

Part E

RISK ASSESSMENT

Quantification

There has been some emphasis recently on "**number crunching**" in the preparation of **Risk Assessment**. This is potentially self defeating. It is more important to understand the laws of science and how these affect the safety process than to crunch numbers. The skill is in getting the correct solution to the problem - this can only be achieved by understanding the "**causes**" of accidents, (**accident causation**), why they occur in the first place, and the contributory factors which lead to escalation or to mitigation. See the "Bow Tie" Figure E 1.1 later.

It is recognised that "numbers" are necessary but on a personal basis I am rarely surprised by the answer which usually provides support for what was **known** to be correct from experience.

These notes on quantification are therefore written not from the stand point of a "here is an equation" but include the "**causes**", "**prevention**", "**mitigations**" plus **calculation**. It is far better to "know" that outflow rate through a hole the size of a 1p coin (decimal) at 20 bars is in excess of one kg per second, that the fire will be large, and steel work will be affected, thereafter the calculation is a nicety! The next feature is that no hole is sharp sided, round and to standard dimensions of 10 mm. The leak size is far from certain, the frequency of the occurrence is open to discussion and these swamp **any** errors with calculation of, say, the flame size. "**Source terms**" or the way the leak source is specified will dominate the answer.

Equally importantly it should be recognised that the models are all very much empirical models and do not stand up to dimensional analysis. They have to be taken on face value and recognised that they have been derived from rigorous physical modelling in research experiments.

E 1 Risk Assessment – An Overview

Introduction

When all of the Design is finished and the Management Systems are in place it is the requirement that the "**risks**" are "**ALARP**". In some cases it may be possible to demonstrate that the design is to "**best practice**" but this may not be the case for a more complex Process Plant. In this case the "**risk**" has to be assessed and "**ALARP**" demonstrated.

It would be wrong to think of Risk Assessment as being accurate or a science - it is approximate, and at best it is an art. As will be shown later the data used has to be treated with care and the calculations are based on empirical formulae, which have many subjective factors. The only certain thing of risk assessment is that final value will be between the extremes of the most **optimistic and pessimistic assumptions!!** Fortunately the theory of uncertainty indicates that the errors tend to be self cancelling. However after each assessment it must be challenged by the question "**Does this reflect reality?**" Some of the simplest (and elegant) risk assessments have been carried out in one side of paper without the use of a

computer! The classic is the assessment of the power of the first nuclear explosion in the Nevada Desert by Oppenheimer. It was based on an experienced judgement and a form of intuitive analysis, it was quick and **it was accurate!**

Everyone carries out some risk assessment every day and while there may be no absolute values of “tolerability” there are some reasonably well-defined bounds which will be used by many people.

Risk Assessment

There are three steps to Risk Assessment, known as:-

- How Big? (is the problem) This requires an assessment of the physical result of the event tempered by the effects of that event.
- How Often? (will it occur)
- So What? (shall I do about it)

This is very blunt but it is a simple guide or aid to the memory. As a means to illustrating this it is worth looking at a risk assessment that everyone carries out every day - crossing the road. No one sits with a calculator in their hand but the mental assessment process will be something like the following:

Car Speed 2 mph - judgement.

How Big?	Impact followed by a bruise or at worst a cut if hit by a car - (judgement).
How often?	2 mph = 0.9 metres per sec. Time to traverse the width of the car (1.7 metres) at a walking speed of 3 mph (1.3m/sec) = Transit time Transit time = 1.34 seconds (accurate value).

If the car is 1.1 metres or more away it will be possible to pass in front without being hit.

So what?	Evidence of being hit tends to zero if 2 metres away - in any case you can walk faster than the car so you could walk away from it and if necessary it should be possible to execute a “rugby hand off!!!
How big?	Bruise, this is a judgement based on the analysis of previous events.
How often?	Very unlikely, once in 10,000 crossings (say).

If the values are now changed such that the speed of the car is now say 40 mph the uncertainty in the speed assessment, the uncertainty in the judgement of distance and the uncertainty in the likely outcome (fatality) are such that the judgement of the risk will tend to err on the safe side. Uncertainty is one of the significant features of risk assessment

A broad definition of **Risk** and **Hazard** was given in the Introduction and other definitions used in Risk Assessment in Part A. These have been repeated here as this is another logical “home”. The following are taken from the IChemE publication *Nomenclature for Hazard and Risk Assessment in the Process Industries*.

Hazard a physical situation with a potential for human injury, damage to property, damage to the environment or some combination of these.

Individual risk The frequency at which an individual may be expected to sustain a given level of harm from the realisation of specified hazards.

Loss prevention A systematic approach to preventing accidents or minimising their effects. The activities may be associated with financial loss or safety issues. It is now being known as “Safety Engineering”!

Redundancy The performance of the same function by a number of identical but independent means.

Risk The likelihood of a specified undesired event occurring within a specified period or in specified circumstances. It may be either a frequency (the number of specified events occurring in unit time) or a probability, (the probability of a specified event following a prior event), depending on circumstances.

Risk assessment The quantitative evaluation of the likelihood of undesired events and the likelihood of harm or damage being caused, together with the value judgements made concerning the significance of the results.

Societal risk The relationship between frequency and the number of people suffering from a specified level of harm in a given population from the realisation of specified hazards.

Please ensure that the words risk and hazard are used correctly.

It is now appropriate to expand on the 3 elements of the assessment process.

How Big?

Models used in the process industry – an Overview

There are many tools and models available to assess the consequence of the event. The effects of heat, thermal radiation and toxics (such as carbon monoxide) are **fairly well** known and understood. Unfortunately the effects change with age, state of health and sensitivity so have to be adjusted from individual to individual.

The main models used in **RISK ASSESSMENT**, as applied to the process industry, are **Gas Dispersion**, **Fires** and **Explosions**. The impact of a toxic gas release involves the calculation of toxic concentrations through dispersion and then the analysis of the physiological effects of those concentrations on the human. In the case of fires it requires an analysis of the rate of build up of temperature on the challenge body (human or structural) and the analysis of the weakening – in the case of structures. In the case of explosions it requires an analysis of the structural response to an imposed loading due to pressure or impulse (pressure times time).

Why were these chosen? **Dispersion** is fundamental to the safe dilution of any gases be they toxic or flammable. Those affected may be on site or off site. They also feed back to the concept of “Hazardous Area Classification” See Part D. **Fires** are possibly the most destructive of the mechanisms but it is often

limited in area. It will destroy steels and injure humans. **Explosions** are probably the next most destructive mechanism but the damage tends to be total and business interruption is major. It can also affect persons off site. The scope of any notes such as these limits the use of sophisticated "**Effects Models**" - such as would be handled by Consultants. This should not be an excuse for not assessing the effects by manual calculations and so gaining a better understanding of the "phenomena" and the variables which might affect the outcomes.

The main types of **dispersion** are:-

- **Jets** - release at high exit velocity.
- **Puff** - the sudden release of a neutrally buoyant gas.
- **Passive** - the release at low exit velocity.
- **Heavy Gas Dispersion** - the dispersion of a sudden release of a heavy (denser than air) gas.

The first, **jet release**, describes the release from a vent or production equipment. The second, **puff release**, describes the release from a burst or ruptured container. The third, passive release, describes releases at low velocity which relies, for the most part, on the internal turbulence within the air. The history of the **plume**, be it a jet or a passive release, depends on:-

- Release rate (kg/sec);
- Release velocity;
- Angle of the release to the wind direction;
- Wind speed;
- Weather;
- Distance;
- Physical properties of the gas;

The concentration at any point beyond the release point will also include an assessment of: -

- The height of the release
- The relative elevation of the receiver point and the release point

In the case of a **puff release** the main parameters are:

- The mass released;
- Weather;
- Distance;
- Physical properties of the gas;

To a lesser extent other parameters which may be assessed for both releases are

- Roughness of the surrounding area, (just as surface roughness in a pipe).
- Relative Humidity of the Air

Only the **passive plume** and **puff releases** are addressed in this part.

The main types of **fire** are:-

- **Torch** (Jet) release of fluids at high velocity.
- **Pool fire**, where the spread is defined by bunds, drains or the rate of release and rate of combustion.
- **Boiling Liquid Expanding Vapour Explosion** (BLEVE) Fireball - the rupture of a vessel in a fire and the sudden release of massive quantities of fuel.
- **Flash Fire** the low pancake like fire lasting only a few seconds as the flame traverses through the cloud of flammable gases at about 3 to 5 m/s without any flame acceleration.
- **Running fire** the cascade of fuel down stairs or a structure. (These are significant following an aircraft fire).

Each describes very different types of fire. The history of a fire depends on the:-

- release rate (kg/sec);
- release velocity;
- wind speed;
- natural confinements of the fluids;
- distance;
- chemical nature of the fluids.

Only the **pool fire** and **BLEVE** are addressed in this part.

The main types of **explosion** are:-

- **Confined** - a pump room or analyser house, a compressor house, an office, warehouse building or a dwelling house.
- **Unconfined** - a vapour cloud explosion in an open plant or structure where flame velocities approach that of sonic.

The history of the explosion depends on the:-

- Release rate (kg/sec);

- Dispersion process;
- Confinement or explosion venting;
- Turbulence generation;
- Chemical nature of the fluids

Only the **vapour cloud explosions (VCE)** are addressed in this part.

Consequences Models

This requires an understanding of the effects on either the **human** or the **physical equipment**. This requires an analysis of physiological data, the analysis of past events or research into the effects. The **effects** are given later in this part.

How Often?

Frequency models used in the process industry

There are three main techniques for assessing the frequency of an event.

1. **Experience.**
2. **Event Outcome Trees**
3. **Fault trees.**

Use of Experience

Individual experience may show that on average pumps have to be overhauled once every three years but this may **not** be the **experience** of someone else. A more reliable source of “experience” is to be found in failure or reliability databases – of which there are many. The database **MUST** be **relevant** to the system under analysis. Data taken from equipment handling water is not relevant to equipment handling corrosive or erosive products! Data must be analysed **very** carefully. Data taken from many databases may cover an order or magnitude (factor of 10); this data may not represent the reality of the problem under study.

Data exists for the likely ignition probability for a specific leak size, human performance and other probabilities. These are based on “global experience” but may require to be adjusted for case specific studies.

Event Outcome trees - Fault Trees

The simplest way of showing the linkage of Fault and Event trees is by “The Bow Tie Diagram”. The LEFT HAND SIDE are the **CAUSES** of the Event where all of the barriers are collapsing and the RIGHT HAND SIDE are the **MITIGATIONS** or Protective Systems built into the design.

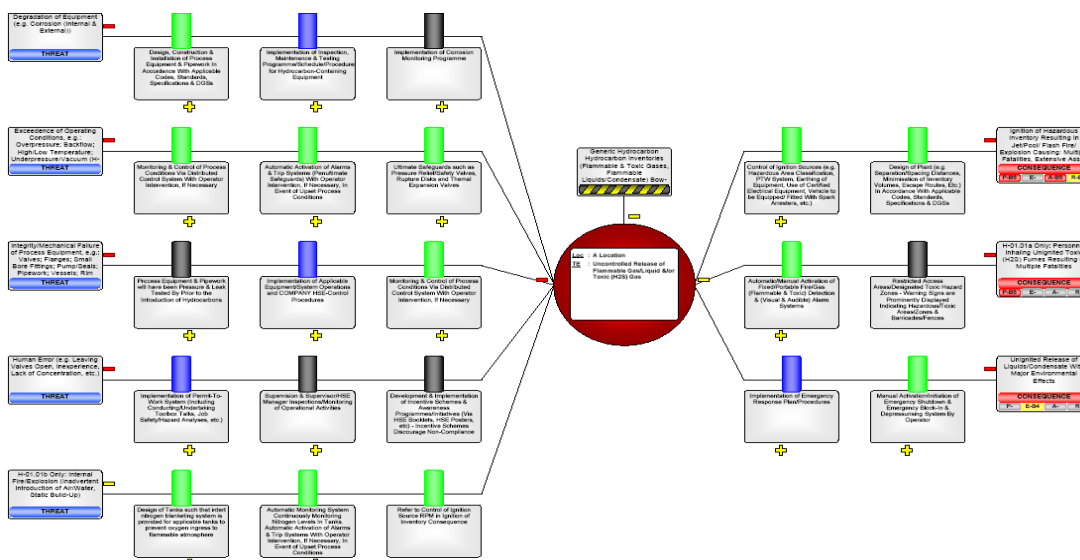


Figure E 1.1 The Bow Tie Diagram

Fault Trees

Fault Trees are the logical analyses of the condition required to create an event and produce results in the form of **probability** or **frequency**. The magnitude is assessed independently. Fault Trees normally start at the end point - such as an explosion - and start to define the exact combination of events that are required to create this event (top down). The structure is very precise and strict rules have to be applied - one rule that must be observed is that of **UNITS (probability and frequency)** have to be analysed carefully and be used consistently. Another is **AVOIDING DOUBLE COUNTS** (see also **Common Mode**). The data used must to be fully justified against references but occasionally "engineering judgement" has to be used and fully justified. **Beware** if the final result depends on that judgement.

The final result **must** be viewed against "credibility": Does the result look credible; does it fit reality/expectation?

Event Outcome Trees

Event Outcome Trees are a variation on fault trees and use a "yes/no" probability logic to define the event flow. The starting point is given – the left hand side of the bow tie - such as a leak every 100 years, The event outcome tree, the right hand part of the bow tie, moderates the frequency of the event to assess the probability of escalation (or control) and as with fault trees the magnitude requires assessment using models.

The values of the probabilities may be based on judgement or data. Once again the values have to be justified.

Once again the final results must be viewed against "credibility".

So What? (See also Part A)

The world can have anything it wants provided the world can afford it! Do you spend £50 million on signal improvements on rails **or** £50 million on road improvements or £50 million on kidney transplants? Someone has to decide! The decision may not be popular!!!!

Criteria are subjective and personal but, where one person may cross 10 metres in front of a car travelling at 20 mph, and another may only cross 15 metres in front, both may have made their judgement against the same objectives/end points - or maybe the first would have missed the last bus home!

Deviation of Criteria

Criteria are not single but are multiple and are not necessarily in sympathy with each other. For example the improvement in safety in one area may have an adverse effect on the environment. The reverse is true, Halons were excellent for fire extinguishers but they had adverse effects on the environment. The environment won and safety lost!

The main criteria are: -

Life/Limb

Health

Environment

(Negative) Public Reaction

Capital Cost

Consequential Losses

Life/Limb

There are many papers on risk to employees and the public. In the historic evolution of risk criteria the first marker used was Fatal Accident Rate (FAR). This was defined as the number of fatalities per 10^8 worked hours, this has now become simplified to the risk to the person which has units of frequency of fatality per person per year. Values were suggested the Introduction and reiterated below under "What values may be used?"

However it must be stated that risk values may not be appropriate elsewhere in the world. Court cases following fatal accidents have also given sightings on what value should be assigned to life. The Piper Alpha accident in July 1988 has shown that in the UK a value of at least £1 Million per life is accepted in courts so it behoves industry to value it higher or else punitive action may be brought. There is no doubt that local or national legislation must be borne in mind as was found out by the Ford Motor Corporation following accidents on the "PINTO" car and The Deepwater Horizon "blow out" where the punitive fines were significantly more than £ 1 million per life.

The Advisory Committee on Major Hazards First Report gave a very guarded comment which suggested that a major accident which occurred once per 10,000 years was **just about** acceptable. This does not suggest that this is acceptable on a global basis, some industries have an inherently higher risk than others, nor does it infer that this value is acceptable - it is just about acceptable so **must** be bettered - nor does it suggest that there can be 100 events each with a period of return of 10,000 years.

The range of "tolerable" "total risk" values ranges from 10^{-3} per person per year for the more hazardous industries such as Nuclear and Offshore Oil and Gas Production to 10^{-4} per person per year for the Chemical and allied industries. This must include the traditional "slips trips and falls". Each industry must set its own criteria. There is a form of logic that suggests that there is a series of decreasing injury with

reducing frequency. A cut hand might be tolerable once per year but a broken arm only once per 10 years. A serious injury, such as amputation, might be tolerable once per 1,000 years and a lesser injury once per 100 years. Again look at the Risk Matrix figure D 12.2.

It must not be forgotten that the public have a criterion which is 10^{-6} per person per year and is evidenced by HSE "Guidance on land use planning". This is two orders of magnitude less than that of the employee. It is an observation on many risk assessments that if the employee criteria satisfied it is likely that the criteria for the public will be also be met this is not an absolute rule but generality.

In the UK it is generally accepted that the risk of the chemical industry is made up half from "slips, trips and falls" (the technical accidents) and half from process or design-influenced accidents. The slips, trips and falls are dealt by design of access and standard of stairs and housekeeping, the process or design-influenced risks are relevant to this part.

Health

Health can be viewed as an extension of Life/Limb. It is now recognised that not only are some chemicals carcinogenic and lead to death but also some produce loss of quality of life.

Values can be applied to the "negative value" of harmful materials which might affect the health.

Values for NO_x and Particulates and other potential carcinogens have been derived.

Environment

There is an international awareness that pollution of the environment is no longer to be tolerated. The Sandoz pollution of the Rhine, Braer pollution in the Shetland, Scotland, Sea Empress in Milford Haven and Exxon Valdez pollution in Alaska have shown that clean-up can be prohibitively expensive and that major pollution is no longer internationally acceptable. (Though it was accepted in principle that the "Industrial Revolution" had to have pollution - "Where there's muck there's brass" - there is clear evidence, as witnessed in East Germany, that it is not now acceptable.) This is a study outwith the scope of these notes but it should be noted that Safety and Environment do not necessarily pull in the same direction and a balance has to be reached!

Values for pollutants are being produced by the day. One is the cost of oil spilled on the high seas; this has a notional value of between £5,000 and £10,000 per tonne if released close to land. It is less easy to ascribe values to some others such as phosphates and nitrogen run-off.

Public Reaction

There are pressure groups within society which are influencing industry so there is a twin pronged attack. The first is at the nuisance level such as smells and visual disturbance which results in adverse press and also letters to the Member of Parliament but the second one is the public aversion to major accidents. This is evidenced by the fact that one accident killing 10 people in one day produces a major press headline but 10 accidents each killing one person each day at different locations get only local press reports. The result is one of "frequency vs. number" criteria which are to be found in Holland and to a lesser extent in the UK Land Use Planning Criteria.

See also the risk Matrix figure D 12.2

Capital Cost and Consequential Loss

This may be viewed as an insurance policy. What do you insure? What do you accept as "self insurance"? As a generalisation it is the consequential loss or loss of sales which is the most punitive.

Are Criteria Absolute?

Criteria can not be absolute values with a clearly defined cut point - it is not realistic. There is clear recognition of a "target" to be aimed at - within that target are the bands of "**the acceptable**" and the "**intolerable or unacceptable**". The latter defines the upper end of the target and the former the lower end of the target. If the **risk** is in the acceptable regime there should be no further effort expended but if the **risk** falls between the two bands there **MUST** be further effort to reduce the risk. In reality the **ALARP** zone is the grey area where money should be spent to reduce the risk **so far as is reasonably practicable**. The **ALARP** dagger has been reproduced as a reminder of this concept.

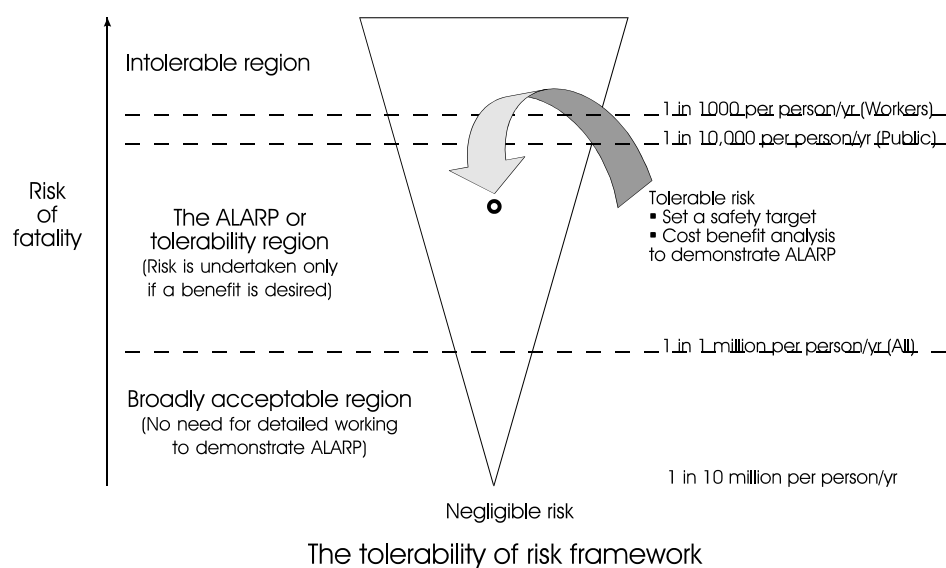


Figure E 1.2 The ALARP Dagger

Disclaimer

It is obvious that no external person should give or set another company's criteria. The values quoted in the text are those quoted elsewhere and used by other companies. Each company must choose its own criteria.

The use of Instrumented Systems to Reduce Risks – the Theory

In many simple risk assessments there is a requirement for a shutdown system, (trip or protective system). This was introduced in the Part D "Design for Safety"

The assessment process is as follows:-

"Is the cost of the protective system likely to be more than the saving"?

Obviously if the answer is "YES" the protective system produces negative cash flow.

The saving of the protective system is easy to assess:-

Cost of losses without protective system minus cost of losses with the protective system

The protective system is not perfect it can fail, and, if there is a human link that link could also fail. The failure is related to the "age" of the equipment. There are three phases, "**wear in**" or "**burn in**" where the failure rate falls with time as the equipment is young and is bedding in, "**beneficial life or use**" where the failure rate is low and constant and is not age related and "**wear out**" or "**burn out**" where the equipment is long overdue for maintenance and the failure rate rises with time as components start to fail.

For most equipment **burn in** takes only a few days, possibly up to a month, **beneficial life** then lasts 4 or 5 years and **aging** sets in at about 5 years.

The probability of a protective system being in operation at any time **T** years assuming random failure - i.e. no "**burn in** or **wear in**" or "**burn out** or **wear out**" is:-

$$e^{-FT} \quad (E 1.1)$$

Where:

F = the **sum** of the failure rates of **ALL** of the elements (per year). This is usually obtained from **Failure Databases**. However many databases give the value of **F** as the **total failure rate**. In reality some of the failures are "**fail safe**" or "**spurious**", that means that the shutdown system fails in a safe manner and shuts the process down. This is often given a failure rate designated as "**S**". The "**fail danger**" is the other failure mode which is the one of interest where the failure results in the non-operation of the system on demand. This is designated "**F**".

T = the test interval - value in years (every 6 months = 0.5 years) -

Note **T** will usually be less than 1

Therefore the probability of the trip being in a failed state or non functional after **T** years is: -

$$1 - e^{-FT}$$

The expansion of the exponential equation - $1 - e^{-FT}$ is:

$$1 - \{1 - (FT)^1 + 2!(FT)^2 - 3!(FT)^3 + 4!(FT)^4 \text{ etc}\}$$

(The divisor is "factorial n" or n!.)

So the final answer is:

$$FT - 1/2!(FT)^2 + 1/6(FT)^3 - 1/24(FT)^4 \text{ etc}$$

Clearly provided **FT** is small the second and subsequent parts of the equation can be ignored.

This reduces to:

$$FT \quad (E 1.2)$$

This is the value after **T** years but it is the average value that is of interest. The probability of failure at time **T** = 0 is obviously **zero** so the mean value between 0 and **T** is the *average* or a half of the bigger value so the probability that the system fails to shut the process down is:

$$\frac{1}{2} FT$$

(E 1. 3)

This only applies when FT is less than about 0.1 as the expansion of e is:

$$1 + 1 + 1/2 + 1/6 + 1/24 + 1/120 + \text{etc}$$

The derivation of this equation has a number of assumptions open to intellectual debate, the equipment is not experiencing wear in/wear out and the equipment is always returned to service "as new". Humans "wear in" over the first 6 months and wear out after 40 years of use!

$\frac{1}{2}$ FT is called the **Fractional Dead Time** or FDT or **Probability of Failure on Demand** PFD. Note FDT and PFD have **NO UNITS** and are a **PROBABILITY**. As the saving is not "perfect" or 100% but only (1- FDT) of the possible maximum the saving will be accrued. So, if losses were £100 per annum without a trip the losses with a trip would be £0.05 x 100 giving a saving of £95 per annum.

In reality testing is not perfect; humans make mistakes during testing and the trip has to be bypassed or taken out of circuit for on-line testing (sometimes it can be tested off line but not always on a continuous plant).

Therefore FDT (PFD) =

$$\begin{aligned} & \frac{1}{2} FT + \text{human error} + \text{Trip Test Dead Time} \\ & = \frac{1}{2} FT + 0.005 + \frac{0.5}{(T \times 8760)} \end{aligned} \quad (\text{E 1. 4})$$

The human failure rate is about 0.005 or 1 in 200 and the trip test dead time is simply the time for the test (in hours here taken as 0.5) times the tested per year (1/T) divided by hours per year.

A shut down system can now be designed and the performance specified. For values of T less than about 0.02 (weekly testing) the last term, called the **trip test dead time**, dominates and the FDT starts to rise for smaller values of T. For values of T over 0.5 (half yearly testing) the $\frac{1}{2}$ FT factor dominates and rises with larger values if T. The FDT derived from equation E 1.4 tends to a flat value for FDT between 0.03 to 0.05 for values of T between 0.1 and 0.25 so test intervals of about two or three months are realistic and economic. A good starting value for the FDT of a simple shut down system in a risk assessment is 0.05.

Unfortunately there is a cost to set against a saving. Each company will have to spend cash to buy the protective system. This in time will incur interest charges and operational costs such as repairs and testing. It is not unusual for this to reach 20% of the capital cost (half being interest charges + half operational costs). It also has to pay off the capital so it is not difficult to see that for a short lived modification the return must be nearer 50% of the capital cost (pay off = 3 years).

Percent Rate of Return / Year	Years to pay off
(Savings per year / Cost of Trip) x 100	
50%	3.0
40%	4.5
30%	7.0

Table E 1 1 Likely Pay-off Times for “Add-on” Safety at 20% annual costs

The usual value used for return = 20% for the first assessment.

The equation now becomes:

$$(\text{Capital Cost} \times 0.20) < \text{Annual Savings} \times 0.95 \quad (\text{E 1. 5})$$

Pitfalls in Assessing Fraction Dead Time

There are always pitfalls in the calculation of risks using "dead time" inherent in the simplifying assumptions. When the probability of a system being failed was assessed it was assumed that FT was less than 0.1 so the next term in the expanded exponential was trivial. FDT is the same as Probability of Failure on Demand - PFD but is used from here on as PFD is sometimes used for **Process Flow Diagram!**)

Simplifying Equation

The simple equation was:

$$\text{HAZARD FREQUENCY} = \text{FREQUENCY OF EVENTS} \times \text{FDT}$$

Where the frequency of the event is the sum of all the plant failures to a danger or hazardous state or another way:-

$$\text{HAZARD RATE} = \text{DEMAND RATE (D)} \times \text{FDT}$$

In reality, if DT or FT is nearer 1 the equation has the following awesome form. This is given for interest only: -

$$\text{HAZARD RATE: } \frac{FD}{F + D} \left(1 - \frac{1}{(F + D)T} \left(1 - e^{-(F+D)T} \right) \right) \quad (\text{E 1. 6})$$

Where:

F = failure rate of components (/ year)

D = demand rate for the process (/ year)

T = test interval (years)

Only use this equation if DT or FT is large

Consider now a car which has brakes which fail once per 10 years.

$$F = 0.1 \text{ per year}$$

Let us assume that the brakes are applied once every 3 minutes; that is

$$T = \frac{3}{8760 \times 60}$$

Clearly FT is definitely less than 0.1 so you would expect the crash hazard rate for a years driving of 1 hour per day to be:-

$$365 \times 1.0 \times 0.5 \times 0.1 \times \frac{3}{8760 \times 60}$$

$$= \text{Demand Rate} \times \text{FDT}$$

$$= 1.04 \times 10^{-3} / \text{year}$$

Well this is **not** true as the first time the brakes are applied after failure there will be a crash. The crash rate tends to the failure rate as shown by equation E 1. 6

Simplifying Assumptions

1. It is implicit that all equipment as tested is returned to "**as new**" – this is not necessarily so. Also some additional failures will still be due to "**burn in or wear in**" or "**burn out or wear out**".
2. FT and DT less than 0.1
3. All other effects such as trip test Dead Time and Human Reliability are added together with the final value.
4. Simplex system
5. No Common Mode Allowance

The **common mode** is that element of a trip system where the failure of the shutdown system is not time-dependant but is a function of design, the operating conditions or some other external effect which might make all or part of a larger system fail at the same time. Instruments are vulnerable to a potential common mode such as a fire or explosion, in this case it is sometimes called common cause, but also multiple shut down valves with a spring close action are likely to have common mode failures with the spring or the release mechanism. Consider also multiple pressure tappings - common modes which might make all of the tappings fail at the same time could be wax, dirt or ice.

As a result the limiting FDT is as follows:-

- 1) 1 of 1 = 0.05
- 2) 1 of 2 = 0.005 - 0.001
- 3) 2 of 3 = 0.001 to 0.0005

Design of Shutdown Systems

The design of shutdown systems, outlined in Part D, and the ability to test them correctly requires skills, which are out with the scope of these notes. Part F gives some markers. It must be noted that a shutdown system is designed with a reliability (Fractional Dead Time or Probability of Failure on Demand) appropriate to the perceived frequency and magnitude of the event (The Risk). In addition, it is essential that the complexity of the shutdown system does not inhibit safe and reliable operation. Shutdown systems sometimes have to be overridden to facilitate start up and they also have a low level shutdown – it will inhibit start up until a level is established – there has to be either an override or a means of establishing the level in a safe manner.

See also Part D Design for Safety where there is a discussion on the use of “**redundant systems**” (a two out of three system designated 2 – o – o – 3).

Hazards in Operation

How do you identify the Hazards Associated with Routine Maintenance and Operations?

Operations are a topic beyond that of a first degree course. However it is appropriate to note that many of the Management Systems described in Parts C & D apply to Operations.

The Incident Studies Part H show where problems were not handled properly and incidents occurred

The identification of hazards that has been applied will still apply to any changes (see Part F Management of Change) but every form of Maintenance will require a special form of Hazard Identification sometimes given the name “**Task Analysis**” where each step of the maintenance work from isolation through to refitting is analysed carefully, the hazards identified and the need for special features (including Personal Protective Equipment) is specified. This becomes part of a Management System called “**Permit to Work**” (PtW) (See Part F.

Physical Models or Phenomenology

E 2.1 Outflow

Before any physical models can be analysed it is necessary to know how much fluid will come out of a hole. The classic systems are Gas and Liquid but flashing fluids behave differently - as would be expected.

Gas Outflow

At high pressures over 200 kPa the classic gas outflow model is

$$M = C_d A P_u \sqrt{\frac{M\gamma}{RT_u} \left(\frac{2}{\gamma+1}\right)^{\frac{\gamma+1}{\gamma-1}}} \quad (\text{E 2.1.1})$$

$$V = \sqrt{\frac{2\gamma RT_u}{M(\gamma+1)}} \quad (\text{E 2.1.2})$$

M	=	Outflow (kg/s)
V	=	Exit Velocity (m/s)
C _d	=	Coefficient of Discharge
P _u	=	Upstream Pressure (Pascals)
A	=	Orifice Area (m ²)
M	=	Molecular Weights
T _u	=	Upstream Temperature (Degrees Kelvin)
R	=	8314 (Joules/Mole/°K)
γ	=	Ratio Specific Heat of the Gas at Constant Pressure and Volume (C _p /C _v)

At very high pressures greater than 1 mega Pa these equations still hold well but the orifice velocity and hence jet mixing velocity is underestimated. In most cases this is not significant.

At upstream pressures less than about 190 K pa, the equations E 2.1 and E 2.2 no longer hold as the gas velocity is subsonic:-

$$M_{LP} = C_d A P_u \sqrt{\frac{\gamma M}{RT_u} \left(\frac{\gamma}{\gamma - 1} \right) \left[\left(\frac{P_u}{P_o} \right)^{\frac{2}{\gamma}} \left(\frac{P_u}{P_o} \right)^{\frac{\gamma+1}{\gamma}} \right]} \quad (\text{E 2.3})$$

P_u = Upstream Pressure (Pascals)

P_o = Atmospheric Pressure (Pascals)

M_{LP} = Outflow (Kg/sec) (low pressure)

Others as above

Liquids (not flashing)

The outflow equation is the standard incompressible fluid flow equation: -

$$M = C_d A \sqrt{2 \rho (P_u - P_o)} \quad (\text{E 2.4})$$

C_d = Coefficient of discharge

A = Orifice Area (m^2)

ρ = Density (Kg/m^3)

P_u = Upstream Pressure (Pascals)

P_o = Downstream Pressure (Pascals)

Liquids (flashing) on the orifice including LPG, Chlorine and other volatile components

The outflow equation has to be modified to somewhere between a gas and an incompressible flow equation: -

$$M = C_d A \sqrt{2 \rho_c (P_u - P_c)} \quad (\text{E 2.5})$$

Where: -

ρ_c = Density of Fluids at $0.55 P_u$ (Kg/m^3)

$$P_c = 0.55 P_u$$

To find p_c it is necessary to use tables of physical properties to assess the fraction of fluid flashed at P_c and then to combine the phases to assess p_c . However, once out of the orifice the fluids will continue to flash to atmospheric pressure conditions. Conventionally when flashing cryogenics into the atmosphere, it has been taken that the mass in the cloud was twice the final flash to make allowances for the aerosol formation. In practise, total volatilisation is more likely when the fluids are more than about 50°C superheated (relative to ambient temperature) at source.

Experimental results suggest that if the effects of flashing are unclear the following are useful correlations:-

$$\text{Flashing flow} = 0.25 \text{ Liquid alone flow}$$

$$\text{Gas flow} = 0.25 \text{ Flashing flow}$$

or Flashing flow

$$(\text{gas flow through the orifice} \times \text{liquid flow through the orifice})^{1/2} \quad (\text{E 2.6})$$

This is a "ready reckoner" when equilibrium data is not available.

There are more reliable equations methods, one of which is called the **Homogenous Equilibrium Method** (HEM). The increased accuracy with the uncertainties in any assessment do not justified the use of HEM in these notes.

Coefficients of Discharge

The best case value for the coefficient of discharge is nearer 1 for a well rounded nozzle entry. It is also the "worst case" for outflow or risk assessment. The value for the coefficient of discharge for a sharp edged orifice can fall to 0.61; a middle ground value of 0.8 is often used for a short pipe stub. If in doubt use the geometric mean for the two values for the coefficient of discharge:

$$(1 \times 0.61)^{1/2} \quad (\text{E 2.7})$$

This tends to 0.8, a value used for outflow from a loss of containment. It is recognised that this is also the arithmetic mean but in the event of uncertainty it is better to use the geometric mean where the error is minimised.

A further complication for ruptured piping is that the flashing may take place inside the pipe upstream of the actual rupture. As a result there is the complication of a two phase flow pressure drop which again arrests the flow rate. Plots of the L/D for the upstream length of pipe and the modifying factor are available but are not part of these notes.

Flash Fraction

If all data is known the physical properties tables should be used, however, the simple formula below is a good approximation.

$$\text{Flash Fraction} = 1 - e^{-\left(\frac{c_p(T_u - T_o)}{c_v}\right)} \quad (\text{E 2.8})$$

Where: -

C_p = Specific Heat (Joules/kg)

C_v = Latent Heat (Joules/kg)

T_u = Upstream Temperatures (Degrees Kelvin)

T_o = Orifice or Downstream Temperature

There are good reasons for believing that based on tests on pressurised sources with upstream pressures >500 kPa plus >50°C superheat (e.g. LPG stored in bullets or even pressurised chlorine) flashed fluids will result in total evaporation due to forced evaporation from the fast moving droplets as they move through the air so resulting in negligible "**rain out**".

Evaporation from Pools

Pools evaporate at the surface and boil due to heat ingress from the soil/substrate. The evaporation is fairly simple but the heat ingress is very complex and involves knowledge of the physical properties of the substrates. This results in a decaying evaporation curve. It is worth being aware of the equations as it closes the gap between spills and evaporation. It is really part of a higher level discussion.

The "peak" evaporation rates E in kg/sec are given by the following equations.

Square Pools

$$E = 2.6 \times 10^{-4} \left[\frac{MW \times P}{T} \right] u^{0.78} x^{0.89} y \quad (\text{E 2.9})$$

Circular pools

$$E = 7.9 \times 10^{-4} \left[\frac{MW \times P}{T} \right] u^{0.78} R^{1.89} \quad (\text{E 2.10})$$

Where:

E = evaporation rates - kg/sec

MW = molecular weight

P = Vapour pressure of the liquid - Pa

T = Absolute temperature of the fluid - $^{\circ}\text{K}$

U = wind speed - m/s

R = radius of pool - m

x = downwind side of rectangle

y = cross wind side of rectangle

The significance of these equations is that of the vapour pressure of the fluid. If the pool can be blanketed with foam or another device which prevents contact between the air and the pool surface the evaporation rate can be reduced greatly as the forced evaporation tends to zero but boiling due to heat gain from the ground will still occur but at a reducing rate as the substrate (soil) chills.

E 3 Gas Dispersion

Gas **dispersion** is chosen as the first phenomenon as in many ways it feeds into some of the other models. Also if it disperses “safely” there should not be a problem.



Photo E 3.1 A Plume of a Continuous (heavy) Gas Release

Note:

1. The “looping” - the plume is not coherent – (See Figures E 3.1 & E 3.6)
2. The “momentum” rise (See Photo E 3.1 and Figure E 3.1)
3. The gravity fall for the gas (which in this case had a density of about 3 kg/m^3)

Introduction and General Background

Dispersion is fundamental to the safety of persons both on and off a site. It is necessary to understand what conditions may increase the risk following a release of toxic or flammable gases. It is also necessary to know how far a release of odoriferous gas may travel and still be smelled. The public have a right to a clean and odour free environment. Dispersion also feeds into “**hazardous area classification**” see Part D.

It is fairly obvious that gas dispersion is an essential feature of the **earth’s boundary layer** and we live in that boundary layer! If it were not so any releases from a site would stay at that unique concentration

until it had gone round the world! Luckily, as experienced in reality, the air is a fairly homogenous mixture and there are strong mixing processes within it. The mixing processes are four fold **Jet**, **Bulk**, **Turbulent** and **Diffusion** - see photo E 3.1 and figure E 3.1. In the case of **Jet Mixing** there is a high velocity jet with high internal turbulence. The action of the jet upon the air produces vortices at the sheared interface of the jet and the air stream, these plus the internal turbulence in the jet are powerful mixing mechanisms. The **Bulk Mixing** (sometimes called **Translation Mixing**) is caused by two gas streams travelling in different directions to each other - one is injected into the other in a shearing or smearing action. During this phase the plume is turned and moves with the air stream at a relative velocity tending to zero. The third and possibly the most important mechanism is **Turbulent Mixing** due to the local vortices within the air stream. The air always has movement within it - this is evident from the study of the movement of water droplets in clouds (fog) and the movement of smoke leaving a chimney. The final effect (and by far the smallest) is **Molecular Diffusion** or the molecular velocities which are random in direction. This results in a uniform concentration of gas - be it light or heavy - within an enclosed room. In theory and in practice hydrogen can appear under the floor boards and hydrogen sulphide behind the ceiling tiles. Be very aware of the potential hazards created by diffusion – it can be a real killer!

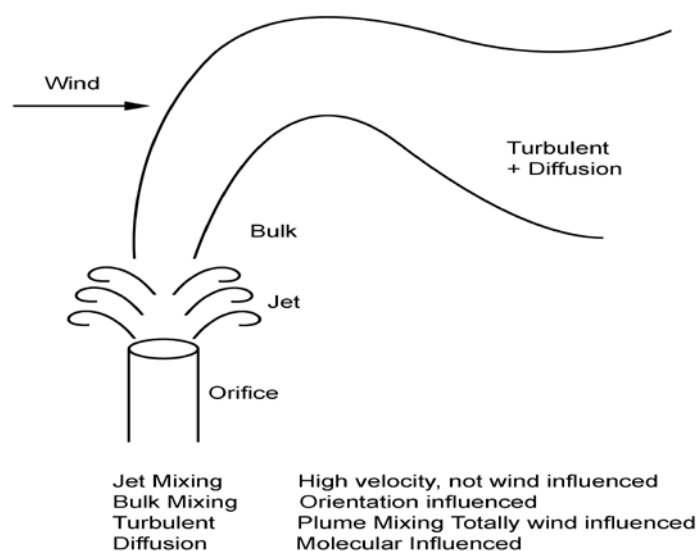


Figure E 3.1 The Plume History

Initially the following two mechanisms are given more detailed analysis:-

- Jet Mixing
- Turbulent Mixing.

Later the **Puff Releases Dispersion** will be discussed

It is reasonable to note that anything which slows up air flow and creates stagnant zones is a hindrance to dispersion (See the formulae later on). An open, uncongested process plant is safer than a congested plant or a confined space which has to be force ventilated. A plant where equipment is well spread out is safer than one where equipment is close together. A plant where there are walls or enclosures are to be resisted as are artificial enclosures such as created by pipe tracks, these all result in a plant which is less than an open. Ventilation rates from fans are often based on 6-12 changes per hour, this produces 'wind'

speeds of about 0.1- 0.2 m/sec, but even on a **very still** day air speeds of 0.5 m/sec are readily achieved in the open air.

The history of any gas plume is at best complex and can only be assessed with tolerable accuracy using sophisticated computer models, these are not available to most Universities – and even these have their own limitations. It is not difficult to see even by visualising this problem that a jet may initially start off up wind but if the wind is in an adverse direction the resultant plume may find its way into a safe area. This is illustrated by the plots below.

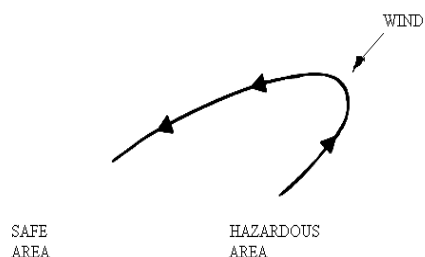


Fig E 3.2 Plume of gas blown by the wind

Heavy Gas Dispersion is even more perverse! LNG (say Liquid Methane) requires heat from the air to complete the evaporation process and at the point of heat balance where all of the un-flashed liquid is fully vaporised the final air temperature is -160°C or 113K. Even though Methane has a density of 0.71 kg/m^3 at atmospheric temperature compared to 1.22 kg/m^3 for air, and while there may be 25% v/v methane in the resultant cloud the true cloud density for methane/air at 113K will be over twice that of air until it warms up. Initially it will sink - not rise and it will flow as a thin cloud slumping under the effects of gravity on the cloud. The same is true for a spill of water onto the floor! There are methodologies for heavy gas dispersion but they are complex.

Beware the perverseness of gas dispersion!

Meteorology

The atmosphere in contact with the earth is in fact a boundary layer and subject to both temperature and velocity gradients. If a small cylinder of gas was raised from the surface of the earth it would expand adiabatically and cool at a temperature gradient of about $-1^{\circ}\text{C}/100$ metres, this is called the adiabatic lapse rate and applies when there are no rising or falling thermals. (In reality it is nearer 0.8°C but the value of -1°C is easier to remember.) If the earth is heated by the sun the thermal gradient may increase to $-2^{\circ}\text{C}/100$ - that is the air at the ground level is more buoyant than the air above so it will rise in "**thermals**". The opposite is true on a starry night, the temperature gradient is zero or even positive so there are descending thermals and the air is trapped at ground level with little dispersion. This leads to fogs. In the case of the adiabatic lapse rate there is no effective buoyancy gradient, so there are no thermals rising or falling.

The rising thermals induced by the sun are therefore a mixing process. In some cases there are inversions, that is, the density of the air above the inversion layer is lower than that below the layer, so, air can not penetrate it by buoyancy alone – and the gases are "trapped". This was shown classically in the photos of the smoke plumes during the Buncefield Tank Farm fires photo E.3.2. (Hydrogen will still rise through the

layer as it has inherent buoyancy but low concentrations of noxious gases such as Sulphur and Nitrogen Dioxides can become trapped with a significant impact on pollution.)



Photo E 3.2 Inversion Conditions during the Buncefield Terminal Fire

Note the flat top of the plume

Stability

The temperature gradient has been discussed earlier and is of importance and is defined by Pasquill "*Stability Levels*".

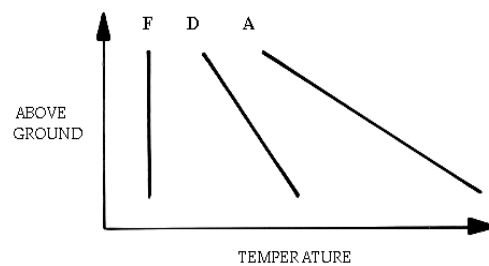


Fig E 3.3 Temperature profile in the boundary layer

Level A equate to a hot bright summer day, temperature gradient over -2°C per 100 m.

Pasquill F equates a cold star lit winter cold/frosty night where the temperature gradient is 0 or maybe 1°C per 100 m.

Between A and F there are 4 levels - the most common being D where the temperature gradient is -1°C per 100 m. This occurs in the UK for almost 80% of the time.

This ratio will be different in any other country round the World.

The main equations worthy of note are the gas dispersion equations for passive and puff releases. But it is of note that there are also equations which describe the mass of fuel between flammable limits. These are not given in this part but it is mentioned as it has some significance when you are looking at Vapour Cloud Explosions.

Within the earth atmosphere there is a velocity gradient as befits any boundary layer for fluid flow. It is not of any major concern and increases, with height and as a result all meteorological references for wind speed are referred to a standard of +10m. This gradient results in wind speeds at the top of mountains which are significantly higher than that at sea level and of course the “jet stream” at 10,000 m. In more complex dispersion calculations elevation has to be taken into account as it modifies the local wind speed.

Dispersion Theory

Continuous Release

The main equations of Turbulent Dispersion are the downwind concentration at ground level or on the centre line equation gives the generalised centre line concentration at x, 0, 0. Where x is the downwind distance, 0 is the cross wind distance (y) and 0 is the vertical distance (z).

$$\chi = \frac{1}{K_1} \frac{Q}{\pi \sigma_y \sigma_z u} \quad (\text{E 3.1})$$

$K_1 = 1$ for a ground level release

$K_1 = 2$ for an elevated release

$\sigma_y \sigma_z$ are dispersion coefficients – see later Table E 3 .1

This equation derives from the full equation

Ground Level

$$\chi_{xyz} = \frac{Q}{\pi \sigma_y \sigma_z u} \exp\left(-\frac{1}{2} \left(\frac{y^2}{\sigma_y^2} + \frac{z^2}{\sigma_z^2}\right)\right) \quad (\text{E 3.2})$$

Where: -

χ_{xyz} = The concentration kg/m³

Q = Release rate kg/s

σ_y and σ_z are the dispersion coefficients in the y (horizontal) and z (vertical axes)

y and z are the horizontal and vertical axes

u is the wind speed m/s

The part of the equation E.3.2 defined by; -

$$\chi_{xyz} = \frac{Q}{\pi \sigma_y \sigma_z u} \quad (\text{E 3.3})$$

represents the concentration along the centre line where it is at the highest. This occurs as at the centre line $y^2 = z^2 = 0$ and $\exp - 0 = 1$ and value of χ_{xyz} is a maximum for any fixed value of x.

The exponential part of the equation: -

$$\exp\left(-\frac{1}{2} \left(\frac{y^2}{\sigma_y^2} + \frac{z^2}{\sigma_z^2}\right)\right) \quad (\text{E 3.4})$$

represents the concentration decay across the plume in the y axis and vertically in the z axes. The equations might suggest that the plume goes on to infinity. It should be limited vertically and horizontally to about 3σ values.

- χ_{xyz} = concentration Kg/m³
- $\sigma_x \sigma_x$ = dispersion coefficients
- x,y,z = are ordinates from source x along, y across, z up/down

Elevated Release

$$\chi_{xyz} = \frac{Q}{2\pi \sigma_y \sigma_z u} \left(\exp\left(-\frac{y^2}{2\sigma_y^2}\right) \left(\exp\left(-\frac{(z-H)^2}{2\sigma_z^2}\right) + \exp\left(-\frac{(z+H)^2}{2\sigma_z^2}\right) \right) \right) \quad (\text{E 3.5})$$

H = release height above the ground

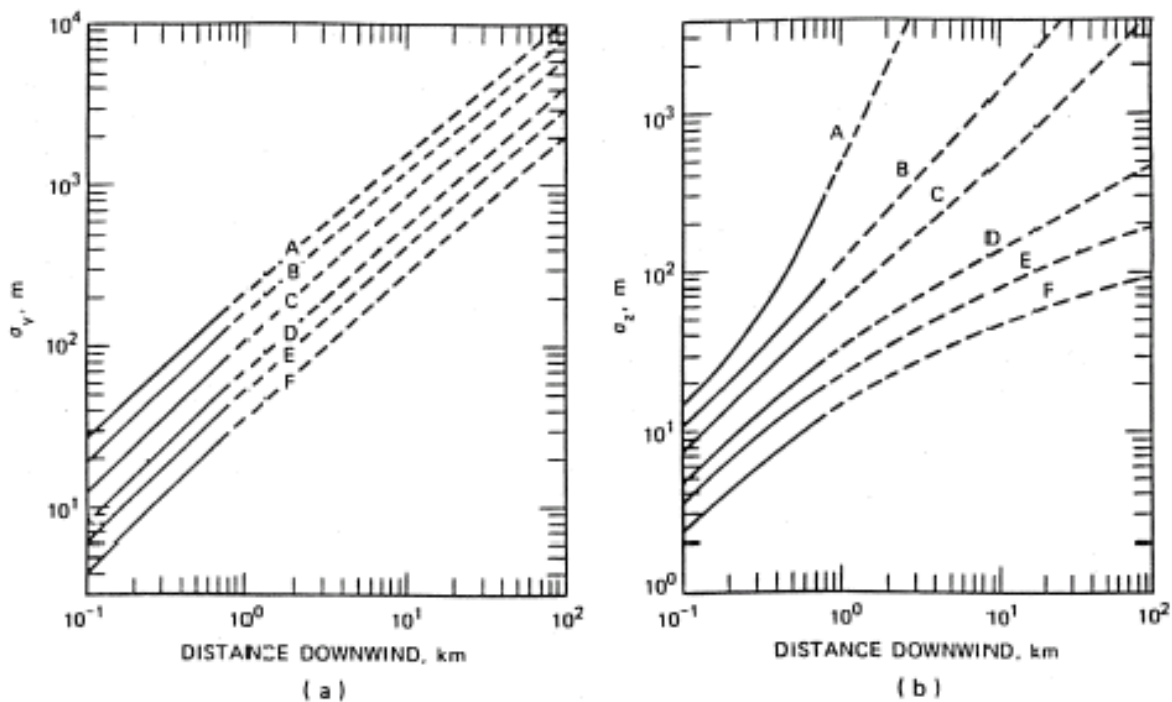
Where the centre line value is required $y^2 = 0$ and $\exp - 0 = 1$, $z = H$ and so $(z - H)^2 = 0$. Where H is more than 3 or 5 metres $\exp - (Z + H)^2$ tends to zero. Note: this is only given for completeness. Use equation E 3.3 with the K_1 moderator.

Pasquill Category	σ_y (m)	σ_z (m)
-------------------	----------------	----------------

A	$\sigma_y = 0.493 x^{0.88}$	$\sigma_z = 0.087 x^{1.10} (100 < x < 300)$
D	$\sigma_y = 0.128 x^{0.90}$	$\sigma_z = 0.093 x^{0.85} (100 < x < 500)$
F	$\sigma_y = 0.067 x^{0.90}$	$\sigma_z = 0.057 x^{0.80} (100 < x < 500)$

Table E 3. 1 - Dispersion coefficients for Passive Plumes

The Dispersion coefficients are only accurate in the range shown - minor errors will result out with these bands. Their origins are shown later in Figure E.3.4 & 5. It will be noted that there are actually 6 stabilities (and a rare one of G) but the three given are the most relevant in simple risk assessment. In more complex assessments it would be necessary to explore ALL stabilities, ALL wind speeds and ALL wind directions. This gives potentially up to 1000 combinations. Fortunately there are some mutual exclusions which can reduce the combinations to nearer 40. For note only the very UNSTABLE – A stability and VERY STABLE – F stability conditions do not occur with wind speeds above about 5 m/s. This helps the assessment process. For simple risk assessments in the UK, and the UK only, reasonably accurate results can be derived from the simplifying assumption that 80% of the year can be characterised by 5 m/s winds, stability D (5D) and 20% of the time it is 2 m/s and stability F (2F).



Figures E 3.4 & E 3.5 Derivation of Dispersion Coefficients σ_y and σ_z given in Equation E 3.4 (As Gifford)

It will be noted that the plots are on a log v log basis and have a limited linearity such that the correlations fall down where x is less than 100 m. There are more sophisticated correlations used in computers but the accuracy is still open to debate where x is small.

For elevated sources two mirror images can be considered - so any value is half what you expect. Consider two parcels of gas in image formation - each of $Q/2$ and inject this into the equations.

The shape of the cloud defined by equation E.3.2 is that of a half tear drop. The maximum dimensions in the y and z axes are about 2/3 along the length of the plume.

Hence for Pasquill Stability D at 2m/sec and 0.5 Kg/second leak rates at an elevated level, the distance to a safe dispersion as measured along the centre line (x) is: -

$$\text{Distance} = 1.75 \sqrt{\frac{0.5}{2\pi \times 2 \times 0.128 \times 0.093 \times \chi}}$$

χ , the concentration should have a modifying factor, described below, and applied to it.

Peak to TWA Concentrations

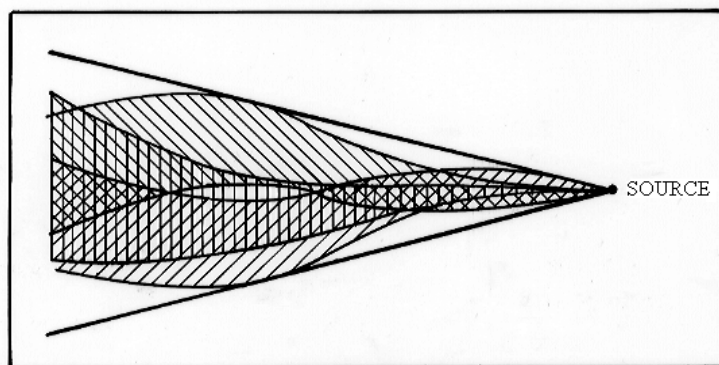


Fig E 3. 6 Shapes of a family of dispersing plumes (see also Photo E 3.1)

In this set you will see there are 3 "snapshots" of a single release taken at different times - say every minute. The equations given, E.3.2, E.3.3 describe the **TIME WEIGHTED AVERAGE (TWA)** taken over the extremes of the three plumes and shown as the two divergent lines forming a triangular envelope around the three plumes in Figure E.3.6. This averaging was typically taken over a 6 minute interval. (Think of the problems of sampling and analysing a time varying plume of gas!) **BUT**, at any one time the **peak** concentration may be 2.5 times the TWA, close in to the release it may be even 4 times that predicted. This means that it is necessary to:

1. Ensure the safety margin is included.
2. Use 40% of the target or desired limiting concentration if using a computer programme. (These programmes usually assess the "time weighted average" (TWA) usually over 3 or 6 minutes - and **not** the instantaneous values). (For example use 40% of the lower flammable limit table E 4.1 later).
3. Multiply any concentration values derived by calculation by a factor of 2.5 to reflect that the peak value could be significantly higher than assessed by the TWA equations.

Conversion vol/vol and weight/vol

Remember that the gas concentrations are often given as vol/vol and the dispersion equations give results in kg/m³. To convert between vol/vol and weight/vol (kg/m³) use the conversion:

$$\left(\frac{\text{Molecular Weight}}{22.41} \right) \quad (\text{E 3.6})$$

$$0.1 \text{ kg/m}^3 \text{ of gas MW} = 40 = \frac{0.1 \times 22.41}{40} = 0.056025 \text{ vol/vol}$$

Likewise a concentration of 10% vol/vol has a mass concentration of 0.1785 kg/m³.

Jet Dispersion

The assessment of a sonic jet concentration is given by the equation:-

$$\frac{2}{C_a} = 0.32 \frac{l}{dj} \left(\frac{\rho_a}{\rho_g} \right)^{0.5} \quad (\text{E 3.7})$$

where:-

- C_a = Axial concentration at distance l along the jet - kg/m³
- l = Distance along the jet - m
- dj = the jet diameter (or effective diameter of a sonic jet if released from over 200 kilo Pascals)
- ρ_a ρ_g Are the air and gas densities in kg/m³ respectively

Whence the jet usually disperses in about 200 equivalent jet diameters dj^* , where dj^* is the equivalent diameter of a jet of gas moving at Mach 1 at Atmospheric Pressure. This is given for completeness.

Limitations in the use of gas dispersion equations

The calculation of gas plumes and the concentration is not as accurate as would be wished. Care must be taken to assess the true variables accurately.

1. The range of results between a hot windy day and a cold still starry night can vary by a factor of greater than 10 - wind speed and atmospheric stability must be assessed as variables.
2. Elevation affects the results.
3. Buoyancy affects the results.
4. Efflux conditioners - velocity and orientation to the wind effects the results.

5. The variables within the computer programme and how these are modelled affect the results.

The answers are probably only accurate to ± 2 .

In order to carry out a simple dispersion calculation, assess the axial concentration using equation E.3.3 and then apply the moderation of equation E.3.4. Note that it is possible to assess the cross wind width (or vertical height) by rearranging the equations to make an equation with an exponential. This is solved by taking natural logs (ln) of both sides of the arrangement.

Instantaneous Releases

Instantaneous releases, sometimes called "Puff Releases" are typified by a "burst vessel". The release is neutrally, positively or negatively buoyant. For the most part the releases tend to be **negatively buoyant**, that is, heavier than air. The reasons for this are that in general the fluids are heavy or of a high molecular weight or are potentially cryogenic or are cold when the pressure is released and the gases expand adiabatically.

Heavy gases mix with air by two mechanisms, first there is the potential energy of the heavy collapsing cloud. The energy produces a "pancake cloud" with a rolling vortex at its edge - this is a powerful mixing mechanism. The analysis of this requires at best long, interactive calculations or computer models but the non buoyant model is easier to handle and probably produces higher concentrations at any point downwind of the release.

The most simple model for the concentration is the following:

$$\chi = \frac{K_1 Q^*}{x^{K_2}} \quad (\text{E 3.8})$$

where:

χ Concentration kg/m^3

Q^* is the mass released (Kg)

x downwind distance (metres)

K_1/K_2 are constants

Remember if Q^* is the resultant release of an air/gas mixture due to a puff release the values of χ need adjusting. **First** the mass of the new mixture Q^* which has both air and gas is needed and **Second** a new mass dilution, χ_{modified} is needed for that mixture. The overall concentration is then the blend of the two diluting mechanisms, the initial mixing and then the dispersion mechanism. Heavy gas dispersion is a totally different and more complex study which is beyond the scope of these notes. The collapsing heavy gas will result in a more vigorous mixing process than the "puff release", as outlined earlier.

Instantaneous Release

The dispersion coefficients are functions of the atmospheric stability and time after release, downwind distance, (as for continuous releases). The equation E.3.8 gives concentration around the cloud centre, which moves with the speed and direction of the prevailing wind.

A common equation for a ground-based instantaneous release comes from the Sutton model:

$$\chi(x, y, z, t) = \frac{2 \cdot Q^*}{(\pi)^{\frac{3}{2}} \sigma_x \sigma_y \sigma_z} \exp\left(-\left(\frac{x^2}{\sigma_x^2} + \frac{y^2}{\sigma_y^2} + \frac{z^2}{\sigma_z^2}\right)\right) \quad (\text{E 3.9})$$

where:

- x, y, z, t = Axes of the cloud downwind, crosswind and vertical (m) for a moving cloud from the centre of the cloud after time t seconds
- t = time after release
- Q^* = size of release (Kg)
- $\sigma_x, \sigma_y, \sigma_z$ = dispersion coefficients in downwind, crosswind and vertical directions
- χ = average point concentration (Kg/m³)

One simple correlation for the dispersion coefficients assumes that:

$$\sigma_x^2 = \sigma_y^2 = \sigma_z^2 = C^2 \cdot (u \cdot t)^{(2-n)} \quad (\text{E 3.10})$$

Pasquill Category	C	n
A	0.20	0.17
D	0.14	0.25
F	0.09	0.35

Table E 3.2 Values for C and n in equation E 3.9.

Please note that the equations use (u.t), this is speed times time which is of course distance. (u.t is the downwind distance travelled by the puff release and is effectively x in the continuous release equations.)

Whence: -

$$\chi = \frac{45 Q^*}{(ut)^{2.745}} \text{ Unstable - (Pasquill A)} \quad (\text{E 3.11})$$

$$\chi = \frac{131 Q^*}{(ut)^{2.625}} \quad \text{Neutral (Pasquill D)} \quad (\text{E 3.12})$$

$$\chi = \frac{493 Q^*}{(ut)^{2.475}} \quad \text{Very Stable (Pasquill F)} \quad (\text{E 3.13})$$

Equation E 3.9 describes a hemisphere, travelling down wind at the speed of the wind but expanding slowly as it moves. The concentration at any point in the cloud, with equidistance measured radially from the centre, will be the same. (It can be looked at as a half onion with the onion rings being the location of equal concentration [isopleths].) It is a simple equation to use but in reality the shape is more of a split rugby ball with the major axis along the wind axis. As with the continuous release the non exponential part of the equation E.3.9 defines the concentration at the centre of the hemisphere as it contacts the ground and the exponential part defines the concentration decay at any other part of the "hemisphere".

E 4 Fires



Photo E 4.1 A Tank Fire

Note:

1. **The downwind lean in a wind of about 7 m/s wind**
2. **The smoky nature of the Fire – smoke = C + CO**
3. **Unburned paint cooled by the stored fluids – possibly petrol**
4. **Minor wind induced flame “drag” down the tank beyond the tank edge**

General Introduction and Background

Fires are one of the major process causes, if not the major cause, of loss of production and life on a process plant. Yet the assessment of fires and their effects is still a somewhat inexact science.

Fires have a potential for major damage. Steel loses **most** of its structural strength at about 450°C and softens. Equipment handling materials under pressure may rupture and spill their contents into the flame, structures may collapse and joints in piping systems may spring which then spill their contents into the flame. This is the “domino” effect.

The combustion of fluids **may** burn to relatively non toxic gases however carbon monoxide is not safe nor is carbon dioxide or carbon particulates. Compounds which contain nitrogen or halogens within the molecule and the mix of compounds in a warehouse **may** burn together to produce toxic materials such as HCN, COCl₂ and HCl.

Fuel/Air/Ignition

The usual graphical representation of the fire process is in the fire triangle - remember fires may also produce an explosion as a precursor to a fire.

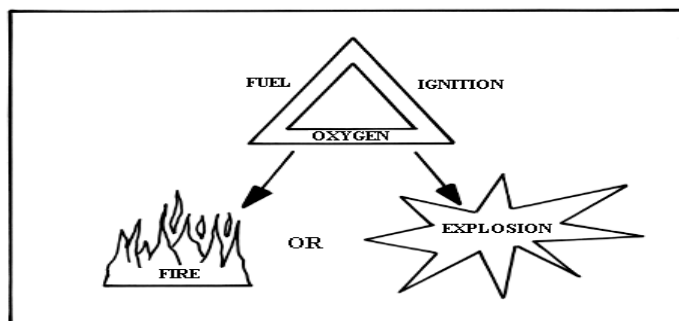


Fig E 4.1 Fire "Triangle"

This triangle is a useful beginning for the discussion.

Fuel	Fuels come in all sizes and phases: solid, liquid and gas.
Solids:	Could be can dust, wood, coal, metal powders - these can also explode.
Liquids:	Could be Petrol, Diesel Oil or others.
Gas:	Could be Hydrogen, Methane, Propane and others.

It is **not** the solid or liquid which burns it is the gas, so, a wooden log requires some form of "lighter", just as does Diesel Oil. (It should be possible to snuff out a match using Diesel oil – but please do not try it!) See figure E 4.2 later.

Oxygen:	Just as with human life, combustion requires oxygen and flames can not exist at low concentration of oxygen typically less than 10% v/v or a partial pressure of 10kPa. See later
Ignition:	This ignition energy is usually low; a spark is all that is usually required.

Given the 3 sides they all have to be together and in the correct balance - take one away - nothing happens.

It may seem odd that metals or dust can explode. A finely divided solid will have a large surface area; these are typified by as corn dust or aluminium powder. If the material is reactive and there is sufficient energy to set off the reaction. Once initiated the exothermic process will take over.

Flammability Envelope

The fuels are **not** flammable across the whole concentration range of 0 - 100% fuel in air. Thankfully, there is only a limited concentration bound in which flames may be supported. This is shown graphically in the flammability diagram or envelope – figure E 4.2, below. The diagonal line represents concentration locus of a mixture of air and a Nitrogen/Methane (80% Nitrogen & 20% v/v methane). At 3 volumes of air to 1 volume of Nitrogen/Methane the mixture is just flammable. At blends of less than 3:1 with air the blend is non flammable (too rich) and with blends more than 4:1 the blends are also non flammable (too lean). Please note that purging out this mixture with air will cut inside the flammable envelope and there may be a point in the purging cycle where there could be a disastrous ignition and explosion.

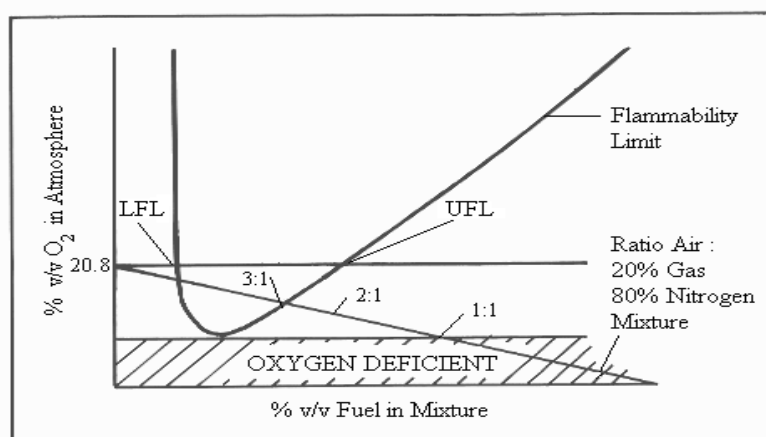
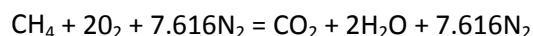


Fig E 4.2 Flammability Envelope

In air (20.8% v/v oxygen) the flammable range is usually only 2 times and 0.5 times the stoichiometric concentration. The two extremes where the flame can **just exist** are the **Upper Flammable Limits** and **Lower Flammable Limits (LFL and UFL)** (sometimes "explosive" is used interchangeably with "flammable"). "Just Exist" means that the flame will just propagate vertically and **only vertically**. The "x 2 and x ½ rule" is not absolute but it is worth remembering if the data is not otherwise available.

The envelope widens as the oxygen concentration increases. In addition the ignition energy falls (table E.4.3). Ultimately at high oxygen concentrations ignition **may** occur spontaneously.

Stoichiometry



$$1 \text{ vol} + 2 \text{ vol} + 7.616 \text{ vols}$$

For methane the stoichiometric concentration is: -

$$\frac{1}{10.616} \text{ vol/vol ratio or } 9.42\% \text{ v/v.}$$

Fuel	LFL% v/v	UFL% v/v	Stoichiometric % v/v
Hydrogen*	4.0	75.0	29.4
Methane	5.0	5.0	9.4
Ethane	3.0	12.4	7.7
Propane	2.1	9.5	4.0
Butane	1.8	8.4	3.1
Ethylene*	2.7	3.4	6.5
Acetylene*	2.4	100.0	12.2

Table E 4.1 Flammability Limits for Selected Hydrocarbons

It should be noted that Hydrogen and unsaturated fuels marked * tend to break the general pattern of 2 times and 0.5 times stoichiometric.

If the gas is in a mixture this flammability limit can be achieved by Le Chatelier's rule: -

$$\frac{1}{LEL} = \sum \frac{N_i}{LEL_i} \quad (\text{E 4.1})$$

Where: -

$$N_i = \text{vol \% of gas } i$$

$$LEL_i = \text{LEL v/v of gas } i$$

The lower apex of the flammability envelope (shown in figure E.4.2) is usually about 10% v/v Oxygen, 90% v/v Nitrogen. This means that a fire can be prevented by keeping the oxygen concentration to a flow level - that is "inerting". An "economic" value of 5% v/v oxygen is often used. Coincidentally "humans" lose consciousness when the oxygen concentration falls to 10% v/v or a partial pressure of 10 kPa.

The bounds of the flammability envelope expand with increased temperature and pressure - that is the minimum oxygen level and the LFL goes down and the UFL goes up in value.

As the oxygen level increases (see figure E.4.2) the energy falls. High pressure, high temperature and oxygen concentrations all enhance the ignition process and require lower ignition levels as would be expected from simple chemistry and the laws of mass action. Ultimately it is possible for liquids and solids to ignite spontaneously with ultra high oxygen concentrations in the supporting atmosphere. This has resulted in disastrous fires, none less than that in one of the NASA Apollo modules (Apollo 1) on ground test in the 1960s.

Diesel oil had a very low vapour pressure - that is the fuel/air mixture at room temperature is "too lean" or below the LFL as the flash point of diesel is well above 20°C. Once the oil is warmed locally above its "**flash point**" it will ignite. (The flash point is that temperature where the vapour pressure is JUST sufficient to provide a flammable vapour in air.) The flash point of petrol is less and 0°C (or your car would not start on

a cold day). Note: - “**flash point**” is a physical ignitability test carried out on potentially flammable fluids using a specially designed piece of apparatus – it has nothing to do with “flashing fluids”.

Typical inerting gas concentrations are as follows for Nitrogen:

N ₂ /Air	N ₂ % in Air	equivalent v/v
Methane	38%	12.9
Ethane	46%	11.2
Propane	43%	11.8
Butane	41%	12.3
Ethylene	50%	10.4

Table E 4.2 Dilutions for inerting with N₂

The flammability diagram is only a means of showing the flammable regime for oxygen and fuel. Yet, ignition is still necessary - it has to be the correct energy and intensity. The energy varies throughout the diagram, at the edge of the flammable regime it is very high but there is a minimum value near to the stoichiometric regime which produces the minimum levels. The following apply for Air:

Fuel	Energy millijoules
Hydrogen*	0.019
Methane	0.29
Ethane	0.24
Propane	0.25
Butane	0.25
Ethylene*	0.12
Acetylene*	0.02

Table E 4.3 Ignition Energy for Various Fuels

Once again hydrogen plus the unsaturates appear to fall out of line.

0.25 millijoules may or may not seem much energy - it is equivalent to dropping 1 bag of sugar about 25 centimetres; the energy in the bag of sugar is present but **not** the intensity so ignition usually requires high temperature localised energy or high intensity. When the car does not start on a cold winter morning - try to think about ignition energy and the flammability envelope. These may help to diagnose the fault - failing this, kick the car!

Taken that the composition has to be correct, the ignition energy has to be correct and at the correct location, it is not surprising that only a small percentage of all leaks actually ignite (fortunately).

Site of Leakage	Possible Cause of Leakage
Piping	Corrosion; erosion; fatigue; metallurgical failure such as <i>stress corrosion cracking</i> ; physical over pressure; physical over load due to lack of adequate supports; impact
Vessels/Equipment	As above
Reactors	Poor ventilation; poor enclosures
Equipment opened for maintenance	Poor preparation; poor isolation; poor procedures
Jointing & Flanges	Fair wear and tear; attack by internal fluids; incorrect fitting; poor joint alignment; poor bolting; corrosion of the bolting
Vents	Use and abuse; inadvertently opened. <u>Vents should be blanked off when not in use.</u>
Drains	As above
Pump Seals	Wear and tear; failure of the seal or radial bearing collapse
Compressor Seals	As above; or seal oil failure
Filters	Use; poor procedures; poor isolations
Sample Points	Sample point left open
Maintenance	Poor preparation; poor standards of isolation
Tank Breathers	Poor location; process upset and volatile fluids
Drum filling	Poor ventilation and controls
Analyser Houses	Inherent with analysers; poor ventilation

Table E 4.4 Sources of Leakage

See also Part D on Design and Hazardous Area Classification

Sources of Ignition

There are many causes of ignition too many to itemise here. The more common ones are formulated below:

Auto Ignition

The ignition of a fuel by contact with hot materials such as superheated steam mains, hot bearings, welded metal or the like. The values are to be found in the literature but usually are over 200°C and many over 350°C. The quoted values are "ideals" and just a wind cooling the surface or a layer of dust or dirt may mean that the actual surface temperatures are lower than anticipated. Don't rely on wind or dirt, but it is fair to note that auto ignition at the quoted temperatures is difficult to achieve.

Chemical Reaction

Oily rags can burst into flames due to slow exothermic reactions. The same is true of oil soaked lagging, iron sulphide (Pyrophoric Iron or FeS₂) and other reactive substances. The build up is exponential and may take some time to be obvious but once it is warm the rate of temperature rise can be quite dramatic at the end of the process and ignition takes place with no warning.

Compression

See every diesel car! But beware if equipment containing air is pressurised with hydrocarbon (pipelines or pigging equipment). The compression of air and fuels to about 20 bars can cause ignition - so ideally purge out the air with inert gas such as nitrogen.

Pressure (Bar)	Temperature (Max) °C
10	280
20	379
30	461
40	525

Table E 4.5. Adiabatic compression of air - Temp °C at P bar.

Flame/Spark

These include welding, chipping, grinding, electric fault - these are only indicative but give the general clue - do not forget bearings (rolling) or hot drive belts. It is for these reasons that there is a requirement for **Permits to Work** and **Hot Work Permits**, see Part F

Permanent Electrical Equipment

Fixed, permanent electrical equipment on plants handling potentially flammable materials has to be "**fit for purpose**". This means that the likelihood of flammable materials has to be assessed, the areas identified, the likelihood of the flammable gases being present assessed and finally the appropriate electrical equipment specified and installed. Fuller discussions are to be found in Part D Design for Safety, **Hazardous Area Classification**.

Static Electricity

Static electricity is formed when there is charge separation by two-phases moving relative to each other.

Solid/Solid
Solid/Liquid
Solid/Gas
Liquid/Gas
Liquid/Liquid

[Gas/Gas is not an apparent source of static]

Static electricity has many causes/sources - in general when high resistivity fluids are sheared due to flow or pumping static can be generated. The threshold resistivity for fluids that might result in the formation of static electricity is **about** 10^{12} ohm centimetres; this includes diesel oil but tends to exclude fuel oils. Two phase flow - solids, flashing flow or water in oil enhances the generation of static electricity. Static can accumulate on unearthed lagging, ball valves, filters and even humans. Static can also be caused by splash filling, jetting water, CO₂ extinguishers, steam lances, water settling through hydrocarbons and many more. The message is that static is a very potent source of ignition.

Ignition by static electricity probably occurs more frequently than most people believe. When all other sources of ignition have been eliminated it is assumed that the ignition source "had to be static electricity" but in reality it may have been the main source even when other sources were present.

The two classic forms of static ignition are:

Lightning – where liquid as rain drops (or solids as hail) are cycled by strong upward thermals in a cloud. Charges accumulate in the cloud and static electricity, as lightning, discharges between clouds, within the cloud or from cloud to earth.

Humans – where solids such as clothing rubs on the body or shoes scuff on carpets (both are solid/solid). In the case of human induced static it should be noted that it only occurs on dry days or in air conditioned buildings where the relative humidity is less than about 60-70%. With higher humidity the charge tends to leak away and can not accumulate.

The human body has a capacitance of about 200 Pico Farads and can be charged to about 10,000 volts. The charge energy ($1/2 CV^2$) is about 10 milliJoules which is more than sufficient to ignite most gas/air mixtures – in fact, as a demonstration; a Bunsen burner was ignited by an experimenter wriggling in a plastic coated chair! The shock when touching a filing cabinet or the 'click' from the discharge means that the discharge was almost certainly an incendive spark with ignition potential.

Another variation is the static generated on a plastic comb following vigorous combing of dry hair (again solid/solid) which can produce small sparks or the charged comb can be used to "lift" small pieces of paper.

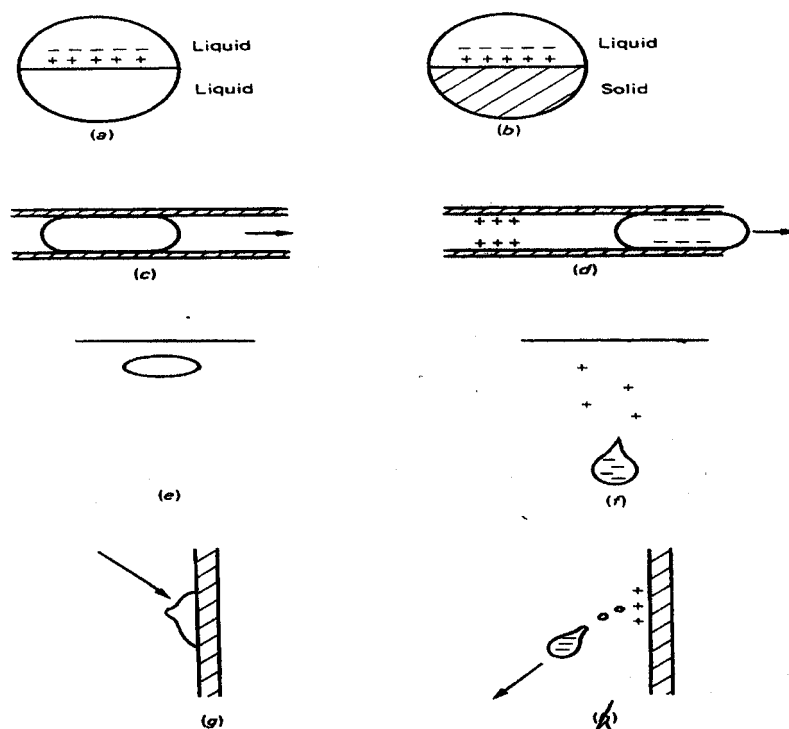


Figure E 4.3 Generation of static electricity in liquid systems: (a) electrical double layer at a liquid-liquid interface; (b) electrical double layer at a liquid-solid interface; (c, d) Charge separation as oil flows through a pipeline; (e, f) charge separation as a water drop falls through oil; (g, h) charge separation as oil splashes on a tank wall. As shown by F P Lees

Liquid systems

The figure above shows the phase and charge separation process and charge accumulation.

Figures (a) and (b) are general models but figures (c) and (d) show fluid flow on a pipeline (solid/liquid), figures (e) and (f) show one phase dropping through another (liquid/liquid), figures (g) and (h) show a droplet hitting the wall of a tank.

The charge can accumulate inside the fluid or be given up to unearthed conductor such as a metal filter in the line or the ball valve with soft plastic seats. In these examples the spark is between an insulated conductor and an earthed conductor such as the pipeline itself. It is normal to earth **all** conductors where charges may accumulate.

A variation of this is the filling of a tank with a non-conducting fluid. The fluid itself is charged and only discharges slowly (charge relaxation) to the walls. If the fluid is “dipped” (to determine its level accurately) with a metal dip tape there could be a local discharge to the tape and a spark causing a fire or worse, a tank explosion. It should be noted that fuel road tankers have a wooden dip rod – not metal – for this reason.

(It is worth noting that “switch filling” tankers are a major source of ignition. In switch filling diesel is put into a tank which has previously contained petrol. The diesel carries the charge and ignition takes place. Think of the filling of Road Tankers and of course the “garden 5 liter containers”!)

Slurry systems, where solids are transported in a turbulent regime on pipes, are even more potent courses of static electricity as there is a complex phase movement:

Solid (slurry) and Liquid
Liquid and Piping
Solid (slurry) and Piping

Splash filling

Splashing organic fluids can generate static charges ((g) and (h) above). It is normal to introduce liquids **below** the liquid surface to prevent this; filler pipes for petrol/diesel are also electrically earthed through the hose to prevent charge build up.

Droplets

Lightning is one obvious source of droplets causing charges.

Water droplets draining through non-conducting fluids can also cause charges to accumulate so it is essential that care is taken when water settles out of non conductive fluids, such as wet diesel oil (figures (e) and (f)).

Flashing process fluids can produce charged clouds. The cloud then induces a charge on an unearthed conductor, which then results in a small spark where it discharges to an earthed body. Flashing fluids bearing solids as slurries are even more powerful sources of static electricity.

Wet steam from steam lances have been known to ignite organic fluids and under certain circumstances carbon dioxide fire extinguishers can produce sparks from the discharge horn.

Variation on a theme

Some critical properties of solids and gases are given below. The properties of organic vapours are similar to solid systems.

Fuel Solid Powder Form	Lower Concentration Kg/m ³ (g/m ³)	Minimum Ignition Energy mJ
Aluminum	0.045 (45)	50
Magnesium	0.03 (30)	40
Sugar	0.045 (45)	30
Wheat flour	0.05 (50)	50
Polystyrene	0.02 (20)	15

Table E 4.6. Some Properties of Flammable Solids

Fuel Gas	LFL V/V%	UFL V/V%	Minimum Ignition Energy mJ
Acetylene	2.5	100	0.02
Ammonia	1.5	28	40.0
Benzene	1.4	8	0.22
Methane	5	15	0.29
Ethane	3	12.4	0.24
Propane	2.1	9.5	0.25
Ethylene	2.7	36	0.12

Table E 4.7 Some Properties of Flammable Gases

Also note vapour is given as v/v and the solid wt/volume, the conversion from vol/vol to wt/vol gives a minimum gas concentration in the range 0.03 to 0.05kg/m³. This is almost identical to that of solids.

The maximum contained over pressure for confined explosions of gases are in the range 900kPa to 1MPa and for solids the range is 500kPa to 800kPa. (See Part E 5)

Grinding/Milling

When solids are milled, solids move relative to solids and charges can accumulate on the solids; these can result in sparks.

Grain silos

Grain is often pneumatically conveyed into storage silos. The silo can also contain grain dust or starch and if the grain is charged there can be a major explosion inside the silo with horrendous consequences.

Belts

Slipping drive belts between a pulley and a motor or slipping conveyor belts can generate incentive sparks either through friction or static electricity.

Filters

Dust filters can become charged during filtration or during discharge of the solids from the filter. If the solids are organic shaking the filter or back blowing the filter can create sparks.

Case Histories to illustrate the perversity of ignition

Case histories can sometimes be serious or humorous.

Fire steam

An operator was trying to inert a leak of hot benzene with fire steam blanket using a fire steam lance to exclude air and so to prevent ignition and a fire. Unfortunately, the fire steam was wet and generated a spark which ignited the benzene, the very event the operator was trying to prevent.

Emptying a polythene sack

An operator was emptying a sack of off-specification polythene into a silo, the sack contained some ethylene (ethene) and the movement of the polythene (solid) out of the polythene sack (solid) resulted in a spark which resulted in a minor fire. The operator was not injured but was left holding the two corners of the sack – the rest of the sack melted in the short lived fire.

Sampling

An operator took a sample of organic fluid in a steel pail – splash filling. The handle was insulated by a plastic coating so a charge built up on the pail and the spark between the pail and sample point caused the fluid to ignite – much to the operator's surprise.

Pump and Seal Leak

The seal of a pump handling hot organic fluid containing a solid catalyst failed catastrophically. The hot fluid flashed and created a small charged cloud, this induced a secondary charge on an unearthed sheet of metal pipe cladding, which in turn discharged to an adjacent earthed section across a gap of only 1 millimeter. The spark across this narrow gap ignited the cloud. (In another incident the bearing of a pump failed due to loss of interference fit, the shaft started to precess so allowing the seal to leak. The fluid then ignited in the hot bearing).

Unearthed Valve

A ball valve was fitted with plastic seats. The ball became charged by the organic fluid flowing in the line and a spark discharged between the ball valve handle and the pipe line. After a period of time the arcing process cut a small hole in the pipe and the leak was ignited.

Drum Filling

A spirit drum (45 gallon) was being filled on a trolley fitted with nylon wheels. The drum was therefore insulated. The filler pump was started but the filling was delayed a few minutes during which time the spirits were heated and sheared in the pump. When the filler nozzle was put into the drum and the filler opened there was an explosion in the drum. It was believed there was a small spark between this pipe and the charged drum. One person was killed.

Steam Leaks

An operator walked through a steam cloud from a steam leak. The operator received a nasty shock when he touched a hand rail.

A Plastic Coated Hand Rail

A hand rail on a stair was coated with plastic to prevent corrosion. The operators found that they became charged as their hand slid down the hand rail and then received a shock when they touched an earthed

structure. The operator found a solution – do not use the hand rail – but one slid on the stair and was injured. This is one example of a poorly specific protective system causing an injury.

Solutions

There are many strategies which can be used to prevent static electricity or the effects of an electrical discharge; each has to be used in a case-specific manner:

- Eliminate the “flammable regime” by inerting the system to exclude oxygen (look at the flammable envelope diagram Figure E 4.1 and Table E 4.2.)
- Modify the fluid conductivity by adding special conductive fluids.
- Earth and bond all equipment to prevent charge accumulation (charges will still take time to relax in large tanks).
- Inert all storage tanks.
- Reduce the transport velocity to less than 1 m/s – this is not always practicable.
- Avoid splash filling and/or setting water.
- Earth all personnel with special footwear.
- Avoid the use of earthed dip tapes.

Flame Shapes and Radiation Assessments

Flames can be considered in two ways. The first is to treat the flames as a point source from which all of the radiant heat emanates. The second is to consider the flame as a solid body which radiates heat uniformly from all over its surface. The surface can be anywhere from a Grey Body to a Black Body radiator - this will be explained later. The main problem in assessing radiation is that the amount of heat radiated is very uncertain and is almost a "guesstimate" although tables of values based on research work using radiometers are available. It is necessary to bring in two small pieces of theory to assist understanding.

Heat Balance

There is a heat balance round a flame.

The heat in is: the fuel burn rate x (heat of combustion - losses due to partial combustion). The heat out is: the heat in the products of combustion + Radiant Heat.

Radiation Laws

The heat radiated from any m² of flame surface (called the **Surface Emissive Power** or **SEP**), is the standard Stefan-Boltzmann equation as follows:-

$$\phi = \frac{W}{m^2} = \sigma \epsilon T^4 \quad (\text{E 4.2})$$

- where σ = a constant ($5.67 \times 10^{-8} \text{ w/m}^2 \text{ K}^4$)
 ϵ = is the emissivity (or "blackness")
 T = is the absolute temperature ($^{\circ}\text{K}$)
 ϵ = 1 for a black body and can be as low as 0.1 for a translucent flame.

All values are in consistent numbers.

There is one other modifier which is not essential at this level of discussion but may be important in major studies. This is the **atmospheric absorption** of heat by the water vapour and carbon dioxide. In most cases it is very much second order.

Figure E 4.4 shows a graphical representation of the heat balance round a flame boundary. The top, horizontal, line is the TOTAL amount of heat available in the ideal combustion process. In reality not all of this heat will be released as there will be some partial combustion resulting in the formation of soot and carbon monoxide. This is indicated by the horizontal 50% line.

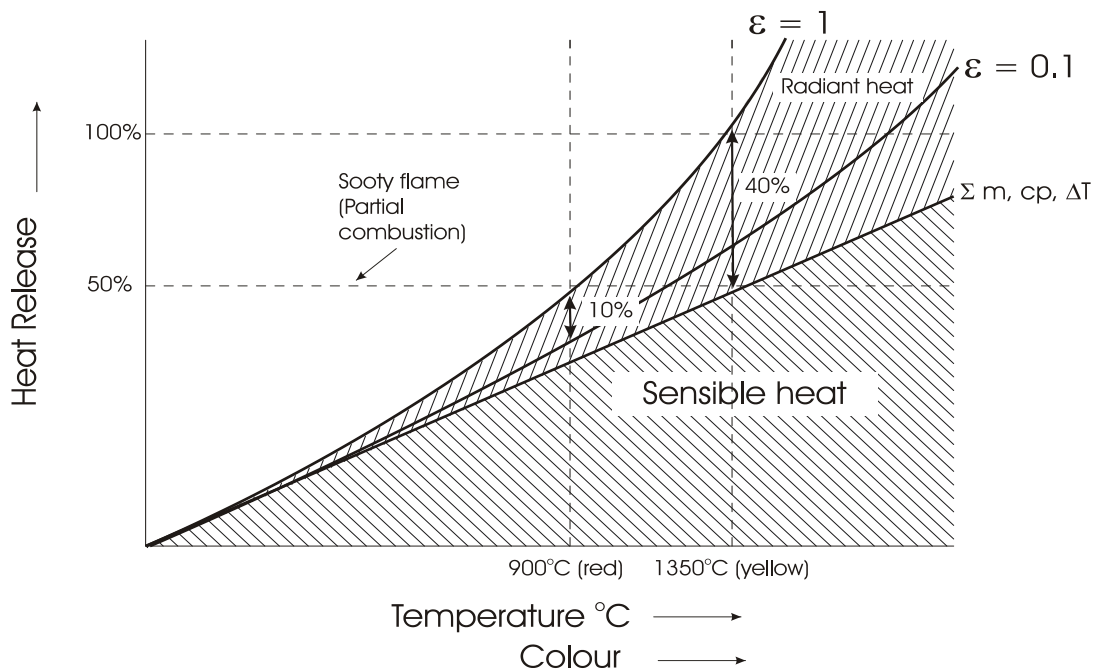


Fig E 4.4 Heat Balance in a Fire

The straight, inclined, line encloses the sensible heat losses which will also be determined by the excess air but to the first approximation can be determined by: $\Sigma [m_1 cp_1 \Delta T]$: where the symbols have the traditional meaning, cp (specific heat) will tend to change with temperature but for this model it can be treated as a constant. The curved line represents the radiative heat losses and follows the equation E 4.2. But T and the emissivity, ϵ , are not fixed.

Taking all of this together the amount of heat radiated from a **perfect** black body flame in **perfect** condition will probably not exceed 40% and the flame temperature will not exceed about 1450°C .

However if there was an imperfect conditions such as incomplete combustion the temperature could be as low as 800°C if it is very sooty due to incomplete combustion. This can be demonstrated by lowering the total combustion line to the 50% line. By changing the shape of the "radiative" curve (E 4.2) the effects of emissivity can be demonstrated. Fuels, which are rich in hydrogen, do not produce soot. It is the soot which is the black body radiator. The emissivity of gases is significantly less than unity (neared 0.1) and so again the area between the radiation line and the sensible heat line is reduced. The flame temperature has to increase to affect the heat balance. There is a slight converse situation in that the *NON-SMOKEY* flames are hotter but have a lower level of radiated heat. The less radiative the flame so the hotter must be the flame to release all of the heat of combustion.

Typically a methane flame will radiate about 10% maximum of the heat of combustion but have a flame temperature of about 1600°C. On the other hand a smoky flame may only radiate 10% of the total heat of combustion but the temperature may be as low as 800°C. On the other side Oxy-propane or Oxy-acetylene cutters do not have the inert Nitrogen to carry the heat of combustion and the burner is designed for good combustion so the flame is well into the blue end of the spectrum (even going clear or white) and the flame temperature is nearer 2,200 °C.

Flame temperatures can be judged by colour:

Straw Yellow = 1350°C

Cherry Red = 850°C

Blue = 1600°C

Remember the colours of the rainbow! Red, Orange, Yellow, Green, Blue, Indigo and Violate. The sequence still works.

Unfortunately the flame surface is not uniform in colour or texture so the TRUE amount of heat radiated from the surface must be assessed using a radiometer. On the other hand some reasonably accurate assessments can be made knowing the amount of flame covered by soot, the visible flame area and the flame colour.

Flame Shape

The flame shape is traditionally a cylinder or tilted cylinder or cone or tilted cone. The base of the cone is traditionally a circle but in the bund of a tank farm or in a gully it must be rectangular and a hydraulic mean diameter "D" applies. This can be calculated from the following

$$D = 4 \times \frac{\text{Pool Area}}{\text{wetted circumference}} \quad (\text{E 4.3})$$

The diameter may also be defined by the balance of the combustion rate and outflow (volume terms) such that

$$\text{Outflow Kg/sec} = \frac{\pi D^2}{4} \times \dot{m} \quad (\text{E 4.4})$$

$$\dot{m} = \text{Burn rate } \text{kg/m}^2/\text{sec}$$

The burn rate (\dot{m}) can be found from tables but is typically about 0.1 Kg/m²/sec but for known fuels the value derived from E 4.5 is more accurate.

$$\text{Burn Rate } (\dot{m}) = 0.001 \frac{C_v}{L_H} \text{ kg/m}^2/\text{sec} \quad (\text{E 4.5})$$

Where C_v is the calorific value and L_H is the latent heat of evaporation of the fuel – kJ/kg.

This correlation is shown below for various fuels.

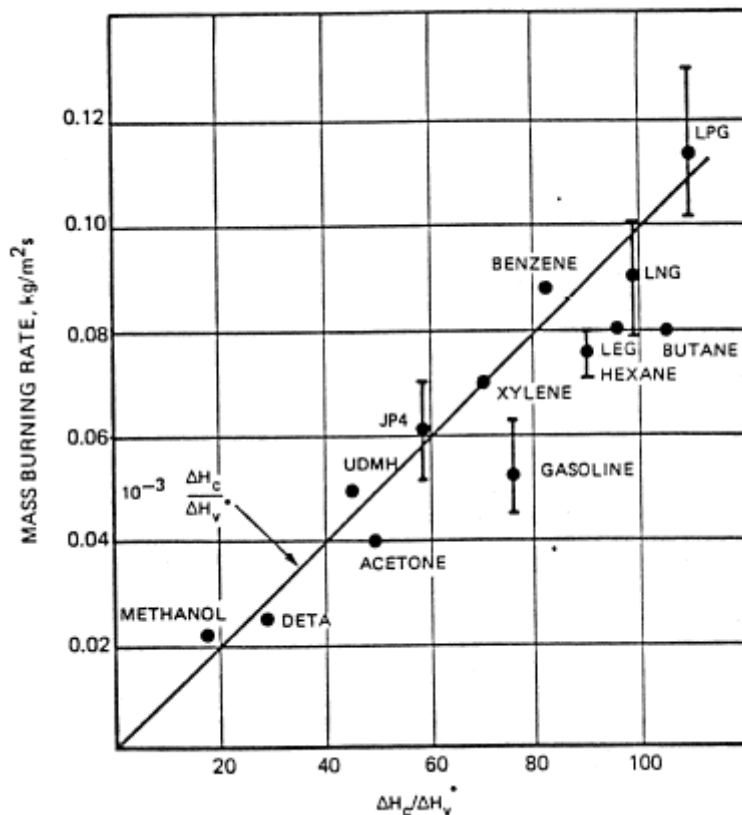


Figure E 4.5 Burn Rate for Various Fuels

Where ΔH_c is heat of combustion kJ/kg and ΔH_l is Latent Heat of vaporisation kJ/kg. **Beware** when there are multi component fuels. ΔH_c and ΔH_l (heat of combustion and latent heat of evaporation - kilojoules/kilogram) may be variables with time. When fuels fall on water ΔH_l may also be higher due to the heat lost to the water heat sink. This will reduce the burn rate. (Note that in the diagram above the latent heat of evaporation is defined as ΔH_v)

The "diameter" of the fire can now be defined either by

- 1) Confines, such as a bund.

- 2) Confines, such as a gully.
- 3) Balance of outflow rate and burning rate.

The height can now be defined by a tried and tested equation:-

Where:

$$Height = Diameter_e \times 42 \times \left(\frac{\dot{m}}{\rho_a \sqrt{g_c D_e}} \right)^{0.61} \quad (E 4.6)$$

ρ_a = is density of air = 1.22 Kg/m³

g_c = 9.81 m/sec²

\dot{m} = Burning rate kg/m²/sec – see equations and figures E 4.5

D_e = equivalent diameter – see E 4.3

This describes a vertical right cylinder which is appropriate these basic notes. There are also equations for tilted cylinders but this is more advanced.

The shape of the flame is now fully defined. But there are some other features which need explanation, so as to deal with the risk assessment.

Fire Balls

The fire ball is typically the **Boiling Liquid Expanding Vapour Explosion** or **BLEVE** caused by a bursting tank or pressurised container. The volume can be defined by a sphere whose Diameter D and life T in seconds is defined by:

$$D = 5.8 W^{\frac{1}{3}} \text{ metres} \quad (E 4.7)$$

$$T = 0.46 W^{\frac{1}{3}} \text{ seconds} \quad (E 4.8)$$

D = Diameter (metres)

T = Duration (secs)

W = wt of fuel in the fire ball (Kg)

Solving these gives a burn rate of
0.2kg/m²/sec

Torches

Torches can be defined as an acute cone:

$$L/D - \text{length/diameter} = 1/10$$

where: -

$$L = 10 W^{0.46} \text{ (gas)} \quad (\text{E 4.9})$$

$$L = 17 W^{.46} \text{ (fluids)} \quad (\text{E 4.10})$$

$$W = \text{outflow (Kg/sec)}$$

This is a fairly advanced study and is only given for completeness.

Utilising Data to Access Heat Fluxes

It is now necessary to assess the "view factor" F_{12} for a flame. Most of the values of F_{12} are obtained from the equation:-

$$A_1 F_{12} = \int_{A_1} \int_{A_2} \frac{dA_1 \cos \theta_1 dA_2 \cos \theta_2}{\pi R^2} \quad (\text{E 4.11})$$

The graphical solution of equation E 4.11 is shown below, where H = Height, R = Radius and X = distance from the flame centre (all in consistent units).

There are a whole series of plots for tilted cylinders but this is a level of complication that is not appropriate to this part.

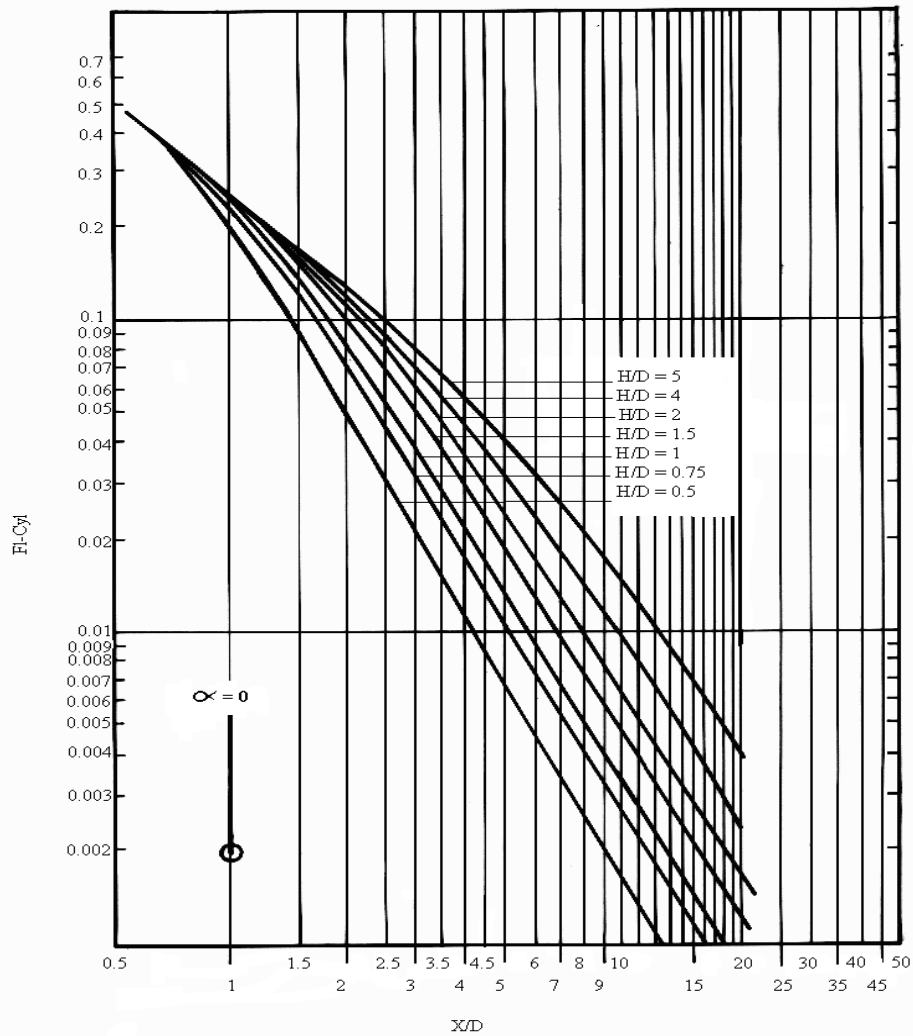


Fig E 4.6 View factor for a vertical cylinder

Note: This plot is in log v log notation not linear.

Fortunately F_{12} can be found as solutions in literature - for a fire ball F_{12} is

$$\left(\frac{D}{2x}\right)^2 \quad (\text{E 4.12})$$

where x is distance and D is the fire ball diameter.

There is one minor extra to be considered - that is atmospheric attenuation - or the reduction of radiation due to water vapour. This value is often called attenuation. For **most** real events it has a value of about 0.9 so it is **not** significant.

With knowledge of F_{12} it is now possible to assess the flux at point x.

$$\epsilon_x = F_{12} \epsilon \quad (E 4.13)$$

where: -

ϵ_x is flux at point x $-(kW/m^2)$

F_{12} is view factor

ϵ is Surface Emissive Power of the source (kW/m^2)

Note: View factors are additive. A flame on a tank can be treated as:-

$$\text{View Factor}_{\text{flame}} = \text{View Factor of (flame + tank)} - \text{View Factor (tank)}$$

At $X/D = 10$ (just for convenience of reading the plots) a tank $H/D = 1$ will have a view factor $f_{12} = 0.004$. Likewise for a tank $H/D = 1$ and a flame on top of the tank $H/D = 2$ (total $H/D = 3$) the view factor $f_{13} = 0.011$. The view factor of the elevated flame $f_{23} = 0.011 - 0.004 = 0.007$. This is nearly in ratio 1:2 but the proper method does take into account the elevation and the change in the angle as viewed by the receiver.

The one missing piece of information is the Surface Emissive Power. This is tabulated in Table E 4.8

Values of F and SEP

Pool Fires			
	F	SEP	T°C
LNG	0.2	200	1300
LPG	0.1-0.2	100	1300
NAPHTHA	0.1	75	1000
FUEL OIL	0.075	50	900
Torch Fires			
LNG	0.15	200	1400
LPG	0.3	350	1350
BLEVE			
Any - other than CH ₄ or H ₂		250	1250

Table E 4.8 Values for F used in equation E 4.15

Limitations

The calculation of thermal radiation profiles has many potential errors:-

1. Wind lean effects under the influence of the wind. (See photo E 4.1)
2. Flame temperatures - locally or overall (See photo E 4.1)
3. Emissivity (Black or Grey Body) (See photo E 4.1)
4. Bunding/Pool confines
5. View Factors for an irregular shape
6. Temperature variations across the flame. (See photo E 4.1)

Are These Models Realistic?

The answer is **yes** and **no**! It is most unlikely that there will ever be a true vertical flame. This only occurs on a completely dead calm day – not very often in Britain! However the reality of a fire attack is that this is a simple model which will probably err on the side of safe. No Fireman would approach a fire from the downwind side they will always approach from the upwind direction. The hot products of combustion would be intolerable if not also toxic, remember the combustion of plastics can produce Hydrogen Chloride, Hydrogen Cyanide and Phosgene and of course Carbon Monoxide, all of which are toxic. Further the lean “away” from the Fireman reduces the view factor so reducing the heat flux. Finally and equally important is that a fire hose/monitor is more effective when the water is carried by the wind and not injected into the wind.

Point Source Model

The most simple model for a fire is the point source model. In this all of the heat is considered to radiate from a "point".

$$Q = 4 \pi x^2 \phi_x \quad (\text{E 4.14})$$

Where:

Q is the total heat radiated (kW)

x is the distance from the point to the receiver

ϕ_x is the flux at the receiver (kW/m²)

and

$$Q = F \times M \times C_v \quad (\text{E 4.15})$$

where: -

- F = fraction of generated heat which is radiated
- M = combustion rate (kg/sec)
- C_v = calorific value, (kJ/kilogram)

This simple model is quite useful where the distance between the source and the receiver is **at least** 2 x main dimension (length) of the flame. It should be noted that the view factor plots shown in figure E 4.6 approximate to the inverse square law (equation E 4.14) at a distance where X/D is over about 5. The limitation is now the values of F and C_v.

Effects on Humans/Equipment

The effects of fires on humans can be viewed two ways; the first is external and the second internal. See Part G for a more detailed description.

External

The limits of thermal radiation are defined by the dose equation

$$\sum_i Flux_i^{\frac{4}{3}} Duration_i = Dose (kw/m^2)^{\frac{4}{3}} secs \quad (E 4.16)$$

This means that as a person runs away the flux, and hence dose, is dropping all of the time but the dose is still being accumulated. Some references have suggested that 1 Kw/m² be removed from the flux to allow for human tolerance - or else no one could go out in the sun! There is also the burning effect of the hot gases.

In the final assessment individual response differs between persons, but the following will apply:-

Dose	Effect
(kW/m ²) ^{4/3} secs	-
250	Pain
1050	1% Fatality
2080	50% Fatality
1400	2nd Degree burns

Table E 4.9 Human Response to heat doses

Flux Kw/m ²	Tolerance
1.9	8 hours still weather
2.5	8 hours gentle cooling breeze

Table E 4.10 Human Response to low level heat fluxes

Internal

Hot products of combustion are potentially toxic if breathed in the heat in the hot products can damage the bronchi and the alveoli. Further smoke particles can clog the alveoli in the lung and require a long recovery period to remove them in the phlegm. The nature of the fire will affect the nature of the toxic; synthetics produce cyanide, phosgene or acid gases such as HCl. Heavy fuel oils may produce 5% v/v CO if poorly aerated or 0.5% v/v CO if well aerated and persons have died as a result of CO poisoning.

Equipment

Fluxes of up to 100 Kw/m² heat flow into flame-licked equipment have been measured. The simple energy balance suggests that flame affected metal will rapidly heat up to 450°C where much of its strength or rigidity will be lost. Values of 300 Kw/m² have been measured for torches.

Woodwork or clothing may spontaneously ignite at radiation levels 12 Kw/m² but steel should be tolerant.

Mitigation

See also under Part D - Design for Safety

The obvious mitigation is to avoid the problem in the first place - this can be achieved by preventing ignition (see earlier) or preventing leakage by good design, maintenance and monitoring - not always as easy as saying it that way.

Thereafter, the obvious mitigations are protection of equipment/structures in the form of lagging or by cooling with water.

There are others, which are practical solutions to theoretical problems:-

- Reduce the pool size by bunding or sandbags (see the equation for flame height)
- Drain pools of flammable fluids **from** the equipment by casting concrete with a slope away from equipment into drains or a safe area.
- Isolation and depressurisation of equipment, no pressure = no flow.
- Apply "**Fire Protection**" active or passive to vulnerable equipment or structures. Active is water deluge and passive is physical insulation. Remember that the physical protection can only last so long as it holds on to the structure. (See table E 4.11 - later.)

Detection/Protection (See also Part D Design)

The following is a list of possible fire protection and detection mechanisms:

Protection	Detection
Water which cools the fire assaulted surface	P.O.C. (Products of Combustion) - the household fire detector is a P.O.C. (The products are electrically charged particles)
Cement which physically insulates the fire assaulted surface	Temperature which cause a bi-metallic strip to change shape and open a circuit
"Insulation" as cement but is usually an industrial heat loss lagging	IR/UV which are properties of the flame and can be detected by the appropriate instrument
Intumesced Paint which intumesces on fire assault and produces an insulation barrier	Plastic Tube which melts or fuses and vents a pressurised air source so activating a low pressure switch at the extreme of the tube
"Epoxy Coatings" – such as Pyrocrete/Chartek which again “chars” and produces an insulating barrier	Eutectic Insulated Wire where the eutectic salt insulates two conductive wires. When heated to about 75°C the eutectic melts and opens an electrical circuit between the two wires
Grading the Concrete which drains the fuel away from the fire site	Eutectic Solder which melts and allows a fire door to close
Bundling which confines the fire spread and reduces the size	Quartzoid Bulb which bursts at about 60°C and vents a pressurised source – often water as a sprinkler
Draining which removes the fuel and reduces the fuel which can be burned	Opacity caused by smoke in a corridor or a warehouse. A light beam is shone onto a distant detector.

Table E 4.11 Fire Detection and Protection

These are essential Design Details See Part D

As the heat flow in a pool fire is about 75 Kw/m² and a fire water rate of 10 litres/min/m² of water is required (the torch flux is 300 kW/m² and requires special water sprays). Common thermal lagging with well designed retention can also be of use.

E 5 Explosions



Photo E 5.1 A Gas Phase Detonation

Note:

- 1. This is actually a detonation as the shock cells are visible**
- 2. The difference between this and an Gas Phase Explosion (below) is marginal other than the peak over pressures**



Photo E 5.2 A Gas Phase Explosion

Note the different flame boundary

Introduction

Explosions are relatively rare but when they do occur the damage is usually extensive and can extend into the public sector. Therefore their potential is not only on site but also off site. The resultant cost of an explosion is major and includes site restoration, loss of sales and probably litigation if property outside the site is damaged.

Explosions are no more than the very fast release of energy. On Piper Alpha in July 1988 about 75 kg of fuel was consumed in less than 0.5 second giving an instantaneous rate of 7500 M Joules/sec or 7500 M Watts, the residual fire prior to the riser rupture released heat at about 65 M Watts. The difference between a fire and an explosion is the rate and density at which the energy is released. This simple statement may seem surprising but it is the key to differentiating between explosions and fires. In a liquid based fire the energy can only be released if the fuel is supplied either as an atomised jet or by the liquid vaporising in (from) the pool due to heat input from the fire itself, therefore the flame has limitations set upon it by the ability of the oxygen to diffuse into the combustion process – these flames are called “*diffusion flames*”. (These are different from a furnace flame where the oxygen (in the form of air) is injected into the combustion zone and the radiation from the furnace assists in the evaporation of the fuel.)

In the case of a flash fire there may be a significant potential cloud of fuel - some kilograms - but the flame does not have a “*feed back*” mechanism so there is no explosion – see later – and the flame may only move at a few metres per second.

It is clear that there are pre-requisites for an explosion. Either there must be some form of **confinement** such as a compressor house or some form of **turbulence** which acts as a *fast feed back* in the reaction process. In reality the two occur together in most process area. Walls round reactors act as confinement and piping and equipment create turbulence. The explosion in the Polythene Plant at ICI Wilton in 1969 and that on Piper Alpha were caused by **confinement** while the explosion at Beek in Holland and Flixborough in England were caused by **turbulence** plus localised **confinement**.

The drawing E 5.1 shows the turbulence (as lines emanating from the rear of pipes) resulting from a rapidly advancing flame front on the right. This turbulence mixes the fuel/air mixture and instead of the flame depending upon diffusion it has turbulence to enhance the combustion process. In addition the turbulence at the flame front mixes in the free radicals from the combustion process, and as a result there is a high feed-back resulting in flame acceleration which eventually runs up to an explosion.

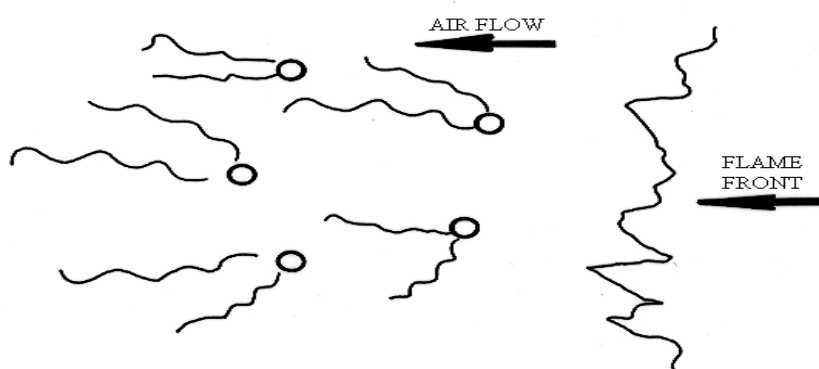


Figure E 5 .1 Massive Turbulence

The explosion history is not as simple as might be expected. See figure E 5.2. which is for a typical military or “condensed phase explosion”. The positive pressure phase is a sharp pressure rise and decay which eventually becomes a negative pressure phase marked “I”. This can produce some rather odd effects such as glass “sucked” out of a window.

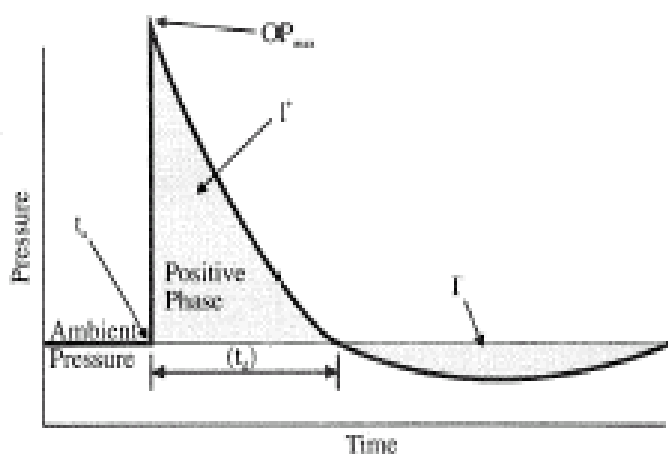


Figure E 5.2 Time v Pressure History for a military Explosion

In a “vapour cloud explosion” the shape of the positive wave is more like a triangle, see figure E 5.3. However there is still the negative pressure phase.

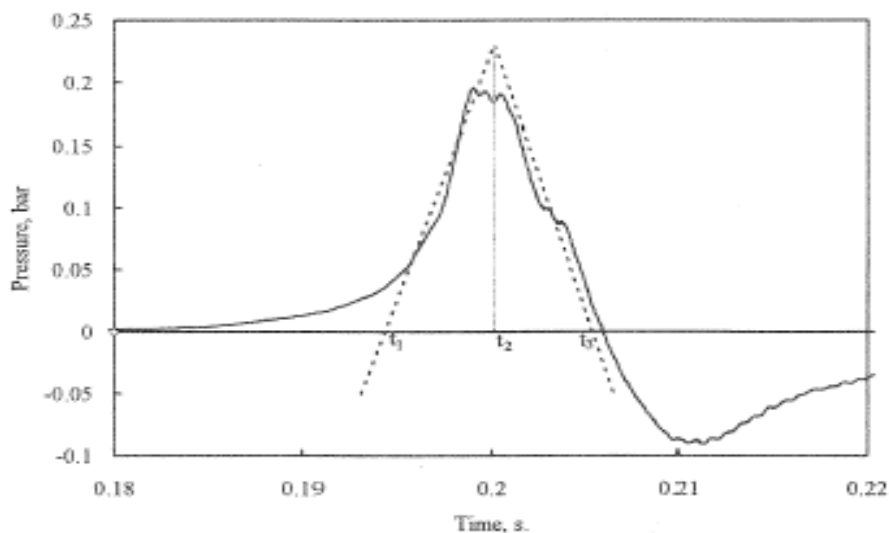


Figure E 5.3 Pressure v Time History of a VCE

There are other forms of explosion which should be recognised. They are not an integral part of these notes but they should be recognised as potential problems. These are given below:

Bursting Vessels

There are equations for the energy released when vessels burst, this needs some interpretation. Essentially about half of the stored energy is released as blast and half is transferred into flying fragments.

Rapid Phase Transformation

This is a rapid, almost instantaneous, boiling of superheated fluids. Consider melted metal quenched in say water, the water may become locally superheated (over 100°C) and then suddenly nucleates on metal sites and boils explosively. Similar effects can be obtained by releasing cryogenic fluids under water, the water is the heat source and the fluids boil explosively. The theory is outside the scope of these notes.

Compressor Houses

The explosion in a compression house (such as Piper Alpha) may contain from 50 to 250 kg of gases and if it were to be ignited it would result in a large explosion (due to the confinement). In this case they are confined and the overpressure is limited by "**blow out panels**". Such a cloud in the open may cause no more than a "pop". Large Vapour Cloud Explosions in open air process plant - now called VCE - may require **1000 kg** or more of fuel before there is an explosion. It is reasonable to assume that this is the threshold and smaller releases are less critical **but only in the "open air"**.

Pressure Piling

Pressure piling may occur with a series of chambers and joining corridors, the pressure in the second chamber is pre-compressed by the explosion formed in first, upstream, chamber. The explosion therefore starts at a higher starting pressure such that if there are a series of chambers the piling may become severe. Eventually walls will fail, so limiting the final effect, but pressures will be larger than expected. Piling can be dismissed from initial studies. This is not part of these notes.

Explosions in Reactors

In this there is a violent exothermic reaction which leads to the overpressure of the vessel containing the reaction. This is a complex study which is discussed in Part D from a process stand point.

Detonations

In a **gas phase detonation** (photo E 5.1) the flame front travels at a speed greater than the sonic velocity of the unburnt gases, velocities may reach 2000 m/sec and the overpressures may reach 20-25 bar. The detonation can be achieved in very rare conditions, but it is most unlikely and can be dismissed for initial studies. Suffice it to know that under very rare conditions of turbulence **and** confinement such as pipes and with some gases detonation can occur but it is an extreme event.

Explosions in Houses

Explosions inside houses have to be modelled in a different manner. Once again the theory is outside the scope of these notes. See Ronan Point below, which was a mixture of pressure piling but then the

structure was destroyed and the building went into systematic collapse). Typically walls will blow out at a fraction of a bar overpressure. (See table E 5.3).



Photo E 5.3 Ronan Point

(See also Pressure Piling and Compressor Houses)

Military Explosives

It is appropriate to make note of the differences between gas phase explosions and the military use of explosives. The military use “condensed phase” explosives where the oxidiser is within the fuel. Peak pressures at the front can be thousands of bars. Various formulations can be made to enhance the fragmentation effects. Of course there are special bombs which carry only the fuel which is then sprayed into the area of attack and then seeded with detonators driving the flame front into a vapour phase detonation.

Explosions

TNT Equivalent Model (1)

The TNT model was the first developed for the process industry in the 1960s. It was a serious attempt to understand the problem and to model the effects of the explosion as data on TNT explosions was widespread from the Military. As with all simplifications the model falls down in the near field - that is close to the explosion and tended to overestimate overpressures so an arbitrary 0.7 to 1 bar ceiling was imposed (Over-pressures of 0.7 to 1 bar will result in massive destruction, as a result this is not a major

weakness in the model. In reality there is a fair amount of evidence to show that under certain conditions the overpressures at the epicentre **can** be well over 1 bar.)

The TNT equivalent model is not appropriate to a confined space.

In its simplest form the TNT equivalent model compares the energy release of the gas/fuel in terms of its calorific values and then equates it to a TNT charge of the same energy. The weight of TNT is then fitted to scaled effects. The process is very simple, fairly accurate, but relies on some very basic assumptions - none less than "what fraction of the release actually explodes or releases energy"? The assumption is that about 4% (usually taken as 4.2%) of the release is involved in this explosion. The weight of TNT with the same energy equivalent as the release is then: -

$$0.042 \times \frac{\text{Mass of fuel released before ignition} \times \text{calorific values of fuel}}{\text{Calorific value TNT}} \quad (\text{E } 5.1)$$

Calorific value of TNT = 4.6MJ/kg

The mass of fuel for flashing fluids has been traditionally 2 x flash percent x total release. Evidence now suggests a more realistic assumption is that low molecular weight fuels are totally vaporised; this may not be true for high molecular weight fuels. The fuel released is either known or can be assessed, for example, it is the capacity of a vessel or the maximum credible outflow for 5 minutes from a severed drain or pipe. Appropriate allowances should be made for the operation of ESD systems or the capacity of the system - the fuel can not exceed the inventory of the system! The assumption that there is a 5 minute discharge is a bit arbitrary and is tied into the arbitrary yield value – 0.042. 0.042 was a “committee number” but has some justification as some of the fuel will have dispersed before ignition, possible over 90%.

The next problem is to assess where will be the epicentre of the explosion? That can only be done by engineering judgement; the rest of the analysis is to be found under "Scale Laws".

Scale Laws

Once the TNT equivalent is known by whatever of the two methods chosen, it is possible to read off the **likely** overpressures for a graph using a scaled distance, where D is the distance from the epicentre to the “target”:

$$\text{Scale Distance} = \frac{\text{Distance } (D)}{\sqrt[3]{TNT}} \quad (\text{E } 5.2)$$

Where D is the distance in metres to the receiver of the blast from the epicentre

TNT is the equivalent charge of TNT in kg with the same potential energy release.

The word "likely" was used because, although the line is there the accuracy is not as well defined - see Figure E 5.4. This curve is based on Military explosions, but slightly modified and so may not be a complete equivalence to VCEs. Particularly the impulse and the duration may not be the same.

Note: This plot is in **Log v Log** notation not **Linear**. Misread the scales and the error potential is significant. Differentiate between the two pressure scales of Pascals and PSI.

Once the TNTe has been assessed using equation E 5.1, the scale distance can be assessed from equation E 5.2 and the overpressure assessed from figure E 5.4. From the overpressure it is now possible to assess the damage from table E 5.3.

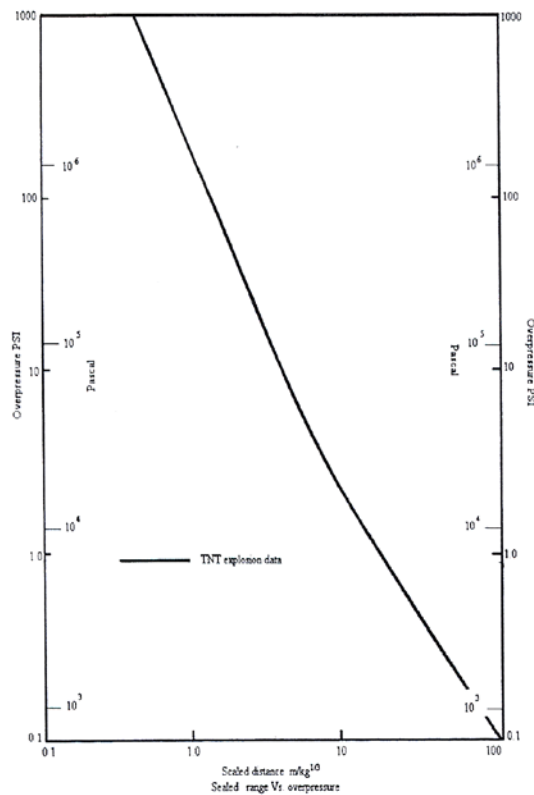


Fig E 5.4 TNT Scale Law Curve

Volume Explosion model (2)

The following is a simple model based on a fair amount of research and tested against real events.

The fuel factor K_1 (or peak, local cloud overpressure (Bar)) is derived from the relationship:

$$K_1 = k \cdot v_{flame}^{2.71} \quad (E 5.3)$$

k in equation E 5.3 is usually about 10 but varies with different researchers. 10 is the best fit for most fuels.

v_{flame} is the fundamental burning velocity of the fuel and lies in the range 0.35 – 0.85 m/s (with the exception of Hydrogen and Ethyne [Acetylene])

If in doubt plot on a log v log the overpressures and fundamental burning velocities for the two following fixed points:

CH₄ = 0.6 bar (Burning Velocity 0.35 m/s)

C₂H₄ = 3.2 bar (Burning Velocity 0.65 m/s)

Vertical axis – flame velocity m/s and the horizontal axis – Factor K_1 . Draw a straight line through these points and read off the value of K_1 for any other fuel with different burning velocity. For mixed gases add by volume/moles.

Damage profiles are shown in the table E 5.3

1) Assess the volume of the likely explosion cells separated by at least 10 metre breaks. These cells are discrete plant volumes. (The 10 m break, as at Flixborough) allows the flame front to slow down and so to disconnect from the upstream cell.)

2) Determine fuel factor – Table E 5.1 - K_1 (or the fundamental burning velocity is known use equation E .5.3)

Fuel	Fuel Factor (K_1 bar)
CH ₄	0.6
C ₂ H ₆ /C ₃ H ₈ /C ₄ H ₁₀	1
C ₃ H ₆	2.0
C ₄ H ₆	1.8
C ₄ H ₈	1
C ₂ H ₄	3.2
H ₂ /C ₂ H ₂	Detonation 20
CO	1
Aromatics	1
Ether	1

Table E 5.1 Likely Maximum Overpressures for Various Fuels Based on the Fundamental Burning Velocity

3) Determine the turbulence factor for each cell (K_2).

Turbulence Factor. K_2

10% blockage = 1, 5% blockage = 0.3

4) Determine the confinement factor (K_3).

Typical Plant	=	1
Pipe Rack with closure overhead + 1 other side	=	2
Compressor house / Analysed house	=	4

Table E 5.2 Overpressure Enhancements for Design Features

Multiply confinements to be taken as 1.25 times the highest value from above.

5) Peak Pressure = P (Peak) = $K_1 \times K_2 \times K_3 = P$ (Peak) bar. For most cases the $K_1 K_2 K_3$ factor will be 1 (E.5.4)

6) Determine the **volume** of the flammable cloud to establish the scale factor.

- **Case 1:** Release unknown. Take the largest vessel capacity – M kg – volume = $15M \text{ m}^3$
- **Case 2:** Release rate known – take the greatest. Release rate over one minute = Q kg – volume = $15Q \text{ m}^3$

7) Determine the **radius** of the **equivalent hemispherical cloud**. $V = 2/3\pi R^3$

(Note this is once again the inverse cube root to be found in the TNTe scale law equation E 5.2.)

$$CASE\ 1 - R = 3\sqrt{\frac{M \times 45}{2\pi}} \quad (E\ 5.5)$$

$$CASE\ 2 - R = 3\sqrt{\frac{Q \times 45}{2\pi}} \quad (E\ 5.6)$$

8) Draw circles radius of R from the plant centre.

9) Assess the volume of each cell within the circle.

10) Calculate the volume of the cells in (9). Volume = length x breadth x height (within areas of turbulence generators) – including pipe tracks as appropriate.

11) Calculate the scale distance “L” for each cell in (9)

$$= L = \left(\frac{3\ vol}{2\ \pi}\right)^{\frac{1}{3}} \quad (E\ 5.7)$$

12) Assess the highest pressure for all sources at distance x metres from edge of the plant and plot isobars. This assumes that the pressure decay from the edge of the plant is proportional to 1/distance

$$P_x = \frac{P_{\max} \times L}{(L + x)} \quad (\text{E 5.8})$$

13) Repeat 6 to 12 with the hemisphere displaced up to 75 M or the radius, whichever is the lesser to assess the maximum damage potential.

Blast Effects (Humans)

Overpressure Effects on humans (kilo Pascals)

O/P (Pa)	Injury probability
7	0
7-21	10% injury
21-24	25% injury
34-48	70%injury
48 +	95% injury
300 +	Internal injury/disruption high probability of fatality

Table E 5.2 Injury profiles for Humans

The injury profiles at low overpressures are influenced by being blown off the feet and impacting on hard/sharp objects. The Military use slit trenches! The injury at higher overpressures is influenced by internal organ damage even in slit trenches.

Damage (Equipment/Houses)

For assessed overpressures it is now possible to assess damage from tables of results from known events. The table may be different to others that might be available. There is a relationship between damage and impulse (overpressure x time – bar.seconds) as well as damage and overpressure. This table uses only overpressure.

Damage	Overpressure kPa
<u>Domestic Houses</u>	
Glass Failure	0.7
	5%
	50%
	90%
Building uninhabitable	7
Severe damage to building	15
Total Destruction of Building	75
<u>Process Buildings</u>	
Serious Damage (major report necessary)	15
<u>Process Equipment</u>	
Instrument displaced (major replacements necessary)	5
<u>Piping Spring</u>	
Storage tank	20
Process piping or Pipe Track	40
<u>Serious Damage</u>	
Storage Tank (90% full)	20
	40
<u>System Displaced</u>	
Fired Heater	50
Major piece of Heavy Equipment	40
<u>System Fails</u>	
Fired Heater	60
Major piece of Heavy equipment	50
<u>Missile Flight</u>	
At pressure of 2ka	
<u>Total Destruction</u>	75

Table E 5.3 Suggested Damage Profiles

Multi Energy Method (3) MEM

The MEM is not part of these notes but is given for completeness.

The MEM is a relatively simple tool, which has elements of the Volume Method (2). It uses a simple graphical correlation with a scale distance similar to that in the Volume Method (2). It starts with the premise that the fuel energy concentration is fuel-dependent but in reality it is 3.5 MJ/m³ for a whole range of fuels with only minor deviations. It has a series of lines, which reflect the turbulence potential,

and the fuel reactivity. Recent extensions of MEM have included equation E 5.3 and more sophisticated tools for the assessment of the turbulence effects.

The MEM graph is shown in figure E 5.5.

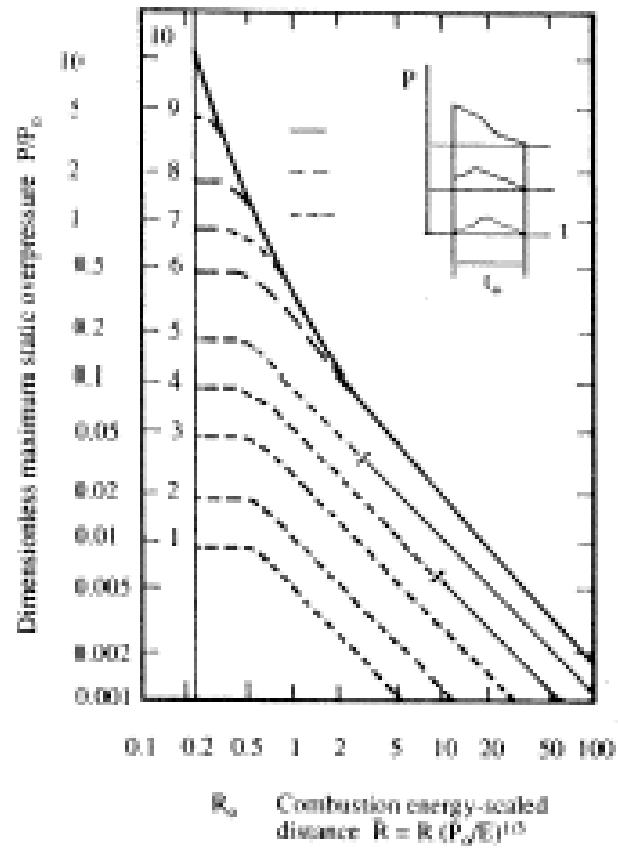


Figure E 5.5 MEM Overpressure v Scale Distance/Turbulence

Explosion mitigation:

Explosion prevention is the best mitigation. These include: -

- Reduce turbulence
- Increase venting and reduce enclosure
- Good dispersion
- Avoid leaks

However it is possible to slow and even arrest a flame front with: -

- Suppression – water mists

- Suppression – powder as with dry powder extinguishers
- (Halons were better but are now proscribed. They mopped up the free radicals which assisted the flame propagation)

E 6 Quantification (The Frequency or Probability of an Event)

The Need

There is a general acceptance that events of catastrophic proportion can and do happen and that improved design/technology can reduce both the frequency and consequence but it cannot be eliminated. It is essential that the balance is drawn between the magnitude and frequency in the form of criteria which have already been discussed but to do that it is essential to know how often an event might occur. This exercise (the assessment of the frequency) has a twin benefit; obviously the frequency has been assessed but also the weak links in the system (physical and numeric) will be identified and the solution may be fairly obvious from this analysis. It is also fair to note that the frequency assessment has the greatest bounds of error in any risk assessment.

Only one technique will be describe in detail, event trees, and the second, fault trees will be outlined. The outline on fault trees will describe what they are and how they are used, as these require experience in execution and it is easy to produce the wrong answer if not used properly. However it is important that there is an understanding of what fault trees are and the potential difficulties with their construction and quantification.

In addition notes on Reliability Theory are given as this can impact on Loss Prevention.

E 6.1 Event Outcome Trees

Event outcome trees are relatively simple to explain so will be introduced first.

The simplest technique for assessing the frequency of an event is the **Event Outcome Tree** (used in the right hand side of the Bow Tie (Fig E.1.1)). It starts with the frequency of an event which is then modified by a branched system which could have 2^n end points (where n is the number of events). Each branch represents a probability of success and failure, where (success + failure) = 1. Each probability (called **conditional probability**) is derived from tables, databases or is calculated as Fractional Dead Times (PFD). In reality the “**outcome**” of the initial event be assessed by “*engineering judgement*” or be based on the rigorous numeric assessment of the effects.

The best way of describing this is to give a word description of a particular event. Take an event such as a toxic release. It is possible to assess the likely spread of leak rates and frequencies knowing the leak frequencies for different sizes of breach, the system pressures. With the wind/weather spectrum it would be possible to assess the likely outcome at any set distance from that release for all combinations of size and weather distribution. The event outcome tree for **one set** of leak rates (kg/s) will be complex as shown in figure E 6.1. (which has only been started).

Hint: When constructing the tree attempt to arrange the logic such that all of the “successes” and “failures” are at the top or bottom by adjusting the logic. This is not always possible but it does help in the

analysis of the outcomes. A little planning before the construction can pay dividends in time, detail and ease of use. In many cases it is quite likely that the event tree can be reduced by a series of manipulations. This requires a little skill and a lot of care. The example of the throw of a head is an easy way to show the reduction process.

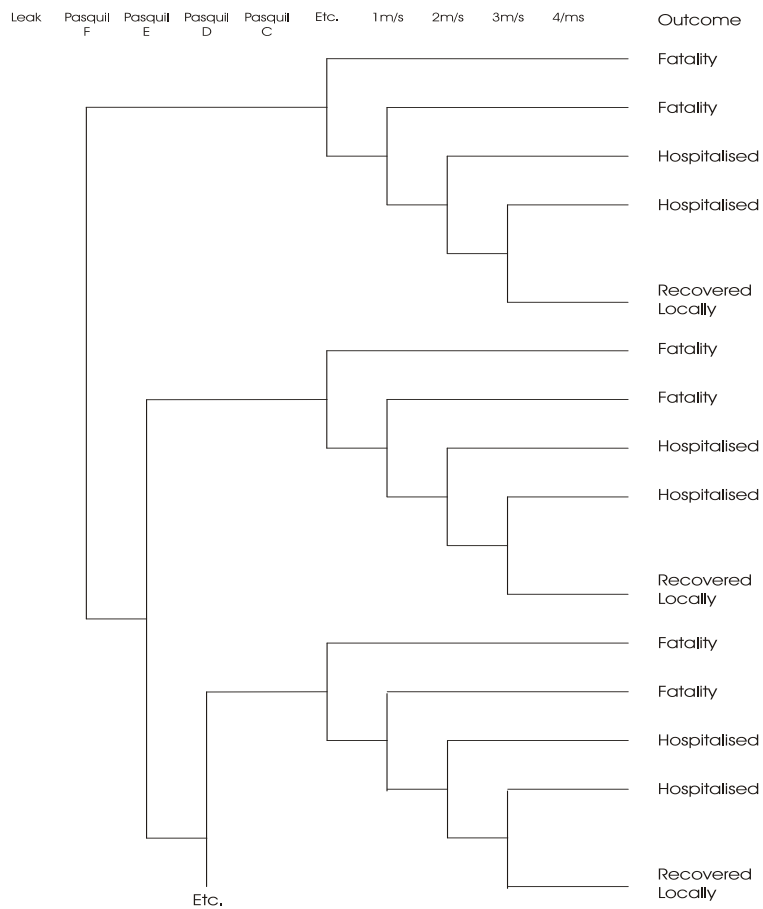


Figure E 6.1.1 Event Outcome Tree for a Toxic Release

It can be seen that this is an enormous tree so it might be simpler to draw it up as a series of trees for either weather stabilities or wind speeds where once again all of the combinations of wind and weather stability **must** be 1 and all leak profiles are described. This could result in 10 or more trees!

An even simpler example is the toss of a coin and the throw of a head. In this example there are only two outcomes, **head** = success and **tail** = non-success. But the outcomes can be predicted quickly as in figure E 6.1.2. In this case the success is 50% per throw. It is possible to take this one stage further. If a head was thrown at the first throw it was a success so the event outcome tree would have just two branches. However if the rules are that you have to throw at least one head some of the outcomes may be redundant and the tree can be tidied up. Figure E 6.1.2 represents the full tree after 4 thrown and figure E 6.1.3 represents the “purged or reduced tree”. It will be noted that the probability of failure to throw a head after n throws is 0.5^n , that is, it is necessary to throw an infinite number of times to be absolutely certain that there will be a head! Think about the event outcome trees when throwing dies in gambling tables!!

dispersion, fire, and explosion. Delayed ignition could create an unsafe dispersion resulting in a toxic gas cloud followed by fire or explosion. The probability of each outcome will be the product of the probabilities of the events (**the Conditional Probabilities**) leading to this outcome. The differences in the various outcomes may require a little judgement but the calculations can be done quite readily. The only difficulty is should there be an overlap of toxic effects prior to injury from an explosion or fire.

This explains why each success (or failure) probability must be **rigorously** justified on **each** and **every** case using references or data.

Now consider a simpler event, the leak of a flammable. Once again from the system pressures and the leak (breach) spectrum frequencies it is possible to assess the likely outcome for any likely duration of leak. The probabilities (**the conditional probabilities**) would now include:

- Immediate ignition (therefore no explosion)
- Delayed ignition (leading to the possibility of explosion)
- Successful operation of the shut down and depressurising system
- Successful operation of the fire protection system (systems)
- Person being present at the time of the event
- Person evacuating from the area

This may require **two** event outcome trees one the Process Plant and one for the Operator. Figure E 6.1.4 represents the Process Plant Event Tree. Once again the probability of ignition, (immediate + delayed + none) = 1.

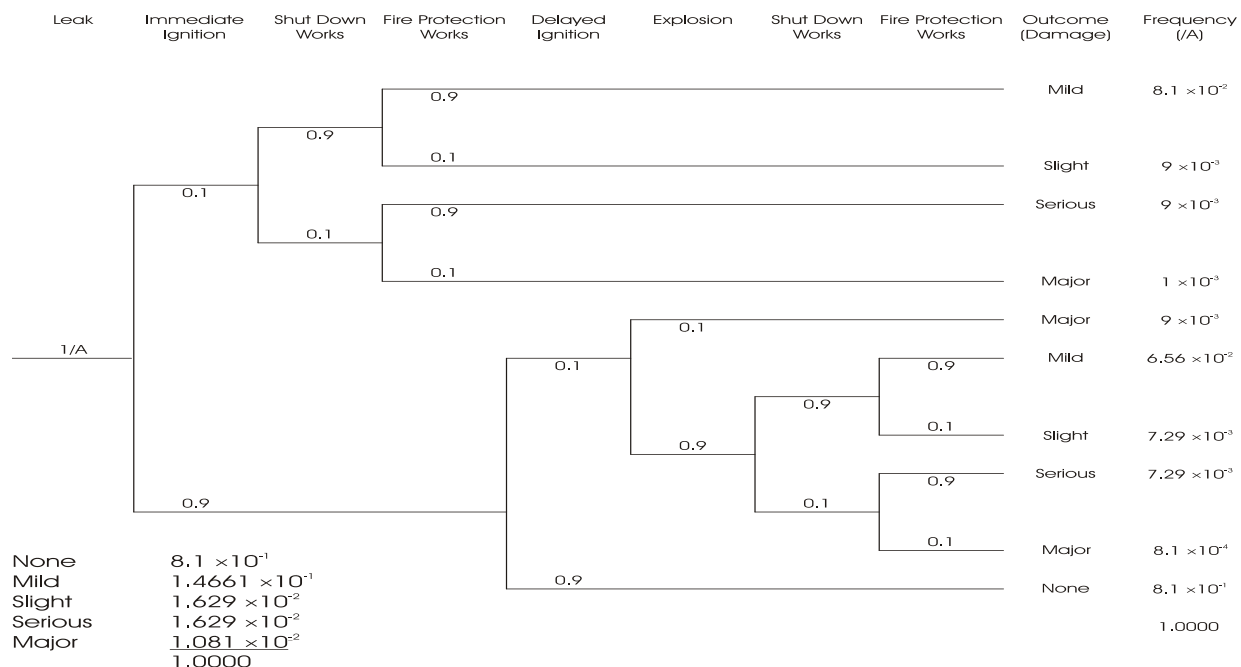


Figure E 6.1.4 Event Outcome Tree for a Flammable Leak and a Process Plant

For the ease of calculation I have taken the leak rate as 1 per annum, the prompt ignition as 0.1 per demand, the delayed ignition as 0.1, if delayed, and the probability of successful operation of the shut down and fire protection systems as 0.9 per occasion. This assists the calculations and **should not** be treated as indicative of real values.

Figure E 6.1.4 results as follows:

All events = 1 /A

No damage = 8.1×10^{-1} /A (fire only)

Mild damage = 1.4661×10^{-1} /A (fire only)

Slight damage = 1.629×10^{-2} /A (fire only)

Serious damage = 1.629×10^{-2} /A (fire only)

Major damage = 1.081×10^{-2} /A (fire plus explosion)

It will also be note that the more defences in place the lower the final frequency of the major event, this proves the benefits of "**defence in depth**".

E 6.2 Fault Trees

Fault trees are an essential part of risk assessment, they are difficult to generate and simplify.

Knowledge of their existence and the difficulties with their use is an essential part of understanding but the ability to use them is not.

The event outcome tree is perfectly acceptable for analysis if all the events leading to an outcome are clearly understood. In general the analysis is more of a "macro size". If the combinations of the events and - worse still - the events themselves are not fully understood it is necessary to develop a fault tree, which then analyses the events in more detail. The fault tree uses logic such as "AND" or "OR" (and sometimes "NEITHER" or "NOR" but these can be converted to AND/OR logic).

Conversion of Units

Frequency data can always be converted into probabilities by the concept of **Fractional Dead Time (FDT or PFD)**. However it is less easy to convert probability data to frequencies without knowledge of the data source, so, a manipulation of the fault tree may be necessary.

Gates

The "**gates**" are the point where a number of items or operations come together for a single operation. These are the "**and**" "**or**" logic. The dimensions that operate in each gate have to be analysed carefully and follow the correct rules.



Figure E 6.2.1 The symbols for “and” “or” logic

The simplest fault tree can be found in Fig 6.2.2. Yet nearly every plant has a small leak somewhere but the leak does not ignites very often so the fault tree is not quite correct and the ignition path requires development, as shown in Figs E 6.2.3 and E 6.2.4.

Clearly, with the exception of pumps there is only a small chance of an ignition source being near enough to ignite a small leak and experience shows that small leaks very rarely ignite.

The Fault Trees have so far demonstrated AND logic - that is events have to occur together. But OR logic requires that either case will satisfy the event, so fuel could be caused by a leaking gland or leaking seal or corrosion.

Combination of Logic in Fault Trees

The combination of logic within fault trees (which will then require numerical evaluation) is of absolute importance. The logic must be correct not only in flow but in dimensions. Data can come in two forms FREQUENCY - 'f' or PROBABILITY - 'p'. **Frequency** has units of "*per unit time*" and **Probability** is a "*number*" lying between 0 and 1.

Development of a Simple Fault Tree

To have a "Fire" what is required?

Fuel	Yes
Oxygen	Yes, but in the correct proportions with the fluid
Ignition	Yes, but at sufficient energy

"Fire" results from "Fuel" + "Oxygen" + "Ignition" remove any element and there is no fire.

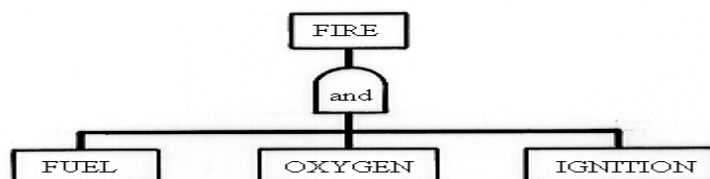


Fig E 6.2.2 Simple Fault Tree for a Fire

It is now possible to develop each step one level more:

Fuel:	Leaks from glands/flanges/corrosion
Oxygen:	Readily available at 20.8% v/v with air
Ignition:	Faulty electric, damaged bearings (etc)

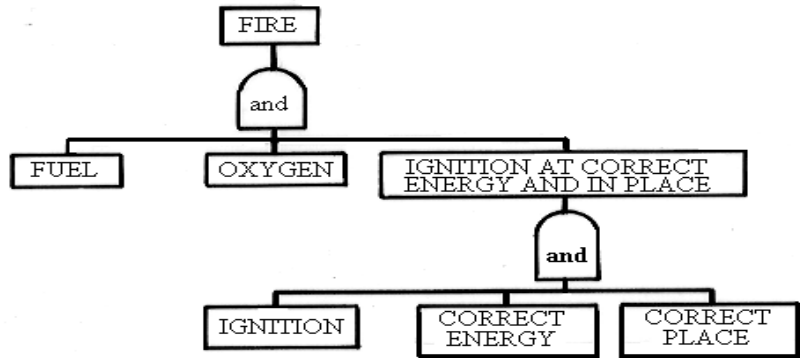


Fig E 6.2.3 Expanded Fault Tree for a Fire

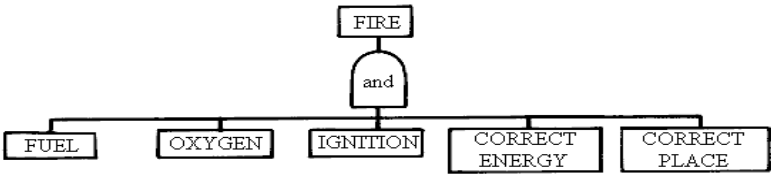


Figure E 6.2.4 Final simple Fault Tree for a Fire

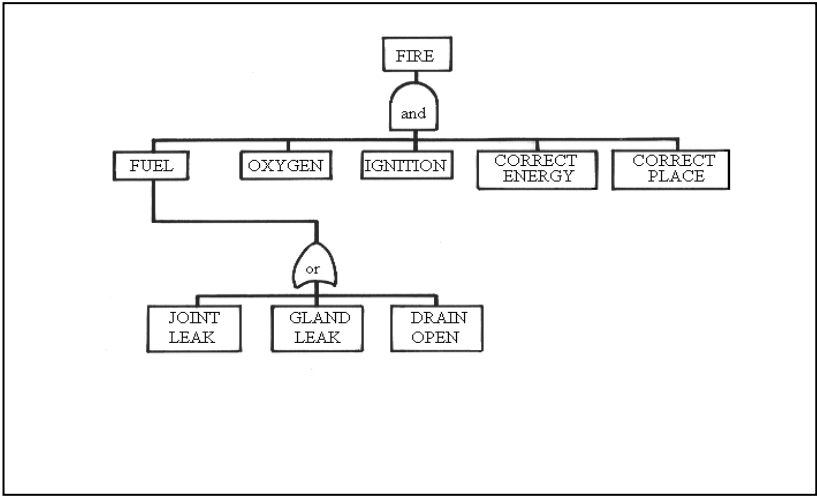


Figure E 6.2.5 Fire Fault tree with and or logic

As it is necessary to **ADD** units of similar nature, **OR** logic must only have **ONE** dimension. However, in AND logic it is necessary to **MULTIPLY** units of different natures, therefore AND logics must **not** contain more than one frequency - (whoever heard of Failures² per Year²?). In the same manner there can be no frequencies if in Fig E 6.2.2 should the answer be the "PROBABILITY OF A FIRE"! On the other hand if the answer is to be in units of "Frequency", put failure in frequencies into **one gate (box) and one gate (box) only** in Fig E 6.2.4 and the rest into probabilities. Fig E 6.2.5 shows **OR** logic coming into the fuel gate, this is addition and the rest, oxygen, ignition etc are now probabilities.

Minimum Cut Sets

The development of a fault tree may produce a slightly anomalous solution if the logic is not cleared of all irrelevant data. In logic terms this is a fairly obvious statement but purging and rearranging the fault tree so that it is correct may require a lot of care, time and effort and it is not unusual to draw a fault tree three or four times before it is "correct".

Consider the single pump circuit in Fig E 6.2.6 and the simple fault tree shown in Fig E 6.2.7. In logic terms this **appears** to be correct. However, it will be noted that "TANK EMPTY" appears at items 1 and 6 so there is the risk of double counting. The logic can be rearranged as shown in Fig E 6.2.8 with all data in terms of probability.

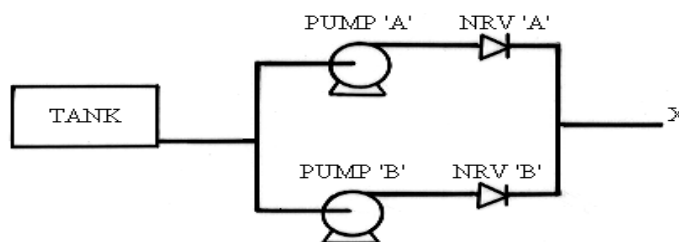


Fig E 6.2.6 Simple Pump Set

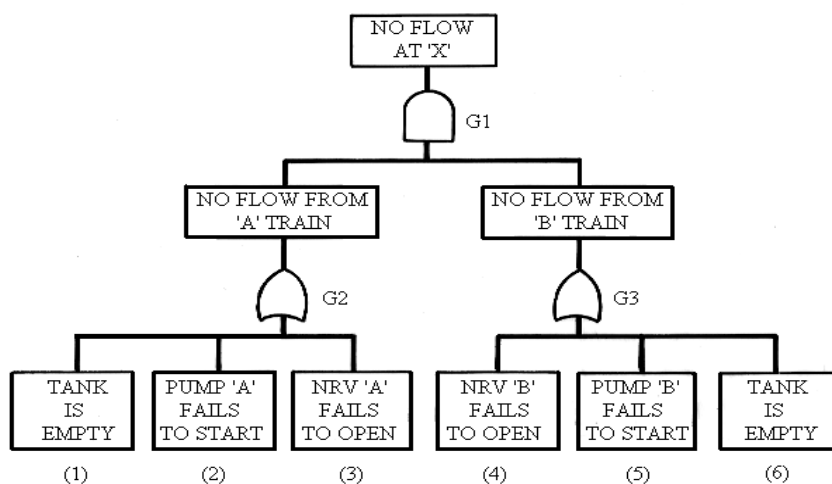


Fig E 6.2.7 Over Simple Fault Tree for Fig E 6 2.6

Gate 1: Gate 2 x Gate 3 - (AND logic)

Gate 2: 1 + 2 + 3 - (OR logic)

Gate 3: 4 + 5 + 6 - (OR logic)

Gate 1: 1.4

1.5

1.6

2.4

2.5

2.6

3.4

3.5

3.6

Or Is It?

1.6 is the same event as 1.1 and if the tank is empty it is irrelevant to whether the pump fails to start!

No fluids = No flow

1.1 AND 1.6 are therefore 1 by Boolean logic (thank you Reverend Bool).

1.4 and 1.5, as well as 2.6 and 3.6, are therefore redundant and the count becomes 1, 2.3, 2.5, 3.4, and 3.5 which is exactly the logic which would be derived from Fig E 6.2.8. The same tree has been drawn in frequency terms in Fig E 6.2.9.

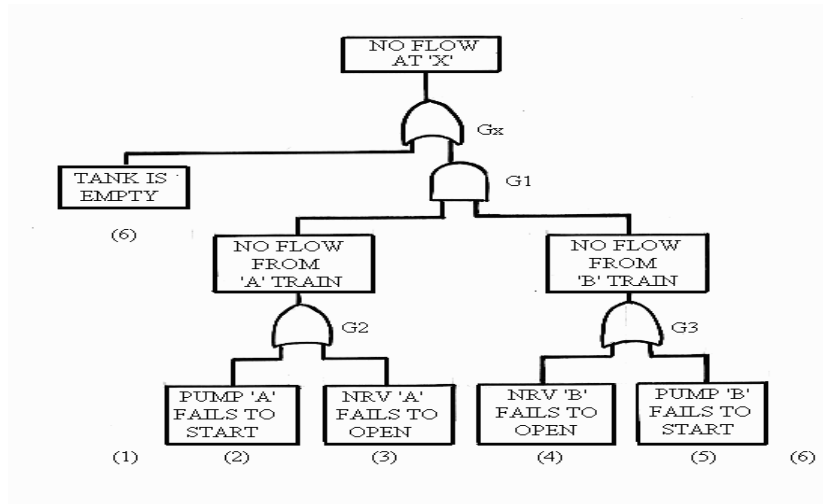


Fig E 6.2.8 Correct Fault Tree for Figure 6.2.6

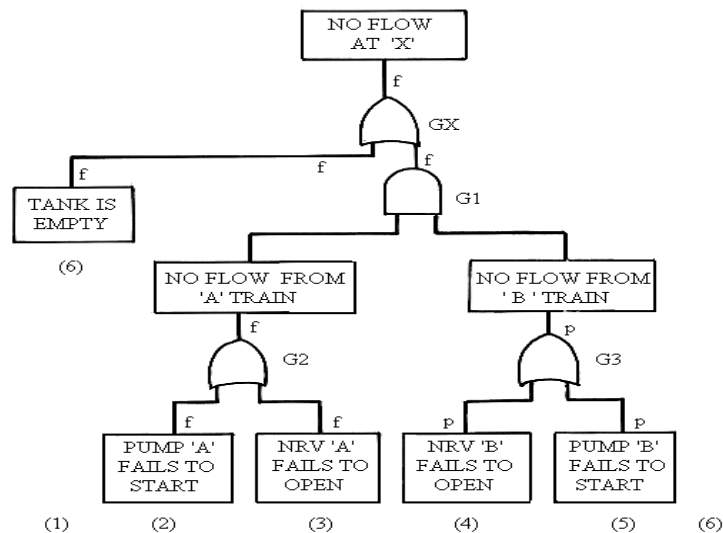


Fig E 6.2.9 Fault Tree for Figures E 6.2.6 and E 6.2.8 with the Answers given as a Frequency (f) or probability (p)

Common Mode Effects

The example of the pump introduces the "*common mode effect*", so at some time a refinement has to be added to fault trees. Common modes, as their name suggests, are the unique events which causes all equipment to fail simultaneously. One common mode effect is clearly the empty tank and it has been taken out separately. There is **at least** one other common mode - electrical power failures. This can be drawn out in the logic tree into the "OR" gate under no flow at Gate X, Fig E 6.2.8 and E 6.2.9. What other common modes can be found? (Hint instrument failure, human error are but two).

In general, common mode effects only have importance when the probabilities are very small; in this case it is the common mode effects which are likely to dominate the answer.

One way of avoiding the common mode effects in the example of the pumps is providing "redundancy" in the power supply. However, even so the switchgear may have a common mode fault such that if a certain condition occurred both items would be put out of operation simultaneously. This could be a fabrication defect, a design defect or simply a fire.

The following are just some of the possible causes of **common mode failures**:

- cable routing,
- design features,
- installation feature,
- maintenance faults,
- operator errors (failure to open or close valves),
- failure of an operator to react to a situation (if he fails to react to the first event he will certainly fail to react to the second).

Human Impact on Fault Trees

Not only are people one of the direct "common mode effects" but also people are sometimes required to complete a control loop. In the example of the pump "*Failure of pump A to start*" does not define how it was to be started. Very often humans are the first level **Protective System** and they must respond to some visual or audible alarm and then take some actions. As fault trees are developed to the final details it is almost certain the human element will appear.

This brief outline is an attempt to show that fault trees are not easy to construct and that it is easy to make mistakes! They are not for the faint hearted!

E 6.3 Reliability Formulae/Protective Systems

Introduction and Background

It may seem odd at first to include 'Reliability' in a Safety and Loss Prevention course, it is, however, quite logical as the reliability of systems do have an impact on Loss Prevention. This can be illustrated very easily. Which is the better arrangement 1 off 150% Fire Water Pump, 2 off 100% pumps or 3 off 50% pumps? There is a big difference in the availability and costs, and the answer is 2 off 100% pumps!

Equipment does not break down on a fixed routine and there is no fixed repair time, failure is usually taken to be a random event and repair time a log normal distribution. Reliability theory used for assessing the performance of process systems can therefore be very complex but it is necessary to challenge the absolute accuracy of the theory in the light of the relatively inaccurate data. This does **not** infer that short cuts should be taken but more that common sense should be used and that the simple formulae may be more appropriate to the crude data available and also allow an order of magnitude result to be arrived at quickly.

Often errors or the accumulation of errors in a calculation do make engineers suspicious of the answer and fear that the answer could be a long way from the truth. To a degree this is true. The potential error is quite large but the probability of the worst-case combination occurring is relatively small. As probabilities are not points but are probability density functions, the uncertainty tends to cancel out in multiplication (AND gates) but accumulate in addition (OR gates). This means that the evaluated probability or frequency of an event occurring in practice is likely to be fairly near to the truth even though the data is subject to a fairly high factor of uncertainty.

Reliability Theory

Reliability can be used to assess the performance of many systems. None less than the availability of the plant or a safety system such as the firewater system.

Calculation of Probabilities

The simplest way of remembering how combinations may occur is to complete a "**TRUTH TABLE**". A very simple example is the toss of an unbiased coin twice. 1/4 of the time there will be two heads, 1/4 there will be two tails and 1/2 there will be 1 head and 1 tail. By the same token it is possible to do the same thing for 3 pumps. If the success probability is S and the failure probability is F ($F = 1 - S$) the truth table is as follows in figure E 6.3.1: -

PUMP			PROBABILITY
A	B	C	
S	S	S	S^3
S	S	F	S^2F
S	F	S	S^2F
S	F	F	SF^2
F	S	S	S^2F
F	F	S	SF^2
F	S	F	SF^2
F	F	F	F^3

Table E 6.3.1 Truth Table for 3 Pump Units

This condenses to the following quadratic equation:

$$S^3 + 3S^2F + 3SF^2 + F^3 = (S + F)^3 \quad (\text{E 6.3.1})$$

As $F = 1-S$ It can be seen that the sum of all states is 1.

Availability

If a unit is not capable of operating (failed F/year) and is under maintenance lasting B hours it is **unavailable** (FB) hours per year. Now divide by 8760 hours per year and the unavailability in probability terms is as below:

$$F B/8760. \quad (E 6.3.2)$$

First start off with a description of the process:

If there is a pump (or any other unit) it can be in one of two states - **AVAILABLE** (S) or **UNAVAILABLE** (F). If the probabilities of each state are S and F it is obvious that

$$S + F = 1 \quad (E 6.3.3)$$

There are many combinations of group arrangements for example 1 out of 1 (1 out of 2, 2 out of 3 and 1 out of 3 etc). This is given as the binomial expansion:

$$(S+F)^n \quad (E 6.3.4)$$

where n is the number of installed pumps

N	STATES								
1	S	+	F						
2	S^2	+	$2SF$	+	F^2				
3	S^3	+	$3S^2F$	+	$3SF^2$	+	F^3		
4	S^4	+	$4S^3F$	+	$6S^2F^2$	+	$4SF^3$	+	F^4
Etc									

Table 6.3.2 The Combinations of States for N Units

Note for the 3 units this is the truth table E 6.3.1

Etc. The numbers come from Pascal's Triangle where each number is derived from the addition of the pair above; The next sequence is 1 : 5 : 10 : 10 : 5 : 1

- a) S, S^2, S^3, S^4 represent the probability that **ALL** units are available
- b) $2SF, 3S^2F, 4S^3F$ represents the probability of 1 unit being unavailable (and the other/others available)
- c) F, F^2, F^3, F^4 represent the productivity of **ALL** units being unavailable.

So to illustrate this and to answer the question posed earlier if there are 2 units and only one is needed to satisfy the demand. The availability = $S^2 + 2SF$ (which is $1-F^2$). For 3 units and needing 2 on line. = $S^3 + 3S^2F$ (or $1-(3SF^2 + F^3)$)

It will be noted that the unavailability of a 2 o o 3 system tends to three times that of a 1 o o 2 system.

It is easier to illustrate this with numbers, if pump is available 90% of the time

$$S = 0.9 \quad \text{and} \quad F = 0.1$$

$$1 \text{ out of } 2 = S^2 + 2 S F = 0.81 + 2 \times .1 \times .9$$

$$1 \text{ out of } 2 = 0.81 + 0.18 = 0.99$$

Or: -
$$= (1 - F^2)$$

$$1 \text{ out of } 3 = S^3 + 3S^2 F + 3SF^2 = (1 - 0.001) = 0.999$$

$$2 \text{ out of } 3 = S^3 + 3S^2 F = 0.729 + 0.243 = 0.972$$

The answer the question posed in the first paragraph is shown in the following table:

	Availability = Success	Unavailability = Failure
1 x 150%	0.9	0.1
1.o.o.2 x 100%	0.99	0.01
2.o.o.3	0.972	0.028

Table E 6.3.2

Note: success plus failure will always equal 1

Now link the units together. 1 out of 3 pumps (1 o o 3) and 2 out of 3 (2 o o 3) heat exchangers are **AND** logic in terms of availability so are multiplied.

If A = 0.9 and U = 0.1 - only as it helps the computation - the availability of each item can be read from above.

$$\text{Availability of a 1 o o 3 pump} = 0.999$$

$$\text{Availability of a 1 o o 2 heat exchanger} = 0.99$$

$$\text{Availability of Both (the product)} = 0.98901$$

As can be seen it is not difficult to erode the availability. The event tree can also be used to combine other conditions such as the probability that a pump is unavailable.

As the numbers coming out of single items have a number of "9s" it is important that there is no attempt to "round-off" until there is the final answer.

Take two fire water pumps Availability = 0.9 - and start on demand = 0.01

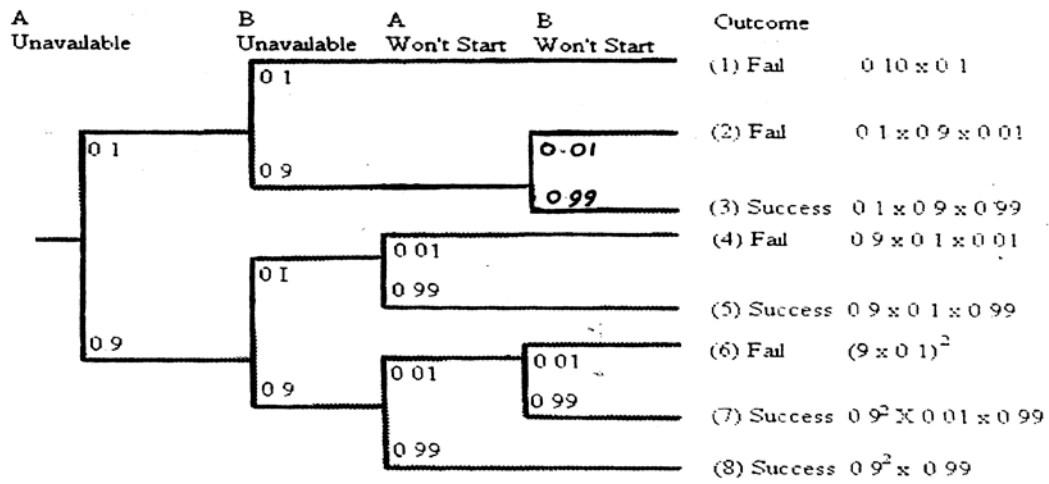


Fig E 6.3.2 Event Tree for Fire Pumps

The success items are 3, 5, 7, and 8, and the failure items are 1, 2, 4, and 6.

1	=	0.1^2	=	0.01
2	=	$0.1 \times 0.9 \times 0.01$	=	0.0009
4	=	$0.1 \times 0.9 \times 0.01$	=	0.0009
6	=	$(0.9 \times .01)^2$	=	0.000081
	=		=	0.011881

Table E 6.3.3 Failure rates from Figure E 6.3.2

Coincidentally this can be derived from logic

State

A	2 pumps unavailable	=	U^2
B	1 pump unavailable the other available but will not start	=	AU, B available but won't start
C	As B but reverse pairing	=	BU, A available but won't start
D	Both available, non starters	=	$(A \text{ Non Start})^2$

This can be drawn as a fault tree.

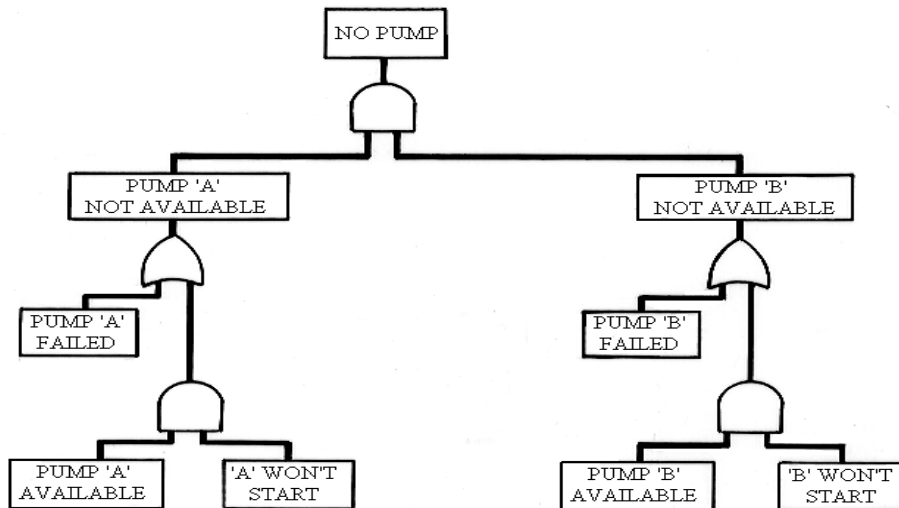


Fig E 6.3.3 Fault Tree for Fire Pump

(Compare the Event Tree E 6.3.2)

Reliability block diagrams

Reliability Block Diagrams (RBDs) are a convenient way of displaying the configuration of the process. Each piece of equipment is described as a single block and where there is more than one piece of equipment the blocks are linked as shown in figure E 6.3.4 below. No attempt is made in the RBD to differentiate between a 2 out of 2 or a 1 out of 2 configurations, this is done by the logic. The overall availability of each set of blocks must be carried out using the binomial expansion and the data given. This in turn can be added to the RBD so that the analogue and data are stored in one document. The RBD can be drawn vertically or horizontally. For LARGE RBDs the vertical configuration may be more appropriate. Eventually the overall availability of the system will be the AND logic which requires multiplication of the individual availabilities. As these may well be 0.999 or less it is essential that the data is stored in a calculator OR the data is recorded manually to at least 6 decimals. If this is not done the systematic rounding will produce an erroneous answer.

In reality there may be outage for “function” or “trip” testing safety systems, this is unlikely to be significant as it can usually be done “on-line” or spare equipment can be fitted. In addition for a large continuous process there will be some major maintenance carried out according to the regulations. This may well involve a month of outage every 3 or thereabouts years. This will affect the overall production of the process. If for example an annual through put of 100 units is required the “peak” design throughput may well be 105 units (or more) to allow for the intermittent outage during maintenance.

Take a simple process of

1 Vessel	Outage 8 hours per 2 years
1 o o 2 Heat Exchangers	Outage 8 hours per year
2 o o 3 Pumps	Outage 4 hours twice per year

The RBD will look as follows:

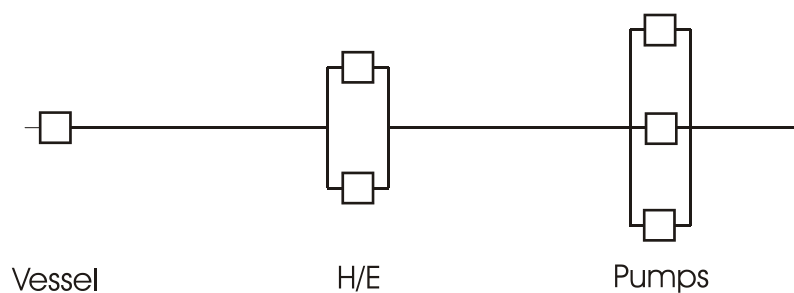


Figure E 6.3.4 The Reliability Block Diagram for the Simple Process

The availabilities of each subset of equipment using the binomial expressions are as follows:

0.999543379 0.99999916 0.9999974996

The values above give the individual reliabilities of the three systems shown in the RBD above.

The OVERALL availability is therefore the product or 0.9995400461. This is dominated by the “unspared” vessel, as would be expected.

E 7 Shutdown Systems (Repeat of Part D)

As discussed in Part D there are three main elements in the shutdown system

- The detector or switch
- A means of converting the signal into a means of shutting an emergency shutdown valve.
- The shutdown valve itself

This section now expands on the “non availability” of the system.

The detector may be a pressure switch, which operates at a present pressure, a level switch that operates at a fixed level or temperature switch, which operates at a present temperature. The common feature of all shutdown systems is that they fail-safe. This means that the interruption of the power or any signal will put the system into the safe condition. This usually means that the system will initiate a shut down. The design of these devices varies between designers and in some cases they are standard control measurements, which are triggered at set points as an on/off signal. The output signal is often electrical and is used to hold a solenoid valve open – loss of power causes the solenoid valve to change its position (fails safe) and interrupt the air (or hydraulic) supply to the Emergency Shutdown valve (ESDV). The ESDV is held open by the air (or hydraulic) signal and is driven closed by a spring: -

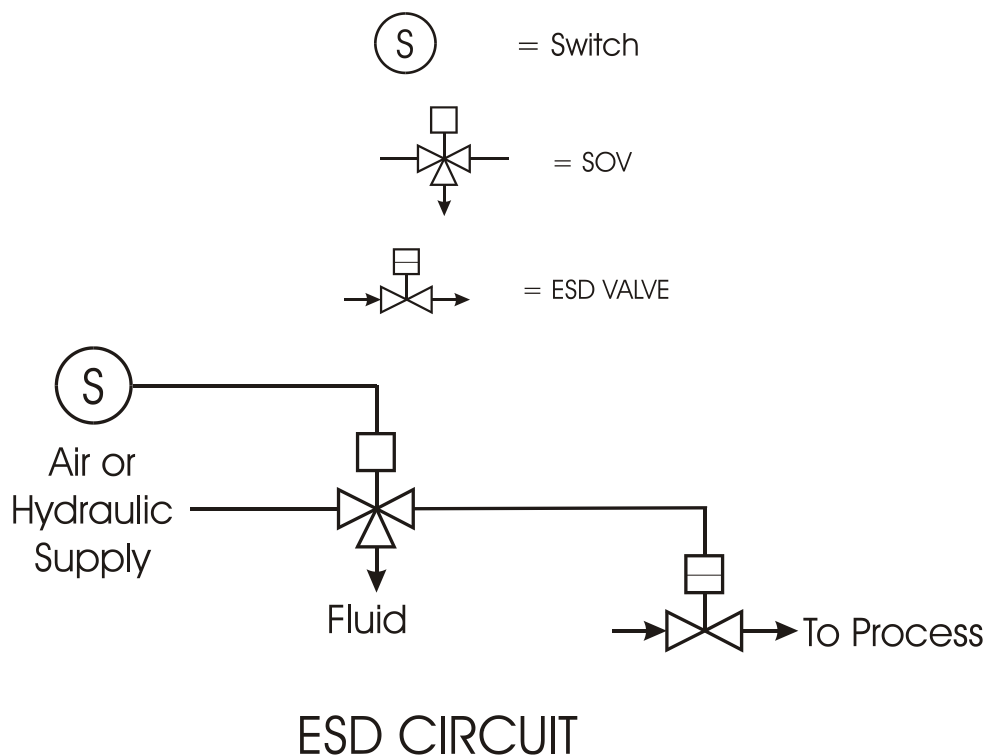


Figure E 7.1 A Simple Shutdown System (Simplex)

The arrow on the ESDV shows it shuts on loss of signal.

In some cases the valve may be held open by a hydraulic supply (instead of air).

As the test must be real and all elements proved to work – including the ESDV, there must be a test facility, which allows all elements to function properly without the plant being shutdown. This is usually achieved by installing a device, which prevents total closure of the ESDV (or plant shutdown). During testing, the shutdown system has to be inhibited leading to **TRIP TEST DEAD TIME**. The design of the test facilities and the test programme requires detailed analysis and obviously consideration has to be given to means of overriding the test facilities, should a genuine plant upset occur during the testing (TRIP TESTING).

As already discussed, sometimes the shutdown has to be bypassed to facilitate the start up of the process. This creates potential hazards if the bypass is left in place. The design can incorporate automatic resets of the shutdown or key controlled bypasses, controlled by rigorous procedures, which can only be operated by senior personnel. In some cases the control may be only be rigorous procedures operated by senior personnel.

In some shutdown systems it may not be acceptable to override the trip for testing purposes. Therefore a fully redundant trip system is installed. Each sensor and valve can be tested on routine with no interruptions to the process.

In more sophisticated systems a failure of the sensor or valve may cause a process upset so new strategy is adopted – “2 out of 3”. Three sensors are fitted and fed into a logic system, which vote any 2 out 3 to cause shutdown. Failure of part of a shutdown system will reduce the system to “1 out of 2”.

The circuit looks as follows:

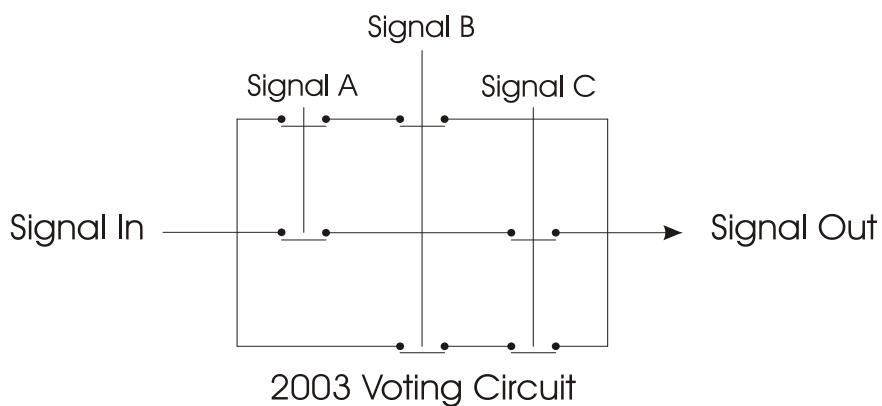


Figure E 7.2 A Simple Two out of Three Voting Circuit

Any 2 sensors operating will cause a shutdown; one sensor operating spuriously will not cause a shutdown and so can be tested on line.

The shutdown valves can now be lined in parallel such that one valve can be closed and tested at any time without causing a full shutdown.

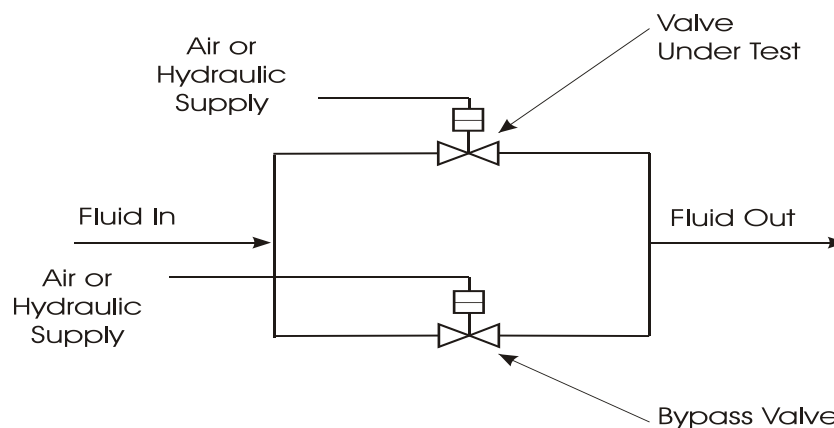


Figure E 7.3 Shut Down Valve with Test By-pass with DMH

The by-pass valve would be controlled by a “*dead man’s handle*” which if released would initiate closure of that valve.

Ultimately, 6 sensors could be used, 3 to close valve A and 3 to close valve B – this is a fully redundant shutdown – typical of nuclear power stations. The whole system can be fully tested without any Trip Test Dead Time.

THE DESIGN AND TESTING OF SHUTDOWN SYSTEMS IS AN ART/SKILL.

Comparison of Protective Systems

Not all failure fail-safe. Some fail “spurious”, that is, they fail in such a way that they initiate a shutdown. Not all protective systems are simplex, some are redundant. The fractional dead time for the system alone then becomes as follows (S = spurious and F = danger): -

System	Fail Safe Fault Rate Faults/Year	Fail to Danger Fault Rate Faults/Years	Fractional Dead Time
1 out of 1	S	F	$\frac{1}{2}FT$
1 out of 2	2S	F^2T	$\frac{1}{3}F^2T^2$
2 out of 2	$2S^2T$	2F	FT
1 out of 3	3S	F^3T^2	$\frac{2}{3}F^3T^3$
2 out of 3	$3S^2T$	$3F^2T$	F^2T^2

Table E 7.1 Fail Safe/Danger Rates for Redundant and non Redundant Protective Systems

However, the typical test dead time for a 2 out of 3 system can tend to zero, as on-line testing is possible. The human element still remains.

No Common Mode Allowance

The common mode (as already discussed earlier in this Chapter) is that element of a trip system which is dependent of itself and not time-dependant. Instruments are vulnerable to a potential common mode such as a fire or explosion, but also multiple shut down valves with a spring close action - are likely to have common mode failures with the spring or the release mechanism. Consider also pressure trappings - common modes could be wax, dirt or ice.

As a result the limiting FDT is as follows: -

1) 1 of 1 = 0.05

2) 1 of 2 = 0.005 - 0.001

3) 2 of 3 = 0.001 to 0.0005

Note: A 1 out of 2 system is almost as good as a 2 out of 3 system - but you avoid spurious trips.

E 8 Vulnerability, Toxics Doses and Effects Models (see part G for more advanced information)

Introduction

It is fairly obvious that the simple calculation of concentrations of gases or thermal radiation or overpressures do not tell the whole story. The next step is the "Effect" on the receiving body, be it mechanical or human. In most cases it will be found that the human is the limiting factor. These effects have been collected together but can be found above but under different headings.

The effect of 10^4 ppm carbon monoxide on a piece of process equipment is negligible but for humans it is different. What happens if it lasts for 1 second? 100 seconds? 1000 seconds? The consequence models will show what is the instantaneous effect but some interpretation is necessary when "total exposure" is the problem. The analysis must therefore consider: -

- How long the incident will last?
- What is the effect of **that** incident, for **that** time?
- The effect models are not absolute values but an analysis of historic data. Further individual responses will vary so that effect on one person of different ages or state of health will be different from another. The final point that must be stated is that all effects, which may affect health of life, are corporate decisions and must be agreed at a corporate level.

E 8.1 The Human

Physical Protection

See Personal Protective Equipment Regs

No one would design "the human" in the same way as evolution. It is poorly designed and very vulnerable to many potential routes for assault. The human cannot perform in heat, cold, reduced oxygen concentration, acceleration dull or bright light and loud noise. Consider the following:

Area	<u>Some Sources of Human Vulnerability</u>
Head (1)	Impact; loose hair being caught in moving equipment
Eyes (2)	Light; – low or high; Ultra Violet Light; Grit; Acid; Alkali; Dust; Projectiles
Ears (3)	Noise – 120 dBA equals pain, 90 dBA for 8 hours equals hearing loss
Nose/lungs (4)	"Toxics"; dust; low oxygen concentration; Hot, dry air; nuclear radiation
Skin (5)	Acid; alkali; nuclear radiation; heat; cold; thermal radiation; projectiles;

	sharp objects; trips/falls
Bone (6)	Heavy objects; trips/falls
Brain (7)	Information overload /cognitive dissonance/mind set/stress/confusion/lack of training/panic
Balance (8)	Wind of 50 m/sec; acceleration of 0.1g laterally
Muscle and Tendon Damage (9)	Poor Ergonomics: stretching; lifting; trying to exert forces by load oblique paths etc. (The list is legion.)

Table E 8.1.1 Areas of Human Vulnerability

The list is not complete and is sufficiently detailed to show that the human needs to be protected with care.

The industrial safety helmet will give reasonable protection against a light impact or light dropped object. (Say 1 kg dropped 10 m.) It will not necessarily protect the wearer against walking into a low beam. (The author can vouch for this!) Nor will it protect against a sharp edged or pointed dropped object.

The industrial safety spectacles and full wrap round goggles give good eye protection against dusts but it **may** be necessary to use a full-face visor for hazardous fluids. "Green Glasses" can protect the eyes against intense light such as experienced in a furnace.

At noise levels over 100dBA communication is difficult and the threshold for noise induced hearing loss is sometimes quoted as low as 85dBA for 8 hours. Hearing loss is accumulative and initially speech appears to be distorted, as the high frequency elements are lost, ultimately there is a problem of "tinnitus" (hissing or ringing sounds in the head). A "disco" is a powerful source of damage. If 90dBA is taken as the threshold, 93dBA for 4 hours is equal to 90dBA for 8 hours, likewise of 96dBA for 2 hours or 99dBA for 1 hour or 102 dBA for ½ hour. Ear plugs give some protection but noise can still be transmitted through the human tissue. Ear muffs are far better protection.

Face masks (filters) are available for dusts but they are of no use for harmful gases. The use of **Breathing Air (BA)** masks is necessary. (It should be noted that the "Gas Masks" used in WW2 were of the activated charcoal absorbent type.) BA may be by a demand valve when the pressure in the mask falls below atmospheric pressure (as a SCUBA air mask so leakage can take place around the face mask seals) or it may be by a supply valve, which keeps the face mask slightly over pressures with respect to the atmosphere.

Gloves come in various forms. Standard gloves will protect against cold and also can have "rubber studs" to enhance grip. "Leather" gloves can protect against "sharps and rags". However special gloves, mitts or gauntlets will be needed for hot, cold or harmful duties. **(Remember to tuck clothing such that the spills shed away from the tucks and not into the gloves)**

Steel tipped safety boots, like helmets, can give protection against light dropped objects. (Again about 1 kg dropped from about 10 m.) Special boots may be needed for harmful fluids. **(Remember to tuck clothing such that the spills shed away from the tucks and not into the boots)**

Industrial "cover-alls" can give good, general, protection against benign fluids but it may be necessary to use acid/alkali resistant clothing. .) In some areas it might be appropriate to wear flame retardant

protection. (Remember to tuck clothing and such that the spills shed away from the tucks and not into them)

Consider the impact of “confusion” and also the dangers of wind load on the body when working in stormy weather.

Dry air at 100°C will damage the mucous membranes in the lung; dusts or smoke can coat the lung (more miners die from silicosis than injury). 10% oxygen will result in unconsciousness.

The skin cannot tolerate temperatures less than -10°C or temperatures over 60°C for more than a few seconds. Thermal radiation of 6Kw/m² for 20 seconds produces *real* pain.

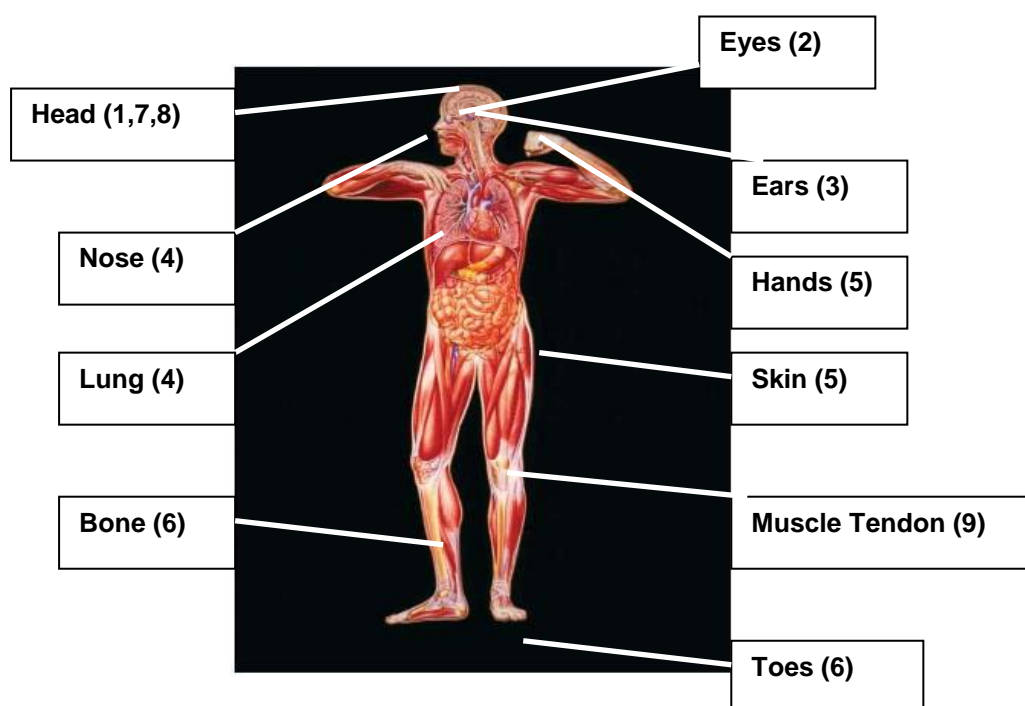


Figure E 8.1.1 The Areas of Human Vulnerability

Toxic Effects: Classes of toxic material Classed in relation to the effects of exposure.

Irritants	Respiratory (Chlorine), skin, eyes. Irritants affect the body tissue at the site of contact. Effects range from discomfort to death. Some make the victims more susceptible to infections such as pneumonia. Panic result from exposure to irritant gases and the response may be unpredictable reaction know as “ <i>fight or flight</i> ”.
Narcotics	Some common chemicals e.g. hydrocarbon vapours produce narcosis, which can interfere with an individual’s ability to look after him/herself.
Asphyxiates	Simple suffocation by reducing oxygen concentration e.g. nitrogen
Chemical	Link, in competition with oxygen, in the blood hemoglobin (carbon monoxide, hydrogen cyanide being two)
Systemic poisons	Cause either temporary or permanent damage to the body system.

Dusts	Many dusts can lodge in the lungs and eventually produce disease. Historically, three times as many coal miners died from lung disease attributed to dust than died in accidents.
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Table E 8.1.2 Some Properties of Toxics

One substance can exhibit a number of different effects (glue sniffing causes narcosis, irritation of skin, mouth and nasal passages and systemic damage).

Carcinogenic Materials

Many industrial chemicals are capable, to some degree, of causing cancer, and the list of potential carcinogens is growing every year. Exposure to these materials produces effects which are delayed, typically, 15-20 years. The handling of some of these materials presents particular problems since it is generally accepted that there may be no 'threshold dose' below which no damage will be done. Exposure limits are set at levels where risks are acceptable. Benzene is one such chemical where the true "safe threshold" is not known accurately and is falling with time. It has been reduced from 25 ppm v/v to 1 ppm v/v over 40 years. Hydrazine and aniline are two other carcinogens.

Toxic Materials

Liquid Toxics can enter the human body via the skin or the mouth. The skin is not imperious and nerve gases can gain entry through this route. There is also the problem of acids and alkalis as well as sensitivity to certain chemicals – even nickel can produce allergic reactions.

Gaseous Toxics enter the human body via the skin or more likely the lung. Some can result in that person becoming susceptible to lung diseases particularly pneumonia.

Solid toxics enter the human body via the stomach, skin or the lung

The human physiology is not the same as that of animals. Certain chemicals produce different effects. 'Thalidomide' is the classic example that animal tests were not a guarantee of the effects on humans.

Doses for chemicals are to be found in data books. Solid and liquid toxics are often given as LD₀ or LD₁₀ or LD₅₀, the dose which will produce 0, 10% or 50% fatality in a group. The values are usually milligrams per kg weight. LC₁₀ is not really sensible – you have killed someone, somewhere!

Other doses are often given as OEL (Occupation Exposure Limits) in tables.

TLV is the threshold value for 8 hours per day. STEL is the Short Term Exposure time for 15 minutes. **BEWARE** the values are often revised downwards every year as the full effects are re-analyzed and re-assessed.

Toxics and Doses

Different toxics have different effects on the Body systems.

Carbon Monoxide produces carboxyhaemoglobin, most other toxics affect the lung for example Chlorine and Ammonia produce Pulmonary Oedema while Nitrogen Oxides can damage the lung and the cilia and so make the invalid more susceptible to Pneumonia. These effects can be delayed for a few weeks!

The toxic dose models have the form: -

$$Dose = \int C^n t \quad (E 8.1.1)$$

Where

C is the concentration of toxic (ppm usually)

t is the exposure in seconds, minutes or hours

n is a constant.

The value of n varies from toxic to toxic for carbon monoxide n = 0.9, for Chlorine and Ammonia n = 2.75 and for NOX or oxides of nitrogen n can be 3 or 4. What this means is that doubling the concentration reduces the exposure limit by a factor of 6-8. Of more importance is that the actual concentration at any point in a plume is varying with time (see the section on dispersion - E .3) so the true value of: $Dose = \int C^n t$ **may** be significantly higher than might be expected from the TWA value of C from the dispersion equations. In fact it is possible that the dose derived from the TWA value of C might be "harmful" but the true dose, taking into account the peaks, could be fatal.

For Carbon Monoxide a dose of **about** 4,000 ppm for 10 mins is likely to be serious.

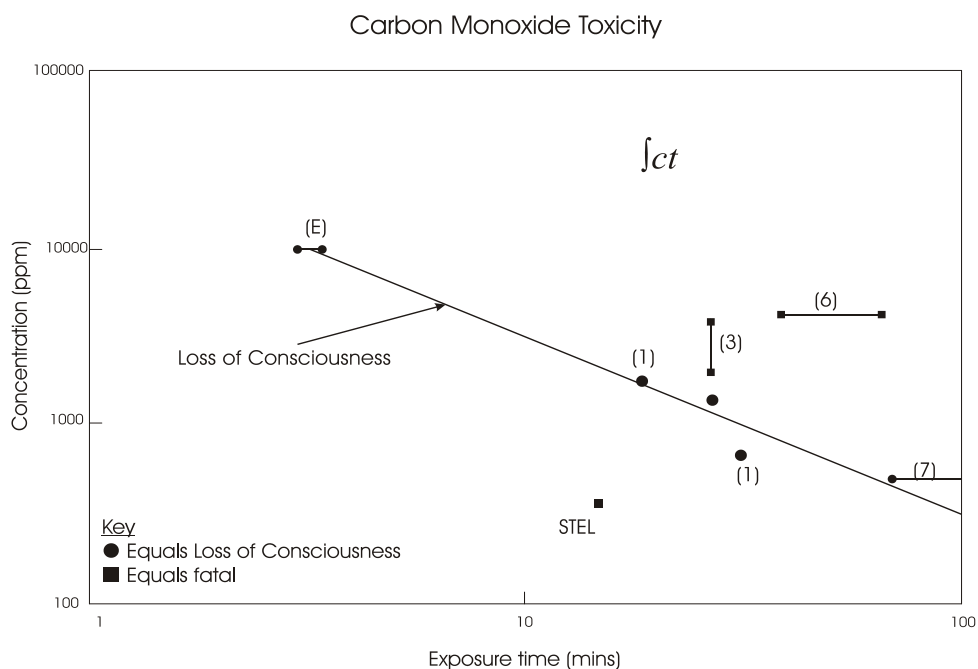


Figure E 8.1.2 Data Used for Toxic Dose of CO (Derived by the Author)

The toxic data is usually derived by plotting the concentration and exposure duration on a log v log graph against time with the recorded physiological effect. The best fit was actually nearer a slope of 0.9 but it was assumed that the fit would be linear. More research shows that the plot is nearer a slope of 0.9.

For Chlorine a dose of **about** 75 ppm for 5 mins is likely to be serious

For Ammonia a dose of **about** 2,500 ppm for 15 mins is likely to be serious

For NOX a dose of **about** 250 ppm for 10 mins is likely to be serious

It is very much a case of horses and courses - or avoid leakage and do not assess the risk by calculation - it is too late once it is out of the piping!

These "numbers" also show the merits of well located "escape" breathing air sets (BA) round plants handling toxic fluids.

As the dose is the area under a curve (the integral of the ingress of toxic into a "refuge" or building produces a rapidly worsening situation. Ultimately if the building is fairly leaky the person may be incapacitated and too weak to help in any rescue.

Probits

There should be some mention Probits for the completeness of "Toxics"- the task is to **avoid** any risks - **prevention** being the objective.

The dose effects can be converted into probability of fatality by a probit equation

$$P = A + B \text{Log}_e (\text{Dose}) \quad (\text{E 8.1.2})$$

$$P = \text{Probit}$$

A and B are constants

$$\text{Dose} = \int C^n t \quad (\text{E 8.1.3})$$

There are equations for Thermal Reaction, Explosions Chlorine, Ammonia, Phosgene etc.

To give some idea of the **likely** range of effects the following are some calculated dose values from probits. These are not quite the same as given earlier as they are derived in a different manner.

Cause	% Fatality		Variable
	1%	50%	
Thermal Radiation	1,000	3000	kW ^{4/3} Secs
Chlorine	120,000	480,000	ppm ^{2.75} Mins
Ammonia	2.6 x 10 ¹⁰	2 x 10 ¹⁵	ppm ^{2.75} Mins

Table E 8.1.3 Some typical hazardous doses based on Probit Values

Thermal radiation is also a "dose effect" that is the effect is a function of time in this case the effect is

$$\text{Flux}^{4/3} \times \text{time}$$

Dose	Effect
Pain	250 (Kw/m ²) ^{4/3} secs
1% Fatality	1050 (Kw/m ²) ^{4/3} secs
50% Fatality	2080 (Kw/m ²) ^{4/3} secs
1st Degree	250 (Kw/m ²) ^{4/3} secs
2nd Degree	1400 (Kw/m ²) ^{4/3} secs
3rd Degree	3000 (Kw/m ²) ^{4/3} secs

Table E 8.1.4 Thermal Doses

(Some of these values were hard to believe as from personal experiments with thermal radiation and it was possible to take 6.3 Kw/m³ for 20 seconds before experiencing **real** pain (230 k^{4/3} secs) - then following a 5 min break the test could be repeated again and again and again. Also a dose of 2000 ppm of ammonia for one lung-full produced a bronchial spasm and the inability to breath. (It is likely that this was the peak (times 2.5 TWA.)

Other Physical Effects

There are many other effects so it is imperative that there is a continuous assessment of the likely injury potential for any task.

E 8.2 Migration of Gas into an Enclosed Volume

The migration of gas into a room can be assessed (as is the concentration profile of a continuously back stirred reactor) as follows:

$$\frac{\text{concentration inside}}{\text{concentration outside}} (1 - e^{-kt}) \quad (\text{E 8.2.1})$$

K = air changes per hour or min - about 1/2 per hour for a modern home and 6 for outdoor building or houses with chimneys and fires

t = time of exposure (hours or min)

E 8.3 Effect Models Humans & Hardware

Heat

Metals lose their integrity when heated. Steel has little strength above a temperature 600°C. Stainless Steels have more integrity but it is not a lot more. Heat gain to a flame-engulfed vessel is of the order of 300 kW/m² from a torch flame and about 75 kW/m² for a pool fire.

The survival damage for humans (as in Triage) is shown in table E 8.3.1.

x body area burned	Age(yr.)																
	0-4	5-9	10-14	15-19	20-24	25-29	30-34	35-39	40-44	45-49	50-54	55-59	60-64	65-69	70-74	75-79	80+
93+	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
88-92	.9	.9	.9	.9	1	1	1	1	1	1	1	1	1	1	1	1	1
83-87	.9	.9	.9	.9	.9	.9	1	1	1	1	1	1	1	1	1	1	1
78-82	.8	.8	.8	.8	.9	.9	.9	.9	1	1	1	1	1	1	1	1	1
73-77	.7	.7	.8	.8	.8	.8	.9	.9	.9	1	1	1	1	1	1	1	1
68-72	.6	.6	.7	.7	.7	.8	.8	.8	.9	.9	.9	1	1	1	1	1	1
63-67	.5	.5	.6	.6	.6	.7	.7	.8	.8	.9	.9	1	1	1	1	1	1
58-62	.4	.4	.4	.5	.5	.6	.6	.7	.7	.8	.9	.9	1	1	1	1	1
53-57	.3	.3	.3	.4	.4	.5	.5	.6	.7	.7	.8	.9	1	1	1	1	1
48-52	.2	.2	.3	.3	.3	.3	.4	.5	.6	.6	.7	.8	.9	1	1	1	1
43-47	.2	.2	.2	.2	.2	.3	.3	.4	.4	.5	.6	.7	.8	1	1	1	1
38-42	.1	.1	.1	.1	.2	.2	.2	.3	.3	.4	.5	.6	.8	.9	1	1	1
33-37	.1	.1	.1	.1	.1	.1	.2	.2	.3	.3	.4	.5	.7	.8	.9	1	1
28-32	0	0	0	0	.1	.1	.1	.1	.2	.2	.3	.4	.6	.7	.9	1	1
23-27	0	0	0	0	0	0	.1	.1	.1	.2	.2	.3	.4	.6	.7	.9	1
18-22	0	0	0	0	0	0	0	.1	.1	.1	.1	.2	.3	.4	.6	.8	.9
13-17	0	0	0	0	0	0	0	0	0	.1	.1	.1	.2	.3	.5	.6	.7
8-12	0	0	0	0	0	0	0	0	0	0	.1	.1	.1	.2	.3	.5	.5
3-7	0	0	0	0	0	0	0	0	0	0	0	0	.1	.1	.2	.3	.4
0-2	0	0	0	0	0	0	0	0	0	0	0	0	0	.1	.1	.2	.2

Table E 8.3.1 Effect of Thermal Radiation on Humans (Triage Table)

Survival probability (0 to 1) is related to age and the area of damage, younger persons may survive 70% burns but older persons may succumb to 30% burns. This is not such a problem as much of the working population have an age 20-40 years old.

Process equipment can tolerate 12 Kw/m² for long periods of time but clothing and cellulose materials (wood or grass) may ignite spontaneously after 30 mins at these fluxes.

Humans

What are not readily described are the effects of hot gases and soot on humans. Above 125°C the lungs can be severely damaged and of course a film of soot in the bronchi and alveoli can be fatal.

There have been a number of unfortunate fires (more particularly in Brazil January 2013) where the fatalities were affected by “smoke”. This may be particulates which clog the alveoli or Carbon Monoxide. However other Products of Combustion (POCs) include Hydrogen Cyanide and Hydrogen Chloride, neither

are safe! It has been noted that partial combustion of hydrocarbons can produce up to 5%v/v Carbon Monoxide. This includes the use of paraffin burning space heaters in buildings!!!

In fires fatality can be caused by trampling as well as internal organic damage leading to “crush syndrome”.

Blast Effects

Overpressure Effects (Mechanical) Effects Human (kilo Pascals)

Overpressure kPa	Effect on Humans
7	0
7 - 21	25% injury due to being blown over
21 - 34	70% injury thrown, physically, against solid objects
34 - 48	95% injury - a worse condition
48 +	Internal injury/disruption, high probability of fatality

Table E 8.3.2 Overpressure Effects - Humans

Once again the values in table E 8 3.3 are to be treated as approximate and indicative. These were derived by the Author from many sources and take into account the “impulse” or pressure multiplied by duration.

Damage	Overpressure kPa
<u>Domestic Houses</u>	
Glass Failure	5% 0.7 50% 2.0 90% 4.5
Building uninhabitable	7
Severe damage to building	15
Total Destruction of Building	75
<u>Process Buildings</u>	
Serious Damage (major report necessary)	15
<u>Process Equipment</u>	
Instrument displaced (major replacements necessary)	5

<u>Piping Spring</u>	20
Storage tank	40
Process piping or Pipe Track	
<u>Serious Damage</u>	20
Storage Tank (90% full)	40
<u>System Displaced</u>	50
Fired Heater	40
Major piece of Heavy Equipment	

Table E 8.3.3 Suggested Damage Profiles for Overpressures