM (MIACC) for the risk level of 1 X 10^{-4} at the company property line. This is the case for all the identified HTEC scenarios. Specifically of note are the storage tank and tank truck BLEVE scenarios where an overpressure shock wave can extend beyond the property line. The actual risk is very low 1 X 10^{-7} and well within the acceptable level of risk of 1 X 10^{-4} at the property line as defined by the CSChE-PSM (MIACC) criteria.

Table 1: Risk Summary

Hazard Scenario	Probability of Occurrence	Resulting Risk at Property Line	**Meets the Risk Tolerance Criteria of 1 X 10^{-4} (Yes/No)**
1	5.0 X 10^{-8}	Less than 1 X 10^{-4}	Yes
2	5.2 X 10^{-3}	Less than 1 X 10^{-4}	Yes
3	1.0 X 10^{-7}	Less than 1 X 10^{-4}	Yes
4	1.0 X 10^{-7}	Less than 1 X 10^{-4}	Yes
5	1 X 10^{-6}	Less than 1 X 10^{-4}	Yes

Note: The probability values do not represent the risk values. Please refer to the "Risk Analysis" section of the report (pages 41-45) to explain the difference.

Hazard Scenarios:

1. A release from a tank connection creating a hydrogen vapour cloud explosion from a liquid hydrogen storage tank release.
2. The failure of a tank truck loading hose leading to a hydrogen vapour cloud explosion due to a liquid hydrogen tank truck release.
3. A fire impinging on one of the storage tanks long enough to create a Boiling Liquid Expanding Vapour Explosion (BLEVE)
4. The ignition of a release of liquid hydrogen leading to a jet fire.
5. Liquid nitrogen will stay close to ground level as a liquid or vapour.

Figure 1: Canadian Acceptable Level of Risk

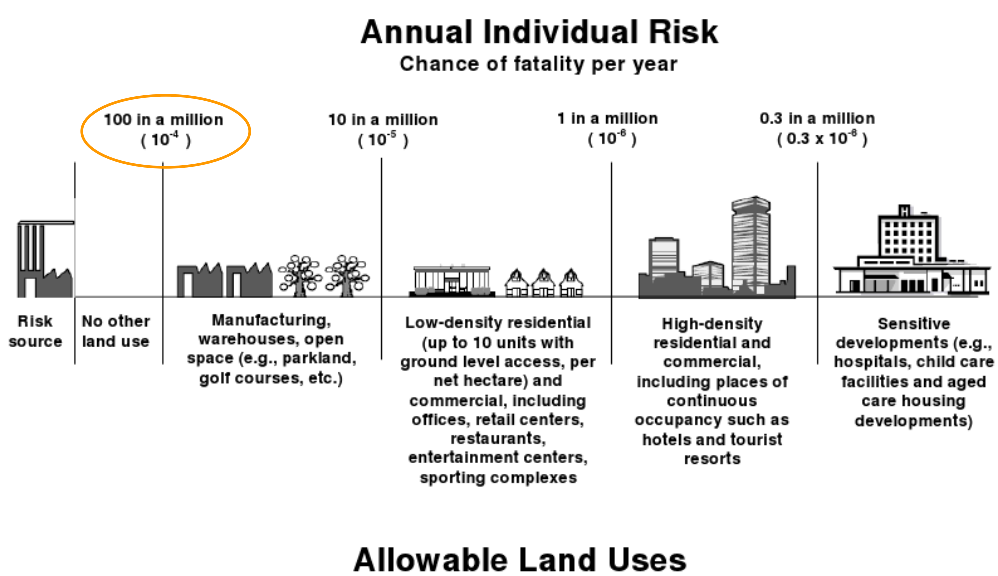
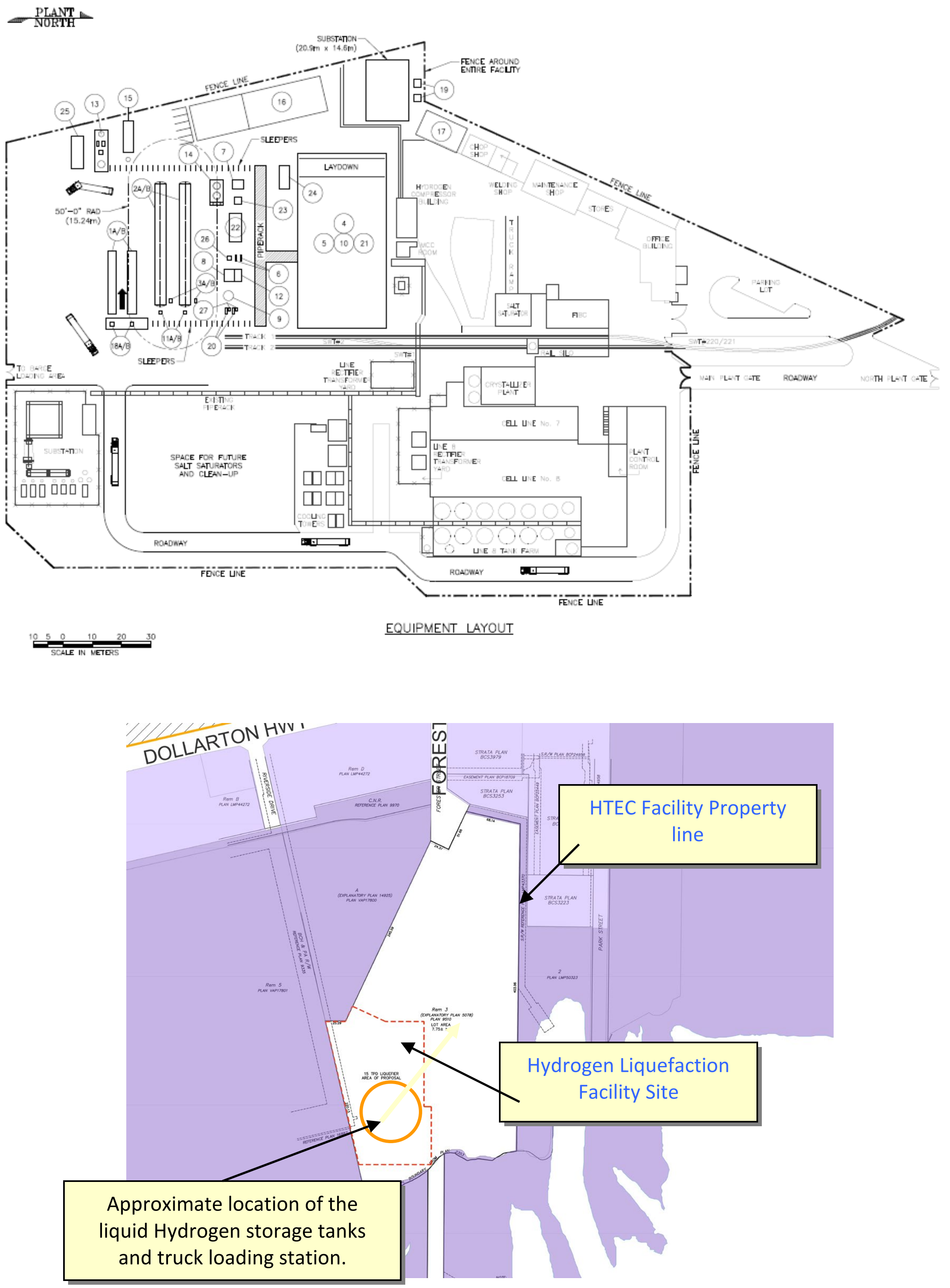


Figure 2: HTEC Site Layout and Site Location



As seen from Figure 1 the site is located in an existing industrial area and zoned by the District of North Vancouver as *Zoning Bylaw 1965 – 3210 Part 7, 750 Employment Zone – Industrial (EZ-I)*. Meeting these zoning criteria indicates the risk to be acceptable for this project.

Specific recommendations for consideration as an outcome of this review include the following. Note that these are not conditions:

1. The risks are acceptable and within the CSChE-PSM (MIACC), criteria for risk based land use planning purposes. Suggested is a good risk management program would see this level of acceptable risk to remain the norm.

2. Suggest considering the addition of a concrete wall between the two storage tanks or equivalent mitigation measure, which will act to prevent a torch effect from one tank leak affecting the other tank, preventing a possible BLEVE, and resulting shrapnel impacts.

3. Suggest consideration be given to a design for a blast resistant control building for at least a 1.0psi overpressure event.

4. Suggest recognizing the potential impact to workers of a liquid nitrogen release and down-wind asphyxiation potential.

5. Cyber Security issues are at the front of controlling unwanted events these days. Although new, there is guidance available to assist companies to build defenses. Consideration to this concern is suggested (See Appendix “3”).

6. Consider focusing on Human Factors issues as part of the risk management plan.

7. Consideration to adding flammable vapour detectors.

8. A static charge can collect on equipment particularly during a release under pressure. In the case of hydrogen, this potential difference can be an ignition source. Sound and robust equipment grounding with annual checking is suggested.

This risk assessment review is conservative. The resulting risk calculations are based on sound and recognized methods meeting Canada’s best practice as defined by the CSChE-PSM (MIACC) criteria and experience with the DNV land use planning requirements. Should there be any questions please ask. I trust you will find the report satisfactory.

D McCutcheon

Doug McCutcheon, P. Eng.

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GENERAL REPORT

This study was commissioned to provide a risk assessment of the proposed HTEC Liquid hydrogen facility in the District of North Vancouver. The resulting study is to meet District requirements as defined through the CSChE-PSM (MIACC) methodology.

The intent of the facility is to accept hydrogen gas from the ERCO Sodium Chlorate facility, purify it and liquefy the hydrogen for market. Liquid hydrogen will be stored in tow 90,000usg tanks for shipment by tank truck only.

The District has not defined the need for risk assessments in their zoning bylaw however; previous hazardous industrial developments have conducted them for specific projects. The risk assessment follows Canadian best practices as defined by the Major Industrial Accidents Council of Canada (MIACC) work, now managed through the Canadian Society for Chemical Engineers – Process Safety Management division, and follows globally accepted methodology

Methodology Used

The risk assessment process used was developed over the many years partly because of the Bhopal India tragedy in 1984. The steps and methods are well established in the industrial world and hence are considered the accepted method for doing risk assessments. See Appendix “1” for a flowchart describing the Risk Management methodology.

A risk assessment begins with identifying the hazards or concerns. The approach taken for this review is to determine the largest realistic worst case scenarios based on the experience and hazards generated internally and by this consultant. *(A HAZARD analysis is a recognized qualitative analysis of hazardous processes designed to identify specific hazards or concerns early in the project design).* This step relies on regulations and management direction to determine what is considered a hazard or not. The possible scenarios are all specific to the release of liquid hydrogen. Included are possible fire and explosions and a Boiling Liquid Expanding Vapour Explosion (BLEVE) because of a fire impinging on a full tank.

The next steps of the risk assessment process are to examine each hazard for the consequence (potential for fatal impacts to the public and on the nearby areas) and the probability of occurrence. Once these two calculations are made the risks can be determined and compared to what is considered acceptable in Canada, as defined by CSChE-PSM (MIACC). It should be noted that risk calculations do not include emergency planning or other mitigating techniques. The calculations are strictly the "realistic worst case" situations. Emergency plans are developed, once the worst case is known.

According to the CSChE-PSM (MIACC) work the "acceptable level of risk" to the public is set at 1.0×10^{-6} for Canada. This number is not regulated but is referred to in standards and regulations in Canada and as Canada’s best practice. It is very much in line with the rest of the world. The value 1×10^{-6} (one in a million) is the annual probability of a fatality to an individual because of an industrial incident.

HAZARD ANALYSIS

Every risk assessment needs to begin with the identification of hazards. These types of hazards are then evaluated in terms of the impact they could have on areas outside the property line of the company. The chart below describes the type of hazard and possible concerns over the impact

Hazard identification involves the identification of specific undesirable consequences. They can be broadly classified as human impacts, environmental impacts, asset damage impact and business damage impacts. These are relatively straightforward and not difficult to identify. However being thorough in the review is necessary in order to ensure all hazards are uncovered.

ADVERSE CONSEQUENCES

Human Impacts

- Consumer injuries
- Community injuries
- On-site personnel
- Loss of employment
- Psychological effect

Environmental Impacts

- Off-site contamination
  - air/ water/ soil
- On-site contamination
  - air/ water/ soil

Asset Damage Impacts

- Property damage
- Stock value
- Insurance premiums
- Negative image
- Lost markets
- Legal liability

Business Damage Impacts

- Production outage
- Inventory loss
- Insurance premiums
- Product quality
- Negative image

POTENTIAL HAZARDS TO CONSIDER INCLUDE

Fire	Explosion	Detonation
Corrosion	Toxicity	Radiation
Noise	Vibration	Noxious Materials
Electrocution	Asphyxia	Mechanical Failure
Environmental Impact	Security Breach	Lost Company Image
Insurance Cost Impact	Impact on the public	Long term exposures

Most hazards are seen as personnel safety issues as they pertain to the workers in the particular company operation and rightfully so as they are exposed to the hazards in their daily work activities. Management must be mindful of this priority and focus on the protection of the workers in the field. However some may have an impact beyond the “fence-line” of the company’s operations.

Here are some calculated values that can be used to understand more clearly the impact of an incident. We have the ability to determine how much energy can be released from almost any incident, having the knowledge of the consequences as shown below makes for better decision making. These represent just some consequences of concern. Other tables can show consequences of other incident types (like electrical, mechanical, etc.).

Table 2: Some Types of Measurable Consequences

TYPE OF INCIDENT	**CONSEQUENCE Odour/Irritation Threshold**	**CONSEQUENCE Irreversible Effects Threshold**	**CONSEQUENCE Life Threatening Effects Threshold**
Toxic Release (concentration - 1 hour exposure)	ERPG-1	ERPG-2	ERPG-3
Fireball - Immediate Ignition (radiation intensity - 60 second exposure)	1st Degree Burns 2 kW/m^2 600 BTU/hr/ft^2	2nd Degree Burns 5 kW/m^2 1600 BTU/hr/ft^2	3rd Degree Burns 8 kW/m^2 2500 BTU/hr/ft^2
Flash Fire - Delayed Ignition (flammable gas dispersion)	NOTE there is no lower level consequence	1/2 of Lower Flammability Limit	1/2 of Lower Flammability Limit
Pool / Jet Fire (radiation intensity - 90 second exposure)	1st Degree Burns 1 kW/m^2 400 BTU/hr/ft^2	2nd Degree Burns 4 kW/m^2 1200 BTU/hr/ft^2	3rd Degree Burns 6 kW/m^2 1900 BTU/hr/ft^2
Unconfined Vapor Cloud Explosion (overpressure)	Window Breakage 0.3 psig 0.02 bar	Partial Demolition of Houses 1.0 psig 0.07 bar	Threshold of Ear drum rupture. Lower limit of serious structural damage 2.3 psig 0.16 bar

Definitions:

kW/m²: are kilowatts per meter squared. A measure of heat energy over a surface area.

Psig & bar: are measures of pressure

ERPG-1: is the maximum airborne concentration below which it is believed that nearly all individuals could be exposed for one hour without experiencing other than mild transient adverse health effects or perceiving a clearly objectionable odour.

ERPG-2: is the maximum airborne concentration below which it is believed that nearly all individuals could be exposed for up to one hour without experiencing or developing any irreversible or other serious health effects or symptoms that could impair their abilities to take protective action.

ERPG-3: is the maximum airborne concentration below which it is believed that nearly all individuals could be exposed for up to one hour without experiencing or developing life-threatening health effects.

Table 3: Specific to Thermal Radiation Incidents

(*CCPS – "Guidelines for Chemical Process Quantitative Risk Analysis, second edition – 2000*")

Intensity **kW/m²**	**Consequential Exposure Damage to People**	**Consequential Damage to Equipment**
37.5	• Significant injury after 10 seconds exposure. • 1% lethality after 10 seconds exposure • 100% lethality after 100 seconds exposure	Sufficient to cause damage to process equipment.
25	• Significant injury after 10 seconds exposure. • 1% lethality after 30 seconds exposure • 100% lethality beyond 100 seconds exposure	Minimum energy to ignite wood at indefinitely long exposures & "unpiloted".
12.5	• Significant injury after 60 seconds exposure. • 1% lethality after 80 seconds exposure	Minimum energy required for "piloted" ignition of wood, melting of plastic tubing
9.5	• Significant injury after 60 seconds exposure. • 1% lethality after 80 seconds exposure	No significant damage
4	• Significant injury after 100 seconds exposure.	No significant damage
1.6	• Pain threshold met after 60 seconds	No significant damage

Key Hazard Scenarios and Assumptions

1. A release from a tank connection creating a hydrogen vapour cloud explosion from a liquid hydrogen storage tank release.
2. The failure of a tank truck loading hose leading to a hydrogen vapour cloud explosion due to a liquid hydrogen tank truck release.
3. A fire impinging on one of the storage tanks long enough to create a Boiling Liquid Expanding Vapour Explosion (BLEVE)
4. The ignition of a release of liquid hydrogen leading to a jet fire.
5. Liquid nitrogen will stay close to ground level as a liquid or vapour.

CONSEQUENCE ANALYSIS

The consequence analysis is focused on the following possibilities:

1. A release from a tank connection creating a hydrogen vapour cloud explosion from a liquid hydrogen storage tank release.
2. The failure of a tank truck loading hose leading to a hydrogen vapour cloud explosion due to a liquid hydrogen tank truck release.
3. A fire impinging on one of the storage tanks long enough to create a Boiling Liquid Expanding Vapour Explosion (BLEVE)
4. The ignition of a release of liquid hydrogen leading to a jet fire.
5. Liquid nitrogen will stay close to ground level as a liquid or vapour.

Atmospheric Conditions

All releases are subject to different scenarios depending on the atmospheric stability at the time of release. Atmospheric stability categories are used to describe turbulence. When modeling differing scenarios assumptions need to be made around time of day, wind speed, cloudiness, and the sun's intensity. There are six (6) categories denoted by the letters "A" through "F", with "A" being very unstable, "D" being neutral and "F" being very stable. "D" and "F" are typically used for the Edmonton area.

Table 4: Climatic Stability Class Categories (*The Pasquill stability classes*)

Stability class	Definition	Stability class	Definition
A	very unstable	D	neutral
B	unstable	E	slightly stable
C	slightly unstable	F	stable

Meteorological conditions that define the Pasquill stability classes

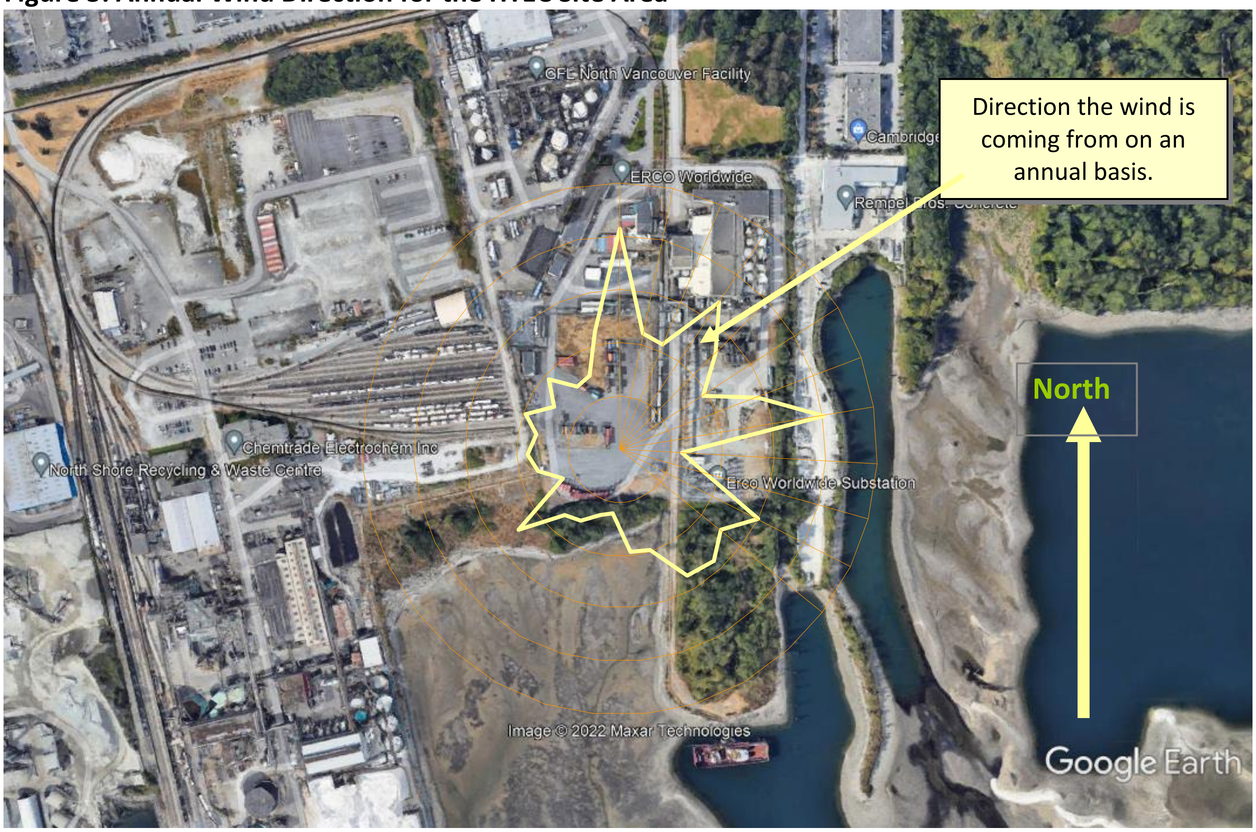
Surface wind speed		Daytime Incoming Solar Radiation			Nighttime cloud cover	
m/s	mi/h	Strong	Moderate	Slight	> 50%	< 50%
< 2	< 5	A	A – B	B	E	F
2 – 3	5 – 7	A – B	B	C	E	F
3 – 5	7 – 11	B	B – C	C	D	E
5 – 6	11 – 13	C	C – D	D	D	D
> 6	> 13	C	D	D	D	D

Note: Class D applies o heavily overcast skies, at any wind speed day or night

North Vancouver Second Narrows Wind Rose

The "annual" wind rose for the area is typically needed to show the potential annual exposure to people from a release of a chemical cloud as it moves downwind. Risk is calculated on an annual exposure basis. As hydrogen vapour will not form low lying clouds that will move with the winds but instead release upwards entirely and not pose a concern for any toxic effect for the public doing such an exposure analysis is not necessary. However, by showing the wind rose for the area does show any cloud would be directed away from the downwind public either towards the Southwest or the Northeast.

Figure 3: Annual Wind Direction for the HTEC Site Area



Source; British Columbia Ministry of the Environment Historical database (Dec 31, 2019 to Dec 31, 2020 with an average wind speed 1.7m/second)

Consequence Modelling Approach for Risk Assessments

Before conducting the consequence analysis for the noted hazards, each hazard will look at the realistic worst-case scenario and a smaller scenario.

The probability numbers for the worst cases are available from reliable sources, however in reality lesser size incidents can occur more often. To be able to describe risk levels for 1 X 10^{-6} up to 1 X 10^{-4} CSChE-PSM (MIACC) uses a referenced method. The method, based on work developed globally for looking at smaller events, which can occur more frequently. Based on global incident history, this method assumes a small release to be 10% of the worst-case scenario and it will happen 100 times more often than the worst-case scenario. *Reference: MIACC "Risk Assessment Guidelines for Municipalities and Industry – Initial Screening Tool - September 1997*).

Each scenario will calculate consequence values for both the worst case and the 10% more likely case. The analysis is shown in the "Risk" section of this study.

SCENARIO ONE: HYDROGEN VAPOUR CLOUD EXPLOSION FOR A LIQUID H_2 STORAGE TANK RELEASE

Each of the two liquid hydrogen storage tanks has a volume of 90,000USG or 340.2m^3.
Liquid hydrogen density = 71.0 kg/m^3
Therefore, one tank containing liquid H_2 would contain 24,154kg (*use 24,000kg for the calculations*).

Scenario

For typically a realistic worst-case release will be from a nozzle failure on the tank. The largest nozzle on either tank is 114mm (4inch) or 168mm (6inch). As a standard for calculating releases any pipe over 50mm (2inch) in diameter uses an opening of 20% of the cross sectional area of the pipe. Therefore for:

114mm pipe area = 10,207mm^2 @ 20% = 2,041mm^2 or a 25mm (1inch) diameter hole
168mm pipe area = 22,167mm^2 @ 20% = 4,433mm^2 or a 38mm (1.5inch) diameter hole

Realistically, the loss of the entire tank liquid contents through a ruptured pipe connection is reasonable. A puncture of the tank wall is an unlikely scenario as the tank is a double walled tank giving additional protection. This realistic worst-case scenario would be the releasing of liquid H_2 to ground forming a pool with flashing to a vapour cloud. Flow-rates are calculated for a pressure drop from the storage tank to atmosphere of 151.7 kPa.

Viscosities and densities lead to a calculated flow rate:

μ_{H2} = .000009kg/ms (*viscosity is not a concern*)
ρ_{H2} = 71kg/m^3

Calculating the liquid release rate (L) for the storage tank operating pressures:

$$L\ (kg/sec) = \left(9.44 \times 10^{-7}\right)\left(D^2\right)\left(\rho_{H2}\right)\left(1{,}000P_g/\rho_{H2} + 9.8\Delta h\right)^{1/2}$$

Where: D^2 = 625mm^2 (1inch hole)
ρ_{H2} = 71kg/m^3
P_g = 151.7 kPa (storage tank pressure)
Δh = 1m (height of liquid above the release point)

Therefore: L = 1.9kg/sec

Where: D^2 = 1,444mm^2 (1.5 inch hole)
ρ_{H2} = 71kg/m^3
P_g = 151.7 kPa (storage tank pressure)
Δh = 1m (height of liquid above the release point)

Therefore: L = 4.5kg/sec

(*Assuming the liquid release is going to take place until a full tank of 24,000kg is emptied the 25mm hole (1 inch) will take 3.4 hours to empty and for 38mm hole(1 ½ inch) 1.5 hours to empty the tank.*)

As the liquid H_2 is released to the ground, some will "flash" to a vapour cloud depending on the temperature of the liquid. Assuming as the worst case a very hot day and a temperature of 30°C outside, the quantity of H_2 flashed to a vapour cloud is:

$$F_V = \frac{C_P}{H_V}(T_S - T_B)$$ **WHERE: F_V = FRACTION FLASHED TO VAPOUR**

C_P = Heat capacity of the liquid H_2 (14,310 J/kg°K)
H_V = Heat of vapourization for H_2 (447,000 J/kg)
T_S = Operating temperature (30°C= 303°K) –maximum
T_B = Normal boiling point (-259°C = 14°K)

The resulting flashed fraction (F_V) of the spill of vessel contents calculated to be 9.25. For a colder temperature of -20°C, F_v = 7.65. If the F_v is greater than 0.2 it is then assumed all of the spilled liquid hydrogen is vapourized and there is no pool developed.

The resulting hazard distances for the flashed H_2 liquid on a hot day (30°C) are. The liquid hydrogen, released through a 25mm (1") opening at a rate of 1.9kg/sec will all flash as it spills from the storage tank. For a 38mm (1 ½ ") opening the spill rate for the same storage tank pressure is 4.5kg/sec.

Vapour Cloud Explosion Impact Distances

(*Based on equations from the EPA's RMP Off-Site Consequence Analysis Guidance (May 24, 1996)*

***NOTE**: As flashing occurs, some liquid will be entrained as droplets. Some of the droplets are quite small and travel with the vapour while the larger droplets fall to the ground and collect in a pool. As an approximation, the amount of material staying in the vapour as droplets is five times the quantity flashed. Therefore, if 20% of the material flashes, the entire stream becomes airborne and there is no pool formed. In this case no pool is formed.*

For vapour cloud explosion, the total quantity of flammable hydrogen released is assumed to form the vapour cloud. The entire cloud is assumed to be within the flammability limits, and the cloud is assumed to explode. As a standard, 10% of the flammable vapour in the cloud is assumed to participate in the explosion. The distance to the one pound per square inch overpressure level is determined using equation:

$$X = 17\left(0.1\, W_f \frac{H_{cf}}{H_{cTNT}}\right)^{1/3}$$

Where if all the liquid hydrogen were to be released in seconds:

X = distance to overpressure of 1 psi = 713m ← Maximum case
W_f = weight of flammable substance 24,000kg
H_{Cf} = heat of combustion of hydrogen = 144,000 kJ/kg
H_{CTNT} = heat of combustion of TNT (4,680 kJ/kg)

Assuming the smaller scenario, the release to be 10% of the worst-case:

$$X = 17\left(\frac{0.1\, W_f\, H_{cf}}{H_{cTNT}}\right)^{1/3}$$

Where if all the liquid hydrogen were to be released in seconds:

X = distance to overpressure of 1 psi = 331m

Minimum case

W_f = weight of flammable substance 2,400kg

H_{Cf} = heat of combustion of hydrogen = 144,000 kJ/kg

H_{CTNT} = heat of combustion of TNT (4,680 kJ/kg)

However, the scenario will likely be a quick ignition meaning the explosion will consume a much smaller amount of hydrogen followed by a fire. In order to determine the impact distances the below calculations describe the quantity of hydrogen released over different times before an explosion occurs. The result is how far the explosion will have a 1.0psi overpressure impact. The calculation does include an explosion efficiency of 10%, which is considered acceptable where not all the hydrogen is involved in the explosion. Not all releases of flammable vapors ignite. It is recognized only 10 – 20% of them do ignite. Hydrogen has a very low minimum ignition energy (0.011mJ) compared to other fuels and susceptible to static discharges as ignition sources. For this review, it is chosen that if the flammable cloud is to explode it will occur very shortly afterward the initial release. This results in an impact felt as far as 39m from the rupture for the 25mm (1") hole and 51m for the 38mm (1 ½") hole.

Table 5: Distances (metres) to a 1.0 Psi Overpressure Impact for a 25mm Diameter Hole

Length of time of release (sec)	1	2	3	4	5		10
Release Quantity	1.9kg	3.8kg	5.7kg	7.6kg	9.5kg		19kg
Worst case (1" hole) at 1.9 kg/sec	31m	39m	44m	49m	52m		66m
(10%) Realistic case (1" hole) at 0.2 kg/sec	14m	18m	21m	23m	24m		31m

Table 6: Distances (metres) to a 1.0 Psi Overpressure Impact for a 38mm Diameter Hole

Time (sec) Release Quantity	1	2	3	4	5		10
Release Quantity	4.5kg	9.0kg	13.5kg	18.0kg	22.5kg		45kg
Worst case (1.5" hole) at 4.5 kg/sec	41m	51m	59m	65m	70m		88m
(10%) Realistic (1.5" hole) at 0.5 kg/sec	20m	25m	28m	31m	34m		42m

Reference: EPA Risk Management Program Guidance for Offsite Consequence Analysis.

Table 7: Effects of Explosion Overpressure on Structures (*CPQRA 1989*[19])

Pressure (psi)	Damage
0.02	Annoying noise (137dB if of low frequency, 10 – 15Hz)
0.03	Breaking of large glass windows under strain
0.04	Loud noise 143dB, sonic boom, glass fracture
0.1	Breakage of small glass windows under strain
0.15	Typical pressure for glass breakage
0.3	**"Safe Distance" (probability of 0.95 no serious damage beyond this value); projectile limits; some damage to house ceilings; 10% window glass broken**
0.4	Limited minor structural damage
0.5 – 1.0	Large and small windows usually shattered; occasional damage to window frames
0.7	Minor damage to house structures
1.0	**Partial demolition of houses; made uninhabitable**
1 – 2	Corrugated asbestos shattered; corrugated steel or aluminum panels fastening fails followed by buckling; wood panel fastenings of standard housing fail; panels blown away
1.3	Steel frames of clad buildings slightly distorted
2	Partial collapse of walls and roofs of houses
2 – 3	Concrete or cinder blocks shattered if not reinforced
2.3	Lower limit of serious structural damage
2.5	50% destruction of brickwork of houses
3	Heavy machines (300lbs), industrial buildings suffered little damage; steel framed buildings distorted and pulled away from foundation
3 – 4	Frameless, self-framing steel panel buildings demolished; rupture of oil storage tanks
4	Cladding of light industrial buildings ruptured
5	Wooden utility poles snapped
5 – 7	Nearly complete destruction of houses
7	Loaded rail cars overturned
7 – 8	Brick panels, 8 – 12 inches thick, not reinforced, fail by shearing or flexure
9	Loaded train boxcars completely demolished
10	Probable total destruction of buildings; heavy machine tools (7,000lbs) moved and badly damaged; heavy machine tools (12,000lbs) survive.
300	Limit of crate lip

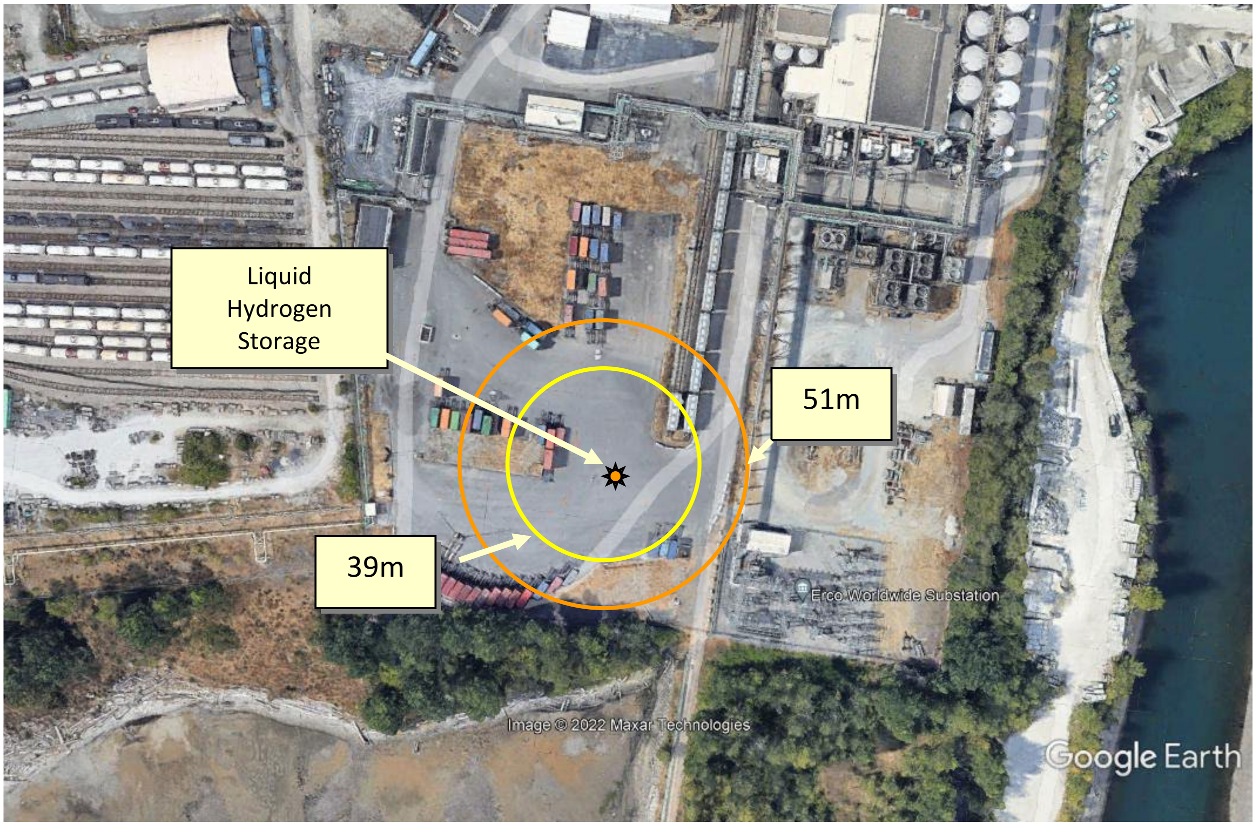
Electrostatic Discharges as an Ignition Source

(*Lees, Frank P., Loss Prevention in the Process Industries – second edition (1995), ISBN 0 7506 1547 8)*

A concern for hydrogen facilities. For a discharge to be incendive a minimum voltage and a minimum energy are required. It is usually reckoned that for a spark discharge from a conductive object into a flammable mixture the minimum voltage is 1,000V (1kV). As far as the discharge is concerned, the incendivity of the discharge is generally assessed by comparing the energy of the discharge with the minimum ignition energy (MIE) of the flammable gas. Usually most often quoted is for a Propane – Air mixture where the MIE = 0.2mJ, (for hydrogen that is 0.011mJ, very low).

The minimum energy potential required in terms of the energy of the discharge and the efficiency of that energy in effecting ignition relative to the energy in the spark. Commonly a value of 100V is used which incorporates a 10 fold safety factor. This means very sound ground grounding and ground monitoring is important.

Figure 4: Distances (metres) to a 1.0 Psi Overpressure Impact



Summary

The explosion will see the shockwave contained with local damage to company and none to neighbouring facilities as a result. It is expected any ignition will take place shortly after the rupture and release. The realistic result is an overpressure of 1.0psi would be felt at a distance of 39m for a 25mm (1") hole up to 51m for a 37.5mm (1 ½") hole release from the source of the explosion.

SCENARIO TWO: HYDROGEN VAPOUR CLOUD EXPLOSION FOR A LIQUID H_2 TANK TRUCK RELEASE

Flashed Vapour Cloud Hazard Distances

Tank Truck volume is generally about 9,000 USG or 34.0m^3.

Liquid hydrogen density = 71.0 kg/m^3

Therefore one tank truck containing liquid H_2 would contain 1,156.7 kg.

Scenario

A typical tank truck carrier connected to one of the two process storage tanks by hose connection for loading purposes. They will be stationary and connected directly to the site storage tanks with a top mounted 25mm (1") loading line. Pressure within the storage tank used to load the liquid hydrogen into the tank truck. A 25mm (1") vent hose connected back to the process is used to remove the vapour from the tank truck. There are no other connections to the tank truck. Realistically, the loss of the entire tank truck liquid contents through a ruptured loading hose is reasonable. A puncture of the tank truck wall is an unlikely scenario as the tank truck is not moving and it is a double walled tank giving additional protection. The more realistic worst case scenario would be the failure of the loading line releasing the liquid H_2 to ground immediately flashing to a vapour cloud. Flow-rates are calculated for a pressure drop from the storage tank to atmosphere of 151.7 kPa.

Viscosities and densities lead to a calculated flow rate:

μ_{H2} = .000009kg/ms (*viscosity is not a concern*)

ρ_{H2} = 71kg/m^3

Calculating the liquid release rate (L) for different storage tank pressures:

(Assuming the liquid release is going to take place for 5 minute until it is shut off)

$$L\ (kg/sec) = \left(9.44 \times 10^{-7}\right)\left(D^2\right)\left(\rho_{H2}\right)\left(1{,}000Pg/\rho_{H2} + 9.8\Delta h\right)^{1/2}$$

Where: D^2 = 625 mm^2

ρ_{H2} = 71kg/m^3

Pg = 151.7 kPa (storage tank pressure)

Δh = 1m (height of liquid above the release point)

Therefore: L = 1.9 kg/sec

Flashed Liquid H_2 Hazard Distances

As the liquid H_2 is released to the ground some will "flash" to a vapour cloud depending on the temperature of the liquid. Assuming as the worst case a very hot day and a temperature of 30^{o}C outside, the quantity of H_2 flashed to a vapour cloud is:

$$F_V = \frac{C_P (T_S - T_B)}{H_V}$$ **WHERE: F_V = FRACTION FLASHED TO VAPOUR**

C_P = Heat capacity of the liquid H_2 (14,310 J/kg°K)
H_V = Heat of vapourization for H_2 (4.47 X 10^5 J/kg)
T_S = Operating temperature (30°C= 303°K) –maximum
T_B = Normal boiling point (-259°C = 14°K)

The resulting flashed fraction (F_V) of the spill of vessel contents calculated to be 9.25. For a colder temperature of -20°C, F_v = 7.65. If the F_v is greater than 0.2 it is then assumed all of the spilled liquid hydrogen is vapourized and there is no pool developed.

The resulting hazard distances for the flashed H_2 liquid on a hot day (30°C) are. The liquid hydrogen, released through a 25mm (1") opening at a rate of 1.9kg/sec will all flash as it spills from the storage tank.

Vapour Cloud Explosion Impact Distances

(*Based on equations from the EPA's RMP Off-Site Consequence Analysis Guidance (May 24, 1996)*

NOTE: *As flashing occurs, some liquid will be entrained as droplets. Some of the droplets are quite small and travel with the vapour while the larger droplets fall to the ground and collect in a pool. As an approximation, the amount of material staying in the vapour as droplets is five times the quantity flashed. Therefore, if 20% of the material flashes, the entire stream becomes airborne and there is no pool formed. In this case no pool is formed.*

For vapour cloud explosion, the total quantity of flammable hydrogen released is assumed to form the vapour cloud. The entire cloud is assumed to be within the flammability limits, and the cloud is assumed to explode. As a standard, 10% of the flammable vapour in the cloud is assumed to participate in the explosion. The distance to the 1.0psi overpressure level is determined using equation:

$$X = 17\left(0.1\, W_f \frac{H_{cf}}{H_{cTNT}}\right)^{1/3}$$

Where if all the liquid hydrogen were to be released in seconds:

X = distance to overpressure of 1.0 psi = 259m ← Maximum case
W_f = weight of flammable substance 1,156.7kg
H_{Cf} = heat of combustion of hydrogen = 144,000 kJ/kg
H_{CTNT} = heat of combustion of TNT (4,680 kJ/kg)

Assuming the smaller scenario, the release to be 10% of the worst-case:

$$X = 17\left(0.1\ W_f \frac{H_{cf}}{H_{cTNT}}\right)^{1/3}$$

Where if all the liquid hydrogen were to be released in seconds:

X = distance to overpressure of 1 psi = 120m ← Minimum case

W_f = weight of flammable substance 2,400kg

H_{Cf} = heat of combustion of hydrogen = 144,000 kJ/kg

H_{CTNT} = heat of combustion of TNT (4,680 kJ/kg)

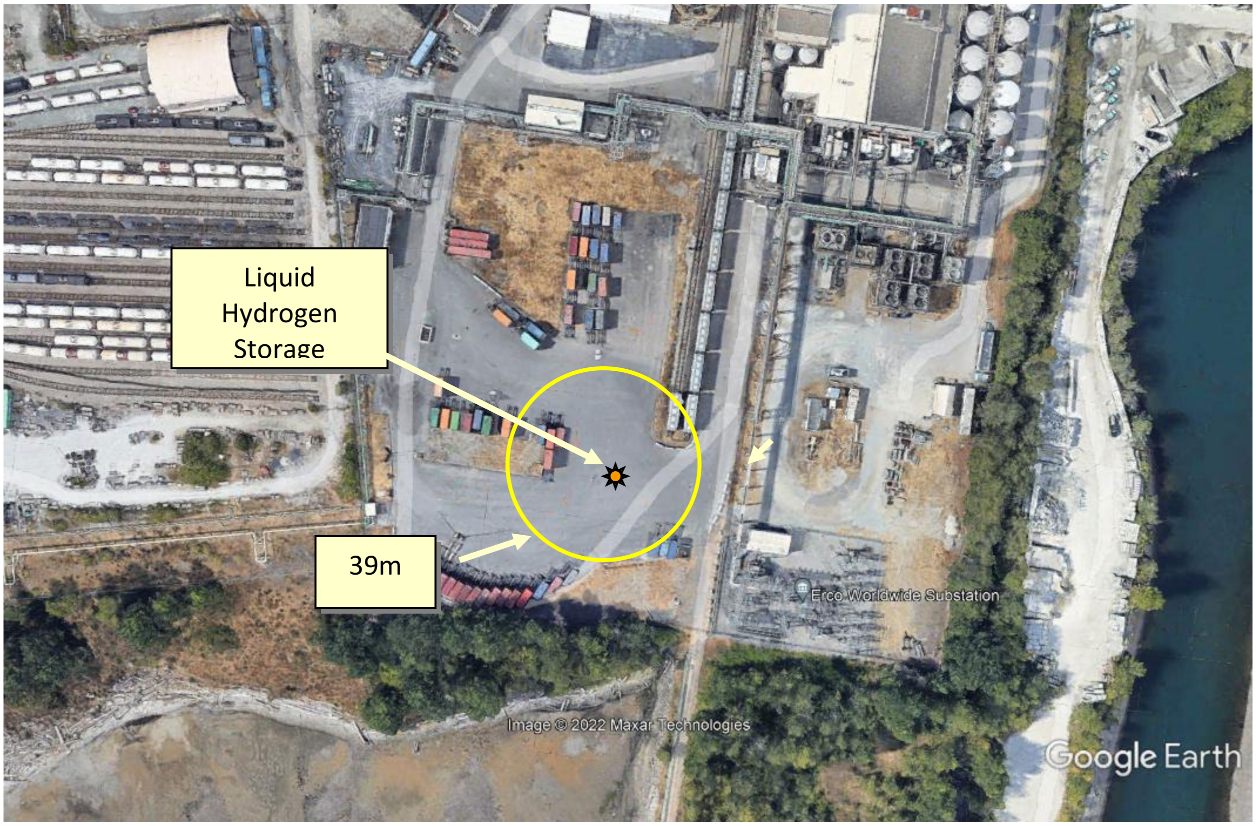
However, the scenario will likely be an ignition shortly after the release happens meaning the explosion will consume a much smaller amount of hydrogen followed by a fire. In order to determine the impact distances the below calculations describe the quantity of hydrogen released over different times before an explosion occurs. The result is how far the explosion will have a 1.0psi overpressure impact. The calculation does include an explosion efficiency of 10%, which is considered acceptable where not all the hydrogen is involved in the explosion. It is expected that if the flammable cloud is to explode it will occur almost simultaneously as the rupture or vey shortly afterwards, less than 1 second. This results in an impact felt as far as 39m from the rupture for the 25mm (1”) hole.

Table 8: Distances (metres) to a 1.0 Psi Overpressure Impact for a 25mm Diameter Hole

Length of time of release (sec)	1	2	3	4	5		10
Release Quantity	1.9kg	3.8kg	5.7kg	7.6kg	9.5kg		19kg
Worst case (1” hole) at 1.9 kg/sec	31m	39m	44m	49m	52m		66m
(10%) Realistic case (1” hole) at 0.2 kg/sec	14m	18m	21m	23m	24m		31m

Reference: EPA Risk Management Program Guidance for Offsite Consequence Analysis.

Figure 5: Distances (metres) to a 1.0 Psi Overpressure Impact



Summary

The explosion will see the shockwave contained with local damage to company and possibly neighbouring facilities as a result. It is expected any ignition will take place immediately after the rupture and release. The realistic result is an overpressure of 1.0psi would be felt at a distance of 39m for a 25mm (1") hole release from the source of the explosion.

SCENARIO THREE: BOILING LIQUID EXPANDING VAPOUR EXPLOSION (BLEVE)

(A result of a jet fire impacting a storage tank or tank truck)

There are two consequences because of a BLEVE to consider. The first one is the immediate overpressure (shockwave) impact from the unconfined vapour cloud explosion followed by a radiant heat impact from the resulting fire.

BLEVE (definition):

The definition of a BLEVE presented by CCPS (1999) (Center for Chemical Process Safety) is "*a sudden release of a large mass of pressurized superheated liquid to the atmosphere*." The sudden release is due to a sudden containment failure caused by fire, a missile, corrosion, a manufacturing defect, internal overheating etc.

1.0PSI OVERPRESSURE SHOCKWAVE

Calculation of the distance to 1.0psi is used as the threshold for the possibility of fatalities to occur. In the calculation, as a standard it is assumed that only 10% of the vapour cloud is involved in the explosion. This is because the released cloud of superheated liquid and vapour (hydrogen in this case) needs to mix with the surrounding air to become flammable. Once ignited the remaining hydrogen is further mixed with air and burns generating a fireball and radiant heat, which is analysed next. (*Reference: EPA Risk Management Program Guidance for Offsite Consequence Analysis).*

$$\text{Distance to 1.0psi} = (17)\left((10\%)W_f \frac{H_{cf}}{H_{CTNT}}\right)^{1/3}$$

W_f = 24,000kg of liquid hydrogen
H_{Cf} = heat of combustion = 144,000 kJ/kg
H_{CTNT} = heat of combustion of TNT (4,680 kJ/kg)

Where if all the liquid hydrogen were to be released in seconds:

X = distance to overpressure of 1 psi = 713m ← Maximum case

W_f = weight of flammable substance 24,000kg
H_{Cf} = heat of combustion of hydrogen = 144,000 kJ/kg
H_{CTNT} = heat of combustion of TNT (4,680 kJ/kg)

NOTE: Typically, risk assessments are calculated on the worst realistic case scenario followed by a lesser event defined as 10% of the worst realistic scenario in order to help to evaluate the reality that smaller events happen more often than the worst case.

In the case of a BLEVE, the event is based on an external heating source impacting the shell of a pressure vessel in the area where there is vapour and no liquid to take away the heat created by the flame. This will happen regardless of the vessel being completely full or partially full. The outcome will be an explosion creating a shockwave.

Therefore, for this risk assessment there is no need to calculate the 10% scenario for a BLEVE.

RADIANT HEAT IMPACT DISTANCE OF 5kW/m²

The impact of the fuel remaining after the initial explosion will be from the resulting fire, which can include the majority of the hydrogen not involved in the explosion. Again using the EPA methodology and equation for a fireball:

$$\text{Distance (X)} = \left(\frac{(2.2)(t_a)(R)(H_c)(W_f)^{0.67}}{(4)(\pi)\left(\frac{3.42 \text{ X } 10^6}{t}\right)^{0.75}} \right)^{½}$$

t_a = Atmospheric transmisity = 1.0
R = Radiating fraction of heat of combustion
= Assumed at 0.4
H_c = Heat of combustion = 144,000kJ/kg
W_f = Weight of flammable material = 24,000kg
t = Duration of the fire = 2.6 $W_f^{1/6}$ = 14.0sec

Substituting into the equation yields a distance of 88m from the source will see a radiant heat exposure of 5kW/m² meaning the potential for fatalities could be within that distance. For a possibility of a fatality the exposure for a fireball for the 5kW/m² radiation intensity is a 60 second exposure. This fireball is calculated to be 14 seconds in duration. The 4kW/m² to 5kW/m² exposure is used as a conservative exposure level to indicate where serious injury and even fatalities can happen provided the exposure time is 60 seconds or longer. This is where extra care is used to ensure a "conservative approach" to describe where fatalities can happen

Figure 6: BLEVE Impact Distance to 1.0 psi and 5kW/m²

SCENARIO FOUR: A JET FIRE AS THE RESULT OF A LIQUID HYDROGEN RELEASE

The source could be a pipeline leak or flange leak off a vessel nozzle or a PSV release that does not close or the case where rupture discs are used.

Pipelines will spew outwards its contents if a leak or rupture were to happen. Assuming for a ruptured liquid hydrogen pipeline the leak is on top of the pipeline (worst case for a Jet fire). The ensuing pressurized hydrogen vapour stream will travel upwards. If ignited (see Appendix “3”) a clean burning jet of fire will rise driven by the pressure inside the pipeline. The heat from the flame will have an impact at a distance that could cause a fatality. The jet fire will be very turbulent creating thorough mixing of the flammable vapour with air leading to an almost complete combustion and an almost smokeless flame. This is the circumspect for hydrogen as it transforms to a vapour form when released.

For a liquid hydrogen pipe rupture and release, the scenario is for a major release from the top of a 60.3mm (2”) *internal diameter = 52.6mm* and 88.9mm (3”) *internal diameter = 78mm,* igniting and forming a jet flame. The method used is from *“AIChE - Center for Chemical Process Safety, Guidelines for Chemical Process Quantitative Risk Analysis 2000, second edition”* which is a recognized source. A standard approach for piping is to use a worst case hole size equal to 20% of the pipe cross sectional area for pipe over 60.3mm (2”). For smaller pipe use the pipe internal diameter.

Jet Flame Length

$$\frac{L}{d_i} = \frac{5.3}{C_T}\left[\frac{T_f/T_j}{\alpha_T}\left(C_T + \left(1 - C_T\right)\frac{M_a}{M_f}\right)\right]^{1/2}$$

Where: L = Flame Length (m)
d_i = Hole Diameter 0.012m & 0.017 m
$M_a = Mw_{air} = 29$
$M_f = MW_{fuel} = 2$
α_T = fuel/air mixture = 1.0
T_f = Flame Temperature = 2,483°K
T_j = Gas Temperature = -195°C = 78°K

Stoichiometric Equation: $2H_2 + O_2 + 3.76N_2 \longrightarrow 2H_2O + 3.76N_2$

(*1 mole of air contains 3.76 moles of N_2*)

Therefore $C_T = \dfrac{1}{2 + 3.76} = 0.174$

Substituting into the equation yields a flame length of 7.2m and 15.0m respectively

Calculating for a Mass Release Rate

$$m = C_D A P_1\left[\left(\frac{k g_c M}{R_g T_1 P_2}\right)\left(\frac{2}{k+1}\right)^{(k+1)/(k-1)}\right]^{1/2}$$

NOTE: *The calculations are for an internal pipeline pressure of 151kPa(a) and a temperature of -250°C.*

Where: m = mass flow-rate of gas through the hole *(discharge rate of methane)*

C_D = Discharge coefficient (*1.0 for choked flow through a hole*)

A = Area of the hole (*0.0001m^2 and 0.0002m^2 respectively)*

P_1 = Pressure upstream of the hole (*151kPa (1.5 X 10^5 N/m^2)*

P_2 = Pressure downstream of the hole (*atmospheric 101,325 N/m^2*)

k = Heat capacity ratio C_p/C_v (*1.41 for hydrogen)*

g_c = Gravitational constant (*1kg m/Ns^2*)

M = MW of the gas (*2 kg/kg mole for hydrogen)*

R_g = Ideal Gas Constant *(0.082057m^3/kg mole ^{o}K)*

T_1 = Initial upstream temperature of the gas (*-250^{o}C pipeline temp 23^{o}K*)

$$m = C_D A P_1 \left[\left(\frac{k g_c M}{R_g T_1 P_2} \right) \left(\frac{2}{k+1} \right)^{(k+1)/(k-1)} \right]^{1/2} = \underline{1.05\text{kg/sec and } 2.10\text{kg/sec respectively}}$$

Radiant Heat Impact of a Hydrogen Gas Jet Fire

Radiant heat impact of 4kW/m^2 over 100seconds represents the distance where serious injury can be expected unless individuals take protective actions. At 9.5kW/m^2 there is a probability of fatalities now happening after 80 seconds of exposure. For this analysis it assumes the pipe is 1m above ground and with the flame height of 7.2m and 15 respectively. With this in mind how far does a radiant heat level of 4kW/m^2 and 9.5kW/m^2 extend from the flame centre?

$F_p = 1 / (4\pi X^2)$ Where F_p = view factor and X = horizontal distance from the flame

$E = \tau_a \eta m \, \Delta H_c F_p$ Where E = Radiant Energy kW/m^2

τ_a = Transmisivity of air = 0.812

η = 0.2 = radiant fraction = for hydrogen

m = 1.05kg/sec and 2.10kg/sec respectively

ΔH_c = 141,584kJ/kg

Solving the equation yields X = 21.9m and 31.0m respectively for a radiant heat energy of 4 kW/m^2, and for an exposure of 9.5kW/m^2, X = 14.2m and 20.1m respectively.

Summary

A jet fire scenario can happen. Generally, they are long in length due to the initiating pressure inside of the pipeline, and since it is hydrogen can burn cleanly exposing nearby facilities. For this ruptured pipe scenario, the 9.5kW/m^2 radiant heat impact will be felt at a distance of 14.2m and 20.1m respectively and means there will be no potential for fatalities to the public.

SCENARIO FIVE: AN ASPHYXIANT CLOUD RESULTING FROM A LIQUID NITROGEN RELEASE

The scenario is a spill of the liquid contents of the largest liquid nitrogen stream, found to be P&ID line number 59-368. Here the operating temperature is just above the boiling point of liquid nitrogen meaning a release will liberate vapour nitrogen reducing the oxygen concentration in the surrounding area. The question is how far down wind will this have an adverse impact on people.

$$F_V = \frac{C_P}{H_V}(T_S - T_B)$$ **WHERE: F_V = FRACTION FLASHED TO VAPOUR**

C_P = Heat capacity of the liquid H_2 (1,040 J/kg°K)
H_V = Heat of vapourization for H_2 (199,000 J/kg)
T_S = Operating temperature (-176.5°C= 96.5°K) –maximum
T_B = Normal boiling point (-195.8°C = 77.2°K)

Substituting, the Fraction Flashed = 10%. Therefore, a pool of liquid nitrogen can form leading to a vapour cloud. Of note is the specific volume for nitrogen vapour (= 0.967) compared to air (= 1.0) is very close meaning the cloud will stay close to the ground mixing with any wind and have a downstream impact.

Largest quantity of nitrogen used for this analysis is stream 59-368 at a liquid nitrogen flow rate of 4,616.3kg/hr, or 1.28kg/sec

Table 9: Liquid Nitrogen

Chemical	Density (kg/m^3)	Contents weight (kg)	Fraction of Liquid Flashed	Airborne Quantity (kg/sec)
nitrogen	808.4 (liquid)	4,616.3 liquid kg/hr	0.10	1.28kg/sec

Definitions Used for Toxic Release Exposures

In the case of nitrogen and specifically liquid nitrogen, the issue is asphyxiation. *The American Conference of Governmental Industrial Hygienists (ACGIH) has defined it as a "simple asphyxiant". When handling liquid nitrogen, make sure that the concentration of oxygen in the breathing air of a person is at least 18% (vol.) or the partial pressure is at least 135Torr. The vapour pressure at the boiling point (-195,8°C) is 760 Torr* which applies to this case.

There are no ERPG values or IDLH values other than ensuring the 18% (vol.) for Normal concentration of O_2 in air is 21% (vol.), meaning there is a difference of 3% (vol.) oxygen content. Using the Dow Chemical Exposure Index methodology the distance (*AIChE, Dow Chemical Company, Chemical Exposure Index Calculation Guide, 2nd edition Sept. 1993):*

Nitrogen MW = 28
Concentration of 3% (vol.) = 30,000ppm
Therefore the Hazard Distance = $6551\left\{AQ/(PPM)(MW)\right\}^{1/2}$
= 8.1m

SUMMARY: Result is no impact will be felt off site.

PROBABILITY ANALYSIS

The probability of failure used for this risk assessment uses the "catastrophic" values for the scenarios describing the worst case and "major failure" for the 10% release scenario describing the more often events. For the most part the database from the Health and Safety Executive (HSE) of the UK is used. This database was developed from incident research and validated through the research arm of the HSE through peer reviews. A valid and respected resource.

Item FR 1.2.3 Hoses and Couplings

ITEM FAILURE RATES

	Probability of failure per transfer x 10^{-6}		
Facility	Guillotine failure	15 mm diameter hole	5 mm diameter hole
Basic facilities	40	1	13
Average facilities	4	0.4	6
Multi safety system facilities	0.2	0.4	6

SPRAY RELEASE FREQUENCY

	Frequency	Effective length of crack
Hose and coupling	1.2×10^{-7} per transfer	Hose diameter

Derivation

91. The hose and coupling probabilities of failure apply to road tanker transfers. The guillotine probabilities of failure are taken from the report by Gould and Glossop, RAS/00/10. An extension of this work by Keeley (RAS/04/03/1) derived the smaller hole probabilities of failure. The work was carried out for chlorine transfer facilities but should be applicable to similar transfer operations. The safety systems applicable to the facilities are pullaway prevention (e.g. wheel chocks, interlock brakes, interlock barriers), pullaway mitigation that stops the flow in the event of pullaway (e.g. short airline, but only if it will separate and activate a shut off valve before the transfer system fails, movement detectors), and hose failure protection (pressure leak test, hose inspection). Facilities have been divided into three categories to typify the range of precautions that might be found in practice:

Basic	These have one pullaway prevention system such as wheel chocks, carry out pressure/leak tests to prevent transfer system leaks and bursts, but have no pullaway mitigation.
Average	Two pullaway prevention systems (one of which should be wheel chocks) as well as inspection and pressure/leak tests to prevent transfer system leaks and bursts but no effective pullaway mitigation.
Multi safety systems	Two pullaway prevention systems, and also an effective pullaway mitigation system and inspection and pressure/leak tests to prevent transfer system leaks and burst.

United Kingdom- Health and Safety Executive "Failure Rate and Event Data for use within Risk Assessments (06/11/2017)

Item FR 1.2.5 Flanges and Gaskets

ITEM FAILURE RATES

Type of event	Failure rate (per year per joint)	Notes
Failure of one segment of a gasket.	5×10^{-6}	The hole size is calculated as the distance between two bolts and the gasket thickness.
Failure of Spiral Wound Gasket	1×10^{-7}	Hole size calculated as gasket thickness multiplied by pipe circumference.

SPRAY RELEASE FREQUENCY

	Frequency	Effective length of crack
Fixed pipe flange	5×10^{-6} per flange per year	Pipe diameter (max 150mm crack length)

Derivation

103. All rates are taken from the MHAU handbook volume 3 (now archived). The 5×10^{-6} value is derived in the Components Failure Rates paper, which is a comparison of 9 sources of joint failure rates derived elsewhere. The values were derived for chlorine duty although the review included LPG, petrochemical, steam/water, nuclear and other data. Assuming a fibre or ring type gasket in a 25 mm pipe, four bolt flange and a 3.2 mm gasket the gasket failure will produce an equivalent hole of 13 mm diameter.

Item FR 1.3 Pipework

ITEM FAILURE RATES

Failure rates (per m per y) for pipework diameter (mm)					
Hole size	0 - 49	50 - 149	150 - 299	300 - 499	500 - 1000
3 mm diameter	1×10^{-5}	2×10^{-6}			
4 mm diameter			1×10^{-6}	8×10^{-7}	7×10^{-7}
25 mm diameter	5×10^{-6}	1×10^{-6}	7×10^{-7}	5×10^{-7}	4×10^{-7}
1/3 pipework diameter			4×10^{-7}	2×10^{-7}	1×10^{-7}
Guillotine	1×10^{-6}	5×10^{-7}	2×10^{-7}	7×10^{-8}	4×10^{-8}

SPRAY RELEASE FREQUENCY

	Frequency	Effective length of crack
Fixed pipework	1×10^{-6} per metre per year	Pipe diameter (max 150mm crack length)

Derivation

106. The original values for pipework diameter < 150 mm were set out in the MHAU handbook volume 3 (now archived). They were derived in the Components Failure Rates paper, which is a comparison of 22 sources of pipework failure rates derived elsewhere. The values were derived for chlorine pipework although the review included LPG, petrochemical, steam/water, nuclear and other data. This information has been updated and augmented in an MHAU Panel discussion and Paper presented by the MHAU Topic Specialist on failure rates. The information presented in the table above is applicable to all process pipework.

107. Failure rates for pipework with a diameter greater than 150 mm are derived in Gould (1997) – Large bore pipework failure rates, which considers data from 9 other references.

108. Further detail on the derivation of the pipework failure rates is given in FRED, Failure Rate and Event Data for Use in Risk Assessment (Betteridge and Gould, 1999).

109. For pipework with diameter greater than 1000mm discussion with the topic specialist is required.

United Kingdom- Health and Safety Executive "Failure Rate and Event Data for use within Risk Assessments (06/11/2017)

Item FR 1.1.2 Refrigerated Ambient Pressure Vessels

ITEM FAILURE RATES

Type of release	Failure rate (per vessel year)
Single walled vessels	
Catastrophic failure	4×10^{-5}
Major failure	1×10^{-4}
Minor failure	8×10^{-5}
Failure with a release of vapour only	2×10^{-4}
Double walled vessels	
Catastrophic failure	5×10^{-7}
Major failure	1×10^{-5}
Minor failure	3×10^{-5}
Failure with a release of vapour only	4×10^{-4}

Item FR 1.1.2.1 LNG Refrigerated Vessels

ITEM FAILURE RATES

Type of Release	Double wall (per vessel year)
Catastrophic	5×10^{-8}
Major failure	1×10^{-6}
Minor failure	3×10^{-6}
Vapour release	4×10^{-5}

RELEASE SIZES

	Hole diameters for Tank volumes (m^3)		
Category	>12000	12000 – 4000	4000 - 450
Major	1000 mm	750 mm	500 mm
Minor	300 mm	225 mm	150 mm

Item FR 1.1.3 Pressure Vessels

43. Failure rates for pressure vessels are further subdivided into those for chlorine vessels, Item FR 1.1.3.1, LPG vessels, Item FR 1.1.3.2, and spherical vessels, Item FR 1.1.3.3. For general pressure vessels the rates below, which are based on those for chlorine vessels, should be used as a starting point.

ITEM FAILURE RATES

Type of release	Failure rate (per vessel year)	Notes
Catastrophic	6×10^{-6}	Upper failures
Catastrophic	4×10^{-6}	Median
Catastrophic	2×10^{-6}	Lower
50 mm diameter hole	5×10^{-6}	
25 mm diameter hole	5×10^{-6}	
13 mm diameter hole	1×10^{-5}	
6 mm diameter hole	4×10^{-5}	

United Kingdom- Health and Safety Executive "Failure Rate and Event Data for use within Risk Assessments (06/11/2017)

BLEVE Probability

A BLEVE scenario involves an outside source for a flame to impinge on the tank surface wall above the liquid level in the vessel. This flame impingement will heat the outer surface wall of the tank and because there is no means to remove the heat through the liquid in the tank the wall will over heat and weaken. As well the temperature of liquid contents inside the vessel will increase and as a result increase the internal pressure of the vessel. Once the vessel wall is weakened enough, it will succumb to the pressure inside eventually opening releasing the heated hydrogen in a very short timeframe (milliseconds) to create a vapour cloud and explosion.

Specific to the two 90,000usg tanks which are double walled and under 50kPa pressure, atmospheric design, the event could happen very quickly. The source of the fuel to provide the flame will be from the neighbouring liquid hydrogen storage tank or tank truck as there is no other fuel source present in the immediate area.

HSE 2017:

The predicted BLEVE frequency of a selected 2000 m^3 butane sphere on a refinery site.	M Selway	August 1988	Determines BLEVE frequency of an LPG tank to be 9×10^{-7} per yr (p 24).
An initial prediction of the BLEVE frequency of a 100 te butane storage vessel.	K W Blything & A B Reeves	1988	Uses fault tree analysis (FTA) to determine BLEVE frequency of a butane tank to be 10^{-8} to 10^{-6} per vessel year

Item FR 3.2.2.1 LPG Road Tanker BLEVE

ITEM FAILURE RATES

Failure Category	Failure Rate (per delivery)	Applicability
Sites with small tanks	1×10^{-7}	Few/no mitigation measures
Sites with large tanks	1.1×10^{-8}	Significant number of mitigation measures present

TYPICAL MITIGATION MEASURES

Fixed water sprays/ deluge system
Passive fire protection coating on vessels
Portable fire fighting equipment
Fire wall
Storage compound protection (e.g. fencing)
Control of ignition sources
Hazard warning notices

Derivation

146. All rates are based on the report by Z. Chaplin, MSU/LET/2011/38. Typical mitigation measures are detailed in the LP Gas Association Code of Practice 1.

147. Small tanks are considered to typically have a capacity of less than 5 tonnes. Such tanks are likely to be found at domestic or educational sites and are unlikely to have any built-in mitigation systems.

148. Large tanks are more likely to be found at larger industrial installations and have capacities of around 25 tonnes or greater. These types of site are likely to contain a significant number of the mitigation measures listed.

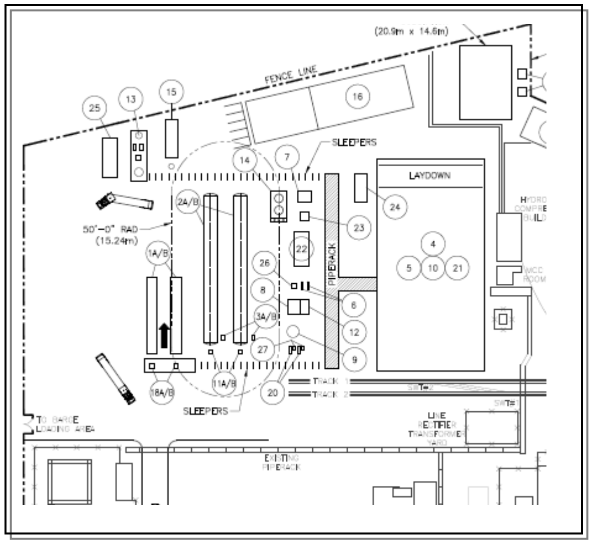
United Kingdom- Health and Safety Executive "Failure Rate and Event Data for use within Risk Assessments (06/11/2017)

SUMMARY

The HSE studies show for Butane storage tanks, which are not under a high pressure the probability, is in the range of 1 X 10^{-6} to 1 X 10^{-8} of a BLEVE. As shown above for sites with large tank truck tanks and mitigation measures the probability of a BLEVE is 1.1 X 10^{-8}. To support this additional studies support a probability of a BLEVE happening as 5.19 X 10^{-8} to 1 X 10^{-8} (*Major Hazards and Their Management, by G. L. Wells, Institute of Chemical Engineers UK – 1997*)

Therefore, for the two hydrogen liquid storage tanks, a worst-case probability of 1 X 10^{-7} is chosen and for the smaller more frequent scenario a probability of 1 X 10^{-7} (as noted above is the same). If the project includes several mitigation measures, the chosen probability is a conservative one.

To note, a BLEVE requires an external source of heat to impinge on the non-wetted portion of the storage tank to weaken the container and allowing the pressure in side to rupture releasing the overheated hydrogen liquid and for it to ignite. In this case, the storage tanks and tank trucks are the source. As the pressure inside the tanks is low the expected "torch" flame that would impinge on a storage tanks or tank truck will not be long in length, therefore simply separating the tanks and the truck loading area with a flame resistant wall (concrete as an example) or ensure the individual tanks have firewater deluge systems would effectively prevent an impinging flame from happening. This could remove the probability of a BLEVE. However, if the probability were reduced to 1×10^{-8} or 1×10^{-9} the outcome would be a lower risk impact.



Human Reliability Analysis

Human involvement has proven to be often at the root of the causes for incidents. In addition, these causes are shown to be management failings for the most part. The human factor component is included in the data below; it just has not been differentiated from the different causes. Suffice it to say management systems focused on people and peoples' actions is important.

Please note the statistics from Alberta Transportation showed 47.5% of highway incidents have some human factor component involved. Also, note the data from the Center for Chemical Process Safety in the US shows a significant incident because of operator error will happen once every 28 years (3.57×10^{-2}).

Table 10: Alberta Highway Dangerous Goods Incident - Causes

Cause	1st Q	2nd Q	3rd Q	4th Q	Total	%
Environment	3	2	1	7	13	7.2
Human Factor	23	18	27	18	86	***47.5***
Insecure	5	4	4	1	14	7.7
Equip. Failure	13	10	8	8	39	21.5
Unknown	3	6	8	6	23	12.7
Vandals	1	0	0	0	1	0.6
Packaging	0	2	1	0	3	1.7
Other	0	0	2	0	2	1.1
Total	48	42	51	40	181	100

(From Hammond & Smith - Table 3 "Causes")

UNCERTAINTY ANALYSIS

The probability data used above is reasonable and straightforward. There will be some uncertainty brought on by:

- Expansion needs in the future may bring in unacceptable products.
- Changes to the operations to include such things as truck and railroad loading/offloading facilities where probabilities of incidents increase.
- The likelihood of ignition happening.
- Impact of human error.

RISK MANAGEMENT PROGRAM

CSChE-PSM (MIACC) "Process Safety Management Standard" 1st edition, 2012 ISBN 978-0-920804-97-1

Introduction

*Expressions appearing in **boldface** in the text are explained in the Glossary*

This Standard identifies the requirements of a **management system** that will address the scope of issues covered by **Process Safety Management** (PSM) for facilities handling or storing potentially hazardous materials. It should be used in conjunction with the *PSM Guide*, which briefly explains the meaning of the elements and components. The approach incorporated in both the *PSM Standard* and the *PSM Guide* is based on that developed by the U.S. Center for Chemical Process Safety (CCPS), based in New York, N.Y. This approach was selected after reviewing several currently available alternatives, and was chosen because it was comprehensive; well-supported by reference materials, tools and an organizational infrastructure; and based on a benchmark of leading or good industry practice rather than on a minimum standard.

Organizations already practising PSM but using a different approach do not necessarily need to switch to the approach given here; however, they should be aware of any items that may not be addressed under their present PSM approach (e.g. human factors) and should be able to demonstrate that they have alternative measures in place that are equivalent in scope and content for proper control of those items. For more information, users should refer to the CCPS book *Guidelines for the Technical Management of Chemical Process Safety* and supporting publications. These are listed in the references given in the *PSM Guide*, which can be found on the PSM page of the Canadian Society for Chemical Engineering website (www.cheminst.ca/PSM).

Purpose

The overarching purpose of this Standard is to identify the performance requirements that can be audited by an organization or a third party to recognize and address gaps that may exist in the overall management system. This Standard identifies the various policies, practices and procedures that will help to ensure the organization achieves the desired results; however, it is not the intent of this Standard to lay out prescriptive solutions that will meet the needs of every organization. Each facility is unique and the user of this Standard will find that a particular policy, practice or procedure that is effective at one site may need to be modified or rewritten for it to be fully effective at another site.

Scope

PSM is the application of management principles and systems to the identification, understanding and control of process hazards to prevent process-related injuries and incidents. This Standard defines the minimum requirements that must be in place to ensure deficiencies are adequately addressed. Such deficiencies can lead to unacceptable risks to safety, health and the environment or losses of assets and/or production. An organization could include the minimum requirements in an integrated health, safety, environmental and risk management program or in a stand-alone PSM program that will be effective in preventing incidents at facilities that manufacture, store, handle or otherwise use potentially hazardous materials.

The PSM system originally suggested by CCPS consists of 12 elements. These elements are shown in Table 1 and are intended to work in conjunction with traditional occupational health and safety programs and applicable federal/provincial/territorial legislation or municipal regulations. A complete framework of PSM elements is recommended for each facility even though some elements or components of PSM may be less applicable to some facilities than to others, depending on the nature and degree of potential hazards involved. A facility should evaluate the applicability of each item before assuming that it does not apply.

Table 1: Elements and Components of PSM[1]

1. Accountability: Objectives and Goals
Continuity of operations
Continuity of systems
Continuity of organization
Quality process
Control of exceptions
Alternative methods
Management accessibility
Communications
Company expectations

2. Process Knowledge and Documentation
Chemical and occupational health hazards
Process definition/design criteria
Process and equipment design
Protective systems
Normal and upset conditions
Process risk management decisions
Company memory

3. Capital Project Review and Design Procedures
Appropriation request procedures
Hazard reviews
Siting
Plot plan
Process design and review procedures
Project management procedures and controls

4. Process Risk Management
Hazard identification
Risk analysis of operations
Reduction of risk
Residual risk management
Process management during emergencies
Encouraging client and supplier companies to adopt similar risk management practices
Selection of businesses with acceptable risk

5. Management of Change
Change of process technology
Change of facility
Organizational changes
Variance procedures
Permanent changes
Temporary changes

6. Process and Equipment Integrity
Reliability engineering
Materials of construction
Fabrication and inspection procedures
Installation procedures
Preventative maintenance
Process, hardware and systems inspection and testing
Maintenance procedures
Alarm and instrument management
Decommissioning and demolition procedures

7. Human Factors
Operator-process/equipment interface
Administrative control versus engineering control
Human error assessment

8. Training and Performance
Definition of skills and knowledge
Design of operating and maintenance procedures
Initial qualifications assessment
Selection and development of training programs
Measuring performance and effectiveness
Instructor program
Records management
Ongoing performance and refresher training

9. Incident Investigation
Major incidents
Third party participation
Follow-up and resolution
Communication
Incident recording, reporting and analysis
Near-miss reporting

10. Company Standards, Codes and Regulations
External codes/regulations
Internal standards

11. Audits and Corrective Actions
PSM systems audits
Process safety audits
Compliance reviews
Internal/external auditors
Corrective actions

12. Enhancement of Process Safety Knowledge
Quality control programs and process safety
Professional and trade association programs
Technical association programs
Research, development, documentation and implementation
Improved predictive system
Process safety resource centre and reference library

[1] The PSM system described here was originally developed by the CCPS of the American Institute of Chemical Engineers (AIChE). This material is copyright 1989 by the AIChE and is reproduced by permission of the CCPS. In 2007 the CCPS moved from the twelve-element system to a risk-based approach using a different set of categories from those shown in this framework, as facilities in the U.S. were generally familiar with PSM due to regulation by the Occupational Safety and Health Administration (OSHA) and various state authorities; however, the CSChE PSM Division decided to retain the original CCPS system for use in Canada, where PSM is not regulated, as it is more self-explanatory for site operators who may be unfamiliar with what PSM comprises.

SUMMARY

Having an ongoing risk management program involving many of the program elements as described will ensure human factors will not be a factor. A human factors probability is not chosen for this review but certainly can be a factor.

CYBER SECURITY PROBABILITY ANALYSIS

Cyber attacks on manufacturing facilities pose a direct threat in today's world. Effective barriers and management procedures are the current methods to prevent unwanted interventions.

Reference: Cybersecurity Risk Assessment According to ISA/IEC 62443-3-2, author Global Cybersecurity Alliance, August 2022

SCENARIO PROBABILITY SUMMARY

(See Appendix 2 "Probability of ignition")

1. Scenario one: hydrogen vapour cloud explosion for a liquid H_2 storage tank release. Probability chosen including probability of ignition= (4×10^{-5}) (1×10^{-1}) = 4×10^{-6} (catastrophic) and = (1×10^{-4}) (1×10^{-1}) = 1×10^{-5} (major). (*NOTE: The tank is double walled but the external wall is not intended to hold the contents pressure*).

2. Scenario two: The failure of a tank truck loading hose leading to a hydrogen vapour cloud explosion due to a liquid hydrogen tank truck release. Probability chosen including probability of ignition = (4×10^{-6}) (1×10^{-1}) = 4×10^{-7} per transfer

3. Scenario three: A fire impinging on one of the storage tanks long enough to create a Boiling Liquid Expanding Vapour Explosion (BLEVE). Probability chosen = 1×10^{-7}

4. Scenario four: The ignition of a liquid hydrogen release leading to a jet fire. Probability chosen including probability of ignition = (1×10^{-6}) (1×10^{-1}) = 1×10^{-7} per metre of pipe.

5. Scenario five: Liquid nitrogen will stay close to ground level as a liquid or vapour. Probability chosen = (5×10^{-7}) (guillotine) and = 1×10^{-6} (major). These are also per metre of pipe.

RISK ANALYSIS

Risk is the combination of consequence and probability. It is often referred to as

"Risk = Consequence X Probability"

The consequences of concern are:

1. Scenario One: The explosion will see the shockwave contained with local damage to company and neighbouring facilities as a result. It is expected any ignition will take place shortly after the rupture and release. The realistic result is an overpressure of 1.0psi would be felt at a distance of 39m for a 25mm (1”) hole up to 51m for a 37.5mm (1 ½”) hole release from the source of the explosion. Probability chosen including probability of ignition= $(5 \times 10^{-7})(1 \times 10^{-1}) = 5 \times 10^{-8}$ (catastrophic) and = $(1 \times 10^{-5})(1 \times 10^{-1}) = 1 \times 10^{-6}$ (major).

2. Scenario Two: The explosion will see the shockwave contained with local damage to company and possibly neighbouring facilities as a result. It is expected any ignition will take place shortly after the rupture and release. The realistic result is an overpressure of 1.0psi would be felt at a distance of 39m for a 25mm (1”) hole release from the source of the explosion. Probability chosen including probability of ignition = $(4 \times 10^{-6})(1 \times 10^{-1}) = 4 \times 10^{-7}$. Assuming sales at 1,500,000kg/yr and 1,157kg/truck, yields 1,300 trucks per year the probability is approximately $(4 \times 10^{-6})(1.3 \times 10^{3}) = 5.2 \times 10^{-3}$.

3. Scenario Three: 713m from the source will see an over pressure of 1.0psi meaning the potential for fatalities will be within that distance. Probability chosen = 1×10^{-7}. Unlike the fire exposure impact of only 14 seconds an explosion, overpressure is immediate.

4. Scenario Three: 88m for a BLEVE from the source will see a radiant heat exposure of $5kW/m^2$ meaning the potential for fatalities could be within that distance. For a possibility of a fatality, the exposure for a fireball for the $5kW/m^2$ radiation intensity is a 60-second exposure. This fireball is calculated to be 14 seconds in duration. Probability chosen = 0 of the possibility a person will be exposed long enough for a fatality.

5. Scenario Four: A jet fire scenario can happen. Generally, they are long in length due to the initiating pressure inside of the pipeline, and since it is hydrogen can burn cleanly exposing nearby facilities. For this ruptured pipe scenario, the $9.5kW/m^2$ radiant heat impact will be felt at a distance of 14.2m and 20.1m respectively and means there will be no potential for fatalities to the public. Probability chosen including probability of ignition = $(1 \times 10^{-6})(1 \times 10^{-1}) = 1 \times 10^{-7}$.

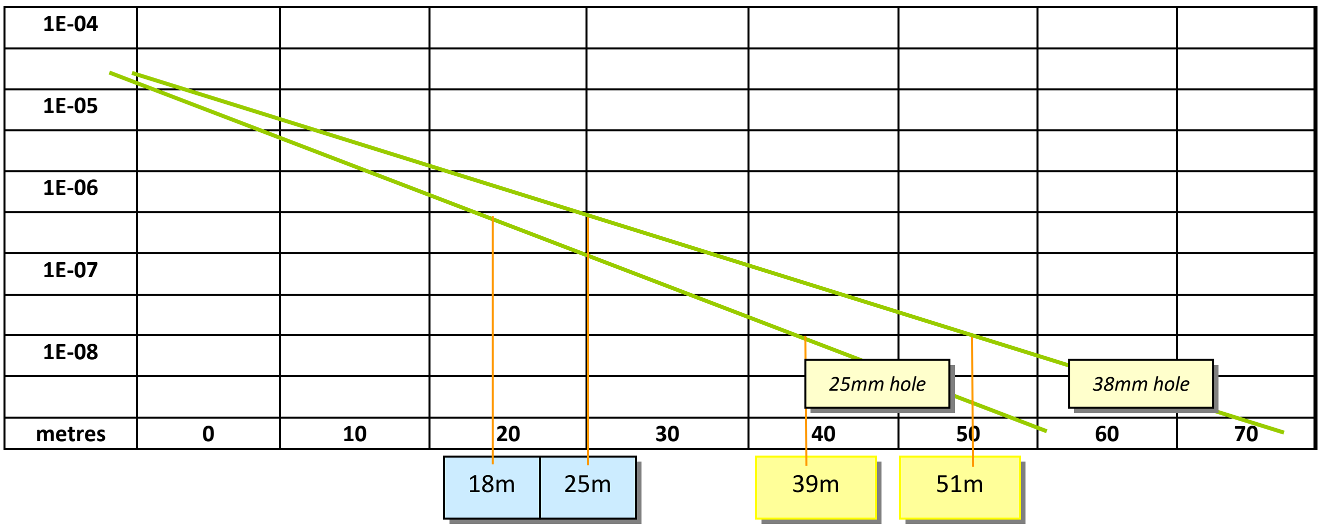
6. Scenario Five: A spill of liquid nitrogen will reduce the oxygen concentration to below 18% for a distance of 8.1m, which will not be felt off site. Probability = 1×10^{-6}.

Risk and Distance Graphs

These probability numbers for the worst cases are acceptable. In reality lesser size incidents can occur more often. To describe risk levels for 1 X 10^{-6} up to 1 X 10^{-4} CSChE-PSM (MIACC) uses a referenced method. The method is based on work developed globally for the purpose of looking at smaller events, which can occur more frequently. Based on global incident history this method assumes a small release to be 10% of the worst-case scenario and it will happen 100 times more often than the worst-case scenario. *Reference: MIACC "Risk Assessment Guidelines for Municipalities and Industry – Initial Screening Tool - September 1997*). Refer to the following "Risk and Distance Graphs".

Figure 7a: Risk and Distance Graph for Scenario 1 (Storage tank spill and explosion)

SCENARIO ONE Hydrogen vapour cloud explosion for a liquid H_2 storage tank release.	**Overpressure Impact (1.0psi) (m)**	**Probability Worst Case**	**Overpressure Impact (1.0psi) (m)**	**Probability Realistic Case**
A 25mm (1”) hole	39m	5.0 X 10^{-8}	18m	1.0 X 10^{-6}
A 38mm (1 ½”) hole	51m	5.0 X 10^{-8}	25m	1.0 X 10^{-6}

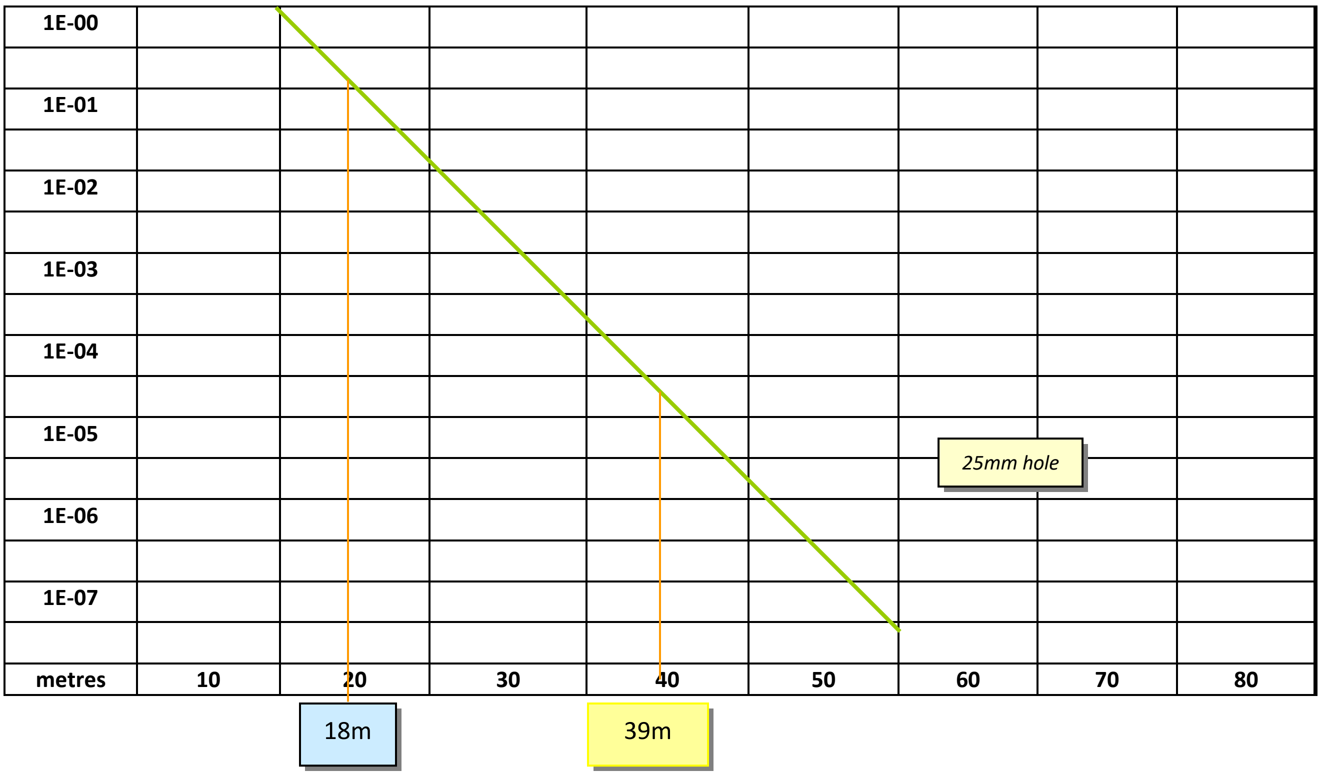


Risk Contour Summary:

- 1 X 10^{-4} - on site impact
- 1 X 10^{-5} - 1m from the source
- 1 X 10^{-6} - 16m for a 25mm hole and 25m for a 38mm diameter hole from the source and on site

Figure 7b: Risk and Distance Graph for Scenario 2 (Tank truck spill and explosion)

SCENARIO TWO Hydrogen vapour cloud explosion for a liquid H_2 tank truck loading hose release.	**Overpressure Impact (1.0psi) (m)**	**Probability Worst Case**	**Overpressure Impact (1.0psi) (m)**	**Probability Realistic Case**
	39m	5.2 X 10^{-3}	18m	5.2 X 10^{-1}



Risk Contour Summary:

- 1 X 10^{-4} on site impact
- 1 X 10^{-5} on site impact
- 1 X 10^{-6} 50m for a 25mm hose diameter hose from the source and on site

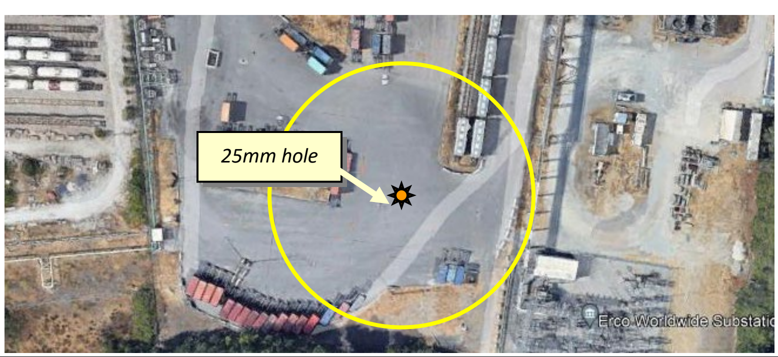
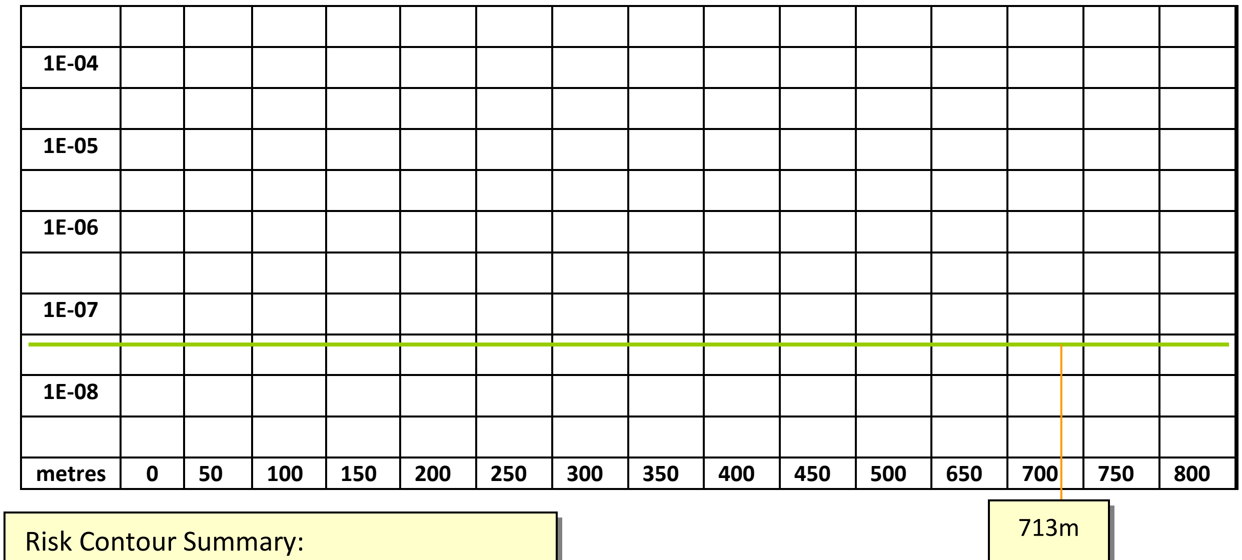


Figure 7c: Risk and Distance Graph for Scenario 3 (BLEVE)

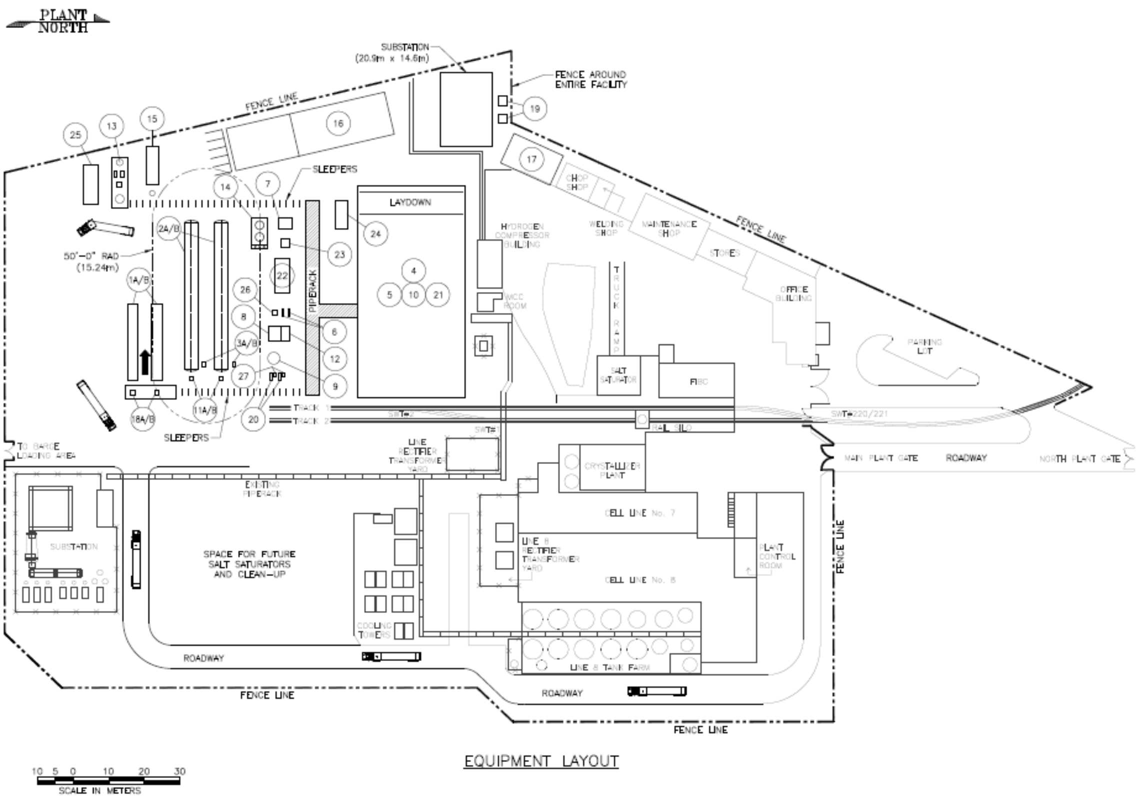
SCENARIO THREE A fire impinging on one of the storage tanks long enough to create a Boiling Liquid Expanding Vapour Explosion (BLEVE)	**1.0 psi Overpressure Impact (m)**	**Probability Worst Case**
	713m	1.0 X 10^{-7}



Risk Contour Summary:

- 1 X 10^{-4} at the source
- 1 X 10^{-5} at the source
- 1 X 10^{-6} at the source

Figure 7c: Site Layout and BLEVE 1 X 10^{-4} Risk Contour



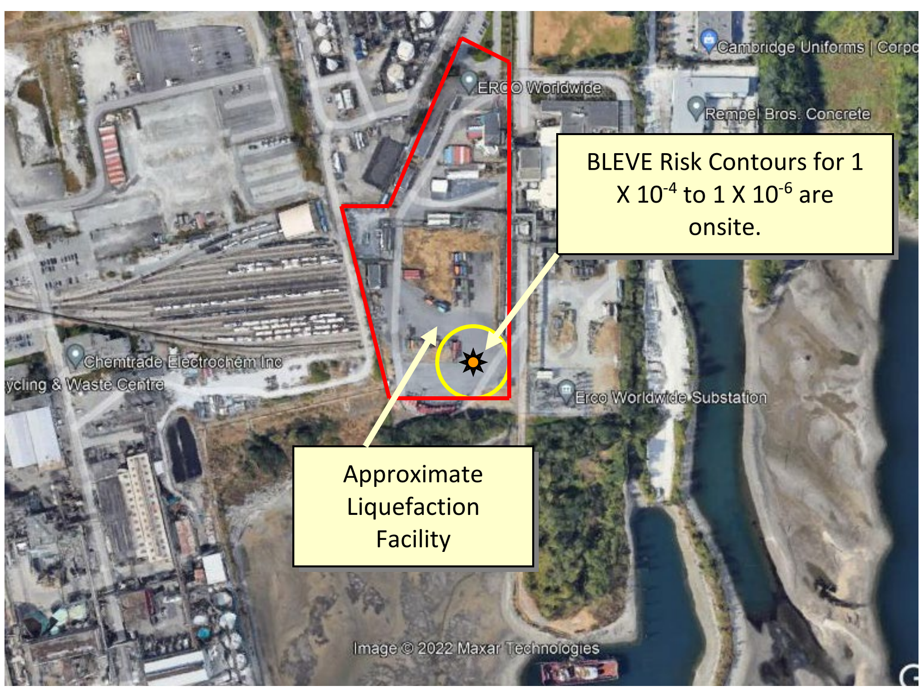
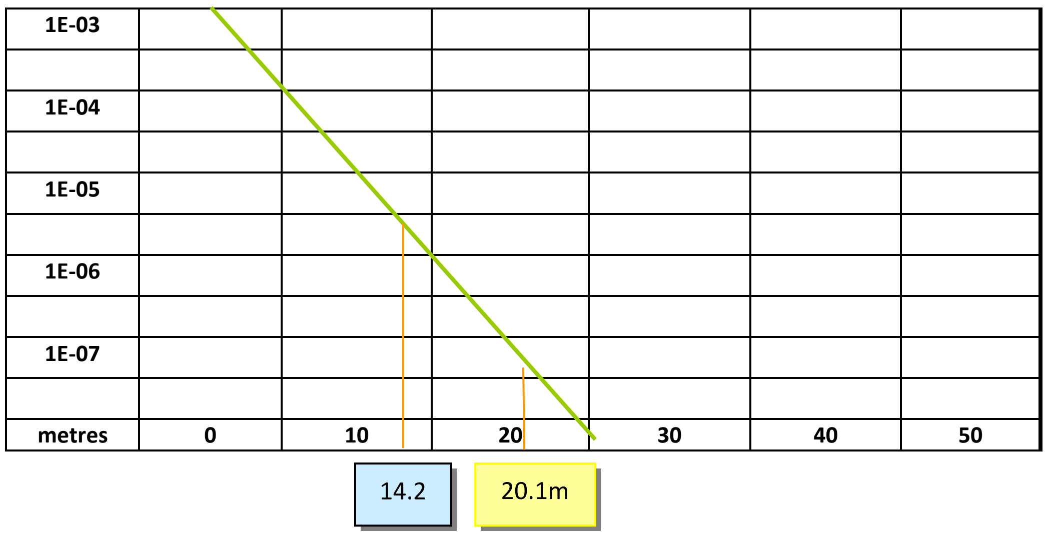


Figure 7d: Risk and Distance Graph for Scenario 4 (Jet Fire)

SCENARIO FOUR The ignition of a liquid hydrogen release leading to a jet fire.	**Radiant Heat Impact (4 kW/m^2) (m)**	**Probability Worst Case**	**Radiant Heat Impact (4 kW/m^2) (m)**	**Probability Realistic Case**
	20.1m	1.0 X 10^{-7}	14.2m	1.0 X 10^{-5}



Risk Contour Summary:

- 1 X 10^{-4} 8m from the source
- 1 X 10^{-5} 14.2m from the source
- 1 X 10^{-6} 17m from the source

(Essentially all risks are on site)

CONCLUSIONS

Discussion of Results

1. All of the scenario risks are within the CSChE-PSM (MIACC) requirements and in all the scenarios, the risk contour of 1 X 10^{-4} is contained on the HTEC site. The risk scenarios indicate all the 1 X 10^{-4} contours remain within the *DNV Industrial Land Use Zone EZ-1* for manufacturing activities along the waterfront.

2. This fireball predicted for the storage tank BLEVE scenario, calculated to have a duration of 14 seconds, would not expose individuals of the public long enough to cause a fatality. Therefore a probability of zero is chosen, which meets the CSChE-PSM (MIACC) criteria.

3. The probabilities used are chosen to represent a conservative view.

4. There are no outstanding issues to be concerned with.

5. Sharing this risk assessment with the DNV's Emergency Services department will aide in emergency planning. This comment is only a courtesy and not suggesting any oversight responsibility. This is a common statement made from risk assessments to help find their results into emergency plans.

RECOMMENDATIONS

Specific recommendations for your consideration as outcomes of this review include the following. Note that these are not conditions:

1. The risks are acceptable and within the CSChE-PSM (MIACC) criteria for risk based land use planning purposes. Suggested is a good risk management program that would see this level of acceptable risk remain the norm.

2. Suggest considering the addition of a concrete wall between the two storage tanks, which will act to prevent a torch effect from one tank leak affecting the other tank, preventing a possible BLEVE, and resulting shrapnel impacts.

3. Suggest consideration be given to a design for a blast resistant control building for at least a 1.0psi overpressure event.

4. Suggest recognizing the potential impact to workers of a liquid nitrogen release and down wind asphyxiation potential.

5. Cyber Security issues are at the front of controlling unwanted events these days. Although new, there is guidance available to assist companies to build defenses. Consideration to this concern is suggested (See Appendix "3").

6. Consider focusing on Human Factors issues as part of the risk management plan.

7. Consideration to adding flammable vapour detectors.

8. A static charge can collect on equipment particularly during a release under pressure. In the case of hydrogen, this potential difference can be an ignition source. Sound and robust equipment grounding with annual checking is suggested.

Appendices

1. Risk Management Process
2. Probability of Ignition
3. Cybersecurity Risk Assessment
4. Risk Analysis & Acceptable Level of Risk
5. Maps of the Area
6. References

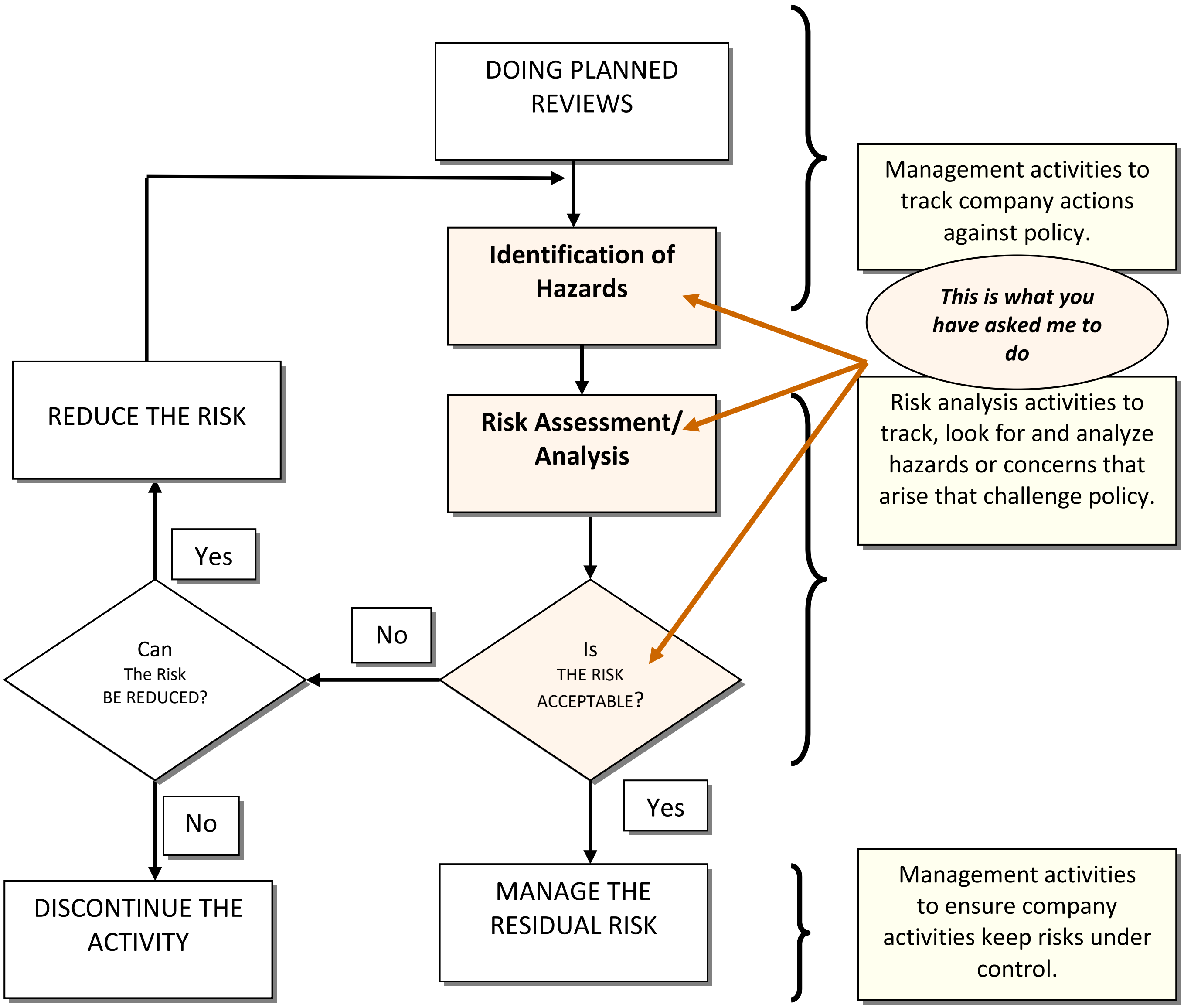
Appendix "1"

Risk Management Process

THE RISK MANAGEMENT PROCESS

This risk management process represents what is practiced around the world particularly for hazardous industries but including others. Each step requires different activities to be conducted in differing formats. The result is a process that has been used successfully globally for over 35 years and is considered to be the best we currently have.

Figure 8: Risk Management Process



What does each box mean?

1. Doing Planned Reviews:

This is a management function. Here you would be conducting whatever reviews you need to do that will provide the data needed to monitor your operations or new project designs. Here is the database for your safety and loss management system. It would include incident investigations, insurance company reviews; regulatory activities (pressure vessel inspections, environmental reporting, asset renewal needs, changes to laws, code updates, etc.). Not to mention the regular data you collect on your business operations and maintenance activities. The point is you want to be proactive so gathering the data and doing trend analyses in conjunction with statistical analyses will keep you ahead of trouble.

2. Identification of Hazards:

One of the outcomes of doing the reviews you mandate as a management team as well as listening to industry activities in general through associations and the news, will be the identification of hazards (or for a better term concerns). Your management team will receive the data and in the wisdom of the team will determine what needs to be further analyzed through doing a risk analysis or analyses.

You may wish to do formal reviews of projects for hazards and this is where a Hazard and Operability Study (HazOp) will come into play. Other tools are available but for the processing industries HazOp's are well thought of. A HazOp can be done on an existing process as well.

It should be noted that legally a hazard analysis is required and once a hazard is identified action to correct the hazard and communicate the concerns is required under the provincial OH&S Act requirements. This emphasizes the need for effective due-diligence by all companies.

3. Risk Assessment/ Analysis:

There are many tools available to help do the risk assessment. There are many tools available to quantify the consequences of all kinds of hazards. Explosions, toxic cloud dispersion models, toxic exposures, lethality, noise, water pollution plumes, etc. etc. All these provide the accurate consequence data you would need to make the right choices.

Probability specifically pertains to the failure of systems, humans, equipment, etc. Data is available generically but the best data is in the company's own database with respect to maintenance records and operational records. Probability (frequency) is also quantifiable.

4. Is the Risk Acceptable?

In order to enjoy the standard of living we as a society would like to have we need to be aware there is a certain amount of risk associated with that. To this end globally, it has been determined it is okay to expose an individual to one chance in a million ($1 X 10^{-6}$) of a fatality on an annual basis due to an industrial activity nearby.

Most company management have developed a risk matrix to describe and communicate company policy. The matrix is used to describe what is a low (acceptable) level risk, medium (acceptable with certain conditions) level risk and high (unacceptable) level risk.

These matrices clarify to employees what they must do and what is acceptable. The low-level risks are usually acceptable without any further management involvement or design additions. Medium risk is the one where management needs to be involved to ensure the risk is kept under control and it is worthwhile noting here management's responsibilities come to the front line as they are assuming the responsibility for taking the risk.

5. Manage the Residual Risk:

Once a risk is determined to be acceptable, it must be managed. This is the largest box in the process as you now have the responsibility for assuming the risk and preventing any incident from happening. This is outlined further in the Process Safety Management systems, which are found around the world as the accepted methods for managing risks.

These consist of 10 – 20 management elements that must be carried out to manage the risks in an acceptable way. Do not forget that once a risk is accepted it does not go away. It is there waiting for an opportunity to happen unless your management systems are actively monitoring your operation for concerns and take proactive actions to correct potential problems.

6. Can the Risk be Reduced?

Often there are ways to reduce the risk once a risk is determined to be unacceptable. The term "Inherently Safe" implies methods, which will eliminate or reduce the risk. Further controls, management systems, protective features, etc. can be added to reduce the risk to an acceptable level.

7. Reduce the Risk:

If the proposed change is viable then do the necessary changes.

Note that once the change is made the process is once again used to evaluate for possible new hazards and risks. Changes in processes often create potential problems upstream or downstream. If they are not uncovered your operational risk may go up unknowingly to yourselves.

8. Discontinue the Activity:

A very important step is to recognize the risk is too high. Management needs to be clear on this one and make the right decisions. Company values, objectives, etc. all come to play in this box including the idea of lost profits, personal promotions, professional defeat, etc.

This statement is a key one because it says you will not do something that is unsafe, pollutes, damages assets, risks your business needlessly, or impacts the public's view of you negatively. Also, your employees are watching your performance and their support for your management decisions is something you need.

There is a psychological component to this too. People will not easily admit defeat when trying to do their jobs. Unless management says and demonstrates that it is okay to stop people will continue to try and succeed which often leads to taking unacceptable risks.

Appendix "2"

Probability of Ignition

By Doug McCutcheon P. Eng.

INTRODUCTION

Not all releases of flammable material ignite. In order to make the risk assessment more in line with this reality the following analysis is provided. Respected sources and evaluation of historical Transportation Safety Board data are used to determine a reasonable and supported approach to include in the probability analysis used in this report.

The release of flammable materials from rail tank cars, highway tank trucks and pipelines are considered. The releases are considered realistic worst cases representing the loss of the entire contents of a rail car or tank truck and a reasonable spill from a ruptured pipeline before it can be isolated. Along with the release of the material necessary consideration is given to the surroundings and the existence of ignition sources. The data used are drawn from a variety of sources specific to the different operational activities. However, the circumstances of the proposed development can be applied to the data in a meaningful way allowing for a reasonable estimation for the probability of ignition.

The literature review includes:

- Frank P. Lees, "Loss Prevention in the Process Industries – Hazard Identification, Assessment and Control, second edition - ISBN 0 7506 1547 8-Section 16.10
- HSE– Research Report 226, "Development of a Method for the Determination of On-site Ignition Probabilities"
- Transportation Safety Board of Canada – "Pipeline Occurrence Data from January 2004"
- Transportation Safety Board of Canada – "Railway Occurrence Data from January 2004"
- Transportation Safety Board of Canada – "Statistical Summary Railway occurrences 2014" – June 2015

The information used from both Lees and the HSE report 226 use research study data from specific industrial activities. It is important to recognize that any release of flammable liquid or gas regardless of the location or facility can ignite. The referenced literature use various categories to make a determination with most of the work focused on industrial activities. There are many ways for flammable materials to be released in many different applications. The industrial activity is the main focus of the literature but the same approach can be used for transportation of flammable materials and other activities. More importantly, regardless of how the flammable material is released, any ignition depends on ignition sources in the area. For open areas the number of ignition sources are limited and in some cases non-existent. For industrial (including offshore drilling) activities the likelihood of ignition sources can be present in large numbers. For rail, truck and pipeline transportation ignition sources in many cases are less likely to be present.

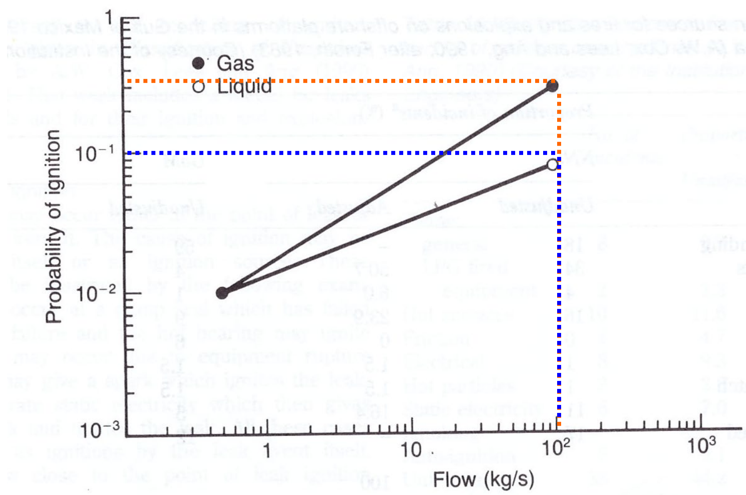
In order to validate the referenced research work the actual data presented by the Transportation Safety Board of Canada with respect to pipelines and railway transport are used. From this a choice as to what probability would be appropriate for this study is made.

FROM LEES LOSS PREVENTION IN THE PROCESS INDUSTRIES

The section in Lees is intended to address the question of probability of ignition for any event regardless of the source of the release. Much of the data involves work conducted using offshore platforms where handling of Crude Oil and Natural Gas are the flammable concerns. A release can be from any source.

Using the figure below from Lees for any release of a flammable material the collective data shows a range of flow rates, which show the likelihood of an ignition, depends on the size of the release. A major pipeline rupture and release and a tank car derailment and spill may be in the order of 10kg/sec to 100kg/sec. resulting in a probability of 3-10%. For a flammable gas release (Natural Gas for example) the likelihood of ignition can be as high as 35%.

Figure 9: Probability of Ignition



FROM THE HSE– RESEARCH REPORT 226

This research report identifies the need to consider the density of ignition sources in the area. It does not assign it to any specific flammable release activity, and assigns a probability range to consider. Here the previous research work is assembled in Table A.4 and a general estimate is provided in Table A.5. The type of incidents could be categorized as "major" at the most for the scenarios identified.

The worst case scenarios for this project include the loss of the flammable liquid contents from a tank car or a tank truck and from Table A.5 would indicate a probability value of up to 0.03 (3%) would be in order. For a Pipeline release the probability that may be in the order of 0.03 (3%) for an Oil pipeline and 0.07 (7%) for the Natural Gas pipeline.

Source	Type of release	Size of release	Location	Probability of ignition	Comments
Kletz (1977)	Polyethylene VCE	mostly small	general - on or near site	10^{-4}	good jet mixing with air
	Hydrogen & hydrocarbons mix (hot, @250bar)	general	general - on or near site	0.033	
		> 10 ton		0.1 to 0.5	
Browning (1969)	LPG release	"massive"	general - on or near site	10^{-1}	Assuming no obvious source of ignition and explosion-proof electrical equipment. Multiply by 10 if strong ignition source present.
	Flammable liquid, flash-point <110°F	general		10^{-2}	
	Flammable liquid, flashpoint 110-200°F	general		10^{-3}	
1st Canvey Report (1978)	LNG vapour clouds	"limited"	general - on or near site	10^{-1}	
		"large"		1	
2nd Canvey Report (1981)	LNG vapour clouds	general	on-site	0.1	"no" sources of ignition
				0.2	"v. few" sources of ignition
				0.5	"few" sources of ignition
				0.9	"many" sources of ignition
Dahl (1983)	Gas	blowouts (massive)	off-shore	0.3	based on 123 incidents
	Oil			0.08	based on 12 incidents

Table A.4 Ignition probability estimates from various incident data reviews by Cox et al (1990)

Source	Type of release	Size of release	Location	Probability of ignition
Cox et al	Gas	minor (<1kg/s)	general	0.01
		major (1-50kg/s)		0.07
		massive (>50kg/s)		0.3
	Liquid	minor (<1kg/s)	general	0.01
		major (1-50kg/s)		0.03
		massive (>50kg/s)		0.08

Table A.5 Summary of ignition probability estimates (Table 1.3) by Cox et al (1990)

Appendix “3”

Cybersecurity

By ISA – Global Cybersecurity Alliance

Cybersecurity Risk Assessment According to ISA/IEC 62443-3-2

By ISA – Global Cybersecurity Alliance

https://gca.isa.org/blog/cybersecurity-risk-assessment-according-to-isa-iec-62443-3-2[2022-08-22 9:15:43 AM]

As cybersecurity for industrial automation continues to evolve, it becomes increasingly important to fundamentally understand, evaluate, and manage cybersecurity risks. Recent attacks such as the one on the Oldsmar Water Treatment Facility further emphasize the need for cybersecurity risk management and demonstrate how cyber incidents have the potential to cause not just financial, but also significant safety and environmental consequences.

The objective of effective cybersecurity management should be to maintain the industrial automation system consistently with corporate risk criteria. In many organizations, ownership for industrial automation cybersecurity concerns falls to controls engineers or similar positions that may have limited time available to focus on security concerns, making it essential that cybersecurity risk is managed in a manner that is both time-efficient and effective.

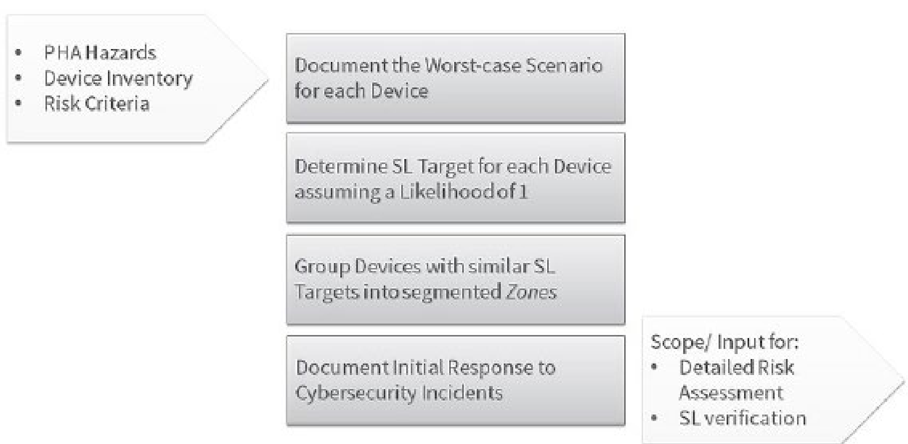
The first step in managing risk is to **understand the current level of risk within a system**. The process for conducting a cybersecurity risk assessment as outlined in the ISA/IEC 62443-3-2 standard is split into two parts:

- Initial Risk Assessment
- Detailed Risk Assessment

Initial Risk Assessment

The Initial Risk Assessment (previously referred to as the High-Level Cybersecurity Risk Assessment) is the starting point for risk analysis activities. Its purpose is to define the scope of future assessments, establish the zone and conduit diagram, establish initial security level targets for devices, and identify high-risk areas for further analysis.

The steps for completing these objectives for a major process area are detailed in the workflow below.



Initial Risk Assessment

The fundamental method behind the Initial Risk Assessment is that it assumes a threat likelihood of one and focuses on evaluating the worst-case scenario if a cyber asset is compromised. This allows a **relatively quick method to determine the highest areas of risk** within an automation system. This method provides easy progression from defining device security level targets to establishing an effective network segmentation strategy by grouping devices with like security requirements into zones and separating zones with boundary devices such as firewalls or data diodes. Combining the results of the Initial Risk Assessment with the operability requirements of the automation system leads to a network architecture that supports both efficient and secure communication between devices.

Although establishing effective network segmentation as described above is easier for new projects, the results of the Initial Risk Assessment still provide benefits to existing facilities by providing an understanding of the highest risk cyber assets in the automation system. This narrows the focus of the Detailed Risk Assessment to the areas that most need it, leading to a reduction in the overall cost and time required for cybersecurity risk assessment activities.

Detailed Risk Assessment

The Detailed Cybersecurity Risk Assessment is the second risk analysis performed for cybersecurity. Its purpose is to gain a definite understanding of the current level of risk within a facility considering potential threat vectors and existing/planned countermeasures ensure that corporate risk criteria are met, and provide detailed cybersecurity requirements for each zone. The steps for completing a Detailed Risk Assessment for a major process area are detailed in the following workflow.

Detailed Risk Assessment

The starting point for the detailed risk assessment is the **output of the Initial Risk Assessment**. In addition to the initial risk assessment results, the full PHA hazards and corporate risk criteria should be available if further questions regarding consequence ranking arise for the site.

The other input for the detailed risk assessment is the **vulnerability analysis**. This can either be done as part of the detailed risk assessment method or before the detailed assessment begins.

The vulnerability analysis reviews the existing network, connected devices, configurations, software versions, and additional factors to identify what vulnerabilities are currently present within a facility and could be targeted by attackers. This provides an important input to the Detailed Risk Assessment when considering the entry points into the system and evaluating how likely a successful attack is—and how easily an attacker can move between devices in the control network.

The first step in the completion of the Detailed Risk Assessment is documenting the potential threat vectors that would provide attackers entry into the system. Depending on the approach taken, this can be a daunting task. I have seen evaluations where asset owners were given a list of more than 300 threat vectors to review and identify which could provide entry into the system.

Although this approach is attempting to be comprehensive by considering detailed threat vectors, it fails to be effective for a couple of fundamental reasons. First, by breaking threat vectors down into so many parts, the amount of time required to complete the assessment is greatly increased, because even for threat vectors that do not apply to the system under consideration, many granularities must be considered. Second, the level of detail completely overwhelms plant personnel because they are not familiar with the detailed ins and outs of cybersecurity analysis, and have now been given hundreds of new terms that they do not understand. Lastly, it does not end up resulting in a more complete analysis of the system because the plant personnel with the knowledge required to evaluate the system cannot speak to the same level of granularity as the selected threat vectors.

Instead of overly confusing the first portion of the risk assessment with hundreds of individual threat vectors, it is helpful to look at manageable categories of attacks. This method helps to provide a complete look at the ways attackers could enter the system, but is still understandable to the plant personnel involved with the risk assessment. The Common Attack Pattern Enumeration and Classification (CAPECTM) database provides common areas of attack that can greatly assist with this process.

If you are thinking that nothing about the "Common Attack Pattern Enumeration and Classification" database sounds less confusing, do not worry. They provide six areas of attack that can be understood by anyone regardless of their level of cybersecurity experience:

- **Social Engineering**: getting into the system by manipulating or exploiting people
- **Supply Chain**: altering the system during production of components, storage, or delivery
- **Communications**: blocking, manipulating, or stealing communications
- **Physical Security**: getting into the system by overcoming weak security measures
- **Software**: getting into the system via vulnerabilities in software applications
- **Hardware**: getting into the system by manipulating the physical hardware of network devices

By starting with broad categories and then moving to the level of detail necessary to evaluate the threats, a Detailed Risk Assessment can be both more efficient and more complete because the personnel with the critical knowledge for the control system will be able to actively contribute to the discussion and provide their valuable knowledge about the system. The level of granularity required is a key difference between the high-level (Initial) and Detailed Risk Assessments.

Another difference from the Initial Risk Assessment, where the likelihood was assumed to be one, is that the **likelihood of a threat** must be considered. When determining the likelihood of an attack, after considering the area of attack, it is typically helpful to start by asking key questions about the threat agent:

What threat agents could execute this attack?

- Internal or external?
- Skilled or unskilled?
- Are nation state-levels resources required?

The above questions can be helpful for understanding how likely the attack would be. It is also important to understand the differences between likelihood from a **functional safety perspective** and **cybersecurity perspective**.

A control engineer must consider a loss of containment event that has a tolerable frequency of 10^{-4} years, whereas IT personnel must consider the hundreds of thousands of attempted cybersecurity intrusions each year. Due to the lack of current well-maintained cybersecurity incident repositories, it is difficult to estimate the likelihood of cybersecurity events with the same level of confidence as causes for a safety risk assessment. As a result, the security community is somewhat split on the best approach for determining likelihood in the Detailed Risk Assessment.

Some experts believe that—because the likelihood cannot be accurately determined—it should be estimated at one, and only consequence severity should be used to prioritize between risks. The other approach is to make conservative estimates that consider the level of skill and access required to execute the attack. There is not one simple answer, but when adopting either approach, it is important to maintain focus on the objective of cybersecurity risk assessment: providing an accurate picture of relative cybersecurity risk to focus resources in the most efficient areas.

In many cases, the consequences identified in the Initial Risk Assessment can be directly applied to the Detailed Risk Assessment, but they should be reviewed to ensure that they are accurate and that no other consequences could potentially result in a higher risk.

After identifying threat consequence pairs for a system, the next step is to identify what **countermeasures** are in place to prevent a successful attack. These countermeasures are any protection that reduces the likelihood of a successful attack.

This step can be achieved by **reducing the potential for an attacker to enter the system** (i.e., by using properly configured firewalls/managed switches, devices with better security capabilities/features, or the least privilege method for assigning access accounts); **increasing the likelihood that an attack would be identified** and stopped before its final objective (i.e., by reviewing firewall logs for unusual access patterns, implementing intrusion detection systems, and verifying code signatures before downloading to the logic solver); or **having measures to stop or mitigate** the end objective of an attack or means of safety risk reduction not susceptible to attack (i.e., hard-coded endpoints in logic solver configurations, pressure relief valves, and pneumatic control loops).

Once the Detailed Risk Assessment has been completed, the zone and conduit diagrams and security level targets from the Initial Risk Assessment should be finalized. The Initial Risk Assessment serves to provide a quick understanding of high-risk areas, and the Detailed Risk Assessment provides a robust understanding of what the threats and countermeasures in those high-risk areas are. The results of the risk assessments provide the key inputs for defining security requirements and the subsequent design phase of the IACS, including the security level verification. They also promote the effective flow of information between lifecycle steps. By understanding the level of risk for a facility, informed decisions addressing cybersecurity concerns can be made to promote safe and secure operation.

Appendix "4"

Risk Analysis

- Acceptable Level of Risk Criteria CSChE-PSM(MIACC)

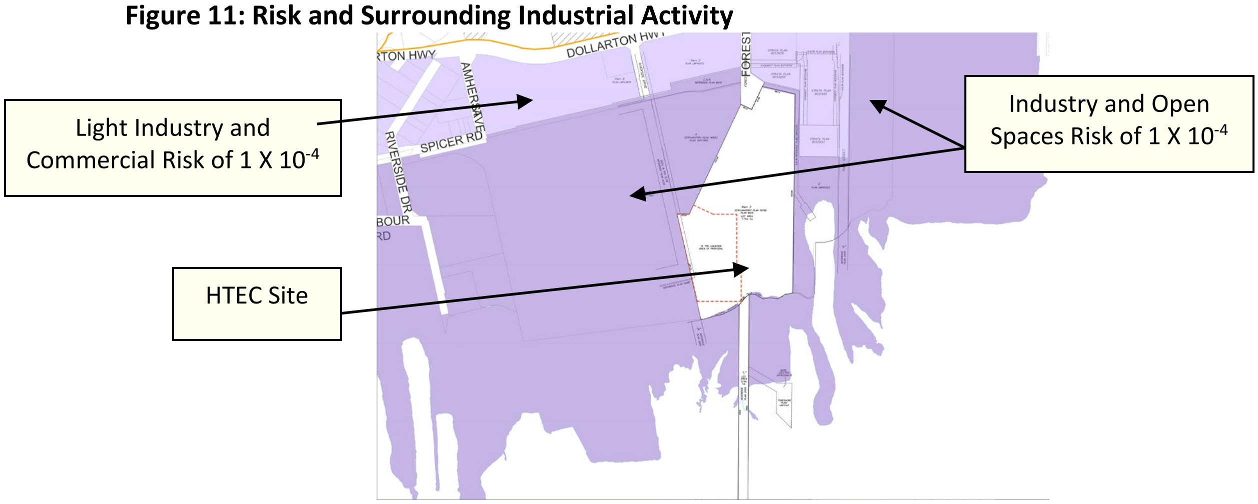
Figure 10: Acceptable Level of Risk Criteria – CSChE-PSM (MIACC)

The CSChE-PSM (MIACC) risk acceptability criteria describes the level of risk for a member of the public who is inadvertently exposed to an industrial incident must be better than a 1 x 10^{-6} chance of a fatality. However, as the risk contour moves towards the source of the risk the risk level increases understandably. However, note that this risk cannot be higher than 1 x 10^{-4} of a fatality. With this in mind, special focus on the workplace is needed to lessen the exposure potential for workers.

This acceptable risk criteria is Canada's approach to a global consensus around industrial risks and land use planning. The concept is developed from a legal conclusion that from a public point of view it is acceptable to have an individual exposed to one chance in a million of being fataly injured over a one year time frame. With this information through the consensus organization called the Major Industrial Accidents Council of Canada now managed through the Canadian Society for Chemical Engineering the above criteria was agreed on.

The type of activity along with the exposure level and density of people all play a part in the determination of the acceptable level for Canada. This is completely in line with the rest of the industrial world.

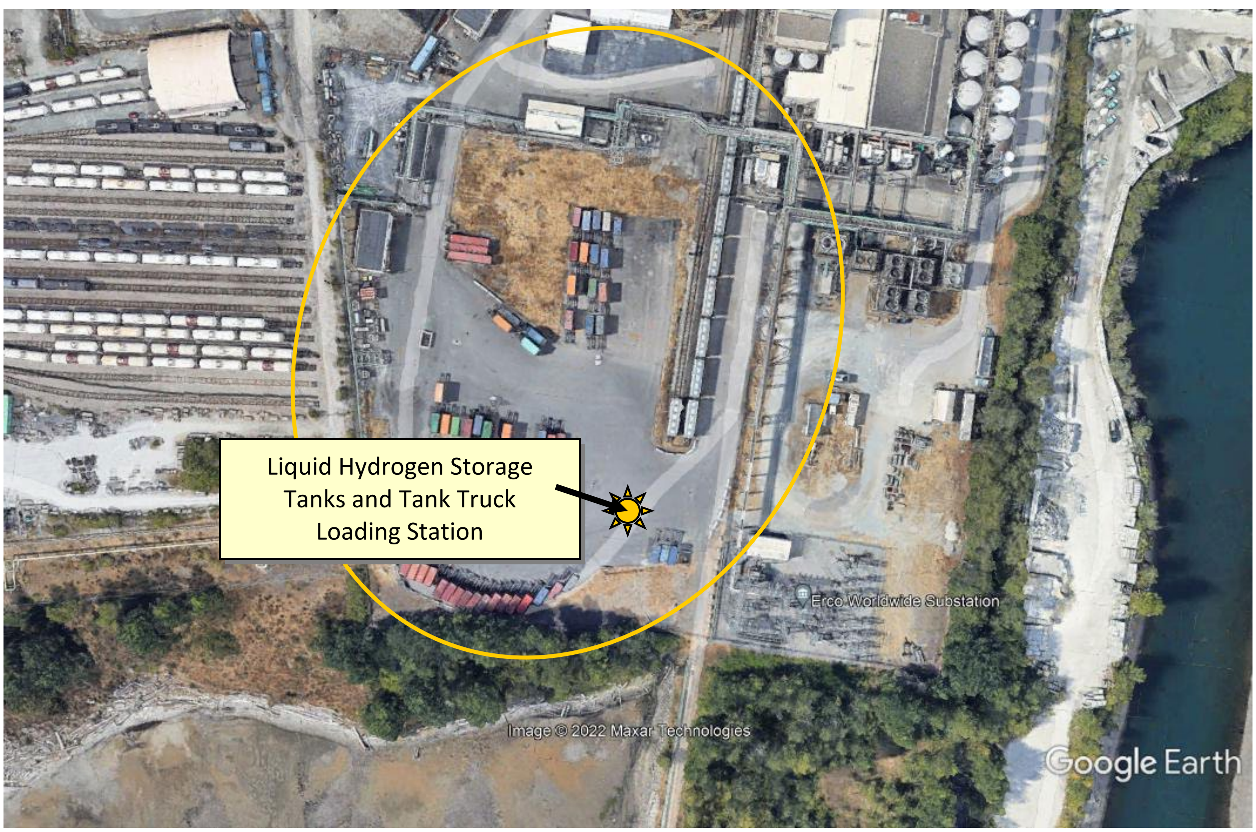
Figure 11: Risk and Surrounding Industrial Activity



Appendix “5”

Maps of the Area

Figure 12: HTEC Liquid Hydrogen Site



Appendix “6”

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"HTEC DNV Liquid Hydrogen Project Risk Assessment Study"

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