Ghent University

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Academic year 2014 – 2015

RISK BASED FIRE PROTECTION STRATEGIES FOR LNG/LPG JETTIES

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Master thesis submitted in the Erasmus Mundus Study Programme

International Master of Science in Fire Safety Engineering

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Henry Eduardo Sánchez Rincón

30 April 2015

“No risks, no rewards”

Ecclesiastes 11:1-6

Summary

A methodology is generated to design fire protection strategies for Liquefied Natural Gas (LNG) and Liquefied Petroleum Gas (LPG) jetties.

The methodology is based on an Event Tree Analysis followed by the design of the fire protection strategies and a cost benefit analysis.

For the Event Tree Analysis, the methodology defines the recommended initiating events, cutsets and corresponding frequencies and probabilities. The likelihood of each scenario is calculated. The damage limits are defined and the impact is estimated using radiation and overpressure contours generated by PHAST 7.1®. The expected consequence is then calculated. The annualised risk is computed from the likelihood and expected consequence of the scenarios.

The fire protection strategies are described together with the associated cost and efficiency when facing a fire.

A parametric study is performed to identify the impact of the different parameters affecting the fire protection strategies design. Finally the methodology is applied to a case study to illustrate its use.

Resumen

Una metodología es desarrollada para diseñar estrategias de protección contra incendio para muelles de Gas Natural Licuado (LNG por sus siglas en inglés) y Gas de Petróleo Licuado (LPG por sus siglas en inglés).

La metodología está basada en un análisis de árbol de eventos seguido por el diseño de las estrategias de protección contra incendio y el análisis costo beneficio.

La metodología define los eventos precursores, los *cutsets* y las frecuencias y probabilidades respectivas para calcular la probabilidad de cada escenario. Los límites de daño son determinados y los contornos de radiación y sobrepresión son generados usando PHAST 7.1® para determinar la consecuencia esperada. El riesgo anual es calculado con la probabilidad y consecuencia esperada de cada escenario.

Las estrategias de protección contra incendio son descritas, incluyendo el costo respectivo y la eficiencia en situaciones de riesgo.

Un estudio paramétrico es llevado a cabo para identificar el impacto de los diferentes factores que afectan el diseño de las estrategias de protección contra incendio. Finalmente, la metodología es utilizada en un caso de estudio para ilustrar su utilidad.

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Acknowledgements

This project is a product of the joint effort of many people, to whom I would like to express my gratitude.

Firstly, to Kris De Troch and Ralf Bruyninckx, who gave me an enormous insight in industrial fire protection engineering and allowed me to acquire new skills and ways of thinking.

To my promotor, Bart Merci, who introduced me to the field of fire safety engineering and gave me the tools to develop this thesis.

To Wouter Van de Velde who gave me fruitful information and insight about the transfer operations.

A special place in my heart will always be occupied by my family. Henry, Clara, Cata, who are my supports and reasons to go ahead. To my extended family, Laura and Andres, who have always been there for me.

To my IMFSE classmates for making this experience unique and exciting. This adventure has been amazing thanks to you.

List of Abbreviations

LNG: Liquified Natural Gas.

LPG: Liquified Petroleum Gas.

$BLEVE$: Boiling Liquid Expanding Vapour Explosion.

RPT: Rapid Phase Transition.

$LOPA$: Layers of Protection Analysis.

$FMEA$: Failure Mode and Effects Analysis.

$F\&EI$: Dow Fire and Explosion Index.

CEI: Dow Chemical Exposure Index.

HRA: Human Reliability Analysis.

IPL: Independent Protection Layer.

PCS: Process Control System.

SIS: Safety Instrumented System.

ERC: Emergency Release Coupler.

D: Diameter of the loading hose or arm.

D_{Max}: Maximum connection diameter.

d_{eq}: Equivalent diameter of the hole

q_S: Release flow rate

C_D: Discharge coefficient [-]

A_h: Cross-sectional area of the hole [m²]

P: Total pressure at opening [N/m²]

P_a: Atmospheric pressure [N/m²]

ρ_L: Liquid density [kg/m³]

$P(t)$: Probability of an ignition in the time t.

$P_{present}$: Probability that the ignition source is present when the cloud passes.

ω: Ignition effectiveness.

F_I: Frequency of the initiating event.

P_D: Probability of direct ignition.

P_V: Probability of delayed ignition.

P_E: Probability of explosion.

F_{Ci}: Frequency of the consequence i.

$C_{i,j}$: Monetary consequence in the asset "i" of the scenario "j".

P_j: Price of the damaged element "j".

K_j: Damage level factor in scenario "j".

AR: Annualised risk.

η_{FPS}: Fire Protection System Efficiency

P_{OR}: Operational reliability

P_{OLA}: On-line availability

P_{RE}: Response Effectiveness

λ: Failure rate

AC_{FPS}: Annualised cost of the Fire Protection System.

AR^*: Recalculated annualised risk.

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1. Introduction

1.1. Liquefied Natural Gas

Liquefied Natural Gas (LNG) is natural gas under ambient conditions, therefore it is mostly methane and ethane, propane and butane can also be found in lower concentrations as well as traces of heavier compounds. For this study, LNG is considered to contain only methane, if the LNG consists of large concentrations of heavier fractions, the results from the modelling have to be re-assessed.

To liquefy the natural gas, it is necessary to bring it down to -162°C at atmospheric pressure, in this state, it is a colour and odourless liquid with half the density of water. The Material Safety Data Sheet can be consulted in Annex 1.

Table 1. Relevant properties and conditions of LNG and LPG

	LNG	**LPG**
Composition	100% Methane	100% Propane
Molecular weight	16 kg/kmol	44 kg/kmol
Temperature	-162 °C	15 °C
Density @ -162 °C	450 kg/m³	500 kg/m³
Flash Point	-188 °C	-188 °C
Boiling Point	-161 °C	-161 °C
Lower Flammability Limit	5%	2.1%
Upper Flammability Limit	15%	10%

The main advantage of liquefying is the high ratio of gas volume over liquid volume (618), which allows transporting large quantities of fuel in a smaller volume.

History and market

LNG usage dates back to the late 19th century, when Karl Von Linde built the first practical compression refrigeration machine. The first LNG installation started operations in 1912 in West Virginia. The first commercial installation was built in 1941 and the first record of ship transportation was in 1959, when it was shown that it was feasible and safe to transport large amounts of LNG [3].

Although the imports of LNG in Europe have been steadily decreasing since 2011, LNG still represents around the 30% of the natural gas market in the world, with emerging markets in Asia and South America. Figure 1 shows the major movements of natural gas through pipeline and LNG for 2013 [4]. The LNG market is expected to grow sustainably in the following decades; reaching 393 Bm³ in 2015 and 758 Bm³ in 2030 worldwide and in Europe the overall trade volume is expected to reach 220 Bm³ by 2020 and 254 Bm³ by 2030 [5].

With the lowest prices in a decade and the available and to be constructed infrastructure, LNG can be a determining factor in the energy outlook for the coming years.

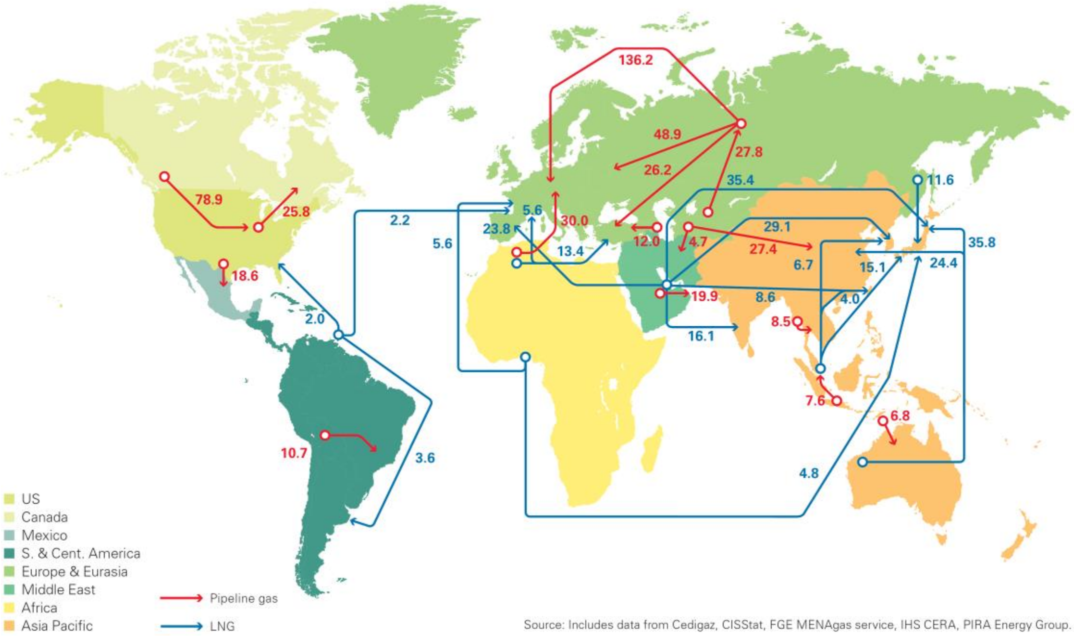


Figure 1. LNG major trade movements in 2013. Values in billion cubic meters. [4]

Production and distribution

The process to produce LNG follows the same steps as for the natural gas depending on the source. When pipeline quality natural gas is obtained it is fed to the cryogenic unit to liquefy it under -162°C using a set of propane, ethane and methane cooling cycles. Subsequently, the LNG is stored for further distribution/transportation.

Most of the transport is made by ship, the carriers are around 300 m in length with a holding capacity between 125,000 and 160,000 m³, and larger ships can carry up to 250,000 m³. In the ship, the LNG is stored in 4 to 6 insulated tanks [6].

Loading of LNG is carried out in the jetty areas, this operation is described further in this chapter, carriers travel from the LNG storage unit to the regasification terminal where unloading operation is carried out. Afterwards the LNG is stored and gasified, odorized and distributed, depending on the future use of the gas. Figure 2 shows a sketch of the general distribution process.

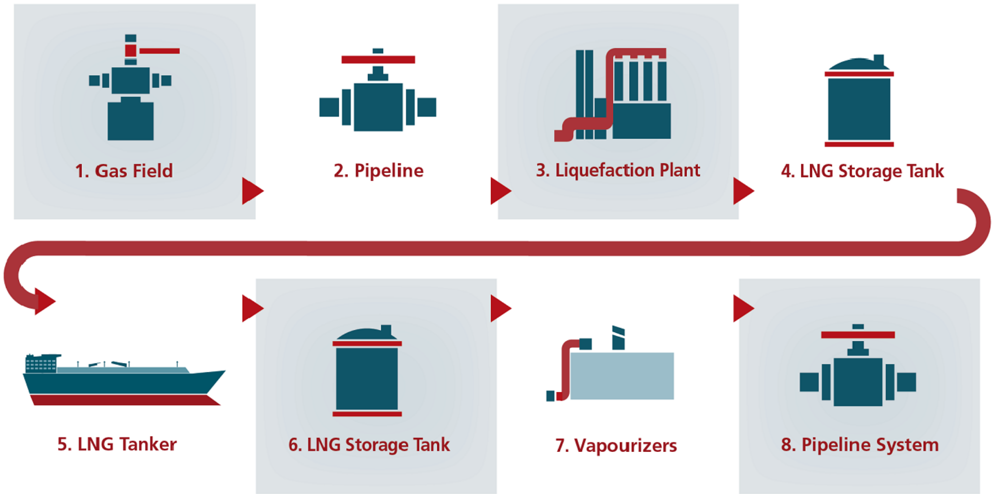


Figure 2. General distribution process of LNG [7]

Hazards

If contained, the LNG is safe since there is no oxidant atmosphere, however if released, LNG is exposed and mixed with air forming a flammable mixture, these releases can be caused either by a rupture, a leak or improper handling.

As the LNG is in cryogenic state before the spill, when released it condensates the moist content in the air forming a visible fog that is dispersed by the wind, however, when it warms up to ambient temperature this cloud is not visible anymore, so the dangerous area can be bigger than the fog.

Because of the turbulence, the mixing of the LNG vapours with air is not uniform, therefore to consider a zone safe, a concentration less than 2.5% is required, which is half of the lower flammability limit. The major potential hazard for unignited LNG is the formation of vapour clouds from a leakage, this cloud can be asphyxiating if the concentration is beyond 50%, but in that case the temperature would also be too low for people. The highest risk is to form a flammable mixture with air.

In case of ignition, the flash fire or fireball will propagate to reach the source of the release leading to a pool fire. This can occur in a time from several seconds to even less than one. If the fireball reaches a confined place in the cloud where there is sufficient mixing with air, an explosion can occur in the form of deflagration and even of detonation if the confinement and mixing levels are sufficient [6].

The heat radiation caused by a burning pool of LNG is 57% higher than the one caused by a similar sized one of gasoline since there is virtually no production of smoke and the combustion efficiency is high.

It is highly unlikely to have an explosion inside an LNG tank, therefore this hazard is disregarded. However, LNG can follow a rapid phase transition (RPT) if the release is on water, which can be considered as an explosion since a sudden overpressure is generated [3][6].

In conclusion, when analysing fire scenarios, it is necessary to consider pool fires, jet fires, vapour cloud fires and explosions [6].

A short review of related accidents can be found in Annex 2.

1.2. Liquefied Petroleum Gas

In a general form, Liquefied Petroleum Gases (LPG) are intermediate compounds between the lighter ones in natural gas and the heavier ones in gasolines, this means compounds with 3 and 4 carbons. It can contain propane, propylene, butanes (n-butane and i-butane), butylenes, and small concentrations of lighter and heavier compounds, like ethane and i-pentane respectively [8]. Table 1 reports the relevant properties of LPG.

History and market

Bottled gas was used as early as 1810 in UK, when compressed gas was sold and distributed, but it was not until the early 20th century that the cryogenic technology was developed and the market registered a steady yearly growth [8]. Its importance lays in the flexibility of the fuel since it can be used in many applications and also the portability as it does not need a fixed network and can be used by users who are not connected to the grid. There is still a high usage in the residential sector with around 47% of the demand [10].

Nowadays the LPG market has a high growing potential due to the lower prices. The fuel, the end of 2014 and beginning of 2015 have seen a decrease of the oil prices to levels seen at the end of the economic crisis in 2009. With growing markets in Africa, Asia and South America, the LPG market is expected to grow.

Production and distribution

Around 75 % of the LPG is extracted from natural gas. Figure 3 shows the basic process to obtain the desired compound; the configuration of the towers determines the fractions to be obtained [9]. The remaining 25% is produced in refineries as a by-product of the several separation processes. LPG is transported by large carriers, pipelines or trains to intermediate storage centres. There it can be bottled in cylinders or loaded to trucks for further distribution [10].

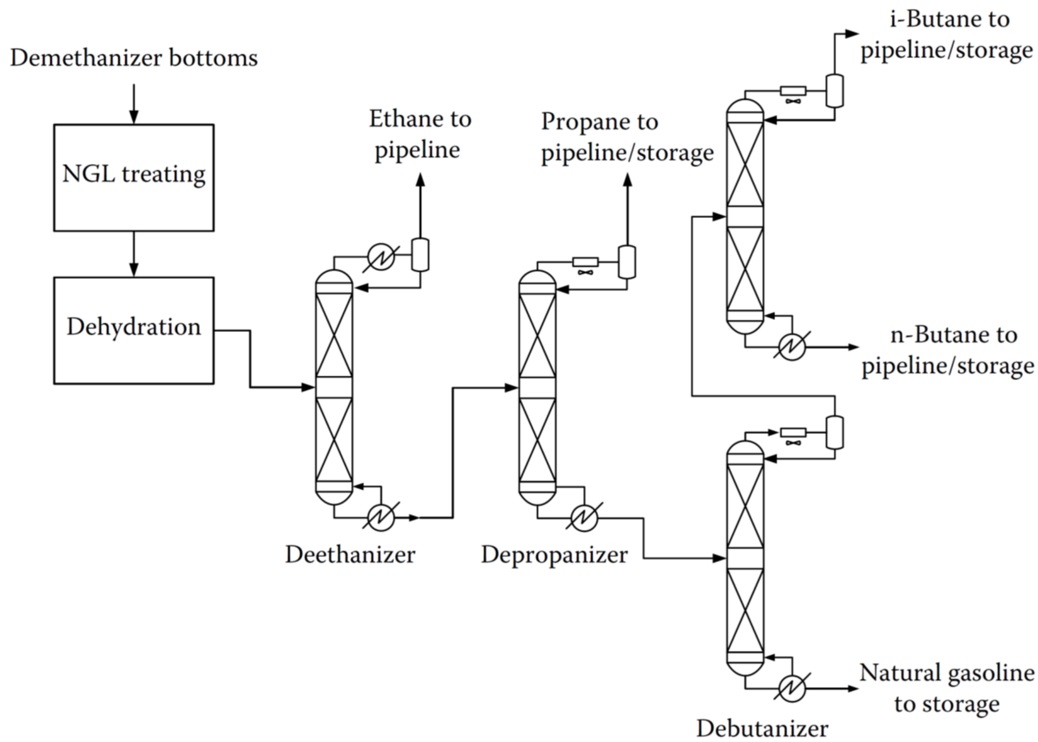


Figure 3. Natural gas liquids fractionation train [9].

Hazards

Similarly as with LNG, when released, LPG is highly dangerous as it quickly forms flammable mixture gaseous clouds with air, for this fuel, the mixture has a considerably high heat of combustion point, therefore the lower flammability limit reaches as low as 1.55% in air [8].

Contrary to LNG, LPG is susceptible to a Boiling Liquid Expanding Vapour Explosion (BLEVE). A BLEVE is a physical explosion due to an increase on temperature followed by an expansion of the liquid and subsequently to the loss of containment. This phenomenon will not be discussed in this document since the outcome is too catastrophic and difficult to control with fire prevention strategies. Many studies have approached this problem. However, this study intends to prevent the occurrence of a BLEVE as a consequence of an ignited release through fire protection mechanisms.

A short review of related accidents can be found in Annex 2.

1.3. Jetty areas

A jetty is a structure that allows berthing operations of ships, ensuring a safe approach and departure navigation; it also contains the necessary structures to allow the unloading of their cargo, in this case liquefied gases. To facilitate these operations the jetty will include fenders for berthing, hooks for quick mooring and monitoring systems to assist the approach and berthing manoeuvres. The (un)loading platform includes the transfer arms and the whole equipment needed for the transfer operations including process and utilities. Support structures are also placed for easy access and safety purposes like road, walkways, monitors and impounding basins among others.

If the jetty is not on the shoreline a trestle is also placed containing the roadways and walkways and the piping rack.

Depending on their configuration jetties can be classified in three classes,

1. Shoreline (Figure 4).

L-type (

2. Figure 5).
3. T-type (Figure 6).

The three of them have space for berthing several ships. Depending on the size of the dock, these ships can be loading and unloading at the same time. Normally L and T-types allow bigger boats to berth since they are further from the shoreline and the depth is bigger.

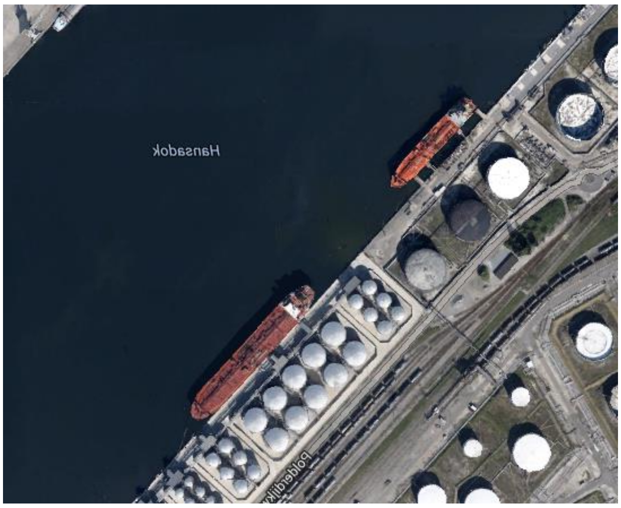


Figure 4. Shore-line jetty. Google Maps [11].

Figure 5. L-type jetty. Google Maps [11].

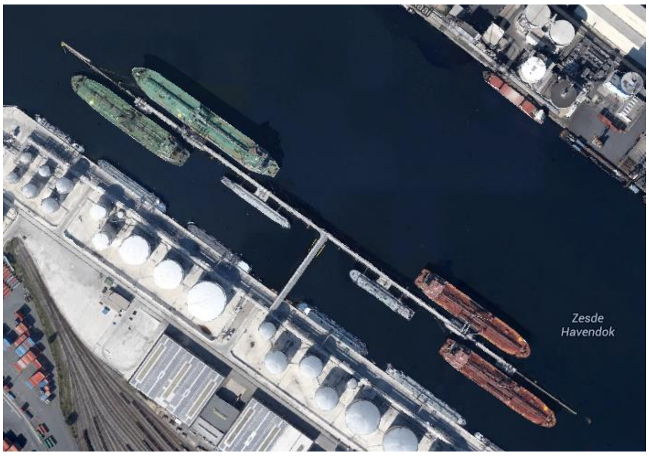


Figure 6. T-type jetty. Google Maps [11].

1.4. Transfer operations

Loading and unloading operations are one of the most critical stages in handling LNG and LPG, since it requires a large level of planning and coordination between the ship and land facilities.

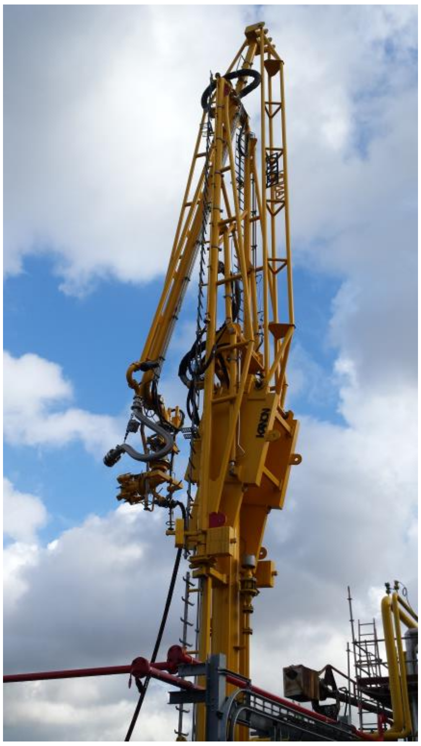


Figure 7. LPG retractile transfer arm at the ATPC facility in Antwerp.

Before the start of the transfer a visible warning sign must be installed and a calibrated gas detector must be available to measure the LPG vapour concentration. Some life safety equipment must be placed on site in case it is needed [12].

In general, the transfer procedure of LNG follows the next procedure:

- For unloading operations, the storage tank must be prepared to receive the boil off gas from the carrier; therefore the operation pressure is reduced so the tank is not affected in case of an overpressure. For loading operations the preparations must be carried out for the ship.
- Berthing and mooring manoeuvres of the carrier ship.
- Establishment of communications.
- Connection of the transfer and the vapour return arms.
- Purging of the transfer lines with nitrogen.
- Cooling down of the transfer arms and auxiliary equipment using the LNG from the carrier.
- Transfer of the LNG using either the send-out pumps at the terminal or the carrier pumps depending on the case. Pressure is balanced in the system using the vapour return arm. Constant checking of the liquid level both in the ship and in the tank to avoid overfilling.
- Pumps decrease their rate before the end of the transfer operation.
- The remaining LNG in the transfer arms is pushed with nitrogen back to the carrier, the rest of it is led to the Jetty KO Drum, from where it is sent to the storage tank.

With some differences, the transfer procedure of LPG follows the procedure described below:

- Berthing and mooring manoeuvres of the carrier ship.
- Establishment of communications.
- Coupling of the articulated arm with the ship.
- Purging and pressurization of the connection using nitrogen.
- Transfer of the LPG using either the pumps at the terminal or the pumps at the ship. Constant checking of the liquid level both in the ship and in the tank to avoid overfilling.
- The remaining LPG in the lines at the end of the transfer operation is pushed using nitrogen partly to the ship and partly to the tank.

1.5. Risk quantification

There are several tools for risk assessment and they must be used with different purposes and for different cases depending on the needs of the project. If the scenario is still considered dangerous after the qualitative and semi-quantitative analysis, a more rigorous quantitative assessment should be carried out.

For quantitative assessments, risk is defined as the likelihood of an accident times the corresponding expected consequence, Equation 1.

Equation 1. Risk definition [14]

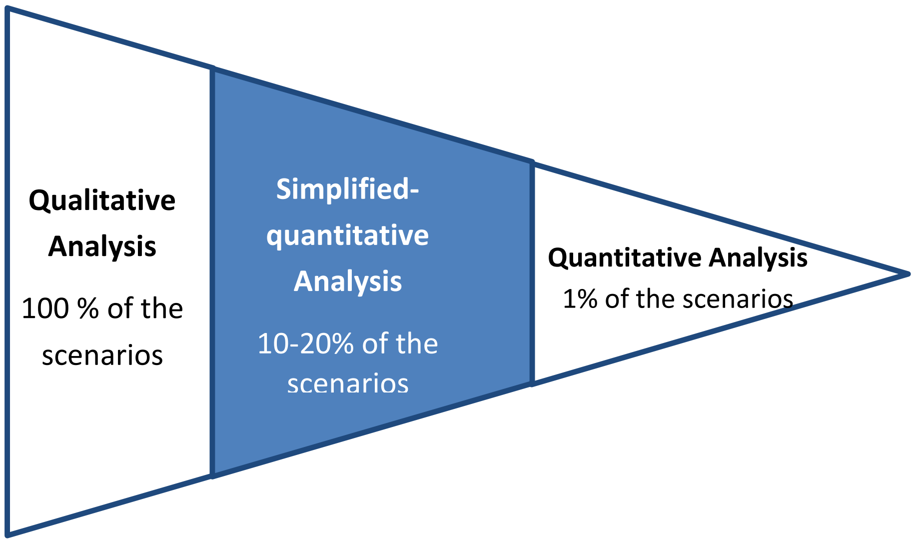
$$Risk = Likelihood\ of\ an\ accident\ \times Expected\ consequences$$

1.5.1. Safety layers of protection

Although this study focuses on the mitigation layer of protection, it is important to keep the other protection layers and the other risk analysis tools and their application in mind. As discussed further in this section, the intention of some layers is to avoid the occurrence of the incident while others will mitigate and avoid the escalation of the consequences.

A global view of the process safety systems can be obtained from the Layers of Protection Analysis (LOPA), which is a global semi quantitative risk assessment method that allows to understand how a process deviation can lead to an incident, accident or catastrophe if not stopped by the available safeguards. This methodology has large and renowned recognition as it has been adopted by the AIChe Center of Chemical Process Safety (CCPS) [15], the International Electrotechnical Commission (IEC) [16] and the International Society of Automation (ISA) [17].

A qualitative analysis is typically the input to a LOPA, which screens the scenarios that must be analysed deeper in a fully quantitative analysis.



Techniques:	HAZOP What-If FMEA	Quantified FMEA F&EI CEI	LOPA	Rough estimate with event tree	Event tree Fault tree HRA
Applicability to simple issues:	Good	Good	Good	Excessive	Excessive
Applicability to complex issues:	Poor	Poor	Fair	Fair	Good

Legend	
FMEA	Failure modes and effects analysis; can be quantified
F&EI	Dow Fire and Explosion Index
CEI	Dow Chemical Exposure Index
HRA	Human Reliability Analysis; uses human success/failure trees to model accident sequences

Figure 8. Spectrum of tools for risk-based decision making [15].

The safety protection of a facility has several types of layer as depicted in Figure 9. For each type of layer there can be 1 or more layers protecting the system. A layer of protection is intended to be independent of the other ones, which is an Independent Protection Layer (IPL). For example, two or more layers cannot rely on a single power source, because in case of failure of the common system, the two layers are useless. This is to avoid the escalation of the incident and to assure the availability of the protection when intended and therefore really needed [15].

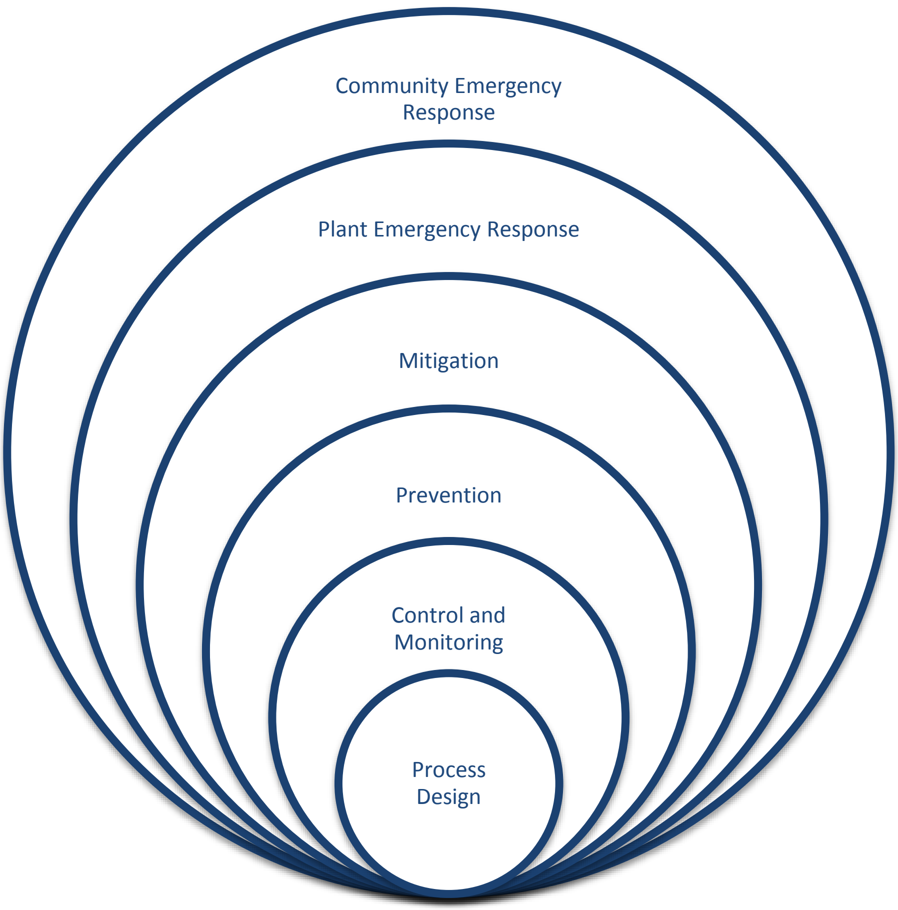


Figure 9. Layers of Protection in an industrial facility

For a deeper knowledge about the uses and procedures of this technique in the process industry, review the AIChe-CCPS [15].

Process design

The main intention of this layer is production and therefore it is the source itself of the risk then it is typically not considered as a risk reduction factor for the system. However, by definition, the process design is inherently safe and is designed to prevent and tolerate certain deviated conditions.

Some process design features can be considered as an IPL if after careful analysis, the feature is proved to actually reduce the risk. Elements to consider in this layer are the process itself, the materials and thickness used in the piping and equipment and the operation philosophy, among others.

Control and monitoring

The control and monitoring system is designed to keep all of the process parameters in the set points or ranges established. This is achieved mainly through valves and instruments, which continuously regulate flows and measure different parameters.

Every process deviation or change in operation is handled through this system. This means that if the control and monitor system is operating according to the design, and the design is inherently safe, no incident should arouse from the process itself. For this reason, the control and monitoring system is normally considered as the first layer of protection, however, additional analysis must be carried out to consider it as an IPL.

As part of this layer we can find the control philosophy and the process control system (PCS), which includes the measurement elements, the connection with the controller, the controller and its software and the final control elements, like valves, regulators, circuit breakers.

Prevention

There are three possible layers in this group, the alarms and human intervention, the action of the Safety Instrumented System (SIS) and the physical protection. All of them are intended to operate before reaching the maximum allowable conditions of the equipment and other elements, therefore they are considered as preventive measures.

The function of the SIS is to detect parameters out of the design range and to prevent the occurrence of a dangerous situation. To achieve this the SIS includes an independent set of measuring devices, controller and final elements.

Normally, alarms are configured to alert in the control room and allow the control staff to intervene. This intervention will depend on the judgement of the person in charge.

If the parameter is still diverging from the set point, the SIS can actuate automatically and lead the process to a safe state, which will vary from equipment to equipment and from process to process.

By definition the SIS is an independent system so it can be considered as an IPL. If the system is more robust, its failure probability is lower.

The physical protection consists of relief devices as relief valves and rupture discs. The intention is to release the pressure affecting the equipment or element by releasing it either to an open or a closed system, by this, the failure and subsequent loss of containment are avoided. Proper maintenance and inspection are crucial for the action of these elements.

Another physical protection used specifically in transfer operations is the Emergency Release Coupler (ERC). This device consists of two valves placed the closest possible together that close and disconnect in case of deviation in the operation parameters or emergency. The appropriate operation of the ERC causes only the release of the volume contained between the two valves, which is virtually nothing.

Mitigation

If the prevention layers are not effective and the accident takes place, the aim is to reduce the extent of the consequences by the installation of the mitigation protection layers. There is a large variety of possibilities in mitigation, like postrelease protection, fire protection and extinguishing systems, foam systems and evacuation procedures among others.

The postrelease protections consist of physical barriers like dikes, impounding basins and blast walls, which will limit the affected area given the loss of containment. If properly designed, these barriers efficiently mitigate the consequences of an event.

The ERC explained in the previous section also plays a role as a mitigation layer, if the emergency compromises the transfer hose / arm, the ERC would cut the flow and limit the extent of the release.

The fixed fire protection systems are included in this category, and they are explained deeper further in the document.

Plant Emergency Response

This category includes the fire brigade, the manual fire protection systems and the evacuation procedures inside the facility. Although these protection layers are effective mitigating the consequence of the accident, they are not normally considered as IPLs since there are many variables affecting their performance.

Community Emergency Response

If the accident scales up and intervention from the community is needed, the community fire brigade and rescue teams play a role in this protection layer. The same as with the Plant Emergency Response, the Community Emergency Response protection layer is not recognised as an IPL.

1.5.2. Event tree analysis

As introduced in the previous section, the event tree analysis is a quantitative risk assessment methodology that identifies and quantifies possible consequences following an initiating event [18].

It can be applied to analyse situations both before and after an incident. The preincident application analyses the systems preventing the initiating events to become incidents or accidents. A typical preincident application is the evaluation of a multilayer protection system to avoid a loss of containment.

On the other hand, the postincident application illustrates the range of possible consequences from an incident [18]. A typical application is the analysis of a loss of containment of a flammable material, which will have different fire consequences depending on the time and conditions of the ignition.

The event tree is constructed from left to right, departing from only one initiating event that splits depending on the cutsets (protection systems or ignition events) to arrive to the branches. Each branch of the tree is a different consequence. A general procedure of an event tree analysis is shown in Figure 10.

This study uses a postincident application of this methodology, and the steps are covered in different chapters, Chapter 3 will cover the first 5 steps, while Chapter 4 will cover the step 6 and Chapter 5 the last step.

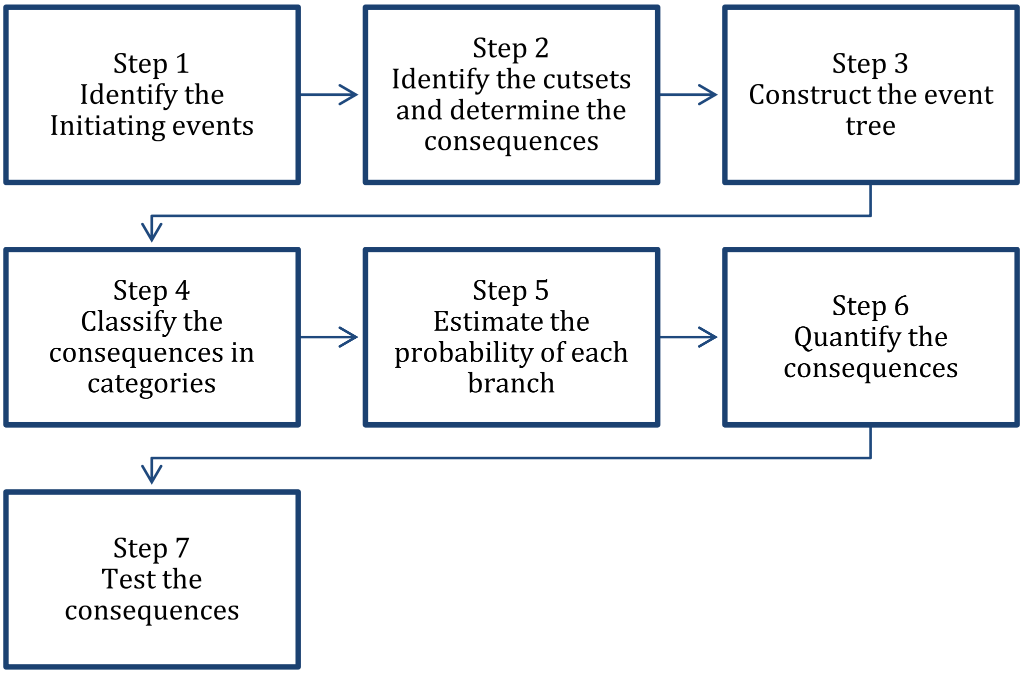


Figure 10. Main steps to perform an Event Tree Analysis [18]

2. Objectives and methodology

2.1. Objectives

- To design a methodology to carry out risk analysis with recommended probability values and scenarios for transfer operations of LNG and LPG.
- To simulate the fire and explosion range for the representative scenarios identified in the risk analysis methodology and perform a parametric study.
- To identify the fire protection strategies pertinent based on the results of the risk analysis methodology and the simulation models.
- To assess the cost-effectiveness of fire protection measures.
- To illustrate the use of the designed methodology by applying it to a case study.

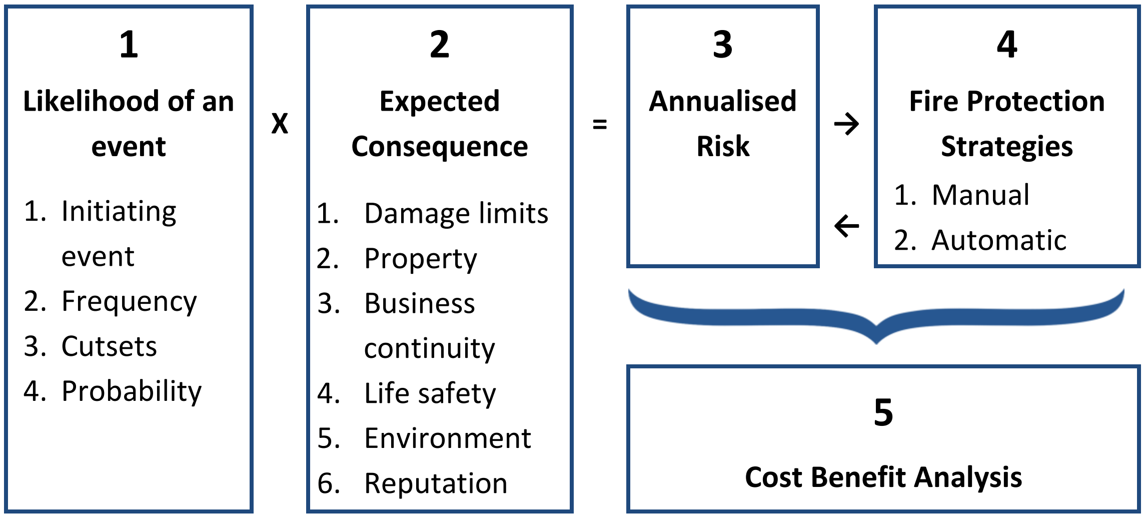
2.2. Scope

To achieve the objectives of this study, six major stages are followed as described below:

1. The determination of the governing fire scenarios in a jetty installation and the corresponding likelihood. To identify the scenarios and for the further discussion, the following parameters are considered:
    - Fuel: LNG and LPG are the focus of this study.
    - Size of the spill: It is directly related to the transfer capacity and the size of the leakage, three spill sizes are studied to indirectly take these factors into account.
    - Type of leak: Given that the associated consequences are different, both continuous and instantaneous releases are considered.
2. The relevant fire scenarios are assessed to determine the expected consequence.
3. Fire protection systems are discussed and their impact to mitigate the negative effects of the fire scenarios evaluated.
4. The expected consequence is compared with the associated cost of the fire protection systems to determine the cost-effectiveness of the strategies.
5. A parametric study is carried out to identify the effect of the different parameters considered in the study on the steps of the methodology. These parameters are the same used to identify the governing scenarios additional to the type of fire and the jetty configuration.
6. A hypothetical case study is carried out to verify and illustrate the practical use of the methodology.

2.3. Methodology

The methodology proposed has five main steps with their corresponding tasks.



The first step is to define the likelihood of an event (Chapter 3). To start, the initiating events must be defined, qualitative and semi quantitative analysis can be used for this step so only useful scenarios are considered. Then, the associated frequency must be determined via statistical data, models or fault tree analysis (Section 3.1). The third task is to define the cutsets and their probability (Section 3.2).

The second step is to calculate the expected consequence in monetary value (Chapter 4). The expected consequence can be caused mainly due to property damage (Section 4.2), business continuity (Section 4.3), life safety losses (Section 4.4), environmental damage (Section 4.5) and reputation effects (Section 4.6). Although these are the main typical aspects to consider, it is also recommended to look into the specifics of the facility to identify additional ones.

The third step is to unify the likelihood together with the expected consequence and calculate the annualised risk (Chapter 5), which is, in few words, the amount of money that would be lost per year with no additional protection measures. It is important to group the consequences according to the corresponding initiating event and the type of consequence, this will allow to identify the scenarios with the highest risk and therefore will facilitate the design of the Fire Protection Strategies.

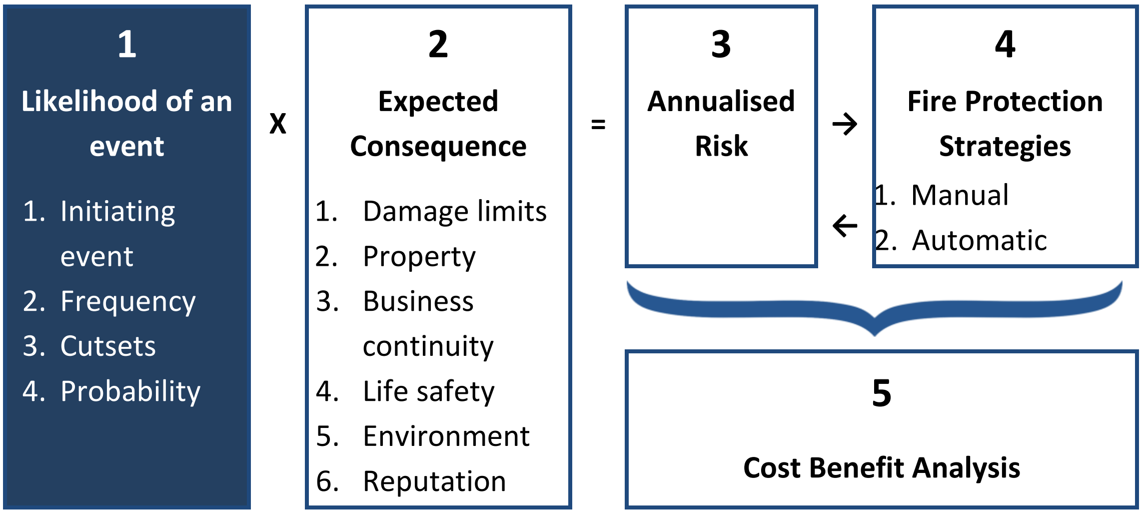
Step 4 gives some general guidelines to design the Fire Protection Strategy (Chapter 6). This design must be made considering the different benefits and disadvantages of the different fire protection mechanisms and also the scenarios with the highest associated risk. This chapter includes a description of the fire and gas detections systems (Section 6.1), a description of the fire extinguishing mechanisms (Section 6.2), a report about typical prices of the solutions (Section 6.3) and a discussion about the efficiency of the systems (Section 6.4).

The cost benefit analysis is outlined in Step 5 (Chapter 6.5). A description of the calculation of the annualised cost of fire protection (Section 7.1) and the recalculation of the expected consequences after the application of the fire protection systems (Section 7.2) are included. If the result of the cost benefit analysis is not satisfactory, the user of the methodology is able to decide whether to reduce the likelihood of an event, reduce the expected consequence or modify the Fire Protection Strategy. In every case, the methodology has to be completely reassessed from the step it was modified.

A parametric study is included in Chapter 8 to evidence the importance on the different steps of the methodology of parameters like transfer means, release size, type of fire and jetty configuration.

Chapter 9 reports a case study where the methodology is used step by step showing its practical use in a realistic environment.

3. Likelihood of an event



The intention of this chapter is to present a methodology to perform an event tree analysis for fire scenarios in jetty installations when transfer operations are carried up. The goal is to define the initiating events and cutsets that should always be included in this kind of analysis and to assign recommended values for frequencies and probabilities

3.1. Initiating events

There are two related steps when defining the initiating events, first, the type of event and its frequency and second, the release size.

3.1.1. Type of initiating event

There are many events leading to a leakage or a rupture of an element in a facility. The range of events that can cause these situations is varied and ranges from merely mechanical impacts to process deviations and severe corrosion problems. If severe enough, these events can cause the partial or total rupture of a process element, leading to a release. The initiating events in this study are the release situations and not their primary cause.

A release can occur in any equipment, piping element or instrument. Each one of the elements has different design intention and characteristics and therefore the probability of release is also different. For this reason the first step in identifying the initiating events is to list the elements in the jetty area studied. With the list of elements the next step is to determine the release events and their corresponding frequency.

In a jetty installation following elements can be found:

- Pipes
- Instruments and valves

- Control Valves
- Pumps
- Compressors
- Loading arms and hoses

There are two approaches when talking about initiating events' frequency. The first one consists of taking generic data and extrapolate it to the project. The advantage of this approach is the high amount of data collected throughout the decades as for example the HCRD by the British Government [19], which reports the generic leak frequencies for typical process equipment and piping elements.

The NFPA 59A 2013 [2] reports, in its new performance based LNG plant siting chapter, the probabilities of failure for elements in an LNG plant, however they only report rupture data, not including the leakage scenarios. Some sources report data specifically for loading arms and hoses and include rupture and leakage events, for example the Flemish Government [20] and the Dutch Government in the Purple Book [21]. However, some studies recommend to use the data for a generic process since the LNG specific values are not sufficiently robust to be used [22].

The second approach is to perform a fault tree analysis specifically for the developed project, there are some examples in the literature for LNG facilities using this approach [23]. Specific fault trees are generated to estimate the frequency of the initiating event from a logic model of the failure mechanism of a system. It is developed using combinations of failures of more basic components and systems [18]. It can be then concluded that in order to know the frequency of the leak it is necessary to have a deep knowledge of the process, its safety systems and its operation philosophy and have the values of their failure frequency.

The advantage of the generic approach is the high reliability of its data and therefore its high recognition. However it is also less representative of the real system, since the data is compiled for typical process. On the other hand, the fault tree analysis approach allows a highly representative result given that all the input data is available and validated. However, its reliability and recognition is low due to the large amount of input data and its corresponding uncertainty.

If possible a fault tree analysis should be carried out, otherwise generic data can be used. The frequencies for release events involving common elements found in a jetty are reported in Table 2 by the Flemish Government [20]. These frequencies consider the instruments typically installed in the equipment or pipe section.

Table 2. Frequencies for initiating events [20]

Element	Rupture	Leakage	Units	Remarks
Loading arm	$3.0*10^{-8}$	$3.0*10^{-7}$	$(\text{h of use})^{-1}$	d_{eq} = 0.1 D (max 50mm)
Hose for LPG	$5.4*10^{-7}$	$5.4*10^{-6}$	$(\text{h of use})^{-1}$	d_{eq} = 0.1 D (max 50mm)
Above ground pipeline	$2.2*10^{-8}*L/D$	$2.8*10^{-7}*L/D$	yr^{-1}	d_{eq} = 0.1 D
	$2.2*10^{-8}*L/D$	$1.2*10^{-7}*L/D$	yr^{-1}	d_{eq} = 0.15 D
	$2.2*10^{-8}*L/D$	$5.0*10^{-8}*L/D$	yr^{-1}	d_{eq} = 0.36 D
Underground pipeline	$2.8*10^{-8}$	$7.9*10^{-8}*L$	yr^{-1}	Crack, d_{eq} = 10 mm
	$2.8*10^{-8}$	$6.9*10^{-8}*L$	yr^{-1}	Hole, d_{eq} = 0.5 D
Pumps with gaskets	-	$4.4*10^{-3}$	yr^{-1}	$d_{eq} = 0.1\ D_{Max}$
Pumps without gaskets	-	$1.0*10^{-4}$	yr^{-1}	$d_{eq} = 0.1\ D_{Max}$
Reciprocating pump or compressor	$1.0*10^{-4}$	$4.4*10^{-3}$	yr^{-1}	$d_{eq} = 0.1\ D_{Max}$

Both for LNG and LPG, the values for an LPG hose are used, since an LNG hose corresponds more closely to an LPG hose than to a generic liquid fuel hose.

If equipment or elements other than the ones listed in Table 2 are installed on the jetty, the designer must include them in the list of the initiating events with the corresponding frequency.

3.1.2. Release size

The release size is part of the definition of the initiating event, and therefore must be calculated in this step.

Continuous leaks

For the continuous leaks, the release flow rate can be calculated using the method described in the Yellow Book [24] and reported in Equation 2.

Equation 2. Release flow rate in orifices

$$q_S = C_D \times A_h \times \sqrt{(2(P - P_a) \times \rho_L)}$$

Where:

C_D: Discharge coefficient [-]

A_h: Cross-sectional area of the hole, crack or rupture [m²]

P: Total pressure at opening [N/m²]

P_a: Atmospheric pressure [N/m²]

ρ_L: Liquid density [kg/m³]

Instantaneous leaks

The instantaneous leaks are normally estimated as the maximum capacity of the tank, vessel or as the volume of the pipe contained between two Emergency Shut Down valves.

3.2. Cutsets

Before the release event evolves into an accident, there are many possibilities called cutsets, which have an associated probability. There are only three possible cutsets whit hazardous consequences when the initiating events are the continuous or instantaneous leak situations. These are immediate ignition, delayed ignition and explosion. The event trees for all of the initiating events are similar and can be seen further in the document in Figure 11. The probabilities associated to the cutsets will depend on the fuel and the size of the release as it is discussed in this section for each cutset.

The combination of the cutsets will determine the type of consequence and therefore the severity of the scenario.

3.2.1. Ignition

In the case of a release, ignition is the final element necessary to have an accident as the fuel is already released in an oxidizing atmosphere, and will generate a flammable mixture unless the evaporation rate is highly reduced, which is not likely in the early stage of the spill. That means that there will always be a time and place in the event when a flammable mixture is present. For flammable liquids ignition probabilities are normally in the order of 0.1 to 0.2, however, in the case of LPG and LNG this is different given their gaseous nature in atmospheric conditions, making them more probable to ignite.

Ignition can be caused by several sources, but not all of them are likely to be present in this specific case. According to the Health and Safety Executive (HSE) [25], the main ones for the bulk transfer of dangerous liquids and gases are:

- Naked flames, including welding and cutting equipment
- Smoking
- Electrical lightning, power circuits and equipment that are not explosion protected
- Processes or vehicles that can cause friction or the generation of sparks
- Radio frequency emissions
- Hot surfaces
- Static electricity

There are three options for the ignition, it might happen either immediately, delayed or not happen at all. Evidently these three options lead to different consequences as can be seen in Figure 11. It is important to state that the term delayed ignition considers both a time and a space delay that are linked through the dispersion behaviour and can thus be modelled.

The ignition probability has been largely researched and documented in several studies, the values depend mainly on three variables:

- The flow rate or amount released
- The substance released
- The surrounding area configuration, which includes the potential sources of ignition in the area [25].

The combination of these three variables gives as a result countless possibilities and therefore the ignition probability ranges from values close to zero to one. Several studies have addressed this matter, giving guidance to face this uncertainty [26].

Some reports are worth to report because of their relevance and completeness:

- The direct ignition probabilities suggested in the Purple Book [21] are reported in Table 3. They group the released substances in 3 groups to consider the different levels of flammability and the size of the spill. For the delayed ignition probabilities, they suggest to use a simple model described by Equation 3. It takes into account the kind of ignition source and the time since the release but not the type of fuel.

Table 3. Probability of direct ignition for stationary installations [21].

Source		**Substance**		
Continuous [kg/s]	**Instantaneous [kg]**	**K1-Liquid***	**Gas, low reactive****	**Gas, average /high reactive****
< 10	< 1000	0.065	0.02	0.2
10 – 100	1000 – 10000	0.065	0.04	0.5
> 100	> 10000	0.065	0.09	0.7

* K1-Liquid: Flammable liquid having a flash point less than 21°C and a vapour pressure at 50°C less than 1.35 bar. (Dutch classification)

**Low reactive gas: Methane and other gases.

*** Average/high reactive gas: Ethane, propane, butane and other gases.

Equation 3. Probability of delayed ignition caused by an ignition source.

$$P(t) = P_{present}(1 - e^{-\omega t})$$

Where:
$P(t)$: Probability of an ignition in the time interval 0 to t.
$P_{present}$: Probability that the source is present when the cloud passes.
ω: Ignition effectiveness.

- The Flemish Government reports the values in Table 4 [20], as shown, they also divide the gas substances (Group 0) in high and low reactive and they split the liquid substances in Groups 1, 2 and 3 depending on their flash point. They also report fixed values both for delayed ignition and explosion, which means they do not consider the specific configuration of the surroundings in their analysis.

Table 4. Probability of direct and delayed ignition [20]

Source		Direct (P_D)or Delayed (P_V)	Group				
Continuous [kg/s]	Instantaneous [kg]		0		1	2	3
			High react	Low react			
< 10	< 1000	P_D	0.2	0.02	0.065	0.02	0.006
		P_V	0.06	0.02	0.07	-	-
10 – 100	1000 – 10000	P_D	0.5	0.04	0.065	0.02	0.006
		P_V	0.2	0.04	0.07	-	-
> 100	> 10000	P_D	0.7	0.09	0.065	0.02	0.006
		P_V	0.7	0.1	0.07	-	-

- Daycock and Rew [25] designed a model specific to LPG for on-site ignition based on the land-use type and the distribution of ignition sources. This model also includes a factor that considers the control of the ignition sources. It is important to mention that this study was funded by the HSE and constitutes one of the most robust models to estimate ignition probability existing nowadays.
- Ronza et al [27] report a summary of ignition probability estimates, the complete set of data can be consulted in the source, Table 5 reports the most relevant ones for this study, that are the ones involving jetty installations or ignition of LPG or LNG. As it can be seen, there are several options with different scope of application and therefore different probability values. The different sets of data report the immediate ignition (P_D), the delayed ignition ($(1-P_D) \times P_V$) and/or the overall ignition probability ($P_D + (1-P_D) \times P_V$).

Table 5. Direct and delayed probability [27].

Scope	Probability data	Source
LPG, no obvious point of ignition with explosion proof equipment	Massive release: $P_D + (1-P_D) \times P_V = 0.1$	[28]
LNG and LPG	Area covered by cloud < 30m²: $P_D + (1-P_D) \times P_V = 0.5223$ 1000 < Area covered < 3000 m² $P_D + (1-P_D) \times P_V = 0.8864$ 3 x 10^6 m² < Area covered < 10^7 m² $P_D + (1-P_D) \times P_V = 0.9992$	[29]

<table>
<tr><th>Scope</th><th>Probability data</th><th>Source</th></tr>
<tr><td>LNG vapour clouds</td><td>Limited release:
$P_D + (1-P_D) \times P_V = 0.1$
Large release:
$P_D + (1-P_D) \times P_V = 1$</td><td>[30]</td></tr>
<tr><td>Ignition at jetty</td><td>After fire or explosion:
$P_D = 0.6$
$(1-P_D) \times P_V = 0.33$
After collision:
$P_D = 0.3$
$(1-P_D) \times P_V = 0.33$</td><td>[31]</td></tr>
<tr><td>LPG, ignition at source</td><td>Large instantaneous release:
$P_D = 0.25$
$(1-P_D) \times P_V = 0.25$ with wind
$(1-P_D) \times P_V = 0.25$ without wind
1000 t:
$P_D = 0.25$
$(1-P_D) \times P_V = 0.25$ with wind
$(1-P_D) \times P_V = 0.1$ without wind
250 kg/s, 50 kg/s:
$P_D = 0.25$
$(1-P_D) \times P_V = 0.25$ with wi
nd
$(1-P_D) \times P_V = 0.1$ without wind
30 kg/s, 16kg/s:
$P_D = 0.15$
$(1-P_D) \times P_V = 0.15$ with wind
$(1-P_D) \times P_V = 0.05$ without wind</td><td>[32]</td></tr>
<tr><td>LPG, vehicle accidents</td><td>$P_D + (1-P_D) \times P_V = 0.24$</td><td>[33]</td></tr>
<tr><td>LPG releases (200 t), industrial area (off-site).

Plant / pipework failure</td><td>$P_D = 0.5$
$(1-P_D) \times P_V =$
<table>
<tr><th rowspan="2">Hole size</th><th colspan="3">Density of sources</th></tr>
<tr><th>Low</th><th>Med.</th><th>High</th></tr>
<tr><td>13 mm</td><td>0.04</td><td>0.14</td><td>0.24</td></tr>
<tr><td>25 mm</td><td>0.05</td><td>0.25</td><td>0.45</td></tr>
<tr><td>50 mm</td><td>0.4</td><td>0.6</td><td>0.8</td></tr>
</table></td><td>[34]</td></tr>
<tr><td>LPG, rail accidents</td><td>Small spills:
$P_D = 0.1$
$(1-P_D) \times P_V = 0$
Large spills:
$P_D = 0.2$
$(1-P_D) \times P_V = 0.5$</td><td>[35]</td></tr>
</table>

<table>
<tr><th>Scope</th><th>Probability data</th><th>Source</th></tr>
<tr><td>LPG, lorry and rail accidents</td><td>P_D =<table><tr><td>Spill size</td><td colspan="2">Type of accident</td></tr><tr><td></td><td>Lorry</td><td>Rail</td></tr><tr><td>Small</td><td>0.25</td><td>0.30</td></tr><tr><td>Large</td><td>0.75</td><td>0.90</td></tr></table></td><td>[36]</td></tr>
</table>

For the ignition probability it can be concluded that the varied range of circumstances leads to a wide range of probability values. Therefore it is recommended either to carry out a specific study or search for the data extensively in the pertinent literature so the closest scenario to reality is taken. The data provided in the Purple Book [21] seems to be reasonable as it does not neglect any of the factors affecting the ignition probability. This data is used in this study.

Although the Flemish Government [20] does not consider the type of installation, they make the difference among fuels with different degrees of reactivity. Equation 3 can be used to calculate the value for different times. The distance from the source point and the time in the equation are linked through the release behaviour. If a plant layout is available, a dispersion simulation can be carried out in order to know the time when the cloud will reach the potential ignition sources (ovens, flares, rotary equipment).

As LNG contains mostly methane it is classified as a low reactive gas. However, if LNG is composed of large amounts of ethane, it has to be considered as a high reactive gas. LPG is classified as a high reactive gas.

For a continuous release, the flow rate for a leakage is lower compared to a rupture, therefore following the classification used in the Purple Book [21], only low and medium releases are considered for the first case and medium and high releases for the second.

3.2.2. Explosion probability

If delayed ignition occurs, there are two possibilities, either that the vapour cloud is confined or not. In the first case the consequence is a mild explosion with built-up pressure and heat radiation, otherwise the ignition will cause a short flash fire turning quickly into a jet or pool fire causing only heat radiation.

- The Flemish Government [20] reports the explosion probability similar to the ignition probabilities, as shown in Table 19. However they completely disregard the location, which is an important factor to consider since confinement is required for an explosion.

Table 6. Probability of explosion according to the Flemish Government [20]

Source	Group

Continuous [kg/s]	Instantaneous [kg]	0		1
		High react	Low react	
< 10	< 1000	0.2	0.2	0.2
10 – 100	1000 – 10000	0.3	0.3	0.2
> 100	> 10000	0.4	0.4	0.2

- The Purple Book [21] reports two sources, DNV [37] and TNO [38]. The DNV source recommends to take a distribution of 60% for the flash fire cases and 40% for the explosion, while TNO recommends 70% and 30%. Again, no consideration is taken to include the surroundings configuration.
- Regarding the explosion probability, Moosemiller [39] developed a calculation represented in Equation 4. The explosion probability is a function of the release flow rate and therefore it can only be calculated for a continuous release event. The study suggests to multiply the outcome of the equation by 0.3, 1, or 3 for low, medium and high reactive fuels respectively. Given that the direct and delayed ignition are taken from the Purple Book [21], three flow rates matching the three ranges are chosen to calculate the P_E. The results are reported in Table 7, LNG is considered as medium reactive and LPG as high reactive following the same reasoning exposed previously. An explosion probability of 1 can be seen in the table, which is not reasonable.

Equation 4. Probability of explosion.

$$P_E = 0.024 \times {q_S}^{0.435}$$

Where:
q_S: Release flow rate [lb/s]

Table 7. Probability of explosion calculated according to Moosemiller [39]

Continuous release range [kg/s]		q_S [kg/s]	q_S [lb/s]	Fuel	Reactivity factor	P_E
Small	<10	5	11	LNG	1	0.07
Medium	10 – 100	50	110	LNG	1	0.19
Large	> 100	200	441	LNG	1	0.34
Small	<10	5	11	LPG	3	0.20
Medium	10 – 100	50	110	LPG	3	0.56
Large	> 100	200	441	LPG	3	1.00

It is important to highlight that the confinement probability depends largely on the plant layout and in the configuration of the equipment, also that in an open space as a jetty, it is largely unlikely to have those conditions. As reported by the Flemish Government [20] and the Purple Book [21] the values vary from 20% to 40%. On the other hand, Moosemiller [39] reports values that range from almost 0 to 100%. Values close to 100% are not reasonable because it means every delayed ignition would happen in a confined space and generate pressure waves.

The values reported by the Flemish Government [20] are recommended for this methodology, as it takes into account both the size of the spill and the reactivity of the fuel with reasonable values.

3.3. Event tree

With the initiating events, their frequencies, the cutsets and the corresponding probabilities, it is possible now to identify the consequences and complete the event tree.

If direct ignition occurs, it can cause a jet fire in the case of a continuous release or a pool fire in an instantaneous one.

If the ignition is delayed, there are two possibilities. That it happens in a confined space or not. In the first case, the ignition would cause a deflagration with pressure build-up; otherwise the deflagration occurs without this phenomena and is called flash fire instead. In both cases the incident can lead to a jet fire if the continuous release is still ongoing, or a pool fire if there is still enough fuel on the ground.

If a release of LPG is never ignited, the vapour cloud will disperse and the only risk is toxicity, which is not a high risk for this gas. For LNG additionally, the cryogenic effect must be considered when regarding life safety since it can cause burns and asphyxia to people in the surroundings.

For each scenario there is an event tree as the one depicted in Figure 11. The frequencies and probabilities (F_I, P_D, P_V and P_E) are then taken from Table 2, Table 4 and Table 6 and the probable consequences from Table 8. It is important to mention that the actual consequences will only be known when the scenarios are modelled.

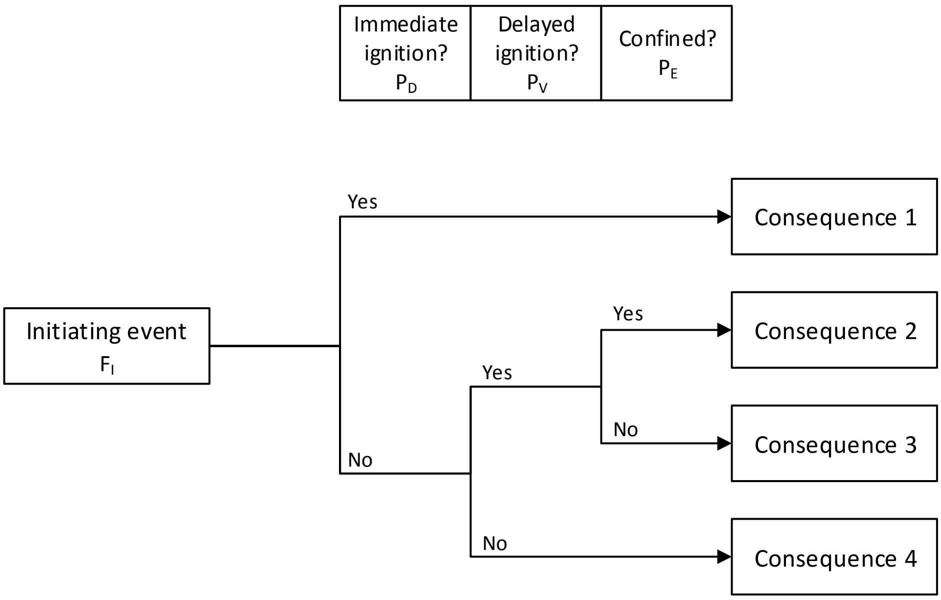


Figure 11. General event tree

For each consequence of every scenario, frequencies are calculated following the event tree in Figure 11 as follows:

Equation 5. Frequency for consequence 1

$$F_{C1} = P_I \times P_D$$

Equation 6. Frequency for consequence 2

$$F_{C2} = F_I \times (1 - P_D) \times P_V \times P_E$$

Equation 7. Frequency for consequence 3

$$F_{C3} = F_I \times (1 - P_D) \times P_V \times (1 - P_E)$$

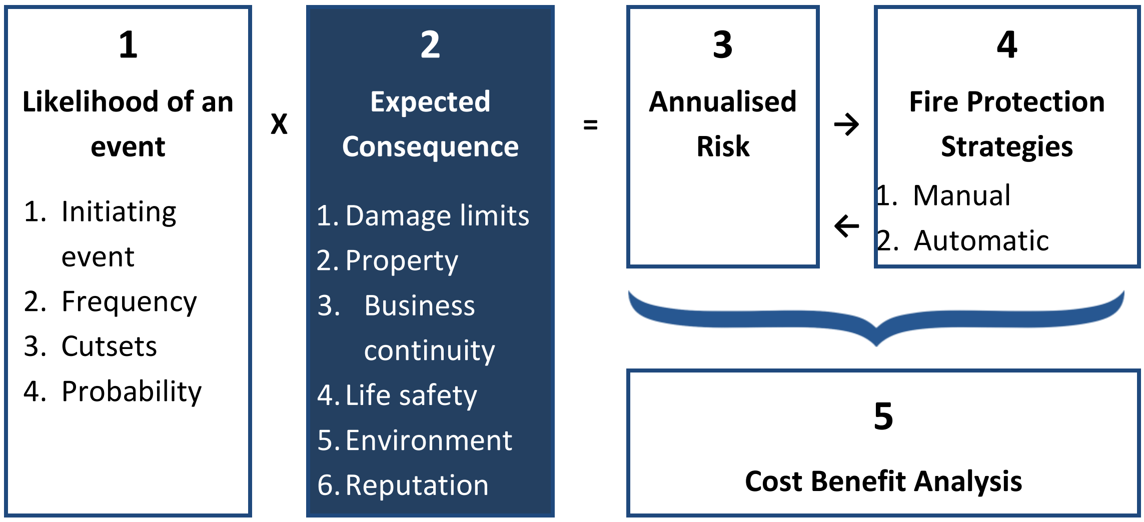
Equation 8. Frequency for consequence 4

$$F_{C4} = F_I \times (1 - P_D) \times (1 - P_V)$$

Table 8. Consequences for the different scenarios.

Consequence 1	**Consequence 2**	**Consequence 3**	**Consequence 4**
Jet fire / Pool fire	Vapour cloud explosion / Jet fire / Pool fire	Flash fire / Jet fire / Pool fire	Toxic / Cryogenic effects

4. Expected consequence



As defined earlier in Equation 1, risk is equal to frequency or likelihood multiplied by the expected consequence. Up to now the methodology to determine the likelihood has been described. The following step is to determine the expected consequence

The expected consequence is a broad concept since the occurrence of an accident can affect different kind of assets for a company. This list is always incomplete but five large groups can be identified; property damage, business continuity, safety, environmental and reputation.

The quantification of the consequences for each scenario must lead to a monetary value that can be used later to calculate the annualised risk. Equation 9 presents the way to calculate the total monetary consequences for each one of the branches of the event tree.

Equation 9. Monetary consequence for each consequence scenario

$$C = C_{Property} + C_{Business\ Cont.} + C_{Life\ Safety} + C_{Environment} + C_{Reputation}$$

The property loss is the focus of this chapter, however general guidelines about the estimation of expected consequences for business continuity, life safety, environment and reputation are included.

It is important to re assess the expected consequence after the fire protection systems are designed so it is possible to recalculate the annualised risk with the protection measures and perform the cost benefit analysis

4.1. Damage limits

The first step to quantify the expected consequence is to calculate the damage limits. The damage limits are the values of defined parameters at which the assets of a company are affected by the consequences of an event.

In the case of a release the damage limits are normally related to toxicology and pollution. In the case of fire and explosion, the parameters are heat radiation, high and low temperatures and overpressure waves. These parameters of an accident affect in different ways the assets of a company.

In this section, damage limits in case of fire or explosion for property (equipment and structures) are presented. The damage limits for life safety, business continuity, environment and reputation can be discussed in a posterior study.

4.1.1. Equipment

In the case of fire equipment can be affected mainly by two factors, heat radiation and extreme temperatures. Barry [14] reports typical threshold damage values for people and common elements in industrial facilities, the effect of the heat flux can be seen in Table 9, while the effect of high temperatures is reported in Table 10. The incident heat flux is based on a 10 minutes exposure.

Table 9. Thermal radiation damage levels for typical elements in a facility [14]

Incident Heat Flux $[kW/m^2]$	Damage to Equipment	Damage to people
35.0 – 37.5	Damage to process equipment.	100% Lethality in 1 min 1% lethality in 10 s
25.0	Minimum energy to ignite wood. Indefinitely long exposure without a flame.	100% lethality in 1 min Significant injury in 10 s
18.0 – 20.0	Plastic cables insulation degrades.	
12.5 – 15.0	Minimum energy to ignite wood with a flame; melt plastic tubing.	1% lethality in 1 min 1st degree burns in 10 s
4.0		Causes pain if duration is longer than 20 s but blistering is unlikely.
1.6		Causes no discomfort for long exposure.

Barry [14] also reports the threshold damage temperature limits for common materials found in a facility, see Table 10.

Table 10. Effect of temperature on common materials [14][40]

Material	Damage conditions	Temperature (°C)
Polystyrene	Collapse Softens Melts and flows	120 120 – 140 150 – 180
Polyethylene	Shrivels Softens and melts Ignites	120 120 – 140 340
PVC	Degrades Fumes Browns Ignites Chars	100 150 200 390 400 – 500
Solder	Melts	250
Lead	Melts, drop formation	300 – 350
Aluminium and alloys	Softens Melts Drop formation	400 600 650
Copper	Melts	100 – 1,100
Cast iron	Melts Drop formation	1,100 – 1,200 1,150 – 1,250
Zinc	Drop formation Melts and flows	400 420
Paint	Deteriorates Destroyed	100 250
Wood	Ignites	240

In the case of an explosion the overpressure can heavily damage the equipment, as reported in Table 11.

Table 11. Damage caused by overpressure [14][40]

Overpressure (mbar)	Characteristic Damage	
	To equipment	To people
170 – 340	Heavy damage to buildings and to process equipment	1% death from lung damage >50% eardrum rupture >50% serious wounds from flying objects
70 – 170	Repairable damage to building and damage to the façades of dwellings	1% eardrum rupture 1% serious wounds from flying objects
35 – 70	Glass damage	Slight injury from flying glass
10 – 20	Glass damage to about 10% of panes	

4.1.2. Structural damage

Additional to the equipment, the structures present in a facility also suffer from the effects of fires and explosions. The main concern in case of fire is the exposure to high temperature, which weakens the typical steel framed structures in industrial plants. Typical damage limits are reported in Table 12 for common structural elements in industrial facilities.

Table 12. Threshold damage limits for common structural elements [41]

Element	Maximum single temperature (°F/°C)	Arithmetic Mean Temperature (°F/°C)
Steel columns (no load)	1,200/650	1,000/538
Steel beams	1,300/705	1,100/594
Alternate (no load)	1,200/650	1,000/538
Reinforcing steel	1,100/594	1,100/594
Prestressed steel	800/427	800/427
Floor-roof slabs	1,300/705	1,100/594

Overpressure product of an explosion also affects the structures. The effect on a typical steel frame structure is illustrated in Table 13.

Table 13. Overpressure effect on a steel frame siding pre-engineered building [14]

Peak side-on overpressure (mbar)	Consequences	Probability of serious injury or fatality
100	Sheeting ripped off and internal walls damaged. Danger from falling objects.	0.4
170	Building frame stands, but cladding and internal walls destroyed as frame distorts.	0.4
340	Total destruction.	1.0

Each facility is different and the threshold damage limits should therefore be different for each case. If critical equipment, instrumentation and/or structure detail is installed it is recommended to carefully assess together with the provider the specific reaction to heat radiation and high and low temperatures.

4.2. Property

With the damage limits defined, it is possible to calculate the actual effect of the fire and explosion scenarios in a facility.

PHAST 7.1® is a software developed by Det Norske Veritas (DNV) that allows the analysis of flammable, fire, explosion and toxic hazards, and therefore it is used to model the severity of such accidents.

To be able to model the magnitude of these accidents, specific conditions of the accident are necessary: the spill size (previously calculated), the surroundings layout and the wind characteristics among others depending on the calculation method used for this purpose.

The intention is to calculate the estimated monetary consequence in property for branch of the event tree ($C_{Property}$).

After performing the modelling of the fire accident for the different scenarios considered in the project, it is necessary to identify the equipment and structures affected and quantify the level of damage.

After the occurrence of an event, the affected equipment and instrumentation must be either repaired or replaced. Typical elements, like small valves and instruments might be available in stock so the replacement is cheaper and considerably faster. Also some spare internal elements for equipment may be available facilitating the repair.

In some other cases, like large and/or customized valves, instruments and equipment, the repair is not straight forward and might include the requisition to the original provider. As well as the cost of the item, the delivery time plays an important role in the aftermath. Additional costs that must be estimated include among others:

- Decommissioning of the damaged equipment
- Construction works
- Transportation
- Installation

Estimated prices and delivery times of typical instruments and equipment are reported in Table 14. The delivery times are accounted since the requisition is formally made to the factory. The prices are from 2012 adjusted to 2015 using a correction factor of 2.0% per year. The delivery times for major equipment is typically between 8 and 12 months. Construction of critical equipment, like LPG spheres or LNG cryogenic storage tanks can last up to 2 years.

Table 14. Estimated prices for typical equipment [42].

Item	**Prices (€)**
Customized control valve 10”	50,000
Loading arm 10”, completely installed	450,000 – 550,000
Loading hose 8”, per meter	7,000
LPG Pump, 150 m^3/h	80,000

Item	Prices (€)
LNG Pump skid, 150 m³/h	500,000
LPG Compressor, 100 m³/h	90,000
LNG Boil Off Gas Compressor, 100 m³/h @ 50 bar	50,000 – 100,000
LNG Atmospheric Vaporizer, 2,000 m³/h	15,000
Hydrocarbon storage tank, 4000 m³	150,000

When the list of equipment in the facility is ready, it is time to generate the radiation and overpressure contours for the fire and explosion scenarios respectively. From Table 9 and Table 11, 35 kW/m² and 12.5 kW/m² and 0.21 bar and 0.14 bar are the radiation and overpressure contours recommended to estimate the expected consequence for equipment. From Table 13, 0.21 bar and 0.14 bar are the overpressure contours recommended to estimate the expected consequence for structures.

The recommended contours must be contrasted with the plant layout to identify the equipment and affected structures and the corresponding degree of damage. As the expected consequence is different depending on the level of damage, a factor (K) is introduced. The values for K are defined based on the damage limits in Section 4.1 and listed in Table 15. For the 35 kW/m² and 0.21 bar contours the factor K is 1.2 as it has to consider the price of the installed new equipment plus the demolishing and dismantling process of the previous equipment. For the 12.5 kW/m² and 0.14 bar contours the K factor is lower given that the equipment can be repaired. Also, for the 12.5 kW/m² contour, the fire brigade intervention can lower the expected consequence.

Equation 10 illustrates the monetary consequence ($C_{Property}$) for the fire scenario "i" because of the damage of the elements "j". P_j is the price of the damaged element, it should consider all of the elements mentioned above, price of the element, construction works, and transportation, among others.

Equation 10. Monetary consequence for fire scenario i.

$$C_{Property,i} = \sum_j K_j \times P_j$$

Table 15. Level of damage factor (K)

Heat radiation level		Explosion	
12.5 kW/m²	35 kW/m²	0.14 bar	0.21 bar
0.1	1.2	0.5	1.2

4.3. Business continuity

A major concern of the decision makers is the business continuity. If the level of damage is negligible the facility is able to operate while performing the minor repair works on the equipment affected. For larger consequences, the damage level can be higher forcing to stop the facility if one of the essential systems is not available, this includes also the utility systems. If the extent of the accident reaches the boundaries of the facility or beyond, the business continuity can be compromised.

The business continuity should include:

- The business interruption or loss in production
- Cost of the alternative supply to provide the customers
- Commercial penalties if no alternative supply to the customer is provided.

An accident of large proportions can lead to situations in which it is not viable to rebuild and continue the operation. The decision makers can decide to declare bankruptcy and stop the operation of the facility. A total business interruption should be considered if the extent of the damage compromises the major part of the facility. Special attention should be given to reduce the risk of these scenarios.

The intention is to calculate the estimated monetary consequence in business continuity for each branch of the event tree ($C_{Business\ Cont.}$).

4.4. Life safety

Life safety is one of the most important assets for a company. This asset is also the most worrying of the assets in a company, since it directly impacts the people, therefore affecting the finances, reputation and if the consequences are large enough, it can also affect the continuation of the business.

The risk tolerance regarding life safety are different for employees of the facility than for the public population. It is considered that the first group of people voluntarily accepts a higher risk than the second one given that they participate in the production activities of the facility. Although this concept is a constant matter of debate, both industry and government make this distinction and report different values for each group of people as shown in Table 16 [14]. The maximum tolerable risk per year must be understood as the maximum number of fatalities that is acceptable in a facility.

The individual risk must then always be calculated for a facility so it accomplishes the maximum tolerable risk defined by the corresponding authority.

Table 16. Individual risk criteria for workers and the public [14]

Authority and scope	Workers or Public	Maximum tolerable risk per year
Health & Safety Executive UK. Existing hazardous industry	Workers	10^{-3}
	Public	10^{-4}
Health & Safety Executive UK. New housing near existing plants	Public	10^{-5}
Shell. Onshore and offshore	Workers	10^{-3}
BP. Onshore and offshore	Workers	10^{-3}
Statoil. Onshore plants	Workers	$8.8*10^{-5}$
VROM, The Netherlands. New plants	Public	10^{-6}
VROM, The Netherlands. Existing plants	Public	10^{-5}
Advisory Committee on Dangerous Substances, UK. Existing dangerous substances transport	Public	10^{-4}
Hong Kong Government. New Plants	Public	10^{-5}
Department of Planning, New South Wales. New plants and housing	Public	10^{-6}

Additional to the individual risk analysis, it is also necessary to quantify the related financial impact. It is a matter of debate and always controversial to assign a value on human lives, but it cannot be disregarded in the expected consequences total and therefore in the design of the risk reduction strategies. Although the population exposed to the risk and the risks themselves are different in every facility, the following costs must be assigned and assessed when forecasting the expected consequence [14]:

- First aid
- Moderate burn injury, with possible hospital treatment
- Severe burn injuries
- Single worker fatality
- Single public fatality
- Several fatalities

Table 9 and Table 11 report the different damage limit values for people in case of fire and explosion respectively. With the monetary value determined, the contours of incident radiation and explosion overpressure and the information about the location of the people it is possible to estimate the expected consequence in financial terms. To stress the value of life safety, the fatalities above the maximum individual tolerable risk can be assigned a higher value.

The intention is to calculate the estimated monetary consequence in life safety for each branch of the event tree ($C_{Life\ Safety}$).

4.5. Environment

Environmental damage is difficult to estimate, however some guidelines can be followed to avoid disregarding aspects in this type of consequence as follows:

1. Consider the three different types of pollution, air, water and soil.
2. Carefully research the different kind of regulation and the respective fine in case of a release.
3. Research the cleaning procedures in case of a release of the materials handled in the facility.
4. Assess the cost of the cleaning procedures for different sizes of releases according to the initiating events.

The environmental damage is highly dependent on the location and conditions of the facility, therefore a careful assessment must be carried out. An environmental impact assessment is normally carried out in the early stage of a new design, this can be used as a starting point for the expected consequence estimation.

The intention is to calculate the estimated monetary consequence in business continuity for each branch of the event tree ($C_{Environment}$).

4.6. Reputation

This is the most difficult aspect to estimate since it involves the abstract concept of reputation. After an accident the reputation of a company is affected and can subsequently have an impact in the future of the company.

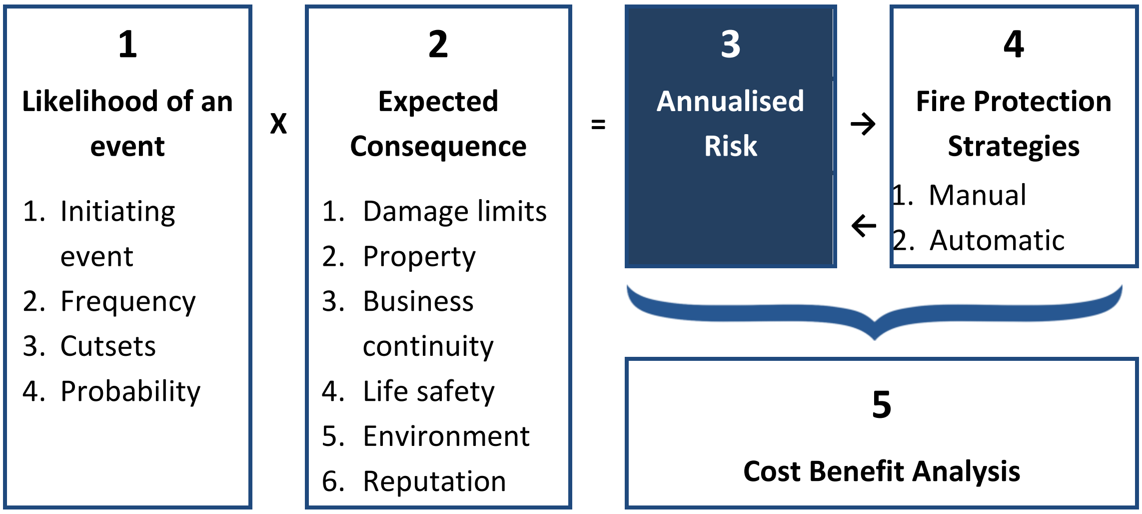
In large scale accidents or disasters, the reputation of the company is highly impacted. Names like Union Carbide or even BP are now associated with industrial disasters in the collective mind set of the public. The two most renowned nuclear disaster, have even led to a reconsideration of the energy policy for the future. Reputation has then a large impact in the activities of a company.

Different classification methods are used to determine the expected damage in reputation. One is to estimate the monetary investment necessary to recover the reputation before the accident. Another can be based on the level of media coverage of the accident. The reputation consequence can also be linked to the property loss and business continuity.

No matter the classification used, different levels of impact must be defined. The respective monetary values can then be assigned to each level in order to quantify the expected consequence because of the impact on the reputation. It is important to remind that the impact in reputation is higher for companies selling to the final customer than for intermediate companies.

The intention is to calculate the estimated monetary consequence in business continuity for each branch of the event tree ($C_{Reputation}$).

5. Annualised risk



The annualised risk as defined in this study is the monetary cost for an enterprise of the negative effects caused by an accident per year.

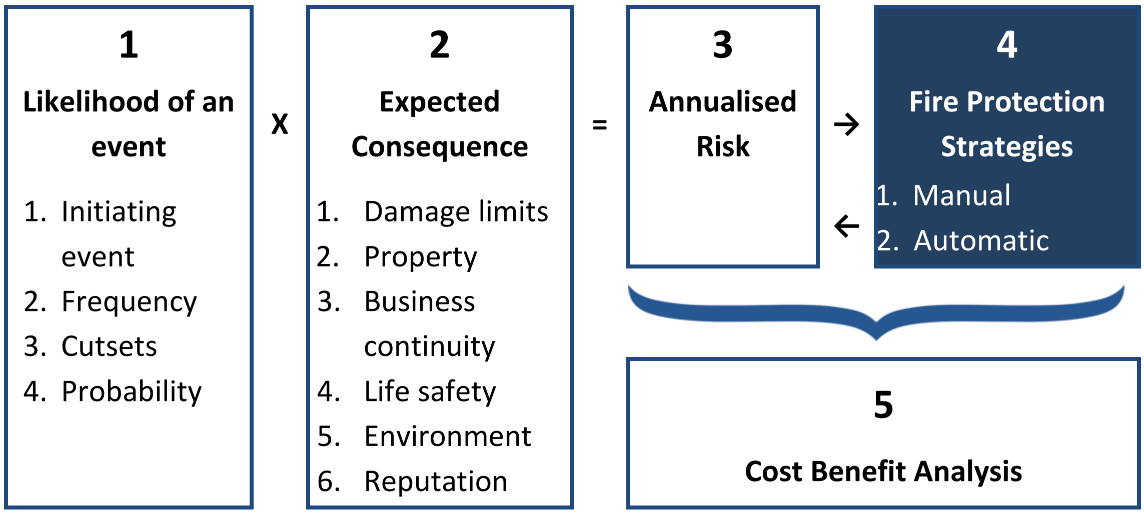
Every monetary consequence estimated for each branch of the event tree is multiplied with the corresponding frequency, all of the results are then added together to obtain the annualised risk value (Equation 11).

Equation 11. Annualised risk for the initiating events considered

$$AR = \sum_{i} F_{Ci} \times C_i$$

It is recommendable to calculate the annualised risk per initiating event and type of consequence, so it is possible to prioritize the efforts and design accordingly to attack the scenario(s) with the highest contribution to the annualised risk.

6. Fire protection strategies



In order to reduce the risk, two approaches can be identified, reducing the likelihood of the initiating event and reducing the expected consequence. These approaches do not exclude each other and are normally applied together to achieve a global risk reduction strategy.

The possible events and risk reduction mechanisms previous to the incident will not be considered and therefore this chapter is dedicated to the post incident risk reduction methods regarding fire protection. However the fire protection strategy must be always regarded as a global solution together with the pre incident risk reduction measures. This safety layers of protection corresponds to the mitigation, as previously discussed in Section 1.5.1, see Figure 9.

The cost benefit analysis and the fire protection strategy are closely related and must be calculated taking each other into account. If the cost benefit analysis is positive, the fire protection strategy can be improved depending on the decision of the designer. If the cost benefit analysis is negative, the fire protection strategy can be modified to more rudimental methods that do not incur in over costs. Always remember the legal and corporative requirements that can overrule the methodology, like the individual risk analysis.

There are mainly two approaches to design the fire protection strategy, manual and automatic intervention. These approaches rely on two complementary fire protection systems, the detection and the fire extinguishment. The factors to define the approach to design fire protection strategy are discussed.

A short description of the different mechanisms of protection is provided additional to the estimated price for different capacities and the overall fire protection system efficiency, necessary further on in the cost benefit analysis step.

6.1. Fire and gas detection systems

The purpose of this systems is to identify the presence of fire or gas to start the mitigation procedures either automatically or manually.

There is a wide range of fire and gas detection systems, from very simple to very sophisticated. The detection systems are external to the process and are able to identify deviations that cannot be detected by the process instruments.

In order to identify a dangerous situation, different deviations can be aimed to be detected like gas, smoke, heat or the flames themselves.

6.1.1. Gas detection

The purpose of this type of system is to detect the combustible gases before ignition. There are mainly two types of gas detectors, the catalytic and the infrared detectors. The first ones are normally used in punctual detection while the second ones are widely used for path detection.

The punctual detection is preferably used close to critical joints like large pressurized flanges, engines, compressors, and any place where a release is more likely to happen or less likely to disperse. The catalytic detectors get poisoned with time by air pollutants, therefore it is necessary to periodically change the catalyst to ensure the detection.

The open path detectors are normally used in open space to cover large areas, the path must be clear of solid objects that can block the detection.

Each detector must be designed and configured to detect a certain type of gas. It is then recommended to configure the system to detect the most dangerous gas present in the vicinity of the risk taking into account the flammability, the volume percentage in the mixture, the ignition temperature, the vapour density and the minimum ignition energy. If two gases are found to be equally and highly dangerous in a facility, it should be considered to install an additional set of detectors.

Both of the detection methods are highly dependent on the location of the instruments. An air path can favour or avoid the flow of the gas and therefore can allow or not its detection. Special attention to places were gas can accumulate.

Additional to the detectors, the systems includes the cabling to the control room and the detection control panel. Depending on the supplier, it can be set on the same detection panel as fire detection.

6.1.2. Heat detection

This kind of method detects the heat emitted by the fire mainly in the form of radiation and convection, which increases significantly the reaction time. If the detection time is crucial, these kind of detectors must be used together with other technology. However its reliability is higher than any other technology, which reduces the occurrence of false alarms.

There are two common types, the fixed temperature and the temperature increment detectors. The first ones are activated when the detector is heated to a certain temperature set point, while the second ones get activated when the rate at which the temperature rises reaches the predefined set point.

Different technologies are used for heat detection, like fusible materials, change in electric currents, heat loads in bimetallic strings, destruction of the device by thermal degradation and thermocouples. A clear advantage of this detector is that it can be fabricated in string shape, which allows to detect over an open path.

6.1.3. Flame detection

The flame detectors sense the presence of light coming from flames usually in the ultraviolet spectrum, the infrared spectrum or a combination of both. A tolerance time in the detection is normally configured to disregard possible false alarms. These are the most expensive detectors in the market.

6.1.4. Human supervision

Although the trend is towards automated detection systems, human supervision still plays an important role in the industry.

Gas detection can be performed using portable gas detectors around the facility. To accomplish the fire detection tasks, the facility has a circuit of cameras pointed to the critical points to visually detect the fire (CCTV). The operator in the control room can then activate the mitigation procedures to avoid the escalation of the incident.

When the operator directly identifies the fire there are mainly two alternatives, to press an emergency push button to start a safety shut down procedure and to report via portable radio to the control room.

This detection method is the cheapest method of all but also the less reliable.

6.2. Fire extinguishing system

When the fire is detected, an automatic or a manual signal can activate a fire extinguishing system. In general the fire extinguishing systems requires of different elements to function as follows:

- Fire water source
- Fire pumps
- Fire piping network
- Flow control and isolation valves
- Extinguishing and protection mechanisms

Additional needs may arise in case of specific technologies.

6.2.1. Fire water source

Depending on the location of the facility and the maximum flow rate required by the fire extinguishing system the water can be taken from the public network, from a fresh water body like a river or a lake, the sea or an underground source. The water treatment depend both on the source of the water and on the requirements of the fire extinguishing system. The quality of the water is different if fed to a monitor or to a foam mixer.

The fire water with the desired quality must be then stored so it is always available in case of need.

6.2.2. Fire pumps

Fire pumps must be sized to feed the fire network in the worst case scenario, which is the one with the highest pressure and flow requirement. It is a recommended practice to install at least two pumps with different power sources must be installed.

6.2.3. Fire piping network

The cost of the piping network is totally dependent on the plant layout and distribution. It is designed depending on the hydraulics of the system and should be able to deliver 150% of the design flow rate.

The fire system control valves should be located within reach of manual activation from a fire but also out of the hazard area.

6.2.4. Extinguishing and protection mechanisms

In the case of a fire there are mainly two options to intervene, one is to attack it and try to suppress it, the second one is prevent the spread of the fire isolating it and cooling down the adjacent equipment and installations. These strategies do not exclude each other.

There are lots of different equipment for fire extinguishing and protection procedures, however only the ones used for LNG and LPG are mentioned in this section:

Water deluge system

A deluge system is a dry nozzle system that in case of detection of a fire opens a deluge valve feeding water to all of the nozzles in the area where the fire is. The actuation of the valve is typically automatic after the detection signal is sent to the fire panel and from there an opening signal is sent to the deluge valve [13].

It is used in situations when the fire spread is a big concern so water is applied in the whole area of the fire to cool it.

Water curtains

Water curtains are normally used to disperse the vapours and avoid the occurrence of a fire or an explosion. This can be achieved through two mechanisms, dispersion and isolation. Dispersion dilutes the vapours to remove the presence of a flammable mixture, the arrangement is normally upwards so the water generates an upward flow that disperses the fuel vapours. This is especially useful for high density vapours. The second mechanism consists on separating areas by the use of the water curtains to avoid the contact of the fuel with potential ignition sources.

On jetties, they are regularly required by law. Here they can serve to cool the equipment, to protect the gangways, and to provide a water curtain between ship and shore.

This mechanism is typically activated after confirmed presence of gas in the area.

Monitors

Monitors are manually operated fire suppression devices.

Monitors are normally located in process areas, and are used to protect different equipment. They can be used both for cooling or fire suppression depending on the judgement of the operator according to the needs of the moment. Monitors must be installed taking elements like pipe racks and structures into account, so the device is able to reach and protect the area it was provided for. Elevated monitors can sometimes be useful to cover longer distances.

When designing and locating these devices, special attention must be provided to the safety of the personal, the required pressure to operate and the wind effect in the water flow. Monitors can be equipped with different nozzles and hose connections, which increases their flexibility.

Foam water systems

Foam water systems can be used either to supress a fire, to prevent a fire to occur or to cool down adjacent equipment. These systems require foam storage vessels, mixers and monitors or towers to operate. The mechanism consists in a mixture of water with a chemical compound and air that forms a foam with air enclosed in bubbles. The chemical compound chosen depends mainly on the type of fuel.

The solution with air causes an expansion of the foam with expansion ratios up to 1000:1. The ones with low expansion ratios are normally used outdoors to apply directly in a fire to extinguish it by covering surfaces of liquid fuel pools. In the case of LNG, high expansion foam is normally applied in the impounding basins when a release occurs. The specific objective here is to reduce the vaporization rate and avoid the formation of flammable vapour clouds. Even after ignited, the high expansion foam reduces flames size and therefore reducing the impact to the surroundings.

Fire truck intervention

A fire truck can be provided instead or together with fixed protection systems.

It becomes more necessary in locations where there is not fire brigade coverage or in facilities where the response times of the fire service are critical either because of the big size of the plant or because of special fire scenarios to be attacked.

Specific training in firefighting must be provided to operate the truck and control the situation. There are different strategies approaches firefighting that give flexibility to the overall fire protection strategy.

6.3. Cost of the fire protection systems

Every facility is different and therefore the fire protection design must take these particularities into account. With the plant layout and main characteristics of the jetty it is possible to estimate the protection mechanisms and corresponding cost of the fire protection systems.

Table 17 reports estimated costs for typical fire and gas detection, fire extinguishing and fire protection systems. As the jetty is only a part of the facility, the fire water storage and fire pumps are not included, since it is supposed these elements are already installed. The costs include the design, delivery, installation, cabling, galvanization and painting.

Table 17. Prices of the fire protection systems [44].

Concept	Price (€)	Remarks
Fire Panel	25,000	
Underground cabling	550	Price per meter of cable installed
Gas punctual detection	2,800	Price per installed detector.
Heat detection	-	Price included in the equipment fire protection prices
Flame detection	6,500	Price per installed detector. Each detector covers a triangular fringe of 50m length * 25 m in the widest point.
Closed circuit TV	4,700	Price per installed camera.
Fixed monitor	24,000	Price per installed fixed monitor.
Protection of a sphere	100,000	1,640 m^3 sphere protected by a water system of 110 nozzles.
Protection of a vertical tank	20,000	500 m^3 vertical tank protected by a water system of 23 nozzles + valve and trimming.
Protection of a horizontal tank	22,000	500 m^3 horizontal tank protected by a water system of 60 nozzles + valve and trimming.
Protection of a pump / compressor	2,000	Normally the protection of these equipment is included in the protection of bigger equipment.
Protection of a loading arm	12,000 – 14,000	10” LPG loading arm protected by a water system of 8 nozzles.
Fire truck	350,000	Carpump with increased capacity monitor

6.4. Fire Protection Systems Efficiency

To be able to perform the cost benefit analysis it is necessary to estimate the efficiency of the fire protection systems. The efficiency has three main components [14]:

- Operational Reliability
- On-Line Availability
- Response Effectiveness

The efficiency of the Fire Protection System (η_{FPS}) is then the multiplication of the three factors as shown in Equation 12. This efficiency is used as a factor to determine the reduction in the expected consequence after the installation of the fire protection system. The extent of the reduction is different when dealing with pool or jet fires than when handling explosions. Then a different efficiency is accounted.

Equation 12. Efficiency of the Fire protection System

$$\eta_{FPS} = P_{OR} \times P_{OLA} \times P_{RE}$$

Where:

P_{OR}: Operational reliability

P_{OLA}: On-line availability

P_{RE}: Response Effectiveness

Performance, maintenance, inspection, testing and weather indicators together with engineering judgement must be used to estimate the overall fire protection systems' efficiency.

6.4.1. Operational reliability

The operational reliability includes many aspects and is normally determined following a fault tree analysis, since it has to consider all of the systems on which the fire protection relies on, like power supply, water supply, mechanical failures and failures on the electronics, among others. If specific information about the reliability of each element in the system and the corresponding relation, a success tree analysis should be used to determine the operational reliability, otherwise databases for generic cases can be used.

Failure rate information for some fire protection systems and devices can be found in the literature or in the device's data sheet provide by the fabricant. As an example, the data set developed by the CCPS [45] is reported in Table 18.

Table 18. Failure rate data for fire protection elements [45]

Element	Period	Lower	Mean	Upper
Fire detection	10^6 hours	0.0198	1.14	4.41

Element	Period	Lower	Mean	Upper
Fire suppression system – water	10^6 hours	0.168	9.66	37.4
Fire suppression system - powder	10^6 hours	0.0245	1.41	5.45
Fire water pumps - diesel	10^3 hours	0.769	18.7	69.8
Fire water pumps - electric	10^3 hours	3.62	42.5	143.0

The operational reliability for a period of time t can be obtained from failure rate data using Equation 13.

Equation 13. Operational reliability [14]

$$P_{OR} = e^{-\lambda t}$$

Where:

λ: Failure rate

Values usually range from 0.75 to almost 1.0 for the operational reliability of fire protection systems.

6.4.2. On-Line Availability

An element is unavailable in a situation in which the fire protection system is acknowledgedly out of service. This unavailability can be caused by three main reasons [14]:

- Scheduled maintenance, inspection and testing activities. This aspect can be easily reported and quantified by checking the schedule of maintenance, inspection and testing activities and the duration per year.
- Unscheduled repairs or spurious triggering. This aspect is difficult to forecast, however a notion can be extrapolated from the failure rate data and the maintenance, testing and inspection periodicity.
- Weather conditions. Extreme low temperatures may avoid the fire protection system to be available, however provisions are provided to avoid it. As an example, glycol can be added to the water in an outdoor sprinkler system to avoid freezing in the pipes.

In general terms, the POLA should be close to 1 if maintenance, inspection and testing operations are performed regularly and the weather conditions are considered in the design.

6.4.3. Response Effectiveness

The residual damage is the fraction of remaining damage under a fire situation received by the equipment and structures if protected by the fire protection strategies. There are several ways the fire protection systems can successfully reduce the risk as follows:

- Some systems reduce the likelihood of the scenarios. It is recommended to give credit to this reduction in the response effectiveness.

- Gas detection can reduce the probability of ignition if ignition sources control is performed and can also reduce the expected consequence of the accident if the plant is sent to shut down earlier.
- Fire detection will allow an automatic reaction of the fire protection system or a faster action of the human intervention procedures like the operation of the monitors or the action of the fire brigade.
- If properly installed, the water systems will reduce the heat radiation to the equipment and structures protected. This prevents further damage and escalation of the incident.
- A well designed water curtain system can highly reduce the probability of ignition as it can block the flow of the vapour cloud toward potential ignition sources.

There is few information about the effectiveness and the extent of protection of the fire protection systems in the industry. Each design must perform a specific analysis considering the risk management policy and the particularities of the system.

Assuming a proper design, it is recommended to use a response effectiveness facing fire events between 0.8 and 0.95. The proposed values are mere guidelines, the designer must consider the specifics of the design and decide accordingly.

6.5. Definition of the strategy

The fire protection strategy has as its main objective to reduce the risk associated to the occurrence of a fire or explosion. Globally, it covers all of the safety layers of protection introduced in the Section 1.5.1. It ranges from a proper design to prevent the release to the mitigation of the consequences and intervention of the community.

A fire protection strategy design must consider the particularities of the facility. Factors important to consider are:

6.5.1. Location

The location of the facility has a high relevance in defining the fire protection strategy.

If the plant is surrounded by residential areas, the impact on life safety could be catastrophic, which should lead to more investment in fire protection systems. On the other hand, if the plant is in a remote location, the impact on life safety is lower, which could loosen the requirements. The expected consequence calculated before this step, gives a view of the impact on life safety.

The location also determines the availability of a community response team or fire brigade.

The availability of the resources is also important. For remote locations it can be more difficult or expensive to provide the fire protection devices or the supplies needed like the firefighting foam or glycol.

6.5.2. Operation philosophy:

Stand-alone facilities are increasing in number. This kind of facilities rely entirely on automatic systems for normal operation. Following this operation philosophy, the fire protection strategy must provide automatic detection that activates the fire suppression systems. In countries with high labour cost, more systems will be automated.

On the other hand, there are the traditional manually operated facilities. Although decreasing in number, this kind of facilities are still the large majority in the industry. These facilities rely highly on the human intervention for the normal operation. Under this philosophy, the detection is done visually by CCTV and the fire protection devices, like monitors, hoses and hydrants are operated manually.

The training of personnel is critical in this approach. The reaction time of the detection and suppression are considerably higher than on the automated approach.

The possibilities can range from fully automated to semi-automatic and to fully manual operated systems. This will depend on the operation philosophy, the cost benefit analysis and other particularities of the jetty.

6.5.3. Cost benefit analysis

As mentioned before, the fire protection strategy design must be carried out together with the cost benefit analysis to correctly meet the needs of the facility. The fire protection strategy must be coherent with the expected consequence in all of the assets; property loss, business continuity, life safety, environment and reputation.

6.5.4. Type of fire:

The annualised risk illustrates the contribution of each type of fire accident on the total. The way to protect the facility against each type of incident is different.

Pool fire

In case of detection of a pool of LNG or LPG, the release must be limited by triggering the emergency shut down system of the section compromised or the facility. To avoid ignition, all of the potential ignition sources must be removed and the operators must be evacuated, except the ones needed to handle the incident. The fire brigade must be informed.

In the case of LNG, the rate of vaporization can then be minimized by the application of foam over the release so the concentration is always below the flammable limit. The monitors and water curtains can be used to separate the release from ignition sources.

If ignition occurs, cool down the equipment compromised and wait until the fire is over.

Jet fire and fire ball

In case of detection of an instantaneous release of LNG or LPG, the release must be limited by triggering the emergency shut down system of the section compromised or the facility. To avoid ignition, all of the potential ignition sources must be removed and the operators must be evacuated, except the ones needed to handle the incident. The fire brigade must be informed.

The monitors and water curtains can be used to separate the release from ignition sources.

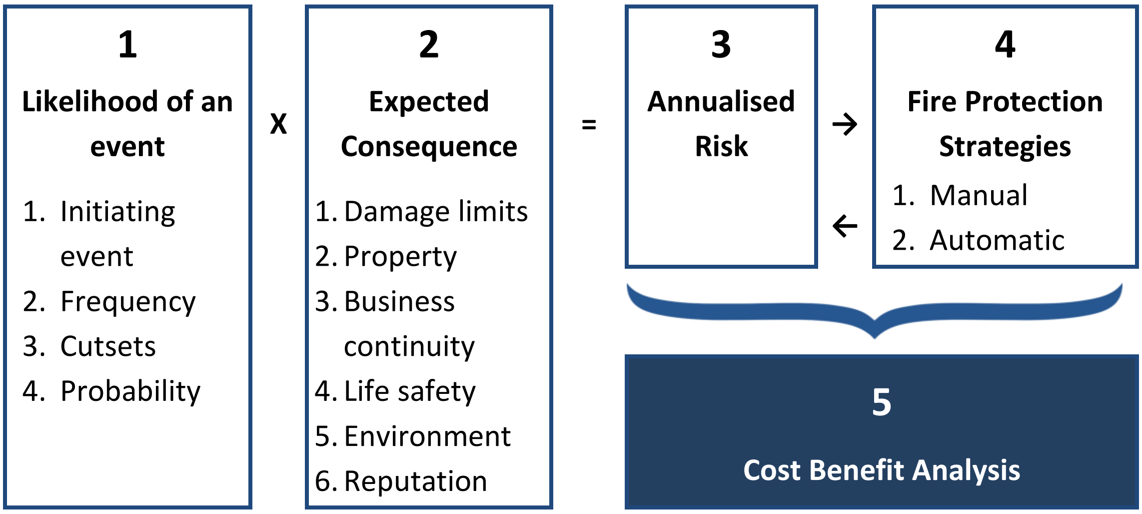
If the ignition occurs, cool down the equipment compromised and wait until the fire is over.

Vapour cloud explosion

In case of detection of an instantaneous release of LNG or LPG, the release must be limited by triggering the emergency shut down system of the section compromised or the facility. To avoid ignition, all of the potential ignition sources must be removed and the operators must be evacuated, except the ones needed to handle the incident. If available, the fire brigade must be informed.

There is not much to be done in the case of an explosion. Special effort must be made to avoid confined places near the jetty and to detect the leak early in the incident timeline.

7. Cost benefit analysis



The fire protection strategies must be evaluated in order to assure the risk reduction and also that the solution is economically feasible. To determine if the strategy is economically viable, the sum of the residual annualised risk and the annualised cost of the fire protection strategy must be lower than the annualised risk in the non-protected case. As bigger the difference as more reduction in the risk and lower cost in the protection systems.

Equation 14. Cost benefit relation

$$AC_{FPS}+AR^{*} < AR$$

Figure 12 shows a successful hypothetical cost benefit analysis where the annualised risk is reduced significantly from 1.0 to 0.2 at a cost of 0.4 per year. The protected scenario is then cheaper than the non-protected scenario and the risk is reduced significantly, accomplishing the goals of the strategy.

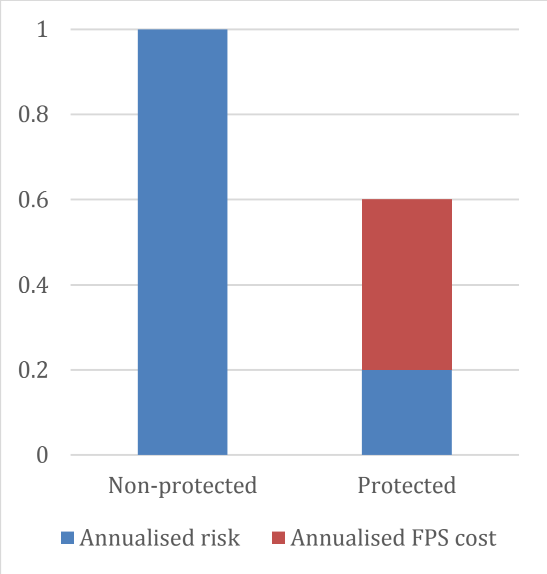


Figure 12. Cost benefit analysis for a successful risk reduction case

The designer must define the acceptable difference between the protected and the non-protected cases. If the protected case is more expensive than the non-protected one or if the difference between the two of them is lower than the defined by the designer, the fire protection system can be redesigned. Take into account other aspects like the individual and societal risk in this decision.

7.1. Annualised cost of the fire protection systems

The yearly fire protection cost must be calculated in order to perform the cost benefit analysis. With the system designed and the preliminary prices estimated, it is possible now to calculate the capital and operational costs in a yearly base.

The capital costs must be depreciated in the life span to find the annual price. Normally the fire protection equipment has a life span similar to the life span of the facility, which is between 10 and 20 years. Include all of the capital costs related to the fire strategy, like:

- Item price
- Delivery cost
- Construction works
- Installation (connection, painting)
- Start-up activities
- Others

In the operational costs the following aspects must be included:

- Maintenance
- Inspection

- Testing
- Human labour
- Power sources
- Supply of extinguishing agents (water, foam)
- Others

Equation 15. Annualised cost of the fire protection

$$AC_{FPS} = \frac{Capital\ costs}{years\ of\ life\ span} + Yearly\ operational\ costs$$

7.2. Recalculation of the expected consequence

The fire protection strategy has as its main purpose to reduce the expected consequence for all of the assets of the company. After the system is designed it is therefore necessary to estimate once again the expected consequence to evaluate the proposed design.

The base for this calculation is the expected consequence for the non-protected case, which was calculated in Chapter 4. In Equation 16, C_i* is the corrected expected monetary consequence for scenario i after the installation of the fire protection systems and η_{FPS} is the fire protection system efficiency accounted for the facility. If not all of the equipment were protected go a step before and recalculate the expected equipment by equipment.

Equation 16. Recalculated expected consequence with fire protection

$$C_i^* = C_i * (1 - \eta_{FPS})$$

Although the likelihood of a scenario can also be changed by the action of the fire protection systems, in the methodology it is considered as unaltered. This change in likelihood must be considered together with the response effectiveness in the recalculation of the expected consequence. The annualised risk is recalculated with the new expected consequence.

Equation 17. Recalculated annualised risk with fire protection

$$AR^* = \sum_i F_{Ci} \times C_i^*$$

8. Parametric study

In order to identify the effect of some variables in the different steps of the methodology, a parametric study is performed. The parameters to evaluate are the release size and type, the type of fuel, the transfer method, the type of fire and the jetty configuration.

8.1. Likelihood of the event

Twenty initiating events are selected combining the two fluids, LNG and LPG, the two release sizes (low and medium for leaks and medium and high for ruptures), the two leak types and the four initiating events. This is expected to cover the range of possibilities sufficiently. Table 19 reports the cutsets and probabilities for all the scenarios considered, while Table 20 the corresponding resulting frequencies.

The initiating events' frequencies are taken from Table 2, while the ignition probabilities are taken from Table 4 and the explosion probability from Table 6.

Table 19. Event trees' cutsets and probabilities

	Fluid	Initiating event	Leak type	Release rate / Size	F_I * (hour^{-1})	P_D	P_V	P_E
1	LNG	Hose leak	Continuous	Low	$5.4*10^{-6}$	0.02	0.02	0.2
2	LNG	Hose leak	Continuous	Medium	$5.4*10^{-6}$	0.04	0.04	0.3
3	LNG	Hose rupture	Continuous	Medium	$5.4*10^{-7}$	0.04	0.04	0.3
4	LNG	Hose rupture	Continuous	High	$5.4*10^{-7}$	0.09	0.1	0.4
5	LNG	Hose rupture	Instantaneous	Small	$5.4*10^{-7}$	0.02	0.02	0.2
6	LNG	Arm leak	Continuous	Low	$3*10^{-7}$	0.02	0.02	0.2
7	LNG	Arm leak	Continuous	Medium	$3*10^{-7}$	0.04	0.04	0.3
8	LNG	Arm rupture	Continuous	Medium	$3*10^{-8}$	0.04	0.04	0.3
9	LNG	Arm rupture	Continuous	High	$3*10^{-8}$	0.09	0.1	0.4
10	LNG	Arm rupture	Instantaneous	Small	$3*10^{-8}$	0.02	0.02	0.2
11	LPG	Hose leak	Continuous	Low	$5.4*10^{-6}$	0.2	0.06	0.2
12	LPG	Hose leak	Continuous	Medium	$5.4*10^{-6}$	0.5	0.2	0.3
13	LPG	Hose rupture	Continuous	Medium	$5.4*10^{-7}$	0.5	0.2	0.3
14	LPG	Hose rupture	Continuous	High	$5.4*10^{-7}$	0.7	0.7	0.4
15	LPG	Hose rupture	Instantaneous	Small	$5.4*10^{-7}$	0.2	0.06	0.2
16	LPG	Arm leak	Continuous	Low	$3*10^{-7}$	0.2	0.06	0.2
17	LPG	Arm leak	Continuous	Medium	$3*10^{-7}$	0.5	0.2	0.3

	Fluid	Initiating event	Leak type	Release rate / Size	F_I * (hour^{-1})	P_D	P_V	P_E
18	LPG	Arm rupture	Continuous	Medium	$3*10^{-8}$	0.5	0.2	0.3
19	LPG	Arm rupture	Continuous	High	$3*10^{-8}$	0.7	0.7	0.4
20	LPG	Arm rupture	Instantaneous	Small	$3*10^{-8}$	0.2	0.06	0.2

* This frequency is per hour of use of the hose or arm

The possible consequences are reported in Table 8. The actual consequences will be known when the simulations are made.

All of the results are shown in Table 20. It is important to remind that the values are frequencies of occurrence per hour of usage of the hose or arm. The difference between using an arm and using a hose is evident. Scenarios 1 to 5 have a higher frequency than scenarios 6 to 10, and they differ at least in one order of magnitude. The same can be seen when comparing scenarios 11 to 15 with the scenarios 16 to 20. This is evident since the determination of the initiating events and their respective frequencies.

The frequency values for all of the fire scenarios are higher for LPG. This is because of the higher reactivity of the LPG, which subsequently leads to a higher probability of direct and delayed ignition, P_D is around 10 times higher for LPG than for LNG, while P_V is in average 5 times higher as reported in Table 3. From this analysis it can be concluded that when handling LPG it is more likely to have a fire related incident.

The situation is similar when comparing leaks with ruptures. Leaks have considerably lower frequency values than ruptures, this is because of their lower P_D and P_V as shown in Table 3.

Table 20. Frequencies per hour of use for all the scenarios analysed.

	Consequence 1	Consequence 2	Consequence 3	Consequence 4
1	1.08E-07	2.12E-08	8.47E-08	5.19E-06
2	2.16E-07	6.22E-08	1.45E-07	4.98E-06
3	2.16E-08	6.22E-09	1.45E-08	4.98E-07
4	4.86E-08	1.97E-08	2.95E-08	4.42E-07
5	1.08E-08	2.12E-09	8.47E-09	5.19E-07
6	6.00E-09	1.18E-09	4.70E-09	2.88E-07
7	1.20E-08	3.46E-09	8.06E-09	2.76E-07
8	1.20E-09	3.46E-10	8.06E-10	2.76E-08
9	2.70E-09	1.09E-09	1.64E-09	2.46E-08

	Consequence 1	Consequence 2	Consequence 3	Consequence 4
10	6.00E-10	1.18E-10	4.70E-10	2.88E-08
11	1.08E-06	5.18E-08	2.07E-07	4.06E-06
12	2.70E-06	1.62E-07	3.78E-07	2.16E-06
13	2.70E-07	1.62E-08	3.78E-08	2.16E-07
14	3.78E-07	4.54E-08	6.80E-08	4.86E-08
15	1.08E-07	5.18E-09	2.07E-08	4.06E-07
16	6.00E-08	2.88E-09	1.15E-08	2.26E-07
17	1.50E-07	9.00E-09	2.10E-08	1.20E-07
18	1.50E-08	9.00E-10	2.10E-09	1.20E-08
19	2.10E-08	2.52E-09	3.78E-09	2.70E-09
20	6.00E-09	2.88E-10	1.15E-09	2.26E-08

8.2. Expected consequence

This section will report and discuss the effect of different variables on the expected consequence. In order to do so, all of the scenarios reported in Table 19 were modelled using PHAST 7.1®.

Table 1 reports the properties and relevant conditions of LNG and LPG to be used in the simulations.

Table 21 reports the assumptions and input values for the simulations to be carried out for the continuous scenarios. The flow rate for the leak scenarios is calculated using Equation 2. For the rupture case the release flow rate is equivalent to the whole flow rate at a velocity of 10 m/s given the diameter. The typical maximum design pressure for transfer is 9 barg. This value is used for the incident modelling.

Table 21. Input values for the simulation of the continuous scenarios

#	Fuel	$P\ [N/m^2]$	$D\ [in]$	$Q\ [m^3/h]$	$A_h\ [m^2]$	$q_S\ [kg/s]$
1	LNG	1000000	4	-	$8.1*10^{-5}$	1.5
2	LNG	1000000	12	-	$7.3 * 10^{-4}$	12.9
3	LNG	-	4	296	-	37
4	LNG	-	12	2628	-	329
6	LNG	1000000	8	-	$3.24*10^{-4}$	5.7
7	LNG	1000000	20	-	$2.03 *10^{-3}$	35.8
8	LNG		8	1162	-	145.3

#	Fuel	$P\,[N/m^2]$	$D\,[in]$	$Q\,[m^3/h]$	$A_h\,[m^2]$	$q_S\,[kg/s]$
9	LNG	-	20	6761	-	845
11	LPG	1000000	4	-	$8.1*10^{-5}$	1.6
12	LPG	1000000	12	-	$7.3 * 10^{-4}$	13.6
13	LPG	-	4	296	-	41.1
14	LPG	-	12	2628	-	365
16	LPG	1000000	8	-	$3.24*10^{-4}$	6.0
17	LPG	1000000	20	-	$2.03 *10^{-3}$	37.7
18	LPG	-	8	1162	-	161.4
19	LPG	-	20	6761	-	939

The size of the instantaneous release is considered as the volume contained in a 50 meters section of hose / loading arm. The geometry of the spill will determine the area of the pool, however for simplification purposes in the model the depth is considered to be 50 cm.

Table 22 reports the assumptions and input values.

Table 22. Input values for the simulation of the instantaneous scenarios

#	Fuel	$D\,[in]$	$L\,[m]$	$V\,[m^3]$	$A\,[m^2]$
5	LNG	12	50	3.65	7.30
10	LNG	20	50	5.14	10.27
15	LPG	12	50	3.65	7.30
20	LPG	20	50	5.14	10.27

Following the damage levels reported in Table 9, thermal radiation contours were generated for each fire case at four levels: 4 kW/m², 12.5 kW/m², 18 kW/m² and 35kW/m². Only the wind class D with wind velocity of 5m/s was considered as it is the worst case for fire and explosion scenarios.

The results of the simulations of some of the scenarios shown that it is not always possible to have different consequences for the branches of the event tree. This is the case of scenarios with low flows, which are not enough to form pools or to form a flammable cloud big enough to have a flash fire, the only consequences possible are therefore a jet fire, a vapour cloud explosion in case of confinement and the toxic effects.

For the instantaneous cases it was found that the vaporization rates are not high enough to generate an explosive cloud as it disperses immediately. As a consequence, all of the hazardous consequences for instantaneous ignition are pool fires.

The actual consequences for each scenario were found through the simulation with PHAST and are reported in Table 23.

Table 23. Consequences found after modelling

Scenarios	**Consequence**			
	1	**2**	**3**	**4**
1, 2, 3, 6, 7	Jet Fire	Vapour Cloud Explosion	Jet Fire	Cryogenic effects
11, 12, 13, 14, 16, 17, 18	Jet Fire	Vapour Cloud Explosion	Jet Fire	Toxic effects
4, 8, 9	Jet Fire	Vapour Cloud Explosion	Pool Fire	Cryogenic effects
19	Jet Fire	Vapour Cloud Explosion	Pool Fire	Toxic effects
5, 10	Pool Fire	Pool Fire	Pool Fire	Cryogenic effects
15, 20	Pool Fire	Pool Fire	Pool Fire	Toxic effects

8.2.1. Fuel

To evidence the influence of the fuel, scenarios 2 and 12 are compared. Figure 13 shows the radiation contours for the consequence 1 of scenario 2, which is a LNG jet fire. On the other hand Figure 14 shows the radiation contours for the consequence 1 of scenario 12, which is a LPG jet fire. Although slightly different in shape, no strong difference is seen in the distances reached by the different radiation contours. The radiation contour for the LNG is broader since the initial temperature is lower and therefore the fuel needs an additional distance to warm up, evaporate and burn, which means that the combustion zone will be further from the release point. In the case of the LPG, although some cooling is generated by the depressurization, this is not as extreme as cryogenic conditions.

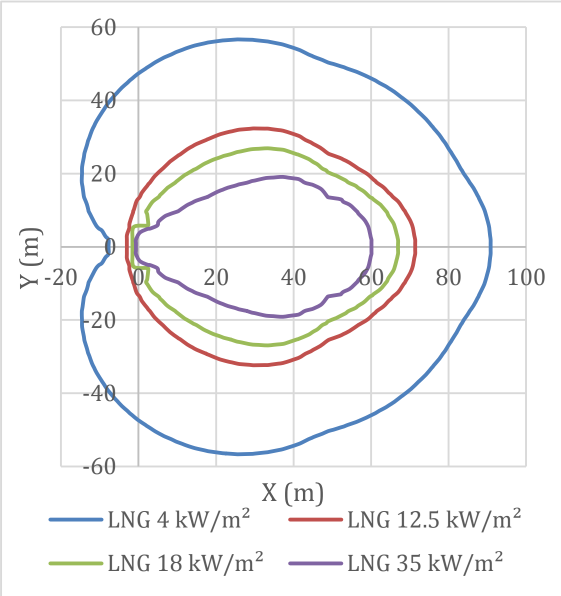


Figure 13. Radiation contours for a LNG jet fire.

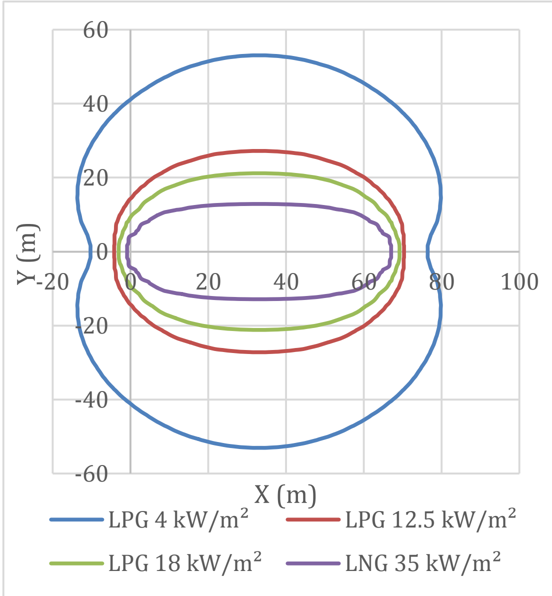


Figure 14. Radiation contours for a LPG jet fire.

Figure 15 reports the 35 kW/m² contours for all of pool fires following an instantaneous release (Scenarios 5, 10, 15 and 20). The difference in the radiation contours can be explained by the fraction of heat that is lost to the fuel, both to heat it up and vaporize it. This fraction is evidently higher for the LNG, given its considerably low temperature.

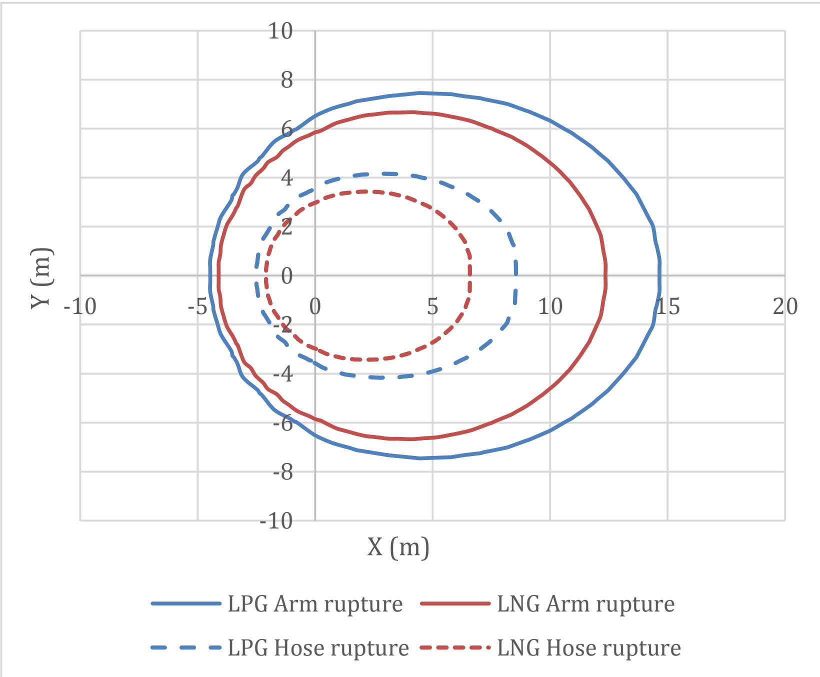


Figure 15. Radiation contour (35 kW/m²) of the instantaneous releases.

8.2.2. Type of fire

Figure 16 shows two of the possible consequences for scenario 19. This is a continuous release of LPG from a loading arm rupture. It was found that the release flow is big enough to form a pool before vaporization.

The turbulence generated in the continuous case is generated by the sudden release of a pressurized liquid turning into a vapour, which causes high velocities and therefore a fast mixing with air, reaching easily flammable conditions. In the case of the pool fire there is no release of pressure since the LPG is already on the ground in liquid form, therefore the mixing process is expected to be slower as well as the burning rate.

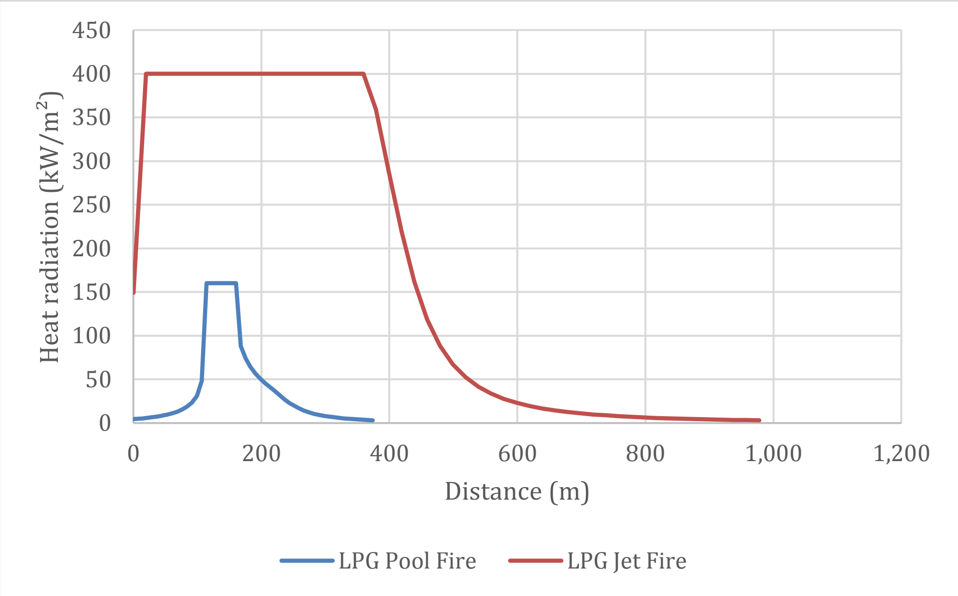


Figure 16. Incident radiation versus distance for a pool and a jet fire.

8.2.3. Size of the release

Figure 17 compares the expected contour of 35 kW/m² of a LPG jet fire from a hose rupture (scenario 14) and an arm rupture (scenario 19). As it can be seen in the Table 21, the hose case has a release rate of 365 kg/s, while the arm case rises up to 939 kg/s. The relation between the release rate and the distance affected by the 35 kW/m² contour is evident. The comparison between a leak and a rupture shows similar results.

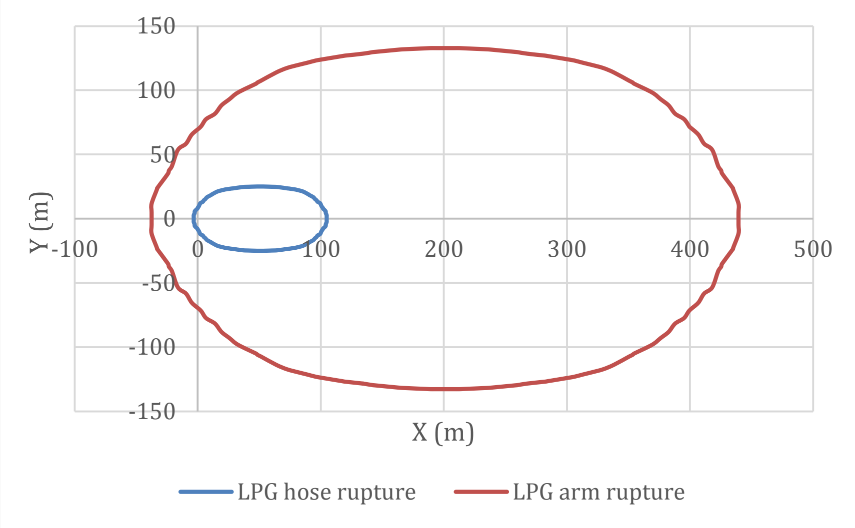


Figure 17. Comparison of the rupture of a hose and the rupture of an arm.

8.2.4. Explosion

The area covered by the explosion contours increases with the time after the release event, since the vapour cloud is fed by the release and grows bigger. The overpressure contours were all taken after 20 seconds of the release start.

Figure 18 shows the overpressure contours for the explosion consequence for scenario 18. The explosion occurs after 60 seconds of the release. The explosion contours are circular because of the even distribution of the pressure waves. As shown, the distance between the 0.14 bar and the 0.21 bar contours is small compared to the size of the wave and the affected area is considerably high. The displacement of the contours along the X axis is due to the wind carrying away the explosive cloud.

It can also be seen that the size of the jet fire contours are considerably smaller, showing the significantly larger negative impact of a vapour cloud explosion.

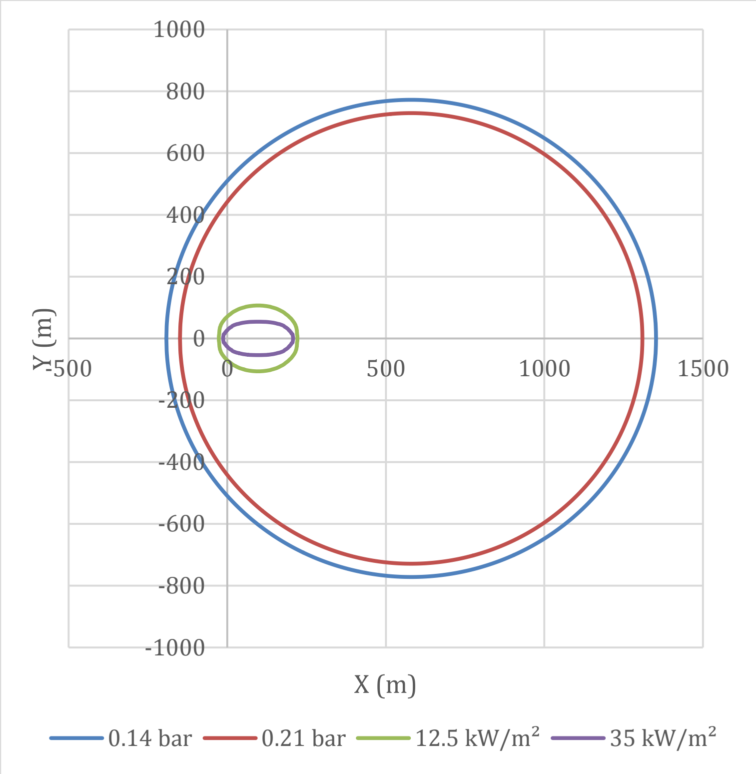


Figure 18. Explosion contours for the scenario 18.

8.2.5. Jetty configuration

The jetty configuration is important regarding the expected consequences and the fire protection systems:

- Dispersion: The L and T type jetties pose an advantage regarding dispersion. As the transfer operation is carried out further from the structures and equipment of the facility, the dispersion is faster and the probability of ignition and confined space is highly reduced.
- Distance from the equipment: Considering only risks from the jetty, it is positive for the equipment in the shoreline to have an L or T-type jetty.
- Cost of the jetty: If an accident happens in an L or T-type jetty, the expected consequences must also include this additional structure.
- Fire protection systems: As longer the jetty or as further the transfer loading arm or hose, as more difficult and expensive to provide fire protection and utility services in general. In an elongated jetty the space is reduced and does not provide much flexibility to install the protection devices.
- Fire brigade: The access of the fire brigade in L or T type jetties is more difficult or significantly more expensive as the jetty should have installed a roadway with the provisions necessary for the fire truck.

8.3. Annualised risk

The expected consequence for the 20 scenarios of Table 19 was calculated in terms of area and the annualised risk was calculated using the likelihood previously calculated. The goal is to identify the different levels of contribution to the total annualised risk when varying the mentioned parameters.

Figure 19 shows the contribution to the total annualised risk per type of fire. As it can be seen, the negative impact of the vapour cloud explosions is more than twice the one of the fires. And now again, the negative impact of the jet fires is more than three times the one of the pool fires. This is a red flag indicating the importance of avoiding confined spaces and performing ignition sources control.

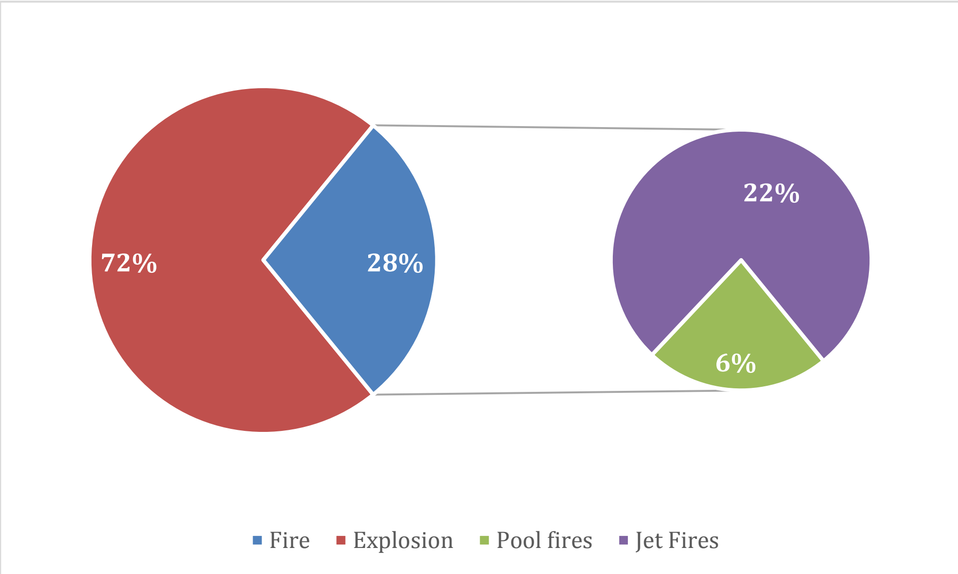


Figure 19. Contribution by type of consequence.

The annualised risk confirms the findings in the expected consequence regarding type of fuel and type of transfer method. In overall the LPG scenarios contribute more than the LNG scenarios. Although the likelihood of leak and rupture is higher for the hoses than for the arms, the flows handled by the hoses is significantly lower, which explains the difference in the contribution to the total annualised risk.

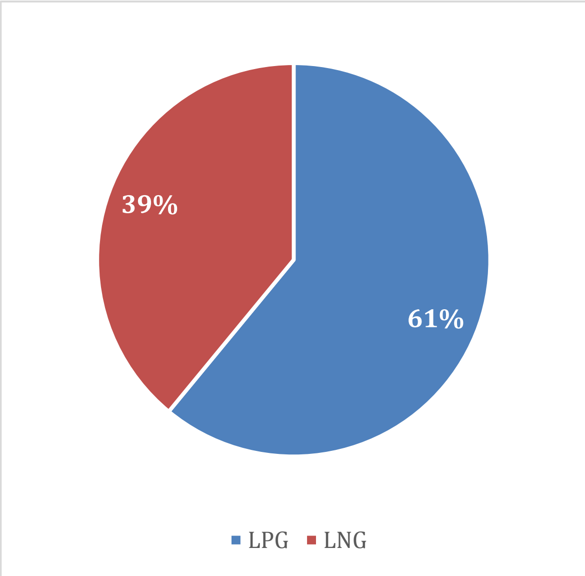


Figure 20. Contribution by type of fuel.

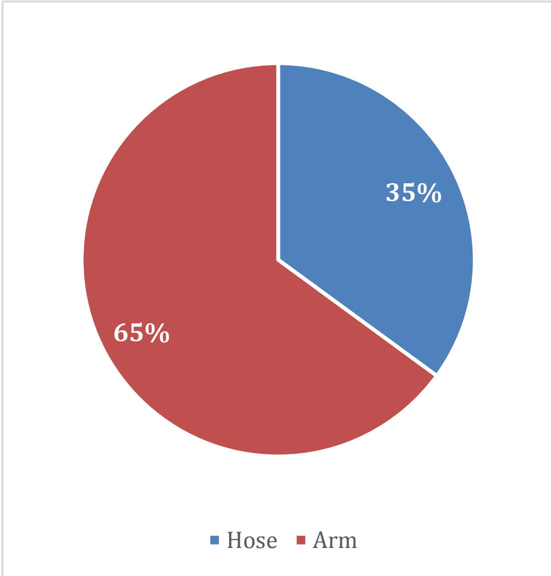


Figure 21. Contribution by type of transfer method.

9. Case study

To illustrate the use of the methodology a hypothetical case study is formulated and the fire protection strategy is designed and analysed.

The methodology is applied to a small shoreline jetty in the port of Beveren, on the left bank of the Schelde River in Belgium. The facility layout is shown in Figure 22.

Figure 22. Plant layout for the case study [11].

The facility handles liquid fuels and LPG. The liquid fuels are unloaded on the L-type jetty, while the LPG is unloaded in the shoreline jetty. The tanks and L-type jetty are already protected, so the company has decided to define the fire protection strategy for the shoreline jetty handling LPG.

The production per month is on average 1’000,000 € for the LPG business section and 3’000,000 € for the liquid fuels business section.

The LPG jetty consists of two loading arms of 8 inches (200 mm) diameter that are used five times per week, each use lasts in average 8 hours. The elements installed in the unprotected jetty are the following:

- 2 retractile loading arms of 8” for LPG in the shoreline jetty.
- Shoreline jetty
- 3 LPG pumps without gaskets with a capacity of 150 m^3/h in the shoreline jetty.
- 3 LPG compressor with a capacity of 100 m^3/h in the shoreline jetty.
- 50 meters of pipe with a diameter of 4 in (100 mm). Above ground.
- 50 meters of pipe with a diameter of 8 in (200 mm). Above ground.

As mentioned before, a large part of the facility is already protected and therefore, the main elements for a fire protection system are already installed and operating. The elements in the protected areas are the following:

- 10 large tanks with a capacity of 4,000 m³.
- 24 small tanks with a capacity of 1,800 m³.
- 2 retractile loading arms 8" for liquid fuels in the L-type jetty.
- L-type jetty.

9.1. Initiating events and cutsets

It is important to remember from Section 1.5.1 that the initiating events should be the result of previous qualitative and / or semi quantitative analysis, only the scenarios whit high levels of risk must be studied in a fully quantitative way.

As no previous analysis has been made for this case study and as the case study intends to focus in jetty related fire scenarios, the initiating events are defined as the continuous releases due to:

1. Full rupture of the loading arm for LPG.
2. Leakage of the loading arm for LPG.
3. Leakage of the 4" intake pipe of the pumps
4. Leakage of the 4" intake pipe of the compressors
5. Rupture of the 8" intake pipe of the compressors
6. Full rupture of the 4" pipe
7. Small leakage of the 4" pipe
8. Medium leakage of the 4" pipe
9. Large leakage of the 4" pipe
10. Full rupture of the 8" pipe
11. Small leakage of the 8" pipe
12. Medium leakage of the 8" pipe
13. Large leakage of the 8" pipe

No instantaneous initiating events are included since the section of pipe between safety valves is too short to be significant. To decrease the difficulty no scenarios from the ship are considered.

The frequencies are taken from the Flemish Government [20]. As the frequency of usage of the loading arm and the length and diameter of the pipes are known, only the yearly frequencies are reported in Table 24.

The release size is now calculated to finish characterizing the initiating events. Equation 2 is used to calculate the release low rate for the leak scenario. The following assumptions are made:

- The maximum operation pressure of the loading arm is 9 barg.
- The velocity at the loading arm is 10 m/s.

The probabilities for the cutsets are taken from the Flemish Government [20]. All of the frequencies and probabilities are reported in Table 24.

Table 24. Frequencies and probabilities for the initiating events of the case study.

	Event	$F_I\ [year^{-1}]$	$q_S\ [kg/s]$	P_D	P_V	P_E
1	Arm Leakage	$3.1*10^{-4}$	6.0	0.2	0.06	0.2
2	Arm Rupture	$3.1*10^{-5}$	161.4	0.5	0.2	0.3
3	Pump leakage	$1.0*10^{-4}$	3.9	0.2	0.06	0.2
4	Compressor leakage	$4.4*10^{-3}$	3.9	0.2	0.06	0.2
5	Compressor rupture	$1.0*10^{-4}$	161.4	0.5	0.2	0.3
6	4” pipe small leakage	$1.4*10^{-4}$	1.5	0.2	0.06	0.2
7	4” pipe medium leakage	$6.0*10^{-5}$	3.3	0.2	0.06	0.2
8	4” pipe big leakage	$2.5*10^{-5}$	18.9	0.5	0.2	0.3
9	4” pipe rupture	$1.1*10^{-5}$	41.1	0.5	0.2	0.3
10	8“ pipe small leakage	$7.0*10^{-5}$	5.8	0.2	0.06	0.2
11	8” pipe medium leakage	$3.0*10^{-5}$	13.1	0.5	0.2	0.3
12	8” pipe big leakage	$1.3*10^{-5}$	75.7	0.5	0.2	0.3
13	8” pipe rupture	$5.5*10^{-6}$	161.4	0.5	0.2	0.3

The event tree is shown in Figure 23, the corresponding frequencies for each fire scenario are reported in Table 25.

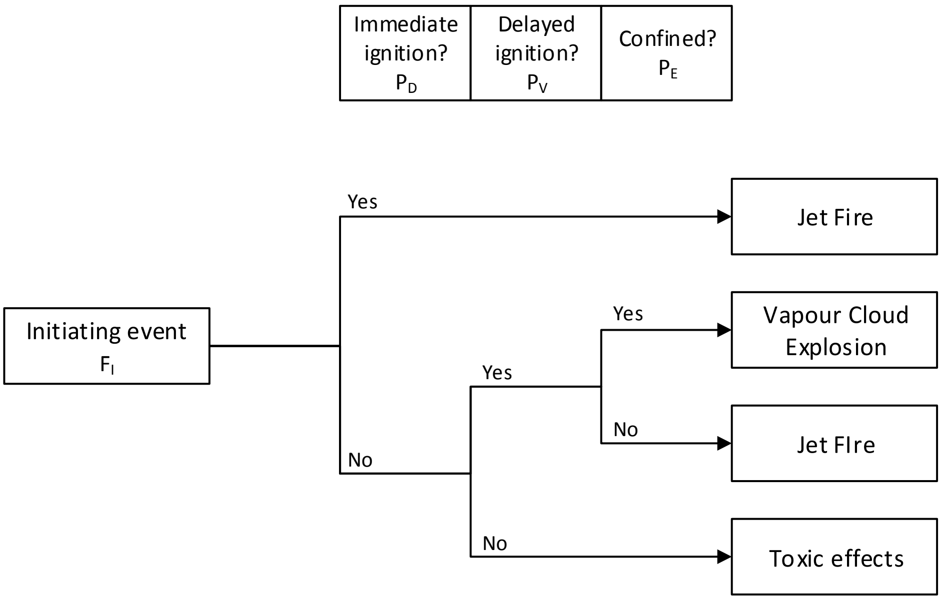


Figure 23. Event tree for the case study

Table 25. Frequencies for the consequences of the case study.

	Consequence 1	**Consequence 2**	**Consequence 3**	**Consequence 4**
	Jet Fire	**Vapour Cloud Expl.**	**Jet Fire**	**Toxic effects**
1	$2.50*10^{-4}$	$1.20*10^{-5}$	$4.81*10^{-5}$	$9.41*10^{-4}$
2	$6.26*10^{-5}$	$3.75*10^{-6}$	$8.76*10^{-6}$	$5.01*10^{-5}$
3	$6.00*10^{-5}$	$2.88*10^{-6}$	$1.15*10^{-5}$	$2.26*10^{-4}$
4	$2.64*10^{-4}$	$1.27*10^{-5}$	$5.07*10^{-5}$	$9.93*10^{-4}$
5	$1.50*10^{-4}$	$9.00*10^{-9}$	$2.10*10^{-5}$	$1.20*10^{-4}$
6	$2.80*10^{-5}$	$1.34*10^{-6}$	$5.38*10^{-6}$	$1.05*10^{-4}$
7	$1.20*10^{-5}$	$5.76*10^{-7}$	$2.30*10^{-6}$	$4.51*10^{-5}$
8	$1.25*10^{-5}$	$7.50*10^{-7}$	$1.75*10^{-6}$	$1.00*10^{-5}$
9	$5.50*10^{-6}$	$3.30*10^{-7}$	$7.70*10^{-7}$	$4.40*10^{-6}$
10	$1.40*10^{-5}$	$6.72*10^{-7}$	$2.69*10^{-6}$	$5.26*10^{-5}$
11	$1.50*10^{-5}$	$9.00*10^{-7}$	$2.10*10^{-6}$	$1.20*10^{-5}$
12	$6.25*10^{-6}$	$3.75*10^{-7}$	$8.75*10^{-7}$	$5.00*10^{-6}$
13	$2.75*10^{-6}$	$1.65*10^{-7}$	$3.85*10^{-7}$	$2.20*10^{-6}$

9.2. Expected consequence

The expected consequence is calculated using contours generated by simulations performed using PHAST 7.1®.

9.2.1. Damage limits

Property

As discussed in Chapter 4, 35 kW/m² and 12.5 kW/m² are the recommended damage limits for radiation heat and 0.21 bar and 0.14 bar are the recommended damage limits for overpressure. These damage limits are applied to the jetties and the equipment installed on this areas.

The level of damage factor (K) is set as recommended in Table 15. The prices of the equipment are the ones listed in Table 14.

No monetary loss is considered because of the impact on the ship.

Business continuity

A business interruption is expected after an incident. The time without production depends on the extent of damage. Additional to the downtime, the compliance with the client and the commercial penalties are accounted as a 120% and 20% of the downtime in production cost respectively. Three levels of damage are defined:

1. If the scenario affects the two LPG loading arms, the downtime is 4 months for the LPG business.
2. If the scenario additionally affects several tanks, the downtime is 8 months for the LPG business.
3. If the damage is general, affecting most of the tanks and the L-type jetty, the downtime is 1 year for the entire facility.

Life safety

A fatality of a worker is assigned a monetary cost of € 1’000,000 while a fatality of a person from the community is assigned a cost of € 1’500,000.

Environment

Three levels of damage are defined:

- Minor: The cost of cleaning plus the respective penalties is € 50,000.
- Medium: The cost of cleaning plus the respective penalties is € 500,000
- Extensive: The cost of cleaning plus the respective penalties is € 5’000,000.

Reputation

Three levels of damage are defined:

- Minor: The cost of the loss in reputation is € 50,000.

- Medium: The cost of the loss in reputation is € 500,000
- Extensive: The cost of the loss in reputation is € 5’000,000.

9.2.2. Contours

As an example, the jet fire contours for the first scenario (leakage of one the loading arms) are shown in Figure 24, while the overpressure contours are illustrated in Figure 25. The vapour cloud explosion contours are generated for an ignition after 20 seconds of the release.

The yellow lines represent the 12.5kW/m² and the 0.14 bar contours respectively, and the red ones represent the 35kW/m² and the 0.21 bar contours respectively.

Figure 24. Case study jet fire contours, rupture case

Figure 25. Case study explosion contours, rupture case

The jet fire contours for scenario 2 (rupture of one of the loading arms) are shown in Figure 26, while the overpressure contours are illustrated in Figure 27. The vapour cloud explosion contours are generated for an ignition after 20 seconds of the release.

Figure 26. Case study jet fire contours, leakage case

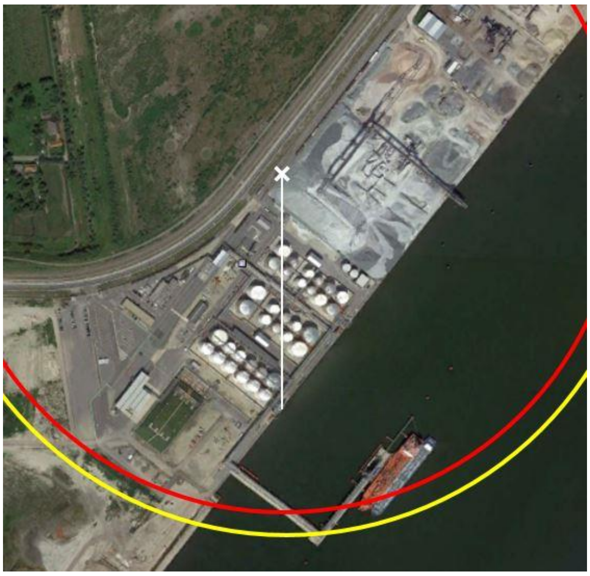


Figure 27. Case study explosion contours, leakage case

9.2.3. Monetary loss

To simplify the calculations, the monetary loss is calculated according to the following assumptions:

- The efficiency of the fire protection system in the protected area is 0.76. The respective monetary loss is calculated using Equation 16. The efficiency is calculated using Equation 12 and the following information:
  - The operational reliability is calculated with Equation 13 using the mean failure rate for water based fire protection systems in Table 18, the result is 0.92.
  - The On-Line Availability is set to 0.97 considering ten days a year for maintenance, testing and inspection works.
  - The Response Effectiveness is set to 0.85 assuming a proper design.
- The fire protection system installed in the protected areas is not able to protect the equipment from the explosion overpressure. The advantage of installing gas detectors is the additional time to act. Since the release is not in the protected areas, when the gas is detected, the explosive cloud is already present and the risk cannot be mitigated.
- A 10% of the property loss due to equipment is added to take the piping into account.
- The workers' and pubic fatalities and the damage levels in environment and reputation are defined qualitatively depending on the extent of the contour.
- The toxicity has no monetary impact in any of the assets.

The detail of the calculations can be found in Annex 3. Table 26 shows the expected consequence for all of the scenarios considered.

Table 26. Expected consequence for the case study.

	Consequence 1	Consequence 2	Consequence 3	Consequence 4
	Jet Fire	Vapour Cloud Expl.	Jet Fire	Toxic effects
1	€ 13,199,725	€ 31,632,700	€ 13,199,725	€ -
2	€ 29,132,964	€ 167,352,000	€ 29,132,964	€ -
3	€ 13,196,828	€ 15,470,900	€ 13,196,828	€ -
4	€ 13,196,828	€ 15,470,900	€ 13,196,828	€ -
5	€ 13,501,807	€ 167,352,000	€ 13,501,807	€ -
6	€ 14,406,548	€ 167,352,000	€ 14,406,548	€ -
7	€ 13,196,828	€ 15,470,900	€ 13,196,828	€ -
8	€ 13,196,828	€ 16,150,700	€ 13,196,828	€ -
9	€ 13,254,769	€ 167,352,000	€ 13,254,769	€ -
10	€ 29,132,964	€ 167,352,000	€ 29,132,964	€ -
11	€ 13,199,725	€ 31,632,700	€ 13,199,725	€ -
12	€ 13,254,769	€ 167,352,000	€ 13,254,769	€ -
13	€ 29,142,708	€ 167,352,000	€ 29,142,708	€ -

9.3. Annualised risk

The Annualised risk without protection is 18,548 €/year. Figure 28 shows the contribution of each type of fire, type of release and element originating the release. A higher risk value is found for scenarios with lower consequence but higher frequency.

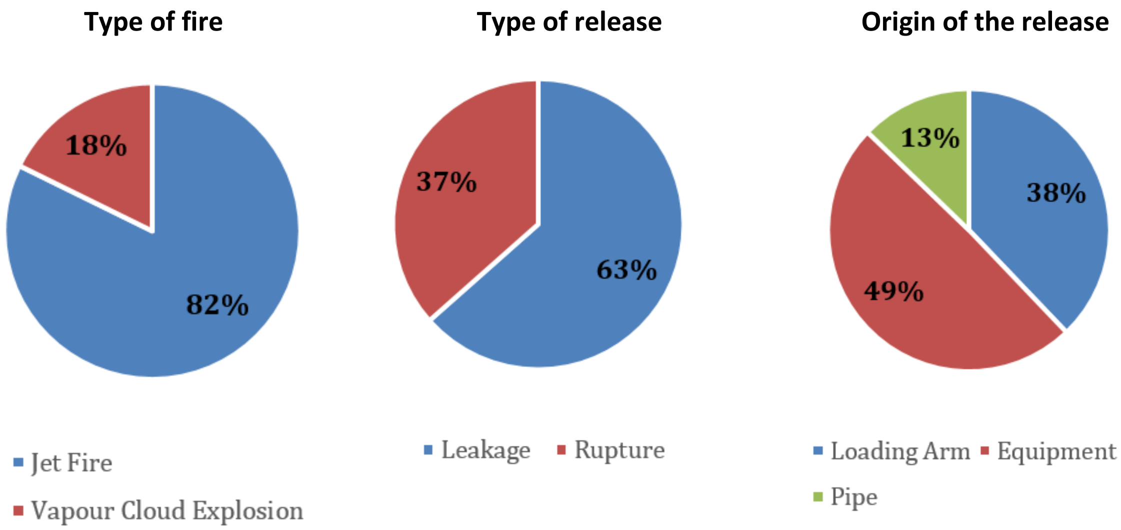


Figure 28. Contribution to the annualised risk in the case study.

9.4. Fire protection strategy

The proposed strategy is semi-automatic. The detection can be made through the CCTV or using the gas and flame detectors installed. Both manual and automatic fire protection systems are installed, like monitors and water spray systems.

The strategy for this jetty will emphasize on detection to avoid escalation of the consequences. Gas detectors will be installed in the couplings of the loading arms and in the discharge pipes of the pumps and compressors. Additionally two flame detectors will be placed to cover the whole jetty.

The fire protection strategies to be considered are the ones to protect the elements in the shoreline jetty area, which are the compressors, the pumps, the control valves and the loading arms. The depreciation time for the fire protection equipment it is 15 years. Table 27 shows the devices to be installed, together with the annualised cost to protect the jetty.

Table 27. Fire protection cost for the case study

Concept	Qty.	Price per unit (€)	Maintenance, inspection and testing (€/year)	Annualised cost (€/year)
Gas punctual detection	8	2,800	300	1,793
Flame detection	4	6,500	300	2,033
Cameras CCTV	2	4,700	300	927
Fixed monitor	2	24,000	600	3,800
Pumps / compressors protection	6	2,000	300	1,100
Loading arm protection	2	12,000	600	3,800
Total				**13,453**

9.5. Cost benefit analysis

The expected consequence is recalculated to take the effect of the fire protection system into account. To do so the following estimations were made:

- The assumptions made to calculate the efficiency of the fire protection systems in the protected area are valid for the shoreline protection. The efficiency (η_{FPS}) is 0.76.
- The fire protection system is able to protect against explosions, given the installation of gas detectors in every the release sources.
- The extent of damage in environment and reputation are recalculated using the efficiency of the fire protection system for fire and explosion cases accordingly.
- The downtime is one week for the jet fire scenarios, given that all of the equipment in the facility are protected and repair operations must be performed. For the explosion the downtime is estimated as half the downtime for the unprotected case.

- The fatalities are recalculated. No fatalities are estimated for the jet fire scenarios. For the explosion scenarios, half the fatalities from the unprotected case are considered. This is to account the detection that allows a quick reaction and earlier evacuation procedures.

The expected consequence is recalculated and the detail can be seen in Annex 3. The likelihood of the scenarios is the same as for the unprotected case reported in Table 25. Table 28 reports the recalculated annualised risk per initiating event, per scenario. The total annualised risk for the protected case is 2,895 €/year.

Table 28. Recalculated expected consequence for the case study

	Consequence 1	**Consequence 2**	**Consequence 3**	**Consequence 4**
	Jet Fire	**Vapour Cloud Expl.**	**Jet Fire**	**Toxic effects**
1	€ 2,159,660	€ 39,889,584	€ 2,159,660	€ -
2	€ 1,510,939	€ 7,517,000	€ 1,510,939	€ -
3	€ 1,508,042	€ 3,665,711	€ 1,508,042	€ -
4	€ 1,508,042	€ 3,665,711	€ 1,508,042	€ -
5	€ 1,813,021	€ 39,889,584	€ 1,813,021	€ -
6	€ 2,033,244	€ 39,889,584	€ 2,033,244	€ -
7	€ 1,508,042	€ 3,665,711	€ 1,508,042	€ -
8	€ 1,508,042	€ 3,849,015	€ 1,508,042	€ -
9	€ 1,565,983	€ 39,889,584	€ 1,565,983	€ -
10	€ 2,159,660	€ 39,889,584	€ 2,159,660	€ -
11	€ 1,510,939	€ 7,517,000	€ 1,510,939	€ -
12	€ 1,565,983	€ 39,889,584	€ 1,565,983	€ -
13	€ 2,169,405	€ 39,889,584	€ 2,169,405	€ -

Figure 29 shows the comparison between the unprotected and the protected cases. There is a slight difference meaning that it is cheaper for the company to protect using a semi-automated strategy than to leave the jetty unprotected.

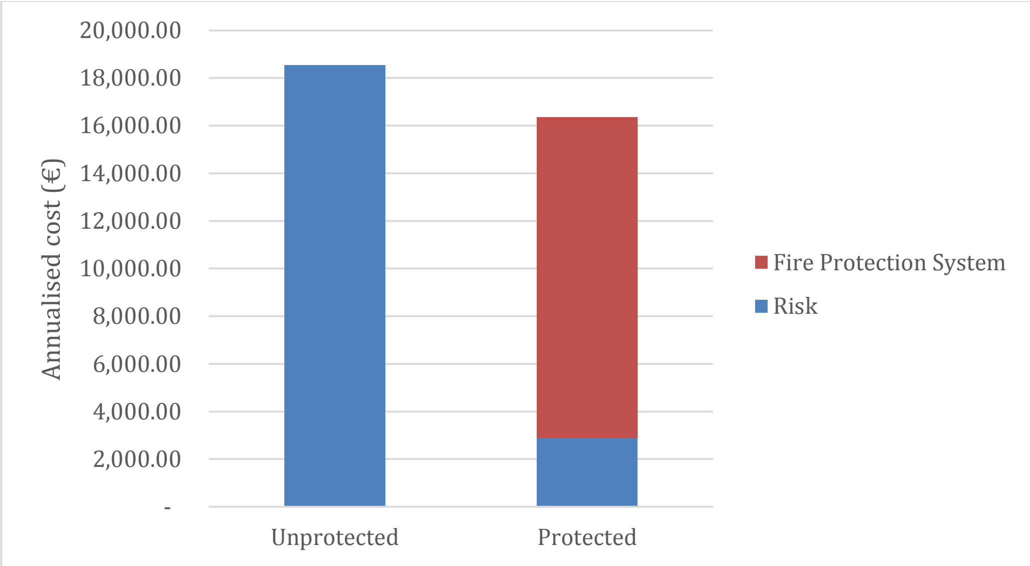


Figure 29. Cost benefit analysis for the case study

This is the result of applying the methodology using realistic values and reasonable assumptions. However the assumptions made can significantly impact the result.

The ship and structures were not considered in the analysis. The inclusion of the ship in the analysis gives rise to additional initiating events to consider and therefore additional fire and explosion scenarios. The impact on the structures increases the annualised risk for the unprotected case. Considering these two elements would increment the gap between the cost of the unprotected and the protected cases. It is recommended then to include the damage on the structures and on the ship on the methodology.

The downtime for an explosion impacting the whole facility was defined as one year. This is assuming that the decision makers choose to rebuild the facility once destroyed. In reality, the chances to rebuild a facility after an explosion affecting its whole area are not high. The monetary losses are so high that the company can go bankrupt. The business interruption is therefore total and the business continuity impact is significantly higher than the one considered in the case study.

The jetty studied is small in capacity, only one ship can load at the same time. For larger jetties, both the likelihood and the expected consequence are higher given the higher amount of equipment that can fail and be lost and the bigger monetary loss in case of a business interruption.

10. Conclusions

The effect of the fuel is seen in the likelihood. The amount of heat generated through combustion is comparable. Fuels with higher reactivity like LPG have larger probabilities of ignition than fuels with low reactivity like LNG. This is evidenced in the parametric study with a larger contribution to the annualised risk from the LPG than from the LNG.

The temperature of the fuel plays an important role in the expected consequence. Part of the heat generated by an LNG fire is lost to warm up the fuel and vaporize it. This fraction is lower for the LPG that is normally stored at ambient temperature. As a consequence, the heat radiation is higher for LPG than for LNG.

The jetty configuration affects the area of the facility affected by the scenarios and the access of the fire protection systems. An L or T-type jetty will have scenarios affecting smaller areas of the facility and therefore the expected consequences are lower. However the access of the fire brigade to the jetty is more difficult and therefore its efficiency could be lower.

The parametric study showed a big impact of the explosion scenarios on the annualised risk compared to the impact of the fire scenarios. On the other hand, the explosion scenarios in the case study have a significantly lower impact in the annualised risk. This is due to the higher proportion of rupture scenarios against leakage scenarios for the parametric study compared to the case study.

As expected, the larger the release, the larger the expected consequence, but also the lower the frequency. Leakages and hoses represent smaller releases than ruptures and loading arms, but are also assigned higher frequencies in the initiating event definition. In the parametric study, the expected consequence is a predominant factor over the likelihood, this is because of the unrealistic proportion between rupture and leakage events. In the case study, the situation is different given the more realistic approach.

The contribution to the annualised risk in the case study is discussed. The explosion scenarios contribution is five times lower than the contribution of the jet fires. Regarding the type of release event, 37% comes from the rupture events in contrast to 63% from the leakage ones. The equipment and piping contribution accounts for 62% in comparison with the 38% of the loading arms. It can be concluded that the scenarios with low expected consequence but high frequency have a higher impact on the annualised risk than the ones with high expected consequence but low frequency. However, the events and scenarios with high expected consequence and low frequency are still relevant and can't be discarded without an initial risk study.

The methodology designed uses a quantitative risk assessment approach. However, it is important to remember that the methodology uses a large amount of assumptions that need to be carefully addressed and analysed. The particularities of the project must be taken into account to make the assumptions as objective as possible.

One of the qualitative steps that affects the results of the methodology the most is the definition of the damage limits for life safety, business continuity, environment and reputation. This is a decision that must be made among the stakeholders of the project in the most objective way possible. It is important to keep in mind that the property loss is most of the time the smallest of the concerns compared to the other assets of the enterprise. This shows the importance of properly estimating these assets.

The post incident risk reduction strategies discussed are mitigating measures that reduce the likelihood and the expected consequence of a scenario. These strategies must be designed together with pre incident risk reduction strategies to get a global picture of the risk management in the facility.

The methodology assists the designer in identifying the initiating events which leads to the determination of the pertinent fire scenarios. After, the methodology gives guidelines to calculate the expected consequence and annualised risk. The calculation of the annualised risk allows the identification of the events and scenarios that contribute the most.

The user can then design the fire protection strategy based on the needs and particularities of the facility. Additional to the information obtained from the annualised risk, the designer should consider aspects like location, operation philosophy and availability of resources to design the strategy.

The recalculation of the expected consequence is based on several assumptions and estimations. Good engineering judgement and strong knowledge of the systems are needed to reach a good estimation. A realistic panorama of the expected consequence is obtained when the fire protection systems are in place.

The comparison of the fire protection strategy cost with the expected consequence allows to determine the level of protection feasible for the facility from an economical point of view.

Other viewpoints than the economy are used to determine the level of protection, like individual and societal risk. These tools should be used together to design the fire protection strategy.

This methodology is highly expensive to carry out in terms of time. Previously a qualitative risk assessment methodologies should be performed to ensure that only the most hazardous cases are evaluated using a quantitative risk assessment.

Under these conditions, the methodology can be highly useful if it is properly integrated in the complete design project.

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