Fire and Explosion Hazards in the Oil Industry:
50 years of Investigations and Research

Chamberlain G.A.[1, 2, 3]

[1] *Waverton Consultancy Ltd, UK*
[2] *Loughborough University, UK*
[3] *Shell Research Ltd, UK*
g.chamberlain@dsl.pipex.com

ABSTRACT

Fifty years of hazard research by the oil and gas industry has seen major advances in understanding the consequences of loss of containment of flammable fluids. This paper charts the progress made, lessons learnt and attempts to identify where gaps remain. The combination of research on potential jet and pool fires and vapour cloud explosions and response to major accidents has led to the development of correlations and physical models that allow engineers to identify and perform consequence analysis in an informed way. However, there is, as yet, no formal procedure to test the reliability of such models. Nevertheless, the aim of making the world a safer place remains a powerful motivation for such studies.

KEYWORDS: Hazard research, jet fires, pool fires, vapour cloud explosions, detonation, flame stability.

INTRODUCTION

Over the past 50 years a great deal of knowledge has been obtained by investigations of the consequences of potential major fire and explosion hazards that exist in the oil, gas and petrochemical industries. Academics, industrialists, regulators and consultants have collaborated in various ways with the aim of making the world a safer place for workers and the public. Several studies have been prompted by major accidents, such as Piper Alpha [1] and Buncefield [2], but thanks to the foresight of business managers and researchers many studies have anticipated possible hazards and protection and mitigation strategies have evolved. It is in conferences such as the present one that valuable information can be passed on for implementation and further examination. I will try to convey a summary of the main lessons learnt from these studies and attempt to identify where gaps remain.

It is worth reminding ourselves where attitudes to hazardous endeavours lay 50 years ago and then to compare with attitudes today. Then, major accidents were accepted as part of industrial progress, but now potential accidents can be identified and prevented. Nowadays the penalties for accidents are severe and can even bankrupt companies. Support for safety was delegated to the responsibility of safety advisors using prescriptive rules and was regarded as a side issue. Nowadays safety is a core issue involving all staff using goal setting rules.

This is all very well, but if major hazards are not identified nor their consequences understood in terms of severity to people and plant then planning strategies can be lost or mismanaged. It has long been recognised that the scaling rules for combustion phenomena are difficult to assess. Early laboratory studies could not be scaled with confidence to the size of the potential large releases of combustible gases and liquids. Therefore, industries dealing with such materials have sought to

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measure the outcomes by conducting tests at scales somewhat larger than could be sustained in laboratories and approaching the scales of actual accidental releases. The results have been captured in correlations and models describing the basic physics to allow for interpolation and extrapolations to other similar events.

EXAMPLES OF SCALING

An early example of when scaling rules went wrong concerns the stability of jet diffusion flames. Laboratory experiments with Bunsen burners found that a natural gas diffusion flame lifted off the burner and blew itself out when the jet velocity approached Mach 0.2. The result was captured in API RP521 and recommended that for stable refinery flares, without the addition of flame holders and flame retention rings, flare diameters should be sized such that the flow velocity for the largest depressurisation rates should not exceed that value. (In later years the requirement was relaxed to Mach 0.5 and then Mach 0.8 [3].) Under normal low pressure flaring in refineries the result was lazy smoky flames which tended to stabilise in the wake of the flare tip in windy conditions. Thermally induced corrosion of the flare became an accepted problem. The guidance was adopted by the offshore oil and gas industry where thermal radiation from the flare presented a major hazard to personnel and equipment because of the much smaller working area compared to refineries. Consequently, large diameter flare pipes were designed to cope with the enormous depressurisation rates in the event of emergencies and to comply with the flame stability requirements.

Researchers began to question the conditions leading to diffusion flame blow-out. Kalghatgi [4] found that there was a critical burner diameter beyond which the flame would not self-extinguish. He derived a correlation, $U_{jb}/S_o = f(R_H)$, relating blow-out velocity U_{jb} to burner diameter d and laminar burning velocity S_o, where R_H is a Reynolds number based on S_o, the distance along the jet axis where the concentration falls to the stoichiometric limit H, and the fuel kinematic viscosity. The critical diameter for vertical methane flames was 30 mm in still air. Tests by Birch et al. [5] at British Gas confirmed the findings. The important observation was that the burner diameter should be replaced by the expanded jet diameter for sonic flow conditions. The boundary of flame stability turned back on itself at the critical diameter as shown in Fig. 1. The expanded jet created a larger source for air entrainment and the flame could stabilise again at the higher source pressures. The more reactive fuels such as ethylene and hydrogen had much smaller critical diameters. Furthermore, Kalghatgi [6] found that the lifted flame became more stable in windy conditions and when the burner was tilted into the wind.

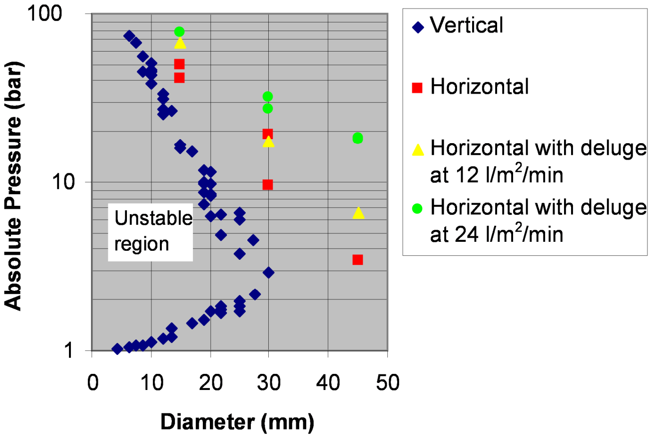


Fig. 1. The stability of natural gas diffusion flames in still air [23].

Kalghatgi concluded that the flame became unstable when the local flow velocity exceeded the local turbulent burning velocity. Other workers [7] have proposed that the criterion is when the local net heat release rate is less than 10 MW/m^3 for propane diffusion flames and 3 MW/m^3 for natural gas. Palacios and Bradley [8] noted the difficulty in predicting diffusion flame blow-off because of strong non-linearities at blow-off involving complexities of mixing with high turbulence, strain rates, flamelet curvatures and localised flame extinctions. Nevertheless, they found that the dimensionless blow-off velocity U_b* correlated approximately with the flame lift-off distance L, and ratio of fuel to air moles for maximum laminar burning velocity, f, as $L/D = U_b^*/10f$.

In summary, high velocity flaring always results in lifted stable flames provided the flare diameter is greater than the critical value. There are many advantages to be gained. Combustion is more efficient than at lower velocities. There is less soot and smoke, thermal radiation is reduced, the flames are stiffer and less prone to be diverted downwards towards personnel and plant by the wind, and corrosion of the flare tip by flame lick is less likely. One drawback is the increased noise as the flow becomes sonic.

Another example of scaling uncertainty involved the consequences of spills of liquified natural gas (LNG). In the early days of bulk transport of LNG by ship concern was raised about a scenario in which the 25,000 m^3 storage was accidentally spilled onto the sea. The extent of the dispersing cloud was estimated between 1.2 and 85 km, and the scale of potential fire and explosion should ignition occur was unknown. It was stated that the combustion energy content of such a release was equivalent to about 100 kilotons of TNT raising fears of catastrophe. To allay these fears and to obtain data on the physical effects, spills up to 20 m^3 were performed at Maplin Sands in the UK in 1980 [9, 10]. The physical models subsequently developed, HEGADAS and HEGABOX [11], demonstrated that the reality of this spill would be 5 km to the flammable limit. Some of these clouds were ignited. The flame burnt slowly as a cloud fire and did not explode. Indeed, even 4000 m^3 of optimally pre-mixed natural gas clouds were shown to burn at speeds of 10-20 m/s and to produce only a few millibar [12].

The industry also considered what would happen if a worst-case event, a detonation of methane/air, took place. During the 1970’s considerable progress was made in understanding gas phase detonations. For all non-pressurised natural gas/air systems it was shown that the conditions needed to initiate detonation, whether by shock, flame-jet ignition or flame acceleration, were too extreme to occur in practice. Indeed, in a detailed survey of historic accidents [13] involving intense vapour cloud explosions none were found to have detonated with natural gas. The same is not true for other more reactive hydrocarbon releases however, and this will be discussed further later.

The tentative conclusion that extreme conditions were necessary for natural gas to detonate was supported by experiments by British Gas in 1989 [14]. In a standard 45 m long rig of repeated 0.18 m diameter grids of 1.5 m spacing, blockage ratio 40%, stoichiometric natural gas/air reached a steady flame speed of 80 m/s whereas cyclohexane continued to accelerate to 230 m/s up to the end of the obstacle array. Further experiments showed that stoichiometric propane/air, given an initial flame speed of 100 m/s from a confined explosion (known as a “bang box”) detonated at the end of the standard array. Natural gas, on the other hand, reached a steady flame speed of around 500 m/s even when the initial flame speed from the confined region was 1000 m/s, under otherwise the same obstacle conditions as with propane. When the initial flame speed was < 500 m/s natural gas reached a steady flame speed of just 30-40 m/s.

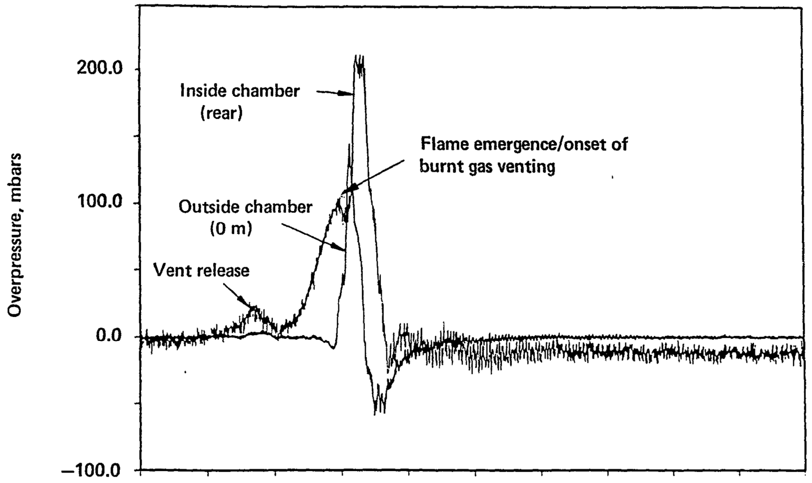
Towards the end of the 1970s an incident occurred on an offshore platform. A severe snow storm had blocked the turbine air intakes. A platform blowdown of the full gas inventory followed. The platform normally vented this gas, but atmospheric conditions led to static ignition. The vent tower was too short and fires from the incident radiation started to break out on the deck. Clearly a review of vent and flare stack design was required. I was tasked with developing a new flare radiation model that could be used to design flare stack heights to avoid excessive radiation levels in the nearby environment. Kalgatghi had already started to develop a surface emitter model based on the frustum of a cone as the flame shape which radiated uniformly from its surface [15]. Several other of my co-workers had fitted flame shape parameters to the frustum derived from medium scale flare studies. My team extended the range of flame sizes by measurements on the Brent C platform and on high velocity flames at the Spadeadam test site in Cumbria UK. Some of the Spadeadam flames were sonic and all were lifted but stable without the use of flame holders. The model was further extended and adopted to describe generalised jet flames for hazard assessment and is still widely used today [16].

Several tests were carried out on vented gas explosions in the 1980s [17]. Propane/air and natural gas/air mixtures were ignited at different positions in an empty 30 m^3 ISO container vented at one end. A main finding was that the external explosion, arising from ignition of unburnt expelled gas, had a significant influence on the internal pressure. Figure 2 shows the overpressure traces from two of the tests, both containing natural gas and air (stoichiometry 1.1). Figure 2a was obtained for rear ignition and ½ area vent. Note that the external overpressure exceeds the internal pressure in Fig. 2a, which turns off the venting process momentarily causing an increase in internal pressure. Figure 2b shows the overpressures for central ignition and a ¼ area vent. The trace is notable for displaying all the features of a vented explosion from an empty compartment. Under these conditions the external explosion can trigger the start of low frequency oscillations in the enclosure. These oscillations have been shown to be of a Helmholtz type and are not standing acoustical modes of the confining enclosure [18]. High speed cine and schlieren photography confirm that the oscillations arise from a bulk motion of the gas within the enclosure and the whole “bubble” of hot products is seen to “bounce” at the observed frequency toward and away from the vent opening. The phenomenon has been observed at all scales up to 60m^3 so far and there is no reason to suspect that it would not occur at even larger scales.

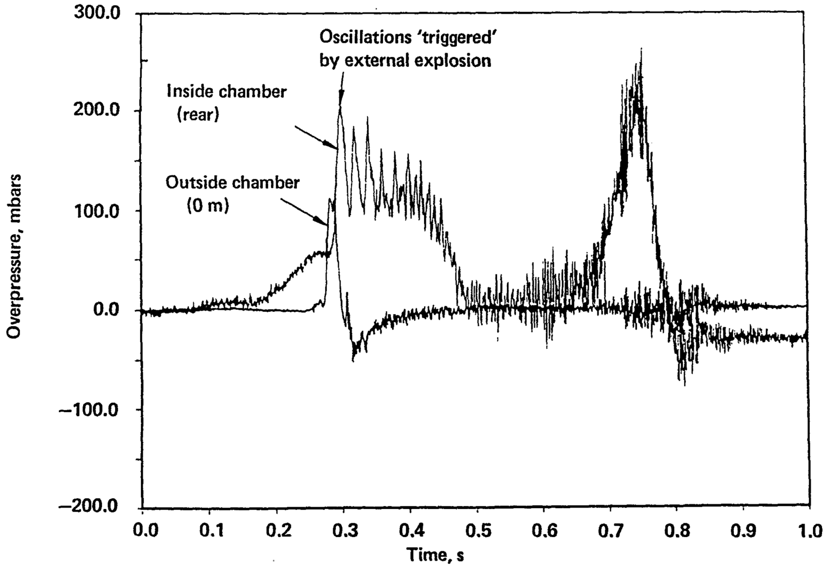
Following the onset of the oscillations the average internal pressure usually rises, implying an increase in rate of volume production by combustion. The oscillations themselves are superimposed on this rising pressure background and become gradually damped and of higher frequency as the flame expands. When the flame reaches the enclosure walls the flame area and hence burning rate and pressure reach a maximum. Thereafter the flame area begins to decrease and the internal pressure starts to fall. Theoretical approaches explain the background pressure rise as the combined effect of two processes. First the burning rate is enhanced by the turbulence generated in the shear layer between the outflowing burned gases and the unburned gas within the enclosure. Second, the flame surface farthest away from the vent satisfies the conditions necessary for instabilities to develop [19].

In general terms, Taylor’s theory says that a density interface becomes unstable when that interface is accelerated in the direction of the higher density medium. A flame surface can be regarded as a density interface between higher density unburnt gas and lower density burnt gas. The Helmholtz oscillations can therefore induce Taylor instabilities on that part of the flame surface that is propagating towards the rear of the enclosure. These instabilities manifest themselves as a wrinkling and break-up of the rear portion of the flame during the half-cycle of oscillation that accelerates the surface away from the vent. This phenomenon is readily observable in high speed films of centrally ignited vented explosions.

The problem for engineering design is to determine the contribution that this phenomenon makes to the explosion hazard. Theoretical studies [18] predict that the amplitude is more readily damped as the vent area is increased and that damping becomes more ineffective as the volume of the enclosure is increased. The implications for scaling are self-evident and should be checked by experiment. However, in most scenarios of practical interest this pressure peak will not define the maximum in a worst-case event. It might be important though in accident investigations, and in designing vent relief from large volumes through ducts and pipes.



(a) Test B5, natural gas, rear ignition, ½ area vent.
Note that external pressure temporarily exceeds the internal pressure



(b) Test B6, natural gas, central ignition, ¼ area vent

Fig. 2. Comparison of pressure signals internal (at rear) and outside the "bang box" [17].

Under the same conditions (ignition near the middle of an empty enclosure with a single vent), an intense high frequency oscillatory pressure peak is generated in the final stages of the explosion. Experimental evidence suggests that this peak is related to acoustically coupled combustion. The phenomenon is observed as an intense final peak on which a high frequency oscillation is superimposed. Experimental studies show that the frequencies match the normal resonant acoustic modes of the enclosure. High-speed film records show that the oscillations are triggered when the flame front approaches and touches the enclosure walls. The normal cellular appearance of the flame front vibrates in sympathy with the oscillations and breaks up into finer structure. The peak is coincident with combustion in the corners of the enclosure accompanied by large flame distortions and an abrupt increase in flame luminosity.

A quantitative prediction of the magnitude of the acoustic peak requires a knowledge of the combustion rate and the rate of venting. At present there are no theoretical expressions for the size of the pressure peak. However, experimental studies clearly show that the magnitude is reduced in venting conditions that hinder (i.e. damp) the feedback process, such as the presence of obstacles, irregularities in the internal surfaces, acoustically absorbent material on the internal surfaces, asymmetric vent locations, asymmetries in the enclosure geometry, and large vent areas that increase the rate of venting.

POST PIPER ALPHA DISASTER

Then the Piper Alpha disaster occurred in 1988, and even more attention was focussed on fire and explosion hazards. Several initiatives were started to address the gaps in knowledge. In summary these were:

- The testing of passive fire protection under jet fire conditions [20] and development of the Standard Jet Fire test [21].
- The efficacy of water sprays protecting pressurised LPG tanks from impinging jet fires [22].
- Jet fire testing to study the internal and external heat fluxes and internal temperatures [23, 24].
- The Shell Offshore Large Vented Explosion (SOLVEX) [25] during which the first blind prediction exercise was held.
- Studies of atomisation of flammable liquids and their explosion hazards [26].
- The Blast and Fire Engineering for Topside Structures (BFETS) involving compartment fire tests, multi-component jet fires, and large-scale confined gas explosions [27].

In addition, the European Union funded the MERGE project [28]. Many other experiments on small to medium scale vapour cloud explosions were carried out by Shell Research [29]. Puttock used the results to develop an empirical model [30] called CAM (the Congestion Assessment Method) based on area blockage of obstacles in the flame path rather than the volume blockage favoured by TNO in their Multi-Energy Method (MEM) [31]. A physical model, The Shell Code for Overpressure Prediction in gas Explosions (SCOPE), was developed [32] to aid assessments of vented explosions in offshore platforms. The CFD codes, EXSIM and FLACS, also saw considerable development in this period.

In some rare circumstances, prediction of a potential explosion in a particular area may be to perform scaled experiments. A theory for scaling explosion experiments was developed by Taylor and Hirst [33], based on Gouldin's [34] fractal model of turbulent combustion.

The fractal scaling theory allows explosion assessments to be performed on scale models by using a more reactive fuel. An alternative approach which can also be used is oxygen enrichment of the fuel/air mixture. Fractal theory predicts that experiments at 1/12th scale using ethylene/air will result in roughly the same flame speeds and overpressures that would occur at full scale with methane/air. The theory can be directly tested by also carrying out experiments at 1/12th scale using propane/air [35]. These are predicted to result in flame speeds approximately one half, and overpressures approximately one quarter, the values measured with ethylene/air. Figure 3 shows excellent agreement in peak overpressure, time of arrival and pulse duration between a propane experiment (scaled appropriately) and an ethylene result.

The example shown in Fig. 4 is of experiments, performed at the design stage, for the wellhead module on the Troll platform and shows the effect on explosion overpressure of reducing the length of the wellhead conductors, a change that was implemented in the final design.

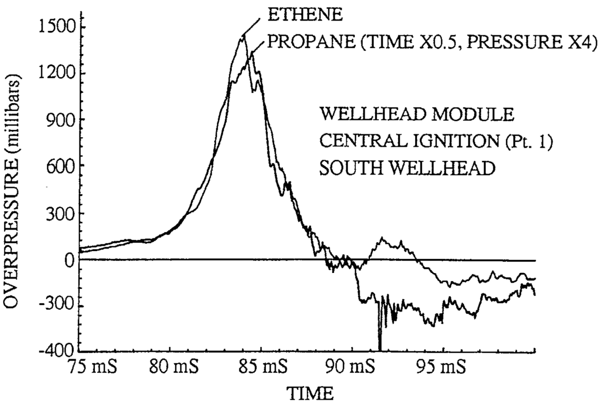


Fig. 3. Peak overpressure, time of arrival and pulse duration between a propane experiment (scaled appropriately) and an ethylene result in the 1/12 wellhead module [35].

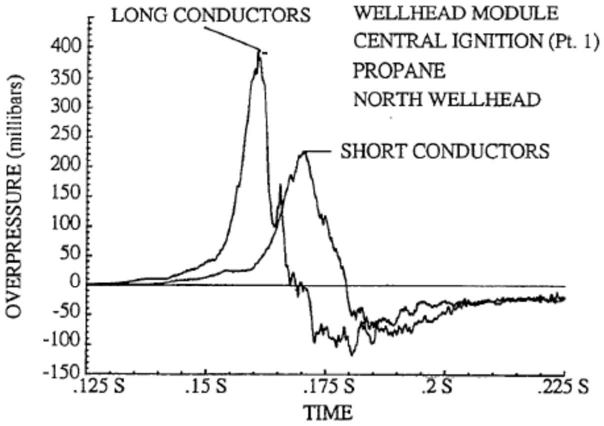


Fig. 4. The effect of a design modification (long wellhead conductors versus short conductors) in the 1/12 scaled Troll wellhead module [35].

The theory requires that overpressure generation is determined by turbulence-induced flame acceleration only. Rapid flame acceleration leading to sonic flame speeds and shock flame interactions, and consequent very high overpressures, are thus beyond the range of applicability of the approach. Nevertheless, the agreement with scaling theory predictions at overpressures close to 1.5 bar (see Fig. 3) is encouraging, since there are relatively few practical situations in which overpressures much greater than 1 bar are tolerable.

Information was required on the internal fluxes and temperatures of jet fires in order to design against rapid rises in temperature of fire engulfed equipment [23]. A 3 kg/s sonic natural gas flame has a flame length of about 22 m, maximum temperatures around 1300 °C towards the end of the flame, and maximum internal fluxes of around 300 kW/m^2. The total heat flux was found to vary over the surface of the engulfed object and the highest convective component is experienced close to the centre of the flame where a combination of high velocities and high temperatures occur. The highest radiative heat load is experienced towards the end of the flame. For effective fire protection passive fire protective coatings should resist such heat loads for a considerable time, usually set to one hour before the back-surface rises above a set level of typically 140°C. A Standard Jet Fire Test was developed [21] using 0.3 kg/s sonic propane gas as the fire source. This jet flame is directed into the bulkhead of a hollow chamber located 1 m from the propane exit nozzle. A fireball is generated in the recirculating gas flows to simulate the high convective and radiative fluxes that have been measured in the 3 kg/s natural gas jet fire. The flame velocity at the front of the chamber reaches around 50-100 m/s. High flame luminosity is achieved in the fireball. An average heat flux of 240 kW/m^2 and maximum 300 kW/m^2 are obtained over the area impinged by the flame. The test reproducibly simulates the thermal, mechanical and erosive forces imparted to PFP coatings by gaseous jet fires.

Nevertheless, the Standard Jet Fire Test does not reproduce the high erosive forces near the source of liquid jets where the jet momentum is high. Temperatures can be considerably lower than the surrounding jet flame due to jet expansion and evaporation of droplets in the source fluid. Such effects are important for jet fires impinging on nearby pipes and equipment. A series of tests were conducted [36] using superheated pentane as solvent for bitumen (a mixture known as Solbit), representing the extraction of heavy hydrocarbons from oil shale. Figure 5 shows an example of the jet fire from a 1.5 kg/s release of pentane. The temperature of the cold core region was -15^{o}C and extended for at least 100 diameters downstream, demonstrating a novel challenge for passive fire protection on nearby equipment.

At the time of the Piper Alpha disaster it was unknown how partial confinement of jet and pool fires would affect the fire properties. A series of experiments were set up at the SINTEF laboratory in Trondheim [27, 36-40]. Medium scale tests, sponsored by Shell, consisted of 0.3 kg/s propane jet fires and diesel pool fires in a 135 m^3 compartment vented at one end. Large scale tests that followed were part of the Blast and Fire Engineering for Topside Structures Project [27], sponsored by several oil companies and regulators. These tests made use of a 405 m^3 compartment vented at one end. A 1 kg/s sonic propane jet fire linked back to the medium scale tests to investigate the effect of scale. Other tests were performed with 1 kg/s condensate (a fraction from crude oil similar to gasoline) jets and pools. Full details are published elsewhere but the main results from the jet fire tests were:

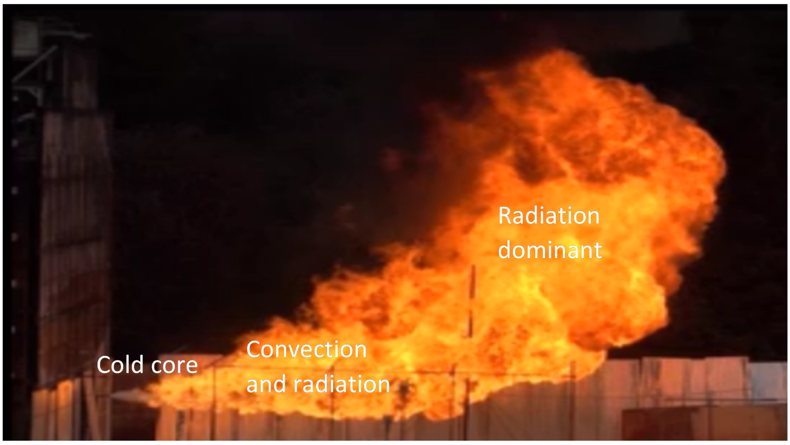


Fig. 5. Jet fire produced from pentane at 74^{o}C from a 10 mm hole at 7.9 barg. Mass flow rate 1.5 kg/s. Curvilinear flame length 13 m. Note the 3 regions of jet fire each presenting different challenges to passive fire protection.

- Quasi steady state conditions were achieved after about 10 minutes.
- The jet fire properties were similar to those observed in open, fuel-controlled fires. Flame temperatures for the gaseous jet fires were measured in the range 1100 – 1300^{o}C and maximum heat fluxes were 300 – 350 kW/m^2 in both fuel-controlled and ventilation-

controlled fires. The liquid releases of condensate had lower total fluxes of around 250 kW/m^2 owing to a reduced convective component.

- There was no significant effect of scale, indicating that realistic scales had been investigated.
- There was no significant dependency on fuel type except within the impingement zone on the ceiling for vertical jets. The hot spot observed for gaseous propane was replaced by a “cold” spot for condensate due to incomplete vapourisation of condensate droplets. Well away from the impingement zone the heat flux was almost 100% radiative for all jets.
- A split vent of equal area to a single vent greatly increased the air flow to the jet fire, turning a ventilation-controlled fire with the single vent into a fuel controlled one.
- Venting through a 5 m^2 hole in the roof in the smaller compartment led to a blue oscillatory flame which extinguished itself after 75 s.

Two important discoveries were made. The first was the measurement of extremely high heat fluxes under the specific conditions of horizontal jet at roughly ½ vent height pointing towards the rear of the compartment. Flame temperatures in the region local to the jet reached >1370^{o}C. Patches of molten steel were observed on a pipe target in the compartment, indicating local temperatures over 1500^{o}C. Favourable access to combustion air and radiative feedback from the hot surroundings may have played a part. The second involved application of deluge water to jet fires. Flame extinction occurred in all the smaller scale tests within seconds of deluge application at typical offshore rates of 10 l/m^2/min, irrespective of whether the flame was fuel or ventilation controlled, sonic or subsonic. Unless the fuel supply is rapidly cut off, these conditions could lead to an explosion hazard on re-ignition of the gas cloud. In contrast at the larger scale, very early application of deluge after seconds did not extinguish the fire, whereas late application after several minutes did so. External flames from the compartment were replaced by copious smoke and steam.

The following results were obtained for the pool fires:

- Fuel controlled pool fires:
  - Diesel pool fires burn at the same rate as open fires, at 0.044 kg/m^2/s.
  - There was no difference in burning rate for pool fires on water or steel.
  - The burning rate was not dependent on the scale of fire source or compartment size.
  - The compartment was fully insulated to protect the fire test hall. This led to rapid rises in fire temperature over 1350^{o}C and high heat fluxes around 350 kW/m^2. Combustion of soot was the cause, resulting in a smokeless flame.
- Ventilation controlled pool fires:
  - The burning rate fell to 0.031 kg/m^2/s. All the available air was consumed and the burning rate settled in the slightly fuel rich regime of 1.15 global stoichiometry.
  - Total heat flux in the fire plume reached 250 kW/m^2 (about 90% radiative) and smoke temperature became uniformly around 1100-1200^{o}C.
  - Roof vented pool fires continued to burn albeit at the lower rate of 0.019 kg/m^2/s.
  - Deluge did not extinguish the pool fires but the burning rate was reduced. Ventilation controlled fires became fuel controlled.

POST BUNCEFIELD EXPLOSION

The Buncefield explosion on December 11, 2005 [2] was allegedly the more intense explosion in the UK since World War 2. Approximately 300 tonnes of winter grade gasoline overflowed from a storage tank on a large storage site over a period of 40 minutes before ignition, which was most probably caused by an electric pump in the firewater pump house. The area covered by the vapour cloud was estimated to be around 120,000 m^2 and the average height of the cloud was around 2 m, giving an approximate volume of 240,000 m^3. The ensuing explosion was of a severity that had not been identified previously in a major hazard assessment of this type of facility.

The event prompted a major inquiry which led to an extensive test and modelling programme to understand the cause of the explosion severity [41, 42]. In early investigation we thought that ignition had taken place at the south end of the site because of the way trees and posts had been displaced to the north. However, after many tests it was concluded that a deflagration to detonation transition (DDT) had occurred in the vapour cloud consistent with ignition to the north of the site and displacement of items in the *opposite direction* to flame propagation. Other mechanisms were considered and investigated but were dismissed [43] because they could not explain the combination of damage levels, directional effects and rapid decay of overpressure from the edges of the cloud.

The investigations [41] discovered several novel features of vapour cloud explosions:

Experiments at large scale using stoichiometric propane/air showed that detonations occurred in a 4.5 m wide row of dense vegetation, whereas no transition to detonation took place in a test using 2 m wide denser vegetation and the flame speed limited itself to around 150 m/s. However, the exact conditions that can give rise to a DDT could not be determined within the limited number of tests performed and the high variability of vegetation. Nevertheless, the presence of foliage and small twigs in the vegetation always resulted in increased flame speed.

Detonation was found to propagate through relatively thin (< 200 mm) layers and paths in a large propane/air vapour cloud. This suggests that once a detonation has started it will propagate through remaining flammable pockets of a large cloud that have developed in relatively calm conditions.

The tests quantified the rapid overpressure decay from the edge of a detonating pancake shaped cloud. Simulations performed using the Fluid Gravity Engineering code, EDEN [41], showed a correlation between the maximum overpressure outside a pancake shaped cloud P (bar) and the ratio of cloud height H (m) to distance from the edge of the cloud D (m). From this work, a simple expression was derived to estimate the maximum overpressure applicable to clouds with a radius ≥ 50 m, $P = 6.57 (H/D)^{0.975}$. For smaller clouds (<50 m radius), it was found that either the Multi-Energy Method or TNT Equivalence was sufficiently accurate within the limitations of the tests and simulations performed.

The negative impulse within the large cloud was observed to be more intense than the positive pulse and causes items inside the cloud to be drawn in a direction opposite to that of detonation propagation. Outside the cloud the positive pulse dominates the impulse and items are displaced outward. Direction indicators, such as how poles and trees have been eroded or tumbled, indicate how the objects encountered the strong shock and subsequent waves that form.

Fig. 6. Damage to some of the items placed inside a propane detonation [41]. Damage created by overpressure > 10 bar. Note the dark markings on the post (extreme right) facing opposite to detonation propagation.

Figure 6 shows damaged items placed inside a propane detonation. These "markers" combined with the directional indicators have played an important role in identifying instances where a DDT has occurred in previous accidents involving vapour cloud explosions. When combined with other effects such as window breakage at > 3 km, Richter Scale measurements > 2, TNT equivalence > 10 tonnes, debeaded tyres, crimped storage tanks, shattered reinforced concrete, severe building damage and availability of reliable information, a review of previous accidents [13 and references therein] has revealed DDTs are highly likely to have occurred at Jaipur, CAPECO, Amuay, Skikda, Brenham, Ufa, Port Hudson, Newark, Flixborough, Pasadena, Decatur and Beek.

A striking feature common to most of these events was release of heavier than air fuel vapour in calm or low wind conditions combined with several minutes delay before an ignition. In such conditions a pancake shaped vapour cloud developed by gravity-driven dispersion over a large area. These incidents are Buncefield, Jaipur, CAPECO, Amuay, Brenham, Ufa, Port Hudson, Newark and Decatur. Large releases of short duration before ignition, on the other hand, remain largely momentum dominated and tend to relate better to turbulent hemispherical clouds. These incidents are Flixborough, Pasadena and Beek. The incident at Skikda appears to be a combination of initial turbulent momentum driven dispersion followed in the far field by passive dispersion. The incidents at Decatur and Beek were powerful events but there is some doubt over whether a DDT occurred due to lack of some key information.

The incidents can also be classified by the general mechanism of DDT. Buncefield, Brenham and UFA underwent DDT by flame acceleration in dense vegetation. Skikda, Flixborough, and possibly Pasadena and Beek suffered from DDT by industrial plant congestion. The incidents at Jaipur, CAPECO, Port Hudson, Amuay, Newark and possibly Decatur are likely to have undergone DDT by jet ignition from confined volumes (“bang box” ignition) into the external flammable cloud. While it is difficult, if not impossible, to assign the fundamental mechanism of DDT to each incident, it is likely that a combination of hot spot formation, energy focussing by coincidence of shock waves and intense turbulence has played its role.

The survey of accidents identified no detonations involving methane or natural gas and very few VCEs involving hydrogen. It seems that the natural buoyancy of these gases does not favour the development of sufficiently large flammable clouds engulfing congested space. Nevertheless, release of these gases in confined surroundings can create the right conditions for powerful explosions. The nuclear power plant at Fukushima is one case in point because evidence points towards at least one detonation having occurred in the build-up of hydrogen/air cloud inside the reactor buildings.

Several recent projects show that a DDT may occur in conditions that are not as extreme as previously thought, given that certain critical requirements are met. Pekalski et al. [44] found that quiescent ethane/air underwent a DDT in a rig of 5.2 m square by 2.6 m high consisting of horizontal and vertical pipes each 0.076 m diameter with a pitch of 0.34 m. The DDT occurred towards the end of the congested region and the detonation propagated through an unconfined part of the rig and back through pockets of unburnt gas at the edges of the congestion. Johnson et al. [45] reported a series of large-scale experiments involving the interaction of vented confined explosions with an external region of pipework congestion (“bang box” jet ignition) that showed evidence of DDT and the continued propagation of the detonation through unobstructed vapor cloud. The DDT occurred in both ethylene/air and propane/air experiments. Davis et al. [46] found that quiescent propane/air at slightly greater than stoichiometric in a “low congestion” rig transitioned to detonation after flame acceleration of about 22 m. The rig consisted of 3.7 m cubes each containing 5.8% volume blockage of uniformly distributed pipes. Lean ($\varphi = 0.9$) and rich ($\varphi = 1.35$) mixtures of propane/air also made the transition to detonation when the rig was changed to the “high congestion” of 10.9% volume blockage.

It seems that detonations in accidentally released vapour clouds may be more common than realised hitherto. One approach to counter such threats is the use of active mitigation systems. The use of solid powders [46] and water sprays [47] have shown promise in this area.

CONCLUSIONS

Over the last 50 years considerable time, expense and resource has been devoted to understanding the chemistry and physics underlying the combustion of large release of hydrocarbons such as might potentially occur in the process industry. The concern is how to pass on this accumulated knowledge to the next generation of safety engineers. As Andrew Hopkins, author of "Lessons from Longford", says in The Chemical Engineer [48], "The best way to maintain the correct level of vigilance (for avoiding major accidents)….is to read accounts of major accidents and identify the causes, both technical and organisational. You should then ask yourself 'could this story be repeated at my workplace?' "

With the rapid development of computing power, many hazard models have been formulated. They are extensively used throughout industry to satisfy consequence and risk analyses. However, whereas scientific papers are peer reviewed there is no formal process to verify or validate computer models. Blind predictions of forthcoming experimental results are occasionally undertaken voluntarily but comparisons between practice and simulation are not always satisfactory. Some degree of overprediction is tolerable as this ensures a "safety net", but too great an overprediction has unacceptable implications for design and maintenance costs. Underprediction and unproven theories lead to mismanagement of the hazard thereby putting life and businesses at risk. In the author's opinion the present situation is not acceptable. I recommend that a formal peer review process should be established for the validation of hazard computer models for use by designers and operators in industry. The process should not limit innovation following future discoveries but should guarantee trust in the simulations that are created.

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