GUIDELINES FOR EVALUATING THE EFFECTS OF VAPOR CLOUD EXPLOSIONS USING A TNT EQUIVALENCY METHOD

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1.0 SCOPE

This document (formerly Data Sheet 7-0S) provides technical guidance to evaluate the physical property damage consequences to buildings and process structures from explosion overpressures and ensuing fires, caused by the outdoor release and delayed ignition of a flammable vapor cloud.

This guideline uses a TNT equivalency methodology to evaluate approximate effects of a worst credible case vapor cloud explosion (VCE). Other refined VCE prediction or estimation methods, while briefly discussed, are beyond the scope of this document.

The techniques and procedures described in the guidelines are a simplified approach to a complex problem. Application is not appropriate for non-chemical plants that may have storage or use of hazardous materials, or for smaller chemical plants that lack the congestion or confined layout needed to produce outdoor VCEs. Loss history supports that VCEs have occurred primarily in large petrochemical or refinery facilities and/or as a result of transportation incidents and that they have occurred on a very low frequency as compared to other major events.

A methodology to determine or factor in the probability of a given VCE event is also beyond the scope of this document. Because probability is not considered, this guideline does not predict overall risk from a VCE.

1.1 Changes

May 2008. Minor editorial changes were made.

2.0 VAPOR CLOUD EXPLOSION THEORY

2.1 Definition

A vapor cloud explosion is defined as an explosion occurring outdoors which produces damaging overpressure. It is initiated by the unplanned release of a large quantity of flammable vaporizing liquid or high pressure gas from a storage tank or system, process vessel, pipeline, or transportation vessel.

2.2 Theory Principles

Generally speaking, for a VCE with damaging overpressure development to occur, several factors must be present. First, the material released must be flammable and processed or held under suitable conditions of pressure or temperature. Examples of such materials are liquefied gases under pressure (e.g., propane, butane,); ordinary flammable liquids at high temperatures and/or pressures (e.g., cyclohexane, naptha); and non-liquified reactive flammable gases (e.g., ethylene, acetylene).

A cloud of substantial size also must form prior to ignition. With most common flammable materials, should ignition occur instantly with release of the material, a large vapor cloud fire may occur, causing extensive localized heat radiation damage; however, significant blast pressures causing widespread damage will likely not occur. (**Exception:** some highly reactive materials, such as ethylene oxide under some conditions, might produce overpressures even with immediate ignition.) Should the cloud be allowed to form over a period of time within a process area and subsequently ignite, blast overpressures away from the cloud center can equal or even exceed those developed from detonation of high explosives and result in extensive damage over a wide area. Ignition delays of from one to five minutes are considered most probable, although major incidents with ignition delays as low as a few seconds and higher than 60 minutes have occurred.

Third, a sufficient amount of the cloud must be within the flammable range of the material to cause extensive overpressure. The percent of the vapor cloud in each region varies, depending upon many factors including type and amount of the material released, pressure at release, size of release opening, direction of release, degree of outdoor confinement of the cloud, and wind speed, atmospheric stability and other environmental effects. The cloud will move over time, changing the flammable regions. For example, a continuous release over a long period of time will generally have a rich region near the source, a lean region at the cloud leading edge, and a flammable region in between. A puff release (essentially instantaneous release) will usually have a rich region at the leading edge with flammable regions following.

Important factors that must be present for an ignited vapor cloud to produce overpressure are outdoor confinement and turbulence generation. Research testing, incident investigation, and computer modeling have demonstrated that the greater the horizontal and vertical confinement and the more turbulence in the gas cloud, the greater the potential for overpressure development. Turbulence can be caused by two primary

mechanisms. First, repeated obstacles in the center of a cloud can accelerate gas mixing (due to eddy and shear layer effects), which in turn can increase flame speeds within the cloud. This highly influences pressure development due to flame instabilities. Second, turbulence can be directly initiated from a high pressure release.

Because obstacles cause turbulence, plant layout plays a principal role in the ability of a released vapor cloud to burn as a flash fire with only radiant heat effects or transit to an explosion with overpressure effects as well as radiant heat effects. The amount of congestion, confinement, the horizontal and vertical spatial arrangement of process structures, and obstacles in the flow path of a cloud, are all important in determining if an ignited cloud of vapor will transit from a fire to an explosion. A plant with mostly open space and minimal or small process structures or partially covered process areas tends not to produce the right conditions for VCE enhancement. This might be typical of a commercial gas distribution or aerosol filling plant where bulk flammable gases may be repackaged or used in small quantities. These facilities would typically involve a small tank farm with minimal piping and spatial congestion. Such plants have a high ratio of open area or enclosed buildings to congested process areas. That is, they could be considered an "open" array. Even though suitable amounts and types of flammable materials may be present and a vapor cloud release could occur, the plant layout would likely not generate the turbulence that contributes to overpressure. If however, the spaces within the gas filling plant or the spaces between tanks or roadways were essentially covered with canopies or dense pipe racks, the plant layout would more closely represent the conditions needed for an explosion.

On the other end of the continuum are highly congested process areas within oil refineries or petrochemical plants. These types of plants have historically been involved in VCE events, and present ideal conditions for turbulence generation due to obstacle induced turbulence. They have large quantities of flammable materials under extreme temperature and pressure conditions. They feature highly congested multi-level open or partially enclosed process structures which include numerous pipe racks, process vessels, and long narrow pump alleyways. The structures may be hundreds of feet (or meters) long and may only be separated from adjacent process units by narrow roadways. Roadways in some cases are also covered by dense multi-level pipe racks, making very large, continuous, partially confined and congested process areas.

Wide open spaces between well separated process areas or around remote tanks farms do not easily promote VCE events unless the area presents unusual conditions of confinement (such as long, narrow ravines) or repeated obstacles (such as dense forests or large railroad staging yards). However, a cloud released in an open area may be of sufficient size and winds may be of suitable velocity and direction to disperse the cloud into a congested process area at great distances from the actual release. While the ignition of the cloud could occur anywhere in the cloud (even in the open space, for example, by a vehicle), the apparent explosion epicenter will be the area where the cloud is confined and where obstacles exist that can cause transition from a cloud fire to a cloud explosion. Remaining portions of the cloud outside the congested area will not contribute to blast effects, although radiant heat effects will occur.

There is no quantitative method to accurately determine if a facility is VCE prone based solely on layout and congestion. While loss history supports that VCE events are more likely in a traditional refinery or petrochemical plant, a VCE exposure cannot be completely ruled out in smaller specialty plants or remote storage terminals.

Ensuing fires and vessel explosions, such as Boiling Liquid Expanding Vapor Explosions (BLEVEs), may produce significant additional damage to a process area subsequent to the pressure effects and radiant heat from the initiating VCE. This is especially true for facilities with large volumes of flammable liquids which may be released over a large area simultaneously with impairment of water supplies and fixed suppression systems. The presence or lack of passive protection such as drainage and steel fireproofing as well as plant spacing, play key roles in the amount of damage occurring in such facilities.

Conversely, plants that utilize primarily gas products, with minimal or negligible amounts of flammable liquids, are not likely to incur as much ensuing damage following a VCE event.

The conditions necessary to produce a VCE are fairly well understood, and a number of calculation methods are available to convert the VCE scenario into an assessment of damage effects. Most calculation techniques include methods for determining amount of material release, cloud size and energy release upon ignition. The energy of the material released is often converted into TNT equivalency (following principles of the so-called Ideal Blast Wave methodology), by assigning an *explosion efficiency number*. Also referred to as explosive yield, explosion efficiency is an estimation of the explosive effect of the mass in the cloud relative

to an equivalent mass of TNT. Once a TNT equivalency is determined, published test data is used to calculate blast overpressures.

Other methods have been developed and published to evaluate VCE overpressure effects. Section 2.3 abstracts several of the more well known methods, including the TNT equivalency method.

Extensive guidelines on pressure resistance of certain structures and process equipment have also been published. From these guidelines, damage occurring from a VCE and ensuing fires can be calculated.

Calculation of vapor cloud releases and consequences can be done by using one of many computer models available. The primary feature of the more sophisticated models is their ability to more accurately calculate source term release, cloud dispersion and drift, primarily because they factor in wind speed and atmospheric stability conditions. Many, however, apply a TNT Equivalency method for determining energy release, which requires the practitioner to determine a credible release scenario and assign variables such as explosion efficiency and blast epicenter.

2.3 VCE Overpressure Predictive Methods

A number of methods to help predict the consequences of an outdoor VCE have developed out of need. The TNT equivalency method is by far the most popular due to its simplicity for approximate property damage prediction. Other methods may have utility for plant siting and design or for refined property damage estimates. A TNT Equivalency method is described in detail in Section 3.0. Other than the following abstracts, specific details on how to apply other methods are not descibed further in this document.

2.3.1 Ideal Blast Wave (TNT Equivalency) Method

The use of a TNT equivalency methodology has developed over a many year period due to research in high explosives conducted by various military agencies worldwide. While the method has drawbacks (described below), it is by far the most common and easily applied approach for approximating accidental explosions, whether outdoor confined gas explosions, vessel ruptures, or other release of stored energy. Principle reasons for this method's popularity have been its simplicity and the amount of data available from testing.

In the TNT equivalency method, a mass of flammable gas is related to an equivalent mass of TNT by comparing heats of combustion and assigning a so-called efficiency factor. The efficiency factor is also known as *explosive yield, explosion equivalancy,* or *yield factor*. Because most common flammable hydrocarbon materials have a heat of combustion of about 20,000 Btu/lb (11,100 kcal/kg) as compared to TNT's heat of decomposition of 2000 Btu/lb (1111 kcal/kg), one can readily see that, at 100% yield, flammable hydrocarbons possess about 10 times the energy per unit weight than TNT. Therefore a 10% yield of a typical hydrocarbon would equal TNT in energy release. However, combustion of flammable gases, especially in non-laboratory conditions, is rarely ideal, and 100% yield cannot be achieved. The actual efficiency of combustion can only be assigned based on the practitioner's best estimate. This is the primary drawback to this methodology.

Because of the need to assign an explosive yield, however, much study has been undertaken by researchers. In addition, great effort has been expended to determine explosive yield in actual incidents. While rarely can a single yield be assigned following an incident (usually due to extreme variance in blast waves and damage patterns), historical evidence has suggested that yields in the 1 to 5% range are appropriate for typical hydrocarbons. However, yields as high as 50% have been recorded, although most have fallen in a much lower range. In fact very low estimated yields—as low as 0.1%—have caused extensive damage, primarily because of congestion and concentrated flammable liquid holdup.

Explosive yield is known to be affected by a number of variables, including the material released; the temperature and pressure of release; the nature of the release (e.g., jet, turbulent, direction); atmospheric conditions (especially wind speed and atmospheric stability); ignition strength; and, most importantly, plant layout. Plant layout—the presence of arrayed obstacles, congestion, and partial confinement in a process area—can influence explosive yield substantially. A release into an open area generally creates very low burning efficiencies and produces low order explosions with little or no significant overpressure generation. Conversely, a release into a highly congested process area, (or a forest with dense trees or a railyard with staged railcars), can create high yield with substantial overpressures.

It cannot be overemphasized that assigning of an explosion efficiency number to a potential gas release incident is, with current technology, an entirely arbitrary exercise. The higher the yield, the higher the TNT equivalency (and usually the greater the property damage), and conversely, the lower the yield, the lower the equivalency. Most organizations, including FM Global, assign fixed yields based on historical evidence and

the degree of conservatism desired in final results. FM Global explosion efficiencies—which are used to define a **worst credible case event** VCE rather than a **normal credible case event** VCE—vary by type of material, and are listed in Section 3.4.3. See Section 3.0 for further discussion and definition of the worst and normal case scenarios.

TNT equivalancy methods have other drawbacks which tend to oversimplify results. A TNT explosion is a point source explosion, while a VCE involves a large mass of material spread over a large area. A VCE does not react like a point source explosion. Because a vapor cloud can not only spread over multiple process areas, but also fill void space between process areas, there can be a great mass of material that does not contribute to overpressure generation. Further, a VCE is normally a deflagration rather than a detonation, which is characteristic of a TNT explosion. An outdoor deflagration produces much lower overpressures (usually less than 1 barg [14.5 psig]) than a detonation (10 to 60 barg [150 to 1000 psig]). There are also significant differences in shape of blast wave between VCEs and TNT detonations. A TNT blast produces a sharp rise in overpressure and has a very short duration (impulse time), while a VCE produces a more gradual rise in pressure but commonly over a longer impulse period. While a lower overpressure is generated, it has a longer decay period than TNT. VCEs thus tend to produce greater overpressures at far field as compared to TNT. In the near field (few hundreds of feet from epicenter) TNT explosions produce more pressure and damage.

Even though TNT equivalancy methods have drawbacks, they remain the preferred method when the practitioner is seeking an *approximate* evaluation of property damage or wishes to screen many scenarios for plant siting or design. Field studies by FM Global[5] have shown that significant differences in yield or release estimation usually result in much less significant differences in predicted property damage, due to the cube root scaling law (Hopkinson's Scaling Law) for overpressures. When the purpose of estimating VCE effects is very refined (instead of approximate) property damage assessment or when very specific design basis building siting is the objective, a generalized TNT equivalancy approach may be inappropriate.

2.3.2 TNO Multi-Energy Model*

This abstract of the TNO Multi-Energy Model (MEM) is not intended to provide sufficient information to use the MEM. Numerous publications are available to aid the practitioner should this method be chosen (refer to the bibliography in Section 6.0).

The MEM is based on the assumption that there is little, if any, correlation between VCE and TNT explosions. Under contract to major petrochemical, refining, and governmental organizations, TNO has conducted small and full scale research to attempt to better quantify the effect of obstacles and spacial confinement on overpressure generation in a VCE. Their work has been widely published and is accepted by many practitioners, especially European government authorities, as state of the art in refined VCE estimation. It is primarily applied to predict effects off-site and for designing and siting new construction. It may have applicability for refined property damage assessment.

The MEM does not primarily determine explosion hazard by the fuel/air mixture itself; rather, it strongly depends on the plant environment into which the cloud has been dispersed. There are several major assumptions and steps in the procedure:

    a. It is assumed a deflagration rather than a detonation will occur.

    b. Potential sources (generators) of blast are determined. That is, it is assumed that only certain spatial configurations such as multilevel process structures, spaces between parallel planes (e.g., alleyways between buildings or storage tanks), closely spaced railcars or automobiles, tunnels and culverts, etc., can produce blast. Only that part of the cloud within one or more of these spaces can generate blast. Remaining portions of the cloud will not contribute to blast.

    c. An estimate of the scale of generated blast is made for each spatial area by calculation of heat of combustion using only the amount of fuel within that spatial area. A stoichimetric mixture is assumed. This step defines charge size and blast dimensions.

    d. An estimate for blast strength (stored energy) from each spatial area is made using published tables and graphs. Scaled distance for overpressures generated are taken from a published graph. Respective blasts from adjacent spatial areas are analyzed separately but overpressure profiles can be superimposed. A strength-of-blast charge value of 1 on the TNO curve represents a weak explosion in an unobstructed or unconfined cloud. A value of 7 represents a strong deflagration in a highly congested, elongated structure. The highest value of 10 represents a VCE detonation.

The principal positive feature of MEM is the use of portions of the cloud that fall only within a congested process area.

The MEM is more difficult to use than a TNT equivalency method because of the need to estimate a volume of a spatial process area and to choose an arbitrary strength-of-blast charge number. The MEM procedure is currently oversophisticated for *approximate* property damage estimation, but may be appropriate for *refined* property damage assessments and for process plant design and siting purposes. Refinements to the MEM are currently ongoing to improve usability.

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2.3.3 TNO "Effects" Model

TNO, in 1979 (revised in 1992), developed what is commonly known as the Yellow Book. Officially entitled *Methods for the Calculation of Physical Effects*, the Yellow Book was developed for the Netherlands government Committee for the Prevention of Disasters. The Yellow Book protocol has been well accepted throughout Europe, (and to a lesser extent in North America), by government agencies that regulate the chemical industry and by the chemical industry itself.

In the Yellow Book, in addition to models for material release, evaporation, dispersion, etc, is a VCE energy prediction method based on fuel air charge rather than TNT equivalency. It is known as the *TNO Effects Model* and can be described as a spherical expanding piston model. A piston model allows for a variable initial strength of the blast. It models the explosion based on two considerations: *scale*, as determined by the amount of combustion energy involved; and *initial strength*, as determined by combustion rate in the explosion process. Blast scale is determined by use of dispersion calculations to estimate fuel quantity within the flammable limits present in the cloud. Initial blast strength is determined assigning a flame speed which is defined by the magnitude of turbulence (mixing) and by the reactivity of the released material.

The most common fuels are classified according their reactivity. The low reactivity group (Class III) includes ammonia, methane and natural gas. Moderately reactive materials —Class II—include propane, ethylene and propylene. Highly reactive—Class I—materials include acetylene, ethylene oxide and hydrogen. Scaled charts are used to develop blast rings. Flame speed values are chosen by the practitioner using guidance given in the procedure.

The TNO Effects piston model does not quantitatively factor in confinement, congestion, and obstructions in the cloud as does the newer and more refined TNO MEM method.

The TNO Effects model produces blast rings approximately 30% smaller than a TNT equivalency model given the same material release amount. It has good applicability for design and siting of new processes and is accepted as meeting European guidelines such as European Directive 82/501/EEC of June 24, 1982 (the Seveso Directive).

2.3.4 Other Predictive Methods

A number of other models, such as the *Baker-Strehlow Method*, have been developed based on fuel air charge and flame speed. Explosively dispersed explosions and exploding jet models have also been developed. These models have been effective in predicting blasts for designing and siting buildings but do not lend themselves to approximate property damage assessments.

3.0 PREDICTING VCE EFFECTS USING A TNT EQUIVALENCY METHOD

This guideline uses a TNT equivalent energy method to predict damage. Each step is covered in detail later in this document. The procedure allows considerable room for judgment. Two decision trees, Figures 1 and 2, are provided to further guide the practitioner through seven steps, which are outlined below.

The use of any TNT Equivalency model (simple or sophisticated) has limited application and is subject to the following considerations:

   a. Only *approximate* consequence effects can be determined.

   b. Variables within the equations must be assigned by the practitioner. Because these variables are primarily based on historical evidence (backed by limited research testing), they are arbitrary and normally assigned based on the intended use of the results by the practitioner.
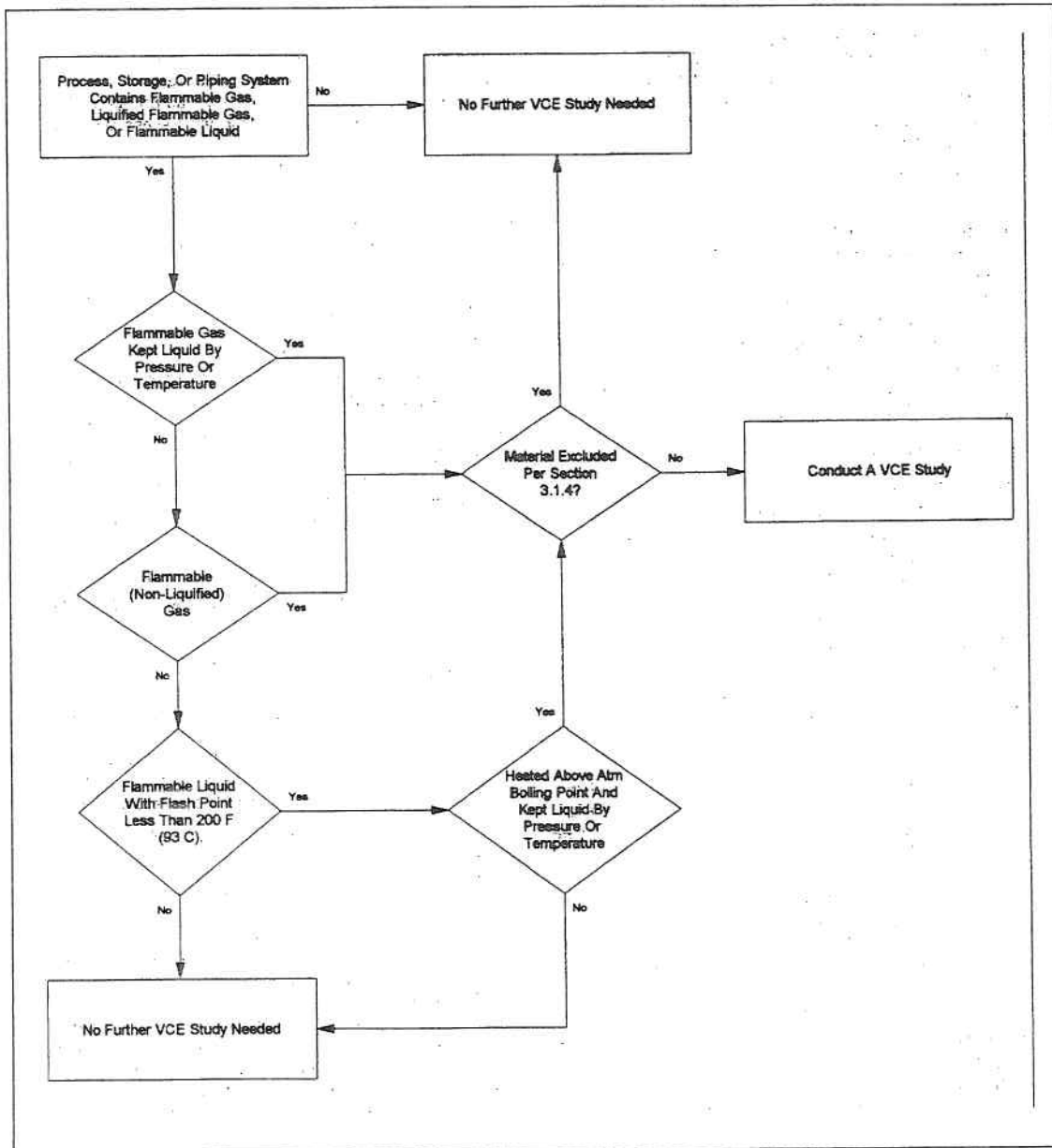
Fig. 1. Preliminary decision tree to determine need to conduct VCE analysis.

    c. Frequency or probability of a given scenario are not predicted, unless the scenario was chosen based on a quantitative risk assessment.

The effects which the FM Global procedure (outlined in Section 3.0 of this document) predicts can be considered conservative. The predicted effects are influenced highly by assumed variables, such as type, location and duration of release; explosion efficiency; and choice of blast epicenter. FM Global has preassigned values to these variables, based on careful review of historical incidents and research literature. The FM Global procedure also predicts major events—**worst credible cases**—rather than smaller events, or **normal credible cases** because the primary use of this procedure is *to identify the largest probable property damage exposures on a given site.*

While definitions of levels of severity vary, **worst credible case** within this document is defined as a scenario of significant events in combination that produce a major, yet foreseeable incident. These types of events may also be called *worst probable* or *worst foreseeable* case events by other organizations. This level of
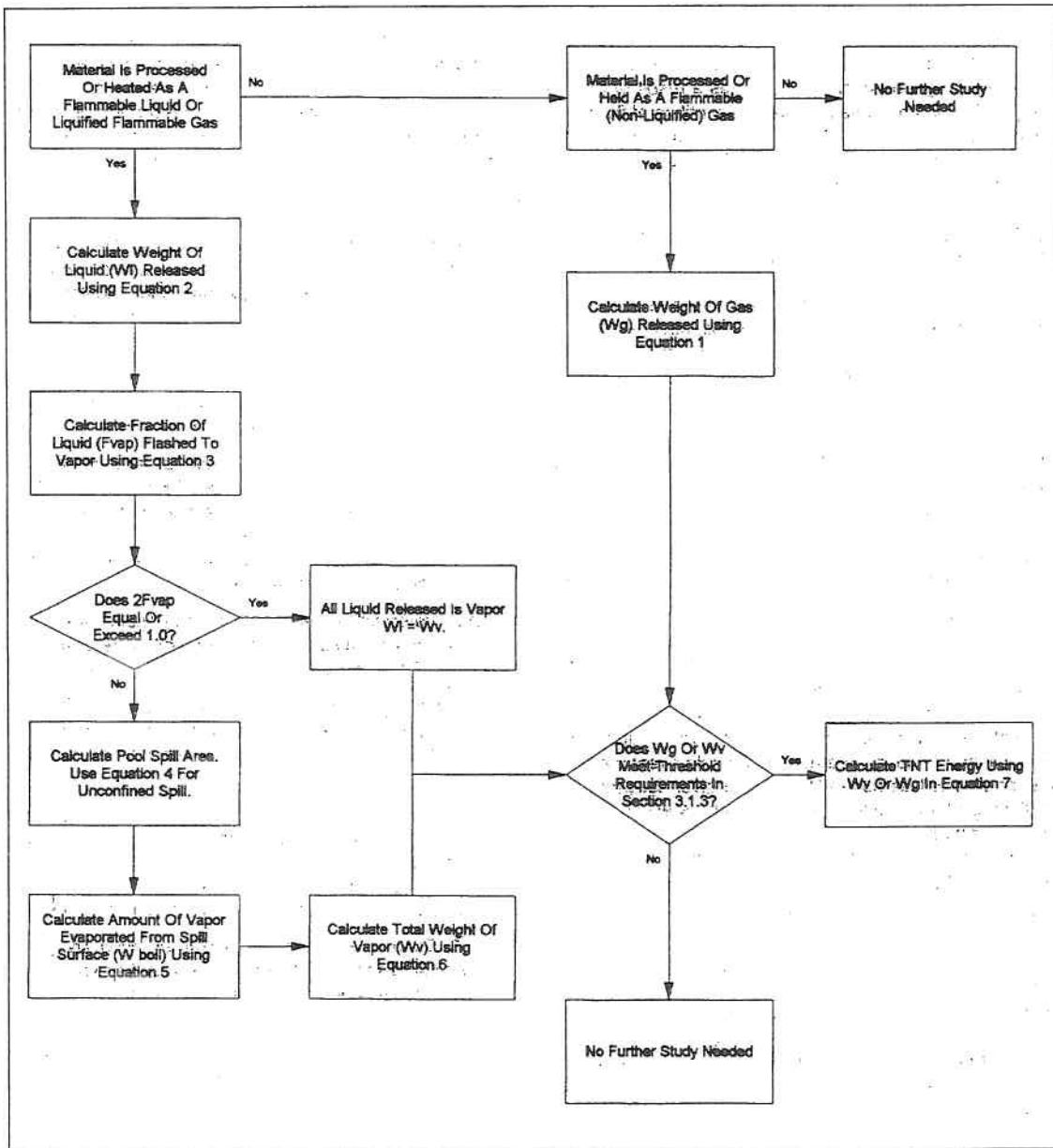
Fig. 2. Material release decision tree.

incident generally assumes large releases which are usually not mitigated by active systems or human intervention.

For example, sudden and catastrophic rupture of a code designed vessel is not considered a credible event, whereas major failure of the largest or highest pressure pipe leading from the same vessel is considered a credible event. Certain ensuing events caused by or as a result of the initating VCE, such as release of large quantities of flammable materials simultaneously with widespread impairments to fixed suppression systems, are considered credible, whereas a series of cascading VCEs spawned by the initiating VCE are not. Passive design features such as fireproofing, containment systems, drainage systems, plant and process area spacing, inherent process design, passive flow limiting or restricting devices, and damage limiting construction, are given credit to minimize the release or the effects of ensuing events. Active systems, such as detection, waterspray systems, automatic or manual block valves, active flow limiting devices, and human intervention, are given little credit.

A *worst credible event* also assumes atmospheric conditions that produce conservative source term release and dispersion results. For example, even though prevailing winds at a facility indicate a high chance of cloud dispersion (drift) away from important process areas, a worst credible case would assume drift into the process area because of its importance, even though this may not be the likely event.

A *normal credible event* is generally a smaller and more likely event than a worst credible case event. The release would more likely be a smaller pipe, such as a routinely opened sample line or a pump seal failure. A normal credible case event assumes both passive and active protection systems function as designed, with some limitations. The duration of release in a normal credible event may also be substantially less as credit might be given for proper functioning of flow limiting devices, such as excess flow valves, and human intervention shortly after detection. A normal credible event might be used during plant design to site or specially design important plant buildings.

These guidelines do not assume or assign frequency (probability) for a given event, small or major. Normal credible (typically smaller consequence) events may occur on a higher frequency than worst credible (typically larger consequence) events. For example, logic would support that a release from a high temperature/pressure process system is more likely than a release from a quiescent, ambient condition storage system because of the more harsh operating conditions of the former. This is supported by loss history. However, the actual likelihood of one release over another is very site and process dependent, and depends principally on local conditions of process design and process control, human element, and process safety management programs in addition to the inherent exposure from the process or storage system itself. These variables can only be completely understood and a true risk assigned by completion of a comprehensive quantitative risk assessment which measures frequency versus consequences.

A TNT equivalency or other VCE model can be used to predict *normal credible* events, however, different variables, such as explosive yield and material release parameters, must be selected based on the practitioner's perception of the hazard and consequence level of the lesser event. That is, credit might be given to operation of process block valves or human intervention; smaller, more susceptible pipes might be chosen as the release point; a shorter duration release might be assumed; and a lesser explosive yield applied. New variables should be assigned, however, based on a full understanding of the intended goal of the study and a full knowledge of the practitioner's defined level of normal credible event.

A more sophisticated computer model may also be substituted for any part of this procedure. A computer model may be especially beneficial in predicting and calculating complicated source term release—such as two phase flashing flow, which is not featured in the equations in Section 3.3—and cloud dispersion.

There are **seven steps** used to evaluate VCE exposure using the FM Global TNT Equivalency method:

Step 1. Determine a need to conduct a vapor cloud explosion study. The need to conduct a VCE study is predicated on having suitable materials and process conditions that can lead to a credible release. These criteria are further described in Section 3.1. Figure 1 will also aid in determining the need to proceed with a VCE study.

Step 2. Determine a material release scenario based on assumptions of release mode, duration of release, and scenario. Refer to Section 3.2.

Step 3. Determine amount of material released. The approximate amount of material released in a given period is converted into mass of vapor by simple calculations in Section 3.3.

Step 4. Determine energy released based on TNT equivalency. Knowing the heat of combustion, the mass of vapor released can be converted to equivalent mass of a TNT charge using a simple equation. Assumptions on explosion efficiency are made in this step. Refer to Section 3.4.

Step 5. Determine overpressure radii at given TNT energy release using scaled distance charts based on vapor cloud incidents, military testing and other research. Refer to Section 3.5.

Step 6. Determine epicenter of explosion. This is an assumption based on worst credible case scenario and guidelines given for various size releases. Refer to Section 3.6.

Step 7. Estimate effects of explosion. See Section 3.7. Estimation of property damage effects to buildings and processing structures is based on overpressure exposure and strength of structure based on published charts. The potential for ensuing fires and explosion is included in the assumptions made in this step.

3.1 Step 1—Determine a Need to Conduct a Vapor Cloud Analysis

The need to conduct a detailed VCE analysis should be evaluated based on the types and quantities of materials in the plant. Refer to Figure 2, *Material release decision tree.*

There are numerous flammable materials in chemical processing facilities which present VCE potentials. Examples of released materials causing known VCEs are: ethylene, butane, propane, butadiene, vinyl chloride monomer, cyclohexane, propylene, ethylene oxide, dimethyl ether and isobutane.

Materials subject to release may be held in storage systems, transported in pipelines, mobile transportation vessels—(both land and water), or be part of process systems. They may be used as feedstock, as intermediates, final product, or fuels for fired equipment and other process equipment such as refrigeration systems. They may be only stored on site and distributed to other facilities.

Some materials, while flammable and representing in some cases severe fire or confined vapor explosion potentials, are not considered to present outdoor VCE exposures as noted in Section 3.1.4.

3.1.1 Materials Subject to VCE

The following materials should be considered to present a VCE exposure unless excluded in Section 3.1.4:

    a. flammable gases maintained as a liquid by refrigeration or pressure;

    b. flammable gases;

    c. flammable liquids with flashpoints below 200°F (93°C) processed above their atmospheric boiling point *and* maintained as a liquid by pressure.

3.1.2 Material Classification

A list of common materials known or suspected to create outdoor VCE potentials is presented in Table 1. This should not be considered a complete list of potential VCE producing materials, however. Other materials may present considerable risk and may need evaluation. Materials in Table 1 are categorized by class based on reactivity. Class I materials, such as propane and flammable liquids, present normal risks. Class II materials such as ethylene are more reactive and present greater risks than Class I materials. Class III materials such as acetylene are highly reactive and may be unstable. Class III materials present the greatest energy potential per unit weight of the three classes.

Table 1. Properties and Classification of Some Potential VCE Prone Materials

Material	Mat'l Class	Molec. Wt.	Net Heat of Combustion ΔH_c		Flash Pt		Boiling Pt		Gas Const K	Heat of Vaporiz ΔH_{vap}	
			Btu/lb	Kcal/kg	°F	°C	°F	°C		BTU/lb	J/kg (x10⁵)
Acetaldehyde	I	44	10,500	5,833	−40	−40	70	21	0.64	245	5.69
Acetone	I	58	12,300	6,833	−4	−20	134	57	0.64	223	5.19
Acetylene	III	26	20,700	11,500	0	−18	−119	−84	0.66	270	6.28
Acrolein	II	56	11,800	6,555	−15	−26	125	52	use 0.68	312	7.24
Acrylonitrile	I	53	13,700	7,611	32	0	171	77	use 0.68	265	6.16 (25°C)
n-Amyl Acetate	I	130	14,400	8,000	77	25	300	149	use 0.68	not available	
Amyl Alcohol	I	88	15,000	8,333	91	33	280	138	use 0.68	211	4.90
Benzene	I	78	17,300	9,611	12	−11	176	80	0.63	169	3.93
1,3-Butadiene	I	54	19,200	10,700	Gas		24	−4	0.63	176	4.10
n-Butane	I	58	19,700	10,940	Gas		31	−1	0.63	167	3.89
1-Butene	I	56	19,500	10,833	Gas		21	−6	0.63	171	3.97
n-Butyl Acetate	I	116	12,200	6,777	72	22	260	127	use 0.68	133	3.10
Carbon Disulfide	II	76	6,100	3,388	−22	−30	115	46	0.72	151	3.51
Cyanogen	I	52	5,000	2,777	Gas		−6	−21	0.66	185	4.31 (0°C)
Cyclohexane	I	84	18,700	10,388	−4	−20	179	81	0.62	155	3.60
n-Decane	I	142	19,000	10,555	114	46	345	174	0.62	113	2.64
o-Dichlorobenzene	I	147	8,100	4,500	150	66	358	181	use 0.68	not available	
Diethyl Ether	II	74	14,500	8,055	−49	−45	95	35	0.62	162	3.77
Divinyl Ether	II	70	14,500	8,055	−22	−30	102	39	use 0.68	not available	
Ethane	I	30	20,400	11,333	Gas		−128	−89	0.65	209	4.94
Ethyl Acetate	I	88	10,100	5,610	24	−4	171	77	use 0.68	184	4.27 (0°C)
Ethyl Alcohol	I	46	11,500	6,388	55	13	173	78	0.63	358	8.33
Ethylamine	I	45	16,300	9,055	0	−18	62	17	use 0.68	263	6.11
Ethyl Benzene	I	106	17,600	9,777	59	15	277	136	0.63	146	3.39
Ethyl Chloride	I	65	8,200	4,555	−58	−50	54	13	use 0.68	166	3.85
Ethyl Cyclohexane	I	112	18,800	10,444	95	35	269	132	use 0.68	133	3.10
Ethylene	II	28	20,300	11,278	Gas		−155	−104	0.66	206	4.77
Ethylene Dichloride	I	99	4,600	2,555	56	13	183	84	use 0.68	140	3.26
Ethylene Oxide	III	44	11,700	6,500	0	−18	51	11	0.65	245	5.69
Ethyl Formate	I	74	8,700	4,833	−4	−20	130	55	use 0.68	175	4.06
Ethylidene Chloride	I	99	4,400	2,444	22	−6	136	58	use 0.68	121	2.80
Ethyl Nitrate	III	91	6,400	3,555	50	10	190	88	use 0.68	not available	
Ethyl Nitrite	II	75	8,400	4,666	−31	−35	61	16	use 0.68	not available	
Ethyl Propionate	I	102	15,200	8,444	54	12	210	99	use 0.68	144	3.35
n-Heptane	I	100	19,200	10,700	25	−4	209	98	0.62	135	3.14
n-Hexane	I	86	19,200	10,700	−7	−22	156	69	0.62	142	3.31
Hydrazine	III	32	7,200	4,000	100	38	236	113	use 0.68	not available	
Hydrogen	I	2	51,600	28,666	Gas		−422	−252	0.68	194	4.52
Hydrogen Cyanide	I	27	10,300	5,722	0	−18	79	26	0.67	378	8.79
Hydrogen Sulfide	I	34	6,500	3,611	Gas		−76	−60	0.67	236	5.48
Isobutyl Alcohol	I	74	14,200	7,888	82	28	225	107	use 0.68	239	5.56
Isobutane	I	58	19,600	10,904	Gas		11	−12	use 0.68	157	3.65
Isobutylene	I	56	19,400	10,777	Gas		20	−6	0.63	169	3.93
Iso-octane	I	114	19,100	10,611	10	−12	211	99	use 0.68	115	2.68
Isopropanol	I	60	13,100	7,278	53	12	181	83	0.63	288	6.69
Isopropyl Nitrate	III	105	no data	no data	no data		216	102	use 0.68	not available	
Methanol	I	32	8,600	4,778	52	11	147	64	0.62	475	11.0
Methyl Acetate	I	74	8,500	4,722	14	−10	140	60	0.64	176	4.10
Methyl Acetylene	III	40	20,000	11,111	Gas		−10	−23	use 0.68	198	4.60
Methyl Amine	I	31	13,200	7,333	Gas		21	−6	use 0.68	354	8.21
Methyl Butyl Ketone	I	100	12,000	6,666	95	35	262	128	use 0.68	148	3.43
Methyl Ether	I	46	12,400	6,888	−42	−41	−11	−24	0.63	202	4.69
Methyl Ethyl Ketone	I	72	13,500	7,500	16	−9	176	80	use 0.68	191	4.44

Material	Mat'l Class	Molec. Wt.	Net Heat of Combustion ΔH_c		Flash Pt		Boiling Pt		Gas Const K	Heat of Vaporiz ΔH_vap	
			Btu/lb	Kcal/kg	°F	°C	°F	°C		BTU/lb	J/kg (x10⁵)
Methyl Formate	I	60	6,400	3,555	-2	-19	90	32	use 0.68	202	4.69
Methyl Mercaptan	I	48	10,000	5,555	0	-18	42	6	use 0.68	not available	
Methyl Propyl Ketone	I	86	14,300	7,944	45	7	215	102	use 0.68	not available	
Monochlorobenzene	I	112	11,300	6,277	84	29	270	132	use 0.68	not available	
Nitromethane	III	61	5,000	2,777	95	35	215	101	use 0.68	243	5.65
n-Pentane	I	72	19,500	10,833	-40	-40	97	36	0.62	148	3.43
Petroleum Ether	I	Varies	21,300	11,833	-57	-49	37	3	use 0.68	not available	
Propane	I	44	19,900	11,055	Gas		-44	-42	0.63	183	4.27
Propionaldehyde	I	58	12,500	6,944	-22	-30	120	49	use 0.68	206	4.77
n-Propyl Acetate	I	102	11,200	6,222	55	13	215	102	use 0.68	144	3.35
Propyl Alcohol	I	60	13,200	7,333	77	25	207	97	use 0.68	295	6.86
Propylene	I	42	19,700	10,944	Gas		-53	-47	0.64	188	4.35
Propylene Dichloride	I	113	6,300	3,500	60	16	205	96	use 0.68	121	2.80
Propylene Oxide	II	58	13,200	7,333	-35	-37	95	35	0.64	205	4.77
Styrene Monomer	I	104	17,400	9,600	90	32	295	145	0.63	155	3.60
Toluene	I	92	17,400	9,666	40	4	231	111	0.63	157	3.64
Vinyl Acetate	I	86	9,700	5,388	18	-8	161	72	use 0.68	148	3.43
Vinyl Acetylene	III	52	19,500	10,833	Gas		42	5	use 0.68	197	4.57
Vinyl Chloride	I	63	8,000	4,444	Gas		7	-14	use 0.68	142	3.31
Vinylidene Chloride	I	97	4,200	2,333	0	-18	99	32	use 0.68	130	3.01
Vinylidene Fluoride	I	64	7,900	4,388	Gas		-122	-86	use 0.68	105	2.45
p-Xylene	I	106	17,600	9,777	81	27	280	138	0.63	146	3.39

3.1.3 Threshold amounts

Incident history demonstates that most damaging VCEs have been the result of significant material releases. Threshold amounts of material subject to release are assigned by most practitioners of VCE analysis to ensure that only credible events are considered. However, just because threshold limits are set does not imply that a VCE cannot occur from smaller releases. Small release VCEs have occurred. When applying a *worst credible case* scenario, however, a VCE study should only be made when W_v or W_g (weight of vapor or gas calculated in Step 3) is:

   a. 5 tons (4.5 tonnes) or greater for Class I materials;

   b. 1 ton (0.9 tonne) or greater for Class II materials;

   c. 1000 lbs (454 kg) or greater for Class III materials.

3.1.4 Excluded Materials

FM Global has interpreted that the following materials do not present a significant or credible outdoor VCE exposure. These exclusions are based on many factors such as heats of combustion, fundamental burning velocities, research testing, ease of dispersal, other experts' opinions and loss history.

3.1.4.1 *Liquified Natural Gas (LNG)* and *Natural Gas (NG) (methane)*, when the ethane component is less than 15% by volume. If NG or LNG has an ethane component in excess of 15%, it may be susceptible to VCE with overpressure development. However, with 85% or greater methane concentration, an explosion outdoors with damaging overpressures is not considered likely. Normal NG or LNG for fuel use is composed of 92–94% methane, 3–4% ethane and other hydrocarbons, and 3% nitrogen. While this varies by region, supplier and time of year, rarely would ethane content exceed 5% due to cost and value of ethane for other uses. The only condition where NG or LNG might have high ethane content would be where it is used as a pure feedstock in chemical processing. For all practical purposes, the ethane content of NG and LNG in pipelines and storage systems will never exceed 5%, and will normally be much lower.

The validity of excluding NG and LNG is supported by testing by the Institution of Gas Engineers (Great Britain)* in small- and full-scale obstacle arrays simulating natural gas processing units. While it was concluded that NG with high methane content could produce moderate overpressures if released into process arrays of extremely close packed obstacles, it was further stated that the tests "demonstrated that both the probability of a vapor cloud explosion occuring and its consequences will be lower for natural gas than with other common hydrocarbons." Further there have been no reported VCEs involving NG or LNG with high methane content, even though numerous ignited vapor releases resulting in flash fires without overpressures have occurred worldwide. Coupling testing and loss history with the low reactivity, the lightness of the vapor, and the relatively low flame speeds of NG and LNG, a VCE involving these materials is considered beyond the scope of a worst credible case scenario.

3.1.4.2 *Ammonia gas* can and has exploded when substantially confined inside equipment or buildings. There is no indication, based on loss history, testing or combustion features, that ammonia will produce overpressures if released and ignited outdoors.

3.1.4.3 *Gaseous Hydrogen* such as in tube trailers, in a pipeline, or in a process, regardless of system pressure, should not be considered to present an outdoor VCE potential. Only one outdoor explosion incident involving hydrogen is widely reported. This involved a small release of high pressure gaseous hydrogen during an acoustical test at the US Department of Energy Nevada Test Site. The extremely turbulent aerial cloud ignited, probably due to static energy caused by the turbulence of release. The overpressures were low and caused minimal damage to test equipment and structures. According to large users of gaseous hydrogen, aerial clouds are occasionally released from hydrogenation units and other processes with slightly delayed ignition. However, these so called "aerial detonations" have not caused significant far-field overpressures and have resulted in localized damage. For this reason, gaseous hydrogen is not deemed an unusually severe or credible VCE exposure.

*R.J. Harris, M.J. Wickens, Understanding Vapour Cloud Explosions—An experimental Study Communication 1408, Presentation to 55th Autumn Meeting, Kensington Town Hall, 27th and 28th November, 1989.

3.1.4.4 *Miscellaneous flammable or combustible gases* such as ammonia synthesis gas (a hydrogen/carbon monoxide mix), coal and blast furnace gases, methylene chloride, and trichloroethylene are excluded from this document, because of low flame speeds and heats of combustion and lack of loss history.

3.1.4.5 *Flammable liquids or gases processed above their autoignition temperature* will immediately ignite on contact with air. A severe flash fire may result, but delayed cloud ignition, necessary for development of significant overpressure, will not occur with these materials.

3.1.4.6 *Flammable liquids having a high viscosity* (greater than 1×10^5 centipoises) will likely not present normal vapor formation and will form pools of non-vaporizing liquid rapidly.

3.1.5 Mixtures

Mixtures of two or more liquid or gaseous materials in a process system may occur, creating difficulty in choosing a material class, calculating release, etc. There is no easy solution for calculating mixtures, even with advanced computer models. Considerable judgment must be used. Generally, the more hazardous material should be selected as if it were the entire volume. If equal amounts exist, several calculations may be necessary to determine a worst credible event. If the actual heat of combustion of the mixture is known, it may be used in energy equations.

3.1.6 Hybrid mixtures

A hybrid mixture is a mixture of dust in a flammable gas medium. VCEs have occurred in hybrid systems, notably at polyolefin manufacturing facilities. Some VCE researchers feel suspended polymer dust in a gas cloud might contribute to overpressure development. While possible, there is no calculation procedure or model known to accurately factor in hybrid mixtures.

Refer to Paragraph 3.2.9 for a discussion of determining volume of vapor or liquid in a hybrid system.

3.2 Step 2—Determination of Material Release Scenario

For purposes of estimating a *worst credible case* event, the following source term criteria for estimating the size of a material release should be used. For a *normal credible* event, these criteria can be reduced based on the results of a hazard analysis and risk assessment.

3.2.1 Process System

The size of a spill from a process system should be based on the contents of the largest vessel or train of vessels connected together that can be released in a 10 min period. Assume rupture of the largest interconnecting pipeline or connection that will allow the most discharge from the vessel or train. Sudden failure of a vessel is not considered. Generally, the presence of shutoff valves in a process system, even if automatic, cannot be given credit for limiting the flow from a multi-vessel system. Considerable judgment is needed on very large systems with redundant automatic block valves. Judgment may also be used when the process system is normally operated below rated volume capacity, as summarized in Section 3.2.9.

3.2.2 Fixed Storage Tanks

To estimate releases from all types of atmospheric and pressurized tanks, consider what can be discharged in 10 min from breakage of the largest pipe connection on the tank, taking gravity and storage pressure as the driving force. Sudden failure of the tank is not considered. The presence of flow limiting devices such as orifice plates, pipe reducers and excess flow valves, may be credited to limit the amount discharged if the criteria in Section 3.2.8 are met. Judgment may also be used when tanks and vessels are not normally kept filled to rated capacity, as summarized in Section 3.2.9.

3.2.3 Pipelines

Leaks in aboveground pipelines at any pressure carrying liquefied or non-liquefied gases to/from remote storage production facilities owned by the insured or others must be considered. For this purpose, assume that the pipeline is completely severed and the vapor cloud will be based on the amount that can be discharged in 10 min considering flow from both ends of the severed pipe, even if automatic or manual block valves are present. Buried high pressure transmission pipelines should be considered only when the pipeline pressure is greater than 250 psig (17 barg). History shows that *high* pressure buried pipelines can create

a crater caused by the gas release, allowing for a massive release of vapor into the environment. However, buried *low* pressure lines have not been responsible for known VCE events, and are considered extremely low probability events.

3.2.4 Mobile Transportation Vessels

Releases from mobile tanks such as rail and truck transportation vessels, and from waterway vessels such as ships and barges, should consider what can be discharged in 10 min from breakage of the largest connected line during transfer operations, considering gravity and storage pressure as the driving force. Sudden failure of the vessel is not considered. The presence of flow limiting devices, such as orifice plates, pipe reducers, and excess flow valves in the transfer line, may be credited to limit the amount released if the criteria in Section 3.2.8 are met.

3.2.5 Ignition Sources

The presence or lack of ignition sources should not be considered when calculating material release. For a worst credible case scenario, the total amount which might be spilled in 10 min should be used in estimating the cloud size. Loss experience has shown that wind patterns may allow the formation of large clouds without ignition by nearby ignition sources. In fact it is an irony of modern chemical plants that strict attention to ignition source control has produced the conditions for a flammable cloud to disperse over a large area without immediate ignition.

3.2.6 Atmospheric Conditions

Atmospheric conditions such as prevailing winds and atmospheric stability should be ignored when applying worst credible event assumptions. Under these assumptions the cloud is assumed to spread and ignite at the largest damage susceptible area of the plant within the constraints of the amount of material released and blast epicenter guidelines presented in Chapter 3.6. Under normal credible event assumptions, a risk analysis can determine prevailing wind patterns and atmospheric conditions, and a sophisticated dispersion model can be used to factor in the conditions that are most likely at the site.

3.2.7 Shape and Height of Cloud

The shape of the cloud is ignored even though it is known that different shapes can produce different overpressure conditions even with similar TNT equivalencies. Inclusion of complicated cloud shape models is beyond the scope of this calculation model. The height of the cloud can also affect overpressure at equal distances. Surface clouds can produce approximately 25-30% greater overpressure than aerial clouds. This guide assumes a *hemispherical, surface cloud*. See Section 3.5.1 for overpressure calculation for surface clouds.

Exception: Ethylene gas, when released at extremely high pressures from a process, system such as a high pressure polyethylene process may be directed upwards (due to barricading or other "confinement") and produce an *aerial cloud* with sufficient turbulence, generated by the jet release, to cause an air burst above the process area. See Section 3.5.2 for further discussion and methods of calculating overpressure from aerial clouds.

3.2.8 Flow Limiting Devices

Flow limiting devices, such as orifice plates, pipe reducers, and excess flow valves (EFV), can be given credit to limit the amount of material released in a worst credible case event if it is assured that the device is reliably installed and protected against mechanical damage, such that a break in section of pipe upstream of the device is very unlikely.

Credit can be given for both *internal devices* (i.e., flow limiting devices installed inside a vessel), and *external devices* (devices installed on a pipe outside of the vessel). When giving credit for externally installed devices, the device must be attached to the first flanged or welded connection to the vessel. Usually this is a nozzle that is factory welded and is delivered as part of the vessel. This first connection is usually very close to the vessel (a foot or less). The flow limiting device should be directly attached (preferably welded) to this first connection or nozzle *and* be reasonably well protected against mechanical damage by barriers or guards, or by the vessel itself.

When giving credit to flow limiting devices, all connections to the tank, excluding those fitted with pressure relief valves, should be installed with a device. Otherwise, the largest unprotected pipe should be used to determine release amount.

When applying flow limiting devices to limit a worst credible case event, credit for active devices, such as EFVs that rely on moving parts, should be carefully evaluated. While EFVs are valuable mitigation devices and desired and often code required, their reliability to stop or even minimize flow under worst credible event conditions is questionable. While such devices generally fail in the closed condition under worst case conditions (perhaps unfavorable from a production standpoint but favorable in an accidental release situation), a foreign object might lodge in the valve body, causing the EFV to fail partially or fully open. For this reason, it is advised in worst case scenario evaluations that only the maximum opening inside the EFV body be considered for reducing flow (i.e., the opening in the body will act only as restriction similar to an orifice plate). However, it is likely that this effective orifice can significantly restrict flow, as much or more than 50% of the pipe within which the EFV is installed. A hydraulic bench test can be used to determine effective orifice size of a wide open EFV.

For *normal credible* events, EFVs, if properly installed and maintained, may be given credit to minimize or completely eliminate release if a hazard analysis has been completed to justify this credit.

In conclusion, for a *worst credible case* event, when the conditions of location and protection against mechanical damage are met, the flow limiting device, whether a simple orifice plate, a reduction in pipe size, or an EFV-can be used to reduce flow rate based on the size of the maximum opening of the device.

3.2.9 Partially Full Systems or Vessels

Process systems or tanks may not be used or filled to capacity due to process or safety considerations. When it can be reliably concluded that this is a normal condition, as verified by operating records, standard operating procedures, etc., and where management systems are in place to ensure that changes are not made without awareness of plant safety professionals, then the *actual amount* in the vessel or system may be used for determining maximum release, rather than the nominal (full capacity) amount.

When solid material represents a significant portion of the volume of a reaction system, then the volume occupied by the solid should not be included in the calculation of the liquid release volume.

In many polymerization processes, the polymerization begins very quickly after catalyst introduction. The liquid or gaseous monomers and comonomers are usually held unreacted in the process system for only a short time (hours or days) (such as during startup) prior to catalyst addition and reaction initiation. At these times, pressures and temperatures will be substantially lower and conditions less severe than under normal operating conditions. Judgment must be used whether to apply the full (non-operating) contents or the partially full (reaction contents) when applying a worst case scenario. Because of the usually very short retention times when full it is generally agreed that the partially full operating conditions are a more reasonable choice for worst case scenario.

3.3 Step 3—Determination of Material Release Amount

The amount of material (gas—W_g or vapor—W_v) in a vapor cloud to be used in the Step 4, *TNT equivalency equation*, can be simply calculated following the steps below. Separate equations are provided for release from a *gaseous*, a *liquid* or *partial liquid-gas* system, and for determining flash fraction and pool spill vaporization.

Equations in Section 3.3 are given only in metric units for simplification. Most were initially derived in the metric system and conversion to English equivalents has, in some cases, compromised the equation. Common conversions, needed for this guideline, are given in Table 2. Additional conversions are available in Data Sheet 9-12/17-14, *Metric System Terms and Conversions*, and other resources. Equations have been simplified, where possible, by combining constants. The first equation in each set is the full equation. The second in each set is the simplified form which has combined certain constants, such as duration of release. The simplified equation may be used for calculation of a worst credible case event as defined in this document, however, the full equation should be used if other than the FM Global recommended constants are used.

Equation 1 is an approximation to the equations for ideal gas flow as presented by IOI, *The Evaluation of Estimated Maximum Loss from Fire or Explosion in Oil, Gas, and Petrochemical Industries with Reference to Percussive Unconfined Vapor Cloud Explosion*, July 1985 Edition, International Oil Insurers, UK. It will

overpredict discharge rates as it does not account for pressure reduction in the vessel during discharge.

The liquid release equation (Equation 2) featured in this guide predicts single phase flow. That is, it is based on a pure liquid release and does not account for two phase flashing flow, a phenomenon which occurs when a boiling liquid begins to vaporize as it suddenly exits a nozzle. This vaporization phenomenon causes a significant choking effect in the nozzle whereby the liquid flow is substantially reduced. Two phase flow can occur when the pipe break is more than a few inches from the vessel. If the break is at the vessel/pipe connection, single phase flow will more likely occur and Equation 2 will be representative.

Two phase flow can be reliably predicted only by using a sophisticated source term computer model. In some cases, it has been shown that prediction of mass released by single phase liquid flow by Equation 2 can be two or more times that predicted by a two phase flow model under the same process conditions. For this reason, the FM Global guide is conservative when predicting liquid release of a boiling liquid. Use of a computer model to more accurately predict liquid release is acceptable in lieu of Equation 2.

Equations 2 and 3 are as presented by World Bank Technical Paper No. 55, *Techniques for Assessing Industrial Hazards*, 1988, The World Bank, Washington, D.C.

Equation 4 is an approximation to a formula proposed by Raj, P.K. (1981), *Models for Cryogenic Liquid Spill Behavior on Land and on Water*, Journal of Hazardous Materials, V5, pp 111-130.

Table 2. *Common Conversions. Also see Data Sheet 9-12.*

1.0 BTU/lb-°F=1.0 Cal/gm-°C	16.04 lbs/cu ft=1 kg/cu m
1 tonne=1.1 tons	7.48 USgal=1 cu ft (H_2O)
2.2 lbs=1 kg	1 USgal=8.34 lbs (H_2O)
psig+14.7=psia	264 USgal=1 cu m
psia×6900=Pa	10.76 sq ft=1 sq m
cal/gm×4190=J/kg	35.3 cu ft=1 cu m
cal/gm °C×4190=J/kg°K	884.26 kg=1 cu m (H_2O)

Refer to Figure 2, *Material Release Decision Tree*, to aid in determining correct calculation procedure.

3.3.1 Gas Release Calculations

If the material exists in the system as a gas, the following can be used to estimate the mass of gas released (W_g) from a break.

Full Equation 1:

$$W_g = K\, C_d\, A_r\, t\sqrt{2\rho_1 P_d}$$

Simplified Equation 1:

$$W_g = K\, A_r\, 849\sqrt{\rho_1 P_d}$$

Where: W_g=Mass of gas released (kg)
    A_r =Area of release opening (sq m)
    C_d=Discharge coefficient (use 1.0)
    P_1 =Process or reservoir pressure (Pa)
    P_a =Ambient pressure (Pa) (absolute) (@ sea level P_a =1.014×10^5 pa)
    ρ_1 =Vapor density at process conditions (kg/cu m)
    t =Discharge duration (sec) (use 600)

    K =Gas Constant. Use K from Table 1. If not listed in Table 1, use 0.68
    P_d =Use actual process or reservoir pressure (P_1) when P_1 is in excess of 20 psi (135 kPa) (sonic flow). For pressures less than this (subsonic flow), $P_d=P_1-P_a$. (Pressure is expressed in Pa absolute.)

The gas constant K usually falls in the range 0.63 to 0.73. Hence, an average value of K=0.68 may be taken with little error when calculating vapor or gas flow rates for most materials.

Once W_g is known, proceed directly to Step 4, *Equation 7*, to determine TNT equivalent energy release.

3.3.2 Liquid Release Calculations

If the material in the vessel exists as a liquid, the amount discharged and the fraction that will flash vaporize can be estimated. To account for entrained aerosol mist, the flash fraction, F_{vap}, from Equation 3, should be doubled (up to but not exceeding 1.0). If this calculation still leaves a sizeable liquid portion, the liquid can be assumed to rain out and form a pool. The amount evaporated in 10 minutes from the resultant pool can be then calculated and added to determine the total vapor amount, W_v, to be used in Step 4. The total amount of vapor cannot exceed the initial amount in the vessel. A reminder that Equation 2 does not predict two phase flow.

3.3.2.1 Calculating Liquid Release

The mass of liquid released (W_l), assuming gravity and/or vessel pressure as the driving force with no vaporization (two-phase flow) in the orifice, is calculated by Equation 2:

Full Equation 2:

$$W_l = C_d A_r \rho_l \ t \sqrt{\frac{2(P_1 - P_a)}{\rho_l} + 2gh}$$

Simplified Equation 2:

$$W_l = 372 A_r \rho_l \sqrt{\frac{2(P_1 - P_a)}{\rho_l} + 19.6h}$$

Where W_l=Mass of liquid released (kg)
  A_r =Area of release opening (sq m)
  C_d=Discharge coefficient (use 0.62)
  g =Gravitational constant (9.81 m/s^2)
  h =Height of liquid in tank above discharge point (m)
  P_a=Ambient pressure (Pa absolute)
  P_1=Process or reservoir pressure (Pa absolute)
  ρ_l =Density of Liquid (kg/cu m) @ process temperature (T_1)
  t =Discharge Duration (sec) (Use 600)

3.3.2.2 Calculating Initial Flash Fraction

The liquid will flash once it has escaped and is at atmospheric pressure. The heat required for vaporization is taken from the liquid itself so that any liquid which is left will have been cooled to its atmospheric boiling point. The initial flash fraction F_{vap} of liquid is given by:

Equation 3:

$$F_{vap} = \frac{C_{P_1}(T_1 - T_b)}{\Delta H_{vap_b}}$$

Where: F_{vap} =Fraction of liquid flashed to vapor
  C_{p1} =Specific heat of liquid at constant pressure (J/kg°K) averaged between process temperature T_1 and atmospheric BP. If range figures not available, use C_{p1} at boiling point to be conservative.
  ΔH_{vapb}=Heat of vaporization (J/kg) at boiling point
  T_b =Atmospheric Boiling point (°K or °C)
  T_1 =Temperature of liquid in vessel (°K or °C)

In the FM Global Guidelines, F_{vap} is doubled to account for aerosol mist in the cloud. Therefore if the initial flash fraction (F_{vap}) is 50% or greater, the entire liquid contents can be assumed to vaporize, and W_l equals W_v. Proceed directly to Step 4, Equation 7 and use W_v to solve for TNT equivalent energy yield.

If F_{vap} is less than 50%, that portion of W_l that is not vaporized will rain out into a ground pool. The next step, Equation 5 in Section 3.3.2.4, is to calculate the amount of this ground pool that is subsequently vaporized due to thermal and atmospheric effects. For example, if the initial flash fraction, F_{vap}, is calculated to 15% and is doubled to 30% to account for aerosol mist, then 70% of the initial liquid release, W_l, will rain out into the liquid pool.

For a more accurate prediction of initial flash fraction a source term computer model can be used.

3.3.2.3 Calculation of Liquid Pool Size

The size of the liquid pool will be a factor in the evaporation rate. The extent of an **unconfined** spill on a relatively non-porous surface can be calculated as follows:

Full Equation 4:

$$A = t\sqrt{gV_o}$$

Simplified Equation 4

$$A = 1879\sqrt{V_o}$$

Where: A = Spill area (sq m)
 g = Gravitational constant (use 9.81 m/s²)
 t = Time (Use 600 even if discharge duration is less) (Sec)
 V_o = Volume spilled (cu m)

For *unconfined spills* on essentially flat surfaces, the area of the spill should be limited so that the pool depth is not less than 6 mm (¼ in.). For a *confined spill*, such as inside a diked or curbed area, use the ground surface as well as vertical surfaces of walls up to the maximum level of liquid. When a dike or process area has drainage the pool calculation becomes difficult. While drainage is favorable and required for controlling pool fires, it may be of little help in limiting vapor formation. The hot liquid can continue to vaporize within the drainage system, often more rapidly due to standing water and vapor can be released to the process area environment through the drainage system openings. For the purposes of this guideline, the presence of drainage within a dike or curbed area should generally be ignored and the full amount spilled used for vaporization calculations. However, for very large capacity drainage systems designed specifically to both rapidly remove spilled material to a safe remote location and to retard vapor formation to the atmosphere, credit can be given after using considerable engineering judgment.

3.3.2.4 Calculation of Liquid Pool Evaporation

Any material discharged as a liquid, if not initially flash vaporized or carried into the cloud as aerosol droplets, will form a pool on the ground and vaporization of the boiling pool may occur. If the atmospheric boiling temperature of the discharged liquid is below the ambient temperature, the liquid will continue boiling after it is spilled on the ground. The heat necessary for boiling is supplied by conduction heat transfer from the ground.

The amount of liquid evaporated (W_{boil}) by ground heat transfer is calculated using Equation 5.

Equation 5:

$$W_{boil} = \frac{2}{\sqrt{\pi}} B \frac{\sqrt{t}}{\Delta H_{vap_b}} (T_a - T_b)A$$

Simplified Equation 5

$$W_{boil} = \frac{27.6 \, A \, (T_a - T_b) \, B}{\Delta H_{vap_b}}$$

Where: W_{boil} = Amount of liquid (kg) evaporated, not to exceed the amount of liquid spilled
 B = Thermal property of spill surface (See Table 3)
 t = Time (Sec) (use 600 even if discharge duration is less)
 A = Area of the spill (sq m) in contact with solid surface.
 T_a = Ambient temperature (°K or °C) (Use actual summer high for ambient temperature)
 T_b = Atmospheric Boiling point of liquid (°K or °C)
 ΔH_{vap_b} = Heat of vaporization (J/Kg) at atmospheric boiling point

If the liquid boiling point is above ambient, this equation fails. In this situation, diffusional evaporation is the controlling factor. Equations to determine this rate are highly subject to atmospheric conditions, particularly wind speed. Since wind speed is not known, this factor is excluded from this simplified approach. A computer model must be used to predict the release under this type of conditions.

Table 3. Thermal Properties of Surfaces (B) for use in Equation 5

Surface	Thermal Property, B $(Ws^{1/2}/sq\ m\ °K)$
Soil (dry)	5400*
Soil (8 moisture)	675
Insulating Concrete	780
Light Concrete	1400
Heavy Concrete	4100
Masonry/Ceramic Tiles	2900
Wood	300
Carbon Steel	13000

*Includes a multiplier of 8 due to soil porosity which increases surface area for vaporization

3.3.2.5 Calculation of Total Amount of Vapor in Cloud

The amount of material used to solve the TNT equivalency equation (Equation 7) in Section 3.4, should be either W_g (contents initially gas) or W_v (contents initially liquid or mixed liquid vapor). W_g was calculated using Equation 1.

W_v can be calculated by adding the amount of released liquid (W_t) (Equation 2) that flashes (Equation 3) and the amount evaporated from the liquid pool (W_{boil}) (Equation 5).

Equation 6:

$$W_v = W_t\ (2\ F_{vap}) + W_{boil}$$

Note that the initial flash fraction (F_{vap}) is doubled to account for aerosolization. $2\ F_{vap}$ cannot exceed unity (1.0) and W_v cannot exceed the total mass of material initially released from the vessel.

3.3.3. Comments

Refer to data in *Physical Properties of Hydrocarbons*, Vols. 1 and 2, by R. W. Gallant or *CRC Handbook of Chemistry and Physics*, for values for C_p, ΔH_{vap} and ρ. Data on materials can also usually be found at the plant site and some data are also provided in Table 1.

3.4 Step 4—Determination of Energy Release

The energy released in an explosion of a vapor cloud is expressed as a TNT equivalent. This methodology is often called the *Ideal Blast Wave* method.

3.4.1 Calculating TNT Equivalency Using English Units

Based on an approximate energy of decomposition for TNT of 2000 Btu/lb, the following may be used to calculate a TNT equivalent for a vapor cloud containing a known weight of flammable gas (W_g) or vapor (W_v) in *English units*:

Equation 7a (English Units):

$$W_e = W\ \frac{\Delta H_c\ f}{4 \times 10^6}$$

Where: W_e = Mass of Equivalent TNT Energy Yield (tons)
   W = Mass of vapor in cloud (lb) of gas (W_g) or vapor (W_v) from Section 3.3
   ΔH_c = Heat of combustion of material (Btu/lb)
   f = Explosive yield (efficiency) factor (See Section 3.4.3)

3.4.2 Calculating TNT Equivalency Using Metric Units

Based on an approximate energy of decomposition for TNT of 1111 Kcal/kg, the following may be used to calculate TNT equivalent in *Metric Units*.

Equation 7b (Metric Units):

$$W_e = W \frac{\Delta H_c \, f}{1.111 \times 10^6}$$

Where: W_e = Mass of Equivalent TNT Energy Yield (tonnes)

    W = Mass of vapor in cloud (kg) of gas (W_g) or vapor (W_v) from Section 3.3

    ΔH_c = Heat of combustion of material. (Kcal/kg)

    f = Explosive yield (efficiency) factor (See Section 3.4.3)

3.4.3 Explosion Efficiency Factors

When applying this guideline to predict a worst credible case scenario, explosion efficiency (f) should be assigned based on the class of material (see Table 1) as follows:

   a. Class I materials, f=0.05 (5%)

   b. Class II materials, f=0.10 (10%)

   c. Class III materials, f=0.15 (15%)

3.4.4 If the material is not listed in Table 1 and it is not a highly reactive material, it should be assumed to be a Class I material.

3.4.5 If the guideline is being applied to predict the effects of a VCE for other than a worst credible case scenario (e.g., a normal credible case scenario), explosion efficiencies must be chosen based on the practitioner's best interpretation of the exposure. This can usually only be done by a systematic process safety review such as HAZOP or quantitative risk analysis.

3.4.6 Explosion efficiencies listed in paragraph 3.4.3 are based on historical evidence and on literature classification of materials by several different sources, including Dave J. Lewis of ICI (England), *Unconfined Vapor Cloud Explosion—Historical Perspective and Predictive method Based on Incident Records*, (1980) Prog. Energy Comb. Sci., VC, pp 151-165.

3.4.7 Heats of combustion (ΔH_c) of common materials are listed in Table 1 or can be found in many reference guides such as *Perry's Chemical Engineers Handbook*.

3.5 Step 5—Determination of Overpressure Radii

Once TNT equivalency (Step 4) is determined, overpressure radii can be determined from scaled distance charts using the Hopkinson Scaling Law. Knowing W_e (from Equation 7), a scaled ground distance factor (Z_g) at selected overpressures can be obtained. This factor is converted into radial distance by Equation 8.

3.5.1 Calculate Overpressure for Hemispherical Surface Cloud

Table 4a lists Z_g at common overpressures for ground burst explosions in hemispherical surface clouds. Surface cloud data should be used for all heavier than air materials, which are essentially all hydrocarbon materials. Overpressures listed in Table 4a are *peak incident (side-on) overpressures*.

3.5.2 Calculate Overpressure for Spherical Aerial Clouds

Table 4b lists Z_g at common overpressures for air burst explosions in spherical aerial clouds. Aerial cloud data should be used for very light gases which would be expected to lift off very rapidly after release. One material of concern is high pressure ethylene which might be jet released directionally into the air from a barricaded high pressure polyethylene process. Overpressures listed in Table 4b represent *peak (side-on) overpressures* from an aerial explosion greater than 10 ft (2.6 m) aboveground.

3.5.3 Scaled distance factors in Tables 4a and 4b are derived from *Structures to Resist the Effects of Accidental Explosions*, Departments of Army, Navy, and Airforce Manual (TM 5-1300, NAVFAC P-397, AFM 88-22), November 1990 Edition. Reference may be made to *Peak Positive Incident Pressure* (P_{so}) curves presented in Figures 2-8 (Spherical Free Air Burst) and 2-15 (Hemispherical Surface Explosion).

Equation 8:

$$R_g = Z_g \, W_e^{1/3}$$

Where: R_g = Radial distance from energy release epicenter (expressed in feet or meters), at given overpressure.

Z_g = Scaled ground distance $(ft/lb^{1/3})$ $(m/kg^{1/3})$ from Tables 4a and 4b)

W_e = TNT equivalent, expressed in *lbs or kgs*

Table 4a. Scaled Ground distance (Z_g) at overpressures (P)
for Hemispherical **Surface** Explosion.

Overpressure (P)		Scaled Ground Distance (Z_g)	
psig	barg	$ft/lbs^{1/3}$	$m/kg^{1/3}$
15	(1.03)	8	3.17
10	(0.69)	9.8	3.89
6	(0.41)	13	5.16
5	(0.34)	14.5	5.75
3	(0.21)	19.5	7.73
2	(0.14)	26	10.0
1	(0.07)	45	17.85

Table 4b. Scaled Ground distance (Z_g) at overpressures (P)
for Spherical Aerial Explosion.

Overpressure (P)		Scaled Ground Distance (Z_g)	
psig	barg	$ft/lbs^{1/3}$	$m/kg^{1/3}$
15	(1.03)	6.5	2.57
10	(0.69)	7.8	3.10
6	(0.41)	10	3.97
5	(0.34)	11.5	4.56
3	(0.21)	16	6.35
2	(0.14)	22.5	8.93
1	(0.07)	40	15.87

3.6 Step 6—Determination of Cloud Drift and Explosion Epicenter

There is no easy or precisely accurate method to predict cloud drift or location of explosion epicenter. This is a qualitative procedure only. While numerous computerized dispersion models are available to approximate cloud drift, the practitioner still has to use judgment to predict atmospheric conditions and to decide where an explosion might originate. Released vapor cloud can disperse (drift) in a wide variety of modes and directions, depending on atmospheric conditions, type of release, nature of the material, and ground terrain features. Stability of the cloud (i.e., propensity to disperse) is highly dependent upon wind and atmospheric stability which is related to cloud cover or darkness (which promote lower levels of atmosphere induced turbulence than sunlight), atmospheric pressure and other meteorological conditions. Clouds under neutral to moderately stable atmospheric conditions will drift further but in a narrower arc than more unstable conditions. Because of this complexity, a quantitative risk assessment and use of a sophisticated computerized dispersion model is needed to fully understand cloud dispersion.

Historically, VCEs have been ignited inside buildings, at process equipment, inside vehicle engine compartments, or at outdoor fired equipment. However, the actual explosion epicenter cannot always be assumed to be the same location as the assumed ignition. Because ignition source location might not be in an area that is congested or confined, the flame in the ignited cloud of gas may not accelerate to deflagration until it reaches a portion of the cloud that is substantially within a congested or confined process area. An example is the 1989 VCE at a refinery tank farm in Baton Rouge, LA (see Section 4.2.2.2 for a description of this incident). The vapor cloud was initially released from a high pressure pipeline at the tank farm but drifted many hundreds of feet toward the parent refinery where it was ignited at a fired heater. However, there was no significant explosion at the heater. Rather, a flash fire in the cloud spread at sub-deflagration speed back through the unburned gas until it reached a substantial pocket of gas confined in an obstacle array
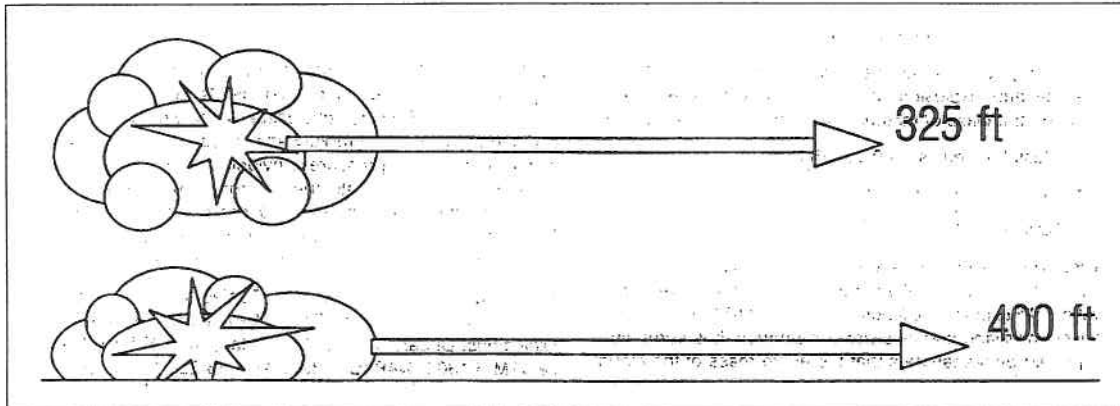
Fig. 3. Effect of cloud height on overpressures. Distance to 5 psig (0.34 barg) ring produced by 10 ton (9.1 tonne) TNT yield, surface explosion compared to aerial explosion.

at the tank farm. The actual epicenter of the explosion was at the tank farm even though the ignition was more than 1500 ft (457 m) away.

3.6.1 Cloud Drift Allowance

A simplified methodology using *cloud drift allowance* has been developed for this guideline. The following procedure should be followed for determining explosion epicenter (EC):

a. A drift allowance based on the class of material (from Table 1) and weight of vapor (W_v or W_g) released (expressed in tons or tonnes) should be chosen from Figure 4a (English) or 4b (Metric). This is given in feet or meters radius from the release point (R).

b. Because probable wind direction cannot be simply determined without a quantitative risk analysis, it must be assumed that the cloud can drift in any direction to the limits of the distance predicted from Figure 4. To define a *worst credible* event a cloud drift circle using the chosen radius should be plotted on a site plan using release point (R) as the centerpoint. The circle is not the actual total area of a cloud. Actually, the cloud will tend to disperse downwind in an elongated arc type shape (similer to a flower petal) away from the release point, or, if conditions are neutral, form a circular, pancake shaped cloud around the point of release. The area enclosed by the circle is called the *explosion vulnerability zone*.

c. It should be assumed that the explosion epicenter can be anywhere within this circular vulnerabilty zone; however, epicenter should not be chosen in an open area, such as a field, wide space, or a roadway between process blocks. Generally the epicenter should be selected within a congested process block. This is commonly the process block where the release originated. If the explosion vulnerability zone significantly extends into (i.e., essentially fills) a congested process area other than the process area where the release originated, the explosion epicenter may be chosen in that area instead. If the explosion vulnerability zone is essentially over open water, such as at a river barge unloading facility, over open land, around a small tank farm or remote product loading facility, it is unlikely that the vapor cloud, if ignited, will produce damaging overpressures because of limited obstacles and confinement. In these cases, radiant heat effects will be the prime concern.

When evaluating *normal credible* events, a risk analysis can determine prevailing wind and atmospheric conditions and an epicenter selected based on the most likely dispersion direction.

3.6.2 Maximum Drift Allowance

A maximum of 1000 ft (305 m) drift should be used when applying Figure 4.

3.6.3 Aerial Clouds

If a cloud, such as high pressure ethylene, is jet released upwards and aerial cloud scaling rules are used, the cloud should *not be drifted*. The explosion epicenter should be chosen directly over the release.
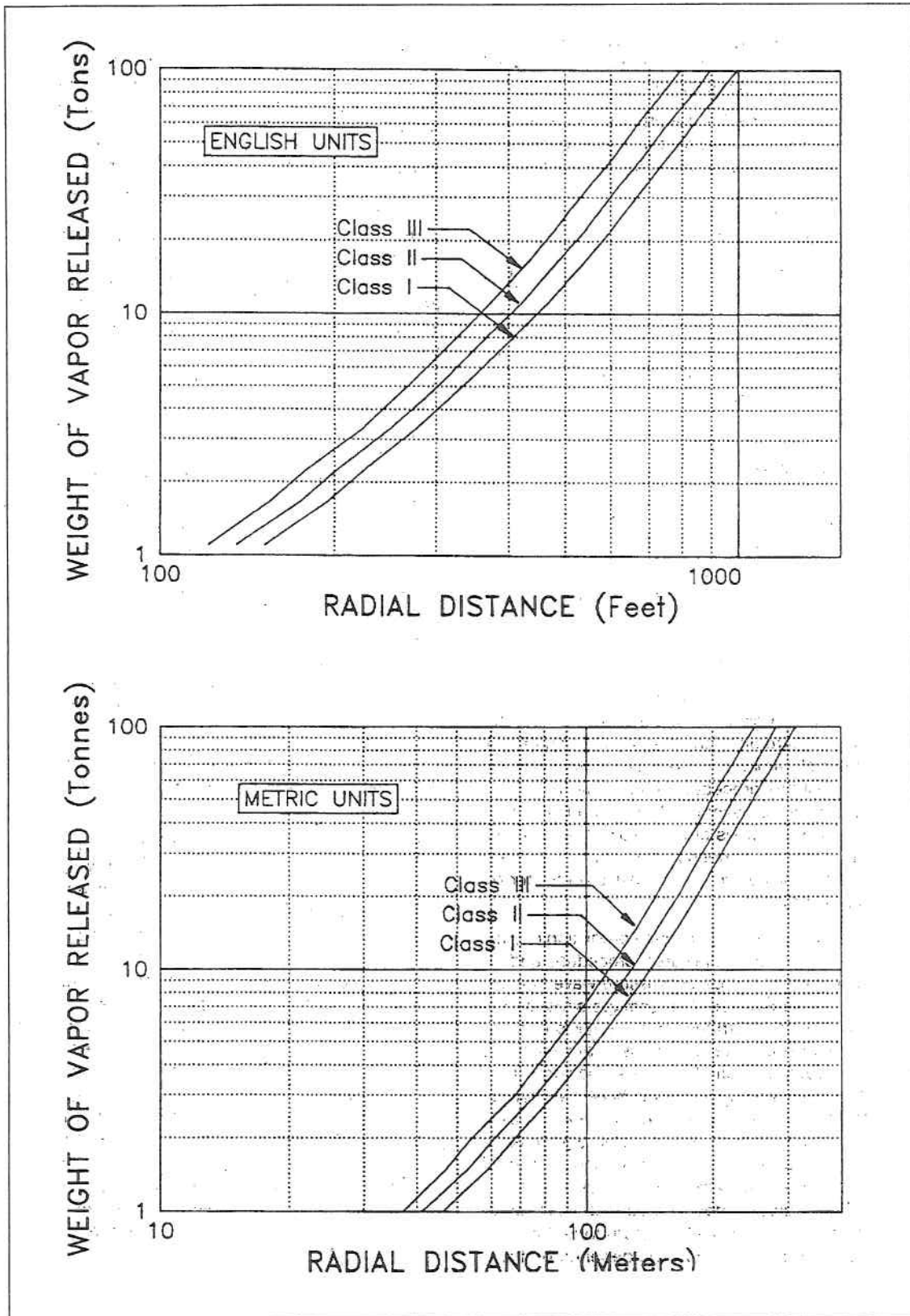
Fig. 4. Cloud Drift Radius

3.6.4 Drift allowances are derived from those developed by David J. Lewis, ICI (Mond Division, Norwich, England) in his 1980 paper, *Unconfined Vapor Cloud Explosions— Historical Perspective and Predictive Method Based on Incident Records.*

3.7 Step 7—Determination of Damage

Physical damage to individual buildings and process equipment is determined for a given pressure profile developed in Step 5. It should be noted that this procedure is very qualitative and gives only an approximate estimate of worst credible case damage. Sophisticated models are available for predicting overpressure and thermal response of structures. Sophisticated modeling techniques may be substituted for these approximate guidelines.

It is important to understand that accurate prediction of blast damage to structures is a complicated procedure and requires fundamental knowledge of pressure wave characteristics and structural response. The damage caused by an overpressure blast is dependent on the peak pressure, the duration of the pressure wave, and the impulse time of the pressure wave. In a detonation type explosion (which is the basis of a TNT model) the blast wave forms a rapid, essentially instantaneous rise in pressure (shock wave) to the peak value. This value is called the peak side-on pressure. (All pressures within this document are based on side-on pressure data.) The side-on pressure wave is followed by a gradual decay through ambient pressure to a negative phase and finally a return to ambient pressure. The positive pressure duration depends on the amplitude of of the explosion and can vary from 1 msec for small events to about 250 msec for very large explosions.

The shape of an explosion can create different damage effects and is dependent on the nature of the explosion. A VCE occurs more slowly than a true TNT detonation. Pressures are at a lower amplitude and build up in a slower, more gradual manner than a detonation. A VCE can have a much longer impulse time, however, than a detonation. Far field damage can be significant in a VCE event, especially to lightweight steel frame buildings with large flat surfaces and to glass.

The principal factors that effect structural performance of a building are incident (side-on) peak pressure; impulse time (duration) of the pressure load; shape and time of pressure rise; resistance or available strength of structural components of the building or structure; mass and stiffness of the structural components which relates to the natural period or frequency of the component; and the ductility or ability to absorb the blast energy without collapse or excessive deformation and damage.

It is not the intent of this guideline to provide quantitative structural data sufficient to estimate with a high degree of certainty the ability of a structure or building to resist overpressure effects from an explosion. However, Tables 5, 6 and 7, which are based on actual incident history, provide qualitative guidance for determining extent of overpressure damage from a VCE. These tables do not factor in ensuing damage from other events such as jet flames from gas releases, fires from flammable liquid pool fires, BLEVEs of vessels, or fragment or missile effects. However, evaluations of historical losses demonstrate that the ensuing events may cause as much or more damage than the initiating vapor cloud explosion. Thus it is of importance to carefully evaluate the possibility of these ensuing events. *It is emphasized that there is considerable room for engineering judgment in damage evaluation.*

When calculating damage estimates, structural damage from pressure should be determined independent of ensuing damage.

For example, using the referenced tables, a degree of pressure damage can be developed for a lightweight steel frame building housing a large, internally lubricated air compressor. Assuming a pressure profile of 5 psi (0.30 bar) at the building, one might conclude essentially 100% damage to the structure yet only superficial damage to the unit inside the building. Because the high flash point lube oil system in this compressor is located internally to the unit and flammable liquids or gas products are not present, an ensuing fire of any significance is not likely, even if released. (Note, however, that the potential for ensuing fire will depend highly on whether the building is exposed to significant radiant heat from the burning cloud or just overpressure effects. Generally the cloud does not burn or cause significant far field heat effects outside the 7 psi [0.50 bar] profile). The overpressure damage may well be the total value of the building, but only a few percent of the value of the contents even through the pressure profile is relatively high.

However, should the this same compressor be used for high pressure flammable gas service, gas piping, may break and ensuing jet flames might impinge upon the compressor, causing greater damage. In this case, building damage would still be 100% but contents damage may increase to 50% or more, depending on how long the gas jet flames were allowed to continue.

Should this same building also house flammable liquid service pumps, tanks and piping, the effects of release of flammable liquids and resultant pool fires or tank BLEVEs would also be considered. In this case both building and contents may suffer 100% damage.

Maximum overpressures in VCEs are generally in the 5 to 10 psi (0.3 to 0.7 bar) range, although near the epicenter, pressures of 10 to 12 psi (0.7 to 0.8 bar) are suspected. Pressures in excess of 15 psi (1 bar) are not likely in outdoor VCEs. Pressures in the 10 to 12 psi (0.7 to 0.8 bar) range can shift or overturn some heavy equipment, such as a reactor or a fractionation column, but would likely not move a steam turbine or a large compressor. Pressures in the 5 to 7 psi (0.3 to 0.5 bar) range have been known to uplift full cone roof tanks and release contents. Surfaces with a broad profile to the pressure wave are more susceptible than round or low profile surfaces. As an example, in one notable VCE event, fin-fan coolers with broad steel panel sides suffered significant distortion damage (deformation, dimpling, and bending) at great distances even though round surface tanks and process columns near the epicenter were only slightly deformed.

Fired vessels such as heaters, reformer furnaces, or cracking furnaces, can be shifted or uplifted with severe results at pressures in the 5 to 7 psi (0.3 to 0.5 bar) range depending on the profile of the unit. Internal refractory in fired vessels can dislodge at pressures as low as 1 psi (0.07 bar). Within 5 psi (0.3 bar) circles, breakage of welded or fitted piping (both process and fire protection) is generally a certainty. Outside of the 5 psi (0.3 bar) pressure circle, breakage of piping is not likely, but is a possibility if the piping is not welded and/or if it is attached to a structure that is likely to fail.

Deluge system piping featuring screw fitted couplings is highly susceptible to damage from overpressure, and will likely fail at 2 psi (0.14 bar) or lower if the structure moves significantly. Welded fittings will provide good resistance to breakage to approx 3 psi (.2 bar). Overpressure structural damage in pressure circles of 3 psi (0.2 bar) or less is generally not great unless the building is of lightweight panel construction or of unreinforced masonry walls with heavy roofs. Interior damage to contents at these pressures are usually not severe unless an ensuing fire develops. Should vapor enter a building and cause a confined vapor explosion, damage will vary based on extent of damage limting construction and amounts, if any, of flammable materials within the building. Outside of the 5 psi (0.3 bar) circle, radiant heat is not likely to ignite ordinary combustibles.

Table 5. Explosion Overpressure Damage Estimates, from Incident Observations.
(Lees, F.P., Loss Prevention in the Process Industries. Vol. 1, Butterworths, London and Boston, 1980.)

Overpressure (P) psig (barg)	Expected (Observed) Damage
0.03 (0.002)	Occasional breaking of large windows already under stress.
0.04 (0.003)	Sonic boom-type glass failures.
0.10 (0.007)	Breakage of small windows under strain.
0.15 (0.01)	Typical pressure for glass failure.
0.30 (0.02)	Some damage to house ceilings; 10 window glass breakage.
0.40 (0.03)	Limited minor structural damage.
0.5-1.0 (0.035-0.07)	Windows usually shattered; some window frame damage.
0.7 (0.05)	Minor damage to wood frame structures.
1.0 (0.07)	Partial Demolition of wood frame structures
1.0-2.0 (0.07-0.14)	Corrugated metal panels fail and buckle.
1.3 (0.09)	Steel frame of clad building slightly distorted.
2.0 (0.14)	Partial collapse of walls and roofs of wood frame structures
2.0-3.0 (0.14-0.21)	Non-reinforced concrete or cinder block walls shattered
2.3 (0.16)	Lower limit of serious structural damage.
2.5 (0.18)	50% destruction of home brickwork.
3.0 (0.21)	Steel frame building distorted and pulled away from foundation.
3.0-4.0 (0.21-0.28)	Frameless steel panel building ruined.
4.0 (0.28)	Cladding of light industrial buildings ruptured.
5.0 (0.35)	Wooden utility poles snapped.
5.0-7.0 (0.35-0.49)	Nearly complete destruction of houses.
7.0 (0.49)	Loaded train wagons (rail cars) overturn
7.0-8.0 (0.49-0.56)	8-12 in. thick non-reinforced brick fail by shearing of flexure.
9.0 (0.63)	Loaded train box cars demolished.
10.0 (0.70)	Probable total building destruction.

Table 6. Explosion Overpressure Damage

Equipment	0.5 (0.03)	1.0 (0.07)	1.5 (0.10)	2.0 (0.14)	2.5 (0.17)	3.0 (0.21)	3.5 (0.24)	4.0 (0.28)	4.5 (0.31)	5.0 (0.34)	5.5 (0.38)	6.0 (0.41)	6.5 (0.45)	7.0 (0.48)	7.5 (0.52)	8.0 (0.55)	8.5 (0.59)	9.0 (0.62)	9.5 (0.66)	10.0 (0.69)	12.0 (0.83)	14.0 (0.97)	16.0 (1.10)	18.0 (1.24)	20.0 (1.38)	>20.0 (>1.38)
Control house: steel roof*	a	c					n																			
Control house: concrete roof*	a	e p	d				n																			
Cooling tower	b		t				o																			
Tank: cone roof		d				k							u													
Instrument cubicle			a			f m						t														
Fired heater				g	i					t																
Reactor: chemical				a				i					p					t								
Filter				h					f	i p									v		t					
Regenerator						j									t						t					
Tank: floating roof						k				u			u												d	
Reactor: cracking							i			t																
Pipe supports							p		q			s o														
Utilities: gas meter										h																
Utilities: electric transformer																		t								v
Electric motor										h				t						t						
Blower											f															
Fractionation column																										
Pressure vessel: horizontal												p i						t								
Utilities: gas regulator												i								m q						
Extraction column													i							v	t					
Steam turbine																					m	s				
Heat exchanger																j							t			v
Tank: sphere															–						–	t				
Pressure vessel: vertical															–					–			–			
Pump																				v	–		v			v

Source: Minimizing Damage to Refineries US Dept. of Commerce, Feb. 1970
Based on actual damage in chemical plant explosions.

* Unless specifically designated to be blast resistant

Code
a. Windows and gauges break
b. Louvers fall at 0.3-0.7 psi
c. Switchgear is damaged from roof collapse
d. Roof collapses
e. Instruments are damaged
f. Inner parts are damaged
g. Brick cracks

h. Debris-missile damage occurs
i. Unit moves and pipes break
j. Bracing fails
k. Unit uplifts (half-filled)
l. Power lines are severed
m. Controls are damaged
n. Block walls fail

o. Frame deforms
p. Frame collapse
q. Case is damaged
r. Frame cracks
s. Piping breaks
t. Unit overturns or is destroyed
u. Unit uplifts (0.9 filled)
v. Unit moves on foundation

Table 7A. Summary of Blast Damage to Structures (Part 1)

Overpressures psig (barg)	Controlhouses		Steel Frame bet. Vessels	Crude Units			
				Atmos/Vacuum Towers		Fractionator Towers	
	Steel Roof Deck and No Frame	Precast Concrete Roof and Steel Frame		Rectangular Conc. Frame	Octagonal Conc. Frame	Rectangular Conc. Frame	Mounted on Conc. Pedestal
0.5 (0.03)	Windows shatter	Windows shatter					
1.0 (0.07)	Roof collapse (switchgear room)	Frame deformation					
1.5 (0.10)	Roof collapse (control room) (West blast) / Partial roof collapse (control room) North and South blast	Roof collapse (all rooms)		Note: Atmospheric & Vacuum Towers	Note: Vacuum Towers only		
3.5 (0.24)	Conc. block walls fail	Conc. block walls fail	Concrete brackets fail causing frame collapse				
4.5 (0.31)							Anchor bolts yielding
5.0 (0.34)							
5.5 (0.38)				Conc. frame cracking		Conc. frame cracking	
7.0 (0.48)				Conc. frame collapse	Conc. frame cracking	Conc. frame collapse	Vessel & foundation overturn
7.5 (0.52)					Vessel anchor bolts fail causing frame collapse		
8.0 (0.55)							
8.5 (0.59)							
10.0 (0.69)		Steel frame collapse					
12.0 (0.83)							
16.0 (1.10)							

Table 7A. Summary of Blast Damage to Structures (Part 1) (continued)

| Overpressures psig (barg) | Fluid Catalytic Cracking Units (FCCU) | | | | |
| | Regenerator Tower | | Reactor Tower | | Fractionator Tower |
	Rectangular Steel Frame	Rectangular Conc. Pedestal	Rectangular Steel Frame	Rectangular Conc. Frame	Mounted on Conc. Pedestal
0.5 (0.03)					
1.0 (0.07)					
1.5 (0.10)			**Note:** Reactor & Fractionator supported by same frame		
3.5 (0.24)					
4.5 (0.31)					
5.0 (0.34)	Leeward columns buckle East or West blast				Anchor bolts yielding
5.5 (0.38)					
7.0 (0.48)	Overturns Easterly blast		Leeward cols. buckle. East blast		Overturns Anchor bolts fail
7.5 (0.52)			Leeward cols. buckle. West blast		
8.0 (0.55)	Overturns Westerly blast			Concrete frame cracking	
8.5 (0.59)		Conc. frame cracking			
10.0 (0.69)					
12.0 (0.83)			Steel frame overturns	Concrete frame collapse	
16.0 (1.10)		Conc. frame collapse			

Source: Minimizing Damage to Refineries U.S. Dept. Commerce, Feb. 1970
Based on actual damage in Chemical Plant Explosions

Table 7B. Summary of Blast Damage to Structures (Part 2)

Overpressure psig (barg)	Light Ends Unit		Furnaces-Pipe Still		Maintenance Building	Water Cooling Tower (Wood Frame)	Flares	
	Deisobutanizer — Mounted on Pedestal and Large Footing	Vapor Recovery Unit — Rectangular Steel Frame	Atmospheric	Vacuum			Tower Supported	Guyed
0.3 (0.02)					Corrugated Asbestos Siding fails	Corrugated Asbestos Louvers fail		
1.5 (0.10)			Moves slightly from original position	Moves slightly from original position				
2.0 (0.14)								
3.0 (0.21)					Steel frame deformation		Steel frame overturns. Blast diagonally oriented	
3.5 (0.24)								
4.0 (0.28)						Tower collapses	Steel frame overturns. Blast squarely oriented	
5.0 (0.34)					Brick walls collapse. Severe frame deformation			
6.0 (0.41)		Steel frame collapse	Stacks collapse	Stacks collapse	Steel frame collapses			
6.5 (0.45)		Steel frame collapse	Steel frame collapse					
7.0 (0.48)								
7.5 (0.52)								
9.0 (0.62)								
9.5 (0.66)	Vessel overturns							
10.0 (0.69)								
10.5 (?)								
11.0 (0.76)								Collapse above middle collar
15.0 (1.03)								Complete collapse
20.0 (1.38)								

Table 7B. Summary of Blast Damage to Structures (Part 2) (continued)

Overpressure psig (barg)	Pipe Bands		Boiler Stack FCC Unit	TEL Building	Bulk Terminal	Storage Tanks		
	Steel Frame	Concrete Frame				Cone Roof	Floating Roof	Spherical
0.3 (0.02)								
1.5 (0.10)				Tile walls fail	Roof of Admin. Bldg. collapses. Cone roofs of tanks collapse	Empty tank uplift	Empty tank uplift	Empty tank uplift
2.0 (0.14)								
3.0 (0.21)								
3.5 (0.24)	Steel frame deformation	Concrete frame cracking				Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio
4.0 (0.28)						Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio
5.0 (0.34)		Concrete frame collapse				Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio
6.0 (0.41)	Steel frame collapse					Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio
6.5 (0.45)			Stack and foundation overturn		Tanks uplift (50% to 90% filled)	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio	Tanks (50% to 90% filled) uplift over this range depending on diameter and height to diameter ratio
7.0 (0.48)				Steel frame deformation				Support deformation (full) support
7.5 (0.52)								Support deformation (empty)
9.0 (0.62)								Overturns (full)
9.5 (0.66)				Steel frame collapse				Overturns (empty)
10.0 (0.69)								
10.5 (?)								
11.0 (0.76)								
15.0 (1.03)								
20.0 (1.38)							Roof Collapses	

Based on actual damage in Chemical Plant Explosions

Source: Minimizing Damage to Refineries U.S. Dept. of Commerce, Feb. 1970

4.0 LOSS EXPERIENCE

4.1 Historical Overview

Historically, the frequency of VCEs is extremely low compared to other events at chemical facilities. This is due to the need in a VCE event for a sudden large release, ideal outdoor confinement, a delay in ignition and proper conditions of explosibility. There are many other explosion events, many much smaller, that are more likely to occur, such as a flammable liquid spill initiating a BLEVE, an indoor (confined) vapor explosion, or a exothermic runaway reaction in a process reactor. Further, there are reportedly many large vapor cloud releases which do not ignite and are safely dispersed. Others ignite but do not have the proper local conditions necessary to generate overpressures, resulting in vapor cloud fires rather than VCEs.

Nonetheless, there have been many VCEs that have caused extensive property damage and production loss in the past 50 years. The frequency and severity of incidents appears to be increasing, attributable to larger production plants operating at greater capacities. The age of the chemical industry may also influence the potential for release, as the older a facility, the greater the likelihood for an unplanned release due to normal wear and tear.

Lenoir and Davenport[3] (1992) list 103 suspected VCE incidents which produced observable overpressures during the period 1921 to 1991. Of these events, 79 occurred in the past 25 years and 27 in the last 10 years.

Marshal[4] reports 50 recorded VCEs in the non-communist world in the period 1950 to 1982.

Strehlow[7] lists 108 accidental VCEs documented over a 42 year period up to 1972.

Since 1987, an estimated eight major outdoor vapor cloud incidents have occurred in the United States. Several others have occurred elswhere in the world, primarily in the former Soviet Union and in developing nations.

Two of the recent US VCE incidents accounted for an estimated $1 billion property damage and $700 million business interruption (time element). The October 1989 Pasadena, Texas VCE is listed as the largest non-marine loss to have occurred in the Western World[3].

FM Global studied 80 well documented VCE incidents of which:

23 (28.8%) involved high pressure pipeline (> 250 psig [17.9] barg)

22 (27.5%) were releases from a process system

16 (20%) involved mobile railcars or tank trucks, of which

    5 occurred onsite of a chemical facility

    11 occurred off site of a chemical facility

6 (7.4%) were emergency relief valve releases

5 (6.3%) initiated from storage tanks or systems

5 (6.3%) were releases from pumps or compressors

3 (3.7%) were miscellaneous incidents, identified as

    1 ship overfilling

    1 salt dome overfilling

    1 dirigible explosion

Of the 80 incidents studied by FM Global, 66 occurred on the site of chemical facilities of which:

33 (50%) were process related (pumps, compressors, reactors, etc)

23 (34.8%) were high pressure pipeline failures

5 (7.6%) were storage tank incidents

5 (7.6%) were railcar or tank truck incidents

The remaining 14 incidents occurred off site of actual processing facilities, primarily in rural areas due to pipeline releases or vehicle accidents.

4.2 Vapor Cloud Explosion Loss Examples

Several noteworthy VCE events are summarized below. When available, property damage (PD) and business interruption (BI or time element) gross estimated loss amounts are provided and are indexed to 1993 US dollars.

4.2.1 Process Area Incidents

4.2.1.1 Isobutane—Pasadena, TX

On October 23, 1989 a release of an estimated 81,000 lbs (36,800 kg) of hot isobutane from a vertical loop reactor occurred at this polyethylene plant. The resulting vapor cloud was estimated by one eye witness to be 600 ft (182 m) long and 150 to 200 ft (45.7 to 61 m) wide. Ignition, which was delayed from 2 to 5 min, probably occurred inside a finishing building or at an outdoor fired heater. The strong initiating indoor (confined) vapor cloud ignited the outdoor cloud in the highly congested process area. The resulting outdoor VCE severed process and fire protection piping, allowing a release of large quantities of flammable liquids and gases into the unprotected process area. Pressure effects were evident at 6 miles (9.6 km) from the epicenter. Damage to the polyethylene plant was total. Damage to remote process areas and other buildings on site was less significant although the main administration building, located in excess of 3000 ft (915 m) from the explosion, was severely damaged and eventually razed. The cloud may have produced as much as 18 tons (16.4 tonnes) TNT equivalent energy, assuming 5% efficiency. The initial release is reported to have occurred due to a failure to properly secure a valve during maintenance under a polyethylene reactor. The incident is estimated at $1 billion property damage (PD) and $700 million business interruption (BI).
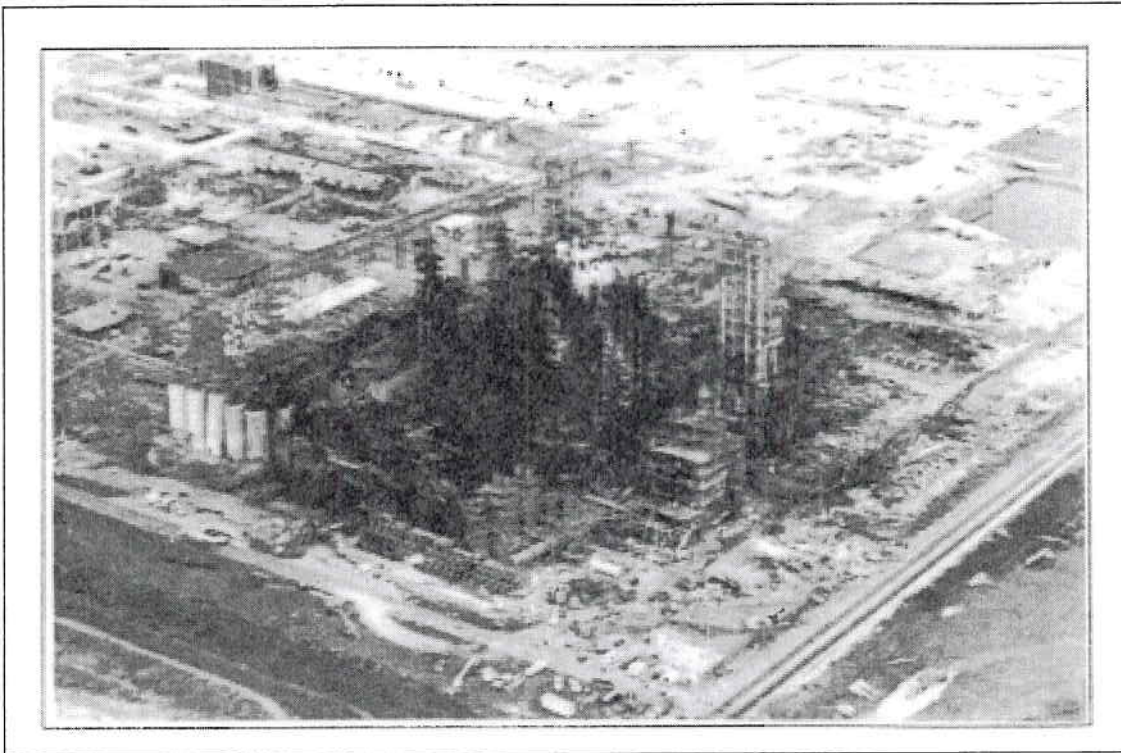

Fig. 5. VCE at Pasadena, Texas

4.2.1.2 Butane—Pampa, TX

On November 14, 1987, a VCE occurred after an estimated 10,000 lbs (4545 kg) of liquefied butane was released from a high pressure liquid phase oxidation reactor used for producing acetic acid. The vapor cloud drifted to a boiler where it was ignited. An ignition delay of 10 sec to 1 min was reported. Some sources estimate a 25% explosion efficiency which would produce approximately 12 tons (10.9 tonnes) TNT

equivalent. Severe ensuing fires occurred with deluge systems out of service. Over 500 buildings on site were destroyed or severely damaged and pressures were felt 25 miles (55 km) away offsite. The loss estimate is in excess of $250 million PD and $200 million BI.

4.2.1.3 C_3 Hydrocarbons — Norco, Lousiania

Information on this refinery incident, which occurred on May 5, 1988 when hot hydrocarbon vapors escaped under high pressure from an 8 in. (200 mm) pipe feeding a catalytic cracking unit, is not well reported. One report indicates a delay in ignition of a vapor cloud of 30 sec. Very extensive damage to the cracking unit and nearby process areas caused a shutdown for nearly a year. Many homes off premises were damaged and overpressures were felt 25 miles (55 km) away. Damage estimates are in excess of $ 300 million PD. BI estimates are not known.

4.2.1.4 Cyclohexane—Flixborough, England

The VCE at Flixborough, England on June 1, 1974 is perhaps the most widely known and reported, because of the immediate government enquiry and public investigation[4] after the event. The explosion caused an estimated $500 million in property damage to the plant, and major damage offsite. Business interruption figures are not reported, however, the plant, which produced caprolactum for the nylon industry, was never rebuilt. The explosion occurred when a 20 in. (500 mm) process pipe carrying cyclohexane at high temperature and pressure catastrophically failed. An estimated 33 tons (30 tonnes) of material escaped in the minute before ignition occurred. The ensuing explosion produced an energy equivalent estimated at 18 tons (16 tonnes) TNT. The VCE was followed by a vapor cloud fire which covered an area of a radius of 600 to 800 ft (180 to 250 m). A flammable liquid fire involving an estimated 450,000 gal (1705 cu m) ensued, destroying the entire facility. Investigations revealed that the 20 in. (500 mm) line that failed was a bypass temporarily installed when one reactor in a series of reactors was removed from service. The temporary bypass line received inadequate engineering study and failed under high stresses and vibration for which it was not designed.

4.2.1.5 Ethylene Releases

Two separate incidents occurred at this plant, one in 1971 and the second in 1976. The first, which occurred in a polyethylene plant, was a result of failure of a ½ in. (12.7 mm) sample line on a high pressure (3200 psi) (221 bar) ethylene gas system. Initially, about 1000 lbs (454 kg) of ethylene released into the highly congested process area and ignited, apparently on an exhaust system of a compressor. Second and third explosions ensued due to broken ethylene lines which released larger quantities of gas. The third explosion, involving "tons of material," caused the greatest far field pressure damage. The vapor cloud was estimated to have been 400 ft (122 m) by 250 ft (76 m) in size prior to the largest explosion. An estimated 42 buildings or structures were damaged, 12 extensively. ($10,000,000 PD, $5,000,000 BI.)

The second incident occurred in an ethanol reactor structure. Ethylene gas at approximately 1200 psi (83 bar) was used in the process. In the incident, an 8 in. (200 mm) process pipe failed, releasing an estimated 2.1 tons (1.9 tonnes) of vapor into the outdoor process area. Ignition occurred approximately 30 seconds after the initial release. A second, less intense explosion may have occurred. The resulting overpressure damaged many buildings and impaired deluge systems, allowing subsequent fires to cause additional damage. The ethanol plant was shut down for about four months. Investigation revealed that the 8 in. (200 mm) process line showed evidence of internal corrosion caused by upstream process conditions. The line ultimately failed under high pressure. ($40,000,000 PD, $10,500,000 BI.)

4.2.2 High Pressure Transmission Pipeline Incidents

4.2.2.1 Natural Gas Liquids—Brenham, Texas

On April 7, 1992 a high pressure pipeline carrying mixed liquefied petroleum gases (LPG)—primarily ethane, propane, and butane—suddenly failed in this remote, rural area. The pipeline was associated with a large underground salt dome storage facility. The release may have occurred at an aboveground valve where a 6 in. (150 mm) pipe joined a 4 in. (66 mm) pipe. The force of the escaping gases cratered the ground around the pipe and the release lasted several hours. The vapor cloud covered a large area and settled into shallow ravines before it was ignited, possibly by a vehicle being driven through the cloud. The explosion was apparently enhanced by the confinement of the ravines and numerous trees. The explosion destroyed many homes in the region, uprooted or snapped trees, and was felt as far away as 120 miles (192 km). The National

Fig. 6. VCE at Flixborough Plant

Transportation Safety Board (NSTB) is investigating the incident. The NTSB also estimated that 1000 bbls or approx 225,000 lbs (103,000 kg) of LPG may have been available for release. No further details or dollar amounts are available.

4.2.2.2 Ethane Propane Mix—Baton Rouge, Louisania

An aboveground pipeline at approx 700 psi (47.6 bar) carrying a mix of ethane and propane from a refinery to nearby, off-site storage facilities, catastrophically failed during record low temperatures on December 24, 1989. The resulting vapor cloud, which was released in a remote tank farm, travelled an estimated 1500 ft (457 m) across open land into the refinery where it was ignited at a fired heater. The resulting burning cloud flashed back to the tank farm where obstacles and confinement in the form of piperacks and a rail underpass were sufficient to cause flame speed transition to deflagration. The VCE ignited storage tanks, causing ensuing fires and produced blast pressures felt six miles (10 km) away. The loss is estimated at $45 million. BI figures are not available.

4.2.2.3 Propane—Port Hudson, MO

This well reported event[5] occurred on December 9, 1970 in rural farmland when a buried 942 psi (64 bar) propane pipeline ruptured, releasing more than 50 tons (46 tonnes) of material in 24 minutes. The cloud filled a narrow, shallow valley and was concluded to have ignited in a rural warehouse. The official report prepared

by the US Bureau of Mines stated that the blast damage could only have occurred if the cloud detonated. If so, this is the only documented vapor cloud detonation on record. Damage was minimal due to the remote area.

4.2.2.4 Propane, Mixed Gas Liquids—Siberia

On June 3, 1989 a buried pipeline failure occurred in a remote region near Ufa, Siberia. Escaping liquefied gases were noticed for several hours, but material was allowed to continue to flow into the system by operators at a remote control station. The massive cloud, estimated at millions of pounds, was ignited by a locomotive on a remote rail line. The cloud was estimated by the military to have produced 10,000 tons TNT equivalent energy, reportedly the largest non-nuclear explosion ever recorded. Trees were uprooted in a several mile radius and windows were broken at 8 miles (14 km) from epicenter. No industrial damage occurred due to the remote location of the release.

4.2.3 Fixed Storage Tank Incidents

There have been very few VCE events initiated as a result of release from a storage tank, none recent.

4.2.4.1 Light Hydrocarbon Liquids—Pernis, Netherlands

From 60 to 120 tons (55 to 109 tonnes) of light hydrocarbon liquids were released January 20, 1968 when a slop oil tank frothed over due to breaking of an oil water emulsion in the tank. The tank failed at the seam due to the frothing action. The resulting cloud diameter was 450 ft (137 m). The explosive yield was estimated at 4 to 8% and 22 tons (20 tonnes) TNT equivalent. The loss exceeded $100 million PD. BI is not reported.

4.2.4.2 Winter Grade Gasoline—Port Newark, New Jersey

On January 7, 1983, gasoline with a high winter grade component of butane was being transferred from a pipeline to a storage tank at this terminal. Due to inattention by operators, the tank was allowed to overfill, forming a cloud which travelled approximately 1500 ft (457 m) into an adjacent warehouse where it ignited. The explosion knocked over locomotives at 600 ft (182 m) and caused extensive blast damage and ensuing fires. Loss is estimated at $50 million PD. BI is not reported.

4.2.4 Mobile Transportation Incidents

VCE incidents causing damage to industrial facilities which have involved transportation vehicles, such as railcars, tank trucks, or waterway vessels have also been infrequent. Most occurred in rural areas or as railyard incidents, causing limited industrial damage.

4.2.5.1 Truck Tanker, ethylene—Baton Rouge, LA

Approximately 8000 lbs (3640 Kg) of ethylene gas was released from a tanker truck on January 19, 1971 in rural Louisiana following failure of a relief valve to open (later determined to be frozen). Two explosions were heard. Damage was felt for about 6 miles (9.6 km) and local industrial buildings of lightweight construction were heavily damaged. Loss estimates are not reported.

4.2.5.2 Railcar, Liquefied Petroleum Gas—East St. Louis, IL

On January 22, 1972, a railcar full of liquefied petroleum gas (LPG) (primarily propylene) rolled at high speed into another railcar during railyard switching operations. The collision occurred at approx 15 mph (24 kph), approx 10 mph (16 kph) faster than normal practice. The coupling on a stationary car punctured the moving LPG car and released 24,300 lbs (53,500 kg) of vapor at 220 psi (15.1 bar) into the railyard. The vapor cloud covered an area of more than 215,000 sq ft (20,000 sq m) and reportedly ignited in two different locations. The explosion was enhanced by the tightly congested arrangement of railcars throughout the large railyard. The US Department of Transportation report[6] indicates an explosion efficiency of 3% and a TNT yield of 2.75 tons (2.5 tonnes). An estimated 1000 buildings, including schools, were damaged to various degrees. The economic loss is estimated at $45 million.

5.0 CASE STUDIES

5.1 Case Study Overview

XYZ Chemical, Inc. produces a plastic polymer in Process Block 1 (Fig. 7a). The primary flammable material is liquefied butane which is used as the carrier solvent and processed at 300 psig (2170 kPa) and 250°F (120°C) within the reactor system. A series of vessels, reactors, and piping in the multi-level open process structure have a combined holdup of 25,000 lbs (5165 gal) (11,360 kg [19.7 cu m]) of butane. Piping, the largest being 6 in. (152 mm), between the reactor and a condenser and located 15 ft (4.6 m) below the highest liquid holdup, interconnects all process vessels.

Approximately 800 feet (244 m) from Process Area 1 is a 250,000 gal (946 cu m) butane sphere pressurized to 40 psig (380 kPa), used for makeup butane solvent carrier lost in the process. Liquid butane from the sphere is pumped at a maximum flow rate of 100 gpm (379 l/min) to Process Area 1 reactors via a 3 in. (75 mm) aboveground pipe. The pipe, which is located beneath the tank and 39 ft (12 m) below the highest liquid level, has an excess flow valve (EFV) connected at the first flanged nozzle. The EFV is well protected against mechanical damage and vehicular traffic. The EFV will allow a maximum flow of 150 gpm (586 l/min) before it automatically closes. The EFV interior orifice is equivalent to a 1.5 in (37.5 mm) opening. The tank is enclosed by a 50 x 50 ft (15.2×15.2 m) dike of earth construction.

Process Block 1 is a highly congested open process structure with multi-levels and extensive piperacks which form a confined area. Essentially the entire process block is full of vessels, pipes, pumps, and other equipment creating a continuous obstacle array.

To solve the problem, the following steps should be followed:

1. Define worst credible case release scenario (refer to Section 3.2)

2. Calculate vapor release amount (W_v) (refer to Section 3.3).

3. Calculate TNT Equivalency (refer to Section 3.4, Equation 7).

4. Calculate overpressure radii (refer to Section 3.5, Equation 8).

5. Determine cloud drift and epicenter of explosion (refer to Section 3.6).

6. Plot overpressure circles.

5.2 Define Worst Credible Case Release Scenario

While many release scenarios possibly exist, two of importance are quickly identified as requiring further evaluation to determine worst case:

Scenario A involves a release under high pressure from a break in the 6 in. (152 mm) pipe on the reactor system in Process Area I. **Scenario B** involves a break in the pipe under the 250,000 gal (946 cu m) butane sphere located remotely from the process area.

5.2.1 Scenario A (Process System Release)

Table 8. Information needed for Scenario A is tabulated as follows.

Data Info	English Units	Metric Units	Source of Data
Process Press. (P_1)	300 psig	2.17×10^6 Pa	Process Info
Process Temp. (T_1)	250° F	120° C	Process Info
Release Height (h)	15 ft	4.6 m	Process Info
Release Area (A_r)	28.3 in²	0.018 m²	Process Info
Ambient Temp. (T_a)	70° F	21° C	Assumed
Ambient Press. (P_a)	14.7 psia	1.01×10^5 Pa	Assumed
ATM BP (T_b)	31° F	-0.5° C	Table 1
Specific Heat (C_{p1})	—	2833 J/kg °K (@ T_1)	Gallant
Specific Weight	4.84 lbs/gal	580 kg/m³	LPDS 7-55
Heat Vap. (H_{vapb})	—	3.89×10^5 J/kg (@BP)	Table 1
Liq. Density (P_t)	—	425 kg/cu m (@ T_1)	Gallant
Heat Comb. (ΔH_c)	19,700 BTU/lb	10,940 Kcal/kg	Table 1

5.2.1.1 Calculate mass of liquid released

Since the material is in the system as a liquid, first solve for W_t, mass of liquid released, using Simplified Equation 2, Section 3.3.2.1:

$$W_t = 372 A_r \rho_t \sqrt{\frac{2(P_1 - P_a)}{\rho_t} + 19.6h}$$

$$= 372(0.0182)(425) \sqrt{\frac{2(2170 \times 10^3 - 101 \times 10^3)}{425} + (19.6)(4.6)}$$

$$= 285,200 \text{ kg}$$

This exceeds the initial vessel contents so the liquid amount discharged, W_t, subject to subsequent vaporization and pool-rain out is 11,360 kg (25,000 lbs) (entire system contents).

The discharge time for the process system may be estimated by interpolation.

$$\frac{11,360 \text{ kg}}{285,200 \text{ kg}} \ 600 \text{ sec} = 24 \text{ sec}$$

5.2.1.2 Calculate flash fraction

To determine the amount of the released hot liquid which will flash vaporize (F_{vap}), use Equation 3 from Section 3.3.2.2:

$$F_{vap} = \frac{C_{p_1}(T_1 - T_b)}{H_{vap_b}} = \frac{2833(120 - [0.5])}{389,700} = 0.88 = 88\%$$

To account for liquid entrainment (aerosol) in the flashing liquid, double F_{vap} (or maximum 100% in this scenario) to use in Step 4, TNT Equivalent Energy equation. Equations for spill pool vaporization and W_{boil} need not be used for this scenario because none of the material will rain out and form a pool spill. The entire initial release, which is the entire contents, will vaporize and contribute to the vapor cloud.

Concluding, total mass of vapor, W_v, in the cloud is 11,360 kg (25,000 lbs), or the entire contents.

5.2.2 Scenario B (Storage Sphere)

The tank scenario represents a spill potential from a break in the line under the tank. Because the EFV has a 1.5 in. (37.5 mm) effective orifice, this restricted opening can be used for solving W_1, mass of liquid released.

Table 9. Information needed for Scenario B is tabulated as follows.

Data Info	English Units	Metric Units	Source of Data
Process Press. (P_1)	40 psig	3.77×10^5 Pa	Process Info
Process Temp. (T_1)	70° F	21° C	Process Info
Release Height (h)	39 ft	12 m	Process Info
Release Area (A_r)	1.76 sq in.	0.001 sq m	Process Info
Ambient Temp. (T_a)	70°F	21°C	Assumed
Ambient Press. (P_a)	14.7 psia	1.01×10^5 Pa	Assumed
ATM BP (T_b)	31°F	−0.5°C	Table 1
Specific Heat (C_{p1})	—	2450 J/kg °K (@ T_1)	Gallant
Specific Weight	4.84 lbs/gal	580 kg/cu m	LPDS 7-55
Heat Vap (H_{vapb})	—	3.90×10^5 J/kg (@BP)	Table 1
Liq. Density (P_1)	—	560 kg/cu m (@ T_1)	Gallant
Heat Comb (ΔH_c)	19,700 BTU/lb	10,940 Kcal/kg	Table 1

5.2.2.1 Calculate mass of liquid released

Since the material is in the system as a liquid, first solve for W_1, mass of liquid released, using Simplified Equation 2, Section 3.3.2.1:

$$W_t = 372 A_r \rho_t \sqrt{\frac{2(P_1 - P_a)}{\rho_t}} + 19.6h$$

$$= 372(0.001)(560) \sqrt{\frac{2(3.77 \times 10^5 - 1.01 \times 10^5)}{560}} + (19.6)(12)$$

$$= 7279 \text{ kg}$$

As can be seen, only a small fraction of the entire contents of the 250,000 gal (943 cu m) sphere is released in 10 min. This liquid amount, W_1, can be further described by the flash fraction calculation below to define what portion will immediately vaporize and what portion will rain out into a pool spill.

5.2.2.2 Calculate flash fraction:

To determine the amount of the released liquid which will flash vaporize (F_{vap}), use Equation 3 from Section 3.3.2.2:

$$F_{vap} = \frac{C_{p_1}(T_1 - T_b)}{H_{vap_b}} = \frac{2450(21 - [-0.5])}{389,700} = 0.135 = 13.5\%$$

To account for liquid entrainment (aerosol mist formation) in the vapor phase, F_{vap} is doubled, to 27%. Therefore 27% of the 7279 kg (16,014 lb) of butane released, or 1965 kg (4324 lbs), will vaporize. The remaining 5314 kg (11,691 lb) of butane will not flash vaporize, but instead will rain out into a liquid pool. To calculate vaporization from the liquid pool, Equations 4, 5 and 6 in Sections 3.3.2.3, 3.3.2.4, and 3.3.2.5 are applied as follows.

5.2.2.3 Calculate pool size

Because the spill is assumed to be contained in the 50× 50 ft (15.2×15.2 m) earthen dike, the spill is confined. Equation 4 could be used to determine spill area, but would approximate a spill greatly exceeding the dike area. Instead, simple mathematics can be used to determine that the 323 cu ft (9.16 cu m) release will fill the 2500 sq ft (231 sq m) dike to a depth of 0.13 ft (0.04 m).

To determine surface area of the spilled liquid subject to vaporization, both the area directly on the ground as well as the area along the dike walls (assumed vertical) must be applied. The dike floor represents 2500 sq ft (231 sq m) and the interior walls an additional 26 sq ft (2.4 sq m). Total area (A) is therefore 2526 sq ft (233 sq m).

5.2.2.4 Calculate Liquid Pool Evaporation

Using Simplified Equation 5, Section 3.3.2.4:

$$W_{boil} = \frac{27.6(233)(21-[-0.5])5400}{3.89\times10^5} = 1919 \text{ kg} = 4222 \text{ lbs}$$

5.2.2.5 Calculate total amount of vapor in cloud

The total amount of vapor in the cloud, W_v, is the product of the initial flash fraction ($2F_{vap}$) and the amount vaporized in the pool (W_{boil}) (Equation 6):

$$W_v = 2(0.135)(7279 \text{ kg}) + 1919 \text{ kg} = 3884 \text{ kg} = 8546 \text{ lbs}$$

5.3 Calculate Vapor Release Amount

Scenario A (process area release) represents a far greater release potential than Scenario B (storage sphere release). Scenario B also does not produce 5 tons (4.5 tonnes) of vapor (W_v) and thus would not qualify for further VCE evaluation as established by threshold values set forth in Section 3.1.3.

Therefore W_v=25,000 lbs (11,360 kg) solved above for Scenario A is used for determining worst case TNT energy release in Equation 7, Section 3.4.

It should be noted that in some cases it is not readily apparent which release in a facility represents the worst credible case event without completing a more thorough evaluation or without screening additional release events. It should also be reminded that this method does not quantitatively establish what event is the more probable or greatest overall risk.

5.4 Solve for TNT Equivalency (W_e)

5.4.1 English units

Using Equation 7a, Section 3.4.1:

$$W_e = \frac{W\Delta H_c f}{4\times10^6}$$

$$= \frac{(25,000 \text{ lbs})(19,700 \text{ Btu/lb})(0.05)}{4\times10^6}$$

=6.2 tons (12,400 lbs) TNT

5.4.2 Metric units

Using Equation 7b, Section 3.4.2:

$$W_e = \frac{W\Delta H_c f}{1.111\times10^6}$$

$$= \frac{(11,360 \text{ kg})(10,930 \text{ Kcal/kg})(0.05)}{1.111\times10^6}$$

=5.6 tonnes (5580 kg) TNT

5.5 Solve for Radial Distance (R_g) at Overpressure (P)

5.5.1 English Units

For example, from Table 4a at 10 psi incident overpressure, the scaled ground distance, Z_g, is 9.8 ft/lb$^{1/3}$. Solving for R_g using Equation 8:

$$R_g = Z_g W_e^{1/3}$$

where $W_e = 12,400$ lbs
and $W_e^{1/3} = 23$ lbs$^{1/3}$

$$R_g @ \text{ 10 psi} = 9.8 \ \frac{ft}{lb^{1/3}} \ (23 \ lb^{1/3}) = 225 \ ft$$

The radial distance at each selected pressure is determined in turn and charted:

Pressure (P) (psig)	Sclaed Ground Distance (Z_g) (ft/lbs$^{1/3}$)	$W_e^{1/3}$ (lbs$^{1/3}$)	Distance From Epicenter (R_g) (ft)
10	9.8		225
6	13		299
5	14.5		334
3	19.5	23	449
2	26		598
1	45		1,035

5.5.2 Metric Units

For example, from Table 4b at 0.75 barg incident overpressure, the scaled ground distance, Z_g, is 3.7 m/kg$^{1/3}$. Solving for R_g using **Equation 8**:

$$R_g = Z_g W_e^{1/3}$$

where $W_e = 5580$ kg
     $W_e^{1/3} = 17.7$ kg$^{1/3}$

$$R_g @ \text{ 0.75 bar} = 3.7 \ \frac{m}{kg^{1/3}} \ (17.7 \ kg^{1/3}) = 66 \ m$$

The radial distance in metric units at other overpressures is calculated and charted:

Pressure (P) (barg)	Scaled Ground Distance (Z_g) (m/kg$^{1/3}$)	$W_e^{1/3}$ (kg$^{1/3}$)	Distance From Epicenter (R_g) (m)
0.75	3.7		66
0.50	4.6		82
0.40	5.2		92
0.30	6.1	17.7	108
0.20	8.0		142
0.10	13.5		239

5.6 Determine Cloud Drift and Explosion Epicenter (EC)

The explosion epicenter of overpressure circles is developed by determining an *explosion vulnerability zone* which is defined by estimating cloud drift as covered in Section 3.6.

Based on a release of 12.5 tons (25,000 lbs) of vapor of a Class I material, Figure 2 produces a cloud drift radius of 500 ft. In *metric units*, a release of 11.4 tonnes (11,400 kg) of vapor produces a cloud drift radius of 150 m (Figure 3).

This drift radius is plotted, using release point R as a center point, on a site plan as shown in Figure 7b. It should be assumed that the actual explosion epicenter is within this vulnerability zone and around the release point. Using judgment, an explosion epicenter which creates the greatest exposure to plant facilities is selected. Care is taken to assure that the assumed epicenter is selected in a congested area rather than in an open area. Figure 7c demonstrates this procedure.

5.7 Plot Overpressure Circles

Using assumed epicenter (EC) as a center point, overpressure circles are drawn on a site plan as shown in figure 7d. Damage to areas (or individual buildings) are assigned using Tables 5, 6, and 7 as guidance.

Damage is estimated, depending on known or assumed ability of buildings or equipment to withstand overpressure, amounts of flammable materials holdup, other combustible loading, and passive features such as spacing and fireproofing of steel structures.
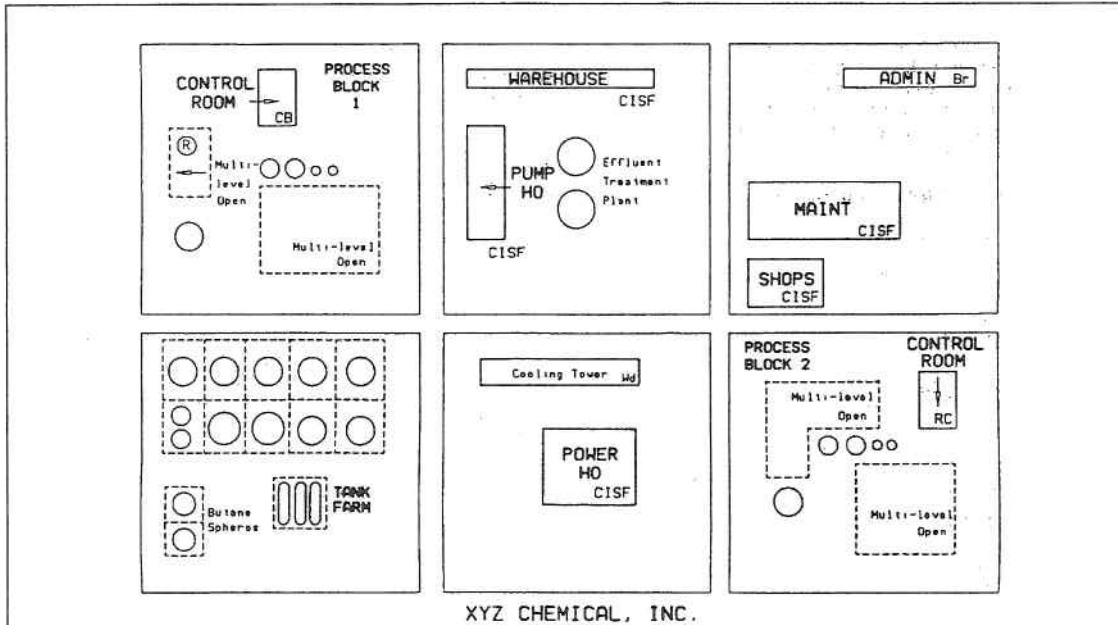
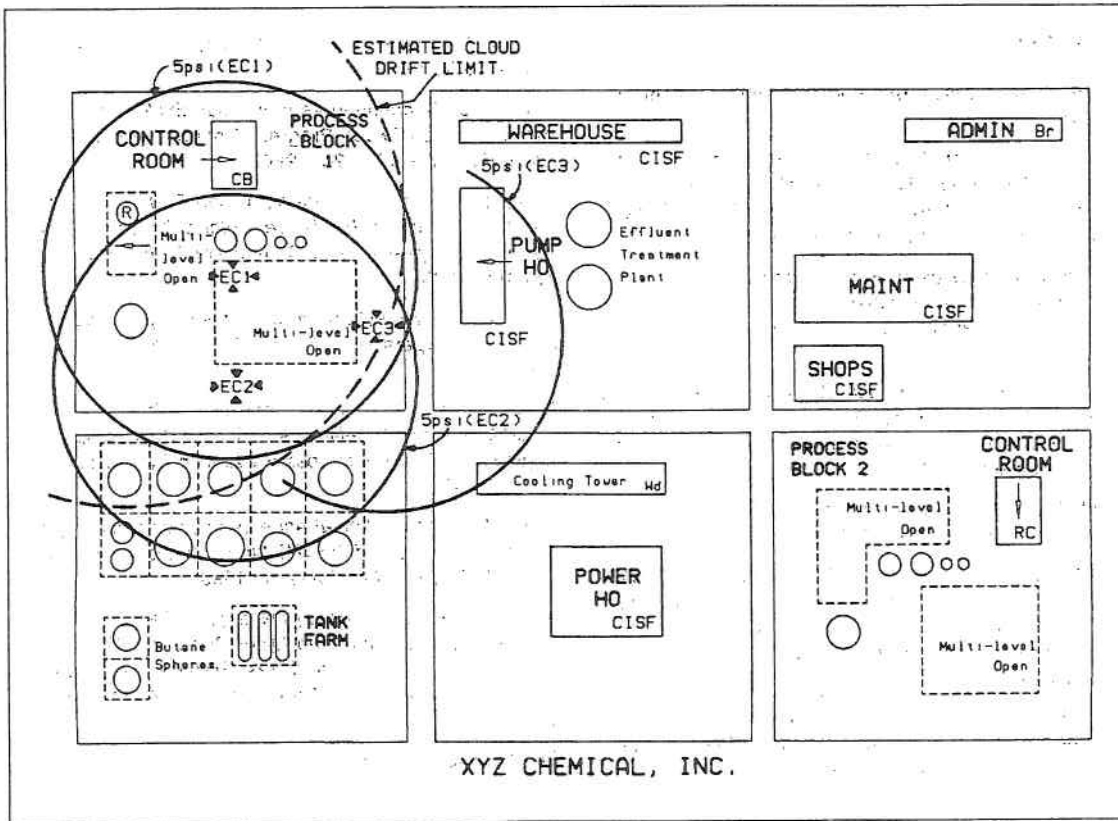Fig. 7a. Site plan.
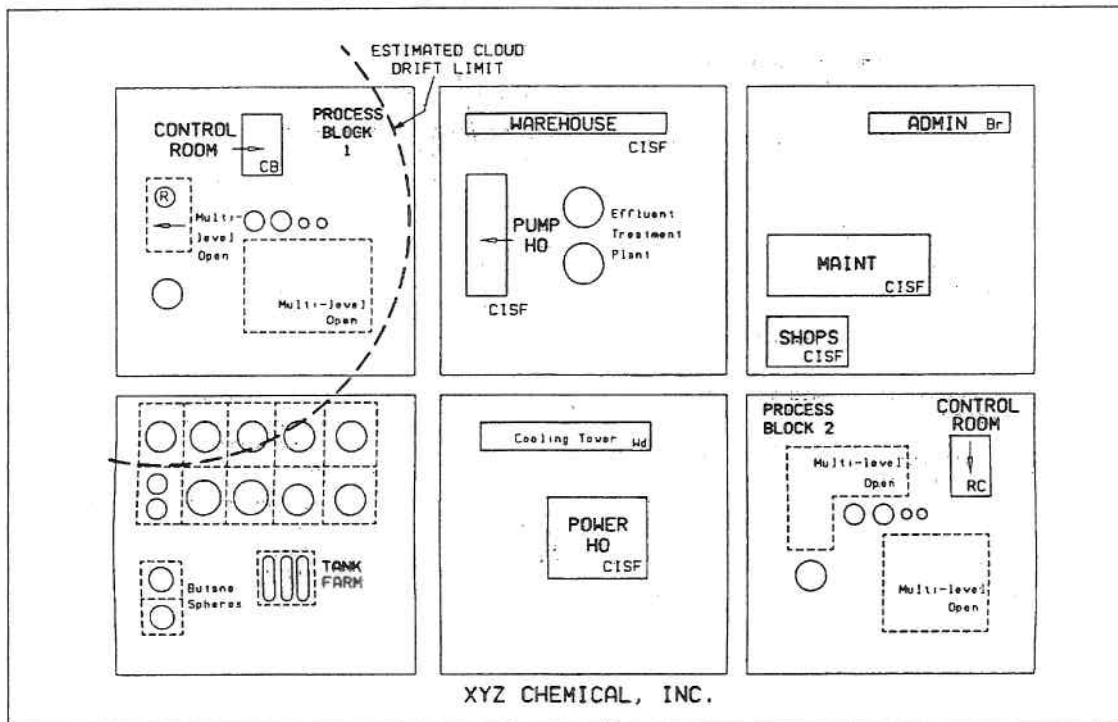
Fig. 7b. Cloud drift limit.

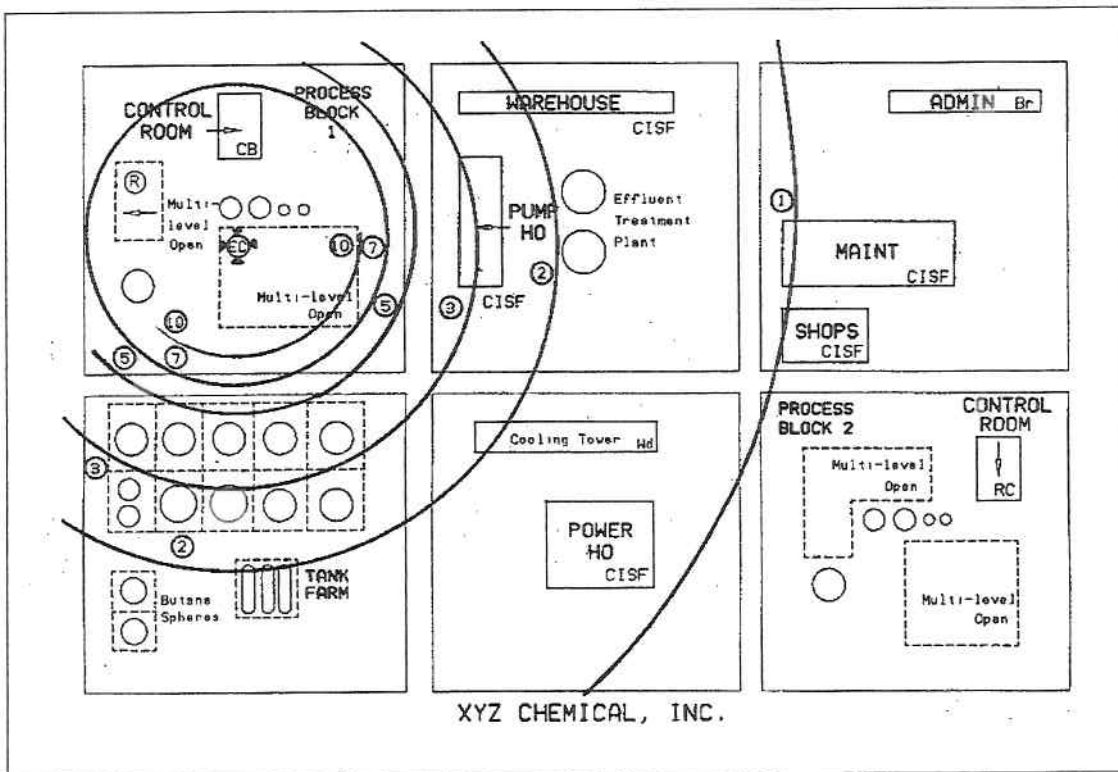Fig. 7c. Possible epicenter locations.

Fig. 7d. VCE blast overpressure profile.

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APPENDIX A DOCUMENT REVISION HISTORY

May 2008. Minor editorial changes were made.

May 2005. Editorial corrections to Table 7B.

January 2001. Editorial corections to Figure 2, Table 1 added to compounds.

September 1998. Minor reformatting and re-issue as Data Sheet 7-42.

January 1998. Editorial corrections.

September 1997. Editorial corrections to Table 1.

April 1994. Initial issue as Data Sheet 7-0S and 7-0SC.