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Preface

This report is intended as a basis for equipment release and accident rates for risk assessments, and in particular, to determine the conditions under which the data given in the "Purple Book" are applicable. It gives modification factors which may be applied, for non standard applications of the data.

Updating history

Issue	Date	Affected	Change
V1I2	Aug 2002	Ch 5	Safety equipment failure data added
V1I3	April 2003	Ch 4	ARIP data added
			Division of report into volumes
	July 2003	All	Detailed analyses and algorithms completed
V1I4	July/Aug 2003		Detailed editing, new contents list
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			Baseline data selection revised, typical/reference data concept introduced
V1I7	June 2005		Editing after review H.Beerens
			Introduction of RBI concepts
	Aug 2006		Elimination of basline concept

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7 Piping Release Rates

The discussion of piping failure rates in this chapter is in fact a model for failure rate prediction. After the introduction, a detailed, fault tree based model is presented, along with a checklist-based algorithm, which allows an operational application of the model. The model is presented initially presented with little justification, for simplicity. This is followed by a discussion of the background for the model, and notes on how each of the adjustment factors were derived.

While it would be possible to make a model which took account of all known failure cause factors, quantification becomes increasingly difficult as causes become rarer. For this account in the algorithms the causes considered have been limited to those arising most frequently.

7.1 Hazards

Although there are several kinds of accidents which can occur internally in equipment, it is generally true that it is safe if you can keep the chemicals inside their containers. Releases from pipes, tanks and vessels are involved in a large proportion of petroleum and chemical industry accidents (Ref 7.1, 7.2).

Inspection or auditing for integrity involves looking not only for existing releases, but also for releases which might occur in the future.

In the following paragraphs, some of the ways in which piping and vessels can fail are described.

1. Internal corrosion

Liquids such as acids or alkalis, or gases dissolved in liquid, attack the surface of the metal. Metal may dissolve or form flakes of rust, which subsequently fall off, leaving a thinned pipe wall or a hole.

2. External corrosion

Rain or leaking process water, often contaminated with salt or acid, attacks the surface of metal. Usually, flakes of rust are formed. Sea spray is a very common cause of corrosive salt solutions on piping up to several kilometres from the coast.

3. Erosion

Sand, crystals, or other solids are carried by a flow of liquid or gas, and wear away the metal. In a few cases, external erosion occurs, due to wind blown sand.

4. Pitting and crevice corrosion

A corrosive solution forms in a hollow, which retains the corrosive liquid. The corrosion deepens the hole, allowing acids to form in an oxygen deficient location and this causes corrosion to accelerate. Stainless alloys can be subject to this kind of corrosion, because in the pits or crevices, oxygen is depleted, and the protective oxide layer may be lost.

Figure 7.1. Pipe with very heavy pitting corrosion, after painting. Note that two adjacent pipe sections had very different susceptibility to pitting corrosion, possibly due to the corrosion occurring under different conditions in storage

5. Materials incompatibility corrosion

Sometimes, materials used for vessels or pipes are not suitable for the task. An example is the use of ordinary carbon steel to hold dilute sulphuric acid.

At other times, the wrong materials may be present due to errors in operation, for example sodium carbonate or hydroxide in an aluminium tank.

Dilution of liquids can be a problem e.g. sulphuric acid can be stored in ordinary carbon steel tanks at high concentrations, because a protective oxide layer is formed on the steel. If water is added corrosion results.

6. Galvanic corrosion

When two different metals are joined, for example at a flange, the different electrical potential of the metals causes current to flow. (see fig 7.2) If a conductive liquid is present in the pipe, or rain is present outside, current may flow, and corrosion rates become very high.

Electrolytic corrosion can occur even if pipe metals are the same or compatible, if there is a leakage of electricity from a power supply elsewhere. One source is a poorly designed or installed cathodic protection system.

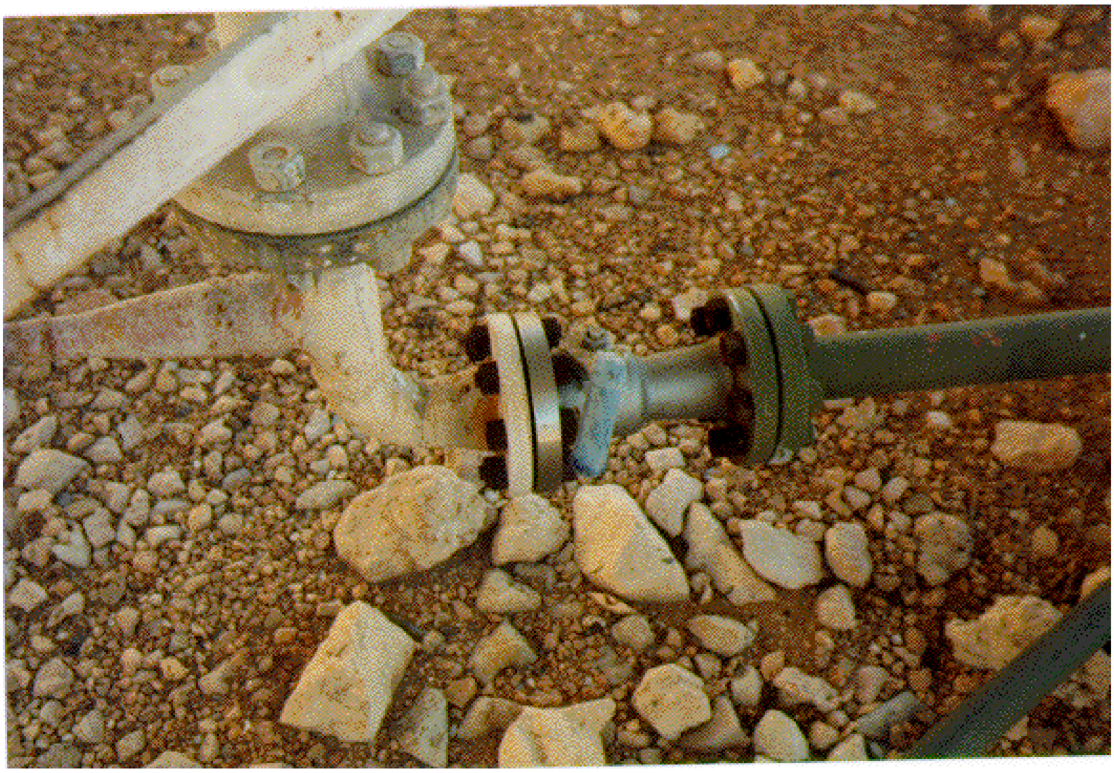


Figure 7.2. Galvanic corrosion on carbon steel bolts in a stainless spool piece

7. Scoring and scratching

Movement of equipment such as agitators inside vessels, and impellers inside pumps can cause scratching of surfaces, if blades or axles are bent. Sometimes the metal can be pushed through a vessel or pump wall. At other times, corrosion is accelerated by the scratching.

Foreign objects such as stones, pieces of tramp metal, broken internals etc. moving inside the pipes can cause this type of damage.

For pipe work, scoring can particularly affect painting or corrosion protection, allowing external corrosion, and can affect internal oxide layer on boiler piping, allowing rapid internal corrosion.

8. Stress corrosion cracking

Some metals are subject to a special kind of corrosion under stress, when small cracks or micro cracks open under stress. Corrosion occurs in the crack, widening it, and at the same time reducing the wall area and increasing the stress.

9. Overstressing

Sometimes pipes will break simply because the stresses on them are too high. This occurs most often because the supports for the pipe are missing, or move. This occurs especially with fibre-reinforced plastic or plastic piping. For metal piping, overstressing due to heat expansion or cooling contraction with insufficient freedom of movement can be a problem.

An example is taken from two steam pipes, which operated in parallel, and joined at a tee. Both were subject to expansion. Since they were used at different times, relative expansion could occur. Proper pipe stress calculations were made, but at the installation, one guide was installed wrongly. As a result, severe stress was placed on a valve in the tee section. As a result, the valve failed, both sticking and leaking

Figure 7.3. Plastic pipe ruptured in torsion due to overstressing

10. Vibration fatigue

Pipes may be destroyed as a result of vibration, if the movement produces high stresses at supports or fixing points vibration can arise due to machinery or conveyors or transmission of pressure pulses from pumps. A particularly vicious form of vibration arises from “vertical two phase flow”, that is, flow of gas and liquid in a mixture up a vertical pipe. Similar problems can arise if mixtures of liquid and airflow down a pipe.

11. Fretting

If two pipes rub against each other, one pipe may eventually rub through the other. This is quite common in badly designed heat exchangers, but is also known to occur between pipes which touch or rest on each other, and between pipes and supports.

12. Scuffing

When pipes move between guides (due to expansion) the guides can bite into the pipe. This can cause heavy gauging.

Fig. 7.4 High quality PVDF pipe damaged by fretting

13. Hammer

When valves are closed rapidly while liquid is flowing, the column of liquid is stopped quickly. The pressure rise, which stops the flow, is hammer pressure. Pressures can be very high, and can often rupture pipes (one of the largest process industry accidents, at Cubatao in Brazil, arose when liquid hit a valve which was partly closed in an 8" pipe with flowing kerosene).

Another hammer mechanism arises when liquid flows into an empty pipe. No hammering occurs until the liquid encounters an obstacle, such as an elbow or a partially closed valve. The increased resistance can then cause a pressure rise, which can burst a pipe, or can tear out pipe supports.

Figure 7.5 . Pipe support distortion arising from hammering during start-up

Particular care is required when filling pipes with liquefied gases such as ammonia, LPG or chlorine with a closed valve at the end. When liquid is pumped into the pipe, the bubble of gas ahead of the liquid is gradually compressed. At the last instant, the bubble collapses and the collapse can be violent. The problem is increased if there is no air, nitrogen, or other incondensable gas to provide a “cushion” to take the shock of bubble collapse.

Draining and relieving pressure from pipes can cause violent hammer effects, if there is pressure, and if liquid is caught at the bottom of a tank or in pockets.

Condensation of steam in pipes can suck water from a sump or pond, so rapidly as to burst a pipe by a hammer effect.

14. Cold embrittlement

Pipes and vessels are generally designed to be tough i.e. to be able to yield if stresses get too high, without cracking. Most steel have a temperature below which they become brittle, so that they crack without significant deformation if stresses get too high.

Most carbon steels can become brittle in very cold weather (most steel has a cold embrittlement point at around -20°C). Pipes can also become cold if cryogenic gases, such as nitrogen from a liquid nitrogen tank are allowed to flow through. Evaporation of liquids such as chorine or propane can cause cold. A typical example arises if a leak of the liquefied gas occurs from a vent, drain or valve seal. When a pipe or fitting is brittle, vibration, or a hard knock can shatter it.

15. Brittle castings

Cast iron is often used for valve and pump bodies. Unless the cast iron alloy is chosen carefully (malleable iron) it can be broken by an impact such as a crash, or by items falling on it. (see figure 7.6)

Figure 7.6 A cast iron valve, broken by impact

Figure 7.7 Overhead pipe, hit by crane and leaking

16. Crashes and crane hits.

Piping may be ruptured by trucks, cranes or tractors that drive off of roads, into pipe trenches, or under low pipe bridges.

If a pipe bridge collapses, the break in piping often occurs not at the bridge itself, but at the bend or anchor point at the side of the bridge. (see figure 7.7)

17. Falling cranes and falling loads.

Cranes are fairly commonly overloaded, and can fall over. When they do so, they can damage both piping and vessels.

Figure 7.6. Pipe elbow broken by impact (PVC pipe)

18. Flange leaks, and breaks.

Flanges may leak because gasket is installed badly, or the wrong gasket material is used. Flanges may also leak because the flange is not adequately tightened.

On pipes and equipment, which change temperatures frequently, bolts tend to stretch, because the pipe and flange material heats up before the bolt. On well-designed systems, bolts are made which can take the extra stress, or with washers which can expand and contract. If the system is badly designed, bolts may rupture, or stretch and require re-tightening. On some systems, re-tightening is so repeated that in the end bolts break.

Flange bolts may also break if over-tightened. Pipe fitters may sometimes use overlong spanners, or spanners fitted with an extension "handle" consisting of a length of pipe, in order to tighten bolts to prevent leaks. When over-tightened, the bolt may breaks, or break later when further stressed during operation.

19. Drain and vent opening

Drains and vents may be opened in error, either because the operator or technician thinks that there is no hazard, or because he or she makes a mistake about which valve to open.

Sometimes an operator will open a valve knowing that there is a hazard, but underestimating it. If there is a valve leak upstream or downstream of the drain, or if the pipeline is pressurised, the operator may get a nasty surprise. For example, an operator may decide to drain a pipe which is in principle not under pressure, but which is in fact pressurised due to valve leakage.

20. Cavitation

Cavitation is usually thought of as a problem of pumps, but sometimes occurs on piping as well. If a liquid is close to its boiling point, and reaches a pipe expansion, flow conditions can cause cavitation.

Figure 7.7 Section from a pipe wall damaged by cavitation on an expansion section

21. Thick walled piping overstress

High-pressure piping sometimes requires very thick pipe walls, especially if temperatures are to be high. Vessels even more often have very thick walls. For example ammonia converters may have walls up to 14” thick. If such steel is heated or cooled too rapidly, stresses become very high, as the inner layer expands.

22. Glass lined piping and vessels.

Glass linings are very vulnerable to rapid heating or cooling. The glass layer expands very rapidly, overstresses and cracks. Rapid cooling damages the lining by the reverse effect.

Glass linings are also very vulnerable to dropped tools or dropped raw materials if these are hard. A small pit allows corrosive substance to get behind the glass or enamel layer and create “rust”. Rust then flakes off the glass layer. Corrosion can become very rapid.

Strong alkalis cause damage to enamel surfaces and can dissolve glass linings.

23. Poor on non-existent heat treatment

Some kinds of piping require heat treatment to relieve stresses arising from welding, or installation stresses. When piping does not “fit”, the job of a pipe fitter sometimes involves making it fit by bending it into place using jacks or chain jacks. Annealing or stress relief should follow such techniques on most heavy steel. Such heat treatment is not always carried out.

Welding causes stress and metal structure changes, particularly in the zone around the weld. With some combinations of steel and liquid, corrosion in the heat affected zone of the metal becomes very rapid unless heat treatment is performed.

24. Fibre reinforced plastic piping.

Plastic piping is vulnerable to fire. It also tends to become brittle with age. This may be due to the effect of chemicals, solution loss of the plasticizer (PVC, not FRP), or to ozone or ultra violet radiation effects from the sun. Embrittlement often leads to cracking at places which are loaded or stressed.

25. Overstressing by valve operation.

Operators can break piping, particularly small diameter pipes up to 1”, and plastic piping, by operating valves using too much force.

26. Climbing on pipes

Operators or technicians climbing on them may break pipes.

27. Using pipes as scaffolding.

Pipes may be broken because riggers use the piping to support equipment during erection or maintenance, or use piping to support gangways. (These practices should be forbidden).

28. Residual contamination,

When acid or alkali enters unsuitable piping by accident, it may be saved by quickly correcting the mistake, and washing out the pipe. There is almost always some residual contamination, though which will continue to cause corrosion problems.

Solutions containing chlorides are especially a problem for stainless steel equipment. Chlorides cause much increased crevice corrosion, and can cause stress corrosion cracking in places where piping is under stress. Once contamination has occurred, it is extremely difficult, and may be impossible, to eliminate it.

29. Screw fittings

Screw fittings are often used to couple pipes, and to attach instruments, valves and fittings such as elbows, to piping. Screwed fittings should preferably be avoided on hazardous materials and especially with flammable materials, because:

- They tend to loosen with vibration
- They are a frequent source of leaks
- The threads are a frequent location for corrosion
- They tend to open up and leak faster when exposed to fire.

30. Pipe supports.

Pipes, which rest on other pipes, on beams, and blocks, tend to corrode at the point of contact. If there is any tendency for external corrosion, a better practice is to fit “shoes”, that is, sections of T beam, welded to the pipe, to provide support.

Supports on beams and blocks should actually provide support. If there is a gap between the pipe (or shoe) and the support, shims (thin strips of steel) should be inserted to ensure support.

Pipes are often hung from rods, rather than being supported on beams. This is particularly the case for boiler piping.

Small-bore piping may be supported on hangers fitted with simple screw turnbuckles, which are tightened to ensure that the support really is present. Tightening should be such that the pipe position is correct, and the pipe does not bend under stress.

Larger piping is provided with spring loaded pipe hangers. It is important that such hangers are adjusted to provide the correct tension. Hangers may need readjustment every few years to correct every few years to correct for settlement or pipe movement.

Figure 7. 8 Pipe supported on an old glove. Eventually the glove will rot away.

31. Pipe sway stops and snubbers

Piping which is hung from support rods may tend to sway or move due to vibration. It can move dramatically under filling, or draining if hammer effects occur.

If pipes sway or move excessively, high stresses are placed on fixed-point supports, or vessel nozzles. Piping which is hung should have end stops or snubbers (shock absorbers) to prevent excessive movement.

32. Electrode burn

Welders sometimes start a welding arc by touching the electrode on a pipe, away from the weld position. The resulting electrode burn is a location for rapid corrosion.

Figure 7.9 Two ends of a pipe section corroded under lagging when water soaked in.

33. Corrosion under lagging

Hot piping under lagging can be subject to extreme corrosion rates. The problem arises because rain often contaminated by industrial salts and acid, penetrates the lagging. Hot acidulated salt solutions provide one of the most rapid corrodants.

34. Lagging fires

Lagging fires or insulation can become wetted with heavy oil or tar. Especially if the pipe is hot, the oil can oxidise, forming acids. These can then oxidise, even more rapidly. The result may be a fire, which can occur even under cladding.

35. Wrong materials

Wrong materials may be delivered to a plant because of mistakes, or deliberate cheating on the part of the supplier (cheating occurs especially if the supplier has a shortfall of the correct grade). Wrong materials can also be installed because of mistakes at the storage yard, or during construction. A particular problem arises when short lengths of piping remain, and on site storage is poorly arranged. Odd pipe lengths can then become mixed.

Wrong materials can be subject to rapid corrosion.

The best practice in use today is to apply spectrometer tests to materials received, and to all suspect materials withdrawn from the stores.

36. Low point corrosion.

If there are low points in a hydrocarbons pipe, this provides a place where water, acids etc. can settle out. If flow is stopped for a long period, the result can be rapid local corrosion.

37. Excessive flow velocity

Erosion and erosive corrosion can occur rapidly if the flow rate in the pipe is too high.

This is especially the case if there are suspended solids in the flow.

38. Ice expansion

Ice formation in water pipes frequently cause cracking in cold climate.

39. Steam impingement

Steam jets from leaking flanges pinholes, and valve seals can cause severe corrosion on adjacent pipes.

40. Foundation subsidence

It is sometimes possible to see significant subsidence under structural steel. When this occurs, piping may become highly stressed, because one end of a span subsides more than the other.

7.2 Case stories

1. Bolts on a kerosene pipe were over-tightened, and later broke due to overstressing and subsequent vibration or expansion stress. A large kerosene fire ensued.

2. During a training course, the use of screw fittings on piping was discussed. As part of the course, a plant inspection was made. Within 5 minutes, a drip of kerosene from a screw fitting onto a hot pump was found.

3. Ethylene under pressure was pumped in a pipe. A high temperature disturbance led to polymerisation and explosion.

4. A run down pipe for crude oil was shut in. It heated in the sun, over-pressured, and burst. The burst was a 3m longitudinal seam opening.

5. Three releases of oil occurred from bottom pumps on different distillation columns. One resulted in a fire. All were due to erosion at the pump discharge line elbow. All were due to use of a wrong grade of steel for the elbow.

6. On a nitric acid plant, a vent line was installed with a wrong grade of steel (carbon steel rather than stainless). The difference was not immediately obvious at audit, because the metal colour was similar when coated with construction dirt.

Six years later, the pipe failed releasing NO_x gases. The corrosion rate for the carbon steel had been lower than might be expected, because only relatively dry gas was involved.

7. A pipeline was thought to be empty but open through to a storage tank, and pumping was begun. In fact, a manual valve was closed. The pumping compressed the air in front of the oil, but this failed to cushion the hammer, when it was relieved through a safety valve. The pipe smashed when the oil hammered into the partially open safety valve.

8. Ice formed under a cold debutaniser downcomer elbow. The plug of ice jammed against a support bracket. Gradual expansion of the ice caused overstressing of the elbow, and caused it to crack.

Ice expansion has also been known to force flanges apart.

9.Compression heating (Kletz, ref. 7.3).

A positive displacement compressor, working against a closed valve, caused compression heating in vapour and air in a fuel pipe. The heating from the air compression resulted in an explosion similar to those in a diesel engine.

7.3 Releases from steel piping

A very large part of the releases in chemical and petroleum plant occur due to failures in piping (a large part of the releases recorded as arising in vessels in fact occur in nozzles and associated piping).

It is convenient in risk assessments to include piping along with unit processes. This means that a certain amount of piping is allocated to each unit process or vessel. This still leaves a certain amount of inter unit piping to be accounted for in the risk assessment. In the later chapters, unit/vessel release frequencies are given both with and without piping included.

The release frequencies for piping are related to hole sizes. Generally, though not always, smaller leaks occur more frequently than large. Determination of hole size distribution is therefore important.

Traditionally, frequencies are given for complete pipe ruptures, for releases of 20 % size, and for small releases of 1-2 mm. Before considering release frequencies, a study was made of the reasonableness of this distribution. Figure 7.10. to 7.13 shows hole size distributions derived from US RMP data, for a number of plant unit types. Table 7.1 shows the percentage break down using "ideal" classifications of the data, using natural groupings as indicated by the data and also the breakdown according to traditional classification. As can be seen, distributions are about 50% "small", under 5 mm, except for the crude units. The explanation for the crude unit exception is that in these units almost all piping is large.

Unit type	Small hole size, < 5 mm	Medium hole size, < 25 mm	Large hole sizes > 25 mm, and ruptures
Ammonia and fertiliser	50	7	43
Alkylation	50	40	10
Crude unit	8	31	61
LPG storage		24	76
Chlor Alkali	44	22	34
Average	40	19	40

Table 7.1 Percentage of failures of different hole sizes, US RMP data (ref. 7.4)

The low number of releases recorded with small hole sizes, for crude units and LPG storage, are without doubt due to the fact that these hole sizes do not lead to offsite consequences for flammables, so the percentage for small holes in table 7.1 id probably low.

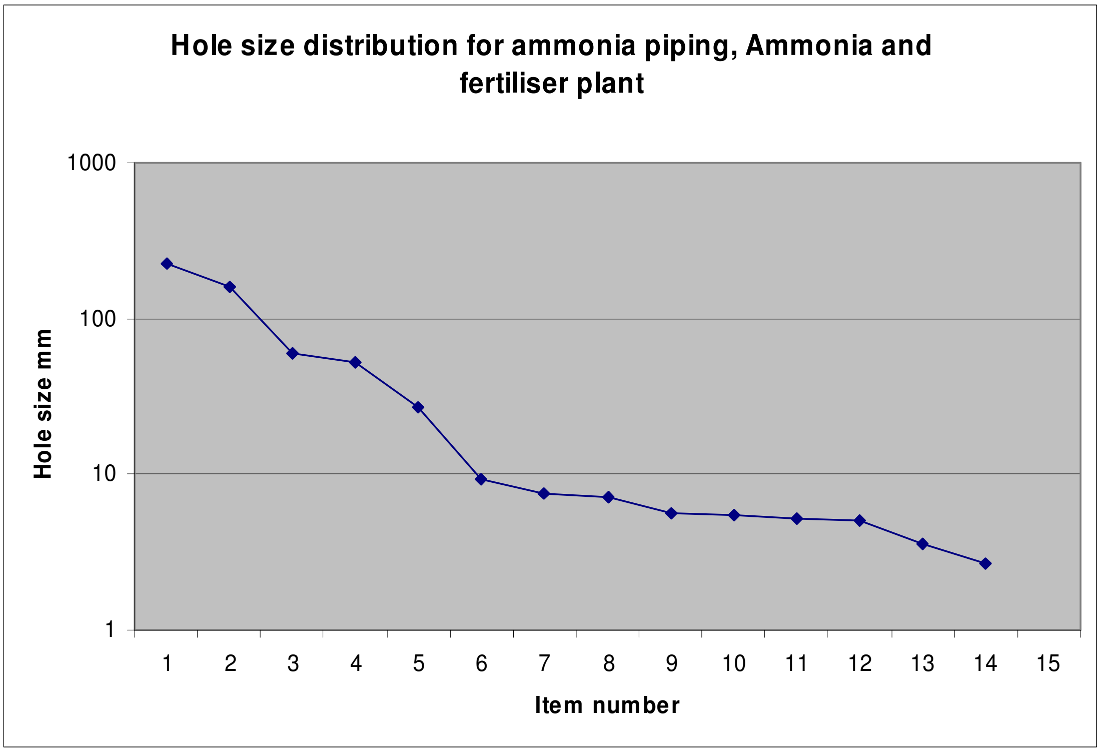


Figure 7.10 Hole sizes in ammonia plant piping, RMP data (ref. 7.4)

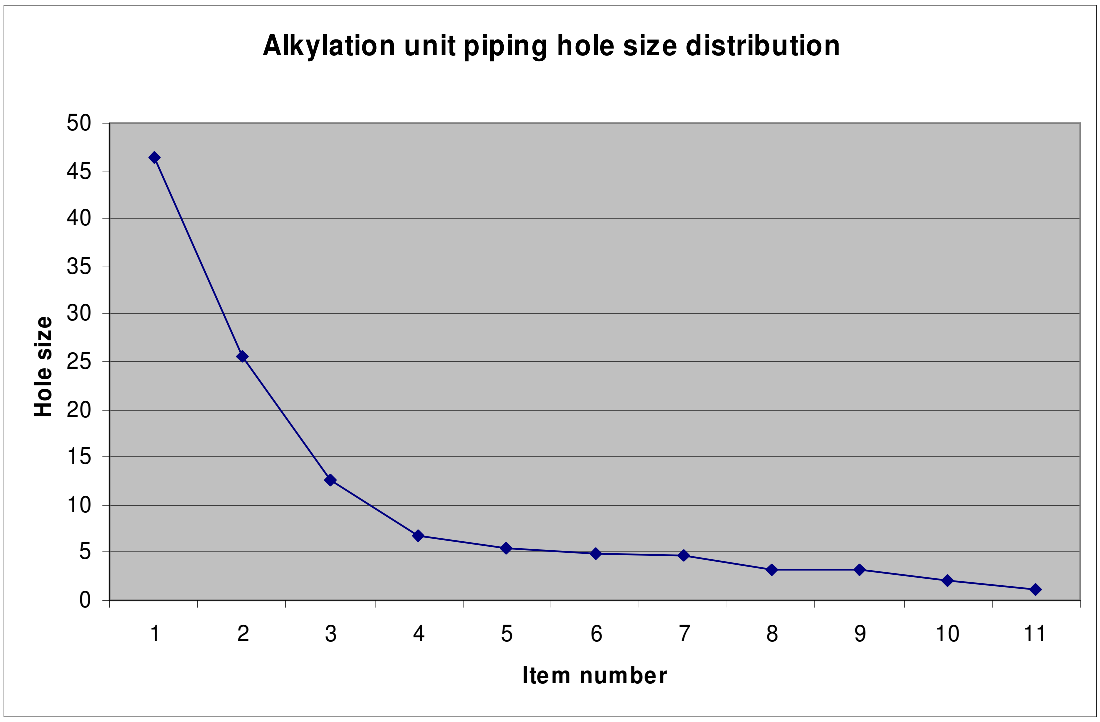


Figure 7.11 Hole sizes in refinery alkylation unit piping, RMP data (ref. 7.4)

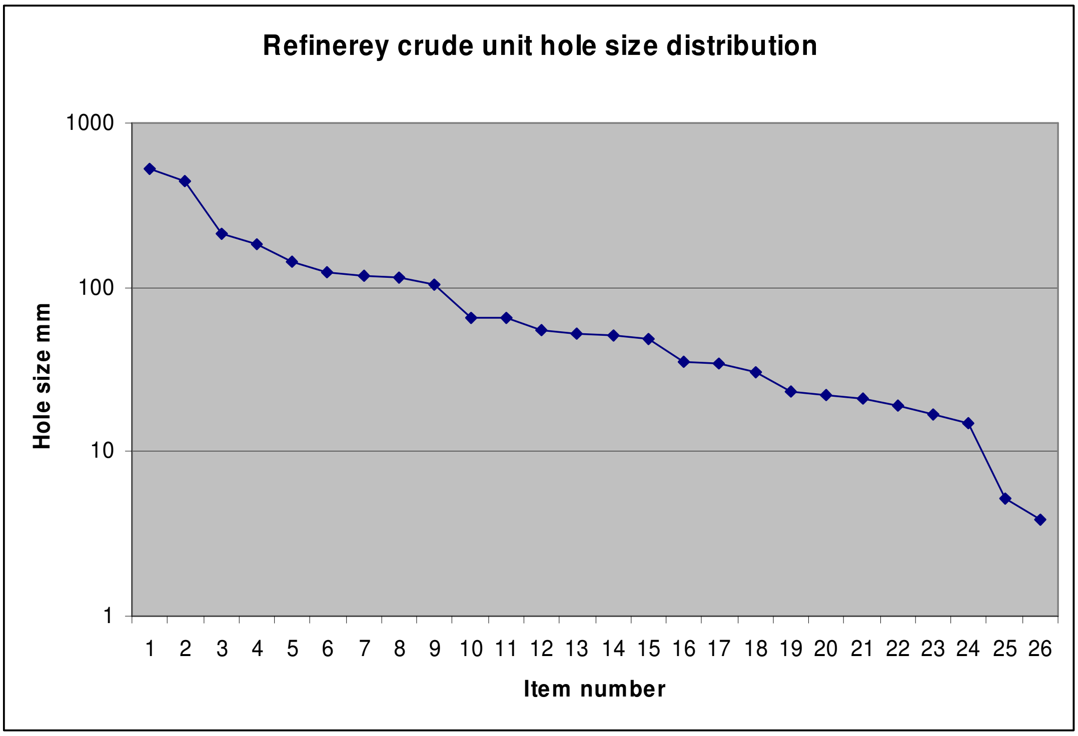


Figure 7.12 Hole sizes in refinery crude unit piping, RMP data (ref. 7.4)

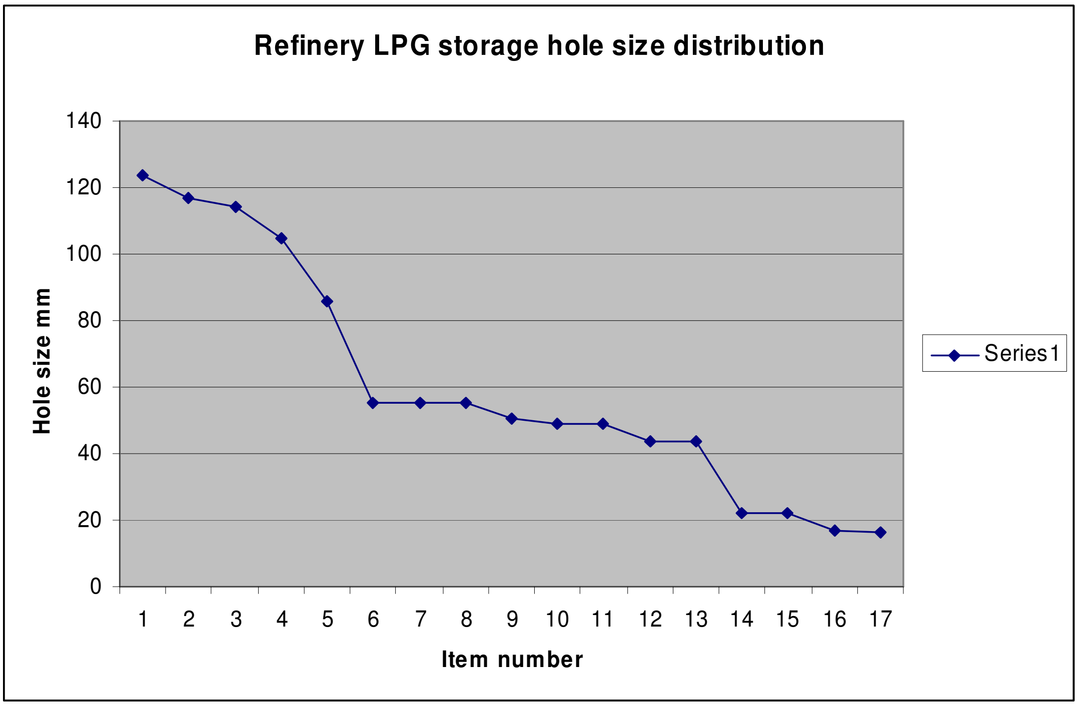


Figure 7.13 Hole sizes in refinery LPG storage piping, RMP data (ref. 7.4)

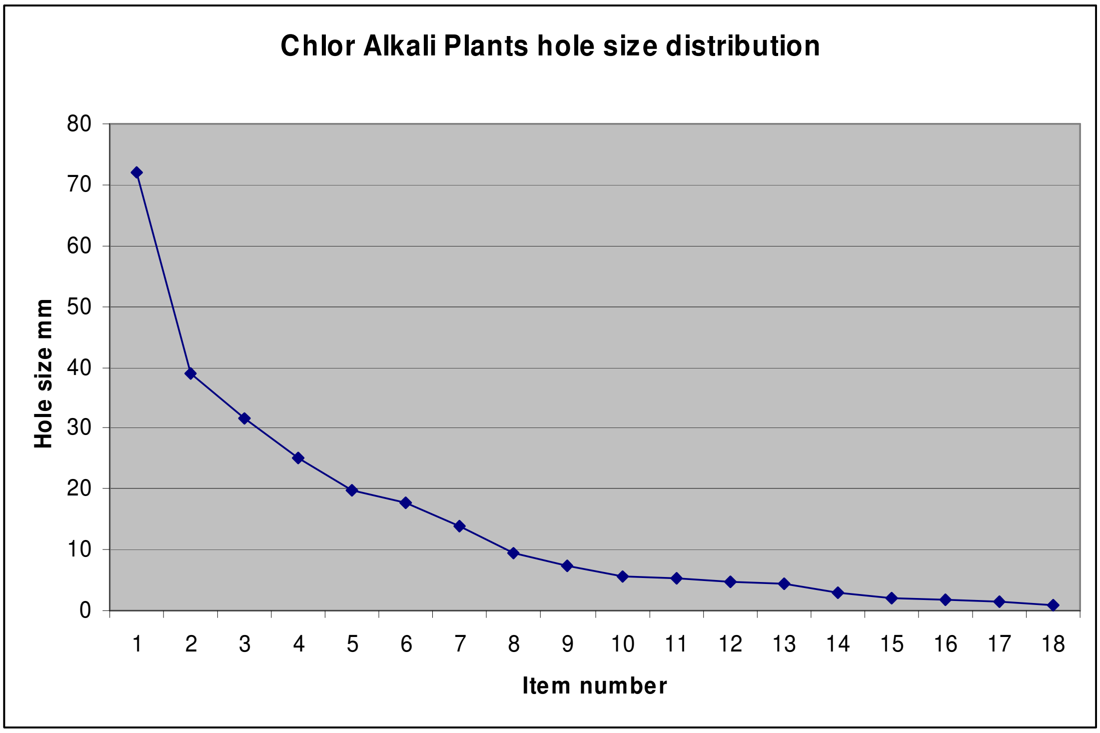


Figure 7.14 Hole sizes in Chlor alkali piping, RMP data (ref. 7.4)

Note that to obtain these data a number of steps were necessary:

1. Groups of reasonably similar unit types, has to be selected from the RMP data. The groups were selected also to be those for which design information such as pressures and typical pipe sizes were available.

2. To determine the frequencies, the length of piping at risk had to be estimated. This was done by taking the average of actual piping lengths for a number of actual units in the given class. For most unit types, values were taken from three units. This step introduces level of uncertainly into the assessment. An estimate of the extent of the uncertainly was derived by looking at the variation of piping lengths and diameters between plants, because unit layouts can vary. For the plant units chosen, the highest variation in unit pipe lengths was 55%.

3. The hole sizes were calculated from release amounts and release period, given in the database. Both of these values are subject to some reporting uncertainty.

7.4 Typical release frequencies

Release frequencies according to the hole size distribution as selected above were calculated for the RMP data. The data are shown in table 7.2 for different plant unit types. The unit types were chosen because the designs for these units is quite standardised, and because there were many of each type in the data base. As can be seen, the variations are wide, much wider than can be explained by uncertainty in the data processing. Note the low small hole frequencies for crude units, LPG strage and gas treatment, which arise from the reporting criteria, with only offsite effercts, injury or severe damage.

Unit type	Total release frequency per m. year	Small hole, < 10 mm per m. year	Medium hole >10mm <25 mm per m. year	Large hole >25 mm <100 mm per m. year	Very large hole, > 100 mm and rupture per m. year
Fertiliser, ammonia	$54 * 10^{-6}$	$27 * 10^{-6}$	$4 * 10^{-6}$	$15 * 10^{-6}$	$8 * 10^{-6}$
Refinery crude unit	$89 * 10^{-6}$	$6.9 * 10^{-6}$	$21 * 10^{-6}$	$31 * 10^{-6}$	$31 * 10^{-6}$
Alkylation unit	$134 * 10^{-6}$	$66 * 10^{-6}$	$54 * 10^{-6}$	$8 * 10^{-6}$	$8 * 10^{-6}$
LPG storage	$42 * 10^{-6}$	-	$9.8 * 10^{-6}$	$22 * 10^{-6}$	$9.8 * 10^{-6}$
Chlor Alkali plant	$280 * 10^{-6}$	$109 * 10^{-6}$	$77 * 10^{-6}$	$31 * 10^{-6}$	-
Refinery light ends	$74 * 10^{-6}$	$37 * 10^{-6}$	$9.2 * 10^{-6}$	$27 * 10^{-6}$	
Gas treatment plant	$66 * 10^{-6}$	$27 * 10^{-6}$	$13 * 10^{-6}$		$27 * 10^{-6}$
HDS, HDT	$370 * 10^{-6}$	$226 * 10^{-6}$	$41 * 10^{-6}$	$103 * 10^{-6}$	
Ammonia distribut'n	$513 * 10^{-6}$	$214 * 10^{-6}$	$171 * 10^{-6}$	$128 * 10^{-6}$	
		< 10 mm	10 to 25 mm	> 25 mm	> 100 mm and rupture
HSE offshore [7.5] <3 inch pipe	$200 * 10^{-6}$	$158 * 10^{-6}$	$26 * 10^{-6}$	$16 * 10^{-6}$	NA
3-11 inch pipe	$58.7 * 10^{-6}$	$44 * 10^{-6}$	$2.9 * 10^{-6}$	$2.3 * 10^{-6}$	$5.9 * 10^{-6}$
>11 inch pipe	$54.8 * 10^{-6}$	$35 * 10^{-6}$	$4.4 * 10^{-6}$	0	$9.3 * 10^{-6}$
Purple book [7.6]	Leak				Rupture
Nom.l dia ≤ 75mm	$5 * 10^{-6}$				$1 * 10^{-6}$
75 mm< Nom. dia <150 mm	$2 * 10^{-6}$				$3 * 10^{-7}$
Nom.l dia / 150mm	$5 * 10^{-7}$				$1 * 10^{-7}$
Pape and Nussey[7.7]	$0.3 * 10^{-6}$				
Lydell[7.8]		$37 * 10^{-6}$		$0.21 * 10^{-6}$	
Cox, Lees and Ang[7.9]				$3 * 10^{-6}$ to $10 * 10^{-6}$	

Table 7.2 Release frequencies, US RMP data and other comparison sources, uncensored.

Also included into the table are data from the Purple Book, for comparison purposes, and data from a number of published sources. Histograms of the data for small releases and leaks, and for large releases and rupture, are given in figures 7.15 and 7.16.

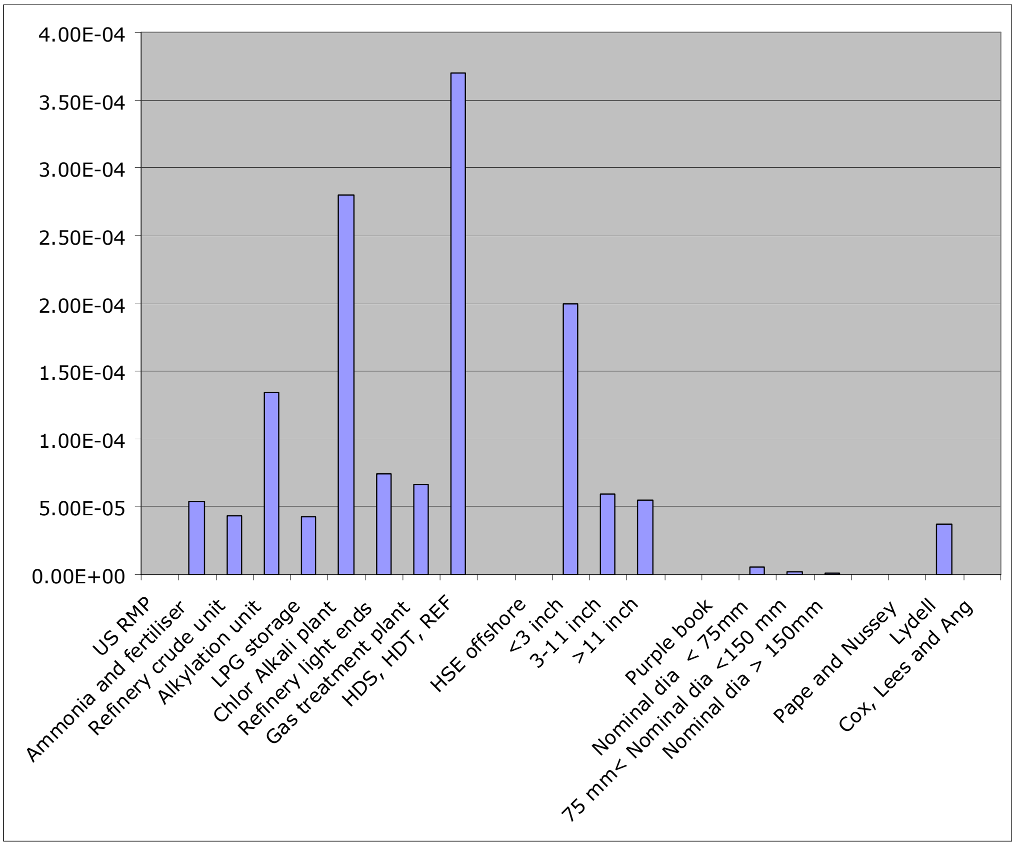


Figure 7.15 Overall leak frequencies, cases per pipe m. year

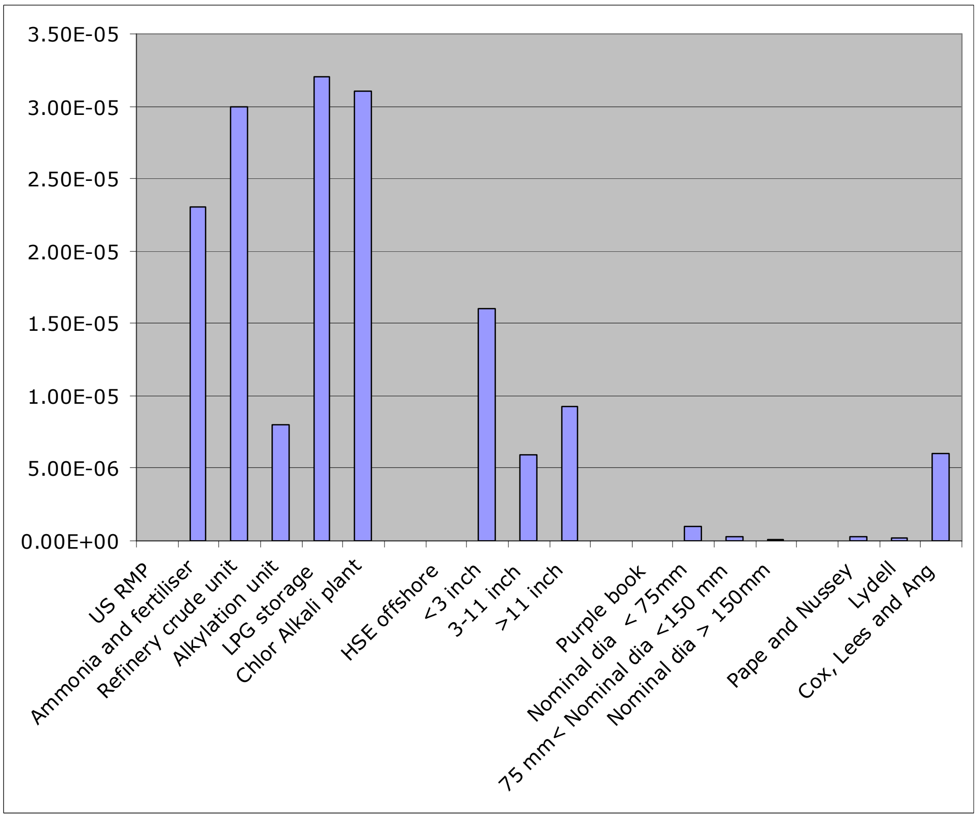


Figure 7.16 Large hole and rupture frequencies, cases per pipe m. year

As can be seen there is a very wide range of frequency values. For large hole sizes, the frequencies vary from 2.6* 10^{-6} per m. year, to 90* 10^{-6} per m. year, i.e. by a factor of 35. This variation is to a large extent real. The Lydell data were derived from a plant with an intense risk centred maintenance program, so the low frequency of large releases is explicable. Many more failures were detected than those used in the statistics here, but all these were recorded as "incipient"i.e. detected by inspection before a release occurred. Note that small leak frequencies are underrepresented for crude units, and LPG. This is due to the reporting criteria, which cover offsite effects, major damage or injury. Small flammable releases are unlikely to cause such effects.

The other main source of variation in the published data is that the RMP data, the HSE offshore data and the Lydell data are derived from actual release records, and these all give relatively high release frequencies. Many of the releases also result from human error or external influences, which are not usually regarded as piping failures when building up failure rate data bases. It seems that the very low values from the Purple Book and from Pape and Nussey, which are widely used is risk assessments, correspond to ideal rates, in the cases where most preventable sources of pipe failure, such as those described in section 7.1, are in fact prevented, and human error effects in piping operation and maintenance are not taken into account.

Some of the releases in the RMP data may also in fact result from flange and gasket failures, since although the data collection instructions identify a category "joints", it is certainly not the case that all engineers would associate the a pipe flange leak with the word "joint", used in the classification scheme.

Source B, which is assessed data gives Table 7.3

Failure size d/D	% dia	Failure rate pr 10^6 m year
< 3"		
V small	5	42
Small	22	18
medium	45	7.0
rupture	100	3.5
4 to 11"		
V small	5	22
Small	22	8.8
medium	45	3.6
rupture	100	1.8
> 12 "		
V small	5	15.8
Small	22	6.8
medium	45	2.7
rupture	100	1.3

Plus failure rates due to secondary failures

Table 7.3 Leak rates for pipes (source B)

Choosing a typical failure rate from these data is largely a question of strategy. Any of the values in table 7.2 or 7.3 could be chosen provided appropriate modification factors were chosen at the same time to account for variations between application. Depending on the value chosen, enough data sets are available to derive suitable modification factors. The best data sets available gave absolute hole sizes rather than sizes as a function of pipe diameter. For this reason, this approach is taken here.

The Purple Book values would be the obvious baseline to choose . The Purple Book values, however, are much lower than all the others, and there is no record of the basis and engineering standard assumed for the data. (The source was described by one of the authors of the Purple book as being initially reference 7.12, but modified over a number of years on the basis of feedback from risk analysis application). The RMP data could well be used as a baseline. However, the RMP data certainly include a large proportion of "special" failure causes, as can be seen by reviewing the cause distributions.

The RMP data has the advantage that it is service specific. The UK HSE data has the advantage that pipe size is recorded (unfortunately, pipe size is not recorded in the US RMP source data, although it can be inferred from the release type, substance and unit type)

From review of piping failure rate data from a number of plants with extensive risk bases inspection programs, it appears that the frequency of failure can be reduced to very low levels by means of frequent audit, non destructive testing, and internal inspections. However, with a reasonable investment of effort, involving yearly visual inspection, and NDT on a few points on each pipe spool once every one to three years, a failure rate of $30*10^{-6}$ is achievable for small diameter pipes in clean non corrosive conditions, and correspondingly $6*10^{-6}$ for pipes over 3 inch diameter. Typical data are given in table 7.4, using this value and using hole size distributions derived from the RMP dtata. The typical values are selected as a weighted average of the values for all of the RMP piping release frequencies calculated, with weighting according to pipe length. The large incidence of crude unit piping ruptures was deleted from the averaging process, since these are anomalous, and a special rule is added for proper treatment of crude unit piping rupture in section 7.9

Unit type	Total release frequency per m. year	Small hole, < 10 mm	Medium hole >10mm <25 mm	Large hole >25 mm <100 mm	Very large hole, >100 mm and rupture
≤3 inch pipe	$129* 10^{-6}$	$55* 10^{-6}$	$52* 10^{-6}$	$22* 10^{-6}$	NA
>3 inch pipe	$49* 10^{-6}$	$18* 10^{-6}$	$17* 10^{-6}$	$6.3* 10^{-6}$	$7.3* 10^{-6}$
cf. HSE < 3 inch	$200* 10^{-6}$	$158* 10^{-6}$	$26* 10^{-6}$	$16* 10^{-6}$	NA
cf. HSE 3-11 inch	$58.7* 10^{-6}$	$44* 10^{-6}$	$2.9* 10^{-6}$	$2.3* 10^{-6}$	$5.9* 10^{-6}$

Table 7.4 Proposed typical failure rates, based on US RMP data

The assumptions which should accompany these failure rates are given in section 7.9.

7.5 Welds

Welds represent one of the weakest points in piping, due to the problems of weld inclusions, material differences, pits and holes, and of unavoidable structural changes to material under the heat of welding. It is probable that many of the failures in piping described earlier in fact occur at welds. Nevertheless, more accurate results are obtained if failure rates specifically for welds are added to the overall failure rates for pipes, at least in the cases where there are many welds. Failure modes concern leakage and breakage. WASH 1400 (ref. 7.10) data, which concerns high pressure, moderately high temperature steam and water service, is reflected in table 7.5.

Failure Mode	**Failure Rate per year**
Leakage (less than 10% of weld area)	$2.6*10^{-4}$
Breakage	$2.6*10^{-5}$

Table 7.5 Weld failure rates for larger pipe sizes (1 ½” to 24”)

7.6 Flanges

The most important failure modes for flanges are leak or breakdown.

The causes may be many

- The flange can be too loose due to the fact that screws or bolts are not tightened, or because heating, overpressuring, vibration, mechanical loads have lengthened bolts or opened the flange .
- Overstressing of bolts can cause breakage. Pipe expansion which has not been allowed for, or jammed hangers, can bow pipes and cause opening of flanges
- Sometimes too few bolts will be fitted or some will be left untightened, during installation or maintenance, as a result of error. The remainder may be overstressed.
- Corrosion products in the flange opening itself can force flanges apart.
- Corrosion or aging can weaken packings.
- Corrosion under a packing, particularly if the packing is scored, can lead to a leakage route around the packing.
- Foreign bodies, mishandling, dirt, or burrs on the flanges can cause damage to seals under installation.
- Bolts can break through overstressing, fatigue or corrosion.

- Sometimes flanges are simply unbolted in error by maintenance staff while the system is under pressure, or remains undrained.

- Flanges can fail as the result of other failures. For example packings can be blown out by overpressure (unless recessed) or be destroyed by fire.

- Overpressuring by hammering is a common cause of flange failure in piping in which gas and liquid flow intermittently. Steam condensate piping, for example, has flange failure rates which are one or two orders of magnitude higher than is typical for other piping, because of the hammering that occurs when steam traps open and close.

Small leaks of corrosive liquors such as hot, salt laden water under pressure, can corrode or erode paths over flange faces, so that once a leak has started, a large leak is formed in the course of a few hours.

Table 7.6.1 gives failure rates for failure modes which do not involve human errors or are not the result of secondary failures.

Source	Failure Mode	No. of failures	Failure Rate
TA	Small leak (0.1-10 mm^2 hole), 2 to 6 inch pipe		0.0035
	Small leak (0.1-10 mm^2 hole), high pressure pipe, 2 to 6 inch pipe		0.00035
	Packing blown out or partially destroyed, or flange opens several mm (not recessed flanges)		0.00026
	Flange breaks open		$8.8*10^{-5}$
Company B	Hole size 0.1* diameter		$880*10^{-6}$
	Hole size 1.0* diameter		$3.5*10^{-6}$
HSE offshore	Pipe < 3 inch	110	
(ref. 7.5)	Small hole < 10 mm		$31*10^{-6}$
	Medium hole 10 to 25 mm		$4.0*10^{-6}$
	Large hole 25 to 100 mm		$4.4*10^{-6}$
	Very large hole, > 100 mm & rupture		$0. 4*10^{-6}$
	Pipe 3 to 11 inch	98	
	Small hole < 10 mm		$47*10^{-6}$
	Medium hole 10 to 25 mm		$28*10^{-6}$
	Large hole 25 to 100 mm		$28*10^{-6}$
	Very large hole, > 100 mm & rupture		$3.6*10^{-6}$
	Pipe > 11 inch		
	Small hole < 10 mm		$84*10^{-6}$
	Medium hole 10 to 25 mm		$4.0*10^{-6}$
	Large hole 25 to 100 mm		$4.0*10^{-6}$
	Very large hole, > 100 mm & rupture		$6.9*10^{-6}$

Table 7.6 Failure rates for flanges

A problem with the first two data collections is that they include some data from steam and condensate piping, which are notoriously hard on flanges. The most representative data available appears to be the HSE data, with values for 3 to 11 inch piping being the most widely applicable. The rounded values are as given in table 7.7:

Unit type	Total release frequency per year	Small hole, < 5 mm	Medium hole 5mm to 25 mm	Large hole >25 mm	Very large hole, > 100 mm and rupture
3-11 inch pipe, hydrocarbon processing up to 50 °C	$56* 10^{-6}$	$47* 10^{-6}$	$0.28* 10^{-6}$	$0.28* 10^{-6}$	$0.36* 10^{-6}$

Table 7.7 Proposed baseline failure rates for flanges

7.7 Small Bore Fittings

Small bore screw and nipple fittings are widely used for piping up to 2 inches, where disassembly for maintenance is desired (primarily by maintenance teams). They are also used instead of welds when repairs or modifications are to be made and hot work needs to be avoided. Screw type fittings should never be used for pipes carrying flammable liquids unless they are seal welded, but unfortunately the often are used especially to avoid the need for welding inside the unit.

Figure 7.17 Small bore fittings alongside flanged connections

Failure modes for small bore fittings include:

- Leakage due to inadequate tightening
- Leakage due to loosening by vibration or impact
- Breakage due to corrosion at the thread
- Breakage due to vibration fatigue

- Breakage due to overloading

- Burst open on pressuring (nipple type) due to inadequate installation or specification error

Failure rates are given in table 7.8

Failure Mode	**Source B**	**CCPS**	**Cox, Lees and Ang**
Leakage 0.1 *D	$6.0 * 10^{-5}$		Pneumatic connector 0.013
Leakage 0.2 *D	$3.3 * 10^{-5}$		$5 * 10^{-3}$
Breakage	$3.6 * 10^{-4}$	$5 * 10^{-3}$	$5 * 10^{-4}$

Table 7.8 Failure rates per year for small bore fittings, per year, D is pipe diameter

These data are quite hard to interpret, showing wide variation, but a value for significant leakage of $5*10^{-4}$ per year seems a reasonable typical value.

7.8 Causes of piping failure

In order to be able to adapt the data to a specific application, it is necessary to know the causes of the failures. For example, if the specific plant has high quality piping, with an additional corrosion allowance beyond that which is normal, then the contribution of corrosion to the failure frequency will be reduced or even eliminated completely.

Figure 7.18 and 7.19 shows safety barrier diagrams for piping failures, showing the causes at two levels – mechanism of failure, and root cause of failure. The diagram was built up from experience, but was then expanded by reviewing accident reports.

The diagram becomes useful once it has been quantified. Two sources of data for this were used. The US RMP reports give root cause indications (as assessed by the companies). For piping in different plant unit types, the root cause distribution is shown in figure 7.20.

The cause classification in the US RMP data base is not very detailed (10 classes plus “other”), and interpretation of the simple yes/no recording in the data base can be difficult. For this reason about 2000 of the original narrative reports were reviewed. About 20% of these include good narrative descriptions of causes of the releases (most just state that root cause analyses were made). The percentage of causal descriptions is not high enough to justify ist use as a statistical database, but it could be used to check that other data bases are qualitatively consistent. Also, the narrative reports were used to check the large hole/rupture releases, in particular for crude units.

The other sources of information, which have been used here, are accident reports in the accident databases. In particular data from the MHIDAS data base is given in figure 7.22

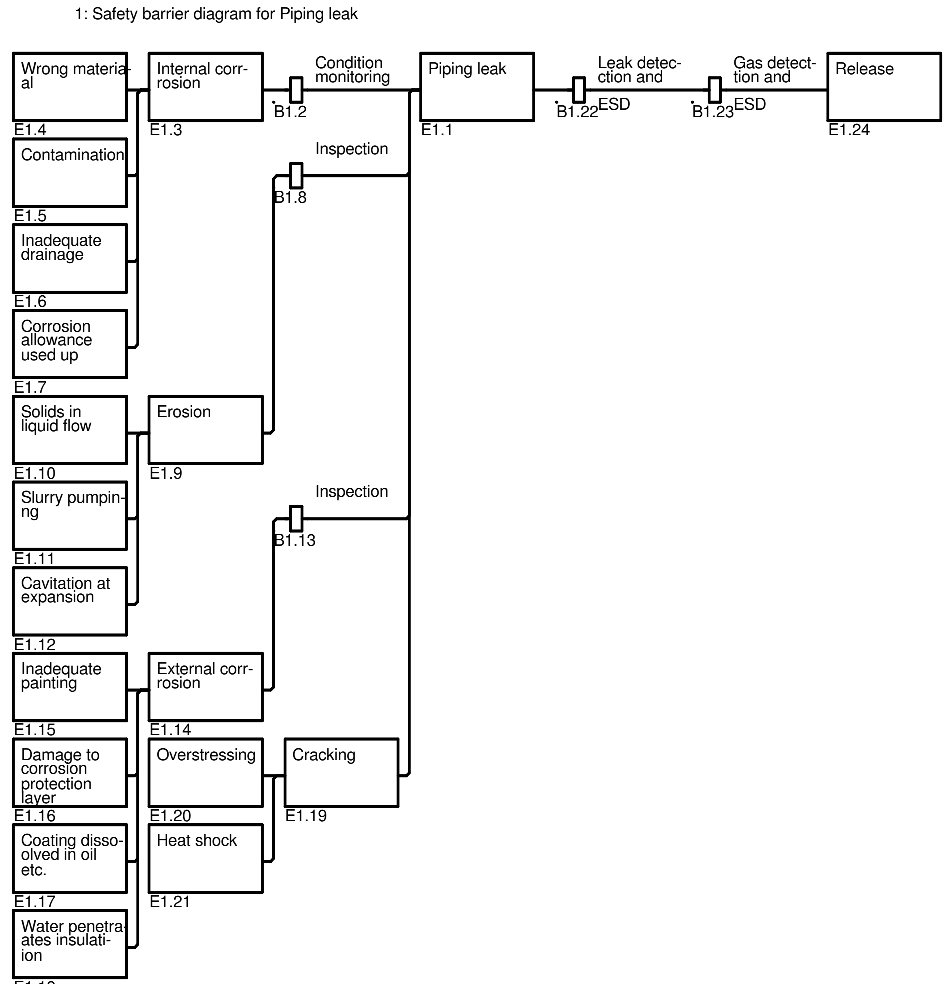


Figure 7.18 Causes of piping leaks

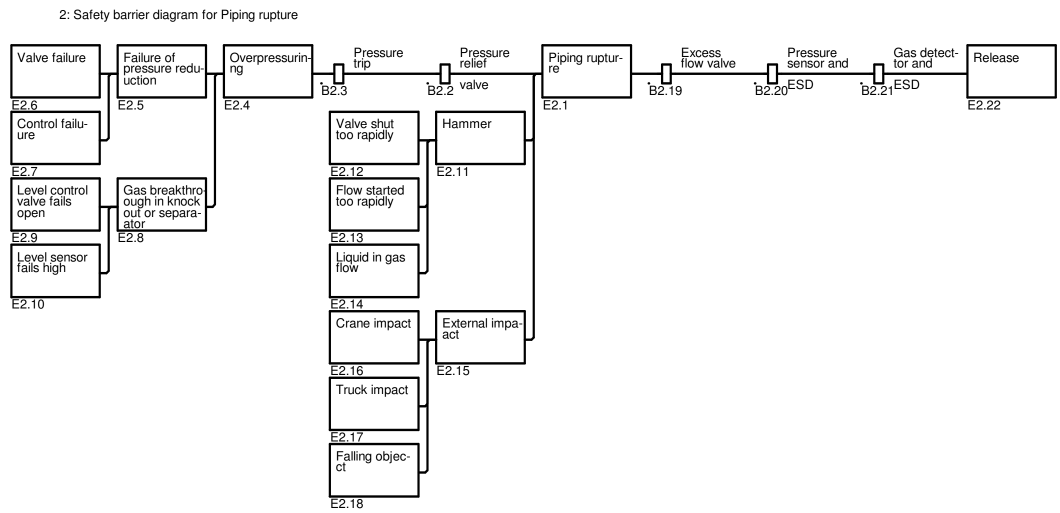


Figure 7.19 Causes of piping breaks (Part 1)

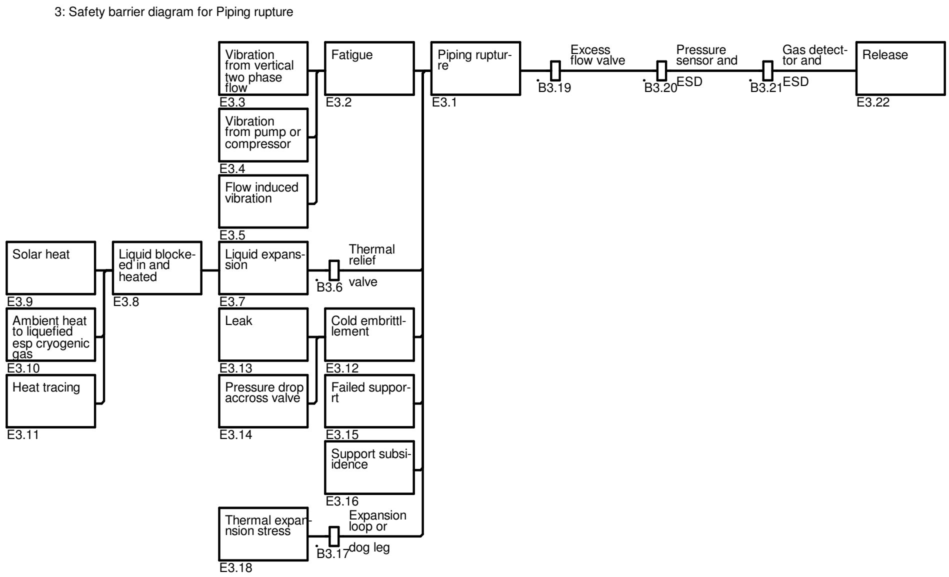


Figure 7.19 Causes of piping breaks (Part 2)

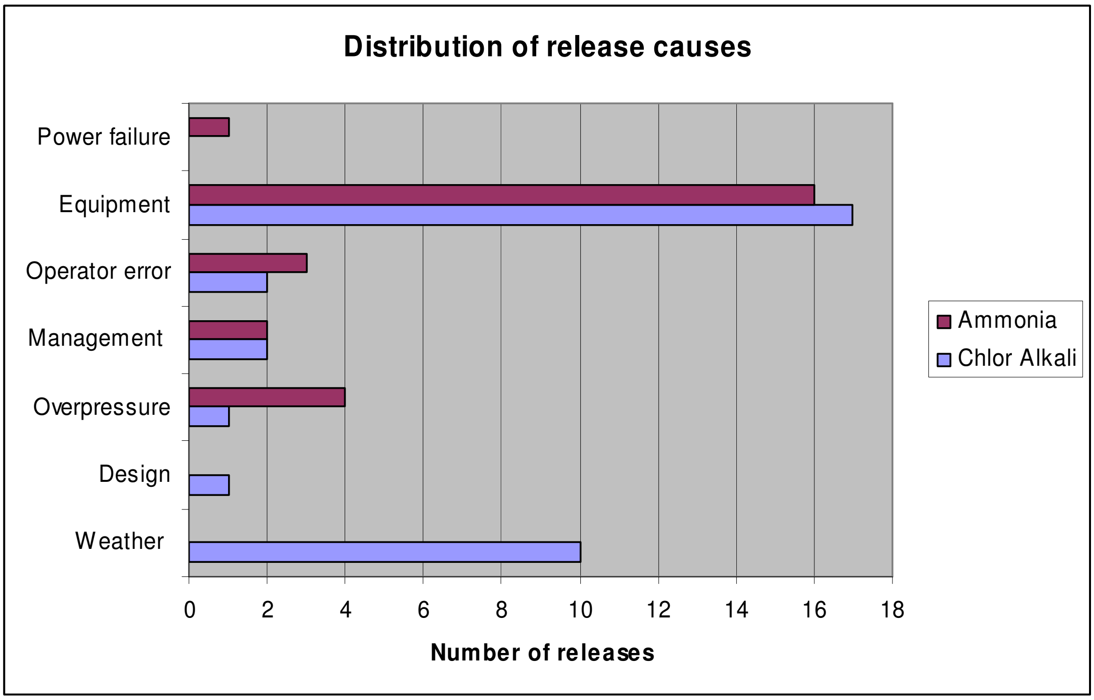


Figure 7.20 Distribution of causes for some piping releases, Ammonia and Chlor Alkali plant , US RMP data. ref 7.4

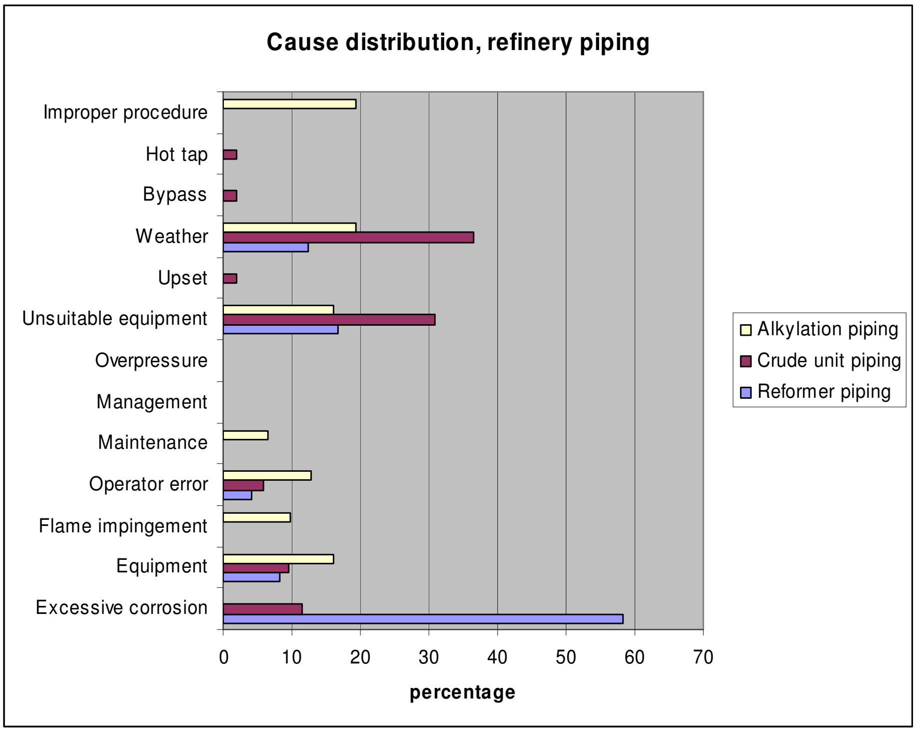


Figure 7.21 Cause distribution for refinery piping releases, US RMP data ref 7.4

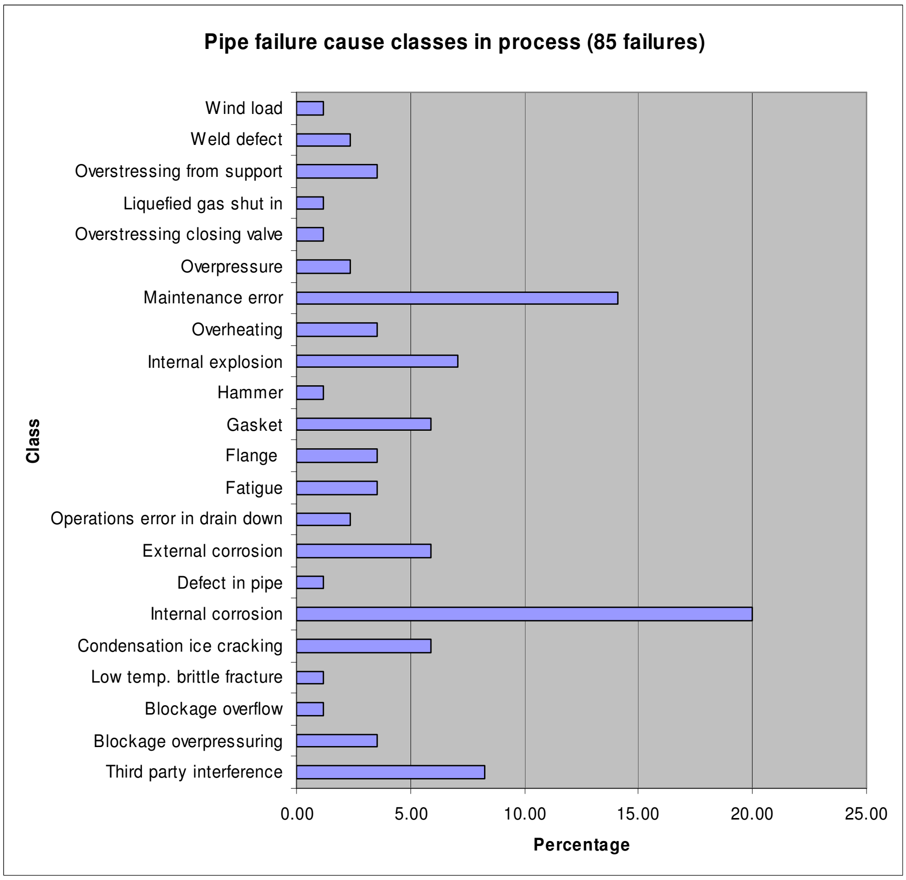


Figure 7.22 Causes of piping failures and releases, MHIDAS data base

As can be seen, it is possible to give causal qualification of the release frequency data. In order for this qualification to become operational, it is necessary to discuss each of the cause types in turn, and to arrive at an algorithm for allocating appropriate release frequencies.

7.9 Assessment of causal factors and susceptibilities

In order to be able to derive release frequencies for a number of different applications and operational conditions, it is necessary to have a baseline release frequency, and a number of additional contributions which will vary according to the different susceptibilities in the installations. The typical values could in principle be chosen in any way, and then contributory values could be added or subtracted as appropriate to account for specific plant types an services. It is most convenient, though, to have a values which are typical for a plant if all of the recognised "special" release frequency contributions have been eliminated, but which still includes the releases which correspond to a normal or typical level of engineering practice.

Typical data will be those associated with the equipment itself, rather than the system into which the equipment is incorporated, and will be those failures which are "inherent" i.e which cannot be avoided except by using unusual and non standard engineering measures. Examples of typical failure causes in piping are those due to inherent variations in corrosion rates, random faults in welding which are not detected by NDT, material flaws, and fatigue failures arising from unnoticed vibration.

Some piping will be especially susceptible to special failure causes. These will have an increased failure rate. Such increases can in many cases be directly estimated by performing a failure probability analysis for the cause itself. For example, a pipe with an upstream pressure regulation, and a pressure specification break will be susceptible to overpressure rupture. The frequency of pressure regulation failure is about 1 per 20 years with modern instrumentation and control. Most such pipes though are protected by safety valves, which reduces risk by a factor from this source by a factor of about 20 (see section 5.5 in volume 1 of this report). The additional risk is then about $2.5*10^{-3}$ per pipe section year from this cause for pipes with no additional protection except the usual safety valve. The risk for pipe rupture from this source for a susceptible pipe is then about $1*10^{-4}$ per metre year, assuming 20 m. pipe sections. This is much higher than the baseline rupture frequency of $6*10^{-6}$ per year (see table 7.4). For this reason, if the problem is recognised, it is usual to provide dual safety measures, such as an ESD valve as well as a safety valve on such pipes. If such additional safety devices are fitted, the rupture frequency for the pipe due to overpressuring will be about $1*10^{-6}$ per m. year i.e. a fraction of the typical failure rate. A judgement about susceptibility for this cause will need to take into account not only whether overpressuring is physically possible, but also the degree to which protection has been provided.

Although such calculations can be made on a case by case basis, the amount of work involved is very large. The work can be carried out using standardised fault trees, such as those in section 7.8 above. A more direct approach is provided in this section and section 7.11, with a table look up and simple check list algorithm to determine modification factors. The method used to develop the modification factors was to use the cause distribution given in section 7.8, primarily from the MHIDAS data base, determine which of these causes apply to the majority of pipelines, and allocate these as baseline causes. Special causes which only apply to a fraction of the pipelines are then regarded as potential reasons for modification of the baseline data. In order to determine the modification factor, it is necessary to know not only the fraction of failures arising from the particular cause, but also, how many items from the data base used to establish the baseline, were actually susceptible to the cause. The susceptibilities were assessed by a review of piping and instrumentation diagrams, a review of audit reports, and a review of photographs of plant equipment for 10 typical plants.

Note that there are several assumptions involved in this process:

1. That the cause distribution from MHIDAS applies also to the RMP and other data used to establish the baseline. This was checked.

2. That the plants used to establish the values have similar susceptibilities to those reviewed in determining the susceptibility factors. Note that this is less uncertain than it sounds, because refinery, ammonia and chlorine plant were used as the primary references. These have reasonably similar designs and operating conditions.

The detailed analysis in the following section gives some insight into the relative importance of the different causes, and provides a basis for checking the modification factors.

Note that special cause contributions are calculated here on the basis of frequency per pipe m. year, even though many of the causes apply to entire pipe sections, rather than to individual pipe metres. For example, the frequency of overpressuring, or of accelerated corrosion due to contamination with a highly corrosive substance will apply to the entire pipe section. The choice of giving the data on the basis of metre years is to provide an easier calculation, and also because but it does mean that the contribution for some special causes will be overestimated for long pipelines, and underestimated for short ones. The actual length of the pipes to which the data apply directly is estimated to be about 20 m.

The overall failure rate for a pipe which is susceptible to all the failure causes, and has an average level of safety protection, will be $200*10^{-6}$ per m. year for pipes over 3 inches, and $200*10^{-6}$ per m. year for small pipes. Much higher values than this have been observed in RBI studies, up to $3000*10^{-6}$ per m. year. Such values can be obtained, for example by increasing the susceptibility value for corrosion to a value greater than 1, for heavily corroded pipes which are beyond their design life age. In general the pipe failure rates will be lower than those recorded here for a secure and well designed and operated modern plant. It this context it should be remembered that the data collected derive from a wide range of plants, and that the largest contribution of data comes from older and less protected plant.

Values for cause specific failure frequencies are given in figure 7.9. The algorithm for determination of the failure frequencies used to derive table 7.9 is:

1. Determine the percentage of failures from each particular cause.

2. Determine the susceptibility to the specific kind of failure for plants surveyed in the causal analysis. The susceptibility is the fraction of plants in the causal survey for which the specific failure type could be possible. This is based on a survey of a representative sample of plants.

3. Determine the safety measures typically present for the plants surveyed in the causal analysis, and the risk reduction factor arising from these safety measures, for the specific accident cause.

4. Determine the release frequency contribution for the specific cause by multiplying the percentage by the overall release frequency and dividing by the susceptibility and safety measure unavailabilities.

The values found in this way should allow the frequencies for typical plants as derived for the RMP data for example to be reproduced when typical safety measures and susceptibilities apply.

Values for small diameter piping less than 3 inch, are given in table 7.10. Note then the susceptibility factors have been adjusted in some cases, as well as the typical failure frequency, when compared with the values for 3 inch an over piping.

The overall frequency of failures calculated in the tables can be very high. The highest values will apply though only to pipes that have every recorded design weakness and susceptibility. The highest single contribution for example will be for a liquefied gas pipe with no thermal relief.

Chapter 23 contains a section validating the values in the tables given here.

Pipes 3 inch +					
Failure rates per year	**Small**	**Medium**	**Large**	**Rupture**	**Total**
Typical	1.80E-05	1.70E-05	6.30E-06	7.30E-06	4.86E-05

No	Failure cause	% of releases	Source	Release size	Suscept-ibility	Safety measure unavail-ability	Failure rate per m. yr.	Basis for susceptibility assessment
1	Internal corrosion	7.4	MHIDAS	small	1	1	4.5E-06	Ammonia not susceptible, LPG low susceptibility
2	Internal corrosion	5.6	MHIDAS	medium	1	1	1.2E-05	All pipes subject to corrosion
3	Internal corrosion	3	MHIDAS	large	1	1	1.8E-06	
4	External corrosion	2.2	MHIDAS	small	1	1	1.3E-06	
5	External corrosion	1.7	MHIDAS	medium	1	1	3.8E-06	
6	External corrosion	0.87	MHIDAS	large	1	1	5.3E-07	
7	Drain lines left open	2.35	MHIDAS	large	0.2	1	7.2E-06	About 1/5 of pipes had drains use by OP's on a selection of refinery P&ID's
8	Maintenance error	14.1	MHIDAS	small	1	1	8.5E-06	All pipes subject to maintenance
9	Corrosion, no inspection	3	Source B	small	0.005	1	3.6E-04	About 2% of companies audited had no effective pipe inspection
10	Corrosion, corrosive liquid, sour gas	3	Source B	small	0.01	1	1.8E-04	Higly corrosive liquids (acid) in about 0.3% of pipes
11	Under lagging corrosion	0.5	Source B	large	0.03	1	1.0E-05	Only for insulated pipe
12	Erosion	0.3	Source B	medium	0.003	1	2.2E-04	About 1 in 300 pipes with erosive conditions
13	Wrong material	0.1	Source B	large	0.03	1	2.0E-06	Poor material control in about 5% of construction
14	Lining failure	0.1	Source B	medium	0.005	1	4.4E-05	About 0.5% of pipes in study lined
15	Support failure	3.5	MHIDAS	large	0.3	1	7.1E-06	Most pipes susceptible
16	Overheating	3.5	MHIDAS	rupture	0.05	1	8.7E-05	About 5% of pipes able to be overheated by fired heaters
17	Overpressure, control failure, single protected	2.35	MHIDAS	rupture	0.2	0.0518	2.8E-04	About 3% of pipes with compressor or PD pump. Safety valve fitted to most.
18	Overpressure, gas breakthrough, single protected	2.4	Source B	rupture	0.02	0.1518	9.9E-04	About 2% of pipes with pressure let down and spec break. Safety valve fitted to most.
19	Overpressure, shut in liquid, no thermal relief	1.17	Source B	medium	0.005	0.1518	3.4E-03	Liquefied gas pipes with no thermal relief. Safety valve fitted to most.
20	External fire	0.5	Special alg	rupture	0.7	1	8.9E-07	Most of the pipes in the study contained flammables
21	Weld crack	2.35	MHIDAS	large	1	1	1.4E-06	Virtually all pipes susceptible
22	Hammer	1.2	MHIDAS	rupture	0.01	1	1.5E-04	Liquefied gas pipes and steam condensate pipes
23	Weather, freezing	5.9	MHIDAS	medium	0.5	1	2.6E-05	Depends on region from which data were obtained
24	Crash, impact	0.3	Source B	large	0.05	1	3.7E-06	About 5% of pipes wer on pipe bridges over roads, or alongside pipe trenches
25	Vibration fatigue	3.5	MHIDAS	large	0.2	1	1.1E-05	Pipes near compressors
26	Thermal expansion	0.2	MHIDAS	large	0.01	1	1.2E-05	Steam and hig temp. Pipes
27	Wind load	1.18	MHIDAS	large	0.005	1	1.4E-04	Depends on region
28	Wrong substance	0.01	MHIDAS	medium	0.01	1	2.2E-06	About 2% of pipes manifolded
29	Earthquake, landslip, flood	0.01	MHIDAS	rupture	0.1	1	1.2E-07	Depends on region, pipes on US West Coast
30	Internal explosion	7.06	MHIDAS	rupture	0.1	1	8.8E-05	Pipes in fine chemicals, vent lines, during maintenance
31	Vandalism, third party	8.24	MHIDAS	rupture	0.1	1	1.0E-04	Pipes exposed to security risk
32	Low temperature embrittlement	0.04	Source B	rupture	0.02	1.0	2.5E-06	Cryogenic systems, liquefied gas systems, high pressure gas regulation with ordinary carbon steel
33	Dropped object	0.1	E&P forum	large	0.05	1.0	1.2E-06	Equipment under crane etc.
34	Design error	13	RMP	large	1	1.0	7.9E-06	All piping, depending on design review and insection
	Total	100.73						

Table 7.9 Modification factors for the release frequencies, based on cause statistics for over 3 inch pipes

Pipes < 3 inch					
Failure rates per year	**Small**	**Medium**	**Large**	**Rupture**	**Total**
Typical	5.50E-05	5.20E-05	NA	2.20E-05	1.29E-04

No	Failure cause	% of releases	Source	Conse-quence	Suscept-ibility	Safety measure unavail-ability	Failure rate per m. yr.	Basis for susceptibility assessment
1	Internal corrosion	7.4	MHIDAS	small	1	1	1.4E-05	Ammonia not susceptible, LPG low susceptibility
2	Internal corrosion	5.6	MHIDAS	medium	1	1	3.8E-05	
3	Internal corrosion	3	MHIDAS	rupture	1	1	4.1E-06	
4	External corrosion	2.2	MHIDAS	small	1	1	4.1E-06	All pipes subject to corrosion
5	External corrosion	1.7	MHIDAS	medium	1	1	1.1E-05	
6	External corrosion	0.87	MHIDAS	rupture	1	1	1.2E-06	
7	Drain lines left open	2.35	MHIDAS	rupture	0.2	1	1.6E-05	About 1/5 of pipes had drains use by OP's on a selection of refinery P&ID's
8	Maintenance error	14.1	MHIDAS	small	1	1	2.6E-05	All pipes subject to maintenance
9	Corrosion, no inspection	3	Source B	small	0.05	1	1.1E-04	About 2% of companies audited had no effective pipe inspection
10	Corrosion, corrosive liquid, sour gas	3	Source B	small	0.01	1	5.6E-04	Higly corrosive liquids (acid) in about 0.3% of pipes
11	Under lagging corrosion	0.5	Source B	rupture	0.003	1	2.3E-04	Only for insulated pipe
12	Erosion	0.3	Source B	medium	0.003	1	6.8E-04	About 1 in 300 pipes with erosive conditions
13	Wrong material	0.1	Source B	rupture	0.03	1	4.5E-06	Poor material control in about 5% of construction
14	Lining failure	0.1	Source B	medium	0.005	1	1.4E-04	About 0.5% of pipes in study lined
15	Support failure	3.5	MHIDAS	rupture	0.3	1	1.6E-05	Most pipes susceptible
16	Overheating	3.5	MHIDAS	rupture	0.05	1	9.5E-05	About 5% of pipes able to be overheated by fired heaters
17	Overpressure, control failure, single protected	2.35	MHIDAS	rupture	0.2	0.0518	3.1E-04	About 3% of pipes with compressor or PD pump. Safety valve fitted to most.
18	Overpressure, gas breakthrough, single protected	2.4	Source B	rupture	0.02	0.1518	1.1E-03	About 2% of pipes with pressure let down and spec break. Safety valve fitted to most.
19	Overpressure, shut in liquid, no thermal relief	1.17	Source B	medium	0.005	0.1518	1.0E-02	Liquefied gas pipes with no thermal relief. Safety valve fitted to most.
20	External fire	0.5	Special alg	rupture	0.7	1	9.7E-07	Most of the pipes in the study contained flammables
21	Weld crack	2.35	MHIDAS	rupture	1	1	3.2E-06	Virtually all pipes susceptible
22	Hammer	1.2	MHIDAS	rupture	0.01	1	1.6E-04	Liquefied gas pipes and steam condensate pipes
23	Weather, freezing	5.9	MHIDAS	medium	0.5	1	8.0E-05	Depends on region from which data were obtained
24	Crash, impact	0.3	Source B	rupture	0.05	1	8.2E-06	About 5% of pipes wer on pipe bridges over roads, or alongside pipe trenches
25	Vibration fatigue	3.5	MHIDAS	rupture	0.2	1	2.4E-05	Pipes near compressors
26	Thermal expansion	0.2	MHIDAS	rupture	0.01	1	2.7E-05	Steam and hig temp. Pipes
27	Wind load	1.18	MHIDAS	rupture	0.005	1	3.2E-04	Depends on region
28	Wrong substance	0.01	MHIDAS	medium	0.01	1	6.8E-06	About 2% of pipes manifolded
29	Earthquake, landslip, flood	0.01	MHIDAS	rupture	0.1	1	1.4E-07	Depends on region, pipes on US West Coast
30	Internal explosion	7.06	MHIDAS	rupture	0.1	1	9.6E-05	Pipes in fine chemicals, vent lines, during maintenance
31	Vandalism, third party	8.24	MHIDAS	rupture	0.1	1	1.1E-04	Pipes exposed to security risk
32	Low temperature embrittlement	0.04	Source B	rupture	0.02	1.0	2.7E-06	Cryogenic systems, liquefied gas systems, high pressure gas regulation with ordinary carbon steel
33	Dropped object	0.1	E&P forum	rupture	0.05	1.0	2.7E-06	Equipment under crane etc.
34	Design error	13	RMP	rupture	1	1.0	1.8E-05	All piping, depending on design review and insection
	Total	100.7301						

Table 7.10 Modification factors for the release frequencies, based on cause statistics for up to 3 inch pipes(part 1)

7.10 The effect of intensive inspection

Intensive inspection and inspection based maintenance has been shown to reduce failure rates significantly. The effectiveness is dependent on the frequency, method, and location of inspection. To be effective, the inspection programme must provide methods which can identify pipe wall thinning and cracks for all piping likely to have a potential for leaks, and the coverage needs to be extensive e.g. every elbow and inside and outside of bend, and every metre of pipe length. With risk based inspection, the inspection intensity is made on the basis of probability of failure and the severity of any failure. The effectiveness is given by the "probability of detection" (POD).

API 581, Risk Based Inspection Source Document, gives one of the most systematic approaches to assessment of inspection. The document defines five effectiveness categories, from highly effective to poorly effective, and three damage state categories for the results, from no worse than expected to considerably worse than expected. These termes are defined with objective and quantifiable criteria. Based on these, the probability of detection of damage can be determined. Depending on these factors the likelihood of detection of the true damage state varies from 0.01 to 0.9. With repeated inspections, the probability of detection of damage increases. The frequencies of failure for piping due to corrosion can reasonably be multiplied by the probability of failure of detection for a high quality inspection programme using a validated method of calculation probability of detection.

7.11 Detailed failure analysis

The causal analysis in section 7.8 and 7.9 is used here to construct a detailed frequency analysis for piping failures. The starting point is the failure rates in table 7.4. These are considered to apply to all "normal" piping, that is carbon steel piping with standard thickness and standard pressure rating. To these " normal" failure frequencies must be added contributions if the piping is especially vulnerable e.g. is used to transport sour waste water. The frequency may be reduced if special measures are put in place. Table 7.10 shows a calculation of failure rates for a pipe carrying solvent, under pressure, with a pressure specification break, and subject to hazards of special corrosion and fire. An safety valve is fitted.

The frequency of failure for the pipe, in the absence of the ESD function, is about 10 times the failure rate for rupture from a pipe which is not particularly susceptible to special failure causes.

Table 7.11 Release frequencies per year Pipe, 3-11 inch, per m	Release size	Metres	Frequency	Susceptibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency	Justification of suceptibility evaluation
Failure cause															
Internal corrosion	small	1	6.78E-06	1		0			0			0		6.78E-06	
Internal corrosion	medium	1	5.36E-06	1		0		ESD	0	0.01		0		5.36E-06	
Internal corrosion	large	1	2.77E-06	1		0		ESD	0	0.01		0		2.77E-06	
External corrosion	small	1	2.02E-06	1		0		ESD	0	0.01		0		2.02E-06	
External corrosion	medium	1	1.63E-06	1		0		ESD	0	0.01		0		1.63E-06	
External corrosion	large	1	8.04E-07	1		0		ESD	0	0.01		0		8.04E-07	
Drain lines left open	rupture	1	1.09E-05	1		0		ESD	0	0.01		0		1.09E-05	
Maintenance error	small	1	6.46E-06	1		0		ESD	0	0.01		0		6.46E-06	
Maintenance error	medium	1	6.75E-06	1		0		ESD	0	0.01		0		6.75E-06	
Corrosion, no inspection, dead legs	small	1	5.50E-04	1		0		ESD	0	0.01		0		5.50E-04	
Corrosive liquid, or sour gas	small	1	2.87E-04	0		0		ESD	0	0.01		0		0.00E+00	
Under lagging corrosion	large	1	1.54E-05	0		0		ESD	0	0.01		0		0.00E+00	ULC and external corrosion alternative
Erosion	medium	1	9.58E-05	1		0		ESD	0	0.01		0		9.58E-05	
Wrong material	large	1	3.08E-06	1		0		ESD	0	0.01		0		3.08E-06	
Lining failure	medium	1	1.92E-05	1		0		ESD	0	0.01		0		1.92E-05	
Support failure	large	1	9.1E-06	1				ESD	0	0.01		0		9.11E-06	
Overheating ++	rupture	1	5.5E-05	1	SV	0	0.05108	ESD	0	0.01		0		5.47E-05	
Overpressure, control failure ++	rupture	1	1.8E-04	1	SV	0	0.05108	ESD	0	0.01		0		1.77E-04	
Overpressure, gas breakthrough ++	rupture	1	0.000617	1	SV	0	0.05108	ESD	0	0.01		0		6.17E-04	
Overpressure, shut in liquid ++	medium	1	0.001476	1	SV	1	0.05108	ESD	0	0.01		0		7.54E-05	
External fire	rupture	1	5.58E-07	1		0		ESD	0	0.01		0		5.58E-07	
Weld crack	large	1	2.17E-06	1		0		ESD	0	0.01		0		2.17E-06	
Hammer ++	rupture	1	9.37E-05	1		0		ESD	0	0.01		0		9.37E-05	
Weather, freezing ++	medium	1	1.13E-05	1		0		ESD	0	0.01		0		1.13E-05	
Crash, impact ++	large	1	5.54E-06	1		0		ESD	0	0.01		0		5.54E-06	
Vibration fatigue ++	large	1	1.62E-05	1		0		ESD	0	0.01		0		1.62E-05	
Thermal expansion ++	large	1	1.85E-05	1		0		ESD	0	0.01		0		1.85E-05	
Wind load ++	large	1	2.18E-04	1		0		ESD	0	0.01		0		2.18E-04	
Wrong substance ++	medium	1	9.58E-07	1		0		ESD	0	0.01		0		9.58E-07	
Earthquake, landslip, flood ++	rupture	1	7.81E-08	1		0		ESD	0	0.01		0		7.81E-08	
Internal explosion ++	rupture	1	5.51E-05	1		0		ESD	0	0.01		0		5.51E-05	
Vandalism, third party ++	rupture	1	6.43E-05	1		0		ESD	0	0.01		0		6.43E-05	
Low temperature embrittlement ++	rupture	1	1.56E-06	1		0		ESD	0	0.01		0		1.56E-06	
Dropped object	rupture	1	1.85E-06	1		0		ESD	0	0.01		0		1.85E-06	
Design error ++	large	1	1.20E-05	1		0		ESD	0	0.01		0		1.20E-05	
Total small														5.65E-04	2.63E+01
Total medium														2.16E-04	1.01E+01
Total large														2.88E-04	1.34E+01
Total rupture														1.08E-03	5.02E+01
													Total	2.15E-03	% of total

Table 7.12 Release frequencies per year, Pipe < 3 inch Failure cause	Release size	Metres	Frequency	Susceptibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency	Justification of suceptibility evaluation
Internal corrosion	small	1	1.37E-05	1		0			0			0		1.37E-05	
Internal corrosion	medium	1	3.78E-05	1		0		ESD	0	0.01		0		3.78E-05	
Internal corrosion	large	1	4.08E-06	1		0		ESD	0	0.01		0		4.08E-06	
External corrosion	small	1	4.07E-06	1		0		ESD	0	0.01		0		4.07E-06	
External corrosion	medium	1	1.15E-05	1		0		ESD	0	0.01		0		1.15E-05	
External corrosion	large	1	1.18E-06	1		0		ESD	0	0.01		0		1.18E-06	
Drain lines left open	large	1	1.60E-05	0.2		0		ESD	0	0.01		0		3.20E-06	
Maintenance error	small	1	2.61E-05	1		0		ESD	0	0.01		0		2.61E-05	
Corrosion, no inspection	small	1	1.11E-04	0.05		0		ESD	0	0.01		0		5.56E-06	
Corrosive liquid, or sour gas	small	1	5.56E-04	0		0		ESD	0	0.01		0		0.00E+00	
Under lagging corrosion	large	1	2.27E-04	0		0		ESD	0	0.01		0		0.00E+00	ULC and external exclusive
Erosion	medium	1	6.75E-04	0		0		ESD	0	0.01		0		0.00E+00	
Wrong material	large	1	4.54E-06	0.03		0		ESD	0	0.01		0		1.36E-07	
Lining failure	medium	1	1.35E-04	0		0		ESD	0	0.01		0		0.00E+00	
Support failure	large	1	1.6E-05	0.3				ESD	0	0.01		0		4.76E-06	
Overheating ++	large	1	1.9E-03	0	SV	1	0.0510798	ESD	0	0.01		0		0.00E+00	
Overpressure, control failure ++	large	1	3.1E-04	0	SV	1	0.0510798	ESD	0	0.01		0		0.00E+00	
Overpressure, gas breakthrough ++	large	1	0.00319628	0	SV	1	0.0510798	ESD	0	0.01		0		0.00E+00	
Overpressure, shut in liquid ++	medium	1	0.03093709	0	SV	1	0.0510798	ESD	0	0.01		0		0.00E+00	
External fire	large	1	9.7182E-07	1		0		ESD	0	0.01		0		9.72E-07	OK
Weld crack	large	1	3.20E-06	1		0		ESD	0	0.01		0		3.20E-06	OK
Hammer ++	large	1	0.00016327	0		0		ESD	0	0.01		0		0.00E+00	
Weather, freezing ++	medium	1	4.00E-04	0		0		ESD	0	0.01		0		0.00E+00	
Crash, impact ++	large	1	8.1633E-06	0		0		ESD	0	0.01		0		0.00E+00	
Vibration fatigue ++	large	1	2.38E-05	0		0		ESD	0	0.01		0		0.00E+00	
Thermal expansion ++	large	1	2.72E-05	0		0		ESD	0	0.01		0		0.00E+00	
Wind load ++	large	1	3.21E-04	0		0		ESD	0	0.01		0		0.00E+00	
Wrong substance ++	medium	1	6.75E-06	0		0		ESD	0	0.01		0		0.00E+00	
Earthquake, landslip, flood ++	large	1	1.36E-07	0		0		ESD	0	0.01		0		0.00E+00	
Internal explosion ++	large	1	9.61E-05	0		0		ESD	0	0.01		0		0.00E+00	
Vandalism, third party ++	large	1	1.12E-04	0		0		ESD	0	0.01		0		0.00E+00	
Low temperature embrittlement ++	large	1	2.72E-06	0		0		ESD	0	0.01		0		0.00E+00	
Dropped object	large	1	2.72E-06	0		0		ESD	0	0.01		0		0.00E+00	
Design error ++	large	1	1.77E-05	1		0		ESD	0	0.01		0		1.77E-05	
Total small														4.94E-05	1.57E-04
Total medium														4.93E-05	1.11E-04
Total large/rupture														3.52E-05	5.88E-05

7.12 Algorithm for piping release frequencies

Base frequency modifications according to application

As can be seen from the data derived from US RMP records, the failure rate for piping varies very much according to the type of application. Some of the reasons for this have to do with special corrosions, temperature, and temperature cycling conditions. It is important for the accuracy of the risk assessments that these factors are not taken into account twice. For example, chlorine piping is often subject to corrosion due to the presence of a relatively high background concentration of hydrochloric acid in the air. If a modification factor for the presence of chlorine is taken into account, it will generally not be necessary to provide an additional correction for corrosion. Some judgement will be needed in selecting the most appropriate factors. The analyst should consider each set of modifications, and should document reasons for rejecting a particular set.

The modification factors apply notionally to a complete pipe section. Generally, for efficiency, the analyst will calculate a modification factor for a complete unit or even a complete plant, except for specially sensitive pipes, such as those with a specification break

#	Question	Action if Yes	Action if No
1	Is the application for a refinery hydrocarbon unit	Fire susceptibility 1 Corrosive liquid susceptibility 2	go to 2
2	Is the application for a reformer	Fire susceptibility 1	go to 3
3	Is the application for a refinery alkylation unit	Fire susceptibility 1 Corrosive liquid susceptibility 1	go to 4
4	Is the application for light ends unit	Fire susceptibility 1	go to 5
5	Is the application for an ammonia vessel	Corrosive liquid susceptibility 0.1 Consider hammer susceptibility	go to 6
6	Is the application for a chlorine pipe	Corrosive liquid susceptibility 1 Consider hammer susceptibility	go to 7
7	Is the application for a bromine pipe	Corrosion susceptibility 1 Corrosive liquid susceptibility 3	go to 8
8	Is the application for an LPG storage unit	Fire susceptibility 1 Corrosion susceptibility 0.1	go to 9
9	Is the application for an acid handling unit	Corrosive liquid susceptibility 2	go to 10
10	Does the unit handle sour gas	Corrosive liquid susceptibility 1.5	go to 11
11	Is the application for a fine chemicals unit	No modification	go to 12
12	Is there an aggressive corrosion inspection programme	See section 7.11.9 Exit	Exit

Table 7.13 Baseline frequency modifications for piping application

Piping failure due to overpressuring

Pipe failure due to overpressuring is in fact quite rare. Only 2 examples could be found in the MHIDAS data base, out of a total of 85 failures. One reason for this is certainly, that much piping is protected by relief valves. The potential for pipe rupture, could be as much as a factor 20 higher, in the case of vulnerable piping i.e. with no pressure relief provided (see Ch. 5).

Much piping is not in fact vulnerable to overpressuring at all, the pipe strength being sufficient to withstand the highest pressure in the system. However over pressuring can occur if there is a pressure specification break in the design, with a pressure regulator reducing the pressure. Some piping is not vulnerable because even when there is a pressure reduction, the downstream piping is chosen for the full pressure specification.

Pressure piping is only rarely under designed (i.e. designed too weak) because pressure-piping design is a specialist activity. In many countries, designers of pressure vessels and piping must have a special authorisation, and in many cases, authorities check the design calculations. Piping may be too weak however, if the application is changed, e.g. using the piping for fluids with higher pressure.

Operators can cause over pressuring by shutting in pipe sections. This is particularly a problem for piping with liquefied gases, but cases have also occurred with oil transfer piping. The problem is prevented with high reliability by fitting thermal relief valves This is common on liquefied gas piping, and is almost universal on cryogenic piping.

A fault tree analysis was carried out for overpressuring, with some of the data derived from the root cause analysis, and some from the failure analysis of typical pressure reduction systems.

The actual calculation of adjustment factors is given in the calculation database, which is described in Volume 1

Note: Many of the additional failure rate contributions which arise, beyond the basis valves, are ones, which apply per pipe section rather than per pipe meter.

These considerations lead to the following questions and frequency modifications:

#	Question	Action if Yes	Action if No
1.	Is there a pressure specification break and pressure reduction in the pipe.	Go to 2	Go to 3
2	If so, is there a pressure relief on the piping or on a vessel connected	Overpressuring, control failure susceptibility = 1 Calculate pressure relief reliability Go to 3	Overpressuring, control failure susceptibility = 1 Go to 3
3	Is the line designed to carry liquid	Go to 4	Go to 7
4	If so, can gas flow enter the pipe	Go to 5	Go to 7
5	Is there a pressure specification break and pressure reduction in the pipe	Go to 6	Go to 7
6	Is there a restriction orifice in the pipe, reducing flow to a safe level	Go to 7	Gas break through susceptibility = 1 Go to 7
7	Does the pipe transfer liquefied gas?	Go to 8	Go to 10
8	Is the liquid cryogenic?	Go to 9	Go to 10
9	Is there a thermal relief valve on each pipe section which can be isolated or shut in	Go to 10	Thermal expansion susceptibility = 1 Exit
10	Does the pipe have a large volume?	Go to 11	go to 13
11	Is it exposed to the sun?	Go to 12	Go to 13
12	Is there a thermal relief on the pipe ?	Go to 13	Thermal expansion susceptibility = 1 Exit
13	Is there heat tracing on the pipe?	Go to 14	Exit
14	Can the liquid in the pipe be shut in	Go to 15	Exit
15	Is there a thermal relief or other relief on the pipe.	Exit	Thermal expansion susceptibility = 1 Exit

Table 7.14 Baseline frequency modifications for overpressuring of pipes

Piping damage due to external impact

Piping can be damaged or broken by vehicle impact (such as back hoes, front loaders and fork lift trucks). Piping can equally be damaged by cranes, or by falling objects such as ice, safety valves or heat exchangers etc.

Impact of vehicles on piping can be prevented by:

- not providing working vehicle access to the plant units.
- providing crash barriers to protect plant.
- providing “headache bars” to protect pipe bridges from damage by high trucks or by front loaders with raised shovels.
- Cranes are often a necessary tool for maintenance or modification of plants. Crane crashes on pipe bridges, and over turning cranes, are typical causes of incidents.

- Releases due to crane crashes can be prevented by:
  - Providing “headache bars” to prevent impact on pipes.
  - Prohibiting access of cranes to plant, which is in operation.
  - Removing the plant inventory completely during crane access.

Release frequency can be reduced by careful plant planning:

- providing good access ways for cranes.
- providing travelling cranes and monorails.
- providing a lifting route in the plant

The quality of crane driving effects release frequencies. In some plants the quality of crane driving is so good that crashes are almost never heard of, where as in others, crashes may occur up to once per year or more. Following the principles stated in chapter 4, such conditions may change with time and are therefore difficult to guaranteed for a risk assessment. However objective measures such as reinforcement of pipe bridges, headache bars etc. can be relied on irrespective of management changes.

Piping damage can occur due to falling objects. Typical problems are:

- Safety valves being taken down for testing.
- Heat exchanger tubing bundles, being pulled for repair or replacement.
- New or replacement piping being installed.
- Upgraded equipment being installed, especially condensers/??, receivers and heat exchangers.
- Cranes, which are not properly braced with outriggers.
- Ice builds up.
- Rocks and other heavy objects falling from conveyors.
- Persons falling from ladders or walkways.
- Tools such as welding sets.

Falling objects can break piping of different sizes, depending on weight and distance. The following table gives some examples of damaged piping.

Damage due to falling objects can be minimised by:

- Placing pipe bridges away from the location of high equipment.
- Providing derricks and lifting/lay down areas on the plot plan.

- Providing adequate work areas for high level activities such as heat exchanger retubing.

Releases from process plant due to falling object damage can be minimised by:

- Avoiding "simultaneous operations" i.e. periods when there is both production and construction or heavy maintenance work.

- Removing in process inventories while simultaneous operations are taking place.

- (These steps are usually required when there is a threat release of highly hazardous substances such as hydrogen fluoride, ammonia, chlorine, methyl isocyanine, phosgene etc.)

The algorithm for determining modifications to baseline data is given below.

#	Question	Action if Yes	Action if No
1	Is the unit a multi storey unit?	Go to 3	Go to2
2	Does the plant unit have high units such as storage tanks, feed tanks, condensers above 10 m high?	Go to 3	Exit
3	Is there as need to remove heavy safety valves.	Dropped object susceptibility = 1 Exit	Go to 4
4	Is it necessary to raise/or lower heavy equipment over pipe racks or pumps?	Dropped object susceptibility = 1 Exit	Go to 5
5	Are simultaneous heavy operations forbidden?	Go to 7	Go to 6
6	Is the inventory removed from the plant during simultaneous operations (either by deliberate drain down, or as a part of the normal shutdown procedure?	Go to 7	Dropped object susceptibility = 1 Exit
7	Can blocks of raw materials, product, or drums fall on piping? (Special study).	Dropped object susceptibility = 1 Exit	Exit

Table 7.15 Base frequency modifications for falling objects, per pipe section

Vehicle impact

Vehicle impact can damage piping, especially at pipe bridges and pipe trenches. Protection can be made by means of bollards, heavy crash barriers, or concrete walls. The algorithm for modification of release frequencies to take into account traffic impact is given below. Note that the actual frequency of damage will depend on the traffic density and actual traffic speeds. These are dependent in turn on the traffic routes in the plant and the degree of separation, and on safety management practices in enforcing speed limits. In accordance with the principles given in Volume 1, safety management influences are not taken into account here, only objective physical protection.

#	Question	Action if Yes	Action if No
1	Are there traffic routes through the plant? Do the routes truck traffic ?	Go to 2	Go to 8
2	Are there well defined protected or safe transport routes through the plant, which protect the relevant pipe sections?	Go to 5	Go to 3
3	Are these heavy crash barriers, or heavy concrete barriers between trucks and critical pipes?	Go to 5	Traffic impact susceptibility = 1 Exit
4	Is it necessary to raise/or lower heavy equipment over pipe racks or pumps?	Traffic impact susceptibility = 1 (cranes) Exit	Go to 5
5	Are there pipe bridges?	Go to 6	Go to 8
6	Have "headache bars" been fitted?	Go to 9	Traffic impact susceptibility = 1 Exit
7	Are the pipe bridges reinforced to protect against vehicle crash?	Go to 8	Add Go to 8
8	Are front loaders used in the plant to transfer materials?´	Go to 9	Exit
9	Are these heavy crash barriers, or heavy concrete barriers between front loaders and critical pipes?	Exit	Traffic impact susceptibility = 1 Exit

Table 7.16 Base frequency modifications for traffic impact, per pipe section

Hammer effects

Water or fluid hammer can cause pipe rupture. Hammer can occur:

- When a valve in a flowing pipe is closed suddenly.
- When liquid is suddenly pumped into an empty pipe.
- When liquefied gas is pumped into a pipe, and the "cushion" of gas ahead of the liquid is compressed. The final collapse of the bubble of gas leads to hammer.
- When a slug of liquid from a vessel drain, a low point in piping, or other trap, is suddenly accelerated by a flow of gas.
- When liquid in a blow down tank or similar vessel is sucked backward due to steam condensation, rapid temperature drop, or solution of a gas such as ammonia.

Hammer effects are more common on larger transfer lines and on pipelines, but are well known also on hazardous in-plant lines such as those for ammonia and for chlorine during plant start up.

Hammer effects can be prevented by:

- Avoiding emptying of pipelines wherever possible.
- Fitting slow opening actuators to valves.
- Starting pumps against closed valves, or on a bypass line, and then slowly filling pipes before regulating up to full pressure.
- Filling liquid gas piping very slowly.
- Providing self-draining slopes for piping.
- Avoiding low points in piping design.
- Always assuming that condensers heat exchangers etc. will have liquid inside and will be pressurised, when opening valves to drain down, and preparing procedures accordingly.

The algorithm for the assessment of hammer risk is given as follows:

#	Question	Action if Yes	Action if No
1	Is there a source of pressure (pump, upstream pressure, pressurised gas), which could accelerate liquid, and is there a sufficient length of pipe, to allow rupture by hammer? (See rules of thumb).	Go to 2	Go to 5
2	Is the potential cause of hammering pumping?	Go to 3	Go to 5
3	Is the pump designed technically for slow start up, with closed discharge, or with bypass?	Go to 4	Hammer susceptibility = 1 Exit
4	Are the procedures designed for slow pump start up or slow discharge opening?	Go to 5	Hammer susceptibility = 1 Exit
5	Is the potential cause of hammering valve closure?	Go to 6	Go to 7
6	Are closure valves designed for slow closing?	Go to 7	Hammer susceptibility = 1 Exit
7	Is the potential cause of hammering the filling of liquefied gas piping?	Go to 8	Go to 9
8	Are procedures written for slow filling?	Exit	Hammer susceptibility = 1 Exit
9	Is the potential cause of hammering liquid collection in piping?	Go to 10	Exit
10	Is the piping designed to be self-draining?	Exit	Hammer susceptibility = 1 Exit

Table 7.17 Base frequency modifications for hammer effects, per pipe section

Internal explosion

Some piping can explode due to reaction taking place inside the piping. Examples are:

- Nitrogen trichloride collecting in piping and evaporators.
- Hydrogen contamination in chlorine gas
- Hot ethylene polymerising.

- Compressed acetylene.

- Vent lines and ducts carrying mixtures of air and solvent vapour

Piping containing oil or solvents can explode when air is admitted during emptying, for maintenance purposes. This can be prevented by purging the line with nitrogen or other non-combusting gas (e.g. flue gas) during pipe emptying. When hot work is to be carried out on a pipe, the pipe may be spaded or plugged, so that only the affected section need be purged and cleaned. The hazards of such limited (compromise) explosion prevention needs to be taken into account.

Compressed air piping can explode or even detonate, if the wall becomes covered in a film of oil, and then the oil is ignited due to overheating. This problem relates only rarely to external major hazards however – the explosion can damage air piping and instrumentation can cause serious accidents to persons, and can in rare cases lead to domino effects. Offsite consequences are unlikely to arise from airline failure however.

An algorithmic approach to determination of pipelines explosion risk is very difficult on actual operating conditions. Generally, a hazop study will be required as a basis for proper analysis.

Pipe explosion during maintenance can be assessed algorithmically as follows:

#	Question	Action if Yes	Action if No
1	Does the piping normally contain both flammable material and air? (Usually a vent)	Susceptibility to explosion = 1 Exit	Go to 2
2	Is the material carried flammable	Go to 3	Go to 7
3	Is there a procedure for purging the pipe with nitrogen etc. during maintenance?	Go to 4	Go to 6
4	Is there a procedure either for cleaning/washing the pipe, or for blocking off a short section, during maintenance?	Go to 5	Go to 6
5	Does the liquid or gas contain hydrogen sulphide, so that there may, while shut down, be pyrophoric sulphide?	Susceptibility to explosion = 1 Exit	Go to 6
6	Would an explosion in the pipe lead to a significant release of hazardous material.	Exit	Susceptibility to explosion = 1 Exit
7	Is there a potential for decomposition or polymerisation in the liquid in the pipe.	Exit	Susceptibility to explosion = 1

Table 7.18 Base frequency modifications for explosion effects, per pipe section

Pipe supports and pipe stresses

Pipe supports are intended to support the pipe in such a way that pipe stresses are controlled within design limits, so that pipe movement is kept under control, and so that the weight of pipe and its contents are supported without excessive stress. Sometimes, the original design of pipe supports is such that the pipe is over stressed from the beginning. This is quite rare however, for main process piping. It is quite common for 1 inch and 1½ inch drain and vent lines, which are quite often not designed at all, but are installed by pipe fitters as "field installations".

Even if piping supports are designed properly they may be installed erroneously, or may fail. Possible causes are:

- Pipe support not properly levelled so that the pipe is supported at some points but not others.
- Pipe shims not installed.
- Pipe shims which have fallen out.
- Pipe spring supports, which are not properly adjusted.
- U bolts which are removed for maintenance and not replaced.
- Supports, which are simply rusted through.
- Collapse of foundations under support.
- Damage to support by vibration.
- Pipe guides, intended to keep pipes in line when they expand or contract, but which "lock up" preventing movement, due to too tight installation, skew installation, or pipe crowding.
- Pipe supports which are "over tight" or which deform pipe, placing unnecessary stress on the pipe (this is especially a problem in the case of plastic and fibre reinforced plastic pipe).

In some plants, there are problems with as much as 50 % of the supports. It is surprising that pipe breakage from this source is not more common. Breakages do occur though. A more common problem is over stressing of flanges or valves, which leads to leakages. Pipe support problems can be prevented by yearly piping audits, and maintenance follow up.

Another group of problems arises from pipe vibration. The source of excitation of the vibration is most frequently rotating or reciprocating machinery, or pulsating flow. Other common causes are long down common lines containing rapid liquid flows together with gas. Vibration stresses and fatigue are a common cause of pipe breakage (fatigue seldom leads to leaks, the most common effect is total breakage).

Vibration in piping can in principle be predicted at design time, but such prediction is affected by uncertainties about the damping coefficients of supports. Most often, no attempts are made to predict vibration. During commissioning, engineers or piping foremen inspect other piping, and provide additional supports to change pipe resonance frequencies or to reduce vibration deflection.

Oversights in curing vibration, or later failure of supports, can lead to excessive vibration and rupture.

Vertical two-phase flow can occur in pipes which transport liquefied gases, liquids with dissolved gas, liquids under pressure and above their boiling point, and mixtures of gas and liquid. Vertical flow often occurs if there is a flow or pressure control valve at the bottom section of a riser or reflux pipe. It can also occur simply due to the hydrostatic pressure drop as liquid flows upward. The formation of bubbles, or plugs of gas, and their release at the top of the pipe, causes vibration, though of fairly low frequency (from a few hertz to a fraction of a hertz).

The vibration can be so extreme that it tears support welds, bends and opens flanges and breaks pipe directly, as well as causing fatigue damage.

Vertical two-phase flow problems can really only be prevented by careful work at the design stage, predicting problems and changing design to prevent them. Once serious two-phase vertical flow vibration begins, safety depends on the awareness of the operators. It is rare that you can rely on auditing to find the problem early enough to act as a preventive measure, so design review and operator awareness are the most effective risk reduction measures.

#	Question	Action if Yes	Action if No
1	Are supports sufficiently closely spaced? (ref. PSE)	Go to 2	Susceptibility to support failure = 1
2	Are pipe supports mostly in good condition? (audit)	Go to 3	Susceptibility to support failure = 1
3	Is there an effective piping integrity audit procedure? (PSM audit)	Go to 4	Susceptibility to support failure = 1
4	Is there a source of vibration (reciprocating or diaphragm pump, large centrifugal pump or compressor)?	Susceptibility to support failure = 1	go to 5
5	Is there any evidence of vibration?	Susceptibility to support failure = 1	Go to 6
6	Are there downcomers which allow vibration?	Susceptibility to support failure = 1	Exit

Table 7.19 Base frequency modifications for vibration effects, per pipe section

Piping at the design stage

Design problems can lead to piping failure. It is especially a significant cause for the larger and more dramatic failures.

#	Question	Action if Yes	Action if No
1	Is there an appropriate piping stress analysis procedure?	Go to 2	Susceptibility to design error = 1
2	Are there company piping design rules?	Go to 3	Susceptibility to design error = 1
3	Is there an effective pre commissioning follow up procedure for pipe supports and vibration?	Go to 4	Susceptibility to design error = 1
4	Are there any column or tank risers, which transfer volatile or pressurised liquids, or gas and liquid mixtures?	Go to 5	Susceptibility to design error = 1
5	Is there a proper procedure or standard for materials selection	Go to 6	Susceptibility to wrong material error = 1
6	Is there an appropriate materials receiving check of the piping quality ?	Go to 7	Susceptibility to wrong material error = 1
7	Are the piping storage facilities used sufficient to ensure that the materials will not be mixed up	Exit	Susceptibility to wrong material error = 1

Table 7.20 Base frequency modifications for design error, per pipe section

Internal corrosion and erosion

Internal corrosion occurs in nearly all steel piping (including stainless steel). The rate of corrosion varies according to the liquid (or gas) in the pipe, the piping material, the velocity of the flow, and the temperature.

Pipes which are heavily corroded generally fail by rupturing, as a result of a pressure transient, which blows open the weakened pipe wall. Pitting corrosion can cause pinhole leaks in pipes.

A special cause of rapid failure of piping is fitting of the wrong material – ordinary carbon steel rather than special corrosion resistant grades for example. This can occur by mistake or as a deliberate substitution “for convenience” during initial construction, or by an erroneous replacements. In some plants, this problem is the dominant cause of serious piping failure, increasing overall failure rates by factors of up to 100, even though only a few sections of pipe are in fact involved. The problem can be reduced by providing for materials receiving inspections.

Erosion occurs in piping if there are abrasive solids, such as sand, in the flow. At very high velocities and with hard solids, piping can be holed in a matter of minutes or hours. Erosion occurs especially a pipe elbows.

Erosion can also occur in piping when there is cavitation. This occurs especially when liquids are close to boiling and there is an expansion in the pipe, such as just after a valve, or at an increase in pipe diameter.

Internal corrosion is generally mitigated by choice of material, and by providing a sufficient corrosion allowance, or increase in wall thickness. When choosing materials, flow rates and temperatures need to be taken into account. Irrespective of how this is done, the piping will have a curtain design lifetime. As the plant approaches its design life, the probability of piping failure will increase.

The probability of failure of piping as a result of corrosion can be reduced by a very large factor by a well-organised programme of condition monitoring. In one such programme, Lydell recorded nearly all pipe failures in a refinery for x years as “incipient”.

Some high-pressure pipes are subject to special failure causes such as crevice corrosion in stainless steel, stress corrosion cracking, hydrogen embrittlement etc. Evaluating the likelihood of these is generally a specialist study.

Welds are often the area of corrosion. Use of the wrong welding rods, and poor heat treatment procedures are typical causes. These problems can be minimised by good quality control.

Erosion can be limited during design by using tees instead of elbows, by limiting flow velocities, and by fitting sand traps and filters. Erosion can also be monitored at likely erosion sites.

A very aggressive inspection policy can reduce the susceptibility to internal corrosion dramatically, with reduction to negligible levels in some cases. The limitation is largely on the number of points at which inspection can be made, and hence on the effort devoted to inspection. If virtually all pipe spools are inspected each two to three years, the release frequency due to corrosion will be reduced by about a factor of 10.

An algorithm for determining internal corrosion frequency is given below.

#	Question	Action if Yes	Action if No
1	Is there a problem of accelerated corrosion due to substances transferred? acids, wet oils, hydrogen sulphide	Go to 2	Go to 6
2	Has a corrosion allowance and maximum operating life been determined?	Go to 3	Go to 6
3	Is the piping within five years of the end of its planned working life?	Go to 4	Go to 6
4	Is there a piping condition-monitoring programme?	Go to 5	Go to 6
5	Does it have a calculated effectiveness target?	Go to 6	Internal corrosion susceptibility = 1 (unless higher already)
6	Is there any special corrosion mechanism possible, such as stress corrosion cracking, hydrogen embrittlement etc.	Go to 7	Go to 8
7	Is there a programme for monitoring these? Is the pipe material a special alloy or specially treated material? Is there a programme for checking piping materials on receiving, with appropriate testing instruments?	go to 8	Internal corrosion susceptibility = 1 (unless higher already)
8	Is the piping pressure piping (above 5 bar). Is there a control program for checking welding consumables and weld heat treatment.	Go to 9	go to 10
9	What percentage of welds is inspected? Is it	Go to 10	Weld crack

	under 10 %		susceptibility = 1
10	Are there any solids in the flow (sand, crystals, product solids, raw materials solids)?	Go to 11	Go to 14
11	Is the flow above the critical velocity for erosion to take place?	Go to 12	Go to 14
12	Is the design made for minimising erosion (erosion tees, erosion plates, erosion resistant ??, special erosion allowance)?	Go to 14	Go to 13
13	Is there a special programme for condition monitoring for the most likely erosion points?	Go to 14	Erosion susceptibility = 1 (unless higher already)
14	Is the liquid close to its boiling point at the operating pressure (or at a non nominal pressure maintained for a long period?)?	Go to 15	go to 16
15	Are there any liquid expansion points such as a pipe expansion, a flow or pressure control valve?	Erosion susceptibility = 1	go to 16
16	Is the pipe operated at high temperature ?	Multiply the internal corrosion susceptibility by 1.4 Go to 17	go to 17
17	Is the piping corrosion rate sufficient to excced the corrosion rate at this age	Multiply the internal corrosion susceptibility by 4 Exit	go to 18
18	Is the piping over 20 years old	Multiply the internal corrosion susceptibility by 4 Exit	Exit

Table 7.21 Base frequency modifications for internal corrosion, erosion, and weld cracks, per pipe section

External corrosion

External corrosion is a significant cause of failure of buried pipes, and a less significant cause of failure of above ground process piping.

Below ground piping may fail because of coating damage, because it is allowed to operate for too long, or because the coating is penetrated by ground contamination. It is not usual to use cathodic protection within plants, because of the complexity of current flow paths, and the difficulties in ensuring that all piping is properly protected.

External corrosion on above ground piping is much more likely in plants which operate in a marine environment (within a few kilometres of the sea), or if hydrogen chloride, bromine or hydrogen bromide are processed. Serious external corrosion should not in principle be a problem, because it is so easy to observe and correct, but not all maintenance organisations know how to evaluate it, so painting may be neglected.

External corrosion is a much more serious cause of pipe failure under lagging. This is especially a problem if cladding to poorly made, is damaged, or if it has been removed for maintenance and improperly replaced. Then the lagging gets wet, either from rain, or from water from leaks, deluge system tests, washing down, or vessel draining. The hot wet condition under lagging ensures rapid corrosion. Inspection is difficult and much experience is necessary to judge the likelihood and location of such corrosion. Surprising, external corrosion under lagging can occur even when piping operates at over 100 ºC, when water in the lagging would be expected to evaporate the explanation may be that the heavy corrosion occurs during shut down periods when temperatures are lower. Salt concentration can be high in logging on this kind of piping, due to accumulation as water seeps in.

Likelihood can be reduced by providing periodic integrity audits, or by providing a condition management programme.

Corrosion can occur under lagging too if the piping is cold, such as in the cold end of LPG, propylene, ethylene and similar plants. Lagging becomes wet by condensation. Even though corrosion rates are lower than with hot piping, they can be fast enough to cause holes in the piping.

#	Question	Action if Yes	Action if No
1	Is the piping below ground?	Go to 2	Go to 4
2	Are there special problems with the ground (salt, contamination)?	Susceptibility to unusually heavy external corrosion = 1	Go to 3
3	Is the pipe tape wrapped ?	Susceptibility to unusually heavy external corrosion = 1	Exit
4	Is there a piping integrity audit programme?	Go to 6	Go to 5
5	Is the location a marine environment ?	Susceptibility to unusually heavy external corrosion = 1	Go to 6
6	Is painting in good condition?	Go to 9	Susceptibility to unusually heavy external corrosion = 1 Go to 9
7	Is piping stainless?	Exit	Go to 8
8	Is there a piping integrity audit?	Go to 9	Susceptibility to unusually heavy external corrosion = 2 Go to 10
9	Is the piping lagged?	Go to 10	Exit
10	Is the lagging in good condition, with properly sealed cladding?	Go to 12	Susceptibility to under lagging corrosion = 1
11	Are the ambient conditions marine or corrosive?	Susceptibility to under lagging corrosion = 3 Exit	Go to 12
12	Is the piping hot (over 60ºC), with some periods of cooling?	Susceptibility to under lagging corrosion = 3 Exit	Go to 13
13	Is the piping cold (under 0ºC)?	Susceptibility to under lagging corrosion = 3 Exit	Exit

Table 7.22 Base frequency modifications for external corrosion, per pipe section

Weld cracking and materials fault

Piping material is sometimes faulty in it self, for example with inclusions or with lamination in the steel. More commonly there are problems with welds, for example due to:

- Inclusions of slag, due to poor welding proficiency, or simple mistakes.
- Cracking in welds due to poor welding procedures.
- Brittle cracking in welds due to inadequate heat treatment.
- Use of the wrong grade of welding rods.
- Cracks due to pipe stressing resulting from weld contraction.

These problems can be reduced by a large factor, by quality assurance procedures, and by good non-destructive testing.

#	Question	Action if Yes	Action if No
1	Is there compulsory inspection of welding for the relevant (pressurised) pipes?	Go to 2	Susceptibility to weld cracking = 1
2	Is the inspection 100%)?	Exit	Susceptibility to weld cracking = 100- inspection %

Table 7.23 Base frequency modifications for weld cracks, per pipe section

Low temperature embrittlement

Carbon steel has a temperature below which it transfers from a ductile to a brittle state. At low temperatures, the brittle state. At low temperatures, the brittle steel may break by vibration or by start up or shut down hammering shocks. Most carbon steels have a brittle transition temperature of about – 20ºC. Special grades of steel such as fine grained aluminium killed steels, have brittle transition temperatures much lower, often –40ºC.

Piping for liquefied gases such as chlorine, ammonia, or LPG will often operate at close to ambient temperatures. Some companies therefore do not try to make piping suitable for low temperature operation. Problems, including shattering of pipe, can then occur when process disturbances occur. Small flange leaks can in some cases develop into complete pipe ruptures in this way.

In many companies, stainless steel piping is used e.g. for ammonia, to avoid these problems. Stainless steel or copper piping is used almost universally for cryogenic systems with temperature below – 40 ºC.

#	Question	Action if Yes	Action if No
1	Is the piping intended for liquefied gases?	Go to 4	Go to 2
2	Is the piping used in a system which contains cryogenic liquids (including liquid nitrogen)	Go to 4	Go to 3
3	Does the system contain high pressure gases, with a potential for Joule Thomson cooling to below -30°C	Go to 4	Exit
4	Does the steel used have a low brittle transition temperature?	Exit	Susceptibility to low temperature embrittlement = 1

Table 7.24 Base frequency modifications for low temperature embrittlement, per pipe section

Piping age

Piping failure rates increase dramatically in corrosive, cyclic loading or vibrating conditions as piping reaches the end of its design working life, due to loss of corrosion allowance or to accumulation of fatigue damage. In a simplified model, if the design corrosion rate is such that the corrosion allowance, and all other safety factors, are used up, the pipe fails with a probability of 1.0. In ral life, the strength of the steel is often known quite accurately, the loading is known less well (especially the combination of pressure and static loading, because of the uncertainty insupport), and the corrosion rate is rarely known accurately. The uncertainty is often expressed as a normal distribution of the corrosion rate parameter, and a similar distribution for load. The probability for failure can be expressed as the probability of the limit state being exceeded, i.e. of the load exceeding the remaining strength of the pipe. The probability of failure can be tabulated if the distributions are known. Table 7.25 gives values of probability for the calculated last five years of pipe life, assuming that load and corrosion parameters are normally distributed, and that the standard deviations for the distributions are 20%. (ref. 14, 15).

Calculated remaining life	Probability of failure per year

Table 7.25 Probability of corrosion failure as a function of calculated remaining life.

Similar tables can be constructed for fatigue life.

7.13Flanges

Many leaks of hazardous materials arise from piping and valve flanges, or from screw fittings typically on 1 inch or 1 ½ inch piping.

Flange designs vary widely, with flat, raised or recessed faces. Higher-pressure flanges have steel seal rings.

Releases from flanges are of various kinds:

- If the flange is in sufficiently tight, a sheet of liquid or gas flow may be created past the gasket.
- Small leaks may create corrosion “wormholes” in the flange face, or may erode “wormholes” in the gasket, causing pinhole like releases.
- Gaskets may be installed incorrectly, so that they are scored or broken from the start.
- Many types of gaskets are supported by friction generated by the compression of the flanges. If the flange is loose, the gasket must support the full pressure of the fluid, and a section of the gasket may burst out.
- In some cases, bolts from the gasket break, or the complete gasket may break, causing rupture.

Causes of flange failure are listed as follows:

- Gaskets can be installed skew, or not properly positioned in recesses.
- The wrong gasket type may be installed.
- Some gasket types may be crushed if the flange bolts are over tightened. Once crushed, they do not seal so well, and do not continue to seal if there are temperature changes.
- Unevenly tensioned, or under tensioned bolts lead to flange leakages.
- If bolts are under dimensioned, they can stretch when pipes are heated by the fluid. When cooling occurs, the bolts then become loose, and can cause leaks.
- Bolts can be stretched if they are over stressed, for examples due to poor pipe support, or due to skewed pipe alignment.
- Bolts can be stretched by forces, which arise from liquid hammering.
- Differential expansion in hot fluid pipes can cause stresses, which damage flanges.
- Fire can damage flanges, so that they begin to leak. This is one of the major causes of domino effects.

Choice of flange and gasket type for a particular application is something of an art. Choice of higher quality gasket types does not always result in better performance. In one company, flange leaks were almost eliminated by switching to a cheaper gasket type.

#	Question	Action if Yes	Action if No
1	Are flanges used on the critical piping?	Go to 2	Exit
2	Is there company guide for flange design and installation?	Go to 3	Susceptibility to design error = 1
3	Are the company standard designs over dimensioned?	Go to 4	Susceptibility to design error = 1
4	Is the company standard followed by maintenance groups?	Go to 5	Susceptibility to maintenance damage = 1
5	Are pipes well supported?	Go to 6	Susceptibility to overstress = 1
6	Is their significant hammering or vibration in the critical piping?	Susceptibility to hammer = 1	Go to 7
7	Is the piping in a system which can be pressurised	Susceptibility to overpressure = 1	Go to 8
8	Is there an established good practice for flange tightening in the company	Go to 9	Susceptibility to flange tightening error = 1
9	Are the flanges difficult to tighten	Susceptibility to flange tightening error = 1	Go to 10
10	Is there a good practice for preparation for maintenance, with good labelling, good permitting, and careful sprnging of flanges	Exit	Susceptibility to erroneous opening = 1

Table 7.26 Base frequency modifications for enhanced flange failure, per flange

The increased failure rate for flanges susceptible to enhanced flange failure is about a factor of 5, based on audit experience.

Table 7.27 provides a breakdown of causes of flange failure, based on 93 failures recorded in the MHIDAS data base.

Cause	#	%
Acid corrosion	2	2.2
Broken bolt	2	2.2
Corrosion	1	1.1
Crash	1	1.1
Design error	1	1.1
Erroneous opening	5	5.4
External fire	1	1.1
Flange tightening error	16	17.2
Freezing	1	1.1
Gasket damage	5	5.4
Hammer	7	7.5
3rd party	1	1.1
Left open	1	1.1
Maintenance damage	2	2.2
Manufacturing fault	2	2.2
Operator damage	1	1.1
Overpressure	6	6.5
Overstress	6	6.5
Rusting	1	1.1
Sabotage	1	1.1
Unbolted	1	1.1
unknown	5	5.4
Unmatched flanges	1	1.1
Vibration	3	3.2
Wrong gasket type	1	1.1
Total	93	

Table 7.27 Distribution of causes for flange failures.

Note the very large contribrution of third party opening of flanges, deriving from fairly rare situations wher contractors or completely unrelated work groups open up piping. In a well regulated plant this contribution is very small, especially if there is goo labelling and a good permit to work procedure.

Flange failure rares	**Small**	**Medium**	**Large**	**Rupture**	**Total**
Flanges 3 to 11 inch .	4.70E-05	2.80E-07	2.80E-07	3.60E-07	4.79E-05

No	Failure cause	% of releases	Source	Release size	Suscept-ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Corrosion, internal	20.4	MHIDAS	small	1	1	3.6E-05	All susceptible except SS
2	Corrosion, external	5.9	MHIDAS	small	1	1	1.1E-05	
3	Acid corrosion	2.2	MHIDAS	large	0.003	1	2.4E-05	Very few acid lines in database
4	Broken bolt	2.2	MHIDAS	medium	1	1	1.1E-07	All susceptible
5	Crash	1.1	Direct experience	medium	0.02	1	2.8E-06	Survey
6	Design error	1.1	Direct experience	medium	1	1	5.6E-08	All susceptible
7	Erroneous opening	5.4	Direct experience	large	1	1	1.7E-07	All susceptible
8	External fire	1.1	Direct experience	medium	0.5	1	1.1E-07	About half of DB is for aqueous production
9	Flange overtightening	17.2	MHIDAS	rupture	1	1	1.6E-07	
10	Freezing	1.1	MHIDAS	rupture	0.05	1	2.1E-07	Only for aqueous production, in north
11	Gasket damage	5.4	MHIDAS	rupture	1	0.0518	9.9E-07	All susceptible
12	Hammer	7.5	Direct experience	rupture	0.4	0.0518	3.5E-06	Survey
13	3rd party interference	1.1	Direct experience	rupture	0.004	0.0518	5.0E-05	Survey
14	Left open	1.1	Special alg	rupture	1	1	1.0E-08	All susceptible
15	Maintenance damage	2.2	MHIDAS	rupture	1	1	2.1E-08	All susceptible
16	Manufacturing fault	2.2	MHIDAS	rupture	1	1	2.1E-08	All susceptible
17	Operator damage	1.1	MHIDAS	large	1	1	3.5E-08	All susceptible
18	Overpressure	2.4	Direct experience	rupture	0.5	1	4.5E-08	Survey
19	Overstress	6.5	MHIDAS	rupture	1	1	6.2E-08	All susceptible
20	Rusting	1.1	MHIDAS	rupture	1	1	1.0E-08	All susceptible
21	Sabotage, vandalism	1.1	MHIDAS	rupture	0.02	1	5.1E-07	Survey
22	Unbolted	1.1	MHIDAS	rupture	1	1	1.0E-08	All susceptible
23	unknown	5.4	MHIDAS	rupture	1	1	5.1E-08	
24	Unmatched flanges	1.1	MHIDAS	rupture	1	1	1.0E-08	All susceptible
25	Vibration	3.2	MHIDAS	rupture	0.05	1	6.2E-07	Survey
	Total	99.69736						

Table 7.28 Modification factors for the release frequencies, based on cause statistics

Table 7.29 Detailed release frequency calculation				
Release frequencies per year	**Small**	**Medium**	**Large**	**Rupture**
Flange, 3-11 inch, per m	4.70E-06	2.80E-07	2.80E-07	3.60E-07

Failure cause	Release size	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year
Corrosion, internal	small	1	3.65E-05	1		0		ESD	0	0.01		0		3.65E-05
Corrosion, external	small	1	1.05E-05	1		0		ESD	0	0.01		0		1.05E-05
Acid corrosion	medium	1	2.37E-05	1		0		ESD	0	0.01		0		2.37E-05
Broken bolt	medium	1	1.12E-07	1		0		ESD	0	0.01		0		1.12E-07
Crash	large	1	2.80E-06	1		0		ESD	0	0.01		0		2.80E-06
Design error	large	1	5.60E-08	1		0		ESD	0	0.01		0		5.60E-08
Erroneous opening	large	1	1.74E-07	1		0		ESD	0	0.01		0		1.74E-07
External fire	medium	1	1.12E-07	1		0		ESD	0	0.01		0		1.12E-07
Flange overtightening	rupture	1	1.65E-07	1		0		ESD	0	0.01		0		1.65E-07
Freezing	rupture	1	2.06E-07	1		0		ESD	0	0.01		0		2.06E-07
Gasket damage	rupture	1	9.93E-07	1		0		ESD	0	0.01		0		9.93E-07
Hammer	rupture	1	3.47E-06	1		0		ESD	0	0.01		0		3.47E-06
3rd party interference	rupture	1	4.96E-05	1		0		ESD	0	0.01		0		4.96E-05
Left open	rupture	1	1.0E-08	1		0		ESD	0	0.01		0		1.03E-08
Maintenance damage	rupture	1	2.1E-08	1		0		ESD	0	0.01		0		2.06E-08
Manufacturing fault	rupture	1	2.1E-08	1		0		ESD	0	0.01		0		2.06E-08
Operator damage	rupture	1	3.48E-08	1		0		ESD	0	0.01		0		3.48E-08
Overpressure	rupture	1	4.4959E-08	1		0		ESD	0	0.01		0		4.50E-08
Overstress	rupture	1	6.1714E-08	1		0		ESD	0	0.01		0		6.17E-08
Rusting	rupture	1	1.03E-08	1		0		ESD	0	0.01		0		1.03E-08
Sabotage, vandalism	rupture	1	5.1429E-07	1		0		ESD	0	0.01		0		5.14E-07
Unbolted	rupture	1	1.0286E-08	1		0		ESD	0	0.01		0		1.03E-08
unknown	rupture	1	5.1429E-08	1		0		ESD	0	0.01		0		5.14E-08
Unmatched flanges	rupture	1	1.03E-08	1		0		ESD	0	0.01		0		1.03E-08
Vibration	rupture	1	6.17E-07	1		0		ESD	0	0.01		0		6.17E-07
Total small														4.70E-05
Total medium														2.40E-05
Total large														3.06E-06
Total rupture														5.59E-05

Table 7.29 Detailed failure calculations for flanges

7.14 Small bore fittings

Small-bore piping (1/2 inch to 1 ½ inch) is used in many plants for instrument piping, drains, vents, and lubricating oil, and in some plants for the main process piping. Small-bore piping is also used widely for water and process airlines. Such piping is often assembled using screwed fittings, such as sleeves and elbows.

Screwed fittings are favoured by maintenance teams because it can be installed without "hot work", in particular welding, and because it is simple to install. It is hated by engineers and operating managers with a strong interest in safety because:

- it is a source of leaks.
- the cord packing and red lead or similar sealant often used ages, and then leaks.
- when Teflon tape is used for sealing, it is vulnerable to melting in a fire. In some cases, screw fittings can unscrew under pressure when lubricated by hot Teflon tape.
- small-bore piping is especially vulnerable to failure due to over stressing due to fatigue or poor support.
- small-bore piping can be destroyed if a person climbs on it.
- small-bore piping is rarely "designed" it is more often fitted by tradesmen. The quality of the installation can vary widely, and mistakes are often made.

Small bore piping can be made safe, even for flammable materials, by providing good support, good material, and by seal welding the screwed fittings (note: welding joints sealed with Teflon tape should be forbidden – the fumes are highly poisonous).

The problem of weakness of small bore piping only seldom apply to stainless steel piping. The problems of screwed joints still apply.

Some kinds of small-bore piping are fitted together using compression fittings. This is especially the case for 4-20 mm stainless instrument piping. These can fail by "pulling out", by over pressuring, or if the joint is poorly made due to poor workmanship, see the following section for a description.

#	Question	Action if Yes	Action if No
1	Are the small bore joints made with screwed fittings?	Go to 2	Go to 3
2	Are the fittings seal welded?	Go to 3	Susceptibility for enhance small bore releases = 1 for small releases Go to 3
3	Does the piping have proper tee junctions to main piping, such as weld o let, or alternatively, it is reinforced with gusset plates?	Go to 4	Susceptibility for enhance small bore releases = 1 for break
4	Is the piping properly supported?	Go to 5	Susceptibility for enhance small bore releases = 1 for break
5	Has the piping been audited and supported to prevent vibration?	Exit	Susceptibility for enhance small bore releases = 1 for break

Table 7.30 Base frequency modifications for small bore piping failure, per pipe section

From audit experience, susceptible small bore piping has a failure rate of about 10 times that of well designed and installed small bore piping

7.15 Compression fitted stainless piping 6-12 mm

Thin stainless steel piping is used widely for instrumentation and for sampling. It is generally very reliable, but suffers from three important failure causes. The tools used for pressure fittings must be of the right type, and must not be excessively worn. Otherwise, there is a strong chance that the compression fittings will blow open under pressure. Overpressuring is a cause of opening, if the piping is underdimensioned. The piping can also fail due to the impact of heavy loads, or to operators climbing on it.

The third typical cause of failure is fatigue. Long runs of instrument piping can suffer from resonant vibration, for example when excited by vibration from compressors. If the vibration is not brought under control, by fitting additional supports, then breakage due to vibration induced fatigue can occur.

#	Question	Action if Yes	Action if No
1	Is there a standard procedure for installation, as approved by the manufacturer?	Go to 2	Susceptibility for enhanced stainless instrument piping releases = 1 for small releases
2	Are the correct tools available?	Go to 3	Susceptibility for enhanced stainless instrument piping releases = 1 for small releases
3	Are the plant maintenance staffs, and original installers trained in installation procedures?	Go to 4	Susceptibility for enhanced stainless instrument piping releases = 1 for small releases
4	Is there a follow up procedure, to limit vibration of the piping, after plant commissioning.	Go to 5	Susceptibility for enhanced stainless instrument piping releases = 1 for small releases
5	Is there heavy vibration in any of the piping (existing plant)	Susceptibility for enhanced stainless instrument piping releases = 1 for small releases	Go to 6
6	Is there a strong source of vibration, such as a compressor (plant in design)	Susceptibility for enhanced stainless instrument piping releases = 1 for small releases	Exit

Table 7.31 Base frequency modifications for instrument piping failure, per pipe section

From accident investigation experience, the frequency of failures for small bore stainless piping is at least 20 times that of ordinary piping.

7.16 References

1. Taylor, J.R. Process Safety Engineering, Designing and Building Safer Process Plant, Taylor Associates, 4th Edition 2001

2. Taylor, J.R. Process Safety management for Managers, Taylor Associates, 1998

3. API, Reported Fire Losses in the Petroleum Industry for 1998

4. See description of US RMP data in Vol. 1

5. Kletz, T, Loss Prevention Bulletin

6. Offshore Hydrocarbon Release Statistics, 2001, UK Health and safety Executive

7. CPR, Guidelines for Quantitative Risk Assessment, CPR 18E, Committee for the Prevention of Disasters, Holland, 1999 (Sdu Uitgivers 1999)

8. Pape, R.P. and Nussey, C. A Basic Approach to the Analysis of Risk from Major Toxic Hazards, In The Assessment and Control of major hazards, I Chem E 1985

9. Lydell, B, Quality Issues in Technical Risk Analysis, RSA Technologies 1996

10. Cox, A.W., Lees, F.P. and Ang, M.L. Classification of Hazardous Locations, I Chem E 1990

11. Rasmussen, N Reactor Safety Study, WASH 1400, Us NRC, 1974

12. API, Safety Digest of Lessons learned, Hanley and Karlowickz ed., undated

13. COVO, Steering Committee, Risk Analysis of Six Potentially Hazardous Industrial Objects in the Rijnmond Area,

14. API, Reccomended Pracice 581.

15. Taylor, J.R. Risk Based Inspection Technical Manual, 2005

8 Pressure vessels for storage and processing of liquefied gas

8.1 Pressure vessels for storage

Pressure vessels are widely used for storage of liquefied petroleum gas (LPG), ammonia, chlorine and others.

Figure 8.1 shows a diagram of a sphere storage, with what today would be regarded as a minimum of protective equipment. LPG, propane, butane etc. are pumped to the vessel or run down from the plant under process pressure. LPG is taken from the base of the sphere for transfer to other vessels. Typical features of the system are:

- an inclined diked area, so that any leakage of LPG runs away from the sphere.
- protected legs (protected by concrete cladding).
- inlet and outlet pipes which are connected into the lower flanged nozzle on the sphere.
- a drain line which has double valves and is piped well away from the sphere.
- transfer lines to other tanks.
- a transfer pump.
- double safety valves with 3 way connecting valve.
- redundant level sensors.
- pressure sensor.
- a water injection pipe at the base of the sphere to allow water to be injected in the case of a leak at the lower nozzle or valves.
- emergency shutdown valves.
- overall deluge for heat radiation cooling
- fixed fire water monitors for jet fire protection of the tank.

In recent years, it has become usual to provide additional fire protection for LPG, propane and butane vessels in the form of mounding, or insulation with wool or glass foam.

Pumps are sited remotely from the spheres and are usually single stage centrifugal pumps, either vertical or horizontal. Seals are double, with intermediate space pressure subject to pressure or flow monitoring. In some cases, the sphere is designed for a lower pressure than the vapour pressure of the LPG. In this case, the LPG must be refrigerated.

Figure 8.1 shows a schematic for a storage sphere. Figure 8.2 shows a corresponding figure for a bullet storage.

The photographs in figure 8.3 and 8.4 additionally show the deluge systems more clearly.

Figure 8.1 LPG storage sphere with minimum equipment

Figure 8.2 Liquefied gas bullet with minimum equipment.

Figure 8.3 LPG spheres

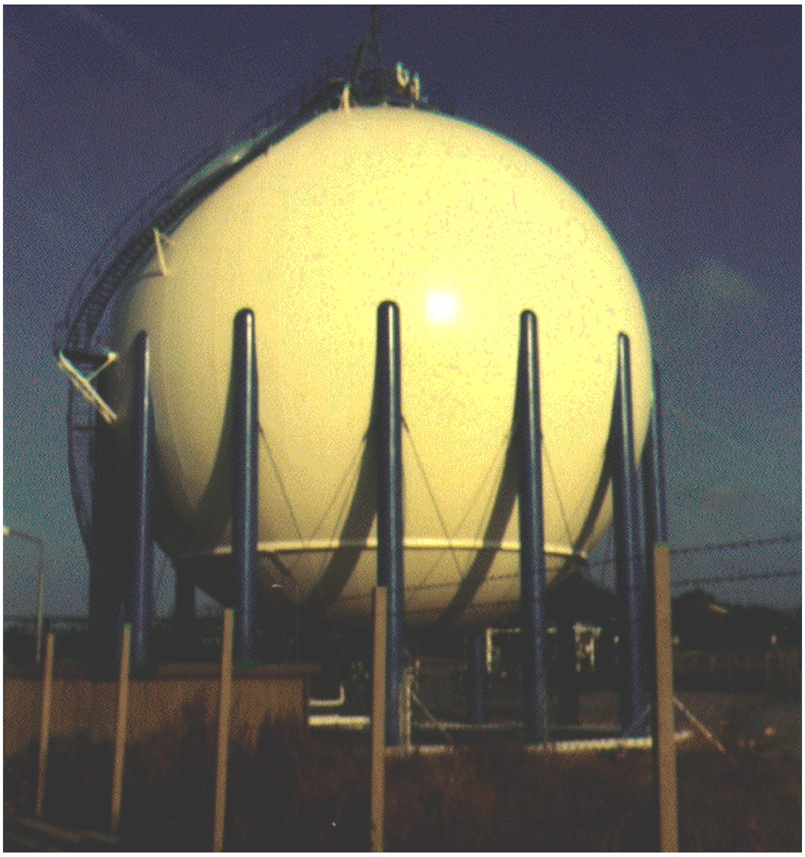


Figure 8.4 Ammonia sphere

Figure 8.5 Highly protected ammonia storage bullet

Figure 8.6 Mounded bullet storage for LPG

Figure 8.7 In process pressure vessels, ammonia plant

8.2 Failure statistics

Pressure vessels are almost always constructed to high standards, with careful choice of materials, class I welding, and full x-ray inspection of weld quality. The design is also carefully controlled in most countries, with specially certified designers, and approved design procedures. As a result, pressure vessel failure is very rare. Whole world statistics are needed in order to form a sufficient basis for determining a release rate (note: failures of mobile pressure vessels such as those on tank trucks and rail tankers are much more common, resulting from crashes)

This section describes some of the studies of pressure vessel failures, which have been reported, around the world.

Phillips and Warwick studied pressure vessel failure extensively during the 1970´s and 1980´s, amassing a large part of the available world data. These studies remain authoritative, and the failure rates determined remain the basis for most risk analyses today.

There is a widely held view among engineers that modern pressure vessels have lower failure rates than those determined by Phillips and Warwick, but there is at present too little statistical evidence to prove beyond doubt that this view is correct. A study of failure causes indicates that many of the possible causes of failure of pressure vessels are actually eliminated by modern design and construction practice.

This is of course only true in the absence of mistakes and oversights. A failure though requires errors both in fabrication and oversights or omissions in non destructive testing.

The reliability of non-destructive testing has been investigated (ref.8.2). The main concern is the probability of not finding a crack above a critical size. The critical size for a crack is determined by the criteria of whether the crack will grow under the influence of stress, material defects, heat treatment and temperature. The assessment from this inspection must be taken into account when determining the expected failure rate.

Note that cracks can continue to grow chemically even if below the critical size, such as in the case of stress corrosion cracking. Problems such as hydrogen embrittlement, blistering etc. (ref. 8.3) can cause problems also, through very rarely in storage vessels.

These considerations have in recent years led engineers to believe that the high rates of vessel failures seen in the past, with frequencies above 10^{-6} per year, have concerned vessels with a relatively low level of quality control in manufacture and non destructive testing. This is confirmed by examination of failure reports from some of the actual failures. Provided that the application is not a challenging one, the vessel is considered:

a) to have a relatively low failure probability

b) to fail in such a way that the failure is not catastrophic

Because they generally fall under regulations and laws originally developed for steam boilers, most pressure vessels are required by law to have a periodic inspection which includes shut down, internal visual inspection, and non destructive testing. These inspections should be designed to reveal significant corrosion, and have a good probability of detecting cracks. 100 % testing of welds is performed only for some vessels however, and there is never 100 % testing of plate material. The testing is therefore more a question of detecting a general tendency to defects. If a cracking is not found, the test protocol is between 4 and 9 years. If small cracks are found, which are below the threshold of concern, but close to it, it is usual to increase the frequency of testing.

Many pressure vessels are subject to pressure testing prior to commissioning, and may be tested in this way on a regular basis, for example at 6 to twelve year intervals, or when there is a change of application. Pressure testing will detect all major flaws which lead to inadequate strength, such as design errors or large cracks in welds. Pressure testing will not detect problems which develop over a two to four year time scale. Frequent pressure testing should be avoided because it stresses vessels at or above the design pressure i.e. much above the operating pressure. This can lead to a form of fatigue known as low cycle fatigue.

In the authors experience, the largest problem with this regime of inspection and testing is the possibility for vessels to be overlooked in periodic testing. Out of 92 vessels for which the author carried out detailed mechanical integrity audits, three had been left out of the inspection list i.e. about 3 % of the total.

In vessels with thin walls, even when the crack penetrates the vessel wall completely crack growth will be limited. This leads to a “ leak before break” criterion, with cracking revealing itself by means of a small release, before a catastrophic failure and vessel rupture can occur.

The leak before break criterion has an impact on design as well as on assessment of risk. If it is the case that most cracks will leak before developing into ruptures, then it is important to provide facilities for draining/depressurising the vessel under all circumstances. On the other hand, if most cracks are considered to lead to vessel destruction, then protective measures such as rapid depress ring and inventory transfer become valueless. In this case only piping need be protected (by means of emergency shutdown valves or excess flow valves). This is an instance where an apparently conservative assumption in risk analysis can have a negative effect on the safety of a design.

These factors determine a typical value of failure rates for pressure vessels due to cracking, with fairly low values for vessels which are guaranteed to be ductile in operation and which have high levels of quality control and non destructive testing.

One issue of concern is that of brittle cracking of vessels after they have become cold. One case is known, arising from liquid nitrogen flowing into a propane transportation tank, which then cracked. In principle, other cases could arise, for a variety of reasons.:

- Cold liquefied gases flowing into an empty vessel at start up.
- A gas phase leak occurring and reducing the temperature of the vessel contents.
- A gas phase leak occurring at a flange and impinging on the vessel.
- The wrong kind of steel or welding material being used on cryogenic storage

Any low temperature cracking of this kind would tend to develop into major cracking or vessel rupture. In actual practice, this kind of effect must be very unlikely, since only one case of complete vessel rupture of this kind could be found, on an LNG vessel.

Ammonia vessels present a special problem, which surfaced in the late 1960's and 1970's. Ammonia can release hydrogen into steel resulting in hydrogen embrittlement. It was found that this could be controlled by adding about 1% of water to the ammonia. Since this was discovered, there has been few problems with this, but there is in principle a potential for failure here, which would affect primarily the vessels, and possibly ships, for a particular ammonia production plant. Most procedures for quality control of ammonia today include a test for water content.

Corrosion presents one of the causes which could in principle lead to vessel failure. Such failures of vessels are unusual and rare. Because of the care taken in design and choice of materials for vessels and because of the periodic inspection mandated in most countries by law. Additionally pressure vessels are by their nature quite robust, and required to have considerable safety margins according to their design standards. Corrosion can occur however and may go unnoticed if it takes place under insulation or if it occurs at a location which is difficult to access.

Destruction of fixed pressure vessels due to overfilling is rare, with only a few cases known. Destruction of pressurised tank trucks, and of transportation vessels, by this cause has more cases. The low frequency of destruction from overfilling reflects in part of the fact that some further circumstances must occur, such as the liquid being shut in and then rising in temperature and also that there are safety devices such as safety valves and level alarms to detect and prevent overpressuring and overfilling .

For these reasons the number of opportunities for vessel overfilling/overpressuring must be at least 10 000 times the frequency of actual rupture. The safety measures prevent most opportunities from developing into actual failures.

How much equipment does a vessel include?

One of the questions which is important in determining vessel failure rates is what equipment is actually included as part of the vessel? Logically piping nozzles up to the first flange, and manholes, must be included. Should piping also be included, for example drain piping, safety valve piping etc. Should bridles for level gauges? Should small bore couplings used for instrumentation?

The answer to these questions is one of choice. It can be very convenient in a risk analysis to have release rate data for "a vessel and all its associated fittings". This avoids having to repeat flange counting, valve counting and estimation of pipe run lengths for each item of equipment. However, such an approach does introduce new uncertainties into calculations, because vessels can have very different amounts of equipment..

8.3 Hazards for pressure vessels (ref 8.1)

Leakage and domino effects.

Leakage of valve and pump seal's occurs relatively frequently. (On some installations there is almost always one or two small leaks, indicated by butane "frost" or "ice". On other plants, leakage may be very rare. The difference lies in the standard of design and maintenance). If such leaks are ignited, the resulting small fire can develop due to damage to the seals. The result is a growing jet fire. If this impinges on other equipment, a major accident can result. (The emergency action is to shut off the flow of LPG).

Pipe breakage.

Pipe or equipment breakage on a vessel can occur as a result of :

- corrosion (e.g. Shell Norco).
- vibration and brittle fracture if the piping is cold (propane temperature).
- over stressing due to misuse or improperly designed piping.
- over stressing due to improperly supported piping.
- sabotage or tampering (e.g. Gothenburg).

Pipe break leads to a two phase gas jet. If the release is large enough and ignites, it can cause an unconfined vapour cloud explosion. Such an explosion will generally destroy deluge piping, and cause major fires.

Vessel rupture.

Vessel rupture has been known to occur due to overfilling. When a vessel containing liquefied gas is overfilled, the vapour space "bubble" collapses. The result can be a shock which breaks the vessel wall (e.g. Texas City). The safety valves do not have sufficiently rapid response to absorb the shock.

Vessel rupture could in principle occur due to poor welds or material, but a few cases are known for ordinary pressure vessels. (One serious case is known on cryogenic tanks, and several cases are known in combination with other problems).

Vessel rupture can occur if, for example propane is pumped to butane tanks, if the tanks are not designed for propane pressure at the designated storage pressure. (Two cases like this known, plus one case for butane mixing with pentane). The problem arises characteristically where there is refrigerated or semi refrigerated storage. It is less common where ambient temperature storage is used, since LPG tanks tend to be dimensioned for the maximum pressure (propane).

Warning: The potential for overfilling should be checked in all cases.

Roll over

Roll over can occur if there are several grades of liquefied gas which are unmixed, and a colder layer below a lighter, warm layer. Mixing can then cause a very rapid boiling, for example of propane in and LPG tank, which is sometimes heard as a "rumble". If the roll over is very violent, the swash can cause tanks to collapse. On semi refrigerated tanks, the boil up can exceed safety valve capacity and cause over pressure rupture.

BLEVE.

Jet fires, for example on bottom piping, or between tanks, can cause flame impingement on vessels. If the impingement is above the liquid level, the metal soon overheats and the result is a major explosion. (Boiling liquid expanding vapour explosion). The resulting explosion power may be measured in kilotons TNT equivalent for a sphere storage.

8.4 Typical frequency for pressure vessel failure

There is an enormous number of potential causes of pressure vessel failure, and of these, only a few can be evaluated in terms of a generic failure rate. For example, airplane crash is a potential cause of tank failure, but it depends more on tank size, distance to airports, and air traffic rate, than on the tank itself.

A list of more common causes of pressure vessel failure is:

- Design errors, including underdimensioning, specification of inadequate materials, specification of wrong welding procedures, overloading of supports and designs in which there are weak points or stress raisers.
- Some pressure vessels require cooling, and multiwall or wound vessel require drains in the walls to protect against leakage - These present special problems.
- Overload, due to too high pressure or temperature, as a result of equipment failures, operation or administrative errors or fire.
- Material faults, welding faults, and faults due to errors in heat treatment.
- Corrosion (especially stress corrosion)
- Corrosion can attack both internally and externally. A frequent cause is improper water treatment, or contamination of water supplies. Stainless steel is especially vulnerable to chloride contamination. For vessels holding liquids other than water high water concentrations are often a problem (e. g. in liquid ammonia).
- Ageing (creep or fatigue)
- Excessive vibration (can cause fatigue or direct overload, usually at flange or weld attachment to vessels.
- Foundation collapse
- Frost heave under foundations (esp. cryogenic tanks)
- Earthquakes
- Crashes (aircraft, ground vehicles, cranes, missiles from explosions.
- Internal explosions and runaway reactions.
- Structural overload of vessels due to external stresses, especially pipe expansion or contraction.
- Liquid expansion when a vessel is completely full of liquid, and is also shut off.

Some failures are the result of several causes combined.

Of these causes, overload, crash, explosion and vibration are properties of the application and not of the pressure vessel. They must be separated out in any treatment of statistical data. In risk analysis, risk from these causes should be assessed separately.

Similar arguments might also be applied to corrosion, since this depends on the environment in which tanks operate. So far, however, this has not been the normal practice in risk analysis, because of the difficulty of obtaining data.

Failure probabilities will be very dependent on the frequency with which pressure tests and inspections are carried out on tanks. Many of the direct causes of failure are small cracks or pits, which develop over a longer period until a critical size is reached. Non destructive testing, using ultrasonic or X-ray or gamma-ray photography can reveal many such flaws, as can surface inspection in some cases, For some applications pressure vessels are inspected regularly, every two or four years (especially transport vessels).

There have been several thorough studies of pressure vessel failure rate data. Phillips and Warwick (PHI 68) studied pressure vessels built to very high standards, and in a following study (SMI 74) the number of "tank years" was brought up to 105402. These studies showed probability for failure of pressure vessels of $4.4 * 10^{-5}$ per year. Of these, catastrophic failures were such a small part that a failure rate for catastrophic damage of $3 * 10^{-6}$ could be given. In all, the study covered 1,700,000 tank years. Boesbeck (1975) undertook an evaluation of these and other data, to find a probability of catastrophic failure which is somewhere between 10^{-5} and 10^{-6}. These studies lead to the results in table 8.1

Failure Mode	**Failure Rate**
Catastrophic failure	$3 * 10^{-6}$ per year
Small leak or small break	$3 * 10^{-5}$ per year

Table 8.1 Failure rates for pressure vessels, from Boesbeck.

A.M. Thomas of Rolls Royce Ltd. has, in a series of articles, used the available data for tanks together with data from pressure vessel testing, to build up a model for pressure vessel reliability. This uses a theory of fracture mechanics , which seems to give a reasonably good correlation with the available data.

The theory is based on the conditions which are necessary for a built in flaw to reach a critical size. The constants in the theory are adjusted to fit data from 700 well studied vessel failures.

The probability for failure of pressure vessels is taken as for a fully protected high quality vessel. The “best” value for the failure frequency is taken to be 10^{-7} per year.

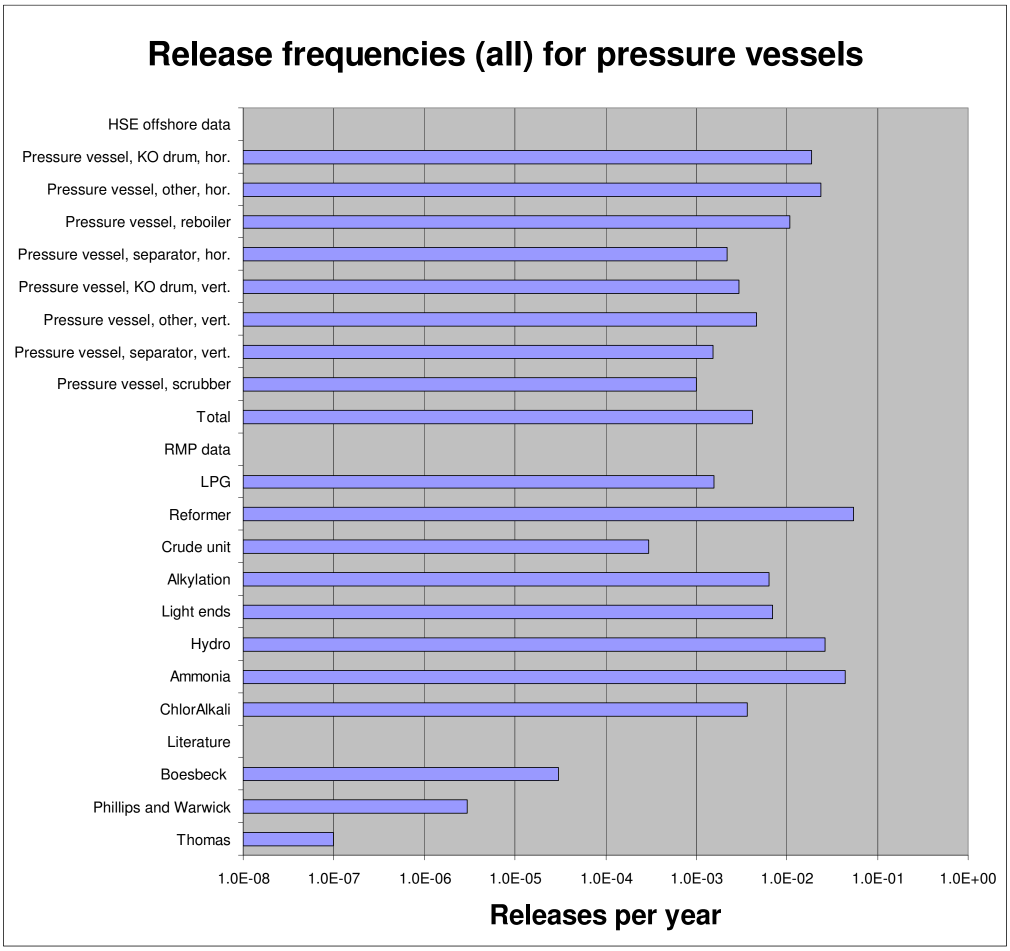
The values for smaller, non catastrophic failure rates can be obtained from the studies given in Ch. 4, particularly the UK HSE offshore data, and the data from the US RMP studies. These are summarised in table 8.2

			Hole sizes				
Equipment type	Failures	Eqip. Years	Failure freq. pr yr	< 10 mm	10-25 mm	25-100 mm	>100 mm
HSE offshore data							
Pressure vessel, KO drum, hor.	5	266	1.88E-02	3.76E-03	3.76E-03	1.13E-02	0.00E+00
Pressure vessel, other, hor.	24	1004	2.39E-02	5.74E-03	5.98E-03	9.56E-04	1.91E-03
Pressure vessel, reboiler	2	183	0.0109	5.45E-03	5.45E-03	0.00E+00	0.00E+00
Pressure vessel, separator, hor.	9	4078	0.00221	4.86E-04	0.00E+00	4.86E-04	0.00E+00
Pressure vessel, KO drum, vert.	5	1697	2.95E-03	0.00E+00	5.90E-04	1.18E-03	0.00E+00
Pressure vessel, other, vert.	8	1458	4.59E-03	1.15E-03	1.15E-03	1.15E-03	1.15E-03
Pressure vessel, separator, vert.	4	2634	1.52E-03	1.52E-03	0.00E+00	0.00E+00	0.00E+00
Pressure vessel, scrubber	3	2956	1.01E-03	1.01E-03	0.00E+00	0.00E+00	0.00E+00
Total	60	14276	4.20E-03	1.29E-03	7.47E-04	6.74E-04	2.52E-04
RMP data							
LPG storage	15		1.57E-03	8.37E-04	0.00E+00	3.14E-04	4.19E-04
Reformer	24		5.50E-02	4.40E-02	1.83E-03	9.17E-03	0.00E+00
Crude unit	13		3.00E-04	2.29E-04	1.76E-05	7.06E-05	1.76E-05
Alkylation	9		6.40E-03	4.27E-03	2.13E-03	0.00E+00	0.00E+00
Light ends	19		6.90E-03	6.90E-03	0.00E+00	0.00E+00	0.00E+00
Hydrotreater, hydrocracker etc.	18		2.67E-02	1.48E-02	0.00E+00	1.19E-02	0.00E+00
Ammonia	25		4.42E-02	8.85E-03	1.59E-02	1.06E-02	7.08E-03
ChlorAlkali	28		3.70E-03	2.38E-03	3.96E-04	7.93E-04	0.00E+00
Average for RMP			4.31E-03	3.14E-03	2.10E-03	1.25E-03	9.40E-04
Boesbeck (leak)							3.0E-5
Boesbeck (catastrophic)							3.0E-6
Phillips and Warwick		170000	3.0E-5				3.0E-6
Thomas							1.0E-7

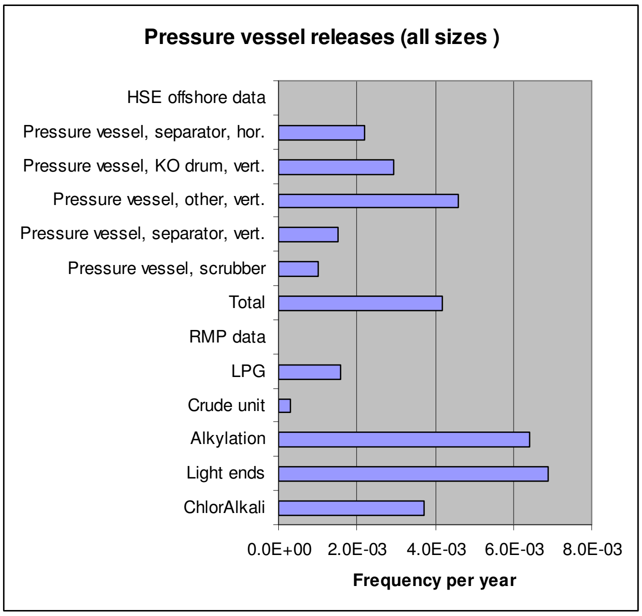
Table 8.2 Pressure vessel release frequencies for HSE offshore and US RMP data

Figure 8.8 shows a distribution for the pressure vessel release frequencies by vessel service. The ammonia and hydrotreater vessels are quite obviously outliers. Eliminating these, figure 8.9 shows the data in more detail.

There is a big discrepancy between the recent (HSE, RMP) observations and the older ones of Boesbeck, Phillips and Warwick, and Thomas. There seems little doubt that the reason for this lies in a difference in objectives and data collection method. The earlier analyses focused on failure of the vessel itself. The later analyses focus on releases from the vessel and associated equipment. There are many causes of release (e.g. an open drain valve) which do not involve vessel failure. Failure rate of greater then $1*10^{-2}$ cannot apply to a steel pressure vessel shell itself. If they did, refineries and petrochemical plants would have to replace several pressure vessels per year, and as a result, it would be necessary to shut down plants for several months each year. For the refineries, 164 plants and roughly 200 pressure vessels per plant, and using the data of Phillips and Warwick, one pressure vessel small failure would be expected every 10 years. Inspection of the US RMP data sources narrative shows that none of the releases occurred through the pressure vessel shell itself. The releases occur at nozzles, flanges, and at instumment attachments.



Figures 8.8 Distribution of release frequencies for vessels from RMP and HSE offshore data, for complete and edited data



Figures 8.9 Distribution of release frequencies for vessels from RMP and HSE offshore data, for censored data (edited to eliminate outliers, ammonia, reformer, hydrotreater and some oil and gas platform PV's).

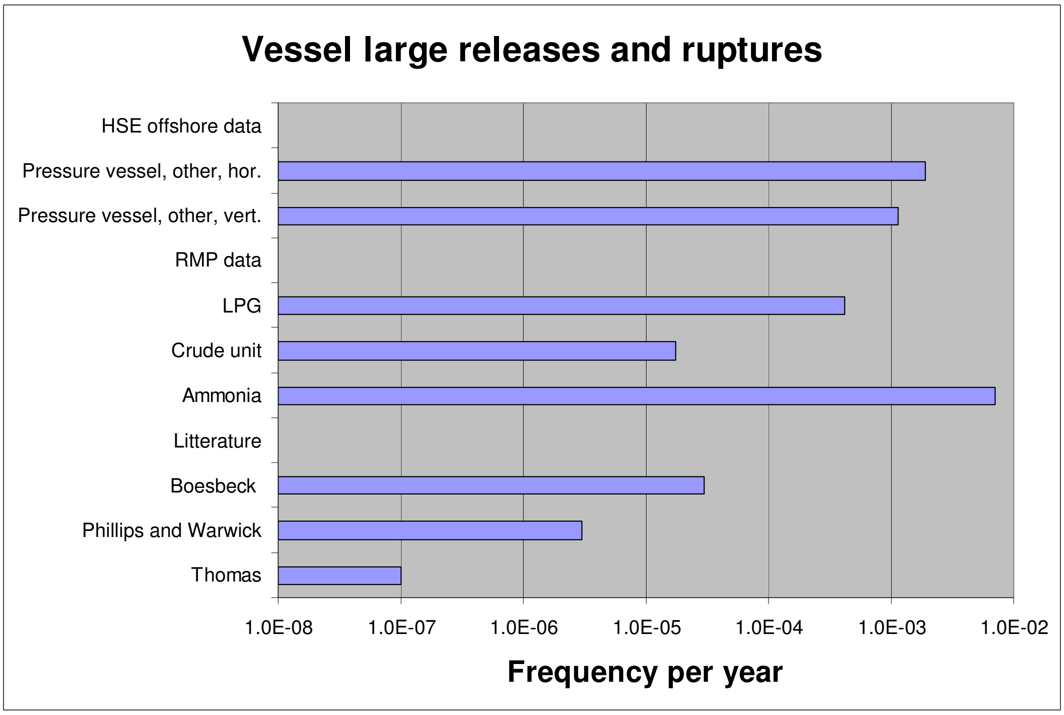


Figure 8.10 Frequency for large releases and ruptures, censored to remove outliers and zero values.

As a minimum value for pressure vessel releases, including nozzles, safety valve piping, first flange, and instrumentation attachments, a value of $3*10^{-4}$ per year, derived for the crude unit, seems appropriate.

A typical release frequency of $4*10^{-3}$ per year seems to be the appropriate, as an average of the RMP release data, with outliers being dealt with by means of modification factors.

For storage vessels, LPG vessel release frequencies are taken as typical

Release frequency per year	Small hole, < 5 mm	Medium hole 5mm to 25 mm	Large hole >25 mm	Very large hole, > 100 mm	Rupture
Averages from table 8.2	4.31E-03	3.14E-03	2.10E-03	1.25E-03	
Typical values, process vessels	4E-3	3E-3	2E-3	1E-3	1E-6
BLEVE (LPG and similar vessels)					1.8E-5
Typical values for storage vessels	2E-3	8E-4	3E-4	1E-4	1E-7

Table 8.3 Release frequency values for vessel failure rates

8.5 Cause distributions

Some information on the distribution of failure causes for vessels can be obtained from RMP data. Unfortunately, the number of classifications is small and only the root cause is given, which makes useful interpretation difficult. Figure 8.10 shows data for refineries, 8.11 for ammonia plant, and 8.12 for chlorosoda plant.

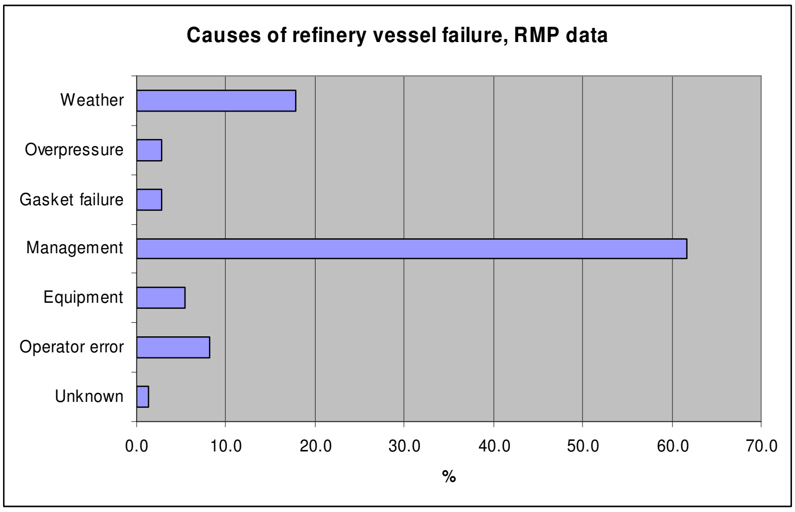


Figure 8.10 Distribution of causes for 73 refinery vessel failures, US RMP data

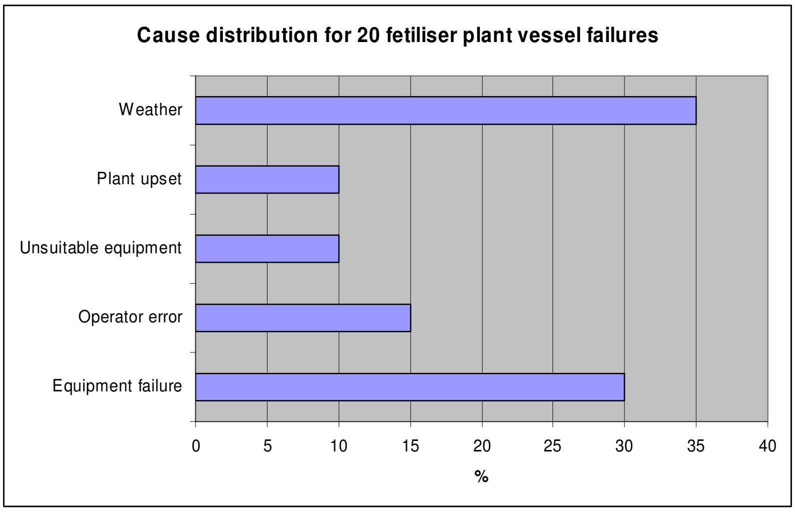


Figure 8.11 Distribution of causes for 20 fertiliser plant vessel failures, US RMP data

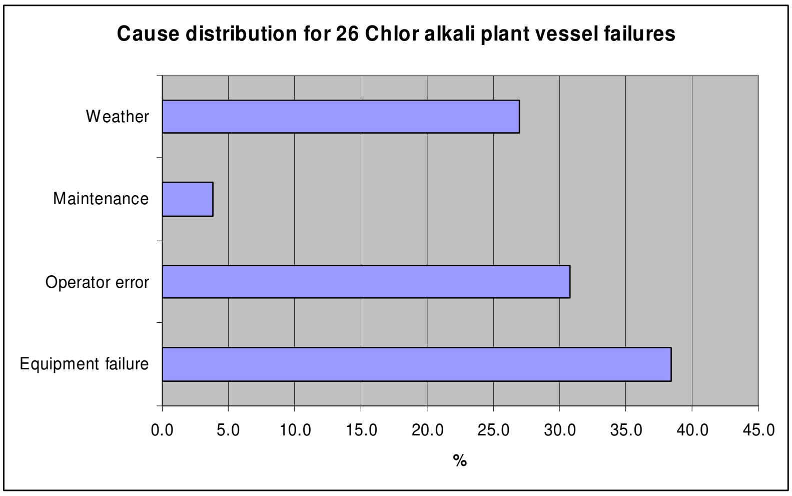


Figure 8.12 Cause distribution for 26 Chlor-alkali plant vessel failures, US RMP data

Examining the distributions, one can see that the refinery data tells much about cause attribution by the companies, with a large preponderance of management caused releases. Weather and operator error are the other main groups, with overpressuring and gasket failure represented.

Weather figures large in the fertiliser and chlor alkali plant vessel failures, as do operator error and equipment failure.

It should be noted that only the equipment failures would normally figure in many equipment failure rate data collections – operator error and weather caused failures are often censored out of the data, or recorded in a different way. This explains a factor of about 3 difference between the calculation of release rates based on detailed analysis, and rates based on incident records.

More specific data on release types were obtained from the MHIDAS data. These are shown for storage and in process vessels. The same assessment derived data for columns and reactors as well, these data are given in subsequent chapters. Both the mechanism and the root cause were assessed by the author, based on the incident report text. Results are given in figures 8.13 to 8.16.

It should be noted that the MHIDAS data is preselected by the data gathering process, to focus on the larger accident types. Some of the accidents covered in the 35 vessel release records involve larger releases than for the entire RMP data base. Domino effects resulted in 33% of the release cases.

Secondly, it is observed that weather does not figure as a cause for any of the releases in the MHIDAS data, in contrast to the RMP data.

Overfilling is a major cause for storage vessel releases. Explosions due runaway reactions are a major cause of in fine chemicals process vessel failures. The incidents recorded in this chapter are for vessels, not reactors, but runaway reactions nevertheless occur in these. A study be the US Chemical Safety and Hazard Investigation Board, *Improving Reactive Hazard Management,* showed that 22 % of reaction hazard incidents occurred in storage.

Overpressuring due to operator error and control failures figure significantly for both process and storage vessels.

Hot work close to or in open (empty) vessels is a significant cause of accidents. These accidents do not in themselves lead to releases, but they are a significant contributor to domino effects, which in turn lead to vessel releases.

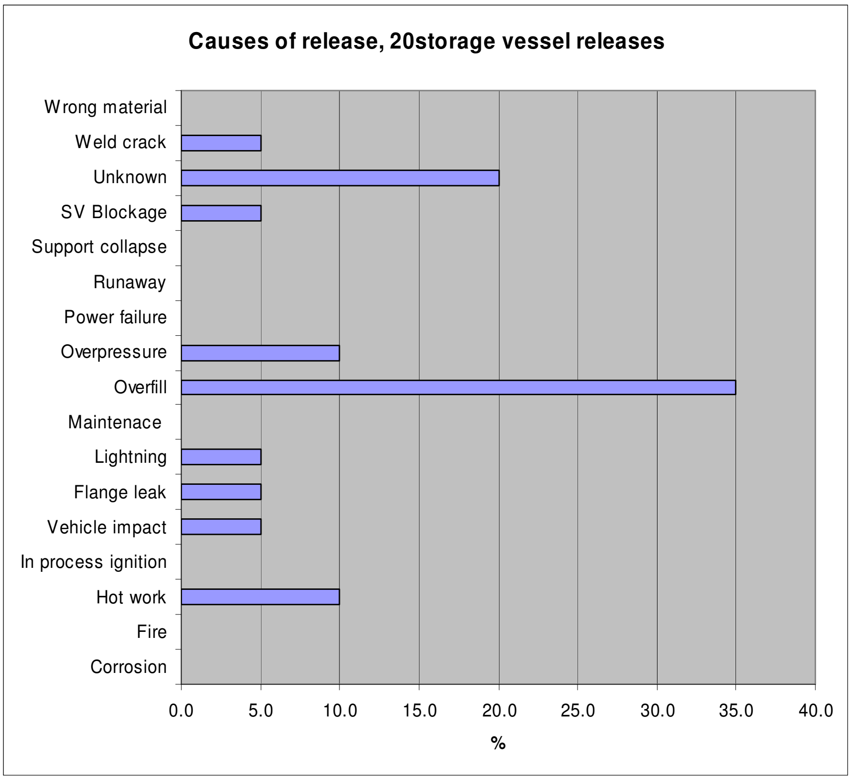


Figure 8.13 Cause distribution for storage vessel releases, MHIDAS data

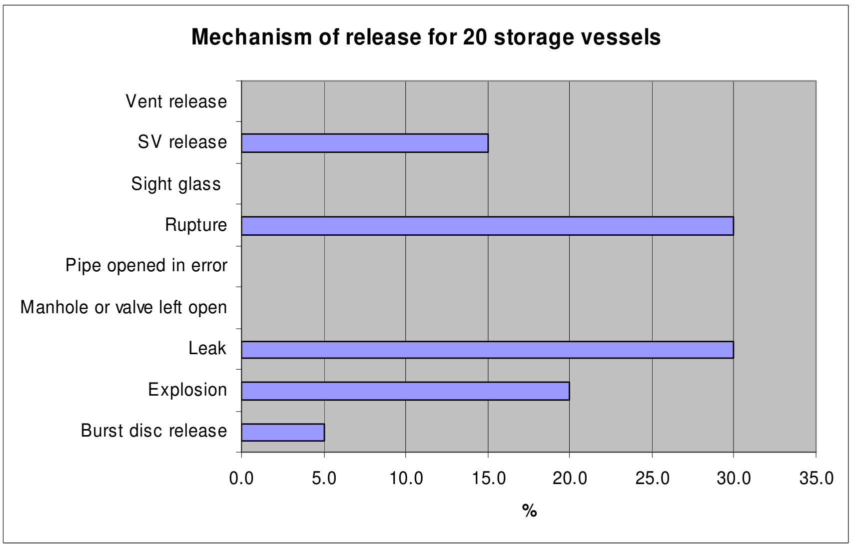


Figure 8.14 Distribution of mechanisms for storage vessel releases, MHIDAS data

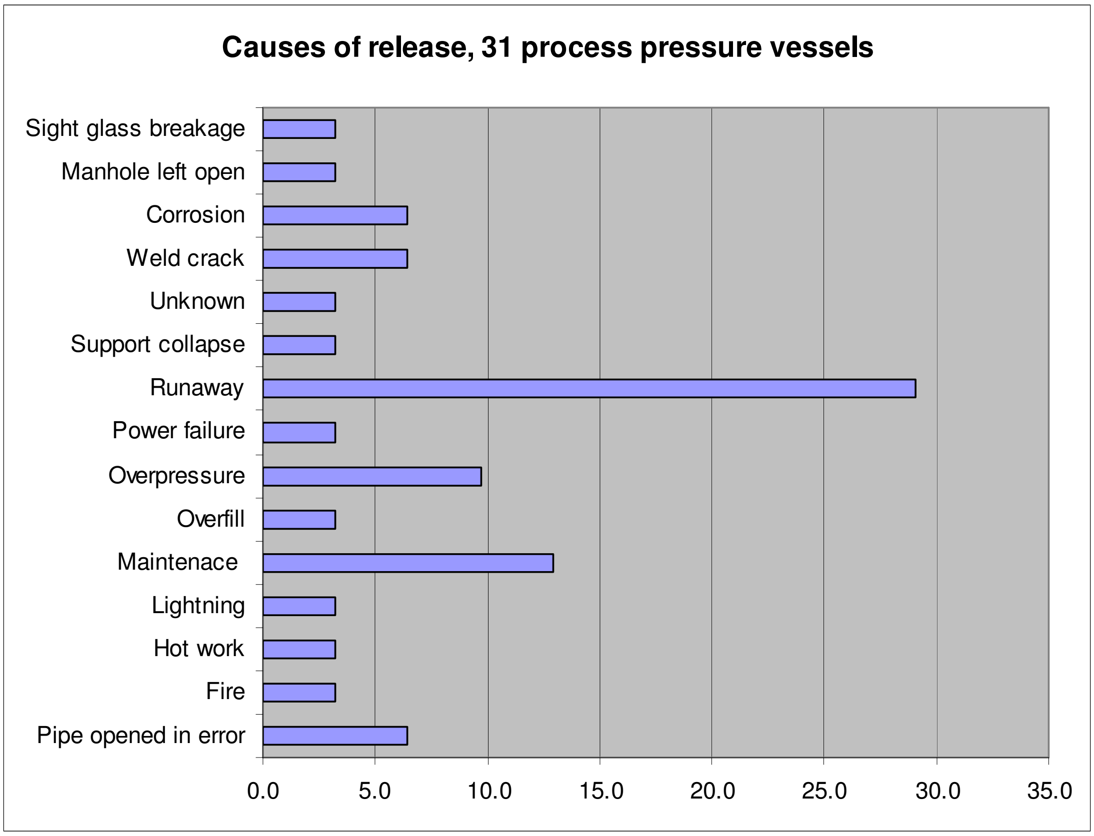


Figure 8.15 Cause distribution for process vessel releases, MHIDAS data

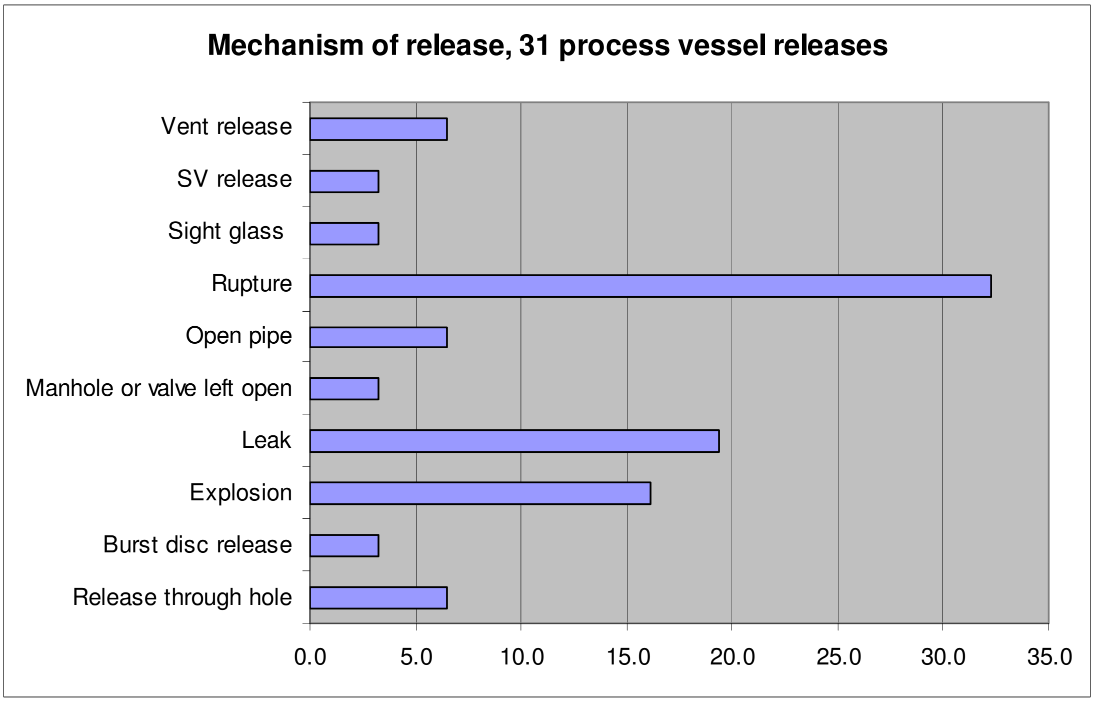


Figure 8.14 Distribution of mechanisms for process vessel releases, MHIDAS data

The MHIDAS data give rise to some conclusions concerning risk assessment practice. The vessel failures as such do indeed appear to be very rare, with ruptures due to weld failure being only about 5 % of the total. Ruptures do occur, however, arising due to overfilling, overpressuring, and especially runaway reaction. These are critical factors, which are often ignored in complete plant risk assessments, in which standard release rates are used for “vessels” without considering the underlying assumptions (see e.g. the Purple Book). Risk analyses should as a minimum take overfilling, overpressuring, and runaway reaction into account wherever there is a physical possibility for this, and should preferable take operator error, sight glass failure, and others of the noted causes into account.

8.6 Assessment of causal factors and susceptibilities

The philosophy underlying the assessment of causal factors, and modification factors is given in section 7.9. The actual assessment for causal factors is given here for pressure vessels, in table 8.4. It is based on the MHIDAS distribution of causes, and on susceptibility factor determinations from LPG and ammonia storage vessels, and from refinery process vessels.

Process vessels

Release frequencies	Small	Medium	Large	Rupture
typical	4.0E-03	3.0E-03	2.0E-03	1.00E-07

No	Failure cause	% of releases	Source	Conse-quence	Suscept-ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Internal corrosion	5.5	MHIDAS	small	1	1.0E+00	8.6E-04	
2	Internal corrosion	1		medium	1	1.0E+00	2.8E-04	
3	External corrosion	0.2	Direct	small	1	1.0E+00	3.1E-05	
4	External corrosion	0.1		medium	1	1.0E+00	2.8E-05	
5	Small bore piping failure	12	Reliability calc	small	1	1.0E+00	1.9E-03	Most vessels have small bore piping
6	Manhole or valve left open	3.2	Reliability calc	large	1	1.0E+00	6.6E-04	Most vessels have drains
7	Process piping, flanges, valves	8	Reliability calc	small	1	2.0E-02	6.2E-02	All vessels have process piping
8	Corrosion, no inspection	0.05	All DB search	medium	0.002	1.0E+00	6.9E-03	One case known
9	Corrosion, corrosive liquid	0.02	All DB search	medium	0.001	1.0E+00	5.5E-03	One case known
10	Sight glass failure	3.2	MHIDAS	medium	0.5	1.0E+00	1.8E-03	Observation
11	Support failure	3.2	All DB search	large	1	1.0E+00	6.6E-04	All vessels susceptible
12	Overfilling	3.2	MHIDAS	large	1	5.1E-03	1.3E-01	All vessels susceptible, most protected
13	Overpressure, control failure	9.7	MHIDAS	rupture	1	5.1E-02	1.3E-06	All vessels susceptible, most protected
14	Overpressure, gas breakthrough (blowby)	1	MHIDAS	rupture	0.5	5.1E-02	2.7E-07	One case known
15	Overpressure, shut in liquid	0.03	MHIDAS	large	1	5.1E-03	1.2E-03	One case known
16	External fire	4	fire statistics	rupture	1	2.1E-02	1.3E-06	All flammables vessels susceptible, but most protected
17	Weld crack, no inspection	6.5	MHIDAS	medium	0.8	1.0E+00	2.2E-03	All vessels susceptible
18	Hammer	0.001	MHIDAS	rupture	0.01	1.0E+00	6.8E-10	Vessels with long filling lines and poor level control (Texas City 1984)
19	Weather, lightning	3.2	MHIDAS	medium	0.5	1.0E-02	1.8E-01	All vessels susceptible
20	Crash, impact	2	MHIDAS	medium	0.3	1.0E+00	1.8E-03	Vessels near roadways, in plant or public
21	Foundation problem	1	MHIDAS	medium	1	1.0E-01	2.8E-03	All vessels susceptible
22	Wrong substance	0.1	MHIDAS	rupture	1	1.0E+00	6.8E-10	No cases known
23	Earthquake, landslip, flood	0.1	Reliability calc	rupture	0.05	1.0E+00	1.4E-08	Equipment on US West Coast
24	Internal explosion, runaway	29	MHIDAS	large	0.5	1.0E-02	1.2E+00	Assessment of MHIDAS DB, only for susceptible vessels
25	Valve opened in error	6.4	MHIDAS	large	0.5	1.0E-02	2.7E-01	Assessment of MHIDAS DB, only for susceptible vessels
26	Vandalism	0	MHIDAS	rupture	0.01	1.0E+00	0.0E+00	Not applicable
27	low temperature embrittlement	1	Litterat-ure	rupture	0.01	1.0E+00	6.8E-07	
	Total	103.701						

Table 8.4 Susceptibilities and modification factors for process vessels

Storage vessels

Release frequencies	Small	Medium	Large	Rupture
typical	2.0E-03	8.0E-04	3.0E-04	1.00E-07

No	Failure cause	% of releases	Source	Conse-quence	Suscept-ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Internal corrosion	1	MHIDAS	small	1	1.0E+00	3.0E-04	
2	Internal corrosion	1	Direct	medium	1	1.0E+00	1.2E-04	
3	External corrosion	0.2		small	1	1.0E+00	6.0E-05	
4	External corrosion	0.1		medium	1	1.0E+00	1.2E-05	
5	Small bore piping failure	0.5	Reliability calc	small	1	1.0E+00	1.5E-04	Most vessels have small bore piping
6	Manhole or valve left open	0.3	Reliability calc	large	1	1.0E+00	2.3E-06	Most vessels have drains
7	Process piping, flanges, valves	5	Reliability calc	small	1	2.0E-02	7.5E-02	All vessels have process piping
8	Corrosion, no inspection	0.05	All DB search	medium	0.002	1.0E+00	3.0E-03	One case known
9	Corrosion, corrosive liquid	5	All DB search	large	0.001	1.0E+00	3.9E-02	Two cases known of carry over of acid
10	Sight glass failure	0.5	MHIDAS	medium	0.003	1.0E+00	2.0E-02	Observation
11	Support failure	3.2	All DB search	large	1	1.0E+00	2.5E-05	All vessels susceptible, one case known
12	Overfilling	30	MHIDAS	large	1	5.1E-03	4.6E-02	All vessels susceptible, most protected, most releases through SV
13	Overpressure, control failure	10	MHIDAS	rupture	1	5.1E-02	1.2E-06	All vessels susceptible, most protected
14	Safety valve blockage	5	MHIDAS	rupture	0.001	5.1E-02	6.1E-04	Some vessels overpressure regularly in filling
15	Overpressure, shut in liquid	0.03	MHIDAS	large	1	5.1E-03	4.6E-05	One case known
16	External fire	1	fire statistics	rupture	1	2.1E-02	3.0E-07	All flammables vessels susceptible, but most protected
17	Weld crack, no inspection	5	MHIDAS	medium	0.8	1.0E+00	7.5E-04	All vessels susceptible
18	Hammer	0.001	MHIDAS	rupture	0.01	1.0E+00	6.3E-10	Vessels with long filling lines and poor level control (Texas City 1984)
19	Weather, lightning	5	MHIDAS	medium	0.5	1.0E-02	1.2E-01	All vessels susceptible
20	Crash, impact	5	MHIDAS	medium	0.3	1.0E+00	2.0E-03	Vessels near roadways, in plant or public
21	Foundation problem	1	MHIDAS	medium	1	1.0E-01	1.2E-03	All vessels susceptible, one incipient case
22	Wrong substance	5	MHIDAS	rupture	0.01	1.0E+00	3.1E-06	One case known (butane into pentane)
23	Earthquake, landslip, flood	0.1	Reliability calc	rupture	0.05	1.0E+00	1.3E-08	Equipment on US West Coast
24	Internal explosion, runaway	11	MHIDAS	large	0.01	1.0E+00	8.6E-03	Assessment of MHIDAS DB, only for susceptible vessels
25	Valve opened in error, draining error	5	MHIDAS	large	0.3	1.0E-02	1.3E-02	Assessment of MHIDAS DB, only for susceptible vessels
26	Vandalism	0	MHIDAS	rupture	0.01	1.0E+00	0.0E+00	
27	low temperature embrittlement	1	Literature, direct.	rupture	0.01	1.0E+00	6.3E-07	One case known, process vessel. Severel near misses observed.
	Total	100.981						

Table 8.5 Susceptibilities and modification factors for storage vessels

8.7 Detailed analysis

A detailed safety analysis for pressure vessels, with safety barrier diagrams, is given in figure 8.10 to 8.15. The safety barrier diagrams are quantified using data from chapter 5, 7 and 8, as above. The results are given in table 8.6.

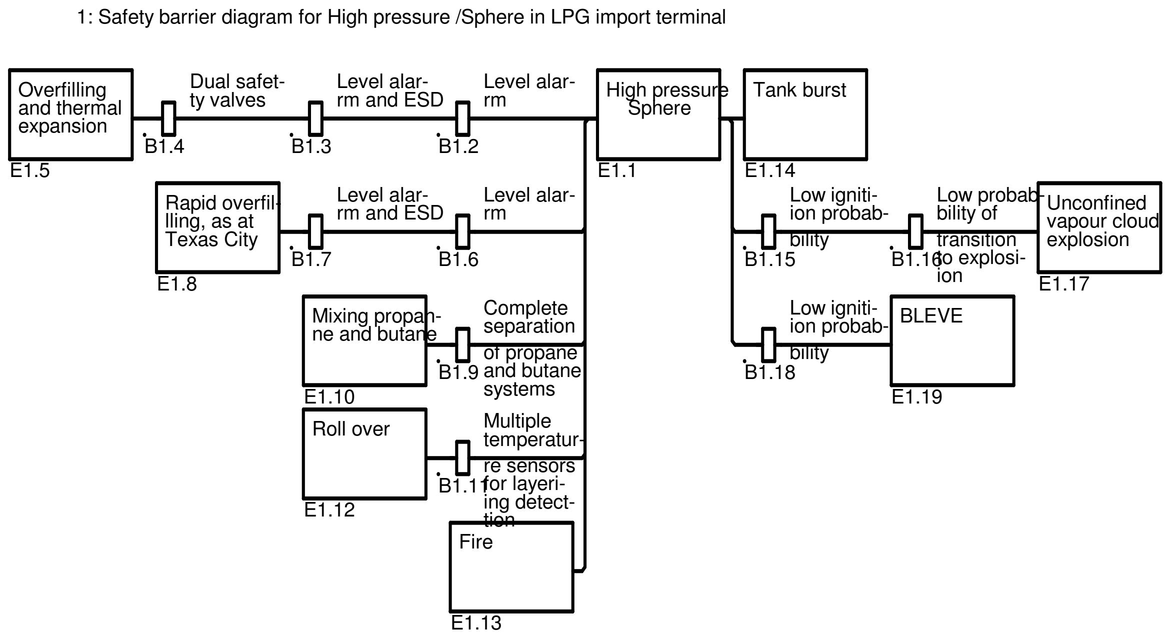


Figure 8.10 Safety barrier diagram, high pressure in pressure vessel

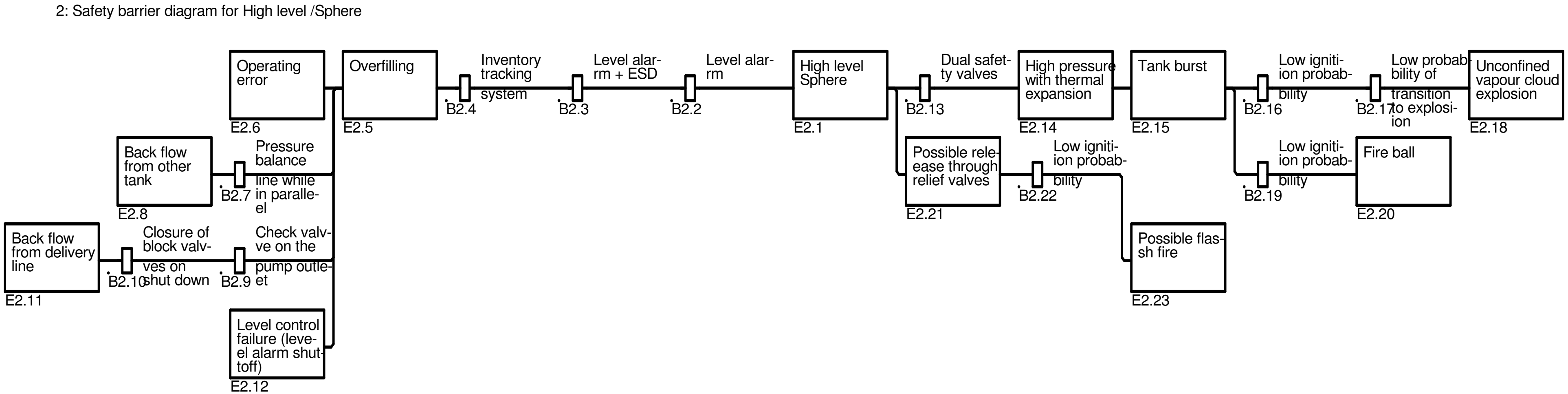


Figure 8.11 Safety barrier diagram, high level in pressure vessel

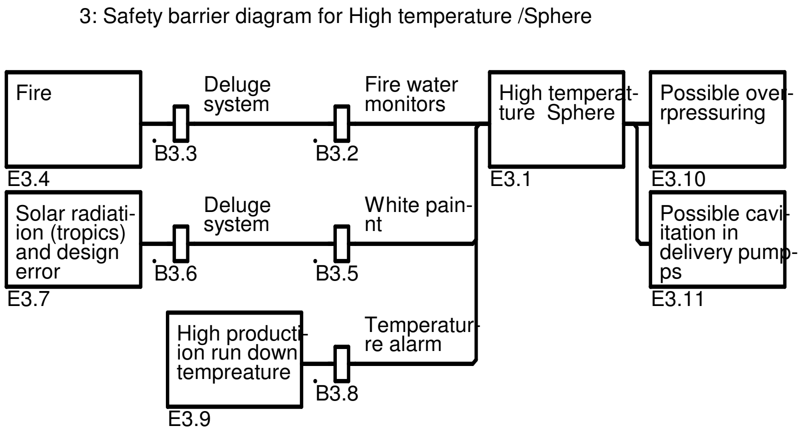


Figure 8.12 Safety barrier diagram, high temperature in pressure vessel

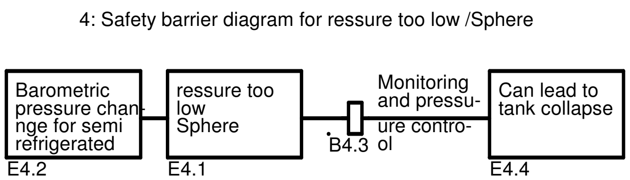


Figure 8.13Safety barrier diagram, low pressure in pressure vessel

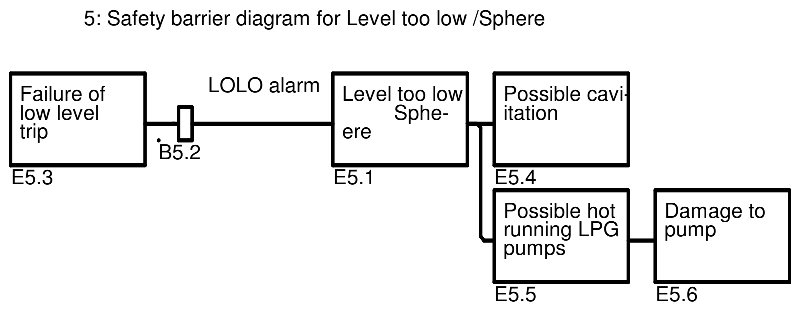


Figure 8.14 Safety barrier diagram, Low level in pressure vessel

6: Safety barrier diagram for Leak/Sphere

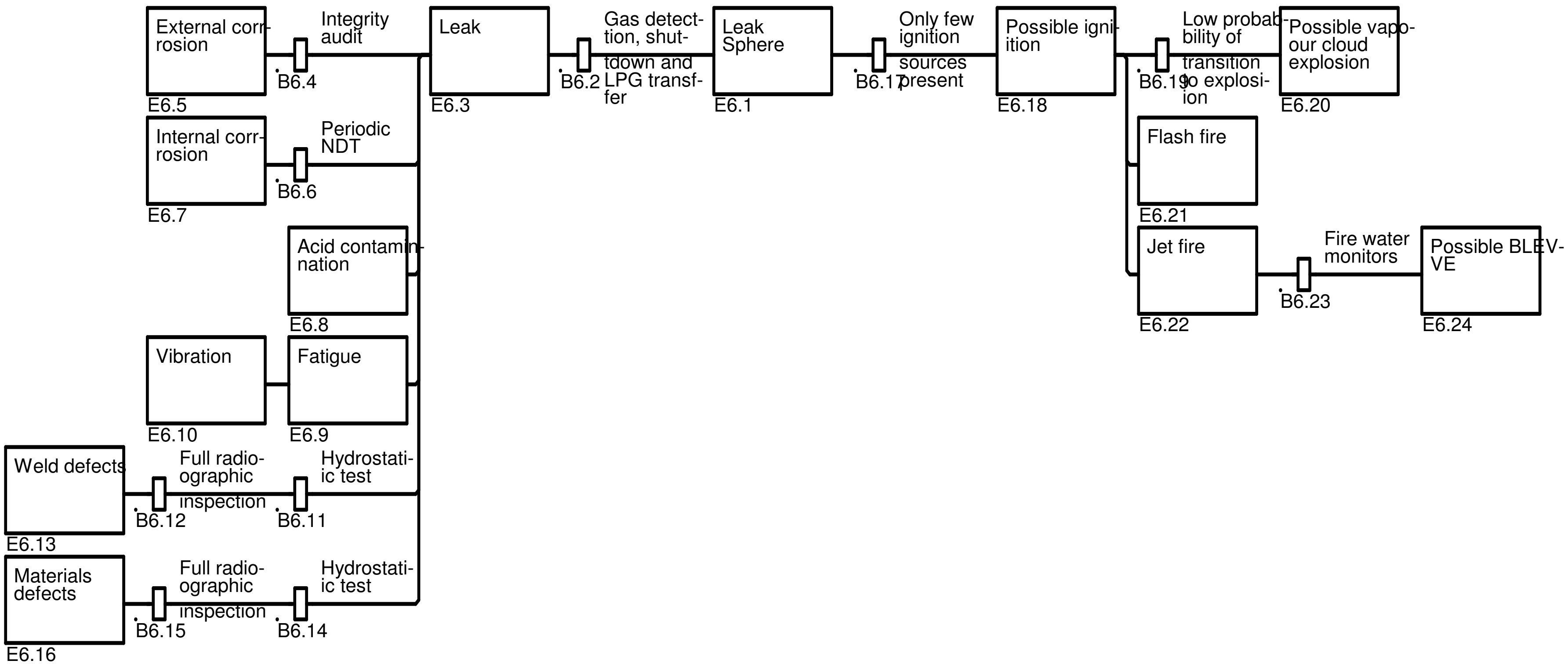


Figure 8.15 Safety barrier diagram, leak in pressure vessel

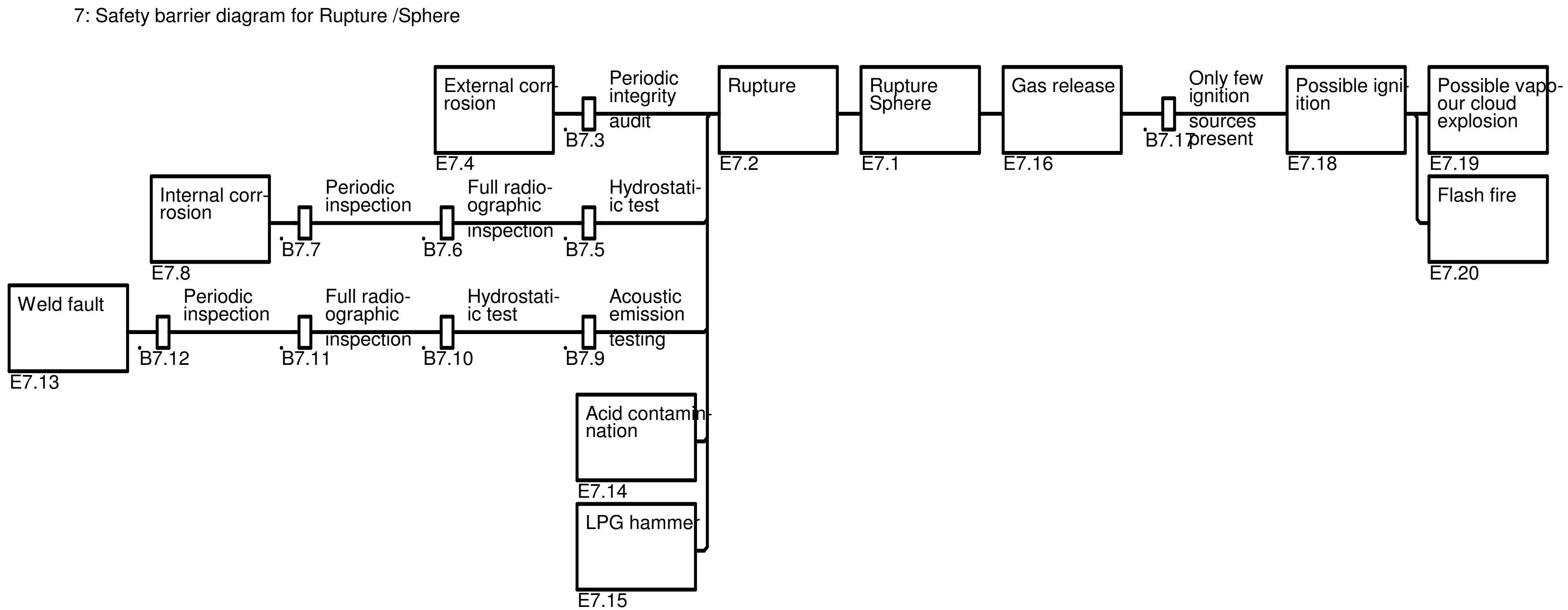


Figure 8.16 Safety barrier diagram, rupture of pressure vessel

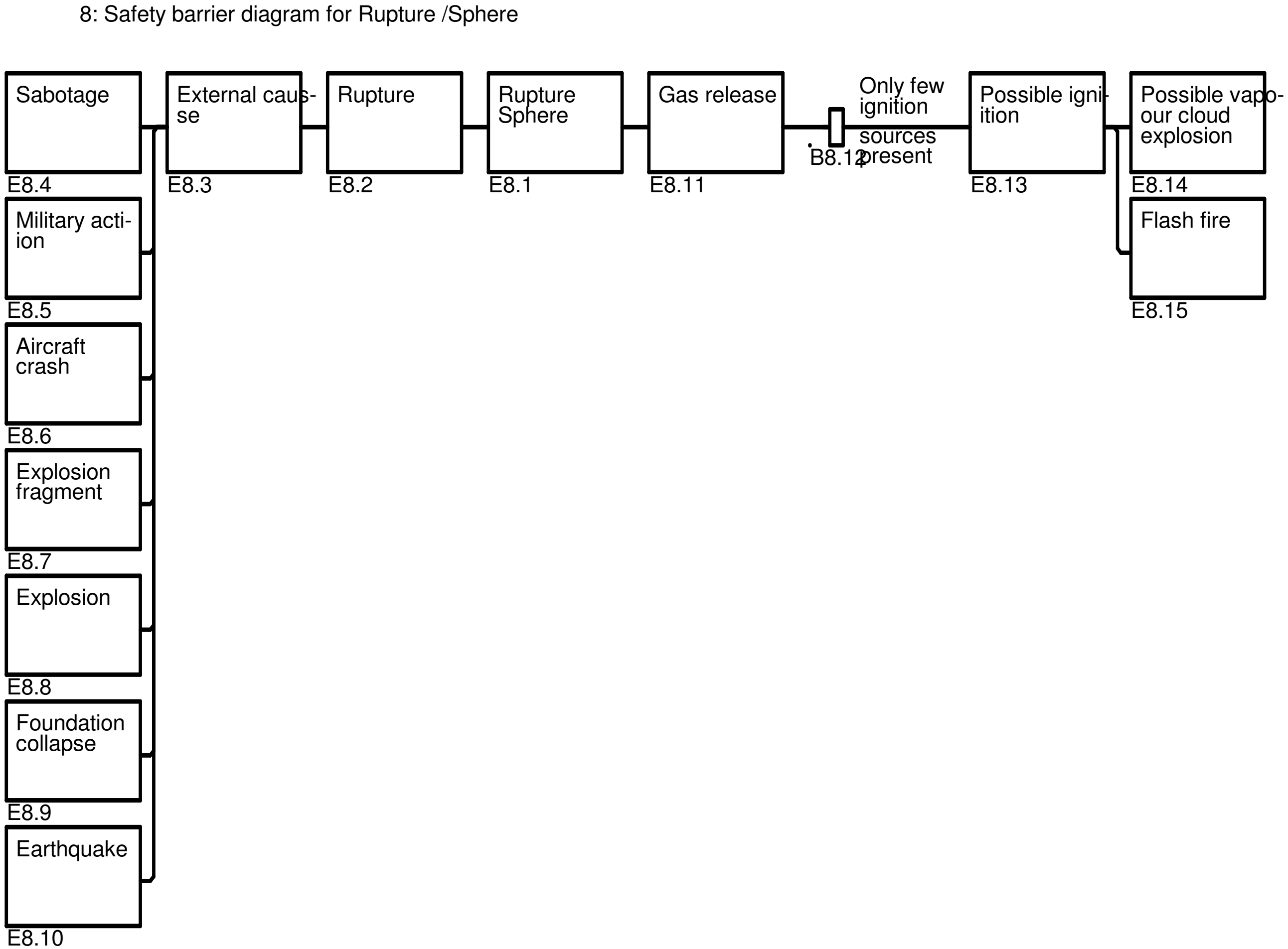


Figure 8.17 Safety barrier diagram, rupture of pressure vessel due to external causes

Release frequencies per year	**Small**	**Medium**	**Large**	**Rupture**	**Base**
Vessel	1.00E-03	3.00E-04	1.0E-05	3.00E-07	1.31E-03

Table 8.6 Process vessel release frequencies Failure cause	Release size	Number of items	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Internal corrosion	small	1	8.56E-04	1		0			0			0		8.56E-04	
Internal corrosion	medium	1	2.76E-04	1		0			0			0		2.76E-04	
External corrosion	small	1	3.11E-05	1		0			0			0		3.11E-05	
External corrosion	medium	1	2.76E-05	1		0			0			0		2.76E-05	
Small bore piping failure	small	1	1.87E-03	1		0			0			0		1.87E-03	
Manhole or valve left open	large	1	6.65E-04	1		0			0			0		6.65E-04	
Process piping, flanges, valves	small	1	6.23E-02	1		0			0			0		6.23E-02	
Corrosion, no inspection	medium	1	6.90E-03	1		0			0			0		6.90E-03	
Corrosion, corrosive liquid	medium	1	5.52E-03	1		0			0			0		5.52E-03	
Sight glass failure	medium	1	1.77E-03	1		0			0			0		1.77E-03	
Support failure	large	1	6.65E-04	1		0			0			0		6.65E-04	
Overfilling	large	1	1.30E-01	1	SV	0	0.0510798	LSHH	0	0.00351596		0		1.30E-01	
Overpressure, control failure	rupture	1	1.29E-06	1	SV	0	0.0510798	PSHH	0	0.00477394		0		1.29E-06	
Overpressure, gas blowby	rupture	1	2.66E-07	1	SV	0	0.0510798		0	0.00477394		0		2.66E-07	
Overpressure, shut in liquid	large	1	1.22E-03	1	SV	0	0.0510798		0	0.00351596		0		1.22E-03	
External fire	rupture	1	1.30E-06	1		0			0			0		1.30E-06	
Weld crack, no inspection	medium	1	2.24E-03	1		0			0			0		2.24E-03	
Hammer	rupture	1	6.80E-10	1		0			0			0		6.80E-10	
Weather, lightning	medium	1	1.77E-01	1		0			0			0		1.77E-01	
Crash, impact	medium	1	1.84E-03	1		0			0			0		1.84E-03	
Foundation problem	medium	1	2.76E-03	1		0			0			0		2.76E-03	
Wrong substance	rupture	1	6.80E-10	1		0			0			0		6.80E-10	
Earthquake, landslip, flood	rupture	1	1.36E-08	1		0			0			0		1.36E-08	
Internal explosion, runaway	large	1	1.20E+00	1		0			0			0		1.20E+00	
Valve opened in error	large	1	2.66E-01	1		0			0			0		2.66E-01	
Vandalism	rupture	1	0.00E+00	1		0			0			0		0.00E+00	
low temperature embrittlement	rupture	1	6.80E-07	1		0			0			0		6.80E-07	
Total small														6.50E-02	
Total medium														1.98E-01	
Total large														1.60E+00	
Total rupture														3.55E-06	

SV = Safety valve LSHH = Level switch high high PSHH = Pressure switch high high

Release frequencies per year	**Small**	**Medium**	**Large**	**Rupture**	**Base**
Storage vessel	8.00E-04	1.00E-04	4.0E-04	3.00E-07	1.30E-03

Table 8.7 Storage vessel release frequencies

Failure cause	Conse-quence	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Internal corrosion	small	1	2.99E-04	1		0			0			0		2.99E-04	
Internal corrosion	medium	1	1.20E-04	1		0			0			0		1.20E-04	
External corrosion	small	1	5.97E-05	1		0			0			0		5.97E-05	
External corrosion	medium	1	1.20E-05	1		0			0			0		1.20E-05	
Small bore piping failure	small	1	1.49E-04	1		0			0			0		1.49E-04	
Manhole or valve left open	large	1	2.34E-06	1		0			0			0		2.34E-06	
Process piping, flanges, valves	small	1	7.46E-02	1		0			0			0		7.46E-02	
Corrosion, no inspection	medium	1	3.01E-03	1		0			0			0		3.01E-03	
Corrosion, corrosive liquid	large	1	3.89E-02	1		0			0			0		3.89E-02	
Sight glass failure	medium	1	2.01E-02	1		0			0			0		2.01E-02	
Support failure	large	1	2.49E-05	1		0			0			0		2.49E-05	
Overfilling	large	1	4.57E-02	1	SV	0	0.0510798	LSHH	0	0.00351596		0		4.57E-02	
Overpressure, control failure	rupture	1	1.22E-06	1	SV	0	0.0510798	PSHH	0	0.00477394		0		1.22E-06	
Safety valve blockage	rupture	1	6.12E-04	1	SV	0	0.0510798		0	0.00477394		0		6.12E-04	
Overpressure, shut in liquid	large	1	4.57E-05	1	SV	0	0.0510798		0	0.00351596		0		4.57E-05	
External fire	rupture	1	2.98E-07	1		0			0			0		2.98E-07	
Weld crack, no inspection	medium	1	7.52E-04	1		0			0			0		7.52E-04	
Hammer	rupture	1	6.25E-10	1		0			0			0		6.25E-10	
Weather, lightning	medium	1	1.20E-01	1		0			0			0		1.20E-01	
Crash, impact	medium	1	2.01E-03	1		0			0			0		2.01E-03	
Foundation problem	medium	1	1.20E-03	1		0			0			0		1.20E-03	
Wrong substance	rupture	1	3.13E-06	1		0			0			0		3.13E-06	
Earthquake, landslip, flood	rupture	1	1.25E-08	1		0			0			0		1.25E-08	
Internal explosion, runaway	large	1	8.56E-03	1		0			0			0		8.56E-03	
Valve opened in error, draining error	large	1	1.30E-02	1		0			0			0		1.30E-02	
Vandalism	rupture	1	0.00E+00	1		0			0			0		0.00E+00	
low temperature embrittlement	rupture	1	6.25E-07	1		0			0			0		6.25E-07	
Total small														7.51E-02	
Total medium														1.47E-01	
Total large														1.06E-01	
Total rupture														6.17E-04	

SV = Safety valve LSHH = Level switch high high PSHH = Pressure switch high high

8.8 Algorithm for pressure vessel release frequencies

Fault trees / safety barrier diagrams for vessel leakage, and one for vessel rupture are shown above. The failure causes shown are causes which could provide additional contributions alongside the base failure frequency.

In previous release frequency data tabulations, a distinction has been made between process vessels and storage vessels. It is hard to know whether this is reasonable. In principle, the fundamental failure frequency should be similar for both process and storage vessels. Process vessels may have a slightly higher fundamental or inherent failure rate, because of the possibility of flows in fitting and welding vessel internals. The main variation, however, is in the type of usage. Such variations are best accounted for by means of modification factors, and this approach is taken here, with the following algorithms.

Fequency modification according to application

#	Question	Action if Yes	Action if No
1	Is the application for a refinery HC unit	Fire susceptibility 1	go to 2
2	Is the application for a reformer	Fire susceptibility 1	go to 3
3	Is the application for a refinery alkylation unit	Fire susceptibility 1 Internal corrosion susceptibility 1	go to 4
4	Is the application for light ends unit	Fire susceptibility 1	go to 5
5	Is the application for an ammonia vessel	External corrosion susceptibility 1 Consider hammer susceptibility	go to 6
6	Is the application for a chlorine vessel	Internal and external corrosion susceptibility 1 Consider hammer susceptibility	go to 7
7	Is the application for a bromine pipe	Internal and external corrosion susceptibility 1	go to 8
8	Is the application for an LPG storage unit	Fire susceptibility 1 Consider hammer susceptibility	go to 9
9	Is the application for an acid handling unit	Corrosion susceptibility 1	go to 10
10	Is the application for a fine chemicals unit	No modification	exit

Table 8.8

External corrosion

External corrosion does not show up often in the list of causes as a major cause of vessel failure. Nevertheless, if it does arise it will change the frequency of failure considerably.

External corrosion is found particularly in areas where there is high humidity, high rainfall, or on cold equipment such as that of a light ends fractionation unit. External corrosion is also increased ehen palns are close to the sea.

One cause of excessive corrosion arises with lagged units for which the cladding is damaged. Water then collects in the lagging and can cause extensive corrosion.

#	Question	Action if Yes	Action if No
1	Is the vessel subject to external corrosion ? (stainless vessels will generally not be)	Go to 2	External corrosion susceptibility 0
2	Is the painting of the vessel in good condition (for existing plant) or is there a good painting specification (for plant in the design stage).	Go to 3	External corrosion susceptibility 1
3	Is the vessel easily accessible ?	Go to 4	External corrosion susceptibility 1
4	Is there an integrity audit system which covers vessels on at least a yearly basis, and integrity standard which does not allow not to go untreated for more than one year.	Go to 5	External corrosion susceptibility 1
5	Is the vessel insulated	go to 6	Exit
6	Is the insulation of high quality, with silicone sealing of the cladding (for vessels in the design stage, is there a standard)	Go to 7	External corrosion susceptibility 1
7	Is there a good standard of maintenance, with proper replacement of cladding removal for maintenance	Go to 8	External corrosion susceptibility 1
8	Are there properly engineered removable panels to allow access for NDT with proper sealing after use.	Exit	External corrosion susceptibility 1

Table 8.9 External corrosion factors

Internal corrosion

Severe internal corrosion occurs typically as a result of aggressive materials such as acids, sour gas etc.

#	Question	Action if Yes	Action if No
1	Is the liquid in the vessel corrosive with respect to the material, with a corrosion rate per year comparable with 1/20 of the corrosion allowance, or more	Internal corrosion susceptibility 1	Go to 2
2	Is there a corrosion monitoring programme with corrosion coupons?	Go to 3	Internal corrosion susceptibility 1
3	Is it possible for the liquid to become contaminated with an especially corrosive component?	Internal corrosion susceptibility 1	Go to 4
4	Is stress corrosion a significant possibility for the vessel material?	Go to 5	Go to 7
5	Is it possible for the contents to become contaminated with substances (e.g. chlorides) which can cause stress corrosion cracking?	Internal corrosion susceptibility 1, go to 6	Go to 6
6	Is there a corrosion control programme which can detect incipient stress corrosion cracking?	Go to 7	Internal corrosion susceptibility 1 for rupture
7	Is there a potential for other types of corrosion e.g. hydrogen blistering?	Internal corrosion susceptibility 1	Exit

Table 8.9 Internal corrosion factors

Overfilling

Overfilling can cause rupture of a vessel if the vessel is heated, or warms in the sun, and pressure relief fails

#	Question	Action if Yes	Action if No
1	Is it physically possible to overfill the vessel?	go to 2	Exit
2	Is there a level control alarm connected to a weigh cell or reliable level sensor	Calculate the level alarm function reliability, see Ch 5	Go to 3
3	Is there a level control filling trip connected to a weigh cell or reliable level sensor	Go to 4	Overfill susceptibility 1 Rethink the design Go to 4
4	Is there an effective pressure relief	Go to 5	Overfill susceptibility 1 for rupture
5	Is it possible for the vessel to be shut in with both entry and exit valves closed?	Overpressuring due to expansion susceptibility 1 Go to 6	Go to 6
6	Is there a long run down line feeding the vessel? (>100 m.)	Hammer susceptibility 1 Go to 7	Go to 7
7	Is there a heater in the vessel?	Overpressure susceptibility 1	Go to 8
8	Are the safety valves dual, with interlocking to ensure that one is always open?	Check SV sizes Exit	Revise safety relief reliability

Table 8.11 Overfilling factors

Drain line leakage

Drain lines are used to separate off water, especially from LPG and similar vessels. Any error, particularly in allowing the drain valve to freeze open, will cause a release

#	Question	Action if Yes	Action if No
1	Is there a drain line for separation of water or for maintenance drainage on the vessel?	Go to 2	Exit
2	Does the drain line pass to a catchment vessel?	Go to 3	Exit
3	Are there two valves on the drain line?	Go to 4	Exit
4	Is one of the valves a spring return type?	Go to 5	Exit
5	Does the drain line vent to atmosphere ?	Go to 6	Exit
6	Is there a long (> 5 m) tail for the drain pipe so that the operator can close the valve safely even when there is a relief?	Susceptibility for drain valve open = 1	Exit

Table 8.12 Drain line leakage factors

Separator vessels

Separator vessels are particularly susceptible to blow by (gas flowing to the liquid system) if the liquid system is protected by level control, and is of a lower pressure specification than the vessel itself.

#	Question	Action if Yes	Action if No
1	Is there a pressure specification change in the liquid out flow line?	Go to 2	Exit
2	Is there a separate level trip and shut off valve, to prevent gas blow by?	Susceptibility for gas blow by overpressuring = 1 Calculate level control reliability	Susceptibility for gas blow by overpressuring = 1
3	Is there a separate level trip to prevent liquid reaching the gas exit line?	Susceptibility for liquid overflow hammer = 1 Calculate level control reliability	Exit

Table 8.13 Blow by factors

Liquid draw down, release from piping

#	Question	Action if Yes	Action if No
1	Is there an excess flow valve on the vessel liquid draw down?	Recalculate safety measure reliability for piping release, Go to 2	Go to 2
2	Is there an ESD valve on the liquid draw down?	Recalculate safety measure reliability for piping release, Go to 3	Exit
3	Is the ESD valve activated by a flow sensor?	Recalculate safety measure reliability for piping release	Exit
4	Is the ESD valve activated by a gas alarm system?	Recalculate safety measure reliability for piping release Go to 5	Exit
5	How good is the gas alarm coverage? 100% of all leaks?	Exit	Go to 7
6	100% of all medium and large leaks?	Modify reliability for small leaks dependent on coverage. Exit	Go to 8
7	75% of all leaks ?	Modify the reliability	Exit
8	Is the liquid flammable?	Go to 9	Set ESD reliability to 1.0
9	Is the ESD valve activated by fire sensor?	Modify ESD reliability	Set ESD reliability for fire scenarios to 1.0

Table 8.14 Draw down line factors

Low temperature embrittlement

In recent years a great deal of discussion has been made concerning low temperature embrittlement of pressure vessels during releases, with a possibility of rupture. However, a search of accident records revealed only one case of release of this kind, on an atmospheric tank, used for cryogenic material, due to defective weld marterial.

#	Question	Action if Yes	Action if No
1	Is it possible for cold liquid or flashing liquefied gas to enter the vessel?	Go to 2	Exit
2	Is the brittle transition temperature of the metal above the atmospheric boiling point of the liquid?	Go to 3	Susceptibility for brittle fracture = 1
3	Is it possible for blanketing nitrogen to enter the vessel in liquid form?	Susceptibility for brittle fracture = 1	Exit

Table 8.15 low temperature embrittlement factors

Overpressuring

#	Question	Action if Yes	Action if No
1	Is the upstream source of liquid or gas at a pressure higher than the maximum working pressure of the vessel?	Overpressure susceptibility = 1 Go to 2	Go to 3
2	Is there a high pressure trip and inlet shut off?	Modify safety measure reliability	Go to 3
3	Is it possible for back flow to occur from a high pressure source?	Overpressure susceptibility = 1 Set safety measure unavailability = 1.0	Go to 4
4	Is it possible for the wrong liquid to pass to the vessel (e.g. propane into a butane vessel or vice versa)?	Overpressure susceptibility = 1 Set safety measure unavailability = 1.0	Go to 5
5	Is roll over possible? *	Go to 6	Exit
6	Is there a mechanism for continual or periodic mixing of liquefied gas to prevent layering?	Exit	Overpressure susceptibility = 1

Table 8.16 Overpressuring factors

* Roll over and rumbling in a liquefied gas vessel is a sign of layering. True roll over requires that there is a difference in density of liquid layers, that allows them to achieve different temperatures. This is possible for example if there are different grades of LPG in a vessel.

8.9 References

1 J.R.Taylor, Process Safety Engineering, Designing and Building Safer Process Plants, ITSA , 3rd edition, 2001

2 Phillips, C.A.G and Warwick, R.G. A Survey of defects in Pressure Vessels, AHSB(s) R 162, UKAEA 1968

3 Smith, T.A. an Warwick, R.G. The Second Survey of Defects in Pressure Vessels, SRD R 30, 1974

4 Boesbeck,

5 Thomas, H.M.A probability Model for the Failure of Pressure Containing parts, Third National conf reliability, UK, 1981

6 Thomas, H.M.A Pipe and Pressure Vessel Failure Probability, Reliability engineering 2 (1981) 83-124

7 Kellerman, O. and Siepel, H.G. Analysis of the Improvement in Safety Obtained by a Containment, in *Containment and Siting of Nuclear Power Plants*, IAEA 1967

9 Fixed roof storage tanks

Storage tanks are important for risk assessment especially because these tanks will contain the largest inventories. The pattern of accidents differs greatly, depending on whether the tank is used of flammables or non flammables, and on whether the tank is a closed roof type (cone or dome roof) or floating roof type.

9.1 Description

Figure 9.1 shows the design of a typical fixed roof tank. The main features are the foundation, the tank base, tank walls and roof.

Figure 9.1 Fixed roof storage tank

The foundation is a built up circle of stone, covered with a layer of gravel, all consolidated by rolling / vibrating.

The materials for the tank base, walls, and roof are:

All welds are made to API 650.

Fittings for the tank are:

- drainage from the bottom of the tank for removal of water.

- a motor driven propeller mixer.

- in the case of heavy oils, a steam heater.

- float level gauge

- dipstick opening (thief hatch) for level gauging.

- ladders inside and outside for access. The inside ladder has rollers to allow it to glide on the floating roof.

- inlet and outlet nozzles, with valves and pumping arrangement.

- high temperature, low and high level alarms.

- fire protection in the form of scum distribution, either at the base of the tank, or with distributors to the tank rim.

The tank is surrounded by a bund or dike, which can take at least 110% of the capacity for the tank (NFPA and many national regulations). If there are two interconnected tanks, the bunded area has at least 75% of the combined capacity (Several oil company standards)

The six major types of releases for storage tanks are:

– Overflow

– Release due to errors or failures during tank drainage

– Tank piping and tank shell leaks

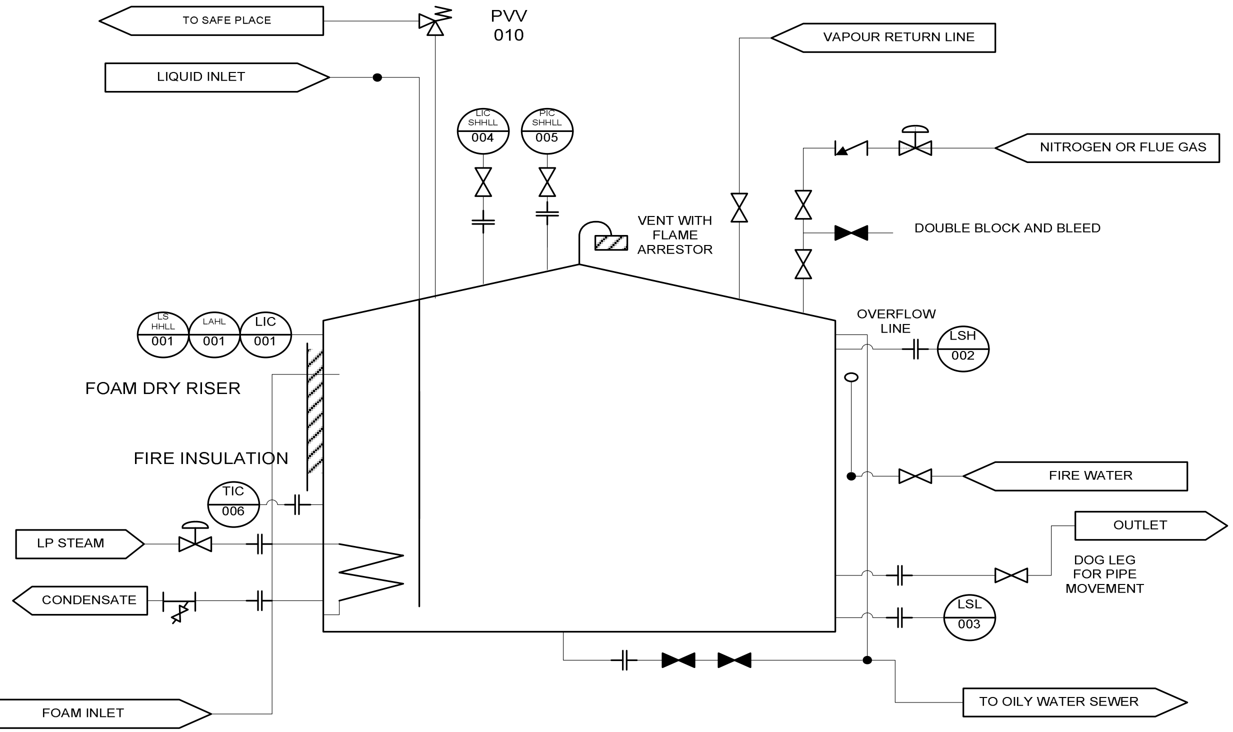
– Tank collapse

– Tank fire

– Tank explosion

Note that fire and explosion are major causes of releases, and must be taken into account, (in contrast to other equipment, where fire and explosion is largely a consequence of release).

SAFETY MEASURE	REFERENCE
LSL 003	API 14C
LSH 002	API 14C
PSV 010	API 14C
PVV 011 (PRESSURE VACUUM	PSR
VALVE ALTERNATIVE TO PSV)	
INERTING	PSR
TICAH 006 (FOR HEATED TANK)	PSR, API 14C
LIC 001 LAH LAL LSHH LSLL	PSR
(ALTERNATIVE TO LS 002, 003)	
REDUNDANT LSHHLL 004	PSR
DUAL DRAIN VALVES	PSR
OVERFLOW LINE	PSR
FIRE INSULATION	PSR
TANK COOLING FOR FIRE FOAM	NFPA 30
DRY RISER	PSR
BASIN	NFPA 30
BASIN DIVIDING WALL	PSR



PROCESS SAFETY ENGINEERING

ORDINARY CONE ROOF TANK
SAFETY FEATURES

SIZE	FSCM NO	DWG NO	REV
A4		100-002	0
SCALE 1:1		SHEET 1 OF 1	

JRT 15 APRIL 2000

9.2 Tank hazards (from ref 9.1)

One feature of storage tanks which is not seen to the same extent in other equipment is that of fires and explosions occurring without a preceding release. Fixed roof tanks (cone roof and dome roof) can also be subject to internal explosions, and subsequent fires. Light products such as gasoline and light solvents will often have such a high volatility that the vapour space in the tank is above the upper explosion limit.

There will then only be sufficient air for the vessel to undergo an internal explosion just after the tank has been drawn down, in order to use the liquid, or to load a tank truck etc. in some cases, the risk is reduced by providing nitrogen blanketing.

In some large tanks, vents are provided so that flammable vapour can be blown away by the wind. Flammable atmosphere then only exists in the tank when winds speeds are low.

Heavier liquids, such as fuel oil, xylene and diesel oil, do not have a vapour pressure high enough, under normal circumstances, to allow the lower explosion limit to be reached. Flammable atmospheres can exist though if:

– The liquid in the tank is contaminated by light fractions or hydrogen

– Hot liquid is pumped into the tank

– There is a steam leak from the tank heater

– The temperature control for the tank heater fails

Some liquids have vapour pressures which give concentrations between the upper and lower explosion limits (UEL and LEL), and will always have an explosive mixture of vapour and air in the tank, unless this is blanketed, or has wind vents.

Ignition of vapour inside closed roof tanks can occur by:

– Electrostatic discharge during sample taking or depth gauging

– Lightning

– Electrical failures on level or temperature gauges

– Pyrophoric sulphide ignition for liquids containing hydrogen sulphide

– Auto catalytic decomposition of residues on tank walls (for vegetable oils and for asphalt

– Runaway reactions, in chemical waste

– Reactions due to admission of the wrong substance e.g. nitric acid into a flammable.

Pumping hot or light fractions into tanks can cause vapour flashing explosions, as can pumping cold, light liquids into tanks with hot oil. And example is that of pumping (or leaking) butane into a pentane tank. Another is pumping hot oil residues into a normally empty gasoline tank, which then exploded.

Inadvertent mixing of substances in tanks is much more common than might be expected. For example, 4.7% of the incidents in tanks recorded in the MHIDAS data base arose due to pumping acid into sodium hypochlorite, and vice versa 2.8% involved pumping of nitric acid into sodium hypochlorite.

Acid tanks may be subject to explosion due to generation of hydrogen when the acid reacts with the tank shell. Usually, the hydrogen is vented away but this does not occur if the vent is blocked. Internal baffles have also been known to trap hydrogen.

One special phenomenon, which is largely unrecognised in the literature but relatively common in practice is fire induced tank explosion (FITE). Here, an external fire causes vapour to be generated, and also provides the ignition source. If the vessel is only partly full, typically 1/4 to 1/3 the tank may be blown into the air like a rocket, travelling typically 50 to 100 m.

Both cone roof and fixed roof tanks can undergo the phenomenon of boil over. This occurs only if there is a) a full surface fire on the tank, (in the case of a fixed roof tank, this requires that the roof has blown off in an explosion, and b) a mixture of liquids with a wide range of boiling points and c) a small residue of water in the tank. The surface reaches water at the steam generated causes hot oil at the top to mix with cold oil from the bottom. The light fractions in the cold oil flash off explosively, ejecting a large fire ball.

9.3 Case stories (ref 9.1)

1. Asphalt tank

An asphalt tank storing heated asphalt caught fire due to autoignition in build up on the tank wall. The fire burned for a day and destroyed the upper part of the tank. (Personal observation)

2. Sour corrosion

A fuel oil tank which had once been charged with high sulphur fuel oil showed heavy corrosion of the roof about two years later. In some areas the tank roof resembled lace. (Personal observation)

3. Mixing of products

Heavy residual oil at about 300 deg C was pumped to a tank considered to be empty. The small amount of naphtha in the bottom of the tank vapourised, and hot resid was thrown out of the tank roof, covering the cars in the manages car park. (Personal observation)

4. Mixing

Chilled butane was pumped to a pentane tank in error. The butane flashed and this forced the roof off the tank. about 20 000 barrels of pentane escaped, but was recovered over about two days, without ignition. (Personal observation)

5. Lightning (API)

Lightning struck a 260 000 barrel cone roof tank containing diesel. The vapour above the diesel exploded. Sections of the tank roof landed on other tanks. One containing gasoline ignited immediately across the whole surface.

On another floating roof tank containing gasoline, fragments landed, and fires started several hours later.

Normally vapour above diesel is not explosive, but gases from the diesel desulphurization had been carried over into the oil

6. Static (Personal observation)

A slop tank exploded when a spark occurred. The spark was apparently generated by floating polystyrene foam that had been used as packing for an instrument.

7. Ice (API)

A column of ice built up between a goose neck vent and the tank roof. in freezing weather. This blocked the vent. Cooling caused the tank to collapse, the roof and the wall buckling. Problems may have arisen due to insulation of the tank roof, which stopped condensation freezing at the tank roof, allowing it to leave the vent even on cold days.

8. Operations error (API)

An instrument technician working to improve system performance lighted fired heaters, thinking that this was normal practice. He did not inform the operator. In fact the tank had a low level of oil.

The operator noticed almost immediately on the control room panel that the heaters were on, and asked a process technician to shut them down. About 5 minutes after the request, the technician reported that the tank was on fire. The technician noted at this time that part of the roof had lifted 18 to 24 inches. The technician pushed all four stop buttons. He heard a rumble, and saw the tank sucked in as if by a vacuum.

The fire was allowed to burn out.

There were four independent shutdown systems on the gas fired heaters. Two were not working, and two were ineffective because they were not designed to take account of start up with low level.

9. Fire induced tank explosions (Loss Prevention and Safety Promotion Conf. 1994)

At Port Edouard Heriot in 1987 an explosion occurred in a 1/ full tank

1. Work on new modifications in

progress

2. 13:30 release near the pumps

3. Flash fire injures 8

4. 13:31 Small explosion

5. 13:40 250 m3 additives vessel

explodes, rockets, main part

travels 60 m.

6. Fire spreads

7. 17:00 Fire begins to diminish

8. 18:45 Tank 6 explodes

9. 6:30 Large foam attack

10 11:00 Fire under control

The fire explosion inside the tank was ignited by fire outside the tank. The bottom seal ruptured.

10. Hydrogen in gas oil tank

Hydrogen was carried over to a gas oil (diesel) tank from a diesel stripper, due to an unusual entrainment phenomenon. The hydrogen was subsequently ignited, presumably by an unearthed sample bottle ignited the hydrogen. Confusion in the standards for sampling may have contributed to the error. The roof was blown off and the tank burned out. A second tank was damaged. Foam piping was severely damaged by the fire in the bund, hanging like spaghetti, but did not break.

9.4 Frequency of tank releases

Large storage tanks have many and varied designs of which the principal ones are cone roofed tanks and floating roofed tanks. Corrosion is a possibility here, and leads to a need for periodic inspection. Failure rates depend very much on the material stored. Apart from various valves, the most likely causes of leaks are either overfilling (this can cause tank roofs to raise) and corrosion attacks around welds.

Catastrophic failures generally arise from overfilling, as a result of operator error, administrative error, or instrument failure, and internal fires or explosions in oil tanks. The reliability of inerting systems used in hydrocarbon storage tanks to prevent explosive atmospheres arising requires a special study in itself. A typical pattern involves overfilling or overpressuring, which then results in a release from a weak, improperly welded, or corroded tank seam. Japanese data gives values of 10^{-5} per year for such failures. (Ref 9.2)

RMP data for tank releases gives, for crude units, a value of 0.0024 per tank year. The amount of data available from the RMP data for tanks is very low. This is not surprising – tank releases occur at close to atmospheric pressure and so release rates are small. Also, the materials stored in such tanks generally have a low vapour pressure. For these reasons, only relatively few release accidents will have offsite consequences, when compared, for example to pressure vessels.

UK HSE offshore data gives the following failure rates:

Equipment type	Failure frequency per year	< 10 mm	10-25 mm	25-50 mm	>100 mm
Crude oil tank	2.57E-03	1.29E-03	0.15E-03	0.80E-03	0.15E-03

Table 9.1 Tank failure rates from UK HSE off shore release frequency data (ref. 9.3)

This data is compatible with the RMP data – in fact the consistency is surprising, and certainly coincidental.

Christensen and Eibert (ref. 9.4)carried out an extensive survey of atmospheric storage tank releases from oil production and refining in the USA. Many of these will be floating roof tanks, so the data is mixed The number of tanks is given in table 9.4.2

Oil industry segment	Surveyed tanks	Estimate of total population	Average age
Marketing	5831	88529	29.4
Refining	11440	29727	34.6
Transportation	5341	9197	31.4
Production	54046	572620	15.1
Total	76708	700073	17.9

Table 9.2 Tanks in the Christensen and Eibert survey.

The release frequency determined for these was $1.5*10^{-2}$ per year. The size of the releases is not given, but E&P Forum (ref.9.5) used data from Lauben and Robinson (ref. 9.6) to calculate a value for major tank release frequency of $6.9*10^{-6}$ per year, based on 92 major tank failures.

The Lauben and Robinson study describes the results of integrity testing carried out by the Hartford Boiler company. The paper describing the study gives about 16000 leaks, and about 92 major release incidents. Their study gives a release frequency of $2.5*10^{-2}$ per year based on a sample of tank inspections of 835 tanks.

A collection of 206 oil and distillates tanks with spill data from a period of 41 years, studied by the author, gave a leak frequency $6*10^{-3}$ per year, with a size distribution as shown in figure 9.4.1. Note that many of these tanks, especially the larger ones, are floating roof types, so the data is mixed.

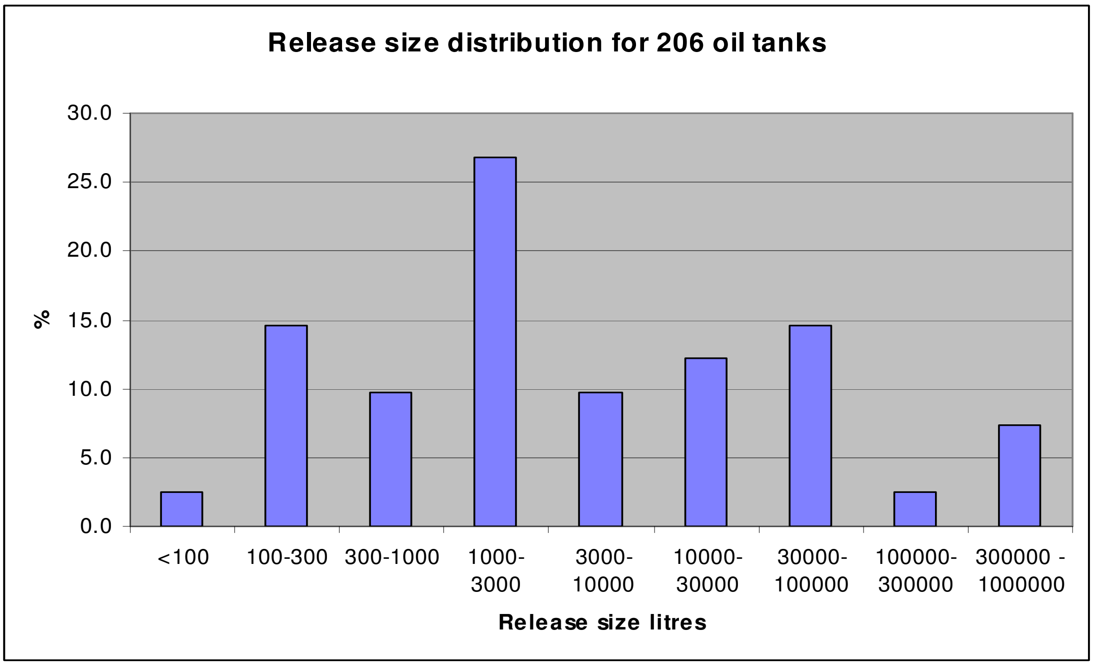


Figure 9.3 Release size distributions for 206 oil tanks

Ruptures are relatively rare for oil tanks, but do apparently occur. By contrast, tank rupture is much more frequent for chemicals and particularly acid tanks with up to 17% of large releases being due to complete rupture. Figure 9.4.2 shows one of the cases where extensive damage has occurred.

As a basis for comparisons later, when selecting a base frequency, the release frequencies for oil tanks are compared in table 9.4.3

Source	Small leaks up to 25 mm	Medium leaks up to 50 mm	Large leaks	Rupture, catastrophic
HSE	$1.44*10^{-3}$	$0.80*10^{-3}$	$0.15*10^{-3}$	
Christensen and Eibert, E&P Forum	$15*10^{-3}$ all releases			
Lauben and Robinson, HSB	$25*10^{-3}$ all releases			$6.9*10^{-6}$
Company G	$0.9*10^{-3}$	$3*10^{-3}$	$2.1*10^{-3}$	$0.047*10^{-3}$

Table 9.3 Frequencies of releases from tanks, data from different sources

Figure 9.2 Sulphuric acid tank explosion/collapse at Motiva

Tank release frequencies for smaller tanks such as those generally used in chemical plant tank yards are very difficult to find data for, not least because it is difficult to determine the tank population. The range of tanks varies from less than a cubic metre to tanks up to 1 m. high by 5 m. diameter, and deciding which tanks are relevant is a difficult issue. One thing that is certain is that there are many more causes of failure for chemicals tanks than for typical oil industry tanks. Table 9.4 below shows many cases of reactions in storage tanks for chemicals, and Table 9.5 many cases of release of chlorine from acid tanks due to pumping of sodium hypochlorite into the wrong tank. Releases from chemicals storage tanks are also much more likely to cause offsite consequences, due to the formation of toxic pools.

Figure 9.3 Leak from an acid tank due to corrosion at the acid, wall, air interface

Actual rupture of tanks is usually considered to be rare, but a few tanks of special manufacture have showed systematic failures, to the extent that US EPA sent out a special warning in 2000. Five tanks from one manufacturer alone give a frequency contribution for rupture of $5*10^{-6}$ per year. Ruptures are relatively rare for oil tanks, but do apparently occur. By contrast, tank rupture is much more frequent for chemicals and particularly acid tanks with up to 17% of large releases being due to complete rupture. Figure 9.2 shows one of the cases where extensive damage has occurred.

Fires on tanks are a separate issue from releases. They may arise as the result of a tank release, or may arise due to internal causes.

API give regular data for refinery fires and explosions. Cox, Lees and Ang quote this data, with 201, 173, 142 and 104 fires in 1982 to 1985 respectively. The distribution of fires in 1985 were:

Size of loss $1000	No of fires	Proportion %	Frequency per refinery year
2.5 - 100	65	60	0.58
100 - 1000	37	34	0.33
> 1000	7	6	0.058

Table 9.4 Frequency of fires, API data quoted by Cox, Ang and Lees (ref. 9.7)

The average frequency of large fires per refinery was estimated to be 0.28 large fires per refinery year (bases on 225 refineries). Very few of these are large tank farm fires, though. Fixed roof tank fires are less frequent than for floating roof types described in the next chapter, due largely to the much lower exposure to the effects of lightning, and to the possibility of blanketing. Table 9.5 gives the frequency for cone roof tank fires quoted from E&P Forum, and supplemented with data from the authors own experience from accident investigations.

Country	Data source	# fires	Tank years	Fire freq. per tank year
USA	API Risk Analysis Task force	270	900,000 est	$3.0*10^{-3}$
Singapore	OPITSC members	2	11125	$1.8*10^{-4}$
	TA experience	4	3880	$1.0*10^{-3}$

Table 9.5 Fixed roof fire frequency rate data reported in ref 9.5, and TA data

Tank explosion data is more difficult to find in the published literature. The author investigated in all 5 tank explosions in a representative group of companies, with a total of 8298 relevant tank years of experience, giving a frequency of $3.6*10^{-4}$ vapour combustion explosions per tank year, and $2.4*10^{-4}$ hot or light fraction inpumping explosions per tank year

9.5 Typical release frequency data for fixed roof tanks

The data in the previous section provides a good background for selection of a baseline failure rates for tanks, in particular table 9. 3 and 9.5. The selected values are given in table 9.6. The baseline values are the minimum of those observed. The typical values are an average from 9.3 and the values from section 9.4. Note that the RMP data does not give a sufficient number of cases to be used as a basis, presumably because there are too few incidents with offsite consequences.

	Small leaks up to 25 mm	Medium leaks up to 50 mm	Large leaks	Rupture, catas-trophic	Fire, tank top	Fire, tank basin	Explo-sion
Baseline	1E-3	0.5E-3	0.1E-3	6E-6	-	2E-4	1E-4
Typical acid, chemical	3E-3	1E-3	0.2E-3	5E-5	-	3E-3 (flammable)	6E-4
Typical, solvent, fuel	4E-3	1E-3	0.2E-3	1E-5	-	3E-3	9E-4

Table 9.6 Typical failure rates per tank year for fixed roof tanks

Tank failure causes

Tables 9.7 to 9.12 show a break down of tank release and fire causes drawn from MHIDAS data base.

Causes	% of total	% of known causes
Tank type		
Cone roof tank	56	73.7
Floating roof tank	10	13.2
Not recorded	10	13.2
	76	
Cause		
Double filling	1	1.3
Drain leak	1	1.3
Earthquake	1	1.3
External fire	1	1.3
Hole in tank	1	1.3
Hose leak	1	1.3
Hose rupture	1	1.3
Leak into sewer	1	1.3
Loading spill	1	1.3
Maintenance	1	1.3
No leak	23	29.5
Overflow	12	15.4
Pipe leak	5	6.4
Pump	1	1.3
Reverse flow	1	1.3
Unknown	1	1.3
Unrecorded	15	19.2
Valve closure	1	1.3
Valve leak	2	2.6
Vapour on hot windless day	2	2.6
Effect		
Roof sank	1	1.3
Spill	1	1.3
Tank leak	1	1.3
Tank rupture	2	2.6
Consequence		
Explosion	34	44.7
Fire	34	44.7
Release, No ignition	8	10.5
	76	

Table 9.7 Causes of releases from gasoline tanks, MHIDAS

Cause	Number	% of total	% of known causes
Arson	2	2.2	3.8
Car	1	1.1	1.9
External fire	2	2.2	3.8
Furnace	2	2.2	3.8
High temperature	1	1.1	1.9
Hot work	4	4.4	7.5
House	1	1.1	1.9
Incinerator	1	1.1	1.9
Lightning	16	17.6	30.2
Maintenance light	2	2.2	3.8
Mobile home	1	1.1	1.9
Nearby facility	1	1.1	1.9
No ignition	8	8.8	15.1
Process heater	1	1.1	1.9
Pump	1	1.1	1.9
Sabotage	1	1.1	1.9
Sparks	1	1.1	1.9
Static	1	1.1	1.9
Static, no bonding	1	1.1	1.9
Tank rupture	1	1.1	1.9
Tractor	1	1.1	1.9
Traffic	1	1.1	1.9
Truck	2	2.2	3.8
Unknown	3	3.3	
Unrecorded	35	38.5	
	91		

Table 9.8 Ignition causes for gasoline tanks, MHIDAS

Accident cause	Number	%
Broken valve	2	4.1
Damage due to theft	3	6.1
Explosion	8	16.3
External fire	4	8.2
Fire	4	8.2
Foundation collapse	1	2.0
Leak cause unrecorded	5	10.2
Leak, cold weather	1	2.0
Overflow	1	2.0
Pipe break	2	4.1
Rupture	2	4.1
Sabotage	10	20.4
Truck crash into piping	1	2.0
Valve left open	4	8.2
Wind damage	1	2.0
	49	

Table 9.9 Causes of releases and fires in diesel tanks, MHIDAS

Accident cause	Number	%
Internal explosion	23	38.3
External fire	3	5.0
Drain, filter etc left open	1	1.7
Fire	9	15.0
Hose leak	3	5.0
Leak, cause unrecorded	1	1.7
Overflow	8	13.3
Fittings broken	1	1.7
Overheating overpressure	2	3.3
Overpressure by clearing blockage with air	2	3.3
Corrosion	1	1.7
Sabotage	2	3.3
Spillage	2	3.3
Valve left open	1	1.7
Vapour fire	1	1.7
	60	

Table 9.10 Causes of releases and fires in solvent tanks, MHIDAS

Cause	Number	%
Corrosion	4	3.8
Fittings broken	3	2.8
Crack	2	1.9
Explosion	9	8.5
External fire	3	2.8
Fire in tank	1	0.9
HCl hypochlorite mixing	5	4.7
HCl nitric acid mixing	3	2.8
Hose fault	2	1.9
Leak cause unrecorded	31	29.2
Overflow	5	4.7
Pipe break	5	4.7
Pipe melted	1	0.9
Sabotage	2	1.9
Spillage during loading	4	3.8
Tank rupture	17	16.0
Loading into leaking tank	1	0.9
Valve failure	8	7.5
	106	

Table 9.11 Causes of releases from acid tanks, MHIDAS

Cause	%
Arson	0.6
Blanketing failure	0.6
Breakage of fitting	0.6
Collapse of tank Total	1.3
Crash	2.6
Tank entry	0.6
Explosion	6.4
External fire	1.9
Fire in tank	7.1
High pressure	6.4
High temperature	7.1
Hot work explosion	4.5
Hot work fire	1.9
Lightning	5.8
Maintenance damage	1.3
Overfilling	8.3
Reaction	9.6
Rupture	3.8
Sabotage	3.2
Tank leak	14.7
Unknown	6.4
Valve leak	3.2
Valve rupture	2.6
Valve left open	1.3
Valve opened by mistake	1.3
Valve stuck open	0.6
Vandalism	0.6
Wrong material put into tank	5.1

Table 9.12 Causes of releases from chemicals tanks, MHIDAS

9.6 Assessment of failure causes and susceptibilities

The philosophy underlying the assessment of causal factors, and modification factors is given in section 7.9. The actual assessment for causal factors is given here for pressure vessels, in table 9.13 and 9.14. It is based on the MHIDAS distribution of causes, and on susceptibility factor determinations from gasoline , chemical and acid tanks.

Release frequencies	Small	Medium	Large	Rupture
Typical	3.00E-03	1.00E-03	2.00E-04	5.00E-05

No	Failure cause	% of releases	Source	Conse-quence	Suscept - ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Internal corrosion	8.7	MHIDAS	small	1	1	9.5E-04	
2	Internal corrosion	1.9	MHIDAS	large	1	1	2.1E-05	
3	External corrosion	2.2	MHIDAS	small	1	1	2.4E-04	
4	External corrosion	1.8	MHIDAS	large	1	1	2.0E-05	
5	Small bore piping	0.6	Reliability calc	small	1	1.0E+00	6.5E-05	Most tanks have have small bore piping
6	Process piping, flanges, valves	16	Reliability calc	small	1	2.0E-02	8.7E-02	All vessels have process piping
7	Drain lines left open	2	MHIDAS	large	0.5	1	4.4E-05	Survey of plant
8	Maintenance error	2	MHIDAS	medium	1	1	4.8E-04	Survey of plant
9	Tank entry	1	MHIDAS	explosion	1	1	1.0E-05	Survey of plant
10	Corrosion, no inspection	0.2	MHIDAS	medium	0.5	1	9.5E-05	Survey of plant
11	Pipes and fittings	10	Calc	large	1	1	1.1E-04	
12	Reverse flow	0.5	MHIDAS	large	0.3	1	1.8E-05	Survey of plant
13	Overfilling	24	MHIDAS	rupture	1	0.03516	1.1E-03	Survey of plant
14	External fire	2	MHIDAS	medium	1	1	4.8E-04	Survey of plant
15	Weld crack	0.5	MHIDAS	rupture	1	1	7.8E-07	Survey of plant
16	Earthquake, landslip, flood	2	MHIDAS	rupture	0.05	1	6.2E-05	
17	Internal explosion	2	MHIDAS	large	0.2	1	1.1E-04	Survey of plant
18	Vandalism, third party	1	MHIDAS	rupture	1	1	1.6E-06	Survey of plant
19	Vapour plume on windless day	4	MHIDAS	explosion	0.2	1	2.0E-04	Survey of plant
20	Wrong substances	0	MHIDAS	rupture	0.1	1	0.0E+00	Survey of plant
21	Rupture, weld, material	4	MHIDAS	rupture	1	1	6.2E-06	Survey of plant
22	Blanketing failure	0	Experience	large	0.2	1	0.0E+00	Survey of plant
23	Crash	0.5	MHIDAS	rupture	0.1	1	7.8E-06	Survey of plant
24	Overpressuring	0.1	MHIDAS	rupture	0.2	1	7.8E-07	Survey of plant
25	High temperature	0.1	MHIDAS	rupture	0.05	1	3.1E-06	Survey of plant
26	Lightning	12	MHIDAS	fire	1	1	5.0E-05	Survey of plant
	Total	99.1						

Table 9.13 Susceptibilities and modification factors for flammable liquid tanks,.

Release frequencies	**Small**	**Medium**	**Large**	**Rupture**
Typical	3.00E-03	1.00E-03	2.00E-04	5.00E-05

No	Failure cause	% of releases	Source	Conse-quence	Suscept - ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Internal corrosion	12.7		small	1	1	1.7E-03	
2	Internal corrosion	3.8		large	1	1	3.4E-05	
3	External corrosion	2.2		small	1	1	2.9E-04	
4	External corrosion	1.8		large	1	1	1.6E-05	
5	Small bore piping	0.6	Reliability calc	small	1	1.0E+00	8.0E-05	Most vessels have small bore piping
6	Process piping, flanges, valves	7	Reliability calc	small	1	2.0E-02	4.7E-02	All vessels have process piping
7	Valve left open,opene by mistake	2.6	MHIDAS	large	0.5	1	4.7E-05	Survey of plant
8	Maintenance error	1.3	MHIDAS	medium	1	1	4.6E-04	Survey of plant
9	Tank entry	0.6	MHIDAS				0.0E+00	
10	Corrosion, no inspection	0.5	MHIDAS	medium	0.5	1	3.6E-04	Survey of plant
11	Reverse flow	2	MHIDAS	large	0.3	1	6.0E-05	Survey of plant
12	Overfilling	8.3	MHIDAS	large	1	0.03516	2.1E-03	Survey of plant
13	Internal fire	1.9	MHIDAS	rupture	1	1	1.4E-05	Survey of plant
14	Weld crack	1	MHIDAS	medium	1	1	3.6E-04	Survey of plant
15	Earthquake, landslip, flood	0.5	MHIDAS	rupture	0.05	1	7.2E-05	
16	Internal explosion	4.5	MHIDAS	rupture	0.2	1	1.6E-04	Survey of plant
17	Vandalism, third party	3.8	MHIDAS	large	1	1	3.4E-05	Survey of plant
18	External fire	1.9	MHIDAS	rupture	0.2	1	6.9E-05	Survey of plant
19	Wrong substances	5.1	MHIDAS	rupture	0.1	1	3.7E-04	Survey of plant
20	Rupture weld material	5.1	MHIDAS	rupture	1	1	3.7E-05	Survey of plant
21	Blanketing failure	0.6	Experience	rupture	0.2	1	2.2E-05	Survey of plant
22	Crash	2.6	MHIDAS	large	0.1	1	2.3E-04	Survey of plant
23	Overpressuring	6.4	MHIDAS	rupture	0.2	1	2.3E-04	Survey of plant
24	High temperature	7.1	MHIDAS	rupture	0.05	1	1.0E-03	Survey of plant
25	Lightning	5.8	MHIDAS	rupture	1	1	4.2E-05	Survey of plant
26	Runaway reaction	9.6	MHIDAS	explosion	0.1	1	5.0E-04	Survey of plant
	Total	99.3						

Table 9.14 Susceptibilities and modification factors for chemicals tanks

Release frequencies	Small	Medium	Large	Rupture	fire	Explosion
Typical	3.00E-03	1.00E-03	2.00E-04	5.00E-05	0.00E+00	8.00E-04

No	Failure cause	% of releases	Source	Release size	Suscept - ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Internal corrosion	4.3	TA exp	small	1	1	4.7E-04	
2	Internal corrosion	0.8	TA exp	medium	1	1.0E+00	9.1E-05	
3	External corrosion	5.2	TA exp	small	1	1	5.7E-04	
4	External corrosion	0.8	TA exp	large	1	1	7.8E-06	Review of acid tanks in 6 companies
5	Small bore piping	0.5	Reliability calc	small	1	1	5.5E-05	Most vessels have small bore piping
6	Process piping, flanges, valves	17.4	Reliability calc	small	1	1.0E+00	1.9E-03	All vessels have process piping
7	Drain lines left open	3.3	MHIDAS	large	0.5	1	6.4E-05	Review of acid tanks in 6 companies
8	Maintenance error	3.3	MHIDAS	medium	1	1	3.8E-04	"
9	Corrosion, no inspection	2	MHIDAS	medium	0.1	1	2.3E-03	Interviews
10	Reverse flow	3.3	MHIDAS	large	0.1	1	3.2E-04	Review of acid tanks in 6 companies
11	Overfilling	10.4	MHIDAS	large	1	0.03516	2.9E-03	"
12	External fire	3	MHIDAS	rupture	1	1	7.7E-06	"
13	Weld crack	2.7	MHIDAS	medium	1	1	3.1E-04	"
14	Earthquake, landslip, flood	0.5	MHIDAS	rupture	0.01	1	1.3E-04	"
15	Internal explosion	12	MHIDAS	rupture	0.2	1	1.5E-04	"
16	Vandalism, third party	2.7	MHIDAS	large	1	1	2.6E-05	"
17	Fire in tank	4	MHIDAS	rupture	0.2	1	5.1E-05	"
18	Wrong substances	10.7	MHIDAS	rupture	0.1	1	2.7E-04	Review of MHIDAS data
19	Rupture,weld, material	12	MHIDAS	rupture	1	1	3.1E-05	7 cases known
20	Hydrogen explosion	-	MHIDAS	explosion	1	1	8.0E-04	4 cases known
	Total	98.9						

Table 9.15 Susceptibilities and modification factors for acid tanks

9.7 Detailed analysis

A detailed analysis for tank releases is given here as a quantified safety barrier diagram. A detailed frequency analysis based on the frequency of individual component failures, is shown in table 9.16 with the individual contributions shown. A comparison is given between the overall tank failure rates, and those derived from detailed analysis.

1: Safety barrier diagram for High temperature /TANK in Generic analysis

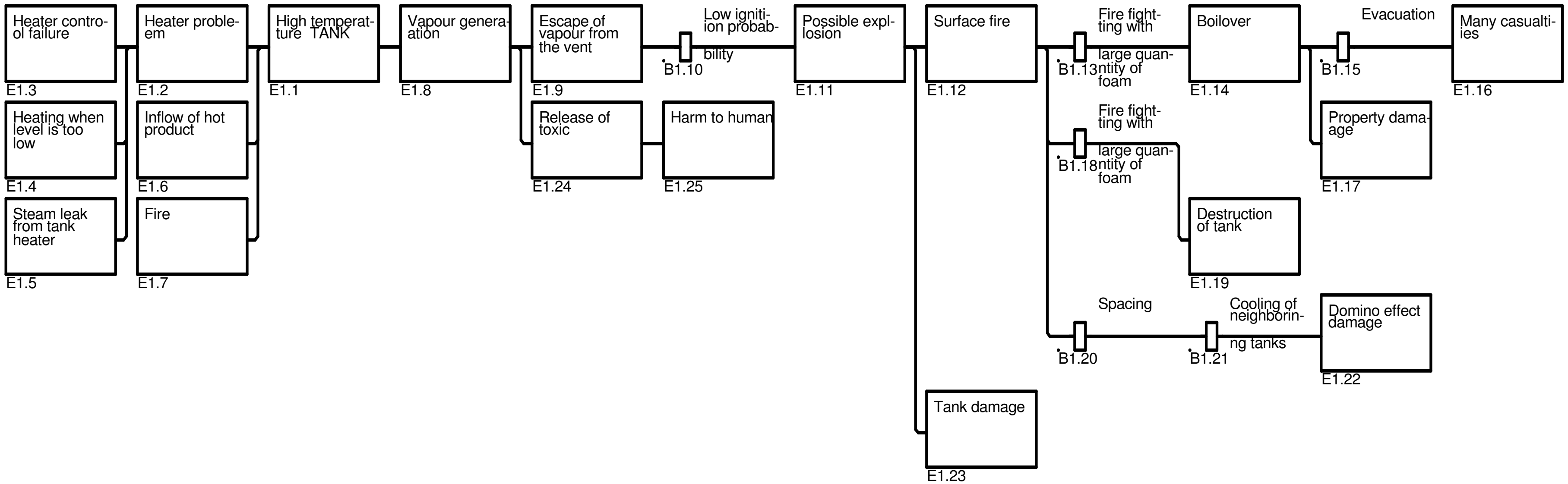


Figure 9.5

2: Safety barrier diagram for High level /TANK in Generic analysis

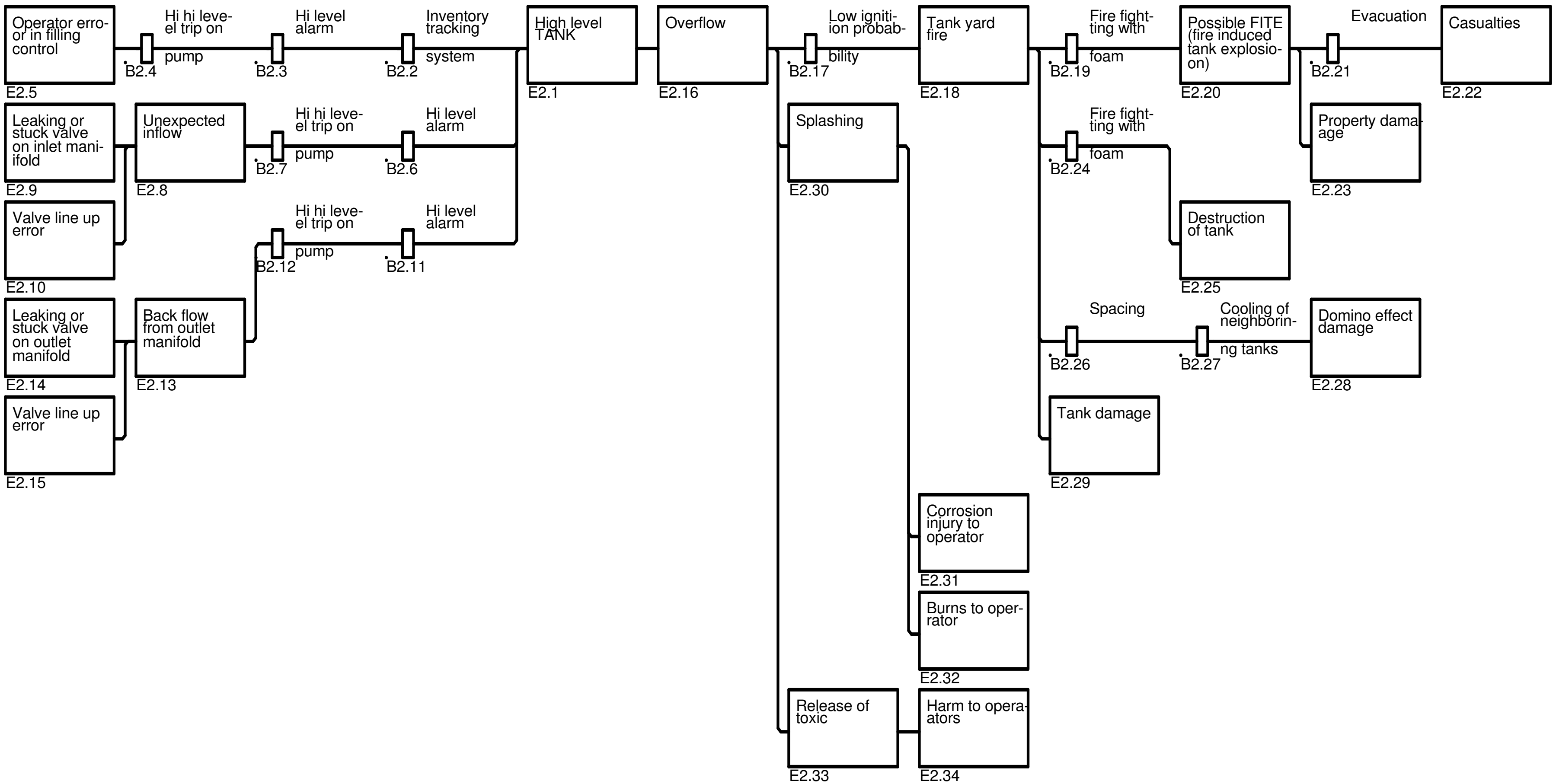


Figure 9.6

3: Safety barrier diagram for High pressure /TANK in Generic analysis

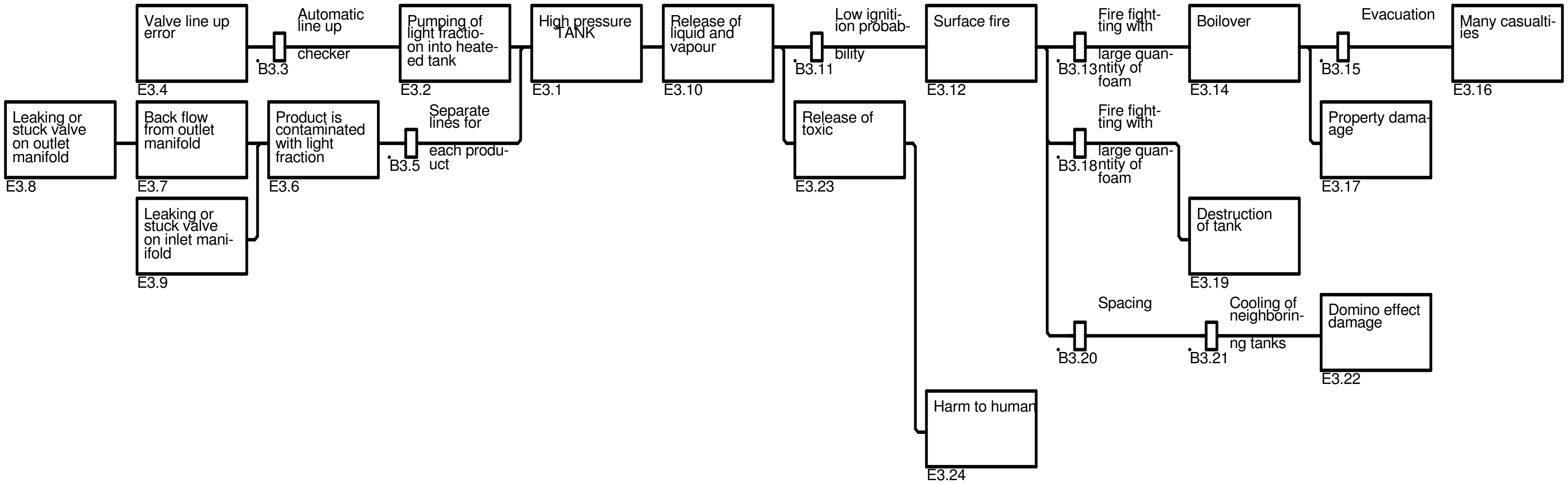


Figure 9.7

4: Safety barrier diagram for Leak /TANK in Generic analysis

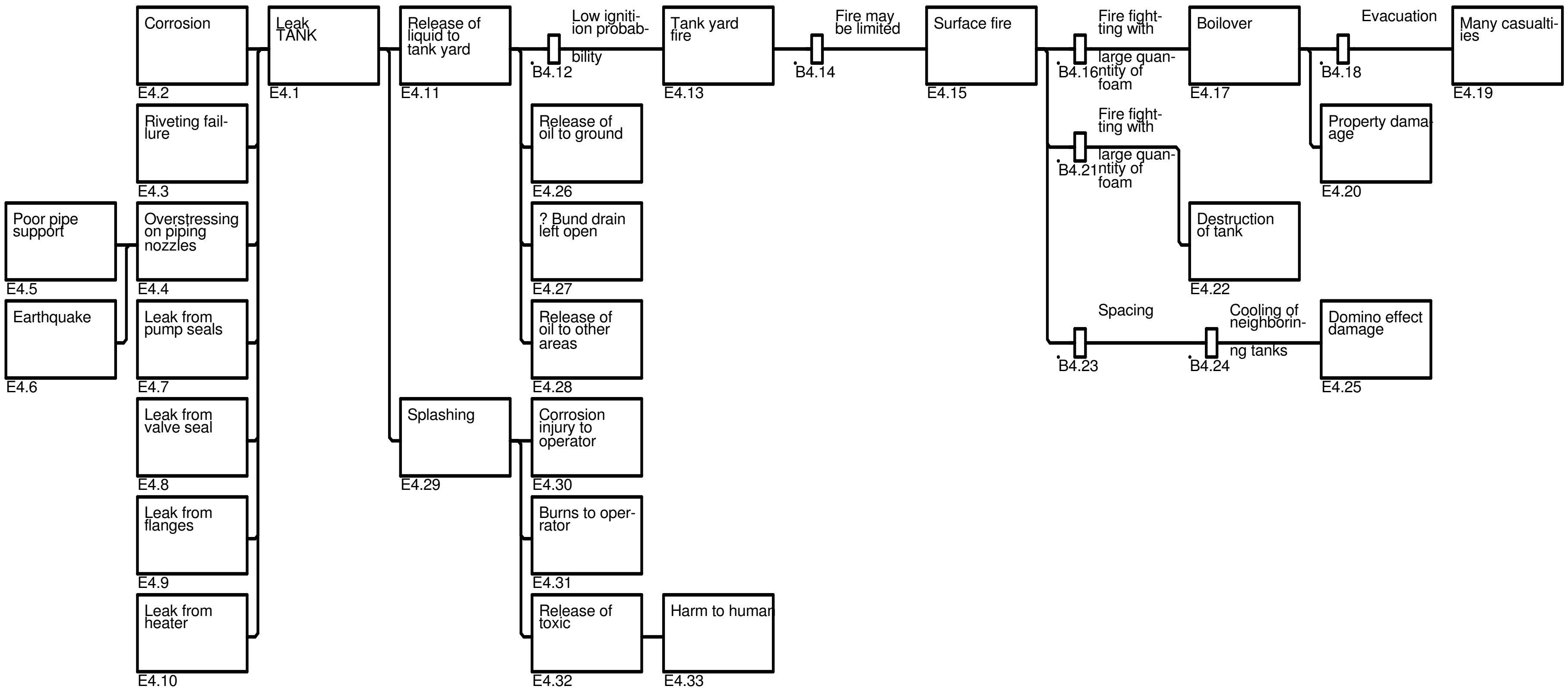


Figure 9.8

5: Safety barrier diagram for Rupture /TANK in Generic analysis

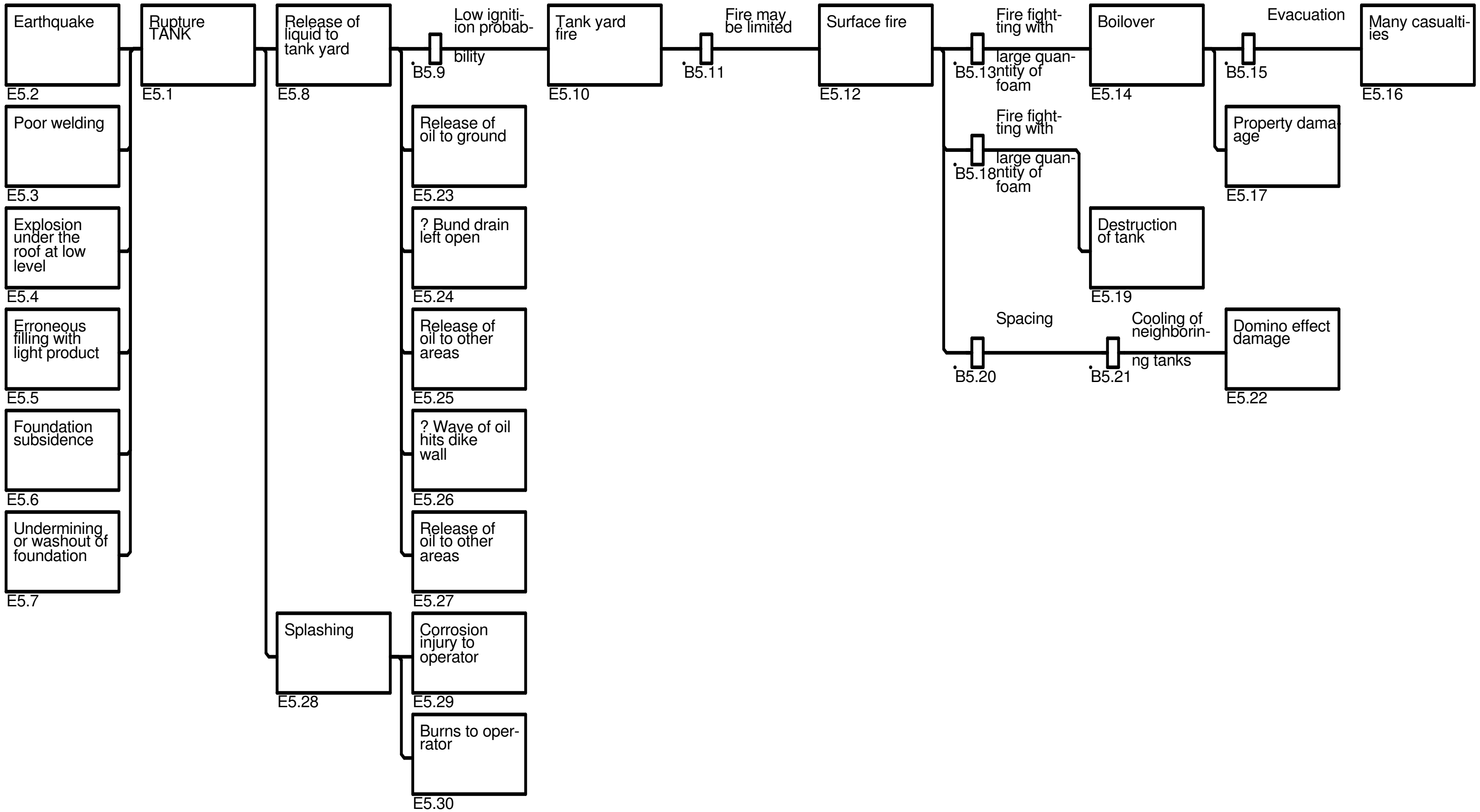


Figure 9.9

Release frequencies per year	Small	Medium	Large	Rupture	Total
Flammable liquids tank	5.00E-04	5.00E-04	5.00E-04	6.00E-06	1.51E-03

Table 9.16

Failure cause	Consequence	Number of items or metres	Frequency per item year	Susceptibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Internal corrosion	small	1	9.49E-04	1		0			0			0		9.49E-04	
Internal corrosion	large	1	2.09E-05	1		0			0			0		2.09E-05	
External corrosion	small	1	2.40E-04	1		0			0			0		2.40E-04	
External corrosion	large	1	1.98E-05	1		0			0			0		1.98E-05	
Small bore piping	small	1	6.55E-05	1		0			0			0		6.55E-05	
Process piping, flanges, valves	small	1	8.73E-02	1		0			0			0		8.73E-02	
Drain lines left open	large	1	4.40E-05	1		0			0			0		4.40E-05	
Maintenance error	medium	1	4.76E-04	1		0			0			0		4.76E-04	
Tank entry	explosion	1	1.00E-05	1		0			0			0		1.00E-05	
Corrosion, no inspection	medium	1	9.52E-05	1		0			0			0		9.52E-05	
Pipes and fittings	large	6	1.10E-04	1		0			0			0		6.59E-04	
Reverse flow	large	1	1.8E-05	1		0			0			0		1.83E-05	
Overfilling	rupture	1	1.1E-03	1	LSHH	0	0.003516		0			0		1.06E-03	
External fire	medium	1	4.8E-04	1		0			0			0		4.76E-04	
Weld crack	rupture	1	7.8E-07	1		0			0			0		7.76E-07	
Earthquake, landslip, flood *	rupture	1	6.2E-05	1					0			0		6.21E-05	
Internal explosion	large	1	1.1E-04	1		0			0			0		1.10E-04	
Vandalism, third party	rupture	1	1.6E-06	1		0			0			0		1.55E-06	
Vapour plume on windless day	explosion	1	2.0E-04	1		0			0			0		2.00E-04	
Wrong substances	rupture	1	0.0E+00	1		0			0			0		0.00E+00	
Rupture, weld, material	rupture	1	6.2E-06	1		0			0			0		6.21E-06	
Blanketing failure	large	1	0.0E+00	1		0			0			0		0.00E+00	
Crash	rupture	1	7.8E-06	1		0			0			0		7.76E-06	
Overpressuring	rupture	1	7.8E-07	1	Vent	0	0.001	PVV	0	0.05107979		0		7.76E-07	
High temperature	rupture	1	3.1E-06	1	TAH	0	0.002		0			0		3.11E-06	
Lightning	fire	1	5.0E-05	1		0			0			0		5.00E-05	
Total small														8.85E-02	
Total medium														1.05E-03	
Total large														8.72E-04	
Total rupture														1.14E-03	
Total fire														5.00E-05	
Total explosion														2.10E-04	

Note: do not multiply by number of tanks for items marked with *

Release frequencies per year	Small	Medium	Large	Rupture	Base
Chemicals tank	2.00E-03	2.00E-03	2.00E-03	6.00E-06	6.01E-03

Table 9.17

Failure cause	Conse-quence	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Internal corrosion	small	1	1.69E-03	1		0			0			0		1.69E-03	
Internal corrosion	large	1	3.41E-05	1		0			0			0		3.41E-05	
External corrosion	small	1	2.93E-04	1		0			0			0		2.93E-04	
External corrosion	large	1	1.61E-05	1		0			0			0		1.61E-05	
Small bore piping	small	1	8.00E-05	1		0			0			0		8.00E-05	
Process piping, flanges, valves	small	4	4.67E-02	1		0			0			0		1.87E-01	
Valve left open,opene by mistake	large	1	4.66E-05	1		0			0			0		4.66E-05	
Maintenance error	medium	1	4.64E-04	1		0			0			0		4.64E-04	
Tank entry		1	0.00E+00	1		0			0			0		0.00E+00	
Corrosion, no inspection	medium	1	3.6E-04	1		0			0			0		3.57E-04	
Reverse flow	large	1	6.0E-05	1		0			0			0		5.98E-05	
Overfilling	large	1	2.1E-03	1	LSHH	0	0.003516		0			0		2.12E-03	
Internal fire	rupture	1	1.4E-05	1		0			0			0		1.38E-05	
Weld crack	medium	1	3.6E-04	1		0			0			0		3.57E-04	
Earthquake, landslip, flood *	rupture	1	7.2E-05	1		0			0			0		7.25E-05	
Internal explosion	rupture	1	1.6E-04	1		0			0			0		1.63E-04	
Vandalism, third party	large	1	3.4E-05	1		0			0			0		3.41E-05	
External fire	rupture	1	6.9E-05	1		0			0			0		6.88E-05	
Wrong substances	rupture	1	3.7E-04	1		0			0			0		3.70E-04	
Rupture weld material	rupture	1	3.7E-05	1		0			0			0		3.70E-05	
Blanketing failure	rupture	1	2.2E-05	0		0			0			0		0.00E+00	
Crash	large	1	2.3E-04	0		0			0			0		0.00E+00	
Overpressuring	rupture	1	2.3E-04	0	Vent	0	0.001	PVV	0	0.05107979		0		0.00E+00	
High temperature	rupture	1	1.0E-03	0		0			0			0		0.00E+00	
Lightning	rupture	1	4.2E-05	1		0			0			0		4.20E-05	
Runaway reaction	explosion	1	5.0E-04	1	TAH	0	0.002		0			0		5.00E-04	
Total small														1.89E-01	
Total medium														1.18E-03	
Total large														2.31E-03	
Total rupture														7.67E-04	
Total fire														0.00E+00	
Total explosion														5.00E-04	

Note: do not multiply by number of tanks for items marked with *

Table 9.18 Detailed release frequency calculation

Release frequencies per year	Small	Medium	Large	Rupture	Base
Fixed roof tank, acid	2.00E-03	2.00E-03	2.00E-03	6.00E-06	6.01E-03

Failure cause	Conse-quence	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Internal corrosion	small	1	4.71E-04	1		0			0			0		4.71E-04	
Internal corrosion	medium	1	9.09E-05	1		0			0			0		9.09E-05	
External corrosion	small	1	5.69E-04	1		0			0			0		5.69E-04	
External corrosion	large	1	7.80E-06	1		0			0			0		7.80E-06	
Small bore piping	small	15	5.47E-05	1		0			0			0		8.21E-04	
Process piping, flanges, valves	small	4	1.91E-03	1		0			0			0		7.62E-03	
Drain lines left open	large	1	6.44E-05	1		0			0			0		6.44E-05	
Maintenance error	medium	1	3.75E-04	1		0			0			0		3.75E-04	
Corrosion, no inspection	medium	1	2.27E-03	1		0			0			0		2.27E-03	
Reverse flow	large	1	3.2E-04	1		0			0			0		3.22E-04	
Overfilling	large	1	2.9E-03	1		0			0			0		2.89E-03	
External fire	rupture	1	7.7E-06	1		0			0			0		7.69E-06	
Weld crack	medium	1	3.1E-04	1		0			0			0		3.07E-04	
Earthquake, landslip, flood*	rupture	1	1.3E-04	1	LSHH	0	0.003516		0			0		1.28E-04	
Internal explosion	rupture	1	1.5E-04	1		0			0			0		1.54E-04	
Vandalism, third party	large	1	2.6E-05	1		0			0			0		2.63E-05	
Fire in tank	rupture	1	5.1E-05	1		0			0			0		5.13E-05	
Wrong substances	rupture	1	2.7E-04	1		0			0			0		2.74E-04	
Rupture,weld, material	rupture	1	3.1E-05	1		0			0			0		3.08E-05	
Hydrogen explosion	explosion	1	8.0E-04	1		0			0			0		8.00E-04	
Total small														9.48E-03	
Total medium														3.05E-03	
Total large														3.31E-03	
Total rupture														6.46E-04	
Total fire														0.00E+00	
Total explosion														8.00E-04	

Note: Fire and explosion values are only relevant if the acid can become contaminated, hydrogen can be generated, runaway reactions can occur, or the wrong substance can be added. **Note: do not multiply by number of tanks for items marked with ***

9.8 Algorithm for fixed roof tank fire and release frequencies

In the following sections, a range of questions are posed, and the variation in release frequency associated with the answers is given.

Application

#	Question	Action if Yes	Action if No
1	Is the material stored flammable	Fire susceptibility 1 Go to 2	go to 3
2	Is the material q chemical	Fire susceptibility 1 Corrosion susceptibility 1	go to 3
3	Is the material an acid	Corrosion susceptibility 1 Reaction susceptibility 1	go to 4
4	Is the material a strong alkali	Corrosion susceptibility 1	go to 5
6	Is the material sodium hypochlorite	Corrosion susceptibility 1 Reaction susceptibility 1	Exit

Table 9.18 Baseline frequency modifications for piping application

Operation

#	Question	Action if Yes	Action if No
1	Is there a log of tank inventory kept continuously, so that the operator always knows the tank content	Go to 2	go to 3
2	Is there a frequent gauging of the tank contents?	Include inventory control as barrier in detailed analysis. Go to 3	go to 3
3	Is there a tank level indicator with a remote reading?	Include level monitoring as barrier in detailed analysis. Go to 4	go to 4
4	Is there a second independent level indicator for safety purposes?	Go to 6	go to 7
5	Is there an independent level switch for safety purposes?	Include switch type level trip as barrier in detailed analysis. Go to 7	go to 7
6	Does the safety instrument, if it exists automatically terminate tank filling? i.e. a level switch.	Include level indicator trip as barrier in detailed analysis. go to 7	go to 7
7	Does the operator always attend the tank during the entire period of tank filling?	Go to 8	Set overfilling susceptibility to 1 and use additional 0.1 per year in detailed analysis go to 8
8	Do the operators drain water from the tank periodically?	Increase drain lines left open percentage by 4*frequency of draining Go to 9	Exit
9	Is there a dual valve on the tank drain?	Include dual valve as barrier in detailed analysis	Exit

Table 9.19 Baseline frequency modifications for operational problems

Flammability

#	Question	Action if Yes	Action if No
1	Is the liquid flammable?	Use table 9.17 for calculation Go to 2	Use table 9.16 for calculation Exit
2	Is the vapour pressure such that the vapour is in the flammable range?	Set susceptibility for internal explosion to 1 Go to 3	Exit
3	Is the tank blanketed with nitrogen, flue gas, or fuel gas?	Include blanketing as safety barrier in the detailed analysis. Go to 4	Go to 4
4	Is the tank sampled or gauged from the top?	Increase internal explosion frequency by factor 10	Go to 5
5	Is splash filling used?	Go to 6	Exit
6	Is the resistivity of the liquid high ?	Increase internal explosion frequency by factor 20	Exit

Table 9.20 Baseline frequency modifications for blanketing

Acid, alkalis etc

#	Question	Action if Yes	Action if No
1	Does the tank contain hydrochloric or acetic acid?	Go to 2	Go to 4
2	Is the tank made of carbon steel?	Go to 3	Go to 4
3	Is the tank well ventilated?	Set susceptibility for internal explosion to 1 for hydrogen explosion. Exit	Set susceptibility for internal explosion to 1 for hydrogen explosion. Increase explosion percentage by factor 5
4	Is sodium hypochlorite used on the site?	Set susceptibility for reaction to 1 in detailed analysis	Go to 5
5	Is the material nitric acid?	Go to 6	Go to 8
6	Are the tanks aluminium ?	Go to 7	Exit
7	Is sodium hydroxide or sodium hypochlorite stored on site	Set susceptibility for internal explosion to 1 for hydrogen explosion. Exit	Exit
8	Is the material sodium hypochlorite	Go to 9	Exit
9	Are acids stored on site	Set susceptibility for internal explosion to 1 for hydrogen explosion. and to 1 for chlorine release Exit	Exit

Table 9.21 Baseline frequency modifications for acid tanks

System

#	Question	Action if Yes	Action if No
1	Is the inlet or outlet to the tank manifolded?	Set susceptibility for reverse flow release to 1 in detailed analysis Go to 2	Go to 4
2	Are light substances handled in the manifold (pentane, butane, hexane, ethyl acetate, solvent spirits, gasoline etc.)?	Set susceptibility to wrong substance explosion to 1 Exit	Go to 4
3	Are hot (above one of the stored component boiling point) substance handled in the manifold system?	Set susceptibility to wrong substance explosion to 1 Exit	Go to 4
4	Is it possible for reactive materials to flow together ?*	Set susceptibility to wrong substance explosion to 1 Exit	Exit

Table 9.22 Baseline frequency modifications for system design * See table 9.8.6

Typical reactive pairs for tankage:

	Acid	Hypochlorite	Sodium hydroxide	Cyanide solution	Hydrogen peroxide	Ammonia liquor
Acid	*	*	*	*	*	*
Nitric acid	*	*	*	*	*	*
Hypochlorite	*			*	*	*
Organics	*	*	Maybe	*	*	*
Sodium hydroxide	*				*	
Sodium cyanide solution	*	*			*	*
Carbonate solution	*					
Hydrogen peroxide	*	*	*	*		*
Ammonia liquor	*	*		*	*	
epichlorhydrin	*	*	*	*	*	*

Table 9.23 Examples of reactive pairs

9.9 References

1. Taylor, J.R. Process Safety Engineering, Designing and Building Safer Process Plant, Taylor Associates, 4th Edition 2001

2. Japanese data, reference Loss Prevention and Safety Promotion

3. HSE, Offshore Release Statistics 2001

4. Christensen, R.A. and Eibert, R.F., Above Ground Tank Storage, EL RM-623, Entropy Limited, MA, 1989

5. E&P Forum, Quantitative Risk Assessment, E& P Forum Report No. 11.8/250, 1996

6. Chemical Safety Board, Refinery Incident, Motiva Enterprises, October 2002

7. Lauben, R.W. and Robinson, D.L. Acoustic Emission Integrity of Above Ground Storage Tanks, PWR Vol 5, Proceedings of Industrial Power Conf, ASME, 1989

8. Cox, A.W. Lees, F.P. and Ang, M.L., Classification of Hazardous locations, I Chem E, 1993

9. OPITSC, Atmospheric Storage Study, OPITSC, Singapore, 1990

10 Floating roof tank releases

Figure 10.1 shows the design of a typical floating roof tank. The main features are the foundation, the tank base, tank walls and floating roof.

Figure 10.1 Floating roof tank

The foundation is a built up circle of stone, covered with a layer of gravel, all consolidated by rolling / vibrating. In some cases a concrete ring foundations is built up to contain the stone bed. The foundation is covered with a layer of clean sand or gravel to transfer the load evenly to the foundation bed.

The materials for the tank base, walls, and floating roof are:

All welds are made to API 650.

The floating roof is either built up of a double layer of sheet steel, with walls about 0.5 m high, divided into sections, or is a single sheet, supported by pontoon which are simply closed vessels containing air.

The floating roof is depressed towards the centre and provided with drainage for rainwater.

The floating roof is provided with legs which support it when the tank is emptied. The support level must be above the mixer and heater coil height, and must be high enough to allow the tank to be cleaned. The floating roof is provided with a seal system to prevent release of vapour a typical seal consists of lower and upper fixed scrapers which retain a flexible reinforced plastic of rubber seal tube, either air or poly urethane foam filled. A hinged flak above the seal tube reduces release of vapour which has passed the seal.

In some cases, the use of a flexible tube seal is prevented due to the corrosive nature of the liquid. In this case, other arrangements such as adjustable steel shoes may be used.

Fittings for the tank are:

- drainage from the bottom of the tank for removal of water.

- a motor driven propeller mixer.

- in the case of heavy oils, a steam heater.

- float level gauge

- dipstick opening for level gauging.

- ladders inside and outside for access. The inside ladder has rollers to allow it to glide on the floating roof.

- inlet and outlet nozzles, with valves and pumping arrangement.

- high temperature, low and high level alarms.

- fire protection in the form of scum distribution, either at the base of the tank, or with distributors to the tank rim.

The tank is surrounded by a bund or dike, which can take at least 75% of the capacity for the tank, and 110% of the volume of the largest tank. If there are two interconnected tanks, the bunded area has at least 75% of the combined capacity .

Figure 10.1.2 Floating roof tanks. The tank at the rear has a sheet plastic dome to reduce vapour loss.

10.1 Operation

Requirements for operation are intended to prevent overfilling and to ensure continued integrity of vessels.

A register is kept of tank filling and draw down flow/quantity, to provide a cross check on the level measurement (depending on activity). Daily or weekly cross checks using a dip stick support the inventory measurement.

Integrity checking is typically:

Each year	-	Visual check.
4 years - 15 years	-	Ultrasonic inspection.
5 years - 15 years	-	Internal cleaning and inspection

10.2 Accident types (ref. 10.1)

Leakage: The most frequent accident types for floating roof tanks are leakage's arising from corrosion, weld defects, or corrosion at weld defects on old tanks. These are by far the most common at the bottom of the tank and at the bottom/wall weld, since water tends to collect here. Corrosion is more severe when the crude which is stored is sour.

The effect of such leaks is generally ground pollution. In some older tank farms, this can be so severe that considerable quantities of oil can be recovered by pumping. Oil can also pass into drainage.

Other sources of leakage are:

- via the roof drain (due to breakage or corrosion). The leakage should then be to the oily water drain.

- via a leak in the floating roof, which passes oil to the roof drain. The leakage should be to the oily water drain.

- via the filling or draw down pipes. In addition to corrosion causes, these can crack due to expansion, or to foundation settling, if the piping is not provided with adequate freedom of movement.

- due to overfilling of the tank. This generally requires a mistake on the part of the operator, accompanied by not seeing, or ignoring a high level alarm, or due to alarm failure.

- due to overfilling as a result of back flow from a high level tank to a low level tank. This can occur if a valve is not closed properly or is not closed at all, so that flow occurs through a transfer pump or manifold.

- overfilling can occur, due to steam or water entry into a full tank. Water can most readily enter as rain water.

- breakage of filling and draw down lines can occur due to severe settlement, foundation collapse, or earth quake. It is common to provide a flexible joint (bellows) and double valves (inboard and outboard of the double joint) when this is likely.

Tank rupture

A few cases of tank rupture are known, one due to improperly specified weld, the other due to failure of a water pipe passing beneath the tank. This failed and washed out the foundation.

Floating roof accidents

Floating roof accidents an occur due to:

- leakage of pontoons, which causes either sinking, or tilting of the roof. A tilted roof can jam, allowing air to enter (explosion risk) and can fall. In falling the roof can generate ignition sparks and break equipment such as instruments which in turn cause.

- mechanical jamming due to inadequate flexibility in the seal, due to rusting of the tank wall, due to poor weld repairs which leave flashes on the wall, and due to failure of centring of floating roof.

The effects are the same as those described for tank roof sinking.

- floating roofs can sink if not drained, so that rain water collects. Causes of blockage of the roof drain are typically valve closure (e.g. to prevent to oil leakage), leaves, and dead animals and birds.

Tank fires

Fires can occur at the tank rim due to the presence of vapour from the tank wall. Such vapour is always present during emptying. Vapour is also present if the tank roof sinks.

In calm weather, vapour can collect above the tank roof, being heavier than air. Entry to a tank roof should always involve an explosimeter reading, if the wind is still.

Ignition sources are:

- sparks from a falling tank roof.

- iron sulphide particles which are pyrophoric. The iron sulphide is formed by crudes with a high sulphur content.

- static changes, especially between tank roof and wall. These should be linked by an earthing cable, but the cable sometimes fails.

- Faults in instruments cables.

- persons entering the tank top e.g. to make a dipstick reading, sampling or for inspection.

- lightning. In some refineries, tank rim fires are a yearly occurrence due to this cause.

Rim fires can usually be prevented by pumping foam to the rim. In some cases, carbon dioxide is piped to the tank rim.

If the rim fire is not put out fast enough the fire can damage the floating roof, so that it sinks. Another cause of sinking is fire fighters who mistakenly try to put out the rim fire with water. When the roof sinks, the fire develops to a whole surface fire. Similarly, a whole surface fire will develop if the fire is started by the roof sinking or collapsing.

Whole surface fires can be extremely difficult to fight. They require the application of large quantities of foam very rapidly. If the tank fire cannot be put out, it is necessary to let the fire burn out. It is possible for the fire to cause slop over or boil over if the fire continues for a long time. Boil over arises when the heat from the fire surface reaches the tank base and causes water to boil. The boiling causes mixing of hot upper layers and colder lower layers, which in turn causes flocking of the low boiling fractions of the crude. The result can be a fire ball of a size measured in kilotons.

The time available before boil over occurs is generally at least 1 or 2 hours, and typically 8 hours to several days, so that there is ample time for emergency measures. Typical measures are:

- draining water from the tank base.

- pumping oil to another tank.

- cooling neighbouring tank walls.

- evacuating persons to a safe distance (0.5 - 1 km.) depending on tank size (Taylor, 1994 gives a review of boilover sizes)

Figure 10.2 Floating roof tank boilover

Tank explosions

Tank explosions can occur on emptying, due to ingress of air. Ignition can occur from inspection access, instrument or equipment failure, pyrophoric iron sulphide etc.

10.3 Case stories (ref 1)

1. Explosion in empty tank

A floating roof tank was emptied, with so much oil pumped out that the tank roof settled on the bottom support pillars. Air then entered the tank. Ignition occurred either from a friction spark, ad the roof settled on the bottom support pillars. Air then entered the tank. Ignition occurred either from a friction spark, as the roof settled, or from an electrostatic spark. The later seems likely because the roof earth cable was broken.

An explosion occurred but did not destroy the tank. The roof was destroyed, though, by the ensuring fire. The tank shell survived with little damage.

Note: API recommends that the roof should be left floating at all times. When tank internal maintenance is to be made, special care needed to ensure that no ignition source exists, and that the tank is free moving for the full range of travel.

2. Tank leak (API)

190 000 bbl of crude were transferred to a 200 000 bbl tank. The operator found the level falling. Checks showed a large leak from under the tank floor. After 5 hours there was 1" of oil in the dike.

Pump out was begun. After probing from above, the leak position was found. Sand bags were dropped and successfully slowed the leak.

After 4 days, clean up was complete. At inspection, no significant corrosion was found. The cause was considered to be a faulty foundation. During construction at 1/4 steel plate was dropped while there was 3" of water in the tank. The water flow disturbed the foundation leading to subsequent overstress of the tank bottom.

3. Roof drain blockage

The floating roof drain became blocked with rags and an old glove. A heavy rainstorm caused the floating roof to flood and sink. Friction at the floating roof seal caused ignition and a full surface fire.

4. Out of round tank

An out of round tank wall caused a floating roof to jam when it turned. (Turning was assumed to arise from uneven wind loads on pontoon manholes). On emptying the floating roof was suspended for a time, but then fell. Friction caused ignition, leading to a flash fire and full surface fire.

5. Lightning

In the course of one summer lightning caused 20 rim fires in a tank yard. All were extinguished quickly and efficiently with foam. Subsequently, lightning conductor towers were installed.

6. Wrong product

A floating roof was designed for cracked naphtha with a gravity of 370 API but service was changed to aviation gasoline, gravity 700 API. The result was that the roof floated 5/8 inch lower than the design level, exerting an upward force on the diaphragm. The diaphragm (the steel plat which floats on the pontoons) tore, and the roof sunk 6” on side.

7. Lightning

A tank was struck by a lightning bolt. The resulting fire destroyed the tank. A recent inspection had revealed vapour leakage from vents in the seal fabric and from openings between the shoes and the tank shell. Some shoes were locked due to corrosion. A plant wide programme of installing grounding shunts between roof and tank shell was in progress.

Foam was supplied through four 4” goose neck outlets at the top of the tank. The stock was unstabilised light naphtha 67.3 API, RVP 10.1 psi.

The fire burned for 1½ hours and began to subside, then flared up. It is believed that the roof sank at this point. The heat became so intense that fire fighters had to retreat. The tank roof may have been torn by the lightning strike.

Figure 10.4 A floating roof tank fire

Figure 10.5 Domino effects from floating roof tank fire

8. Multiple lightning fires

Three tank fires were caused almost simultaneously by a lightning strike, apparently by induced electrical charges.

9. Lightning strike not at tanks

Six tanks were ignited when lightning struck a radio tower outside the tank area. Each of the tanks had a steel cable between floating roof and tank shell. Some tanks are now being fitted with multiple earth bonding.

10. Hurricane

In a hurricane, wind sucked crude spray from full tanks.

11. Swing line jams

The drain line from the tank roof was a swing line supported under the floating roof. In two instances, a chain supporting the swing line corroded and fell off. In this case the swing line stayed in the vertical position because it was dead centre. The floating roof hung up in a canted position.

12. Vapour under the roof

Steel pan floating roofs in covered tanks sank when slugs of vapour entered the flow due to filling with unstabilised crude, or lines were blown with air.

Figure 10.6 Splashing around a tank roof vent, due to filling with unstabilised crude.

13. Wax build up on roof

High wax crude was stored in a 240 foot diameter floating roof tank. Two 50Hp mixers and a small coil were fitted. The roof sank during filling due to uneven floating resulting from wax build up.

14. Lightning

Lightning caused a fire in a 244 ft dia tank and power shortage occurred. Cooling water was applied to the shell. Foam was connected to subsurface injection nozzles. Connection took longer than expected, about two hours. Final extinguishing required lowering a hose to the seal. Final extinguishing occurred after about 4 hours.

The tank was out of operation for several months:

- the seal and many shunts were blown off by the initial explosion
- shoes were damaged and loosened from hangers.
- some of the roof pontoon plates buckled
- the paint was scorched
- the automatic gage tape was broken

Lessons learned were:

- fires at seals which have weather shields or wax scrapers cannot be fought effectively
- injected foam may channel and pocket through viscous stock
- traffic control was inadequate too many vehicles led to congestion
- a longer supply of foam was needed
- eight bolt blind flanges on foam injection lines required too much time to remove.

15. Roof sank

A floating roof on the tank sank, and the space between floating roof and cone roof cover filled with product. The product was covered with foam, and the refiner believes that ignition occurred as a result of static from an unbonded foam line.

16. Lining ignites

A welding spark caused a freshly laid fibreglass/epoxy lining to ignite.

17. Catastrophic failure

A tank failed by catastrophic brittle failure. The steel was ASTMA 10-39 with 0.3% Mn. The tank was 66 000 bbl capacity with 63 000 bbl content. The fracture originated at a repair weld near the bottom of the tank and spread to the bottom and top of the shell in a straight line through the plates not the welds. The shell pulled away from the bottom as it split.

Oil was carried 400 to 600 ft.

On investigation, it was found that the vertical weld in a patch ring did not have complete penetration, and suffered from excessive stress.

Ambient temperature was -3^0F, and wind velocity 15mph over 30 hr. The steel had a charpy impact strength of 2½ ft lb of 0^0F, compared with a design criterion of 15 ft lb at operating temperature. Transition temperature was as high as 70^0C.

10.4 Frequency of floating roof tank releases

Large storage tanks have many and varied designs of which the principal ones are cone roofed tanks and floating roofed tanks. Corrosion is a possibility here, and leads to a need for periodic inspection. Failure rates depend very much on the material stored. Apart from various valves, the most likely causes of leaks are either overfilling (this can cause tank roofs to raise) and corrosion attacks around welds.

Catastrophic failures generally arise from overfilling, as a result of operator error, administrative error, or instrument failure, and roof fires.

RMP data for tank releases gives, for crude units, a value of 0.0024 releases per tank year. The amount of data available from the RMP data for tanks is very low. This is not surprising – tank releases occur at close to atmospheric pressure and so release rates are small. Also, the materials stored in such tanks generally have a low vapour pressure. For these reasons, only relatively few release accidents will have offsite consequences, when compared, for example to pressure vessels.

Christensen and Eibert (ref. 10.2)carried out an extensive survey of atmospheric storage tank releases from oil production and refining in the USA. The number of tanks is given in table 10.1 A large proportion of these will be floating roof tanks.

Oil industry segment	Surveyed tanks	Estimate of total population	Average age
Marketing	5831	88529	29.4
Refining	11440	29727	34.6
Transportation	5341	9197	31.4
Production	54046	572620	15.1
Total	76708	700073	17.9

Table 10.1 Tanks in the Christensen and Eibert survey.

The release frequency determined for these was $1.5*10^{-2}$ per year. The size of the releases is not given, but E&P Forum (ref. 10.3) used data from Lauben and Robinson (ref. 10.4) to calculate a value for major tank release frequency of $6.9*10^{-6}$ per year, based on 92 major tank failures.

The Lauben and Robinson study describes the results of integrity testing carried out by the Hartford Boiler company. The paper describing the study gives about 16000 leaks, and about 92 major release incidents. Their study gives a release frequency of $2.5*10^{-2}$ per year based on a sample of tank inspections of 835 tanks.

A collection of 206 oil and distillates tanks with spill data from a period of 41 years, studied by the author, gave a leak frequency $6*10^{-3}$ per year, with a size distribution as shown in figure 10.7.

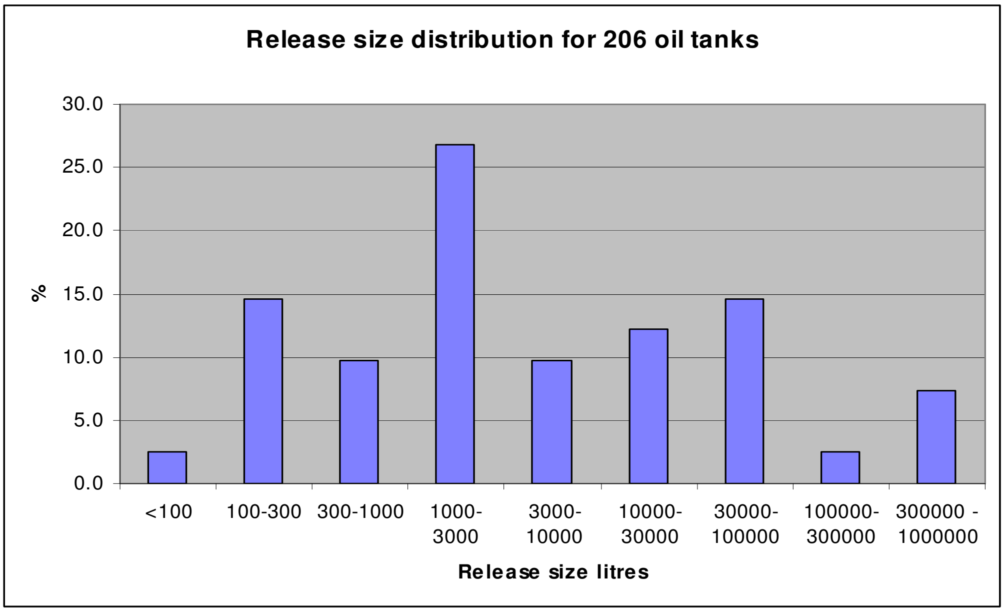


Figure 10.7 Release size distributions for 206 oil tanks

As a basis for comparisons later, when selecting a base frequency, the release frequencies for oil tanks are compared in table 10.2

Source	Small leaks up to 25 mm	Medium leaks up to 50 mm	Large leaks	Rupture, catastrophic
Christensen and Eibert, E&P Forum	$15*10^{-3}$ all releases			
Lauben and Robinson, HSB	$25*10^{-3}$ all releases			$6.9*10^{-6}$
Company G	$0.9*10^{-3}$	$3*10^{-3}$	$2.1*10^{-3}$	$0.047*10^{-3}$

Table 10.2 Frequencies of releases from tanks, data from different sources

Fires on tanks are a separate issue from releases. They may arise as the result of a tank release, or may arise due to internal causes.

API give regular data for refinery fires and explosions. Cox, Lees and Ang (ref.10.5) quote this data, with 201, 173, 142 and 104 fires in 1982 to 1985 respectively. The distribution of fires in 1985 were:

Size of loss $1000	No of fires	Proportion %	Frequency per refinery year
2.5 - 100	65	60	0.58
100 - 1000	37	34	0.33
> 1000	7	6	0.058

Table 10.4 Frequency of fires, API data quoted by Cox, Ang and Lees (ref. 10.5)

The larger fires here will tend to be those in floating roof tanks, partly because these are more prone to fire due to lightning, and in part because most of the large tanks in a refinery are floating roof types. The average frequency of large fires per refinery was estimated to be 0.28 large fires per refinery year (bases on 225 refineries). Very few of these are large tank farm fires, though rim fires are fairly frequent at some locations due to lightning. The proportion of refinery fires which occur at storage tanks is typically about 2% (API). In any case, a full surface fire would give above $1 million on costs. Tank top fire frequencies, are therefore estimated to occur with a frequency of less than 0.0012 per year. For rim fires, the author's own observations of rim fire frequency have ranged from 12 per year per refinery due to lightning in one very lightning prone plant, through 1 per year for a refinery, to less than 10^{-4} per year per tank.

Data quoted by E&P Forum give the fire frequencies for floating roof tanks as in table 10.5.

Country	**Data source**	**# fires**	**Tank years**	**Fire freq. per tank year**
Netherlands	Saval-Kronenburg	1	673	$1.5*10^{-3}$
USA	Company name confidential	10	3883	$2.6*10^{-3}$
Scotland	N Sea Oil Terminals	1	461	$2.2*10^{-3}$
Total		12	5017	$2.4*10^{-3}$
	Full surface fires			
Netherlands	Saval-Kronenburg	0		
USA	Company name confidential	1	3883	$2.6*10^{-4}$
Scotland	N Sea Oil Terminals	0		

Table 10.5 Floating roof tank fire frequency (ref. 10.3)

The LASTFIRE project, carried out by Resource Protection International for a large consortium of oil companies, gives data from a large number of observations of floating roof oil tank fires. The number of tank years covered was 33909. The data from www.resprotint.co.uk is reproduced here

Type of Incident	**Frequency pr 10^5 tank-year**
Spill onto roof	160
Sunken roof	110
Spill into bund	280
Rim seal fire	160
Spill on roof fire	3
Small bund fire (mixers, pipes, valves or flanges)	9
Large bund fire (major spillage)	6
Full surface fire following sunken roof	3

Table 10.6 Floating roof failure rate data from the LASTFIRE project

Boilover is the most extreme effect resulting from crude oil and similar tank fires. It requires that there is a liquid such as crude oil, with a wide range of boiling points, that there is a small amount of water in the tank, and that the fire continues for several hours.

Boilover is more or less inevitable if a full surface fire occurs in a tank such as a crude oil tank, and the fire cannot be extinguished. Historically, about one in two such fires have developed into BLEVE's. However, if insufficient supplies of foam are available, boilover is virtually inevitable for crude oil tanks. The LASTFIRE project recorded one boilover, giving a probability of 1 in 6, (6 full surface fires). However, the source of data for the LASTFIRE statistics is primarily from large oil companies, which tend to have good fire fighting resources. In all 26 large boilovers could be identified by the author from the literature and private communications, over a period of 40 years.

10.5 Typical frequencies for floating roof releases and fires

The basis chosen for the typical values for floating roof spills and fires is that given in table 10.4.3 and 10.4.4. These sets of data are reasonably compatible, except for the full surface fire data, which differ by an order of magnitude. The data in table 10.4.3, however, is based on just one incident. Note that in any case, such data are very dependent on the quality of fire fighting systems at the plant, so that it should be expected that there will be a wide range of data. The LASTFIRE data in table 10.4.4 is assumed to represent a good standard of safety practice. The preferred typical data is then that given in table 10.7.

Type of Incident	**Frequency pr 10^5 tank-year**
Spill into bund	280
Spill on roof fire	3
Small bund fire (mixers, pipes, valves or flanges)	9
Large bund fire (major spillage)	6
Full surface fire following sunken roof	3
Boilover	1

Table 10.7 Typical line release and fire frequencies for floating roof tanks

The boil over frequency can be compared with those given for refinery fires, in ref 10.5 for 255 refineries in W. Europe and USA, and extended by the author with a survey of S. American refineries and US oil storage terminals, giving a frequency of about $2.3*10^{-3}$ per refinery/terminal year, or about $2*10^{-4}$ per tank year (assuming 10 floating roof tanks per refinery or terminal, on average. Some refineries have as few as two floating roof tanks, some terminals have many tens of them). This observed value gives a frequency which is about a factor of 20 higher than the LASTFIRE value. This presumably reflects the difference in the scope of the data collection. The data collected in the LASTFIRE project was from large oil companies, all with well developed safety programmes. The data collected by the author was from a wide range of installation types, some of them with very poor safety arrangements, and in particular, with only minimal or no facilities for extinguishing full surface fires.

Typical values are selected as the average of observed values here. The values used are given in table 10.7a

	Small leaks up to 10 mm	Medium leaks up to 25 mm	Large leaks	Rupture, catas-trophic	Fire, tank top	Fire, tank basin	Boil-over
Typical	2.8E-3	1E-3	0.2E-3	5E-5	3E-5	6E-5	2E-4

Table 10.7a Typical release frequencies per tank year

10.6 Tank failure causes

Tables 10.8 shows a break down of tank accident causes drawn from MHIDAS data base. Table 10.9 shows the frequency of fire and explosion ignition sources.

Accident cause	**Number**	**%**
Unignited releases		
Hose failure	1	3.8
Leak cause unrecorded	1	3.8
Overflow	1	3.8
Pipe rupture	4	15.4
Pump house fire	1	3.8
Tank rupture	2	7.7
Rupture	2	7.7
Sabotage	5	19.2
Spill cause unrecorded	3	11.5
Valve left open	3	11.5
Vandalism	3	11.5
Total	26	100.0

Table 10.8 Causes of crude oil tank releases, MHIDAS data base

Accident type	**Ignition source**	**Number**	**Number**	**%**
Boil over	Contamination	1		
Boil over	Flare	1		
			2	7.1
Broken valve			1	3.6
Damage due to theft			1	3.6
Earthquake			1	3.6
Explosion	Hot work	2		
Explosion	Power line	1		
Explosion			6	21.4
External fire			1	3.6
Fire	Cigarette	1		3.6
Fire	Hot work	1		3.6
Fire	Lightning	20		71.4
Fire	Roof damage	1		3.6
Fire, ignition source unknown		5		17.9
Fire, total			28	100.0

Table 10.9 Causes of crude oil tank fires, MHIDAS data base

10.7 Assessment of causal factors and susceptibilities

The technology for floating roof tanks is fairly standard around the world, the major differences being in the standard of maintenance. Since management factors are not taken into account here, the range of cause factors is very limited. There are large differences in the degree of protection of floating roof tanks. For many very large tanks it is impossible to fight a full surface fire. For others, it is possible in principle, but there may be insufficient foam or foam projection capacity.

One factor which affects the susceptibility to accidents at a site is that of lightning – frequency of lightning strikes can vary by three orders of magnitude, depending on location. This variation can occur over a distance of as little as a few kilometres. Even though those areas where lightning strikes are very frequent tend also to have better protection, the difference in frequency is reflected in a difference in probability.

Table 10.10 Assessment of floating roof accident susceptibilities

10.8 Detailed analysis

A detailed analysis for tank is given here as a set of quantified safety barrier diagrams. A detailed frequency analysis based on the frequency of individual component failures, is shown in table 10.11 with the individual contributions shown. A comparison is given between the overall tank failure rates, and those derived from detailed analysis. As can be seen, the detailed analysis does not give so high a frequency as the whole tank historical data. However, it does give a distribution of causes, which can be used as modification factors for the historical data.

1: Safety barrier diagram for High temperature / Floating roof tank in Generic analysis for floating roof tank

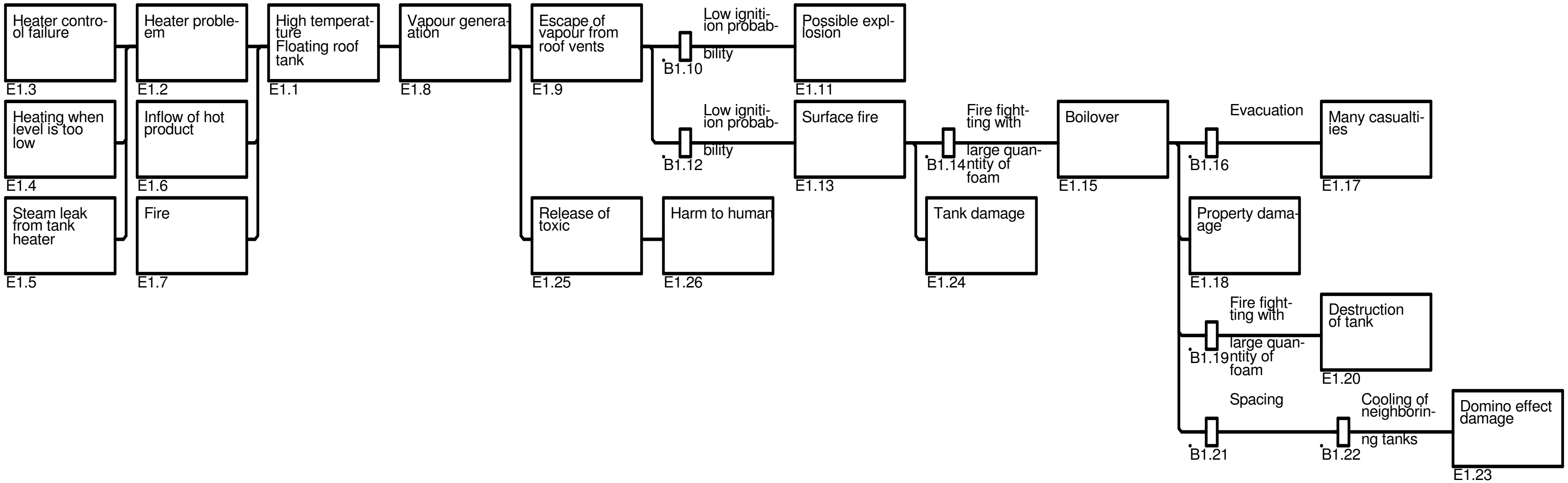


Figure 10.8

2: Safety barrier diagram for High level / Floating roof tank in Generic analysis for floating roof tank

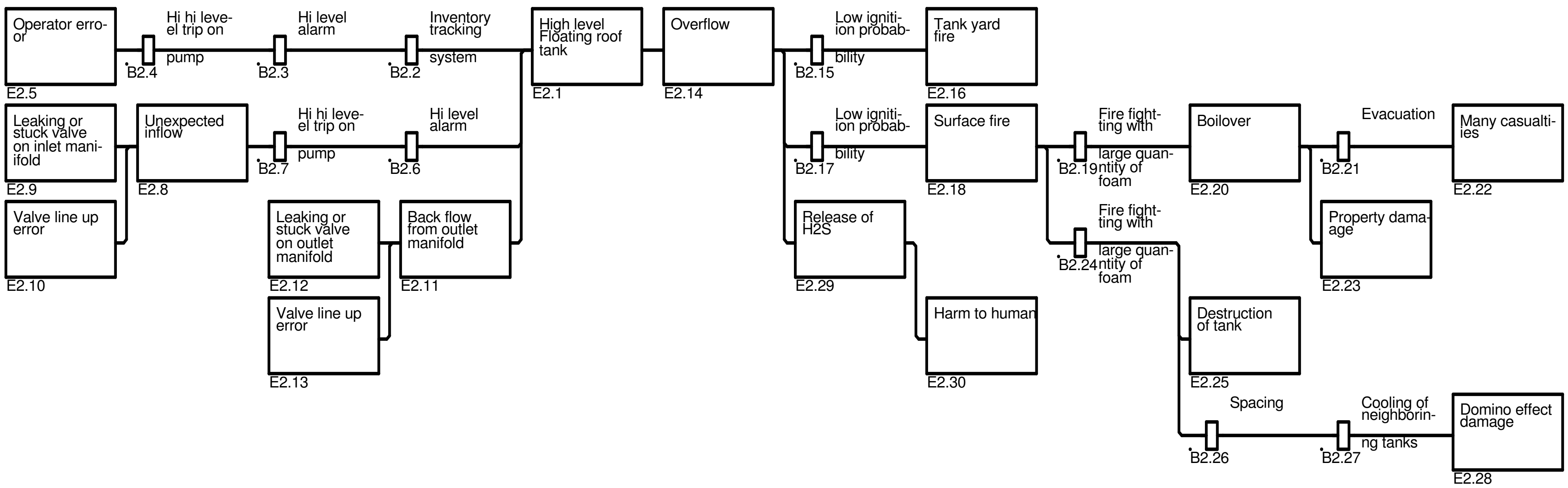


Figure 10.9

3: Safety barrier diagram for High pressure / Floating roof tank in Generic analysis for floating roof tank

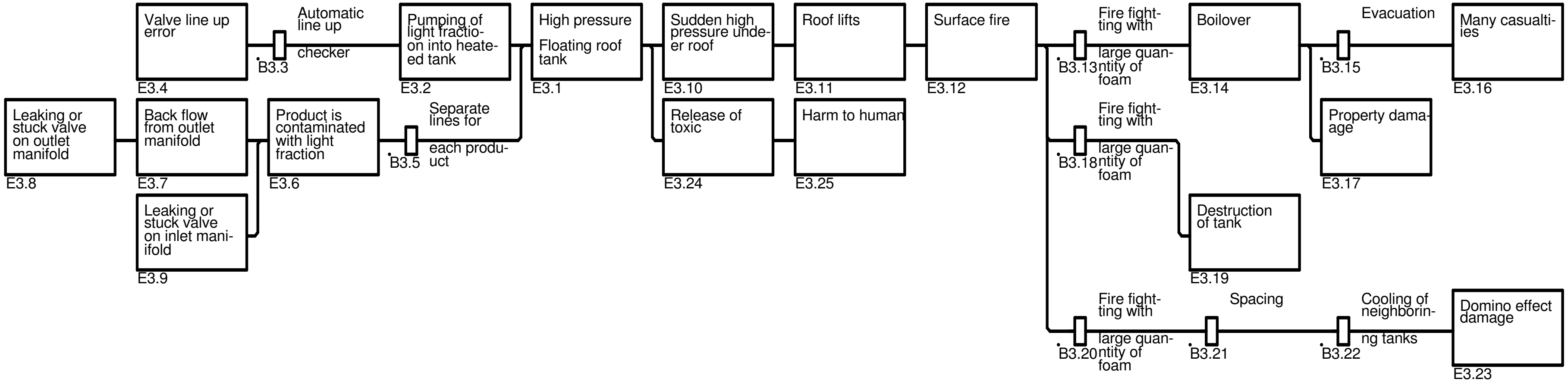


Figure 10.10

4: Safety barrier diagram for Leak / Floating roof tank in Generic analysis for floating roof tank

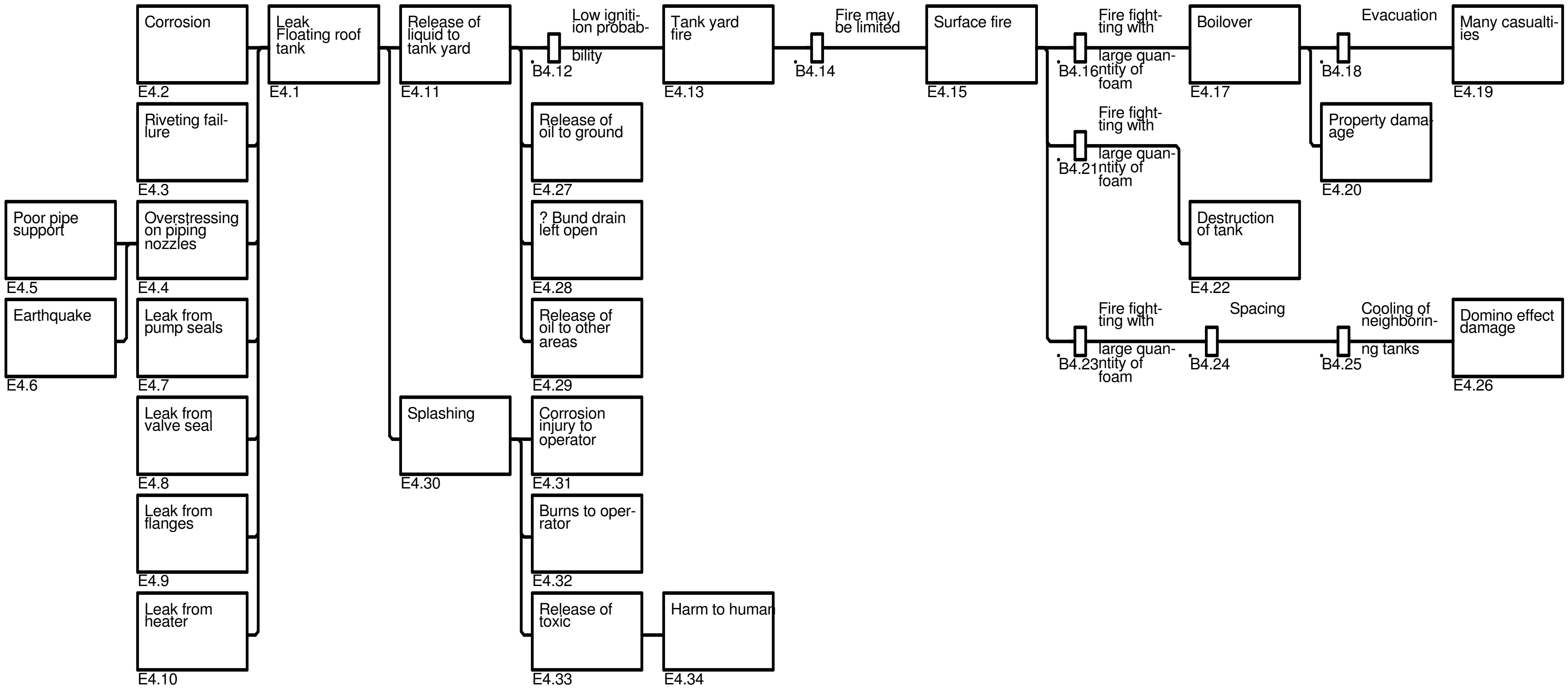


Figure 10.11

5: Safety barrier diagram for Rupture / Floating roof tank in Generic analysis for floating roof tank

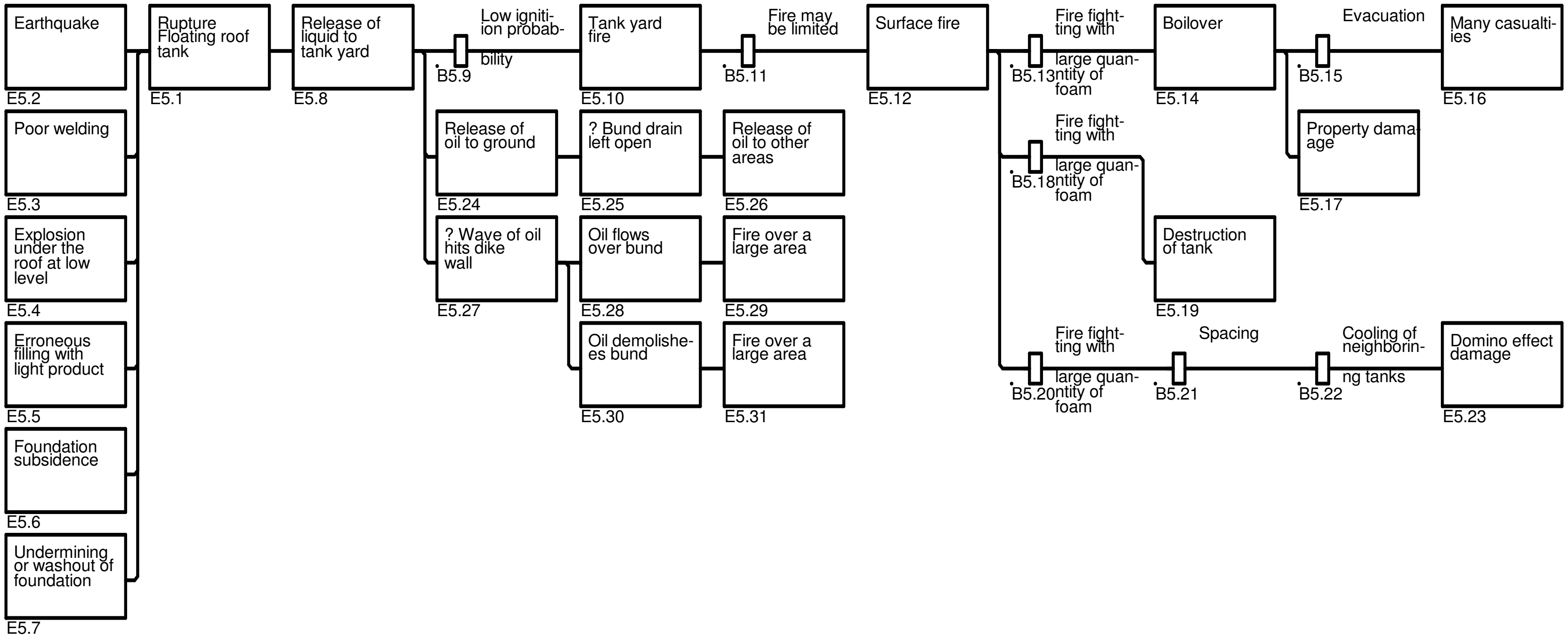


Figure 10.12

6: Safety barrier diagram for Surface fire / Floating roof tank in Generic analysis for floating roof tank

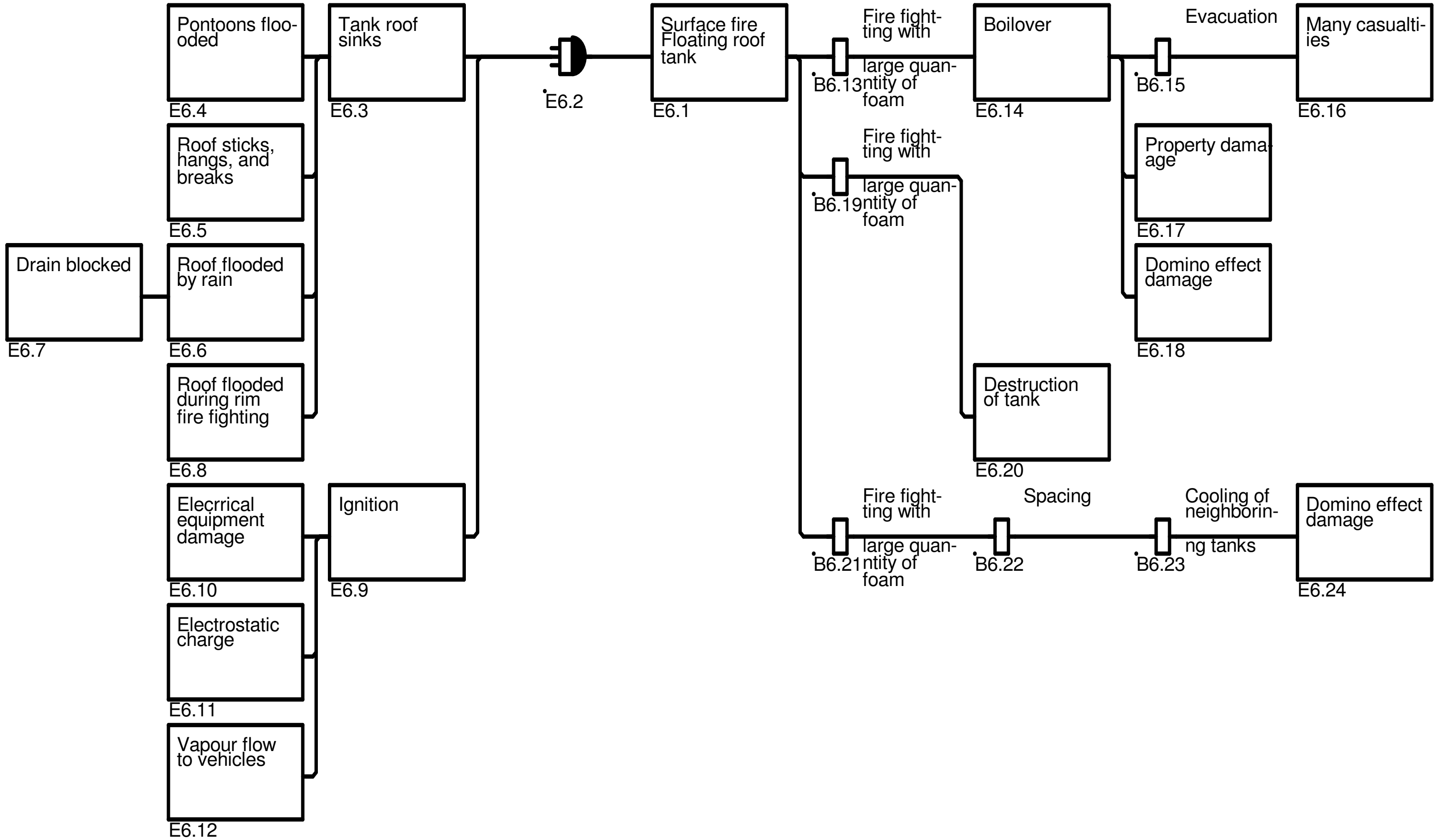


Figure 10.13

Floating roof tank Failure cause	Consequence	Number of items or metres	Frequency per item year	Susceptibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year
Tank bottom leak	small	1	1.78E-03	1		0			0			0		1.78E-03
Tank bottom leak	large	1	5.51E-05	1		0			0			0		5.51E-05
Tank wall or base weld leak	small	1	1.02E-03	1		0			0			0		1.02E-03
Tank wall base weld hole	large	1	4.13E-05	1		0			0			0		4.13E-05
Overflow	rupture	1	1.79E-03	1	LSHH	0	0.003516		0			0		1.79E-03
Pipe leak	medium	1	1.00E-03	1		0			0			0		1.00E-03
Pipe rupture	large	1	2.05E-03	1		0			0			0		2.05E-03
External fire	rupture	1	6.25E-06	1		0			0			0		6.25E-06
Tank rupture	large	1	2.75E-05	1		0			0			0		2.75E-05
Rupture	large	1	2.8E-05	1		0			0			0		2.75E-05
Drain valve left open	rupture	1	1.9E-05	1		0			0			0		1.88E-05
Valve left open	rupture	1	1.9E-05	1		0			0			0		1.88E-05
Vandalism	large	1	4.1E-04	1		0			0			0		4.13E-04
Lightning	fire	1	4.8E-05	1		0			0			0		4.76E-05
Roof damage	fire	1	2.4E-06	1					0			0		2.38E-06
Rim fire escalates	fire	1	3.0E-05	1	Foam pour	0	0.1		0			0		3.00E-05
Escalation to boilover	boil over	1	1.0E-05	1		0			0			0		1.00E-05
Bund fire	fire	1	6.0E-05	1	Foam mon.	0	0.1							6.00E-05
Total small														2.80E-03
Total medium														1.00E-03
Total large														2.61E-03
Total rupture														1.83E-03
Total fire														1.40E-04
Total boilover														1.00E-05

Table 10.11 Detailed release frequency analysis for a crude tank   Note: do not multiply by number of tanks for items marked with *

10.9 Algorithm for floating roof tank fire and release frequencies

In the following sections, a range of questions are posed, and the variation in release frequency associated with the answers is given.

Operation

#	Question	Action if Yes	Action if No
1	Is there a log of tank inventory kept continuously, so that the operator always knows the tank content	Go to 2	go to 3
2	Is there a frequent gauging of the tank contents?	Include inventory control as barrier in detailed analysis. Go to 3	go to 3
3	Is there a tank level indicator with a remote reading?	Include level monitoring as barrier in detailed analysis. Go to 4	go to 4
4	Is there a second independent level indicator for safety purposes?	Go to 6	go to 7
5	Is there an independent level switch for safety purposes?	Include switch type level trip as barrier in detailed analysis. Go to 7	go to 7
6	Does the safety instrument, if it exists automatically terminate tank filling? i.e. a level switch.	Include level indicator trip as barrier in detailed analysis. go to 7	go to 7
7	Does the operator always attend the tank during the entire period of tank filling?	Go to 8	Set overfilling susceptibility to 1 and use additional % from table 9.8.1, 9.8.2 or 9.8.3 Go to 8
8	Do the operators drain water from the tank periodically+	Increase drain lines left open percentage by 4*frequency of draining Go to 9	Exit
9	Is there a dual valve on the tank drain?	Include dual valve as barrier in detailed analysis	Exit

Table 10.12 Baseline frequency modifications for operational problems

Location

#	Question	Action if Yes	Action if No	Value	Reference
1	Is the area especially lightning prone	Multiply by factor for rim fire depending on the frequency of lightning strikes	exit		

Table 10.13

Note that the frequency for lightning strikes can vary from many per year to less than one per thousand years – local knowledge is needed to determine the frequency

Fire protection

#	Question	Action if Yes	Action if No	Value	Reference
1	Is there a rim fire detection system	Divide fire frequency factor 10 go to 2	go to 2		
2	Is there an effective foam pouring system for the tank rim	Insert foam pouring system risk reduction factor in detail analysis	Go to 3	Unavailability of foam flood system = 0.024	
3	Is there a foam projection, pouring, or foam monitor system with sufficient capacity to deal with the largest fires.	Insert foam system risk reduction factor in detail analysis	Exit	Unavailability of foam projection system = 0.02	

Table 10.14

10.10 References

1. Taylor, J.R. Process Safety Engineering, Designing and Building Safer Process Plant, Taylor Associates, 4th Edition 2001

2. Christensen, R.A. and Eibert, R.F., Above Ground Tank Storage, EL RM-623, Entropy Limited, MA, 1989

3. E&P Forum, Quantitative Risk Assessment, E& P Forum Report No. 11.8/250, 1996

4. Lauben, R.W. and Robinson, D.L. Acoustic Emission Integrity of Above Ground Storage Tanks, PWR Vol 5, Proceedings of Industrial Power Conf, ASME, 1989

5. Cox, A.W. Lees, F.P. and Ang, M.L., Classification of Hazardous locations, I Chem E, 1993

6. OPITSC, Atmospheric Storage Study, OPITSC, Singapore, 1990

7. LASTFIRE Project www.resprotint.co.uk

11 Hoses and Loading Arms

11.1 Construction

A significant proportion of the releases which take place in chemical plants arise as a result of failures of hoses and loading arrangements, when liquids transferred between tank truck, tank wagons, tank ships and storage tanks.

Figure 11.1 A loading arm for coupling ship to shore terminal piping.

Hoses used for liquids transfer are generally of high quality rubber composition. They may be armoured even when intended for atmospheric pressure transfers, and are almost always armoured when used for pressurised gases and liquefied gases. (exceptions are for example small diameter hoses for LPG.)

For cryogenic applications, special hoses able to withstand low temperatures are used. This applies also for hoses used for ammonia and LPG hoses, which reach temperatures around - 35^{o}C when the hose is vented after loading.

It is important that hoses of different types are kept segregated, and preferably should be labelled according to purpose.

For transfer of cryogenic media, bellows formed stainless steel hoses may be used. For liquefied gases such as chlorine in small quantities, copper tubing with up to 8 mm diameter may be used. For faster transfer carbon steel coils, which have some flexibility, may be used. Such coils have sufficient movement to allow connections to be made, provided that the tank truck or tank wagon is placed very close to its intended position.

Hoses will be damaged if trucks run over them. The damage may include deformation of reinforcement, so that it is weakened. The damage may also cause cracking of rubber.

In operation, hoses must be coupled up to mobile tankers using a flanged coupling, screw coupling, or rapid release coupling. The quality and condition of these couplings is critical, in that they can be damaged or incorrectly fitted, usually with a spray release as a result.

Many loading systems for liquefied gases or volatile liquids are fitted with nozzles to allow vapour return lines to be used. The liquid pumped from tank A to B displaces vapour which flows from tank B to A. if the operator forgets to connect up the line, vapour may be released depending on whether the vapour return line is opened.

Some unloading systems have a manifold, so that several tankers can unload at the same time. Failures and errors in valve line up can result in liquid being pumped via the same distribution or manifold pipe to the wrong tank, with overfilling and overflow as a potential problem.

Most tankers which transport pressurised liquefied gases are fitted with excess flow valves. These close if the hose is ruptured and a release begins. Operators sometimes become annoyed with these valves, however, because a) the tend to close if transfer is started too quickly and b) they sometimes stick. It is not unknown for operators to wire or jam the excess flow valves open.

The way in which the hose is connected to the couplings can be critical for safety. For low pressure hoses up to about 4" in diameter, screw clips are often used, but these are very unreliable. Proper, machine applied pressure clips are about 30 times more reliable for low pressure.

Hoses used for ship loading and unloading can be snapped if the ship moves too much during bad weather or if there is an excessive tidal range or a flood wave (sieche, tsunami etc.

Hoses used for loading and unloading rail tank wagons can be ruptured if the wagon is shunted, for example by uncontrolled movement of other wagons.

Tank truck hoses can be snapped if the tank truck driver drives away without uncoupling the hose. (This is relatively frequent, see table 11.3). it can also happen if the tank truck rolls away on an incline, with the brakes not being applied. Most companies insist that drivers apply chocks under the tank truck wheels before transferring hazardous liquids, for this reason.

Hoses may have safety release couplings, which close if the coupling is pulled away. It is important that the piping used to connect up to hoses is reinforced with supports, so that is can withstand the forces from hoses which are pulled away.

11.2 Operation

1. The driver drives the truck to the loading station.
2. The truck is stopped and put into free gear.
3. All electrical equipment and the motor are shut off.
4. Chocks are placed under the wheels.
5. Earthing cables are fitted.
6. The hose is coupled.
7. The tank valve is opened.
8. The delivery valve is opened.
9. Level in the tanker is checked on the tank level gauge, and co-ordinated with the metered content.
10. When filling is complete, valves are closed.
11. The interface and hose content is vented to flare or a safe place.
12. The hose is uncoupled and hung or laid correctly.
13. Earthing cable is disconnected.
14. The chocks are removed.
15. The tanker is driven away.
16. The truck is driven of for weighing. Maximum filling for liquefied gases is about 80% of volume, depending on substance and regulations.
17. Inspections of piping is visual, once per year.
18. Hoses are inspected daily.

11.3 Hazards

Four main accident types have been observed.

1. Leakage of flanges, couplings, or the hose itself, leading to fire. Impingement of flames can then lead to a BLEVE. Hoses and couplings are particularly vulnerable to leakage.

2. Trucks have been driven away without decoupling the hose.

3. Truck crash accidents have been known at loading racks.

4. Overfilling can give release of liquid through the safety valves. One case of rupture due to overfilling is known. Again a fire, with a following BLEVE, can result.

11.4 Case stories

1. A truck at an LPG station had finished loading, and valves were shut off by the station operator. The truck drove away, before the hoses had been uncoupled. LPG was released from the hoses themselves, and from leaks caused by damage to the piping.

2. A driver started filling of a gasoline truck, and departed to the rest room. He was delayed, and the truck overflowed through the top manhole. The gasoline ignited, and destroyed the truck and the loading racks.

3. The installation was provided with double hose connections, to allow filling two sections of a divided tank at the same time. A driver for a single tank section truck did not understand the system, and started filling with the valve for the second hose open. LPG escaped. An ESD operated only slowly so that significant quantities escaped. The system was later upgraded with interlocks providing nitrogen pressurisation and leak detection before loading could begin.

4. An operator connected up a hose for ammonia transfer, with a flat face flange coupling. The flange was not properly tightened, so that when transfer was started, ammonia sprayed out around the entire flange. The gas was detected, and the ESD valves shut off within 15 seconds. The hose was still pressurised, however, and continued to release ammonia. The operator, in protective clothing and with a self contained breathing apparatus, tightened the bolts. (personal observation by the author)

11.5 Hose and loading arm failure frequencies

Data for the shore to ship transfer of oil were collected by Technica, based on a 5 year study using data from UK HSE, with a reported oil transfer spill frequency of $1.8 * 10^{-4}$ per loading station year with causes as given in the following table:

Ship shore connection failure	**Number**
Hose broke	2
Hose leak	0
Loading arm failed	2
Loading arm leak	0
Pipe broke/valve failed	4
Pipe/flange/pump leak	5
Operator error	1
Ranging failure	
Mooring fault	3
Grounding	1
Passing ship	1
Overflow	
Overflow in loading	9
Backflow in discharge	0

Table 11.1 Distribution of failures causes for ship shore releases

For hoses the available data is quite sparse. Frequencies for hose loading releases are given by UK HSD in *Major Hazards Aspects of the Transport of Dangerous Substances*, 1991. The frequency derived is $1.3*10^{-5}$ per transfer. For comparison, CCPS gives a frequency of $5*10^{-3}$ failures per year for hoses. This CCPS frequency does not take into account automatic shutdown systems. Many systems have an extensive automatic shutdown system, based on pressure drop and gas detection. The dominant cause of failure for this system is the excess flow valve on the truck. The frequency of failure for this is taken from Ch. 5 (CCPS data) as $3*10^{-3}$ per year (check valve), and $2.2*10^{-3}$ per activation of the excess flow valve. This gives an overall frequency of hose failure continued releases of $3 * 10^{-8}$ per transfer, taking ESD into account.

Unit type	Release frequency per year
Alkylation	$10*10^{-3}$
Ammonia storage	$6.0*10^{-3}$
Ammonia production	$15*10^{-3}$
Gas plant	$3.1*10^{-3}$
Water treatment plant (Cl_2)	$12*10^{-3}$

Table 11.2 Hose failure frequencies from RMP data

Table 11.2 shows hose failure frequencies from the US RMP data base. Note that these data are for actual releases. However, the frequencies are so high, at about 1 per 100 years, that it is hard to believe that there could have been excess flow valves fitted in these cases. What is problematic is that there may be many more hose release which are never reported because they are stopped very quickly by means of ESD valves and excess flow valves. For this reason, the data in table 11.2 was calculated for plants which have actually had releases recorded, and not for plants with zero releases. Since the plants which did have releases had often 2 or 3, it seems reasonable to split the population in this way.

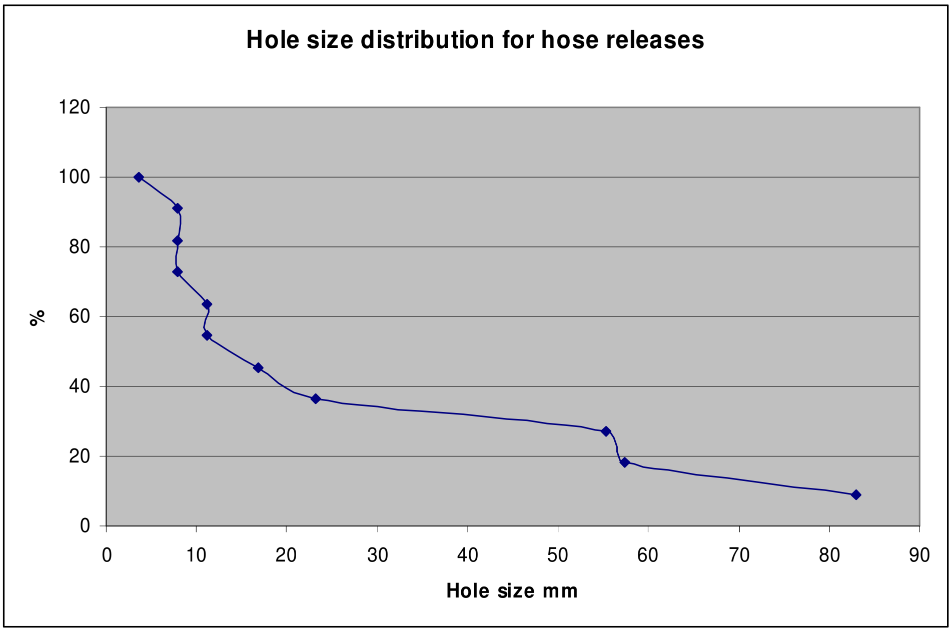


Figure 11.2 Hole size distribution for hose releases

Figure 11.2 shows the distribution of hole sizes for hose releases in the RMP data base. for alkylation unit, ammonia, gas plant and chlor alkali units (other data were either too sparse, or untrustworthy, for example with only one cause type recorded. The amount of data was insufficient to provide a reasonable breakdown of causes.

11.6 Typical frequencies

As can be seen from the previous section, there is a very wide range of values for loading and unloading release frequencies. This is hardly surprising, since the amount of usage and the quality of loading station installations varies widely. For high quality loading arm couplings the HSE/Technica value seems representative, with a value of $1.8*10^{-4}$ per loading station year. For general hose transfer of $5*10^{-3}$ per loading station year appears appropriate, being compatible with the US RMP data, and allowing for modification for variation in integrity. This value includes most of the factors for low integrity, including poor hose couplings, worn gaskets, and poor hose storage

Of course, the frequencies will vary, according to the amount of loading activity. For some stations, there will only be one or two loadings per year, while a practical maximum is about 10 per day. The UK HSD report value of $1.3*10^{-5}$ per transfer is well documented for gasoline deliveries, (it includes small spills). This would correspond to 300 transfers per year for a loading/unloading station. For loading arm releases, a connection frequency of about 60 per year on average can be assumed, based on chemical port experience from three ports. This gives an release frequency per loading or unloading of $3*10^{-6}$ for high quality loading arms. Table 11.3 gives baseline values for failure frequency.

Frequency basis	Equipment type	Total	Leaks up to 25 mm	Rupture, catastrophic
Per year	Hose	$5*10^{-3}$	$4*10^{-3}$	$1*10^{-3}$
	Loading arm	$1.8*10^{-4}$	$0.5*10^{-4}$	$1.3*10^{-4}$
Per transfer	Hose	$1.3*10^{-5}$	$0.8*10^{-5}$	$0.5*10^{-5}$
	Loading arm	$3*10^{-6}$	$0.86*10^{-6}$	$2.1*10^{-6}$

Table 11.3 Typical release frequency per year for loading arms and hoses

11.7 Failure causes

Table 11.4 gives the breakdown of failure frequency by causes, based on the MHIDAS data base.

Cause	**Number**	**%**
Connection came loose	4	2.4
Compression clip failure	1	0.6
Connecting error	1	0.6
Connection broke	1	0.6
Connection error	1	0.6
Connector leak	2	1.2
Coupling fault	9	5.5
Damage due to theft	1	0.6
Disconnection by operator while pressurised or filled	5	3.0
Drive away accident	7	4.2
Drive shaft damage	1	0.6
External fire	1	0.6
Hose contaminated	1	0.6
Hose cut by propeller	1	0.6
Hose slipped and snapped	1	0.6
Impact damage, truck hit hose	4	2.4
Leak on hose	23	13.9
Loading arm failure	1	0.6
Nozzle broke	1	0.6
Overpressuring	4	2.4
Piping break	6	3.6
Rail tanker roll away accident	4	2.4
Return line not coupled up	1	0.6
Roll away	4	2.4
Rupture, hose fault	54	32.7
Snapped hose	1	0.6
Spillage	1	0.6
Splash fill hose pulled from tank	1	0.6
Tank ship pull away, weather or mooring failure	13	7.9
Unrecorded	3	1.8
Valve break	2	1.2
Valve left open	4	2.4
Vandalism	1	0.6
Total	165	

Table 11.4 Causes of hose failure, from MHIDAS data base

Table 11.5 gives data on loading arm failure causes, taken primarily from the IChemE Accident Database (The MHIDAS database contained only 5 records on loading arms, whereas the I Chem E records contain a series from a special loading arm study)

Collision	3	6.7
Drive away	2	4.4
Earthing inadequate	2	4.4
Equipment failure	16	35.6
Fill pipe lifted out of filling hole	1	2.2
Ignition	3	6.7
Operator error	3	6.7
Ranging at mooring	6	13.3
Splash	1	2.2
Unknown	5	11.1
Vessel moorings broken	3	6.7
Total	45	100.0

Table 11.4 Loading arm failure causes, IChemE Accident Database

11.8 High integrity design

There is a wide range of variation in the release rates for hoses and loading arms depending on the quality of the design. The main measures are:

1. High quality hoses are used, with specifications appropriate to the liquid
2. Hoses are inspected regularly, and replaced whenever signs of wear appear.
3. Hoses are connected up without bending , pulling, or kinking.
4. Hoses are not mixed, so that the wrong type of hose can be used
5. The laydown or storage area for the hoses allows the hoses to be stored straight, or with smooth bends, and does not twist the hose.
6. Loading arms have appropriate ranging, so that they cannot be stretched or pulled sideways.
7. Hoses conveying liquefied gases have provision for depressurising, via a filling line
8. Connections have emergency shutdown valves and or excess flow valves to prevent releases.
9. There are separate connections for each hose, so that reverse flow from one tanker to another is prevented in the case of blockage of flow or valving errors. Check valves are not regarded as effective safety measures for reverse flow from tank to tank, because of the difficulty of testing them.
10. Automated leak testing is used before pumping or transfer commences

11. Locations are provided with drive away protection, so that trucks and rail cars cannot leave before couplings are disconnected.

12. Couplings are made robust, so that even if drive away occurs, it at most affects the hose, and does not damage piping.

13. If the material is flammable, there is good provision for grounding, and equalising voltages between tank and loading point.

Figure 11.3 High integrity loading arm coupling, designed by the author.

Figure 11.4 Properly stored hoses

Figure 11.5 A poor use of a hose, with a very tight bend at the coupling

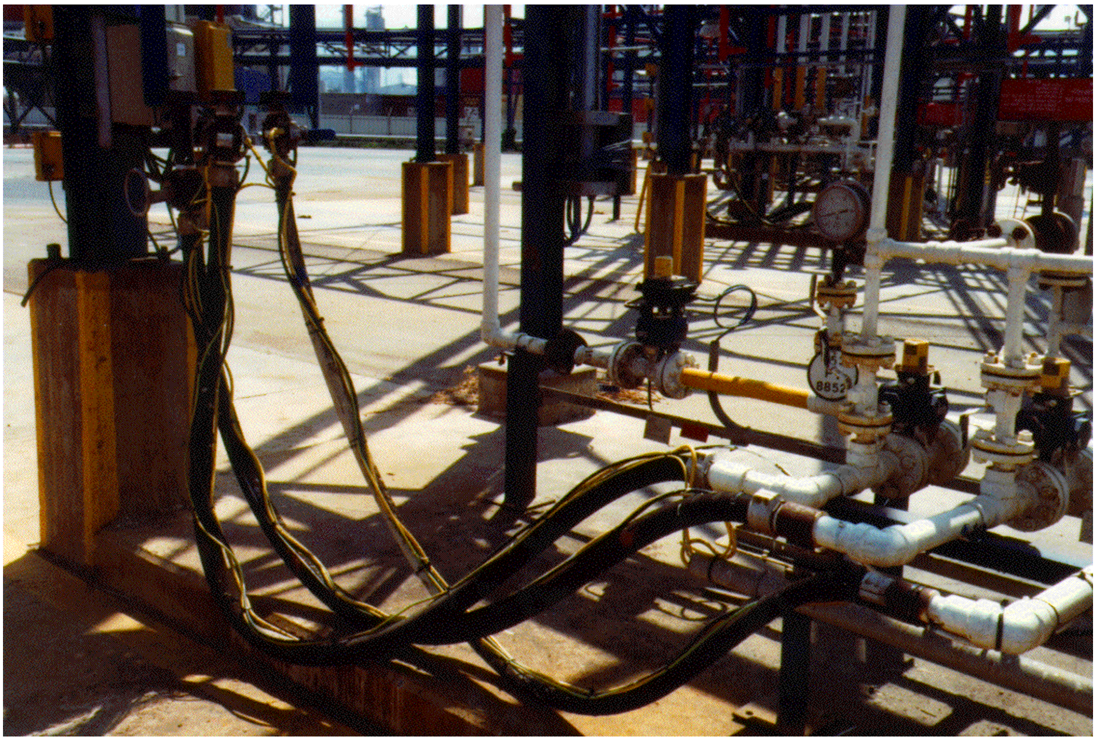


Figure 11.6 Hoses in an LPG loading rack, with redundant earthing cables.

11.9 Assessment of causal factors and susceptibilities

The causal factors and additional frequency contributions are assessed below, on the basis of MHIDAS database data. The variation in failure rates due to design differences is not so large as in other types of equipment. It should be noted that there are large differences in failure rates due to safety management factors, especially training and discipline in following procedures. These factors are not taken into account here.

No	Failure cause	% of releases	Source	Conse-quence	Suscept - ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Small bore piping	1	Reliability calc	small	1	1	5.6E-04	
2	Process piping, flanges, valves	5	Reliability calc	medium	1	1	4.4E-04	
3	Connection loose	18.1102	MHIDAS	medium	0.5	1	3.2E-03	Review of 7 loading racks
4	Connection broken	3.93701	MHIDAS	rupture	1	1	7.7E-05	
5	Hose impact damage	1.5748	MHIDAS	medium	0.5	1	2.7E-04	
6	Drive away	7.87402	Calc	small	1	1	4.4E-03	
7	Hose pulled out	0.7874	MHIDAS	rupture	0.3	1	5.1E-05	
8	Valve not closed etc	3.93701	MHIDAS	rupture	1	0.03516	2.2E-03	
9	Leak	18.1102	MHIDAS	medium	1	1	1.6E-03	
10	Overpressuring	3.14961	MHIDAS	medium	1	1	2.7E-04	
11	Hose rupture	42.5197	MHIDAS	rupture	0.2	1	4.2E-03	
	Total	106						

Table 11.5 Special causes and susceptibilities for loading and unloading tank truck hoses

No	Failure cause	% of releases	Source	Conse-quence	Suscept - ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Small bore piping	1	Reliability calc	small	1	1	2.7E-04	
2	Process piping, flanges, valves	5	Reliability calc	medium	1	1	1.1E-04	
3	Connection loose	17.6923	MHIDAS	medium	1	1	3.8E-04	
4	Connection broken	2.30769	MHIDAS	rupture	1	1	6.0E-05	
5	Hose impact damage	1.53846	MHIDAS	medium	1	1.0E+00	3.3E-05	
6	Ranging episode	10	MHIDAS	small	1	1.0E-01	2.7E-02	
7	Hose pulled out	0.76923	MHIDAS	rupture	0.1	1	2.0E-04	Review of 4 terminals
8	Valve not closed etc	3.84615	MHIDAS	large	1	1	1.0E-03	
9	Leak	17.6923	MHIDAS	medium	1	1	3.8E-04	
10	Overpressuring	3.07692	MHIDAS	medium	1	1	6.6E-05	
11	Connector broken	1.53846	MHIDAS	medium	1	1	3.3E-05	
12	Hose rupture (wear)	35.5385	MHIDAS	rupture	1	1	9.2E-04	
	Total	100					3.1E-02	

Table 11.5a Special causes and susceptibilities for loading and unloading ship loading hoses

Release frequencies	Small	Medium	Large	Rupture
Typical	1.80E-04	5.00E-05	1.30E-04	1.30E-04

No	Failure cause	% of releases	Source	Conse-quence	Suscept-ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Small bore piping	1	Reliability calc	small	1	1	1.8E-04	
2	Process piping, flanges, valves	5	Reliability calc	medium	1	1	1.0E-05	
3	Collision	9.7	MHIDAS	large	0.5	1	1.3E-04	Review of 3 chemical ports
4	Drive away	6.5	MHIDAS	rupture	1	1	1.5E-05	
5	Earthing inadequate	6.5	MHIDAS	medium	0.5	1	2.6E-05	
6	Equipment failure	48.0	Calc	rupture	1	1	1.1E-04	
7	Fill pipe lifted	3.2	MHIDAS	rupture	0.3	1	2.4E-05	
8	Ignition, fire	9.7	MHIDAS	medium	1	0.03516	5.7E-04	
9	Operator error	9.7	MHIDAS	large	1	1	6.5E-05	
10	Splash	3.2	MHIDAS	medium	0.2	1	3.3E-05	
		102.3871						

Table 11.6 Special causes and susceptibilities for loading arms, truck loading

Release frequencies	Small	Medium	Large	Rupture
Typical	1.80E-04	5.00E-05	1.30E-04	1.30E-04

No	Failure cause	% of releases	Source	Conse-quence	Suscept-ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Small bore piping	1	Reliability calc	small	1	1	4.8E-06	
2	Process piping, flanges, valves	5	Reliability calc	medium	1	1	5.0E-05	
3	Earthing inadequate	6.7	MHIDAS	small	0.5	1	6.4E-05	Review of 6 plants
4	Equipment failure	53.3	MHIDAS	rupture	0.3	1	3.6E-04	
5	Operator error	10.0	MHIDAS	small	1	0.0351596	1.4E-03	
6	Ranging at mooring	20.0	MHIDAS	small	1	1	9.6E-05	
7	Vessel moorings broken	10.0	MHIDAS	rupture	0.2	1	1.0E-04	
	Total	100.0						

Table 11.6a Special causes and susceptibilities for loading arms, ship loading

11.10 Detailed analysis

An example of a detailed analysis for a loading arm used for filling a gasoline tank truck is shown in figures 11.7 ff. The analysis goes beyond analysing the failures considered in section 11.8 and 11.9, since it takes into account all of the causes, of which overfilling of the truck is an important example.

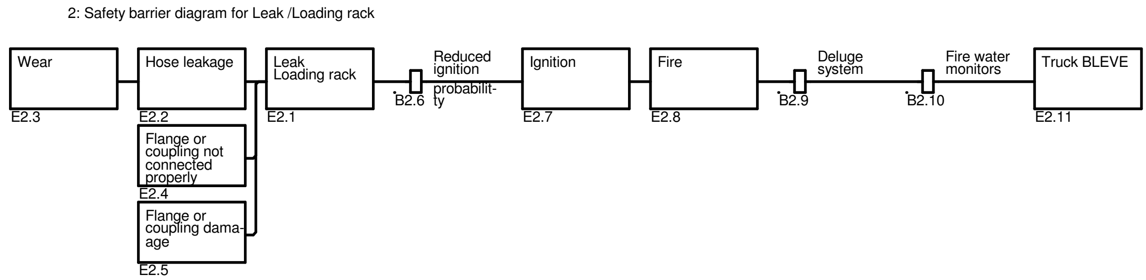


Figure 11.7 Failures leading to leakage at a loading rack with hose coupling

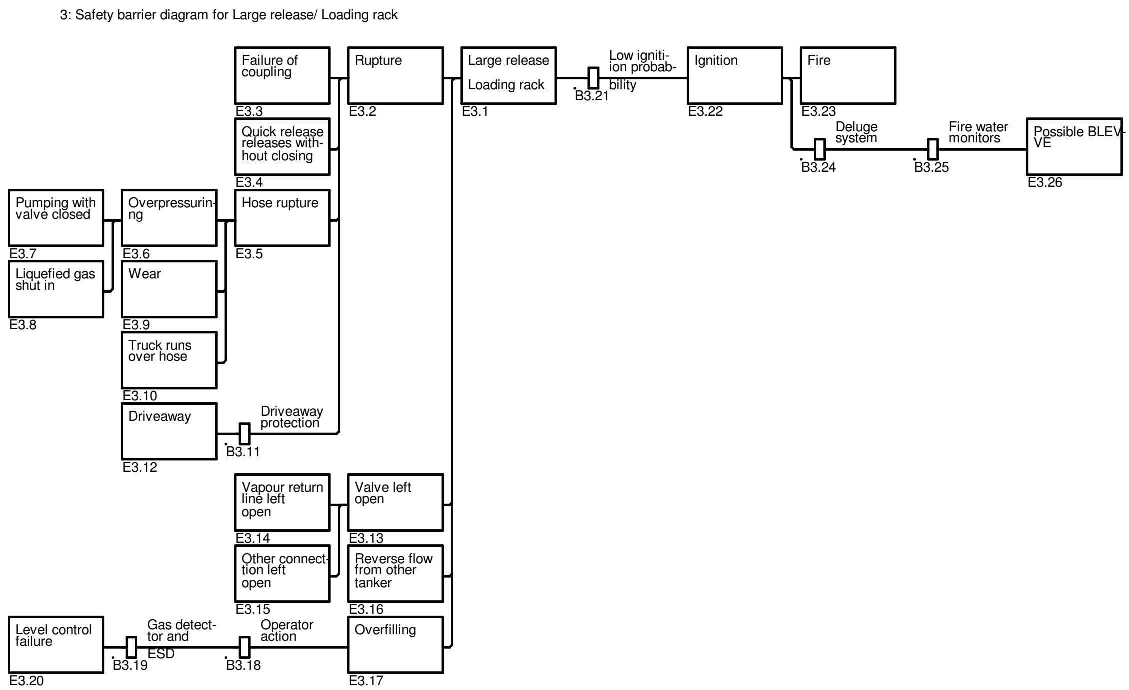


Figure 11.8 Failures leading to rupture at a loading rack with hose coupling

Release frequencies per year	Small	Medium	Large	Rupture	Total
Typical	5.00E-03	4.00E-03	0.00E+00	1.00E-03	1.00E-02

Truck and rail tanker hose Failure cause	Conse-quence	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Small bore piping	small	1	5.63E-04	1		0			0			0		5.63E-04	
Process piping, flanges, valves	medium	1	4.35E-04	1		0			0			0		4.35E-04	
Connection loose	medium	1	3.15E-03	1		0			0			0		3.15E-03	
Connection broken	rupture	1	8.33E-05	1		0			0			0		8.33E-05	
Hose impact damage	medium	1	2.74E-04	1		0			0			0		2.74E-04	
Drive away	small	1	4.44E-03	1		0			0			0		4.44E-03	
Hose pulled out	rupture	1	5.56E-05	1		0			0			0		5.56E-05	
Valve not closed etc	rupture	0	2.19E-03	1		0			0			0		0.00E+00	
Leak	medium	1	1.6E-03	1		0			0			0		1.58E-03	
Overpressuring	medium	1	2.7E-04	1		0			0			0		2.74E-04	
Hose rupture	rupture	1	4.5E-03	1		0			0			0		4.50E-03	
Total small														5.00E-03	
Total medium														5.71E-03	
Total rupture														4.64E-03	

Table 11.7 Release frequencies for tank truck and rail tanker loading hoses

Release frequencies per year	Small	Medium	Large	Rupture	Total
Typical	3.00E-03	1.00E-03	1.00E-03	1.00E-03	**6.00E-03**

Ship hose Failure cause	Conse-quence	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Small bore piping	small	1	2.73E-04	1		0			0			0		2.73E-04	
Process piping, flanges, valves	medium	1	1.07E-04	1		0			0			0		1.07E-04	
Connection loose	medium	1	3.80E-04	1		0			0			0		3.80E-04	
Connection broken	rupture	1	5.98E-05	1		0			0			0		5.98E-05	
Hose impact damage	medium	1	3.31E-05	1		0			0			0		3.31E-05	
Ranging episode	small	1	2.73E-02	0		0			0			0		0.00E+00	
Hose pulled out	rupture	1	1.99E-04	1		0			0			0		1.99E-04	
Valve not closed etc	large	1	1.00E-03	1		0			0			0		1.00E-03	
Leak	medium	1	3.8E-04	1		0			0			0		3.80E-04	
Overpressuring	medium	1	6.6E-05	1		0			0			0		6.61E-05	
Connector broken	medium	1	3.3E-05	1		0			0			0		3.31E-05	
Hose rupture (wear)	rupture	1	9.2E-04	1		0			0			0		9.20E-04	
Total small														2.73E-04	
Total medium														1.00E-03	
Total large														1.00E-03	
Total rupture														1.18E-03	

Table 11.8 Release frequencies for ship loading and unloading hoses

11.11 Algorithm for hose failure rates

From study of the accident records, the following factors were found to be significant in determining failure rates.

– How well the hoses are stored, and protected particularly from damage by trucks, and by abrasion..

– Whether there is drive away protection (gate, barrier, engine key interlock etc.)

– Whether the hoses are bent at a sharp angle.

– For liquefied gases whether there is an excess flow valve or emergency shut down valve to stop the release.

– For hazardous substances, whether there is a hose drain system, allowing the hose to be emptied before uncoupling.

Condition of storage

#	Question	Action if Yes	Action if No
1	Are the hoses stored off the ground, on racks, to prevent abrasion?	Go to 2	Impact damage susceptibility = 1
2	Are the hoses stored in such a way that trucks cannot drive over them?	Go to 3	Impact damage susceptibility = 1
3	Are the hoses inspected regularly and replaced if damaged?	Exit	Impact damage susceptibility = 3

Table 11.9

Conditions of use

#	Question	Action if Yes	Action if No
1	Are the hoses bent at a tight angle?	Hose leak susceptibility = 5	Go to 2
2	Are the hoses subject to stresses to reach nozzle locations?	Hose leak susceptibility = 5	Go to 3
3	Is the hose coupled to a manifold system?	Valve line up susceptibility = 1	Go to 4
4	Is there a vapour return line?	Add contribution for vent line release failure, in the detailed analysis	Exit

Table 11.10

Hose break protection

#	Question	Action if Yes	Action if No
1	For liquefied flammable gases, is there an ESD or excess flow valve on each end of the hose	Include ESD as a safety measure in the detailed analysis Go to 2	Go to 3
2	Is the ESD function activated by a gas alarm	Include automatic ESD as a safety measure in the detailed analysis ?	Go to 3
3	For liquids, is there a remote pump shut down ?	Include remote pump shutdown as a safety measure in the detailed analysis ?	Exit

Table 11.11

Drive away/roll away protection

#	Question	Action if Yes	Action if No
1	Is the loading station on a slope?	Go to 2	Go to 3
2	Are chocks fitted under the wheels before loading/unloading?	. Go to 3	Drive away susceptibility = 1. Go to 3
3	Is there a drive away protection system (gate, engine key interlock) barrier or traffic light with hose interlock?	Add drive away protection as safety barrier in detailed analysis	Drive away susceptibility = 1. Go to 4
4	Does the hose have a safety release coupling or the truck ?	Add safe release coupling as barrier in the detailed analysis. Go to 6	Go to 5
5	Is the piping and support reinforced sufficiently to be able to take loads from a drive away/hose snapping?	Go to 6	Record susceptibility to drive away accidents damaging piping as 1 in the detailed analysis. Exit (this assumes drive away will damage both piping and ESD function)
6	Is there an ESD or excess flow valve on each end of the hose	Include ESD as a safety measure for drive away in the detailed analysis Go to 7	Exit
7	Is the ESD function activated by a gas alarm	Include automatic ESD as a safety measure in the detailed analysis	Exit

Table 11.12

Loading station

#	Question	Action if Yes	Action if No
1	Is the road tanker loading station protected from other vehicle traffic with passive barriers?	Set susceptibility for crash incidents to 0 in detailed analysis. Go to 2	For road tankers, Set susceptibility for crash incidents to 1, go to 3 For rail tankers, go to 2
2	Are the rail tank wagons protected from shunting or runaway cars by means of wheel brakes or derailers?	Set susceptibility for collision incidents to 0 in detailed analysis. Go to 3	Set susceptibility for collision incidents to 0 Go to 3
3	For flammables, liquids is there a foam deluge system or fire water monitors capable of projecting from both sides of the truck or tank wagon?	Add foam deluge system as safety barrier in the detailed analysis. Go to 4	Go to 4
4	For liquefied flammable gases, are there fire water monitors capable of projecting an intense jet of water at both sides of the truck?	Add fire water monitors as safety barrier in the detailed analysis.	Exit

Table 11.13

11.12 Detailed analysis for loading arms

Loading arms, or Chicksan's ™ are used as alternatives to hoses, especially when the loading/unloading frequency is high. Flexibility is provided with two articulation seals and one rotation seal.

Loading arms are often protected by excess flow valves and emergency shut off valves, as for hoses.

Loading arms can fail

– By wear on the rotating seals

– By damage to the seals and articulated links if these freeze by leakage of compressed liquefied gases, or solidification of heavy oil, pitch etc.

– By pull away so that the linkages are stretched beyond their working range.

All the problems of coupling leakage described for hoses apply equally for couplings for loading arms

An example of a detailed analysis for a loading arm used for transferring propane from a ship is shown in figures 11.14 ff. The analysis goes beyond the failures considered in section 11.8 and 11.9, since it takes into account all of the causes of releases.

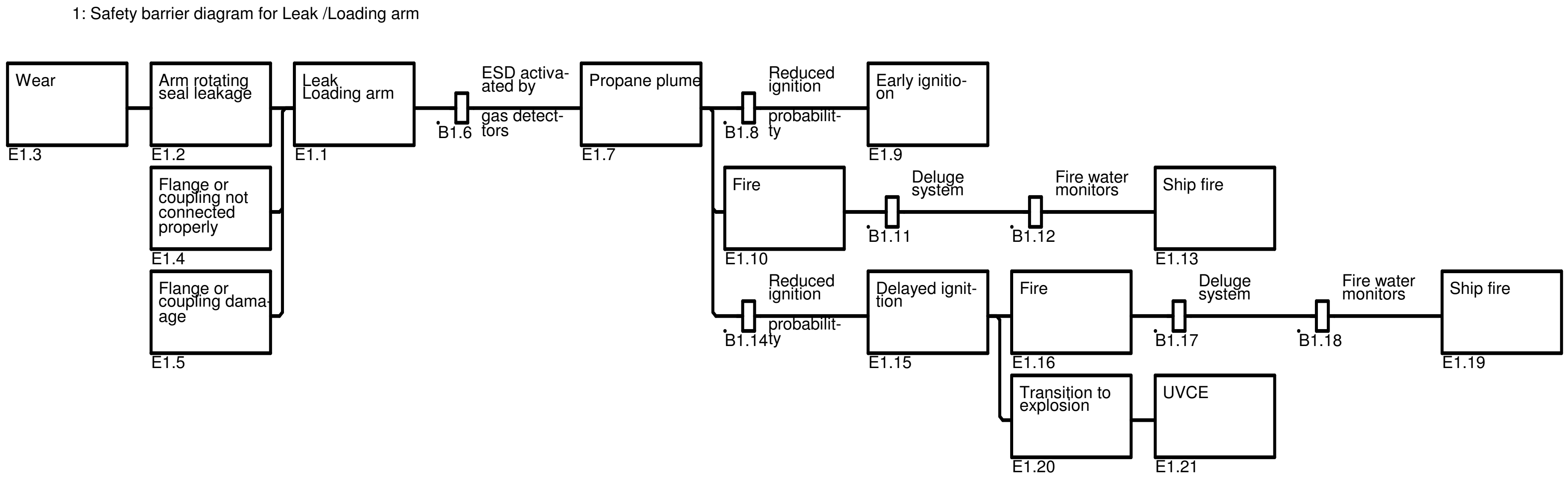


Figure 11.9 Failures leading to leakage at a loading arm

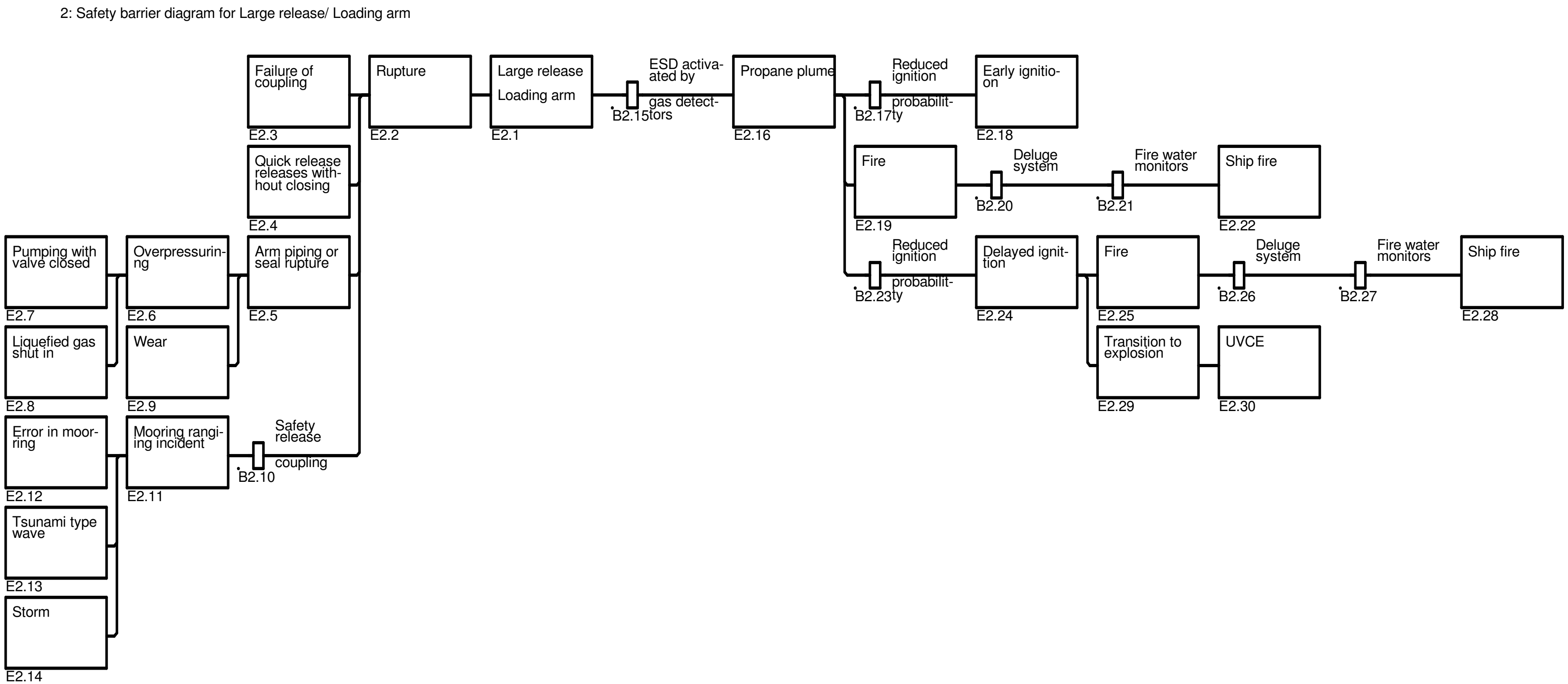


Figure 11.10 Failures leading to rupture at a loading arm

Release frequencies per year	Small	Medium	Large	Rupture	
Typical	1.80E-04	1.50E-04	0	1.30E-04	**4.60E-04**

Truck and rail tanker load arm Failure cause	Conse-quence	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Small bore piping	small	1	1.80E-04	1		0			0			0		1.80E-04	
Process piping, flanges, valves	medium	1	1.03E-05	1		0			0			0		1.03E-05	
Collision	large	1	1.30E-04	1		0			0			0		1.30E-04	
Drive away	rupture	1	1.45E-05	1		0			0			0		1.45E-05	
Earthing inadequate	medium	1	2.65E-05	1		0			0			0		2.65E-05	
Equipment failure	rupture	1	1.08E-04	1		0			0			0		1.08E-04	
Fill pipe lifted	rupture	1	2.42E-05	1		0			0			0		2.42E-05	
Ignition, fire	medium	1	5.65E-04	1		0			0			0		5.65E-04	
Operator error	large	1	6.5E-05	1		0			0			0		6.50E-05	
Splash	medium	1	3.3E-05	1		0			0			0		3.31E-05	
Total small														1.80E-04	
Total medium														6.35E-04	
Total rupture														1.47E-04	

Table 11.14 Detailed release frequency analysis for tank truck or rail tanker loading arm

Release frequencies per year	**Small**	**Medium**	**Large**	**Rupture**
Typical	1.80E-04	1.50E-04	0	1.30E-04

Ship load arm Failure cause	Conse-quence	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Small bore piping	small	1	4.78E-06	1		0			0			0		4.78E-06	
Process piping, flanges, valves	medium	1	5.00E-05	1		0			0			0		5.00E-05	
Earthing inadequate	small	1	6.37E-05	1		0			0			0		6.37E-05	
Equipment failure	rupture	1	3.65E-04	1		0			0			0		3.65E-04	
Operator error	small	1	1.36E-03	1		0			0			0		1.36E-03	
Ranging at mooring	small	1	9.56E-05	1		0			0			0		9.56E-05	
Vessel moorings broken	rupture	1	1.03E-04	1		0			0			0		1.03E-04	
Total small														1.52E-03	
Total medium														5.00E-05	
Total rupture														4.68E-04	

Table 11.15 Detailed release frequency analysis for ship loading arm

11.13 Algorithm for loading arm failure rates

From study of the accident records, the following factors were found to be significant in determining failure rates.

– How well the loading arm is maintained

– How well the arm is protected from damage by trucks.

– For truck or rail tanker loading stations, whether there is drive away protection (gate, barrier, engine key interlock).

– For liquefied gases whether there is an excess flow valve or emergency shut down valve to stop the release.

– For hazardous substances, whether there is a piping drain system, allowing the loading arm to be emptied before uncoupling.

Because of the safety measures associated with loading arms, the release frequencies associated with the arm will tend to underestimate the actual damage frequencies, except in the case of rupture of large ship to shore connections. For this reason, a base rate of failures has been taken here which is the actual rate of releases. Special release causes which can occur, such as those arising from drive away are then added to the base rate.

Arm break protection

#	Question	Action if Yes	Action if No
1	Is there an ESD or excess flow valve on each end of the hose	Include ESD as a safety measure in the detailed analysis Go to 2	Go to 3
2	Is the ESD function activated by a gas alarm	Include automatic ESD as a safety measure in the detailed analysis ?	Exit

Table 11.16

Drive away/roll away protection (truck and rail tanker loading/unloading)

#	Question	Action if Yes	Action if No
1	Is the loading station on a slope?	Go to 2	Go to 3
2	Are chocks fitted under the wheels before loading/unloading?	Set susceptibility for drive away accidents to 0.1 Go to 3	Set susceptibility for drive away accidents to 1. Go to 3
3	Is there a drive away protection system (gate, engine key interlock) barrier or traffic light with hose interlock?	Add drive away protection as safety barrier in detailed analysis	Go to 4
4	Does the hose have a safety release coupling or the truck ?	Add safe release coupling as barrier in the detailed analysis. Go to 6	Go to 5
5	Is the piping and support reinforced sufficiently to be able to take loads from a drive away/hose snapping?	Go to 6	Record susceptibility to drive away accidents damaging piping as 1 in the detailed analysis. Exit
6	Is there an ESD or excess flow valve on each end of the hose	Include ESD as a safety measure for drive away in the detailed analysis Go to 7	Exit
7	Is the ESD function activated by a gas alarm	Include automatic ESD as a safety measure in the detailed analysis ?	Exit

Table 11.17

Mooring ranging protection (ship loading/unloading)

#	Question	Action if Yes	Action if No
1	Is there a possibility of tsunami, seiche, tidal wave?	Add the frequency of tsunami per loading period to the failure frequency for rupture. Go to 2	Go to 2
2	Is there a possibility for the ship to slip its moorings during a storm ?	Add the frequency of violent storms per loading period to the failure frequency for rupture. Go to 3	Go to 3
3	Is there a significant tide	Record susceptibility to mooring errors as 1	Exit

Table 11.18

Loading station

#	Question	Action if Yes	Action if No
1	Is the road tanker loading station protected from other vehicle traffic with passive barriers?	Set susceptibility for crash incidents to 0 in detailed analysis. Go to 2	For road tankers, Set susceptibility for crash incidents to 1, go to 3 For rail tankers, go to 2
2	Are the rail tank wagons protected from shunting or runaway cars by means of wheel brakes or derailers?	Set susceptibility for collision incidents to 0 in detailed analysis. Go to 3	Set susceptibility for collision incidents to 0 Go to 3
3	For flammables, liquids is there a foam deluge system or fire water monitors capable of projecting from both sides of the truck or tank wagon?	Add foam deluge system as safety barrier in the detailed analysis. Go to 4	Go to 4
4	For liquefied flammable gases, are there fire water monitors capable of projecting an intense jet of water at both sides of the truck?	Add fire water monitors as safety barrier in the detailed analysis.	Exit

Table 11.19

12 Pumps

Pumps provide one of the primary causes of releases within plants. Ordinary pump seals fail typically once every year per or two years. The range is, for the most difficult services, a failure every month, up to one failure every 20 years, for the most carefully designed seals.

Leaks from seals are only rarely sufficiently large to cause hazardous concentrations outside plants. If such hazardous releases are possible, it is probable that special measures such as tandem seals, or “canned” pumps with magnetic drives, will be used.

Pumps generate much larger releases if the seal is completely destroyed. A possible cause of this is overheating of the seal, or mechanical damage.

The other major cause of releases from pumps is complete destruction of the pump casing. This can happen from overpressuring, from erosion or corrosion of the pump casing, or from overheating by pumping against a closed discharge valve (dead heading).

Positive displacement pumps are especially sensitive to the problem of closed discharge valves, or blocked discharge piping. It is rare for the pump itself to explode, but the piping can be ruptured explosively.

12.1 Equipment description

Centrifugal pumps are very widely used in the process industry, for pumping liquids of varying types, from light liquefied gases to sludge and mud.

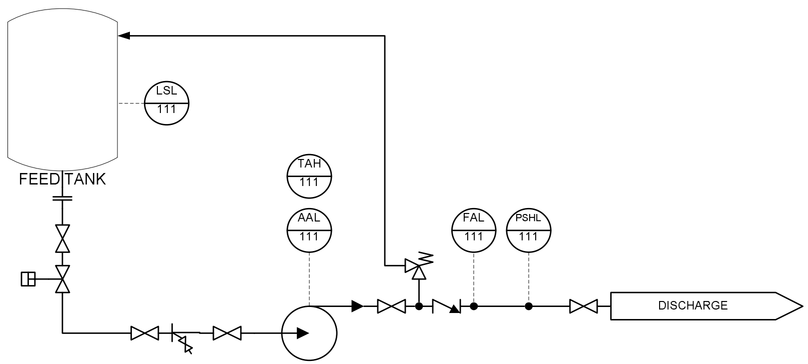
Two basic design are available, - horizontal and vertical in line. Vertical in line pumps are less sensitive to stresses derived from piping, they can generally “float” in the piping system. The motor is generally rigidly attached to the pump body, so that the axle remains aligned. Capacities of in line pumps may be limited, for some applications.

Centrifugal pumps are the equipment which causes accidents most frequently in process plant. The problems arise particularly from seals, which prevent release of liquid around the pump axle. Seals are very delicate mechanical devices, and are vulnerable to vibration solids in the pumped medium, wear, and temperature changes.

The devices which exist around the pump can be as important for safety as the pump itself. A minimum arrangement for a pump pumping non hazardous material is shown in fig. 12.1.

The pump may be preceded by a filter, and followed by a check valve to prevent back flow. The filter may be unnecessary if the liquid flow can be guaranteed to be clean.

SAFETY MEASURE	REFERENCE
LSL	API 14C
LSH	API 14C
PSV	API 14C
ESD	API 14C
FAL	API 14C
PSHL	API 14C
TAH	API 14C
AAH High current alarm	PSR



PROCESS SAFETY ENGINEERING

CENTRIFUGAL PUMP
SAFETY FEATURES

SIZE	FSCM NO	DWG NO	REV
A4		100-002	0

JRT 15 APRIL 2000

SCALE 1:1 SHEET 1 OF 1

Spectacle plates have been added to allow positive shut off of flow during maintenance. Also a spill valve has been added, to ensure that there will be a minimum flow even when outlet valves are closed.

Secondary support systems are important for safety. Large pumps operate with cooling systems, providing water or oil to bearings, seals. The seals will also have forced lubrication/sealing fluid on larger pumps.

For hazardous liquid service, double mechanical seals or tandem seals may be fitted. The inboard seal receives liquid from the process itself as lubricant/sealing fluid. The secondary outboard seal is provided with fluid from a special buffer fluid tank, using either a small pump or static head to ensure that the seal is supplied. If pressure increases between the seals this means that the primary seal is beginning to foil, and an alarm is given.

Pumps are generally driven by an induction motor, (some by a steam turbine, not considered here). The motor itself is subject to hazards, primarily overheating and short circuiting. Overheating may result from poor ventilation or cooling, but the most frequent cause is bearing failure.

The alignment between pump and motor is also critical. If the motor is misaligned with the pump, or becomes misaligned, vibration occurs which destroys seals and bearings, and may cause axle breakage. Many pumps are fitted with flexible couplings between motor and pump, in order that they can tolerate a level of misalignment.

Positive displacement pumps are used when very high pressures are to be achieved, or for pumping high viscosity liquids or slurries. There are many types, including reciprocating pumps, Mono pumps and similar types, with a screw formed rotor in a screw formed housing, and gear pumps.

12.2 Pump hazards (ref. 1)

1. Seal leaks

Seal leaks are the primary hazard on pump. When the seal begins to be damaged, it can overheat, and metal inclusions can be generated, so that the seal is destroyed.

If the liquid is flammable, it is quite likely that a fire will occur at the pump, ignited by the heat generated at the failed seal.

Once a fire has started, the heat input to the pump can cause further damage. For example, the axle can deform gaskets can be destroyed and the pump motor can catch fire.

If the pump housing is damaged, then much larger releases are likely to occur.

2. Bearing fires

Bearings can fail as a result of wear and breakage of rollers or balls through fatigue; or due to cooling failure; or due to failure of lubrication.

Once bearings burn, other flammable may be ignited, for example liquids flowing from seals if the flow is free to a catch try.

Bearings which overheat can seize (jam). In this case, the axle or the bearing housing can break and the pump may break up.

3. Pump housing failure

Pump housings may fail due to corrosion, or due to erosion wear. Sometimes a pump housing will break due to solids jamming in the impeller. Examples are welding rods, bolts, stones, and solids crystallised from solution.

Pump housings can also burst due to overpressuring and overheating arising when the pump is operated at zero flow. This is a problem particularly when pumping substances which may decompose, such as ammonium nitrate.

Pump housings may fail due to vibration. There are two main types of vibrations affecting pumps. One arises from imbalance or misalignment. An impeller can become imbalanced because of corrosion, or due to deposits. Axle misalignment can arise because the pump is not installed properly, or because the pump moves. Pumps may move because of stressed imposed by the piping, or because of hydraulic vibrations and hammering.

The other type of vibration can arise due to pulsations in flow. Pulsation can occur due to resonance in the flow. Another cause is vapour locks in either discharge or suction piping, which move backwards and forwards at a high point. An even worse cause is vertical two phase flow, in which bubbles of vapour flow up a pipe, and then allow liquid to crash back.

4. Turbining

A special problem which arises occasionally is that of a pump which acts like a turbine, due to back flow of liquid from a high pressure vessel on the discharge side. This can occur if the pump is tripped off, and the check valve fails or is not fitted. Such turbining can cause damage to seals, and over speeding of the pump. Cases are known in which a pump has generated voltage in a motor which was undergoing maintenance. An electrocution resulted.

5. Pump break up

Pumps can break up due to seal and bearing failure, cracking of bearing housings etc. and due to overpressuring especially by pumping with the discharge blocked, so that overheating occurs.

6. Overpressuring with positive displacement pumps

Positive displacement pumps of the reciprocating and gear types can often generate very high pressures. If the discharge is blocked, or has a valve which has not been opened, the piping can often rupture. One advantage of the positive displacement pumps is that they act like a valve, so that the amount of material released cannot be greater than the pumping rate, plus the inventory in the piping and any back flow.

12.3 Case stories

1. A multistage boiler feed pump broke up due to heavy vibration. The axle was thrown like a spear, and penetrated into the control room and through a control panel.

2. A pump conducting hydrogen fluoride liquid was subject to pump seal failure on a double seal! The cause was blockage in the seal flushing system. Acid escaped, and gave hazardous HF concentrations at up to 300m.

3. A vertical LPG transfer pump suffered seal failure. The escaping LPG ignited, probably from the seal damage or from a transformer station at 10m distance. The fire destroyed the seal and the pump motor, and damaged piping flanges at up to 15m distance piping was undamaged, because it was shrouded with foam glass insulation.

4. (Sparks and Wauchel, Hydrocarbon processing July 1977).

A four stage centrifugal pump suffered repeated failure at the splitter between pump stages. The cause was found to be resonance in a long cross over between second and third stage.

5. (ibid) A 10 stage centrifugal pump was driven by an 8 cylinder engine through a gear box. Engine speed was 690 to 800 rpm with pump speeds of 3880 to 4500 rpm. Repeated crack shaft failures occurred. Measurement of natural frequencies of vibration revealed torsional frequencies of 1020 and 4260 cpm for the shaft and 70 Hz for the piping. Fitting an acoustic filter eliminated the failures.

6. A pump for ammonium nitrate solution became blocked by crystallisation during a shutdown. Continued pumping resulted in overheating and burst of the casing.

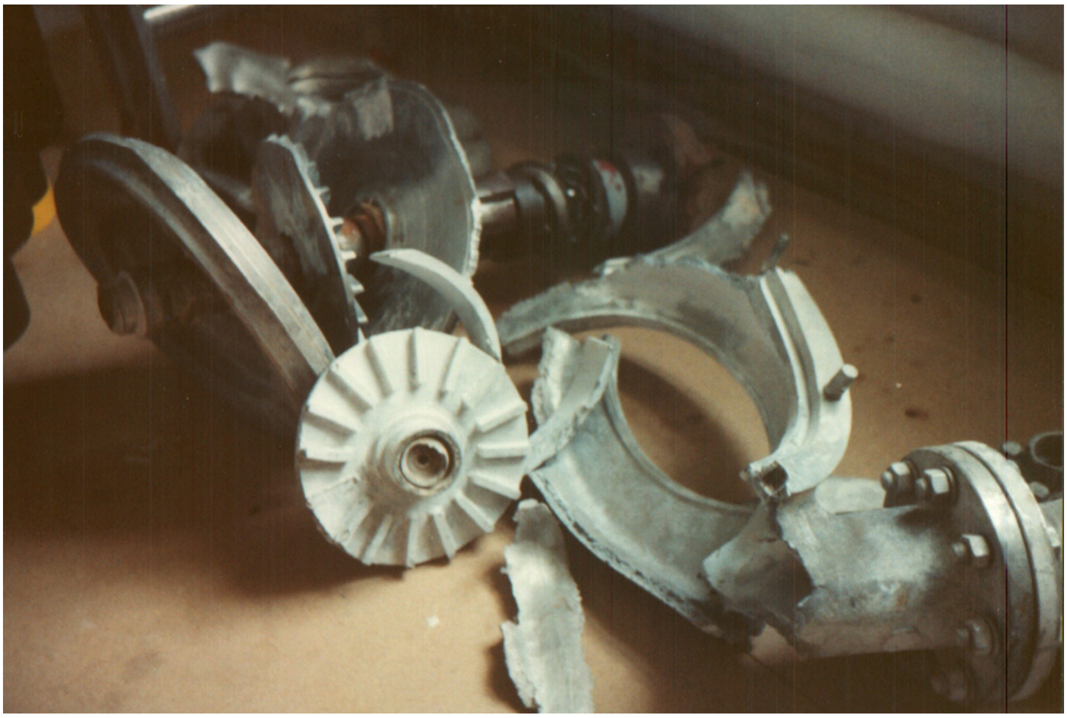


Figure 12.2 Remains of an ammonium nitrate solution pump after rupture.

Pumps provide one of the primary causes of releases within plants. Ordinary pump seals fail typically once every year or two years. The range is, for the most difficult services, a failure every month, up to one failure every 20 years, for the most carefully designed seals.

Leaks from seals are only rarely sufficiently large to cause hazardous concentrations outside plants. If such hazardous releases are possible it is probable that special measures such as tandem seals, or "canned" pumps with magnetic drives, will be used.

Pumps generate much larger releases if the seal is completely destroyed. A possible cause of this is overheating of the seal, or mechanical damage.

The other major cause of releases from pumps is complete destruction of the pump casing. This can happen from over pressuring from erosion, corrosion of the pump casing, or for overheating by pumping against a closed discharge valve (dead heading).

Positive displacement pumps are especially sensitive to the problem of closed discharge valves, or blocked discharge piping. It is rare for the pump itself to explode, but piping can be ruptured explosively.

12.4 Frequency of releases from pumps

Pump releases are generally one of three types:

- Pump seal releases
- Pump casing bursts
- Pump bursts due to external fire, sometimes as a result of earlier seal leaks

Of these, seal releases are the most difficult to define. Some pump seal types require a small bleed of liquid, all the time, either to lubricate the seal, or to keep the seal faces apart by pressure, or both. During a failure, a seal will often gradually develop, with an increasing leak rate, until finally judged to require maintenance, or until the damage is so severe that the seal breaks. Generally failure rate data are collected for seals when they need to be replaced. The actual size of the release at this time will usually be unknown, since such data are not usually recorded. It is important, in any case, when collecting pump failure rate data, to exclude the cases where seals are replaced as part of a preventive maintenance programme.

Some seal leaks will not be recorded as releases, because the seals are double or tandem types. Although a failure occurs, there is no release, the failure being detected while the second seal is still working. Unfortunately no data have so far been obtained in which dual seals have been treated separately. In by far the majority of cases, failure of dual or tandem seals does not lead to a release. However, failures arising from loss of lubrication, heavy vibration, or overheating, will generally affect both seals within a short time.

The data obtained for pumps from the US RMP records is given in table 12.4.1. Data from UK HSE offshore release surveys were given in ch. 4. The results are given alongside the RMP data, for comparison purposes. The data seem to be reasonably consistent within a factor of 5, and the differences are explicable in terms of the different levels of concern i.e. the size of leak which is considered recordable. There are two outliers in the data though. LPG storage and ammonia production pumps have relatively low release rates, and it is speculated that these values reflect the increased use of tandem and dual seals in these units.

Unit type	Releases	No of equipment years	Release frequency per 10^4 years
US RMP data			
Crude unit	5	19680	2.5
Alkylation unit	27	1560	173
Reformer	5	350	143
LPG storage	2	6800	6
Light ends unit	35	5600	62
HDS	2	600	100
Ammonia production	4	200	3
Ammonia storage	5	250	303
Nitric acid	5	200	183
Ammonia distribution	5	325	384
UK HSE Offshore data			
Pump, centrifugal, double seal	40	6624	60.4
Pump, centrifugal, single seal	16	4457	51.6
Pump, reciprocating, double seal	10	1230	81.3
Pump, reciprocating, single seal	3	760	39.5
Company C			
Centrifugal pump leaks	27	437	78

Table 12. 1 Pump release frequencies

Berezowski, of Clark Refining, collected extensive data over a number of years. His results, including none release failures are given in table 11.3 and table 11.4

Pump Failure Categories	**1994**	**1995**	**1996**	**1997**
Seal Failures	141	151	102	116
Ball/Roller Bearing Failures	30	33	37	22
Case Gasket Leak	8	6	11	6
Overhaul of Packed Pump	13	4	5	2
Material in Pump	12	6	15	6
Corrosion/Erosion	5	4	10	4
Internal Rubbing	6	4	9	1
Sleeve Bearing Failure	0	0	2	4
Infrequent	0	9	3	3
Vertical Pump Failures	18	22	25	24
Geared High Speed Pumps	12	5	6	7
Positive Displacement Pumps	15	6	3	5
High Vibration	4	26	13	10
Total	**264**	**276**	**241**	**210**
Non-Pump & Non-Failure Categories	**1994**	**1995**	**1996**	**1997**
Coupling Failures	27	24	24	36
Minor Repairs	19	70	80	30
Packing Adjustment or Re-Packing	68	54	42	39
Metering (Controlled Volume) Pumps	65	77	59	68

Table 12.2 Actual number of pump failures (ref. 11.3, Berezowski)

Categories	1994	1995	1996	1997	No. of Pumps
Seal Failures	2.5	2.3	3.4	3	346
Ball/Roller Bearing Failures	9.7	8.8	7.8	13.2	290(1)
Overhaul of Packed Pump	5.2	16.8	13.4	33.5	67
Vertical Pump Failures	2.6	2.1	1.8	1.9	46
Geared High Speed Pumps	2.2	5.2	4.3	3.7	26
Positive Displacement Pumps	2.3	5.7	11.3	6.8	34
All Failures	1.7	1.6	1.9	2.1	449(2)
Coupling Failures	12.7	14.3	14.3	9.5	342

Notes:
(1) Vertical In-line pumps, geared pumps were not included.
(2) Metering pumps, submersible pumps were not included.

Table 12.3 Failure frequencies per year for different kinds of pumps, (ref. 11.3, Berezovski)

The release frequency from Berezovski's data show the difference between the actual frequencies observed by maintenance teams (where a release frequency of 2 per year is not unexpected) and values for significant releases which give offsite effects. The author worked on improving one pump seal which had a failure rate of about one per three weeks, with no real success (The service was hot wet acid chloride residue, so a high failure rate is not unexpected). The data illustrate the importance of experienced and alert plant maintenance staff to keep plant running reliably.

12.5 Hole sizes

Hole sizes for pump releases were derived from US RMP data on specific substances, from release time and release quantity data. Values are shown in figure 12.5.1 for refinery pumps. Note that complete pump ruptures are not taken into account. If this is done, the hole size is equivalent to the pipe diameter.

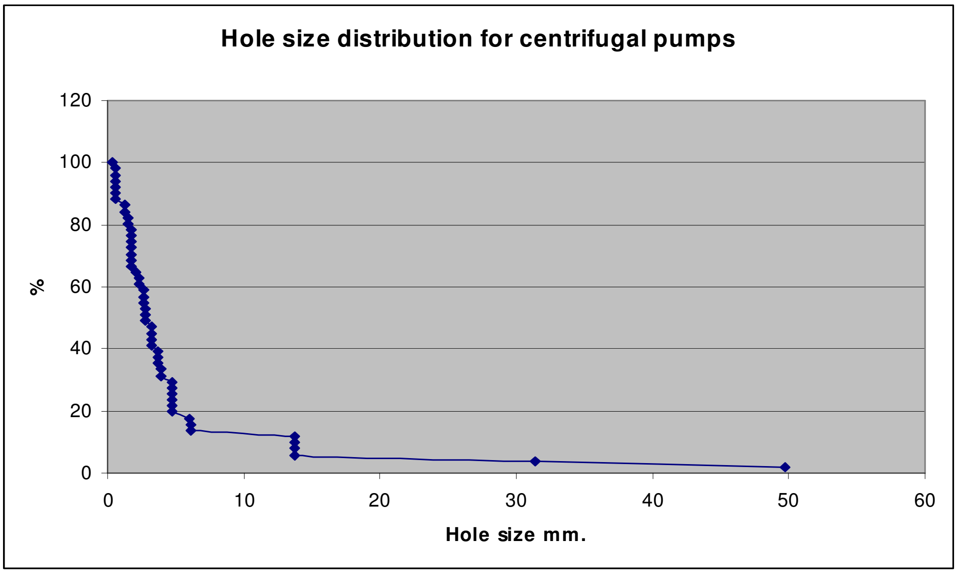


Figure 12.3 Hole size distribution for refinery pumps.

12.6 Typical frequency for pump releases

From the data in section 12.4 and 12.5, a value for typical release frequency of $100*10^{-4}$ per year seems a reasonable basis, with a size distribution as shown in table 11.4. The HSE offshore data is also shown for comparison. Note that the double seal release rates in the HSE data seem curious, with double seals contributing very little to risk reduction. (Also, the HSE size contributions do not add to 100% in the original document)

		Hole size distribution %			
Equipment type	**Failure frequency per 10^4 year**	**< 10 mm**	**10-25 mm**	**25-50 mm**	**NA**
HSE Offshore data					
Pump, centrifugal, double seal	60.4	83	13		15
Pump, centrifugal, single seal	51.6	87	9	14	9
US RMP data, assessed					**Rupture**
Pump, centrifugal, single seal	100	88	8	4	15
Typical,					**Rupture**
	100	88	8	2	2

Table 12.3 Typical and baseline release frequencies for pumps, excluding seal leaks

These actual values are very low, at least as far as seal leaks are concerned, and must reflect the effect of protection systems. Seal failures occur with frequencies from 1 per year to 1 per 10 years for most pump seal types. Seal leaks represent only a small part of the RMP data, because small leaks are unlikely to give releases which can reach the plant fence. Similarly, seal flows on platforms are often either closed in to flushing systems or piped up to oily water drains. A typical frequency for single seal very small leaks of 0.3 per year is assumed here if the seal flow drain is open, with a frequency of $60.4*10^{-4}$ per year for dual or tandem seals, based on HSE data. It will generally be better to make a separate analysis of the seal system reliability as a unit, if a sealing fluid system is provided.

12.7 Causes of pump releases

Causes of pump releases based on selected RMP data from refinery, ammonia and chlorine units are shown in table 12.4 The data cannot be said to be very informative.

Cause	Number	%
Equipment failure	30	65.2
Operator/maintenance error	7	15.2
Unsuitable equipment	9	19.6

Tale 12.4 Pump release cause distribution, from US RMP data

Table 12.5 gives a much more extensive list of causes, broken down by effect type and mechanism of failure. The table is based on 116 pump failures with significant consequences, taken from the MHIDAS data base.

Cause		#	%	%
Breakdown		3		2.9
Explosion	13		12.5	
	seal leak	4		3.8
	overheat, dry running	2		1.9
	procedure error	1		1.0
				0.0
Fire	32		30.8	
	broken pipe	2		1.9
	overheat	2		1.9
	sabotage	1		1.0
	seal leak	10		9.6
	seized bearing	1		1.0
	unknown	14		13.5
Leak	16		15.4	
	corrosion on nipple	1		1.0
	drain left open	1		1.0
	dry running	1		1.0
	flange bolts not tightened	1		1.0
	gasket failure	2		1.9
	plug mounted wrongly	1		1.0
	valve leak	1		1.0
	unknown	8		7.7
Seal leak		18		17.3
Rupture	15		14.4	
	seal rupture	1		1.0
	age	1		1.0
	dead head pumping	2		1.9
	wrong material	1		1.0
	mechanical damage	1		1.0
	vibration	1		1.0
	overpressure	1		1.0
	pipe rupture	1		1.0
	unknown	6		5.8
Spill	11		10.6	
	broken pipe	2		1.9
	disconnected, erroneous start	2		1.9
	frozen pipe	1		1.0
	maintenance, not isolated	1		1.0
	drain left open	1		1.0
	unknown	4		3.8
	control failure	1		1.0
	left running	1		1.0
Release through SV		1		1.0
Unknown		1		1.0

Table 12.5 Causes of pump failure, MHIDAS data base

12.8 Assessment of causal factors and susceptibility

Susceptibilities to special release types, and associated modification factors, are given in table 12.6. This is based on table 12.5, with unknown causes eliminated. Note that the majority of pumps are vulnerable to most types of failure, so that the susceptibility analysis gives fewer corrections than for other equipment types. The values for susceptibility were determined by examining pumps at two refineries and two fine chemicals plants.

Release frequencies	Small	Medium	Large	Rupture	Fire	Explosion
Typical	8.8E-03	8.0E-04	2.0E-04	2.0E-04	1.3E-03	1.06E-03

No	Failure cause	% of releases	Source	Release size	Suscept - ibility	Safety measures	Failure Rate	Basis for susceptibility assessment
1	seal leak	7.6		explosion	1	1	1.1E-04	6 plants
2	overheat, dry running	3.8		explosion	0.5	1	1.1E-04	
3	procedure error	2		explosion	1	1	3.0E-05	
4	broken pipe	3.8		fire	1	1	2.5E-05	
5	overheat	3.8		fire	0.5	1	4.9E-05	
6	sabotage	2		fire	1	1	1.3E-05	
7	seal leak	19.2		fire	1	1	1.2E-04	
8	seized bearing	2		fire	1	1	1.3E-05	
9	corrosion on nipple	2		small	1	1	1.1E-03	
10	drain left open	2		small	1	1	1.1E-03	
11	dry running	2		small	1	1	1.1E-03	
12	flange bolts not tightened	2		small	1	1	1.1E-03	
13	gasket failure	3.8		small	1	1	2.1E-03	
14	plug mounted wrongly	2		small	1	1	1.1E-03	
15	valve leak	2		small	1	1	1.1E-03	
16	seal rupture	2		rupture	1	1	1.0E-04	
17	age	2		rupture	1	1	1.0E-04	
18	dead head pumping	3.8		rupture	0.5	1	3.8E-04	
19	wrong material	2		rupture	1	1	1.0E-04	
20	mechanical damage	2		rupture	1	1	1.0E-04	
21	vibration	2		rupture	1	1	1.0E-04	
22	overpressure	2		rupture	0.5	1	2.0E-04	
23	pipe rupture	2		rupture	1	1	1.0E-04	
24	broken pipe	3.8		large	1	1	3.9E-05	
25	disconnected, erroneous start	3.8		large	1	1	3.9E-05	
26	frozen pipe	2		large	0.3	1	6.8E-05	
27	maintenance, not isolated	2		large	1	1	2.0E-05	
28	drain left open	2		large	1	1	2.0E-05	
29	control failure	2		large	1	1	2.0E-05	
30	left running	2		large	1	1	2.0E-05	
31	release through SV	2		large	0.1	1	2.0E-04	
	Total	97.4						

Table 12.6 Susceptibility and modification factors for centrifugal pumps

12.9 Detailed analysis

A detailed analysis is included here for a medium sized centrifugal pump, with a full set of safety measures, pumping flammable solvent.

1: Safety barrier diagram for High pressure /Centrifugal pump in Centrifugal pump generic release analysi

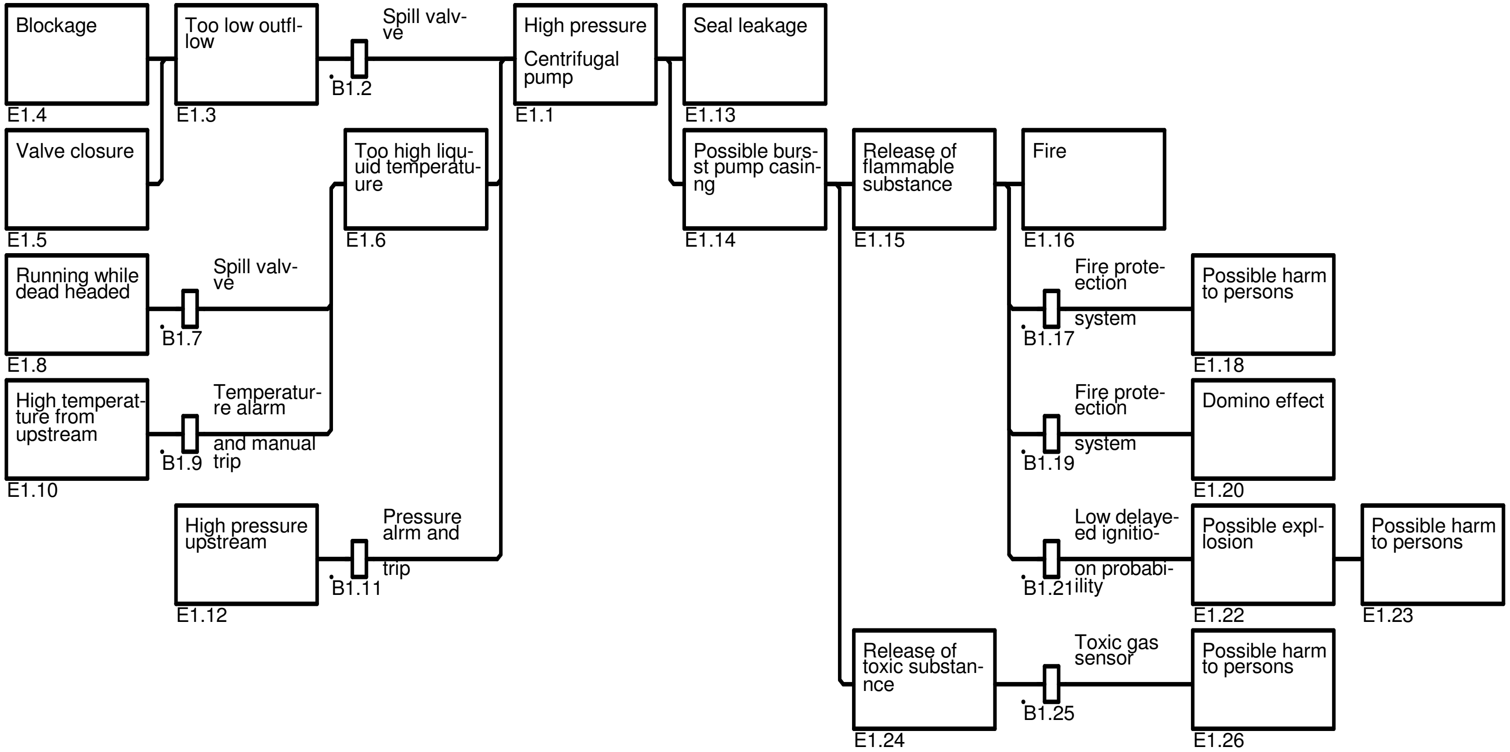


Figure 12.4

2: Safety barrier diagram for Pressure too low /Centrifugal pump in Centrifugal pump generic release analysi

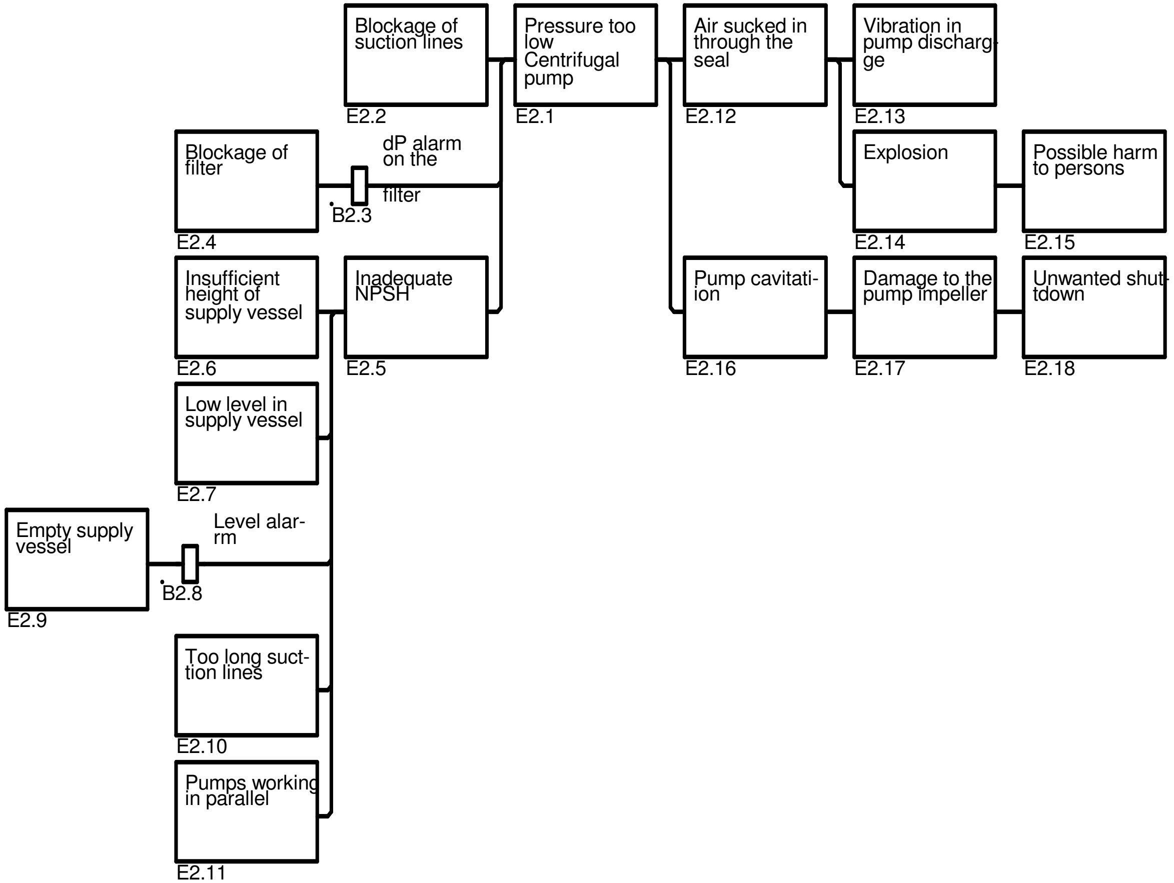


Figure 12.5

3: Safety barrier diagram for Too low flow /Centrifugal pump in Centrifugal pump generic release analysi

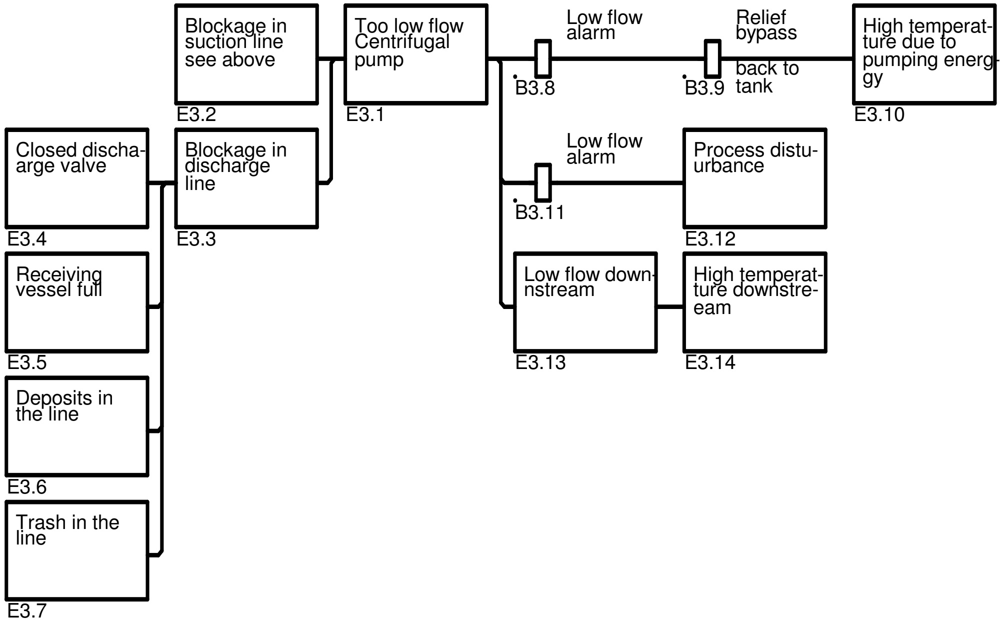


Figure 12.6

4: Safety barrier diagram for High flow /Centrifugal pump in Centrifugal pump generic release analysi

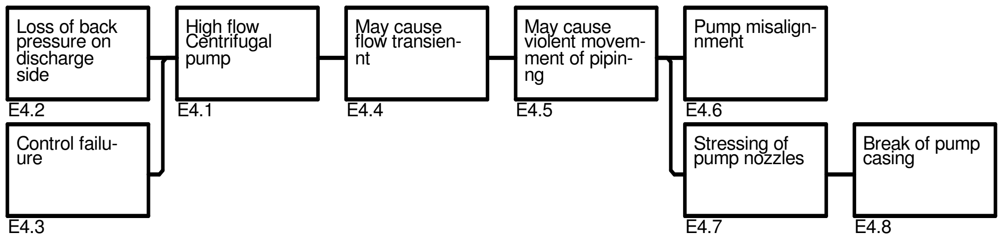


Figure 12.7

5: Safety barrier diagram for High temperature /Centrifugal pump in Centrifugal pump generic release analysi

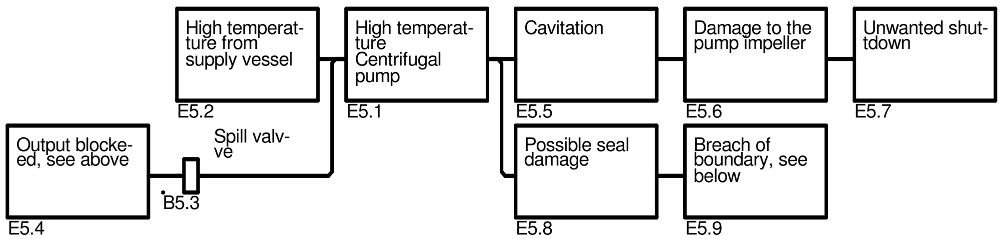


Figure 12.8

11: Safety barrier diagram for Seal leak /Centrifugal pump in Centrifugal pump generic release analysi

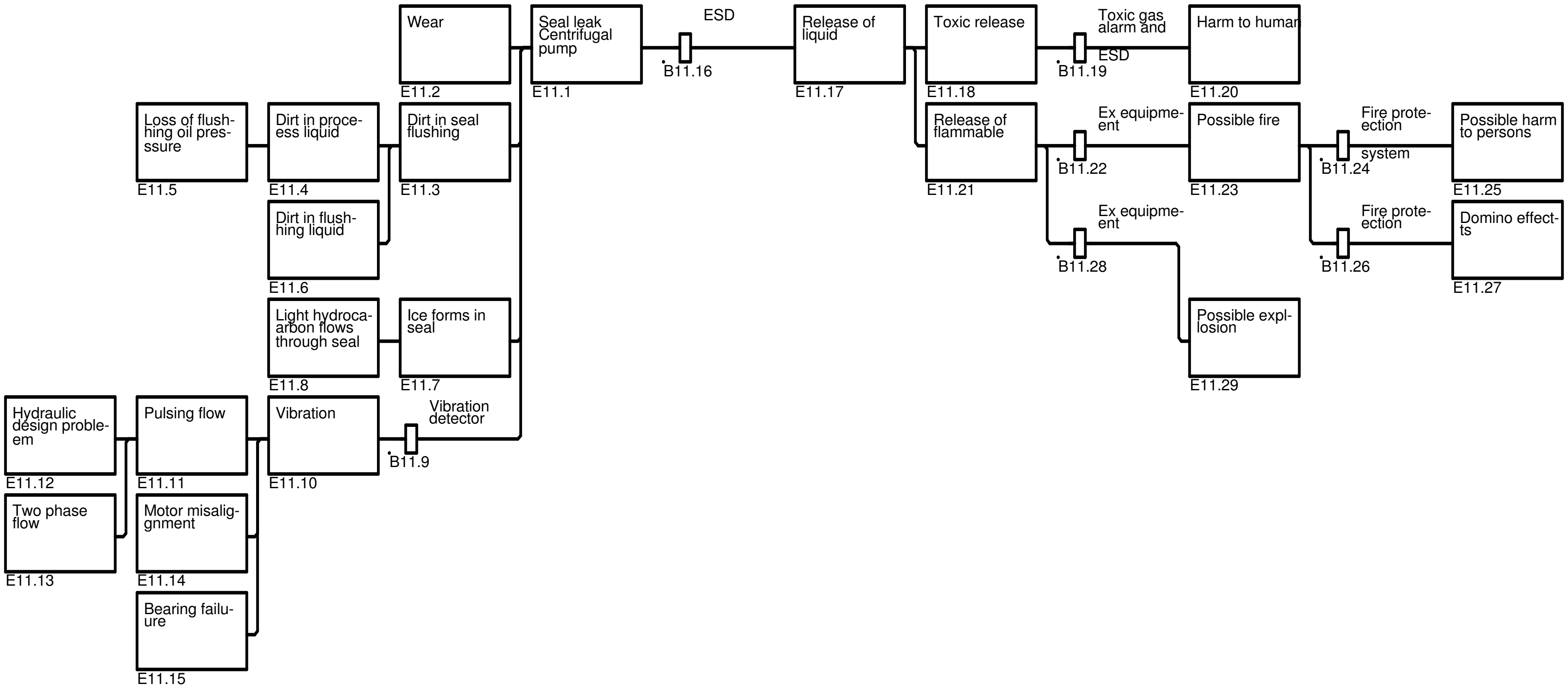


Figure 12.9

7: Safety barrier diagram for Leak /Centrifugal pump in Centrifugal pump generic release analysi

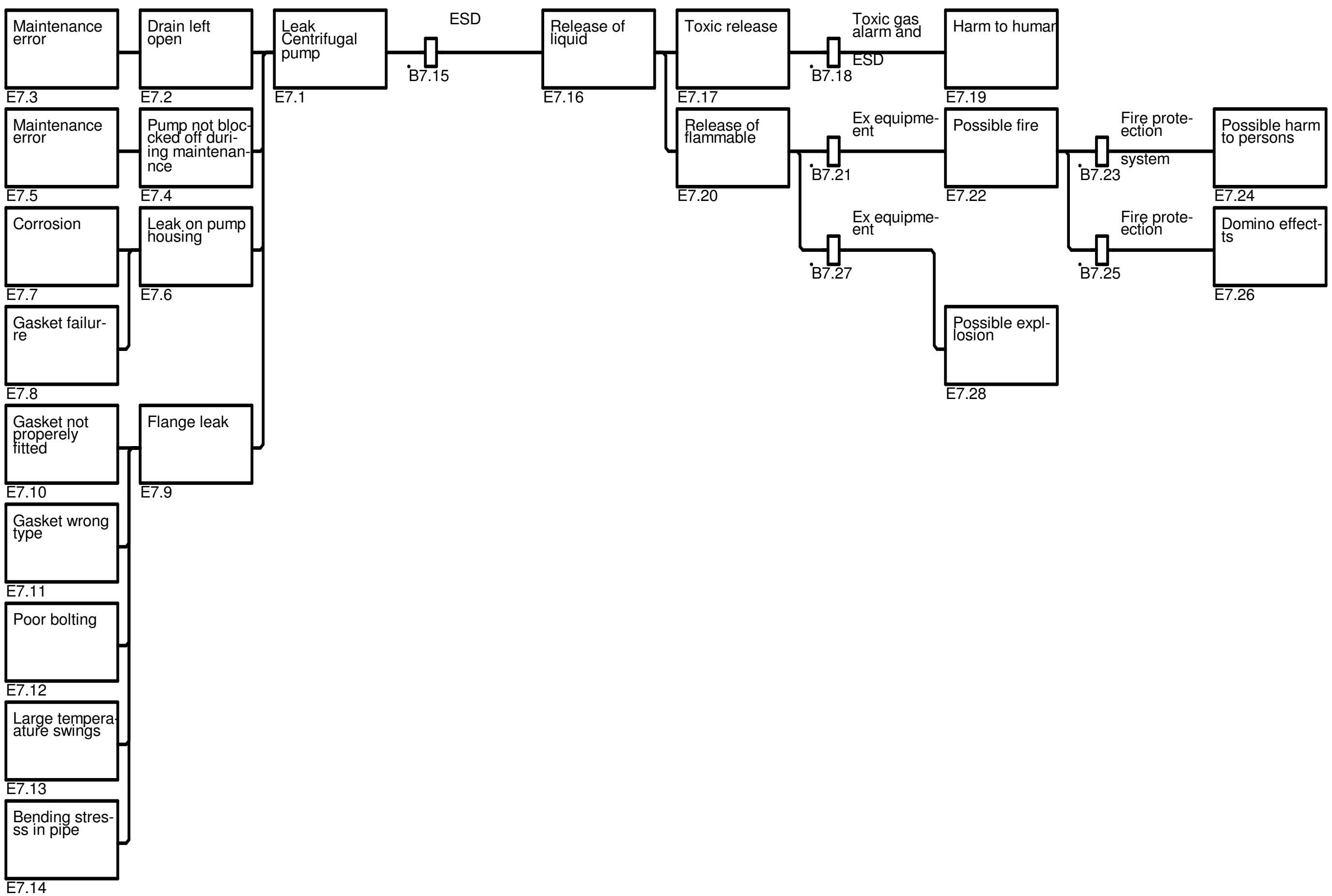


Figure 12.10

8: Safety barrier diagram for Rupture /Centrifugal pump in Centrifugal pump generic release analysi

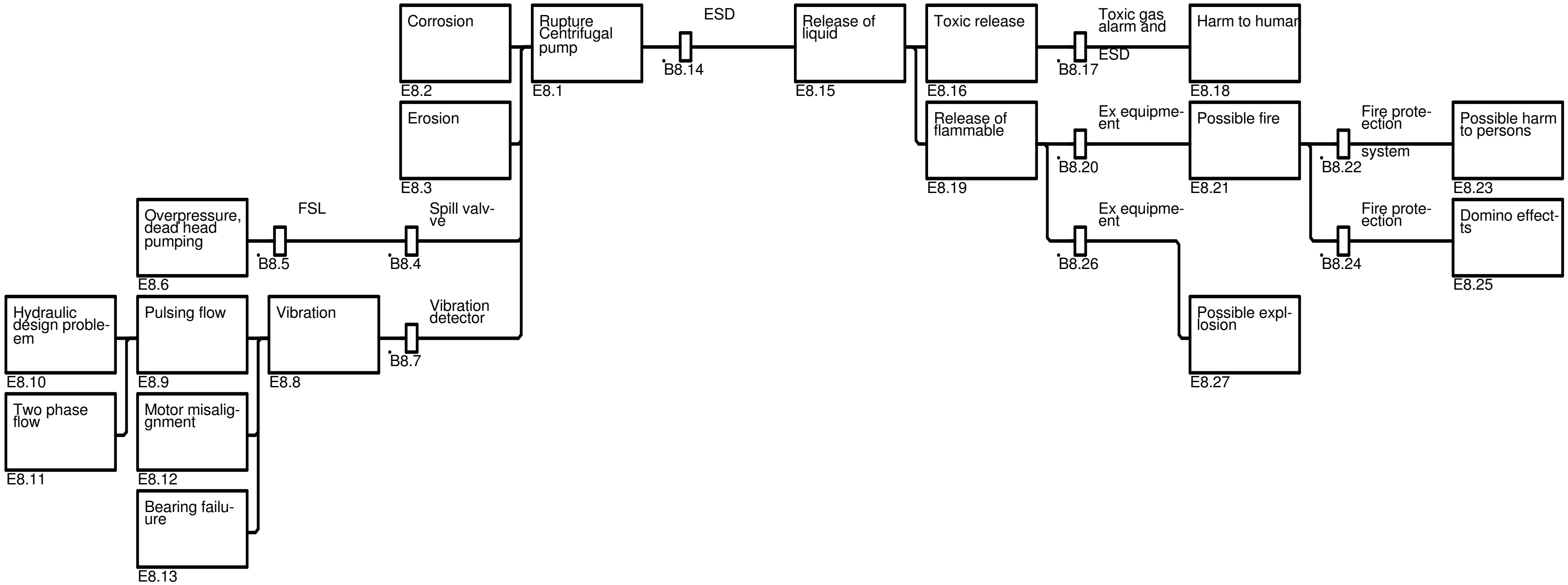


Figure 12.11

Figure 12.12

10: Safety barrier diagram for Reverse flow /Centrifugal pump in Centrifugal pump generic release analysi

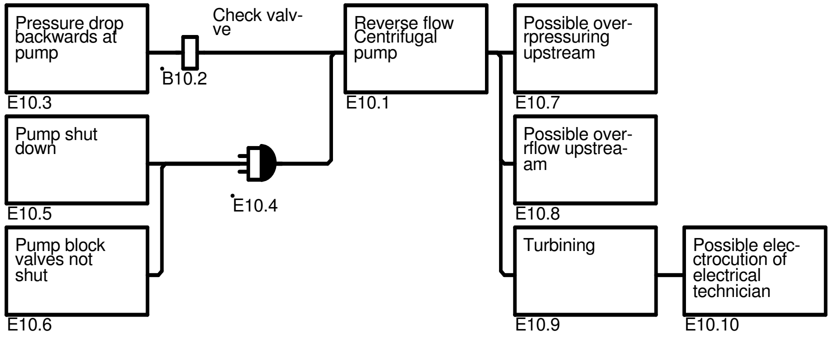


Figure 12.13

12: Safety barrier diagram for Explosion / Centrifugal pump in Centrifugal pump generic release analysi

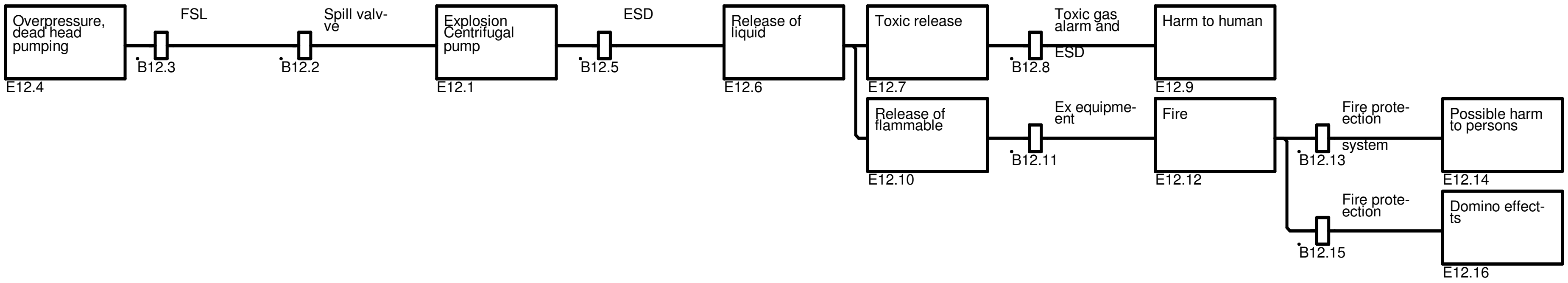


Table 12.7 Detailed release frequency calculation, centrifugal pump

Failure cause	Conse-quence	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
seal leak	explosion	1	1.13E-04	1		0			0			0		1.13E-04	
overheat, dry running	explosion	1	1.13E-04	1		0			0			0		1.13E-04	
procedure error	explosion	1	2.99E-05	1		0			0			0		2.99E-05	
broken pipe	fire	1	2.47E-05	1		0			0			0		2.47E-05	
overheat	fire	1	4.94E-05	1		0			0			0		4.94E-05	
sabotage	fire	1	1.30E-05	1		0			0			0		1.30E-05	
seal leak	fire	1	1.25E-04	1		0			0			0		1.25E-04	
seized bearing	fire	1	1.30E-05	1		0			0			0		1.30E-05	
corrosion on nipple	small	1	1.11E-03	1		0			0			0		1.11E-03	
drain left open	small	1	1.1E-03	1		0			0			0		1.11E-03	
dry running	small	1	1.1E-03	1		0			0			0		1.11E-03	
flange bolts not tightened	small	1	1.1E-03	1		0			0			0		1.11E-03	
gasket failure	small	1	2.1E-03	1		0			0			0		2.12E-03	
plug mounted wrongly	small	1	1.1E-03	1		0			0			0		1.11E-03	
valve leak	small	1	1.1E-03	1		0			0			0		1.11E-03	
seal rupture	rupture	1	1.0E-04	1		0			0			0		1.00E-04	
age	rupture	1	1.0E-04	1		0			0			0		1.00E-04	
dead head pumping	rupture	1	3.8E-04	1		0			0			0		3.80E-04	
wrong material	rupture	1	1.0E-04	1		0			0			0		1.00E-04	
mechanical damage	rupture	1	1.0E-04	1		0			0			0		1.00E-04	
vibration	rupture	1	1.0E-04	1		0			0			0		1.00E-04	
overpressure	rupture	1	2.0E-04	1		0			0			0		2.00E-04	
pipe rupture	rupture	1	1.0E-04	1		0			0			0		1.00E-04	
broken pipe	large	1	3.9E-05	1		0			0			0		3.88E-05	
disconnected, erroneous start	large	1	3.9E-05	1		0			0			0		3.88E-05	
frozen pipe	large	1	6.8E-05	1		0			0			0		6.80E-05	
maintenance, not isolated	large	1	2.0E-05	1		0			0			0		2.04E-05	
drain left open	large	1	2.0E-05	1		0			0			0		2.04E-05	
control failure	large	1	2.0E-05	1		0			0			0		2.04E-05	
left running	large	1	2.0E-05	1		0			0			0		2.04E-05	
release through SV	large	1	2.0E-04	1		0			0			0		2.04E-04	
Total small														8.80E-03	
Total medium														0.00E+00	
Total large														4.31E-04	
Total rupture														1.18E-03	
Total fire														2.25E-04	
Total explosion														2.57E-04	

12.10 Algorithm for determining pump release frequencies

Pump service

#	Question	Action if Yes	Action if No
1	Is the pump in flammable liquids service?	Set susceptibility for fire and explosion to 1 in the detailed analysis Exit	Set susceptibility for fire to 1
2	Is the pump in flammable liquefied gas service ?	Set susceptibility for pump overpressuring explosion to 1 in the detailed analysis Exit	Go to 3
3	Is the pump in acid or highly corrosive service ?	Set susceptibility for corrosion to 1 in the detailed analysis	Go to 4
4	Is the substance pumped unstable at high temperature ?	Set susceptibility for pump explosion to 4 in the detailed analysis	Exit

Table 12.8

Pump construction

#	Question	Action if Yes	Action if No
1	Is the pump vulnerable to overpressuring with dead head pumping?	Set susceptibility for overpressuring to 1 in detailed analysis	goto 3
2	Is there an overpressure SV or spill valve ?	Add spill valve as safety barrier in detailed analysis Go to 3	Go to 3
3	Is there an ESD valve to protect against pump releases ?	Add an ESD valve as safety barrier in detailed analysis Go to 3	Go to 4
4	Is there a possibility of release by back flow ?	Set the susceptibility for reverse flow to 1 Go to 5	Exit
5	Is there a check valve to protect against reverse flow releases ?	Add a check valve as safety barrier in detailed analysis	Exit

Table 12.9

12.11 References

1. Taylor, J.R. Process Safety Engineering, Designing and Building Safer Process Plant, Taylor Associates, 4^{th} Edition 2001

2. Offshore Hydrocarbon Release Statistics, 2001, UK Health and Safety Executive

3. Berezowski

13 Centrifugal and Axial Compressors

Compressors are used to raise the pressure of gas, and come in many different forms and sizes. The main types are reciprocating compressors which use pistons in cylinders to compress gas, with valves to let gas flow in and to let gas out: and centrifugal compressors which use fans to raise the pressure. Centrifugal and axial fan compressors can in turn be divided into two groups, high pressure compressors, which generally have multiple sets of fan blades, in a housing up to 1m in diameter, though generally much smaller; and large diameter single fans, typically used to move large quantities of air or gas into large incinerator or boiler furnaces. Each of these types have different failure characteristics.

13.1 Hazards

Centrifugal and axial compressors for gases such as ammonia or hydrogen can cause major releases from leaking seals. The effects through are usually quite local. Major accidents occur when rotors shed a blade and often, as a result, rupture the casing.

Gases can then be released. In one case investigated, for example, a wing from a large air blower providing air to a sulphur burner in a sulphuric acid plant was shed. Since the fan rotated at 3000 r.p.m., this resulted in the fan rotor running skew, ripping the bearing from the housing, and starting a fire. Sulphur dioxide from the sulphur burner was then released backwards through the fan.

13.2 Release frequencies

US RMP data for compressors could be evaluated for gas treatment plants, with a total of 2 releases in 5 years for 48 plants, giving a release frequency of $42*10^{-4}$ per year, assuming two compressors per plant on average. The release sizes corresponded to a hole size of 300 mm and 13 mm respectively.

OREDA gives a failure rate for external releases for centrifugal compressors of 2.2 per year. These are almost certainly seal leaks – no offshore installation could operate with this frequency of larger releases. The value is based on 142 releases. Of these 8 are regarded as critical (presumably larger), with a failure rate of 0.12 per year. This may be regarded as the frequency for medium size releases.

UK HSE gives a frequency of $9.2*10^{-3}$ releases per year, based on 22 releases, with a distribution of sizes as follows:

< 10 mm	10-25 mm	25-50 mm	NA
73%	14%	5%	9%

Table 13.1 Release frequency distribution

The US RMP and UK HSE data are within a factor of 2.2 of each other, and are regarded as the preferred values. This leads to the baseline data values of table 13.2

< 10 mm	10-25 mm	25-50 mm	Rupture
$6.7*10^{-3}$	$1.3*10^{-3}$	$0.46*10^{-3}$	$0.83*10^{-3}$

Table 13.2 Typical release frequencies per year for a centrifugal compressor

13.3 Release causes

The range of causes for centrifugal compressor releases is given in table 13.3, based on MHIDAS data. The explosions recorded are almost certainly the result of an initial leak or pipe rupture, followed by an ignition in the confined or semi confined space of a compressor house.

Centrifugal compressor releases		
Cause	**#**	**%**
Explosion	9	42.85714
External fire	3	14.28571
Leak	6	28.57143
Pipe rupture	1	4.761905
Compressor rupture	1	4.761905
Seal leak	1	4.761905
	21	100

Table 13.3Centrifugal compressor release causes

13.4 Detailed analysis

While a detailed analysis is possible, the amount of data available does not seem to justify quantification at the detailed causal level. The UK HSE data given in section 13.2 are regarded as baseline data.

13.5References

1. Taylor, J.R. Process Safety Engineering, Designing and Building Safer Process Plant, Taylor Associates, 4th Edition 2001

2. Offshore Hydrocarbon Release Statistics, 2001, UK Health and safety Executive

14 Reciprocating Compressors

Reciprocating compressors are used to raise gas pressure when the pressure requirement is very high or when the desired flow rate is low. The compressor has a number of pistons which move in a cylinder block. Gas is allowed to enter, and is released from the cylinder, by means of special valves.

It is especially important that no liquid enters the cylinders, because they will rupture. For this reason, a knock out drum is usually fitted in the inlet pipe to a compressor.

14.1 Hazards

Reciprocating compressors suffer from shattering accidents relatively frequently, due to liquids entering the cylinder. When the compressor attempts to compress the liquid, the pressure rises to a much higher level than for the gas, and usually either the cylinder head is blown off, or a hole is punched in the cylinder itself. Figure 14.1 shows a compressor shattered in a similar way to this, due to a cylinder head bolt becoming loose and being trapped between the cylinder and cylinder head.

Such accidents do not usually lead to off site consequences, because flow to and from compressors can generally be shut off by emergency shut down valves.

Reciprocating compressors can generate very high pressures. If they are started with blocked discharge valves, or if the discharge valve is closed during operation, the piping will usually be ruptured.

Reciprocating compressors require care in lubrication. Leaks of lubricating oil are a fairly frequent cause of fires. Additionally, release of gas to the compressor sump can occur when piston rings fail, with a release of gas as a result. Special ventilation systems are required, if the gases are flammable or toxic.

Reciprocating compressors generate heavy vibration. Piping at the compressors is vulnerable to fatigue cracking and rupture, unless the piping is very carefully supported. Both gas piping and lubrication oil piping represent fatigue cracking hazards.

14.2 Frequency of releases

US RMP data for reciprocating compressors could be evaluated for chlorine plants, with a total of 4 significant releases in 5 years for 163 plants, giving a release frequency of $1.6*10^{-3}$ per year, assuming an average of 3 compressors per plant. The release sizes corresponded to a hole size of 100 to 300 mm and 13 mm respectively.

OREDA gives a failure rate for critical external releases for reciprocating compressors of 2.6 per year. The value is based on 34 releases. These must be small seal leaks, since this rate would otherwise be operationally unacceptable.

UK HSE gives a frequency of $6.5*10^{-2}$ releases per year, based on 5 releases, with a distribution of sizes as follows:

< 10 mm	10-25 mm	25-50 mm	NA
81%	19%	-	-

Table 14.1 Distribution of hole sizes for reciprocating compressors

All three sets of data vary widely in their values. This probably reflects the actual variation between different compressor applications.

For ruptures, direct experience is given in table 14.2.

Company	#Compressors	Years	Cmp. Years	Ruptures	Fires
U	6	24	144	2	4
U2	2	10	20	0	0
V	1	2	2	1	0
W	4	30	120	1	2
X	6	24	144	0	0
			430	4	6
			Frequency	0.009302	0.013953

Table 14.2

These values lead to the reference release frequencies of table 14.3

< 10 mm	10-25 mm	25-50 mm	Rupture	Fire
$5.3*10^{-2}$	$1.2*10^{-2}$	-	$9.3*10^{-3}$	$1.4*10^{-2}$

Table 14.3 Typical release frequencies per year for a centrifugal compressor

14.3 Causes of reciprocating compressor releases

From personal experience, the most important issues determining the accident rate for reciprocating compressors are:

1. The quality of level control and alarm instrumentation on the knock out drum, and the freedom from blockage of the drum liquids drain.

2. The provision of gas detectors and rapid shutdown in the case of gas leaks.

The cause distribution derived from MHIDAS is given in table 14.3.1. It is less informative than might be desired. The “explosions” are almost certainly the result of gas leaks which ignite. The “compressor ruptures” are almost certainly the result of liquid entering the compressor. The pipe and valve ruptures are almost certainly the result of blockages or valve closures in the discharge piping.

Reciprocating compressor releases		
	#	%
Explosion	6	18.2
External fire	2	6.1
Leak	8	24.2
Pipe rupture	6	18.2
Compressor rupture	6	18.2
Seal leak	3	9.1
SV release	1	3.0
Valve rupture	1	3.0

Table 14.4

Figure 14.1 Ruptured horizontal compressor

14.4 Algorithm for reciprocating compressor Release frequencies

Protection

#	Question	Action if Yes	Action if No
1	Is there a knock out drum ?	Go to 3	Go to 2
2	Can there be liquid or condensation in the gas flow ?	Multiply the rupture frequency by 30.	Go to 4
3	Is there a reliable high level alarm on the KO drum ?	Add the high level alarm as a safety barrier in the detailed analysis.	Go to 4
4	Is the gas flammable ?	Go to 5	Exit
5	Is there a reliable gas alarm and ESD system ?	Add the gas alarm system as a safety barrier in the detailed analysis. Exit	Exit

Table 14.4

15 Chemical Batch Reactors

Batch reactors are used in fine chemical plants, typically those producing pharmaceuticals, pesticide, special resins and adhesives, for example. Batch reactors are a significant sources of releases and explosions. They generally have a relatively short consequence range, but on the other hand, fire chemicals plant are often situated in cities, or quite close to dwellings.

A typical batch reactor consists of a vessel with a bolted on cover, a jacket welded onto the outside to provide heating and/or cooling, an agitator (stirrer) to provide mixing, piping to supply liquids, and piping to take away products. Additionally, there will be a reflux or distillation column integrated into the reactor, and provision for adding solids trough a man hole or through a silo and sluice pipe. Sometimes the heating/cooling jacket is replaced by an internal heating or cooling coil. Where one of the reactants is a gas, this is fed into the reactor generally through a sparger i.e. a pipe with holes in it, at the base of the reactor.

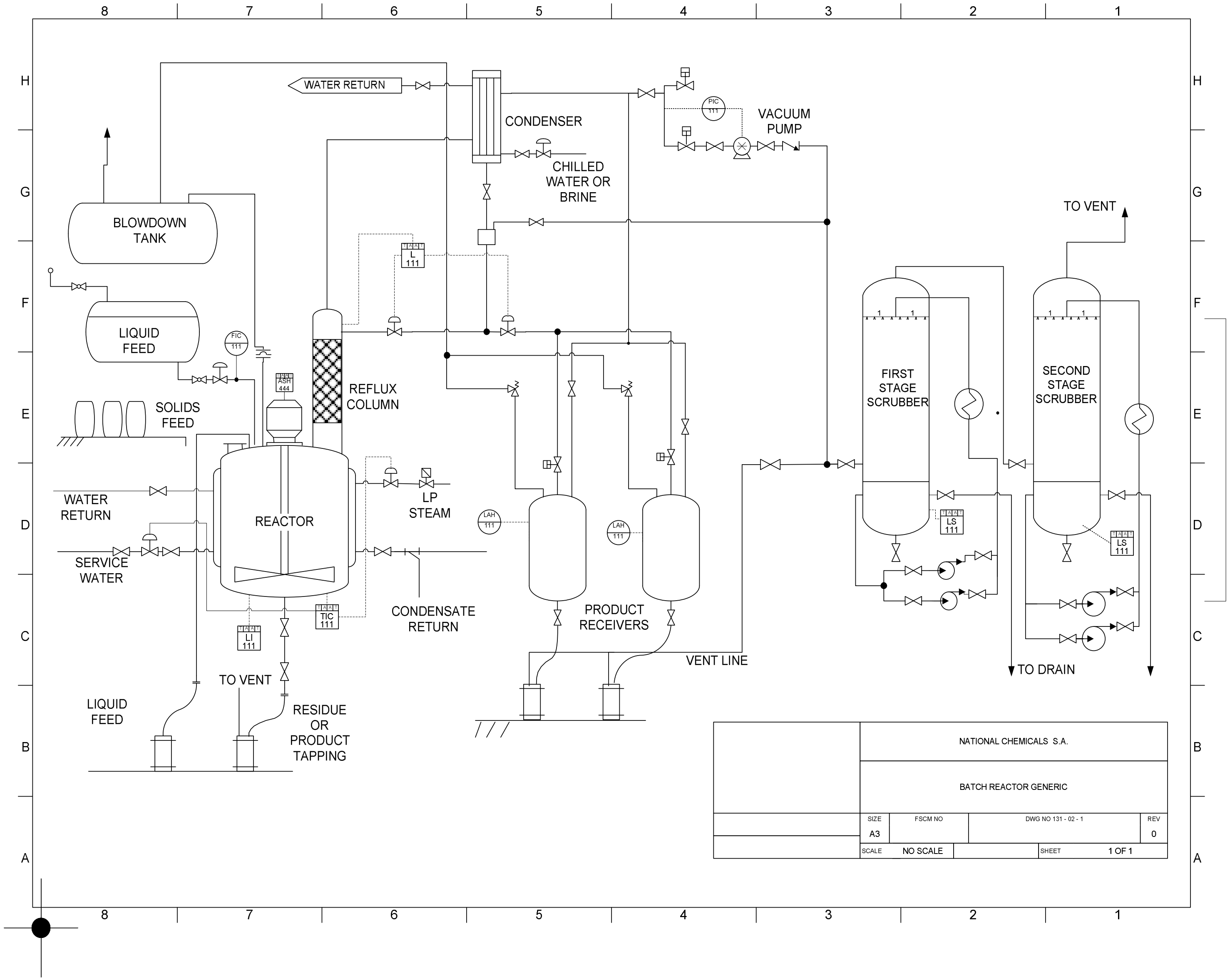
15.1 Reactor accident frequencies

It is very difficult to derive data for chemical batch and semi batch reactor release and explosion frequencies because the processes vary so widely chemically. Only a few have a potential for runaway reactions which cause explosions. More can cause reactions which give vapour or gas releases through burst discs and blow down systems.

Table 13.1 shows data for a number of reactor accidents for which the reactor population is known, and for which the reaction type and accident descriptions were available. US RMP data have not been used here, because the reactor population could not be determined, but accident cause distributions are given in table 13.3 (from the author’s own investigations).

Company	Years	Reactors	Reactor years	Explos-ions	Serious fires	Explosion frequency per 10^4 reactor year	Fire frequency per 10^4 reactor year
J	4	7	28	1	0	357	0
K	15	5	75	5	0	667	0
L	20	8	160	1	0	62.5	0
M	10	10	100	1	1	100	100
N	20	12	240	2	0	83.3	0
P	24	6	144	3	1	208	69
C	15	6	90	0	0	0	0
Q	40	22	880	14	5	159	56.8
	4	20	80	0		0	
R	24	10	240	0	0	0	0
S	6	20	120	0	0	0	0
Total			2157	27	7	125	32.5

Table 15.1 Frequency of batch reactor fires and explosions, reactors with exothermic reactions only.


WATER RETURN
CONDENSER
CHILLED
WATER OR
BRINE
VACUUM
PUMP
PIC
111
TO VENT
BLOWDOWN
TANK
L
111
LIQUID
FEED
FIC
111
ASH
444
REFLUX
COLUMN
SOLIDS
FEED
FIRST
STAGE
SCRUBBER
SECOND
STAGE
SCRUBBER
WATER
RETURN
REACTOR
LP
STEAM
LAH
111
SERVICE
WATER
LS
111
TIC
111
LI
111
CONDENSATE
RETURN
PRODUCT
RECEIVERS
VENT LINE
TO DRAIN
TO VENT
LIQUID
FEED
RESIDUE
OR
PRODUCT
TAPPING
NATIONAL CHEMICALS S.A.
BATCH REACTOR GENERIC
SIZE
A3
FSCM NO
DWG NO 131 - 02 - 1
REV
0
SCALE
NO SCALE
SHEET
1 OF 1

Marrs and Lees investigated reactor accidents in UK for 1970 to 1981, with 66 incidents, in 2100 reactors, giving an incident rate of $2.6*10^{-3}$ per reactor year (ref. 13.1), which is about one fifth of that recorded by the author. The difference can be quite readily be explained by differences in the population of reactor types and processes. Marrs and Lees investigated all reactors, not just those with exothermic processes.

Most exotherm incidents will not lead to accidents, either because emergency cooling is effective, or because pressure relief provisions prevent explosion. (Not all reactions can be vented though). Marrs and Lees estimated the frequency of runaway to be $5*10^{-2}$ thermal excursions per year. The probability of failure of venting and pressure relief is estimated to be 0.05, with the distribution of causes as in table 13.2.

A. Behaviour of relief system	**No of cases**
Burst disc operated	8
Relief valve operated	2
Relief fitted , but failed	13
Relief not, or probably not, fitted	25
Total	48
B. Relief systems fitted but failed	
Relief valve, vent part closed	1
Relief valve failed (inc 4 know to be underdimensioned)	7
Burst disc failed to rupture	1
Burst disc too small	1
Burst disc, details unknown	1
Relief valve and burst disc failed	1
Relief valve and burst disc failed, known to be large	1
Total	13
C. Burst disc failure	
Poor reliability	1
Inadequate capacity	12

Table 15.2 Failure of reactor relief systems

Runaway incident	No. cases	%
Vessel open, hazardous release	18	27.3
Glasswork shattered, hazardous release	16	24.2
Vessel ruptured, hazardous release	19	28.8
Vessel ruptured only	1	1.5
Explosion	5	7.6
Hazardous release	5	7.6
Catch pot ruptured	1	1.5
Catch pot fire	1	1.5
Total	**66**	**100**

Table 15.3 Thermal runaway incident consequences, (ref 15.1)

By far the most common accident types for chemical batch reactors is thermal runaway, which as seen above. From table 13. it can be seen that explosions occur roughly once per 100 reactor year, with a variations from 8 per 1000 to 7 per 100 years, i.e. by an order of magnitude. If the highly energetic reactions (those with an exotherm of more than 100 kJ per kg) the explosion frequency rises to one per 12 years per reactor, based on 25 runaways (i.e. the frequency of explosion is determined almost solely by the most energetic reactions.)

The other major group of accident types is fire. These can occur where flammable reagents or solvents are used, and this includes most chemical batch reactors. Frequencies for accidents such as these are even more difficult to determine, because reporting is quite incomplete. This is natural because many of the fires have purely local effects. The frequency is, from table 13.1, about 3 per 1000 reactor years.

Some chemical batch reactions are made in the gas phase, by “sparging” bubbles of gas into the liquid. Others actually generate gases. These systems can result in gas releases if the gas admission rate is too high or the vent scrubber becomes blocked. The quantities can be quite significant in a risk analysis context, especially if liquefied gas overflows from an evaporator into the reactor.

Figure 15.2 Explosion results from dry distillation of solvent and residue

Figure 15.3 Explosion results from dry distillation of solvent and residue

	Number	(%)
Mischarging of reactants or catalysts	32	26
Little or no study of the reaction chemistry / thermochemistry	18	15
Inadequate temperature control	17	14
Inadequate maintenance	9	7
Inadequate agitation	16	13
Raw material quality	12	10
Operator error	5	4
Other	13	11
The underlying causes were identified as: • inadequate understanding of the reaction chemistry/thermochernistry leading to badly designed plant. • under-rated control and safety back-up systems. • inadequate operational procedures, including training.		

Table 15.4 Causes of reactor accidents according to Nolan and Barton

Code	Cause	MCA	All
HO	Not known	3	
H1	Change of pressure	2	1
H2	Change of temperature	14	19
H3	Electrical energy	1	1
H4	Breach/opening of boundary	4	3
H5	Sparks	2	1
H6	Wrong substances mixed	26	19
H7	Correct subs. but incorrect mixing conditions	21	19
H8	Insufficient mixing/stirrer stopped	14	13
H9	Impure or contaminated chemicals used	3	6
H10	Contaminated vessel used	9	7
H11	Stray catalyst	5	6
H12	Hot spot	4	6
H13	Accumulation of reactants or intermediates	8	9
H14	Leak into system	5	6
H15	Other initiation mechanism	1	2

Table 15.5 Causes of batch reactor accidents, Rasmussen

The causes of accidents in batch reactors have been well researched. Table 13.3 gives the causal distribution derived from examples in the 1970´s and 1980´s by Rasmussen (ref. 15.2)

A more extensive list of possible reactor accident causes is:

- Lack of knowledge of reaction thermodynamics
- Lack of knowledge of side reactions
- Unexpected side reaction
- Impurities
- Stray catalysts
- Overdosing of a reactant
- Double charging of reactant
- Distillation to dryness
- Wrong substance added
- Temperature control error
- Timing error and batch cooled too long
- Hot spots
- Lack of agitation, too low cooling
- Too rapid dosing
- Failure to “ignite”
- Failure of mixing r agitation
- Adding too little
- Adding too late
- Vapour explosion
- Dust explosion
- Nitrogen asphyxiation
- Manhole cover leaks
- Manhole cover blown off
- Blockage of outlet vent
- Reflux cooling low temperature
- Entry of high boiling products to reflux
- Steam explosion
- Adding liquids to a hot reactor
- Open drain valve
- Emptying valve sticks open
- Overflow
- Overpressuring

An extensive study of reactive chemical accidents was undertaken by the US Chemical Safety and Hazard Investigation Board in 2001-2002 (Report No. 2001-01-H). This gives some interesting information from 167 incidents from 1980 to 2002. Among other things, the number of fatalities was determined, in all 108 fatalities occurred. There were 12 accidents with 3 or more fatalities. The largest number was 17 fatalities at ARCO Chemical, Channel View, Texas in 1990. The breakdown of consequences was as in table 13.6

Consequence type	%
Hazardous liquids spill	5
Fire/explosion and toxic release	16
Toxic gas release	37
Fire / explosion	42

Table 13.6 Hazardous reaction consequences in 166 incidents

Nearly 50 of the accidents affected the public, with one public fatality.

The chemicals involved were as shown in table 15.7:

Chemical class	No. of incidents	%
Acid	38	16.7
Oxidiser	20	8.8
Monomer	15	6.6
Water	14	6.2
Base	12	5.3
Organic peroxide	12	5.3
Hypochlorite	10	4.4
Alcohol	8	3.5
Hydrocarbon	7	3.1
Inorganic/Metal	6	2.6
Hydrosulphite	6	2.6
Other	79	34.8

Table 15.7Chemicals involved in 167 hazardous reactions

Reaction type	%
Decomposition	26
Acid/base	11
Water reactive	10
Polymerisation	10
Oxidation	6
Decomposition initiated by other reaction	5
Oxidation/reduction	4
Chlorination	1
Undetermined	23

Table 15.8 Reaction types involved in 167 hazardous reactions

Not only reactors are involved in hazardous reactions. Table 15.9 shows the equipments involved

Equipment type	%
Reactor	25
Storage equipment	22
Waste handling	3
Transfer equipment	5
Separation equipment	5
Storage drum	10
Other	22
Unknown	8

Table 15.9. Equipment involves for the 167 hazardous reaction incidents.

The causes recorded for the hazardous reactions were as given in table 15.10

Reaction cause	No. of incidents	%
Inadequate hazard identification	9	24.3
Inadequate hazard evaluation	16	43.2
Inadequate storage and handling procedures	17	45.9
Inadequate training for storage and handling	10	27.0
Inadequate management of change	6	16.2
Inadequate process design for reactive hazards	6	16.2
Inadequate design to prevent human error	9	24.3
Inadequate company wide communication of hazards	5	13.5
Inadequate emergency relief design	3	8.1
Inadequate safe operating limits	3	8.1
Inadequate near miss/ incident investigation	2	5.4
Inadequate inspection/maintenance/monitoring of safety critical devices	2	5.4
Previously unknown reactive hazards	1	2.7

Table 15.10 causes of 37 hazardous reactions.

15.2 Protective measures

For modern reactors, or ones which have been back fitted with up to date safety systems, the actual accident rates should be much lower than those shown in table 15.1. A typical set of safety devices is:

- a properly dimensioned burst disc and blow down system
- a blow down tank
- a highly reliable temperature and temperature rate of rise alarm and shut down system, typically designed to SIL 3 (following standard IEC 61508/IEC61511)
- a cooling system with adequate capacity or emergency cooling provisions
- agitator function monitoring and shut down also to SIL 3
- wherever possible, use of the semi batch feed concept
- use of a full set of interlocks to prevent errors in sequence control (see ref.15.3)

In the newest reactors, accurate weigh cells are used to check that the amounts of substances added are correct.

15.3 Typical release frequency

A typical release frequency from chemical batch reactors is taken from table 15.11.The basis for the value is a kettle type batch reactor with a reflux column, and a minimum of safety devices. The values are as follows:

Accident type	Frequency per year
Large spill (solvent or reagent handled in barrels)	0.08
Release from bottom valve	0.006
Fire	$32.4*10^{-4}$
Explosion	$125*10^{-4}$
Typical, large spill	0.01

Table 15.11 Typical frequencies for reactor accidents

As a typical frequency for explosions in batch reactors, a value for one specific plant with a low release frequency (source T) was taken.

15.4 Assessment of causal factors and susceptibilities

Susceptibilities to different causes of releases are given in table 15.12

Release frequencies	Small	Medium	Large	Rupture
Typical	6.00E-02	1.00E-02	0.01	1.25E-04

No	Failure cause	% of releases	Source	Release size	Suscept-ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Bottom valve opened	2.2	TA	large	0.5	1.0E+00	4.6E-03	Inspection of
2	Piping failure	1.6	TA	small	1	1.0E+00	2.1E-02	10 fine chemicals plants
3	Piping failure	0.4	TA	large	1	1.0E+00	4.2E-04	plants, 64 reactors
4	Hose failure	4.0	TA	large	1	1.0E+00	4.2E-03	
5	Leak	3.0	Rasmussen	small	1	3.0E-01	1.3E-01	
6	Fire	3.0	TA	large	1	1.0E+00	3.1E-03	
7	Overheating	10.6	Rasmussen	explosion	0.01	1.0E+00	1.5E-03	
8	Overpressure, control failure	1.5	Rasmussen	explosion	0.003	1.0E+00	7.4E-04	
9	Inward leak	3.8	Rasmussen	explosion	0.03	1.0E+00	1.8E-04	
10	Accumulation of reactants	6.1	Rasmussen	explosion	0.3	5.0E-01	5.9E-05	
11	Hot spot	3.0	Rasmussen	explosion	0.2	1.0E+00	2.2E-05	
12	Stray catalyst	3.8	Rasmussen	explosion	0.2	1.0E+00	2.8E-05	
13	Contamination	9.1	Rasmussen	explosion	0.5	1.0E+00	2.7E-05	
14	Wrong substance	19.7	Rasmussen	explosion	0.2	1.0E+00	1.4E-04	
15	Wrong mixing conditions, double charging	15.9	Rasmussen	explosion	0.2	1.0E+00	1.2E-04	
16	Internal explosion, ign	1.5	Rasmussen	explosion	0.5	1.0E+00	4.4E-06	
17	Stirrer failure, stopped	10.6	Rasmussen	explosion	0.8	2.0E-02	9.7E-04	
		99.86667						

Table 15.12 Batch reactor susceptibilities

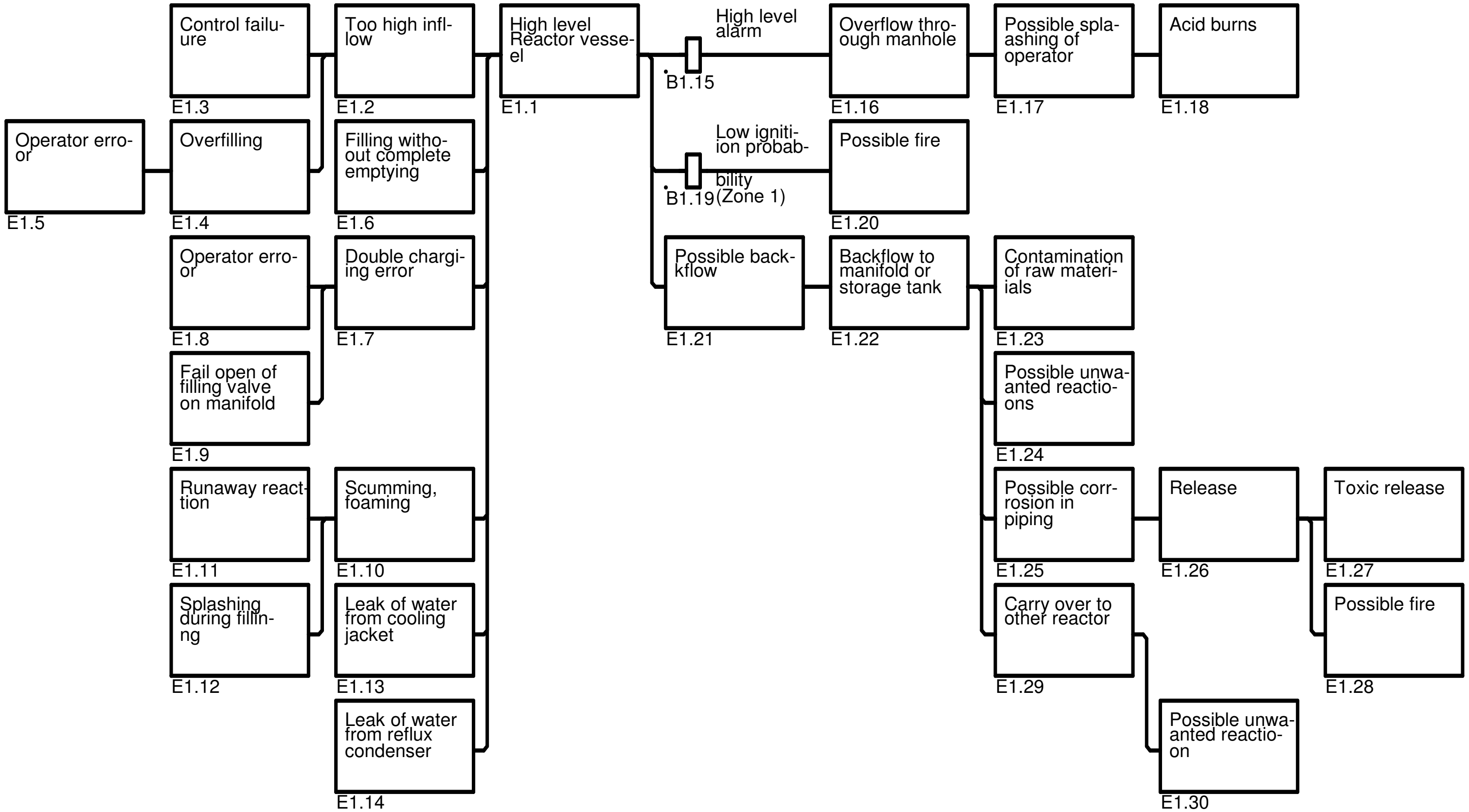
The actual release frequencies for batch reactors are almost all “unusual” – the reactors themselves have virtually no significant failure modes, being for the most part unpressurised or low pressure vessels with a large safety factor. Also, the reactors used for establishing the typical release frequencies are almost all susceptible to all the failure causes. The algorithms for modifying failure rates are therefore mostly concerned with eliminating a certain percentage of the overall failure rate, for those failure types which are impossible for the particular reactor.

15.5 Detailed analysis

A detailed analysis is given here in diagrams 15.3 ff., showing the range of causes for incidents and the types of safety measures which can be used. The list is not necessarily complete, and the reader may find other causes or safety barriers. The model should nevertheless give a good basis.

Table 15.13 gives a summary frequency evaluation, using the baseline release frequencies, the distribution of causes, assumptions about the degree of protection for the plants in the typical accident frequency study, and assumptions about susceptibility to the individual accident types for the typical frequency study.

1: Safety barrier diagram for High level /Reactor vessel in Batch reactor generic analysis



2: Safety barrier diagram for High level /Reactor vessel in Batch reactor generic analysis

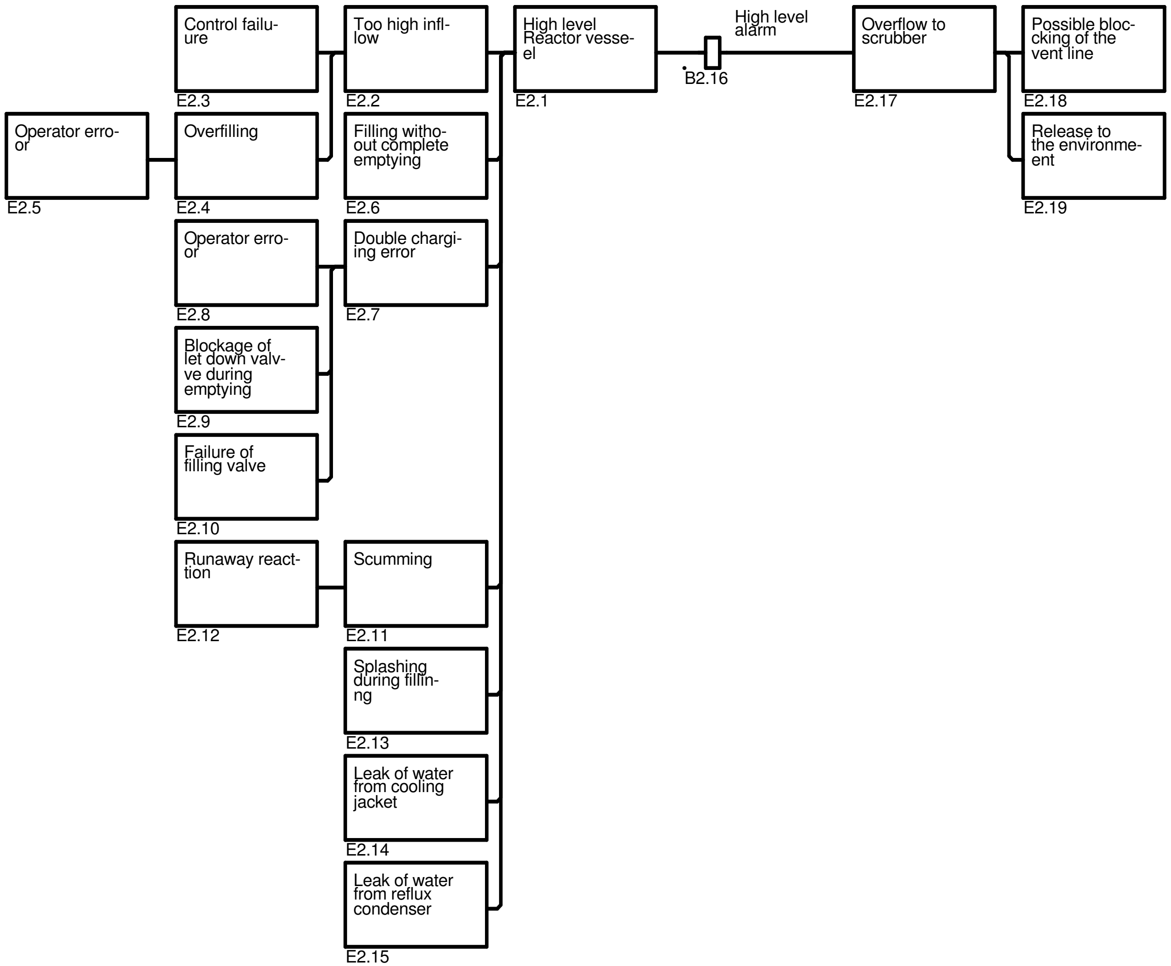


Figure 15.4

3: Safety barrier diagram for Level too low /Reactor vessel in Batch reactor generic analysis

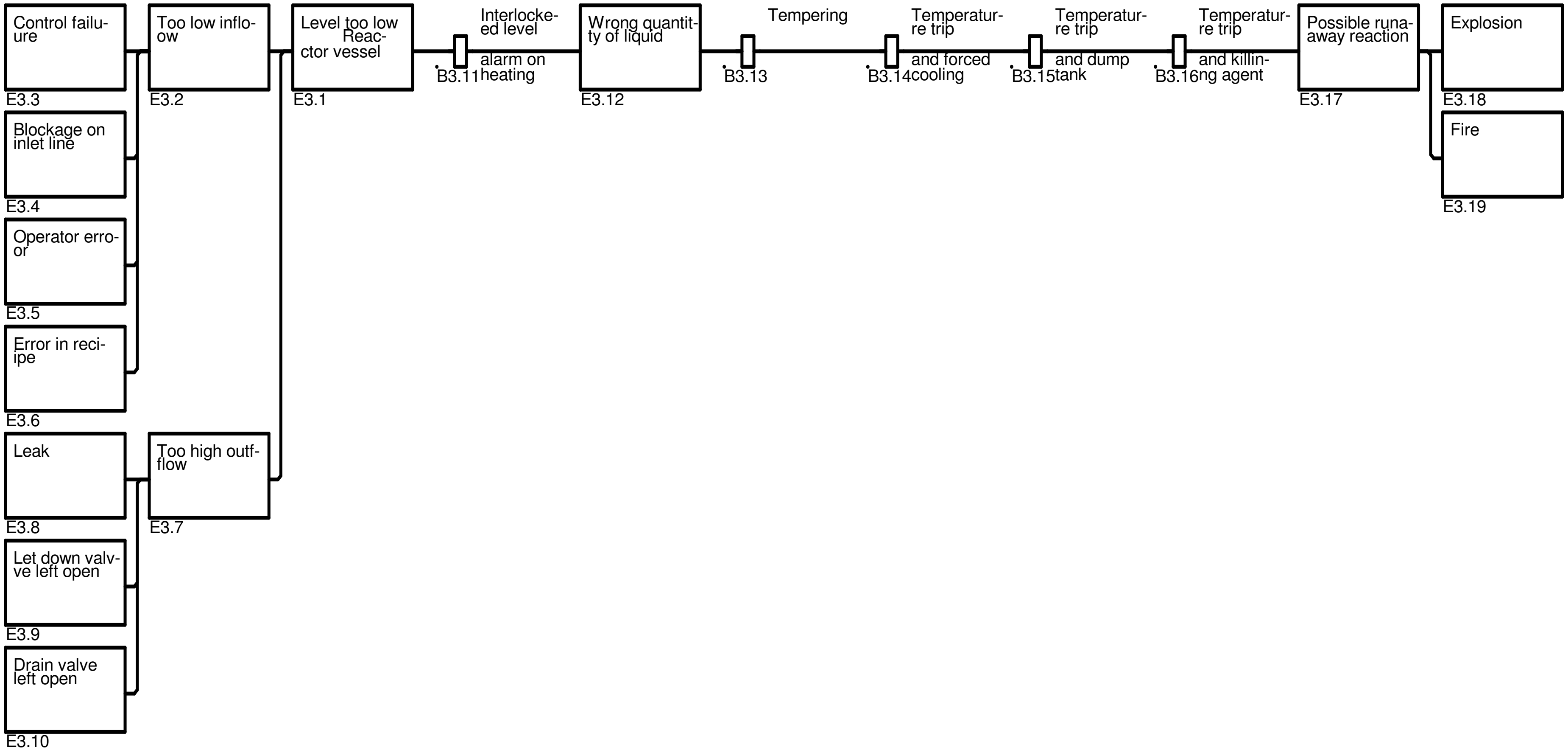


Figure 15.5

4: Safety barrier diagram for High temperature /Reactor vessel in Batch reactor generic analysis

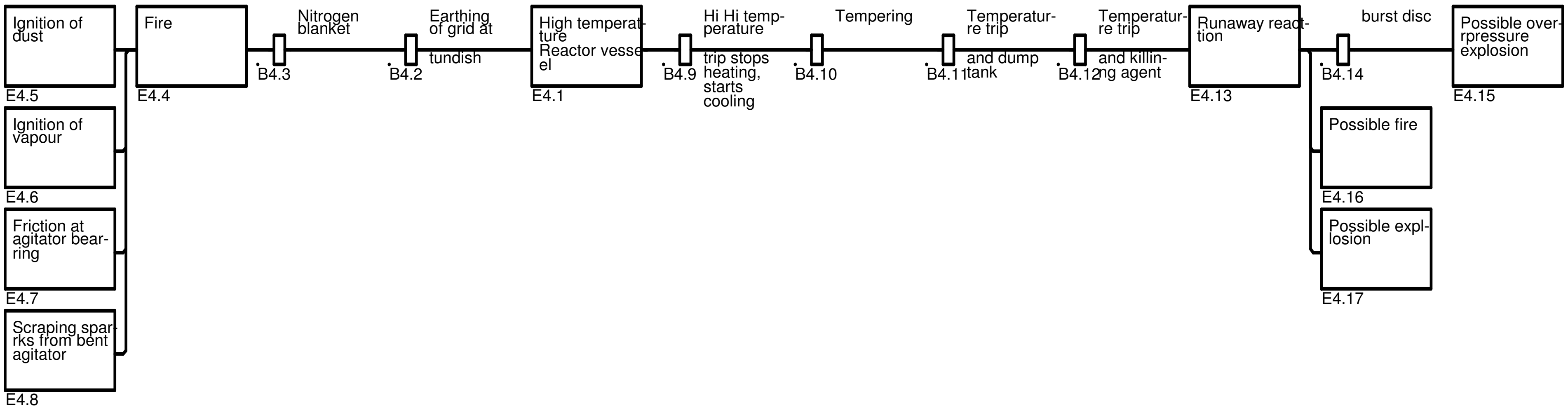


Figure 15.6

5: Safety barrier diagram for High temperature /Reactor vessel in Batch reactor generic analysis

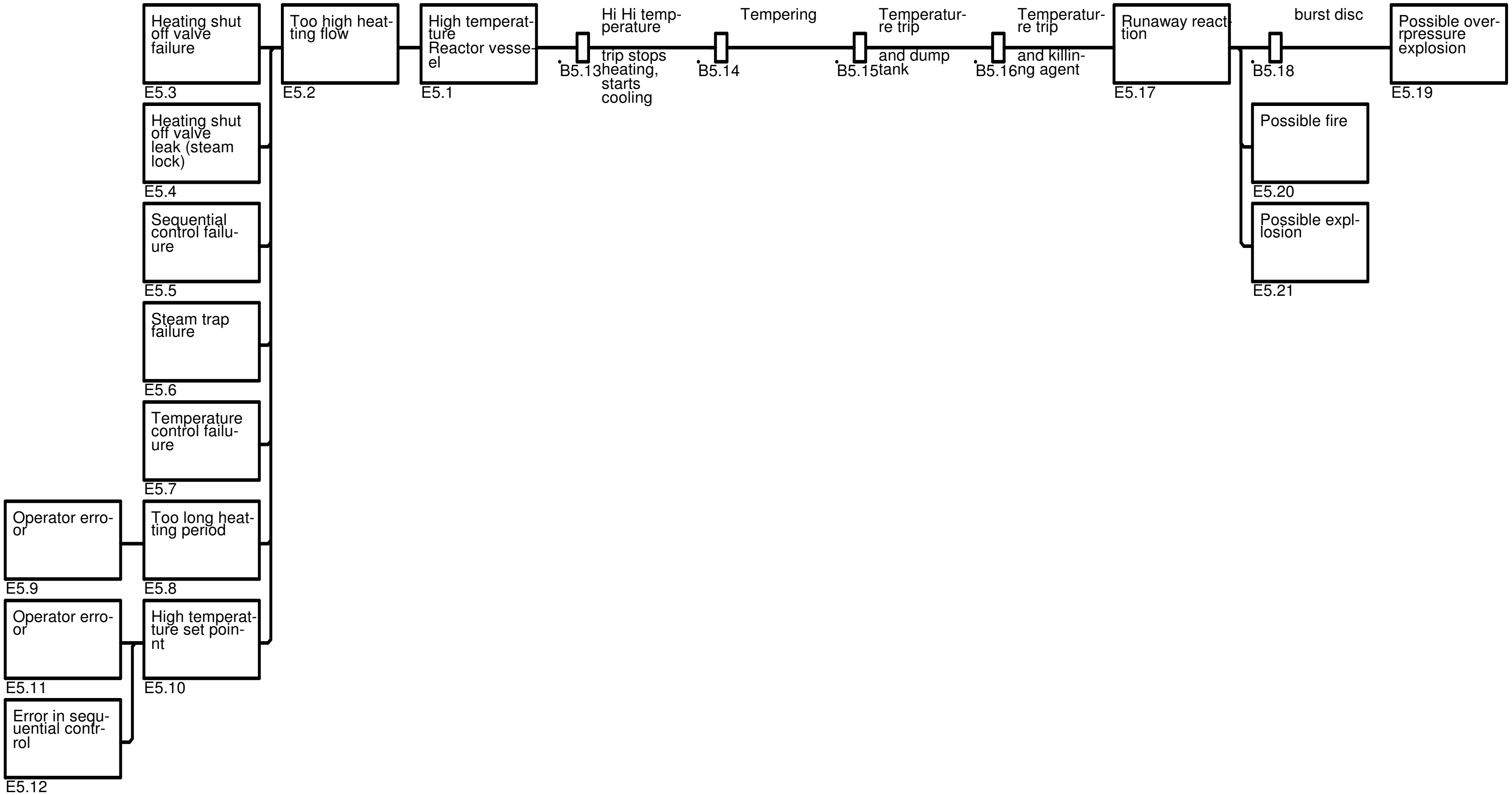


Figure 15.7

6: Safety barrier diagram for High temperature /Reactor vessel in Batch reactor generic analysis

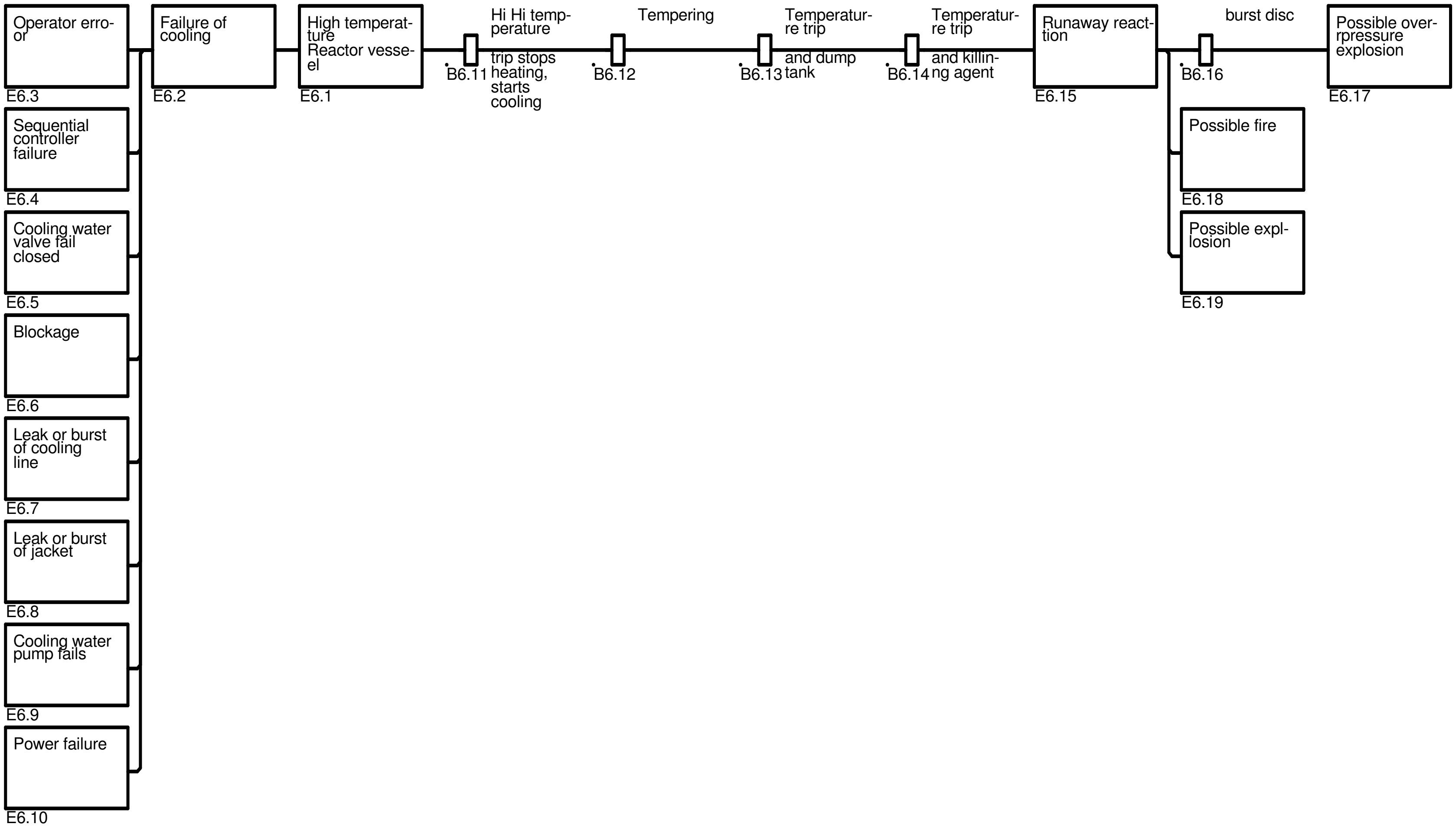


Figure 15.8

7: Safety barrier diagram for Runaway reaction /Reactor vessel in Batch reactor generic analysis

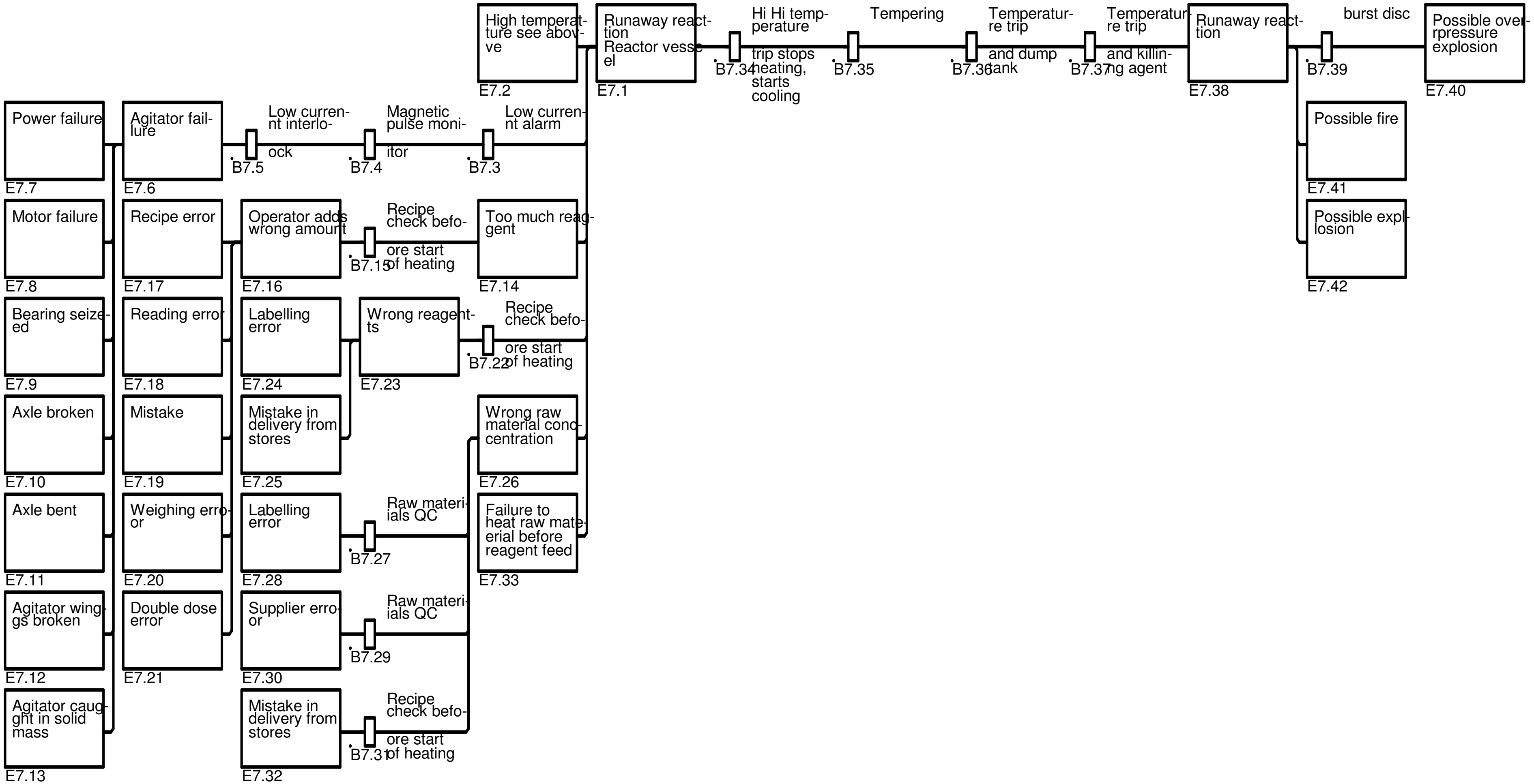


Figure 15.9

8: Safety barrier diagram for Pressure too high /Batch reactor in Batch reactor generic analysis

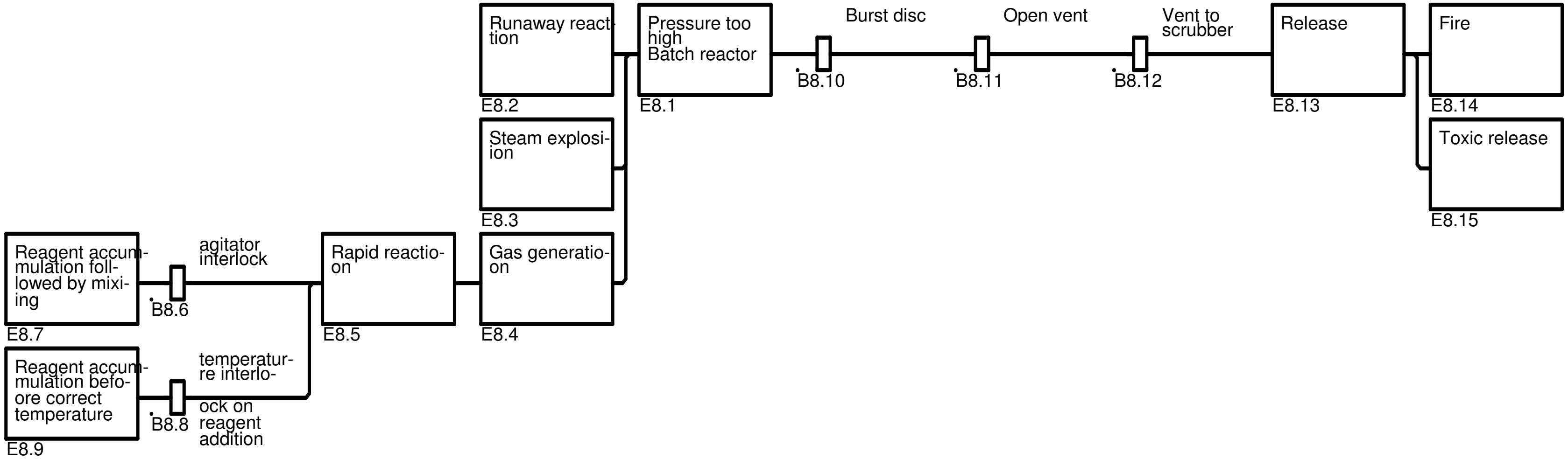


Figure 15.10

9: Safety barrier diagram for Pressure too low /Reactor vessel in Batch reactor generic analysis

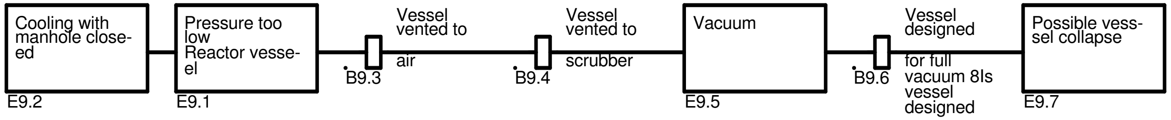


Figure 15.11

10: Safety barrier diagram for Wrong concentration /Reactor vessel in Batch reactor generic analysis

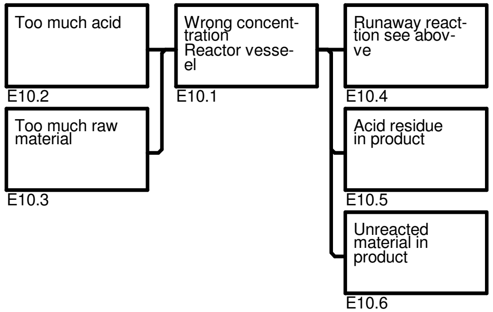


Figure 15.12

11: Safety barrier diagram for Breach of boundary /Reactor vessel in Batch reactor generic analysis

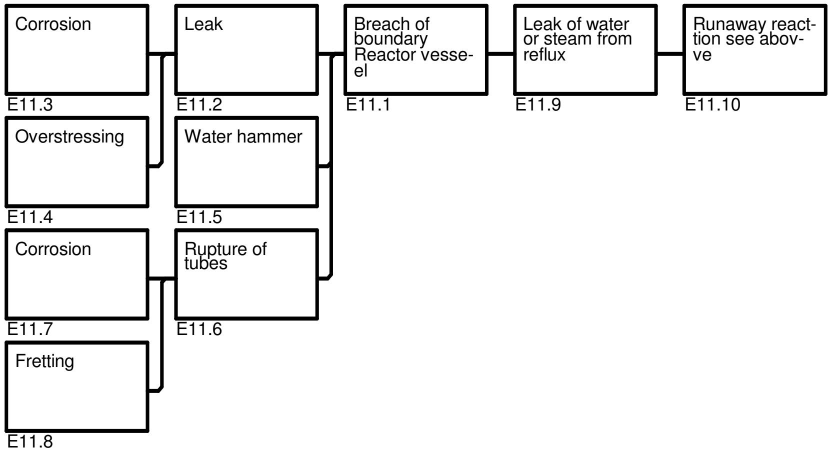


Figure 15.13

12: Safety barrier diagram for Breach of boundary /Reactor vessel in Batch reactor generic analysis

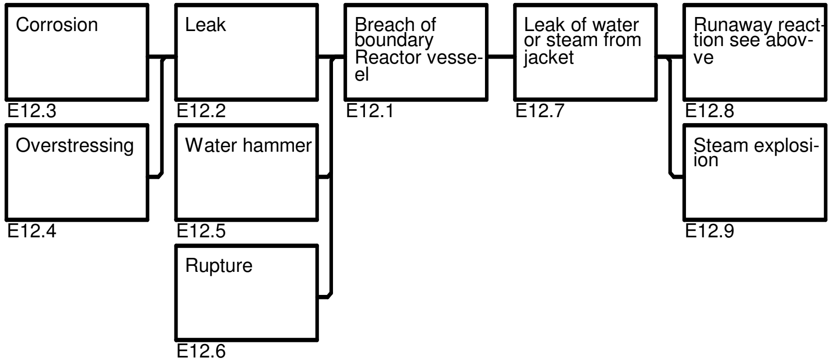


Figure 15.14

13: Safety barrier diagram for Breach of boundary /Reactor vessel in Batch reactor generic analysis

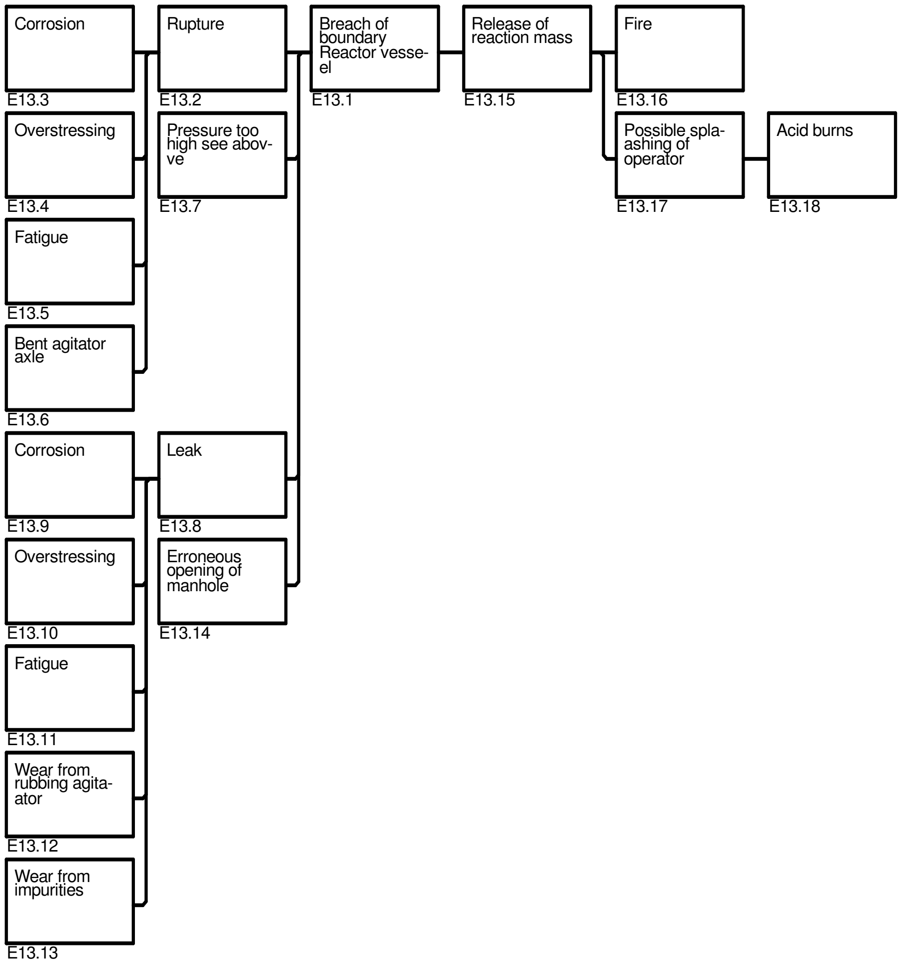


Figure 15.15

Release frequencies	Small	Medium	Large	Rupture	Fire	Explosion
Batch reactor	1.00E-02	5.00E-03	0.00E+00	0.00E+00	3.24E-03	2.00E-05

Failure cause	Conse-quence	Number of items or metres	Base frequency per year	Suscept-ibility	Risk contribution %	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year
Bottom valve opened	large	1	4.57E-03	1			0	1		0	1		0	1	4.57E-03
Piping failure	small	1	2.07E-02	1			0	1		0	1		0	1	2.07E-02
Piping failure	large	1	4.15E-04	1			0	1		0	1		0	1	4.15E-04
Hose failure	large	1	4.15E-03	1			0	1		0	1		0	1	4.15E-03
Leak	small	1	1.31E-01	1			0	1		0	1		0	1	1.31E-01
Fire	large	1	3.15E-03	1			0	1		0	1		0	1	3.15E-03
Overheating	explosion	4	1.55E-03	1			0	1		0	1		0	1	6.19E-03
Overpressure, control failure	explosion	10	7.37E-04	1			0	1		0	1		0	1	7.37E-03
Inward leak	explosion	30	1.84E-04	1			0	1		0	1		0	1	5.53E-03
Accumulation of reactants	explosion	10	5.90E-05	1			0	1		0	1		0	1	5.90E-04
Hot spot	explosion	6	2.21E-05	1			0	1		0	1		0	1	1.33E-04
Stray catalyst	explosion	2	2.77E-05	1			0	1		0	1		0	1	5.53E-05
Contamination	explosion	4	2.65E-05	1			0	1		0	1		0	1	1.06E-04
Wrong substance	explosion	1	0.000144	1			0	1		0	1		0	1	1.44E-04
Wrong mixing conditions, double charging	explosion	1	0.000116	1			0	1		0	1		0	1	1.16E-04
Internal explosion, ign	explosion	1	4.42E-06	1			0	1		0	1		0	1	4.42E-06
Stirrer failure, stopped	explosion	1	0.000968	1			0	1		0	1		0	1	9.68E-04
Total small															1.52E-01
Total medium															0.00E+00
Total large															1.23E-02
Total rupture															0.00E+00
Total fire															0.00E+00
Total explosion															2.12E-02

Table 15.18 Detailed release frequency calculations for a batch reactor

15.6 Algorithm for batch reactor accident rates

Runaway

#	Question	Action if Yes	Action if No
1	Is the reaction carried out exothermic ?	Set the susceptibility to 1 for overheating, accumulation of reactants, hot spot, stirrer failure. Go to 3	
2	Is there an exothermic side reaction, decomposition, or polymerisation possible ?	Set the susceptibility to 1 for overheating, accumulation of reactants, hot spot, stirrer failure. Go to 3	
3	Can a stray catalyst cause runaway.	Set the susceptibility to 1 for stray catalyst. Go to 4	Go to 4
4	Can a contamination cause runaway.	Set the susceptibility to 1 for contamination. Go to 5	Go to 5
5	Can wrong substance addition cause runaway	Set the susceptibility to 1 for wrong substance. Go to 6	Go to 6
6	Can addition of the wrong amount lead to runaway	Go to 7	Go to 8
7	Is the addition via a manifold feeding several vessels	Set the susceptibility to 1 for wrong amount. Add 0.003 to the frequency for wrong amount Exit	Set the susceptibility to 1 for wrong amount Exit

Fire

#	Question	Action if Yes	Action if No
1	Are the reactants or solvent flammable ?	Set the susceptibility for fire to 1	

Overpressuring

#	Question	Action if Yes	Action if No
1	Is the reactor pressurised?	Set the susceptibility for rupture and overpressuring to 1	

15.7 References

1. Taylor, J.R. Process Safety Engineering, Designing and Building Safer Process Plant, Taylor Associates, 4th Edition 2001

2. Marrs and Lees

3. Barton, J.A. and Nolan, F.N. Runaway Reactions in Batch Reactors, in The Protection of Exothermic Reactors and Pressurised Storage Vessels, I Chem E Symposium Series No. 85,1984

4. Rasmussen, B., Unwanted Chemical reactions in the chemical process industry, Risø-M-2631, 1987

5. CSB, Hazard Investigation, Improving Reactive Hazard Management, Report No. 2001-01-H, 2002, US Chemical Safety and Hazard Investigation Board.

16 Scrubbers

16.1 Construction

Scrubbers are used for removal of gases from vent streams from reactors, acid tanks etc. A typical construction consists of one or two high non pressurised vessels, into which absorbent is sprayed at the top. The vent gas flows upward through the falling spray, and the reactive or soluble components are absorbed. The tanks used are often of fibre reinforced plastic.

In some cases scrubbers are packed, to form a packed column. These have a much higher absorbing surface than simple falling droplets, but the pressure drop through the scrubber is higher. Packed scrubbers are also much more susceptible to blockage.

High pressure scrubbers, constructed of steel, are used in the petroleum and petrochemical industries, to separate liquid droplets from gas flows, and to absorb acid gases. One characteristic type is amine absorbers, used to separate hydrogen sulphide from gases such as propane, butane and natural gas. These are really a different kind of equipment, built to different standards, and will not be considered here.

16.2 Releases from scrubbers

Most batch reactors release vapours or gases of one sort or another. Some produce significant amounts, such as ammonia from amination processes, and sulphur dioxide and hydrogen chloride from the production of acid chlorides using thionyl chloride. Such releases can constitute a major accident. However it is virtually impossible to obtain representative statistics for these incidents, firstly for the usual reason, that it is difficult to determine the size of the population of scrubbers; and secondly, because the actual incidents are very much under-reported. In most cases the scrubber discharge is via a high vent, and with good wind conditions, the release may not be noticed at all. From personal experience of 22 highly critical scrubbers (MIPA, hydrogen sulphide, thionyl chloride/sulphur dioxide/hydrogen chloride, and bromine scrubbers), releases which required reporting to environmental authorities have occurred with a frequency of about one per 8 scrubber years.

The main causes of failure have been pump failure, blockage of scrubber liquid nozzles due to crystallisation, power failure, and failure to replace absorbent due to operator error. Because of the high dependency on pumping, scrubbers in critical locations have in recent years been built using high integrity engineering techniques, such as use of dual pumps, vent gas concentration monitoring, and use of standby power supplies for the pumps.

Because of the uncertainties in reporting, and the relative simplicity of probabilistic failure analysis for scrubbers, the preferred value for release frequency used here is derived from the fault tree analysis below.

As can be seen from the cause statistics seen below, scrubbers are susceptible to internal explosion, if the gas streams contain flammable materials, or if the liquids become contaminated with flammable materials. They are also susceptible to overpressuring rupture , due blockage with absorbent crystals

16.3 Assessment of causal factors and susceptibilities

Table 16.3.1 gives a breakdown of causes of scrubber releases, based on data from MHIDAS

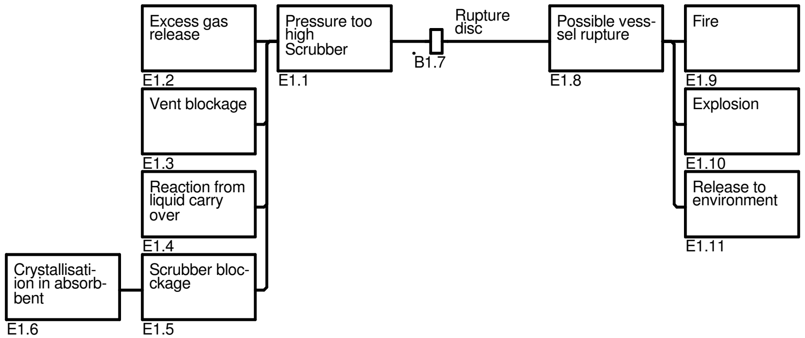
Cause	Number	%
Corrosion	2	8.3
Crack	2	8.3
Drain leak	1	4.2
Explosion	4	16.7
Inflow of liquid	4	16.7
Overflow	1	4.2
Lack of absorbent	3	12.5
Overload	4	16.7
Overpressure	3	12.5
	24	

Table 16.1 Scrubber release causes

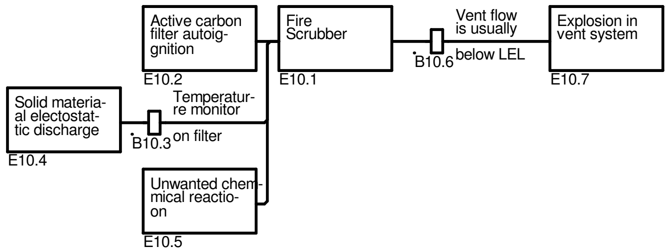
16.4 Detailed analysis

A detailed analysis is given in figures 16.1 to 16. 10 below. A summary quantitative analysis is given in table 16.2, based entirely on the detailed risk analysis.

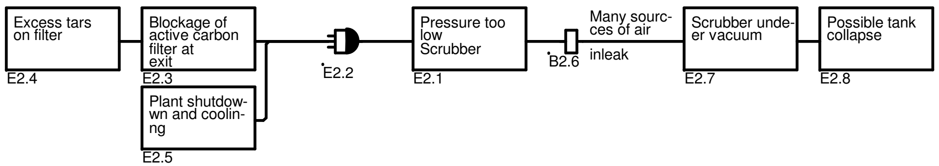
1: Safety barrier diagram for Pressure too high /Scrubber in Scrubber system



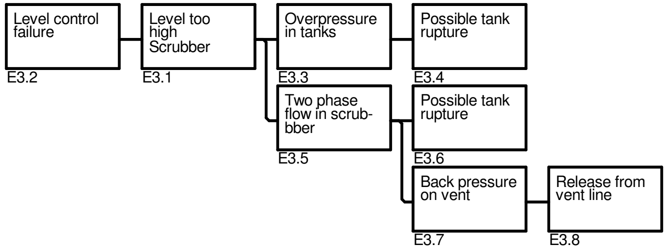
10: Safety barrier diagram for Fire /Scrubber in Scrubber system



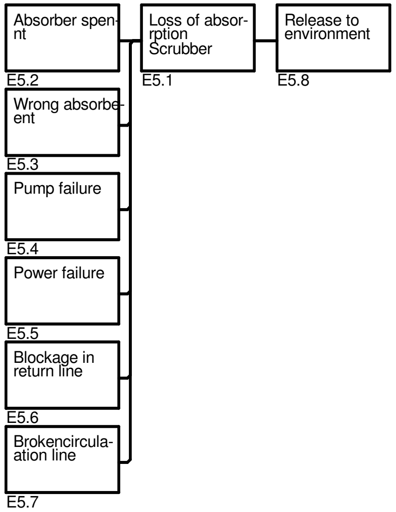
2: Safety barrier diagram for Pressure too low /Scrubber in Scrubber system



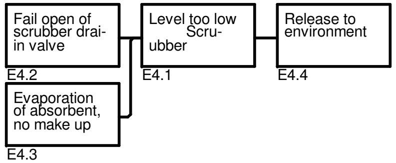
3: Safety barrier diagram for Level too high /Scrubber in Scrubber system



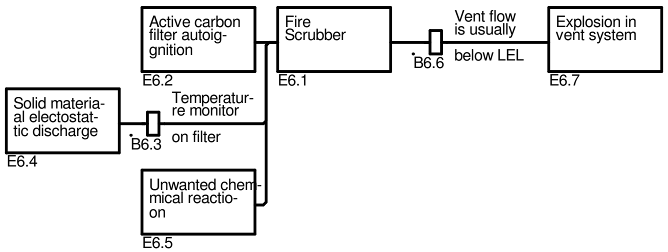
5: Safety barrier diagram for Loss of absorption /Scrubber in Scrubber system



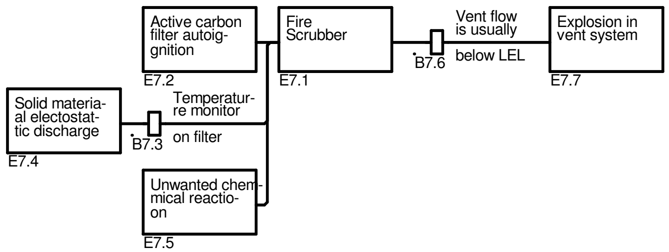
4: Safety barrier diagram for Level too low /Scrubber in Scrubber system



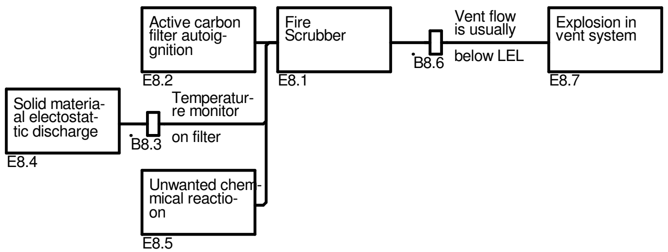
6: Safety barrier diagram for Fire /Scrubber in Scrubber system



7: Safety barrier diagram for Fire /Scrubber in Scrubber system



8: Safety barrier diagram for Fire /Scrubber in Scrubber system



9: Safety barrier diagram for Fire /Scrubber in Scrubber system

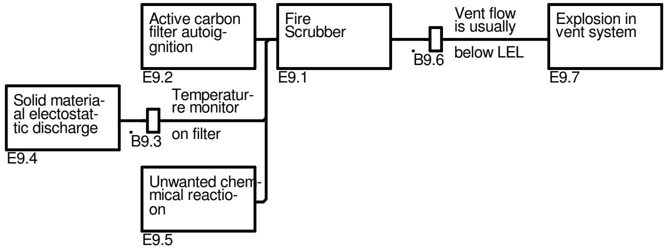


Table 16.2 Detailed release frequency calculation

Release frequencies	Small	Medium	Large	Rupture	Fire	Explosion
Scrubber			1.25E-01			

Failure cause	Conse-quence	Number of items or metres	Base frequency per year	Suscept-ibility	Risk contribution %	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	Susceptibility assessment
Small bore piping failure	small rel.	8	1.58E-04	1			1	1		1	1		1	1	1.26E-03	
Process piping, small	small rel.	2	4.00E-05	1			1	1		1	1		1	1	8.00E-05	
Process piping, med	med. rel.	20	3.00E-06	1			1	1		1	1		1	1	6.00E-05	
Process piping, Large	lg. rel.	5	2.50E-06	1			1	1		1	1		1	1	1.25E-05	
Flanges	small rel.	6	4.70E-05	1			1	1		1	1		1	1	2.82E-04	
Instruments	small rel.	2	4.00E-04	1			1	1		1	1		1	1	8.00E-04	
Valves	med. rel.	4	2.00E-03	1			1	1		1	1		1	1	8.00E-03	
Drain lines left open	lg. rel.		7.00E-04	1		Interlock	0	0.0013774		1	1		1	1	7.00E-04	
External fire	Rupture		1.00E-03	1		Deluge	0	0.0210319		1	1		1	1	1.00E-03	
Overpressure,blockage	rupture		5.00E-04	1		Burst disc	1	0.002	SV	0	0.05107979		1	1	1.00E-06	
Pump failure	lg. rel.		0.2	1		Standby	1	0.2	PTL	1	0.00477394		1	1	1.91E-04	
Absorbent depleted	lg. rel.		0.4	1		Trip	1	0.0013774		1	1		1	1	5.51E-04	
Power failure	lg. rel.		0.1	0		Trip	1	0.0013774		1	1		1	1	0.00E+00	
Insufficient capacity	lg. rel.		2.00E-02	1		Mass flow	0	0.0013774		1	1		1	1	2.00E-02	
Blockage of absorbent lines	lg. rel.		0.1	1			0	1		1	1		1	1	1.00E-01	
Explosion **	Explosion		0.0001	0			0	1		1	1		1	1	0.00E+00	
Total small															2.43E-03	
Total medium															6.00E-05	
Total large															1.22E-01	
Total rupture															1.00E-03	
Total fire																
Total explosion															0.00E+00	

**Explosion only if the scrubber handles (or can come to handle) flammable gases or vapour

16.5 References

1. Taylor, J.R. Process Safety Engineering, Designing and Building Safer Process Plant, Taylor Associates, 4th Edition 2001

17 Continuous High Temperature Reactors.

Continues high temperature reactors such as reformers crackers etc. take liquids or gases, often through a solid bed of catalyst, sometimes simply through a vessel mixed with suspended or dissolved catalyst. Such reactors have surprisingly low release rates. Table 171 shows the release frequencies for hydrodesulphurization and reformer reactors. The most common release types are small ones from flanges, instruments etc. However, when large releases do occur, the results are often Catastrophic. Many continuous reactors work at temperatures such as 250° C to 350°C, and at high pressures. If they operate in the liquid phase, the inventories are large. A typical catalytic reformer for example is 12 m high and 3 m in diameter. A release from such a reactor will typically be about 5 tons of gas. Rupture of a liquid reactor will release typically 30 to 50 tonnes. Ignition is virtually guaranteed, because most large scale continuous reactors are heated by fired heaters, and these are usually within the dispersion zone of released gas. The liquid being processed is often above its autoignition temperature.

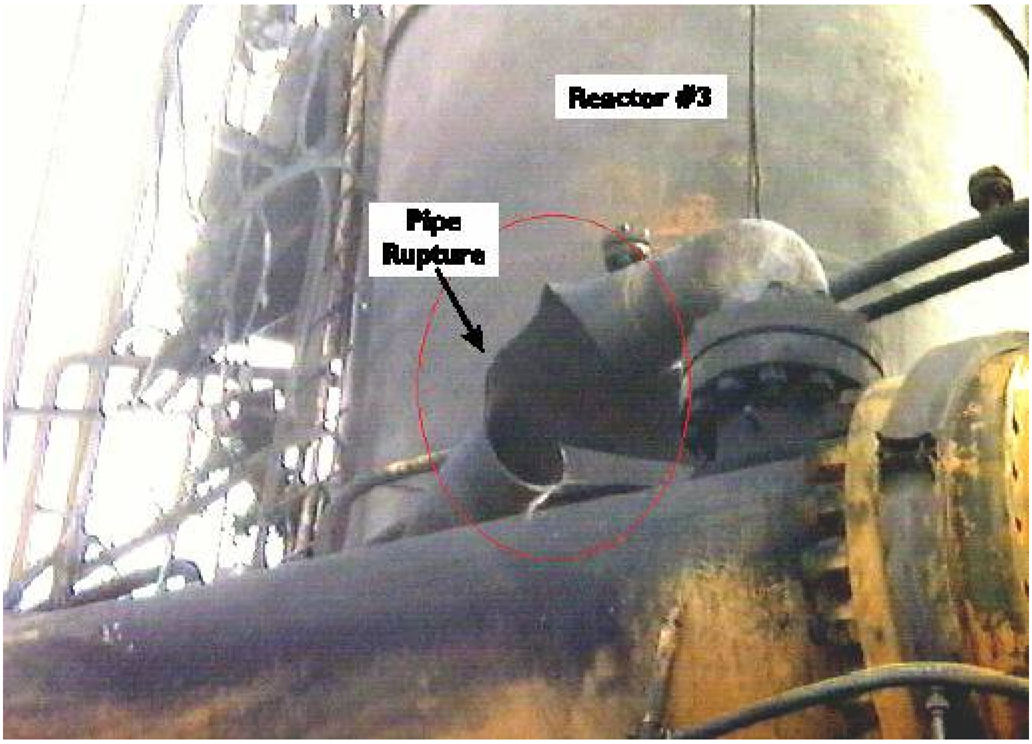


Figure 17.1 Pipe rupture resulting from a reactor overheating, ref 17.1

Liquid filled reactors are vulnerable to overpressuring due to overheating. The reactors are usually provided with safety valves. However, unless special provisions are made, the safety valve may release liquid. For this reason, relief is generally made to a flare system, if gas is being processed. This is not possible if there are significant quantities of liquid in the process, because the flare would then become a “fiery fountain”. For these cases, the relief is usually directed to a blowdown drum.

17.1 Release frequencies and hole sizes

No cases of catastrophic failure could be found in the ARIP or RMP data. This means that the frequency of such failures is probably less than $2*10^{-4}$ per year. However, one such failure, in the main piping, from a reactor, has been reported in chemical safety board reports in 2000 i.e. after the RMP reporting period(ref. 17.1), indicating that a value of $2*10^{-4}$ per year is of the right order of magnitude.

The release records for hydrotreating and platforming units in the US RMP data gave a frequency of 267 per 10^4 years, based on 18 releases. The hole size distribution is shown in figure 13.2.

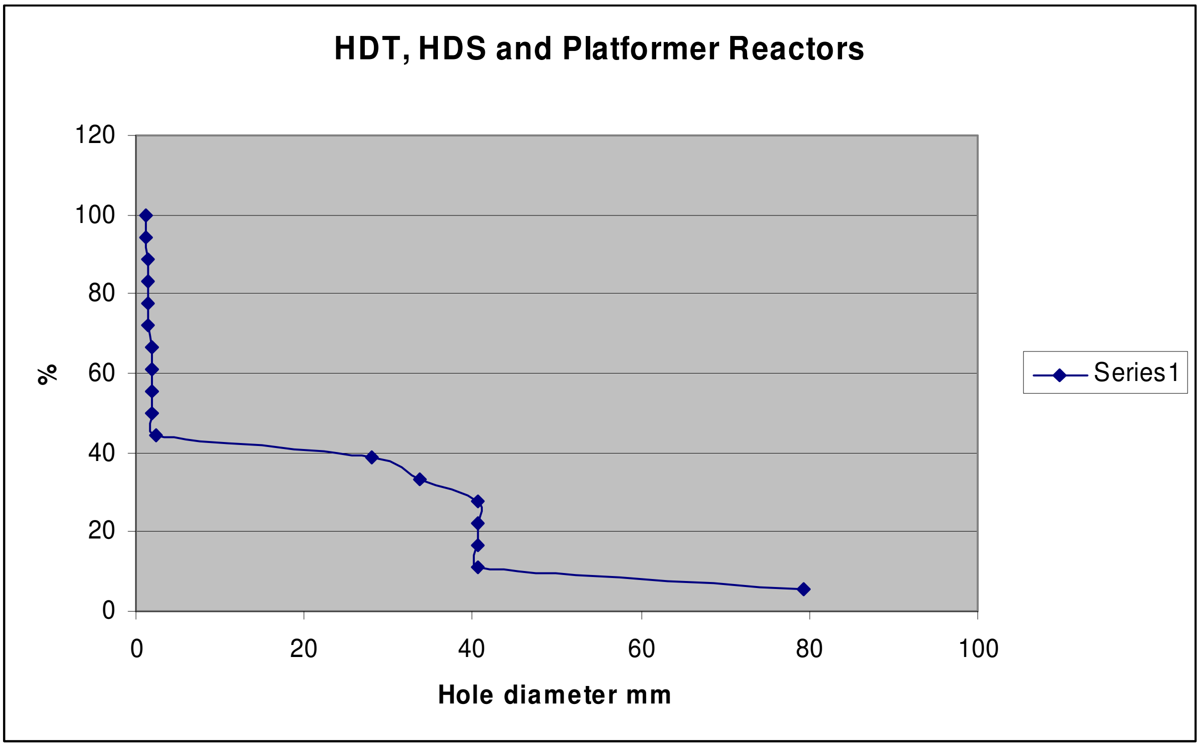


Figure 17.2 Reactor release hole size cumulative distribution

The frequency of continuous reactor explosions is significant in the overall picture of major hazards. The data from Marsh, (ref 17.2) shows that of the 100 largest loss accidents over 30 years, 12% arose in continuous reactors which make use or produce hydrogen, all hydrocrackers. Four of these were in USA. Seven hydrocracker accidents are registered in the RMP data base. In all there are 125 hydrocrackers, hydrotreaters, and hydrofining units registered. Many of these have more than one reactor vessel. Depending on the basis for analysis, the frequency of large UVCE's for these types of units is then calculated to be between $50*10^{-4}$ per year and $190*10^{-4}$ per year.

17.2 Baseline and typical frequencies for continuous reactor releases

Based on the above data the following typical and baseline frequencies are selected:

Value	< 10 mm	10-25 mm	25-50 mm	>50 mm	Rupture/ Explosion
Typical	55%	10%	25%	5%	
	$147*10^{-4}$	$27*10^{-4}$	$67*10^{-4}$	$14*10^{-4}$	$90*10^{-4}$

Table 17.1 Typical values for failure frequencies

17.3 Causes of reactor failures and releases

Table 17.2 gives causes of continuous reactor failures based on 58 failures from the MHIDAS data base.

17.4Detailed analysis

The causes of failure for continuous reactors do not show any “special” causes – the causes apply to virtually all reactors. A percentage reduction may be made based on table 17.3, for reactors which do not carry out exothermic reactions (10% reduction due to no runaway) for explosions. An increase of about 50% should be made for explosions in oxygenation reactors. The large vapour cloud explosion accidents recorded are almost all for hydrotreaters, and the frequency for other reactor types for explosions could be made by at least a factor of 3.

Figures 17.3 ff. give a detailed analysis of hazards.

		#	#	%	%
Explosion			24		41.4
	Air entry	1		1.7	
	Compressor slow	1		1.7	
	Error in gas ratio	1		1.7	
	Excess concentration	1		1.7	
	Excess oxygen	3		5.2	
	Flammable vapour	1		1.7	
	Gas air mix	1		1.7	
	Heat damage	1		1.7	
	Hot spot	1		1.7	
	Hydrogen generation	1		1.7	
	Ignition	2		3.4	
	Overheating	2		3.4	
	Runaway	4		6.9	
	Unknown	4		6.9	
External fire			1		1.7
Fire			2		3.4
Gas release			7		12.1
	Catastrophic corrosion	1		1.7	
	Crack	1		1.7	
	Insufficient feed	1		1.7	
	Leak	4		6.9	
Liquid release			10		17.2
	Corrosion	2		3.4	
	Flange failure	2		3.4	
	Gasket failure	1		1.7	
	Valve rupture	3		5.2	
	Crack	1		1.7	
	Seal leak	1		1.7	
Vapour release			10		17.2
	Crack	2		3.4	
	Gasket failure	1		1.7	
	Leak	1		1.7	
	Pipe rupture	2		3.4	
	Sight glass failure	1		1.7	
	Thermocouple pocket failure	1		1.7	
	Valve opened	1		1.7	
	Valve removed	1		1.7	
Vessel rupture			7		12.1
	Valve opened	1		1.7	
	Valve removed	1		1.7	
	Corrosion	1		1.7	
	Hot spot	1		1.7	
	Overpressure	1		1.7	
	Rapid tube corrosion	1		1.7	
	Refractory failure	1		1.7	
		58			

Table 17.2 Causes of continuous reactor releases (MHIDAS)

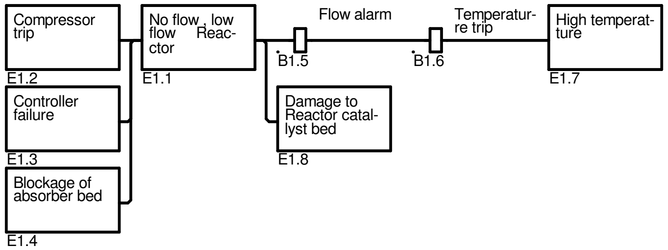
Continuous reactor

Release frequencies	Small	Medium	Large	Rupture	Explosion	Total
Typical	1.5E-02	2.7E-03	6.7E-03	1.4E-03	9.0E-03	2.55E-02

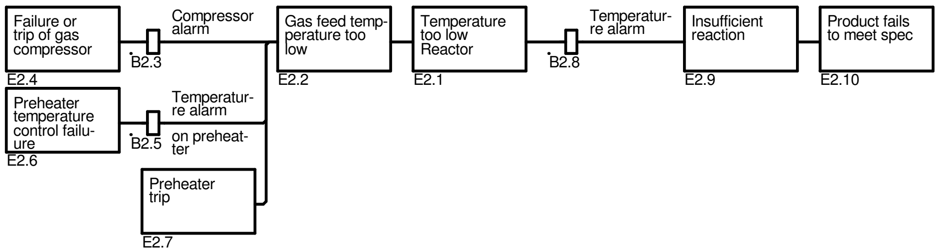
No	Failure cause	% of releases	Source	Release size	Suscept-ibility	Safety measures	Failure rate	Basis for susceptibility assessment
1	Internal corrosion	10.5	RMP	small	1	1.0	5.0E-03	Inspection of
2	Internal corrosion	2	RMP	medium	1	1.0	5.1E-04	3 refineries
3	External corrosion	1	RMP	large	1	1.0	4.1E-04	
4	External corrosion	0.1	RMP	rupture	1	1.0	7.8E-05	
5	Catastrophic corr.	1.7	MHIDAS	medium	0.5	1.0E+00	8.7E-04	
6	Corrosion	12		small	1	1.0E+00	5.7E-03	
7	Crack	6.9		small	1	1.0E+00	3.3E-03	
8	Flange failure	3.4		medium	1	3.0E-01	2.9E-03	
9	Gasket failure	3.4		medium	1	1.0E+00	8.7E-04	
10	Insufficient feed	1.7		rupture	0.01	1.0E+00	1.3E-01	
11	Pipe rupture	3.4		large	1	1.0E+00	1.4E-03	
12	Seal leak	1.7		small	0.03	1.0E+00	2.7E-02	
13	Sight glass failure	1.7		large	0.3	5.0E-01	4.6E-03	
14	Thermocouple pocket failure	1.7		large	0.2	1.0E+00	3.5E-03	
15	Valve opened	1.7		large	0.2	1.0E+00	3.5E-03	
16	Valve removed	1.7		large	0.5	1.0E+00	1.4E-03	
17	Valve rupture	5.2		large	0.2	1.0E+00	1.1E-02	
18	Hot spot	1.7		rupture	0.1	1.0E+00	1.3E-02	
19	Overpressure	1.7		rupture	0.05	1.0E+00	2.6E-02	
20	Rapid tube corrosion	1.7		rupture	0.1	1.0E+00	1.3E-02	
21	Refractory failure	1.7		rupture	0.5	1.0E+00	2.6E-03	
22	Air entry	1.7		explosion	1	1.0E+00	7.3E-05	
23	Error in gas ratio	1.7		explosion	0.2	1.0E+00	3.7E-04	
24	Excess concentration	1.7		explosion	1	1.0E+00	7.3E-05	
25	Excess oxygen	5.2		explosion	0.1	1.0E+00	2.2E-03	
26	Flammable vapour	1.7		explosion	1	1.0E+00	7.3E-05	
27	Gas air mix	1.7		explosion	0.1	1.0E+00	7.3E-04	
28	Heat damage	1.7		explosion	0.5	1.0E+00	1.5E-04	
29	Hot spot	1.7		explosion	1	1.0E+00	7.3E-05	
30	Hydrogen generation	1.7		explosion	0.5	1.0E+00	1.5E-04	
31	Ignition of gas mix	3.4		explosion	0.2	1.0E+00	7.3E-04	
32	Overheating	3.4		explosion	0.5	1.0E+00	2.9E-04	
33	Runaway	6.9		explosion	0.8	2.0E-02	1.9E-02	

Table 17.3 Assessment of susceptibilities for continuous reactors

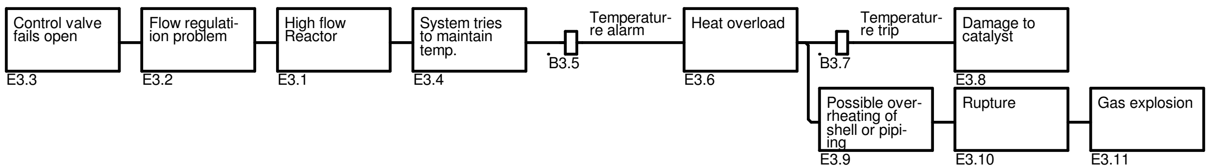
1: Safety barrier diagram for No flow , low flow /Reactor in Continuous reactor



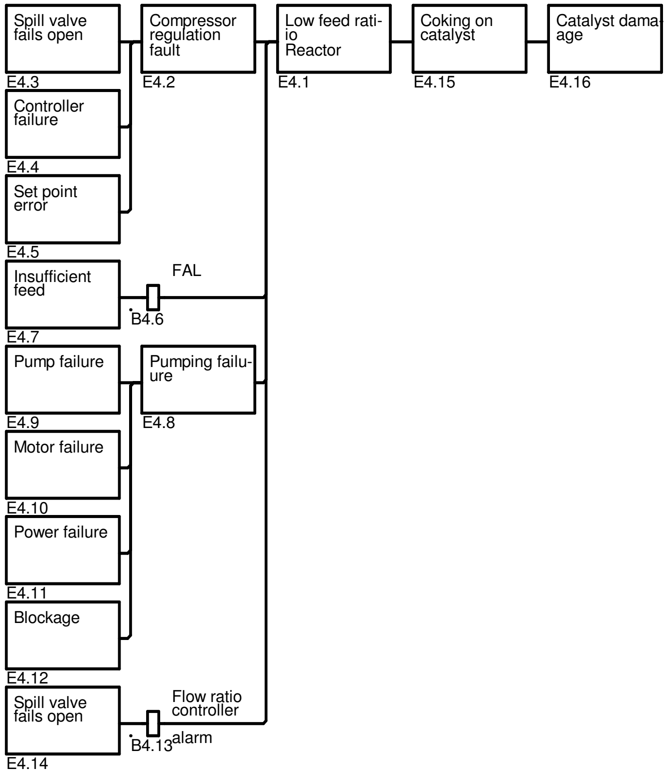
2: Safety barrier diagram for Temperature too low /Reactor in TCR



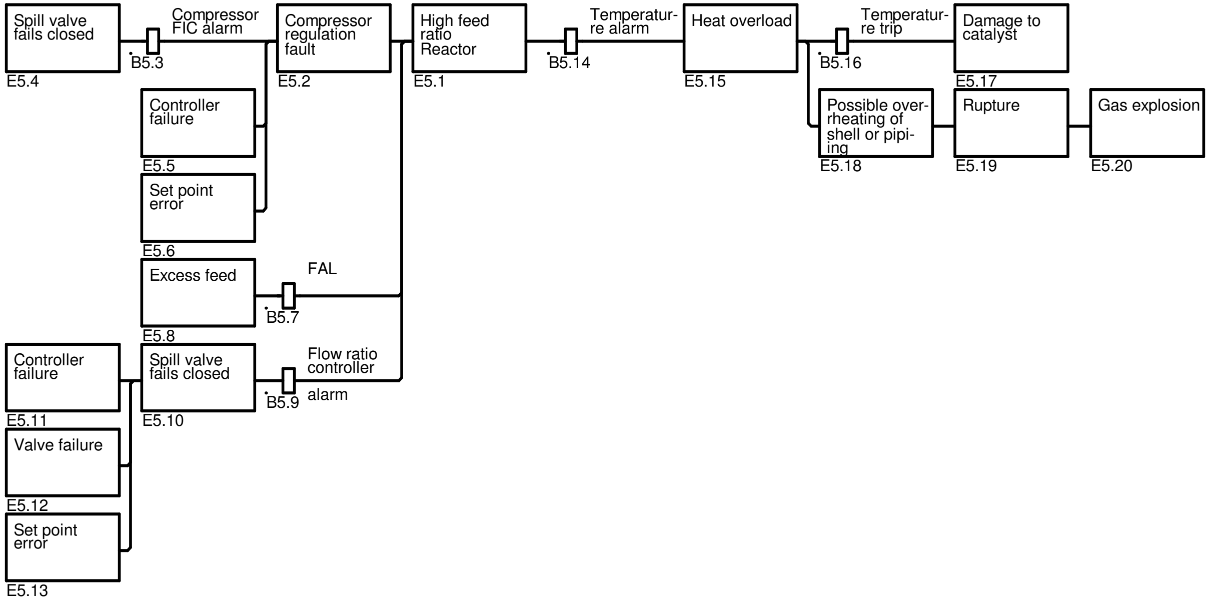
3: Safety barrier diagram for High flow /Reactor in TCR



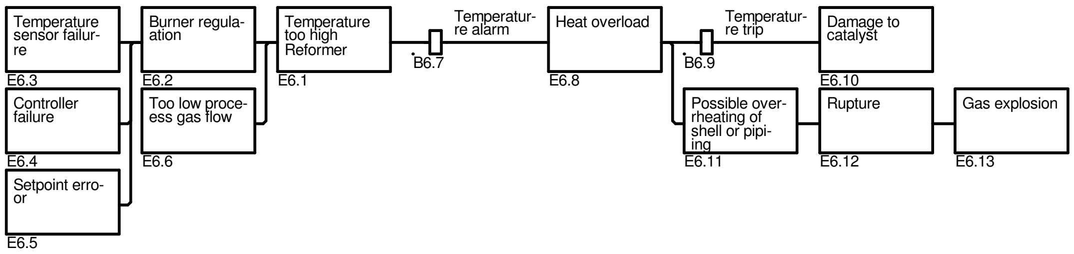
4: Safety barrier diagram for Low feed ratio /Reactor in TCR



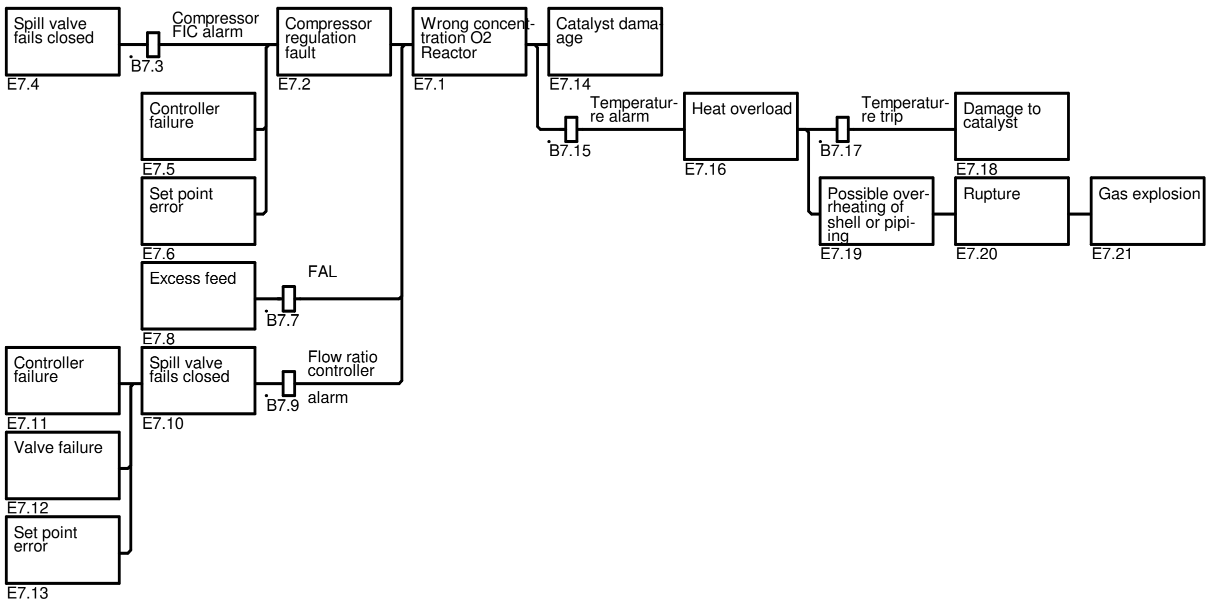
5: Safety barrier diagram for High feed ratio /Reactor in TCR



6: Safety barrier diagram for Temperature too high /Reformer in TCR



7: Safety barrier diagram for Wrong concentration O2 /Reactor in TCR



Failure cause	Conse-quence	Number of items or metres	Base frequency per year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction factor	Safety barrier 2	Y/N	Risk reduction factor	Safety barrier 3	Y/N	Risk reduction factor	Assessed frequency per year
Internal corrosion	small	1	4.96E-03	1		0	1		0	1		0	1	4.96E-03
Internal corrosion	medium	1	5.14E-04	1		0	1		0	1		0	1	5.14E-04
External corrosion	large	1	4.09E-04	1		0	1		0	1		0	1	4.09E-04
External corrosion	rupture	1	7.78E-05	1		0	1		0	1		0	1	7.78E-05
Catastrophic corr.	medium	1	8.74E-04	1		0	1		0	1		0	1	8.74E-04
Corrosion	small	1	5.67E-03	1		0	1		0	1		0	1	5.67E-03
Crack	small	1	3.26E-03	1		0	1		0	1		0	1	3.26E-03
Flange failure	medium	1	2.91E-03	1		0	1		0	1		0	1	2.91E-03
Gasket failure	medium	1	8.74E-04	1		0	1		0	1		0	1	8.74E-04
Insufficient feed	rupture	1	1.32E-01	1		0	1		0	1		0	1	1.32E-01
Pipe rupture	large	1	1.39E-03	1		0	1		0	1		0	1	1.39E-03
Seal leak	small	1	2.68E-02	1		0	1		0	1		0	1	2.68E-02
Sight glass failure	large	1	4.63E-03	1		0	1		0	1		0	1	4.63E-03
Thermocouple pocket failure	large	1	3.47E-03	1		0	1		0	1		0	1	3.47E-03
Valve opened	large	1	3.47E-03	1		0	1		0	1		0	1	3.47E-03
Valve removed	large	1	1.39E-03	1		0	1		0	1		0	1	1.39E-03
Valve rupture	large	1	1.06E-02	1		0	1		0	1		0	1	1.06E-02
Hot spot	rupture	1	1.32E-02	1		0	1		0	1		0	1	1.32E-02
Overpressure	rupture	1	2.64E-02	1		0	1		0	1		0	1	2.64E-02
Rapid tube corrosion	rupture	1	1.32E-02	1		0	1		0	1		0	1	1.32E-02
Refractory failure	rupture	1	2.64E-03	1		0	1		0	1		0	1	2.64E-03
Air entry	explosion	1	7.32E-05	1		0	1		0	1		0	1	7.32E-05
Error in gas ratio	explosion	1	3.66E-04	1		0	1		0	1		0	1	3.66E-04
Excess concentration	explosion	1	7.32E-05	1		0	1		0	1		0	1	7.32E-05
Excess oxygen	explosion	1	2.24E-03	1		0	1		0	1		0	1	2.24E-03
Flammable vapour	explosion	1	7.32E-05	1		0	1		0	1		0	1	7.32E-05
Gas air mix	explosion	1	7.32E-04	1		0	1		0	1		0	1	7.32E-04
Heat damage	explosion	1	1.46E-04	1		0	1		0	1		0	1	1.46E-04
Hot spot	explosion	1	7.32E-05	1		0	1		0	1		0	1	7.32E-05
Hydrogen generation	explosion	1	0.000146	1		0	1		0	1		0	1	1.46E-04
Ignition of gas mix	explosion	1	0.000732	1		0	1		0	1		0	1	7.32E-04
Overheating	explosion	1	0.000293	1		0	1		0	1		0	1	2.93E-04
Runaway	explosion	1	0.018577	1		0	1		0	1		0	1	1.86E-02
Total small														4.96E-03
Total medium														1.39E-03
Total large														4.09E-04
Total rupture														7.78E-05
Total fire														0.00E+00
Total explosion														2.35E-02

Table 17.4 Detailed frequency calculation for continuous reactor

17.5 Algorithm for continuous reactor release frequency

Runaway

#	Question	Action if Yes	Action if No
1	Is the reactor a hydrocracker ?	Set susceptibility for release and explosion to 1 in detailed analysis	Go to 2
2	Does the reactor handle hydrogen ?	Set susceptibility for release and explosion to 1 in detailed analysis	Go to 3
3	Is the reaction an oxygenation ?	Set the susceptibility for internal explosion to 1 There should be a high integrity shutdown system. If there is, include it in the analysis	Go to 4
4	Is the reactor liquid phase ?	Set the susceptibility for overpressuring to 1	Exit
5			
6			
7			

18 Distillation Columns

Distillation columns generally consist of a tall narrow vessel, inside of which there are either trays used to retain liquid, with holes to allow vapours to bubble through the liquid; or with packing of steel or ceramic, used to ensure good liquid/vapour contact. Liquid flows down through the column, and vapour flows upward. Typically the content of liquid in a column is about 5 % of the total volume. Since column volumes are high, this quantity itself can be quite significant. Many distillation columns work at pressure up to 30 bars or so, and occasionally, very high pressure columns are used. Most columns are designed to work at pressures from one to five bars however.

Figure 18.1 Distillation columns in a BTX plant

18.1 A simple column

Distillation columns are used in oil refineries, petrochemical plants, and chemical plants to separate liquids with different boiling points.

The classical design for a petroleum distillation column is shown in figure 7.1 vessel, with a number of trays. Heat is provided at the bottom of the column, so that vapour passes upward through the column. The trays serve to hold up liquid flowing down the column, and provide intimate contact between the vapour and liquid. As a result, each tray provides effectively a new distillation.

At the top of the column, vapour passes to a condenser or dephlegmator. Some of the condensed liquid is generally returned to distributor, or sprays or nozzles at the top of the column as a reflux flow. This serves to provide more liquid/vapour contact in the column.

Heating for the columns may be provided by steam in a heat exchanger known as a reboiler; by a reboiler heated by hot liquid from another part of the process (heat recovery); or by a fired heater.

Circulation of the bottoms liquid through the pump may be by natural convection (thermosyphon type) or there may be a reboiler circulation pump.

The overhead condenser may be a heat exchanger type, cooled by water or by another process stream; or it may be a fin fan type air cooler.

There will generally be a vessel to collect the condensed overhead flow, though this may be part of the condenser, if this is of the heat exchanger type. This vessel, the reflex drum, ensures that there will be adequate reflex flow in the case of disturbances.

The reflux drum often serves as the feed drum for the next column in a series, as is the case for example in a deethaniser, depropaniser, and debutaniser column train.

The bottom of the column is filled with an amount of liquid which serves to keep the reboiler filled. The volume is designed to be sufficient to even out upsets in the feed or operation of the column.

The bottom stream is pumped out to the next stage in the processing.

Variants on the simple column

Columns may be much more complex than that just the simple arrangement described above.

Many columns have packed sections. These are section filled with packing such as rings or saddles of steel or ceramic, contained between two layers of steel mesh. The rings or saddles become wetted with liquid, and so serve to bring the vapour into intimate contact with the liquid.

Large columns are often used to make multiple separations. An example is the atmospheric column in a refinery. To allow separation of multiple products trap out trays are provided. These collect liquid part way down the column.

Trapped out liquid may be cooled, and then returned to the column as a "pump around", which is a kind of reflux. In this way, a large column becomes the equivalent of several small distillation columns.

Fine chemicals plants often make use of semi batch distillation. A column is mounted on top of a kettle type vessel. The kettle is initially charred with feed, which is then heated. As top product is distilled off, more feed is added until the kettle is completely filled with bottoms product.

Such semi batch columns are often used as multi product columns. First a light fraction such as solvent, is distilled off. Then the temperature is raised and the main product is distilled off. The residue, which is often tar, remains in the kettle.

Vacuum columns

Vacuum columns are used in refineries to extract remaining light and medium fractions from atmospheric column residue. Vacuum columns are also used in fine chemicals plant to allow distillation of products which decompose with excessive heating.

Refinery vacuum columns use large steam ejector pumps to achieve a vacuum. The steam is condensed along with the distillation vapour, and a separator then serves to give the top distillate liquid stream.

Fine chemicals plant columns typically use water ring pumps to achieve a vacuum. The Vacuum pump is generally placed after the overhead condenser, so that vapour passes to the pump in only limited quantities.

Physical arrangement of columns

Large distillation columns are usually free standing vessels, placed on a steel skirt. The skirt has a man hole, which also serves for ventilation.

The shirt should be provided with fire protection, typically in the form of a concrete coat. The shirt is very vulnerable to fire.

The bottoms line, or a drain, will often be located at the lowest point of the column end shell.

Very large columns, over 50 m, may be guyed.

The column skirt is tied into the base foundation with solid tie down bolts.

The column should have an earthing strap to take current from possible lightning strikes.

18.2 Safety equipment

The prime equipment for column safety is the safety valve, protecting against overpressure. Level alarms and trips in the bottoms and reflux drum serve to reduce the chance of vapour transfer to pumps. Level trips are particularly important to prevent transfer of liquid to vacuum pumps.

Delta P alarms are used to indicate onset of column flooding, which can cause column damage.

Modern columns have emergency shutdown valves on the bottoms liquid line, to prevent release of the liquid inventory.

Fire fighting on columns is generally difficult, for example flange fires. High, automatics or manually remotely controlled monitors are used to keep columns cool while depressurisation takes place.

Skirts and columns up to a certain height should have fire protection in the form of concrete coatings.

18.3 Hazards (ref 14.1)

Hazards on columns themselves have generally fallen into three main groups.

1. Leaks on piping, flanges, fittings, manhole covers etc. These have often developed into fires, which have often further damaged equipment, sometimes developing into large releases which damage large parts of plants.

In some cases released vapour has exploded, causing damage or injury.

2. Overpressuring of columns has occurred but has usually led to releases via safety valves. These in turn have caused problems, such as liquid discharge, overpressuring of blow down lines, and liquid release via the flare stack.

3. Overheating and decomposition of reactive and unstable substances has occurred in chemical distillations. This has in some cases led to explosions.

More hazards occur in equipment associated with distillation columns.

4. Pump seal leakages, releases during pump maintenance, and product sampling errors leading to releases, have occurred relatively frequently.

5.Overflow of liquid to ejector or vacuum pumps from the dephlegmator or reflux drum has occurred with fire or toxic release as a result.

6.Fired heaters have leaked with sometimes drastic consequences.

7.Heat exchanger leakages have caused fires and explosions

See the appropriate chapter for more examples

Distillation column instabilities or failures tend to have wide spread effects in a process plant. These instabilities can cause accidents at a considerable distance from the column itself.

18.4 Case stories

1.Texaco Refinery, Milford Haven 24 July 1994.

Lightning started a fire at the crude distillation unit. This affected vacuum distillation, alkylation and butamer units. Time was 09:00.

Hydrocarbon flow was lost to the deethaniser column in the FCC unit. The deethaniser bottoms valve closed, stopping feed to the debutaniser. The debutaniser bottoms valve to the naphtha splitter also closed.

The debutaniser column was effectively blocked in, but heating continued. Debutaniser pressure rose, and pressure relief valves opened three times to the relief line and KO drum.

A little later, liquid level was restored in the de-deethaniser, and flow to the debutaniser resumed. The debutaniser bottoms valve should have opened, but stuck. The operators received an erroneous feedback indicating the debutaniser bottoms valve was open. To relieve debutaniser pressure, operators opened from the debutaniser reflux drum to the wet gas compress or interstage drum.

The wet gas compressor compresses vapour from the FCC. The interstage compressor drum overflowed and caused the wet gas compressor to trip.

The debutaniser vented to flare.

The trip of the compressor caused very large quantities of FCC gas and vapour to be released to flare.

Operators improvised a drain from the interstage drum using steam hoses. The wet gas compressor restarted at 12:28. The debutaniser vented again to flare at 12:46 and continued venting.

The wet gas compressor tripped again at 13:22 because of flooding. This vented large volumes of gas to flare. However, by this time, the flare knock out drum was overfilled, due to a design error/procedure error in the knock out drum pump out. Liquid passed to the flare line, which could not take the load. It ruptured, and released liquids ignited at a location 110 m away. The ensuing explosion had the TNT equivalent of 4 tonnes (from a 20 tonne release) and damage amounted to £ 34 million.

Note that the operators were confused in the period because of inadequacies in the instrumentation and man machine interface.

3. Dutch State Mines, Beek November 1975

A naphtha cracker consisted of fired heater, cracker, wet gas compressor, deep cooling unit, demethaniser, deethaniser, depropaniser and debutaniser.

The plant was being brought on line from 06:00, and compressed gas sent to the deep cooling system. Gas escaped from the depropaniser section and exploded, killing 14 and injuring 106.

The release was 3 to 5 tonnes of propylene.

A weld had been made on the depropaniser column feed drum 40 mm connection to a safety valve, using gas welding. The brittle transition temperature was suspected, after the accident, to have risen to about 0^0C as a result of the welding. Normal temperature at this location is 65^0C.

At the time of the accident the deep cooler was bypassed. The reboilers on the depropaniser column, which used hot water, were not operating properly. It is suspected that the improper operation was due to liquid viscosity. As a result, demethaniser (deethaniser?) bottoms were flashed from 32 bar to 23 bar, giving a temperature of about 0^0C or lower, with a C_2 content of about 35%.

Discharging deethaniser bottoms to the flash drum could have caused flashing and a temperature as low as -10^0C. This discharge occurred coincidentally with an interruption of flow in the condensate stripper. As a result an amount of cold liquid with considerable C_2 content was disposed on top of the warm C_3 + liquid in the feed drum. Gas flow through the pressure control valve, and pressure build up, prevented flow from the feed drum. The drum filled up with a layer at -10^0C. The safety valve may have opened. Laboratory tests showed that an impulse of as little as 5 kg m could then rupture the safety valve pipe.

3.Butadiene refining at Union Carbide, Texas City October 23 1969

The butadiene unit recovered product from a crude C_4 stream from the olefins cracking. Butadiene was separated by absorption in dimethylacetamide. Stripped gases from the absorption were recycled. Heavy components of the feed stream were removed as bottoms product. This included vinyl acetylene at about 80% concentration.

Normal operation was proceeding, but the unit was to be shut down for stripper make compressor repairs. Reduction of feed began at about 09:00 and all feed was stopped by 11:00. The absorber and stripper columns were shut in under methane pressure. The fore column and refining column were placed on total reflux. Shut in was by manual valves on the feed and kettle lines, and by motorised valves on the overhead lines.

The refining column operated erratically, but this was normal, since it operated under very high reflux.

Subsequent examination of records showed that the refining column was slowly losing material through a leaking seat on the overhead valve. Reflux and steam flow fell accordingly, indicating loss of material. The make flow meter showed a continuous flow. The operator assumed that the flow meter was off calibration since the make motor valve was closed, and the chart showed a continuous straight line.

Vinylacetylene concentration in one tray rose to an estimated 60 %. Loss of liquid uncovered reboiler tubes. (Calandria tubes). After about 9 hours an explosion occurred. The lower 40 ft of the column fragmented. Thereafter, released vapour exploded. The butadiene column was replaced. Others had major repairs for deformed shells, shrapnel holes and distorted or broken nozzles. Auxiliary equipment including piping, cables and instruments required replacement. Costs exceeded $ 6 million.

18.5 Release frequencies

Distillation columns have proved to be surprisingly robust, with very few accidents giving releases. In the US RMP data refineries there were no records of major leaks or ruptures of distillation columns with off site consequences. However, this may be a result of the recording practice – columns are recorded as “vessels”

In the Marsh data for largest losses for the last 30 years (ref 14.1) 3 accidents are recorded involving columns or reboilers (3% of the total), all of them involving explosions in reactive compounds being processed or purified in the column.

Some dramatic accidents are known from the literature, for high pressure columns

In all, although distillation columns represent active process equipment, it seems most reasonable to regard them simply as vessels, from the point of view of failure frequencies. The lower section, where the liquid is stored in order to provide suction pressure for pumps, is often subject to corrosion significantly above that expected for a storage vessel, but columns are relatively straight forward to inspect, and corrosion monitoring programmes seem to be able to prevent dramatic failures.

Earth quake has proved to be one of the more significant causes of catastrophic damage to columns around the world, with major accidents at Iznat, Turkey, in China, and in California. Operational problems in control of columns have proved to be the cause of failures in other equipment and piping, see for example accidents at Milford Haven in 1996, where level control problems over a period of hours led to release of butane from a distillation receiver to a flare system, in which piping rupture when the flare separator drum overflowed.

Table 18.2 gives the assessed release frequencies for columns.

Figure 18.3 Fragment fro the explosion of a high pressure column, after accumulation of acetylene.

18.6 Frequency of column releases and accidents

Distillation columns show a surprisingly low frequency of releases and explosions. The original IFAL, TA, and API data (see section 4.4) seems to be the best available:

< 10 mm	10-50 mm	25-50 mm	>50 mm	Rupture/ Explosion
$25*10^{-4}$	$3*10^{-4}$	$0.1*10^{-4}$	$0.1*10^{-4}$	$0.05*10^{-4}$

Table 18.2 Typical release frequencies per year for columns

19 Heat Exchangers

Heat exchangers are used for heating or cooling gases and liquids and are found in many different forms.

Figure 19.1 shows a typical two pass shell and tube heat exchanger with a sliding tube sheet (right end) to allow for tube expansion.

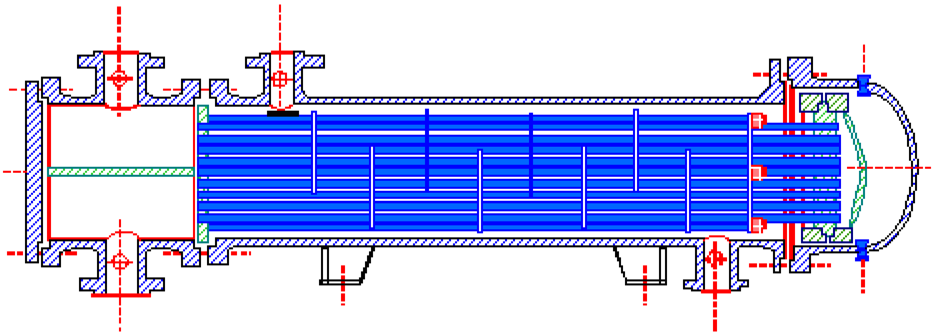


Figure 19.1 Shell and tube heat exchanger

19.1 Hazards

Heat exchangers are pressure vessels, and suffer the same hazards as any pressure vessel. Overpressuring leakage, and problems with external fire. Additionally, heat exchangers suffer from problems of leakage, from the high pressure to the low-pressure side.

Such leakage can cause problems of overpressuring of the low-pressure side. In some cases this can be so violent as to cause bursting.

Passage of high pressure into a liquid stream can cause problems of contamination, corrosion (see case stories), and heavy vibration due to two-phase flow.

Passage of liquids into liquid causes contamination. If the high-pressure liquid has a low boiling point, it can boil and cause serious overpressuring of the low-pressure side.

Heat exchangers can leak because of improper design. Expansion of tube due to heating can cause deformation of tube plates. Tube plates should be dimensioned for this, with expansion rods as support if necessary. Expansion can also overstress the heat exchanger shell. Depending on design, the stress may be borne by the head flange, with leakage as a result.

Heat exchanger tubes may also buckle, due to expansion as temperatures increase.

Corrosion is a significant cause of heat exchanger leakage. Heat exchanger tubes are usually made with a minimum of thickness to reduce cost, and to limit heat flow resistance. A corrosion allowance is applied, to give acceptable life. Unusual or unexpected conditions, or contamination, accelerate corrosion, as a result, leakage can occur long before the design lifetime is approached.

The tube plate is also a frequent location for leakage. The tubes are inserted into the tube plate, and rolled into place using expanding rollers. The tubes should be deformed so well as to block any entry of corrosive liquid. Poor workmanship during manufacture, unevenness in the bored hole, or chips of metal and burrs can prevent proper sealing. The result can be direct leakage, but also gaps which allow corrosion.

Tube/tube plate seals made in stainless steel are vulnerable to corrosion, because the gap is low in oxygen. The protective oxide film can therefore not form.

Fretting is an important cause of failure in heat exchangers.

If liquid flow rates in the shell are high, and particularly if cross flow rates are high, the tubes can begin to vibrate and sing. This can cause fatigue cracking in extreme cases. More frequently, leaks arise as the tubes rub against each other or against the flow baffles.

In order to limit fretting, vibration must be prevented. This can be achieved with low shell flow rates, stiff heat exchanger tubes, or by making flow baffles function as supports. (The shorter the unsupported tube length, the higher the frequency of vibration, and the higher the forces required to cause vibration; also, the lower the maximum displacement for a given force (flow rate)). Since vibration is usually a self-amplifying resonance effect, shortening unsupported tube length can often eliminate vibration.

Heat exchangers are also subject to blockage. This is rarely a cause of accidents in the exchanger itself (in principle if blockage could lead to overheating or overpressure, but designs which allow this are rare). More important is that loss of cooling in a heat exchanger can lead to overheating and resultant overheating accidents (see e.g. chapter on chemical reactors).

Large heat exchangers have large heads, with resultant large gaskets. Placing such gaskets is difficult. Also, bolting of the gaskets can be difficult. High temperature heat exchangers usually have leaks, which lead to corrosion or flange fire. The problem can be solved by overdimensioning gaskets.

19.2 Case stories

1. A heat exchanger cooling overhead gas suffered a tube rupture. Gas entered the cooling flow. The liquid was forced out, and the hammer ruptured a riser pipe. The pipe bent over, across a transformer station. The sparking ignited the gas.

2. An operator opened the drain valve for a condenser. The condenser was pressurised since its temperature was a little over 100°C. Condensate in the bottom of the condenser hammered into the first elbow on the drain line and ruptured it.

3. A fire in a heat exchanger for a crude distillation vessel caused an interruption to plant operation of 7 days.

4. Four people killed and one injured when an explosion occurred in a heat exchanger in the thermal cracker unit.

5. Titanium tetrachloride leaked from tank in production area. 40 fire-fighters tackled leaking vapour cloud, containing it on site. Initial reports blame leaky heat exchanger which caused pressure drop.

6. Small fire within a visbreaker heat exchanger caused 2-day disruption at refinery.

7. Fire in heat exchanger in refinery desulphurisation unit also affected reformer operations. 1 injured as result of incident.

8. Pipeline burst during tank transfer as heat exchanger tube broke & leaked water coolant into chlorine which travelled into storage tank & corroded hole in discharge line. 55tons released. Huge chlorine cloud. Gas dispersed over desert.

9. Explosions and fire following a propane leak at a refinery north of Leixoes. Fire contained and burnt out. 7 workers injured. Cause reported to be leak from plug from heat exchanger in propane deasphalting unit.

10. Fire at Petroleos Mexicanos Madero refinery caused when one of the walls of a heat exchanger collapsed. Damage amounted to half a million pesos and took 2 weeks to repair.

11. During the commissioning of an ethylene cracker, a low-pressure heat exchanger over-pressurised & fractured. Flammable vapour escaped & ignited. Fire burned for 12hours. Alum heat exchangers, 30m above the ground were destroyed & other equipment damaged.

12. Gas oil heat exchanger in synthetic crude oil process badly damaged when fan caught fire. Damage halted production for 6 weeks.

13. During heat exchanger repair on benzene plant mixed methane/hydrogen contained in cold box leaked from improperly spaded valve. Welding spark ignited gas

14. Heat exchanger used to cool LPG before injection into underground caverns failed violently. Exchanger apparently valved off in liquid full condition & had no pressure relief valve. The tank ruptured & caused pipe failures, a fire & the death of 2 workers.

15. Leak in heat exchanger let air into heat transfer system during start-up following maintenance. Top blew off transfer oil surge tank. Fire lasted 4 hours causing extensive damage to 6 reactors and associated equipment

16. Fire at hydrocarbons plant of saras-chimica broke out on early 11/11 in 1 heat exchanger of the reforming system. Fire extinguished after a couple of hours. Apparently fire caused by leakage of hydrogen.

17. Heat exchanger on an LNG plant ruptured violently during start-up operation. Investigation showed valve on blowdown line not opened after shut down. Fragments of 170te column thrown 160ft. Ensuing fire extinguished in 30mins.

18. The steam piping in a gas treatment plant, only 4 years old, was found to have hundreds of pinholes from corrosion. The investigation showed that sour gas had leaked into the boiler feed water from a leak in a heat recovery heat exchanger.

Figure 19.2 Holes in heat exchanger due to fretting corrosion, widened by pressurised flow.

Figure 19.3 Heavy external corrosion on a heat exchanger due to the use of an external water drench to increase cooling capacity

19.3 Release causes

Crozier (Chemical Engineering Dec 15 1980) gives causes of exchanger failure:

- Water and steam hammer.
- Corrosion.
- Erosion.
- Vibration (fatigue and fretting).
- Overpressure.

Water hammer occurs when a quick acting valve closes. One example of this occurs when a check valve closes to stop a reverse flow. Steam hammer occurs when a steam bubble collapses. This can occur, for example, in a condenser, when steam or vapour feed stops. Condensers should have a vacuum breaker on the liquid outflow line to prevent suck back of liquid.

In vertically installed heat exchangers, gas can collect at the upper end. The tubes will then not be cooled by liquid with the result that salts can concentrate at the hot surface.

Shell side flow velocity needs to be maintained 0.7 m/s to prevent deposition of the suspended solids.

Leakage of liquefied gas or liquid above its boiling point from tube to shell or shell to tube, can cause serious over pressuring if the pressured in the primary and secondary sides are different. Leakage of sulphuric acid across a tube plate or through a tube can cause similar over pressuring. Excessive temperature on the high temperature side can also cause rapid boiling and over pressuring.

A typical relief valve opens 10 to 50 milliseconds. The worst flashing liquid over pressuring incidents can overpressure a vessel in less than this time, so speed of opening needs to be considered. Rupture disc can open more rapidly.

19.4 Release frequencies

The US RMP data does mot distinguish heat exchangers fro other vessel type equipment.

The HSE offshore data described in Volume I gives the best published set of data available for hydrocarbon applications. The data are repeated here:

		Hole size distribution						
Equipment type	Failure frequency per year	< 10 mm	10-25 mm	25-50 mm	50-75 mm	75-100 mm	>100 mm	NA
Heat exchanger HC in tube	4.94E-03	0.75	0.08	0.08				0.08
Heat exchanger HC in shell	2.92E-03	0.85	0.08					0.08
Heat exchanger, plate	1.03E-02	0.85	0.1		0.05			

Table 19. 1 Heat Exchanger release frequencies, HSE

Anyakora and Lees gave release frequencies between 0.17 and 0.5 per year for small and large heat exchangers, respectively, based on data from the 1960's and 1970's, but these definitely do not apply to modern refinery and oil installations, nor to typical modern fine chemicals installations.

The author's own observations in petrochemicals plant over a period of 3 years, gave one significant (large) release, on 85 exchangers, or $4*10^{-3}$ per year, which accords with the HSE data, except perhaps for the size of the hole (60 mm) Observations on a number of fine chemicals plants over two years are given in table 19.2

Equipment type	Failure type	No of items	No. of failures	Frequency per year
Shell and tube heat exchanger, water cooled, carbon steel	Leak across tubing	44	3	0.034
	Gasket leak	44	1	0.011
Stainless shell and tube	Gasket leak	172	2	0.0058

Table 19.2 Heat exchanger release frequencies, fine chemicals

20 Warehouse and Storage Fires

Many chemical production plants store raw materials in drums, sacks, and big bags, in storage sheds and warehouses. The main hazards arising from these are fires, which can start from leaks, subsequently ignited by electrical systems, by fork lift trucks, or by electrostatic sparks for example from clothing. The other main problem arises from falling drums and packages due to collapse of pallets, racks etc, or as a result of fork lift truck crashes and stacking accidents. Some substances can ignite spontaneously, for example with a release of fire nickel powder or decomposition of peroxides, which create both a spill and an ignition at the same time. Given such an ignition, fire can spread rapidly by rupturing drums, or spreading to other packages in a rack. The off site consequences are in by far the largest number of cases a smoke plume, with possibly toxic smoke depending on warehouse content.

In some cases, drums rupture in the fire and can cause domino effects, if the storage walls or roof are not sufficiently strong to hold the drum back. (corrugated steel roofs in a warehouse will often collapse in the fire). Drums have been observed to travel up to 80 m from their original location, and to pass almost unobstructed through the warehouse roof.

Figure 20.1 A fire in a chemicals drum storage.

Determination of frequencies of warehouse fires is unusually difficult, because not only is the population of warehouses and storages difficult count, but also the definition of a chemical product warehouse is difficult to define. Many food processing plants, for example, have a considerable storage of solvents and oils used for flavourings, and most blending plants for soap, shampoo etc. have chemicals and solvents for aroma substances. Substances such as lemon oil may be regarded as chemicals and hazardous substances, but elsewhere are regarded as food ingredients.

Very little systematic information is available on the causal patterns of accidents in warehouses and storage. From audit information it is known that crowding and high levels of fork lift traffic are indicators of high accident rates, and that risks are reduced by fire protection systems. The quality of electrical equipment in classified areas, proper classification, and proper segregation of substances according to hazard class, all affect .

Table 19.1 gives the results of observations from one company with a good safety management system, and which handled about 400,000 drums of chemicals per year through its warehouse.

		Releases per year	Releases per drum
Drums handled per year	400,000		
Transport operations per drum	2.2		
Observations	32		
Releases	20		0.000025
Release with person present	17	0.85	
Ignitions, including unwanted reactions	4	0.2	0.000005
Ignitions with no one present	1	0.05	
Releases due to human error	10	0.38	0.0000125

Table 20.1 Frequency of handling incidents in a chemical warehouse

20.1 Bulk warehousing of fertilizers

Fertilizers at production plants are generally stored initially in piles, in covered storage. The amounts can be tens of thousands of tones. After the initial storage fertilizer may be prilled, (turned into small spheres) and coated. Even if not, the fertilizer is generally bagged typically in 50 to 70 kg plastic sacks.

NPK (nitrogen phosphorous potassium) fertilizer may be based on ammonium sulphate, but more often is based on either potassium or ammonium nitrate. It may burn if it is ignited in contact with organic material, such as wooden pallets or sacking, will generate nitric oxide.

Ammonium nitrate fertilizer, if involved in a fire, can explode extremely widely with serious damage at large distances.

The frequency of fires in bulk fertilizer storage was estimated using the number of installations as listed in the US RMP data base, and the fires for the same installations as derived from the I Chem E accident data base and MHIDAS. The result found was a frequency of fires and explosions in ammonium nitrate storage of greater than $30*10^{-4}$ per year (some of the facilities may not actually have ammonium nitrate stored, so that the actual frequency for susceptible storages may be higher. From inspection of the data, it is not possible for the frequency to be as much as twice as high).

From audits it is fairly easy to see the difference between a well run and a poorly run storage, but there are few technical differences known which can differentiate easily between high risk and low risk storages. Weighting factors are therefore not given here.

Figure 20.2 Fire in bulk fertiliser storage

20.2 Release and accident frequencies

Releases of liquids in storage are of two types. The dominant cause in most storages is release due to damage in handling. Additionally, releases occur due to corrosion of containers, usually leading to small leaks. If the leak occurs in a pile of barrels, however, the amount released can be significant, before the leak is noticed and something is done about it.

Hymes and Flynn (1992) (reproduced by Willis) reported between 1 and 5 fire incidents annually in warehouses, mostly controlled by manual intervention. 1 in 250 developed into reportable incidents, with 1 in 750 becoming serious fires, with roof collapse.

21 Other Equipment Types

There are several special types of equipment in which large accident types can occur, but which are not widely used in industry. As a result, the population of equipments is small and the statistical basis for determining an accident frequency is limited. Some types are given in the following sections.

21.1 Oil seed extraction plants

Vegetable oil is extracted from oil seed such as soya beans, first by crushing and pressing, and then by extraction with solvent naphtha (usually purified hexane). The extraction is usually carried out in a multi tray extraction vessel with the crushed seed pulp and solvent falling from tray to tray, and steam passing upwards.

After extraction, the solvent is evaporated from the seed oil, the residue of solvent in the oil seed pulp is evaporated off, typically using heated air.

Most extraction plants are operated indoors. Leaks and releases are presumably less frequent in oil seed extraction plants than in other oil refineries. However, because any release indoors will accumulate the hazard becomes much higher. Ignition of leaks from an oil seed extraction plant will in almost all cases lead to an explosion except in the case of the smallest (small flange leaks).

Incident location	date	Type	Cost US $	Fatalities	Injuries	Damage radius m
Glasgow	1971	Dust explosion	500000	0	8	
Chicago	1957	Dust explosion	700000		1	1200
Hamburg	1983	Vapour explosion		2	12	400
Copenhagen	1980	Vapour explosion	17000000	0		400
Antwerp	1986			1	5	

Table 21.1 Oil seed extraction plant explosions

Figure 21.1 Remains of an oil seed plant after a UVCE, Copenhagen 1982

21.2Driers

Rotary kiln driers are often used for drying fertilizer, since the action of the kiln not only produces a dry product, but also forms the product into pellets or prills which are easy to handle and popular with the users of the fertilizer (they flow reasonably freely through fertilizer spreading equipment).

If the fertilizer includes nitrates, and becomes contaminated in any way with organic material there is a good chance that a fire will occur, which can release nitric oxide gas. If the fertilizer contains ammonia nitrate, or ammonium salts and potassium nitrate, the amount of nitric oxide released will be much larger.

The frequency of such accidents was determined from direct experience and from MHIDAS data base. The observed frequency was 0.032 incidents per drier year, though this frequency is based on just two accidents.

21.3 Flare systems

Flare systems have historically been responsible for a surprising fraction of the large accidents, including those at Milford Haven, Wales in 1994, and La Plata, Argentina in 1995. The problem arises if liquids enter the flare line in large quantities, in which case hammer effects can rupture the line. Other accident types involve:

- release of liquefied gas into the flare, with freezing and possible brittle cracking as a result. (not all flare lines of this type are designed for low temperatures).
- blockage of the flare line with ice forming from condensed steam
- liquid entering the flare stack and falling as a “fountain” of burning oil.

- liquid hammering rupture of the line as described above.
- air entering the line and leading to an explosion, due to maintenance errors.

Frequency data for such incidents are very difficult to derive systematically since many such incidents go unrecorded unless they lead to major damage. From incidents investigated by the author (and the accompanying search through operating logs and internal incident reports) the frequency of sufficient liquid entering a flare to cause ejection of burning liquid is 0.015 per flare year, based on four incidents.

21.4 Waste water and slops tanks

Waste water tanks are often considered not to present a major hazard threat. However, more actual explosions have occurred with these than with oil and gasoline tanks. Such tanks do not often figure in risk assessments, because the inventory of flammable material is nominally zero records, but such tanks should be included in risk assessments. One such accident was that at Signal Hill, California, in 1950, when hot oil and water frothed over from a tank and flowed don a hillside, killing two.

In Pernis, Netherlands in 1968, hot oil entered a slops tank, with a large release of vapour as a result, and a subsequent vapour cloud explosion.

Again it is difficult to determine frequencies for such accidents because they are rarely recorded unless they lead to large losses. From personal accident investigations, and by taking inventory of the tanks at these and other similar sites, a frequency of 0.008 per tank year for explosion or violent overflow due to water/oil interaction is estimated, based on three incidents.

21.5 Sewer systems

Sewer systems have proved to be surprisingly hazardous, if volatile flammable liquids are allowed to enter, or if hot water entering causes heavy solvents to evaporate. The MHIDAS data base contained 52 records of such accidents, and many smaller ones are known to the author from direct experience. Determining a frequency for such accidents “per sewer” is virtually impossible, since the size and coverage of a sewer is not deteminate. Therefore in each case of possibility of flammable materials entering a sewer , i.e. most storages and uses of solvents and gasoline, and virtually all transportation of these substances, needs to be analysed in detail to determine the actual local frequency. Figure 20.3 shows the extent of such accidents.

Figure 21.2 Debris after a gasoline vapour explosion in a sewer, Guadelajra, 1994

22 Domino effects

Domino effects occur especially in petroleum and petrochemical plant, and can occur in other plants, serving to convert a small accident into a much larger one. The mechanisms known for domino effects are:

- Release of gas or especially liquefied gas in large quantities from one piece of equipment, followed by a vapour cloud explosion and damage to other equipment.
- A BLEVE in one vessel which causes explosion damage to others, usually with fires and often with further BLEVE's to follow.
- Jet fires from one small release, which impinge on other equipment, and cause BLEVE like accidents.
- Pool fires under vessels, arising from leaks and which lead to overheating and BLEVE like accidents.
- Small and medium sized fires in tank basins, which cause fire induced tank explosions in partially empty tanks.
- Runaway reaction explosions in batch and semi batch reactors and distillation kettles, followed by fires in the production unit.
- Small fires from leaks in a production unit, which melt plastic or rubber hoses, or fibre reinforced plastic piping and vessels, and release more liquid to add to the fire.

Table 22.1 shows a classification of the one hundred largest accidents, published by the Marsh company. The table records which accidents have involved domino effects, which of these have been caused by an initial small release (i.e. escalations), and those in which the accident has been increased in size but not in scale by the domino effects.

As can be seen, domino effects involving escalation from small accidents have been important in the causal pattern in about 44% of large accidents, and have increased the amount of damage in accidents which started as large releases in about 93%. In other words, fire and explosion risk analyses which don't take domino effects and escalation into account will be severely defective.

One simple way of dealing with this in a risk calculation would be:

a) to take the initially calculated frequency of large accidents and double it

b) to calculate the damage area for large accidents, (BLEVE's, UVCE's, building explosions and fires) and to increase the material damage values for the accident.

This crude approach is simple enough but is not in the spirit of the work in this report. The objective here is to try to reflect the real risk, to give credit for risk reduction measures, and to penalize plants where risk factors are very high. Some plants, for example, have domino effect risk levels up to a factor 100 higher than average. Just taking an industry average value for escalation would grossly underestimate risk for example, in plants with very close equipment spacings.

Detailed approaches to domino effect calculation methods are given in ref. 22.1 and ref. 22.2. These methods can be used but require computer aids to support the calculation (that is, if the calculation is to be completed in a reasonable amount of time).

Table 22.1 gives the a record of the largest process plant accidents, including location and date, plant and equipment type, release and explosion mechanism, and losses. The column “escalation” indicates whether the accident resulted from a small fire which developed into a large accident. The column “Domino” indicates whether there was a domino effect or not. The column “Domino mechanism” indicates the type of accident (fire, explosion etc) which initiated the domino effect.

UVCE is unconfined vapour cloud explosion

FIE is fire induced explosion

BI is business interruption

Location	Date	Plant type	Unit	Equipment	Mechanism	Accident	Loss m$	BI loss m$	Esca-lation	Domino	Domino mechanism
Deer Park, Texas	22/06/1997	refinery		Valve	Mechanical failure	UVCE	?		N	Y	UVCE
Okinawa Japan	26/04/1996	refinery	HDS furnace	Furnace tube	tube rupture	Fire	12		Y	N	Fire
La Plata, Argentina	12/01/1995	refinery	flare KO	Flare line	line rupture	Fire	18.9		N	Y	Fire
Cilacap, Indonesia	24/10/1995	refinery	FR tank	Tank seal	tank FR	Fire	34.7		Y	Y	Fire
Rousebville, Pennsylvania	16/10/1995	refinery	Pipe racks	Piping	Leak, fire	Fire	42		Y	Y	Fire
Pembroke, UK	24/07/1994	refinery	flare KO	Flare line	line rupture	UVCE	83.7		N	Y	Explosion
Belpre, Ohio	27/05/1995	refinery		Storage		Fire	10		Y	Y	Fire
Kawasaki, Japan	25/02/1994	refinery	Turbine expander	Turbine seal	mech fail	Fire	37.8		N	Y	Fire
Baton Rouge, Louisiana	02/08/1993	refinery	Delayed coker	Pipe elbow	Pipe rupture	Fire	72.4		Y	Y	Fire
La Mede, France	09/11/1992	refinery	FCC light end		Pipe rupture	UVCE	260		N	Y	Explosion
Sodegaura, Japan	16/10/1992	refinery	HDS unit		Pipe rupture	UVCE	78.3		N	Y	Explosion
NRW, Germany	10/12/1991	refinery	Hydrocracker		Pump seal	UVCE	57.1		Y	Y	Explosion
Beaumont, Texas	03/11/1991	refinery	Crude unit		Pump seal	Fire	17		Y	Y	Fire
Sweeny, Texas	13/04/1991	refinery	RDS		?	Explosions	26	225	N	Y	Explosion
Lake Charles, Louisiana	03/03/1991	refinery	FCC light end		Op error	Steam explosion	26	44	N	N	Explosion
Port Arthur, Texas	12/01/1991	refinery	Crude unit		Pump seal	Fire	28.8	76	Y	Y	BLEVE
Ras Tanura Saudia Arabia	30/11/1990	refinery	Kero/gas oil		?	Fire	37.1	20	?	Y	Fire
Chalmette, Louisiana	11/03/1990	refinery	Hydrocracker	Heat exchanger	Shell crack	UVCE	23.2		N	Y	Explosion
Baton Rouge, Louisiana	24/12/1989	refinery	Light ends	Pipe	Pipe rupture	UVCE	81.3		N	Y	Explosion
St. Croix, Virgin Islands	18/09/1989	refinery	Storage	Crude tanks	Wind damage		73.2		N	N	Wind damage
Martinez, California	04/09/1989	refinery	Hydrotreater	Pipe	Leak, fire	UVCE	56.9		N	Y	Explosion
Richmond, California	10/04/1989	refinery	Hydrocracker	Pipe, 2 inch	Jet fire	Jet fire	102.9		Y	Y	Jet fire, column collapse
Port Arthur, Texas	08/06/1988	refinery	Propane storage	Pipe, 6 inch	Pipe rupture	UVCE	13.3		N	Y	Explosion
Norco, Louisiana	05/05/1988	refinery	FCC light end	Pipe elbow	Pipe rupture	UVCE	308.2		N	Y	Explosion
Grangemouth, UK	22/03/1987	refinery	Hydrocracker	Separator	Overpressure	Equipment explosion	98.1		N	Y	Explosion

Location	Date	Plant type	Unit	Equipment	Mechanism	Accident	Loss m$	BI loss m$	Esca-lation	Domino	Domino mechanism
Las Piedras, Venezuela	13/12/1984	refinery	HDS	Separator	Pipe rupture	Jet fire	81.9		Y	Y	Jet fire
Ft McMurray, Canada	15/08/1984	refinery	Fluid be coker	Pipe, 10 inch	Pipe rupture	UVCE	100.3		N	Y	Explosion
Romeoville, Illinois	23/07/1984	refinery	Amine	Column	Crack	UVCE	252.1		N	Y	Fragment, explosion, BLEVE
Milford Haven, UK	30/08/1983	refinery	Crude tank	Tank roof	Crack	Boilover	15.2		N	Y	Boilover
Avon, California	07/04/1983	refinery	FCC	Slurry line	Erosion	Jet fire	67.1		N	Y	Jet fire
Shuaiba, Kuwait	20/08/1981	refinery	Pump manifold	?	?	Fire	156		Y	Y	Fire
Borger, Texas	20/01/1980	refinery	Alkylation	Flare system	Freeze, overpressure	UVCE	59.7		N	Y	Explosion
Geelong, Australia	11/12/1979	refinery	Crude unit	Pump	Mech rupture	Fire	20.9		Y	Y	Fire
Deer Park, Texas	01/09/1979	refinery	Tanker offloading		Lightning	Tank explosion	126.5		N	Y	Explosion
Texas City, Texas	21/07/1979	refinery	Alkylation	Pipe elbow 12 in.	?	UVCE	42.9		N	Y	Explosion
Denver, Colorado	03/10/1978	refinery	Polymerisation	Reboiler	Pipe failure	UVCE	43.8		N	Y	Explosion
Texas City, Texas	30/05/1978	refinery	Tank farm	?	?	Fire	110		N	Y	BLEVE, fragments
Baton Rouge, Louisiana	17/10/1977	refinery	FCC		Short circuit	Fire	20.4		Y	Y	Fire
Philadelphia, Pennsylvania	17/08/1975	refinery	Storage tank	Tank	Vapour release	Flash fire	31.7		N	Y	Tank fire
Avon, California	16/03/1975	refinery	Coker	Coker vessel	Implosion !	Fire	25.3		N	Y	Fire
St. Croix, Virgin Islands	24/08/1973	refinery	HDS	Pipe	Crack	Fire	31.5		Y	Y	Running fire
Billings, Montana	14/08/1972	refinery	Alkylation	Strainer	Maintenance error	Flash fire	16		Y	Y	BLEVEs
Linden, New jersey	05/12/1970	refinery	Hydrocracker	Reactor	Overheating	Vessel explosion	104.6		N	Y	Explosion
Pernis, Netherlands	20/01/1968	refinery	Slop tank	Tank	Hot and cold mixing	Froth over	129.4		N	Y	Boilover
Lake Charles, Louisiana	08/08/1967	refinery	Transfer pipeline	Valve	Mechanical failure	UVCE	84.8		N	Y	Fire, BLEVE
Port Neal, Iowa	13/12/1994	fertiliser	ammonium nitrate	?	Solid phase explosion	Explosions	129.6	60	N	Y	Explosion, fragments, ammonia release
Baytown, Texas	17/10/1994	PE	All units		Flooding						
Baton Rouge, Louisiana	08/08/1994	Petrochm	Ethylene	Fractionator		Fire	27		N	N	

Location	Date	Plant type	Unit	Equipment	Mechanism	Accident	Loss m$	BI loss m$	Esca-lation	Domino	Domino mechanism
Belpre, Ohio	27/05/1994	Petrochm	Rubber	Reactor	Runaway	Explosions	108		N	Y	Explosion, Fire
Westlake, Louisiana	28/07/1992	Petrochm	Urea	Reactor	Weld crack	Vessel explosion	28		N	N	
Tarragona, Spain	08/05/1992	Petrochm	Ethylene oxide	Reactor	Seal failure	UVCE	14.6		Y	Y	Explosion
Alvin, Texas	13/01/1992	Petrochm	Chemical	Feed tank	Runaway	Vessel explosion	36.2		N	N	
Dhaka, Bangla Desh	20/06/1991	Petrochm	Urea	Reactor	Weld crack	Vessel explosion	80.2		N	Y	Explosion
Sterlington, Louisiana	01/05/1991	Petrochm	Nitroparaffin	Compressor	Leak, small fire	FIE	118.7		Y	Y	Explosion
Seadrift, Texas	12/03/1991	Petrochm	Ethylene oxide	Column	?	Vessel explosion	90.4	90	N	Y	Fragments, fire
Coatzacoalcos, Mexico	11/03/1991	Petrochm	Pipe rack	Pipe	Leak	UVCE	103.2		Y	Y	Explosion
Nagothane, India	06/11/1990	Petrochm	LDPE	Transfer pipe	Leak	UVCE	25.5		N	Y	Explosion
Channelview, Texas	05/07/1990	Petrochm	Propylene oxide	Wastewater tank	Tank explosion	Tank explosion	13.9		N	Y	Explosion
Pasadena, Texas	23/10/1989	Petrochm	HDPE	Valve	Maintenance error	UVCE	796.5		N	Y	BLEVE, Fire
Morris, Illinois	07/06/1989	Petrochm	depropaniser	Flare valve open	Maintenance error	UVCE	38.4	55	Y	Y	Explosion
Antwerp, Belgium	07/03/1989	Petrochm	Aldehyde	Piping	Fatigue crack	Lagging fire	90.9	270	Y	Y	Fire, Explosion
Pampa, Texas	14/11/1987	Petrochm	Gas treatment	Inlet separator	Hydraulic overpressure	UVCE	269.1	140	N	Y	Explosion
Pascagoula, Mississippi	15/06/1986	Petrochm	Aniline	Column	Overheating	Vessel explosion	12.9		N	Y	Fragments, fire
Priolo, Italy	19/05/1985	Petrochm	Ethylene	Column	Leak, Ssmall fire	Fire	85.2		Y	Y	Jet fire
Duluth, Minnesota	06/05/1982	Petrochm	Fumaric acid	Grinding	Dust explosion	Dust explosion	20.2		N	Y	Explosion, fire
Edmonton, Canada	18/04/1982	Petrochm	LDPE	Compressor	Instrument pipe break	Confined explosion	30.2		Y	Y	Explosion, fire
Philadelphia, Pennsylvania	09/03/1982	Petrochm	Phenol	Holding tank	Overheating	Liquid phase explosion	36		Y	Y	Fire
New castle, Delaware	21/10/1980	Petrochm	PP	Reactor	Maintenance error	UVCE	101.9		N	Y	Explosion
Cactus, Mexico	26/07/1996	Gas	Cryogenic	Pump	Maintenance error	UVCE	252.5		Y	Y	Explosion
Ras Tanura Saudia Arabia	15/08/1987	Gas	depropaniser	Flange	Flange fail	UVCE	75		Y	Y	Explosion

Location	Date	Plant type	Unit	Equipment	Mechanism	Accident	Loss m$	BI loss m$	Esca- lation	Domino	Domino mechanism
Basile, Louisiana	30/09/1984	Gas	Absorber	Drain line	Pipe failure	UVCE, jet fire	39.6		Y	Y	Explosion, Column collapse, jet fire
Bontang, Indonesia	18/04/1983	Gas	Cryogenic	Heat exchanger	Overpressure	OP explosion	68.5		Y	Y	Projectiles
Abqaiq, Saudi Arabia	15/04/1988	Gas	Transfer pipeline	Pipeline	Corrosion	UVCE	107.4		Y	Y	Projectiles
Abqaiq, Saudi Arabia	11/05/1977	Gas	Transfer pipeline	Crude oil line	Pipe failure	Jet impact	117.2		Y	Y	Jet fire
Umm Said, Qatar	03/04/1977	Gas	Cryogenic	Tank	Weld failure	Fire (propane)	164.2		N	Y	Pool fire
Melbourne, Australia	21/08/1991	Terminal	Storage	Tank		Tank explosion	12.7	40	N	Y	Explosion, projectiles
Denver, Colorado	25/11/1990	Terminal	Storage	Pump	Leak	Pool fire	37.1		Y	Y	Fire
Naples, Italy	21/12/1985	Terminal	Storage	Tank	Overfill	Flash fire	55		Y	Y	Flash fire, projectiles
Mont Belvieu, Texas	05/11/1985	Terminal	Pipeline	Pipe	Maintenance error	UVCE	56.3		N	Y	Explosion
San Juan Ixhuatapec, Mexico	19/11/1984	Terminal	Pipeline	Pipe	Rupture	BLEVE	26.3		N	Y	Jet fire, BLEVE
Newark, New Jersey	07/01/1983	Terminal	Storage	Tank	Overfill	UVCE	47.9		Y	Y	Explosion
Cincinnati, Ohio	19/07/1990	Resins	Acrylic	Reactor	Solvent vapour	UVCE	26.7		Y	Y	Explosion
Urdingen, Germany 14/2/89	14/02/1989	Paint	Resin unit	Reactor	Runaway	UVCE	47.2	85	N	Y	Fire
Scunda, South Africa	30/01/1989	Synfuel	Quench	Pipe	Corrosion	Fire	15.3		Y	Y	Fire
Prudhoe Bay, Alaska	26/05/1983	Oil	NGL	Surge drum	Rupture	OP explosion	43.3		N	N	
Brooks, Alberta	26/02/1980	Gas	Pipeline	Valve		Gas jet	68.4		N	Y	Jet reaction, Jet fire
St John, New Brunswick	09/06/1998	refinery	Hydrocracker	Feed heater		Fire	62.4				
Berre l'Etang, France	06/10/1998	refinery	Crude unit	Air cooler	Corrosion	Jet fire	22		Y	Y	Jet fire
Richmond, California	25/03/1999	refinery	Hydrocracker	Valve		UVCE	87		Y	Y	Explosion

Location	Date	Plant type	Unit	Equipment	Mechanism	Accident	Loss m$	BI loss m$	Esca-lation	Domino	Domino mechanism
Korfez, Turkey	17/08/1999	refinery	Crude unit	Chimney	Earthquake	Fire	200		N	Y	Collapse
Siracha, Thailand	02/12/1999	refinery	Storage	Tank	Overfill	Fire	25.8		Y	Y	Fire
Wuppertal, Germany	08/06/1999	Pesticide	Reactor	Reactor	Runaway	Reactor explosion	75		N	Y	Explosion
Victoria, Australia	25/09/1998	Gas	depropaniser	Heat exchanger	Rupture	UVCE	200		N	Y	Explosion
Ras Gharib, Egypt	10/05/1998	Terminal	Storage	Tank	Lightning	Fire	30		N	Y	Fire
Port Heriot, France	10/06/1987	Terminal	Storage	Pump	Maintenance error	Fire			Y	Y	FITE
Delaware City, Delaware	17/07/2001	refinery	Storage	Acid tank	Corrosion	Explosions	50		Y	Y	FITE

22.1 References

1. Marsh Mclennan, 100 Largest Accidents, 1985 and 2002 editions

23 Validation

In order to check the methods described her, a number of validation studies were cariied out, for equipments for which a global release frequency is known.

23.1 RMP Data

An elementary requirement is that the method reproduces the original data on which it was based. The first table, 23.1, shows the detailed calculation sheet for piping, with all the susceptibilities values set to 1, and all the safety measure unavailabilities as in table 7.9. The resulting release frequencies are compared with the typical values from table 7.8. As can be seen, the original values are reproduced with moderate accuracy which indicates only that the calculations are internally consistent. The differences lie in assessment of susceptibilities for the population of pipes as a whole, and the assessment for a "typical" pipe.

Unit type	Total release frequency per m. year	Small hole, < 5 mm	Medium hole 5 to 25 mm	Large hole 25 to 100 mm	Very large hole, > 100 mm and rupture
Typical data					
≤3 inch pipe	$129*10^{-6}$	$55*10^{-6}$	$52*10^{-6}$	$22*10^{-6}$	NA
>3 inch pipe	$49*10^{-6}$	$18*10^{-6}$	$17*10^{-6}$	$6.3*10^{-6}$	$7.3*10^{-6}$
Predicted					
≤3 inch pipe	$134*10^{-6}$	$49*10^{-6}$	$49*10^{-6}$	$35*10^{-6}$	
>3 inch pipe	$60*10^{-6}$	$25*10^{-6}$	$16*10^{-6}$	$10*10^{-6}$	$8.4*10^{-6}$
predicted/ typical					
	103.88%	89.09%	94.23%	159.09%	
	122.45%	138.89%	94.12%	158.73%	115.07%

Table 23.1 Comparison between prediction algorithm and typical data

More interesting are the calculations for refinery crude unit, ammonia, chlor alkali and light ends piping in tables 23.2 . Susceptibility values have been selected following the algorithm in section 7.10 for a "typical" refinery and petrochemical complex. These plants operate a generally good standard of operation, but have no special programs such as intensive risk based inspection. The table compares predicted values with the observed values from section 7.3. The correspondence is not perfect. The reason for this can be traced to the fact that the data in section 7.5 are for all piping diameters pooled together, whereas the calculation sheets are for typical pipes. For the ammonia piping the correspondence is very good (coincidentally, the actual correspondence is better than the uncertainty would on aveage allow), except for medium size holes.

For the crude unit, the overall correspondence is reasonable, with a small discrepance for medium and large hol sizes discrepancy, and with about a factor 2 overprediction for ruptures.It is clear that the actual small release frequency is under reported in the RMP data. This is in accordance with the RMP reporting criteria, because small releases would not cause off site consequences.

For the alkylation plant, the match is reasonable except for the large hole size frequencies, which are overpredicted by about a factor of 3, and rupture, which is underpredicted by about 40%. Again, it is the judgement of hole sizes which affects the prediction most.

For the LPG storage, it is clear that the actual small release frequency is under reported. This is in accordance with the RMP reporting criteria, because small releases would not cause off site consequences.

Unit type	Total release frequency per m. year	Small hole, < 5 mm per m. year	Medium hole 5mm<x<25 mm per m. year	Large hole >25 mm per m. year	Very large hole, > 100 mm and rupture per m. year
Ammonia and fertiliser	$54* 10^{-6}$	$27* 10^{-6}$	$4* 10^{-6}$	$15* 10^{-6}$	$8* 10^{-6}$
Predicted	$55* 10^{-6}$	$22* 10^{-6}$	$9.2* 10^{-6}$	$15* 10^{-6}$	$9.5* 10^{-6}$
Refinery crude unit	$89 * 10^{-6}$	$6.9 * 10^{-6}$	$21 * 10^{-6}$	$31 * 10^{-6}$	$31 * 10^{-6}$
Predicted	$142 * 10^{-6}$	$70 * 10^{-6}$	$25 * 10^{-6}$	$30 * 10^{-6}$	$17 * 10^{-6}$
Alkylation unit	$134* 10^{-6}$	$66* 10^{-6}$	$54* 10^{-6}$	$8* 10^{-6}$	$8* 10^{-6}$
Predicted	$136* 10^{-6}$	$57* 10^{-6}$	$36* 10^{-6}$	$28* 10^{-6}$	$5.1* 10^{-6}$
LPG storage	$42* 10^{-6}$		$9.8* 10^{-6}$	$22* 10^{-6}$	$9.8* 10^{-6}$
Predicted	$60* 10^{-6}$	$20* 10^{-6}$	$16* 10^{-6}$	$24* 10^{-6}$	$12* 10^{-6}$
Chlor Alkali plant	$280* 10^{-6}$	$109* 10^{-6}$	$77* 10^{-6}$	$31* 10^{-6}$	-
Predicted	$287* 10^{-6}$	$155* 10^{-6}$	$81* 10^{-6}$	$51* 10^{-6}$	

Table 23.2 Predicted verses observed values for over 3 inch piping failure frequencies with different sevices.

Table 23.3 and 23.4 give values for a refinery pressure vessel (a crude unit distillation receiver) and for a cone roof kerosene tank, with appropriate values taken from real life conditions. The release frequencies roughly match the corresponding RMP data. More interesting, the actual values for the refinery, drawn from 30 years of operating experience are shown alongside.

Table 23.3 shows a reasonable comparison except for the small releases and rupture. As above, there may be an absolute underreporting in the RMP data, because small and medium releases in a crude unit would be unlikely to give offsite consequences. The overall frequency is dominated by small and medium size internal corrosion releases

Table 23.4 for a gasoline storage tankshows a nearly perfect comparison, but note that this is a comparison with the typical data used as the basis for determining thhe detail calculation reference.

	Hole sizes				
Equipment type	Failure freq. pr. yr	< 10 mm	10-25 mm	25-100 mm	>100 mm
Crude unit receiver, RMP	4.50E-04	2.29E-04	1.76E-05	7.06E-05	1.76E-05
Predicted	12.0E-04	8.9E-04	3.04E-04	2.34E-05	1.30E-06
Observed, 4 refineries, 420 vessel years	47.6E-04				

Table 23.3 Crude unit receiver vessel failure frequency comparison.

	Small leaks up to 25 mm	Medium leaks up to 50 mm	Large leaks	Rupture, catas-trophic	Fire, tank top	Fire, tank basin	Explo-sion
Typical, solvent, fuel	4E-3	1E-3	2.0E-4	1E-5	-	3E-3	9E-4
Predicted, gasoline	1.3E-03	9.5E-04	7.6E-04	7.7E-06	5.0E-05	4.7E-04	2.1E-04

Table 23.4 Comparison, typical and prdicted values, gasoline tank

23.2 HSE data

Data from UK HSE in ref. 12 has been refered to frequently in the earlier chapters, but was not used directly in determining the typical values for release frequencies, only in helping to understand dependencies and tendencies. The data can therefore be used as a comparison basis for the calculation methods described here.

Table 23.5 gives an assessment for a 4 inch high pressure pipe used in crude oil production. Sweet oil (without excessive sulphur content) is assumed. Susceptibilities are set for a typical offshore oil pipe.

Source	Failure Mode	Failure Rate, HSE	Predicted failure rate	Predicted / HSE value
HSE offshore	Pipe 3 to 11 inch, total	5.9E-5	7.6E-5	1.3
(ref. 7.5)	Small hole < 10 mm	4.4E-5	4.3E-5	0.97
	Medium hole 10 to 25 mm	2.9E-6	1.4E-5	4.8
	Large hole 25 to 100 mm	2.3E-6	1.5E-5	6.5
	Very large hole, > 100 mm & rupture	5.9E-6	5.2E-6	0.88

Table 23.5 Comparison with HSE data for 4 inch pressure pipe

The comparison is not good for medium and large holes but is reasonably good overall, and for small holes and rupture. The HSE data will in any case be preferred for off shore projects, but it appears that the causal distribution may be useful

The assumptions for these calculations are:

- the environment has potential for serious external corrosion
- the external corrosion inspection and painting program thorough.

A more extensive comparison is for an oil/gas/water separator, equipment which is very common in oil production. Table 23.6 gives an analysis, with the same assumption as for piping alone. The separator is regarded as a pressure vessel with a total of 30 m of product piping, 10 m of small piping, 30 flanges, two control valves, and a pair of safety relief valves. The piping and flange contributions for the vessel itself were set to zero, and the piping and flange contributions calculated separately and added to the total. The comparison is between the prediction and UK HSE releases frequency data for horizontal separators.

		Hole size distribution				
Equipment type	Failure frequency per year	< 10 mm	10-25 mm	25-50 mm	50 mm to rupture	
Pressure vessel, separator, hor., HSE	2.21E-03	4.86E-04	0.00E+00	0.00E+00	4.86E-04	
Predicted	5.89E-03	8.87E-04	2.2E-03	2.2E-03	6.88E-04	
Ratio predicted/HSE	2.7	1.8			1.4	

Table 23.6 Comparison for a oil and gas separator

The comparison for small and very large to rupture releases agrees reasonably, considering the different sources. The HSE failure rates are based on few cases for this type of equipment, so that there is no data for medium and large holes, and no comparison of frequencies is possible, other than to say that it will be poor. This also affects the overall comparison.

23.3 Release frequency for an LPG storage vessel

BLEVE frequencies have been determined for LPG storage based on world wide data (see Vol. 1, ch.5) To provide a basis for comparison, an actual risk assessment must be made, which takes into account failure in the storage vessel itself, associated piping, ESD valves, safety relief valves, and also the probability of ignition.

Such an analysis was carried out, with the following underlying assumption.

- A good standard of protection with deluge systems able to cool the vessel in case of fire
- Above ground, uninsolated construction.
- The level of external corrosion is not especially high
- The level of internal corrosion is very low

Figure 23.1 shows the hazards identified, the safety measures which apply, the frequency of release, the probability of ignition, and the probability of flame impingement on the vessel.

Note that the BLEVE frequency input in table 8.10 was not used as part of the input to the calculation since this would have resulted in double counting.

Total release frequency per year	Small	Medium	Large	Rupture
2.47E-03	1.86E-03	1.85E-04	4.23E-04	4.77E-06

Table 23.7 Frequency of releases from LPG vessel, as calculated

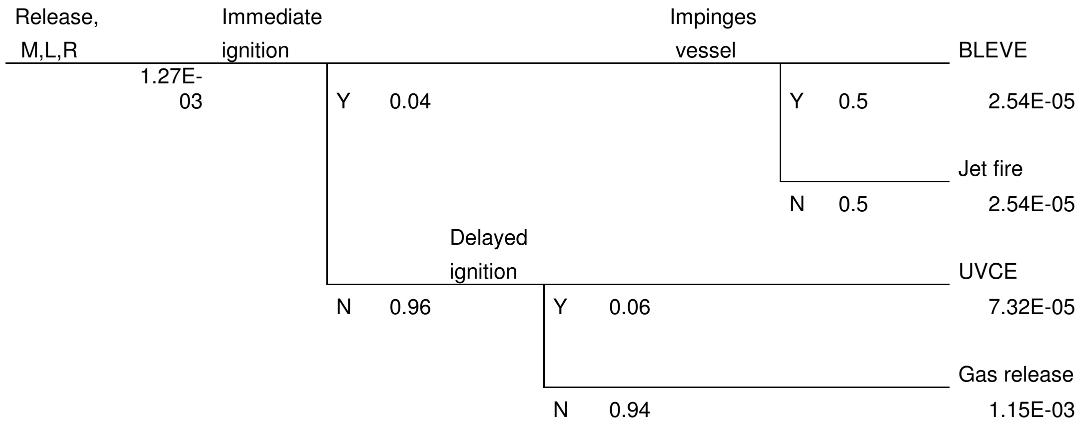


Figure 23.1 Event tree for LPG storage sphere releases

Cox, Lees and Ang (ref. 3)	$1.6*10^{-5}$ BLEVE's per vessel year.
Hurst et al. (ref. 2)	$2.45*10^{-5}$ BLEVE's per vessel year
Predicted here	$2.54*10^{-5}$ BLEVE's per vessel year

Table 23.8 Comparison of BLEVE frequency estimates

Table 23.8 shows the overall result, compared with the world data. The difference of a factor 1 to 1.6 is regarded as acceptable, given that the actual "average condition" for LPG storage around the world is only approximately known, particularly for the data collected in the 1970´s and early 1980´s. (Most of the inspections and surveys which were the basis for the underlying assumptions in thes report were made between 1984 and 2003)

23.4 An application – two fuel pipes

An analysis was made as part of an environmental study for two 10 inch 200 m.gas oil pipes, carrying fuel from a tank farm to a power plant. The lines had a number of branch lines leading to different fuel tanks, and the line was sectioned by two block valves. The frequency assessment calculations are given as table 23.9.

The values calculated do not apply during the early years of operation because there will be little contribution from corrosion during the early years of operation. The values apply fairly precisely (within a factor 2) after 12 years. The values assume 1-2 mm corrosion allowance. Corrosion failure is likely to increase significantly after 25 years depending on the internal corrosion rate

The distribution of causes for the pipelines themselves is shown in the table. Under lagging corrosion and dead leg corrosion (primarily on the drain line stubs) dominate. Under lagging corrosion can be effectively eliminated if periodic inspection under the insulation is performed. For convenience inspection plugs should be fitted, because otherwise there is a risk that inspectors will damage the lagging seal. Under lagging corrosion can be reduced significantly by using high quality cladding (good beading) and silicone sealing. Care is needed at valves, because water tends to enter around valve stems (if valves are insulated) or around the flange seal (if valves are not insulated).

The dead legs feature high on the list of causes too. This source of leaks can be reduced by providing for periodic inspection.

Flange release frequencies are taken from Chapter 7. Valve release frequencies are taken from UK HSE offshore releases data base, with values appropriate for manual valves, 3 to 10 inches.

The most important part of the table is the explanation and justification of the risk reduction measures, which explains why the values are selected. This provides traceability back to the conditions in the plant. It also allows the designer to see the effect which improved integrity engineering can have on safety, in this case about a factor of 10 improvement in the frequency of large releases.

Failure cause	Release size	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Assessed frequency per year	
Internal corrosion	small	400	1.19E-05	0.2		0			0		9.56E-04	Clean petroleum product
Internal corrosion	medium	400	2.71E-05	0.2		0		ESD	0	0.01	2.17E-03	Clean petroleum product
Internal corrosion	large	400	5.53E-06	0.2		0		ESD	0	0.01	4.43E-04	Clean petroleum product
External corrosion	small	400	3.55E-06	0		0		ESD	0	0.01	0.00E+00	Under lagging corrosion applies
External corrosion	medium	400	8.24E-06	0		0		ESD	0	0.01	0.00E+00	Under lagging corrosion applies
External corrosion	large	400	1.61E-06	0		0		ESD	0	0.01	0.00E+00	Under lagging corrosion applies
Drain lines left open	large	4	2.17E-05	1		0		ESD	0	0.01	8.67E-05	Four drains, not normally used maintenance only
Maintenance error	small	400	2.28E-05	0.1		0		ESD	0	0.01	9.10E-04	Very low maintenance line, not routinely opened
Corrosion, no inspection, dead legs	small	4	9.68E-04	1		0		ESD	0	0.01	3.87E-03	Pipe cannot normally be inspected, insulated
Corrosive liquid, or sour gas	small	400	4.84E-04	0		0		ESD	0	0.01	0.00E+00	No, fluid in line is a clean product
Under lagging corrosion	large	400	3.07E-05	1		0		ESD	0	0.01	1.23E-02	Yes
Erosion	medium	400	4.85E-04	0		0		ESD	0	0.01	0.00E+00	No, no solids present
Wrong material	large	2	6.15E-06	0		0		ESD	0	0.01	0.00E+00	No, all carbon steel confirmed and photographed
Lining failure	medium	400	9.69E-05	0		0		ESD	0	0.01	0.00E+00	No, no lining
Support failure	large	400	2.2E-05	0				ESD	0	0.01	0.00E+00	No, high quality supports, inspected
Overheating ++	rupture	2	9.3E-05	0	SV	0	0.05108	ESD	0	0.01	0.00E+00	No, not physically possible, tank provides a buffer
Overpressure, control failure ++	rupture	2	3.0E-04	0	SV	0	0.05108	ESD	0	0.01	0.00E+00	Doubtful, pipe dimensioned for > max pump pressure
Overpressure, gas breakthrough ++	rupture	2	0.001049	0	SV	0	0.05108	ESD	0	0.01	0.00E+00	No, No high pressure gas in system
Overpressure, shut in liquid ++	medium	2	0.007471	0	SV	0	0.05108	ESD	0	0.01	0.00E+00	No, pipeline is insulated, valves almost always open
External fire ++	rupture	2	9.48E-07	0		0		ESD	0	0.01	0.00E+00	No, pipeline is insulated
Weld crack	large	2	4.34E-06	1		0		ESD	0	0.01	8.67E-06	Yes
Hammer ++	rupture	2	0.000159	0		0		ESD	0	0.01	0.00E+00	No, if hammer rupture occurs it is at the end of the line
Weather, freezing ++	medium	2	5.72E-05	0		0		ESD	0	0.01	0.00E+00	No, very little water in pipe - check drain line
Crash, impact ++	large	2	1.11E-05	0.05		0		ESD	0	0.01	1.11E-06	Low probability, crash barriers in place
Vibration fatigue ++	large	2	3.23E-05	0		0		ESD	0	0.01	0.00E+00	No, centrifugal pump, stable flow (Check)
Thermal expansion ++	large	2	3.69E-05	0		0		ESD	0	0.01	0.00E+00	Thermal expansion loops and sliding guides OK, see design error below
Wind load ++	large	2	4.35E-04	0		0		ESD	0	0.01	0.00E+00	No, not physically possible
Wrong substance ++	medium	2	4.85E-06	0		0		ESD	0	0.01	0.00E+00	No, only fuel in tank
Earthquake, landslip, flood ++	rupture	2	1.33E-07	0		0		ESD	0	0.01	0.00E+00	Earthquake very unlikely in Denmark, no real slope, above flood level
Internal explosion ++	rupture	2	9.37E-05	0		0		ESD	0	0.01	0.00E+00	Only with"empty" pipe
Vandalism, third party ++	rupture	2	1.09E-04	0.1		0		ESD	0	0.01	2.19E-05	Yes, but so far low probability in Denmark
Low temperature embrittlement ++	rupture	2	2.65E-06	0		0		ESD	0	0.01	0.00E+00	Not physically possible.
Dropped object ++	rupture	2	3.69E-06	1		0		ESD	0	0.01	7.38E-06	Yes in principle, low probability at actual location
Design error ++	large	2	2.40E-05	1		0		ESD	0	0.01	4.80E-05	Yes, in principle
Total small											5.74E-03	
Total medium											2.17E-03	
Total large											1.29E-02	
Total rupture											3.46E-05	

Table 23.9 Pipe release frequencies for the two pipes, with normal design. ++ indicates failures per pipe section.

Table 23.10 Pipe release frequencies for the two pipes, with special design of cladding and with enhanced inspection. ++ indicates failures per pipe section

Failure cause	Release size	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	
Internal corrosion	small	400	1.19E-05	0.2		0			0			0		9.56E-04	Clean petroleum product
Internal corrosion	medium	400	2.71E-05	0.2		0		ESD	0	0.01		0		2.17E-03	Clean petroleum product
Internal corrosion	large	400	5.53E-06	0.2		0		ESD	0	0.01		0		4.43E-04	Clean petroleum product
External corrosion	small	400	3.55E-06	0		0		ESD	0	0.01		0		0.00E+00	Under lagging corrosion applies
External corrosion	medium	400	8.24E-06	0		0		ESD	0	0.01		0		0.00E+00	Under lagging corrosion applies
External corrosion	large	400	1.61E-06	0		0		ESD	0	0.01		0		0.00E+00	Under lagging corrosion applies
Drain lines left open	large	4	2.17E-05	1		0		ESD	0	0.01		0		8.67E-05	Four drains, not normally used maintenance only
Maintenance error	small	400	2.28E-05	0.1		0		ESD	0	0.01		0		9.10E-04	Very low maintenance line, not routinely opened
Corrosion, no inspection, dead legs	small	4	9.68E-04	0.01		0		ESD	0	0.01		0		3.87E-05	Pipe cannot normally be inspected, insulated
Corrosive liquid, or sour gas	small	400	4.84E-04	0		0		ESD	0	0.01		0		0.00E+00	No, fluid in line is a clean product
Under lagging corrosion	large	400	3.07E-05	0.01		0		ESD	0	0.01		0		1.23E-04	Careful cladding design, special inspection
Erosion	medium	400	4.85E-04	0		0		ESD	0	0.01		0		0.00E+00	No, no solids present
Wrong material	large	2	6.15E-06	0		0		ESD	0	0.01		0		0.00E+00	No, all carbon steel confirmed and photographed
Lining failure	medium	400	9.69E-05	0		0		ESD	0	0.01		0		0.00E+00	No, no lining
Support failure	large	400	2.2E-05	0				ESD	0	0.01		0		0.00E+00	No, high quality supports, inspected
Overheating ++	rupture	2	9.3E-05	0	SV	0	0.05108	ESD	0	0.01		0		0.00E+00	No, not physically possible, tank provides a buffer
Overpressure, control failure ++	rupture	2	3.0E-04	0	SV	0	0.05108	ESD	0	0.01		0		0.00E+00	Doubtful, pipe dimensioned for > max pump pressure
Overpressure, gas breakthrough ++	rupture	2	0.001049	0	SV	0	0.05108	ESD	0	0.01		0		0.00E+00	No, No high pressure gas in system
Overpressure, shut in liquid ++	medium	2	0.007471	0	SV	0	0.05108	ESD	0	0.01		0		0.00E+00	No, pipeline is insulated, valves almost always open
External fire ++	rupture	2	9.48E-07	0		0		ESD	0	0.01		0		0.00E+00	No, pipeline is insulated
Weld crack	large	2	4.34E-06	1		0		ESD	0	0.01		0		8.67E-06	Yes
Hammer ++	rupture	2	0.000159	0		0		ESD	0	0.01		0		0.00E+00	No, if hammer rupture occurs it is at the end of the line
Weather, freezing ++	medium	2	5.72E-05	0		0		ESD	0	0.01		0		0.00E+00	No, very little water in pipe - check drain line
Crash, impact ++	large	2	1.11E-05	0.05		0		ESD	0	0.01		0		1.11E-06	Low probability, crash barriers
Vibration fatigue ++	large	2	3.23E-05	0		0		ESD	0	0.01		0		0.00E+00	No, centrifugal pump, stable flow
Thermal expansion ++	large	2	3.69E-05	0		0		ESD	0	0.01		0		0.00E+00	Thermal expansion loops and sliding guides OK, see design error below
Wind load ++	large	2	4.35E-04	0		0		ESD	0	0.01		0		0.00E+00	No, not physically possible
Wrong substance ++	medium	2	4.85E-06	0		0		ESD	0	0.01		0		0.00E+00	No, only fuel in tank
Earthquake, landslip, flood ++	rupture	2	1.33E-07	0		0		ESD	0	0.01		0		0.00E+00	Earthquake very unlikely in Denmark, no real slope, above flood level

Failure cause	Release size	Number of items or metres	Frequency per item year	Suscept-ibility	Safety barrier 1	Y/N	Risk reduction	Safety barrier 2	Y/N	Risk reduction	Safety barrier 3	Y/N	Risk reduction	Assessed frequency per year	
Internal explosion ++	rupture	2	9.37E-05	0		0		ESD	0	0.01		0		0.00E+00	Only with"empty" pipe
Vandalism, third party ++	rupture	2	1.09E-04	0.1		0		ESD	0	0.01		0		2.19E-05	Yes, but so far low probability
Low temperature embrittlement ++	rupture	2	2.65E-06	0		0		ESD	0	0.01		0		0.00E+00	Not physically possible.
Dropped object ++	rupture	2	3.69E-06	1		0		ESD	0	0.01		0		7.38E-06	Yes in principle, low probability at actual location
Design error ++	large	2	2.40E-05	1		0		ESD	0	0.01		0		4.80E-05	Yes, in principle
Total small														1.90E-03	
Total medium														2.17E-03	
Total large														7.10E-04	
Total rupture														3.46E-05	

23.5 Refinery unit UVCE's and very large fires

Unconfined vapour cloud explosions are quite well registered for European and US refineries and petrochemical plants (ref. 6, 7). Very large fires are also quite well registered (ref. 1 to 5), as part of insurance company publications, as well as several individual collections. Since the number of refineries is known from RMP data, it becomes relatively straightforward to calculate the frequency of very large fires and explosions in refineries, and with a greater degree of uncertainty, in petrochemical plants.

Making predictions which can be compared with the staistics is more problematic, because it involves:

- Defining typical units
- Making release frequency calculations for the releases
- Calculating the extent of vapour clouds, and flammable pools
- Calculating the probability of ignition
- Estimating the probability that a vapour cloud ignition will lead to a UVCE

In effect, a full quantitative risk assessment must be made for each type of process plant considered. The prediction will depend on the UVCE model and the ignition probability model.

In order to make the predictions here a hig quality GIS based process plant risk assessment package (ref. 8), has been applied to an example refinery (based on actual refinery experience, with a few simplifications, and with a layout specified for simplicity, but following good plant layout practice. The UVCE model used is the TNO multi energy model with the GAMES empirically based parameter selection, using a map of congested refinery areas. Three ignition probability models were investigated, the Purple Book values (ref 10.), the Atkins/HSE ignition sorce density model (ref. 11), and a model based on data originally developed for the IFAL method (ref. 12).

Manual application of the release data given in Ch. 7 to 21 would be far too heavy for practical risk analyses of the scale of a refinery (194 vessels and tanks and 240 inter equipment pipe sections are included in the analysis). Instead software for scenario management (SCENEx) was developed. This draws on the above data, and allows scenarios to be generated and release frequency values to be allocated automatically for vesssls, tanks, pumps and piping etc. Default values for susceptibilities and safety measure unavailabilities can be set on the basis of plant type, unit type, equipment type, service type (fluid, temperature, pressure). Each value can also be set manually. The program allows the questions given in the earlier chapters to be displayed and answers recorded, so that the assumptions underlying the calculations can be documented. Full details of the calculation are given in ref. 13 .

In order to be able to provide some details here, an approximate calculation was made for some of the refinery units. For presentation purposes, the calculation was prformed by:

1. Making an inventory of the equipment

2. Calculating release frequencies using the RELBASE values from the earlier chapters.

3. Estimating the fraction of releases that could give releases large and hot enough to give a vapour cloud (using the analysis in ref 13 as a basis)

4. Making an overall event tree for the releases, using probability data from the literature to determine probabilities for ignition, fire fighting success etc

The assumptions underlying the event tree probabilities are standards from the 1980's and early 1990's, so that, for example, very few emergency shutdown valves exist in the units, just those at the battery limits. In the full analyses in ref 13, several event trees are given for each vessel and all the larger pipes. Here, for brevity, just a single summary event tree is given here. This means that some fairly large averaging approximations are made here. The results are nevertheless reasonably compatible with those in ref. 13.

For the crude unit, the event tree is given in figure 23.2 (see below). The unit considered is just the atmospheric column, a prior flash column, a kerosene and gas oil column, three receivers, a fired heater, eleven heat recovery/cooler heat exchangers, three fin fan coolers, associated transfer and reflux pumps, and piping. No desalters are included in the analysis, which may affect the large fire frequencies, but is unlikely to affect the UVCE frequency.

For the alkylation unit, a UOP style design is assumed, with an isobutane and propane feed drum, a reactor vessel, a hydroflouric acid settler vessel, unreacted naphta stripper column, product receiver, HF stripper column, KOH treater, and associated piping and pumps. Details of hydrogen fluoride handling equipment have been omitted here, because the only validation base available. A very high standard of maintenance and inspection is assumed for the plant, even when the standard is assumed to be from the 1980's. The reason for this is that hardly any alkylation plant has ever been operated with a poor standard, staffs who do not maintain a high standard on these plants do not live long. There is a possibility of thermal runaway in an alkylation reactor, due to cooling failure. The probability of this was assessed separately, in a fault tree analysis.

The resulting frequencies for UVCE are compared in table 23.10 with those derived by Fryman in ref. 6. Note that the comparison is very dependent on assumptions about ESD provision, on layout of the plants, and on the ignition model used. Precise correspondence is therefore not to be expected, and one of the values must be regarded as coincidentally good.

Unit type	Frequency of UVCE, per year		
	Statistics from Fyman, ref 6	Predicted	Ratio Predicted/observed
Crude unit	$4.9*10^{-4}$	5.45E-04	1.1
Alkylation unit	$5.1*10^{-4}$	1.56E-04	0.31

Table 23.10 Observed versus predicted UVCE frequencies for refinery units

In all, the prediction seems to be within a factor of 3.

As a cross check, the frequency for a major (prolonged) fire in a refinery crude unit is calculated in the fault tree to be $1.05*10^{-3}$ per year. Cox, Ang and Lees give values for large fires, with over $1 million in losses, of $6*10^{-2}$ per refinery year. With typically 10 to 20 units in a refinery, the resulting frequency for fire in a single unit is $4*10^{-3}$ per year. This comparison is not good, but not surprising considering the number of differences in assumption involved, and the differences in definition of “large fire”.

	Medium or large release, or rupture	Release hot enough to give vapour cloud	ESD	Early ignition	Delayed ignition	Transition to UVCE	Fire fighting effective	Impingement on vessel	Consequence	Frequency
	Frequency per year	Fraction of total	No ESD assumed for 80's, 90's unit PFD	Probability, from IFAL	Probability, from Purple Book, IFAL	Probability from Games model used in ref. 13	Review of cases in ref	Geometric probability (estimate)		
	0.065	0.16	1	0.22	0.28	0.24	0.95	0.2		

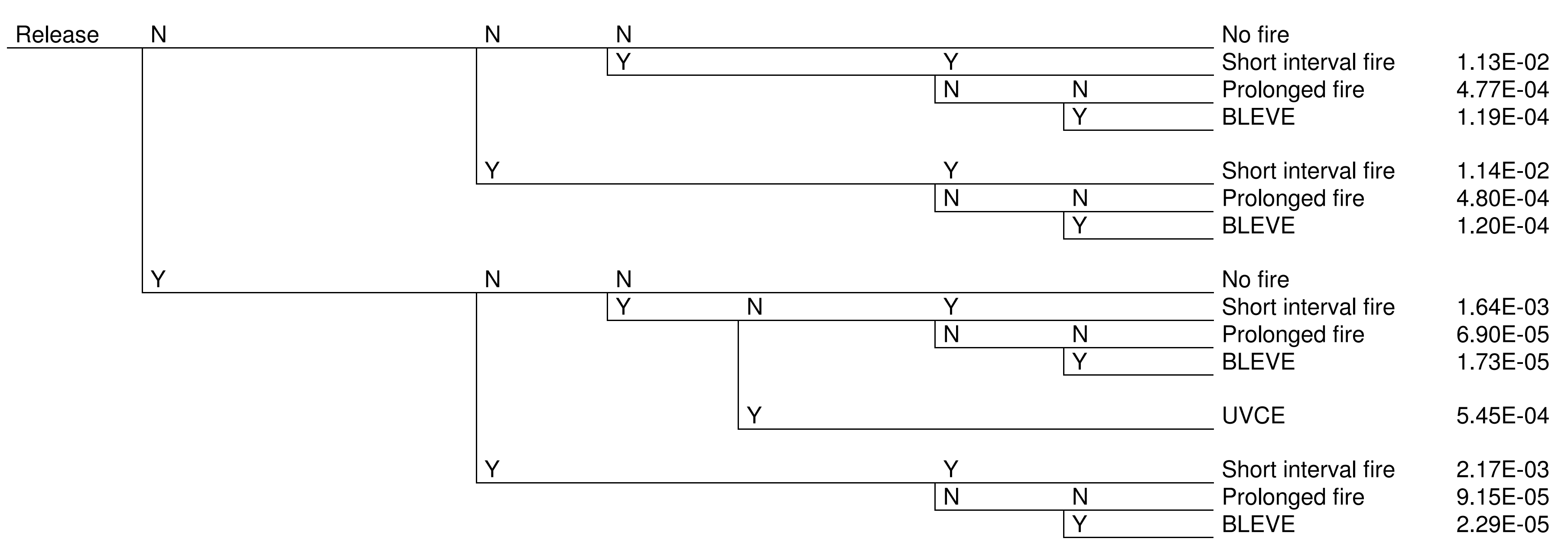


Figure 23.2 Event tree for crude unit large fires and UVCE

Medium or large release, or rupture	Release hot enough to give vapour cloud	ESD	Early ignition	Delayed ignition	Transition to UVCE	Fire fighting effective	Impingement on vessel	Consequence	Frequency
Frequency per year	Fraction of total	ESD on all vessels with light naphtha SIL1	Probability, from IFAL	Probability, from Purple Book, IFAL	Probability from Games model used in ref. 13	Review of cases in ref	Geometric probability (estimate)		
9.26E-02	0.5	0.1	0.22	0.18	0.24	0.95	0.2		

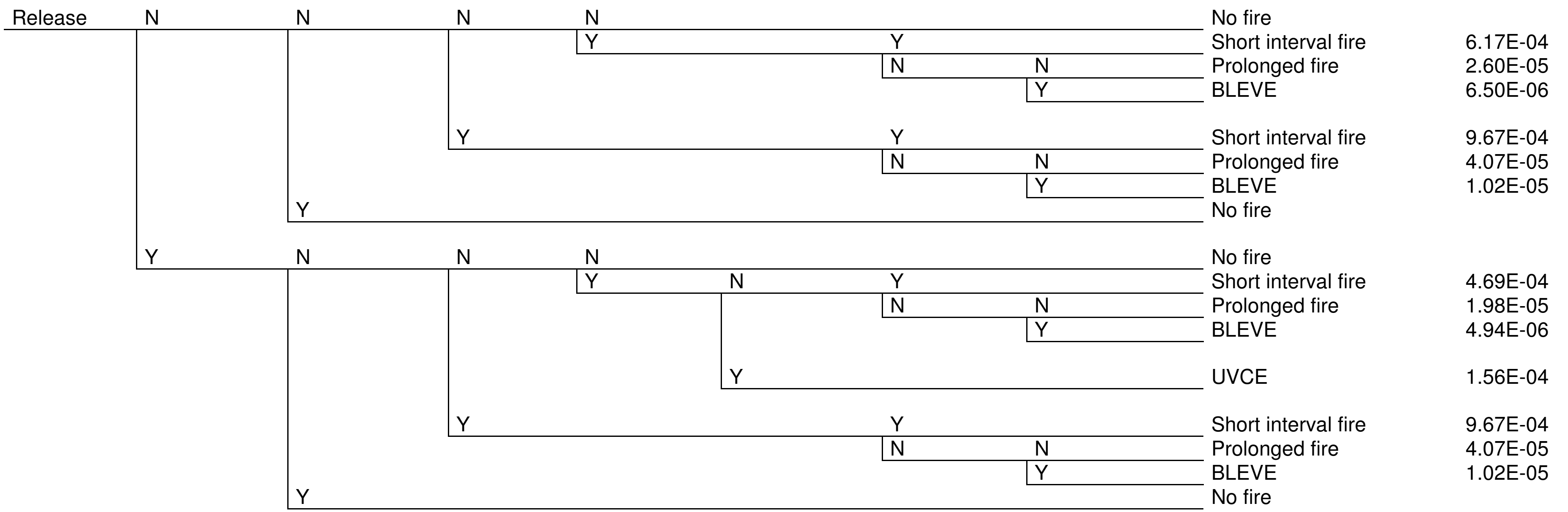


Figure 23.3 Event tree for large fires and UVCE in an alkylation unit

23.6 Ammonia synthesis

Ammonia synthesis trains are very similar in design, with a few variations in absorption of carbon monoxide from the sysnthesis gas, and variations in the degree to which carbon dioxide is recovered in order to synthesise urea. These differences do not affect risk to any large degree. Three designs were analysed in detal in full scale QRA's, yielding a difference in release frequencies of only about 20%. Risks differed by more than this, largely due to differences in layout.

Risks as calculated in QRA were compared with the risks arising from 38 well documented ammonia plants in USA.

It is very difficult to verify the full range of frequency data for an analysis such as this one, particularly for failures such as vessel rupture which are rare. However accident data are available in the RMP data base for 5 years from 38 ammonia trains in USA. The comparison with the total frequency estimate in the QRA is given in table 4.3.

Scenario	Frequency from 38 plants in USA, per year	Frequency as calculated in QRA
Ammonia release, all	0.21, all sources 0.12, excluding the 3 worst plant 0.07 excluding 3 worst plant and ammonia truck loading	0.036
Ammonia releases, large	0.057	0.04
Explosions	0.053	0.033

Table 23.11 Incident frequencies per ammonia train. Note that there were no ammonia loading facilities for the plants in the QRA

The historical values agree with prediction to better than a factor 2 when like is compared with like. This agreement is very good when compared with earlier risk assessment benchmarks. The result should not be overinterpreted however. All that has been done is to take data from US RMP registered ammonia plants and find compaonent equipment failure frequencies, then to carry out a risk assessment and to use this to predict accident frequencies for similar plant, and finally to compare the predicted frequencies with the major hazards part of the original data. This means just that the data are well tuned to ammonia plants in USA. Whether the same data is applicable elsewhere, to ammonia plants run under different safety management regîmes is a matter for judgement. The assumption algorithms given in chapters 7 to 20, though should provide some aid in adapting the data to new circumstances. Note that the frequencies as calculated assume standard ammonia plant safety measures on older plant i.e. the numbers as calculated are before application of modern risk reduction techniques such as release minimisation ESD's (i.e. only battery limit ESD's for plant sections), non SIL rated instrumentation and ESD, and no measures such as water curtains or ammonia suppression deluge.

23.7 Practical application

The methods developed as part of this project are in practice quite heavy, at least if they applied to a plant consisting of much more than a few tanks and pipes. Since the intention is tht the work can be applied even to large petrochemical plant, it is obvious that use of the data requires some degree of automation.

Additionally, it is the case that there are so many adjustable parameters that a user can produce almost any result. If the underlying assumptions are not well documented, quality control of the results becomes a very burdensome task.

To overcome these problems, the data given here has been transferred into what I have called, (with some exaggeration), an intelligent data base. The data base (RELBASE) allows the names, locations, type, service, and similar data to be recorded for individual plant equipments and pipe spools. The generic values for release frequencies can then be looked up automatically. (Figure 23.4) Assumptions (the answers to the algoritm questions) are also looked up automatically, with conservative values. It is possible for the user to provide his or her own sets of generic assumptions, to be selected, so that assumptions for example for different service types such as clean product, sour service etc. can be stored and retrieved. Assumptions concerning company practice, such as inspection intervals etc. can be set for a unit as a whole, for the whole plant, or for individual equipments. Assumptions can be selected based on service type, pressure etc. and can be selected by pattern matching on unit parameters.

It will often be the case that generic values apply "in general", but that there are a few exceptions. Such exceptions can be recorded explicitly into the database, with clear marking, so that exceptions are highlighted. Fields are provided as to allow written justification of the exception. The assumptions also be related to the company safety critical equipment and systems administration, providing tracability between underlying assumptions, calculated values, and demonstration of performance. (figure 23.5). The assumptions and calculations can be printed or filed in certificated form, which can then be used in audits of risk assessments.

The data base is linked to a risk analysis scenario management system (SCENIX), which provides automated generation of risk assessment scenarios, again with generic scenario event sequences, and with an automated generator for safety barrier/butterfly diagrams. The scenarios can of course be supplemented manually, for example from HAZOP or HAZID studies, so that special accident causes can be taken into account. The scenarios can be passed directly to a QRA risk calculation package.

The tool allowed the potential accident scenarios in ref. 13 to be generated with about 6 man days of effort. In other words, despite the additional detail in selection of release frequencies, the overall risk assessment could be made in a relatively effective fashion.

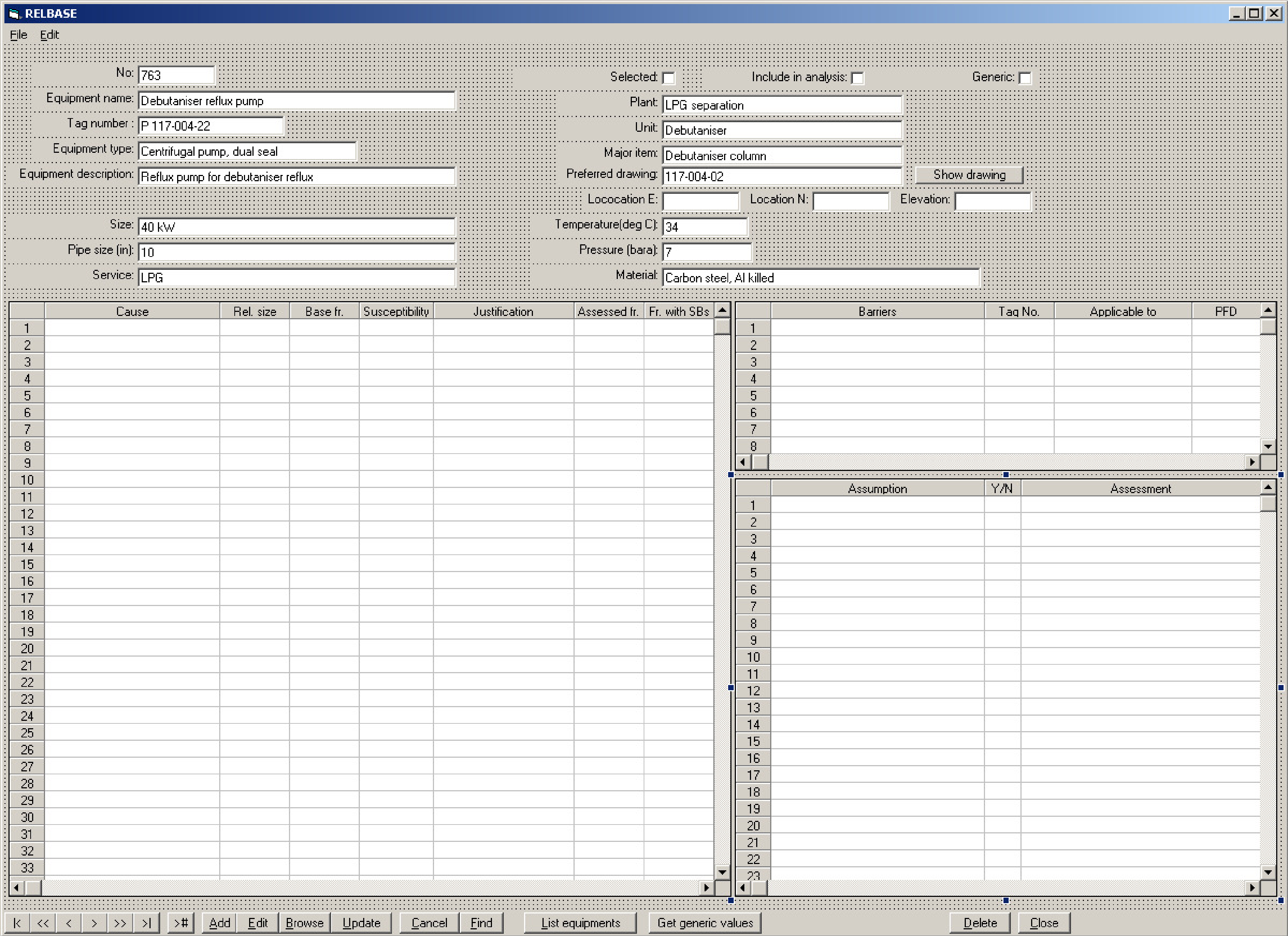


Figure 23.4. RELBASE database and assumption checker

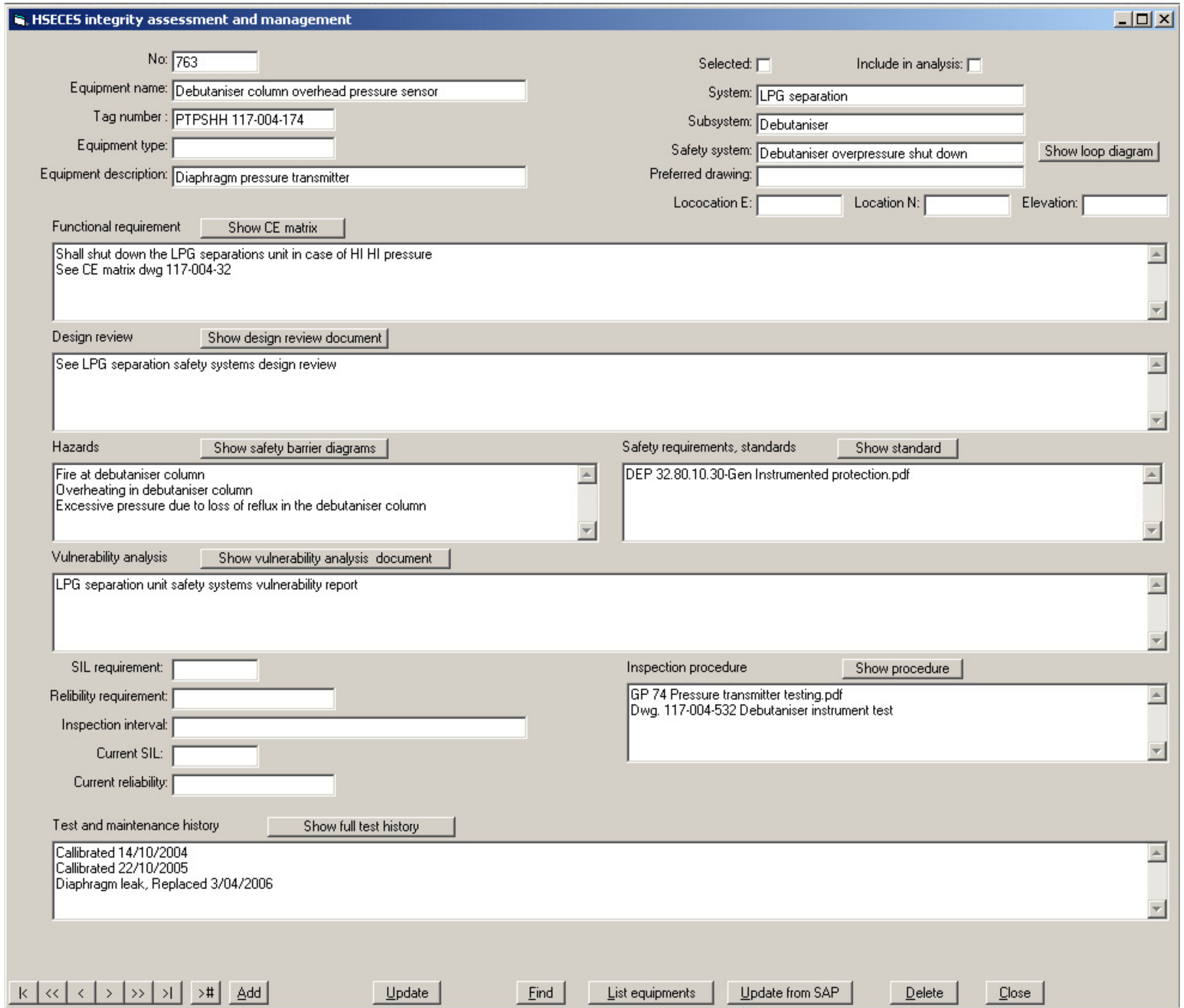


Figure 23.5. HSE CES safety critical equipment and systems management

23.8Conclusions

From these analyses, it appears possible to make predictions of major accident release frequencies which agree with average industry values to within a factor of about 2 to 3. Considering the uncertainties involved in deriving the underlying data, this is surprising. One hidden source of improved agreement may be the central limit theorem. Risk analyses involve the summation of very many numbers, and extreme uncertainty or inaccuracy will tend to be evened out, overestimates against underestimates. A full investigation of this effect remains part of our future program of work.

The final result of the present study should be not just to be able to calculate industry average accident rates, but also to be able to provide plant specific analyses, taking into account the age of plant equipment, the quality of design, and details of specific processes. The present data allows this to be done, but validation of such plant specific results is difficult to achieve. Nevertheless, it appears that at least some work can be done in this direction, given adequate detailed accident reports. This too remains part of the future investigations.

23.9References

1. API, Reported Fire Losses in the Petroleum Industry for 1998

2. Pape, R.P. and Nussey, C. A Basic Approach to the Analysis of Risk from Major Toxic Hazards, In The Assessment and Control of major hazards, I Chem E 1985

3. Cox, A.W., Lees, F.P. and Ang, M.L. Classification of Hazardous Locations, I Chem E 1990

4. Marsh & McClennan Protection Consultants (M&M PC), Large Property Damage Losses in the Hydrocarbon Chemical Industries. A Thirty Year Review, 14th Edition, M&M PC, 1992.

5. Dennis P. Nolan, Handbook Of Fire And Explosion Protection Engineering Principles For Oil, Gas, Chemical, And Related Facilities, Noyes Publications, 1996

6. Fryman, C.E., A Screening Methodology for Assessing Adequacy of Blast Protection for Control Rooms and Other Onsite Buildings, Report No. 1993, 221031, BP International Ltd, Sunbury, 1993, cited in Guidelines for Evaluating Process Plant Buildings for External Explosions and Fires, CCPS, 1996

7. Davenport, J.A. A study of vapour cloud incidents, Loss Prevention and Safety Promotion, Vol 4, I Chem E

8. Taylor, J.R., QRA Pro User’s manual, Version 7, Taylor Associates 2006

9. Taylor J.R. The QRAQ Project - Quality of Risk Assessment for Process Plant, Volume 6, The Effect of Methodology on Risk Analysis, Version 1 Issue 1, September 2006

10. Purple book

11. Atkins HSE ignition

12. IFAL

13. Taylor J.R. The QRAQ Project - Quality of Risk Assessment for Process Plant, Volume 6, A Reference QRA, National Refining S.A. , Version 1 Issue 1, November 2006

Notes