This chapter provides examples of major incidents that have occurred releasing flammable materials. Incidents in which flammable mass estimates are reported are included in Section 3.2.

3.1. Property Losses from Vapor Cloud Accidents

Vapor cloud explosions and fires are responsible for most of the largest property loss events worldwide in the hydrocarbon industry as illustrated in Figure 3.1. This figure analyzes the 100 largest property losses for the period 1966–1996. Vapor cloud explosions account for the largest proportion of losses both by number of events and by dollar loss (inflation adjusted).

The type of equipment involved in the largest property loss events is shown in Figure 3.2. Piping systems are most frequently involved, and reactor events have the largest average dollar loss. Failure of any of the listed type of equipment could result in a flammable vapor cloud.

3.2. Examples of Vapor Cloud Events

The following examples have been selected to illustrate a variety of settings, materials, and conditions that occur, to represent the situations in which flammable mass estimation could be needed. The focus of this discussion is on flammable mass. Further details of the events are provided in referenced material.
3.2.1. Bangkok, Thailand, LPG Vapor Cloud

At approximately 10:30 p.m. in the evening of September 24, 1990 a flatbed truck carrying two LPG tanks “recklessly and at high speed” careened off an expressway exit ramp, apparently trying to beat the traffic light at the base of the ramp [Hazardous Cargo Bulletin, 1990]. The truck crashed into an automobile on the six lane New Petchburi Road in the center of Bangkok (see Figure 3.3). The LPG tanks on the truck bed were interconnected by a 2-in. line at the bottom of the tanks. The truck tipped over immediately and the 2-in. line severed, discharging...
some 5000 kg of LPG. The LPG formed a white vapor cloud that spread west along New Petchburi Road. The wind speed at the time was relatively calm. An eyewitness in a car near the Wireless Road intersection reported, “We could smell gas everywhere. It had filtered into the car.” The cloud passed over numerous ignition sources without igniting, possibly because its concentration was above the UFL (although the edge of such a cloud always contains a flammable

Figure 3.3. Schematic layout at the scene of the Bangkok LPG truck accident (Reproduced by permission of DNV Technica, Inc.)
zone). After a considerable delay it ignited, with one report saying the flame spread east up New Petchburi Road from the corner of Soi 37. There was a flash fire, and at least one explosion, probably from gas that had entered a nearby building. Forty-eight of the shop houses on both sides of the street were destroyed, as were 57 cars. Some 68 people died and over 100 were injured. Evidently, the explosion(s) occurred in confined areas, inside shops. Flash fire occurred in the unconfined street area (Lees, 1996, p. A162; Shaw, 1990).

This event illustrates how both flash fires and explosions can result from a single release and that a flammable cloud can pass over several ignition sources before igniting because not all the cloud is in the flammable range and not all ignition sources have an immediate effect.

3.2.2. Saint Herblain, France, Gasoline Cloud, October 7, 1991

A petroleum depot with a fuel storage capacity of approximately 80,000 m³ at Saint Herblain, France adjoined a parking lot used by numerous petroleum trucks. At approximately 4 a.m. the daily activities were just starting and the atmospheric conditions were stable (Pasquill Stability Class E or F) with wind speeds less than 1 m/s. The temperature was 15 °C with nearly 100% relative humidity. A leak occurred on a transfer line. Gasoline leaked continuously and developed a large aerosol cloud of about 23,000 m³ that covered the road and parking lot and part of the storage area. Four tank trucks in the area of the cloud had their motors running. About 20 min later, the aerosol cloud was ignited and a gas explosion developed. Storage tanks were damaged and tank trucks were overturned and burned.

This event was extensively analyzed to see why the conditions led to flame acceleration and an explosion. The computational fluid dynamics (CFD) model REAGAS was set up with six parallel tank trucks inside the vapor cloud, as shown in Figure 3.4 (Lechaudel and Mouilleau, 1995). The flame progression is predicted as shown in Figure 3.5 for the plan view and end view. Interestingly, the simulation showed that it required at least six trucks to develop enough flame acceleration to produce explosive overpressures. Figure 3.6 shows a large increase to damaging levels of 100 kPa when the flame passed under the sixth tank truck. When the simulation is repeated using only five trucks in the vapor cloud, the peak overpressures are predicted to be only around 10 kPa, not enough to cause serious structural damage.

The importance of this example is that it confirms findings from small-scale experiments showing that repeated plant structural elements can cause flame acceleration to explosion velocities. The Saint Herblain incident indicates that a small difference in the number of parallel-parked trucks is enough to develop a critical degree of partial congestion leading to an explosion. It also is an event
3.2. Examples of Vapor Cloud Events

Figure 3.4. Arrangement of tank trucks inside vapor cloud as modeled by REAGAS model. (Reproduced by permission of INERIS Parc Technologique Alata.)

Figure 3.5. Temperature fields predicted by REAGAS model near tank trucks at Saint Herblains, France. Light color is >2500 K, dark is <1750 K. (Reproduced by permission of INERIS Parc Technologique Alata.)
occurring under meteorological conditions giving poor dispersion (F stability, < 1 m/s wind speed).

3.2.3. Pampa, Texas, Hoechst-Celanese Explosion, November 17, 1987

The Pampa plant used liquid phase oxidation of butane with air about 49 bar (700 psig) to produce acetic acid and acetic anhydride. The plant was shut down because of a boiler outage, but each reactor contained about 38.0 m$^3$ (10,000 gallons) of reactants in the reactor and 26.6 m$^3$ (7000 gallons) in the decanter. Upon starting up one of the three 38 m$^3$ reactors, a flammable mixture in a reactor manifold reached hot catalyst and exploded, breaking the reactor feed line. Butane, acetic acid and methyl ethyl ketone discharged through the broken line and formed a vapor cloud for about 30–45 s. It was ignited by gas-fired boilers about 60 m away. The resulting explosion extensively damaged the plant and broke windows in the town of Pampa. Three people were killed and 37 injured. It also ruptured the underground fire water main, rendering the pressure inadequate to fight the subsequent fire (Mahoney, 1990; Lees, 1996, p. A1/59).

Flammable-mass modeling was used in the accident analysis to estimate the energy of the explosion and to set design parameters for blast-resistant buildings for the rebuilt plant.

![Figure 3.6. Overpressures predicted between the last two tank trucks in the vapor cloud at Saint Herblains, France. See Figure 3.4 for location of points P4 and P5. (Reproduced by permission of INERIS Parc Technologique Alata.)](image-url)
3.2.4. Monsanto Ethanol Explosion, Autumn, 1970

Ethanol is considered one of the less dangerous flammable materials to handle, yet an explosion of ethanol vapors destroyed a large building. A brick building at Monsanto’s Queeny, Missouri plant was 18 m (60 ft.) wide, 43 m (140 ft.) long, and 10 m (34 ft.) high with a flat roof constructed of hollow clay tiles. Originally the building had numerous large windows containing multiple glass panes, but these had all been removed and the areas left open (with 496 m$^2$ area). Boiling liquid ethanol overflowed from a vessel opening that was flush with the mezzanine level, intermediate between ground level and the second story. Most of the estimated 0.23–0.57 m$^3$ (60–150 gallons, 167–417 kg) released flowed over the edge of the mezzanine and through the open grating floor to the ground level. It cascaded over piping, reaction vessels, and structural steel on the way, vaporizing a sizable fraction.

Upon ignition there was a weak explosion, sufficient to blow out the north wall, some 6 m (20 ft.) from the spill, but not the south wall, which was cracked and had to be replaced. Roof tiles were blown upward and fell back essentially vertically. Six-meter (20 ft.) sections of the side walls were also blown out. It appeared that the pressure developed was just sufficient to exceed the strength of the building, which was estimated to fail at 69 mbar (1 psi). An explosion developed because the vapor cloud was confined and venting was inadequate. The venting provided by the open window area partially, but inadequately reduced the explosion effects. In fact, this incident provides a datum point for explosion venting design.

3.2.5. Mexico City Vapor Cloud and Explosion, November 19, 1984

When PEMEX originally constructed their LPG storage and handling facility in San Juan Ixhuatepec, or San Juanico, a suburb of Mexico City in 1962, there were no buildings or residents near the site. At the time of the incident approximately 100,000 people lived in the surrounding hills. The nearest residential building was only about 130 m south of the storage tanks. Around 5:30 a.m. on November 19, 1984, an 8” feed line delivering LP gas (20% butane, 80% propane) ruptured inside the PEMEX tank area. A 2-m-high vapor cloud spread over an area of 150 × 200 m, as shown in Figure 3.7. There were no gas detectors, and apparently no mitigation action was taken before the vapor cloud reached an ignition source, likely the flare, and exploded at 5:45 a.m. This began a series of major “knock-on” events (Pietserson, 1985, 1986, 1988; Skandia, 1985; Lees, 1996, p. A4/1-8).

After a flash fire burned away, a ground fire was left as well as a jet fire from the yet discharging 8” line fracture. This jet fire was directed toward one of the two large 1600 m$^3$ storage spheres, number F:4. Both of these nearly empty large
spheres exploded and ruptured other tanks. Burning LPG flowed around and through nearby houses and ignited secondary fires in the residential quarters. Several more explosions occurred, and 12 propane “bullets” rocketed up to 1000 to 1200 m away. There were 560 fatalities and 7000 injured. A large fraction of the people indoors within 300 m of the plant perished. The total destruction of the facility occurred because the protection system was destroyed, including emergency isolation and water spray systems. Traffic chaos developed as residents sought to flee while the emergency services tried to get in.

This illustrates that knock-on events can be far more destructive than the initial vapor cloud explosion. Automatic detection and protection systems are important when the release event need last only seconds to generate such a destructive series of events.

3.2.6. Pasadena, Texas Fire and Explosion, October 23, 1989

The Phillips complex at Pasadena, near Houston, Texas produced high-density polyethylene using a loop reactor with ethylene dissolved in isobutane at 49 bar (700 psi) and elevated temperature. The dissolved ethylene polymerizes to form polyethylene particles that gradually come to rest in settling legs shown in Figure 3.8 where they are removed periodically through valves at the bottom. At the top of each of these legs is a single air-actuated ball valve which is kept open during production so that the polyethylene particles can settle into the leg.
The settling legs frequently plugged with plastic material. When this happened, the ball valve was closed, the leg disassembled, and the blockage removed. If the ball valve were to open during a clean-out operation, there would be nothing to prevent the escape of flammable material. On the day of the accident, Monday, October 23, 1989, at 8:00 a.m. work began to clear the second of three plugged settling legs on Reactor No. 5. The maintenance team partially disassembled the leg and were able to remove part of the line plug, but part remained lodged in the pipe 12–18 in. below the ball valve. One of the team was sent to the control room to seek assistance. Shortly after, at 1:00 p.m. a release occurred through an open ball valve. Because of the high operating pressure, over 99% of the contents of the reactor, more than 38,550 kg (85,000 lb.), emptied in a matter of seconds.

It was subsequently established that the air hoses to the ball valve actuator had been improperly cross-connected so moving the actuator to the closed position would actually have opened it.

Figure 3.8. Typical settling leg on polyethylene reactor at Pasadena plant. (U.S. Dept. of Labor, A Report to the President, April 26, 1990.)
Reactors 4 and 5 were in a highly congested area. A huge vapor cloud, a mixture of isobutane, ethylene, hexene, and hydrogen, formed almost instantly, and was well-mixed with air by the influence of the surrounding congested piping. About 1.5 to 2 min later the cloud contacted an ignition source and exploded. A second explosion occurred 10 to 15 min later when two isobutane storage tanks exploded. Other knock-on explosions developed in a chain-reaction of explosions. One witness reported hearing ten separate explosions over a 2-hour period. Fire hydrants were sheared off by the explosion, and water pressure dropped too low to fight the fires.

Twenty-two people onsite were killed immediately, and one later died of injuries. The number injured is variously given between 130 and 300. Polyethylene plants 4 and 5 were totally destroyed. The property loss was nearly $750 million and business loss was around $650 million. Much of the damage was highly directional. For example, the administration building, located at the end of a length of pipe rack, was heavily damaged 400 m (¼ mile) from the explosion, although there was negligible damage in process areas only 30 m (100 ft) from the center of the explosion.

This illustrates the effect of congestion and confinement on an explosion as well as the highly directional nature of some explosions. The plant was rebuilt with particular attention to avoiding the high congestion of the original plant.

3.3. Examples with Postaccident Determination of Flammable Mass

The following incidents illustrate how flammable mass calculations were used in the analysis of the event.

3.3.1. Flixborough Vapor Cloud Explosion, June 1, 1974

The Flixborough Works of Nypro (UK) Ltd. was largely demolished in an explosion at about 4:53 p.m. on Saturday, June 1, 1974. This was by far the most serious accident in the chemical industry in the UK in many years, and brought about far-reaching changes in governmental oversight of plant safety throughout Europe (Lees, 1996, Appendix 2).

The plant produced caprolactam used in making nylon 6. In this process making caprolactam required making cyclohexanone by liquid-phase oxidation of cyclohexane. The oxidation was done in a train of six reactors by air injection in the presence of a catalyst at 155 C and 8.6 bar(g) (125 psig). This is above the normal boiling point of cyclohexane of 80.7 C so the pressure is some 3.5 bar(g) (51 psig) above the vapor pressure of cyclohexane [of 5.1 bar(g) (73.8 psig)]. The product from the reactors contained cyclohexanone, cyclohexanol, and 94% unreacted cyclohexane.
On the evening of March 27, 1974 it was discovered that Reactor 5 was leaking cyclohexane and the plant was shut down. Inspection on the following morning revealed a crack some 6 ft. long in the \( \frac{1}{2} \) inch mild steel plate. Reactor 5 was removed for repairs. Reactors 4 and 6 were connected with temporary piping so the plant could continue in production.

The reactors were normally connected by 28-inch openings with bellows for thermal expansion, but the largest pipe available on site suitable for the bypass connection was 20 in. in diameter. The two flanges were at different heights, so the connection had to take the form of a dog-leg of three lengths of 20-inch pipe welded with flanges at both ends and bolted to the existing flanges and bellows on the reactors. Figure 3.9 sketches the plant flow sheet and the 20-in. bypass. No calculations were made to check whether the bellows would withstand the twisting moments caused by the dog-leg bypass, which was not restrained in place aside from some minimal support by scaffolding.

On the morning of June 1, 1974 startup began. The sequence of events is complex and uncertain, but at a point the plant was subjected to pressures slightly higher than the normal operating pressure of 8.63 bar. During the late afternoon while the plant was on hot recycle a large quantity of cyclohexane was accidentally released. Although alternate views were put forward, the court of inquiry (Dept. of Employment, 1995) attributed the release to the rupture of the bellows and the falling away of the 20-in. bypass pipe. The resulting release formed a vapor cloud, and between 30 and 90 s after the release there was a massive vapor cloud explosion.

According to the court of inquiry report, the mass of cyclohexane discharged is estimated as 30–50 tonne from an initial inventory of 120 tonne.
estimated dimensions of the vapor cloud are shown in Figure 3.10, and formed a lenticular-shaped cloud ranging between 75 and 210 m from the source. The mechanism of failure was taken to be the simultaneous failure of both bellows and buckling of the dog-leg pipe. This failure would produce a discharge from the two 0.7-m diameter nozzels on Reactors 4 and 6. However, the court accorded this conclusion only a low probability, saying it would be readily displaced if some [other event of] greater probability could be found.

From evidence of carbon deposition and from eyewitness accounts there was a flash fire occurring at the edge of the cloud before the explosion. The cloud was expanding faster than the flame speed at first, but when the cloud expanded to a certain point, calculations show that the expansion speed would fall below the flame speed. The flame would then travel toward the center of the cloud, analogous to a flashback from a Bunsen burner. There is evidence on seismic (ionospheric) records of two explosive events, a smaller precursor followed after 45 s by the main event (Jones & Spracklen, 1974; T. B. Jones, 1975).
An alternative detailed analysis conducted by Venart (1999) and backed by numerous eye witnesses is that only one bellows ruptured perhaps because of work hardening from flow-induced oscillations of 2–4 Hz. This is deduced from hydraulic jumps likely in the new dog-leg flow path and from calculating the natural frequency of the piping. The upstream R4 bellows was most susceptible to oscillations since it was unsupported and only partially full of fluid. When this bellows ruptured, the pipe attached to Reactor 6 was kinked back and upward in a “gooseneck” and the resulting crimp in the pipe largely prevented discharge from Reactor 6. According to this theory, the main discharge was from Reactor 4, and perhaps only 16 tonnes discharged. The flammable vapor cloud predicted by a computational fluid dynamics model depicted in Figure 3.11 envelopes the administration building in the upper corner of the figure. An indoor ignition in this building is believed to have resulted in an early explosion that destroyed the administration building. It also acted as a high-energy ignition source triggering the detonation of the surrounding vapor cloud. Such events have been observed (Harris, 1983).

Of the 28 people who died on site, 18 were in the caprolactam plant control room and were cut by flying glass and crushed by the collapsing roof. Among the general public in the surrounding area there were 53 injuries; 1821 houses and 167 shops and factories were damaged.

**Figure 3.11.** Three-dimensional plot of LFL contours of flammable vapor cloud at the Nypro (UK) Ltd. Plant at Flixborough by alternative discharge theory (Venart, 1999). Southeast view looking toward main office building (in upper corner). Discharge from R4 only at 12 s.
Estimates of explosion energy have been made, including estimates from ionospheric readings (T. B. Jones, 1975) and from the barograph of a glider in the vicinity (Gugan, 1975). The most detailed estimates are derived from damage surveys by Gugan (1976), Munday (1975), and Sadee, Samuels, and O’Brien (1976–1977). Several estimates are based on the TNT equivalence model, which range 15–45 tonne of TNT. Sadee, Samuels, and O’Brien use an alternative two-parameter model in which the height of the explosion is the second parameter. They estimate the explosion was equivalent to 16 ± 2 tonne of TNT at a height of 45 ± 24 m. Using 4.648 MJ/kg for the energy of TNT and 43.44 MJ/kg for the heat of combustion of cyclohexane, for a release of 30 tonne of cyclohexane, the gross explosion efficiency would be 5.7% based on the TNT equivalence model (Section 4.8.3). Marshall (1987) plots both near and far field damage effects and finds wide scatter in the results.

This accident illustrates that high-momentum releases can generate a vapor cloud that flows upwind. It also illustrates the possibility of an indoor detonation providing a high-energy ignition source for the outdoor vapor cloud.

### 3.3.2. Piper Alpha North Sea Platform Fire, July 6, 1988

The Piper Alpha platform was destroyed and 167 died when an initial explosion escalated into fires that caused the main risers to rupture. This incident was the worst accident that has occurred on an offshore platform. The inquiry that followed, presided over by Lord Cullen, produced The Public Inquiry into the Piper Alpha Disaster (the Piper Alpha Report or Cullen Report, Cullen, 1990), the most comprehensive inquiry conducted in the UK into any process industry disaster, onshore or offshore (Lees, 1996, pp A19/1–16).

The platform was a 10-legged steel jacket in 140 m of water. The jacket top was 62 m × 33 m. It had 36 wells, of which 26 were producing. The platform was designed with fire protection, but not with explosion protection. It was a congested, multilevel platform.

The release is attributed to miscommunication that left an open flange after removing the pressure relief valve to a condensate injection pump. The pump was isolated only by closing the gas-operated valves on the suction and discharge sides. About 9:50 p.m. the lead maintenance man organized electricians to deisolate this pump, but it is not clear what action they took. About 9:55 p.m. on July 16, 1988 signals came up in the main control room for the tripping of two centrifugal compressors in Module C. Then a third compressor tripped, followed by a high gas alarm. Personnel in the adjoining D module heard a loud screeching sound which lasted about 30 s, and then the first explosion occurred.

The inquiry concluded, using modeling, that the explosion had been caused by ignition of a gas cloud containing about 45 kg of light hydrocarbon within the flammable range arising from a two-stage leak. In the first stage, the leak rate
was probably 4 kg/s, and in the second stage some 1.83 kg/s leaked through an orifice of equivalent diameter of some 8 mm for 30 s.

Figure 3.12 portrays that the flammable cloud covered only a relatively small portion of the platform area. These predictions are consistent with readings from gas sensors. Explosion damage analysis showed that 45 kg involved in an explosion would produce 0.2 bar overpressures, enough to account for observed damage. Dynamic failure analysis showed the fire-walls fail at 0.1 and 0.12 bar overpressure.

The first explosion in the C Module destroyed most of the B/C and C/D fire-walls and severed the main communications. Crews on the adjacent platforms continued pumping which contributed to the later fires. There followed almost immediately a large fireball which issued from the west side of the B module and

Figure 3.12. Contours of flammable gas cloud predicted by BMT wind tunnel tests: LFL contours at 30 s for leak rate of 100 kg/min. (Reproduced by permission of Controller, Her Majesty’s Stationery Office.)
a large oil pool fire at that point. The large pool fire gave rise to a massive smoke plume that enveloped the platform and cut off escape routes from the living accommodation section. Some 68 men escaped by climbing down knotted ropes or jumping into the sea.

About 20 min from the initial explosion, the riser from the Tartan field ruptured. This resulted in a massive jet fire that destroyed much of the platform. About 10:50 p.m. the MCP-01 riser ruptured, and about 11:18 p.m. the Claymore riser ruptured. The pipe deck collapsed and the living quarters area tipped. By 12:15 a.m. on July 7 the north end of the platform had disappeared. By the morning only the wellhead module remained standing.

Flammable mass calculations supported numerous design changes in offshore platforms. Platform designs now try to keep firewater systems and communication systems less vulnerable to an explosion in one location. Designs are made to prevent accumulation of an oil pool. Temporary safe refuge areas and escape routes are designed accounting for flammable clouds and heavy smoke using wind tunnels and computational fluid dynamics models. Platform enclosure is being limited to allow for cross-ventilation to disperse flammable clouds, and this makes use of flammable mass modeling. Better explosion models are being developed to assist in the design of blast walls.

### 3.3.3. DSM Naphtha Cracker, Beek, the Netherlands, 7 November 1975

The Dutch State Mines (DSM) Naphtha Cracking Plant II for producing ethylene had been out of service for some time and was being started up. At 9:48 a.m. a release occurred, estimated to be 5500 kg of hydrocarbon. The most likely cause of the escape was low temperature embrittlement at a weld in the feed drum of the depropanizer section (Ministry of Social Affairs, 1976). According to witnesses, the vapor cloud enveloped two sides of the control house, as shown in Figure 3.13 (Gugan, 1979). After about 2 min, an explosion occurred, ignited by one of the cracking furnaces, which destroyed the plant and a nearby tank farm and broke windows up to 4.5 km away. Fourteen people were killed, six of whom were in the control room, and eight in the open air outside the control room. Inside the plant, 104 people were injured, and 3 were injured outside.

The Dutch research organization TNO (in Ministry of Social Affairs, 1976) found 800 kg of hydrocarbon exploded, equivalent to 2200 kg of TNT. The net explosion efficiency on this basis is 28%. (See Section 4.8.3.) The investigators pinpointed the epicenter of the explosion as point E in Figure 3.13. It took a considerable distance from this point for the flame speed to build up to give damaging overpressures.

The Beek explosion was modeled in considerable detail using the FLACS model by van Wingerden et al. (1995). Their model produces overpressure contours that move through the vapor cloud as illustrated in Figure 3.14. The
influence of the control room is seen in keeping the cloud partially confined and less dilute. Their model predicts very strong explosions for an ethylene cloud if the cloud height is taken to be 3.90 to 5.73 m, but very low-pressure effects if the cloud height is reduced to 2.36 m. This illustrates an aspect of partial confinement and congestion and bears on the issue of the minimum cloud size needed for an explosion. With a vapor cloud of height <3.9 m, the mixing of air in the top of the cloud dilutes a substantial fraction of the flammable material outside of the flammable range. Furthermore, during combustion, the combustion products are readily vented out the top of the cloud and do not contribute much turbulent flame acceleration. Also, small vapor clouds may not present enough run-up distance for flame acceleration to reach critically high flame speeds.

Van Wingerden et al. also predict mild overpressures for a propylene cloud, emphasizing how the reactivity of the hydrocarbon strongly influences results. They also find a stronger explosion for a rich ethylene cloud (equivalence ratio

Figure 3.13. Plan view of Beek naphtha cracking plant showing vapor cloud contours reported by witnesses (Reproduced by permission of IChemE.)
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Figure 3.14. Pressure field predictions during explosion at Beek by FLACS model. Ethylene–air well-mixed at stoichiometric concentration; initial height of 5.73 m; pressure contours at 0.5 m above ground; scale at right in bar
1.5) than for a stoichiometric mixture (equivalence ratio 1.0). This prediction supports the assertion that some dilution occurs during the explosion.

Good quality flammable mass estimates for the events described above would help to improve our understanding of the energies involved, and how to design the appropriate mitigation measures. Control rooms can be moved away from processing areas or hardened. Buffer spaces can be specified between process areas and public area. Blast walls can be designed for offshore platforms to be both effective and minimal in weight. Accident analysis can better pinpoint the source of releases and of ignition. Explosion and fire models should be more successful in predicting effects or in matching observed effects if the important input values are better known. In short, the application of scientific principles to improving plant safety will be advanced.