National Iranian Petrochemical Co.
HSE Directorship

1 Process Safety

Design Basis Safety Concepts For Petrochemical Plants and Projects

IN THE NAME OF GOD

1 PROCESS SAFETY

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پیشگفتار

تولید پایدار و ایمن همراه با توجه ویژه به مقوله سلامت کارکنان و حفظ محیط زیست جزء اولویت های اساسی در واحدهای صنعت پتروشیمی می باشد، که یکی از عوامل مهـم در تحقق اهـداف فوق، لحاظ نمودن اصـول ایمنی فرآیند و طراحـی ذاتاً ایمن این تاسیسات می باشد.

با توجه به روند توسعه صنعت پتروشیمی توسط بخش خصوصی در کشور و تنوع صاحبان لیسانس و مشاوران/ طراحان بعضاً شاهد سطوح متفاوتی از طراحی ذاتاً ایمن توسـط این مجریان هستیم. از آنجائیکه شرکت ملی صنایع پتروشیمی (NPC) نقش حاکمیتی در توسعه این صنعت را بعهده دارد، لذا بمنظور اطمینان از یکپارچگی و لحاظ شـدن حداقل الزامات در طـرح ها و پروژه ها در طی مراحل مختلف انتخاب لیسـانس، طراحی پایه و مفهومی و عملیات سـاختمان و نصب، این مدیریت نسبت بـه گردآوری و تدوین الزامات پایه ایمنی فرآیند شـامل ۱۲ عنـوان مدرک به عنوان سـند بالادستی در قالب مجموعه ای سـه جلدی اقدام نموده است که لازم است کلیه مشـاوران/ طراحان و پیمانکاران جهت ایجاد محیطی ایمن و پیشـگیری از حوادث احتمالی از این الزامات پیروی نمایند.

قدرت ا... نصیری
مدیر بهداشت، ایمنی و محیط زیست

National Iranian Petrochemical Co. HSE Directorship.	**Design Basis Safety Concepts for Petrochemical Plants & projects** DOCUMENT COVER SHEET

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NPC-HSE-S-01

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1.Scope

This document describes the general safety considerations (as minimum requirement) that shall be considered in overall plant design.

2. References

- API RP 500, Classification of Locations for Electrical Installations at Petroleum Facilities
- API 650, Welded Steel Tanks for Oil Storage
- API 2000, Venting Atmospheric and Low Pressure Storage Tanks
- API 2003, Protection Against Ignition Arising out of Static, Lightning, and Stray Currents
- API 2216, Ignition Risk of Hydrocarbon Vapors by Hot Surfaces on the Open Air
- NFPA 30, Flammable and Combustible Liquids Code
- NFPA 68, Explosion Venting Guide
- NFPA 70, National Electrical Code
- NFPA 325, Fire Hazard Properties of Flammable Liquids, Gases and Volatile Solids
- NFPA 496, Purged Enclosures for Electrical Equipment in Hazardous Locations, National Fire Protection Association
- NFPA497, Classification of Flammable Liquids, Gases or Vapors and of Hazardous (Classi-fied) Locations for Electrical
- Installations in Chemical Process Areas
- NFPA 499, Classification of Combustible Dusts and of Hazardous (Classified) Locations for Electrical Installations in Chemical
- Process Areas
- NFPA505, Powered Industrial Trucks Including Type Designations, Areas of Use, Mainte-nance, and Operation
- Exxon mobile engineering standards

3. Definitions

Auto-Ignition Temperature

The Auto-Ignition Temperature, AIT is the lowest temperature required to cause self-sustaining combustion, without initiation by spark or flame. Auto-ignition temperatures of the majority of hydrocarbons fall in the range of 400 to 1000°F (200 to 500°C) **(see appendix 1)**. In certain phases of safety design, an arbitrary auto-ignition temperature of 600°F (315°C) is used. This is judged to be a conservative estimate in the absence of experimental data. When experimental data is available, the actual value of the AIT or 600°F (315°C), whichever is lower, should be used as the governing criterion for equipment spacing and any other design features where AIT is a consideration.

BLEVE

BLEVE is standing for Boiling Liquid Expanding Vapor Explosion. This type of explosion occurs if a vessel containing superheated hydrocarbon liquid fails catastrophically when fire exposure results in overheating and yielding of a pressure vessel.

Contingency

Contingency is an abnormal event which causes an emergency. A single contingency is a single abnormal event causing an emergency. A remote contingency is the result of a single extremely low probability event, or of two remotely related events which may happen to occur simultaneously. A double contingency would be the simultaneous occurrence of two or more unrelated events between which there is no process, mechanical, or electrical inter-relationship.

Critical Exposure Temperature

When the metal temperature of equipment or piping is below the CET, there is a risk of brittle fracture if the stresses from the operating pressure or other loads exceed some percentage of the design allowable stress.

Deflagration

Deflagration is an explosion in which the burning process of the flammable mixture (vapor or dust) is governed by heat and mass transfer. The resulting flame front propagates below sonic velocity and the pressure increase (if in a vessel) is typically between 8 to 10 times the initial pressures.

Vapor Cloud Explosion (VCE)

VCE can occur if a gas or vapor release forms a large flammable cloud in a congested area. Ignition may be caused by the cloud drifting to a fired heater or ignition source outside the plant. The resulting explosion can be devastating.

Detonation

Detonation is a high severity explosion. In this type of explosion, the propagating flame front travels as a supersonic shock wave that is closely followed by a combustion zone which releases the energy needed to maintain the shock wave. The resulting pressure rise may be between 100 to 1000 times the initial pressure. Highly destructive energies are produced if pressure wave formation, reflection, and impact reinforce one another in confined areas. Detonations are unlikely to develop in unconfined areas.If a highly confined flammable vapor/air mixture is ignited, the probability of the resulting explosion being a detonation is increased if:

- *Fuel is rich in hydrogen, acetylene, ethylene, ethylene oxide or propylene oxide.*
- *Pure oxygen, or oxygen enriched air, is present.*
- *Vapor/air mixture is turbulent.*

- *The mixture is contained in equipment of long narrow shape, e.g., piping.*
- *The mixture is initially at high pressure and temperature.*
- *The mixture is in detonable range (narrower than flammable range and approximately equal to stochiometric conditions).*

Detonation arresters can be used to design against the propagation of a detonation in low pressure (near atmospheric) piping systems such as vapor recovery systems. Another way of stopping detonations in such systems is to rapidly close a valve which then prevents passage of the detonation wave. To do this, the pressure wave must be remotely sensed far enough upstream so that the valve has time to shut. If pressures higher than 2 -3 psig (14 -20 kPa) are involved it is not feasible to mechanically design against detonations and the only valid approach is to avoid conditions that favor detonations.

Emergency

An emergency is an interruption from normal operation in which personnel, equipment, or the environment may be endangered or harmed in some manner.

Exposure Limits

Occupational exposure limits are used for evaluating exposures to toxic hazards and assessing the potential need for control. Occupational exposure limits usually refer to airborne concentrations of substances and represent conditions under which it is believed that nearly all workers may be repeatedly exposed day after day without adverse health effects. A widely accepted set of exposure limits are the Threshold Limit Values (TLVs), which are regularly updated and published annually by the American Conference of Governmental Industrial Hygienists (ACGIH). Three categories of TLVs are used: Time Weighted Average (TLV-TWA), Short-Term Exposure Limit (STEL), and Ceiling Limit (TLV-C):

- ***Threshold Limit Value - Time-Weighted Average (TLV-TWA)** - The time-weighted average concentration for a conventional 8-hour workday and a 40-hours workweek, to which it is believed that nearly all workers may be repeatedly exposed, day after day, without adverse effect.*
- ***Threshold Limit Value - Short-Term Exposure Limit (TLV-STEL)** - A STEL is defined as a 15-minute TWA exposure that should not be exceeded at any time during a workday even if the 8-hour TWA is within the TLV-TWA. Exposures above the TLV-TWA up to the STEL should not be longer than 15 minutes and should not occur more than four times per day. There should be at least 60 minutes between successive exposures in this range. An averaging period other than 15 minutes may be recommended when this is warranted by observed biological effects.*
- ***Threshold Limit Value - Ceiling (TLV-C)** - The concentration that should not be exceeded during any part of the working exposure. TLVs are based on an 8-hour workday and a 40-hour*

work week. For unusual work shifts or work weeks, the exposure limits should be adjusted. To determine the appropriate exposure limit, consult your local Industrial Hygiene contact, regional Industrial Hygiene Coordinator.

Flammable (or Explosive) Limits

Flammable (or explosive) limits are the minimum and maximum concentrations (expressed as volume fraction or %) of a flammable vapor (or aerosol/mist) in air which is capable of supporting combustion. These limits are usually abbreviated to LFL (Lower Flammable Limit) and UFL (Upper Flammable Limit) or LEL/UEL. Flammable limits are normally given at atmospheric conditions. An increase in oxygen contents will widen the flammable range (i.e., the LFL will be lower and the UFL will be higher). A decrease in oxygen contents (adding inert gas) will narrow the flammability range to the point where combustion stops. Increase in pressure or temperature will also widen the flammable range (or reduce the amount of oxygen required to support combustion). Typical Flammable Limits are listed in Appendix 1. In the case of combustible dusts, the lower flammable limit is the minimum concentration of the dust in the air (usually as weight per unit volume) that will propagate a flame. There is no corresponding upper flammable limit and any suspension of a combustible dust above the LFL should be considered flammable. Flammable limits are normally recorded at atmospheric pressure and temperature conditions. Increased temperature and/or pressure will lower the LEL.

- *Flammable Liquids are liquids with a closed-cup flash point below 100°F (38°C), and liquids having a closed-cup flash point of 100°F (38°C) or higher when handled at temperatures above or within 15°F (8°C) of their flash points. The National Fire Protection Association (NFPA) classifies liquids with a closed-cup flash point below 100°F (38°C) as Class I liquids. Class I liquids are further subdivided into Class IA, IB and IC depending on their actual flash point and boiling point. Refer to NFPA 30, Flammable and Combustible Liquids Code for additional information on the classification of flammable liquids.*
- *Combustible Liquids are liquids having a closed-cup flash point of 100°F (38°C) or higher when handled at temperatures lower than their flash point minus 15°F (8°C). The NFPA classifies liquids having a closed-cup flash point of 100°F (38°C)or higher as Class II, Class IIIA or Class IIIB liquids depending on their actual flash point. Refer to NFPA 30, Flammable and Combustible Liquids Code for additional information on the classification of combustible liquids.*

Flash Point

Flash point is the lowest temperature at which a liquid exposed to the air gives off sufficient vapor to form a flammable mixture near the surface of the liquid, or within the test apparatus used, which can be ignited by a suitable flame. The flash point is related to the lower flammable limit (LFL). The composition at LFL can be calculated from the vapor pressure curve and the flash point temperature. Typical flash points can be found in appendix 1.

International Practices distinguish between high flash and low flash stocks, particularly for atmospheric storage. Note that other codes may use different limits.

- *Low-Flash Stocks are those having a closed-cup flash point below 100°F (38°C) and any other (high flash) stocks if handled at temperatures above, or within 15°F (8°C) below, their flash point. For example, a stock with a closed-cup flash point of 150°F (65°C) at a temperature of 135°F (57°C) or higher is treated as a low-flash stock.*
- *High-Flash Stocks are those having a closed-cup flash point of 100°F (38°C) or higher when handled at a temperature not higher than their flash point minus 15°F (8°C).*

Hazard

A chemical or physical condition that has the potential for causing harm to people, property or the environment.

High Integrity Protective System (HIPS)

An arrangement of instruments and other equipment, including sensors, logic controllers and final control elements used to isolate or remove a source of pressure from a system or to trip a shutdown or Depressuring device such that the design pressure and/or temperature of the protected system will not be exceeded. HIPS are a safety critical system and must be independent from all other control schemes and from shutdown systems whose failure can lead to an event requiring HIPS activation. Functionally, a HIPS must provide equal or lower (better) unavailability on demand than a typical pressure relief valve. This can be achieved by specifying a Safety Integrity Level (SIL) of 3.

Light Ends

Light ends are volatile flammable liquids which are significantly vaporized at normal ambient conditions. This indicates a type of material of greater fire hazard than heavier hydrocarbons because of the large volume of vapor generated by a liquid leak or spill. For the purposes of this document, the definition of light ends is a material having a Reid Vapor Pressure (RVP) of 15 psia (103 kPa) or greater, as determined by the standard ASTM D-323 test. By common usage this covers the following:

- *Pentane and lighter hydrocarbons (either pure hydrocarbons or mixtures).*
- *Unstabilized naphthas which meet the RVP criterion.*
- *Flammable chemicals which meet the RVP criterion.*

When used as a criterion of hazardous properties for safety design purposes, the term is applied to the above materials only when they are in the liquid phase or a combination of liquid and vapor phases. A process unit is considered to be a light ends unit when a significant part of the equipment handles light ends. Pure hydrogen is not considered a light ends material.

Pyrophoric Material

Pyrophoric materials are those that rapidly oxidize when exposed to air and which may incandesce, thus forming a source of ignition. Pyrophoric materials are known to form inside equipment and in the absence of oxygen. Typical examples are Iron Sulfide and Carbonaceous Materials. Ni-catalysts also become pyrophoric after use in process and precautions have to be taken when unloading the catalyst.

During storage, butadiene forms peroxides (which may be spontaneously explosive), and which in turn May lead to the formation of plastic polymer and/or pyrophoric "popcorn" polymer. These reactions may be limited by inhibitors and by controlling storage temperatures as low as practical, but precautions must be taken when opening equipment for inspection or repair.

Reid Vapor Pressure (RVP)

The vapor pressure of a liquid at 100°F (37.8°C), determined by a standard laboratory procedure, expressed in psia, is called the Reid Vapor Pressure. A listing of Reid Vapor Pressures for some materials can be found in **appendix 1**.

Risk

The combination of the probability of an abnormal event or failure and the consequences of that event on people, the community and the plant.

Safety Critical Device

A device or system is considered safety critical if it is the last line of defense to prevent an uncontrolled major breach of containment, severe personal injury or death or a major environmental incident, or is otherwise essential in the control or mitigation of such incidents. For example, heat tracing systems (steam or electric) used to prevent plugging of pressure relief devices due to solidification of process fluids are considered safety critical and should be identified as such. Check valves can also be safety critical under certain conditions. **Appendix 2** lists some examples of safety critical check valve applications. Other examples of safety critical devices include restriction orifices that limit the flow rate to a pressure relief device and Emergency Block Valves (EBVs).

Safety critical devices should be identified as such in relevant documentation such as Piping and Instrumentation Diagrams (P&IDs), operating manuals, and equipment files. For safety critical instrumentation, reliability targets (Safety Integrity Levels) must be specified, a testing and maintenance program must be in place to ensure that the reliability target is achieved, and a system to control deactivation of the device must exist. All safety critical devices or systems must be subject to periodic inspection and maintenance, and Management of Change (MOC) protocols must control their removal, alteration or replacement.

Safety Integrity Level (SIL)

One of three possible discrete levels used to characterize the reliability of instrument-based safety systems. SILs are defined in terms of Probability of Failure on Demand (PFD). The PFDs for various SIL levels are as follows:

Probability of Failure On Demand (PFD)	Safety Integrity Level (SIL)
Between 1 in 10 and 1 in 100	SIL-1
Between 1 in 100 and 1 in 1000	SIL-2
1 in 1000 or better	SIL-3

4. General

The basic principles upon which safety is incorporated into a plant design can be summarized in the three following steps:

1. Managing Control of Hazards

Hazards in process, storage and transportation should be managed by the following strategies:

- Potential hazards that are associated with a process should be identified and evaluated by a thorough study during the conceptual stage.
- Efforts should be made to eliminate or reduce the hazard through the application of **inherent safety principles.**
- The focus of inherent safety is on the **avoidance** of potential hazards rather than on their **control** by the addition of protective equipment.

The basic principles involved in designing for inherent safety are:

- *Intensification: Use or store less quantity of a hazardous material*
- *Substitution: Use a less hazardous material*
- *Attenuation: Use hazardous material under less hazardous conditions (e.g., reduced pressure and/or temperature)*
- *Simplification: Minimize opportunity for mistakes*
- *Error Tolerance: Make incorrect action impossible or more difficult than correct action.*

Examples of the application of inherent safety principles to plant design include:

- *Elimination or reduction of intermediate storage or surge capacity (intensification example)*
- *Use of aqueous ammonia in lieu of anhydrous ammonia (substitution example)*
- *Use of refrigerated instead of pressurized storage (attenuation example)*
- *Use of steam reheat instead of reheat burners (simplification example)*
- *Use of unique colors and fittings for air, nitrogen, steam and water at utility stations (error tolerance example)*

For the concept and application of inherent safety, refer to **Design Basis for Inherent Safer Plant Design in NPC-HSE-S-02.**

After the process design specification of a facility is completed, a risk assessment should be performed. This is best done through an evaluation of the risks by an experienced team. Members should not have been previously associated with the process design specification and should have experience in safety, operations and design / instrumentation. The proposed design should be assessed in detail (line by line) and findings should be documented for design follow-up. Prevention of loss of containment is by far the most important concept during this review. The Design and International Practices present the knowledge on how to prevent the initiating event, facilitate detection before the occurrence, and mitigate in case it occurs. The design should incorporate adequate facilities such that the likelihood an uncontrolled release of flammable or toxic materials is minimized and that sources of ignition are always under control.

The residual risks in a plant should be managed by a combination of design, mechanical and operating procedures, and emergency response capabilities. Once the plant is running, the compliance with mechanical, operating, and emergency response procedures will have to be managed by operational integrity systems. Particular attention should be given to changes to plant design or operating procedures since this has been a frequent reason for unwanted events.

2. Minimizing Damage from Fire or Explosion

The plant shall be designed to minimize the resulting damage, if a fire, explosion, or other accident occurs. This can be achieved by providing detection and mitigation systems, e.g., Depressuring and emergency isolation facilities, adequate spacing, blast resistant control rooms, adequate fireproofing, and good fire fighting, and drainage facilities.

3. Special Considerations

Special factors should be considered, such as local codes, or large inventories of liquid petroleum gas (LPG) in close vicinity to population areas which may require design safety features which are different from or beyond those normally provided.

5. Control of Hazards in Plant Design

5.1. Overstressing

Excessively high or low pressures or temperatures on equipment may result in yield stresses being exceeded and catastrophic failure occurring. Pressures above normal operating pressure can be generated by heat transfer from external fire or by various forms of operating failure, such as instrument malfunctions, utility failure, overfilling, thermal expansion, closure of equipment outlets, tube rupture in a heat exchanger, etc. Excessively high temperatures may also result in equipment being pressured beyond safe limits, since yield stresses are a function of temperature. Excessively low temperatures may cause the metal to lose its strength by embrittlement. Subjecting equipment to internal pressure at temperatures below the Critical Exposure Temperature (CET) may result in brittle fracture. Vacuum is another potential overstressing mechanism. If a vessel is subjected to, but not designed for sub-atmospheric pressure, this may result in vessel collapse.

Plant design must include protective features to prevent equipment from being overstressed by mechanisms such as the above. The essential steps in these design procedures may be summarized as follows:

- ***Consideration of Contingencies*** *- Based on the Single/Remote Contingency concept all possible contingencies which could cause overstressing and equipment failure shall be considered. The resulting relieving rates shall be evaluated in order to establish a basis for the sizing of a relief system. Where over-pressuring can occur, relieving facilities are provided; alternatively, a basis is established for designing the equipment to withstand the highest pressure and temperature that may occur. Protective facilities are normally sized on the basis of handling the largest release resulting from a single contingency, without exceeding the design pressure or temperature of the equipment.*
- ***Where temperature excursions can result in overstressing,*** *the resulting failure cannot be precluded by the installation of pressure relief devices. A basis for appropriate protection in the form of high or low-temperature alarms/cut-outs, control instrumentation, isolation, Depressuring, quenching, material selection, and/or other means must be developed. If instrumentation is relied upon to provide protection, it may have to be configured as High Integrity Protective System (HIPS).*
- ***Selection of the Appropriate Type of Overpressure Protection*** *- Normally this will be a pressure relief valve, or another pressure relief device. Temperature control devices for overpressure protection may only be used where overstressing may occur only due to high temperature below design pressure.*
- ***Sizing of Pressure Relief Devices*** *to handle the required relieving rate. This must be based on the various contingency considerations.*
- ***Designing Pressure Relief Device Installations*** *should include location, associated piping, and disposal systems.*

5.2. Fire and Explosion

The basic approach to minimizing the risk of fire and explosion is by incorporating features into plant design which would reduce the probability of such events.

External- To avoid external fire and explosion, it is necessary to prevent loss of containment, and to locate controlled release points in a safe location or into an enclosed system. In addition, ignition sources must be minimized/controlled.

1.Prevention of Loss of Containment and Location of Controlled Releases - Loss of containment of flammable material usually is the result of equipment failure or operating error. Equipment failure may be due to the exposure of construction materials to operating conditions more severe than those which they are capable of withstanding- or may be associated with the inherently more vulnerable nature of certain components such as fired heater tubes, machinery, and small piping systems.

Legislation meanwhile has banned venting of flammable gases to atmosphere which means that all releases must go into enclosed (blow down) systems. Where pressure relief valve releases of flammable materials to atmosphere is permitted they must be all vapor and comply with location and velocity criteria to ensure adequate dispersion, per design standards and international design practices.

Atmospheric vents should discharge to a “safe location.” This means that personnel should not be endangered by toxic or otherwise hazardous materials, and that accumulations of flammable mixtures near potential ignition sources should not be allowed. Environmental considerations are separate and additional. Unless other considerations apply, a “safe location” means a minimum of 10 ft. (3 m) above any equipment or platforms within a 50 ft. (15 m) radius. In addition, vents discharging flammable materials should not discharge around fired heaters or other ignition sources. Vents that are not tied into closed systems, yet could discharge liquid droplets on occasions (condensate), should not end near locations where persons may be present. A typical case would be the vent on a bellows pressure relief valve. Such vents should be carefully designed by providing a small disengaging drum to separate the droplets from vapor. The drain pipe of the drum should be installed such that droplets do not discharge above walkways or hot piping, or soak insulation.

Adequate equipment drainage and safe disposal facilities must also be provided to avoid dangerous accumulations of flammable material when equipment is taken out of service.

Atmospheric tankage is a potential source of release of flammable material, due to the possibilities of over-filling, excessive vapor evolution, corrosion, tank settling, and boil-over.

2. Minimizing Ignition Sources - The common ignition sources occurring in a chemical plant has been in NPC-HSE-S-02, together with methods by which they may be minimized and controlled.

Internal - To avoid internal fires and explosions, it is necessary to prevent flammable vapor/air mixtures within process equipment, and to minimize internal ignition sources.

- ***Elimination of Flammable Mixtures*** **-** *The formation of a flammable vapor/air mixture within process equipment represents one of the most dangerous situations that can exist in a process plant. Many processes and procedures involve the use of air within the equipment, or potential entry of air by leakage or entrainment, e.g., air oxidation reactions, regeneration and decoking systems, all types of combustion equipment, plant startup and shutdowns, vacuum processes, etc. Design procedures to prevent fire and explosion in these processes shall be considered in detail.*
- ***Control of Oxidants*** *- Materials that are not readily ignitable in air under normal atmospheric conditions may ignite spontaneously when exposed to pure oxygen or to oxygen-enriched air. In addition, materials other than oxygen will chemically oxidize, i.e., react with flammable or combustible materials. For instance, most flammable or combustible materials will burn in chlorine as they do in oxygen. Flammable gases (hydrocarbons, hydrogen, ammonia, and alcohols) may form explosive mixtures with chlorine. Carbon steel ignites in chlorine above 480°F (250°C), or at lower temperatures if traces of hydrocarbons are present. Mixing of hydrocarbons and liquid chlorine should be prevented. Location of chlorine cylinders in process areas should follow the requirements of* **NPC-HSE-S-03**. *Inventories should be kept as low as reasonably possible.*

5.3. Operational Factors Affecting Safety

A plant shall be designed so that the operating and maintenance personnel can carry out their duties effectively and safely, without exposing themselves or the plant to the risk of fire, explosion, toxic materials above allowable levels or accidents. To achieve this, the following features must be included:

1. Access and escape paths for emergencies, adequate platforms, ladders, guards, safety showers *and similar facilities which are basic requirements for safe working conditions.*

2. Instrumentation, alarms and controls *sufficient to enable the operating crew to operate the plant safely and effectively.*

3. Safe Startup/Shutdown *facilities and procedures should be available to permit plants or individual items of equipment to be safely started up and shut down, e.g., purge connections, drainage systems, etc.*

5.4. Process Factors Associated with Safety

Certain types of processes, process conditions, or fluids handled introduce factors which affect the safety of the plant. These factors must be taken into consideration in the design. They include:

1. High-severity operating conditions, *e.g., extremes of temperature or pressure.*

2. Batch or cyclic processes *or processes undergoing frequent startup and shutdown, where the*

opportunities for operating error are greater than normal.
***3. Processes subject to frequent upsets** by integration with other plants or where dangerous conditions may arise from utility failures.*
***4. Unstable processes,** in which decompositions, temperature runaways, or other unstable reactions are possible.*
***5. Fluid solids processes,** in which stable and safe operations depend on the effectiveness of fluidization of solids to prevent reverse flow.*
***6. Fluid properties and characteristics** such as flammability, vapor pressure, auto-refrigeration, corrosion, erosion, toxicity, and chemical reactivity, including the variations in these properties which may occur at abnormal operating conditions.*
***7. Start up or shut down** is an infrequent activity. Therefore, startup and emergency/normal shutdown procedures must be as simple and logical as possible. This must be incorporated into design considerations.*
***8. High noise evolution** may pose communications problems and impair operator performance by creating additional stress. Background information on factors such as the above should be investigated to ensure that the actual and potential hazards of a process are identified.*

5.5. Environmental Factors Affecting Safety

Environmental or climatic hazards which may apply must be recognized in a plant design. These include low temperatures which may cause icing-up of safety devices, dust or sand storms which create machinery lubrication problems, lightning as a source of ignition of atmospheric vents, potential for plugging atmospheric vents (e.g., by bird nests), and the possibilities of flood, high winds, or earthquake. In addition to appropriate International Practices, special design measures may have to be applied when these problems exist.

6. Designing Plant to Minimize Damage from Fire or Explosion

Although the design philosophy is intended to control all foreseeable risks of fire, explosion, exposure to toxic materials or other accidents, it is recognized that such incidents may still occur. A major contributing factor to this is human fallacy, e.g., inadequate communication, not following procedures, inadequate control, or mistakes. The plant design must therefore aim to minimize the damage resulting from such incidents. This is achieved by providing means to stop the release of flammable or hazardous materials as quickly as possible, by enabling the plant to withstand fire exposure without further failure while a fire is being extinguished, and by providing effective firefighting facilities.

The overall objective of these considerations is to enable a major fire to be brought under control as quick as possible preferably less than one hour. "Under control" is defined as a situation where the fire is still burning but further equipment failures and uncontrolled releases are unlikely. This "under control" situation is achieved when the heat being released is balanced by the factors of water cooling, fireproofing and spacing, such that equipment within reach of the fire is no longer in danger of failure through heat exposure. Achievement of the "under control" condition is also a function of the inventory of fuel feeding the fire, and the speed with which it can be reduced. The essential components of a plant design which are used to minimize the damage resulting from fire and explosion are listed below:

1. Spacing and Layout *- A well laid-out plant (including adequate equipment spacing, adequate drainage, "fire breaks" to establish separation between fire risk areas), limits the geographical extent of a fire and allows effective firefighting access. Equipment location and spacing standards are covered in* ***NPC-HSE-S-03*****.**

2. Fireproofing *- Fireproofing of structural steelwork, vessels, and vessel supports provides protection against failure from fire exposure and additional release of fuel. Fireproofing is also employed to ensure the continued functioning of certain emergency systems under fire exposure. Details are covered in* ***NPC-HSE-S-07*****.**

3. Blast Protection *- control/computer rooms, electrical substations, certain instrument houses, and other plant buildings are designed to withstand a certain size explosion in the plant. Details are covered in* ***NPC-HSE-S-05*****.**

4. Fire Fighting Facilities *- Adequate fixed and mobile firefighting facilities must be provided and be capable of meeting extinguishing and equipment cooling requirements for fires in all processing and offsite areas. The design basis for firefighting facilities is described in* ***NPC-HSE-S-07*****.**

5. Emergency Facilities *- Emergency facilities are required to reduce the release of flammable material feeding a fire as rapidly as possible. These facilities comprise remote shutdowns for certain items of equipment, emergency isolation and means of depressuring and removal of flammable inventory and water flooding capability. Details are specified in* ***NPC-HSE-S-10****.*

7. Special Considerations in Safety Design

7.1. Special Factors

Special factors should be considered during the course of a plant design. Such factors may be local regulatory requirements, hazards of special concern, or special incentives to minimize the probability of any incident occurring. These factors may justify safety design features in addition to the normal requirements of Design Practices and International Practices. Items for consideration include:

1. Governmental/regulatory statutory requirements, *codes and standards.*

2. Hazards associated with (large) plants, *large inventories of flammable or toxic materials, high density of large rotating equipment, process instability or high severity conditions.*

3. Limited availability of local assistance *in the event of a major fire, e.g., in isolated locations.*

4. Hazards associated with pressure storage *– Liquid petroleum gas (LPG) is normally stored in pressurized* ***vessels****. There is a potential for catastrophic failure if an adjacent fire causes such vessels to overheat and yield. Such an event is called a Boiling Liquid Expanding Vapor Explosion (BLEVE). To protect against such events, in close vicinity to population areas, for some installations it may be recommended to provide passive protection (mounding or fireproofing).*

5. Hazards associated with enclosed storage *(warehouses) of flammable, combustible, or toxic materials.(see also* ***NFPA 30****).*

7.2. Additional Design Safety Features

A number of methods are available by which Design Practice and International Practice standards may be supplemented, where justified by special considerations. Sound engineering judgment is necessary in selecting appropriate features from the following:

1. Designing to eliminate failures from low probability events which would normally be discounted. The "1.5 Times Design Pressure Rule" may be applicable to such cases of remote contingencies. Refer to ***NPC-HSE-S-09.***

2. Reducing the potential for uncontrolled release of flammable / toxic materials by:

a. Selection of superior quality machinery or materials of construction.

b. Selection of special machinery features such as seal-less pumps, submerged pumps, canned pumps, or oil mist lubrication.

c. in some cases it may be appropriate to further reduce the risk of failure of small piping connections or vulnerable equipment by specifying features such as:

- *Minimizing the number and extent of small piping connections.*
- *Increasing mechanical strength by using larger pipe sizes [say 2 in. (50 mm)].*
- *Combining multiple connections into a single valved nozzle of larger size at the vessel.*
- *Replacing gauge glasses with level indicators.*
- *Provision of excess flow valves or restriction orifices in small piping such as instrument* connections.

3. Provision of additional instrumentation, alarms, and surveillance devices (e.g., closed circuit television, vibration alarms, toxic gas detectors, combustible gas, or fire detectors) to identify potential emergency situations and actuate alarm or corrective devices.

4. Designing safety equipment for on-stream maintenance, so that maintenance can be carried out on it while keeping the plant fully protected at all times (e.g., provision of installed spare pressure relief valves, spare fire water pumps, etc.).

APPENDIX 1
FIRE HAZARD PROPERTIES OF SOME GASES AND LIQUIDS

MATERIALS	RVP			FLAMMABLE LIMITS			FLASH POINT			AIT	
	Psia		kPa	LFL.%	-	UFL.%	°F		°C	°F	°C
Acetylene		Gas		2.5	-	100.0		Gas		581	305
Hydrogen		Gas		4.0	-	75.0		Gas		932	500
Ammonia		Gas		15.0	-	28.0		Gas		1204	651
H_2S		Gas		4.0	-	44.0		Gas		500	260
CO		Gas		12.5	-	74.0		Gas		1128	609
CS_2	11.1		76.4	1.3	-	50.0	-22		-30	194	90
Methane		Gas		5.0	-	15.0		Gas		999	537
Ethane		Gas		3.0	-	12.5		Gas		882	472
Ethylene		Gas		2.7	-	36.0		Gas		842	450
Propane		Gas		2.1	-	9.5		Gas		842	450
Propylene		Gas		2.0	-	11.1		Gas		851	455
i-Butane		Gas		1.8	-	8.4		Gas		860	460
1-Butene		Gas		1.6	-	10.0		Gas		725	385
Pentane	15.6		107.0	1.5	-	7.8	-40		-40	500	260
Cyclohexane	3.3		22.5	1.3	-	8.0	-4		-20	473	245
Benzene	3.2		22.2	1.2	-	7.8	12		-11	928	498
Toluene	1.0		7.1	1.1	-	7.1	40		4	896	480
O-Xylene	0.3		1.8	0.9	-	6.7	90		32	867	463
Gasoline	14.0		96.0	1.2	-	7.1	-50		-45	500	260
MTBE	8.0		55.3	2.0	-	15.0	-18		-28	815	435
Ethanol	2.3		15.9	3.3	-	19.0	55		13	685	363
Diesel			< 3.5		Mist Ignitable		150		65	500	260
Crude oil	>1		>7	1.3	-	6.0	- 45		- 43	540	282
Hydrocarbons, Rule of Thumb:				1	-	10	-		-	600	315

Source:
NFPA 325, 1994 Edition

APPENDIX 2
EXAMPLES OF SAFETY CRITICAL CHECK VALVE APPLICATIONS

1. Check valves for which credit has been taken for the prevention or reduction of backflow from high-pressure systems to low-pressure systems when sizing the pressure relief device protecting the low-pressure system.
2. Check valves used to separate portions of a system with different design temperatures, CETs, shock chilling potential or construction materials if reverse flow or leakage through the check valve could lead to a loss of containment incident.
3. Check valves on the discharge of centrifugal pumps and centrifugal or axial compressors rated at or above 500 BHP. For compressors with one or more interstage feeds, there should also be a safety critical check valve at the suction of the first stage and at each interstage feed except the highest pressure one. For compressors with interstage products, the check valve on each interstage product is also considered safety critical. The reason for this is that failure of a check valve in these services can result in severe damage to the machine due to back spinning.
4. Check valves on the discharge of spared centrifugal pumps and centrifugal or axial compressors supplying a utility (e.g., cooling or boiler feed water, utility or instrument air) where failure of the check valve on a machine being taken off-line to close could lead to total loss of a utility.
5. Check valves on the discharge of firewater pumps and in connections to firewater systems.
6. Check valves with drilled flappers for protection against thermal expansion of trapped liquids or water freeze-up.
7. Check valves intended for emergency isolation of fired heaters from downstream inventory in the event of a tube rupture.
8. Check valves intended to prevent the uncontrolled mixing of air with combustible or flammable materials. For example, the check valve in the air injection line or the check valve in the ammonia injection line, unless there is protective instrumentation such as a low-flow or low-pressure-differential cut-out valve.
9. Check valves in compressed air starting systems for diesel and gas engines.
10. Check valves in utility connections where the normal operating pressure of the process exceeds the normal operating pressure of the utility. For example, the check valve in a steam line to a heat exchanger where the process pressure is higher than the steam pressure. For cases where the operating pressure of the process exceeds the operating pressure of the utility only during abnormal conditions, the need to treat check valves as safety critical should be based on a risk assessment.
11. Check valves intended to prevent the backflow of vapor or mixed phase streams into atmospheric tankage.
12. Check valves in articulated pipe drains for floating roof tanks to prevent a large spill in the tank area.

13. Check valves in motive fluid accumulators for Emergency Block Valves.
14. Check valves intended to prevent a release of toxic material near grade, such as combustion air to a CO boiler or Claus plant.

Installation Requirements for Safety Critical Check Valves

1. Safety critical check valves should be installed such that they will close depending only on gravity, not on flow reversal, spring action or external actuators. For example, swing and tilting plate check valves should be installed in vertical up flow or horizontal lines, straight-through ball and dual-plate check valves should be installed in vertical up flow lines, and globe type ball check valves should be installed in horizontal lines.
2. The use of "wafer-type" (flangeless) check valves is forbidden for service temperatures above 600°F (315°C).
3. Safety critical check valves should be installed such that they can be tested on line, without the need for removal or disassembly. A bleeder should be provided upstream of the check valve for this purpose.
4. Where two or more check valves are installed in series for safety reasons, at least two different types of check valves should be specified to minimize the risk of a common mode failure.
5. Safety critical check valves should be clearly identified in unit Piping and Instrumentation (P&I) Diagrams and in the field to minimize the risk of changes without appropriate review, and should be subject to regular inspection and testing programs similar to those applicable to other safety critical devices.

Design Basis For Inherent Safer Plant Design
NPC-HSE-S-02

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1. Scope

This document describes minimum requirements for Inherent Safer Design principles and considerations that shall be incorporated in plant design.

2. References

- API RP 500, Classification of Locations for Electrical Installations at Petroleum Facilities
- API 650, Welded Steel Tanks for Oil Storage
- API 2000, Venting Atmospheric and Low Pressure Storage Tanks
- API 2003, Protection Against Ignition Arising out of Static, Lightning, and Stray Currents
- API 2216, Ignition Risk of Hydrocarbon Vapors by Hot Surfaces on the Open Air
- NFPA 30, Flammable and Combustible Liquids Code
- NFPA 68, Explosion Venting Guide
- NFPA 70, National Electrical Code
- NFPA 325, Fire Hazard Properties of Flammable Liquids, Gases and Volatile Solids
- NFPA 496, Purged Enclosures for Electrical Equipment in Hazardous Locations
- NFPA 497, Classification of Flammable Liquids, Gases or Vapors and of Hazardous (Classified) Locations for Electrical Installations in Chemical Process Areas
- NFPA 499, Classification of Combustible Dusts and of Hazardous (Classified) Locations for Electrical Installations in Chemical Process Areas
- NFPA 505, Powered Industrial Trucks

3. General Considerations

The first one basic principle underlying the incorporation of safety into plant designs are the recognition and elimination of hazards include the consideration of fire, explosion and other accident potentials. To minimize the risks of fire, explosion and other accidents, the general design approach consists of the following:

- Preventing uncontrolled releases of flammable or toxic materials, such as those which may result from equipment failure, improper materials or incorrect procedures.
- Minimizing the number of ignition sources.
- Preventing the formation of hydrocarbon/air mixtures in the flammable range within process equipment.
- Ensuring that the plant is safe and operable for personnel.
- Preventing runaway exothermic reactions.

The risk associated with flammable and toxic materials is a combination of the probability and the consequences of their release. To reduce risk, design must reduce the probability or the consequence of a release, or both. Risk is reduced by inherent and extrinsic safety features. There are five inherent safety principles which can reduce the risk associated with hazardous material:

1. Substitute a less hazardous material for a more hazardous one, where feasible. The substitution can involve process raw materials or utility materials. Process raw materials usually have to be considered early in the development stage, but utility materials may be substituted even in the later design stages. For instance, in selecting water treatment chemicals, chlorine can be replaced by sodium hypochlorite.

2. Reduce the inventory in storage and process vessels to a practical minimum. Inventory can be reduced in the obvious way by reducing hold up or surge capacity. Less obvious ways include increasing the reaction rate, possibly through better mixing or increasing yield to reduce recycle rate, both reduce reactor volume.

3. Reduce the severity (attenuation) of the conditions under which hazardous material is used. For exothermic or decomposition reactions, heat input might be limited by selection of an appropriate temperature heating medium. The reactant might also be diluted to reduce the size of the exotherm.

4. Simplify the process to reduce the potential for operating error. When flexibility is desired, it should be weighed against the increased complexity which it creates and the increase in the associated risk of operating error.

5. Avoid or mitigate potential operating errors or equipment failures through fault tolerant design. For example, incompatible connectors are used on plant air and nitrogen utility stations to avoid cross connection and appropriate spacing between process units will mitigate the impact of a problem from one unit on another.

Extrinsic safety features are devices which are added on to the process. Extrinsic devices are used to reduce the probability of a loss of containment, such as alarms, cut-outs, and deluge systems. They also mitigate the consequences of a release, such as emergency isolation valves, water curtains to disperse vapor clouds, and firewater systems.

Inherently safe features may be considered to be passive and built in while extrinsic features are devices which are active and added on. Extrinsic devices must be maintained and tested to be reliable, while designed in inherent safety remains reliable with no continuing maintenance or cost.

To get the maximum benefit, inherent safety features should be incorporated as early as possible in the design. However, even after inherent safety principles are applied, extrinsic safety devices will still be required to further reduce the risk of a loss of containment.

4. Containment

Containment is the key strategy for preventing fires, explosions, and toxic releases. Uncontrolled releases are likely to be the result of equipment failure or incorrect procedures. The following aspects of plant designs are particularly important in preventing such releases.

4.1. Materials of Construction

Appropriate materials of construction must be specified in plant designs, taking into consideration all anticipated service conditions, including variations in temperature and corrosive conditions that may occur during operation. Reference should be made to the plant Construction Materials Manual for guidance. When materials of construction which are more than usually vulnerable to failure (e.g., low melting point alloys, Karbate, etc.) are used, the need for additional protection, such as emergency isolation valves or fireproofing shall be considered.

The selection of materials of construction for refrigerated or cryogenic processes, or where volatile materials may auto refrigerate, must be such as to avoid brittle fracture under all anticipated operating conditions. Abnormal situations such as emergency shut down, loss of heat exchange, depressuring, etc., must be analyzed to determine the most severe low temperature conditions to which equipment can reasonably be expected to be exposed. Safety valve discharge headers, flare lines and vent stacks are especially vulnerable to low-temperature exposures during upset or emergency conditions, if liquid is released.

Where ambient conditions may be below -20°F (-29°C), the design should recognize the potential for brittle fracture in carbon steel pipe during startup.

4.2. Safe Machinery Installation

Operating machinery is potentially hazardous, because of: (a) the kinetic energy which may be released in highly destructive form in the event of mechanical failure, and (b) the vulnerability of shaft seals to failures which may release large quantities of flammable or toxic fluids. Further details of safe design and application may be obtained from the API Standards 610, 611, 612, 613, 616, 617, 618 and 619.

4.3. Piping Systems Mechanical Design

Good design and construction of piping systems, in accordance with the Piping International Practices, will eliminate many of the points of potential failure which may cause fires and emergencies. Pipe dead-legs in critical service, e.g., flammable or toxic fluids in combination with a corrosive material, should be eliminated or minimized. Dead-legs are piping sections which normally contain a stagnant fluid. The stagnant zone may experience different conditions, than the flowing fluid. Lower temperatures due to heat loss can cause condensation of corrosive vapors, or condensation and freezing of water.

Higher temperatures, often caused by heat tracing, can increase corrosion rates. Where dead-legs are a required part of the process, e.g., bypass lines in critical service, the dead-leg shall be inspected on a regular basis. Small bore piping, typically used for machinery and instrumentation connections shall be strengthened to prevent mechanical or vibration damage. Instruments themselves must be capable of withstanding the system design temperature and pressure, and be compatible with the system fluids.

4.4. Fired Heater Hazards

Features which minimize the possibilities of fired heater tube failure and entry of fuel gas condensate into fired heaters, both of which are frequently the cause of fires, are required, as follows:

4.4.1. Minimizing Potential for Fired Heater Tube Failure

- **Instrumentation for Flow Stoppage -** Provide all process heaters with a low flow fuel shutdown (i.e., low process flow actuating shut-off valves in all fuel streams to the fired heater, excluding pilot gas).
- **Reliable Fired Heater Feed Supply -** Provide the fired heater feed pump with a reliable spare. In most cases it is satisfactory to specify electric motor drives for a fired heater feed pump and its spare, supplied from the two sides of a secondary selective power distribution system. If a secondary selective system is not available, or if the source of power supply is unreliable, a steam turbine spare is normally provided. When the fired heater is fed from a tower or other process vessel, provide sufficient holdup time below the vessel low level alarm point to allow operating personnel time to prepare the fired heater for no feed.
- **Equalized Flow Distribution in Multi-pass Fired Heater**
  a. For fired heaters in liquid or mixed-phase service, a means of controlling flow in the inlet to each pass shall be provided.
  b. For multi-pass, vertical tube fired heaters in vapor service, where the possibility of blockage exists from liquid slugs entering or being left in the coil, individual pass flow or temperature detection shall be provided.
  c. Provide process temperature measurement.
- **Emergency Steam to Fired Heater Coils -** The use of steam injection **<u>into fired heater coils</u>** in the event of tube failure is generally not recommended and connections for this purpose are not normally provided. However, steam connections shall be installed for purging the firebox, snuffing header box fires, or for injection into the coils to reduce coke formation. Purge steam is not recommended for smothering a “fire” which results from a tube failure in the firebox. If the steam pressurizes the firebox, unburned gas or liquid could exit and ignite outside the box.
- **High Temperature Alarm -** All fired heaters shall be provided with a high temperature alarm, independent of the temperature controller, in the coil outlet.

- **Toe-Walls -** Spills resulting from tube failure should be contained as follows:
  a. The grade area below a fired heater should be paved at least 10 ft. (3 m) beyond the extremities of the heater and provided with a toe-wall at that point.
  b. Where two or more fired heaters are located adjacent to each other, dividing toe-walls should be provided.
  c. Catch basins should not be located directly below fired heaters. Catch basins necessary for draining fired heater areas shall be located next to the toe-wall.
  d. Drains from other process equipment which could contain hydrocarbons should not be routed to the catch basins within fired heater areas.

4.4.2. Preventing Entry of Fuel Gas Condensate into Fired Heaters - Fuel gas supply streams to any combustion device must pass through a knockout drum for removal of condensate, sized to handle slug of liquid in the supply line from the fuel gas header. The drum must be fitted with a high level alarm and a condensate dump line to a continuously available closed disposal system (safety valve header preferred). In all locations where ambient temperatures can fall below the dew point of the fuel gas (i.e., where condensation downstream of the knockout drum can occur), gas header from the knockout drum to the burners and pipe shall be insulated and heat traced, the low point drains shall be directed to the same condensate disposal system. Where fired heaters are widely spaced, such that tracing and insulation of the entire gas header is not practical, separate knockout drums should be provided. Process vessels discharging gas streams into fuel gas headers must be provided with a separate high level alarm independent from the normal process level controller alarms.

If the fuel is a low-pressure gas, that is fed directly from the same process unit, a fuel gas knockout drum is normally not provided. In such cases, the condensate is removed from the fuel gas manifold to an independent steam atomizing condensate burner located in the fired heater. This burner should be kept in continuous service while the heater is in operation. Since this arrangement cannot handle a large liquid spillover from the distillate drum, a high level-cut-out should be provided on the drum, to shut off the fuel gas line.

4.5. Safe Failure Mode for Instruments

Process instrumentation shall be designed such that, upon loss of the control signal or the actuating medium, the final control element (usually a control valve) fails in the safest position. Depending on the specific application, this may require the control valve to fail open, fail closed or fail stationary in its last controlling position. For pneumatically operated valves that fail stationary, the direction (open or closed) in which the control valve should move as the air pressure in the actuator decays (due to leakage) should also be specified. The analysis must consider individual control loop failures as well as total loss of the actuating medium (instrument air, power, etc.).

The choice of failure position for control valves frequently requires considerable engineering judgment since general rules may not be applicable to all situations. The main objectives are:

1) To bring the unit to a safe condition with minimal operator intervention,
2) To minimize upsets to other units, and
3) To facilitate the return to normal operation once the original problem is corrected.

Meeting these objectives generally requires that control valves that bring energy or material into the unit fail closed, control valves that remove material from the unit fail closed and control valves that remove energy from the unit fail open.

Some examples of the application of these principles are listed below:

- *Fired heater stack dampers, used to control draft, should fail to the open position.*
- *Tower reflux control valves should fail to the open position.*
- *Fired heater and boiler fuel control valves should fail to the closed position.*
- *Compressor anti-surge control valves should fail open or stationary drifting open.*
- *Fired heater feed control valves should fail open or stationary drifting open to minimize risk of tube rupture due to overheating. The use of fail stationary drifting open should be considered for fired heaters having few parallel passes (4 or less) and in which the individual pass flow controllers are reset by a total feed flow controller; under these conditions, failure of the flow control valve at the inlet of a single pass in the full open position could lead to starvation of the remaining passes, since their flow rates would be reduced by the action of the total feed flow controller.*
- *Unit feed and product control valves should fail to the closed position.*

It is recognized that deviations from these practices are sometimes necessary. In such cases, the rationale for the deviation must be documented and formally accepted. Consultation with Engineering's Department safety specialists is recommended.

4.6. Equipment Drainage, Contaminated Effluents and Vents

Safe means of disposal must be provided for the various drainage, effluent and vent streams, so that they do not constitute a safety, health or environmental risk. Vents through which flammable materials are released to the atmosphere must be safely located with respect to adjacent equipment and personnel working areas, including consideration of heat radiation in the event of ignition.

4.7. On-Stream Blinding of Process Equipment

When a blind is installed on one side of an isolation valve, and where the other side of the valve is exposed to process conditions, valve leakage can result in pressurization of the space between the valve and the blind. Such leakage cannot be detected and can lead to a hazardous situation when the blind is removed. Any such blinding points that will be used for on-stream isolation of equipment (e.g., heat exchangers) should be identified during design and should be provided with drains and vents required to ensure that the space in question can be safely depressured prior to removal of the blind.

5. Minimizing Sources of Ignition

Typical sources of ignition which are commonly found in petrochemical plants with design features which are used to minimize their potential for initiating fires or explosions are as follows:

5.1. Fired Heaters and Hot Equipment

The risk of a vapor release being ignited by fired heaters or hot equipment can be minimized by good plant layout and spacing. These requirements are detailed in **NPC-HSE-S-03**.

5.2. Auto Ignition

The Auto-Ignition Temperature (AIT) is the lowest temperature required to cause self-sustaining combustion, without initiation by spark or flame. Auto-ignition temperatures of the majority of hydrocarbons fall in the range of 400 to 1000°F (200 to 500°C). In certain phases of safety design, an arbitrary auto-ignition temperature of 600°F (315°C) is used. This is judged to be a conservative estimate in the absence of experimental data. When experimental data are available, the actual value of the AIT or 600°F (315°C), whichever is lower, should be used as the governing criterion for equipment spacing and any other design features where AIT is a consideration.

5.3. Pyrophoric Materials

Pyrophoric material (e.g., iron sulfide, acetylene polymers, spent Ni - catalyst) can rapidly oxidize in air. Oxidation can elevate the material temperature to its auto ignition point or to the ignition temperature of any surrounding hydrocarbon deposits. Fires have occurred on-line in storage tanks and off-line in equipment prepared for, or undergoing maintenance.
Practical experiences and procedures may recommend:

- *Assume deposits in hydrocarbon systems are pyrophoric.*
- *Continuously wet the material with water or maintain an inert atmosphere until work has been completed or the deposits are removed from the vessel.*
- *Keep any material wet during transit to a safe disposal area.*

These Pyrophoric materials are process byproducts which can be exposed to air during turn around or regenerations. Pyrophoric raw materials such as metal alkyls used in polymerizations, which can be reactive with water, should be handled so that there is no exposure to oxygen.

5.4. Maintenance Activities

The risk of ignition of flammable materials is always present during plant maintenance involving welding, burning, use of cranes, vehicles, etc., and other hot work. However, the risk can be reduced by providing adequate spacing between units that shut down for turnaround separately, per **NPC-HSE-S-03**.

5.5. Vehicle Traffic

The potential for ignition from vehicle traffic (including all mobile equipment drivers) from the vehicle electrical system, exhaust system, engine over speed/valve bounce, can be minimized by traffic restrictions. Guidelines for the electrical classification of fork lift trucks are given by **NFPA 505**, Powered Industrial Trucks Including Type Designations, Areas of Use, and Maintenance and Operation.

5.6. Diesel Engines

Diesel engines (stationary machinery drivers) can produce ignition mechanisms from the electrical system, exhaust system, engine over speed/valve bounce. The risk of ignition can be minimized by complying with **NFPA 505** or local code requirements for diesel engines in classified areas (spark arrestors, flame traps, over speed shut down, explosion-proof electrical equipment, etc.).

5.7. Lightning

Lightning is a potential ignition source for flammable vapors, and appropriate design features shall be provided as follows:

- *Lightning protection and equipment grounding per **NPC-HSE-S-08**.*
- *Installation of shunts for bonding between roof and shell of floating roof tanks having steel-shoe roof seals with the hanger mechanism in the vapor space below the seal (pantograph type).*
- *Provision of pressure-vacuum vents for cone roof tanks holding materials with flash points below 100°F (38°C). This type of vent functions as a flame arrester, preventing an internal explosion even though the vapor space is within the explosive range.*
- *Steam snuffing connections are specified for atmospheric vents from process equipment, as described later in this section and **NPC-HSE-S-09**.*

5.8. Static Electricity

Petrochemical materials are first classified as static accumulators or non-accumulators. They are then classified as high, intermediate or low vapor pressure products, according to their flammability characteristics at the conditions applying in the equipment under consideration.

For a full description of static electricity mechanisms, definitions and classification procedure, and design and operating requirements, see NPC-HSE-S-08 and **API 2003.**

Agitation of Static Accumulators in Storage Tanks - Where agitation of static accumulators in storage tanks is required and where the products can generate a flammable vapor space, the electrostatic ignition hazard shall be eliminated by:

1. Selecting a floating roof tank to eliminate the vapor space, provided that the liquid level remains high enough to keep the roof afloat.

2. Controlling the vapor space atmosphere in a cone/dome roof tank by inerting or enrichment well above the upper flammable limit.

Static accumulators should not be agitated in cone roof tanks if the vapor space could be flammable. Agitation by means of air, gas, steam, jet nozzles or mechanical mixers shall be avoided. Recirculation is permissible subject to the following inlet velocity limitations:

1. Velocity less than 3 ft/s (1 m/s) until inlet nozzle is well covered (height of liquid at least one pipe diameter above the inlet).

2. After the inlet is well covered, velocity less than 20 ft/s (6 m/s) for tanks of less than 10,000 gallons (40m^3) capacity; less than 35 ft/s (11 m/s) for larger tanks.

Filtration of Static Accumulators - Micropore filters, filter/separators, and contamination monitors are powerful electrostatic generators. The high electrostatic charge must be reduced to a safe level by allowing a 30 second residence time for charge relaxation in the pipe between the filter and the receiving tank. This requirement applies to both metal and non-metal devices. Metal screens having a pore size larger than 300 microns do not require downstream residence time.

Pressure Relief Valve Riser Design for Hydrogen and Methane Service - To minimize the possibility of incendiary electrostatic discharge during release, all PR valve risers in hydrogen or methane service should be provided with a torroidal ring.

5.9. Electrical Equipment Sparking

The widespread application of electric power in petrochemical plants presents many opportunities for the generation of electrical sparks. Sparking may be the result of a fault in an electrical component, or may occur as a normal feature of the working of an electrical device, e.g., the sparking at the contacts of a switch when opening or closing. In order to specify the appropriate electrical equipment for a plant design which will minimize the ignition hazard, it is necessary to consider the probability of flammable vapor/air mixtures occurring, and the specific flammable vapors involved in the various plant areas.

Electrical area classification is based upon the definitions of the National Electrical Code, NEC Article 500. In applying these definitions, the guide rules incorporated in **API RP 500**, Recommended Practice for Classification of Locations for Electrical Installations at Petroleum Facilities are applied.

After plant areas have been classified according to the above definitions, following the interpretations and guide rules of **NEC, API RP 500, NFPA 497**, and according to the group of the gases or vapors involved, the selection of electrical equipment for different zones /Divisions shall be followed as per the requirements of **NEC Article 501** .

The principle upon which equipment selection is based is that in the Class I Division 1 locations, electrical equipment must not constitute a source of ignition for a flammable atmosphere in the area, either by the sparking associated with its normal functioning or by sparking caused by failure or faults. On the other hand, in Class I Division 2 locations, it is only required that electrical equipment must not constitute a source of ignition by the sparking associated with its normal functioning. In locations where flammable vapors are regularly present, such as within the vapor space of a tank, only electrical equipment classified as intrinsically safe may be used. These same principles are used for dusts in Class II areas.

6. Flammable Vapor-Air Mixtures Elimination

The third general design approach in the Inherent Safer Design is to prevent the formation of flammable vapor-air mixtures in the explosive range within process equipment. When considering safety of plant designs from the aspect of flammability of hydrocarbon/air mixtures within equipment, it is important that one use flammability values corresponding to actual operating conditions. Explosive limits are normally quoted for atmospheric temperature and pressure, but an increase in temperature or pressure usually widens the explosive range.

Flammability data for selected chemicals are presented in the appendix of **NPC-HSE-S-01**. Additional flammability data may be obtained from **NFPA 325**.

Note: In the case of certain materials, e.g., butadiene, explosive hazards arise from the presence of air (causing formation of unstable materials, such as peroxides) at concentrations very lower than the explosive limit of the butadiene, and exclusion of air in design and operation is accordingly very critical.

There are many situations where air is deliberately or inadvertently present within process equipment. The appropriate design procedures to prevent internal fires or explosions in such equipment are described as follows:

6.1. Fire and Explosion Prevention in Fired Heaters, Boilers and Gas Turbines

All types of fired heaters and other combustion devices have the inherent potential hazards of:
1. Flame-out with subsequent explosive re-ignition of the hydrocarbon/air mixture.
2. Explosion during initial lighting-up if prior purging of gas from the combustion chamber and downstream equipment has not been completed satisfactorily.

Therefore, all combustion devices require flameout protection and a means of purging the fire box as per specific design standards and international practices.

6.2. Flammable Vapor-Air Mixtures Elimination in Regeneration Systems

Many processes involve the removal of carbon, coke, etc. from catalysts or equipment by controlled combustion.

6-2-1. Continuous Regeneration Systems - These are particularly hazardous by virtue of the air always present in the unit and separated from the process hydrocarbons only by valves. This group includes:

1. Catalytic Cracking Plants - Design safety features are well established which prevent reverse flow (and hence mixing of air and hydrocarbons) in the spent or regenerated catalyst piping, by closing catalyst slide valves when low differential pressures across them occur.

2. Cyclic POWERFORMING Units - Double block valves and bleeders are provided to separate the air-containing regeneration circuit, which is manifolded to each of the reactors in turn, from the main reaction process. Correct operation of these valves is essential and safety interlocks are provided.

6-2-2. Intermittent Regeneration Systems - Examples are regenerative Units, and steam/air decoking operations on steam crackers, fired heaters, etc.

- *Intermittent regeneration in these plants is often carried out on the specific item of equipment to be decoked, while the remainder of the plant remains in operation or continues to contain process fluids. Positive means of isolation by blinds or swing elbows must therefore be provided to isolate the equipment containing air for regeneration from the remainder of the plant. Every connecting line must be included. The requirements for isolation valving depend on the regeneration preparation procedure, if the equipment changes directly from operation to regeneration without Depressuring, then double block valves are necessary in order that isolation blinds can be inserted between them.*
- *During normal operations, when regeneration is not in progress, there must be no possibility of air and hydrocarbon mixing at the connection where regeneration air is piped to the unit. The connection must be provided with double block valves, check valve and bleeder, or breakaway spool.*
- *If regeneration effluent gases are routed to a fired heater stack or other shared vent system, means of positive isolation by a blind or swing elbow must be provided in this line to prevent leakage of process fluids during non-regeneration periods.*
  The requirement for valving is dependent on the regeneration preparation procedure.

6.3. Flammable Mixtures Elimination in Processes using Air as a Reaction Component

In a number of processes, air is deliberately injected as a reaction component (oxidizing agent). Air is strongly preferred over oxygen as the oxidizing agent, since oxygen is much more dangerous than air. The fundamental design principle of these units is that hydrocarbon-air mixtures should not be allowed to enter the flammable range.

6.4. Flammable Mixtures Elimination in Vacuum Processes

Plants processing flammable materials under vacuum have the inherent hazard of air entry through any leak, with the possible formation of a flammable hydrocarbon-air mixture, and operating temperature may be high enough to initiate ignition. Vacuum units must therefore be provided with the following protection:

1. The ejector and ejector condenser system should be provided with a means of preventing backflow of air from the atmosphere in the event of ejector steam failure.
2. Vacuum recorder and alarm actuated by loss of vacuum.
3. Means by which the unit may be rapidly shut down, the vacuum broken and the equipment purged.

6.5. Flammable Mixtures Elimination in Compressed Air Systems

Compressed air systems are subject to internal fire and explosion when reciprocating compressors with oil-lubricated cylinders are used. Ignition is provided by the heat of compression or oxidation of coke formed by lube oil decomposition.

Design features to eliminate these fire and explosion hazards are detailed below:

6.5.1. Air Compressors

1. For air compression services, compressors of the type which are inherently free of oil contamination are preferred. These include centrifugal, oil-free rotary and non-lubricated reciprocating machines. The majority of pressure and capacity requirements can be met by appropriate selection from one of these groups.
2. However, selection of a conventional lubricated reciprocating compressor over the non-lubricated type may in some cases, be dictated by considerations of lower cost, commercial availability and reduced maintenance requirements. In such applications, the following design features is recommended to minimize the fire and explosion hazard:

a. Limit actual stage discharge temperatures to 350°F (175°C), to minimize lube oil decomposition.
b. Provide a high temperature alarm in the compressor discharge, immediately upstream of the after cooler [if no after cooler is included, install 20 ft. (6 m) downstream of the compressor].
c. Install a temperature indicator in each compressor stage discharge as an index of mechanical condition.
d. If an after cooler is included, install an in-line centrifugal oil separator downstream of it.
e. Provide low point drains in all discharge piping and vessels to permit regular removal of accumulated lube oil.
f. Design compressor discharge piping for ease of inspection and cleaning of coke deposits.
g. Select compressor mechanical designs which minimize the use of cylinder lubricant.
h. For cylinder lubrication, select naphthenic based oil with minimum coking tendency. Alternatively, synthetic lube oil may be used, provided that the compressor and downstream equipment are designed for its special properties. However, it should not be assumed that the fire and explosion hazard is eliminated solely by the use of these so-called non-flammable lubricants. Synthetic lube oils should not be used for instrument air service. Lubricating oil viscosity should be selected as low as possible, consistent with an acceptable maintenance level.

6.5.2. Air Circulation

Process compressors, typically centrifugal, should not be used to circulate air in a closed loop unless special design features are incorporated. Closed-loop air recirculation might be used during check out, air drying, or regenerations. Leakage of oil through compressor seals into the air can build up and be ignited by the heat of compression.

Applications of this type shall be reviewed with a rotating equipment specialist. Reciprocating compressors with lubricated cylinders should never be operated in a closed loop using air or oxygen, since an explosive mixture could build up in the circuit.

6.5.3. Compressed Air as Motive Power

1. Restrictions on the use of air blowing for tank mixing are discussed later in this document.
2. Compressed air should not be used for line clearing or for driving pipeline pigs when the line contents are in any of the following categories:

a. Flammable material with a flash point below 100°F (38°C).

b. Combustible material heated to within 15°F (8°C) of its flash point.

c. High-flash material [flash point above 100°F (38°C)] which may be contaminated by low-flash material.

6.5.4. Air Intakes - Location of air intakes in relation to surrounding equipment is critical to their safety. Gas present in the vicinity may be drawn in, introducing the potential for an internal explosion, since ignition sources are present in many cases. These apply to air intakes for all plant purposes, including combustion in fired heaters, internal combustion engines, pressure ventilation of control rooms, regeneration and oxidation processes, cooling and compressed air systems, etc.

The fuel source for these explosions is leakage of flammable gas past piston rings or piston rod glands (e.g., from engine power cylinder to crankcase, gas compressor cylinder to crankcase, or gas compressor cylinder to engine scavenger air compressor cylinder). The source of ignition is usually local frictional overheating of worn rod packing, bearings, or piston rings. Incidents of this type are normally associated with poor mechanical condition.

In flammable or toxic gas services, distance pieces should be well ventilated or purged to prevent cross-leakage of flammable gas between crankcase, scavenge and compressor cylinders.

6.6. Flash Back Prevention in Vent Lines and Flare Systems

A small or intermittent flow of flammable vapor into an atmospheric vent stack introduces the potential for a flammable vapor-air mixture to be formed in the vent line by buoyancy and/or diffusion effects at the open end. An alternative mechanism for the formation of potentially flammable mixtures exists when air as well as flammable vapor can enter a system vented to atmosphere or to flare. Although designs should not permit such air discharges into vent or flare headers, the inadvertent entry of air may occur under some conditions; e.g., by leakage if there should be draft in the header system. Ignition of such flammable mixtures in vent systems may occur by lightning, causing internal explosion and possibly detonation.

In the case of flares, the pilot constitutes a continuous ignition source for any flammable vapor-air mixture which may be formed in the stack or upstream system. Deposits of pyrophoric materials in vent or flare systems should also be considered as possible ignition sources. Protection against such internal explosions shall therefore be provided.

6.7. Sewer Explosion Elimination

Plant sewers have the inherent potential for accumulation of flammable mixtures, since hydrocarbons can enter as a result of spillage, equipment draining, entrainment in effluent water streams, instrument failure, etc. Particularly dangerous are volatile liquids meeting warm water or hot condensate streams in the sewer; the rapid evolution of vapor constitutes an overpressure potential as well as a fire and explosion hazard.

Hot work in the area above manholes and catch basins is a potential source of ignition for flammable mixtures in sewers. Therefore careful covering and sealing of sewer openings is an essential part of the work permit control system. To minimize sewer fire and explosion hazards, design requirements shall be followed to minimize risk.

6.8. Atmospheric Tankage

Tanks storing flammable petroleum or petrochemical materials at atmospheric pressure are subject to the inherent hazards of (a) uncontrolled release and (b) internal explosion, if a flammable mixture in a tank vapor space is ignited.

Tankage hazards may arise as the result of the following operating mechanisms:

1. Frothover - A frothover occurs when the temperature of a tank is sufficient to boil any water that is present in it. The generation of steam below the surface of the oil results in the formation of oil froth, often with sufficient violence to rupture the weak shell-roof weld seam of roof. The resulting spill of low density oil froth may rapidly fill and overflow the tank dike and spread to a source of ignition, causing a major fire. Frothovers are usually caused by:

a. Routing water or a stream containing water into a "hot" tank [i.e., a tank operating above 265°F (130°C)]. This may be the result of incorrect stream routing, or heat exchanger tube leakage if the plant stream to the hot tank passes through a water cooled exchanger with higher pressure on the water side.

b. Light hydrocarbons inadvertently routed to "hot" tanks have the same effect as water. Incorrect routing or exchanger leakage may be the basic cause, as above.

c. Routing a hot stream to a "cold" tank [i.e., a tank normally operating below 200°F (93°C)] long enough to raise the tank temperature to the boiling point of the water present in the tank bottom. This may be the result of incorrect stream routing, or the loss of cooling capacity on a plant product stream.

2. Rollovers - The rollover phenomenon is characterized by a sudden rapid generation of vapor and has occurred in LNG and slop tanks where streams of different densities have been added, forming layers. Good mixing during tank filling and capability for recirculation are needed to prevent rollover. In the case of slop, thermal convection between light and heavy layers may be limited by emulsions until heating causes the lower layer to rise suddenly and produce rapid vaporization of the lighter layer. For this reason, storage or heating of light and heavy slop in the same tank shall be avoided.

3. Overfilling and Other Tank Spills - Fire hazards may also arise from other types of spillage of tank contents into the surrounding area. These may include:

a. Overfilling.

b. Failure of tank connections or associated piping.

c. Unattended withdrawal of water from the tank bottom.

d. Failure of a floating roof water drain, with the external drain valve left open.

e. Sinking of a floating roof, due to mechanical failure or accumulation of water, resulting in the hazardous exposure of the whole liquid surface.

4. Excessive Vapor Evolution - A limit of 13 psia (90 kPa abs) true vapor pressure at storage temperature is established to define the lightest petroleum fraction which can be safely stored in atmospheric tankage without risk of excessive vapor evolution. If, however, lighter hydrocarbons are routed into an atmospheric tank, or if the tank temperature is allowed to rise such that the true vapor pressure of the contents approaches atmospheric pressure, then a hazardous uncontrolled release of hydrocarbon vapor from the tank vent will result. If the vapor evolution exceeds the vent capacity on a floating roof tank, the roof can tilt or sink. The basic cause of such incidents may be incorrect stream routing, faulty upstream blending operations, exchanger tube leakage, loss of level or loss of stabilization in upstream process plant, or loss of cooling capacity on a plant product stream. When loss of level in an upstream process vessel would allow pentane or lighter materials to enter tankage, a low level cut-out shall be installed in the rundown line, set below the setting of the low level alarm.

5. Internal Explosion - Except during initial filling operations, before the roof is floated, external floating roof tanks do not contain a vapor space and are, therefore, not subject to internal explosion. Routine landing of a floating roof on its legs should be avoided since this would create a potentially explosive vapor space each time it occurs. This is particularly applicable to internal floating covers or pan roofs, since displaced vapors beneath the cone-roof take longer to dissipate.In a cone-roof atmospheric tank, however, the vapor space consists of a vapor-air mixture, the composition depending on the quality and temperature of the liquid in the tank and the physical operations involved (e.g., filling, emptying, mixing).

To minimize the hazards described above, the design of plant tankage facilities must include the features listed below :

6.8.1. Minimize Explosion Hazard by Appropriate Tank Selection - Selection of the type of atmospheric tank for a particular service, in addition to considerations of cost, evaporation loss, pollution, etc., must be such as to minimize the internal explosion hazard. Where products can generate a flammable vapor space, the hazard may be minimized by selecting a floating roof or inert gas blanketed roof design, or by following the “static rules” to ensure that electrostatic sparking mechanisms are avoided. The correct classification of the material to be stored is an essential part of the selection procedure. Rundown, storage, and ambient temperatures, liquid composition, and the effect of contamination or upstream plant upsets must be considered.

6.8.2. Restrict use of Air Blowing for Tank Mixing - Air blowing shall be avoided, since it may alter the flammability condition of the vapor space, and may thus contribute to a potential for fire or explosion.

6.8.3. Eliminate Ignition Sources - In addition to the elimination of electrostatic discharges, as described above, protection against ignition by lightning or electrical equipment sparking must be provided. Exposed surfaces of tank heaters when the liquid is at low level may also constitute an ignition source or promote the oxidation of pyrophoric deposits. Design should prevent the exposure of heating surface over the normal range of the tank liquid level.

6.8.4. Control Rundown Streams to Tankage - The routing of streams to atmospheric tankage must comply with the following:

a. A stream must not be routed to atmospheric tankage if its true vapor pressure at rundown temperature is greater than 13 psia (90 kPa abs), or if it can result in the true vapor pressure of material in the tank exceeding 13 psia (90 kPa abs).

b. As an exception to (a), light materials with true vapor pressure greater than 13 psia (90 kPa abs) may be used as components for line blending systems discharging into atmospheric tankage, subject to the following conditions:

- *The true vapor pressure of the blend must not exceed 13 psia (90 kPa abs).*
- *Controls must be provided to automatically shut off the flow of light component in the event of low flow of the base or heavy stream, or high temperature of either stream.*
- *A back pressure controller is also required to ensure good mixing and to minimize flashing prior to entry into the tank.*

c. A stream must not be designed for routing to “cold” atmospheric tankage if its temperature is above 200°F (93°C).

d. High-temperature alarms shall be provided on all plant rundown lines to tankage where loss of heat exchange or cooling capacity could result in any of the following:

- *The rundown stream exceeding 13 psia (90 kPa abs) true vapor pressure.*
- *The rundown stream to a “cold” tank exceeding 200°F (93°C).*

e. Appropriate alarms and/or cut-out shall be provided to give warning of process plant conditions which could result in light materials being routed to atmospheric tankage. Such alarms and/or cut-out shall be treated as “safety critical” devices. These considerations may also be applicable to streams routed to underground pits operating at atmospheric pressure.

f. Where possible, plant product stream heat exchangers and coolers should be designed with the product as the higher pressure fluid. This will minimize product quality contamination problems as well as tankage hazards that might otherwise result from exchanger leakage.

g. Particular consideration should be given to the effects of process upsets, emergencies or partial shutdown operations in highly integrated plants on the quality of product streams to tankage.

h. Piping systems upstream of tankage should be designed to minimize the potential for hazardous contamination by water or light materials.

i. Streams normally routed directly from one unit to another at elevated temperatures may require diversion to tankage in the event of shut down of the downstream unit. Design must include emergency coolers and appropriate instrumentation to ensure that such streams, when diverted, are cooled below 200°F (93°C) before being routed to a "cold" tank.

6.8.5. Minimize Potential for Spillage in Tank Areas

a. The high level switch shall be independent of the tank gauging facilities, but the alarm signal may be transmitted through the tank gauging transmission system provided that the data transmission system is self-checking to alarm upon a major component malfunction.

The alarm originating from independent high-level switches shall be both audible and visible to the operator. It should also be of the type that must be acknowledged to silence. The audible independent high-level alarm (horn or voice synthesizer) must be distinctive from lower priority and audible alarms. This requirement exists for hard-wired alarms and for alarms transmitted via a data collection system. Each tank in alarm condition must have a maintained display indicating the tank identification.

b. Design tankage piping systems to minimize vulnerability to mechanical damage or fire exposure. Within plant limits, piping systems should be above ground to minimize corrosion and avoid underground leaks which create safety and environmental concerns.

c. Provide safe means of water withdrawal and disposal from tank bottoms.

d. In viscous services, a closed drainage system should be provided, with heated separating tank and oil recovery facilities.

e. Provide safe and reliable means of water drainage from the roofs of floating roof tanks.

f. Ensure elevations of take-offs to a vapor recovery unit (VRU), if provided, are designed such that any tank overfill will not run preferentially to the VRU rather than into the diked area via the tank vents.

6.8.6. Include Protection Features to Handle Tankage Hazards

a. Install adequate vent capacity in accordance with **API Standard 2000**.

b. Include overpressure protection of all cone roof tanks. **API Standard 650** permits this to be provided by either a weak roof-shell weld seam, or by emergency venting devices in accordance with **API RP 2000**.

However, the **API 650** weak roof seam feature can be incorporated only into a cone roof tank of specified configuration and dimensions. There is a potential problem with obtaining frangible roof designs on small cone roof tanks under about 50 ft. (15 m) diameter since shell weight may be inadequate to counteract the uplifting force due to tankage overpressure.

If a frangible roof is not available in a cone roof tank handling low flash material, some possible solutions to reduce the chance of tank failure are:

- *Install anchor bolts to the shell.*
- *Change to high flash service.*
- *Install explosion hatches per **NFPA 68** requirements.*
- *Inert or gas blanket the tank and install emergency venting per **API 2000**.*
- *Convert to internal floating roof design.*

Other tanks such as dome roof designs, where the **API 650** weak roof seam cannot be included, must therefore be protected against overpressure by emergency vents per **API Std 2000**, but must not be used for "hot" services or in services where internal explosion is possible, e.g., by the formation and ignition of a flammable mixture.

c. Provide spacing and diking in accordance with **NPC-HSE-S-03**.

d. Provide fixed fire fighting facilities for tanks also see **NPC-HSE-S-07**. Cone roof tanks in low flash service are limited to 150 ft. (45 m) maximum diameter, because of the reduced effectiveness of foam attack on larger tank fires.

6.8.7. Special Tank Services - For safety features required in certain special tank services, reference should be made to the following:

a. Hot Tanks.

b. Slop Tanks.

c. Ballast Water Tanks

e. Fuel Oil Tanks (air blowing).

f. Internal Floating Roof Tanks: Covered floating roof tanks, as defined by the API, i.e., fixed metal roof tanks with steel pan-type internal floating roofs, are considered equivalent to cone roof tanks for spacing, diking and fire protection purposes. This includes 100% foam coverage for the floating roof.

Tanks with internal floating covers (i.e., of any type other than steel pan construction) should have spacing and diking in accordance with the requirements for open-top pontoon-type floating roof tankage.

Venting of the vapor space above the floating cover must comply with **API Standard 650**.

This type of tank has an inherent degree of risk during commissioning after the roof has been landed, since vapors are displaced into the space above the roof until it is floating. Tank sizing should therefore be based upon the roof remaining afloat during normal operation, and only being landed on its legs when the tank is to be taken out of service.

g. Acid Tanks.

6.9. Contamination between Plant Systems

Inadvertent leakage or flow reversal between different process systems may result in hazardous situations. The following design requirements should be included where appropriate, to minimize cross contamination hazards:

6.9.1. Utility Connections, Injection of Solutions

1. Utility connections to piping and equipment must comply with proper standards.
2. A check valve is required in any connection where wash water, caustic soda solution, inhibitor or other treating agent is injected into a process stream. There must be no other branches or recycle lines connected between the check valve and the injection point.
3. In cases where a particularly serious hazard could result from reverse flow of process fluid into a utility or chemical injection line (e.g., hydrocarbon vapors backing into a water injection header and being released from an atmospheric tank in a utilities area), then additional protection shall be provided, such as a second check valve, low flow alarm, or automatic reverse-flow shut-off valve.

6.9.2. Potable Water Systems - Direct connections must not be made between process equipment and potable water systems. If it is necessary to use potable water for the process, it should discharge, with an air break, into a vented tank and be pumped from the tank into the process equipment.

6.9.3. Waste Heat Boilers - Where hot process streams are routed through heat exchangers to preheat boiler feed water or to generate or to superheat steam, with the process side design pressure higher than the steam side operating pressure, the possibility of steam contamination by hydrocarbons as a result of exchanger tube leakage must be considered. In such cases the steam generated must be manifolded to a branch of the steam header such that it is consumed only within the same plant. In addition, the steam generated must not be used for purposes where it will contact air, e.g., regeneration, or purging.

6.9.4. Heat Exchanger Leakage - Contamination of product streams as a result of heat exchanger tube leakage may introduce a hazardous change of conditions in the receiving tankage. Protection against overpressure that may result from heat exchanger tube failure is covered in NPC-HSE-S-09.

6.10. Explosion Elimination in Start-up / Shut-Down

Adequate purging (a) of air at start up and (b) of process fluids at shut down, from process equipment is essential to avoid formation of air-gas mixtures which might pass through the explosive range. Design must cover consideration of the purging methods to be used (steam, water, inert gas, evacuation) and include adequate purging and venting connections.

If water flooding is to be used for purging, checks should be made that the equipment is mechanically capable of handling any increased loads resulting from the weight of water. Similarly, if steam purging is to be used, equipment may be exposed to temperatures higher than normal operation, and design temperatures should be specified accordingly. Even small items of equipment, e.g., pumps or exchangers, which traditionally have not been purged before commissioning must be reviewed to determine whether explosive mixtures and adiabatic compression heating to ignition temperature could, occur. If so, then purging should be required.

6.11. Explosion Elimination in Air Eliminators

The following factors should be considered to minimize this risk:

- For low *flash point hydrocarbons, and for high flash point hydrocarbons within 15°F (8°C) of their flash point, the set pressure of the air eliminator pressure relief valve should be based on keeping the maximum possible adiabatic air compression temperature below the hydrocarbon auto ignition temperature.*
- *If the hydrocarbon is a static accumulator, or if it is a non-static accumulator with the potential for forming fine spray or mist in the air eliminator, precautions should be taken to minimize the potential for the accumulation of static and electrostatic ignition within the air eliminator.*
- *If the hydrocarbon is sour, pyrophoric iron sulfide deposits may form in an unlined air eliminator. For such situations, materials of construction should be selected to minimize the potential for the formation of iron sulfide. Linings such as modified epoxy phenolic should be considered and any internal demister screens should be constructed of stainless steel.*

7. Operability and Personnel Safety/Design Features

The fourth general design approach in the Inherent Safer Design Approach is to design the plant so that it can be operated and maintained properly and safely by plant personnel. To prevent accidents it is also important to consider specific chemicals and incorporate known safeguards for those chemicals.

Plant design should permit the operating and maintenance personnel to carry out their duties effectively and safely without exposing themselves to the risk of fire, explosion, or accident. To accomplish this, the design must consider how operating and maintenance personnel will interact with individual pieces of equipment and the whole unit or plant. This interaction of personnel and equipment is referred to as Human Factors. The conditions of personnel interaction with equipment in plants is very diverse because of the multitude of tasks which are carried on. Process control, process sampling, and equipment maintenance, among others, all involve human interaction with equipment.

Human factors should be considered during the design to reduce the probability of human error and the impact on the personnel and the plant of those human errors which do occur. The designer should first consider the operations which are expected of the personnel and assess whether they are reasonable expectations, then should incorporate sufficient and accessible features so that there is no difficulty to proper operation.

Because of the numerous operations in the plant, it is helpful to separate interactions into two categories. The first is tasks which might result in explosion or loss of containment leading to fire, either of which could result in very serious injury. The other is tasks which could result in poor operation of the plant or less severe injury to personnel. The distinction is not always clear because multiple errors, which individually result in poor operation, could lead to a serious incident.

In addition to the above which may lead to a loss of containment, etc., human factors should be considered as they affect general safe working conditions or the ability to produce the required amount of on-spec product. Examples of these are listed below:

- *Requirements for platforms, stairways, headroom, and ladders, including escape routes.*
- *Avoid locating equipment requiring routine maintenance in the upper platforms of fired heaters.*
- *Safety shields for piping and equipment.*
- *Steam trap discharges, when not connected to a closed system, should be routed such that they do not constitute a hazard to personnel.*
- *Provide safety showers in locations where hazardous materials are handled. Examples of hazardous materials include caustic, acids and amine solutions.*
- *Insulate equipment for personnel protection.*

- *Winterize instruments, piping, and equipment to ensure operability of the plant.*
- *Provide necessary feed and product manifolding, unit bypasses, purge connections, and disposal routes for diversion of process streams and off-spec products for startup, shut down, and upset operations.*
- *Provide good accessibility to equipment for efficient and ergonomic operation and maintenance.*
- *Balance the desire for flexibility which can lead to increasing complexity via bypasses and jump overs with the need for simplification to reduce the chance of error.*

Often the answers to human factor issues are based on the experience of operating personnel and may be subjective.

8. Design Considerations for Specific Materials

Selected materials can present special fire or explosion hazards or acute toxic exposure hazards. These hazards are different than chronic or environmental hazards. General design guidelines for specific materials are (not limited to the following):

8.1. Ammonia

1.For installations handling anhydrous liquid ammonia. Copper and copper alloy fittings are not permitted.

2.Safety valves and flare headers which receive continuous ammonia-bearing streams must be segregated from continuous hydrogen sulfide or carbon dioxide flows, to avoid blockages resulting from the formation of solid ammonium sulfide or ammonium carbonate.

8.2. Butadiene

1.C4 acetylenes are present in raw butadiene and are removed in the butadiene extraction process. They are highly reactive, particularly vinyl acetylene which is thermally unstable and may undergo explosive decomposition in high concentrations if heated.

2.Copper acetylides are unstable complexes of copper and C4 acetylenes, and are often present as solid deposits when equipment in the CAA extraction process is opened up. These acetylides may be spontaneously explosive.

3.Acetylene polymers are formed as solid and sludgy pyrophoric materials in acetylene removal sections, particularly in heated equipment.

4. Butadiene reacts, as a function of temperature, to form a dimer, but this is not specifically a hazardous material. In contact with air, butadiene forms peroxides which may be spontaneously explosive, and which in turn may lead to the formation of plastic polymer and/or pyrophoric "popcorn" polymer. These reactions may be limited by inhibitors and by controlling storage temperatures as low as practical.

5.Operations which allow C4 acetylenes to concentrate at elevated temperature should be avoided.

6.The plant should be designed for ease of washing out, draining, and dismantling for cleaning, particularly CAA acetylene removal sections. Eliminate dead-legs and liquid traps, and slope safety valve inlets and outlets to avoid liquid accumulations.

7.Butadiene-producing units should be designed to minimize oxygen contamination of the product before being routed to storage.

8.Butadiene storage drums and spheres should be designed according to LPG practice. They should include vapor space sample points and the facility to control oxygen content below 0.2 vol% by purging to atmosphere, flare or back to the process unit.

8.3. Chlorine

1. Chlorine is non-combustible in air, but most combustible materials will burn in chlorine as they do in oxygen.
2. Flammable gases will form explosive mixtures with chlorine. Chlorine may react explosively with hydrocarbons, hydrogen, ammonia, and alcohols.
3. Carbon steel ignites in chlorine above 480°F (250°C), or at lower temperatures if traces of hydrocarbons are present.
4. Chlorine is highly corrosive to carbon steel and all common metals in the presence of moisture.
5. Mixing of hydrocarbons and liquid chlorine should be prevented during the injection of chlorine for catalyst activation.
6. Temperatures in the injection system exceeding 250°F (120°C) should be avoided (this allows a margin of safety against the combustion of carbon steel in chlorine).
7. Specific design details of the injection system shall be described in the plant Construction Materials Manual. Location of chlorine cylinders in process areas should follow the requirements of **NPC-HSE-S-03**.

8.4. Copper Acetylides

1. Under certain conditions, acetylene in process streams and copper in copper alloys in the equipment will react to form copper acetylide, which is potentially explosive.
2. Alloys containing more than 50 wt% copper must not be used in services where acetylene is present in amounts greater than 1000 ppm.
3. In general, the use of copper alloys in contact with acetylene-bearing streams should be avoided.

8.5. Hydrogen Sulfide

1. Special facilities are necessary for the disposal of extended releases of concentrated H_2S. These facilities may consist of either a special incinerator furnace, or a special H_2S flaring system, as described in **NPC-HSE-S-11**.
2. Intermittent releases of H_2S (such as safety valve discharges) Shall be routed to the regular flare system.
3. Flare headers containing concentrated H_2S should be constructed for ease of isolation, washing out and dismantling for cleaning.
4. Flame-out protection must be provided for the thermal combustor and, if applicable, for the tail gas incinerator burner(s).
5. Water drawoff streams from distillate drums and overflow water from flare seal drums may require further treatment, if the water has been in contact with H_2S bearing liquids or vapors.
6. Other sulfiding agents, such as dimethyl sulfide and dimethyl disulfide should be considered as substitutes for H_2S as possible.

8.6. Methyl Ethyl Ketone (MEK)

1. The properties of MEK, as they affect safety consideration, are similar to those of typical petroleum hydrocarbons.However, a condensable blow down drum is normally provided as a means of recovering the high value MEK from safety valve discharges.
2. Alcohol-resistant foam is required for fires with appreciable depth, e.g., tank or dike fires. For spill fires with only a thin layer of MEK, water is also effective.

8.7. Nickel/Cobalt Catalysts and Nickel Carbonyl

1. Unregenerated catalyst is highly pyrophoric. Therefore, the catalyst normally should be regenerated before dumping.

8.8. Nitrogen

1. Permanently piped nitrogen connections into process equipment should be restricted to services where continuous flow is required; breakaway connections should be used for all others.
2. In addition to the check valve at each individual connection, a check valve should also be installed at unit battery limits.
3. When nitrogen is supplied to plant by pipeline, a pressure reducing valve and check valve should be installed at the plant fence line to prevent back flow contamination of the nitrogen supply line.
4. **If nitrogen connections are provided at utility stations, a type of connector different from that for air shall be used, so that air hoses cannot inadvertently be connected to the nitrogen system.**

8.9. Phenol

1. Separate closed pump-out and drain header should be connected to equipment handling liquid phenol, and routed to the condensable blow down tank or back into the process.
2. Condensable blow down tank for phenol service is required.

8.10. Sodium Hydroxide

Specifications of appropriate construction materials and fabrication methods must be applied not only to equipment containing caustic soda but also to downstream equipment into which caustic may be entrained and where embrittlement may result.

8.11. Sulfuric Acid

1. A separate safety valve header and blow down drum is required for safety valves discharging acid or acidic hydrocarbons, with facilities to safely dispose of acidic materials and hydrocarbons in the blow down drum.

2. Equipment handling sulfuric acid should be located within a segregated area of the unit plot, with separate acid-resistant sewer, and neutralizing lime pit designed for ease of recharging.
3. A separate pressure relief header and blow down drum is required for safety valves discharging acid or acidic hydrocarbons, with facilities to safely dispose of acidic materials and hydrocarbons in the blow down drum.
4. A separate closed pump-out and drain header should be connected to equipment handling acid or acidic liquid hydrocarbons, and routed to the acid blow down drum.
5. Corrosion test probes and a multiple sampling pH meter may be warranted at strategic locations in the plant for early identification of corrosive conditions.

8.12. Tetra Ethyl Lead (TEL) and Tetra Methyl Lead (TML)

1. Both of these materials are subject to decomposition and violent explosion at elevated temperatures. Adequate spacing, water sprays, and fire monitor coverage are therefore essential fire protection features.
2. TEL and TML suppliers have comprehensive standards for design and operation of unloading and storage facilities and these should be closely followed. Design must prevent exposure of personnel to these highly toxic materials.

9. Run Away Reaction Prevention

Exothermic reactions may result in runaways if reaction conditions are not carefully controlled. To prevent runaways, the designer must understand the conditions under which they might occur, while operating near design conditions and when operating under abnormal conditions caused by credible plant upsets or failures. As a result of laboratory and pilot studies, the designer is usually well aware of exotherms and decompositions near standard operating conditions. Processes which have a long operating history usually have accumulated enough abnormal operating incidents that they are well understood. However, process changes that significantly alter operating conditions and new chemical reactions may require analysis of the potential for runaway.

Reactive chemistry techniques can be used to evaluate the impact of abnormal conditions on reactions through modeling and testing. The use of models and analytical tests can replace evaluation at the laboratory or pilot scale. This can save time and expense and is safer because the volume of material tested is very small.

Table 1 - Auto Ignition Temperatures (1)

Component or Stream	Autoignition Temperatures, °F	Autoignition Temperatures, °C
Hydrogen(pure) (2)	932	500
Carbon Monoxide	1128	609
Hydrogen Sulfide(3)	500	260
Carbon disulfide(3)	194	90
Methane(2)	999	537
Ethane(2)	882	472
Ethylene	842	450
Propane	842	450
Propylene	851	455
n-Butane	550	288
i-Butane	860	460
Buten-1	725	385
n-Pentane	500	260
i-Pentane	788	420
Pentene -1	527	275
n-Hexane	437	225
i-Hexane	506	266
N- Heptane	399	204
i-Heptane(4)	536	280
N-Octane	403	206
i- Octane	784	418
N-Decane	410	210
Benzene	928	498
Toluene	896	480
O-Xylene	867	464
P-Xylene	984	529
Cyclohexane	473	245
Naphta (BP: 212-290°F)(100-143°C)	450	232
Acetone	869	465
Methyl ethyl ketone	759	404
Methyl alcohol	725	385
Isopropyl alcohol	750	400
kerosene, fuel oil#1	410	210
Fuel oil#2	494	257
Fuel oil #4	505	263
Fuel oil#8	765	407
Gas oil	640	338

Notes:

(1) AITs taken from NFPA 325. When an auto ignition temperature for a specific hydrocarbon is not available a design value of 600°F (315°C) should be used. AIT's are approximate values because test volume, material, and residence time can all affect ignition temperature. Tests are done in closed vessels. Flammable clouds in the open atmosphere of a plant generally require hot surface temperatures higher than these AITs to ignite, see API 2216. In cases where hydrocarbons soak insulation and are confined, the ignition temperature can be lower than those above due to very long residence times.

(2) Static ignition must be considered.

(3) These materials are usually flared

(4) Unavailable from NFPA 325, taken from Hilado, CJ and Clark S.W., Autoignition Temperatures of Organic Compounds, Chemical Engineering.

Basic Safety Concepts For Plant siting & Lay out
NPC-HSE-S-03

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Appendixes

1. Scope

This document describes minimum requirements for equipment spacing in the design and layout of new plants and expansions of existing plants.

2. References

- Loss Prevention in Process Industries
- NFPA 30, Flammable and Combustible Liquid Code
- GAP 2.5.2, 2007, Oil and Chemical Plant Layout and Spacing
- Guidelines for plant siting and lay out, CCPS
- Exxon mobile engineering standards

3. Definitions

Battery Limits - The boundaries of a process area that establish the outer limit for location of equipment associated with a complete process or a group of integrated processes that must be shut down together for turnaround or isolated in the event of an emergency.
Combustible Liquids - High-flash liquids (closed-cup flash point 100°F (38°C) or higher) when handled at a temperature lower than the flash point minus 15°F (8°C).
Flammable Liquids - Low flash liquids (closed-cup flash point less than 100°F (38°C), and high-flash liquids (closed-cup flash point 100°F (38°C) or higher) when handled at a temperature above or within 15°F (8°C) of their flash points.
Flammable Materials - Flammable liquids, hydrocarbon vapors, and any other vapors that are readily ignitable when released to atmosphere (e.g. hydrogen, carbon disulfide).
High-Flash Stocks - Stocks with a closed-cup flash point equal to or greater than 38°C.
Light Ends - Hydrocarbons having a Reid Vapor Pressure (RVP) of 15 psia (103 kPa) or greater, Includes liquids with pentane and lighter components. Pure hydrogen is excluded from the definition.
Low-Flash Stocks - Stocks with a closed-cup flash point less than 38°C.
Pipe band - A pipe way, pipe rack, or sleeper way containing piping.
Process Vessel - Any vessel, drum, column, tower or reactor associated with the processing or handling of hydrocarbons in a process area.

4. General

The objectives of the spacing recommendations are as follows:

- *To permit access for firefighting, fire trucks and other emergency equipment.*
- *To permit access for operators to perform emergency actions in a fire situation.*
- *To minimize involvement of adjacent facilities in a fire and prevent further equipment failures.*
- *To ensure that critical emergency facilities are not subject to fire damage.*
- *To separate continuous ignition sources from probable sources of release of flammable materials.*
- *To minimize exposure of facilities to adverse consequences resulting from events external to the site.*
- *To permit access for normal operation and maintenance.*
- *To permit access for turnaround maintenance activities.*
- *To permit turnaround maintenance activities to be carried out without impact to adjacent, non-involved units*
- *To enhance site security.*

The equipment spacing distances detailed in this document, are recommended minimum figures. However, special considerations for a particular plant or local factors may justify deviations from the recommended standards, Reductions below recommended spacing should not be made, however, unless reviewed and approved in accordance with the plant processes for management of change or accepting Design Practice deviations, as appropriate. Factors which affect layout and spacing and which may justify deviations include the following:

- **Physical Limitations** on plot space available, i.e., roadways or boundaries of existing facilities.
- **Special Hazards -** Increased spacing may be justified where special hazards or incentives to minimize fire losses apply, or where continuity of operations is a prime consideration.
- **Flexibility -** Space requirements for future expansion, which may be a complete unit or individual equipment items.
- **On-stream Maintenance -** This requires access space for personnel and equipment to individual items of equipment, and also separation of units within the plant which may be individually shut down for turnaround.
- **Topography** has a major influence on safety of layout design in sloped areas. For example, location of separators depends on sewer drainage, and the possibility of major tank spills gravitating downhill to process areas must be avoided.
- **Local conditions** may dictate that site security be greater than normal, requiring increased spacing between the plant facilities and the property lines.
- **Safety and public relations factors** associated with the type of adjacent property must be

taken into consideration in the plant layout design. Distance from the plant facilities to the property line must be sufficient to prevent danger or nuisance to the neighbors and vice-versa. The spacing required depends on the type of plant nearest to the boundary, and the nature of the adjacent property, i.e., industrial or residential. In residential areas, additional space for screening may be required. Generous spacing would also be advisable where the adjoining land is undeveloped but is subject to future development.

- **Special Cases -** In certain cases, such as small independent chemical plants handling materials with flash points above 100°F (38°C), reductions below the recommended spacing distances may be justified where the associated risk is limited.

5. Design Considerations

5.1. Spacing Charts and Tables

Required spacing distances are presented in the Onsite Spacing Chart (**Figure1**) and the Offsite Spacing Chart (**Figure 2**). For each facility, the Basic Spacing is indicated in the diagonal column. Basic Spacing is the recommended horizontal distance between a given facility and other general processing equipment. This basic spacing is not required between items of equipment in similar services, e.g., between a pump and it's spare.

The spacing requirement between two facilities is obtained from **Figure 1** or **Figure 2** at the intersection of the appropriate horizontal and vertical columns. For spacing between two facilities of the same type, refer to the first square adjacent to the appropriate reference letter in **Figure 1** or **Figure 2**.

It is essential that the Onsite and Offsite Spacing Charts be used in conjunction with their respective guides, i.e., **Table 1** or **Table 2**.

Figures 3A, 3B represent additional spacing recommendations for Inter/Intra -Units in chemical plants. **Figure 3C** represent spacing requirements for Storage Tanks.

For buildings, additional spacing beyond these guidelines may be required for blast protection. The size and congestion of process units determine the overpressure that could be developed from ignition of a potential process area vapor cloud, which will impact the building spacing requirements.

5.2. General on-Site Layout

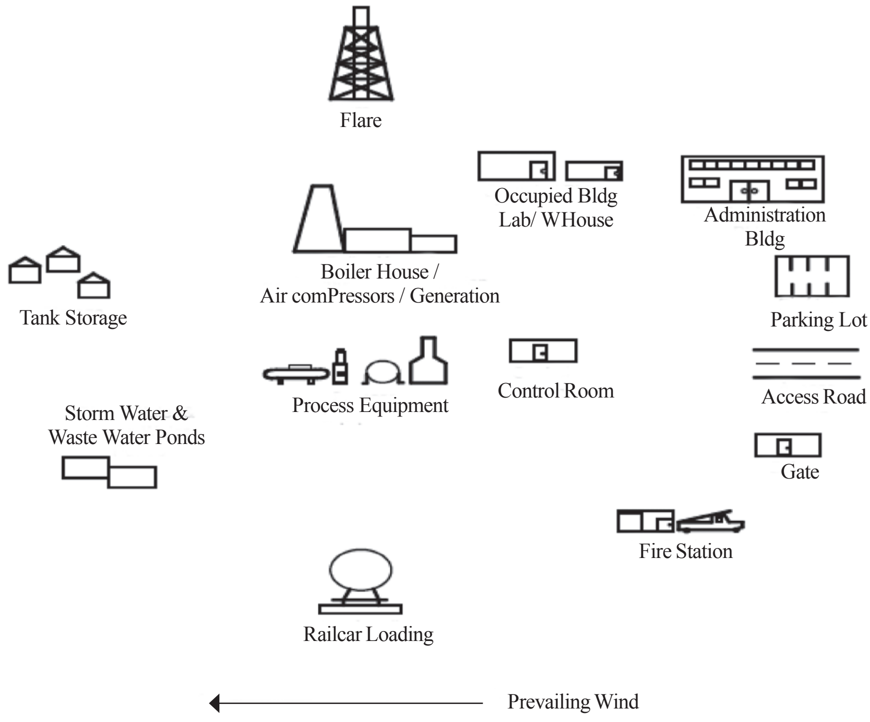
Within the unit plot, the location of the major facilities, i.e., control house, electrical substation, fired heaters, compressors, etc., shall be designed not only in accordance with the spacing standards, but also with consideration of proximity to adjoining facilities, location of roadways, prevailing wind direction, site topography, etc.

Whenever practical, locate process units, utility, flare and other areas with similar open flames at a higher elevation than tank farms and other bulk quantities of hydrocarbons; this minimizes the potential for ignition of hydrocarbon vapor releases or liquid spills.

Where it is not feasible to locate tank farms at elevations lower than process areas, increased fire protection and Gas detection measures may be required to offset the increased potential for ignition. Similar precautions for spills and vapor releases are needed when siting plants that will contain extensive quantities of toxic materials.

Prevailing wind shall be considered when locating ignition sources such as fired heaters and flares. Areas with high concentrations of personnel, such as office buildings, shop areas and existing neighboring community areas, shall also be considered in reference to prevailing wind to reduce potential exposures.

A typical practice for general site lay out shown as follow:



Reference: G.A.P. 2.5.2

The overall layout of the process area must be subdivided with access ways for firefighting, turnarounds, fire risk areas, and maintenance in accordance with the following:

- **Access for Firefighting** - 20 ft. (6 m) minimum width access ways for firefighting are required and these are normally arranged in a rectangular pattern such that the shorter dimension of each rectangle does not exceed 100 ft. (30 m). Consideration should be given to limiting the longer dimension of each subdivision to 400 ft. (120 m) by the inclusion of additional firefighting access ways where necessary. 20 ft. (6 m) of clear space under an overhead pipe band (including the case where air-fin exchangers are located above the pipe band) is acceptable as firefighting access on one side only of an area subdivision.
- **Turnaround Isolation** - 50 ft. (15 m) separation between groups of equipment which shut down separately for turnaround [75 ft. (22.5 m) for light ends units], as detailed in ITEM KK in Table 2.

- **Process Area Subdivisions for Determining Fire Water Rates -** 50 ft. (15 m) minimum separation between process area subdivisions which are used as the bases for determination of fire water application rates, as described in NPC-HSE-S-07.
- **Access for Mobile Equipment for On-Stream and Turnaround Maintenance -** 20 ft. (6 m) minimum access ways are required to allow mobile equipment to reach equipment for maintenance.
- **Fire Risk Areas for the Sizing of Safety Valve Headers -** In laying out a unit plot plan, 20 ft. (6 m) access ways are provided for maintenance. These unit areas, surrounded by at least 20 ft. (6 m) clear spacing on all sides, are the maximum areas which can reasonably be expected to be totally involved in a single fire. These areas are used to determine the combined requirement for pressure relief due to fire exposure. This should not be confused with the areas used to determine fire water and sewer capacities, which are defined as plot subdivision areas in NPC-HSE-S-07. Where the size of the fire risk areas determined by other criteria in Paragraphs 1 through 4, result in an abnormally large flare header size, additional 20 ft. (6 m) access ways may be provided to reduce the fire risk areas.

5.3. Equipment Stacking

A degree of equipment stacking is necessary to achieve a reasonable utilization of plot space, but the potential heavy fire involvement of equipment stacked in several layers must be avoided. The following restrictions apply:

- **Maximum Height -** Avoid stacking more than three levels of equipment.
- **Special Restrictions -** Do not locate equipment over any of the following:
  a. Pumps or compressors handling flammable materials.
  b. Air fin exchangers.
  c. Heat exchangers or drums containing flammable material above 600°F (316°C) or above its auto ignition temperature, whichever is lower.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM A - Gas Compressors		
General - Where there are several large compressors in a unit, it is usually economical for operation and maintenance to locate them in one area. Associated intercoolers, knockout drums, etc., may be located in the compressor area and need not comply with the 25 ft. (7.5 m) spacing to non-associated equipment, provided that they do not restrict access for firefighting and maintenance (generally, spacing of 10 to 15 ft. (3 to 4.5 m) is sufficient). Access for firefighting must be available on at least two sides. Compressor orientation should consider the possibility of massive mechanical failure in relation to surrounding equipment. Compressors handling inert gases or air may be spaced closer than is shown for gas compressors, unless they are in a service critical to plant operations such as main plant instrument and industrial air systems.		
Basic	25 (7.5)	Protects high investment compressor equipment from fires involving other equipment and vice-versa. Small steam or electric motor driven compressors [below 200 hp (150 kW) drivers] may be treated as pumps for spacing and location purposes.
From A (Gas Compressor) From B (Drivers Other Than Steam or Motor)	As required for operations and maintenance	Provide access for operation and maintenance. Compressors may be grouped in the same risk area.
From W (Onsite Pipe bands)	15 (4.5)	Additional spacing (above the basic spacing for onsite pipe bands) is intended to protect piping from potentially large compressor fires. Pipe bands do not pose a significant risk to compressor and reduced spacing (below the basic spacing for compressors) is appropriate. Where the compressor is equipped with provisions for automatic shutdown, isolation and blow down in the event of a fire, spacing can be reduced to 10 ft. (3m)

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM B - Drivers Other Than Steam or Motor		
Basic	35 (10.5)	Provides spacing between equipment handling hydrocarbons from sources of ignition on gas turbines, diesel or gas engines. Steam or motor drivers require maintenance spacing only.
From M (Fired heaters)	25 (7.5)	Minimizes damage to driver in case of fired heater fire.
From W (Onsite Pipe bands)	15 (4.5)	Additional spacing (above the basic spacing for onsite pipe bands) is intended to protect piping from potentially large fires involving drivers. Pipe bands do not pose a significant risk to drivers and reduced spacing (below the basic spacing for drivers) is appropriate. Where the driver is equipped with provisions for automatic shutdown, isolation and blow down in the event of a fire, spacing can be reduced to 10 ft.(3m)
ITEM C - Central And Multi-Unit Control Houses		
General - Location should be next to a roadway. The spacing guidelines on this Table are intended to minimize the exposure of central and multi-unit control houses to fires. Additional spacing beyond that given on this Table may be required depending on blast resistance. Refer to NPC-HSE-S-05 for additional information.		
Basic	100 (30)	Provides space between equipment with vapor release or fire risk potential and important centralized or multi-unit control houses. Additional distance may be required depending upon type of construction and the location of the boundaries of Potential Explosion Domains. See NPC-HSE-S-05 for details.
From W (Onsite Pipe bands)	25 (7.5)	Pipe bands pose a lower fire risk than general process equipment, provided that there are no large concentrations of valves, flanges or other potential release sources within a horizontal radius of 100 ft. (30 m) of control center. Where pipe bands are part of a Potential Explosion Domain, additional spacing may be required. Refer to NPC-HSE-S-05 for details.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM D - Single-Unit Control Houses		
General - Location should be at unit battery limits, next to a roadway. Single unit control houses are rarely used on new construction, but may be present in existing, older plants. The spacing guidelines on this Table are intended to minimize the exposure of single-unit control houses to fires. Additional spacing beyond that given on this Table may be required depending on blast resistance. Refer to NPC-HSE-S-05 for additional information.		
Basic	50 (15)	Provides space between equipment with vapor release or fire risk potential and single-unit control houses. Since the control house is associated with a single unit, less space is required than for central or multi-unit control houses. Additional distance may be required depending upon type of construction and the location of the boundaries of Potential Explosion Domains. **NOTE:** In no case shall a unit control house be located less than 100 ft. (30 m) from the nearest boundary of a Potential Explosion Domain associated with a process unit not controlled by the control house. See NPC-HSE-S-05 for details.
From W (Onsite Pipe bands)	15 (4.5)	Pipe bands pose a lower fire risk than general process equipment, provided that there are no large concentrations of valves, flanges or other potential release sources within a horizontal radius of 50 ft. (15 m) of control house. Where pipe bands are part of a Potential Explosion Domain, additional spacing may be required. Refer to NPC-HSE-S-05 for details.
ITEM E - Air Fin Coolers And Plate Heat Exchangers		
General - Air fin exchangers, with their extensive heat transfer surfaces, are highly vulnerable to failure under fire exposure. A preferred location for air fins is at the opposite side of fractionating towers, away from fired heaters. They should not be located over any of the following: 1. Pumps or compressors handling flammable materials. 2. Drums or heat exchangers containing flammable materials, unless they are part of the same process unit as the air fin cooler.		

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM E - Air Fin Coolers And Plate Heat Exchangers		

3. Drums or heat exchangers containing flammable materials above 600°F (316°C) or above auto ignition temperature, whichever is lower.
4. Preferably, air fins should not be located over or within 10 ft. (3 m) horizontally of major pipe bands except when a better layout is not available, because of the maintenance difficulties that are introduced, in addition to the fire involvement hazard.
If they are so located, then the following requirements must be applied:
a. There should be at least 8 ft. (2.5 m) of head room between the lowest part of the air fin cooler and the top of the pipe rack.
b. Flanges, valves and extraneous manifolds not associated with the air fin cooler should not be located directly below the air cooler or within 10 ft. (3 m) horizontally beyond it in all directions.
c. Normal horizontal spacing requirements from both the air fin cooler and the pipe band to adjacent equipment must be met.
d. Access must be provided for mobile equipment for removal and replacement of motors and air fin units.
e. Supports for the air cooler and pipe band must be fireproofed.
For gasketed plate exchangers, the following precautions are recommended to ensure a safe and reliable plate heat exchanger unit:
1. Spacing requirements for air cooled heat exchangers should be applied to plate exchangers handling flammable fluids.
2. If in flammable service and located less than 20 ft. (6 m) horizontally from equipment with a high potential of fire, plate exchanger should be fireproofed.

Basic	10 (3)	Minimizes damage to air fins, since they are more vulnerable to damage from fire exposure than shell and tube exchangers.
From T (Main Equipment Structures)	As required	Air discharged from air fins should not create operating or maintenance problems on other equipment, in structures.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM F - Cooling Towers		
Basic	50 (15) or 100 (30)	Provides spacing for vapors to disperse. Discharge of vapors can be anticipated from exchanger leakage unless a water disengaging drum is installed upstream. Equipment that does not represent a source of ignition and is lower than the cooling tower may be spaced 50 ft. (15 m) away, and equipment that represents a potential source of ignition or is higher than the tower should be spaced 100 ft. (30 m) away. Spacing and location should also be selected to minimize corrosion, ice formation and visibility problems that may result from impingement of the cooling tower exhaust plume on adjacent high structures.
ITEM G - Process Vessels Below 600°F (316°C) And Below Auto ignition		
Basic	5 (1.5)	Primarily to provide access for operation and maintenance. Also allows some access for firefighting. For very large vessels, consider providing a low toe-wall around the vessel a minimum of 10 ft. (3m) from the vessel walls. The area inside the toe wall should be graded to drain to one or more sealed catch basins located near the toe wall. No other equipment should be installed within the toe-wall.
ITEM H - Electrical Unit Load Centers or Unit Substations		
General - The electrical transformers, circuit breakers, switches, etc., for process units are normally grouped together in a load center or unit substation. Location should be at the edge of the unit.		
Basic	50 (15)	To meet electrical area classification requirements and to protect electrical equipment from fires in the process area. Greater spacing may be required under certain conditions. Greater spacing may also be required for substation buildings depending on blast resistance, criticality of substation, and location of the boundaries of Potential Explosion Domains. See NPC-HSE-S-05 for details.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM H - Electrical Unit Load Centers or Unit Substations		
From W (Onsite Pipebands)	15 (4.5)	Pipe bands pose a lower fire risk than general process equipment, provided that there are no large concentrations of valves, flanges or other potential release sources within a horizontal radius of 50 ft. (15 m) of the load center or substation. Where pipe bands are part of a Potential Explosion Domain, additional spacing to substation buildings may be required depending on blast resistance and criticality of substation. Refer to NPC-HSE-S-05 for details.
ITEM I - Critical Electrical Switch Racks		
General - Critical electrical switch racks consist of electrical supply or control equipment for an individual piece of equipment or small group of equipment the loss of which would interfere with the function of an emergency system, or would result in downtime exceeding week duration. Reduced spacing may be considered for non-critical switch racks based on a case by case evaluation. All switch racks, whether they are considered critical or not, must meet the applicable electrical area classification requirements.		
Basic	50 (15)	To meet electrical area classification requirements and to protect electrical equipment from fire in the process area. Greater spacing may be required under certain conditions. For non-critical radially fed switch racks containing only motor controllers (no associated power transformer or power distribution gear) the spacing may be decreased to 15 ft. (4.5 m) from potential release sources other than fired heaters and to 25 ft. (7.5 m) from fired heaters. The applicable electrical area classification requirements must be met if reduced spacing is used.
From W (Onsite Pipe bands)	25 (7.5)	Pipe bands pose a lower fire risk than general process equipment, provided that there are no large concentrations of valves, flanges or other potential release sources within a horizontal radius of 50 ft. (15 m) of the switch rack.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM J - Emergency Valves For Shut-Off, Isolation, Water Sprays, Etc.		
General - Refer to NPC-HSE-S-10 for information regarding the location of Emergency Shutoff and Isolation Valves for various services.		
ITEM K - Exchangers Handling Flammable Materials Above 600°F (316°C) or Above Auto ignition		
General - The following spacing requirements apply to any exchanger processing flammable materials at an inlet temperature above 600°F (316°C) or above the auto ignition temperature, whichever is lower, on either side.		
Basic	15 (4.5)	Provides some access for firefighting and minimizes damage to other equipment. Probability of fire is greater than with exchangers operating below 600°F (316°C) or below auto ignition, since auto-ignition is likely if leakage of process fluids occurs.
From K , L (Other heat exchangers regardless of temperature)	15 (4.5)	Exchangers that are interconnected by a common process stream may be spaced 3 ft. (1m) apart regardless of operating temperature.
From M (Fired Heaters)	25 (7.5)	Allows reduced spacing considering auto-ignition is likely to occur upon exchanger leakage regardless of fired heater ignition source.
ITEM L - Exchangers Handling Flammable or Combustible Materials Below 600°F (316°C) And Below Auto ignition		
Basic	5 (1.5)	Provides access for operation and maintenance.
From K, (Exchangers operating above 600°F (316°C) or above Auto ignition temperature)	15 (4.5)	Exchangers that are interconnected by a common process stream may be spaced 3 ft. (1 m) apart regardless of operating temperature.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM L - Exchangers Handling Flammable or Combustible Materials Below 600°F (316°C) And Below Auto ignition		
From L (Exchangers operating below 600°F (316°C) and below Auto ignition temperature)	5(1.5)	Exchangers that operate below both 600°F (316°C) and auto ignition temperature may be spaced 5 ft. (1.5 m) apart even if they are not interconnected by a common process stream.
ITEM M - Fired Heaters		
General - Since fired heaters are a constant source of ignition, their location must be carefully selected. Where practical they should be located on the upwind side of a unit, near battery limits, with consideration given to the effects of adjacent units as well as the equipment in the same unit. The top of a fired heater stack should be at least 10 ft. (3 m) higher than any equipment within a horizontal distance of 50 ft. (15 m) and at least 10 ft. (3 m) higher than any working platform that is regularly used by operating or maintenance personnel (once per day or more), and is within a horizontal distance of 100 ft. (30 m).		
Basic	50 (15)	Separates equipment handling flammable material from a constant source of ignition. Spacing also minimizes damage to other equipment in case of a fired heater fire.
From G (Process Vessels Below 600°F (316°C) and below AIT)	50 (15)	As an exception, fuel gas knockout drums should be spaced as close as possible to their respective fired heaters, subject to a minimum of 10 ft. (3 m).
From M (Fired Heaters)	As required	Fired heaters that handle flammable or combustible liquids in the tubes or that shut down individually for turnaround require 25 ft. (7.5m) spacing if coil inlet pressure is less than 1000 psig (6900 kPa). Fired heaters having coil inlet pressures equal to or greater than 1000 psig (6900 kPa) require 50 ft. (15 m) spacing regardless of fluid handled or turnaround considerations. If none of these conditions apply, only normal maintenance spacing is required.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM M - Fired Heaters		
From N (Pumps handling flammable liquids above 600°F (316°C) or above AIT)	25 (7.5)	Minimizes damage to equipment in case of fired heater fire. Fired heaters do not represent a source of ignition, since the stocks in the pumps are already above their auto-ignition temperature.
From P (Equipment handling non-flammables)	15 (4.5)	Minimizes damage to equipment handling non-flammables in case of fired heater fire. Equipment higher than 10 ft. (3 m) should be spaced 25 ft. (7.5 m) from fired heaters having liquid in the tubes.
From Q (Externally Insulated vessels operating above 600°F (316°C) or above AIT).	15 (4.5)	Externally insulated catalytic reforming reactors and their associated fired heaters which handle only vapor in the tubes may be spaced 8 ft. (2.4 m) apart.
From R (Internally insulated vessels operating above 600°F (316°C) or above AIT).	25 (7.5)	Internally insulated catalytic reforming reactors and their respective fired heaters which handle only vapor in the tubes may be spaced 15 ft. (4.5 m) apart.
From W (Onsite pipe bands)	20 (6)	Minimizes involvement of pipe band in fired heater fire. This spacing does not apply between a fired heater and its own individual process and utility lines.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM N - Pumps Handling Flammable Liquids Above 600°F (316°C) or Above Auto ignition		
General - Hot pumps handling flammable materials present an increased fire risk because of the high likelihood of auto ignition when a leak occurs. The general considerations for Item O, Pumps Handling Flammable or Combustible Materials Below Both 600°F (316°C) and auto ignition temperature are also applicable to hot pumps.		
Basic	15 (4.5)	Separates pumps with higher fire potential and provides access for firefighting. Also minimizes damage to other equipment. When a pump and its spare are exposed to a common fire hazard, additional separation between them may be justified to minimize the risk of losing both of them during a fire.
From W (Onsite Pipe bands)	15 (4.5)	Spacing may be reduced to 10 ft. (3 m) if fire water spray coverage is provided on the pumps.
ITEM O - Pumps Handling Flammable or Combustible Liquids Below 600°F (316°C) And Below Auto ignition		
General - Pumps handling flammable or combustible materials are relatively frequent sources of leakage and should therefore be located as far as practical from continuous sources of ignition. They should not be located beneath other equipment, such as towers, drums, or pipe bands in order to avoid fire involvement of such equipment. Also, they should be located clear of overhead obstructions so that they can be reached by a small crane. A preferred location for pumps is on the opposite side of towers, away from fired heaters. In-line pumps should be treated the same as regular pumps for spacing purposes.		
Basic	10 (3)	Provides access for firefighting and minimizes damage in case of fire. Pumps below auto-ignition temperature do not represent as great a fire risk as those above auto-ignition temperature.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM O - Pumps Handling Flammable or Combustible Liquids Below 600°F (316°C) And Below Auto ignition		
From O (Pumps handling flammable or combustible liquids below 600°F (316°C) and below AIT)	3 (1)	Provides access for operation and maintenance.
From W (Onsite Pipe bands)	10 (3)	Spacing may be reduced to 5 ft. (1.5 m) if fire water spray coverage is provided on the pumps.
ITEM P - Equipment Handling Non-Flammables		
Basic	As required for operations and maintenance	No basic spacing is required for this equipment. However, if toxic or corrosive material is handled, special consideration must be given.
ITEM Q - Externally Insulated Vessels Above 600°F (316°C) or Above Auto ignition		
General - Vessels operating above 600°F (316°C) or auto ignition temperature (whichever is lower) represent an increase risk of fire.		
Basic	15 (4.5)	Provides access for firefighting and minimizes damage to other equipment. The external insulation protects the vessel from fires on other equipment.
From Q (Externally insulated vessels above 600°F (316°C) or above AIT)	15 (4.5)	Reactors in the same service may be spaced a minimum of 5 ft. (1.5m) from each other.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM R - Internally Insulated Vessels Above 600°F (316°C) or Above Auto ignition		
General - Internally insulated vessels are more vulnerable to overheating and potential failure when exposed to external fires. Increased spacing is justified.		
Basic	25 (7.5)	Provides access for firefighting and protects exposed shells of internally insulated vessels. Internally lined vessels are more susceptible to fire damage than unlined vessels. Spacing also minimizes damage to other equipment in case of a vessel fire.
From P (Equipment handling non-flammables)	15 (4.5)	Equipment handling non-flammables does not represent a fire risk to other vessels.
From R (Internally insulated vessels above 600°F (316°C) or above AIT)	25 (7.5)	Reactors in the same service may be spaced a minimum of 5 ft. (1.5m) from each other.
From T (Main Equipment Structure)	15 (4.5)	Provides some access for firefighting around reactor and minimizes damage to structure.
ITEM S - Onsite Pressure Storage Vessels		
General - Even though an offsite location is preferred, it is sometimes necessary for process reasons to provide onsite surge capacity for light ends liquids. The inventory of such onsite pressure storage should be held to a minimum and location should be at the unit boundary, as far as possible from fired heaters and hot equipment.		
Basic	75 (22.5)	Minimizes exposure to unit equipment from potential source of severe fire.
From S (Onsite Pressure Storage Vessels)	1 Diameter	Provides nominal spacing between vessels to give access for operation, maintenance and firefighting.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM T - Main Equipment Structures		
Basic	N.A.	It is impractical to set a basic spacing requirement, since the purpose of the structure is to support equipment. In most cases, equipment-to-equipment spacing will govern.
ITEM U - Remote Instrument Modules (RIM's)		
Basic	100 (30)	Provides space between equipment with vapor release or fire risk potential and remote instrument buildings. For single-unit RIM's, the minimum basic spacing may be reduced to 25 ft. (15m) from any equipment containing flammable liquids or vapors. Additional distance may be required depending upon type of construction and the location of the boundaries of Potential Explosion Domains. See NPC-HSE-S-05 for details.
From W (Onsite Pipe bands)	25 (7.5)	Pipe bands pose a lower fire risk than general process equipment, provided that there are no large concentrations of valves, flanges or other potential release sources within a horizontal radius of 100 ft. (30m) of RIB. For single unit RIM's, the minimum spacing may be reduced to 15 ft. (4.5m) provided that there are no large concentrations of valves, flanges or other potential release sources within a horizontal radius of 50 ft. (15m). Where pipe bands are part of a Potential Explosion Domain, additional spacing may be required. Refer to NPC-HSE-S-05 for details.
ITEM V - Blow down And Water Disengaging Drums		
Basic	25 (7.5)	Locate in a separate unit area, to avoid involvement in unit fires. Greater spacing to certain higher risk equipment is required (see chart). Water disengaging drums are spaced as for blow down drums if vented to the flare. If vented to the atmosphere, normal process drum spacing is adequate unless separate turnaround considerations apply.

Table 1
On-Site Spacing, Guide to Figure 1

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM W - Onsite Pipe bands		
Basic	15 (4.5)	Provides spacing from pipe bands to equipment for operation, maintenance, and to reduce involvement from adjacent fires. *Process equipment should not be located under onsite pipe bands.*
From N (Pumps Handling Flammable or Combustible Materials Above 600°F(316°C) or Above AIT)	15 (4.5)	These distances refer to the horizontal spacing between the edge of an overhead pipe rack and the nearest part of the pump. There are no limitations on spacing pump drivers from pipe racks. Spacing may be reduced to 10 ft. (3m) if fire water spray coverage is provided.
From O (Pumps Handling Flammable or Combustible Materials Below 600°F (316°C) and Below AIT)	10 (3)	These distances refer to the horizontal spacing between the edge of an overhead pipe rack and the nearest part of the pump. There are no limitations on spacing pump drivers from pipe racks Spacing may be reduced to 5 ft. (1.5m) if fire water spray coverage is provided.

6. Other On-Site Spacing Requirements

6.1. Portable Gas Cylinders

The vulnerability to fire exposure of portable gas cylinders located in plant areas (e.g., chlorine or ammonia for injection into the process, nitrogen for pressuring lube-oil reservoirs, hydrogen for continuous analyzers, etc.) must be considered in selecting their location.

1. High-risk cylinder installations (volatile, flammable or toxic liquids, e.g., Cl_2, NH_3; and multiple flammable gas cylinders, e.g., H_2) should be located at the edge of the plant area, 25 ft. (7.5 m) from other process equipment, with good access for firewater cooling. Bulk storage of such cylinders shall be minimized and located in a remote offsite area at least 200 ft. (60 m) from process areas.
2. Lower risk installations (single cylinders containing flammable or inert gases, e.g., H_2, N_2) shall be located at the edge of a fire risk area, and at least 25 ft. (7.5 m) from high fire risk equipment such as pumps and fired heaters.

6.2. Chemical Injection Facilities

If the onsite inventory of chemicals being injected exceeds 200 gallons (0.75 m^3), the location of chemical injection facilities should conform to the following:

1. Flammable Materials - Follow the spacing guidelines for ITEM O, Table 1.
2. Non-Flammable Materials - No basic spacing is required, but if toxic or corrosive material is injected, special considerations must be given. If inventory being injected is 200 gallons (0.75 m^3) or less, no special spacing requirements are needed except to provide access for operations and maintenance.

6.3. Nitrogen Storage and Vaporizer

Cryogenic storage of nitrogen should be located at least 25 ft. (7.5 m) from other process equipment, preferably at edge of corner of process area where fire involvement is minimized and where access for truck deliveries is available.

6.4. Bulk Oxygen Storage

Locate at least 50 ft. (15 m) from process equipment, preferably across the road from process area, since oxygen is reactive with hydrocarbons and risk of contact should be minimized.

6.5. Safety Showers

In areas where corrosive chemicals are handled and personnel exposure may occur, causing skin burns or eye injury, safety showers with eye baths shall be provided. They should be located so that any such equipment is directly accessible and not further than 50 ft. (15 m) from a safety shower. Closer spacing should be considered for highly corrosive chemicals such as HF.

6.6. Steam Turbine and Engine Exhausts

Atmospheric exhausts from steam turbines, gas turbines, and internal combustion engines should be located such that they do not present a hazard to personnel on adjacent working platforms. They should be at least 10 ft. (3 m) higher than any platform or access within a horizontal distance of 25 ft. (7.5 m).

6.7. Sewer Vents

These should be located with respect to ignition sources as required by hazardous area classification.

6.8. Electrical and Instrument Cable and Wiring

Location of electrical and instrument cable and wiring is determined as per its specific standards, respectively.

6.9. Air Intake Location

The potential for contamination of the air drawn from the atmosphere introduces the possibility or internal explosion if a flammable vapor/air mixture is formed, or of a hazard or nuisance to personnel if toxic or malodorous vapors enter a force-ventilated working space. The locations of air intakes in relations to adjacent equipment (safety valve and other vents, exhaust from engines or steam turbines, air fins, equipment liable to uncontrolled release such as pumps, etc.) must, therefore, be carefully selected, taking into account factors such as prevailing wind direction, probable horizontal and vertical dispersion patterns and the degree of hazard which would result from contamination. As a minimum requirement, the location of any air intake in a chemical plant should be in an **unclassified** or **safe** Area as defined by the electrical area classification procedure. Air intakes should be located a minimum of 25 ft. (7.5m) above grade when within 50 ft. (15 m) of process units or within 100 ft. (30m) of process units containing light ends. The following additional requirements apply in certain cases:

1. Air intakes to inert gas generators, gas turbines, internal combustion engines, air compressors, or other equipment located within enclosed or partially enclosed buildings should be located outside the building.

2. Hydrocarbon contamination of air intakes to air separation plants cannot be tolerated and minimum spacing of 150 ft. (45m) to possible hydrocarbon sources is necessary. The specific design of the plant should be evaluated to ensure air contaminants are detected and captured by mole sieves prior to processing.

3. Air intakes to control house pressure ventilation and air conditioning systems shall be elevated at least 40 ft. (12 m) above grade. For centralized control rooms the addition of a combustible gas detector and alarm in the intake is recommended.

6.10. On-Site Atmospheric Storage Tanks

Atmospheric storage tanks within process unit battery limits shall be limited to a maximum size of 80 m^3 when they contain flammable liquid or a combustible liquid heated above its flash point. Diked enclosures that meet NFPA 30 shall be provided for all storage tanks placed in process units. Process equipment and pumps should not be located in diked areas. Diked areas shall be sized for the largest tank in the enclosure and shall be provided with fire sealed drains piped to the oily water sewer or chemical drain. Drains shall be equipped with an indicating type shut-off valve located outside of the diked area. Tanks shall be spaced so that at least 30 ft. (9 m) separation is provided between enclosure dikes and process unit

6.11. Paving in Onsite Areas

Concrete paving is required under any equipment where flammable or hazardous liquids may be spilled during routine operating or maintenance functions. This includes areas under pipe bands where low point draining, sampling, insertion of blinds, etc., are frequently carried out. Toe-walls are required around fired heaters, per NPC-HSE-S-02, and at the edge of areas paved with special chemical resistant materials or which have segregated drainage for special chemicals, etc. They may also be used around other equipment subject to spillage, e.g., pump groups, to direct drainage flow to catch basins.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM AA - Boiler Houses, Power Generating Facilities		
Basic	150 (45)	Protects critical steam and power facilities feeding major portions of the plant from a possible fire or explosion in equipment handling hydrocarbons. Non-critical steam generators, which supply a small proportion of the total steam demand (less than 25%) or which supply steam only to a particular plant, may be located in a process area with spacing as for process fired heaters.
From AA, EE, GG (Boiler Houses, Power Generating Facilities, Major Electrical Distribution Centers, Main Fire Pumps)	Provide access for operation and maintenance	Boilers, generators and electrical distribution centers may be located in the same area with no basic spacing between similar facilities. Fire pumps may also be located in this area, since they are often operated by boiler house personnel.
From BB, FF, NN (Other Occupied Buildings, Fire Houses , Railroad Main Lines)	100 (30)	Boilers and generating stations constitute only a minor risk to these other facilities which do not represent a significant hazard to the boiler or power facilities, hence reduced spacing is acceptable.
From LL (Property Lines)	150 (45)	100 ft. (30 m) is acceptable if the adjoining property has only buildings or other low-risk facilities on it.
From OO (Railroad spurs)	25 (7.5)	Space provides protection against collision in case of derailment.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM BB - Other Occupied Buildings (Offices, Workshops, Laboratory, Etc.)		
Basic	150 (45)	Protects personnel from possible fires in major plant equipment. Additional distance may be required depending on type of construction and location of the boundaries of Potential Explosion Domains. Refer to NPC-HSE-S-05 for details.
From AA, (Boiler Houses, Power Generating Facilities, Cooling Towers)	100 (30)	Boiler Houses, Power Generating Facilities and Cooling Towers represent a lower risk to buildings than most processing facilities and reduced spacing is acceptable. Buildings do not represent a risk to such facilities.
From DD, KK, and VV (Docks, Process Areas, and LPG Loading Racks)	200 (60)	Provides additional spacing from these higher-risk facilities for personnel protection. Additional spacing may be required depending on type of construction and location of the boundaries of Potential Explosion Domains. Refer to NPC-HSE-S-05 for details.
From JJ (Major Offsite Pipe bands)	50 (15)	Offsite pipe bands represent a relatively low risk to buildings. Other occupied buildings may be closer to pipe bands than to general process equipment provided that the lines are all welded and there are no large concentrations of flanges, valves or other potential leak sources within 200 ft. (60 m).
From LL (Property Lines)	As required for operations and maintenance	Minimum operational and maintenance access spacing is acceptable if facilities on adjoining property are not harmful. Greater spacing, similar to internal spacing requirements, should be used if process plants, storage tanks, or other risks are located adjacent to our property line.
From NN, OO (Railroad Main Lines, Railroad Spurs)	25 (7.5)	Railroads represent minimum risk. Space provides protection against collision damage in case of derailment.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM CC - Cooling Towers		
Basic	50 (15) or 100 (30)	Provides spacing for vapors to disperse. Equipment that does not represent a source of ignition and is lower than the cooling tower may be spaced 50 ft. (15 m) away, and equipment that represents a potential source of ignition or is higher than the tower should be spaced 100 ft. (30 m) away
ITEM DD - DOCKS (LOADING MANIFOLD AREAS)		
Basic	200 (60)	Protects major plant equipment from fires on docks or tankers and vice-versa. Spacing also allows dispersal of vapors emitted during loading operations.
From DD (Docks)	As required by operations and maintenance	It is impractical to set the spacing between docks, since this is governed by tanker operational requirements.
ITEM EE - MAJOR ELECTRICAL DISTRIBUTION CENTERS		
Basic	150 (45)	To meet electrical area classification requirements and to protect against potential damage from fire or explosion on equipment in hydrocarbon service, this could result in loss of critical electrical facilities and shut down of major parts of the plant. For distances between 150 ft. (45 m) and 700 ft. (215 m), blast protection may be required per NPC-HSE-S-05.
From JJ (Major Offsite Pipe bands)	25 (7.5 m)	Offsite pipe bands represent a relatively low risk to electrical distribution centers. Electrical distribution centers may be closer to pipe bands than to general process equipment provided that the lines are all welded and there are no large concentrations of flanges, valves or other potential leak sources within a horizontal radius of 100 ft. (30 m).
From LL (Property Lines)	As required for operations and maintenance	Minimum operational and maintenance access spacing is acceptable if facilities on adjoining property are not harmful. Greater spacing, similar to our internal spacing requirements, should be used if process plants, storage tanks, or other risks are located adjacent to property line.
From NN, OO (Railroad Main Lines, Railroad Spurs)	25 (7.5 m)	Railroads represent minimum risk to electrical distribution centers. Space provides protection against collision damage in case of derailment.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM FF - Fire Houses		
General - See NPC-HSE-S-07 for further details.		
Basic	150 (45)	Provides separation from major equipment in hydrocarbon service to prevent loss of fire vehicles and equipment. Additional spacing may be required depending on type of construction and location of the boundaries of Potential Explosion Domains. Refer to NPC-HSE-S-05 for details.
From LL (Property Lines)	As required for operations and maintenance	Minimum operational and maintenance access spacing is acceptable if facilities on adjoining property are not harmful. Greater spacing, similar to internal spacing requirements, should be used if process plants, storage tanks, or other risks are located adjacent to our property line.
From NN, OO (Railroad Main Lines, Railroad Spurs)	25 (7.5)	Railroads represent minimum risk to buildings. Spacing prevents collision damage in case of derailment.
ITEM GG - Main Fire Pumps		
Basic	150 (45)	Provides spacing between fire pumps and major plant equipment in hydrocarbon service to prevent loss of fire pumps from fire exposure. Special purpose fire pumps, e.g., for specific plants, may be spaced closer to non-related facilities.
From CC (Cooling Towers)	100 (30)	Cooling tower fires are relatively rare, hence spacing may be reduced to 100 ft. (30 m).
From MM (Major Pump Areas)	100 (30)	Fires in offsite pump areas are probably less severe; hence, spacing may be reduced to 100 ft. (30 m).
From NN, OO (Railroad Main Lines, Railroad Spurs)	25 (7.5)	Railroads represent minimum risk. Spacing prevents collision damage in case of derailment.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM HH - Fire Training Grounds		
General - See NPC-HSE-S-07 for further details.		
Basic	150 (45)	Provides separation from equipment handling hydrocarbons, since training fires are a source of ignition. Also allows for smoke dispersal.
From SS (Atmospheric Storage of High Flash Stocks)	100 (30)	Reduced risk to high-flash tankage, hence reduced spacing is acceptable.
ITEM II - Loading Racks Except For Light Ends		
General - The main loading and unloading racks for road tank trucks and rail tank cars should be consolidated at one location at the periphery of the main plant facilities, and close to an access gate, so that traffic through the plant is minimized and high risk areas are avoided. Refer to item VV for loading racks where high flash products are handled at the same loading racks as low flash products. Single road truck racks for low volume chemicals (e.g., sulfuric acid liquid sulfur) may be located at the edge of a process area. A slip road to the rack should be provided such that: (a) the rack is at least 15 ft. (4.5 m) from the nearest process equipment [20 ft. (6 m) if the adjacent equipment contains light ends] and (b) a discharging vehicle is at least 15 ft. (4.5 m) from the adjacent roadway. Tank truck loading and unloading rack areas must be provided with adequate space and roadways for safe truck maneuvering and parking, as well as safe access to and from the loading racks and weigh-scale (where included). Rail loading and unloading rack areas must also have adequate allowance of rail spur track for parking and shunting tank cars. For LPG Loading Racks, refer to item VV.		
Basic	150 (45)	Allows dispersal of vapors that may be released at the rack during loading or in case of a liquid spill. Also, minimizes damage to other equipment in case of fire at the racks and vice-versa.
From II (Loading Racks), OO (Railroad Spurs)	N.A.	Not applicable. Spacing between racks is set by operational requirements covering rail cars and road trucks.
From JJ (Major Offsite Pipe bands)	50 (15)	Lower risk of affecting pipe bands; hence, reduced spacing is acceptable.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM JJ - Major Offsite Pipe bands		
General - Piping which interconnects units, or units and storage, must be routed such that it will not remain in service through plant areas which are shut down for turnaround.		
Basic	150 (45)	Prevents damage to other equipment in case of fire or explosion .
From KK (Process Areas)	50 (15)	Process-to-process spacing to allow independent shudowns for turnaround. Light ends units must be separated by 75 ft. (22.5 m). Where a number of process units are integrated into one process area and shut down together, the internal spacing should be based on operation and maintenance requirements, access for firefighting, and spacing between subdivision and fire-risk areas.
TEM LL - Property Lines		
Basic	200 (60)	Spacing is required between plant facilities and property line to minimize exposure to the public. Spacing may have to be increased depending upon: explosion overpressure considerations (see NPC-HSE-S-05); facilities immediately beyond the fence line; public relations; fire exposure; toxic exposure; noise considerations; and site security considerations.
From BB (Other Occupied Buildings) EE (Major Electrical Distribution Centers) FF (Fire Houses) GG (Main Fire Pump)	As required for operations and maintenance	Minimum operational and maintenance access spacing is acceptable if facilities on adjoining property are innocuous. Greater spacing, similar to our internal spacing requirements, should be used if process plants, storage tanks, or other risks are located adjacent to our property line.
From MM (Major Pump Areas)	150 (45)	

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM MM - Major Pump Areas		
Basic	50 (15)	Provides spacing to other major equipment to minimize damage in case of fire. Also allows access for firefighting.
From LL (Property Lines) NN (Railroad Main Lines)	150 (45)	Pump areas represent reduced risk to property line or main line railroads; hence, spacing can be less than the basic 200 ft. (60 m) for these facilities.
From QQ (Pressure Storage) RR (Low Flash Atmospheric Storage) SS (High Flash Atmospheric Storage) TT (Refrigerated Storage)	-	As a minimum spacing, pump areas must be located outside tank dikes.
ITEM NN - Railroad Main Lines		
Basic	200 (60)	Minimizes exposure of personnel and railroad equipment to fire or explosion on plant equipment.
From NN (Railroad Main Lines) OO (Railroad Spurs)	N.A.	Not applicable. Spacing is set by railroad operational requirements.
From SS (High flash atmospheric storage)	150 (45)	High-flash tankage is minimum risk to railroad; hence, less than basic spacing.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM OO - Railroad Spurs		
General - When a railroad spurs runs along a boundary of a process unit, the preferred layout has the fired heaters and other continuous ignition sources placed at the same side of the plant as the rail track.		
Basic	50 (15)	Provides spacing from major plant equipment handling flammables to reduce the likelihood of ignition from railroad engines.
From OO (Railroad Spurs)	N.A.	Not applicable. Spacing between spur lines is set by railroad requirements.
ITEM PP - Main Separators or Skimming Ponds		
Basic	150 (45)	Separation from processing areas and other sources of ignition to prevent vapors discharged at the separator from igniting. Distance to lower risk facilities that do not represent a source of ignition is based on basic spacing for those facilities.
ITEM QQ - Pressure Storage		
General - If possible, do not locate uphill from a process plant or other critical facilities or populated areas. Equipment other than associated piping must not be located within diked areas of storage vessels. Pumps must be at least 25 ft. (7.5 m) from the vessel shell. The vessel shell should be at a least 10 ft. (3 m) from the dike wall.		
Basic	200 (60)	Provides spacing to contain a tank fire and prevent involving other facilities or personnel. Also, protects tankage in case of fire at other facilities. As a minimum spacing, dikes must be located 50 ft. (15 m) from process units.
From JJ (Major Offsite Pipe bands)	15 (4.5)	Pipe bands do not represent a significant risk to pressure storage. Reduced spacing is acceptable.
From OO (Railroad Spurs)	50 (15)	
From PP (Main Separators and Skimming Ponds)	150 (45)	

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM RR - Atmospheric Storage of Low-Flash Stocks		
General - Equipment other than associated piping must not be located within diked areas of storage vessels. If possible, do not locate uphill from a process plant or other critical facilities or populated areas.		
Basic	150 (45)	Provides spacing to contain a tank fire and prevent involving other facilities or personnel. Also, protects tankage in case of fire at other facilities. As a minimum spacing, dikes must be located 50 ft. (15 m) from process units.
From JJ (Major Offsite Pipe bands)	15 (4.5)	Pipe bands do not represent a significant risk to atmospheric storage.
From OO (Railroad Spurs)	150 (45)	Atmospheric storage does not represent a significant risk to railroad spurs.
ITEM SS - Atmospheric Storage Of High-Flash Stocks		
General - Equipment other than associated piping must not be located within diked areas of storage vessels. If possible, do not locate uphill from a process plant or other critical facilities or populated areas.		
Basic	100 (30)	Provides spacing to contain a tank fire and prevent involving other facilities or personnel. High-flash stocks represents low fire risk; hence, less spacing is required than for low-flash materials. Also, protects tankage in case of fires at other facilities.
From JJ (Major Offsite Pipe bands)	15 (4.5)	Pipe bands do not represent a significant risk to atmospheric storage.
From OO (Railroad Spurs)	50 (15)	Atmospheric storage does not represent a significant risk to railroad spurs.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
ITEM TT - Refrigerated Storage		
General - If possible, do not locate uphill from a process plant or other critical facilities or populated areas. Equipment other than associated piping must not be located within diked areas of storage vessels. Pumps must be at least 25 ft. (7.5 m) from the vessel shell. The vessel shell should be at a least 10 ft. (3 m) from the dike wall.		
Basic	200 (60)	Provides spacing to contain a tank fire and prevent involving other facilities or personnel. Also, protects tankage in case of fire at other facilities. As a minimum spacing, dikes must be located 50 ft. (15 m) from process units.
From JJ (Major Offsite Pipe bands)	15 (4.5)	Pipe bands do not represent a significant risk to refrigerated storage.
From OO (Railroad Spurs)	50 (15)	Refrigerated storage does not represent a significant risk to railroad spurs.
From PP (Main Separators and Skimming Ponds)	150 (45)	Refrigerated storage does not represent a significant risk to main separators and skimming ponds.

Table 2
Off-Site Spacing, Guide to Figure 2

SPACING	DISTANCE, ft. (m)	REMARKS
Item VV - Light Ends Loading Racks		

Light Ends Loading Racks represent a higher risk to other facilities than loading racks for other products due to the higher potential for vapor cloud formation in the event of an accidental spill. A flammable vapor cloud may travel long distances than a liquid spill and could potentially be ignited by a remote source of ignition. Because of this, increased spacing is justified between Light Ends loading racks and other facilities.

Basic	200 (60)	Allows dispersal of vapors that may be released at the rack during loading or in case of a liquid spill. Also, minimizes damage to other equipment in case of fire at the racks and vice-versa.
From CC (Cooling Towers)	150 (45)	Cooling towers represent a lower ignition risk than other facilities.
From JJ (Major Offsite Pipe bands)	50 (15)	Offsite pipe bands do not represent a risk of ignition.
From OO (Railroad Spurs)	150 (45)	Railroad spurs represent a lower risk of ignition than railroad main lines due to less traffic and better traffic control. The minimum spacing does not apply to railroad spurs serving LPG loading racks. This is set by operational requirements.

7. Other Off-Site Spacing Requirements

7.1. Flares

Minimum flare spacing from other equipment is 200 ft. (60 m) and depending on radiant heat calculations may exceed this distance. Height of elevated flares is determined by radiant heat considerations. Public relations, air pollution, and noise factors are also taken into account and may influence spacing from the property line.

7.2. Plant Roadways

Four types of plant roads and access ways are used:

1. Major plant roads:

These are roads that have unrestricted traffic. They are a minimum of 25 ft. (7.5 m) wide and include a 25 ft. (7.5 m) clearance on each side to prevent vehicle / equipment collisions. Pipe bands may be run within the 25 ft. (7.5 m) clear space on each side of major plant roads provided that the distance from the edge of the road and the nearest edge of the pipe band is a minimum of 15 ft. (4.5 m).

2. Secondary plant roads:

These are roads between process units that are used mainly for maintenance and turnaround activities, and for firefighting access. They are a minimum of 20 ft. (6 m) wide and include a 15 ft. (4.5 m) clearance on each side to prevent vehicle /equipment collisions. These roads set the minimum separation requirement between different turnaround areas (50 ft. (15 m)). Pipe bands may be run within the 15 ft. (4.5 m) clear space on each side of secondary plant roads provided that the distance from the edge of the road and the nearest edge of the pipe band is a minimum of 10 ft. (3 m).

3. Access roads:

These roads provide unrestricted single vehicle width access to outlying areas and tank fields. They are a minimum of 10 ft. (3 m) wide and include a 5 ft. (1.5 m) clearance on each side.

4. Plant access ways:

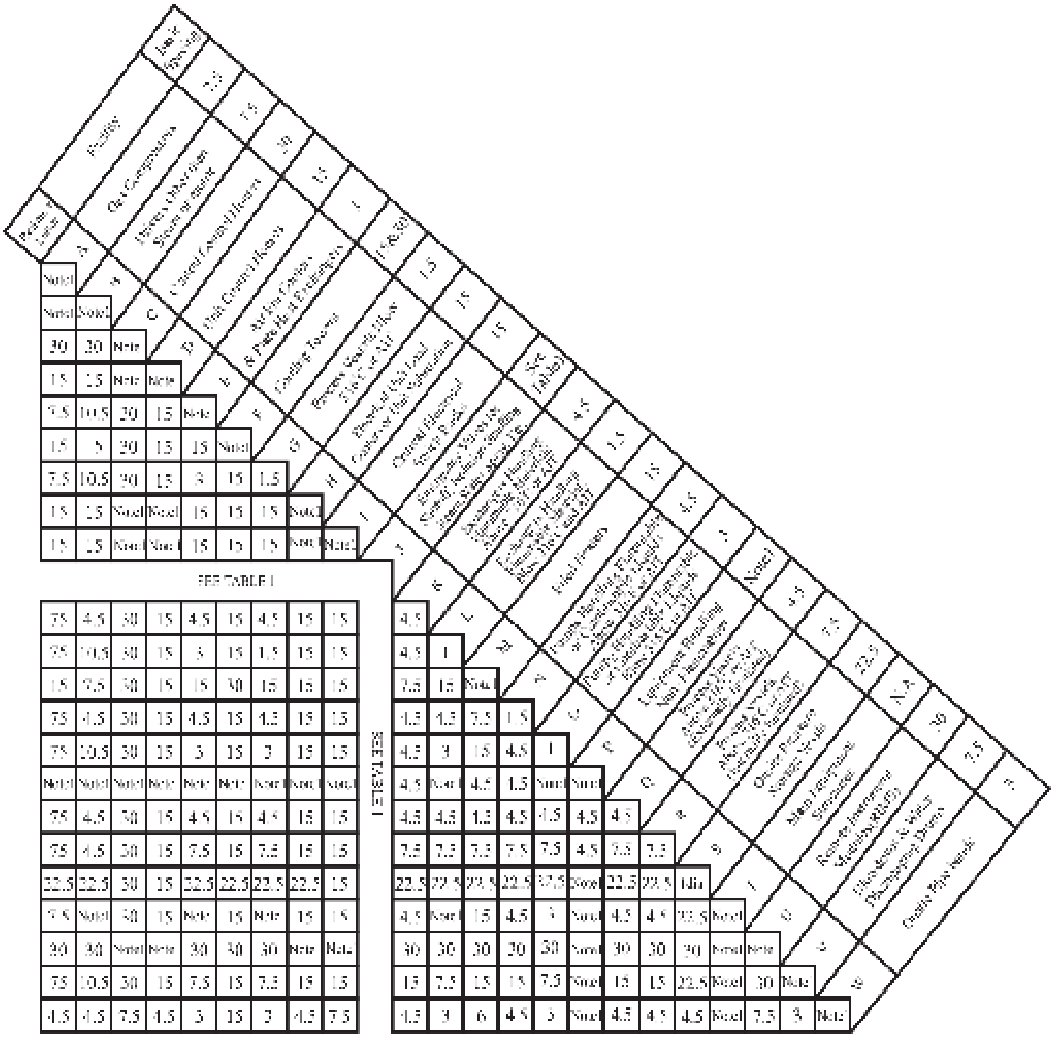
Plant access ways provide controlled access for maintenance vehicles into process areas. They are a minimum of 10 ft. (3 m) wide and include a 5 ft. (1.5 m) clearance on each side. A 10 ft. (3 m) access road may be formed along the top of the dike wall of an atmospheric tank (but not a .hot. tank) if economically attractive, with 5 ft. (1.5 m) level clearance on each side and a safety curb. The radius of curvature at bends must be such as to permit safe passage of fire trucks. Plant access ways are included between adjacent units and fire risk areas in a process plant area, located centrally in the clear space provided.

7.3. Paving in Off-Site Areas

In offsite areas, concrete paving is required under any equipment where flammable or hazardous liquids may be spilled during routine operating or maintenance functions, e.g., at loading or unloading racks, loading pumps, filters, water draw offs, etc.

Paved areas must be graded and drained to a suitable drainage system, in accordance with common practices. Grading must be such as to direct spills away from high-risk areas such as loading/unloading racks.

Figure1
On-Site Spacing Chart (Metric)



NA:Not Applicable

Note 1: Provide spacing based on access for operation and maintenance

Reference: Exxon-Mobil Engineering Standards)

Figure 2
Offsite Spacing Chart (Metric)

Refrence Letter	Facility	Basic Spacing	AA	BB	CC	DD	EE	FF	GG	HH	II	JJ	KK	LL	MM	NN	OO	PP	QQ	RR	SS	TT	UU	VV
AA	Boiler Houses, Power Generating Facilities	45	Note1																					
BB	Other Occupied Buildings (Main Office, Workshops, Laboratory, Etc)	45	30	Note1																				
CC	Cooling Towers	15&30	45	30	Note1																			
DD	Docks(Manifold Loading Areas)	60	60	60	30	NA																		
EE	Major Electrical Distribution Centers	45	Note1	Note1	45	60	Note1																	
FF	Fire House	45	30	Note1	30	60	Note1	NA																
GG	Main Fire Pumps	45	Note1	Note1	30	60	Note1	Note1	Note1															
HH	Fire Training Grounds	45	45	45	45	45	45	45	15	NA														
II	Loading Racks For (except Light Ends)	45	45	45	45	60	45	45	45	45	NA													
JJ	Major Offsite Pipebands	4.5	4.5	15	4.5	4.5	7.5	4.5	4.5	45	15	NA												
KK	Process Areas	45	45	60	30	60	45	45	45	45	45	4.5	15											
LL	Property Lines	60	45	Note1	30	60	Note1	Note1	Note1	60	45	4.5	60	NA										
MM	Major Pump Areas	15	45	45	15	15	45	45	30	45	15	4.5	15	45	Note1									
NN	Railroad Main Lines	60	30	7.5	30	60	7.5	7.5	7.5	60	45	7.5	60	NA	45	NA								
OO	Railroad Spurs	15	7.5	7.5	15	15	7.5	7.5	7.5	15	NA	4.5	15	NA	15	NA	NA							
PP	Main Seperators or Skimming Ponds	45	45	45	30	45	45	45	45	45	45	4.5	45	60	15	60	30	Note1						
QQ	Pressure Storage	60	60	Note2	60	60	60	60	60	60	60	4.5	60	Note2	SEE TABLE 2	60	15	45	Note2					
RR	Atmospheric Storage of Low-flash Stocks	45	45	Note2	Note2	45	45	45	45	45	45	4.5	45	Note2		60	15	45	Note2	Note2				
SS	Atmospheric Storage of High-flash Stocks	30	30	Note2	Note2	30	30	30	30	30	30	4.5	30	Note2		45	15	30	Note2	Note2	Note2			
TT	Refrigerated Storage	60	60	Note2	60	60	60	60	60	60	60	4.5	60	Note2		60	15	45	Note2	Note2	Note2	Note2		
UU	TEL, TML & Cetane Improver Facilities	30	45	45	30	30	45	30	30	45	45	4.5	45	45	30	60	15	30	30	30	60	60	NA	
VV	Light Ends Loading Racks	60	60	60	60	60	60	60	60	60	60	15	60	60	15	45	45	60	60	60	60	60	60	Note1

NA:Not Applicable

Note 1: Provide spacing based on access for operation and maintenance

Note 2: Refer to design standards and Figure 3C

Reference: Exxon-Mobil Engineering Standards)

Figure 3A

Inter-Unit Spacing Recommendations for Oil and Chemical Plants.

	Service Buildings	Motor Control Centers And Electrical Substations	Utilities Areas	Cooling Towers	Control Rooms	Compressor Buildings	Large Pump Houses	Process Units Moderate Hazard	Process Units Intermediate Hazard	Process Units High Hazard	Atmospheric Storage Tanks	Pressure Storage Tanks	Refrigerated Storage Tanks Dome Roof	Flares	Unloading And Loading Racks	Fire Water Pumps	Fire Station
Service Buildings	/																
Motor Control Centers And Electrical Substations	/	/															
Utilities Areas	50	50	/														
Cooling Towers	50	50	100	50 ①													
Control Rooms	/	/	100	100	/												
Compressor Buildings	100	100	100	100	100	30											
Large Pump Houses	100	100	100	100	100	30	30										
Process Units Moderate Hazard	100	100	100	100	100	30	30	50									
Process Units Intermediate Hazard	200	100	100	100	200	50	50	100	100								
Process Units High Hazard	400	200	200	200	300	100	100	200	200	200							
Atmospheric Storage Tanks	250	250	250	250	250	250	250	250	300	350	*						
Pressure Storage Tanks	350	350	350	350	350	350	350	350	350	350	*	*					
Refrigerated Storage Tanks Dome Roof	350	350	350	350	350	350	350	350	350	350	*	*	*				
Flares	② 300	300	300	300	300	300	300	300	300	300	300	400	400	/			
Unloading And Loading Racks	200	200	200	200	200	200	200	200	200	300	250	350	350	300	50		
Fire Water Pumps	50	50	50	50	50	200	200	200	300	300	350	350	350	300	200	/	
Fire Station	50	50	50	50	50	200	200	200	300	300	350	350	350	300	200	/	/

1 ft. = 0.305 m
/ = No Spacing Requirements
* = Spacing Given in Table 3

Examples:
① 50 ft. Seperation between Two Cooling Towers
② 300 ft. Seperation between Service Building and Flare
(Reference: G.A.P. 2.5.2)

Figure 3B
Intra-Unit Spacing Recommendations for Oil and Chemical Plants.

	Compressors	Intermediate Hazard Pumps	High Hazard Pumps	High Hazard Reactors	Intermediate Hazard Reactors	Moderate Hazard Reactors	Columns, Accumulators, Drums	Rundown Tanks	Fired Heaters, Incinerators, Oxidizers	Air Cooled Heat Exchanger	Heat Exchanger	Pipe Racks	Unit Block Valves	Analyzer Rooms
Compressors	30													
Intermediate Hazard Pumps	30	5												
High Hazard Pumps	50	5	5											
High Hazard Reactors	50	10	15	25										
Intermediate Hazard Reactors	50	10	15	25	15									
Moderate Hazard Reactors	50	10	15	25	15	15								
Columns, Accumulators, Drums	50	10	15	50	25	25	15							
Rundown Tanks	100	100	100	100	100	100	100	100						
Fired Heaters, Incinerators, Oxidizers	50	50	50	50	50	50	50	100	25					
Air Cooled Heat Exchanger	30	15	15	25	15	15	15	100	50	/				
Heat Exchanger	30	10	15	25	15	10	10	100	50	15	5			
Pipe Racks	30	10	15	25	15	10	10	100	50	/	10	50		
Unit Block Valves	50	50	50	100	50	50	50	100	50	50	50	200	/	
Analyzer Rooms	50	50	50	50	50	50	50	100	50	50	50	200	/	/

1 ft. = 0.305 m
/ = No Spacing Requirements
(Reference: G.A.P. 2.5.2)

Figure 3C
Storage Tank Spacing Recommendations for Oil and Chemical Plants.

	Floating & Cone Roof Tanks < 3000 Barrels	Floating & Cone Roof Tanks > 3000 < 10.000 Barrels	Floating & Cone Roof Tanks > 10.000 < 3000 Barrels	Jumbo Floating Roof Tanks > 300.000 Barrels	Cone Roof Tanks, Class II, III Product > 10.000 < 300.000 Barrels	Cone Roof Tanks, Inerted Class I Product** > 10.000 < 150.000 Barrels	Pressure Storage Vessels Spheres And Spheroids	Pressure Storage Vessels Drums And Bullets	Refrigerated Dome Roof Storage Tanks
Floating & Cone Roof Tanks < 3000 Barrels	0.5 D *								
Floating & Cone Roof Tanks > 3000 < 10.000 Barrels	0.5D	0.5D							
Floating & Cone Roof Tanks > 10.000 < 3000 Barrels	1 X D	1 X D	1 X D						
Jumbo Floating Roof Tanks > 300.000 Barrels	1 X D	1 X D	1 X D	1 X D					
Cone Roof Tanks, Class II, III Product > 10.000 < 300.000 Barrels	0.5 D	0.5 D	1 X D	1 X D	0.5 D				
Cone Roof Tanks, Inerted Class I Product** > 10.000 < 150.000 Barrels	1 X D	1 X D	1 X D	1 X D	1 X D	1 X D			
Pressure Storage Vessels Spheres And Spheroids	1.5D 100' MIN	1.5D 100' MIN	1.5D 100' MIN	2 X D	1.5D 100' MIN	1.5D 100' MIN	1 X D 50' MIN		
Pressure Storage Vessels Drums And Bullets	1.5D 100' MIN	1.5D 100' MIN	1.5D 100' MIN	2 X D	1.5D 100' MIN	1.5D 100' MIN	1 X D 100' MIN	1 X D	
Refrigerated Dome Roof Storage Tanks	2 X D 200'MIN	2 X D 200'MIN	2 X D 200'MIN	2 X D	2 X D 200'MIN	2 X D 200'MIN	1 X D 100'MIN	1 X D 100'MIN	1 X D 100'MIN

d = Largest Tank Diameter
1 Barrel = 42 Gallons = 159 L
$^{\circ}C = (^{\circ}F - 32) \times 0.555$
1 ft. = 0.305 m

* For Class II, III Products, 5 ft. Spacing Is Acceptable.
** or Class II or III Operating at Temperatures > 200 °F
(REFERENCE: G.A.P. 2.5.2)

Basic Concepts for Risk Management in design
HAZOP Study Procedure
NPC-HSE-S-04-A

REVISION INDEX SHEET

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CONTENTS

1. Introduction

This document constitutes the procedure for the HAZOP study and has been prepared to ensure a common understanding existing prior to commencement.

2. Abbreviation

C&E	Cause & Effect
LOPA	Layer Of Protection Analysis
ESD	Emergency Shutdown
F&G	Fire and Gas
HAZOP	Hazard and Operability
P&ID	Piping and Instrumentation Diagram
PCS	Process Control System
PSD	Process Shutdown
FAL/LL	Flow alarm low/low low
FAH/HH	Flow alarm high/high high
PAL/LL	Pressure alarm low/low low
PAH/HH	Pressure alarm high/high high
TAL/LL	Temperature alarm low/low low
TAH/HH	Temperature alarm high/high high
LAL/LL	Level alarm low/low low
LAH/HH	Level alarm high/high high

3. Background.

A HAZOP study identifies hazards and operability problems. The concept involves investigating how the plant might deviate from the design intent. If, in the process of identifying problems during a HAZOP study, a solution becomes apparent, it is recorded as part of the HAZOP result; however, care must be taken to avoid trying to find solutions which are not so apparent, Although the HAZOP study was developed to supplement experience-based practices when a new design or technology is involved, its use has expanded to almost all phases of a plant's life. HAZOP is based on the principle that several experts with different backgrounds can interact and identify more problems when working together than when working separately and combining their results. The "Guide-Word" HAZOP is the most well known of the HAZOPs; however, several specializations of this basic method have been developed. These specializations will be discussed as modifications of the Guide-Word approach, but they are not to be regarded as less useful than the Guide-Word approach. Indeed, in many situations these variations may be more effective than the Guide-Word approach.

4. Concept

The HAZOP concept is to review the plant in a series of meetings, during which a multi-disciplinary team methodically "brainstorms" the plant design, following the structure provided by the guide words and the team leader's experience.

The primary advantage of this brainstorming is that it stimulates creativity and generates ideas. This creativity results from the interaction of the team and their diverse backgrounds. Consequently the process requires that all team members participate (quantity breeds quality in this case), and team members must refrain from criticizing each other to the point that members hesitate to suggest Ideas.

The team focuses on specific points of the design (called "study nodes"), one at a time. At each of these study nodes, deviations in the process parameters are examined using the guide words. The guide words are used to ensure that the design is explored in every conceivable way. Thus the team must identify a fairly large number of deviations, each of which must then be considered so that their potential causes and consequences can be identified.

The best time to conduct a HAZOP is when the design is fairly firm. At this point, the design is well enough defined to allow meaningful answers to the questions raised in HAZOP process. Also, at this point it is still possible to change the design without a major cost. However, HAZOPs can be done at any stage after the design is nearly firm. For example, many older plants are upgrading their control and Instrumentation systems. There is a natural relationship between the HAZOP deviation approach and the usual control system design philosophy of driving deviations to zero; thus It Is very effective to examine a plant as soon as the control system redesign is firm.

The success or failure of the HAZOP depends on several factors:

- The completeness and accuracy of drawings and other data used as a basis for the study
- The technical skills and insights of the team
- The ability of the team to use the approach as an aid to their Imagination in visualizing deviations, causes, and consequences
- The ability of the team to concentrate on the more serious hazards which are identified.

The process is systematic and it is helpful to define the terms that are used:

STUDY NODES - The locations (on piping and Instrumentation drawings and procedures) at which the process parameters are investigated for deviations.

INTENTION - The intention defines how the plant is expected to operate in the absence of deviations at the study nodes. This can take a number of forms and can either be descriptive or diagrammatic; e.g., flow sheets, line diagrams, P&IDs.

DEVIATIONS - These are departures from the intention which are discovered by systematically applying the guide words (e.g., "more pressure").

CAUSES - These are the reasons why deviations might occur. Once a deviation has been shown to have a credible cause, it can be treated as a meaningful deviation. These causes can be hardware failures, human errors, an unanticipated process state (e.g., change of composition), external disruptions (e.g., loss of power), etc.

CONSEQUENCES - These are the results of the deviations should they occur (e.g., release of toxic materials). Trivial consequences, relative to the study objective, are dropped.

GUIDE WORDS - These are simple words which are used to qualify or quantify the intention in order to guide and stimulate the brainstorming process and so discover deviations. The guide words shown in Table 1 are the ones most often used in a HAZOP; some organizations have made this list specific to their operations, to guide the team more quickly to the areas where they have previously found problems. Each guide word is applied to the process variables at the point in the plant (study node) which is being examined. For example:

Guide Words	Parameter	Deviation
NO	FLOW	NO FLOW
MORE	PRESSURE	HIGH PRESSURE
AS WELL AS	ONE PHASE	TWO PHASE
OTHER THAN	OPERATION	MAINTENANCE

These guide words are applicable to both the more general parameters (e.g., react, transfer) and the more specific parameters (e.g., pressure, temperature).

TABLE 1: HAZOP GUIDE WORDS AND MEANINGS

Guide Words	Meaning
No	Negation of the Design Intent
Less	Quantitative Decrease
More	Quantitative Increase
Part Of	Qualitative Decrease
As Well As	Qualitative Increase
Reverse	Logical Opposite of the Intent
Other Than	Complete Substitution

With the general parameters, meaningful deviations are usually generated for each guide word. Moreover, it is not unusual to have more than one deviation from the application of one guide word. For example, "more reaction" could mean either than a reaction takes place at a faster rate, or that a greater quantity of product results. With the specific parameters, some modification of the guide words may be necessary. In addition, it is not unusual to find that some potential deviations are eliminated by physical limitation. For example, if the design intention of a pressure or temperature is being considered, the guide words "more" or "less" may be the only possibilities.

There are other useful modifications to guide words such as:
SOONER or LATER for OTHER THAN when considering time WHERE ELSE for OTHER THAN when considering position, sources, or destination HIGHER and LOWER for MORE and LESS when considering elevations, temperatures, or pressures.

Finally, when dealing with a design Intention involving a complex set of interrelated plant parameters (e.g., temperatures, reaction rates, composition, or pressure), it may be better to apply the whole sequence of guide words to each parameter individually than to apply each guide word across all of the parameters as a group. Also, when applying the guide words to a sentence it may be more useful to apply the sequence of guide words to each word or phrase separately, starting with the key part which describes the activity (usually the verbs or adverbs). These parts of the sentence usually are related to some impact on the process parameters. For example, in the sentence "The operator starts flow A when pressure B is reached", the guide words would be applied to:
Flow A (no, more, less, etc.)
When pressure B is reached (sooner, later, etc.)

5. Guidelines for Using Procedure

The concepts presented above are put into practice in the following steps:
1.Define the purpose, objectives, and scope of the study
2.Select the team
3.Prepare for the study
4.Carry out the team review
5.Record the results.

some of these steps can take place at the same time. For example, the team reviews the design, records the findings, and follows up on the findings continuously. Nonetheless, each step will be discussed below as separate items.

5.1. Define the Purpose, Objectives, and Scope of the Study.

The purpose, objectives, and scope of the study should be made as explicit as possible. These objectives are normally set by the person responsible for the plant or project, assisted by the HAZOP study leader (perhaps the plant safety specialist). It is important that this interaction take place to provide the proper authority to the study and to ensure that the study is focused. Also, even though the general objective is to identify hazards and operability problems, the team should focus on the underlying purpose or reason for the study. Examples of reasons for a study might be to:

- *Check the safety of a design*
- *Decide whether and where to build*
- *Develop a list of questions to ask a supplier*
- *Check operating/safety procedures*
- *Improve the safety of an existing facility*
- *Verify that safety instrumentation is reacting to best parameters.*

It is also important to define what specific consequences are to be considered:

- *Employee safety (in plant or neighboring research center)*
- *Loss of plant or equipment*
- *Loss of production (lose competitive edge in market)*
- *Liability*
- *Insurability*
- *Public safety*
- *Environmental impacts.*

For example, a HAZOP might be conducted to determine where to build a plant to have the minimal impact on public safety. In this case, the HAZOP should focus on deviations which result in off-site hazards.

5.2. Select the Team.

Ideally, the team consists of five to seven members, although a smaller team could be sufficient for a smaller plant. If the team is too large, the group approach fails. On the other hand, if the group is too small, it may lack the breadth of knowledge needed to assure completeness. The team leader should have experience in leading a HAZOP.
The rest of the team should be experts in areas relevant to the plant operation. For example, a team might include:

- *Design engineer*
- *Process engineer*
- *Operations supervisor*
- *Instrument design engineer*

- *Maintenance supervisor*
- *Safety engineer.*

The team leader's most important job is to keep the team focused on the key task: to identify problems, not necessarily to solve them. There is a strong tendency for engineers to launch into a design or problem-solving mode as soon as a new problem comes to light. Unless obvious solutions are apparent, this mode should be avoided or it will detract from the primary purpose of HAZOP, which is hazard identification.

In addition, the team leader must keep several factors in mind to assure successful meetings: (1) do not compete with the members; (2) take care to listen to all of the members; (3) during meetings, do not permit anyone to be put on the defensive; (4) to keep the energy level high, take breaks as needed.

5.3. Prepare for the Study

The amount of preparation depends upon the size and complexity of the plant. The preparative work consists of three stages: obtaining the necessary data; converting the data to a suitable form and planning the study sequence; and arranging the meetings.

a. Obtain the necessary data:

Typically, the data consist of various drawings in the form of line diagrams, flow sheets, plant layouts, isometrics, and fabrication drawings. Additionally, there can be operating instructions, instrument sequence control charts, logic diagrams, and computer programs. Occasionally, there are plant manuals and equipment manufacturer's manuals. The data must be inspected to make sure they pertain to the defined area of study and contain no discrepancies or ambiguities.

b. Convert the data into a suitable form and plan the study sequence:

The amount of work required in this stage depends on the type of plant. With continuous plants, the preparative work is minimal. The existing, up-to-date flow sheets or pipe and instrument drawings usually contain enough information for the study, and the only preparation necessary is to make sure that enough copies of each drawing are available.

Likewise, the sequence for the study is straight forward. The study team starts at the beginning of the process and progressively works downstream, applying the guide words at specific study nodes. These nodes are established by the team leader prior to any meetings. The team leader will generally define the study nodes in pipe sections. These nodes are points where the process parameters (pressure, temperature, flow, etc.) have identified design intent. Between these nodes are found the plant components (pumps, vessels, heat exchangers, etc.) that cause changes in the parameters between nodes. While the study nodes should be identified before the meetings, it is to be expected that some changes will be made as the study progresses due to the learning process that accompanies the study.

With batch plants, the preparative work is usually more extensive, primarily because of the more extensive need for manual operations; thus, operation sequences are a larger part of HAZOP. This operations information can be obtained from operating instructions, logic diagrams, or instrument sequence diagrams. In some circumstances (e.g., when two or more batches of material are being processed at the same time), it may be necessary to prepare a display indicating the status of each vessel on a time basis. If operators are physically involved in the process (e.g., in charging vessels) rather than simply controlling the process, their activities should be represented by means of process flow charts.

The team leader will usually prepare a plan for the sequence of study before the study starts to make sure that the study team approaches the plant and its operation methodically. This means the team leader must spend some time before the meetings to determine the best study sequence, based on how the specific plant is operated.

The team leader will often have to prepare a representation of the equipment (logic diagram, flow chart, etc.) tailored to suit the application of the HAZOP technique to the equipment. This may include a display of the relationship of the equipment with operators and with other plant equipment. The preparative work will often involve a lengthy dialogue between the project engineer and the team leader and sometimes involves the component manufacturers as well. The team leader will prepare a plan for the study and discuss the equipment representations and the plan with the team before starting the study.

c. Arrange the necessary meetings:

Once the data have been assembled and the equipment representations made (if necessary), the team leader is in a position to plan meetings. The first requirement is to estimate the team-hours needed for the study. As a general rule, each individual part to be studied, e.g., each main pipeline into a vessel, will take an average of fifteen minutes of team time. For example, a vessel with two inlets, two exits, and a vent should take one and a half hours for those elements and the vessel itself. Thus, an estimate can be made by considering the number of pipelines and vessels. Another way to make a rough estimate is to allow about three hours for each major piece of equipment. Fifteen minutes should also be allowed for each simple verbal statement such as “switch on pump”, “motor starts”, or “pump starts”.

After estimating the team-hours required, the team leader can arrange meetings. Ideally, each session should last no more than three hours (preferably in the morning). Longer Sessions are undesirable because their effectiveness usually begins to fall off. Under extreme time-pressures, sessions have been held for two consecutive days; but such a program should be attempted only in very exceptional circumstances, (for example, when the team is from out of town and travel every day is not acceptable).

With large projects, it has been found that often one team cannot carry out all the studies within the allotted time. It may therefore be necessary to use several teams and team leaders.

One of the team leaders should act as a coordinator to allocate sections of the design to different teams and to prepare time schedules for the study as a whole.

5.4. Carry Out the Team Review.

The HAZOP study requires that the plant schematic be divided into study nodes and that the process at these points is addressed with the guide words. As shown in Figure 1, the method applies all of the guide words in turn and either of two outcomes is recorded: (1) more information is needed, or (2) the deviation with its causes and consequences. If there are obvious remedies, these too are recorded.

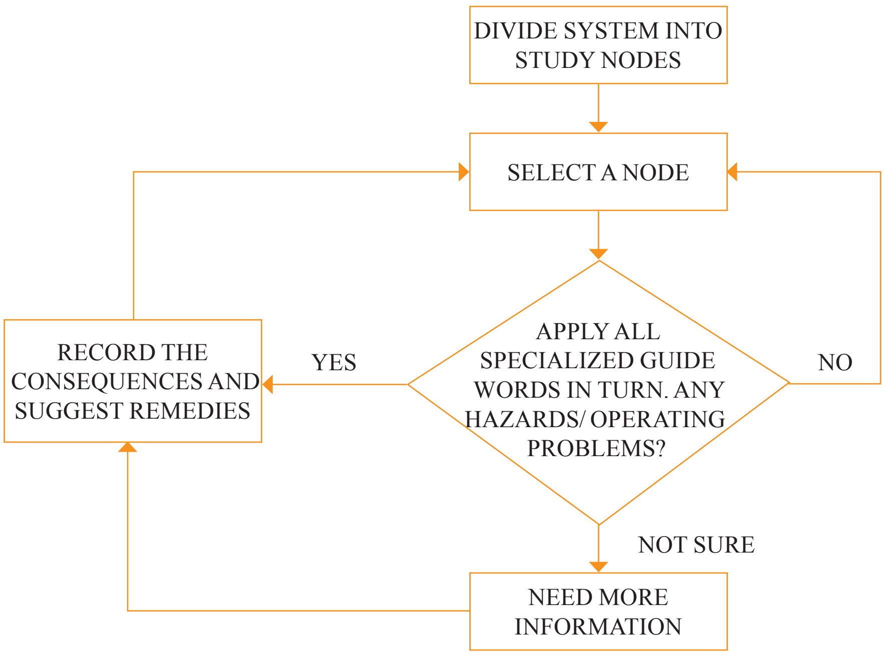


Figure 1: HAZOP Method Flow Diagram

As hazards are detected, the team leader should make sure that everyone understands them. As mentioned earlier, the degree of problem-solving during the examination sessions can vary. There are two extreme positions:

- *A suggested action is found for each hazard as it is detected before looking for the next* *hazard*

- *No search for suggested actions is started until all hazards have been detected.*

In practice, there is a compromise. It may not be appropriate or even possible for a team to find a solution during a meeting. On the other hand, if the solution is straight forward, a decision can be made and the design and operating instructions modified immediately.

To some extent, the ability to make immediate decisions depends upon the type of plant being studied. With a continuous plant, a decision made at one point in a design may not invalidate previous decisions concerning upstream parts of the plant which have already been studied but this possibility always has to be considered. For batch plants with sequence control, any alteration in the design or mode of operation could have extensive implications. If a question is noted for future evaluation, a note is also made of the responsible for follow-up.

Although the team leader will have prepared for the study, the HAZOP technique may expose gaps in the available plant operating information or in the knowledge of the team members. Thus it may sometimes be necessary to call in a specialist on some aspects of how the plant is intended to operate or even to postpone certain parts of the study in order to obtain more information.

Once a section of pipeline or a vessel or an operating instruction has been fully examined, the team leader should mark (e.g., “yellow out”) his or her copy to that effect. This action ensures comprehensive coverage. Another way of doing this is that once every part of a drawing has been examined, the study leader certifies that the examination has been completed in an appropriate box on the flow sheet.

5.5. Record the Results.

The recording process is an important part of the HAZOP. It is impossible to record manually all that is said, yet it is very important that all ideas are kept. It is very useful to have the team members review the final report and then come together for a report review meeting. The process of reviewing key findings will often fine-tune these findings and uncover others. The success of this process demands a good recording scheme.

First, a HAZOP form should be filled out during the meeting. This form is best filled out by an engineer who can be less senior than the team members. This recorder is not necessarily part of the team but, as an engineer, can understand the discussions and record the findings accurately. Other means of recording can be developed as best suits the organization. Some have found that when insufficient information is available to make a decision, cards are filled out so that the responsible individual is reminded of the action item. It has also been found useful to tape-record the sessions and have them transcribed. This saves the only complete record of the discussions and the reasoning behind the recorded findings, and it can be invaluable later in the Plant life when the plant is modified, or if an event occurs which the result of a deviation is.

6 . Team composition

There will be teams performing the HAZOPs and each team will be provided with the support as detailed below:

- *HAZOP Manager*
- *Chairmen and scribes*

Each HAZOP team shall comprise the following functions/individuals as a minimum:

- *HAZOP Chairperson*
- *HAZOP Secretary*
- *Operations Engineer, (OWNER)*
- *Process Engineer (OWNER)*
- *Instrument Engineer and safety Engineer*

The requirement for any additional attendance will be established at the start of the HAZOP Kick-off and Introduction session, this will include any requirement for part-time representatives from other disciplines who will be called into the meeting as required.

7 . Process Description

Process description is "FOR INFORMATION ONLY" and should not be used for design purposes.

8 . Facilities

The following facilities shall be provided:

- *Sufficient room and table top lay down area for team members to deploy drawings and Documents,*
- *A screen for projected action sheets.*
- *Wall area to pin up A1or A2 master copies of drawings.*
- *PC projector*

9 . Reference Documents

Plot Plans of the Installation and P&ID's shall be used for the study.
The team will require the following documentation to be available, including but not limited to:

- *Process P&ID's Diagrams*

- *Process Flow Diagram(PFD)*
- *Utility P&ID's Diagrams*
- *Plot Plans*
- *Process Description*
- *Control philosophy*

10. HAZOP Study Guidewords

The list of guide-words to be applied to all process elements during the study shall be selected by the HAZOP Leader.

11. Reproting

The minutes will be recorded using HAZOP software. The minutes will be displayed as written for each node on a screen using an overhead projector. The benefit is that all the team can see the minutes as they are being generated, which helps with the derivation of the action items and prevents changes before formal team approval.

If there are no findings associated with a particular "guideword" this will be identified on the particular work sheet.The findings of the HAZOP study will be recorded using the PHA PRO software. Hardware requirements have been identified as follows:

- *Laptop PC.*
- *Projector unit to enable HAZOP team to view record sheets.*

12. Issue of Draft/final HAZOP worksheets & study report

HAZOP worksheets will be reviewed on screen by team to ensure that they are complete and readily understood. Agreed comments will be incorporated within the provisional worksheets, which will be issued with the HAZOP Report for action and close out by the Project.

Following the final review session, the Hazop Engineer shall issue a full set of HAZOP Worksheets to the Project within specified working days not more than one month.

The HAZOP Manager and Chairman shall finalise the HAZOP actions into the formal HAZOP Report, this report shall be issued within 15 days of the last HAZOP being completed. The HAZOP Manager shall issue the report as hard copy and soft copy on a CD. Recommendations/findings originating from the HAZOP will be tracked and managed by the Project via the project action tracking system.

13. HAZOP Report Format

The HAZOP Report shall conform to the following structure:

- *Contents List*
- *Introduction*

This section shall also cover the study methodology, identify any difficulties in meeting, and include recommendations for further studies to be addressed.

- *HAZOP Worksheets*

This section shall include if used:

- *List of Flow Diagrams studied (including revision numbers)*
- *List of nodes references*
- *All other documentation referenced during the study*

- *Principal Recommendations*
- *Master P&ID's:*

Colour copies of all the HAZOP master P&ID's, plus any other drawings reviewed during the study will be included in the report with the nodes studied clearly indicated.

APPENDIX A – PARAMETERS AND GUIDEWORDS

The following parameters and guidewords shall be used during the HAZOP to assist the team to identify hazards and operability problems associated with the design.

PARAMETER	GUIDEWORDS
Flow	No, Less, More, Reverse
Pressure	High, Low
Temperature	High, Low
Level	High, Low
Composition	Change
Corrosion	More
Deposition	More
Erosion	More
Services	Failure
Start-up and Shutdown	Problems and Requirements
Maintenance and Inspection	Problems and Requirements
Other	

APPENDIX B – RISK MATRIX

	LIKELIHOOD					
SEVERITY		A	B	C	D	E
	1	1	1	1	1	2
	2	1	1	1	2	3
	3	1	1	2	2	3
	4	1	2	2	3	4
	5	3	4	4	4	4

Risk Ranking	Description	Required Mitigation
1	Critical	Mitigate the hazard by implementing engineering controls and if necessary administrative controls to reduce the risk ranking to an acceptable level
2	Undesirable	Mitigate the hazard with administrative and / or engineering controls to reduce the risk ranking to an acceptable level
3	Acceptable with controls	Verify hazard has administrative controls in place
4	Acceptable	Identified hazard is acceptable and does not require further mitigation

Severity Table

Severity	Description
1	Multiple fatalities Process unit Damage EI:Major / Extended Duration / Full Scale Response / Long Term Effect
2	Potential on-site Death Multiple Major Equipment Damage EI:Serious/ Significant Resource Commitment / Long Term Effect
3	Potential On-site Severe Injury or Severe Health Effects Single Major or Multiple Minor Equipment Damage EI:Serious/ Significant Resource Commitment
4	Potential On-site Minor Injury or Severe Health Effects Single Minor Equipment Damage EI: Moderate/Limited response of short Duration
5	No Injury or Health Effect No Equipment damage EI: Minor/Limited or no response needed

Likelihood Table

Likelihood	Description
A	Could Occur more often than Once Per Month
B	Could Occur more often than once per year
C	Could occur several times in plant life
D	Could occur once in plant life
E	Not expected in plant life

APPENDIX C –Typical HAZOP WORKSHEET

Project Title:	
Node Number:	Node Description:
Drawing Numbers:	
Node Design Intention:	
Parameter:	

Deviation	Causes	Consequence	Safeguards	Risk Ranking			Recommendation	Action by:
				S	L	R		

APPENDIX D – Failure Rate Data

Components	Failure Mode	CCPS *1)		OREDA *2)		likelihood
		Failure Data		Failure Data		
		fail rate (1/10 ^ 6hrs)	MTBF (years)	fail rate (1/10 ^ 6hrs)	MTBF (years)	
Compressor(Turbine)	Trip	-	-	118	1.0	B
Pump(Moter - Centri)	Trip	104	1.1	52.64	2.2	B
Fan(Include Controller / Motor)	Trip	9.09	12.6	-	-	C
Sensor/ Transmitter	Abnormal Output	-	-	19.76	5.8	C
	Fail to Operate	-	-	1.56	73.2	E
	Spurious Operation	-	-	0.78	146.4	E
Sensor / Switch	Function Without Signal	1.16	98.4	-	-	E
	Fail to Function when Signaled	4.2	27.2	-	-	D
CV	Fail to Close/ Open (Actuator Failure)	-	-	4.66	24.5	D
	Spurious Operation	3.59	31.8	2.87	39.8	D
MOV	Spurious Operation	1.36	83.9	-	-	E
SOV	Spurious Operation	0.409	279.1	-	-	E
EMV	Fail to Close/ Open (Actuator Failure)	-	-	3.0	38.1	D
	Spurious Operation	-	-	0.26	439.1	E
Check Valve	Catastrophic	3.18	35.9	-	-	D
Shell / Tube H / EX	Internal Leakage	-	-	3.53	32.3	D

* 1) CCPS : Guidelines For Process Equipment Reliability Data. Center For Chemical Process Safety(CCPS), American Institue Of Chemical Engineer(AICHE),1989
*2) OREDA : Offshore Reliability Data 3Rd Edition. Prepared By SINTEF Industrial Management , 1997.

Basic Concepts for Risk Management in design
SIL Study Procedure
NPC-HSE-S-04-B

REVISION INDEX SHEET

PAGE \ REV.	1	2	3	4	5	6	PAGE \ REV.	1	2	3	4	5	6	PAGE \ REV.	1	2	3	4	5	6

CONTENTS

1. Introduction

The purpose of this document is to outline the procedure to determine the required safety integrity levels (SILs) for the project safety instrumented systems.

2.Background

The application of safety instrumented systems within the Process Industries requires that a process hazard and risk assessment be carried out to enable the specification for safety instrumented systems to be derived. Other safety systems are only considered so that their contribution can be taken into account when considering the performance requirements for the safety instrumented systems. The safety instrumented system includes all components and subsystems necessary to carry out the safety instrumented function (SIF) from sensor(s) to final element(s).

The International Standard IEC 61511 is process industry specific within the framework of IEC 61508. It sets out an approach for safety life-cycle activities to achieve minimum standards. In most situations, safety is best achieved by an inherently safe process design but, if necessary, this may be combined with a protective system or systems to address any residual identified risk.

3.SIL Definition

Safety Integrity: the likelihood of a safety related system satisfactorily performing the required safety functions under all the stated conditions within a stated period of time (IEC 61508 definition)
Safety integrity level: a discrete level for specifying the safety integrity requirements of safety functions (IEC 61508 definition)

4.Abbreviations

C&E	Cause & Effect
ESD	Emergency Shutdown
F&G	Fire and Gas
HAZOP	Hazard and Operability
LOPA	Layers of Protection Analysis
P&ID	Piping and Instrumentation Diagram
PCS	Process Control System
PSD	Process Shutdown
QRA	Quantitative Risk Analysis
SIF	Safety Instrumented Function
SIL	Safety Integrity Level

5. SIL Target

The required failure probability of the Safety Instrumented Function is determined by:

- Acceptable hazard frequency
- Hazard frequency assessed without the safety instrumented system

Therefore in the SIL process the hazards are first evaluated as if the safety instrumented system under discussion does not exist, although credit may be claimed for other mitigating systems. To obtain an acceptable hazard frequency, improvements can be made to any of the systems that contribute to the hazard frequency not only the safety instrumented system under discussion. The choice of where to make the improvements is one of safety optimisation.

The safety integrity level (SIL) is a measure of the failure probability or frequency of a safety related instrumented function. IEC-61508 has defined the SIL values in reliability terms as:

Table 1: SIL Rating

Safety Integrity Level (SIL)	Failure Probability (for low demands)	Failure Frequency (for high demands)
4	$1x10^{-5} – 1x10^{-4}$ per demand	$1x10^{-5} – 1x10^{-4}$ per year
3	$1x10^{-4} – 1x10^{-3}$ per demand	$1x10^{-4} – 1x10^{-3}$ per year
2	$1x10^{-3} – 1x10^{-2}$ per demand	$1x10^{-3} – 1x10^{-2}$ per year
1	$1x10^{-2} – 1x10^{-1}$ per demand	$1x10^{-2} – 1x10^{-1}$ per year

The SIL target is taken as the highest value arising from the three categories; safety, environment and economic.

The results of the SIL study are used as follows.

The determined SILs are generally taken as design requirements.

- In cases where the assessed SIL levels have Intolerable design implications, more detailed consideration of the risks will be carried out to determine if the SIL may be reduced. This will use Layers of Protection Analysis (LOPA) or Quantitative Risk Assessment (QRA)
- For loops that have been identified as requiring SIL 2/3/4 rating, the system design (i.e. the equipment under control) will be re-examined to assess whether it can be changed to avoid or reduce reliance on an instrumented safety system (e.g. made more inherently safe). Where changes are possible, the SIL will be reassessed to ensure that the equipment under control and the instrumented safety system are engineered in a way to meet the required level of performance and comply with the requirements of IEC 61508.

6. Scope

The scope of the SIL assessment will be all instrumented control loops with a potential protective function. These loops will be identified by the HAZOP or by the system designers. Therefore the loops subject to analysis will be all those included in the Terminal Shut down Systems. It will not include the loops assigned to the Process Control System (PCS) as these do not have a protective function.

The scope of SIL analysis includes assessment of asset-related risks (such as equipment damage and loss of production) and environment risks (such as pollution), as well as safety risks.

7. Methodology

It is important that the distinction between risk and safety integrity is fully appreciated. Risk is a measure of the frequency and consequence of a specified hazardous event occurring. This can be evaluated for different situations (process risk, tolerable risk, residual risk-see Figure 2). The tolerable risk involves consideration of societal and political factors. Safety integrity is a measure of the likelihood that the SIF and other protection layers will achieve the specified safety functions. Once the tolerable risk has been set, and the necessary risk reduction estimated, the safety integrity requirements for the SIS can be allocated. The role that safety functions play in achieving the necessary risk reduction is illustrated in Figure 1 and 2.

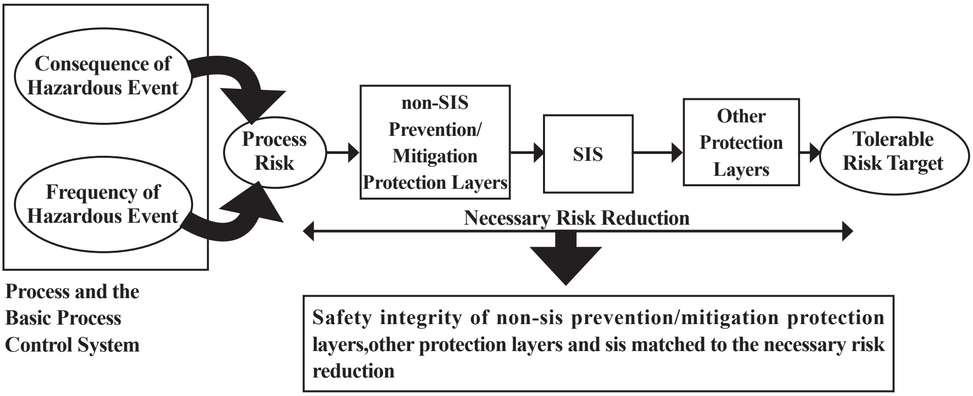


Figure 1: Risk and safety Integrity concept

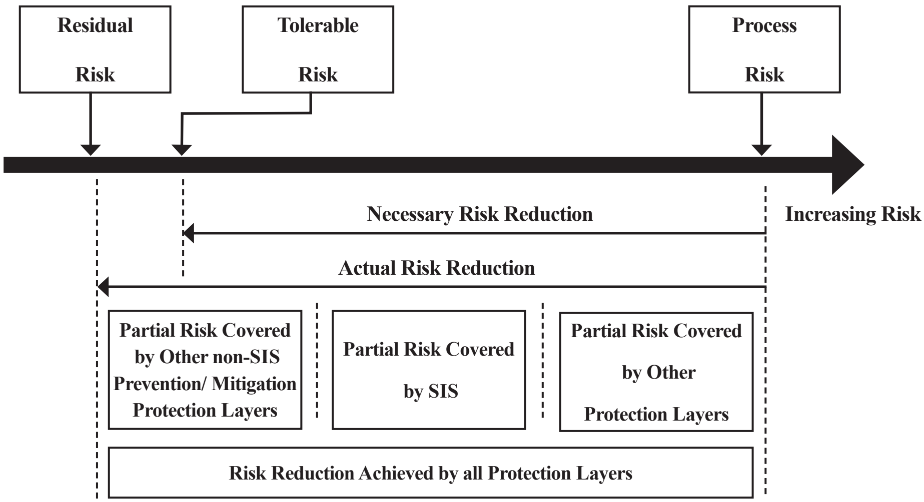


Figure 2: Risk reduction general concept

Safety function is implemented by an SIS, other technology safety related system or external risk reduction facilities, which is intended to achieve or maintain a safe state for the process, with respect to a specific hazardous event. The safety functions in process industries are more often delegated to electrical, electronic or programmable electronic (E/E/PE) Safety Instrumented Systems (SIS).

The functional safety standards IEC 61508 and IEC 61511 propose guidelines which can be used in order to define the requirements for achieving a specified Safety Integrity Level (SIL) and in order to evaluate the actual availability of a SIS.

There are several methods that can be used for SIL determination for a specific safety instrumented function. IEC 61511-3 presents information on a number of methods that have been used. The following are basic and generic steps to determine a safety function SIL rating based on IEC 61511:

- *Perform a hazard and risk analysis to evaluate existing risk*
- *Identify safety function(s) needed*
- *Allocate safety function(s) to independent protection layers*
- *Determine if a SIF is required*
- *Determine required SIL of the SIF.*

The methods presented in this document are based on IEC 61511 and utilize a Workshop approach:

- *Risk Graph*
- *Layer of Protection Analysis (LOPA)*
- *SIL Matrix*

In some applications more than one method may be used. A qualitative method may be used as a first pass to determine the required SIL of all the SIFs. Those which are assigned a high SIL by this method should then be considered in greater detail using a quantitative method to give a more rigorous understanding of their required safety integrity.

In order to determine the required SIL of the safety instrumented functions (SIFs), it is necessary to define tolerable risk target in terms of probability and consequence of the process potential incidents. This would take place by discussion and agreement between the interested parties before the workshop (for example safety regulatory authorities, those producing the risks and those exposed to the risks).

7.1. Review team

The review team will include representatives from the Safety, Process and Instrument disciplines from both the design team and the client. Further individuals will attend as applicable to the area under analysis. Members of the multi-disciplinary SIL determination team shall generally be as follows.

- *Chair (familiar with the methodology and from outside the project)*
- *Safety Engineer*
- *Process Engineer*
- *Instrument Engineer*
- *Operations Representative*
- *Other as necessary (e.g. package engineer, vendor engineers).*

There must be people on the team who can explain the purpose of each loop and either know or have documents which give the input and output actions. This is necessary to enable the descriptive entries on the worksheets to be made without holding up the study. It is preferred that the SIL review runs concurrently with the HAZOP, and team members who were also involved in the HAZOP.

7.2. Study documentation

The study shall work through all the relevant instrumented loops on the installation with a protective function and so documentation is needed to identify these loops. Typically the documentation will be either P&IDs or cause and effect diagrams. This will include information on vendor packages when this becomes available. In addition to the above, reference will be made to relevant Layouts, HAZOPS, QRA and other safety studies, as required.

7.3. Loop identification

The loops identified for analysis are protective functions identified in the HAZOP and the ESD/PSD Cause & Effects diagrams. For the purposes of SIL determination a loop shall be defined as an individual input device and all its associated outputs. As loop tag numbers are based on the input devices this makes it very straight forward to identify all the loops. Using this method the SIL is determined for the most significant output action. This is usually straight forward to identify. However, in some cases the consequences of two different output actions may be quite different and this may justify determining the SIL of the two outputs separately.

7.4. Reporting

Findings shall be recorded during the analysis on a computer, with use of an overhead projection during the Study. Where action is required this shall be identified and allocated to a specific discipline.

8. Study Procedure

8.1. Risk Graph Method

8.1.1. Procedure

1. Identify the loop to be examined, and record the tag and P&ID number.
2. Determine the function of the loop (i.e. what is it for?).
3. Determine the cause of demand on the loop (most commonly, control failure), drawing on the HAZOP knowledge where available.
4. Identify the output actions (e.g. close specified valves) referring to the HAZOP and/or C&E diagrams.
5. Determine the consequence if the loop fails on demand, generally drawing on the HAZOP. At this point no credit is taken for any other relevant risk reduction measures.
6. Having gathered the above information, use judgement to determine the four parameters C, F, P and W on the safety risk graph (see Appendix 1). The demand rate W is the sum of the frequency of the various causes of demand identified in Step3.
7. Apply the safety risk graph to determine the SIL required on safety risk considerations.
8. Determine the environmental loss consequence and use the environmental risk graph (see Appendix 2) to determine the SIL required on environmental risk considerations.
9. Determine the economic loss consequence, in terms of asset loss and terminal unavailability, and use the economic risk graph (see Appendix 2) to determine the SIL required on economic risk considerations.
10. Determine the SIL required for the function identified in Step 2 as the highest of the three SILs determined in Steps 7, 8 & 9.
11. Identify any other protective measures, e.g. relief valves, which have the same function as the loop under consideration.
12. Determine the SIL required for the loop under consideration.
13. Record any recommendations for further work. Following the review meeting these should be transferred onto tracking action sheets to ensure they are properly addressed.

If the measure identified in Step 11 is completely independent of the loop under consideration, the loop SIL is the function SIL minus 1. If the measure identified in Step 11 is a relief valve, the loop SIL is the function SIL minus 2. If there are no completely independent measures, the loop SIL is the same as the function SIL.

8.1.2. Discussion of Key Issues

Steps 2 & 4.

It is important to distinguish between what the loop is for - the function - and what it does - the output actions. This is because the main SIL determination Steps 6 - 10 are applied to the function. The study determines how important it is that the function is achieved. (The loop SIL may or may not be the same as the function SIL depending on whether other independent safety devices share the same function: see Steps 11 & 12.

Step 6.

Care is needed when taking credit for other safety systems or risk reduction factors which are in addition to the instrument loop. This is done in two places: the avoid consequence parameter P in this Step 6, and in Step 11. Since Step 11 is reserved for measures with the same function as the loop under consideration, parameter P is used to take credit for any other factor which reduces the likelihood that the consequence C will actually occur. For example, in a gas release, low ignition probability and the likelihood that people will escape are relevant factors that justify a rating Pa rather than Pb. It can be seen that the method is extremely coarse here: the probability of consequence C occurring may be well below 1 %, but credit can only be taken for a 10 % risk reduction.

Steps 10-12.

A key feature of the assessment process is that a SIL is first determined for the safety function, e.g. the function of preventing a vessel being over-pressured. If this is high, then it is generally preferable to achieve this by having independent layers of protection rather than one high integrity device. Provided these are independent, the individual unavailability may be multiplied together to give the unavailability for the combined systems. This is the basis for Step 12. A relief valve is considered to have an availability equivalent to a SIL 2 device: other devices are taken to be SIL 1.

8.1.3. Input Data for SIL Assessment

The assessment requires the following factors to be estimated to the nearest order of magnitude:

1. *The consequences such as the expected injuries or costs.*
2. *The demand rate for the hazard*
3. *The failure probabilities for all mitigating systems other than the system for which the SIL is being assessed.*
4. *The occupancy factor or proportion of time at risk.*

8.1.3.1. Consequences

- **Safety consequences (C)**

The relevant consequence is the consequence which can be avoided by the use of the system rather than the absolute consequences. This point can best be illustrated by taking for example a fire detection system which initiates ESD valves and a blow-down system. It could be postulated that a fire may kill some operators in the immediate vicinity but the operation of the fire detection system would have no effect on immediate fatalities. In this example the relevant consequences would be any additional deaths which might occur should the fire escalate due to failure of the fire detection system.

The consequences (number of fatalities) is determined by assessing the numbers of people likely to be present when the area exposed to the hazard is occupied and multiplying this by their vulnerability to the identified hazard. The vulnerability is determined by the nature of the hazard being protected against. The following vulnerabilities have been used:

Small release of flammable or toxic material	0.01
Large release of flammable or toxic material	0.1
Large release of flammable material with a high probability of catching fire or highly toxic material	0.5
Rupture or explosion	1.0

The consequence parameter, C, selection is based on the following criteria:

Minor injury	Ca
Serious injury	Cb
Multiple serious injuries and / or a single fatality	Cc
More than one fatality on site and / or serious injury offsite	Cd

Overpressure of hydrocarbon containing systems in excess of 1.3 x design pressure will generally be considered to have the potential for more than 1 fatality (Cd). Overpressure of non hydrocarbon containing systems, releases of toxic gas (other than through a significant overpressure event), overpressure less than test pressure, and releases arising from high temperature will generally be considered to have the potential for serious injury to more than one person or possibly a single fatality (Cc). The consequence field will be selected based on the combined judgement of the multidiscipline team.

- **Environmental consequences (E)**

The consequence parameter, E, selection is based on the following criteria:

A release with minor environmental damage but which is large enough to be reported to plant management, e.g: • A moderate leak from a flange or valve • Small scale liquid spill • Small scale soil pollution without affecting ground water	Ea
A release with significant environmental damage within the site boundary, e.g. A cloud of obnoxious vapour travelling beyond the unit following flange gasket blow-out or compressor seal failure	Eb
Release outside the fence with major damage which can be cleaned up quickly without significant lasting consequences, e.g. A vapour or aerosol release with or without liquid fallout that causes temporary damage to plants or fauna	Ec
Release outside the fence with major damage which cannot be cleaned up quickly or with lasting consequences, e.g: • Liquid spill into a river or sea • A vapour or aerosol release with or without liquid fallout that causes lasting damage to plants or fauna • Solids fallout (dust, catalyst, soot, ash) • Liquid release that could affect groundwater	Ed

- **Economic consequences (L)**

The consequence parameter, L, selection is based on the following criteria:

Minor operational upset and / or minor damage to equipment (totalling less than £1m)	La
Moderate operational upset and / or moderate damage to equipment (totalling £1 to £10m)	Lb
Major operational upset and / or major damage to equipment (totalling £10m to £100m)	Lc
Major damage to essential equipment causing major economic loss (totalling greater than £100m)	Ld

8.1.3.2. Demand Rate For The Hazard Sequence (W)

The demand rate is the frequency of the fault sequence initiation, which may not correspond to the frequency that the SIF will be called to operate because mitigating systems may intervene. The table below gives the type of fault and typical frequencies of the initiating events. These, or alternative frequencies, should be agreed amongst the team prior to the start of the study.

Table 2: Initiating Event Demand Frequencies

Event	Frequency	Notes
Mechanical Items (Individual Pumps, compressors, cranes)	1 / Year	
Human Error (Routine, once per day opportunity)	1 / Year	
Loss of Cooling	1 / 10 Years	1, 2
Loss of Power	1 / 10 Years	1, 2
Loss of Instrument Air	1 / 10 Years	1, 2
Human Error (Routine, once per month opportunity)	1 / 10 Years	
Human Error (Non-routine, low stress)	1 / 10 Years	
Control Loop failure	1 / 10 Years	

NOTES

1. System includes back-up of mechanical items, otherwise failure rate would be higher.

2. Should be fail safe design.

Simultaneous failures and obviously rare events will be taken as less than 1 per 30 years (W1). Extremely rare events will be assigned W0, i.e. less than 1 per 300 years.

8.1.3.3. Personnel Exposure Factor (F)

This is the likelihood that when the accident occurs there will be someone present. Credit is taken for a low probability of exposure to the hazard where this probability is less than 10%. Generally personnel exposure will be taken as Fb except where manning levels and distribution indicate occupancy of less than 10% of the time (Fa). Large releases will tend towards Fb as their effect zone will be greater.

8.1.3.4. Avoidance Probabilities (P)

Avoidance of the consequences of the hazard depends on the speed with which the hazard develops. Where the event does not go immediately to the consequences and there is time for corrective action or escape, it is assumed that there is a 90% chance of avoiding the consequences.

The probability of avoiding consequences will be taken as Pb except where it is considered than the failure / event will not go immediately to the consequences and there is time for corrective action or escape (Pa).

Pa will only be selected if all the following are true:

- *facilities are provided to alert the operator that the Safety Instrumented Function has failed;*
- *independent facilities are provided to shut down such that he hazard can be avoided or which enable all persons to escape to a safe area;*
- *The time between the operator being alerted and a hazardous event occurring exceeds 1 hour or is definitely sufficient for the necessary actions.*

8.2. LOPA methodology

The Layers of Protection Analysis (LOPA) method as other method requires that the tolerable risk level (e.g. per scenario or cumulative) be stated explicitly as a numerical target. Once the tolerable risk frequency target is known, the required risk reduction - in terms of Probability of Failure on Demand (PFD) of the SIF - can be determined. LOPA evaluates risk in order of magnitude of selected unwanted event scenarios.

The information required for the LOPA is contained in the data collected and developed in the HAZOP study. Table below shows the relationship between the data required for LOPA and the data developed during the HAZOP study.

Table 3

LOPA Required Information	HAZOP Developed Information
Impact event	Consequence
Impact event severity level	Consequence severity
Initiating cause	Cause
Initiating likelihood	Cause frequency
Protection layers	Existing safeguards
Required additional mitigation	Recommended new safeguards

LOPA provides basis for specification of Independent Protection Layers (IPLs) and support compliance with good process safety practices as per IEC 61508 and IEC 61511.

8.2.1. Protection Layers

In a typical chemical process various layers of protection against incidents are in place. The main purpose of the layers is to reduce the frequency of undesired consequences.
These layers consist of preventive, protective or mitigating measures. Examples are:

- *Inherently safe design features;*
- *Basic Process Control System (BPCS);*
- *Critical alarms and Operator intervention;*
- *Safety Instrumented System (SIS) or Emergency Shutdown System;*
- *Pressure Relief Device;*
- *Mechanical Integrity of Vessel;*
- *Fire Suppression System;*

The layers of protection identified must be considered to be sufficiently independent to avoid common cause failure. An Independent Protection Layer (IPL) is a device, system, or action that is capable of preventing a scenario from proceeding to its undesired consequence independent of the initiating event or the action of any other layer of protection associated with the scenario to control, prevent and/or mitigate process risk.

8.2.2. LOPA Steps

The method starts with data developed in the Hazard and Operability analysis (HAZOP study) and accounts for each identified hazard by documenting the initiating cause and the protection layers that prevent or mitigate the hazard. The total amount of risk reduction can then be determined and the need for more risk reduction analyzed. If additional risk reduction is required and if it is to be provided in the form of a SIF, the LOPA methodology allows the determination of the appropriate SIL for the SIF. The method is illustrated in the figure below.
Steps are:

1. Select a SIF identifier (tag number) from the Cause & Effect Tables.
   - *Develop an `impact event scenario ‘ based on the HAZOP workshop records. The consequences’ identified in the HAZOP records are listed as `impact events’. Each `hazard and consequence’ is a single `impact event scenario’.*
   - *For each impact event scenario evaluate the severity consequences on HSE, and Assets*

2. Set the impact event scenario ` Target Likelihoods ‘ after mitigation to meet the HSE and Assets tolerable risks on the basis of severity of consequences on HSE and Assets

3. Initiating Cause(s)

Determine the initiating causes of each impact event, i.e. all of the Initiating Causes of the hazard determined in the HAZOP are listed.

4. Select an initiating cause and its Frequency

Calculate the enabled initiating event(s) frequency. The hazard initiating cause likelihood (in events per year) is agreed on, i.e. likelihood is estimated for each initiating cause.

5. Independent Protection Layers ‘IPLs’

Independent Protection Layers (IPLs) are listed. Each IPL is assigned a Probability of Failure on Demand (PFD) value.

Among IPLs are:

- *General Process Design / Inherent Safety: The general process design to reduce the likelihood of hazard manifesting itself, when an Initiating Cause occurs. An example of this would be a jacketed pipe or vessel. The jacket would prevent the release of process material if the integrity of the primary pipe or vessel were compromised.*
- *BPCS: If a control loop in the BPCS prevents the impacted event from occurring when the Initiating Cause occurs, credit based on its PFD is claimed.*
- *Operator Intervention (Alarms): This takes credit for alarms that alert the operator and utilize operator intervention. Ensure that the alarm is independent of the cause, and the BPCS*

6. Other Protection Layers. For each event the following probabilities are also determined:

- *Occupancy - The probability of a person being in the area.*
- *Ignition - The probability that a release of flammable material will ignited / explodes (given that it has already released). The probability that a release will be ignited depends on a number of factors, including the chemical’s reactivity, volatility, auto-ignition temperature, and physical state as well as the potential sources of ignition that are present. For a blast resulting from vapor cloud combustion, a reasonable amount of obstructions and confinement must exist to cause the flame front to burn turbulently and reach sonic velocity.*
- *Fatality - The probability that a person will die given a release of hazardous material and a person is already there. Allow for escape and/or avoidance.*

7. Intermediate Event Likelihood

The Intermediate Event Likelihood is calculated by multiplying the Initiating Likelihood by the PFDs of the protection layers and mitigating layers. The calculated number is in units of events per year. If the Intermediate Event Likelihood is less than the Corporate Criteria for Events of this Severity Level, additional PLs are not required. Further risk reduction should, however, be applied if economically appropriate.

8. Mitigated Event Likelihood

Mitigated event likelihood is calculated by multiplying the initiating cause likelihood by the

PFDs for the applicable IPLs. The mitigated event likelihood is then compared to a criterion linked to the corporation's criteria for unacceptable risk levels. Additional IPLs can be added to reduce the risk. The mitigated event likelihoods are summed to give an estimate of the risk for the whole process. Mitigated event likelihood is calculated by multiplying the initiating cause likelihood by the PFDs for the applicable IPLs. The mitigated event likelihood is then compared to a criterion linked to the corporation's criteria for unacceptable risk levels. Additional IPLs can be added to reduce the risk. The mitigated event likelihoods are summed to give an estimate of the risk for the whole process.

9. Select other initiating causes and their Frequencies. Repeat all the previous steps

10. Safety Integrity Level Selection

The SIFs required Integrity Level can be calculated by dividing the Corporate Risk Criteria for the event by the Required Event Likelihood (for all causes). A PFD for the SIF below this number is selected as a maximum for the SIS and entered.

Required Event Likelihood = Intermediate Event Likelihood x *(Probability of Ignition*
x *Probability of Occupancy* * *Probability of Fatality)*

11. Environmental Integrity Level `EIL' Selection

Exposure factor for Environmental effects and consequences are determined and inserted in corresponding cell. As a result the Environmental Integrity Level `EIL' will be determined. If a new SIF is needed to prevent environmental consequences, the Required Integrity Level can be calculated by dividing the Corporate Risk Criteria for the event by the Required Event Likelihood. A PFD for the SIF below this number is selected as a maximum for the SIS and entered.

Required Event Likelihood = (Intermediate Event Likelihood) x *(Exposure factor)*

12. Asset / Economical Integrity Level `AIL' Selection

Exposure factor for Asset / Economical effects and consequences are determined and inserted in corresponding cell. As a result the Asset / Economical Integrity Level `AIL' will be determined. If a new SIF is needed, the Required Integrity Level can be calculated by dividing the Corporate Criteria for the event by the Required Event Likelihood. A PFD for the SIF below this number is selected as a maximum for the SIS and entered.

Required Event Likelihood = Intermediate Event Likelihood x *(Probability of Ignition*
* *Probability of Occupancy* * *Probability of Fatality)* x *(PFD of safety instrumented function)*

13. Select another SIF identifier (tag number) from the Cause & Effect Tables
Repeat the process above

8.3. SIL Matrix Method

One common technique, among international refining, chemical and petrochemical companies, is to use a risk matrix, which provides a correlation of risk severity and risk likelihood to SIL. The method allows the probability of the potential event to be considered during the assignment of SIL.

At the first a hazard analysis to identify hazards, potential process deviations and their causes, available engineered systems, initiating events, and potential hazardous events that may occur should be performed for the process.

One such technique that is widely applied is a Hazard and Operability (HAZOP study) analysis. A qualitative approach can be used to assess process risk. Such an approach allows a traceable path of how the hazardous event develops, and the estimation of the likelihood (approximate range of occurrence) and the severity.

A risk matrix can be used for the evaluation of risk by combining the likelihood and the impact severity rating of hazardous events. A similar approach can be used to develop a matrix that identifies the potential risk reduction that can be associated with the use of a SIS protection layer. Such a risk matrix is shown in Figure 3. In Figure 3, the safety target level has been embedded in the matrix. In other words, the matrix is based on the operating experience and risk criteria of the specific company, the design, operating and protection philosophy of the company, and the level of safety that the company has established as its safety target level.

Figure 3: Example of Safety Layer Matrix

SIL Level Required

Number of PL'S									
3							c)	1	1
2	c)	c)	1	c)	1	2	1	2	b) 3
1	c)	1	2	1	2	b) 3	b) 3	b) 3	a) 3
Hazardous Event Likelihood	LOW	Med	High	LOW	Med	High	LOW	Med	High
	Minor			**Serious**			**Extensive**		
	Hazardous Event Severity Rating								

a) One level 3 safety instrumented function does not provide sufficient risk reduction at this risk level. Additional modifications are required in order to reduce risk (see d).
b) One level 3 safety instrumented function may not provide sufficient risk reduction at this risk level. Additional review is required (see d).
c) SIS independent protection layer is probably not needed.
d) This approach is not considered suitable for SIL 4.

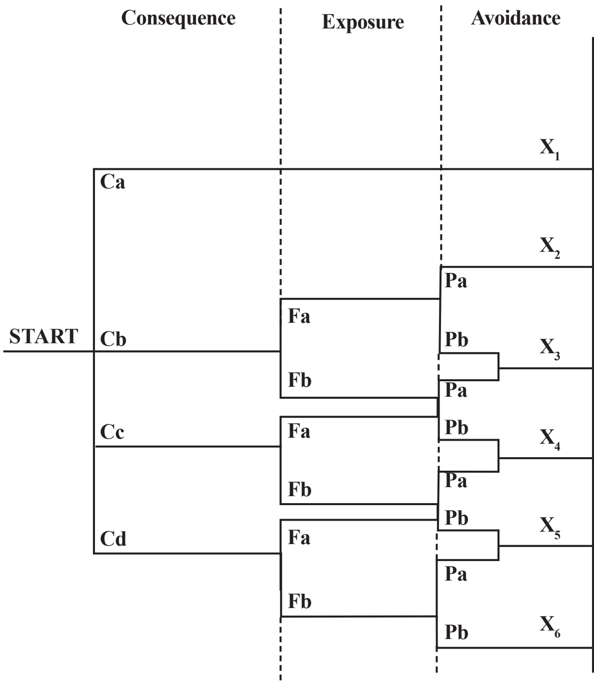
Total number of PLs – includes all the PLs protecting the process including the SIS being classified.
Hazardous event likelihood – likelihood that the hazardous event occurs without any of the PLs in service.
Hazardous event severity – the impact associated with the hazardous event.

8.3.1. Steps

1) Establish the process safety target level.
2) Perform a hazard identification (for example, HAZOP studies) to identify all hazardous events of interest.
3) Establish the hazardous event scenarios and estimate the hazardous event likelihood using company specific guidelines and data.
4) Establish the severity rating of the hazardous events using company specific guidelines.
5) Identify existing PLs. The estimated likelihood of hazardous events should be reduced by a factor of 10 for every PL.
6) Identify the need for an additional SIS protection layer by comparing the remaining risk with the safety target level.
7) Identify the SIL from Figure 3.

NOTE : The user should assess the possible level of dependency between protection layers and attempt to minimize any such occurrence.

Appendix 1. Safety risk graph



Demand Frequency

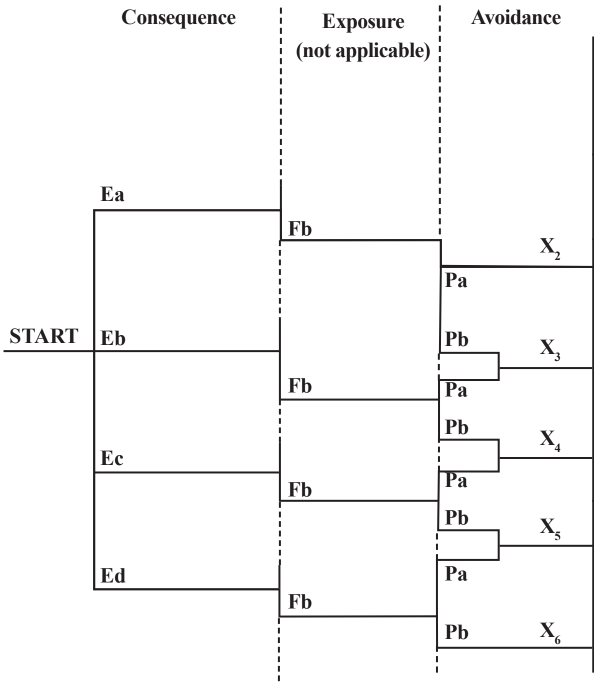
	W3	W2	W1	W0
	> 0.3 Per Year	0.3 to 0.03 Per Year	0.03 to 0.003 Per Year	< 0.003 Per Year
X_1	a	---	---	---
X_2	1	a	---	---
X_3	2	1	a	---
X_4	3	2	1	a
X_5	4	3	2	1
X_6	b	4	3	2

C	Consequence parameter
F	Exposure time parameter
P	Probability of avoiding the hazardous event
W	Demand frequency in the absence of the Safety Instrumented Function
---	No safety requirements
a	No special safety requirements
b	A single safety instrumented function is not sufficient
1,2,3,4	Safety Integrity Level
X	Row
Ca	Minor injury
Cb	Serious injury
Cc	Multiple serious injuries and / or a single fatality
Cd	More than one fatality on site and / or serious injury offsite
Fa	Probability of occupation of the area exposed to hazard less than 10%. This must take account of the possibility of personnel being present to assess upset conditions
Fb	Probability of occupation of the area exposed to hazard greater than 10%.
Pa	All the following conditions must be satisfied

• facilities are provided to alert the operator that the Safety Instrumented Function has failed;
• independent facilities are provided to shut down such that the hazard can be avoided or which enable all persons to escape to a safe area;
• The time between the operator being alerted and a hazardous event occurring exceeds 1 hour or is definitely sufficient for the necessary actions.

Pb	Any one of the conditions for Pa not satisfied
W3	Greater than 1 per 3 years
W2	1 per 3 years to 1 per 30 years (o.3 to 0.03 per year)
W1	1 per 30 years to 1 per 300 years (0.03 to 0.003 per year)
W0	Less than 1 per 300 years (0.003 per year)

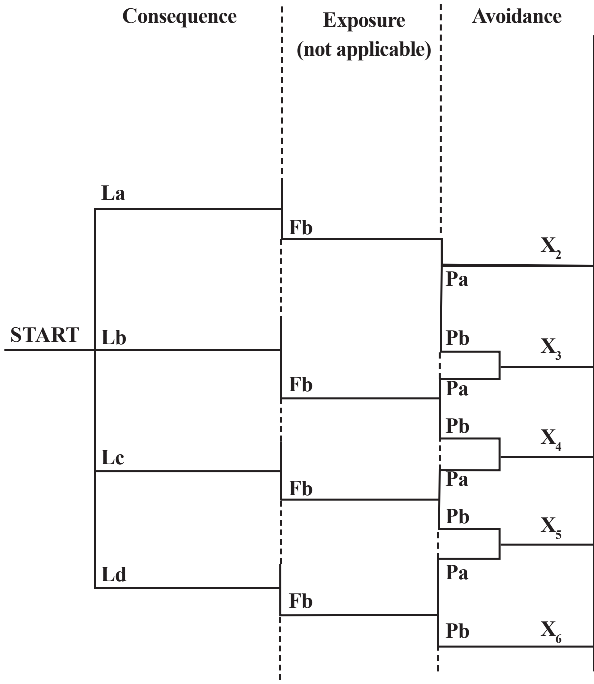
Appendix 2. Environmental risk graph



Demand Frequency			
W3	W2	W1	W0
> 0.3 Per Year	0.3 to 0.03 Per Year	0.03 to 0.003 Per Year	< 0.003 Per Year
a	---	---	---
1	a	---	---
2	1	a	---
3	2	1	a
4	3	2	1
b	4	3	2

L	Economic consequence parameter (includes asset loss and downtime costs)
F	Not applicable
P	Probability of avoiding the hazardous event
W	Demand frequency in the absence of the Safety Instrumented Function
---	No safety requirements
a	No special safety requirements
b	A single safety instrumented function is not sufficient
1,2,3,4	Safety Integrity Level
x	Row
La	Minor operational upset and / or minor damage to equipment (totalling less than £1m)
Lb	Moderate operational upset and / or moderate damage to equipment (totalling £1 to £10m)
Lc	Major operational upset and / or major damage to equipment (totalling £10m to £100m)
Ld	Major damage to essential equipment causing major economic loss (totalling greater than £100m)
Fa / Fb	Not applicable
Pa	All the following conditions must be satisfied • facilities are provided to alert the operator that the Safety Instrumented Function has failed; • independent facilities are provided to shut down such that the hazard can be avoided; • The time between the operator being alerted and a hazardous event occurring exceeds 1 hour or is definitely sufficient for the necessary actions.
Pb	Any one of the conditions for Pa not satisfied
W3	Greater than 1 per 3 years
W2	1 per 3 years to 1 per 30 years (o.3 to 0.03 per year)
W1	1 per 30 years to 1 per 300 years (0.03 to 0.003 per year)
W0	Less than 1 per 300 years (0.003 per year)

Appendix 3. Economic risk graph



Demand Frequency

	W3	W2	W1	W0
	> 0.3 Per Year	0.3 to 0.03 Per Year	0.03 to 0.003 Per Year	< 0.003 Per Year
	a	---	---	---
X_2	1	a	---	---
X_3	2	1	a	---
X_4	3	2	1	a
X_5	4	3	2	1
X_6	b	4	3	2

L	Economic consequence parameter (includes asset loss and downtime costs)
F	Not applicable
P	Probability of avoiding the hazardous event
W	Demand frequency in the absence of the Safety Instrumented Function
---	No safety requirements
a	No special safety requirements
b	A single safety instrumented function is not sufficient
1,2,3,4	Safety Integrity Level
x	Row
La	Minor operational upset and / or minor damage to equipment (totalling less than £1m)
Lb	Moderate operational upset and / or moderate damage to equipment (totalling £1 to £10m)
Lc	Major operational upset and / or major damage to equipment (totalling £10m to £100m)
Ld	Major damage to essential equipment causing major economic loss (totalling greater than £100m)
Fa / Fb	Not applicable
Pa	All the following conditions must be satisfied • facilities are provided to alert the operator that the Safety Instrumented Function has failed; • independent facilities are provided to shut down such that the hazard can be avoided; • The time between the operator being alerted and a hazardous event occurring exceeds 1 hour or is definitely sufficient for the necessary actions.
Pb	Any one of the conditions for Pa not satisfied
W3	Greater than 1 per 3 years
W2	1 per 3 years to 1 per 30 years (o.3 to 0.03 per year)
W1	1 per 30 years to 1 per 300 years (0.03 to 0.003 per year)
W0	Less than 1 per 300 years (0.003 per year)

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