LOSS ESTIMATION FOR REFINERIES AND CHEMICAL PLANT AND RISK IMPROVEMENT

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The subject of this paper is concerned with the prediction of losses in oil refineries and chemical plants. Losses may occur through fire, explosion, machinery breakdown or outside influences such as absence of feedstock or failure to meet environmental standards. Insurance claims for 1989 amounted to over US$ 3 billion for material damage and loss of profits. Most chemical plants are complex consisting of highly integrated process, utility and storage units and the analysis of all the variables can lead to a complicated assessment. The objective is to find the EML as quickly as possible whilst ensuring that this estimate is the worst loss.

Insurance, Risk, Peril, Fire, Explosion, Loss

INTRODUCTION

In setting the objectives for loss assessment of refineries and chemical plant we must fix three important criteria. These are the range of possible events that can cause a loss, their probability of occurrence and the applicability of the loss to the Insured. In this paper the term Estimated Maximum Loss (EML) only will be used. All losses should not be given the same probability of occurrence, those associated with complex process units such as an ethylene plant, for example, may be higher than those associated with a storage area because the plant is continually undergoing parameter changes through its normal operational life, or perhaps it is undergoing modification or expansion. It is unfortunate but true to state that wherever there is a high level of human interaction above that of monitoring, a high degree of risk should be allowed for. Losses during startup and shutdown of a facility are often high and it is worthy of note that 25 percent of recorded losses occur during these periods. Finally the EML has to be based on the business aspect i.e. a company may operate a facility and draw profits from its sales but may not own the plant and would therefore not include the material damage loss in the EML.

DEFINITION OF EML

In this paper the term EML shall include consideration of the effect of spillage of flammable substance or inventory from the largest discrete circuit. For a discrete circuit automatic isolation or remotely operated hand valves may be considered as

a boundary element in defining the volume. In the EML analysis, no prediction of the ignition source location may be made in order to reduce the damage level. No time limit is placed on the discharge of materials in the discrete circuit unless it can be demonstrated that the contents <u>cannot</u> be released in a matter of minutes. The term catastrophic loss shall be given to those events which are either of extremely low probability such as the leakage from an LPG storage sphere or that due to an earthquake. Furthermore, the catastrophic loss ignores the effectiveness of the process isolation or shutdown valves, emergency blowdown valves or vents to flare and all active fire protection. Pipeline releases from remote storage may also be considered in a catastrophic loss for a period of 30 minutes during which time it is reasonable to assume that the flow can be terminated. The use of the term 'Probable Maximum Loss' has caused some considerable confusion in industry and should be discouraged since the specific interest is the 'ceiling loss'. A probable maximum tends to encourage lower insurance coverage because it selects an incident which can be expected rather than one which is worst case.

EXPOSURE TO PERILS

The fire and explosion risk to a refinery or a chemical plant is generally by far the most probable cause of loss. In the processing of flammable hydrocarbons often at high temperatures and pressures, the storage of both feedstocks and products and the transfer for export, there is a degree of risk of material leakage, several probable sources of ignition, for example, furnaces and electrical drivers and therefore the potential for fire and explosion.

In any analysis prior to quantification of such losses, the risk assessor should examine the plant for those unit inventories which are large and volatile enough to provide the potential for an estimated <u>maximum</u> loss. It is possible by qualitative analysis to eliminate those events which result in a low combination loss either because the physical damage is low and fairly quickly repaired or perhaps the loss of profits is of low magnitude. In the EML evaluation there will often be a vapour cloud explosion calculation because this form of incident can have the potential to release a high quantity of energy which may cause extensive damage. The other types to be examined are vessel explosions releasing the inventory due to corrosion or a rapid chemical reaction but these do not usually produce the controlling VCE.

Explosions and fires in refineries and petrochemical plants are more extensive than in other industries. Fires are often difficult to fight and cannot be accessed easily.

Surrounding Exposures

A risk assessment of surrounding exposures at all sites should be undertaken. This is particularly important because many chemical plants are located adjacent to upstream suppliers and

the effects of an explosion in a neighbouring plant can provide a controlling EML.

LOSSES DUE TO NATURAL HAZARDS

The risks presented by natural hazards can be extensive, for example, a hurricane can destroy plants, storage areas and buildings (refer to loss number 26 in table 5). There is also a serious risk of the emergency shutdown of a plant causing damage, loss of products, catalysts and chemicals. In the adverse freezing conditions in 1989 in the United States of America, for example, loss of profits in chemical and refining plants were experienced due to shutdowns through inoperable conditions, (references 28 and 29 in table 5). Other risks such as lightning which could result in damage of tall structures, subsidence and collapse, tsunami and flood, storm and water damage need to be considered. The estimation of storm damage is difficult. Flash flooding should be catered for in the plant drainage design. The assessor should satisfy himself that existing site drainage is adequate to handle the maximum firewater discharge plus stormwater. The possibility of drainage blocking should also be evaluated. The layout is an important criterion. Plants built on basically flat ground with little slope difference in any direction are generally safer. Spillage from the neighbouring plants could also be a problem particularly where there is an absence of curbing around equipment with large liquid inventories.

The subject of earthquake risk assessment is particularly complex and is closely related to probability. For those areas of the globe which are subject to the risk of earthquake, the loss assessment should be included in any insurance evaluation. This type of loss cannot be estimated without a thorough examination of the plant structural aspects onsite. The ability to withstand a major seismic movement is not only dependent on the design criteria but also on the current condition of the plant. In estimating earthquake losses, line fracture should be considered particularly where there are many units which can release the quantity of flammable material necessary to produce a VCE. Furthermore, little outside assistance should be considered, blockage of roads may prevent the arrival of the municipal fire brigade and line fracture may also reduce the effectiveness of the onplot firefighting systems.

Recommendations on the inclusion of remedial measures to reduce earthquake exposure are often fairly inexpensive and the corrective action can be carried out quickly with the units on-line.

LOSSES DUE TO THE HUMAN FACTOR

In addition to losses through leaving valves open and errors in operating the plant, there are the external risks of aircraft collision damage, road vehicle impact, riots, strikes and civil commotion. There have been a few instances of terrorist attacks on oil installations but the losses to date are of low order. Security is normally of high standard in most plants because of the value of raw materials, products, machinery and spares.

Methodology for Estimation of Maximum Losses

In the Estimation of Maximum Losses for a plant all physical damage and consequential loss scenarios are evaluated for each processing area. It is possible to discount those associated with small inventories which either result in low damage explosions or those which result in higher physical damage but minimal consequential losses.

Vapour Cloud Explosions

With many plants involving the handling of flammable components, the loss of containment can result in a leak of sufficient volume and rate to result in the formation of a vapour cloud. The mixture of the cloud with air to form a flammable mixture can lead to a composition in the explosive range which will find a source of ignition and provide a serious loss. The vapour cloud explosion (or VCE as it is commonly termed) often provides the most severe damage level within a refinery or chemical plant and therefore the largest monetary loss. It is worthy of note that approximately a third of all major losses to date have been caused through VCEs. The main sources of loss are the process units themselves, storage such as the tank farm or transfer piping connecting the two areas. There is a tendency to search the plant for the largest inventory and therefore establish the loss based on a vapour cloud explosion following leakage from a large storage area and subsequent ignition. However, it should be borne in mind that storage areas are deliberately located away from active plant areas and therefore present a lower risk of loss of containment and there are limited sources of ignition. There is evidence[1] that a vapour cloud explosion requires a minimum accumulation of 2 to 5 tonnes of material to ensure that deflagration does not occur as opposed to an explosion, the basic difference being that in a deflagration the overpressures generated are so low that the forces exerted on adjacent plant do not cause significant damage. The vapour cloud burns rather than explodes. Secondly and more importantly, with reference to the above comments on tank farms, it is not the quantity of material released that matters, it is the quantity that is in the explosive range, since it is only that quantity that contributes energy to the explosion. However, the larger the cloud the higher the probability that a significant portion of it will be in the explosive range. The acceleration of the flame front is particularly important with respect to the magnitude of the explosion and therefore materials such as ethylene which are highly reactive present the worst risk with respect to damage ratio. The flame speed is a prime factor in the establishment of the characteristics of a blast wave since it is the expansion of the gas cloud through production of combustion products (i.e. the burning of gas) which produces a force on the unburnt mixture thereby

causing it to move away from the flame zone. Therefore the assumption of the various methods in evaluating the loss is that a portion of the spill will be in the explosive range at a particular time.

The subject of rate of release is often overlooked. A very large leak may promulgate turbulence and enhance mixing with air. A very slow leak will be diluted outside the explosive range before the mass of vapour for a VCE can be accumulated. However for indoor explosions the concentration of material within the building will not be diluted and the ensuing explosion damage is enhanced by confinement. In the ranking of hazards in the hydrocarbon industry it is apparent that flammable gases containing 2-6 carbon atoms have been responsible for practically all vapour cloud explosions and further those with 3-4 atoms possess those properties to account for more than 40% of all industrial plant accidents. The olefin (unsaturated chain hydrocarbon) is also to be ranked above that of the corresponding paraffin. Ethylene, for example, is a material which does not deflagrate due to its reactivity.

The propane, propylene, butane and butylene range have the following properties which make them particularly hazardous:-

1) The molecular weight is above that of air and therefore the gas cloud will stay close to ground level as opposed to dispersing. Note: the dangers of materials such as methane or hydrogen in outdoor plants are reduced by their dispersion into the atmosphere. This point is particularly relevant in the damage analysis of a plant from an explosion point of view but it should be stated that serious local explosions and fires are possible with respect to hydrogen because of its flame speed and its high reactivity and with methane because of its flammability. Methane is not a reactive gas and there are few reports of it being involved in VCEs.

2) The volatility of the C_3 and C_4 hydrocarbon range requires these hydrocarbons to be stored under pressure unless they are refrigerated and the former provides the mechanism for high leakage rates. In vessels where a BLEVE (Boiling Liquid Expanding Vapour Explosion) could occur, this range of components are the most hazardous.

The analysis of EML in all plants should include for potential explosions based on the release of any of the above components i.e. LPG. This will often yield the EML.

The method of calculation most recognised in industry is the IOI method[2] to establish the controlling scenario. The Institution of Chemical Engineers has published 'The effects of explosions in the Process Industries'[3] and this method uses equipment operating conditions but requires a method correlating blast damage against loss. There are methods of prediction of vapour cloud shape and modelling is also discussed in this paper.

The compactness of units is a particularly important evaluation criteria. There are two important aspects to note in this respect; as the distance between adjacent plants decreases the probability of a higher damage factor to the surrounding plant greatly increases and where plants are constructed on a high density ratio, the degree of destruction is also increased. Therefore, the lower risk plants are built on unconfessed plot areas adequately spaced from adjacent units. The factors determining layout are seldom process related; they are economically based on minimising land usage. Furthermore the spacing criteria adopted in design are rarely capacity related (reference confidential procedure Suregrove Limited[4]).

The majority of estimations on damage can be applied satisfactorily to onshore 'open-air' plants because the density factor is low but those in buildings should be evaluated carefully. The confinement of an explosion will have disastrous effects and this has been confirmed by recent evaluations on offshore platforms noticeably, the Piper Alpha platform, where significantly higher overpressure values have been experienced than those predicted by the methods described above. The upgrading of blast wall protection is necessary to combat the damage potential of a confined explosion.

TNT EQUIVALENCE METHOD

In the classification of explosions the damage potential is usually defined in terms of TNT equivalence i.e. the equivalency is established by comparing against the heat of combustion of TNT with that of the fuel involved. This is primarily because there is a large amount of statistical data available for the detonation of TNT and therefore in the prediction of damage. It should be added at this point that the equality of a point source TNT explosion is not rigidly valid to a VCE with its cloud volume and its relatively slow flame speed. This method tends to over predict the blast wave created and therefore the damage ratio. However it is important to attempt to find a relationship between the blast wave and a mass of exploding vapour. Also the explosion durations are longer for VCEs as opposed to TNT explosions and it is this fact which compensates for damage ratio since the VCE equivalent of longer duration can produce damage similar to that of a higher TNT mass explosion. The amount of energy which can be converted into blast energy is only a small proportion of the value of the heat of combustion, perhaps as low as 4 to 6% for an explosion involving masses of fuel in the range of 10-40 tonnes. It should be noted that there is little evidence of any vapour cloud explosion resulting from an accumulation of over 100 tonnes and calculations using the available empirical methods should not be carried out for inventories above this figure.

INSTITUTION OF CHEMICAL ENGINEERS METHOD

The method described in the document 'The effects of explosions in the Process Industries'[3] uses equipment operating conditions and considers the flashed fraction of a release coupled with an explosion efficiency factor. This method requires an interpretation of the overpressure levels in terms

of empirical correlation with damage levels. The decay of overpressure with distance is best estimated by the simplified model developed by Hopkinson[5] which is the scaling law:-

$$D = D_1 \times (M_{tnt}^{1/3}) \qquad \text{....(A1)}$$

where D = the scaled distance

D_1 = the distance from the explosion centre

M_{tnt} = the mass of fuel

The method of calculation is illustrated for the ethylene plant shown in Figure 1 where an incident in the propylene refrigerant section of the plant results in the release of 36 tonnes of liquid propylene.

TABLE 1 - Calculation of overpressure circles by I Chem E method

Propylene Properties

Where the vaporisation factor F is given by:-

$$F = 1 - e^{-C_p(T_1 - T_2)/H_{vc}} \qquad \text{....(A2)}$$

C_p is the mean specific heat(liq)(kJ/kg^{o}C)
T_2 is the boiling point (^{o}K)
T_1 is the ambient temperature(^{o}K)
H_{vc} is the latent heat of vaporisation(kJ/Kg)

The thermal properties are extracted from reference[6] page 3-218

$$F = 1 - e^{-2.375 \times (293 - 235)/427} = 1 - 0.7243 = \underline{0.276}$$

M_{tnt} (the TNT equivalent) is calculated by considering the mass vaporised and multiplying by the TNT equivalency of 10 for hydrocarbons and twice the vaporisation factor and by the explosion efficiency (taken as 0.042).

$$M_{tnt} = 36 \times 10 \times 2 \times 0.276 \times 0.042 = \underline{8.35 \text{ tonnes}}$$

Scaled distance D_1 (using equation A1)

$$= \frac{81}{(8.35 \times 1000)^{-1/3}} = 4.01 m/kg^{-1/3}$$

Using fig. 2: At 81 m there is an overpressure of 0.7 bar
At 111 m there is an overpressure of 0.41 bar
At 172 m there is an overpressure of 0.21 bar

The mass which vaporises and forms a vapour cloud is estimated to be 10 tonnes (36 x 0.276) using the equation A2 (Table 1). The TNT equivalency of the explosion is based on the assumption that the energy ratio for hydrocarbons to TNT is 10:1. This value and other materials are given in reference[3] Table 4. The combustion energy explosion factor is taken to be 0.042 or 0.03 if the cloud curvature is ignored.

In the example given the cloud is ignited in the compressor house. However it is easy to see that the damage predicted is severe in all directions of cloud drift. This particular unit has many sources of ignition and therefore a high probability of explosion on a significant releases of process fluids. In Table 1, the I Chem E method is used to to evaluate the overpressure radii at three different values i.e. 0.7 bar (10 psi), 0.41 bar (6 psi), 0.21 bar (3 psi) and the level of damage expected within the plant is described in Table 4. The overpressure circles are drawn on Figure 1 for the above calculation. A damage value may be established by using a value close to 100% for the plant within the 0.7 bar envelope with perhaps 80% for the 0.41 bar circle and 40% for the 0.21 bar circle. These values are based on the damage factors predicted in Table 4 of this paper. For this unit the decay of overpressure will not influence the peripheral loss unless the unit borders other plants and therefore the selection of the secondary plant damage values is not critical. For a 150,000 tonnes per year ethylene unit the 1991 end of year financial loss calculated was US$ 240 million for fire and explosion. This loss is termed the 'Material Damage' Loss and for the example given is part of the controlling EML.

It should be noted that values may vary for the parameters given in the equations A1 and A2, in particular the energy ratio, the efficiency factor and therefore the TNT Equivalence factor are considerably different for other types of chemicals and substances.

US BUREAU OF MINES EVALUATION OF TNT EQUIVALENCE

This calculation method[7] assumes that an explosivity factor of 0.02 is applied to the cloud based on the assumption that 100% of the spill vaporises. It uses the heat of combustion of the gas and, for the example given in the previous method in Table 1, the results are shown in Table 2. The calculation of vapour cloud mass in the past has been made on one quarter of the spill mass for materials relating to C_3 type or one fifth based on materials relating to C_4 type hydrocarbons but this method can obviously lead to inaccuracies. The explosion efficiency factor is not representative of other materials varying from the C_3 to C_4 range. The best use for the method is in evaluating a potential unit loss in the absence of plant operating and design conditions. It should be superseded by the previous method when sufficient plant data is available. The overpressure circles are at much higher radii from the explosion centre because the entire gas cloud mass was used in the ethylene unit example.

TABLE 2 - Calculation of the overpressure radii by the TNT method

$$M_{tnt} = \frac{M_{vc} \times H_c \times f}{4 \times 10^3} \qquad \text{....(A3)}$$

M_{vc} is the mass of the vapour cloud(tonnes)

M_{tnt} is the TNT yield of the cloud(tonnes)

H_c is the heat of combustion (kJ/kg)

f is the explosion efficiency

The thermal properties are extracted from reference[6] page 3-156

$$M_{tnt} = \frac{36 \times 47200 \times 0.02}{4 \times 10^3} = \underline{\underline{8.50 \text{ tonnes}}}$$

Using Fig.3: 0.7 bar overpressure circle has a 128 m radius
0.41 bar overpressure circle has a 167 m radius
0.21 bar overpressure circle has a 250 m radius

Compatible radii are calculated for this method if the fraction vaporised from the I Chem E method is used to reduce the mass included in the vapour cloud. The yield of TNT is much lower than that of the I Chem E method but the explosion factor appears to compensate.

IOI METHOD

The circle method is based on a concept developed by International Oil Insurers[2] in which the leakage of hydrocarbon from a discrete circuit occurs and a percentage of the material is assumed to vaporise giving rise to a vapour cloud explosion (or VCE). Again the forces generated by a known mass equivalent of TNT are used but these are equated against two damage contours, one the 80% circle in which this percentage is anticipated as a loss and the second larger diameter circle in which a 40% damage factor is expected between the inner circle and the outer circle (in the annulus). If losses are equally distributed the overall loss is 50% within the large circle. The IOI method defines a single discrete circuit by considering the limiting flow rates to be greater than 10kg/min for remotely operated valves and greater than 100kg/min for circuits with automatic remotely operated valves.

In the ethylene example given the mass vaporised into the gas cloud is assumed to be 5% of the discrete propylene circuit and the 80% and 40% damage circles are located at 72 and 144 metres respectively for a mass of cloud equal to that of the first method. The IOI method also assumes that these circles can drift across the unit plot to a maximum distance of 180 metres in any direction. The IOI method also takes into consideration the account of fire damage from assessments of known occurrences. It agrees with the I Chem E method predictions fairly well. VCE Damage Radii for various overpressures are given by :-

TABLE 3 - Damage Radii for the IOI method

Mass of Fuel (tonnes)	Inner Radius (metres)	Outer Radius (metres)
1	33	66
2	42	84
4	53	106
6	61	122
8	67	134
10	72	144
20	92	182
30	104	208
40	115	230
50	123	246
60	132	264
70	139	278
80	144	290
90	151	302
100	156	312

In all the above methods the difficulty in estimating the yield of an explosion is in the prediction of:-

a) The symmetry of the cloud at the point of ignition. The TNT methods make no allowance for cloud shape characteristics other than those included by way of the damage analysis.

b) The vapour-air mixture will vary from above the flammable limit to stoichiometric to being below the flammable range. Each method assumes a percentage of the cloud will be in the flammable/explosive range.

c) The location, number and energy strength of sources of ignition cannot be readily determined. With large releases, the cloud may fragment or there may even be several sources of ignition.

It should be stressed that the calculations are estimates only and the following key assumptions are critical in the analysis:-

d) The leakage mass or the rate from a discrete circuit includes an estimate of inventory discharged before a valve can be closed or prime movers such as compressors or pumps can be shutdown on a continuously operated plant or the circuit drains (batch plant or closed circuit).

e) An explosion results from the vaporisation of part or all the contents of a discrete circuit.

f) The damage factor estimated for overpressure circles is taken to be of the same magnitude (resulting from combustion of different substances) and the effects on all structures and plant is the same.

VAPOUR CLOUD MODELLING

In the modelling of vapour clouds the prediction of blast wave shape, duration and overpressure are based on hemispherical cloud shapes or other recognised volumes. The main problem in defining the shape of clouds produced in process units is the degree of confinement. The shape of a cloud will be influenced by the equipment local to its formation point and also any drift due to environmental conditions will change its characteristics. Calculations applied to individual sections of a vapour cloud which use a higher blast value strength for a confined portion of the cloud and then add a contribution for the unconfined burning to give a resultant specification may not predict the potential damage to any degree of increased accuracy. The more sophisticated model techniques are better applied to building explosions where the external influences are not so great and the shape is more easily defined. Modelling should only be carried out in situations where there are less degrees of freedom and unfortunately the employment of more parameters cannot necessarily predict the degree of air mixing in open-air locations (reference 8).

HISTORICAL CALCULATION

The prediction of damage to a plant by comparison with records from other losses appears to merit the highest potential for obtaining order of magnitude estimates. Recorded data is becoming more accurate through the deployment of sophisticated computer systems and and software. In the future these systems may be capable of data logging to permit the evaluation of the mass released and therefore the size of the cloud, and perhaps the use of CCTV will even record the shape. Study of damage will then produce accurate modelling of explosion effects. It is also to be hoped that new plants will be built in more generous plot spacing and therefore the degree of confinement of vapour clouds and their release of energy can be more accurately evaluated (reference confidential procedure Suregrove Limited[4]).

DAMAGE

The prediction of damage to plants and surrounding property is the objective of the analyses. The accidental nature of the VCE makes it extremely difficult to obtain reliable data and as such the quantification of damage is not easy. No two VCEs are alike. Localised damage is usually intense and this often could mean values close to 100% of loss but the peripheral areas are not easily estimated. Glass breakage, for example, is highly variant depending on the pane strength. The correlation between glass breakage and overpressure gives the values in the range of 0.03 to 0.07 bar (refer Table 4). However the correlation between TNT and VCE explosion is less accurate at low overpressures.

Overpressure occurs as a consequence of the combustion process which creates a pressure volume increase dependent on flame speed, oxygen fuel concentration and the reactivity of the fuel. Initially the pressure is high and the wave velocity is fast. This decays as the blast propagates radially away from its point of origin or ignition. The generation of positive overpressure is followed by a negative phase (i.e. below atmospheric) which can cause damage in the some cases as extensive as the overpressure phase. Buildings which are designed to withstand positive overpressure blast may collapse in the negative phase.

The damage produced by a blast wave occurs as a result of the instantaneous rise from ambient pressure to a peak at incident pressure. The blast wave compresses and heats the air in front of it. The wave front travels at a velocity greater than ambient sonic velocity. As the blast wave expands into an increasingly larger volume away from the source of ignition, the peak pressure in the blast wave decreases but the duration of the overpressure increases. Classification of damage is based on the peak incident pressure known as side-on pressure. The prediction of peak overpressure[9] is not quite as simple as it seems. Theoretical studies[9] suggest a peak overpressure of 0.2 bar for a 100 millisecond duration is possibly more representative of the of a VCE and this corresponds to the overpressure contour at the 40% level for the IOI method[2]. The IOI inner circle corresponds to 0.4 bar at 80% damage.

The forces acting on a structure consist of static and dynamic pressures imposed over a period of time and these consist of three components:-

1) Force resulting from the incident side-on pressure.

2) Force resulting from the dynamic pressure.

3) Reflected pressure resulting from the overpressure.

Diffraction loading contributes the greatest damage in refinery and chemical plant explosions. Small structures suffer less from diffraction loading because the time interval in which the blast wave envelopes the object may be less than the plastic response of the object to the differential loading. For large buildings such as multi-level plants and tall structures the differential loading may cause complete destruction. In an explosion the maximum pressure differential between the front and the back of the structure exists at the moment of reflection. The translational forces developed tend to try and move or overturn the structure in the direction that the blast wave is travelling.

TABLE 4 - Relationship between Damage Criterion and Overpressure

TABLE 4

DAMAGE CRITERION	OVERPRESSURE(Bar)
Glass Breakage	0.03 - 0.07
Minor Brick Damage Shelter collapse Empty storage tanks fail	0.07 - 0.14
Blast wall failure Equipment uprooted Pipes fractures(not designed for blast) and surface controls are damaged	0.14 - 0.28
Structural frames collapse Debris hurled, severe structural damage to full tanks	0.28 - 0.41
Pipes fracture on heat exchangers, spheres and equipment overturns	0.41 - 0.55
Total destruction	Above 0.55

The key feature to blast resistant construction is the ability of the structural elements to absorb large amounts of blast energy without deformation which leads to destruction (reference 9).

COMMERCIAL INSURANCE

The above sections have dealt with material damage to plant and the methods available for estimation. In a combination loss the EML will also include Loss of Profits (or Business Interruption Losses), which are becoming increasingly important to many operators for a variety of reasons:-

1) The outage of units in recent years has lengthened due to intensive periods of industrial activity in which the delivery time of equipment and materials has increased and activities take longer in the construction phase.

2) The restructuring of industrial companies providing a more integrated approach has resulted in a higher degree of risk.

3) The amalgamation and concentration of plants into larger and more profitable centres has increased the risk.

Business Interruption (BI) Losses do not necessarily arise from the physical damage to the plant. Machinery breakdown with the absence of a spare or its unavailability often leads to shutdown and loss of profits. The supply of feedstocks and the reception of products are covered via policies for 'Suppliers and Customers Extensions' of the common elements between two plants or companies. This form of loss is often not easily predicted, for example, the loss of steam sales from a refinery to the ethylene plant (described earlier in this paper in a nearby chemical complex) through an incident at the Customer's plant can mean a high loss value for long-term shutdown. The degree of integration must be evaluated very carefully. In the ethylene unit example reconstruction of the plant may take two years with a loss of sales of perhaps US$ 100 million in sales revenue and the overheads costs adding perhaps another 60%. Therefore an EML approaching US$ 400 million for the combination loss of material damage and business interruption could have to be underwritten.

Table 5 provides a characteristic breakdown of 40 of the largest losses in the last 10 years with a breakdown of the claims submitted both for material damage and that for business interruption. It should be noted that only claim figures are reported and therefore the fact that no business interruption is given may result from no insurance taken out or no loss of profits. Furthermore BI is purchased on a time dependency basis and therefore where the policy limit expires the value lost at the termination point is stated. For example, an outage of 18 months is paid only in a particular policy even though the plant may have taken two years to rebuild. The estimate of construction durations in fluctuating markets is difficult. The most critical point in the analysis is the value given to the assets. Many plants can be undervalued in terms of replacement for new. Valuation of all industrial facilities is a function of the current engineering, procurement and construction market situation, the location, currency purchasing power for home and overseas materials, equipment and services. There is no such revaluation on a 10% uplift per year basis.

The Current Industry Status with Respect to Losses

This section reviews the current industry status with respect to large losses in refinery, petrochemical plants and chemical plants. The plants listed in table 5 are selected from a minimum value criteria of US$ 50 million (combination loss if applicable).

FACTORS DETERMINING FINANCIAL LOSS

In recent times the tendency to construct large capacity single train process units to maximise profitability has created a situation in which a large inventory leads to a large quantity of release and therefore a large diameter loss radius. The relative spacing of equipment is not connected with capacity and perhaps it should be for highly hazardous units. The single train aspect means that any interruption to the process caused by main compressor outage, for example, will lead to total shutdown.

The age of plants is also an interesting subject in terms of

TABLE 5 - Insurance Losses over US$ 50 million for the refining/chemical industries

- - - - - - - - Loss Details- - - - - - - - - - - - - - - -					—— Millions of US Dollars)——		
No.	YEAR OF LOSS	COUNTRY	PROCESS	TYPE OF LOSS	MATERIAL DAMAGE	BUSINESS INTERR	TOTAL LOSS MILLIONS
1	1983	Indonesia	LNG	Explosion	33	150	183
2	1984	USA	Refinery	Explosion/Fire	203	0	203
3	1984	Canada	Synth.Oil.	Explosion/Fire	75	95	170
4	1984	Italy	Ethylene	Fire	52	0	52
5	1985	Germany	Ethylene	Explosion/Fire	40	38	78
6	1985	GB	Peroxide	Fire	62	60	122
7	1985	USA	Refinery	Fire	13	38	51
8	1985	Italy	Ethylene	Fire/Explosion	74	0	74
9	1987	Ecuador	Pipeline	Earthquake	120	0	120
10	1987	GB	Refinery	Explosion/Fire	68	0	68
11	1987	Belgium	Ethyl.Oxide	Explosion/Fire	78	0	78
12	1987	Canada	Refinery	Explosion/Fire	14	62	76
13	1987	USA	Acetic Acid	Butane VCE	200	160	360
14	1987	USA	Acetic Acid	Explosion/Fire	15	37	52
15	1987	USA	Refinery	Pollution	90	0	90
16	1987	USA	Refinery	Propane VCE	300	0	300
17	1988	Norway	VCM Plant	Explosion/Fire	17	67	84
18	1989	Africa	Synth.Oil.	Fire	18	39	57
19	1989	India	Paints	Explosion/Fire	63	76	139
20	1989	Japan	Refinery	Explosion/Fire	31	44	75
21	1989	Belgium	Ethyl.Oxide	Explosion/Fire	79	208	287
22	1989	USA	Refinery	Explosion/Fire	79	0	79
23	1989	Africa	Synth.Oil.	Fire	6	104	110
24	1989	USA	Ethylene	Fire	30	60	90
25	1989	USA	Ethylene	Explosion/Fire	8	56	64
26	1989	US Virgin	Refinery	Hurricane	134	138	272
27	1989	USA	Polyethylene	Explosion/Fire	718	675	1393
28	1989	USA	Chemical	Freeze	17	64	81
29	1989	USA	Refinery	Freeze	0	63	63
30	1990	France	Ethylene	Explosion/Fire	9	85	94
31	1990	USA	Chemical	Explosion	20	300	320
32	1990	Saudi	Refinery	Fire	57	0	57
33	1991	USA	Refinery	Fire	25	70	95
34	1991	Germany	Chemical	Explosion/Fire	53	54	107
35	1991	S Korea	Refinery	Fire	51	0	51
36	1991	USA	Refinery	Explosion	20	37	57
37	1991	Mexico	Chemical	Explosion	97	0	97
38	1991	USA	Ethyl.Oxide	Explosion	90	0	90
39	1991	USA	Refinery	Fire/Explosion	90	125	215
40	1991	USA	Chemical	Explosion	60	35	95

risk. The loss is not always of higher probability with an older plant, although the probability of leak is higher. The use of deductibles to reduce claims for short periods or for small amounts is common to the insurance industry. Both time and money are offset against the loss. Shutdown for a period of two weeks is not uncommon and therefore a 14 day deductible is included to offset multiple claims for small repairs. The cost of cleanup, accessibility, and removal of wreckage after an incident requires careful allowance. The reusability of the residual plant is often overestimated in analyses. Tanks which are damaged may have to be totally rebuilt, structures deformed by overpressures pulled down and new ones erected.

The cost of rebuild is influenced by the configuration of the new plant. Valuations on a plant which has been gradually expanded should take into consideration that a new plant of equal capacity would be designed and constructed in the most optimal manner i.e. with the minimum equipment and therefore may be of lower value. It will also be modernised with the incorporation of new control systems such as digital in place of analogue. Delays in the construction phase are not uncommon due to engineering update and production of project drawings and market conditions.

RISK IMPROVEMENT MEASURES

Design Change

A major turning point in the design history of refinery and chemical plants was undoubtedly the tragic Flixborough disaster in which over 50 tonnes of cyclohexane produced a 100 metre diameter vapour cloud which ignited in a reformer furnace to produce an explosion which destroyed the entire site. The loss of life in the main control centre was a heavy percentage of those in the plant.

Operations Personnel & Control Centres

Modern control rooms are constructed of blast resistant design with small aperture windows or none at all. With the sophistication of the control system it is now less and less necessary for operations personnel to be on site or located close to the plant. The use of CCTV cameras has become a standard with observation permissible from a safe area. On the other hand many companies still believe that there is no substitute for a walkround. It is claimed that 50% of leaks that occur are still detected by operators. Recommendations issued by the Chemical Industries Association[9] for control centres are based on the primary criterion of 0.7 bar with a positive duration of about 20 milliseconds assuming an explosion at ground level. In the setting of objectives for new plant design:-

a) The location of any control facility within an enclosed structure should be away from the key areas of risk particularly

within damage circles where the structure will not withstand an overpressure of 0.7 bar. There are many instances today where control rooms overlook the plant so the operators can get a bird's eye view. However these buildings, which were constructed with glass windows and contained analogue computer control systems, offer minimum resistance to damage by explosive forces.

The optimum construction for a control room is a flat one story building because it experiences less blast loading and therefore the overturning moment is smaller.

b) The orientation of the building is important and the smaller wall area should face the most probable source of explosion to minimise the exposure.

c) The location of heavy objects such as air conditioning equipment on the roof is to be discouraged. There is also no reason why a control room should be located close to a large heavy piperack or process equipment which is elevated or of a tall nature.

d) Internal fittings and fixtures should be designed to prevent their dislodgement when the structure is deformed by a large explosion force. The integrity of personnel safety and the control system is of paramount importance.

d) It is also apparent that many older plants have sited the office block close to the processing area. Engineering and administration should be located away from the the process units. The utilisation of blast walls around large vessels containing volatile components, or reactors operating at high pressures is also a suitable method of risk reduction.

e) A serious explosion will undoubtedly cause the loss of field instrumentation systems and therefore partial or total loss of controllability. Thus the control area should be primarily designed to protect the operators as opposed to the computer systems.

f) The development of emergency plans for plants by safety officers and management,will do much to assist in the promotion of safety and even the reduction of damage, instructions for actions to be taken during:

(i) An earthquake
(ii) A serious unit incident involving release of potential vapour clouds and subsequent explosions on the refinery site
(iii) A serious incident at any of the neighbouring plants.

The External Factor

The workforce is another area where the human factor can introduce serious problems. The use of contract labour unfamiliar with the plant for maintenance or operation presents the owners with an unnecessary risk. The placement of such contracts with outside companies should never be based on purely financial grounds. All staff should be trained in their job function, external contractors should be assessed and their standards examined in addition to the provision of comprehensive supervision at all times. The key is the placement of those jobs which are hazardous or can lead to a dangerous situation with those who recognise the consequences of error. The 'Permit to work' systems are of paramount importance to the safety of any plant.

The introduction of advanced plant monitoring techniques has contributed significantly to minimising the downtime of units by reducing the number of shutdowns due to trips caused by the failure of individual plant components. However, it must also be stated that this has tended to reduce the number of inspections of equipment and the path to failure may go undetected.

The Process

Minimisation of the storage of highly flammable materials and reduction of process equipment inventories to a minimum in the design phase should be a prime objective. In particular, with large capacity plants, the deployment of process shutdown valves needs to be addressed more stringently. The reduction of inventory by increasing the number of discrete circuits is a positive feature towards risk improvement. Where large quantities of flammable hydrocarbon products or feedstocks have to be stored in the tank farm, the location is critical and the prevailing wind direction usually determines the positioning of LPG spheres or bullets on a probability basis. The selection of refrigerated or pressurised (non or partially refrigerated) storage is made on economic grounds but there is no doubt that refrigeration reduces the release both in terms of the the leakage rate and secondly in terms of vaporisation. The autorefrigeration effect is not taken into account in the calculation methods described above but it is known that it will restrict the vapour cloud size. In the IOI method 1% of the storage volume is taken to be the release in the cloud for pressurised storage and 0.01% for atmospheric storage illustrating the difference in risk.

Protective Systems

The protections provided in the design of most industrial plants are firewater, foams for particular chemicals or hydrocarbons, passive fire protection in the form of fire resistant coatings and concretes, inert gases such as Halon and carbon dioxide, and dry powders. The deployment of fire and gas detection systems is increasing in industry with the latest advances in detector design. The important criterion is the location of sufficient activation points.

However it must be stated that the effectiveness of these systems is dependent on the human factor i.e. the response time in the decision for application. In serious losses the firefighting systems have often been rendered inoperable by fire or explosion

damage. Their effectiveness has often been in the cooling of adjacent facilities, storage tanks or columns and not for extinguishment. In the examination of the safety system it is important to check whether a system can operate to the capacity intended, for example, can the capacity of the firewater pumps be carried by the piping. There is no doubt that the deployment of spray systems above pump seals promotes the possibility of extinguishing the fires but the best protection method is to avoid the location of pumps handling fluids above their autoignition temperature directly below piperacks containing heavy vessels, aircoolers and significant piping weights.

The upgrading of firefighting systems should also be fully investigated to ensure that the additional equipment, lines and fittings actually add safety to the plant. Poorly routed lines or non fire resistent materials may actually hinder the operation by loss of integrity of the system during a fire.

Maintenance of all protective systems must be continuous and this includes repairing passive fireproofing and active system testing.

Catastrophic Loss

The protection of plants against earthquake loss is made by enhancing the robustability of support structures to seismic shocks. The use of cross bracing on the legs of table or plinth supporting structures and LPG sphere legs and the ring beam foundations below cylindrical tanks will prevent collapse. The correct anchorage of vessels, particularly those in elevated structures is an important aspect.

Concluding remarks

The methods described in this document are empirical correlations in order to evaluate the EML for insurance purposes only. Refinements are made from time to time and calculations modified to produce financial assessments of expected losses.

On the business front recent lean years in the refining and petrochemical industry have lead to a philosophy of keeping the overhead costs down. Such overheads have undoubtedly been extended to encompass the enhancement of safety systems, maintenance expenditure and technical services. This is reflected quite clearly in the trends shown for the last decade of industrial losses.

One final comment is made on the modification of refineries and chemical plants. There will always be debottlenecking and revamping involving the addition of plant or the removal of obsolete equipment. This activity should be very carefully matched with the existing plant design to ensure compatibility and integrity is maintained and safety is always of the same high standard. There have been a number of incidents connected with changes of plant configuration.

The expenditure on intensive safety analyses such as HAZOPS and HAZANS is to be encouraged to promote a higher understanding of risk.

The need for risk awareness is obvious from the tabulated losses given in Table 5 and in order to reduce the upward trend in incidents and losses, both technical and business aspects need to be closely evaluated. In particular the interdependency of plants will have a measured effect on the loss values arising in the 1990s as well as the material damage through fire and explosion or natural peril.

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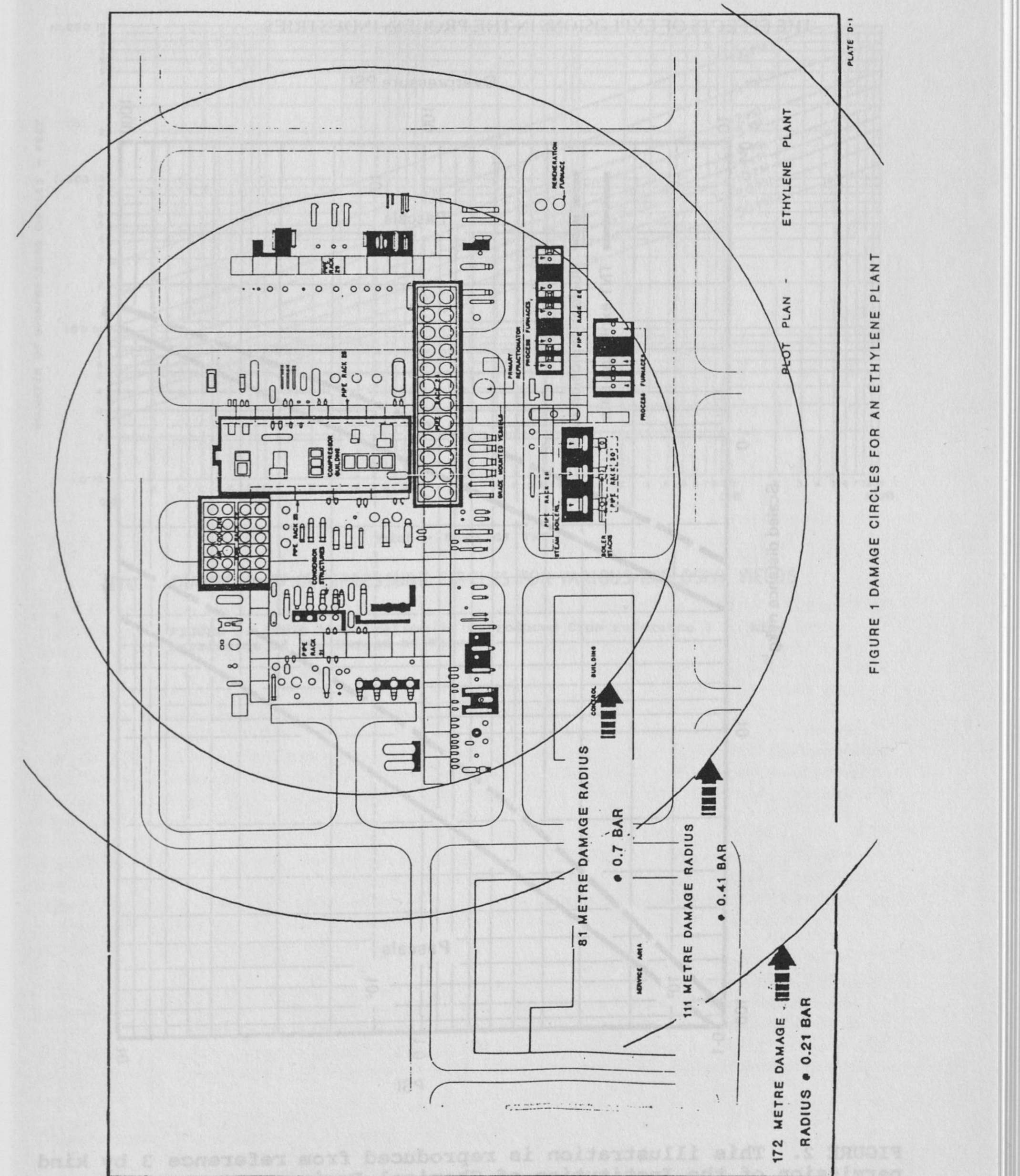

FIGURE 1 DAMAGE CIRCLES FOR AN ETHYLENE PLANT

THE EFFECTS OF EXPLOSIONS IN THE PROCESS INDUSTRIES

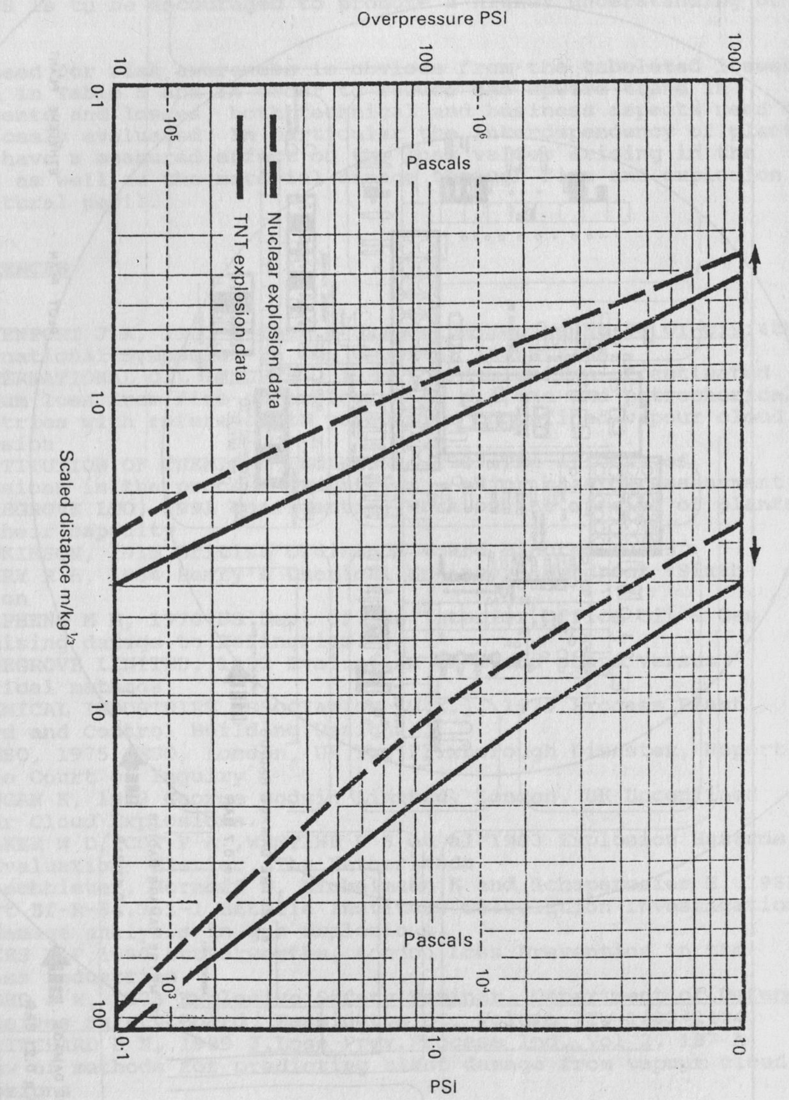

FIGURE 2. This illustration is reproduced from reference 3 by kind permission of the Institution of Chemical Engineers.

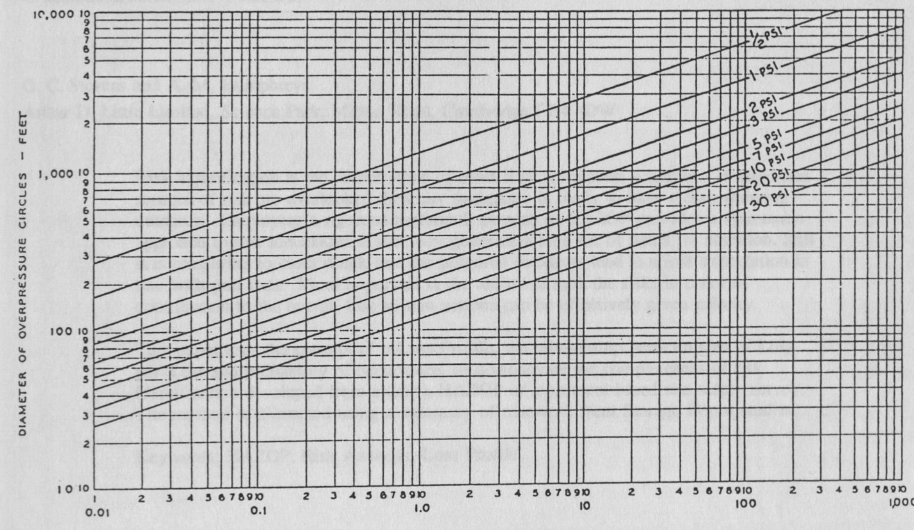

FIG. DIAMETER OF OVERPRESSURE CIRCLES FOR VARIOUS EXPLOSIVE YIELDS

FIGURE 3. This illustration is reproduced from reference 7 by kind permission of the Bureau of Mines.