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MANAGEMENT OF LARGE LNG HAZARDS

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ABSTRACT

This paper describes the background to the Hazard and Effects Management Process that is routinely applied by Shell when assessing major fire and explosion hazards that could potentially occur in LNG plant operations. The heart of the process relies on identification and assessment of potential accident scenarios that can harm people or damage plant. A hierarchy of assessment tools and knowledge are available, and in constant process of improvement, to assess design events and worst cases. A review of the capabilities and limitations of these tools and techniques is discussed in the context of large LNG hazards. The objective is to design plant so that major LNG hazards such as large fires, explosions and escalation of these events are considered and minimized.

The focus is on large LNG hazards that might arise from major loss of containment. The paper also refers to renewed concerns over the proposed installation of LNG import terminals in the USA.

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1. INTRODUCTION

Liquefied Natural Gas (LNG) is an important energy source that contributes to energy security and diversity. LNG as with all hydrocarbons needs to be handled with the appropriate safety measures. The LNG industry, with Shell in a leading role, has spent many millions of dollars understanding the implications of HSE issues in LNG and has produced codes, standards and guidelines for its safe handling. Indeed, it must be taken into account that LNG boasts an impressive track record, having never experienced a significant loss of an LNG tank cargo.

LNG is cold, boiling at –161^{o}C, and its vapours are flammable. Special low temperature metals and materials are used during liquefaction, transfer and storage. Industry best practice is used to design LNG plants. LNG is usually stored as a liquid at atmospheric pressure in special steel inner tanks with outer concrete tanks with no bottom connections. Pressurised storage and transport is not used so the dangers of catastrophic pressure vessel failure, such as boiling liquid expanding vapour explosions (BLEVES), are avoided. The release and subsequent dispersion, and fire and explosion hazards have received special attention through extensive safety research programmes and computer modelling. These topics are discussed here.

Natural gas is less reactive than other fuels and potential plant explosions (deflagrations) are less severe than with, say, hydrogen, propane or ethylene. Detonations of natural gas/air mixtures are not considered to be credible design accidental events, because of the restrictive conditions under which they could happen.

Appropriate measures are taken to control and mitigate the consequences to protect personnel and assets, through application of Safety Management Systems. At the heart of a good management system is a Hazard and Effects Management Process (HEMP). This should be applied throughout the lifecycle of a project, from concept to decommissioning. The general form of HEMP is to identify the hazards, assess their likelihood and consequences, control the process effectively with instrument and material protection and recover from any loss of containment with the minimum of consequences. The effectiveness of risk reduction measures has also received considerable attention and entered into HEMP. This allows the residual risk to be benchmarked against risk tolerability criteria, or to be minimised by considering alternative strategies until the risks are tolerable or As Low As Reasonably Practicable (ALARP). A hierarchy of engineering experience and methods are used throughout this decision making process:

- Codes and standards
- Assurance reviews
- Hazard Identification (HAZID) and Hazard and Operability (HAZOP) studies
- Physical effects modelling (consequence analysis)
- Quantitative Risk Assessment (QRA)
- Cost benefit analysis of risk reduction options

The HEMP can also be directly linked to operational roles and responsibilities, training and other human factors during operation and maintenance of the plant.

One of the key safety drivers is layout of the plant. This is most cost-effectively optimised during the early design stage of a project. Unfortunately this coincides with the stage when detailed information is lacking, so special techniques are applied to predict major accident scenarios and to devise conservative rule sets for avoiding them. Increasingly sophisticated, and less conservative, consequence models can then be applied as the level of detail or safety criticality increases. The paper discusses the application of this hierarchical approach.

A major objective behind optimised plant layout is to avoid escalation of fires and explosions to protect people, assets and reputation, using the principle of inherent safety. Some examples are good separation between hazardous and vulnerable areas, and minimise: equipment, liquid hydrocarbon inventory in process equipment, vulnerability through selection of equipment type, and exposure to people through reduced process complexity and maintenance.

With the continued expansion of the LNG business and ever-more sophisticated operations, the safety engineer is faced with a continuous challenge to meet the demands of tolerably safe design. The application of safety research via interdisciplinary teams plays an important role in providing robust solutions to manage LNG risks.

This paper will present several examples of applications of these techniques and new safety-related research to the design and construction of LNG plants, regas terminals and floating LNG plants. In particular, the paper will refer to techniques which can be used to demonstrate how the design of blast protection can be optimised, the importance of natural ventilation in reducing risk, how the separation between units can be specified to tolerable levels of risk, and how to find the best placement of occupied buildings on onshore processing facilities.

2. LNG MAJOR HAZARDS

An essential part of the HEMP procedure is to identify and quantify the physical effects of potential accident scenarios. Once the effect distances are known then safe separation distances and/or cost effective mitigation systems can be defined. A hierarchy of predictive tools are brought to bear on the identified scenarios with increasing sophistication and less conservatism depending on the tolerability of risk to the effects. Empirical models which employ the essential physics tend to be formulated in a conservative way to enable screening of the worst case and design events, and indeed have been used extensively with some success. Integral or Computational Fluid Dynamics (CFD) models tend to be less conservative but require longer computer times and are less useful for screening, but can be valuable if the hazards predicted by empirical models become critically important for safety. Nevertheless it is important that validation for any model is carried out by experiment or large scale test to check the physics and provide confidence in sensitivity studies. We review here some of the capabilities and shortcomings of these consequence models.

An accidental release of LNG results in a variety of consequences. Outflow can lead to spreading, boiling, evaporation and dispersion of flammable vapours in the absence of ignition. Delayed ignition might lead to explosion if the vapours disperse into congested or confined areas, followed by burn back (sometimes called a flash- or cloud fire) to the site of the original release and a sustained pool fire, depending on the pool area and release rate. Early ignition would create a pool fire but no vapour cloud explosion. Existing knowledge rules out the formation of a rising fireball, as in an LPG BLEVE, because of the relatively shallow nature of the dispersing cloud and lack of significant upward momentum. For a comprehensive overview the reader is referred to [1] and [2].

- **Pool Spread and Evaporation**

Pool spread and evaporation of LNG define the source terms for subsequent fire and explosion, yet there is still plenty of scope for improved prediction. If the LNG spreading is on land the type of substrate will control the rate of boiling and evaporation after an initial film-boiling phase. Empirical models for boiling and evaporation are available, but they have yet to be tested at large scale. Of critical importance is the limited experimental evidence that shows greatly enhanced heat transfer from porous substrates, such as sand and gravel, where the boiling regime allows the liquid to permeate. This is further complicated in the case of LNG by enrichment in ethane within the substrate as the liquid ages, and other effects such as foaming that also depend on composition [3].

A recent study of the effect of composition on the behaviour of LNG on water described the experimental evidence on pool spreading on water as scant [3,4,5]. Note that boiling can initially be particularly violent resulting in liquid ejection, and a greater interaction with the surrounding air. The following relationship appears to be generally accepted to describe the gravitationally driven rate of pool spread and is sourced from many authors [for example 6,7,8].

$$\frac{dR}{dt} = 1.64 \frac{1}{R} \sqrt{\frac{Mg(\rho_w - \rho)}{\pi \rho \rho_w}}$$

Where R is the radius of the pool, ρ is the density of LNG, subscript w refers to water, and M is the mass of liquid in the pool. The report sponsored by the Federal Energy Regulatory Commission (FERC) into LNG safety [2] recommends use of the Webber model [7] because it has a more sound theoretical basis, and accounts for friction effects, compared to simple gravitational models. The model is described in the TNO “Yellow Book” 1997 [6] as implemented in the ‘Gas Accumulation over Spreading Pools’ (GASP) computer model. An alternative source is the MathCAD pseudo-code listed in the FERC report [2].

Spread and evaporation on water has, with the exception of the work by Quest [9], ignored the effect of waves and currents. It is not known even if the waves would increase or decrease the total evaporation rate. There has been some debate about the formation of ice and/or hydrates when LNG is spilled. In general it is thought that ice only forms if the water is quite shallow, so is not included in the model. ABS report [2] a body of opinion that ice will not form with such releases. Hydrate is not mentioned. For the Shell Maplin Sands spills [10], significant quantities of ice and/or hydrate were formed. The suggestion that this would not happen in deeper water may be correct, but it needs to be confirmed as ice/hydrate formation could have a significant effect on pool spread.

- **Rapid Phase Transitions**

Also there may be Rapid Phase Transition events (RPTs) which can also be quite violent and physically eject liquid. RPTs occurred for large LNG releases at Maplin Sands, although this may be only due to the fact that the LNG had “aged” before release, i.e. there was time for methane to boil off, increasing the concentration of ethane and higher hydrocarbons. If RPTs do happen, a large amount of LNG will be propelled up into the air, evaporating as it falls back down. The source of vapour would be very different from that emanating from a gently spreading pool, with potentially major effects on the subsequent dispersion. In such an event, validating models of gentle pool spread and evaporation by large scale experiments might not be possible or even relevant.

RPTs create blast waves that might cause damage to the nearby surroundings. However the indications are that there should be no major problem associated with single RPT events for either a steel or concrete LNG vessel hull. As an example, consider the rare scenario of spillage of LNG caused by loading arm breakaway, pumping rate 10,000 m^3/hr, followed by 30 second manual reaction time before initiation of a 30 second shutdown period. The total spill volume is then about 150 m^3. The spill occurs between an floating production unit (FPSO) and LNG bulk carrier separated by 6m. The estimated blast energy density onto the surface of the hull is calculated to be $60 kJ/m^3$, sufficient to deform the hull but not to rupture it. There are of course considerable uncertainties associated with this calculation, but it is based on conservative blast and response assumptions.

- **Dispersion**

Even for a known rate of evaporation from a liquid pool, there are uncertainties in current dispersion models. They usually model the sea as equivalent to a land surface of low roughness; it is known that the marine boundary layer is different from this. There are also potentially strongly non-uniform temperature gradients over the sea near land, which can affect dispersion significantly. The dispersion process itself is governed by gravity slumping but the effects of humidity and heat transfer from the underlying surface are important. It is unlikely that the flammable cloud, i.e. distance up to the Lower Flammable Limit (LFL), will become buoyant before the LFL is reached and in test cases of release over the sea [10] the LFL was contained within the visible cloud. For releases below water, however, heat transfer to the evolving vapour did make the plume buoyant before the LFL was reached. The largest tests done so far however have only been up to about 100 kg/s continuous [11] or $20m^3$

instantaneous [10], and appropriate scaling rules should be considered when extrapolating models to larger releases. CFD modelling is proceeding apace and should be watched closely for new developments. For example fairly good results were obtained [12] for dispersion distances with the FLACS code but only when the rate of evaporation was known.

An alternative development is the application of “random walk” of particles to represent dispersion, described by Chynoweth [13]. The equations of Brownian motion are used to calculate the motions of particles, representing pockets of vapour, in a flow field modified by the wind velocity, and appropriate turbulence intensity. For pressurised releases, the initial momentum of the particles can also be added. The technique is less computationally demanding than CFD and appears to be on a par with the accuracy obtained. Although further validation will be required, the technique implemented in a code called DICE (Dispersion In Congested Environments) has already found application where many dispersion scenarios are required, such as in calculation of the source cloud sizes for explosion modelling and explosion exceedance [14].

- **Pool Fires**

The thermal radiation hazard from a pool fire is largely determined by the visible flame size and flame brightness. Both these characteristics vary with pool diameter, but not in a simple way. A typical pool fire of small or intermediate diameter burns, on average, as a cylinder tilted over by the wind with a periodic necking as large turbulent eddies entrain large quantities of air before the visible flame tip. This type of fire is referred to as an “intermittent” fire owing to the periodic nature of its upper structure. As the pool fire diameter increases there is a change in flame behaviour when necking no longer occurs and air entrainment occurs from above via convection cells. This type of fire is called a “mass” or “percolating” fire. (Forest fires are extreme examples of this type of fire.)

From the series of LNG pool fire tests carried out thus far, the flame brightness, or Surface Emissive Power (SEP), appears to have reached a plateau, about 170 kW/m^2 for the largest diameter tested so far (35m Montoir tests [15]). There is uncertainty how the SEP might vary thereafter for bigger pools, but there is strong theoretical evidence that there would be no further increase in SEP and indeed the SEP might start to decrease due to the shielding by soot from incomplete combustion. The limit seems to be about 20 kW/m^2 typical of hot smoke. Pool fires larger than 35m diameter would be required to confirm this trend.

The flame height depends on how well the flame can entrain air for combustion. This in turn depends on the upward momentum and buoyancy of the upward flow of fuel and hot combustion gases. A characteristic non-dimensional grouping, the fire Froude number, is often used to quantify the relative magnitude of these effects [16]. However a more common characteristic grouping, based on the square root of the Froude number, is often used to quantify the different burning regimes. This is Q* which given by,

$$Q^* = \frac{Q}{\rho_a C_p T_a D^2 \sqrt{gD}}$$

where Q is the theoretical heat release rate (m ΔH, where m is the mass rate of burning, ΔH is the heat of combustion per unit mass of fuel), ρa is the density of air, Ta is the ambient temperature, Cp is the heat capacity of air, D is the pool diameter, and g is the acceleration due to gravity.

For methane, the flame heights L for the different burning regimes are given by,

Intermittent fires, $\frac{L}{D} = 2.6 Q^{*2/3}$ 0.23 < Q* < 1.9

Mass fires, $\frac{L}{D} = 18.8Q^{*2}$ 0.23 > Q*

We now have the scaling rules that determine the transition from one type of fire to another. If a pool fire test can be designed large enough to cross this regime then we will have sufficient information to model any size of pool fire. Calculations however suggest that the mass fire regime will not be reached even for extremely large spill sizes. A large LNG spill therefore is predicted to burn in the intermittent fire regime.

The burning rate, m, is however, somewhat uncertain for large LNG spills on water. In the 35m diameter Montoir tests, the LNG burning rate was found to be 0.14 $kg/m^2/s$ or 0.00035 m/s, given the density of LNG is 411 kg/m^3. However this test series was on land. Smaller LNG pool fires on land had burning rates between 0.00021 and 0.00024 m/s. The largest LNG pool fire experiments on water were conducted by Raj, Mudan and Moussa (RMM) [17]. They report burning rates of 0.0004 to 0.001 m/s. This range is larger than the values measured for land-based pool fires. A sensitivity study was therefore carried out by varying the burning rate between from 0.00035 to 0.001 m/s. The same conclusion was reached i.e. a large LNG spill is predicted to burn in the intermittent fire regime. At the highest burning rate we found that the Q* exceeded the bounds of the intermittent fire and moved into a regime typical of turbulent jet fires. It seems unlikely therefore that this high burning rate would be sustained throughout the fire duration and probably relates more to initial measurements of burning rate when the pool was boiling vigorously.

For physical effect calculations a value must be assigned to the burning rate, which could be derived from large experiments. However, given that the mass burning rate is 0.14 $kg/m^2/s$ on land and that the steady evaporation rate for non-burning pools is around 0.08 $kg/m^2/s$, the burning rate on water is likely to be around 0.22 $kg/m^2/s$ or about 0.0005 m/s. This is within the range reported by RMM.

- **Vapour Cloud Explosion**

Explosion modelling, particularly for methane air mixtures has improved greatly, from the days when the Shell Maplin tests in 1980 demonstrated that large methane clouds in the open would not explode [18]. It is now understood that a vapour cloud in congested or confined regions of plant is necessary for the flame acceleration mechanisms to become established for the generation of blast. Empirical models, such as CAM [19] and the Multi-Energy Method, and CFD models such as EXSIM, FLACS and AutoReaGas have been well validated over typical ranges of accident scenarios. Some uncertainties remain concerning flame front/blast wave interactions, fuel/air mixing during the combustion and the sub-grid modelling of turbulent combustion, so research continues. For a review, see Bull [20]. Nevertheless many informed decisions can now be made concerning the blast loading to personnel and plant assets.

In particular the technique known as explosion exceedance can be applied to investigate the risk distribution from all foreseeable explosion scenarios [14]. This combines the many thousand possible dispersion scenarios predicted by DICE with ignition probabilities to generate a library of possible explosion scenarios, using Monte Carlo techniques. The results can be expressed as the probability of exceeding a given impulse or overpressure at a given location, including occupied buildings on an LNG site. In most cases empirical explosion models are used because of the speed of computation (and lower cost of study) and general lack of 3-D information on plant layout at early design stages or for older plant. In some cases though a full CFD, and less conservative, approach can be used if the safety risk information is deemed critical to the site.

An important consideration for the layout of onshore plant and Floating LNG (FLNG) design is the effect of gaps between process units on explosion severity. A small gap may not be sufficient to slow the flame accelerating through a large gas cloud engulfing two adjacent units and the space between them. Severe explosion hazards may then result which escalate the hazards or exceed the design of hull and nearby structures. A gap of sufficient size could act as an “explosion break”, but the drawback for FLNG is extended hull length and more extensive hook-up operations. It is important to design a gap that is large enough to be safe yet small enough not to impose large cost penalties. This work also demonstrated that it is the escalation potential from blast that governs the gap requirement not that from fires.

TNO carried out several small scale tests to investigate the effect of separation distance between areas of congestion on the severity of explosion hazards. The Project was called RIGOS [21]. An extended series of tests called ERGOS sponsored by Shell are currently being analysed. The data have been used to develop empirical guidelines on the critical gap size. The controlling parameters are thought to be the flame speed on emergence of the flame from the unit in which ignition first occurs (the donor unit), the representative obstacle size, the donor size, and the separation distance between adjacent congestion areas (donor and acceptor units). The donor size and flame speed control the extent of the turbulent region beyond the donor. The scale of the turbulence is related directly to the obstacle size, and the separation distance is related to the decay of turbulence. A full parametric study is underway at present to understand these effects and to validate numerical calculations using EXSIM.

Whilst uncertainties remain, it is clear that the larger the separation distance between modules the smaller is the risk of escalation of explosion severity between modules. The empirical analysis can then combined with an explosion exceedance analysis of all possible explosions that could occur from delayed ignition of flange, valve, and pipework leaks. In one study [22], the 10-4/yr escalation risk level corresponded to a minimum clear gap of about 24 m. Using the same analysis, a clear gap of 47m was required for complete explosion isolation of the units. The same study demonstrated the large benefit in explosion risk reduction, up to two orders of magnitude, by allowing natural ventilation to disperse accidental vapour clouds.

The TNO experiments showed that the presence of pipe bridges between process units where the bridge cross-sectional area is ¼ that of the donor module had no effect in propane and methane experiments. The effect of more powerful explosions is currently being analysed.

3. LNG HAZARDS AT IMPORT TERMINALS

Worldwide there are approximately 40 LNG import terminals. With demand for energy rising faster than production of natural gas in North America, the United States is looking for more ways to diversify its energy supplies. The flurry of proposed terminals (latest count around 55, of which 13 are regulatory approved in the United States, Canada or Mexico) has brought LNG back to attention.

LNG has given rise to concern in the US (most other countries have accepted it as another source of energy) ever since its appearance as an alternative energy source on the world market. The existing four LNG terminals (Everett-1971, Cove Point -1978, Elba Island-1978 and Lake Charles-1982) were constructed at a time when LNG specific consequence models were first developed but large scale tests to validate these models for design spills were still to be performed.

By the time the tests were done, models validated and a LNG specific code (NFPA-59A) published, low gas prices had resulted in a drop of LNG imports and mothballing of equipment, some of the terminals started to operate as peak shavers. With the cause gone, the US public interest in LNG safety diminished as did the reason to follow developments in LNG and the implementation of updated codes of practice.

At the start of this century, gas prices picked up and a renewed interest in LNG imports resulted in a flurry of proposed new terminals. The public reaction to these started at the point where the issue was left in the early eighties of the last century. Only this time the focus was no longer the design accidental spills but worst-case incidents.

As none of the consequence models for dispersion, fire and explosion have been validated at this large scale, wildly different results from studies were published. This is understandable given the difference in maturity of models available and different levels of conservatism applied. As discussed earlier, most models have been validated at the much smaller scale needed for design accidental spills. Most of these models are correlations describing the underlying physics of the relevant processes observed at this smaller scale. It is here where most of the uncertainty lies when extrapolating to larger scale and the tendency is to deliberately over-predict the effect distances. The fact that there are no experimental data on LNG for validation at large scale fails to recognise that parallels can be drawn from the wealth of data on the dispersion, fire and explosion behaviour of similar fluids. However, there is still some scope for performing key demonstration tests to verify any predictions at large scale and if necessary to reduce any model conservatism.

4. CONCLUSIONS

The Hazards and Effects Management Process (HEMP) is a valuable technique for accounting for all foreseeable major hazards, whether these be design accidental events or worst cases. HEMP allows for the application of a hierarchy of modelling tools with increasing sophistication, and correspondingly greater accuracy, depending on safety criticality. Much effort has already gone into experimental validation of design accidental events. However, validation of any modelling tool is key to demonstrating tolerability to safety risk. There remain some uncertainties and possible over-prediction in extrapolating to the larger scale demanded by identification of major loss of containment scenarios.

Demonstration tests at an appropriate scale would be beneficial on the spreading, vapour dispersion and pool fire behaviour of LNG to verify the relevant physics and to validate existing, and if necessary to develop improved mathematical models. LNG spreading and vapour cloud dispersion should be investigated both over land and over water. The influence of porous substrates should also be studied. A large enough scale should be chosen to give confidence to the results by including the important physical processes. Pool fires of sufficient size should be performed to investigate the burning regime and to verify the thermal radiation behaviour. A joint industry project would be an ideal platform to perform these large scale experiments.

Experimental work and model development in the field of natural gas vapour cloud explosions at large scale should be monitored for new relevant developments: in particular new insights on the critical spacing of process units, the effect of pipe bridges, and development in sub-grid turbulent combustion modelling.

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