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Experiences in planning process changes on the basis of consequence and risk assessment

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Abstract

The procedure used in a quantitative top-down consequence and risk assessment is described and evaluated. Two examples of practical applications of a quantitative consequence and risk assessment are introduced. These studies have been made in the feasibility or preliminary design stage of a new installation. Both cases concern the storage of liquefied natural gas (LNG).

1. Introduction

Safety analysis can be used as a part of the design process as early as the first stages of a feasibility study. In the development and process design engineering stage, the safety of a plant as a whole is considered to ensure that the process is, as far as possible, inherently safe and that the system can cope with all abnormal operations, material release, etc. [1]. Furthermore, the suitability of potential plant locations can be screened by carrying out preliminary consequence analyses and then comparing different siting options and technological alternatives.

Once the number of plant items and their capacities has been established, it is possible to consider the layout of the total plant. Detailed planning of the layout is not the responsibility of the process designer, but early discussions must be made on the siting of the plant. The plant must fit on the designed site, and the acceptability or tolerability of the potential hazards to the public and environment has to be considered [1]. This requires identification of the hazards and assessment of the consequences of major releases. For example, there can be a number of different storage options in the same process area,

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and an evaluation of their safety can be made by assessing the consequences and/or risks of each option.

This paper begins with a short overview of chemical risk assessment and its limitations. The use of safety analysis in siting and layout design is also discussed. Experiences in planning process changes on the basis of consequence and risk assessment are highlighted with two example cases.

2. Assessment of chemical risks

2.1 Common procedure in risk assessment

In carrying out risk assessment of a chemical process plant containing toxic and/or flammable gases it is necessary to:

(1) Identify and subsequently quantify the potential hazardous events, e.g.

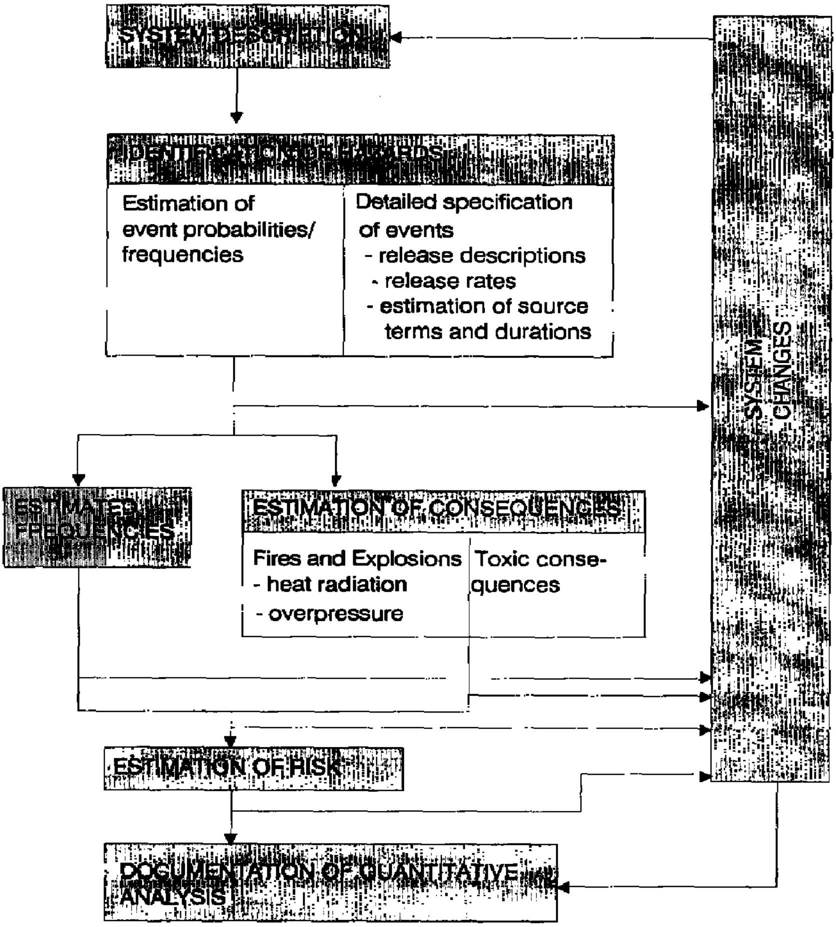

Fig. 1. Main phases of safety analysis on chemical installations and their connections with system changes.

release modes and source terms, and frequencies of events leading to identified hazards, and the durations of releases.
(2) Determine the behaviour of the material after its release, e.g. gas concentrations versus distance from release point.
(3) Calculate the risks of the release cases. Apply toxicity data, information about ignition points and ignition probabilities, and take account of the wind and weather data. Sum up the risks of significant releases separately for the toxic and the flammable cases.

Figure 1 shows the main stages in assessing the potential consequences and risks of toxic and/or flammable gases. The decision as to the level at which the analysis is stopped is made on the basis of the complexity of the activity for analysis and the potential risk. The objective of a consequence analysis is to quantify the harmful impacts of potential events. In consequence analysis, the two main harmful characteristics associated with a gas are its toxicity and/or flammability. Potential consequences of hazardous releases are: exposure to

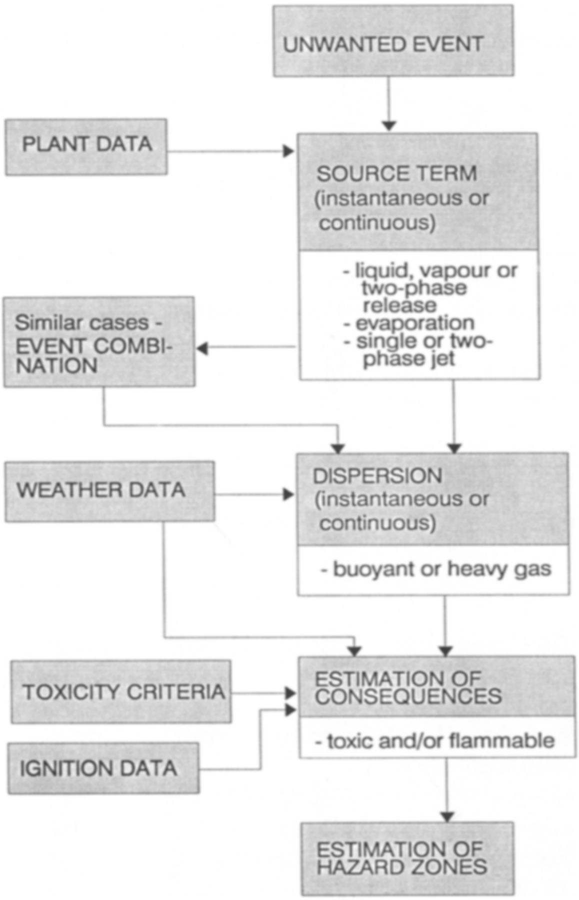

Fig. 2. Main phases of consequence analysis concerning chemical releases.

toxic materials; missiles and overpressure from explosions; and thermal radiation. Such a consequence analysis requires data on chemical, physical and toxic properties of the chemicals, weather statistics and locations of ignition sources. Figure 2 shows the main phases of consequence analysis.

The final result for a given toxic material is an exposure estimate for each event of concern and the different levels of severity of these events. Fire and explosion models convert the information on the cloud for flammable releases into hazard potentials such as thermal radiation and explosion overpressures. Effect models convert these case-specific results into effects on people (injury or death) and structures. Additional refinement is provided by mitigation factors, such as sheltering or evacuation, which tend to reduce the magnitude of potential effects in real accidents.

2.2 Criticism of consequence analysis

A substantial amount of research has been conducted to reduce the uncertainty which exists in many of the models applied to consequence analysis. Nevertheless, because much is still unknown about the phenomena behind the

TABLE 1

A summary on the main problems of the models in consequence analysis [3–8]

Phase of the consequence analysis	Deficiencies and restrictions
Source term	Only a limited number of spill scenarios have been modelled. There are no models for multicomponent materials Some models have had limited testing and their uncertainties are not well known
Dispersion	Some variation in the results of heavy gas dispersion models Only few models exist for estimation of the effects of topography and obstacles
Estimation of consequences	
A. Toxicity	Great variation in existing toxicity data and in the application of the data
B. Flammable	Unconfined vapour cloud explosions (The effects of obstacles and partial confinement are not well enough understood)
Estimation of hazard zones	Escape of people difficult to include, because of unpredictable behaviour of people in accidents The effects of obstacles or topography cannot be estimated

models and because many of the phenomena are intrinsically random, significant uncertainties remain in most consequence analyses.

In a consequence analysis, an important feature of many of the models should be the ability to extrapolate from the results of relatively small-scale experiments to large-scale accidental releases. All models suffer to some degree from errors arising from extrapolation, but those that are the best from this point of view are the ones that have a sound fundamental basis and the minimum of empirical constants [2]. Table 1 presents a summary of the main deficiencies in models employed in the different phases of consequence analysis.

In order to estimate the final consequences of various accident cases, it is necessary to use exposure–response relationships for determining the degree of harm to people, property or the environment. For flammable substances, such relationships are associated with thermal radiation and/or explosion overpressure, and they have been identified by experiment and/or experience. For toxic substances such relationships cannot readily be identified and the experience of the effects on people and the environment is very limited.

2.3 "Worst case" or "design base cases"

In applying risk assessment techniques to loss prevention in chemical plants, there are several theoretical problems encountered with some of the calculations. One of the major problems is the time variability of a failure case. Another problem is connected with the so-called worst case approach, which is often used in safety analysis. This approach is, in many cases however, questionable because of the extremely low probability of the "worst case". In practice, the event frequency often increases rapidly as the release rate decreases. Often so-called design base accidents are used instead of the worst case approach in order to define the extent of consequences and the types of accidents considered as reasonably possible and to be included when planning risk-reducing measures in a plant.

The respective contributions of frequency and consequence to the total risk can be illustrated with an example concerning a pipeline transferring anhy-

TABLE 2

Source terms and event frequencies of a pipeline release [8]

Length of the pipe (m)	Hole size	Source term (kg/s)	Basic event frequency ($\times 10^{-6}$/m year)	Event frequency ($\times 10^{-6}$/year)
680	D	52	2.5	1 700
	$D/2$	26	3.8	2 580
	$D/4$	11	17.5	11 900

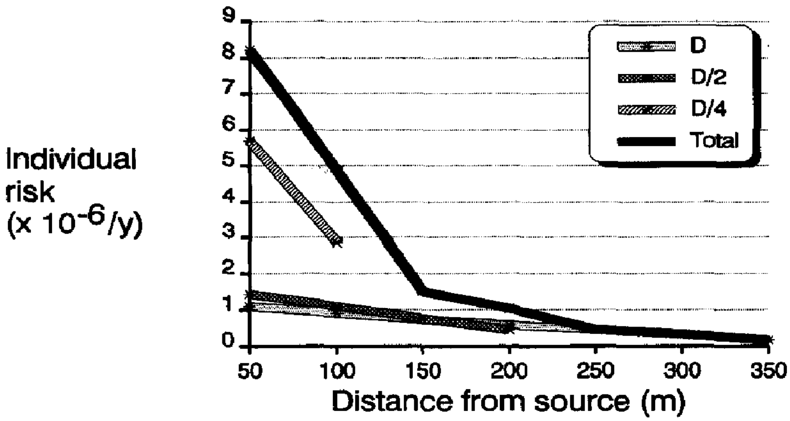

Fig. 3. Individual risk values for different rupture diameters (D, $D/2$, $D/4$) when the pipelength is 680 m [8].

drous ammonia. The pipeline's diameter D is 305 mm and its length 680 m. Ammonia temperature and pressure are, respectively, 20°C and 7.6 bar. Three hole sizes whose estimated frequencies are given in Table 2 are considered as possible ruptures.

The individual risk of a person outdoors receiving a lethal toxic dose was calculated with the program RISKIT [9,10]. A 10 min exposure to a 5,000 ppm concentration was taken as the minimum toxic dose causing death. The example case was calculated for the Pasquill stability category D and wind speed 5 m/s, only. all wind directions were assumed to be equally probably and the isorisk curves were, consequently, concentric circles.

The total individual risk and the contributions of the three hole sizes calculated at 50 m intervals are given in Fig. 3. The curves for the hole sizes $D/4$ and $D/2$ end at 100 m and 200 m, respectively, since the toxic dose attains the minimum value at these distances. It is seen from Fig. 3 that the risk at distances up to 100 m is dominated by the smallest hole size $D/4$ since it has the largest frequency (Table 2). Typically, the high probability–low consequence events dominate the risk close to the plant.

3. Safety analysis in siting and layout design

3.1 The design process

Figure 4 describes the main phases of process engineering design. Process engineering design can be divided into three phases; namely, feasibility, main and detailed studies.

The feasibility study includes the following steps [11]:

(1) An evaluation of the need for a new system or a modification of an existing system.

(2) A definition of the fundamental requirements and boundaries of the designed system.

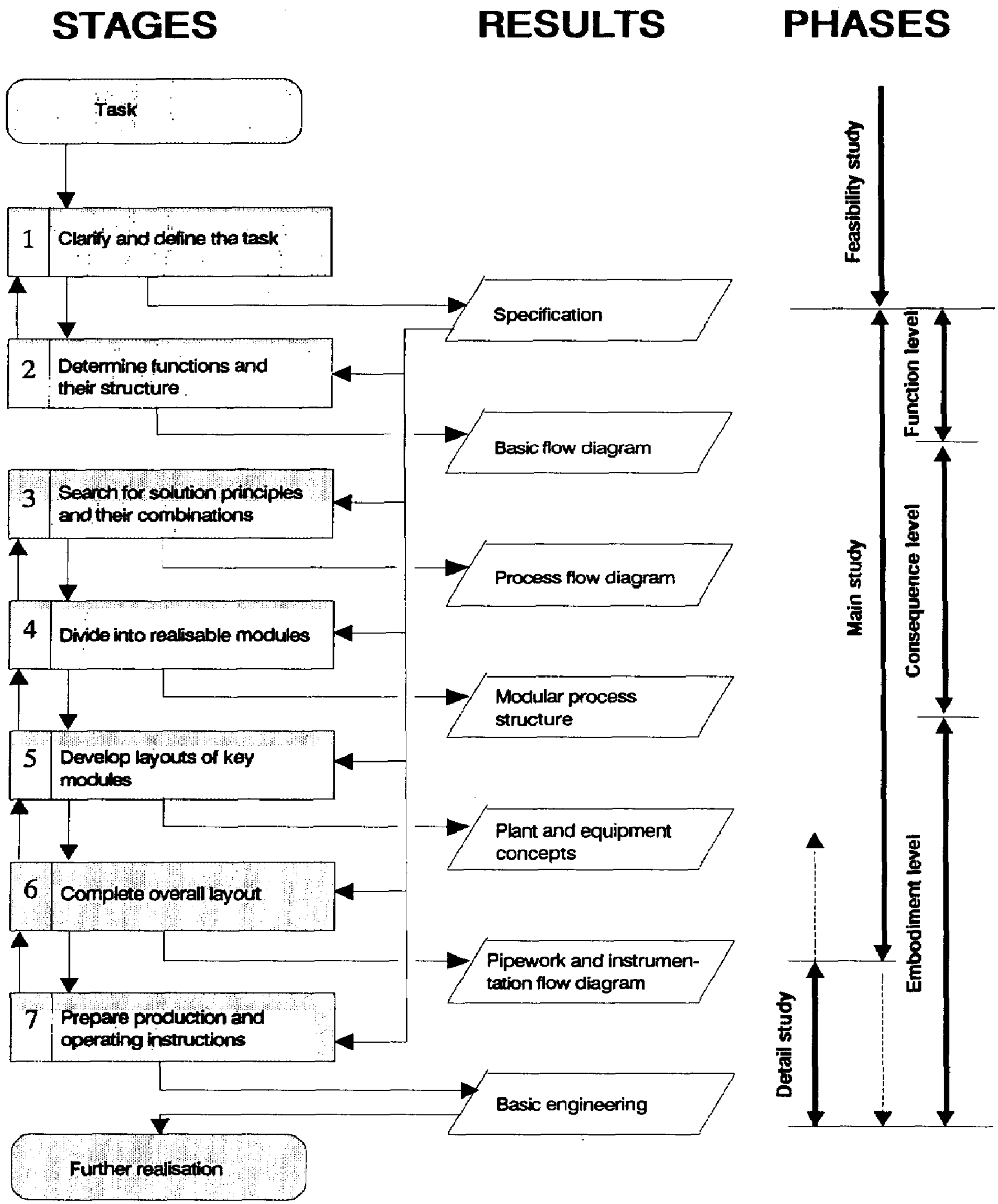

Fig. 4. General approach to design in process engineering [11].

(3) A feasibility study from technical, economical, political, social, psychological and ecological points of view.
(4) A selection of the most promising solution principle(s).

The feasibility study also determines the practicability of the chemical or biological reactions, including the selection of raw materials and an examination of the resulting consequences.

In the main study, attention is concentrated on the whole process engineering system, developing the results of the preliminary study in order to produce an overall concept on which to base the investment decision. In addition, detailed studies are defined in the course of this main study, and priorities are set for their execution. Figure 4 shows the levels of the main study.

In detailed studies, the course of the main study is defined and priorities are set for the execution. The detailed studies may overlap the main study in some cases.

3.2 Consequence assessment in plant design process

Once the number of plant items and their capacities have been established it is possible to consider the layout of the total plant. Detailed planning of layout is not the responsibility of the process designer, but early discussions must be made on the siting of the plant. The plant must fit on the designed site, and the acceptability or tolerability of the potential hazards to the public and environment has to be considered [1]. This means the identification of hazards and the assessment of consequences of major releases.

In the following, examples are presented of recent consequence analyses for two proposed LNG storage systems. One of the analyses was made during the feasibility study and the other during the main study of the new installation.

4. Study of siting options of an LNG storage

4.1 Aim of the study and description of the system

The aim of the study was to investigate the safety of different siting options, to support both siting design and decision-making during the feasibility study phase. The safety of the storage options was evaluated by investigating the possible consequences and probabilities of hazardous events which could cause harm to people and the environment. The three LNG storage options in the same process area were the following:

Option 1: A double-wall steel tank in a rock excavation, $\frac{1}{3}$ of the tank is above the ground level. The storage tank is filled once a year from a tanker.

Option 2: A tanker moored to a dock, LNG is pumped with ship pumps to vaporizers on land.

Option 3: A full containment tank with a concrete outer tank. The storage tank is filled once a year from a tanker.

4.2 Estimation of event probabilities

Table 3 presents a summary of the estimated event probabilities of each siting option. The probabilities presented by Baker [12,13], Atallah [14] and Pelto [15] were adjusted by using fault tree analysis and site-specific data on environmental and operating conditions to estimate frequencies of significant accidental events. It was found that the event probabilities were small except in the case of tank underpressurization. Option 3 seems to be much better protected against external causes because of the concrete outer tank.

Data on event probabilities of an LNG ship used as a storage ship in arctic conditions were not found in the literature.

TABLE 3

Estimated probabilities of possible hazardous events of Options 1 and 3

Event	Comments and references	Estimated event probabilities/year
1. Rupture of unloading arm during unloading	The estimated unloading time is 30 hours/year In Baker [12] the unloading time was 1500 hours/year	4×10^{-7}
2. Rupture of main transfer line during unloading	See comments of Event 1	3×10^{-8}
3. Catastrophic storage tank failure	A probability estimate was found for Option 1 only. The tank is better protected against external causes than those considered by Baker [12,13] and Atallah [14]. (The event probability of fire is estimated to be 3×10^{-6}/year)	1×10^{-5}–1×10^{-6}
4. Storage tank is overfilled	Overfilling of the storage tank when safety systems do not work In this case the storage tank is filled only once a year In Baker [12] the tank was filled 100 time/year	2×10^{-5}
5. Storage tank is overpressurized	See comments of Event 4	1×10^{-6}
6. Storage tank is underpressurized	Probability estimate is the same as in Baker [12]	2×10^{-3}
7. Break of the feedline of a tank	See comments of Event 1	8×10^{-10}
8. Break of the unloading line of a tank	The unloading line is used 24 days/year In Baker [12] the unloading line was used continuously	3×10^{-8}

4.3 Estimation of consequences

In the consequence analysis, the hazard ranges of large spills were estimated; LNG was assumed to be spilled into the sea from a fractured or punctured ship tank or a leaking unloading arm or pipeline. The spill is assumed to form a semicircular evaporating pool. For the heat flux from the sea to the pool, the two extreme values given in the literature (25 kW/m^2 and 100 kW/m^2) were used [16].

If the spill is not ignited immediately, a flammable vapour cloud will be formed. The length and width of the flammable cloud was calculated with the program, RISKIT for the weather category Pasquill D–5 m/s, only. The dimensions of the flammable cloud are indicative of the area affected by the resulting flash fire. The possible semiconfined spaces and obstacle configurations leading to a vapour cloud explosion were not identified.

If ignited, the fire is assumed to flash back to the pool. For this case, the ranges of thermal radiation of the pool fire to selected flux values were calculated. The most probable large pool fire events considered were:

Option 1: A spill into the rock excavation due to tank failure or overfilling or into the sea during tanker unloading.

Option 2: A spill into the sea.

Option 3: A spill into the sea during tanker unloading.

The following experimental values were used for the surface emissive power of a LNG poolfire [17]:

- pool on ground: 153 kW/m^2
- pool on water: 203 kW/m^2

The following flux values were used to indicate the effects of thermal radiation [18,19]:

- 38 kW/m^2: damage to process equipment
- 12 kW/m^2: ignition to vegetation
- 6 kW/m^2: second degree burns after 20–60 s exposure
- 3 kW/m^2: emergency work possible
- 1.5 kW/m^2: safe evacuation distance.

A summary of the hazard ranges is given in Table 4 and drawn on the schematic layout in Figure 5. For those ranges that depend on the heat flux from the sea (25 or 100 kW/m^2) only the value that gives the larger range is given. It is seen from Fig. 5 that a punctured or fractured ship tank (Option 2) leads to the largest hazard ranges. A spill during tanker unloading (Options 1 and 3) leads to hazard ranges that are only about one fifth of those of Option 2. A spill from the storage tank was considered possible for the double-wall steel tank (Option 1) but remote for the full containment tank (Option 3).

This study supported the designers and decision-makers by providing information about the hazards and their eventual causes and consequences in each three options under certain specified conditions. It also assisted in finding effective means of reducing risks and provided a structured view of the main

TABLE 4

Summary of the results connected with each storage option

Case	Pool fire hazard ranges
Option 1	
Spill from unloading arm or pipeline •LNG flows to the sea → pool fire •source term 2,500 $m^3/h \approx 300$ kg/s •the length of the flammable cloud is 200–300 m	130 m, ignition of vegetation (12 kW/m^2) 190m, slight second degree burns, 20–60 s (6 kW/m^2) 275 m, emergency work possible (3 kW/m^2) 390 m, safe evacuation distance (1.5 kW/m^2)
Storage tank is overfilled/catastrophic storage tank failure •LNG flows to the rock excavation → in worst case the whole LNG tank ignites	130 m, ignition of vegetation (12 kW/m^2) 190 m, slight second degree burns, 20–60 s (6 kW/m^2) 270 m, emergency work possible (3 kW/m^2) 390 m, safe evacuation distance (1.5 kW/m^2)
Option 2	
Spill from unloading arm or pipeline •LNG probably flows to the sea (pool on land not considered) •source term 220 $m^3/h \approx 26$ kg/s •the length of the flammable cloud is 60–80 m	35 m, ignition of vegetation (12 kW/m^2) 50 m, slight second degree burns, 20–60 s (6 kW/m^2) 75 m, emergency work possible (3 kW/m^2) 110 m, safe evacuation distance (1.5 kW/m^2)
Failure in one of the tanks in LNG ship •half tank content (=12 500 m^3) flows to the sea in few minutes •the whole amount of spilled liquid evaporates in 5–10 minutes •the average source strength is 9,400–16 400 kg/s •estimated fire duration is 5 min •the length of the flammable cloud is 1,400–2,100 m	680 m, ignition of vegetation (12 kW/m^2) 1,000 m, slight second degree burns, 20–60 s (6 kW/m^2) 1,450 m, emergency work possible (3 kW/m^2) 2,000 m, safe evacuation distance (1.5 kW/m^2)
Option 3	
Spill from unloading arm or pipeline •see Option 1 Catastrophic inner storage tank failure •this case is remote	

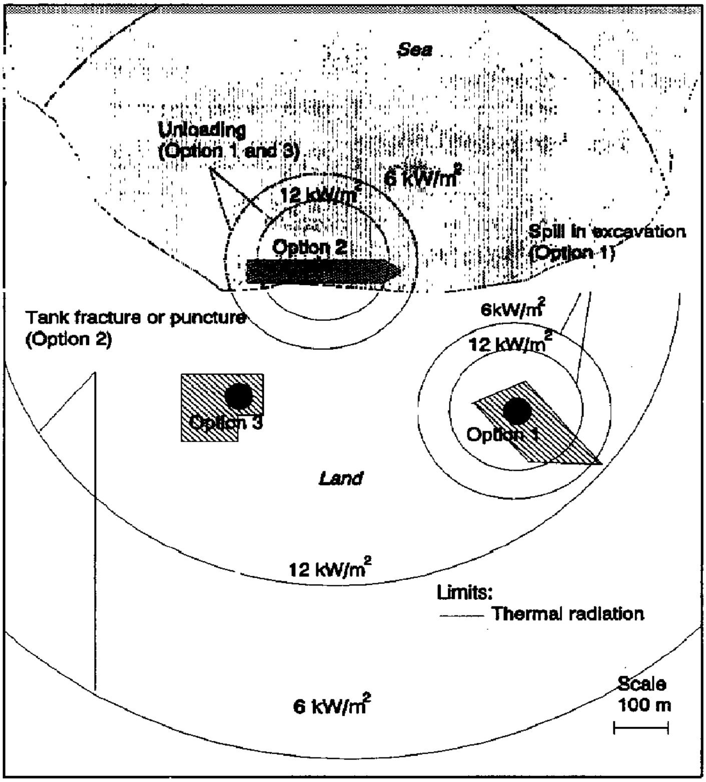

Fig. 5. Consequences connected with different options. Consequences of Option 3 are connected only with unloading of the ship.

hazards. In this study, each option had similar consequences but considerable differences in the hazard ranges were observed (Fig. 5). The results of the analysis produced a more sound basis for making decisions between the different options and for planning further actions to be taken.

5. Study of a storage system in a rock cavern

5.1 The aim and content of the study

The aim of this study was to investigate the safety of a proposed LNG storage system to be designed in the main study phase (consequence level in Fig. 4) in a rock cavern. This analysis was based on a planned system consisting of unloading facilities in harbour, pipelines and pumps, rock cavern storage system and vaporizers. The analysis included events that may occur to an LNG tanker approaching the harbour. This study concentrated on modelling and estimation of consequences.

5.2 Identified hazards and their consequences

LNG spills into the sea could occur during sea transport and during unloading. A puncture or fracture of the ship tank while the ship is approaching the

TABLE 5

Summary of the consequences in different cases

Description of object	Summary of the identified consequences

Sea transport
Failure in one of the tanks ($25\,000$ m^3) in the LNG ship; leakage does not ignite near the ship – formation of a semicircular evaporating pool:
- estimated evaporation rate is 16 500 kg/s and
- radius fo the pool 460 m

Flammable cloud

Weather case	Length of flammable cloud (km)	Cloud area (km^2)
D–5 m/s	1.3–1.9	1.3
E–2 m/s	4.9–8.5	42.5

Pool fire

Thermal load (kW/m^2)	38	12	6	3	1.5
Fore-and-aft (m)	360	775	1125	1600	2200
On the beam (m)	560	1125	1600	2200	3000

Unloading the ship
Leakage on the sea – formation of rectangular pool:
- average unloading rate is 10 500 $m^3/h \approx 1\,240$ kg/s
- size of the pool is 70×360 m

Flammable cloud

Weather case	Length of flammable cloud (km)	Cloud area (km^2)
D–5 m/s	0.2–0.3	0.014
E–2 m/s	1.2–1.9	2.1

Pool fire

Thermal load (kW/m^2)	38	12	6	3	1.5
Range (m)	135	265	380	525	730

LNG storage
Small spill in pipeline shaft – possible explosion in the cavern:
- assumed spill rate is 1 kg/s

Confined explosion

Damage type	Heavy (0.30 bar)	Repairable (0.10 bar)	Damage of glass (0.03 bar)	Crack of windows (0.01 bar)
Range (m)	23–45	72–140	135–265	370–720

TABLE 5 (continued)

Description of object	Summary of the identified consequences					
Pump area Spill on the diked area – pool formation to embankment on land: •flow to vaporizers is 220 $m^3/h \approx 26$ kg/s •circular pool with radius 9 m	*Flammable cloud*					
	Weather case	Length of flammable cloud (km^2)			Cloud area (km^2)	
	D–5 m/s	0.03–0.04			0.0007	
	E–2 m/s	0.13–0.21			0.03	
	Pool fire					
	Thermal load (kW/m^2)	38	12	6	3	1.5
	Range (m)	15	40	55	80	115
– formation of semicircular pool on sea: •flow to vaporizers is 220 $m^3/h \approx 26$ kg/s •semicircular pool with radius 10 m	*Flammable cloud*					
	Weather case	Length of flammable cloud (km^2)			Cloud area (km^2)	
	D–5 m/s	0.02–0.03			0.0009	
	E–2 m/s	0.13–0.20			0.024	
	Pool fire					
	Thermal load (kW/m^2)	38	12	6	3	1.5
	Range (m)	25	45	65	90	130

harbour would form a semicircular spreading and evaporating pool alongside the ship. A spill during unloading would form a rectangular pool between the ship and the shore. If not ignited immediately, the vapour will form a large flammable cloud.

Consequences of a spill on the ground in the diked area surrounding the LNG pumps were estimated. Another interesting spill was one that would occur in the underground part of the pipeline leading from the cavern to the vaporizers. The maximum length and area of the flammable cloud were calculated with the program RISKIT for two weather categories, Pasquill D–5 m/s and E–2 m/s. The results of the calculation are given in Table 5. Considering the enormous maximum size of the flammable cloud resulting from tank puncture or fracture it is very probable that the cloud will ignite while it is still

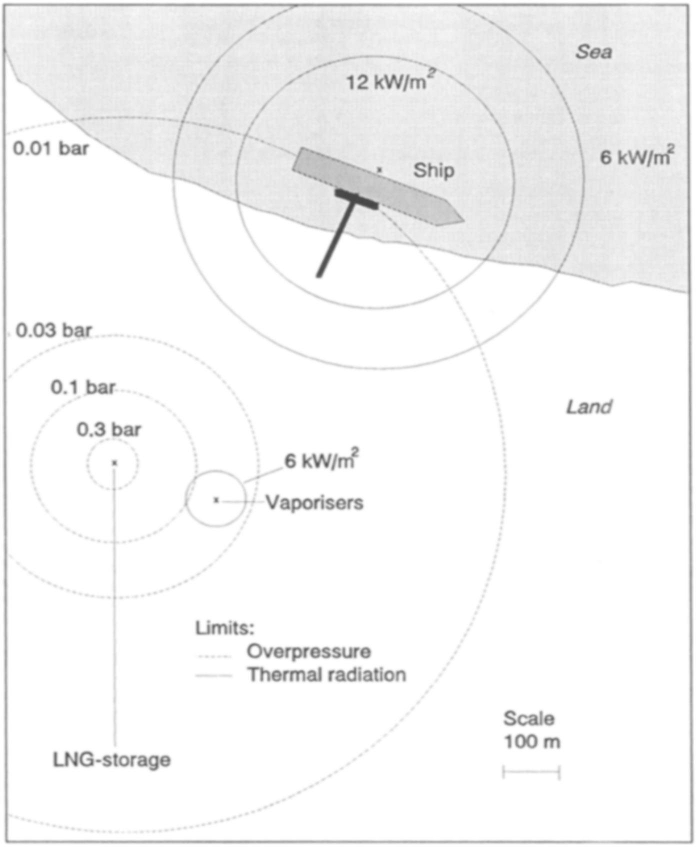

Fig. 6. Consequences of a cavern storage system.

spreading and thus the area affected by the flash fire will be much smaller than the maximum area given in Table 5.

A flammable cloud formed while the ship is unloading could drift towards the process area. There are many potential ignition sources in the area and the ignition would result in a flash fire. It is also possible that the flammable mixture would enter a building and ignite inside, causing a confined explosion. If the cloud is ignited, the fire will flash back to the pool. Ranges ~~of~~ selected values of the thermal flux calculated for the pool fire are given in Table 5.

If there is a leak in the underground part of the pipeline leading from the storage cavern to the vaporizers, LNG will be spilled in the pipeline shaft. Even a relatively small spill rate (say 1 kg/s) may fill part of the shaft with flammable mixture creating the prerequisites for a confined explosion. The explosion would be vented through the shaft entrance. On the basis of large-scale experiments reported by Pappas [20] the peak overpressure in the shaft could be up to 1 bar. This value was used to calculate the hazard ranges of overpressure outside (Table 5 and Fig. 6). The overpressure could cause severe damage to the storage facility and less damage to other process equipment and buildings. People working in the office building could receive injuries from flying glass fragments.

This study improved the company's awareness of the hazards, and their causes and consequences. It also provided a structured view of the main hazards to be taken into account when deciding whether, and under what conditions, to continue with the investment project.

6. Conclusions

Consequence analysis can reveal effective measures to reduce the risk. The sensitivity of the results to inventory size, storage temperature, storage pressure and material properties can point the way to inherently safer plants. Furthermore, consequence analysis can be a very significant and useful tool when applied, for example, to the design and siting of new plants, preparation of emergency plans, and control of releases during normal operation.

Consequence and risk analyses are increasingly employed in the chemical process industries to ensure the safety of new installations. Consequences and risk estimates are valuable aids in planning and engineering decision-making. They are gradually becoming the "norm" and it is probably only a matter of time before they are fully accepted in the public forum of regulation and legislation. This, however, requires improvements in the accuracy of the quantitative results of consequence and risk assessment. The quality problems mainly involve the level of the identification and consequence analysis, and the assessment of their effects. Areas in which risk analysis is being applied are rapidly expanding and there is reason to believe that this will continue for some time to come.

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