

*PROCESS DESIGN ENGINEERING
MANUAL:
Part 1*

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1. GETTING STARTED

1.1 PROCESS ENGINEERING AND ROLE OF PROCESS ENGINEER

Process design is the design of processes for desired physical and or chemical transformation of materials. Process design is central to chemical engineering and it can be considered to be the summit of chemical engineering, bringing together all of the components of that field.

Process Engineering involves design of unit operations & equipment design.

Role of Process Engineer:

Chemical engineers (or process engineers) are responsible for developing new industrial processes and designing new process plants and equipment or modifying existing ones. The processes that they come up with are used to create products ranging from oil and gas, chemicals, petrochemicals, speciality chemicals to food and drink.

It is a vocation wherein the process engineer is supposed to perform any one or all of the activities as mentioned below to provide documentation for a safe, reliable and profitable design

- Design new equipment/unit/plant as per good and internationally accepted engineering practices (Greenfield)
- Rate or check adequacy of existing equipment/unit/plant for changed operating conditions (e.g. pressure, temperature, flow etc.) as per good and internationally accepted engineering practices (Brownfield)

➤ **Principal accountabilities:**

- High level design option selection including economics.
- Optimization/evaluation of existing processes, including understanding feedstock composition, heat/material balance implications.
- Troubleshoots existing plant systems.
- Understands, uses and champions engineering standards and work processes.
- Modelling of systems.
- Plant support when needed, Process Hazard analysis, Documentation of work.

➤ **Responsibilities:**

- Develops and evaluates plans and criteria for a variety of projects and problem solving activities, including scale up design, optimization of chemical manufacturing processes and facilities.
- Develops/assesses the feasibility, soundness, and safety of engineering designs and solutions. Provides conceptual design of process equipment systems.
- Identify and apply design and simulation tools for design, optimization and troubleshooting of processes.
- Based on Client's requirements, develops Heat & Material Balances using simulation software.
- Develops Process Flow diagrams to show operations necessary to convert feedstock into on specification products, in the most economical manner.
- Defines and sizes Equipment and Instrumentation to be implemented and issues relevant Process Data sheets.
- Elaborates Piping and Instrumentation Diagrams.

- Performs hydraulic and thermal calculations on piping systems, pipelines, flow lines in steady state and transient operation.
- Defines the utilities requirements and designs the associated systems.
- Defines quantities and characteristics of effluents to be disposed to the environment.
- Participates to the preparation of Operating Manual, under Pre-commissioning department supervision.
- Identify process hazards and participate in process safety review.

1.2 VARIOUS PACKAGES USED IN INDUSTRY

➤ Conceptual And Feasibility Studies

When a client is thinking of investing money in a chemical process plant or chemical technology he has more questions than answers. Some such questions would be:

- Is it technically feasible?
- What is the best possible chemistry and reaction for the process?
- Should it be a batch or continuous production process?
- How much would it cost to set up the unit?
- Is there a ready market available for selling the product?
- Is the technology environment friendly and safe?

What kind of process engineering activities and deliverables are expected during this phase?

- Evaluating the various process routes from economic, environment and safety view point and selecting the best option.
- Develop preliminary process flow or block diagrams
- Develop preliminary material and energy balances based on a selected economic plant capacity.
- List of major equipment.
- Estimate cost of the project with a deviation of +50% / -30%.
- Identify weak areas in process design such as experimental design, insufficient laboratory or pilot plant data, scale-up from pilot plant design and benchmarking against existing designs.

➤ Front End Engineering Design (FEED)

Process Engineering Information

- Raw Material Specifications
- Product Quality Specifications & Deviation Limits in case of specific products with specific end-use
- Utility Specifications

Typical Process Engineering Deliverables:

- Process Design Basis
- Extensive Heat & Material Balance generally using process design simulators
- Process Flow Diagrams (PFDs) including major process control automation
- Utility Balance and Utility Flow Diagrams (UFDs)
- Piping & Instrumentation Diagrams (P&IDs)
- Process Safeguarding Diagrams

- Material Selection Diagrams
- Line Designation List, Equipment List, Tie-in List
- Hazardous Area Classification Schedule
- Equipment (Static / Rotating) Process Datasheets
- Instrument Process Datasheets
- Relief Devices Datasheets and Relief Devices Summary Report
- Process Description
- Operating, Control and Safeguarding Philosophy, Process Control Narrative
- Hydraulic Calculations and Line Sizing Report
- Trip and Alarm Summary Report
- PHA / HAZID / HAZOP Reports
- Piping Material Specification
- Insulation Specification

➤ **Detail Engineering Design**

All the deliverables prepared during the FEED stage will be updated and upgraded during this phase. In addition to this many documents and drawings specific to this phase are prepared which involve engineering from other engineering disciplines such as Mechanical, Piping, Instrumentation, Electrical and Civil.

List of Specific Deliverables Prepared during Detail Design

- Pre-commissioning and Commissioning Manual
- Operation and Maintenance Manual
- Pre-Construction HAZOP Review and Closeout Report
- Effluents and Emissions Summary
- Single Line Diagrams (SLDs)
- Instrument Loop Diagrams (ILDs)
- Material Requisitions (MRs) for Static Equipment (Tanks / Vessels / Columns)
- MRs for Rotary Equipment (Pumps / Compressors / Fans / Blowers)
- MRs for Electrical Items (Transformers / DG Sets / Switchgear / Cables / UPS System)
- MRs for Instrument Items (Flow / Pressure / Level / Temperature / Control Valves / Shut-down Valves / DCS / PLC)
- MRs for Civil / Structural
- Piping Key Plan, Piping GA Drawings, Piping Isometric Drawings
- Piping Bulk Material Take-Off (MTO) (Valves / Pipe fittings / Pipe Supports)
- Instrument Bulk MTO (Tubing & Fittings / Cable Trays / Stanchion / Instrument Cables/ Cable Glands / Multi Cable Transit)
- Electrical Bulk MTO (Cables / Conduits/ Cable Trays / Cable Ties / Connectors / Lugs)
- Bar Bending Schedule (Civil)
- Civil Bulk MTO (Earthwork Cut/Fill / Piles / Structural Steel / Concrete)
- Review of Vendor Documents / Drawings (All Disciplines)
- Technical Bid Analysis
- Specifications for Active and Passive Fire Protection
- Fire Fighting MTO

As-built design

During the construction phase of the project changes can occur due to the following reasons:

- Errors in the design phase.
- Changes demanded by the client due to changed business scenario and economics during the construction phase (this is rare, but can happen).
- Cost-cutting and / or time-saving exercise for bulk items generally items such as valves and field instrumentation.
- Field changes demanded by the commissioning / operating experts to facilitate plant commissioning.

Engineering Activities and Deliverables during As-Built Design

- Maintaining a site “Management of Change” dossier wherein the construction phase changes are planned, reviewed from design, cost and most importantly from the HSE view point, approved and actioned. (This is a construction engineering activity).
- P&ID updation based on red-line mark-ups done by the construction / commissioning team.
- Piping Gas / Isometrics updation by the piping contractor based on red-line mark-ups
- Instrument Loop Drawings again based on red-line mark-ups.
- Pre-commissioning / Commissioning Manual Updation.
- Plant Operation Manual Updation.

1.3 OVERVIEW OF VARIOUS UNIT OPERATIONS IN USED PROCESS INDUSTRIES

- Fluid flow processes, including fluids transportation, filtration, and solids fluidization.
- Heat transfer processes, including evaporation.
- Mass transfer processes, including gas absorption, distillation, extraction, adsorption, and drying.
- Thermodynamic processes, refrigeration.
- Mechanical processes, sieving.

➤ Fluid flow processes, including fluids transportation, filtration, and solids fluidization

○ Fluid Transportation

Pipeline transport is the transportation of goods or material through a pipe. Liquids and gases are transported in pipelines and any chemically stable substance can be sent through a pipeline. Pipelines exist for the transport of crude and refined petroleum, fuels – such as oil, natural gas and biofuels – and other fluids including sewage, slurry, water, and beer. Pipelines are useful for transporting water for drinking or irrigation over long distances when it needs to move over hills, or where canals or channels are poor choices due to considerations of evaporation, pollution, or environmental impact. Pneumatic tubes using compressed air can be used to transport solid capsules.

Oil pipelines are made from steel or plastic tubes which are usually buried. The oil is moved through the pipelines by pump stations along the pipeline. Natural gas (and similar gaseous fuels) are lightly pressurised into liquids known as Natural Gas Liquids (NGLs). Natural gas pipelines are constructed of carbon steel. Hydrogen pipeline transport is the transportation of hydrogen through a pipe. District heating systems use a network of insulated pipes which transport heated water, pressurized hot water, or sometimes steam to the customer.

Pipelines conveying flammable or explosive material, such as natural gas or oil, pose special safety concerns and there have been various accidents. Pipelines can be the target of theft, vandalism, sabotage, or even terrorist attacks. In war, pipelines are often the target of military attacks.

- **Filtration:**

Filtration is any of various mechanical, physical or biological operations that separate solids from fluids (liquids or gases) by adding a medium through which only the fluid can pass. The fluid that passes through is called the filtrate. In physical filters oversize solids in the fluid are retained and in biological filters particulates are trapped and ingested and metabolites are retained and removed. However, the separation is not complete; solids will be contaminated with some fluid and filtrate will contain fine particles (depending on the pore size, filter thickness and biological activity).

Filtration occurs both in nature and in engineered systems; there are biological, geological, and industrial forms. For example, in animals (including humans), renal filtration removes wastes from the blood, and in water treatment and sewage treatment, undesirable constituents are removed by absorption into a biological film grown on or in the filter medium, as in slow sand filtration.

- **Fluidization**

Fluidization (or fluidisation) is a process similar to liquefaction whereby a granular material is converted from a static solid-like state to a dynamic fluid-like state. This process occurs when a fluid (liquid or gas) is passed up through the granular material.

When a gas flow is introduced through the bottom of a bed of solid particles, it will move upwards through the bed via the empty spaces between the particles. At low gas velocities, aerodynamic drag on each particle is also low, and thus the bed remains in a fixed state. Increasing the velocity, the aerodynamic drag forces will begin to counteract the gravitational forces, causing the bed to expand in volume as the particles move away from each other. Further increasing the velocity, it will reach a critical value at which the upward drag forces will exactly equal the downward gravitational forces, causing the particles to become suspended within the fluid. At this critical value, the bed is said to be fluidized and will exhibit fluidic behavior. By further increasing gas velocity, the bulk density of the bed will continue to decrease, and its fluidization becomes more violent, until the particles no longer form a bed and are "conveyed" upwards by the gas flow.

When fluidized, a bed of solid particles will behave as a fluid, like a liquid or gas. Like water in a bucket: the bed will conform to the volume of the chamber, its surface remaining perpendicular to gravity; objects with a lower density than the bed density will float on its surface, bobbing up and down if pushed downwards, while objects with a higher density sink to the bottom of the bed. The fluidic behaviour allows the particles to be transported like a fluid, channelled through pipes, not requiring mechanical transport (e.g. conveyor belt).

➤ **Heat transfer processes, including evaporation and heat exchange**

- **Evaporation**

Evaporation is a type of vaporization that occurs on the surface of a liquid as it changes into the gaseous phase before reaching its boiling point. The surrounding gas must not be saturated with the evaporating substance. When the molecules of the liquid collide, they transfer energy to each other based on how they collide. When a molecule near the surface absorbs enough energy to

overcome the vapor pressure, it will "escape" and enter the surrounding air as a gas. When evaporation occurs, the energy removed from the vaporized liquid will reduce the temperature of the liquid, resulting in evaporative cooling.

On average, only a fraction of the molecules in a liquid have enough heat energy to escape from the liquid. The evaporation will continue until equilibrium is reached when the evaporation of the liquid is equal to its condensation. In an enclosed environment, a liquid will evaporate until the surrounding air is saturated.

➤ **Mass transfer processes, including gas absorption, distillation, extraction, adsorption, and drying**

○ **Absorption:**

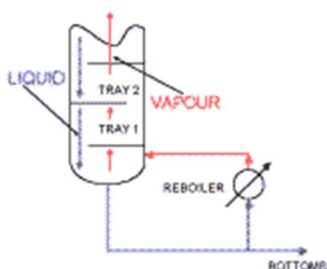
Absorption is the one of the most commonly used separation techniques for the gas cleaning purpose for removal of various gases like H_2S , CO_2 , SO_2 and Ammonia. Cleaning of solute gases is achieved by transferring into a liquid solvent by contacting the gas stream with liquids that offers specific or selectivity for the gases to be recovered. Unit operation and is mass transfer phenomena where the solute of a gas is removed from being placed in contact with a non-volatile liquid solvent that removes the components from the gas.

Solvent: Liquid applied to remove the solute from a gas stream.

Solute: Components to be removed from entering streams.

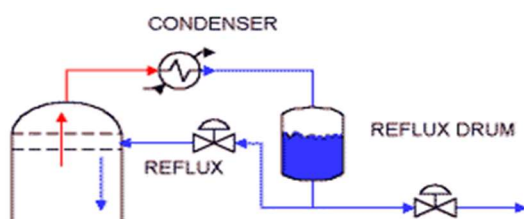
○ **Distillation:**

The liquid mixture that is to be processed is known as the feed and this is introduced usually somewhere near the middle of the column to a tray known as the feed tray. The feed tray divides the column into a top (enriching or rectification) section and a bottom (stripping) section. The feed flows down the column where it is collected at the bottom in the reboiler.



Heat is supplied to the reboiler to generate vapour. The source of heat input can be any suitable fluid, although in most chemical plants this is normally steam. In refineries, the heating source may be the output streams of other columns. The vapour raised in the reboiler is re-introduced into the unit at the bottom of the column. The liquid removed from the reboiler is known as the bottoms product or simply, bottoms

The vapour moves up the column, and as it exits the top of the unit, it is cooled by a condenser. The condensed liquid is stored in a holding vessel known as the reflux drum. Some of this liquid is recycled back to the top of the column and this is called the reflux. The condensed liquid that is removed from the system is known as the distillate or top product.



Thus, there are internal flows of vapour and liquid within the column as well as external flows of feeds and product streams, into and out of the column.

- **Adsorption**

Adsorption technology is now used very effectively in the separation and purification of many gas and liquid mixtures in chemical, petrochemical, biochemical and environmental industries and is often a much cheaper and easier option than distillation, absorption or extraction. Some of the major applications of adsorption are gas bulk separation, gas purifications, liquid bulk separation, and liquid purifications.

One of the most effective method for recovering and controlling emissions of volatile organic compounds is adsorption. Some of the commercial adsorbents are silica gel, activated carbon, carbon molecular sieve, charcoal, zeolites molecular sieves, polymer and resins, clays, biosorbents. Some of the key properties of adsorbents are capacity, selectivity, regenerability, kinetics, compatibility and cost. Some of the methods used for regeneration of adsorbent are thermal swing, pressure swing, vacuum (special case of pressure swing), purge and gas stripping, steam stripping [Crittenden, 1988].

Some of the important criteria of good adsorbent are as below.

- It must selectively concentrate one or more components called adsorbate to from their fluid phase levels
- The ability to release adsorbate so that adsorbent can be reused,
- As high as possible delta loading the change of weight of adsorbate per unit weight of adsorbent between adsorbing and desorbing steps over a reasonable range of pressure and Pressure swing adsorption (PSA) is based on the principle of relative adsorption strength, is a milestone in the science of gas separation. Some of the commercial application of PSA are air drying, hydrogen purification, bulk separation of paraffins, air separation for oxygen and nitrogen production,
-

Chromatography is a sorptive separation process. In chromatography feed is introduced in column containing a selective adsorbent (stationary phase) and separated over the length of the column by the action of a carrier fluid (mobile phase) that is continually supplied to the column following the introduction of the feed. The separation occurs as a result of the different partitioning of the feed solutes between the stationary phases. The separated solutes are recovered at different time in the effluent from the column.

Liquid-Liquid Extraction

Liquid-liquid extraction has been commonly used in petroleum and petrochemical industry for separation of close boiling hydrocarbons. Some of the major applications are:

- Removal of sulphur compound from liquid hydrocarbons
- Recovery of aromatics from liquid hydrocarbon
- Separation of butadiene from C4 hydrocarbons
- Extraction of Caprolactam
- Separation of homogeneous aqueous azeotropes
- Extraction of acetic acid
- Removal of phenolic compounds from waste water
- Manufacture of rare earths
- Separation of asphaltic compounds from oil

- Recovery of copper from leach liquor
- Extraction of glycerides from vegetable oil. Some of the important property of a good solvent
- High solvent power/capacity
- High selectivity for desired component
- Sufficient difference in boiling points of the solvent and the feed for effective separation
- Low latent heat of evaporation and specific heat to reduce utility requirement
- high thermal and chemical stability
- Low melting point
- Relatively inexpensive
- Non-toxic and non –corrosive
- Low viscosity low interfacial tension
- Drying

Drying may be defined as the vaporization and removal of water or other liquids from a solution, suspension, or other solid-liquid mixture to form a dry solid. It is a complicated process that involves simultaneous heat and mass transfer, accompanied by physicochemical transformations. Drying occurs as a result of the vaporization of liquid by supplying heat to wet feedstock, granules, and filter cakes and so on. Based on the mechanism of heat transfer that is employed, drying is categorized into direct (convection), indirect or contact (conduction), radiant (radiation) and dielectric or microwave (radio frequency) drying.

Heat transfer and mass transfer are critical aspects in drying processes. Heat is transferred to the product to evaporate liquid, and mass is transferred as a vapor into the surrounding gas. The drying rate is determined by the set of factors that affect heat and mass transfer. Solids drying is generally understood to follow two distinct drying zones, known as the constant-rate period and the falling-rate period. The two zones are demarcated by a break point called the critical moisture content.

In a typical graph of moisture content versus drying rate and moisture content versus time (Figure 1), section AB represents the constant-rate period. In that zone, moisture is considered to be evaporating from a saturated surface at a rate governed by diffusion from the surface through the stationary air film that is in contact with it. This period depends on the air temperature, humidity and speed of moisture to the surface, which in turn determine the temperature of the saturated surface. During the constant rate period, liquid must be transported to the surface at a rate sufficient to maintain saturation.

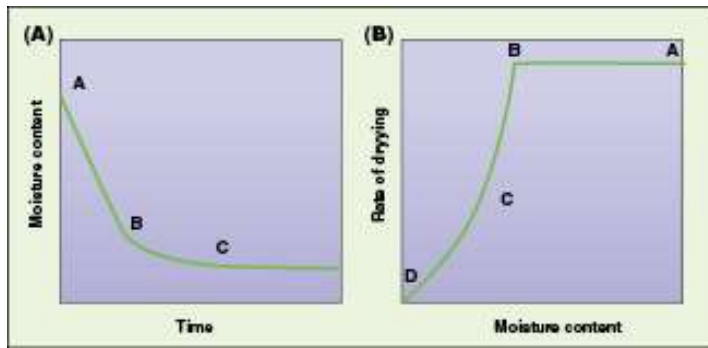


Figure 1. Segment AB of the graph represents the constant-rate drying period, while segment BC is the falling-rate period.

At the end of the constant rate period, (point B, Figure 1), a break in the drying curve occurs. This point is called the critical moisture content, and a linear fall in the drying rate occurs with further drying. This section, segment BC, is called the first falling-rate period. As drying proceeds, moisture reaches the surface at a decreasing rate and the mechanism that controls its transfer will influence the rate of drying. Since the surface is no longer saturated, it will tend to rise above the wet bulb temperature. This section, represented by segment CD in Figure 1 is called the second falling-rate period, and is controlled by vapor diffusion. Movement of liquid may occur by diffusion under the concentration gradient created by the depletion of water at the surface. The gradient can be caused by evaporation, or as a result of capillary forces, or through a cycle of vaporization and condensation, or by osmotic effects.

The capacity of the air (gas) stream to absorb and carry away moisture determines the drying rate and establishes the duration of the drying cycle. The two elements essential to this process are inlet air temperature and air flow rate. The higher the temperature of the drying air, the greater its vapor holding capacity. Since the temperature of the wet granules in a hot gas depends on the rate of evaporation, the key to analysing the drying process is psychrometry, defined as the study of the relationships between the material and energy balances of water vapor and air mixture.

➤ Thermodynamic processes, refrigeration.

○ Refrigeration

The job of the refrigeration cycle is to remove unwanted heat from one place and discharge it into another. To accomplish this, the refrigerant is pumped through a closed refrigeration system. If the system was not closed, it would be using up the refrigerant by dissipating it into the surrounding media; because it is closed, the same refrigerant is used over and over again, as it passes through the cycle removing some heat and discharging it. The closed cycle serves other purposes as well; it keeps the refrigerant from becoming contaminated and controls its flow, for it is a liquid in some parts of the cycle and a gas or vapor in other phases.

Let's look at what happens in a simple refrigeration cycle, and to the major components involved. Two different pressures exist in the cycle - the evaporating or low pressure in the "low side," and the condensing, or high pressure, in the "high side." These pressure areas are separated by two dividing points: one is the metering device where the refrigerant flow is controlled, and the other is at the compressor, where vapor is compressed.

The metering device is a point where we will start the trip through the cycle. This may be a thermal expansion valve, a capillary tube, or any other device to control the flow of refrigerant into the evaporator, or cooling coil, as a low-pressure, low-temperature refrigerant. The expanding refrigerant evaporates (changes state) as it goes through the evaporator, where it removes the heat from the substance or space in which the evaporator is located.

Heat will travel from the warmer substance to the evaporator cooled by the evaporation of the refrigerant within the system, causing the refrigerant to "boil" and evaporate, changing it to a vapor. This is similar to the change that occurs when a pail of water is boiled on the stove and the water changes to steam, except that the refrigerant boils at a much lower temperature.

Now this low-pressure, low-temperature vapor is drawn to the compressor where it is compressed into a high-temperature, high-pressure vapor. The compressor discharges it to the condenser, so that it can give up the heat that it picked up in the evaporator. The refrigerant vapor is at a higher temperature than the air passing across the condenser (air-cooled type); or water passing through the condenser (water-cooled type); therefore that is transferred from the warmer refrigerant vapor to the cooler air or water.

In this process, as heat is removed from the vapor, a change of state takes place and the vapor is condensed back into a liquid, at a high-pressure and high-temperature.

The liquid refrigerant travels now to the metering device where it passes through a small opening or orifice where a drop in pressure and temperature occurs, and then it enters into the evaporator or cooling coil. As the refrigerant makes its way into the large opening of the evaporator tubing or coil, it vaporizes, ready to start another cycle through the system.

The refrigeration system requires some means of connecting the basic major components - evaporator, compressor, condenser, and metering device - just as roads connect communities. Tubing or "lines" make the system complete so that the refrigerant will not leak out into the atmosphere. The suction line connects the evaporator or cooling coil to the compressor, the hot gas or discharge line connects the compressor to the condenser, and the liquid line is the connecting tubing between the condenser and the metering device (Thermal expansion valve). Some systems will have a receiver immediately after the condenser and before the metering device, where the refrigerant is stored until it is needed for heat removal in the evaporator.

There are many different kinds and variations of the refrigeration cycle components. For example, there are at least a half dozen different types of compressor, from the reciprocating, piston through a screw, scroll and centrifugal impeller design, but the function is the same in all cases - that of compressing the heat laden vapor into a high-temperature vapor.

The same can be said of the condenser and evaporator surfaces. They can be bare pipes, or they can be finned condensers and evaporators with electrically driven fans to pass the air through them, or with a condenser pump to pump the water through a water-cooled condenser.

There are a number of different types of metering devices to regulate the liquid refrigerant into the evaporator, depending on size of equipment, refrigerant used, and its application.

The mechanical refrigeration system described above is essentially the same whether the system be a domestic refrigerator, a low-temperature freezer, comfort air conditioning system, industrial chiller, or commercial cooling equipment. Refrigerants will be different and size of the equipment will vary greatly, but the principle of operation and the refrigeration cycle remains the same. Thus, once you understand the simple actions that are taking place within the refrigeration mechanical cycle you should have a good understanding how a refrigeration system works.

➤ **Mechanical processes, sieving**

○ **Sieving**

Sieving is one of the oldest and most commonly used method of sorting materials. Yet, when improperly carried out, sieving can provide misleading information or biased separation. Sieving has widespread application to industries as diverse as mining, pharmaceutical production and agriculture. The goal is typically to control or measure the particle size distribution. Sieving may be a direct part of a production process, a quality control procedure or a sample characterization. Regardless of the purpose, an understanding of sieving is necessary to optimize and accurately use this technique. Sieving has the clear advantage of being a simple, readily understood and relatively inexpensive method. This method also has the ability to provide reproducible results. This has tended to boost confidence in the method, even where it is not warranted.

Some of this confidence is a result of ignorance as to the actual errors involved in a given sieving operation or with a particular set of sieves. While these errors can be described and measured, the difficulty of doing so largely detracts from the attractiveness of the method. Thus, most people rely on standard procedures and the reproducibility of the method to reassure themselves as to the quality of the data obtained.

Sieving was the earliest means of particle size fractionation. Basically, the process of sieving is that of placing the particles to be fractionated on a pattern of openings or holes. The individual openings are referred to as the aperture. Small particles may fall through or the sieve retains the larger particles. Separation requires agitation and time. A variety of mechanisms exist to provide agitation either to the sieve or to the particles to be fractionated. Typically, commercial equipment varies in the manner in which agitation is created or the fluid used to support the particles. Either air or water may be used to support the particles as they sort on the sieve. Dry sieving has a lower practical limit of 50 μm , while wet sieving can separate smaller particle sizes when using special sieves or small volumes of particles.

1.4 HAZOP AND HAZID

HAZOP study is to carefully review a process or operation in a systematic manner to determine whether deviations from the design or operational intent can lead to undesirable consequences. This technique can be used for continuous or batch processes and can be adopted to evaluate written procedures. The HAZOP team lists potential causes and consequences of the deviation as well as existing safeguards protecting against the deviation. When the team determines that inadequate safeguards exist for a credible deviation, it usually recommends the action be taken to reduce the risk.

Objective of carrying out a HAZOP study:

- To check a design
- To decide whether and where to build
- To decide whether to buy a piece of equipment
- To obtain a list of questions to put to a supplier
- To check running instructions
- To improve the safety of existing facilities

Risks to be evaluated:

Major risks to be evaluated are probable incidents which may impact any of the following:

- Risk to Personnel: Health and Safety risk to own or public personnel.
- Risk to Environment: Risk of damage to environment due to any emissions causing pollution, from solid, liquid, vapour, vibration/shock, radiation noise, odour.
- Risk to Material: Risk to own or public material in terms of equipment, installation, work in process, utilities, etc. causing substantial losses affecting profitability.
- Risk to Reputation: Risk to own reputation as a responsible and reliable organization in terms of product quality, statutory and social obligation, ethics and goodwill.

Recommended stages at which Hazop study should preferably be conducted:

Stage 1-Preliminary: During Basic/Detailed Engineering when the process and P&I drawings and layouts are finalized.

Stage 2-After Installation: After site installation and water trials, prior to or during commissioning.

Stage 3-Need Based: On major modifications of installation or operational procedures.

HAZID:

HAZID (Hazard Identification) is a qualitative technique for the early identification of potential hazards and threats effecting people, the environment, assets or reputation. The major benefit of a HAZID study is to provide essential input to project development decision. It is a means of identifying and describing HSE hazards and threats at the earliest practicable stage of a development or venture.

Objective of the study:

- To identify the potential hazards and to reduce the probability and consequences of an incident at site that would have a detrimental impact to the personnel plant, properties and environment.

Benefits of Carrying out HAZID

- Identify opportunities for inherent safety.
- Identify Fire, explosion, toxic release scenarios and measure to prevent it.
- Any special preparations required to be taken to handle these can be pre-planned.
- Any specific process modifications if required can be established at an early stage.
- Prepares the system & team, ready and confident to go ahead for commissioning. Avoids major surprises.
- Hazards involved in operating each equipment's can be enlisted at the beginning, leading to better process mapping & better control in future to getting OHSAS / ISO approvals.
- The major benefit of HAZID is early identification and assessment of the critical health, safety and environmental hazards provides essential input to the project development decisions.

1.5 INTERNATIONAL CODES, STANDARDS AND PRACTICES OVERVIEW

➤ **CODE**

Code is group of general rules or systematic procedure or guidelines required for design, fabrication, installation and inspection and is prepared in such a manner that it can be adopted by legal jurisdiction and made into law.

➤ **STANDARDS**

Standards prepared by a professional group or committee which are believed to be good and proper engineering practice and which contain mandatory requirements and dimensions.

Standards are mainly two types:

- Dimensional Standards
- Pressure Integrity Standards

➤ **DIMENSIONAL STANDARDS**

They provide configuration control information for components. The main purpose of Dimensional Standards is to assure similar components manufactured by different supplier will be physically interchangeable.

➤ **PRESSURE INTEGRITY STANDARDS**

They provide performance criteria. The components designed and manufactured to the same standards will run an equivalent manner.

MAJOR ORGANIZATION FOR STANDARDS

Sr. No.	Country	Organization
1	India	BIS: Bureau of Indian standards
2	U.S.A	ASME: American Society of Mechanical Engineers
3	United Kingdom	BSI: British Standards Institute
4	Germany	DIN : Deutsches Institute for Normung
5	Japan	JIS: Japanese Industrial Standards
6	Europe	CEN : Europe Community for Standardization
7	Canada	CSA: Canadian Standards Association
8	France	AFNOR: Association Francoise

➤ **AMERICAN STANDARDS**

Commonly used American Standards:

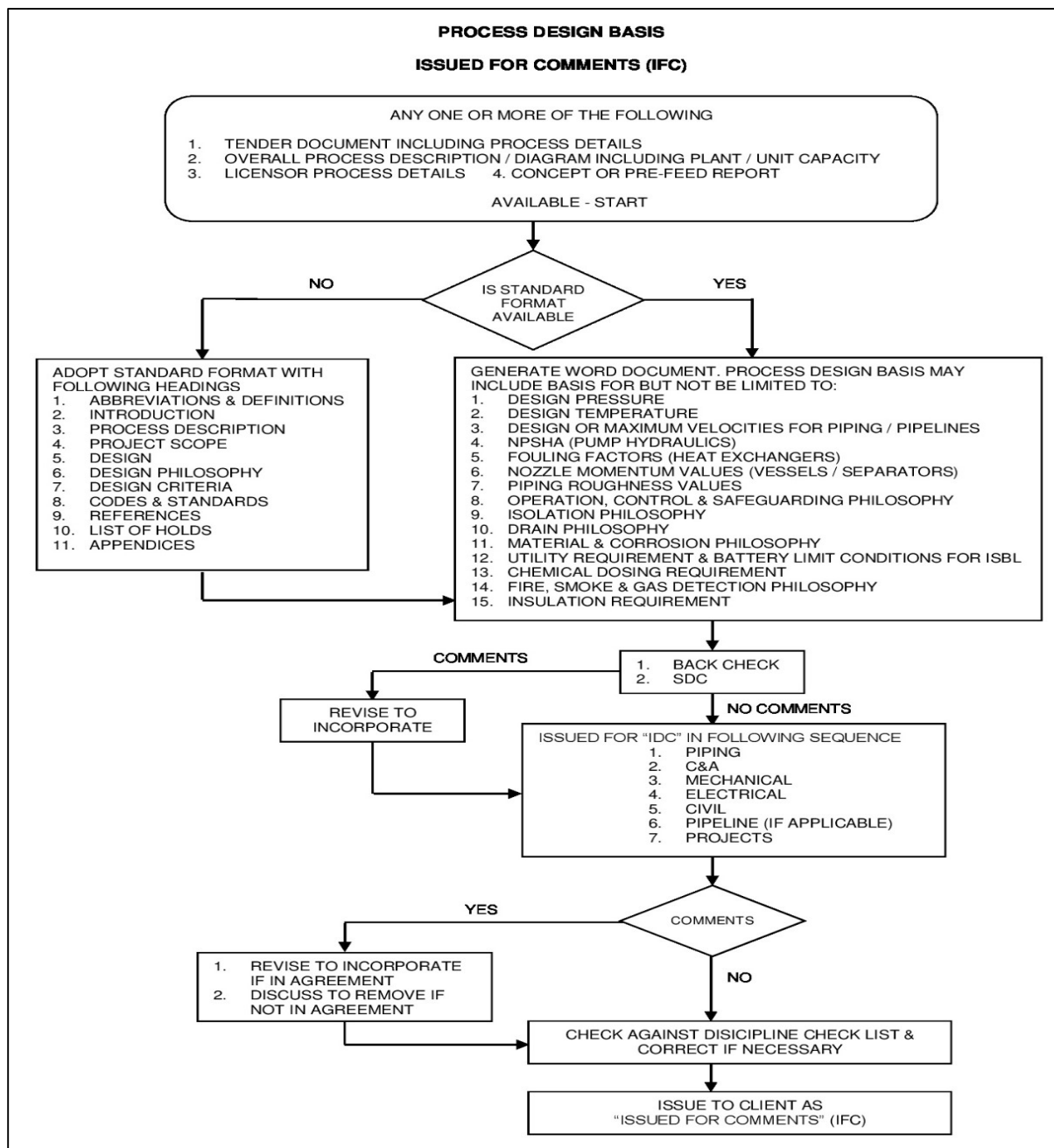
- API 12J: Specification for Oil and Gas Separators
- API RP 14E: Recommended Practice for Design and Installation of Offshore Production Platform Piping Systems
- API 520: Sizing, Selection, and Installation of Pressure-relieving Devices
- API 521: Pressure relieving and depressuring systems.
- API 2000: Venting Atmospheric and Low Pressure Storage Tank
- ASME SEC VIII DIV1/TEMA_/: Heat Exchanger Design

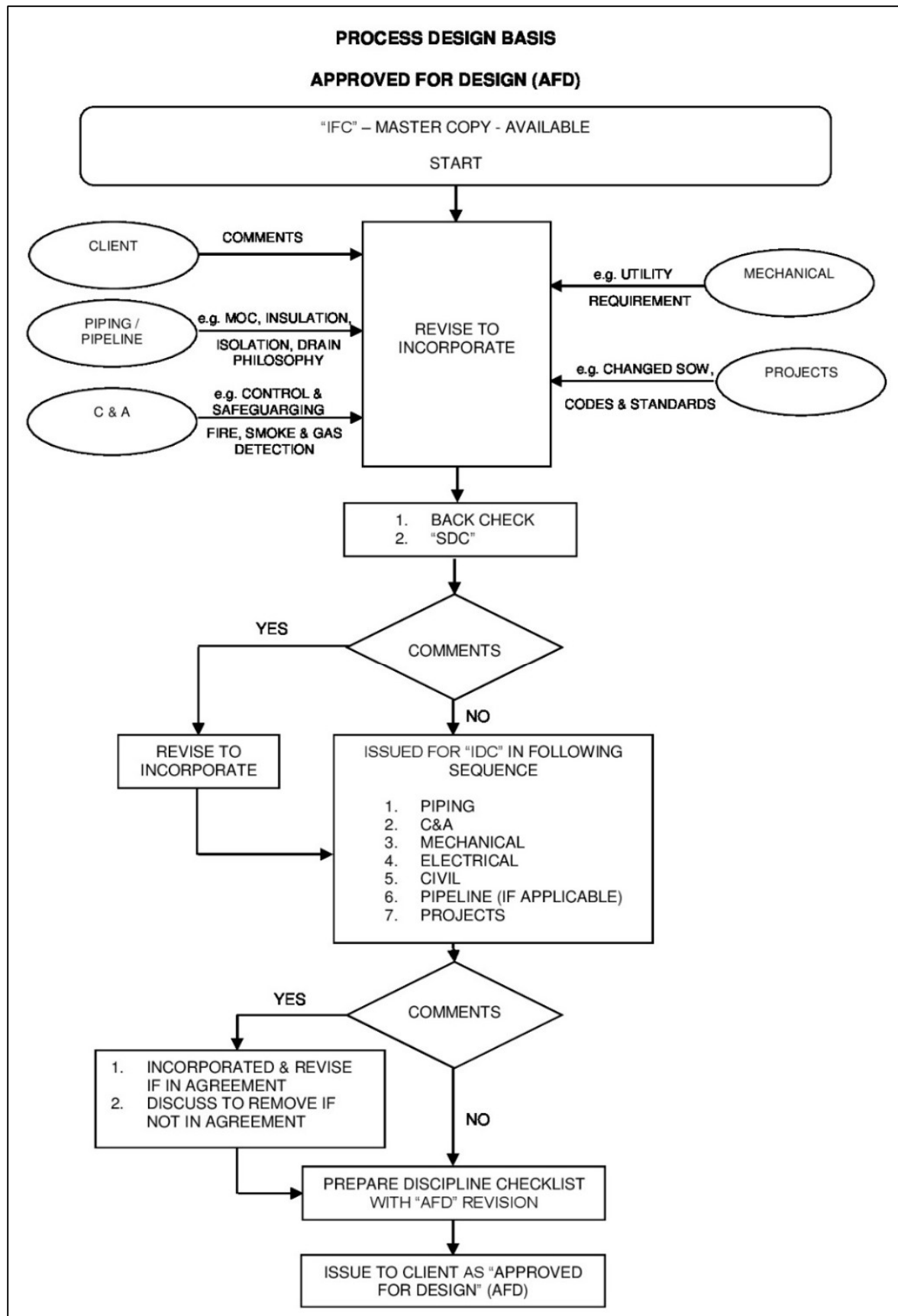
- ANSI/ISA–75.01.01–2002 (Flow equations for sizing control valves)
- ANSI/API 610 - Centrifugal Pumps for General Refinery Service
- ISO 5167-2 First Edition 2003-03-01: Orifice plates.

2. PROCESS DOCUMENTS

2.1 PROCESS DESIGN BASIS

Refer the flow chart below for preparing a process design basis document. The flowchart provides generic guidelines for the structure of and contents in the process design basis document. However, the process design basis needs to be prepared on a case-to-case basis depending on the type of project such as for oil & gas, petrochemicals, chemical storage and transport terminals, industrial water treatment plants etc. Some guidelines on information to be provided in the process design basis is given in the subsequent sections.





Document Revision Status

- 1st issue - Issued For Comments (IFC).
- 2nd issue – Approved for Design (AFD).

Responsibilities

- Prepared: Designated Process Engineer.
- Checked: Designated Senior Process Engineer and / or Lead Process Engineer.
- Approved: Discipline Head – Process.

Effect on other Documents / Drawings / Deliverables

- Document Register.
- PFD
- P&ID.
- Equipment List.
- Pump Hydraulic Calculations.
- Heat Exchanger Thermal Design.
- Equipment Datasheets.
- Hazardous Area Classification.
- Relief Valve Sizing & Datasheets
- Electrical Load List

Quality Check

- Definitions
 - SDC: It is defined as “single or self-discipline check”, where the person from the concerned discipline who is generating the document will first self-check his or her own document for correctness. The second part of SDC includes a peer check (e.g. Senior / Lead Engineer / Discipline Manager) from the same discipline for verifying the correctness and quality of the document.
 - IDC: It is defined as “Inter-Discipline Check” where other disciplines such as “Piping”, “Instrumentation”, and “Mechanical”, “Electrical” etc. will check the document from their perspective and ensure that any information related to their discipline is correct, and make corrections if required by marking up the document.
- IDC check for PDB
 - Process design basis shall have IDC check duly stamped, a scheduled return date and sent to other department lead engineers (LEs) for review as per master IDC matrix.
 - IDC comments to be duly incorporated. Any comments requiring clarification need to be discussed with the concerned LE and then incorporated.
- SDC check for PDB
 - Process design basis shall have SDC check duly stamped and signed before issue.
 - Signed process design basis is to be handed over to Projects for further distribution and original to be taken back.
- Checklist for PDB

- It is a list that guides the engineers to perform a check of the quality of the document and ensure that major criteria for process design are covered. A PDB checklist will be specific to a specific type of project. Many companies have their own checklist form specific to their projects. ***As a homework problem, you are requested to prepare a checklist for the PDB.***

2.2 OPERATING AND CONTROL PHILOSOPHY

Scope

The purpose of this document is to establish the guidelines for preparation of Operating, Control and Safeguarding Philosophy (OCSF).

Input Document

Following input documents are required during preparation of Operating, Control and Safeguarding Philosophy:

- Basis of Design – Process Description
- Piping & Instrument Diagrams (P&IDs)
- Process Safeguarding Flow Scheme (PSFS) (only for PDO projects)

Procedure

OCSF is a Microsoft Word document.

Structure

The OCSF is a structured document with a particular narrative style. The following is the sequence of the document:

- Cover sheet with project title and document name i.e. OCSF
- 1st sheet with project title, document name i.e. OCSF, Engineering Consultant job No., Document No., Revision block (this contains Revision, Date, Originator, Checker, Approver, Description, Client Authorization).
- List of Contents which includes:
 - Summary
 - Abbreviations and Definitions
 - Introduction which includes 'Document Purpose' and 'Project Background'
 - Operating Philosophy
 - Control Philosophy
 - Safeguarding Philosophy
 - References
 - Appendix: Documents / Drawings for Reference

Summary

This section provides a brief outline of the project.

Abbreviations and Definitions

The abbreviations and definitions of terms used in the entire document are summarized here.

Introduction

The first sub-section under this heading provides what is the purpose of the document and whether it is linked to any other document and whether this document needs to be read in conjunction with any other document.

The second sub-section under this heading provides the project background giving a brief overview of the project. This section essentially describes the basic plant / unit operations in short to provide an overall picture of the process.

Operating Philosophy

In this section the operating philosophy of the plant / unit is described. This section describes in further detail the operational aspect of the plant / unit.

The mode and sequence of operations is mentioned. Equipment description, tag numbers, normal operating parameters of flow, pressure and temperature and stand-by equipment availability may be mentioned in this section. The description provided in this section should be sufficient to cover all the units in the plant including utilities and off-sites.

Control Philosophy

This section provides the description of all controls required for the safe, reliable and uninterrupted operation of the plant / unit. This could include flow, pressure, and temperature and level control of the plant / unit for the smooth operation of the plant / unit. Controls required for start-up, planned shutdown and to change plant / unit capacity should be mentioned.

High and Low alarms for process operating parameters are also described in this section.

Safeguarding Philosophy

This section provides the description of the safeguarding functions that may be available for the following:

- Prevention of uncontrolled process (e.g. run-away reaction leading to unacceptable pressure and / or temperature rise, overflow from equipment such as tanks and vessels, build-up of static electricity).
- Risk reduction or mitigation in case of external hazards such as external fire, water flooding, earthquakes. Among the external hazards mentioned external fire is one of the most commonly encountered in plants / units handling hydrocarbons. Other hazards may be applicable on a Case-to-case basis.
- Manually initiated emergency shutdown of the plant / unit by operator intervention.

In this section a brief description of all safeguarding instruments / safety interlocks should be provided which can prevent or mitigate any of the above. Some examples would be:

- Shutdown of pump due to low suction pressure and / or high discharge pressure.
- Overpressure / vacuum protection of tanks and vessels by relief devices such as pressure Safety valve, rupture disks, pressure-vacuum relief valve.
- Closure of process shutdown (PSD) or emergency shutdown (ESD) valves to isolate plant / Unit / section by initiation of remote process shutdown or emergency shutdown.
- Stop of rotating machinery due to excessive vibrations, low level and / or flow of lube oil, mechanical seal failure, and over-current.

The above are just a few typical examples and the actual description of the safeguarding functions will be project specific.

References

References would include reference to an existing Plant Operating Manual and / or OCSP if the project relates to de-bottlenecking / revamp / modification of an existing plant / unit.

Appendix

This section will contain a list of documents and drawings in a tabular form utilized in the generation of the OCSP. The table should include the document / drawing number, title or description and the revision. Typical examples would be the 'Basis of Design', Piping and Instrument Diagrams (P&ID's), Process Safeguarding Flow Scheme (PSFS).

Interaction with Others

- C&A discipline for obtaining any clarifications on the control and safeguarding systems
- Designated Lead / Senior Process Engineer / Process Discipline head for resolving any queries and clarification of any doubts

Output Document

- Operating, Control and Safeguarding philosophy

Document Revision Status

- 1st Issue – Issued for comments (IFC)
- 2nd Issue – Approved for Design (AFD)

Responsibilities

The responsibilities for the generation and approval of the document shall be as follows:

- Prepared: Designated Process Engineer
- Checked: Lead Process Engineer
- Approved: Discipline Head – Process

Effect on other Documents / Drawings / Deliverables

The following documents / drawings would be impacted by the OCSP:

- Cause and Effect diagrams
- Document Register

Quality Checks

- SDC and IDC Check
 - OCSP shall be subjected to SDC check and duly signed on SDC check stamp by the originator and Process Lead engineer before IDC.
 - After SDC, OCSP will have to undergo IDC check by C&A, the IDC stamp being duly signed by C&A Lead engineer and the project engineer.
- Check List

The OCSP document shall be checked against the standard checklist for any errors and omissions and if necessary to be corrected. The checklist and the document shall be signed off before distribution.

2.3 LINE LIST

A line list (or line schedule) is a database created to communicate between the process and mechanical engineering teams when designing piping in a plant or process unit. At its core, it should help provide justification on why certain decisions were made in the design process. If the maximum pressure is 200 psi but the piping code is an ANSI 600 system, it should be documented on the line list. At a minimum, the information on a line schedule is as follows:

- Line Identifier (Line Designation, Line Number) - a unique number that identifies the line in the plant or process unit. The author of this article cannot stress enough the use of the word unique.
- Start and ending point of the line - includes equipment connections or connections to other lines.
- Service or commodity - materials that are flowing through the piping
- Fluid Phase - Liquid, vapor, two phase, etc.
- Piping Specification and/ or piping code
- Design and operating pressure and temperature of the service
- P&ID reference numbers
- Piping Isometric reference number
- Corrosion Allowance
- Insulation type and thickness
- Heat Tracing Type
- Special calculation requirements such as pipe stress
- Special Information & requirements such as post weld heat treatment, pickling, hydro test requirements, etc.

Line ID	Start	End	Service	Phase	Design		Operating		Insulation	Heat Tracing	Drawings		Notes
					Press	Temp	Press	Temp			P&ID	Isometric	
10-6"-WW-4867-A-040-A-2"HC-ST	10-6"-WW-4866-A-040-A-2"HC-ST	10-6"-WW-4868-A-040-A-2"HC-ST	Waste Water	Liquid	200	150	250	200	2"	Steam	PD-25091	ISO-10489	PWHT & Stress Calc Required.
10-6"-WW-4868-A-040-A-2"HC-ST	10-6"-WW-4867-A-040-A-2"HC-ST	V-202A	Waste Water	Liquid	200	150	250	200	1"	N/A	PD-25091	ISO-10490	

		LINE LIST CHECK LIST SHEET 1 OF 1 SHEET		αβχ	
JOB No:					
PROJECT:					
DOCUMENT NUMBER					
DOCUMENT TITLE					
REVISION	DATE	PREPARED BY	CHECKED BY	APPROVED BY	

CHECK LIST	YES	NO	N/A	COMMENTS	REV
1. Correct drawing titles and revision	☐	☐	☐		
2. Drawing number included	☐	☐	☐		
3. Consultant standard title block and Job Number	☐	☐	☐		
4. Consultant Disclaimer description block	☐	☐	☐		
5. Standard document front sheet included	☐	☐	☐		
6. Standard Holds List and action party (in brackets)	☐	☐	☐		
7. All lines/equipment included as per P&ID/PDS	☐	☐	☐		
8. All lines/equipment numbering matches P&ID/PDS	☐	☐	☐		
9. All design information included and as per P&ID/PFD	☐	☐	☐		
10. Operating and design conditions clearly specified	☐	☐	☐		
11. For line list, "from" and "to" locations match with P&ID	☐	☐	☐		
12. Requirement for references and are they clearly specified	☐	☐	☐		
13. Check that references to P&ID/PDS match with the appropriate drawing/document	☐	☐	☐		
14. SDC comments incorporated	☐	☐	☐		
15. IDC comments incorporated	☐	☐	☐		
16. Client comments incorporated	☐	☐	☐		

2.4 PROCESS DATASHEET

The process data sheets (PDS) are generally A4 documents with number of sheets. A PDS relates to a single item of equipment and contains the essential process data for initiating the detail design of an item. It includes the overall size, number, approximate geometry and identification of the connections, material of construction and the full range of operating conditions. The process data sheet generally includes a simple diagram. The process design also generally includes for Instrument process datasheets which provides a similar level of detail. Refer the example below:

		PROJECT NAME				ABC CONSULTANTS	
		PROCESS DATASHEET FOR STORAGE TANK				Project	
						Doc. No.	DS-R-001
						Sheet No.	2 of 3
1	GENERAL						
2	Client						
3	Plant/Unit						
4	Location						
5	Item No.						
6	Service						
7	No Req'd	Working		Standby		Total	
8	Rev.	PROCESS DATA				STORAGE TANK DATA	
9	Pressure	Top	mmH ₂ O		Dimension (mm)	Diameter	
10						T-T Length	
11	Temp.	Min.	°C		Volume (m ³)	Nominal	
12		Nor.	°C			Working	
13	Test Pressure	Hydrostatic	barg		Design and Constr. Code		
14		Pneumatic	barg		Roof Type		
15	Ambient Temp.	Min	°C		Bottom Type		
16		Max.	°C		Slope	Roof	
17	Fluid Handled				Bottom		
18	Density Liquid	kg/m ³			Requirement of Agitator/Mixer		
19	Density Vapour	kg/m ³			Insulation (Cold/Hot/PP/No)		
20	Vapour Pressure	bar			Fire Proofing (Yes/No)		
21	Viscosity	cP			Tank (Heating/ Cooling)		
22	Pump In Rate	m ³ /hr			Tank Gauging Required (Yes/No)		
23	Pump Out Rate	m ³ /hr			Vortex Breaker Required (Yes/No)		
24	Blanketing Gas Press.	barg			Tank (Heating/ Cooling)		
25	Flash Point/Pour Point (if any):						
26	Corrosive (Yes/No)						
27	MECHANICAL DESIGN CONDITION				MATERIAL SPECIFICATION		
28	Pressure	Internal	mmH ₂ O		Material		Corrosion Allowance (mm)
29		External	barg		Shell		
30	Temperature		°C		Roof		
31					Bottom		
32	Notes:						
33							
34							
35							
36							
37							
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	PROJECT NAME	ABC CONSULTANTS	
	PROCESS DATASHEET FOR STORAGE TANK	Project	
		Doc. No.	DS-R-001
		Sheet No.	3 of 3

Nozzle Connection Summary				
Rev.	Mark	Quantity	Size (in)	Services
				Liquid In
				Liquid Out
				Drain
				Overflow
				Vent
				Minimum Flow Return
				Manhole 1
				Manhole 2
				LT Connection 1
				LT Connection 2
				TG Connection
				Pressure Connection
				Safety Valve
				Stand Pipe
				Level Switches
				Ventilation Nozzel
				Blanketing Connection In
				Blanketing Connection Out

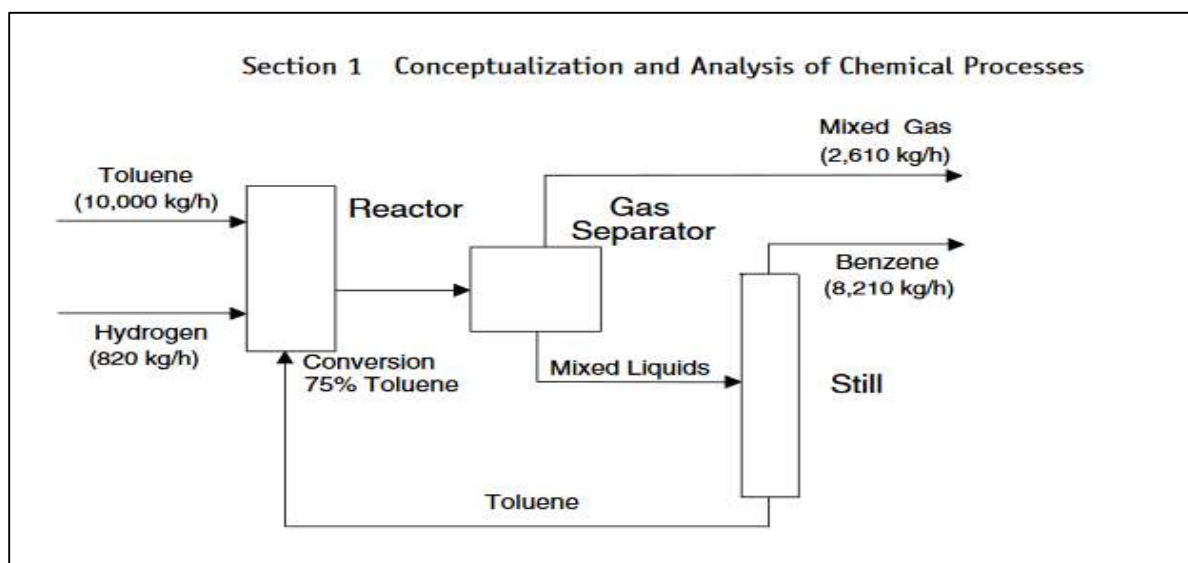
SKETCH

Notes:
1) Sizes of all instruments will be finalised during detail engineering .

3 PROCESS DRAWINGS

3.1 BLOCK FLOW DIAGRAM (BFD)

This is to show the essentials of an installation in sufficient block outline to indicate the main design features. This will use line and simple geometric shape indicating major material and energy flows. Figure 1:



3.2 PROCESS FLOW DIAGRAM (PFD)

A process flow diagram (PFD) is a diagram commonly used in engineering to indicate the general flow of plant processes and equipment. The PFD displays the relationship between major equipment of a plant facility and does not show minor details such as piping details and designations. Another commonly-used term for a PFD is a flow sheet.

PFD is normally finalized during the FEED stage and is generally issued only once during DD after incorporating vendor information. Typically, process flow diagrams of a single unit process will include the following:

- Process piping, major streams
- Major bypass & recirculation lines
- Major equipment symbols, names and identification numbers; e.g. for tanks/ vessels their type and orientation, for rotating equipment such as pumps / compressors their type / capacity / head, for distillation columns the number of trays or packing height
- Flow directions
- Control loops that affect the operation
- Interconnections with other systems
- Stream Identification using numbers and / or alphabets

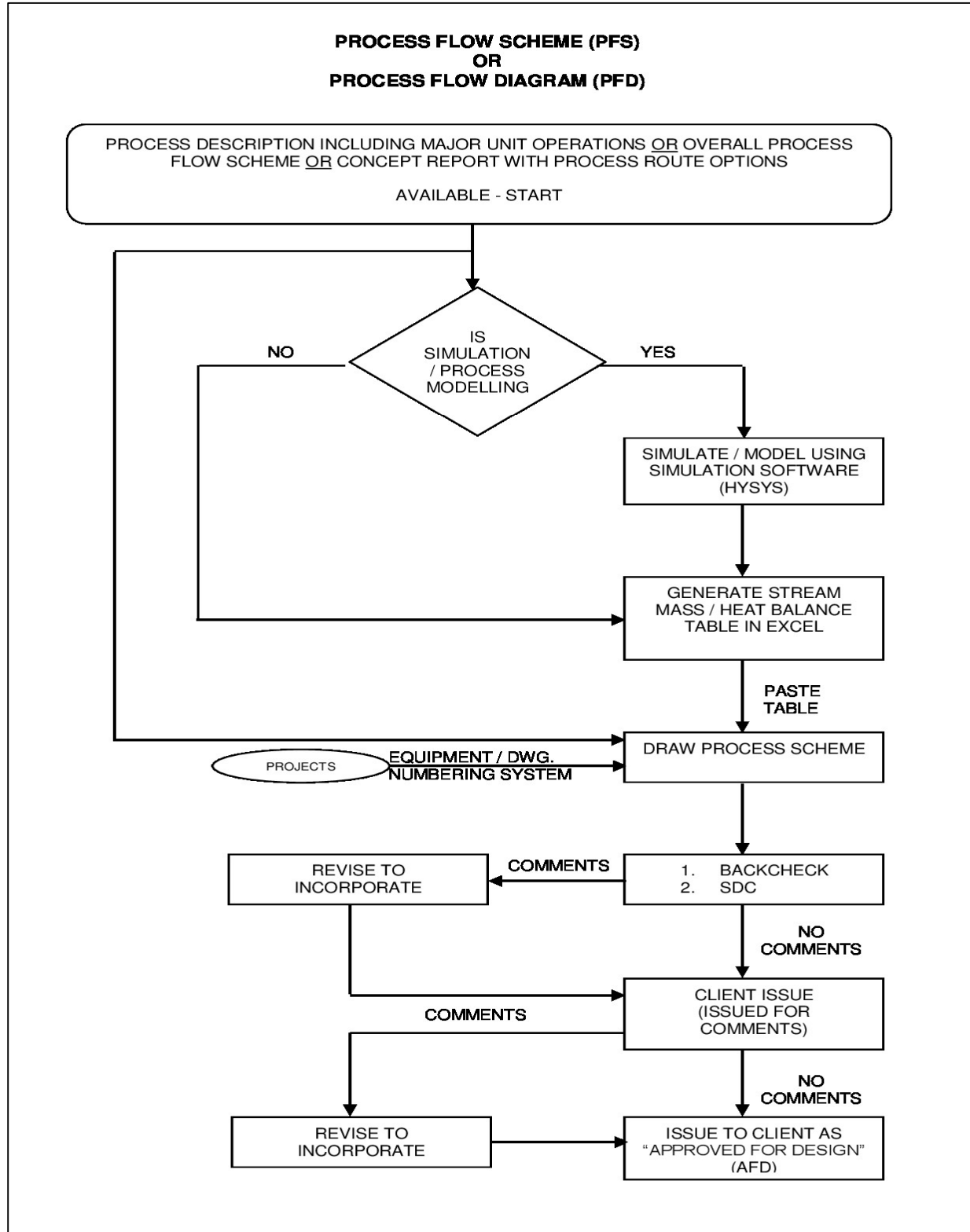
- Material or Mass Balance table for the identified streams along with operating pressure, temperature and composition. In some cases properties such as density and viscosity are also provided for the stream.

PFDs generally do not include the following:

- Piping classes or piping line numbers
- Line size
- Instrumentation not used for plant or unit control (non-critical instrumentation)
- Minor bypass lines
- Isolation and shut-off valves
- Maintenance vents and drains
- Relief and Safety Devices
- Pipe fittings and flanges

Typical mass balance and PFD schematic are shown on the next two pages

FLOW CHART FOR PREPARATION OF PROCESS FLOW DIAGRAM (PFD)



Heat & Material Balance Sheet

NFIS : 117 deg C (Summer) (Note 3): PHASE I									
NFIS suction 90 barg									
Mixture mole fraction									
Methane	7.84E-01	7.84E-01	7.84E-01	7.84E-01	7.84E-01	7.84E-01	7.84E-01	7.84E-01	7.84E-01
Ethane	1.13E-02	1.13E-02	1.13E-02	1.13E-02	1.13E-02	1.13E-02	1.13E-02	1.13E-02	1.13E-02
Propane	3.39E-03	3.39E-03	3.39E-03	3.39E-03	3.39E-03	3.39E-03	3.39E-03	3.39E-03	3.39E-03
i-Butane	9.94E-04	9.94E-04	9.94E-04	9.94E-04	9.94E-04	9.94E-04	9.94E-04	9.94E-04	9.94E-04
n-Butane	1.29E-03	1.29E-03	1.29E-03	1.29E-03	1.29E-03	1.29E-03	1.29E-03	1.29E-03	1.29E-03
i-Pentane	6.89E-04	6.89E-04	6.89E-04	6.89E-04	6.89E-04	6.89E-04	6.89E-04	6.89E-04	6.89E-04
n-Pentane	1.48E-03	1.48E-03	1.48E-03	1.48E-03	1.48E-03	1.48E-03	1.48E-03	1.48E-03	1.48E-03
n-Hexane	9.50E-04	9.50E-04	9.50E-04	9.50E-04	9.50E-04	9.50E-04	9.50E-04	9.50E-04	9.50E-04
n-Heptane	1.21E-03	1.20E-03	1.20E-03	1.20E-03	1.20E-03	1.20E-03	1.20E-03	1.20E-03	1.20E-03
n-Octane	3.12E-04	3.06E-04	3.06E-04	3.06E-04	3.06E-04	3.06E-04	3.06E-04	3.06E-04	3.06E-04
n-Nonane	2.60E-04	2.50E-04	2.50E-04	2.50E-04	2.50E-04	2.50E-04	2.50E-04	2.50E-04	2.50E-04
n-Decane	1.46E-04	1.35E-04	1.35E-04	1.35E-04	1.35E-04	1.35E-04	1.35E-04	1.35E-04	1.35E-04
n-C11	6.71E-05	5.83E-05	5.83E-05	5.83E-05	5.83E-05	5.83E-05	5.83E-05	5.83E-05	5.83E-05
n-C12	3.53E-05	2.67E-05	2.67E-05	2.67E-05	2.67E-05	2.67E-05	2.67E-05	2.67E-05	2.67E-05
n-C13	1.48E-05	8.67E-06	8.67E-06	8.67E-06	8.67E-06	8.67E-06	8.67E-06	8.67E-06	8.67E-06
n-C14	7.89E-06	3.65E-06	3.65E-06	3.65E-06	3.65E-06	3.65E-06	3.65E-06	3.65E-06	3.65E-06
n-C15	4.15E-06	1.92E-06	1.92E-06	1.92E-06	1.92E-06	1.92E-06	1.92E-06	1.92E-06	1.92E-06
n-C16	2.15E-06	9.50E-07	9.50E-07	9.50E-07	9.50E-07	9.50E-07	9.50E-07	9.50E-07	9.50E-07
n-C17	1.09E-06	4.85E-07	4.85E-07	4.85E-07	4.85E-07	4.85E-07	4.85E-07	4.85E-07	4.85E-07
n-C18	4.04E-07	1.66E-07	1.66E-07	1.66E-07	1.66E-07	1.66E-07	1.66E-07	1.66E-07	1.66E-07
n-C19	2.34E-07	9.35E-08	9.35E-08	9.35E-08	9.35E-08	9.35E-08	9.35E-08	9.35E-08	9.35E-08
n-C20	1.33E-08	5.35E-09	5.35E-09	5.35E-09	5.35E-09	5.35E-09	5.35E-09	5.35E-09	5.35E-09
H2O	2.50E-03	2.50E-03	2.50E-03	2.50E-03	2.50E-03	2.50E-03	2.50E-03	2.50E-03	2.50E-03
Nitrogen	1.52E-01	1.52E-01	1.52E-01	1.52E-01	1.52E-01	1.52E-01	1.52E-01	1.52E-01	1.52E-01
CO2	3.79E-02	3.79E-02	3.79E-02	3.79E-02	3.79E-02	3.79E-02	3.79E-02	3.79E-02	3.79E-02
H2S	1.80E-03	1.80E-03	1.80E-03	1.80E-03	1.80E-03	1.80E-03	1.80E-03	1.80E-03	1.80E-03
Mixture									
Vapour Fraction	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Temperature	60.12	60.12	60.12	60.12	60.12	60.12	60.12	60.12	60.12
Pressure	90.00	90.00	90.00	90.00	90.00	90.00	90.00	90.00	90.00
Mass Flow	98558.52	98494.69	98494.69	98494.69	98494.69	98494.69	98494.69	98494.69	98494.69
Molecular Weight	19.75	19.74	19.74	19.74	19.74	19.74	19.74	19.74	19.74
Heat Flow	-106.36	-103.49	-103.49	-103.49	-103.49	-103.49	-103.49	-103.49	-103.49
Mass Heat Capacity	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29
Mass Density	71.24	70.83	70.83	70.83	70.83	70.83	70.83	70.83	70.83
***** Vapour *****									
Std Gas Flow	100.00	99.99	99.99	99.99	99.99	99.99	99.99	99.99	99.99
Actual Volume Flow	1383.37	1380.57	1380.57	1380.57	1380.57	1380.57	1380.57	1380.57	1380.57
Molecular Weight	19.74	19.74	19.74	19.74	19.74	19.74	19.74	19.74	19.74
Mass Heat Capacity	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29
Mass Density	71.20	70.83	70.83	70.83	70.83	70.83	70.83	70.83	70.83
Compressibility	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91
Thermal Conductivity	0.04	0.04	0.04	0.04	0.04	0.04	0.04	0.04	0.04
Viscosity	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
***** Light Liquid *****									
Mass Flow	60.12	60.12	60.12	60.12	60.12	60.12	60.12	60.12	60.12
Standard Ideal Liquid Volume Flow	0.08	0.08	0.08	0.08	0.08	0.08	0.08	0.08	0.08
Actual Volume Flow	0.09	0.09	0.09	0.09	0.09	0.09	0.09	0.09	0.09
Molecular Weight	110.46	110.46	110.46	110.46	110.46	110.46	110.46	110.46	110.46
Mass Heat Capacity	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29
Mass Density	676.26	676.26	676.26	676.26	676.26	676.26	676.26	676.26	676.26
Thermal Conductivity	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10
Viscosity	0.45	0.45	0.45	0.45	0.45	0.45	0.45	0.45	0.45
Heavy Liquid									
Liquid Vol Flow @Std Cond	-	-	-	-	-	-	-	-	-
Mass Density	-	-	-	-	-	-	-	-	-
Mass Flow	-	-	-	-	-	-	-	-	-
Mass Heat Capacity	-	-	-	-	-	-	-	-	-
Molecular Weight	-	-	-	-	-	-	-	-	-
REFERENCE DRAWINGS:									
GENERAL NOTES:									
1. FOR DRAFTING STANDARD REFER TO QP ESD-10 Rev.6									
2. THIS NEW MATERIAL BALANCE IS PREPARED FOR HUFF GAS FEED FROM WHIP-H & L.									
3. THIS IS FEED GAS TEMPERATURE.									
4. DELETED.									
THIS ENTIRE DRG. IS NEW & FORM PART OF PROJECT CODE 2712									
REV. DESCRIPTION DATE (mm) (dd) (yy) APP									
4	ISSUED FOR EPC TENDER	27/04/11	MCH	SAP	JRD				
3	ISSUED FOR APPROVAL	26/04/11	MCH	SAP	JRD				
2	ISSUED FOR HAZOP	26/04/11	MCH	SAP	JRD				
1	ISSUED FOR APPROVAL	06/07/10	MCH	VTD	SAP				
0	ISSUED FOR COMMENTS	01/07/10	MCH	VTD	SAP				

PFD CHECK LIST (PER PROJECT) SHEET 1 OF 2 SHEETS		$\alpha\beta\chi$
JOB No:		
PROJECT:		
DRAWING NUMBERS		
REVISION	DATE	PREPARED BY

CHECK LIST	YES	NO	N/A	COMMENTS	REV
GENERAL					
1. Correct drawing titles, numbers and revision	☐	☐	☐		
2. Standard Holds List and action party (in brackets)	☐	☐	☐		
3. Consultant standard title block and Job Number	☐	☐	☐		
4. Consultant Disclaimer description block	☐	☐	☐		
5. Drawing continuation flags match between drawings	☐	☐	☐		
6. Scope of project clearly defined by clouding (but use of clouding kept to a minimum)	☐	☐	☐		
7. Original Master Drawing Number listed	☐	☐	☐		
8. Specific Notes included and numbers matching	☐	☐	☐		
9. In line with Standard drawing legend	☐	☐	☐		
10. SDC comments incorporated	☐	☐	☐		
11. IDC comments incorporated	☐	☐	☐		
12. Client comments incorporated	☐	☐	☐		
EQUIPMENT AND PACKAGES					
1. All Equipment (operating and standby) included	☐	☐	☐		
2. Battery limits shown for Packages	☐	☐	☐		
3. Equipment and Package names and numbering	☐	☐	☐		
4. Design information data included and as per datasheet	☐	☐	☐		

	PFD CHECK LIST (PER PROJECT) SHEET 2 OF 2 SHEETS	αβχ
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CHECK LIST	YES	NO	N/A	COMMENTS	REV
EQUIPMENT (CONT.....)					
5. Directional arrows on major process & utility lines	☐	☐	☐	_____	
6. Line sizes shown on major lines	☐	☐	☐	_____	
PROCESS CONTROL					
1. Control valves and selection (pneumatic, motor etc.)	☐	☐	☐	_____	
2. Control valve failure positions	☐	☐	☐	_____	
3. All control loops shown including any logic	☐	☐	☐	_____	
4. All key Instrumentation shown and tagged	☐	☐	☐	_____	
5. Control logic as per Operating & Control Philosophy	☐	☐	☐	_____	
SAFEGUARDING					
1. Overpressure protection	☐	☐	☐	_____	
2. Process Shutdown Valves (PSD)	☐	☐	☐	_____	
3. Emergency Shutdown Valves (ESD)	☐	☐	☐	_____	
4. HP / LP Interfaces	☐	☐	☐	_____	
5. Non-Return valves	☐	☐	☐	_____	
6. Emergency blowdown requirements	☐	☐	☐	_____	
MISCELLANEOUS TECHNICAL					
1. Stream numbers on major process streams	☐	☐	☐	_____	
2. Stream numbers match with Process Model	☐	☐	☐	_____	
3. Heat and Mass Balance Table	☐	☐	☐	_____	
4. Future facilities (shown dotted if required)	☐	☐	☐	_____	
5. Revision triangles and Revision clouds shown on drawing updates	☐	☐	☐	_____	

3.3 PIPING & INSTRUMENT DIAGRAM (P&ID)

A P&ID is a pictorial representation of the piping of the process flow together with the installed equipment (tanks, vessels, pumps, compressors, etc.) and the instrumentation (flow, pressure, temperature, level, etc.) required to control and operate the process. Preliminary P&IDs are prepared during FEED and are further updated during Detail Design (DD). P&IDs are normally issued in 3 to 4 revisions during DD wherein information such as vendor data, Design Review, HAZOP changes and As-Built changes are incorporated.

For processing facilities, it is a pictorial representation of:

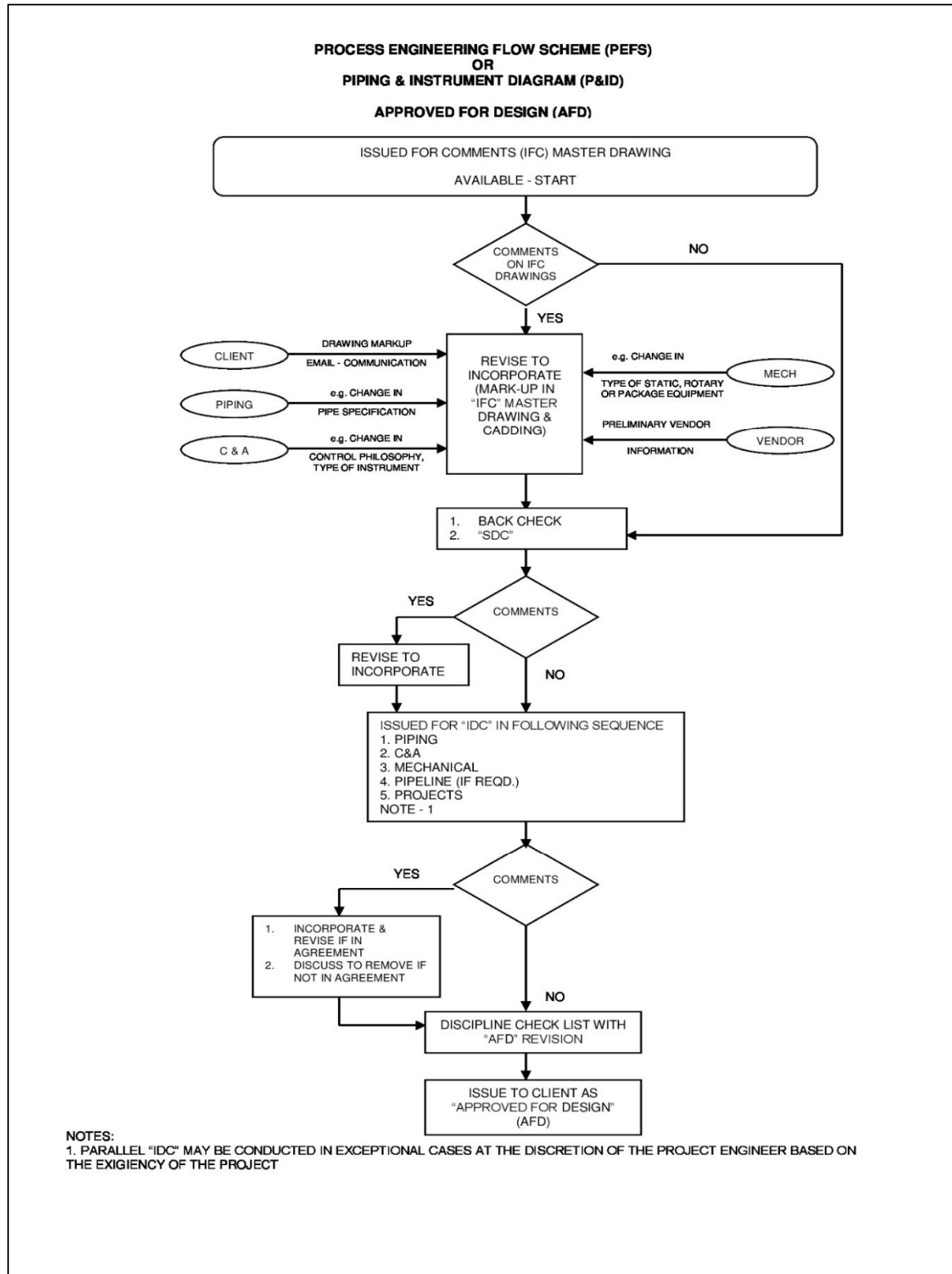
- Key piping and instrument details
- Control and shutdown schemes
- Safety and regulatory requirements
- Basic startup and operational information

The Institute of Instrumentation and Control defines a P&ID as follows:

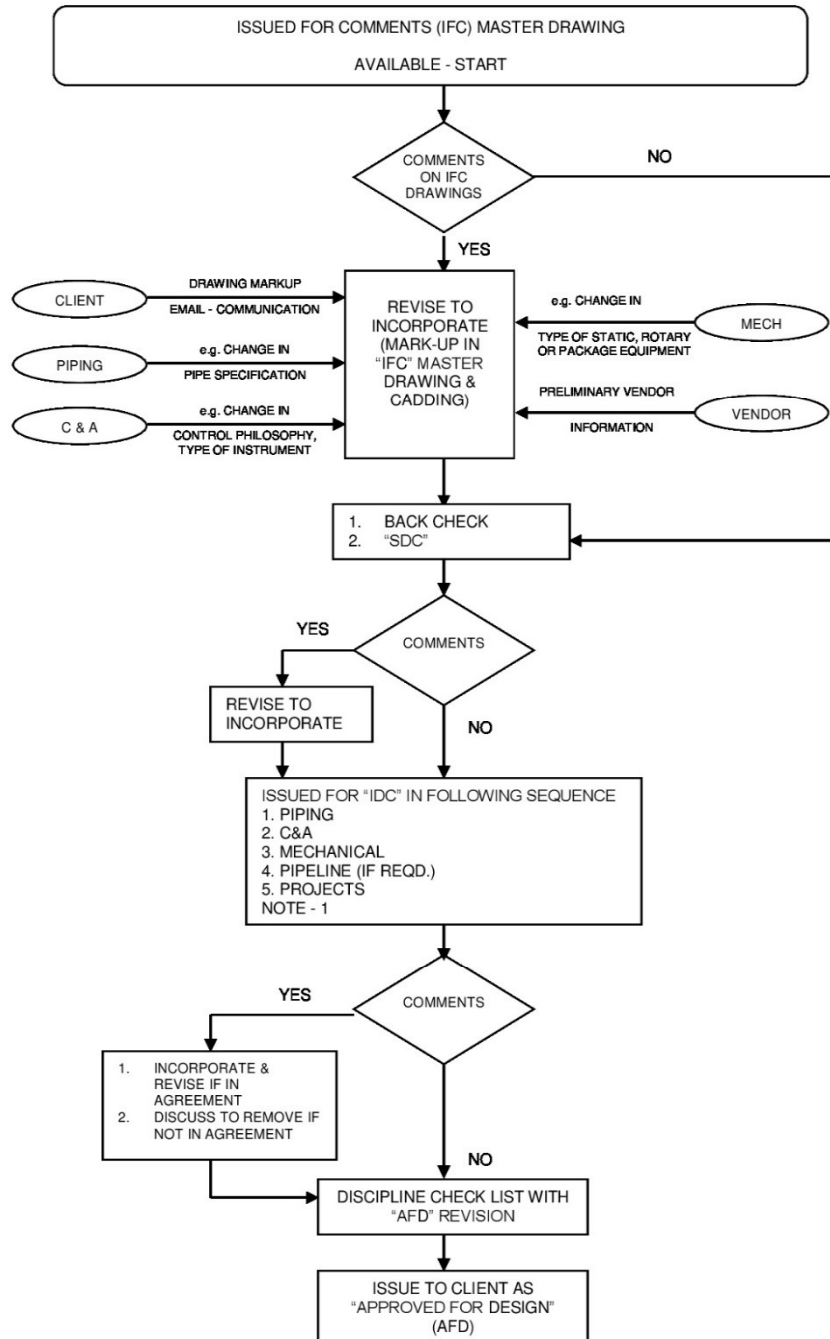
“A diagram which shows the interconnection of process equipment and the instrumentation used to control the process. In the process industry, a standard set of symbols is used to prepare drawings of processes. The instrument symbols used in these drawings are generally based on International Society of Automation (ISA) Standard S5.1”.

- CAD system is employed for developing the P&IDs and is not vendor, hardware or software specific.
- It is intended to provide a balance between the desire to show all data on P&IDs with the need to make P&IDs legible and easy to read. Most details that are available from other types of documentation (e.g., instrument loop diagrams, vessel data sheets) are not recommended for inclusion on P&IDs.
- P&ID shows equipment with simple outline representation.
- Equipment's are not drawn to scale, however equipment relative to one another both in size and general orientation has to be maintained.
- P&ID shows nozzles on equipment, including spares, as single lines.
- P&ID indicates nozzle sizes unless the size is implied by piping connections.
- P&ID indicates equipment Item Number and Title/Service as a minimum.
- P&ID indicates internals for equipment as dashed lines. Details of internals that have no significant bearing on the piping design and layout or equipment operation are omitted.
- P&ID does not indicate equipment elevations unless they are necessary to specify process requirements for associated equipment location or orientation relative to one another.
- P&ID indicates associated trim (e.g., vent and drain valves, instrument bridles) for equipment.
- P&ID indicates auxiliary system requirements for individual pieces of equipment (e.g., lube oil systems, seal flush systems, turbine gland leak-off piping, sample systems) on auxiliary P&IDs.
- P&ID indicates jacketing requirements for equipment and jacketing / tracing requirements for piping.
- P&ID indicates the type of insulation (e.g., personnel protection, heat conservation) for equipment as part of the equipment data.
- P&ID indicates insulation thickness where applicable.
- P&ID indicates piping serial number and piping class

FLOW CHART FOR PREPARATION OF PIPING & INSTRUMENT DIAGRAM



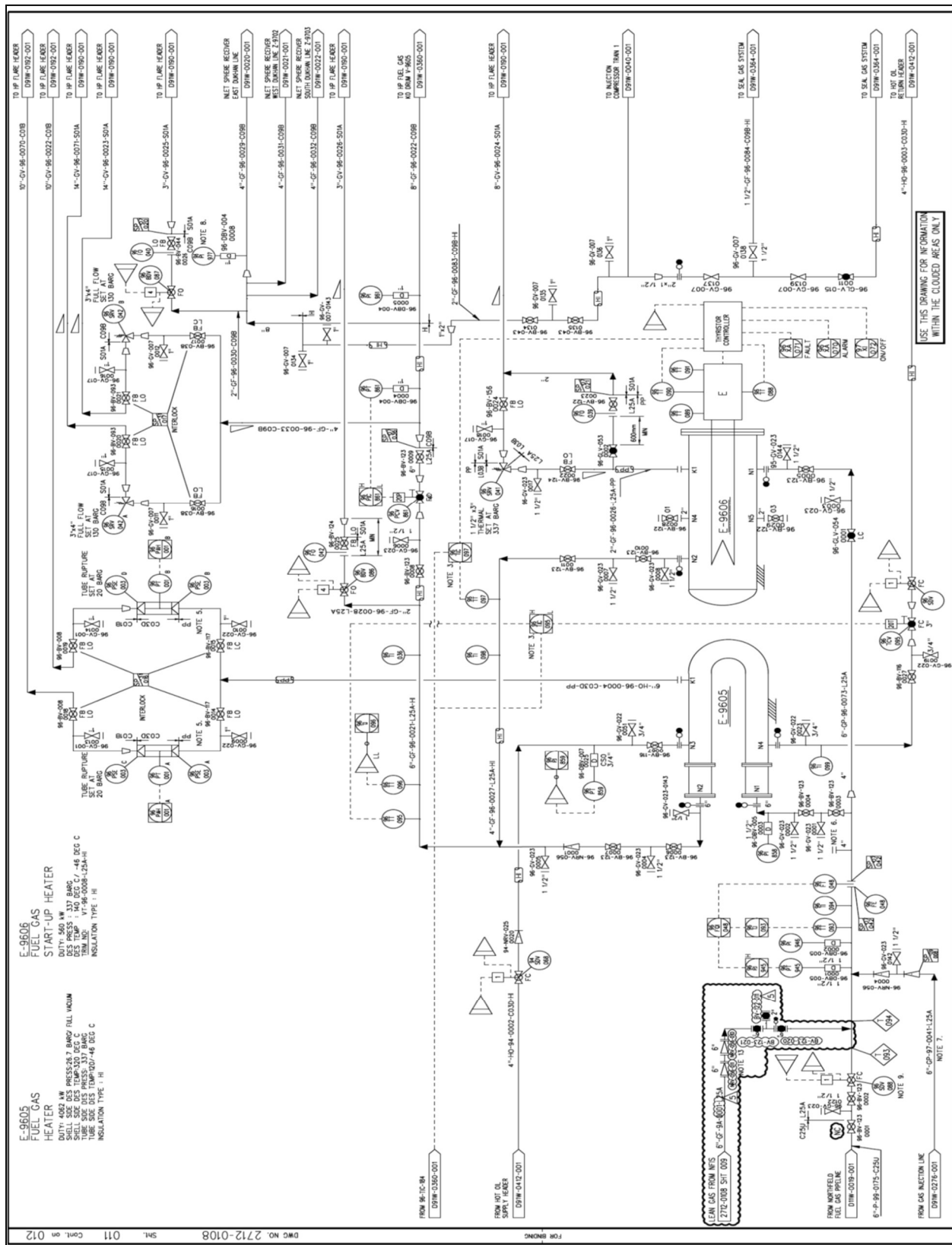
**PROCESS ENGINEERING FLOW SCHEME (PEFS)
OR
PIPING & INSTRUMENT DIAGRAM (P&ID)
APPROVED FOR DESIGN (AFD)**



NOTES:

1. PARALLEL "IDC" MAY BE CONDUCTED IN EXCEPTIONAL CASES AT THE DISCRETION OF THE PROJECT ENGINEER BASED ON THE EXIGENCY OF THE PROJECT

A typical P&ID is shown:



		P&ID CHECK LIST (PER PROJECT)			
		SHEET 1 OF 3 SHEETS		αβχ	
JOB No:					
PROJECT:					
DRAWING NUMBERS					
REVISION	DATE	PREPARED BY	CHECKED BY	APPROVED BY	

CHECK LIST	YES	NO	N/A	COMMENTS	REV
GENERAL					
1. Correct drawing titles, numbers and revision	☐	☐	☐		
2. Standard Holds List and action party (in brackets)	☐	☐	☐		
3. Consultant standard title block and Job Number	☐	☐	☐		
4. Consultant Disclaimer description block	☐	☐	☐		
5. Construct and Demolition drawing numbers match	☐	☐	☐		
6. Drawing continuation flags match between drawings	☐	☐	☐		
7. Scope of project clearly defined by clouding (but use of clouding kept to a minimum)	☐	☐	☐		
8. Original Master Drawing Number listed	☐	☐	☐		
9. Specific Notes included and numbers matching	☐	☐	☐		
10. In line with Standard drawing legend	☐	☐	☐		
11. SDC comments incorporated	☐	☐	☐		
12. IDC comments incorporated	☐	☐	☐		
13. Client comments incorporated	☐	☐	☐		
EQUIPMENT					
1. All Equipment (operating and standby) included	☐	☐	☐		
2. Equipment names and numbering shown	☐	☐	☐		
3. Equipment design information data included and as per datasheet	☐	☐	☐		

	P&ID CHECK LIST (PER PROJECT)	
	SHEET 2 OF 3 SHEETS	αβχ

CHECK LIST	YES	NO	N/A	COMMENTS	REV
EQUIPMENT (CONT.....)					
4. Equipment and internals/externals as per datasheet	☐	☐	☐	_____	
5. Nozzles, manways and numbering as per datasheet	☐	☐	☐	_____	
6. Insulation and Trace heating	☐	☐	☐	_____	
7. Draining, Venting and Purging	☐	☐	☐	_____	
8. Equipment Isolation including valves and spades	☐	☐	☐	_____	
9. Overpressure protection, relief valves, bursting discs	☐	☐	☐	_____	
10. Equipment trim numbers	☐	☐	☐	_____	
PACKAGES					
1. Package names and numbering	☐	☐	☐	_____	
2. Package information data blocks completed	☐	☐	☐	_____	
3. Piping & Instrument Connections as per datasheet	☐	☐	☐	_____	
4. Battery limits shown	☐	☐	☐	_____	
5. References to Supplier PEFS's included	☐	☐	☐	_____	
PIPING					
1. Directional arrows on major process & utility lines	☐	☐	☐	_____	
2. Line sizes on all lines	☐	☐	☐	_____	
3. Line numbers on all lines including pipe specs	☐	☐	☐	_____	
4. Requirement for slopes and avoiding pockets shown	☐	☐	☐	_____	
5. Tie-In Point numbers including spec breaks	☐	☐	☐	_____	
6. Tie-in points between Construct and Demolition drawings match	☐	☐	☐	_____	
7. Insulation and Trace heating	☐	☐	☐	_____	
8. Pipe specification breaks shown	☐	☐	☐	_____	
9. Piping special items included and numbered	☐	☐	☐	_____	
10. Overpressure protection, TRV's, bursting discs	☐	☐	☐	_____	
11. Valve selection in line with Pipe Specs	☐	☐	☐	_____	

P&ID CHECK LIST (PER PROJECT)				αβχ	
SHEET 3 OF 3 SHEETS					
12. Draining, Venting and Isolation requirements		☐	☐	☐	
CHECK LIST		YES	NO	N/A	COMMENTS
REV					
PIPING (CONT.....)					
13. Sample points and corrosion monitoring		☐	☐	☐	
PROCESS CONTROL AND INSTRUMENTATION					
1. Correct control valves selection (pneumatic, motor etc.)		☐	☐	☐	
2. Control valves sizes, tag numbers & failure positions		☐	☐	☐	
3. Control valve isolation block valves & drains/vents		☐	☐	☐	
4. Control valve bypasses (to be avoided where possible)		☐	☐	☐	
5. All control loops shown including any logic		☐	☐	☐	
6. All Instrumentation shown and as per PFS including tag numbers		☐	☐	☐	
7. Alarm and trip set points		☐	☐	☐	
8. Vessel level alarm and trip set points		☐	☐	☐	
9. Tapping size on equipment/piping		☐	☐	☐	
SAFEGUARDING					
1. Relief valves and numbering		☐	☐	☐	
2. Relief valve set pressures and line sizes		☐	☐	☐	
3. Relief valve sparing, isolation and interlocks		☐	☐	☐	
4. Emergency shutdown valves and numbering		☐	☐	☐	
5. Non-Return valves		☐	☐	☐	
6. Emergency blowdown requirements		☐	☐	☐	

4 DETAILED DESIGN CALCULATIONS

The training manual covers the basic information required for a new or trainee process engineer to understand the minimum scope of Process Engineering in terms of activities done, design calculations performed and deliverables generated by the process engineer. It is important to note that Process Engineering is not limited to the activities, calculations and deliverables mentioned in this training manual. The manual in its present form does not cover the entire scope of Process Engineering. However, the most basic Process Engineering activities and deliverables are covered for any trainee process engineer to start with. The following topics have been covered:

- ❖ Heat & Material Balance
- ❖ Determination of Process Conditions (Design Pressure / Temperature)
- ❖ Emergency depressurization Systems
- ❖ Pressure Protection and Overpressure Relief Devices
- ❖ Line Sizing
- ❖ Storage tanks
- ❖ Venting of Atmospheric and Low-Pressure Storage Tanks
- ❖ Control Valves (Design Guidelines)
- ❖ Orifice Sizing
- ❖ Pumps
- ❖ Fractionation Column
- ❖ Separator Sizing
- ❖ Heat Exchanger
- ❖ Flare System
- ❖ Software Tools

4.1 HEAT & MATERIAL BALANCE (H&MB)

Heat Balance

A heat balance or in other words energy balance is essentially following the first law of Thermodynamics which says that energy can neither be created nor destroyed, only modified in form. In its most simplified form a heat balance for transfer of sensible heat can be written as:

$$Q_{\text{sensible}} = m \cdot C_p \cdot \Delta T$$

If the heat released or absorbed involves a phase change (solid $\leftrightarrow$ liquid or liquid $\leftrightarrow$ vapor) then the heat balance introduces the term latent heat of melting / solidification for phase change from solid to liquid and vice versa and latent heat of vaporization / condensation for phase change from liquid to vapor and vice versa.

$$Q_L = m \cdot \lambda$$

If the process includes sensible heat transfer and phase change then the total heat transfer would be:

$$Q_{\text{total}} = Q_{\text{sensible}} + Q_L$$

The terms in the equations provided above are:

$$Q_{\text{sensible}} / Q_L / Q_{\text{total}} = \text{heat flow rate, kJ/h}$$

m = mass flow rate, kg/h

C_p = Specific heat, kJ / kg-K

ΔT = Temperature Difference from Initial State to Final State, K

λ = Latent heat, kJ / kg

Mass Balance

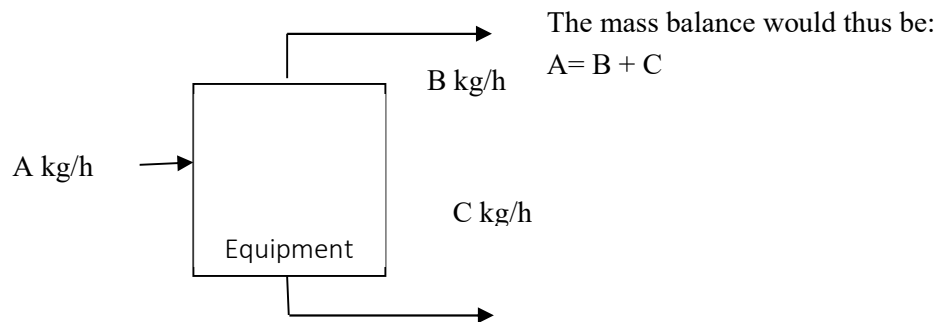
A mass balance applies the law of conservation of mass and in its simplest form can be written as:

Input = Output + Accumulation

For non-reactive systems the mass balance can further be simplified as:

Input = Output

If represented in a pictorial form the mass balance for any equipment would be:



4.2 DETERMINATION OF PROCESS CONDITIONS

Determination of design pressure and temperature for equipment and piping to be used for petroleum refineries, petrochemical plants and similar plants. It should be noted that specific instructions given in the project specifications take precedence over the requirements described in standard practice.

The data to be used to determine the design conditions should be clarified concerning the following:

Source of the data

The data will be obtained from, but not limited to, the following documents;

- Process Flow Diagram (PFD)
- Basis of Design (BOD)
- Piping & Instrument Diagram (P&ID)
- Plot plan
- Operation case to be taken as the design base
- Related documents to be used as a part of the input data

Related documents to be checked or revised based on the results of the design

Design Pressure Estimation

Design pressure is based on the maximum operating pressure or maximum operating vacuum in the system. The maximum operating pressure for an equipment item shall take into account all modes of operation. In the event that there is only one mode of operation, i.e. the normal operating mode, the maximum operating pressure shall be 5% above the normal pressure. Equipment that can see vacuum under normal mode of operation or under abnormal conditions or during maintenance operations such as steam-out shall also consider vacuum for design of the equipment. Where it is difficult to exactly specify the vacuum conditions during normal operation and for abnormal operating conditions and maintenance operations (steam-out) "Full Vacuum" (FV) should be specified as the design condition. General and specific guidelines related to various types of equipment and piping is provided below:

Design Pressure for Equipment (General)

Design pressure is generally for the top of the equipment or vessel. It shall not include the liquid static head, as this will be added by the vessel design group based on high level. Pressure drop across vessel internals shall be included, if it is significant. Design pressure = maximum operating pressure plus 10% or 1.7 bar, whichever is greater.

Storage Tanks

- For non-pressure storage tanks with fixed roof and open vent design pressure should be considered as (+) 1.0 kPag / (-) 0.5 kPag
- For low-pressure storage tanks with fixed roof and pressure and vacuum breather valves design pressure should be considered as (+) 2.5 kPag / (-) 0.85 kPag.
- For high-pressure storage tanks with fixed roof and pressure and vacuum breather valves design pressure should be considered as (+) 6 kPag / (-) 0.85 kPag.
- For open top tanks or external floating roof tanks design pressure should be considered as (+) 0 kPag / (-) 0.5 kPag.
- For fixed roof tanks with internal floating roof the same categories i.e., non-pressure, low-pressure and high-pressure as mentioned in the first three bullet points are applicable and the same design pressures should be applied.
- For dome type fixed roof tanks the same design pressures are applicable as mentioned in the first four bullet points.

Flare Knockout Drum

The DP for flare knockout drum shall be evaluated on a case to case basis. However DP shall not be less than 3.5 barg for any case.

Shell & Tube (S&T) Heat Exchangers

A S&T heat exchanger has a shell side and a tube side which could have different operating pressures i.e. lower operating pressure for shell side as compared to tube side or vice versa. The selection of the DP for the lower pressure side should be done in such a manner that the corrected hydro test pressure of the low pressure side of the heat exchanger equals or exceeds the design pressure of the heat exchanger high pressure side. This obviates the requirement of providing a relief valve on the low-pressure side of the S&T exchanger for the case of tube rupture. The term corrected hydro test pressure as defined by API STD 521 is the hydro test pressure multiplied by the ratio of allowable stress at the test temperature to the allowable stress at design temperature. It

should be noted that the corrected hydro test pressure is higher than the uncorrected value. Example of corrected hydro test pressure is provided in section 4.3.2 of API STD 521.

DP for Components in Centrifugal Pump Discharge Circuit (Static Lift)

The DP for components in the pump discharge circuit is based on the shut-off pressure of the pump. The shut-off pressure of the pump is defined as the pressure at the discharge connection of a centrifugal pump with the suction pressure at the maximum possible value and the discharge system closed (no flow).

The shut-off pressure (SOP) for any component may be calculated as follows:

$$SOP = MSP + HSPS + PDP - HSPD$$

Where

MSP = MSP is the maximum system pressure at suction side. If the suction system has a vessel then the vessel PSV set pressure shall be used as MSP.

HSPS = hydrostatic liquid head above the pump suction

PDP = pump differential pressure at no flow and maximum pump speed and highest relative density (specific gravity) of the liquid.

HSPD = hydrostatic liquid head above the pump discharge up to the component

The SOP calculated as above may be used as DP for a component in the pump discharge circuit. Normally the entire discharge system of a centrifugal pump is designed for one SOP calculated by considering HSPD value as zero.

DP for Components in Centrifugal Pump Discharge Circuit (Static Drop)

The shut-off pressure (SOP) for any component may be calculated as follows:

$$SOP = MSP + HSPS + PDP + HSPD$$

Where

MSP = MSP is the maximum system pressure at suction side. If the suction system has a vessel then the vessel PSV set pressure shall be used as MSP.

HSPS = hydrostatic liquid head above the pump suction

PDP = pump differential pressure at no flow and maximum pump speed and highest relative density (specific gravity) of the liquid.

HSPD = hydrostatic liquid head below the pump discharge up to the component

The SOP calculated as above may be used as DP for a component in the pump discharge circuit. The DP of the pump suction up to and including the individual pump suction isolation block valves for all pumps operating with a spare shall be equal to the discharge design pressure.

Design Pressure for Centrifugal Compressor

The DP of a compressor is based on its "maximum operating pressure" (MOP) or shut-in pressure. DP for a compressor could be the shut-in pressure or more.

MOP (Shut-In Pressure) for centrifugal compressors should be determined as maximum operating suction pressure plus 1.3 times the normal differential pressure at the rated flow developed by the compressor. This will ensure that conditions such as pressure rise at surge condition and maximum speed are accounted for when specifying the MOP. The PSV set pressure from upstream equipment such as separators/suction KOD to the compressor determines the maximum operating suction pressure for a compressor.

Variation in process conditions (molecular weight, suction pressure and temperature, driver RPM) may affect determination of the design pressure. If operating conditions of rare occurrence require high design pressure, it is advantageous to limit the design pressure to a lower pressure by protecting the system with a pressure relief valve.

Design Pressure for Piping

The design conditions shall be determined in the same way as for equipment. The design pressure for piping shall be no less than the pressure at the most severe conditions of coincident internal or external pressure and temperature during service.

Design Temperature (DT) Determination

The DT is the highest temperature to which the equipment may be subjected to at the design pressure.

The DT is based on the “maximum operating temperature” (MOT) which is defined as the highest temperature, which provides sufficient flexibility for the control of the intended operation. While determining the MOT, the temperatures encountered during start-up, shutdown & process upset conditions should be studied carefully before finalizing the MOT.

Where the maximum operating temperature (MOT) can be ascertained accurately, this temperature should be used as design temperature, without adding a safety margin.

Where the maximum operating temperature cannot be ascertained accurately, the DT should be determined by adding 25°C to the “normal operating temperature” (NOT).

Normal operating temperature (NOT) is defined as the temperature which prevails inside equipment and piping during any intended operation.

Design Temperature for Compressor Discharge Components

For determining the DT for compressor discharge side the MOT needs to be arrived at.

The maximum operating temperature (MOT) on a compressor discharge should be determined as follows:

- When a compressor curve is not available it can roughly be defined as 15°C above the predicted design point temperature to allow for lower efficiency and higher pressure ratio at compressor surge conditions.
- When compressor curves are available, the temperature at surge conditions and maximum compressor speed for normal and start up cases. The DT can then be determined by adding 25°C to the MOT.

Design Temperature for Compressor Suction KOD

Compressor suction KOD design temperature should be the higher of the following:

- Maximum operating temperature at the compressor suction in the event of cooling medium failure, the maximum operating temperature can be limited by a high temperature shutdown function on the compressor suction or discharge.
- Maximum recycle temperature (maximum discharge (temperature trip) minus temperature drop across anti-surge valve) in the event of cooling medium failure.
- Maximum temperature due to settle out conditions.
- Operating temperature plus a margin as defined under the heading “Design Temperature (DT) Determination”.

Design Temperature for Heat Exchangers

For all heat exchangers, both sides can have the same design temperature determined by the hottest of the fluids on either side. This should be decided on a case-to-case basis after careful evaluation.

Design Temperature for Fired Heaters

For tubes in fired equipment the DT is normally the calculated maximum tube skin temperature.

Lower or Minimum Design Temperature

The lower design temperature determines the low temperature characteristics of the material, and shall be the more stringent of the following:

- Minimum operating temperature (obtained during normal operation, start-up, shutdown or process upsets) with a margin of 5°C
- Minimum ambient temperature, the lowest temperature should be based on available weather data, and safety factors should be selected based on the quality of such weather data.
- Minimum temperature occurring during depressurizing with a margin of 5°C, the temperature calculations shall as a minimum include heat transfer between fluid and vessel, and the most conservative starting conditions for depressurizing shall be used including the following considerations as a minimum.
 - cool down to minimum ambient temperature after shut-in at PSV set pressure and corresponding temperature (including reduction in pressure during cool down)
 - Conditions during a start-up operation following a depressurization.
 - Minimum operating temperature and maximum operating pressure.

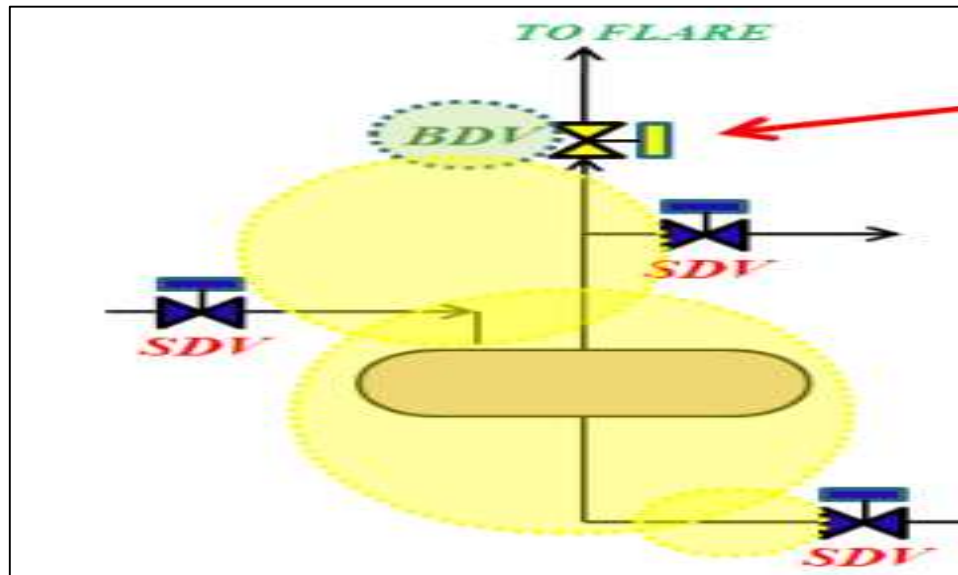
4.3 EMERGENCY DEPRESSURING SYSTEM

Depressurisation is crucial for safe operation of chemical plant and protection of its equipment. It is to carry out to accomplish at least one of the following

- To reduce the risk of catastrophic equipment failure and/or BLEVE during fire exposure. For this case, the project design philosophy shall clearly state whether emergency depressuring is being employed to mitigate the effect of pool fires, jet fires or both.
- To reduce the risk of equipment failure during an internal exothermic runaway reaction.
- To reduce the amount of material released if there is a loss of containment.
- To rapidly move the facility into safe state, in the event of other emergency scenarios such as loss of instrument air or power.
- Manual controls near the vessel may be inaccessible during a fire.

Depressuring may be automatically or manually initiated and may be caused by a variety of triggering events and initial conditions and some reasons to depressuring some equipment/plant are as follows:

- Recycle gas compressor failure.
- Charge pump failure.
- Power failure
- Instrument Air Failure.
- Fire around Equipment/Unit



Depressuring design criteria

- **High-rate depressuring**

Events may lead to severe consequences in a relatively short time frame require high rate emergency depressuring. High rate emergency depressuring is to reduce the equipment pressure to 7 bar (g) (100 psig) or 50% of design pressure whichever is lower within 15 minutes or lower.

- **Low-rate depressuring**

Processes where the reactive hazard can be mitigated using a depressuring rate lower than the high-rate emergency depressuring, may have "low rate operational depressuring valve. Low rate operations depressuring shall depressure the equipment to 50% of design pressure within 60 minutes.

4.4 PRESSURE PROTECTION & RELIEF

Relief Design Philosophy

Reference is made to Section 4.2 of API STD 521 for philosophy related to pressure protection and relief.

Causes of Over-pressurization

Reference is made to Section 4.3 of API STD 521 for the Causes of Over-pressurization.

Approach to Pressure Protection

Three main approaches are possible for pressure protection systems

Full Pressure Rated Mechanical Design

In this approach, the system design pressure is higher than the maximum possible pressure at design temperature, including the case of process upset, and with due allowance for corrosion being made.

Relief System Integrated Design

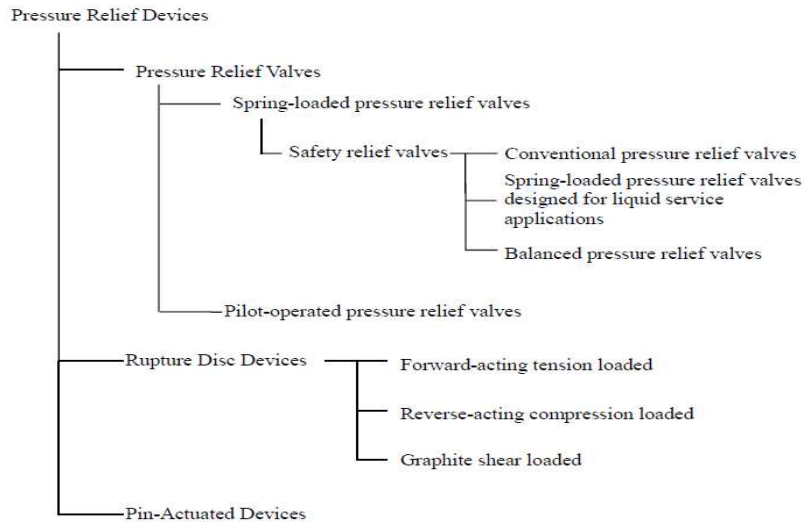
In this approach, although the design pressure has a safety margin on the maximum operating pressure, this margin is not sufficient in case of a process upset and the pressure system under consideration needs to be fitted with relief devices designed to open for relieving over-pressure in case of upset conditions.

High Integrity Pressure Protection System (HIPPS)

These are basically Safety Instrumented Systems (SIS) for overpressure protection. HIPPS typically involve an arrangement of instruments, final control elements (e.g. valves, switches etc.), and logic solvers configured in a manner designed to avoid overpressure incidents by removing the source of overpressure or by reducing the probability of an overpressure contingency to such a low level that it is no longer considered to be a credible case.

With appropriate levels of redundancy, a HIPPS can be designed to achieve a level of availability/reliability equal to or greater than a mechanical relief device.

Reference is made to API STD 521 for further details on HIPPS.



Spring Loaded Relief Valves

- **Conventional Spring Loaded Relief Valves**

These shall be installed where back-pressure does not exceed 10% of the set pressure. These are the recommended type for Thermal Relief Valves (TRVs).

- **Balanced Pressure Relief Valves**

These are suitable for back-pressures ranging from 10% to 50% of the set pressure. They can be of two main types: balanced piston and balanced bellows. Balanced bellows shall be given preference where the fluid is corrosive or fouling.

- **Pilot Operated Relief Valves**

Pilot-Operated valves are recommended for applications where the back- pressure on the valves are very high (typically 50% or higher). The other criteria for selecting pilot-operated valves are as follows:

- Low accumulation rates are required
- The set pressure is required to be close to the operating pressure.
- Calibration is required without removing the valve.
- Handling of large relief flows are required.
- The service is non-fouling i.e. they are not recommended for fouling fluids

PSV Sizing

The spreadsheet PSV Sizing.xls has been developed for estimating the relief valve loads and the relief valve sizes.

The spreadsheet estimates the relief loads for the contingencies listed and also sizes the relief valve orifice size based on the designations in API 526.

The contingencies considered for estimation of relieving loads are

- External Fire – Liquid Vaporization
- External Fire – Gas Expansion

- Tube Rupture – Single phase

The Relief valves sizes are estimated for

- Liquid Relief
- Vapor relief

The equations used in the spreadsheet are detailed in the sections below.

2.0 NOMENCLATURE:

Symbol	Description	Units	Symbol	Description	Units
A	Total surface area	m ²	L	Length of a horizontal vessel	m
A _C	Wetted surface area for cylindrical portion of a vessel	m ²	M	Molecular weight	Kg/kg-mole
A _H	Wetted surface area for head portion of a vessel	m ²	P ₁	Upstream relieving pressure	kPa (g)
A _L	Required effective discharge area for liquid service	mm ²	P ₂	Total back pressure	kPa (g)
A _T	Total wetted surface area of vessel	m ²	P _{cr}	Critical flow pressure	kPa (g)
A _v	Required effective discharge area for vapour service	mm ²	P _H	Pressure at HP side of heat exchanger	kPa (g)
A _{ws}	Final wetted surface area of vessel	m ²	P _L	Pressure at LP side of heat exchanger	kPa (g)
C ₁	Constant		P _n	Normal operating pressure	kPa (g)
C _p	Specific heat capacity at constant pressure	kJ/kmole-K	P _s	Set pressure	kPa (g)
C	Co-efficient for critical flow		Q	Total heat absorption to the wetted surface	W
D	Internal Diameter of Vessel	m	R	Radius of a vessel	m
d ₁	Internal diameter of tube	mm	r ₁	Ratio of LP side to HP side pressures	
F	Environmental factor		r	Ratio of u/s to d/s pressures	
F'	Constant		T ₁	Relieving temperature	K
F ₂	Co-efficeint of sub-critical flow		T _h	Temperature at HP side of heat exchanger	K
G	Specific gravity		T _l	Temperature at LP side of heat exchanger	K
H	Vertical vessel height	m	T _n	Normal operating temperature	K
H ₁	Height of liquid in the vessel	m	T _w	Maximum wall temperature	K
H ₂	Internal diameter – Height of liquid in the vessel	m	W	Mass Flow rate	kg/h
k	Ratio of specific heats		Y	Expansion factor	
K _b	Correction factor due to back pressure for vapour service		Z	Compressibility factor	
K _c	Combination correction factor		μ	Viscosity	cP
K _d	Rated co-efficient of discharge		β	Ratio of orifice diameter to inlet diameter	
K _p	Correction factor due to over pressure		λ	Latent heat of vaporization	kJ/kg
K _v	Correction factor due to viscosity		ρ	Density	kg/m ³
K _w	Correction factor due to back pressure for liquid service		Θ	Angle	radians

Calculation Methodology for Fire Case:

Sizing For Liquid Vaporization:

Parameter (Note: All units in spreadsheet's input cell are in SI unit, spreadsheet will convert it to English unit as required by formulas)

INPUT DATA		CODE	UNITS	RUN	REMARKS
SELECTION	VESSEL ORIENTATION			NOTE:1	
	BOOT			YES (or) NO	
	VESSEL HEAD TYPE			NOTE:2	
	LEGS (or) SKIRT			YES (or) NO	
	ADEQUATE DRAINAGE & FIRE FIGHTING EQUIPMENT EXIST			YES (or) NO	
	TYPE OF DEVICE			NOTE:3	
	TYPE OF SAFETY VALVE			NOTE:4	
CONSTANTS	FIRE HEIGHT CRITERIA		m	7.60	as per API 521, section 5.15.1.1
	PIPING ALLOWANCE FOR WETTED AREA	minimum 1.1			
	ENVIRONMENTAL FACTOR	F			as per Table 6, API 521
	CORRECTION FACTOR DUE TO BACK PRESSURE	Kb			for balance bellow type refer fig.31 of API 520 part I
DIMENSIONS	INTERNAL DIAMETER OF VESSEL	D	m		
	HEIGHT (or) LENGTH (TL to TL) OF VESSEL	H (or) L	m		
	MAXIMUM LIQUID LEVEL	HLL	m		above bottom tan line(BTL), in general 80% of ID (or) height
	INTERNAL DIAMTER OF BOOT		m		
	HEIGHT OF BOOT		m		
	HEIGHT OF LEG (or) SKIRT		m		Elevation from grade (or) ground.
CONDITIONS	VESSEL NORMAL OPERATING PRESSURE	Pn	kPa (g)		
	VESSEL NORMAL OPERATING TEMPERATURE	Tn	K		
	PSV SET PRESSURE	Ps	kPa (g)		
	OVER PRESSURE		%		as per API 520
	BACK PRESSURE**	P2	kPa (g)		
	RELIEVING TEMPERATURE	T1	K		
PROPERTIES	FLUID				
	DENSITY	ρ	kg/m3		(Either of "p" or "G" should be given)
	SPECIFIC GRAVITY	G			
	MOLECULAR WEIGHT	M			
	COMPRESSIBILITY FACTOR	Z			
	SPECIFIC HEAT AT CONSTANT PRESSURE	CP			(Either of "CP " or "k" should be given)
	RATIO OF SPECIFIC HEATS	k			
	HEAT OF VAPORIZATION	λ	kJ/kg		

NOTES:

1. Options available for orientation: "HORIZONTAL" (or) "VERTICAL".
2. Options available for head type: "HEMISPHERICAL", "ELLIPSOIDAL" (or) "DISHED"
3. Options available for device type: "PRV", "RUPTURE DISC" (or) "COMBINATION"
4. Options available for type of relief valve: "CONVENTIONAL", "BALANCED", (or) "PILOT-OPERATED"

Relief Load Calculation:

(Reference is from API 521 5th Edition)

Step 1 > Calculation of Vessel Height in the fire zone:

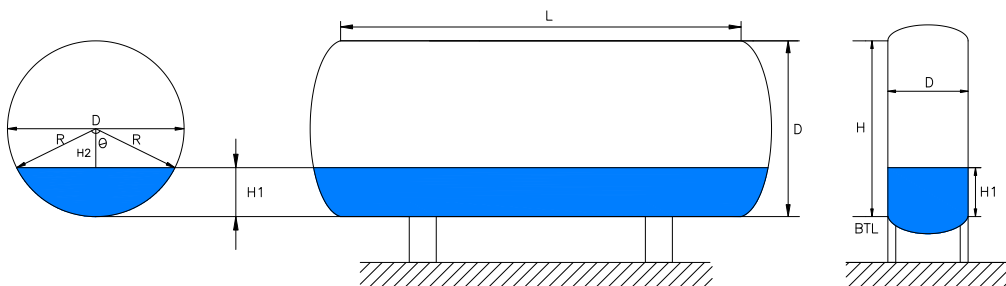
As per API 521 (Clause 3.15.1.1) only that portion of the vessel that is wetted by its internal liquid and is equal to or less than 7.6 m above the source of fire should be considered for calculation of the area exposed to fire.

If (HLL+ELEVATION) is less than 7.6 m then H1 (vessel height in the fire zone) is considered as HLL, else it is considered as (7.6 m – ELEVATION).

Also please note that HHLL should be less than vessel diameter for horizontal vessel and less than length for vertical vessel.

Step 2 > Calculation of wetted Area:

H₁ = vessel height in fire zone calculated in step 1



For horizontal Vessel:

$$H_2 = R - H_1$$

$$\theta = 2 \times \cos^{-1}(H_2/R)$$

Chord L = radius of the cylinder $\times \theta$

$$AC = \text{Chord L} \times L = 2 \times R \times L \times \cos^{-1}((R - H_1)/R)$$

$$AT = AC + 2 \times (A_w / A) \times A_{\text{head}}$$

For Vertical Vessel:

$$A_C = \pi \times H_1 \times D$$

$$A_T = A_C + A_H$$

Head surface area A_H is calculated based on the following equation:

Hemispherical : $(\pi / 2) \times D^2 (=1.57 \times D^2)$

2:1 Semi-Ellipsoidal : $1.084 \times D^2$

Tori spherical : $0.9286 \times D^2$

Final wetted area is then calculated considering piping area and 10% design margin.

$$A_{WS} = A_T \times 1.1$$

Step 3 > Calculation of Heat absorbed:

$$Q = C_1 \times F \times (A_{WS})^{0.82} \quad (\text{API 521, Equation 6 \& 7})$$

As per clause 5.15.2.2.1, API 521 if adequate drainage and fire-fighting facilities exist, then C_1 is 43,200 or else 70,900. For environmental factor “F” refer Table-6 of API 521.

Step 4 > Calculation of Vapor Load

$$W = 3.6 \times Q / \lambda$$

For fire liquid vaporization the latent heat needs to be determined and entered into the spreadsheet.

If there are a number of design cases, the highest molecular weight and lowest molecular weight cases need to be considered for the estimation.

Refer the topic “*Calculation of Latent Heat of Vaporization*” in Section 4.7 of the course book.

Relief Valve Sizing:

Sizing For Vapor Expansion:

Parameter (Note: All units in spreadsheet’s input cell are in SI unit, spreadsheet will convert it to English unit as required by formulas)

INPUT DATA		CODE	UNITS	RUN	REMARKS
SELECTION	VESSEL ORIENTATION			NOTE:1	
	VESSEL HEAD TYPE			NOTE:2	
	TYPE OF DEVICE			NOTE:3	
	TYPE OF SAFETY VALVE			NOTE:4	
	FLOW CONDITION (only for super critical condition)			YES (or) NO	
CONSTANTS	FIRE HEIGHT CRITERIA		m	7.6	as per API 521, section 5.15.1.1
	PIPING ALLOWANCE FOR WETTED AREA	Min. 1.1			
	CORRECTION FACTOR DUE TO BACK PRESSURE	Kb			for balance bellow type refer fig.31 of API 520
DIMENSIONS	INTERNAL DIAMETER OF VESSEL	D	m		
	HEIGHT (or) LENGTH (TL to TL) OF VESSEL	L	m		
	ELEVATION (from grade to BTL)	h	m		
CONDITIONS	VESSEL DESIGN PRESSURE	PD	kPa (g)		
	VESSEL OPERATING PRESSURE	Pn	kPa (g)		
	VESSEL OPERATING TEMPERATURE	Tn	K		
	PSV SET PRESSURE	Ps	kPa (g)		
	OVER PRESSURE		%		as per API 520
	BACK PRESSURE **	P2	kPa (g)		
	RELIEVING TEMPERATURE	T1	K		as per GPSA P-H chart
	MAXIMUM VESSEL WALL TEMP. UNDER FIRE	Tw	K		
PROPERTIES	FLUID				
	DENSITY	ρ	kg/m ³		
	SPECIFIC GRAVITY	G			
	MOLECULAR WEIGHT	M	kg / kg-mole		
	COMPRESSIBILITY FACTOR	Z			
	SPECIFIC HEAT CAPACITY	Cp	kJ/kmol-K		
	RATIO OF SPECIFIC HEATS	k			
	HEAT OF VAPORIZATION	λ	kJ/kg		as per GPSA P-H chart

NOTES:

- Options available for orientation: "HORIZONTAL" (or) "VERTICAL".
- Options available for head type: "HEMISPHERICAL", "ELLIPSOIDAL" (or) "DISHED"
- Options available for device type: "PRV", "RUPTURE DISC" (or) "COMBINATION"

4. Options available for type of relief valve: "CONVENTIONAL", "BALANCED", (or) "PILOT-OPERATED"

Relief Load Calculation:

Step 1 > Calculation of Vessel Height in the fire zone:

Horizontal vessel: If (7.6 m – Elevation) greater than Diameter of the vessel then $H_1=D$ else 7.6 m-Elevation

Vertical vessel: If (7.6 m – Elevation) greater than Length of vessel then $H_1=L$ else 7.6 m –Elevation.

Step 2 > Calculation of area exposed to fire:

H_1 = vessel ht. in fire zone calculated in step 1

The surface area is identical to that used for determining the wetted area in Section 2.1.1

Step 3 > Vessel wall temperature:

For CS: $T_w = 593\text{ }^{\circ}\text{C}$ (866 K)

For SS: $T_w = 816\text{ }^{\circ}\text{C}$ (1089 K)

Other: T_w = user to specify

Step 4 > Relieving temperature:

$$T_1 = T_n \times \left[\frac{P_1 + 101.325}{P_n + 101.325} \right] \quad (\text{API 521, Equation 11})$$

Step 5 > Relieving pressure:

$$P_1 = P_s + (\% \text{overpressure} \times P_s)$$

Step 6 > Gas Relief Rate:

$$W = 0.454 \times 0.1406 \sqrt{M \times (0.145 \times (P_1 + 101.325)) \left(\frac{(10.76 \times A_{WS}) \times (1.8 \times (T_w - T_1))^{1.25}}{(1.8 \times T_1)^{1.1506}} \right)}$$

Relief Valve Sizing:

Calculation Methodology for Tube Rupture Case:

Parameter (Note: All units in spreadsheet's input cell are in SI unit, spreadsheet will convert it to English unit as required by formulas)

INPUT DATA		CODE	UNITS	RUN	REMARKS
SELECTION	FLUID			NOTE: 1	
	TYPE OF DEVICE			NOTE: 2	
	TYPE OF SAFETY VALVE			NOTE: 3	
	CAPACITY CERTIFICATION REQUIRED			YES (or) NO	(for ASME code equipment only)
CONSTANT	CORRECTION FACTOR DUE TO BACK PRESSURE	Kw			for balance bellow type refer fig.31 of API 520 part I
	CORRECTION FACTOR DUE TO OVER PRESSURE	KP			as per fig.38 of API 520, part I,(for over pressure other than 25% only)
	CORRECTION FACTOR DUE TO VISCOSITY	Kv			as per fig.37 of API 520, Part I
CONDITIONS	INTERNAL DIAMETER	D	mm		
	HPS MAX. OPERATING (or) DESIGN PRESSURE	PH	kPa (g)		
	(HP) HIGH PRESSURE SIDE OPERATING (or) DESIGN TEMPERATURE	TH	K		
	(LP) LOW PRESSURE SIDE DESIGN (or) PSV SET PRESSURE	Ps	kPa (g)		
	OVER PRESSURE		%		
	BACK PRESSURE	P2	kPa (g)		
	LPS DESIGN TEMPERATURE	Tl	K		
	UPSTREAM RELIEVING TEMPERATURE		K		
PROPERTIES	LIQUID (or) VAPOUR DENSITY	ρ	kg/m ³		
	SPECIFIC GRAVITY				
	COMPRESSIBILITY FACTOR	Z			
	MOLECULAR WEIGHT	M	kg/kg-mole		
	SPECIFIC HEAT CAPACITY AT CONST. PRESSURE	CP	kJ/kmole-K		
	RATIO OF SPECIFIC HEATS	k			

NOTES:

- Options available for fluid: "LIQUID" (or) "VAPOUR".
- Options available for device type: "PRV", "RUPTURE DISC" (or) "COMBINATION"
- Options available for type of relief valve: "CONVENTIONAL", "BALANCED", (or) "PILOT-OPERATED"

Relief Load Calculation:

As per API 521 (clause 5.19.3), for tube rupture case, double the discharge rate through an orifice should be consider. Crane (Technical Paper # 410) for single phase calculations and DIERS method for two phase flow to be followed.

For calculating the flow from a ruptured tube an orifice with a discharge co-efficient of 0.7 for liquid and two-phase flow, and 0.9 for vapors is assumed.

For Liquids:

$$W = 2 \times 1.265 \times C \times d_1^2 \times \sqrt{0.01 \times (P_H - P_1) \times \rho} \quad (\text{CRANE, TP \# 410M, Section 3-5})$$

For Vapors:

$$W = 2 \times 1.265 \times C \times Y \times d_1^2 \times \sqrt{0.01 \times \Delta P \times \rho} \quad (\text{CRANE, TP \# 410M, Section 3-5})$$

$$Y = \sqrt{\left(\frac{k}{k-1}\right) r_1^{\frac{2}{k}} \left[\frac{1 - r_1^{\frac{k-1}{k}}}{1 - r_1} \right]} \quad (\text{Perry's Handbook, Equation 10-26, taking } \beta = 0)$$

$$r_1 = (P_1 + 101.325) / (P_H + 101.325)$$

$$P_{cf} = P_H \times \left[\frac{2}{k+1} \right]^{\frac{k}{k-1}}$$

If $P_{cf} \geq P_1$, the $\Delta P = P_H - P_{cf}$; or else $\Delta P = P_H - P_1$

Relief valve sizing:

Follow sections 5.1 and 5.2 for liquid and vapor valve sizing respectively.

For relieving temperature calculation, considering the process as adiabatic, following equation to be used:

$$T_1 = T_n \times \left[\frac{P_1 + 101.325}{P_n + 101.325} \right]^{\frac{1-k}{k}}$$

Rupture Disks

Rupture disks shall be only used in following cases:

- Where fast response time is required
- Where operation of pressure relief valve may be affected by corrosion or corrosion products or by deposition of material that may prevent lifting of relief valve. In such cases the rupture disk may be placed upstream of the PSV.
- Where fluid to be relieved is highly toxic and its leakage from the relief valve cannot be tolerated.
- Where it is necessary to provide for rapid depressurization and which may lead to freezing of liquids.

Types of Rupture Disks

- **Conventional Rupture Disks**

Suitable when operating conditions are stable and do not exceed 70% of the rated burst pressure. If vacuum or back pressure can be present, rupture disks shall be fitted with an adequate support to prevent reverse flexing or implosion.

- **Scored Tension-loaded Rupture Disks**

They shall be given preference over conventional RD's when the system operating pressure reaches 85% of the rated burst pressure and/or when debris resulting from disk burst is to be avoided.

- **Reverse-acting Rupture Disks**

Recommended when operating pressure reaches up to 90% of the rated burst pressure. As compared to scored tension loaded bursting disks, they present additional advantages which must be contemplated for selection such as their increased material thickness which provides improved resistance to corrosion and in most cases, they can withstand full vacuum without additional support.

- **Composite Rupture Disks**

They are selected when resistance to corrosion is of primary concern.

Domed types are suitable for operating pressure, reaching 80% of the rated burst pressure.

Flat type are particular suitable for low rated bursting pressure and shall be typically used as corrosion barriers in which case they may typically operate at 50% of the rated burst pressure.

- **Relief Device Selection & Setting**

The relief device selection shall depend on the service which may include the fluid composition, pressure, temperature, level, phase, hazards etc. Refer Section 4 of API STD 520 Part-1 for detailed description of various pressure relief devices and their applications.

The set pressure and relieving pressure shall be as per API STD 520 Part-1, Section 5.4 for process equipment, utilities and pressure vessels for storage of liquefied hydrocarbon.

For liquid petroleum product tanks (atmospheric and low-pressure) or crude oil storage tank the set pressure shall be according to API STD 2000. Typical set pressure range for venting devices is given in Appendix C of API STD 2000.

4.5 LINE SIZING

General Evaluation for Piping

When sizing piping, the following constraints shall be addressed:

- Required Capacity / Available Driving Pressure
- Flow Induced Forces
- Noise / Vibration
- Pressure Surges
- Material Degradation – Erosion, Corrosion, Cavitation's
- Liquid Accumulation / Slug Flow
- Solids Accumulation

Pipe Roughness

For all calculations of pressure drop, the following pipe roughness values should be used:

- Carbon steel (CS) corroded: 0.46 mm (0.018 inch) (Note-1)
- Carbon steel (CS) non-corroded (for relief system piping): 0.15 mm (0.006 inch)
- Carbon steel (CS) non-corroded (other systems): 0.046 mm (0.0018 inch)
- Stainless steel (SS): 0.046 mm (0.0018 inch)
- Titanium and Cu-Ni: 0.046 mm (0.0018 inch)
- Glass fiber reinforced polyester (GRP): Vendor to provide
- Polyethylene, PVC: Vendor to provide

Notes:

1. The value of 0.46 mm (0.018 inch) shall be used when hydraulic calculations are performed for existing pipe installations including relief piping.

Liquid & Gas Line Sizing

- Velocity Limitations

The velocities shall in general be kept low enough to prevent problems with erosion, water-hammer pressure surges, noise, vibration and reaction forces. In some cases a minimum velocity is required, such as in liquid lines carrying solids (e.g. slurries, suspensions) to prevent settling of solids in the lines.

	<i>Select lower value of the two columns</i>	
Service	Velocity (1st column)	Velocity (2nd column¹)
Pump suction piping	Based on adequate NPSH ≤ 3 m/s	
Pump discharge piping: $\leq 4''$ 6'' and 8'' 10'' and 12'' 14'' and 16'' $\geq 18''$	3.1 m/s 4.6 m/s 5.8 m/s 6.4 m/s 7.3 m/s	
Gas or Vapor Piping (including superheated steam)	38 m/s	$V = 180 \cdot \rho^{-0.5}$ m/s where: V = Erosional Velocity limit ρ = flowing density, kg/m ³
Steam (other than superheated steam) piping	23 m/s	
	<i>Select lower value of the two columns</i>	
Service	Velocity (1st column)	Velocity (2nd column¹)
Liquid-gas (or liquid-vapor) mixture piping, except heater transfer and thermo-syphon reboiler outlet piping.	23 m/s	
Thermo-syphon reboiler outlet piping.	12 m/s	
C.S. phenol or Sulphuric acid piping	1 m/s	
Slurry piping, except as noted below	4.6 m/s	
Slurry piping where erosion or crystal breakage is of concern	3 m/s	
Notes: (1) Refer to API RP 14E paragraph 2.4(a) and ISO 13703 paragraph 5.5.1 for further guidance		

- **Maximum Pressure Drop Recommendations**

In addition to the velocity limits as given above maximum pressure drop criteria should be checked during line sizing.

For sizing pump suction lines the following maximum pressure drops should be adhered to:

- Sub-cooled liquids: 0.25 bar / 100 m
- Boiling Liquids: 0.05 bar / 100 m

For “single-phase gas process lines” where pressure drop is critical (e.g. when it results in unacceptable liquid drop out in suction lines between scrubber and compressor suction, inlet lines to turbo expanders and contactors etc.) the pressure drop guidelines are according to Table below:

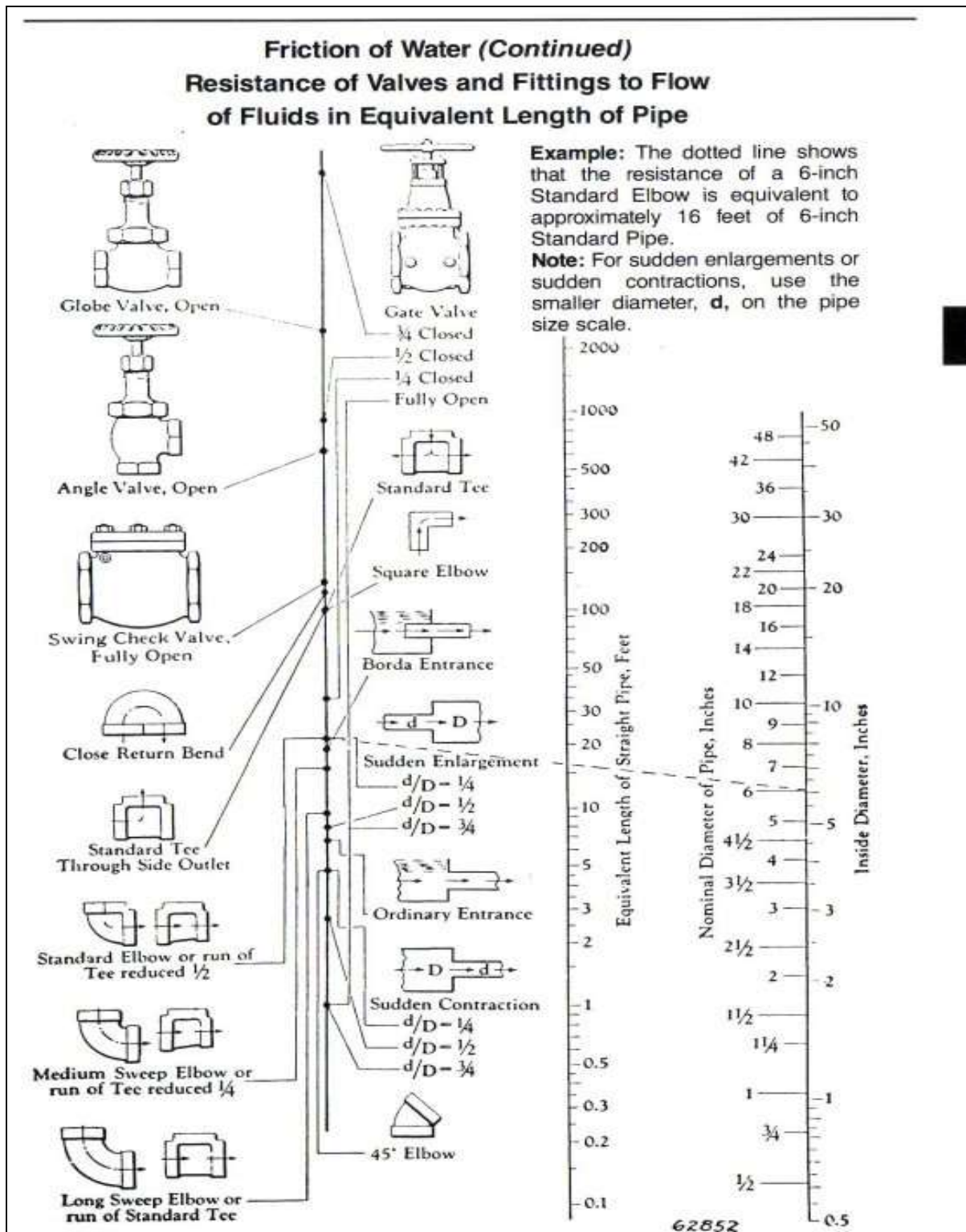
Operating Pressure, barg	Pressure Drop, bar / 100 m
0 to 35	0.001 to 0.11
35 to 138	0.11 to 0.27
Over 138	$P / 500$ (where P is operating pressure in bar (abs))

When determining the length of the pipe for pressure drop calculations, besides considering the straight length of the pipe, pipe fittings and valves need to be considered for the purpose of pressure drop. Defining pipe fittings and valves in terms of straight pipe leads to the concept of “equivalent length”.

Equivalent length is a method to establish the friction losses in pipe fittings and valves which when added to the straight length provides the total length to be considered for pressure drop calculations.

For a very instructive article on equivalent length refer to the link below:
<http://www.cheresources.com/eqlength.shtml>

A nomograph for equivalent length estimation is provided below. Equivalent length calculations can be programmed in MS-Excel in order to determine the total length of the system to be considered for pressure drop



➤ Sizing of Gas Relief Lines

General

In general, all flare lines shall be designed to keep the $\rho V^2 < 200,000 \text{ kg/ms}^2$ criteria (where ρ is the fluid density or mixed density for two phase conditions in kg/m^3 and V is the velocity in m/s).

Further, the selection of piping specification shall consider the effect of acoustic fatigue, which is affected by factors such as

- Relative differential pressure in upstream restriction device
- Temperature in the flowing gas
- Mole weight of flowing gas
- Pipe diameter and wall thickness
- Mass flow rate

Pressure Safety Valve line, Flare Sub-header & Flare Main Header

The PSV discharge line up to the sub-header/main header shall be sized to limit the velocity to Mach 0.8.

New flare sub-headers and main header should be designed to limit the velocity to Mach 0.5.

Existing flare sub-header and main header should be checked to ensure that the velocity does not exceed Mach 0.7 when new connections are made to them. While making new connections the impact of back pressure on existing relief devices should be duly taken care.

Controlled Flaring Lines

Flaring lines downstream of control valves shall be designed for a maximum velocity of Mach 0.5.

Depressurization Lines

The maximum flowing velocity in the lines downstream the reducer shall be Mach 0.7.

Relief Lines with Slug / Plug Flow

For potential slug/plug flow, line sizing shall be based on slug velocity and slug density. These slug characteristics shall form the basis for stress calculations and design of piping support.

Vent Lines for Atmospheric Tanks

Maximum allowable backpressure shall be 0.07 barg.

Excel Programmable Formulas for Incompressible Gas Flow and Liquid Flow Line Sizing

Gases are compressible. Essentially it means that gas density changes as gas flows through the pipe and the gas pressure drops mainly due to friction losses. Gas density is a function of the gas pressure considering isothermal gas flow. The property that dominates friction losses in pipe flow is the fluid viscosity. Since gas viscosities are low, the pressure drop in single-phase gas lines are generally very low for short pipe lengths. This allows for all practical purposes the gas flow to be considered as incompressible for plant piping. Hence, the formulas applicable for liquid flow can be applied for gas flow in plant piping for all practical purposes. The formulas referred below are mentioned for liquid flow but can be applied for gas flow occurring in process plant piping. It is important to note that gas flow in long distance pipelines (length in kilometers) requires the density change to be accounted for in pressure drop calculations, and such gas flow needs to be treated as compressible flow. A different set of formulas are applied for compressible gas flow in pipelines and the formulas provided below are not applicable.

Nomenclature in formulas:

W = Liquid Flow Rate, kg/h

ρ = Liquid Density, kg/m³

μ = Liquid Viscosity, cP

ϵ = Absolute Pipe Roughness, mm

d = Inside Pipe Diameter, mm (Refer Note below)

Re = Reynolds number, dimensionless

f = Darcy friction factor, dimensionless

ΔP = Pressure Drop, kPa / 100m

v = Liquid velocity, m/s

Note: As per global industry practices, pipes are specified as nominal pipe size (NPS) identifying the outer diameter in inches and a pipe schedule which indicates the wall thickness. The pipe inside diameter needs to be calculated by subtracting 2-times the wall thickness from the outer diameter of the pipe. For a given NPS, but with different pipe schedules, the pipe inside diameter will change according to the pipe schedule. Refer the link below:

<https://hardhatengineer.com/pipe-class-piping-specifications-pipeend/pipe-schedule-chart-nominal-pipe-sizes/>

a) Formula for Initial Estimate of Pipe Diameter

$$d = ((125060 * W^2) / (\delta P * \rho))^{1/5}$$

Use next largest diameter, in mm, for initial calculations

b) Formula for Reynolds Number

$$Re = (354 * W) / (\mu * d)$$

c) Formula for Friction factor

$$f = 0.25 / (\log ((\epsilon / 3.7d) + (5.74 / Re^{0.9})))^2$$

d) Formula for pressure drop

$$\delta P = ((6253000 * f * W^2) / (d^5 * \rho))$$

e) Formula for velocity

$$v = ((353.7 * W) / (d^2 * \rho))$$

Line Sizing Problem:

Inputs:

Fluid: Water @30°C

Viscosity: 1cP

Flow: 100,000 kg/h

Inlet Pressure: 500 kPag

Pipe Straight Length: 200 m

Pipe Fittings and Valves: 4 gate valves and 20 90° Elbows

Pipe Material: Carbon Steel (New)

Allowable Pressure Drop: 45 kPa / 100 m

Problem Statement: Find the pipe inside diameter and the corresponding nearest nominal pipe size and pipe schedule

*PROCESS DESIGN ENGINEERING
MANUAL:
Part 2*

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4.6 STORAGE TANKS

This topic deals with storage tanks of metallic construction using metals such as carbon steel, stainless steel, aluminum, alloy steels.

In a chemical process plant such as an oil refinery, petrochemical plant, bulk organic or inorganic production plant, or any other chemical production plant storage of the raw materials, intermediate products and finished products is necessitated due to the following reasons:

- Uncertainty in availability of required quantity of raw materials on a continuous basis to sustain Plant operation and production, necessitates keeping an inventory of raw materials in bulk storage tanks.
- Batch processes where different chemicals or reactants are required to be mixed in a discontinuous manner require that the chemicals be stored in storage tanks for addition in the reaction as per the required reaction step at the appropriate time.
- Finished products may not have an immediate market and are required to be stored in large quantities prior to sale and dispatch.

Storage tanks are to be found constructed above ground, in ground and below ground. In shape they are mostly usually of vertical cylindrical form, but also come in horizontal cylindrical, spherical and rectangular forms. Products range from gases, liquids, solids and mixtures thereof. Tanks for the storage of particulate solids are more usually known as silos. Temperatures range from high temperature heated storage tanks (for products such as bitumen) through to -163°C for the storage of LNG and -196°C for liquid nitrogen.

A wide variety of storage tank configurations exist, including those with fixed roofs, floating roofs, internal roofs, with single walls, double walls and insulated tanks to name but a few.

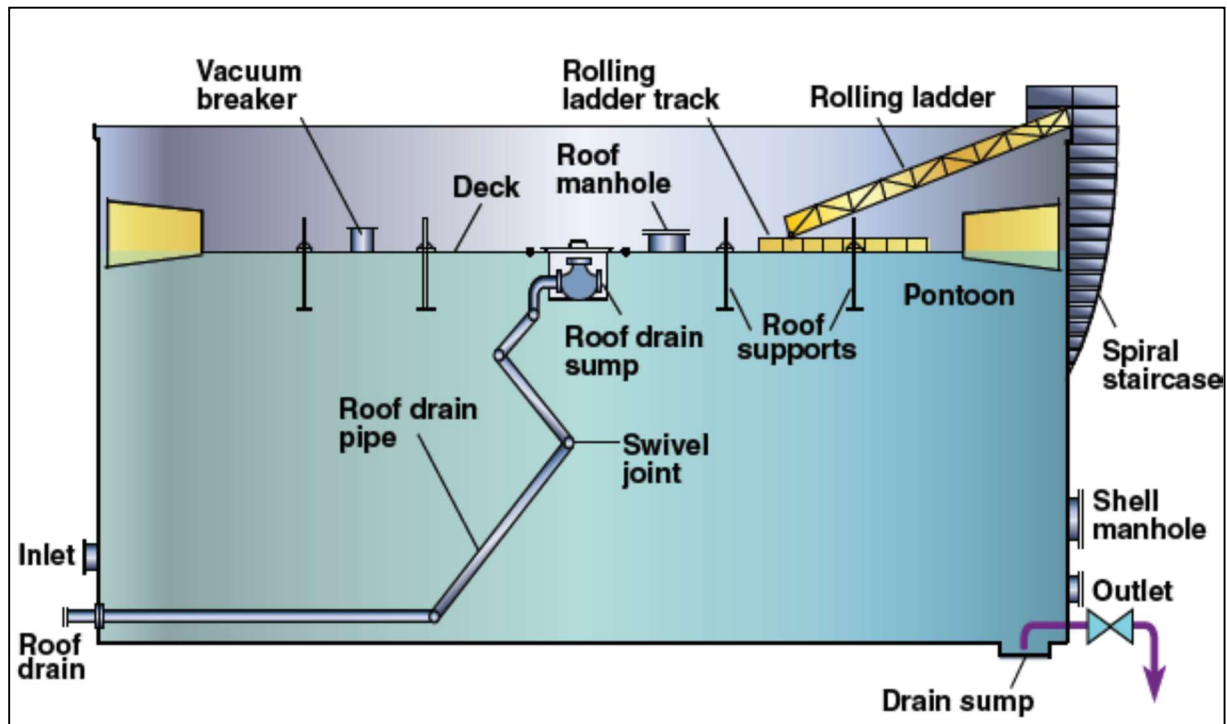
It is important to distinguish between storage tanks and pressure vessels. European Directive called the PED 2014/68/EU defines pressure vessels as those vessels with a design pressure greater than 0.5 barg, which implies that storage tanks are vessels with a design pressure less than 0.5 barg. The American Petroleum Institute (API) standard API STD 620 (Design and Construction of Large, Welded, Low-Pressure Storage Tanks) defines a storage tank as vessel whose design pressure will not exceed 15 psig (1.034 barg). Another API standard for storage tanks API STD 650 (Welded Tanks for Oil Storage) limits the design pressure to 0.18 barg.

Design criteria for storage tanks also includes the design temperature for storage tank design, where API STD 620 and API STD 650 differ on the limits of design temperature. Whereas API STD 620 limits the design temperature range from a minimum design temperature of -198°C to a maximum of 121°C, API STD 650 provides the design temperature range from a minimum of -40°C to 260°C. Both API STD 620 and API STD 650 are popular standards for design and fabrication of storage tanks worldwide.

EN-14015(Specification for the design and manufacture of site built, vertical, cylindrical, flat-bottomed, above ground, welded, steel tanks for the storage of liquids at ambient temperature and above) is a European standard for design and fabrication of storage tanks.

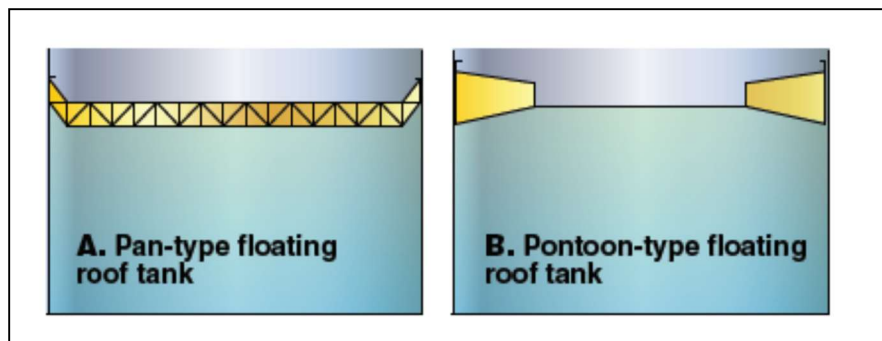
Basic Storage Tank Configurations

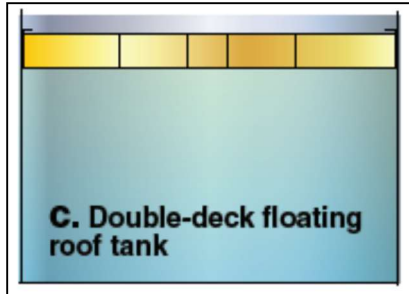
Open-Top Tank with Floating Roof (External)



Details of an Open-Top Tank with External Floating Roof (Pontoon Type)

In external floating-roof storage tanks, the roof is made to rest on the stored liquid and is free to move with the level of the liquid. These tanks reduce evaporation losses and control breathing losses while filling. There are principally three different types of external floating roofs. The same are illustrated below:





The pan-type roof (Figure A) is a single-deck roof and has the following characteristics:

- Full contact with liquid surface
- Has a deck, hence any leak through the deck will cause it to sink
- Has no buoyancy other than that provided by the deck
- Rain or snow may cause deformation
- Is the least expensive of the floating roofs

The pontoon-type roof (Figure B) is a significant improvement over the pan roof. It has the following characteristics:

- Increased buoyancy and stability
- Pontoons occupy about 20 – 40% of roof area

The double-deck roof (Figure C) comprises upper and lower decks separated by bulkheads and trusses.

These roofs have the following characteristics:

- The space between the decks is separated into liquid-tight compartments
- Superior loading capacity
- Recommended for tank diameters below 12 m and above 60 m

Fixed Cone Roof Tanks



This is the most common configuration for storage tanks and is generally the least expensive storage tank design.

The main components for a storage tank are as follows:

- Foundation
- Bottom
- Shell
- Roof

The mechanical design of these components and the entire tank is done by a static equipment engineer. However, the process data for the storage tanks is provided by the process engineer, which facilitates the mechanical design by the mechanical designer. Refer the figure below for the data that is typically provided as an input to the mechanical designer:

General	Status:	Tender		Internal diameter (M)	45
	Service	Condensate / Fuel Oil		Tank Height (M)	25
	Location	Fujairah		Net capacity required (M ³)	37,500 (approx) per tank
	Number of tanks	03		Area classification/ IP Class	Refer Notes on Page 7
	Equipment tag no/s	TK-601, 603,606		Type of Bottom	Cone Down
	Ambient Temperature	Min	5.0	Type of Roof	Aluminium Geodesic Roof with Aluminium internal floating deck
Process		Max	50.0		
	Medium	Liquid		Design liquid Sp. Gr.	1.00
	Properties:-	Condensate	Fuel Oil	Design liquid level (MM)	Full Height
	Specific gravity (Max)	0.74	0.990	Design pressure (kPa)	Atmospheric
	Viscosity Max (cSt)	1.0	180 cSt at 50°C	Operating level (MM)	TBA by Contractor
	Flash point °C	Below 38 °C	Above 60 °C	Tank heating / cooling	Heating Required
	Vapour pressure (Psia)	13.0	0.2	Nozzle heating	Not Applicable
	Medium Corrosiveness	Non-Corrosive		Heating/cooling medium	Thermal Oil
	Filling rate (M ³ /hr)	4500		Heat load KW/hr	Refer Note :26
	Emptying rate (M ³ /hr)	4500		Tank mixers	No
	Storage Temp. °C	Ambient		Water draw off sump	Yes, at centre
	Storage pressure (Psig)	Atmospheric		Tank gauging	Yes, Gauge Hatch required
	Design temperature °C	Shell	90	Nitrogen (N ₂)blanketing	No
		Roof	90	Mixing using N ₂ Web	Required
Accessories	Tank Level & Temperature instruments	Radar Gauge with Temperature Element		Level switches	Independent High Level Alarm
	Level & Temperature display station locations	a) Tank top near gauging hatch b) Tank bottom near stairway entry c) Control room		Emergency vent	As per code
				Pressure/vacuum vent	Required & as per code, In & out breathing Capacity-4500 m ³ /hr
				Gauge hatch	Required – Periphery
				Sample hatch	Required – Centre
				Stilling well	Required
	Vortex breaker	Required		Side entry mixers	Not Applicable
	Inlet diffuser	Required		Foam pipe supports	Required
	Floating suction	Not Applicable		Deluge pipe supports	Required

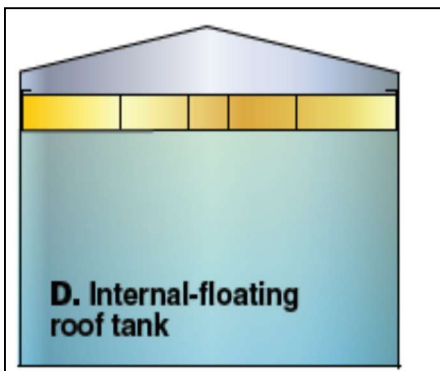
Fixed Dome Roof Tanks

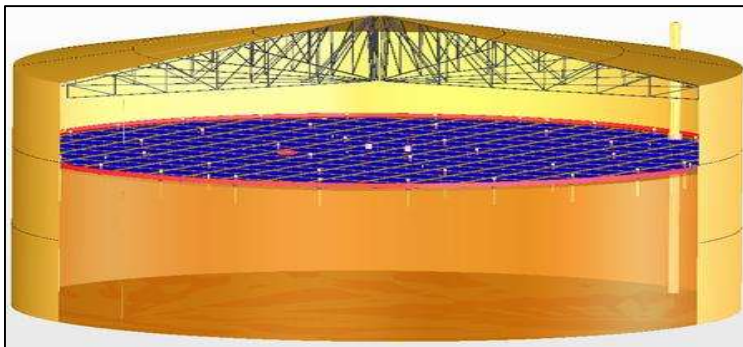
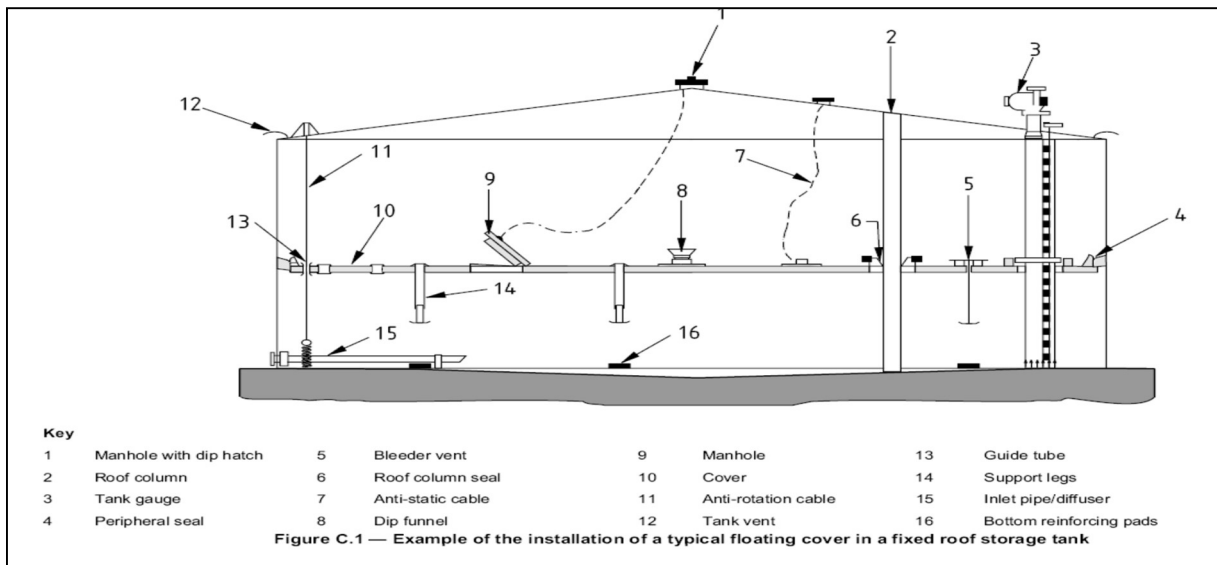
These fixed roof tanks have a self-supporting dome and are preferred for storing higher vapor pressure product in comparison to fixed cone roof tanks. The dome roof design also provides advantage in terms of less snow accumulation compared to cone roof design, in locations where snowfalls are a regular occurrence. The fixed dome roof design is more expensive than fixed cone roof design.



Fixed Roof Tank with Internal Floating Roof

A fixed roof tank with internal floating roof can either have a fixed cone or fixed dome roof. Annex H of API STD 650 provides details of design for fixed roof tanks with internal floating roofs. Refer the figures below for a better understanding of fixed roof tanks with internal floating roof:





Cutaway view of a Fixed Roof Tank with Internal Floating Roof

Selection Guide for Vertical Storage Tanks

Abbreviations	
HPCRT:	High-Pressure Cone Roof Tank
LPCRT:	Low-Pressure Cone Roof Tank
NPCRT:	Non-Pressure Cone Roof Tank
HPDRT:	High-Pressure Dome Roof Tank
LPDRT:	Low-Pressure Dome Roof Tank
NPDRT:	Non-Pressure Dome Roof Tank
OTFR:	Open-top Tank with Floating Roof

Pressure Rating for Cone Roof Tanks

Type	Pressure/Vacuum (mbar)
NPCRT	7.5 / 2.5
LPCRT	20 / 6.0
HPCRT	56 / 6.0

Pressure Rating for Dome Roof Tanks

Type	Pressure/Vacuum (mbar)
NPDRT	7.5 / 2.5
LPDRT	20 / 6.0
HPDRT	56 / 6.0

Note: This type of roof is supported by a frame of internal curved rafter beams. In general dome roof tanks are used for special cases only.

Fixed Roof Tanks with Internal Floating Roof

A fixed roof tank may be provided with an internal floating roof to reduce the vapor losses and air pollution. It is recommended that their use is limited to tanks with a maximum diameter of 39 m. Design of fixed roof tanks with an internal floating roof shall be in accordance with Annex H of API STD 650.

	Tank Diameter in Meters																									
Class of Product	3	4	6	8	10	12.5	15	17.5	20	22.5	25	27.5	30	33	36	39	42	45	48	54	60	66	72	75	78	
Class I, Flash Point less than 21°C	HPCRT, LPCRT						HPCRT, LPCRT, HPDRT, LPDRT, OTFR				Preferably OTFR, LPCRT, LPDRT						OTFR			OTFR						
Class II, Flash Point 21°C and higher but less than 55°C	LPCRT						LPCRT, LPDRT, OTFR				Preferably OTFR, LPCRT, LPDRT						OTFR (on bad sub-soils or for other special conditions NPCRT may be preferred)			OTFR						
Class III + unclassified, Flash Points 55°C and higher	NPCRT						NPCRT, NPDRT				NPCRT, NPDRT						NPCRT			NPCRT special cases						

HEIGHT OF TANK IN m	TANK DIAMETER IN m						
	3	4	6	8	10	12.5	15
3							
4							
6							
8							
10							
12							
14							
16							
18							
20							
22							

TANKS IN SIZES BELOW HEAVY LINE SHOULD NOT BE USED.

Determination of tank stability in high winds (1-minute wind gusts)

Notes for tank stability in high winds.

- As a rough indication it may be assumed that NPCRT, LPCRT, NPDRT and LPDRT tanks in sizes above the dotted line are stable in winds less than 160 km/h (100 mph) when at least 0.5 m of product is present in the tank. Such NPCRT, NPDRT and LPDRT in sizes marked with hatching are unstable in winds of 160 km/h (100 mph) or more.
- Tanks classified as HPCRT and HPDRT shall always be fitted with anchor bolts and concrete foundation rings or shell-to-bottom brackets.
- Detailed stability calculations shall be made to determine whether a tank of a certain size is stable for a specified wind speed.

Nominal capacities of standard vertical cylindrical tanks

Height in m	Tank diameter in m																	
	3	4	6	8	10	12.5	15	17.5	20	22.5	25	27.5	30	33	36	39	42	45
Nominal capacities in m ³																		
1	7	12	28	50	78	122	176	240	314	397	490	593	706	855	1017	1194	1385	1590
2	14	25	56	100	157	245	353	481	628	795	981	1187	1413	1710	2035	2389	2770	3180
3	21	37	84	150	235	358	530	721	942	1192	1472	1781	2120	2565	3053	3583	4156	4771
4	28	50	113	201	314	490	706	962	1256	1590	1963	2375	2827	3421	4071	4778	5541	6361
5	35	62	141	251	392	613	883	1202	1570	1988	2454	2969	3534	4276	5089	5962	6927	7952
6	42	75	169	301	471	736	1060	1443	1884	2385	2945	3563	4241	5131	6107	7167	8312	9542
7		87	197	351	549	859	1237	1683	2199	2783	3436	4157	4948	5987	7125	8362	9698	11133
8		100	226	402	626	961	1413	1924	2513	3180	3926	4751	5654	6842	8142	9556	11083	12723
9			254	452	706	1104	1590	2164	2827	3578	4417	5345	6361	7697	9160	10751	12468	14313
10			282	502	785	1227	1767	2405	3141	3976	4908	5939	7068	8552	10178	11945	13854	15904
11				552	863	1349	1943	2645	3455	4373	5399	6533	7775	9408	11196	13140	15239	17494
12				603	942	1472	2120	2886	3769	4771	5890	7127	8482	10263	12214	14335	16625	19085
13					1021	1595	2297	3126	4084	5168	6381	7721	9189	11118	13232	15529	18010	20675
14					1099	1718	2474	3367	4398	5566	6872	8315	9896	11974	14250	16724	19396	22266
15					1178	1840	2650	3607	4712	5964	7363	8909	10602	12829	15268	17918	20781	23856
16					1256	1963	2827	3848	5026	6361	7853	9503	11309	13684	16285	19113	22167	25446
17						2086	3004	4088	5340	6759	8344	10097	12016	14540	17303	20308	23552	27037
18						2208	3180	4329	5654	7156	8835	10691	12723	15395	18321	21502	24937	28627
19						2331	3357	4570	5969	7554	9326	11285	13430	16250	19339	22697	26323	30218
20						2454	3534	4810	6283	7952	9817	11879	14137	17105	20357	23891	27708	31808
21							3711	5051	6597	8349	10308	12473	14844	17961	21375	25086	29094	33399
22							3887	5291	6911	8747	10799	13067	15550	18816	22393	26280	30479	34989

Notes:

1. Nominal capacities of tanks are calculated on the assumption that the tank has a flat bottom and is filled to the top of the shell plates.
2. The net capacity of fixed roof tanks should take into a dead stock at the bottom of approx. 0.3 m and the fact that the maximum safe working level at the top is taken 0.2 m below the top curb angle.
3. The diameter of a floating roof tank should at least be equal to its height
4. The net capacity of a floating roof tank is less than the tabulated nominal capacity by approx. 2.0 m of its height. This is because of a dead stock of approx. 1.2 m to keep the roof floating and the fact that the maximum working level at the top is approx. 0.8 m below the top curb angle.

When a wind skirt is applied at the top of the shell the net capacity is less than the tabulated nominal capacity by approx. 1.2 m of its height.

5. The net capacity of a fixed roof tank with a floating cover is considerably less than the tabulated nominal capacity.

A dead stock of approx. 1.6 m is needed to keep the cover floating. For dome roofs and cone roofs up to 12.5 m a maximum safe working level at the top is taken 0.6 m below the top curb angle and for cone roofs over 15 m diameter this max. safe working level is 0.8-2 m below the top curb angle, as the roof trusses protrude below the top of the shell.

4.7 VENTING OF ATMOSPHERIC AND LOW-PRESSURE STORAGE TANKS

Reference: API Standard 2000, 7th Ed., March 2014 and API STD 521, 6th Ed., 2014

➤ Terms, Definitions and Abbreviations

Accumulation

Pressure increase over the design pressure of the vessel during discharge through the pressure-relief device.

NOTE Accumulation is expressed in units of pressure or as a percentage of or design pressure. Maximum allowable accumulations are established by pressure-design codes for emergency operating and fire contingencies.

Bubble Point

Temperature at which the first vapor bubble is produced from a liquid mixture of two or more components heated at constant pressure. For single component systems the bubble point is referred to as the boiling point.

Emergency Venting

Venting required for external fire or other abnormal conditions.

Full Open Position

Position where lift of the pallet is sufficient for the nozzle to control the flow or where the pallet or main valve seat lifts against a fixed stop.

Latitude

In geography, latitude is a geographic coordinate that specifies the north-south position of a point on the Earth's surface. Latitude is an angle (defined below) which ranges from 0° at the Equator to 90° (North or South) at the poles. Lines of constant latitude, or parallels, run east–west as circles parallel to the equator.

Example: Mumbai is at a latitude of 19°N

Non-Refrigerated Tank

Tank that stores material in a liquid state without the aid of refrigeration, either by evaporation of the tank contents or by a circulating refrigeration system.

NOTE generally, the storage temperature is close to, or higher than, ambient temperature.

Normal cubic meters per hour (Nm³/h)

SI unit for volumetric flow rate of air or gas at a temperature of 0 °C and pressure of 101.3 kPa, expressed in cubic meters per hour.

Normal Venting

Venting required because of operational requirements or atmospheric changes.

Overpressure

Pressure increase at the PV valve inlet above the set pressure, when the PV valve is relieving.

NOTE 1. Overpressure is expressed in pressure units or as a percentage of the set pressure.

NOTE 2. The value or magnitude of the overpressure is equal to the value or magnitude of the accumulation when the valve is set at the design pressure and the inlet piping losses are zero.

PV Valve

Weight-loaded, pilot-operated, or spring-loaded pressure vacuum valve used to relieve excess pressure and/or vacuum that has developed in a tank.

Rated Relieving Capacity

Flow capacity of a relief device expressed in terms of air flow at standard or normal conditions at a designated pressure or vacuum.

NOTE Rated relieving capacity is expressed in Nm³/h.

Refrigerated Tank

Tank that stores liquid at a temperature below atmospheric temperature with or without the aid of refrigeration, either by evaporation of the tank contents or by a circulating refrigeration system.

Relief Device

Device used to relieve excess pressure and/or vacuum that has developed in a tank.

Relieving Pressure

Pressure at the inlet of a relief device when the fluid is flowing at the required relieving capacity.

Required Flow Capacity

Flow through a relief device required to prevent excessive pressure or vacuum in a tank under the most severe operating or emergency conditions.

Set Pressure

Gauge pressure at the relief device inlet at which the device is set to start opening under service conditions.

Thermal Inbreathing

Movement of air or blanketing gas into a tank when vapors in the tank contract or condense as a result of weather changes (e.g. a decrease in atmospheric temperature).

Thermal Outbreathing

Movement of vapors out of a tank when vapors in the tank expand and/or liquid in the tank vaporizes as a result of weather changes (e.g. an increase in atmospheric temperature).

Vapor Pressure

The pressure exerted when a liquid is in equilibrium with its own vapor. Vapor pressure is a function of the substance and temperature.

Wetted Area

Surface area of a tank exposed to liquid on the interior and heat from a fire on the exterior.

➤ Causes of Overpressure or Vacuum in Tanks

General

When determining the possible causes of overpressure or vacuum in a tank, consider the following:

- Liquid movement into or out of the tank;
- Weather changes (e.g. pressure and temperature changes);
- Fire exposure;
- Other circumstances resulting from equipment failures and operating errors.

Liquid movement into or out of a tank

Liquid can enter or leave a tank by pumping, by gravity flow, or by process pressure.

Vacuum can result from the outflow of liquid from a tank. Overpressure can result from the inflow of liquid into a tank and from the vaporization or flashing of the feed liquid. The user is cautioned that a flashing feed stream can create an outbreathing load that is much greater than the volumetric inflow of liquid.

Weather Changes

Vacuum can result from the contraction or condensation of vapors caused by a decrease in atmospheric temperature or other weather changes, such as wind changes, hailstorm, rainstorm, etc. Overpressure can result from the expansion and vaporization that is caused by an increase in atmospheric temperature or weather changes.

Fire Exposure

Overpressure can result from the expansion of the vapors and vaporization of the liquid that may occur when a tank absorbs heat from an external fire.

Other Circumstances

When the possible causes of overpressure or vacuum in a tank are being determined, other circumstances resulting from equipment failures and operating errors shall be considered and evaluated. Overpressure or Vacuum protection due to such other circumstances needs be done on a case-to-case basis and the overpressure or vacuum protection device to be selected accordingly. For description of these other overpressure and vacuum scenarios, students are recommended to refer API STD 2000, 7th Edition.

Normal Venting Requirements

- Normal inbreathing resulting from a maximum outflow of liquid from the tank (liquid-transfer effects),
- Normal inbreathing resulting from contraction or condensation of vapors caused by a maximum decrease in vapor-space temperature (thermal effects),
- Normal out-breathing resulting from a maximum inflow of liquid into the tank and maximum vaporization caused by such inflow (liquid-transfer effects),
- Normal out-breathing resulting from expansion and vaporization that results from a maximum increase in vapor-space temperature (thermal effects),

Total inbreathing requirement will be the sum of the inbreathing due to liquid outflow (also known as pump-out) and the thermal inbreathing as defined in b) above.

Total outbreathing requirement will be the sum of the outbreathing due to liquid inflow (also known as pump-in) and the thermal inbreathing as defined in d) above.

Outbreathing Calculations

The out-breathing shall be determined as follows. In these calculations, the vapor/gas being displaced will be at the actual pressure and temperature conditions of the tank vapor space. Out-breathing flows shall be converted to an air-equivalent flow at normal or standard conditions for tanks operating above 49 °C.

Non Volatile Liquids

Non-volatile liquids are products with a vapor pressure equal to or less than 5.0 kPa.

The out-breathing volumetric flow rate V_{op} , expressed in SI units of cubic meters per hour of vapor/gas at the actual pressure and temperature conditions of the tank vapor space, shall be as given by

Equation (1):

$$V_{op} = V_{pf} \text{ ----- (1)}$$

Where:

V_{pf} = Maximum volumetric filling rate (pump-in) of non-volatile liquid in the tank, m³/h

Volatile Liquids

Volatile liquids are products with a vapor pressure greater than 5.0 kPa. The flow of volatile liquids into a tank will result in higher out-breathing flow (compared to the same inflow with a non-volatile liquid) due to changes in liquid-vapor equilibrium.

The out-breathing volumetric flow rate V_{op} expressed in SI units of cubic meters per hour of vapor/gas at the actual pressure and temperature conditions of the tank vapor space, shall be as given by

Equation (2):

$$V_{op} = 2.0 * V_{pf} \text{ ----- (2)}$$

Where:

V_{pf} = Maximum volumetric filling rate (pump-in) of volatile liquid in the tank, m³/h

Flashing Liquids

Flashing liquids can cause the venting requirement to be many times greater than the Volumetric in-flow of the liquid. Flashing will occur when the vapor pressure of the entering stream is greater than the operating pressure of the tank. For products that can flash due to high

temperature or because of dissolved gases (e.g. oil spiked with methane), perform an equilibrium flash calculation and increase the out-breathing venting requirements accordingly.

Thermal Outbreathing

The thermal out-breathing (i.e. the maximum thermal flow rate for heating up) VOT expressed in SI units of normal cubic meters per hour of air, shall be given as per Equation (3):

$$V_{OT} = Y \cdot V_{tk} \cdot 0.9 \cdot R_i \text{ -----(3)}$$

Where:

Y = Factor for latitude (*Refer terms, definitions and abbreviations for definition of latitude*)

V_{tk} = Tank volume, m³

R_i = Reduction factor for insulation

$R_i = 1$, if tank is not insulated

$R_i = R_{inp}$, if tank is partially insulated (*Refer eqn 5 below*)

$R_i = R_{in}$, if tank is fully insulated (*Refer eqn 4 below*)

The Y-factor for the latitude in Equations (3) can be taken from Table 1

Latitude	Y-factor
Below 42°	0.32
Between 42° and 58°	0.25
Above 58°	0.2

$$R_{in} = \frac{1}{1 + \frac{h \cdot l_{in}}{\lambda_{in}}} \text{ -----(4)}$$

Where:

h = Inside heat-transfer coefficient, W/m²-K

NOTE An inside heat-transfer coefficient of 4 W/(m²-K) is commonly assumed for typical tanks.

l_{in} = Wall thickness of the insulation, m

λ_{in} = Thermal conductivity of the insulation, W/m-K

$$R_{inp} = \frac{A_{inp}}{A_{TTS}} \cdot R_{in} + \left(1 - \frac{A_{inp}}{A_{TTS}} \right) \text{ -----(5)}$$

Where:

A_{TTS} = Total tank surface area (shell and roof), m²

A_{inp} = Surface area of the insulated part of the tank, m²

Total Outbreathing

The total outbreathing is the sum of the outbreathing calculated due to pump-in (equations 1 or 2), vapor generated due to flashing (flash calculations), and the thermal outbreathing.

Inbreathing Calculations

The inbreathing venting requirement V_{ip} , expressed in SI units of normal cubic meters per hour of air, shall be the maximum specified liquid discharging capacity for the tank as given by Equation (6).

$$V_{ip} = V_{pe} \text{ ----- (6)}$$

Where:

V_{pe} = maximum volumetric rate of discharge (*pump-out*) from the tank, m^3/h

Thermal Inbreathing

The thermal in-breathing (i.e. the maximum thermal flow rate during cooling down) V_{IT} expressed in SI units of normal cubic meters per hour of air, shall be given as per Equation (7):

$$V_{IT} = C * V_{tk} * 0.7 * R_i \text{ ----- (7)}$$

Where:

C = factor that depends upon vapor pressure, average and latitude (Refer table below)

C-factors				
Latitude	C-factor for various conditions			
	Vapor pressure similar to Hexane		Vapor pressure higher than hexane, or unknown	
	Average storage temperature °C			
	< 25	≥ 25	< 25	≥ 25
Below 42°	4	6.5	6.5	6.5
Between 42° and 58°	3	5	5	5
Above 58°	2.5	4	4	4

V_{tk} and R_i are as defined in the Thermal Outbreathing section.

The calculated inbreathing assumes ambient air flow through the tank vent. It is typical practice to assume the ambient air is at normal or standard conditions.

Total Inbreathing

The total inbreathing is the sum of the inbreathing calculated due to pump-out (equation 6) and the thermal outbreathing.

Converting Tank Outbreathing Vapor flow to Nm³/h of Air flow

Capacities for relief devices are expressed as air at either standard or normal conditions. In order to provide reasonable accuracy in the sizing of relief devices for storage tanks it may be necessary to adjust the calculated relief requirements into equivalent air flow.

If the maximum operating temperature of the tank is ≤49°C, then the out-breathing vapor volumetric flow rate V_{op} as calculated from equations (1) and (2) can be directly considered as volume flow rate of air in Nm³/h and specified as such for the tank overpressure relief device.

However, if the tank maximum operating temperature is >49°C then the outbreathing vapor flow V_{op} needs to be corrected for the Nm³/h of air flow using equation (8).

$$q_{air} = 0.326 \times V_{op} \times \sqrt{\frac{T_i}{M}} \text{----- (8)}$$

Where:

q_{air} = Equivalent Air flow rate, Nm³/h

T_i = Temperature of the vapor in the tank vapor space (>49 °C), K

M = Molecular weight of the vapor

NOTES

- 1) For practical purposes it is assumed that the vapor in the tank vapor space is in equilibrium with the liquid in the tank and the vapor molecular weight is the same as the liquid molecular weight
- 2) The constant 0.326 in the above equation is derived from air properties and normal temperature of 0 °C (273.15 K). The factor “Sq. root of (M_{air} / T_{air})” when inserted values of 28.97 (mol. wt. of air) with 273.15 K as air temperature provides the constant value.

Emergency Venting due to External Fire

When storage tanks are exposed to fire, the venting rate can exceed the rate resulting from normal outbreathing.

Tanks with Weak Roof-to-shell Attachment

On a fixed-roof tank with a weak (frangible) roof-to-shell attachment, such as that described in API 650, the roof-to-shell connection will fail prior to other tank welds, allowing relief of the excess pressure if the normal venting capacity proves inadequate. For a tank built to these specifications, it is not necessary to consider additional requirements for emergency venting; however, additional emergency vents may be used to avoid failure of the frangible joint. Care should be taken to ensure that the current requirements for a frangible roof-to-shell attachment are met, particularly for tanks smaller than 15 m (50 ft) in diameter.

Fire Relief Requirements

When a tank is not provided with a Weak Roof-to-Shell attachment the procedure and equations given below shall be used in evaluating the required relief flow capacity for fire exposure.

The required flow capacity q , expressed in SI units of Nm³/h of air, for tanks subject to fire exposure as given by Equation (9):

$$q = 906.6 \times \left(\frac{Q \times F}{L} \right) \times \left(\frac{T}{M} \right)^{0.5} \text{----- (9)}$$

Where:

Q = the heat input from fire exposure as given in table below, W

Heat Input, Q
(Expressed in SI Units)

Wetted Surface Area, A_{TWS} m ²	Design Pressure kPa (gauge)	Heat Input, Q W
<18.6	≤103.4	$63,150 A_{TWS}$
≥18.6 and <93	≤103.4	$224,200 \times (A_{TWS})^{0.566}$
≥93 and <260	≤103.4	$630,400 \times (A_{TWS})^{0.338}$
≥260	>7 and ≤103.4	$43,200 \times (A_{TWS})^{0.82}$
≥260	≤7	4,129,700

F = the environmental factor from table below (credit may be taken for only one environmental factor)

L = the latent heat of vaporization of the stored liquid at the relieving pressure and temperature, J/kg

T = the absolute temperature of the relieving vapor, K

NOTE:

It is normally assumed that the temperature of the relieving vapor corresponds to the bubble point of the stored fluid at the relieving pressure

M = the molecular weight of the vapor

The fire relief requirements presented are expressed as air-equivalent flow rates based on the conversion of the actual vapor generation rate into an equivalent volume of air (at either normal or standard conditions).

Calculation of Latent Heat of Vaporization

The latent heat of vaporization of pure single-component liquid decreases as the temperature at vaporization increases and becomes zero at the critical temperature and pressure for that liquid.

The Figure A-6 provided below shows the vapor pressure and latent heats of the pure, single-component paraffin hydrocarbon liquids. This chart is directly applicable to such liquids and applies as an approximation to paraffin-hydrocarbon mixtures composed of two components whose molecular weights vary no more than by the difference between propane and butane or butane and pentane. Reference is made to Section A.4, Annex A of API STD 521, 6th edition, 2014.

The chart can also be applicable to isomer hydrocarbons, aromatic or cyclic compounds, or paraffin-hydrocarbon mixtures that have slightly different molecular weights. The equilibrium temperature needs to be calculated. Using the relationship for the calculated temperature versus vapor pressure, one can obtain the value of latent heat from the figure in the next page. The molecular weight relationship as shown in the chart is not to be used in such cases. In such cases, the molecular weight of the vapor should be determined from the vapor-liquid equilibrium by performing a flash calculation at that temperature and pressure.

For cases that involve mixture of components that have a wide boiling range or significantly different molecular weights a rigorous series of equilibrium calculations can be required to estimate the latent heat.

Near the critical point, if no accurate latent heat value is available for hydrocarbon mixtures a minimum value of 115 kJ/kg may be used as an approximation.

When using HYSYS as a simulation tool the liquid composition is used from HYSYS simulation results and a 10% flash calculation is simulated in HYSYS at the relief pressure (relief valve set pressure + overpressure) and corresponding bubble temperature. The latent heat obtained thus is used in the calculations.

In some cases three stages of flash each of 10% is used and the lowest latent heat of vaporization value and lowest molecular weight vapor stream is used for estimating the relief load.

In the case water is present along with the hydrocarbons; it is recommended that the water free composition is used in determining the latent heat since this will result in more conservative design.

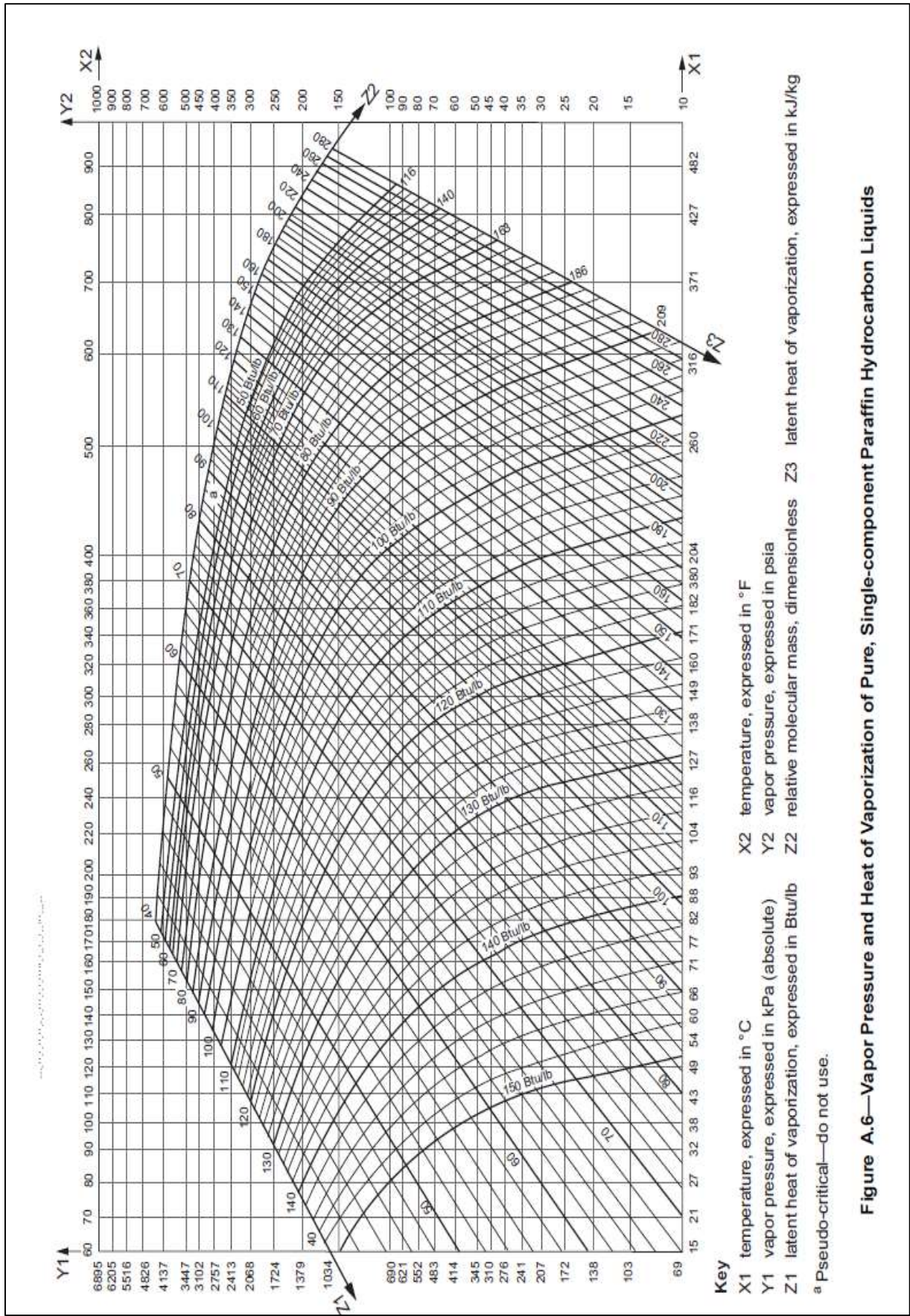


Figure A.6—Vapor Pressure and Heat of Vaporization of Pure, Single-component Paraffin Hydrocarbon Liquids

Environmental Factors for Nonrefrigerated Aboveground Tanks
(Expressed in SI and USC Units)

Tank Design/Configuration	Insulation Conductance		Insulation Thickness		<i>F</i> -factor ^b
	W/m ² ·K	Btu/(h·ft ² ·°F)	cm	in.	
Bare metal tank	—	—	0	0	1.0
Insulated tank ^a	22.7	4.0	2.5	1	0.3 ^b
	11.4	2.0	5.1	2	0.15 ^b
	5.7	1.0	10.2	4	0.075 ^b
	3.8	0.67	15.2	6	0.05 ^b
	2.8	0.5	20.3	8	0.0375 ^b
	2.3	0.4	25.4	10	0.03 ^b
	1.9	0.33	30.5	12	0.025 ^b
Concrete tank or fireproofing	—		—		^c
Water-application facilities ^d	—		—		1.0
Depressuring and emptying facilities ^e	—		—		1.0
Underground storage	—		—		0
Earth-covered storage above grade	—		—		0.03
Impoundment away from tank ^f	—		—		0.5

^a The insulation shall resist dislodgment by fire-fighting equipment, shall be noncombustible and shall not decompose at temperatures up to 537.8 °C (1000 °F). The user is responsible for determining whether the insulation can resist dislodgment by the available fire-fighting equipment. If the insulation does not meet these criteria, no credit for insulation shall be taken. The conductance values are based on insulation with a thermal conductivity of 9 W/m²·K/cm (4 Btu/h·ft²·°F/in.) of thickness. The user is responsible for determining the actual conductance value of the insulation used. The conservative value of 9 W/m²·K/cm (4 Btu/h·ft²·°F/in.) of thickness for the thermal conductivity is used.

^b These *F*-factors are based on the thermal conductance values shown and a temperature differential of 887.9 K (1600 °F) when using a heat input value of 66,200 W/m² (21,000 Btu/h·ft²) in accordance with the conditions assumed in ISO 23251. When these conditions do not exist, engineering judgment should be used to select a different *F*-factor or to provide other means for protecting the tank from fire exposure.

NOTE For the purposes of this provision, API 521 is equivalent to ISO 23251.

^c Use the *F*-factor for an equivalent conductance value of insulation.

^d Under ideal conditions, water films covering metal surfaces can absorb most incident radiation. The reliability of water application depends on many factors. Freezing weather, high winds, clogged systems, undependable water supply, and tank surface conditions can prevent uniform water coverage. Because of these uncertainties, no reduction in environmental factors is recommended; however, as stated previously, properly applied water can be very effective.

^e Depressuring devices may be used, but no credit shall be allowed in sizing the venting device for fire exposure.

^f The following conditions shall be met.

- A slope of not less than 1 % away from the tank shall be provided for at least 15 m (50 ft) toward the impounding area.
- The impounding area shall have a capacity that is not less than the capacity of the largest tank that can drain into it.
- The drainage system routes from other tanks to their impounding areas shall not seriously expose the tank.
- The impounding area for the tank, as well as the impounding areas for the other tanks (whether remote or with dikes around the other tanks), shall be located so that when the area is filled to capacity, its liquid level is no closer than 15 m (50 ft) to the tank.

^g Local regulations may specify or allow different *F*-factors. For example, OSHA 1910.106 allows different credit for water spray and insulation.

Where fluid properties are similar to those of hexane, the required venting capacity can be Determined from Table 5 of API STD 2000 as given below:

Table 5—Venting Capacity (Expressed in SI Units)		
Wetted Surface Area, A_{TWS}^a m ²	Design Pressure kPa (gauge)	Required Venting Capacity Nm ³ /h of air
<260	≤103.4	See Table 7 and 3.3.3.3.4
≥260	≤7	19,910 (see 3.3.3.3.4)
≥260	>7 and ≤103.4	Use Equation (16) ^b
^a The wetted area of a tank or storage vessel shall be calculated as follows. — For spheres and spheroids, the wetted area is equal to 55 % of the total surface area or the surface area to a height of 9.14 m above grade, whichever is greater. — For horizontal tanks, the wetted area is equal to 75 % of the total surface area or the surface area to a height of 9.14 m above grade, whichever is greater. — For vertical tanks, the wetted area is equal to the total surface area of the vertical shell to a height of 9.14 m above grade. For a vertical tank setting on the ground, the area of the ground plates is not included as wetted area. For a vertical tank supported above grade, it is necessary to include a portion of the area of the bottom as additional wetted surface. The portion of the bottom area exposed to a fire depends on the diameter and elevation of the tank above grade. It is necessary to use engineering judgment in evaluating the portion of the area exposed to fire. ^b Calculate the venting requirement q, expressed in normal cubic meters per hour of air as given in Equation (16), which is based on the total heat absorbed Q, expressed in watts, equal to $43,200 A_{TWS}^{0.82}$ [see Equation (B.7)]: $q = 208.2 F A_{TWS}^{0.82} \quad (16)$ where F is the environmental factor from Table 9 (credit may be taken for only one environmental factor); A_{TWS} is the wetted surface area, expressed in square meters.		

The total heat absorbed, Q , is expressed in watts for Equation (16). Table 7 and the constant 208.2 in Equation (16) are derived from Equation (14) and Figure B.1 of Annex B of API STD 2000 by using the latent heat of vaporization of hexane, equal to 334,900 J/kg at atmospheric pressure, and the relative molecular mass of hexane (86.17) and assuming a vapor temperature of 15.6 °C. This method provides results within an acceptable degree of accuracy for many fluids having similar properties

Table 7 and Clause 3.3.3.3.4 of API STD 2000 are shown below:

Table 7—Emergency Venting Required for Fire Exposure vs Wetted Surface Area
(Expressed in SI Units)

Wetted Area ^a m ²	Venting Required Nm ³ /h		Wetted Area ^a m ²	Venting Required Nm ³ /h
2	608		35	8,086
3	913		40	8,721
4	1,217		45	9,322
5	1,521		50	9,895
6	1,825		60	10,971
7	2,130		70	11,971
8	2,434		80	12,911
9	2,738		90	13,801
11	3,347		110	15,461
13	3,955		130	15,751
15	4,563		150	16,532
17	5,172		175	17,416
19	5,780		200	18,220
22	6,217		230	19,102
25	6,684		260	19,910
30	7,411		>260 ^b	—

^a The wetted area of a tank or storage vessel shall be calculated as follows.

- For spheres and spheroids, the wetted area is equal to 55 % of the total surface area or the surface area to a height of 9.14 m above grade, whichever is greater.
- For horizontal tanks, the wetted area is equal to 75 % of the total surface area or the surface area to a height of 9.14 m above grade, whichever is greater.
- For vertical tanks, the wetted area is equal to the total surface area of the vertical shell to a height of 9.14 m above grade. For a vertical tank setting on the ground, the area of the ground plates is not included as wetted area. For a vertical tank supported above grade, it is necessary to include a portion of the area of the bottom as additional wetted surface. The portion of the bottom area exposed to a fire depends on the diameter and elevation of the tank above grade. It is necessary to use engineering judgment in evaluating the portion of the area exposed to fire.

^b For wetted surfaces larger than 260 m², see Table 5.

NOTE This table and the constant 208.2 in Equation (16) are derived from Equation (14) and Figure B.1 by using the latent heat of vaporization of hexane, equal to 334,900 J/kg at atmospheric pressure, and the relative molecular mass of hexane (86.17) and assuming a vapor temperature of 15.6 °C. This method provides results within an acceptable degree of accuracy for many fluids having similar properties (see Annex B).

3.3.3.3.4 The total rate of venting determined from Table 7 or Table 8 may be multiplied by an appropriate environmental factor, *F*, selected from Table 9; credit may be taken for only one environmental factor.

Refer Annex B “Basis of Emergency Venting for Table 7 and Table 8” of API STD 2000

➤ Means of Venting

Normal Venting

Normal venting for pressure and vacuum shall be accomplished by a PV valve with or without a flame arresting device or by an open vent with or without a flame-arresting device.

Protect atmospheric storage tanks against flame transmission from outside the tank if

- The stored liquid has a low flash point, i.e. less than 60 °C (140 °F) or in accordance with the applicable regulations, whichever is higher; or
- The storage temperature can exceed the flash point; or
- The tank can otherwise contain a flammable vapor space.

A discussion of the types and operating characteristics of venting devices can be found in Annex C of API STD 2000.

PV valves are recommended to be used on atmospheric storage tanks to avoid product loss.

If open vents are selected to provide venting capacity for tanks that can contain a flammable vapor space as defined above, a flame-arresting device should be used. Open vents without a flame-arresting device may be used for tanks that do not contain a flammable vapor space.

In the case of viscous oils, such as cutback and penetration-grade asphalts, where the danger of tank collapse resulting from sticking pallets or from plugging of flame arresters is greater than the possibility of flame transmission into the tank, open vents may be used as an exception or heat traced vents that ensure that the vapor temperature stays above the dew point may be used.

In areas with strict fugitive emissions regulations, open vents might not be acceptable and vent-device selection should consider the maximum leakage requirements during periods of normal tank operation.

Emergency Venting

Emergency venting may be accomplished by the use of the following:

- a) Larger or additional open vents
- b) Larger or additional PV valves,
- c) A gauge hatch that permits the cover to lift under abnormal internal pressure,
- d) A manhole cover that lifts when exposed to abnormal internal pressure,
- e) Weak (frangible) roof-to-shell attachment
- f) A rupture-disk device,
- g) Other forms of construction that can be proven to be comparable for the purposes of pressure relief.

➤ **Basic Principles of Operation of Pressure-Vacuum Relief Valves (PVRVs)**

General

Two basic types of PVRVs, direct-acting vent valves and pilot-operated vent valves, are available to provide overpressure or vacuum protection for low-pressure storage tanks. Direct-acting vent valves can be weight-loaded or spring-loaded. These venting devices not only provide overpressure protection but also conserve product. Direct-acting vent valves are sometimes referred to as conservation vents.

Direct-Acting Vent Valves (PVRVs)

Direct-acting vent valves are available to provide pressure relief, vacuum relief, or a combination of pressure and vacuum relief. Combination vent valves may be of a side-by-side configuration (see Figure C.2 referenced from API STD 2000, Annex C). Side-by-side vent valves or pressure-relief vent valves are available with flanged outlets for pressure discharge when it is necessary to pipe the pressure-relief vapors away.

Larger, direct-acting vent valves are available to provide emergency relief and can provide access to a tank's interior for inspection or maintenance. They are typically available in sizes from 400 mm (16 in.) to 600 mm (24 in.) (See Figure C.3 referenced from API STD 2000, Annex C).

The principle of operation of a direct-acting vent valve is based on the weight of the pallet or the spring force acting on the pallet to keep the device closed. When tank pressure or vacuum acting on the seat sealing area equals the opposing force acting on the pallet, the venting device is on the threshold of opening. Any further increase in pressure or vacuum causes the pallet to begin to lift off the seat.

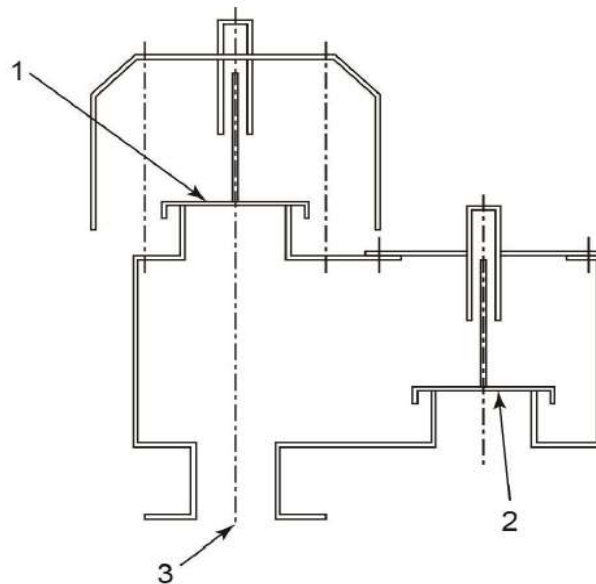
Some weight loaded PVRVs require significant overpressure (or vacuum) to achieve the required flows and as a result may need to be set well below the design pressure/design vacuum of the tank. The user should consult the PVRV manufacturer to verify that the PVRV will have sufficient capacity to keep tank pressure/vacuum within limits specified by the tank's design code of construction. Where the overpressure/vacuum required for the PVRV to satisfy the tank flow requirements exceeds the pressure/vacuum limits of the storage tank, a larger PVRV or multiple PVRVs can be used at reduced lift and provide sufficient capacity. Fewer large PVRVs instead of more small PVRVs are usually preferred to minimize the number of tank penetrations.

Direct-acting vent valves are typically available in sizes from 50 mm (2") to 350 mm (14"); however, vent valves in a stacked configuration are available in sizes up to 700 mm (28").

Typical set pressure ranges for weight-loaded vent valves are up to 6.9 kPag and up to a vacuum of -4.3 kPag. Spring-loaded vent valves should be used for pressure or vacuum settings that exceed these values because the supporting structure and space for the added weights is not available.

Verification of the set pressure of a venting device after it has been installed on a storage tank can be accomplished by increasing the tank pressure or vacuum. To change the set pressure, weights

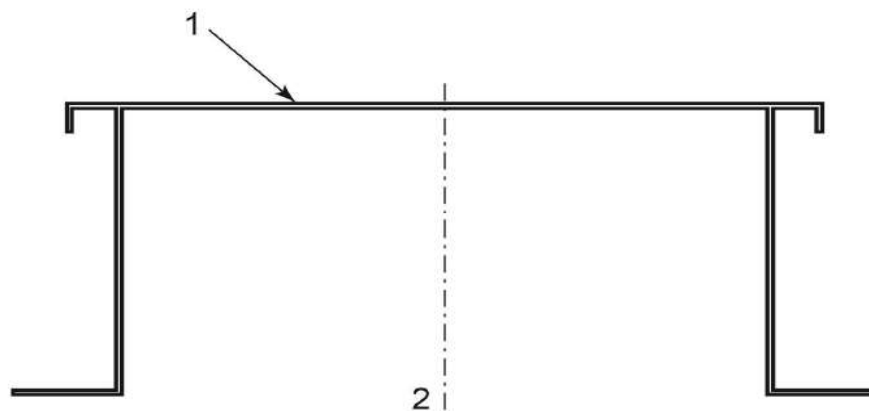
shall be added or removed from the pallet, or a new pallet shall be used, or the spring shall be adjusted (if a spring-loaded vent valve is being used).



Key

- 1 pressure pallet
- 2 vacuum pallet
- 3 tank connection

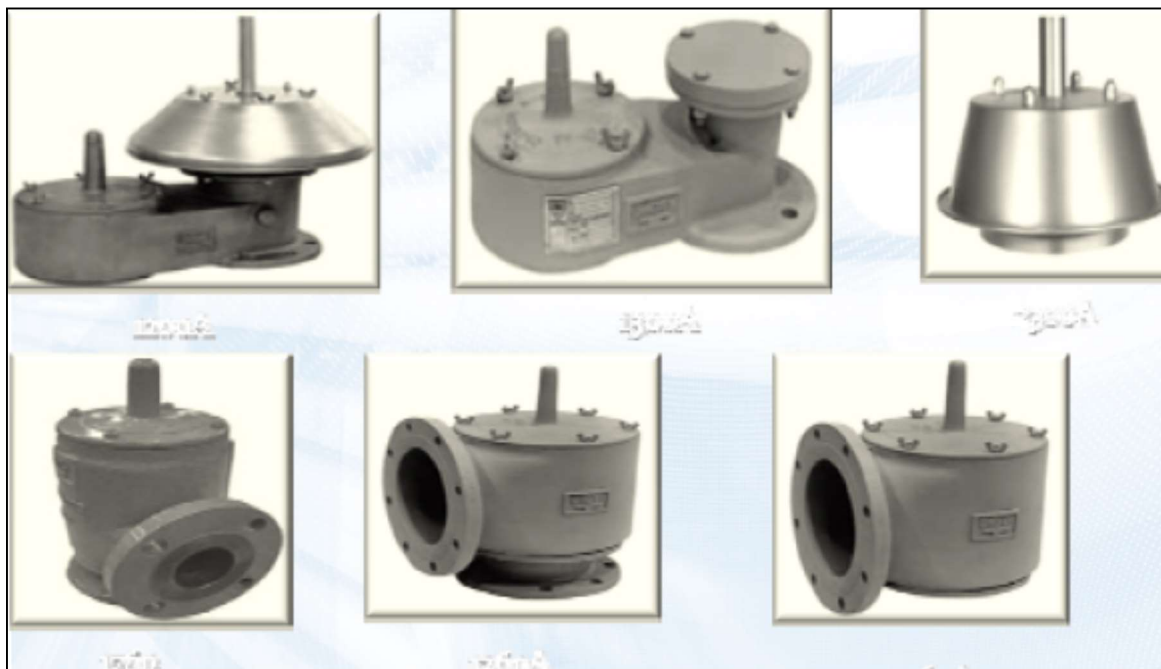
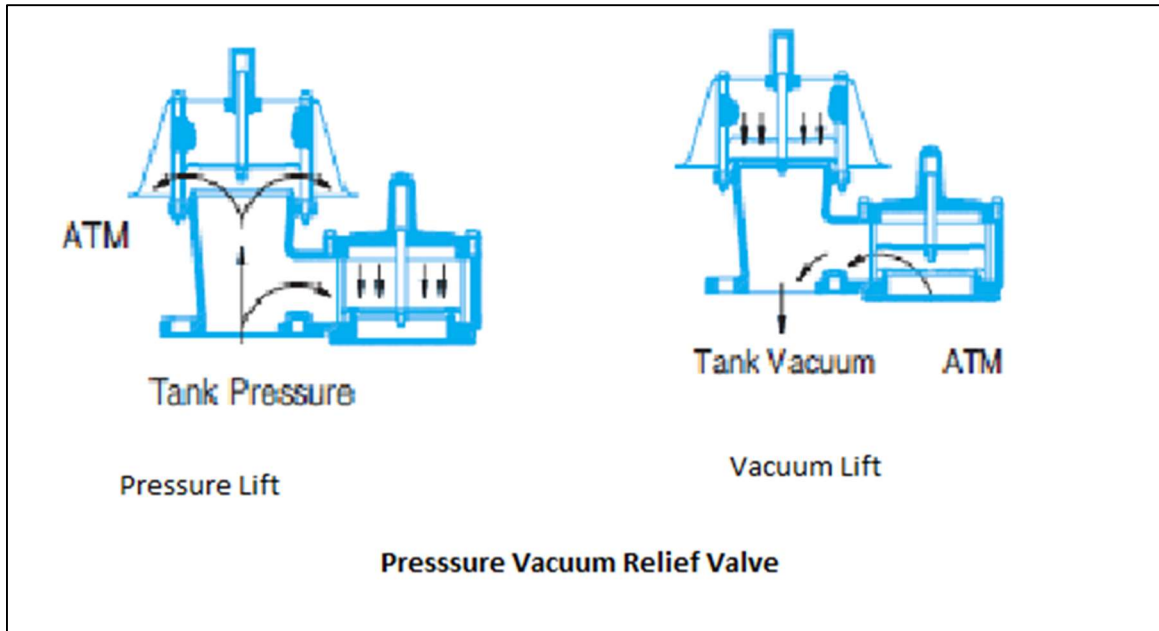
Figure C.2—Side-by-side Pressure/Vacuum Vent



Key

- 1 pressure pallet
- 2 tank connection

Figure C.3—Large, Weight-loaded Emergency Vent



Pressure Vacuum Relief Valves

Pilot-Operated Vent Valves (Pilot Operated PVRVs)

Pilot-operated vent valves are available to provide pressure relief, vacuum relief, or a combination of pressure and vacuum relief. Some vent valves can be equipped with flanged outlets if it is necessary to pipe pressure-relief vapors away. Unlike side-by-side direct-acting vent valves, pilot-operated vent valves relieve pressure or vacuum through the same opening to atmosphere (see Figure C.6 referenced from Annex C, API STD 2000).

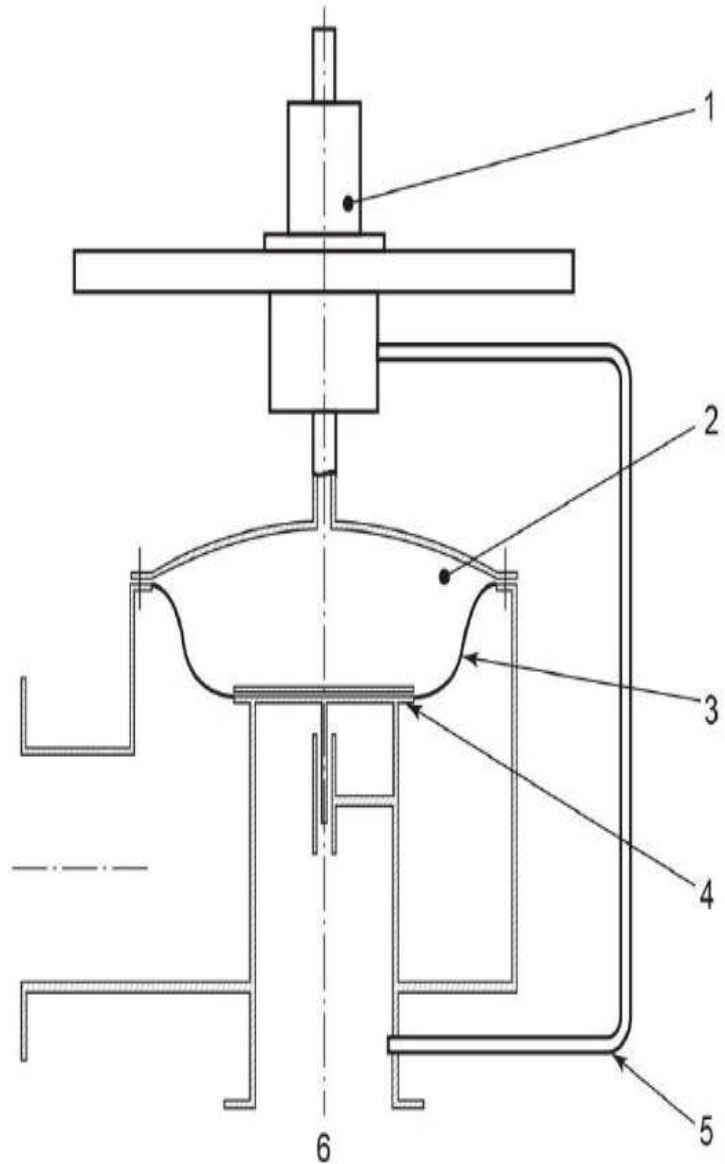
A pilot-operated vent valve for pressure relief uses tank pressure, not weights or a spring, to keep the vent valve seat closed. The main seat is held closed by tank pressure acting on a large-area diaphragm. This tank pressure covers an area greater than the seat sealing area, so the net pressure force is always in a direction to keep the seat closed. The volume above the diaphragm is called the dome. If the diaphragm fails, the dome pressure decreases and the vent valve opens.

The pilot is a small control valve that continuously senses tank pressure. When the tank pressure increases to set pressure, the pilot actuates to reduce the pressure in the dome volume, the force holding the seat closed is reduced and the seat lifts to permit tank pressure to discharge through the vent valve. When the tank pressure decreases, the pilot closes, the dome volume re-pressurizes, and the main seat closes. Two types of pilot actions are available: modulating and snap action. For modulating action, the main vent valve opens gradually with increasing pressure and achieves rated relieving capacity at relieving pressure. Modulating valves reseal very close to the set pressure. For snap action, the main valve opens rapidly at set pressure and achieves rated relieving capacity at relieving pressure.

A pilot-operated vent valve achieves full lift at or below 10 % overpressure (see Figure C.5 referenced from Annex C, API STD 2000). This lift characteristic permits overpressure protection to be accomplished with smaller or fewer venting devices. In addition, relative to direct-acting vent valves, pilot-operated vent valves can have a tank-operating pressure closer to the set pressure.

A pilot-operated vent valve for vacuum relief uses atmospheric pressure to keep the seat closed. The force holding the seat closed is equal to the seat sealing area times the pressure differential across the seat. This pressure differential is equal to atmospheric pressure plus the tank vacuum. When the tank vacuum equals the pilot set, the pilot opens to apply tank vacuum to the large dome volume above the diaphragm. Atmospheric pressure acting on the downstream side of the diaphragm forces the diaphragm and seat up. Little or no increase in tank vacuum beyond the vent valve setting is required to obtain full lift of the seat. When the tank vacuum decreases, the pilot closes and atmospheric pressure enters the dome to close the main seat.

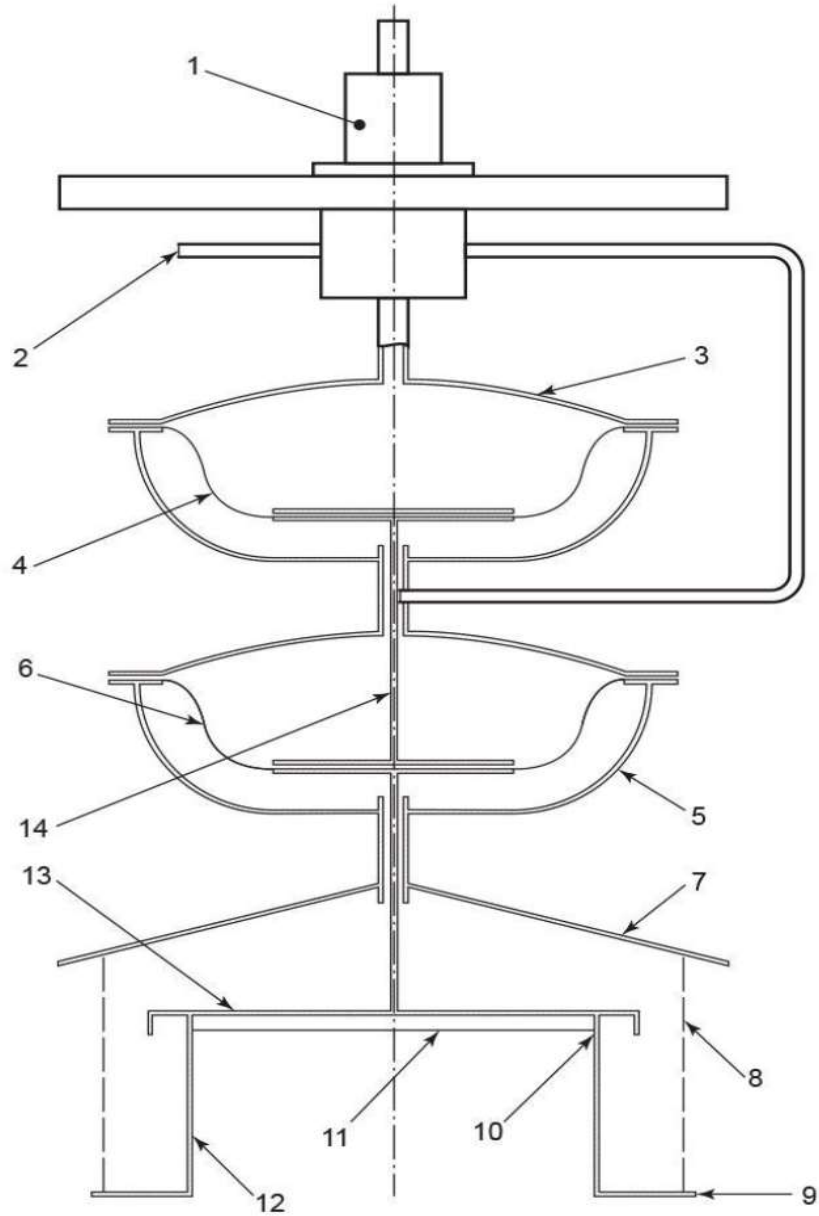
If the diaphragm fails, atmospheric air enters the dome and prevents the tank vacuum from creating a force differential to lift the seat. Double-diaphragm vent valves are available to prevent such a failure (see Figure C.6 referenced from Annex C, API STD 2000): one diaphragm is for pressure actuation and one is for vacuum actuation. Each diaphragm is isolated and protected from the flow stream and fully supported to minimize stress. The vacuum diaphragm moves only to provide vacuum relief to extend its service life.



Key

- 1 pilot
- 2 dome
- 3 diaphragm
- 4 seat
- 5 pressure probe
- 6 tank connection

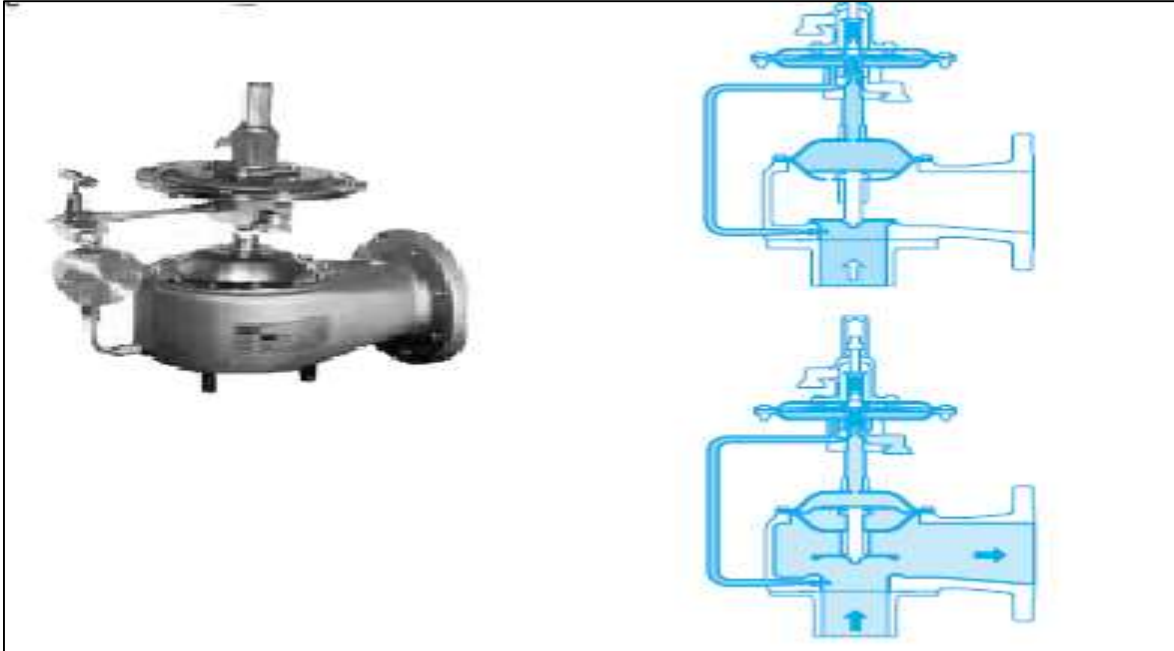
Figure C.5—Pilot-operated Pressure Vent (Single Diaphragm)



Key

- | | |
|------------------------------|-------------------------|
| 1 pilot | 8 screen |
| 2 pilot inlet | 9 base flange |
| 3 vacuum actuator assembly | 10 film seat |
| 4 vacuum diaphragm | 11 film protector plate |
| 5 pressure actuator assembly | 12 nozzle |
| 6 pressure diaphragm | 13 seat plate |
| 7 shield | 14 vacuum rod |

Figure C.6—Pilot-operated Pressure/Vacuum (Double Diaphragm)



Low-pressure, pilot-operated vent valves are typically available in sizes from 50 mm (2 in.) to 600 mm (24 in.). Available set pressures range from 103.4 kPag to –101.3 kPag vacuum. The minimum opening pressure is typically a 0.5 kPag (2" wcg) or –0.5 kPag (–2" wcg) vacuum.

4.8 CONTROL VALVE

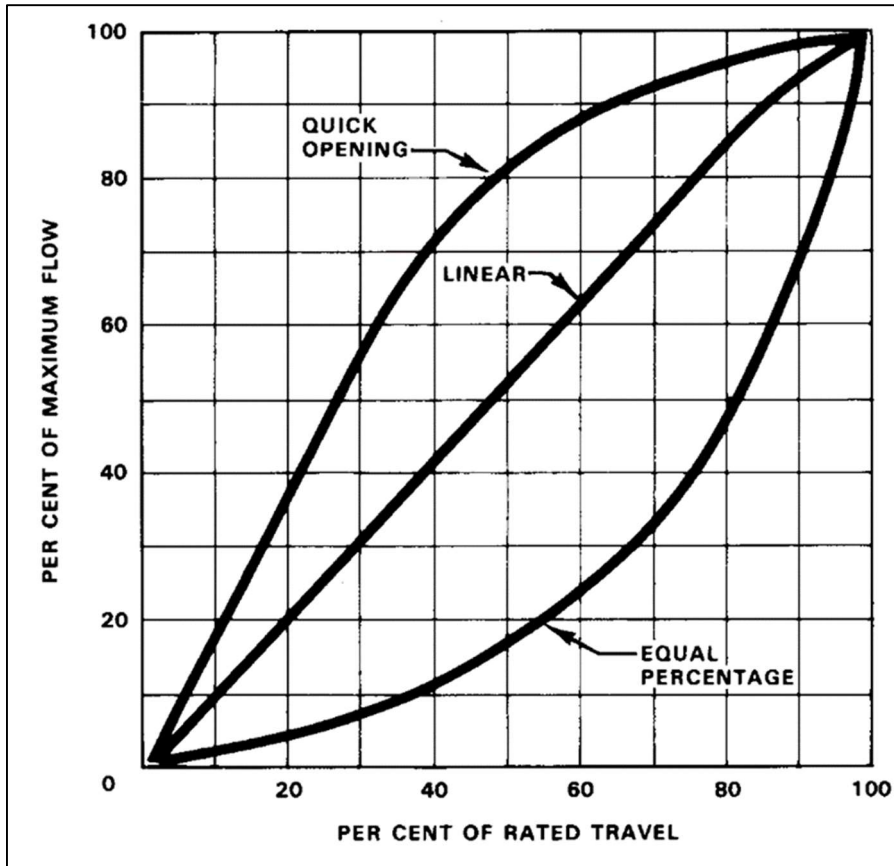
General

- Valve shall be sized for maximum, minimum, and normal flow at corresponding differential pressures and process conditions.
- Calculated noise level shall not exceed 85 dB.
- Final valve sizing shall be confirmed by the valve vendor.

Pressure Drop Criteria

For initial sizing purpose of control valves in pumped liquid lines, a general guideline is to consider a pressure drop of 30% of the total system pressure drop or 0.7 bar whichever is higher.

For control valves on gas lines the minimum pressure drop specified shall not be less than 0.2 bar at design flow unless otherwise dictated by specific process requirements.



Equal Percentage Characteristics:

An inherent flow characteristic that, for equal increments of rated travel, will ideally give equal percentage changes of the flow coefficient C_v .

Linear Characteristic:

An inherent flow characteristic that can be represented by a straight line on a rectangular plot of flow coefficient, C_v versus rated travel. Therefore equal increments of travel provide equal increments of flow coefficient, C_v .

Quick Opening Characteristic:

An inherent flow characteristic in which a maximum flow coefficient is achieved with minimal closure member travel.

Guideline for characteristic selection:

PRESSURE CONTROL SYSTEM	
Service	Best characteristic
Liquid Process	Equal %
Gas Processes: Small volume - Less than 10 ft of pipe between control valve and flow destination	Equal %
Gas Processes: Large volume - Receiver volume tank in system; Greater than 100 ft of pipe volume. Decreasing ΔP with increasing volume. ΔP at max. load > 20% of min. load	Linear
Gas Processes: Large volume - Decreasing ΔP with increasing volume; ΔP at max. load < 20% of min. load.	Equal %

LIQUID LEVEL CONTROL SYSTEM	
Service	Best characteristic
Constant ΔP	Linear
Decreasing ΔP with increasing flow. ΔP at max. flow > 20% of min. flow ΔP	Linear
Decreasing ΔP with increasing flow. ΔP at max. flow < 20% of min. flow ΔP	Equal %
Increasing ΔP with increasing flow. ΔP at max. flow < 200% of min. flow ΔP	Linear
Increasing ΔP with increasing flow. ΔP at max. flow > 200% of min. flow ΔP	Quick Opening

MISCELLANEOUS SYSTEM	
Service	Best characteristic
3 way valve or 2 way valve used as a 3 way valve	Linear
Compressor gas recycle	Linear
ΔP = constant	Linear
Temperature control valve ΔP < 50% of total system ΔP	Equal %
pH control: ΔP < 50% of system ΔP	Equal %
pH control: ΔP > 50% of system ΔP	Linear
Heating - Cooling	Linear
Split - ranged	Linear
On - Off	Quick Opening

4 Symbols

Symbol	Description	Unit
C	Flow coefficient (K_v , C_v)	Various (see IEC 60534-1) (see Note 4)
d	Nominal valve size	mm (in)
D	Internal diameter of the piping	mm (in)
D_1	Internal diameter of upstream piping	mm (in)
D_2	Internal diameter of downstream piping	mm (in)
D_o	Orifice diameter	mm (in)
F_d	Valve style modifier (see Annex A)	Dimensionless (see Note 4)
F_F	Liquid critical pressure ratio factor	Dimensionless
F_L	Liquid pressure recovery factor of a control valve without attached fittings	Dimensionless (see Note 4)
F_{LP}	Combined liquid pressure recovery factor and piping geometry factor of a control valve with attached fittings	Dimensionless
F_P	Piping geometry factor	Dimensionless
F_R	Reynolds number factor	Dimensionless
F_T	Specific heat ratio factor	Dimensionless
M	Molecular mass of flowing fluid	kg/kmol (lbm/lbm-mol)
N	Numerical constants (see Table 1)	Various (see Note 1)
P_1	Inlet absolute static pressure measured at point A (see Figure 1)	kPa or bar (psia) (see Note 2)
P_2	Outlet absolute static pressure measured at point B (see Figure 1)	kPa or bar (psia)
P_c	Absolute thermodynamic critical pressure	kPa or bar (psia)
P_r	Reduced pressure (P_1/P_c)	Dimensionless
P_v	Absolute vapor pressure of the liquid at inlet temperature	kPa or bar (psia)
ΔP_{actual}	Differential pressure between upstream and downstream pressure taps ($P_1 - P_2$)	kPa or bar (psia)
ΔP_{choked}	Computed value of limiting pressure differential for incompressible flow	kPa or bar (psia)
ΔP_{sizing}	Value of pressure differential used in computing flow or required flow coefficient for incompressible flows	kPa or bar (psia)
Q	Actual volumetric flow rate	m ³ /h (ft ³ /h)
Q_s	Standard volumetric flow rate (see definition 3.2)	m ³ /h (scfh)
Re_v	Valve Reynolds number	Dimensionless
T_1	Inlet absolute temperature	K (°R)
T_c	Absolute thermodynamic critical temperature	K (°R)
T_r	Reduced temperature (T_1/T_c)	Dimensionless
t_b	Absolute reference temperature for standard cubic meter	K (°R)
W	Mass flow rate	kg/h (lbm/h)

Symbol	Description	Unit
x	Ratio of actual pressure differential to inlet absolute pressure ($\Delta P/P_1$)	Dimensionless
x_{choked}	Choked pressure drop ratio for compressible flow	Dimensionless
x_{sizing}	Value of pressure drop ratio used in computing flow or required flow coefficient for compressible flows	Dimensionless
x_T	Pressure differential ratio factor of a control valve without attached fittings at choked flow	Dimensionless (see Note 4)
x_{TP}	Pressure differential ratio factor of a control valve with attached fittings at choked flow	Dimensionless
Y	Expansion factor	Dimensionless
Z_1	Compressibility factor at inlet conditions	Dimensionless
ν	Kinematic viscosity	m ² /s (cS) (see Note 3)
ρ_1	Density of fluid at P_1 and T_1	kg/m ³ (lbm/ft ³)
ρ_1/ρ_0	Relative density ($\rho_1/\rho_0 = 1,0$ for water at 15 °C)	Dimensionless
γ	Specific heat ratio	Dimensionless
ζ	Velocity head loss coefficient of a reducer, expander or other fitting attached to a control valve or valve trim	Dimensionless
ζ_1	Upstream velocity head loss coefficient of fitting	Dimensionless
ζ_2	Downstream velocity head loss coefficient of fitting	Dimensionless
ζ_{B1}	Inlet Bernoulli coefficient	Dimensionless
ζ_{B2}	Outlet Bernoulli coefficient	Dimensionless

NOTE 1 To determine the units for the numerical constants, dimensional analysis may be performed on the appropriate equations using the units given in Table 1.

NOTE 2 1 bar = 10² kPa = 10⁵ Pa

NOTE 3 1 centistoke = 10⁻⁶ m²/s

NOTE 4 These values are travel-related and should be stated by the manufacturer.

- **Control Valve Sizing for Single Phase-Gas Flow**

The fundamental flow model for incompressible fluids in the turbulent flow regime is given as:

$$Q = CN_1 F_P \sqrt{\frac{\Delta P_{sizing}}{\rho_1 / \rho_o}} \quad (1)$$

The fundamental flow model for compressible fluids in the turbulent flow regime is given as:

$$W = CN_6 F_P Y \sqrt{x_{sizing} P_1 \rho_1} \quad (5)$$

To calculate Reynolds number:

$$Re_v = \frac{N_4 F_d Q}{\nu \sqrt{C F_L}} \left(\frac{F_L^2 C^2}{N_2 d^4} + 1 \right)^{1/4}$$

x_{TP} is the pressure differential ratio factor of a control valve with attached fittings at choked flow. To meet a deviation of $\pm 5\%$ for x_{TP} , the valve and attached fittings shall be tested as a unit. When estimated values are permissible, the Equation (22) shall be used:

$$x_{TP} = \frac{\frac{x_T}{F_P^2}}{1 + \frac{x_T \zeta_i}{N_5} \left(\frac{C}{d^2} \right)^2} \quad (22)$$

The fundamental flow model for incompressible fluids in the non-turbulent flow regime is given as:

$$Q = CN_1 F_R \sqrt{\frac{\Delta P_{actual}}{\rho_1 / \rho_o}} \quad (A.2)$$

The fundamental flow model for compressible fluids in the non-turbulent flow regime is given as:

$$W = CN_{27} F_R Y \sqrt{\frac{\Delta P (P_1 + P_2) M}{T_1}} \quad (A.3)$$

- **Control Valve Sizing for Single Phase-Liquid Flow**

The fundamental flow model for incompressible fluids in the turbulent flow regime is given as:

$$Q = CN_1 F_P \sqrt{\frac{\Delta P_{sizing}}{\rho_1 / \rho_o}} \quad (1)$$

The fundamental flow model for compressible fluids in the turbulent flow regime is given as:

$$W = CN_6 F_P Y \sqrt{x_{sizing} P_1 \rho_1} \quad (5)$$

To calculate Reynolds number:

$$Re_v = \frac{N_4 F_d Q}{\nu \sqrt{C F_L}} \left(\frac{F_L^2 C^2}{N_2 d^4} + 1 \right)^{1/4}$$

x_{TP} is the pressure differential ratio factor of a control valve with attached fittings at choked flow. To meet a deviation of $\pm 5\%$ for x_{TP} , the valve and attached fittings shall be tested as a unit. When estimated values are permissible, the Equation (22) shall be used:

$$x_{TP} = \frac{\frac{x_T}{F_P^2}}{1 + \frac{x_T \zeta_i}{N_5} \left(\frac{C}{d^2} \right)^2} \quad (22)$$

The fundamental flow model for incompressible fluids in the non-turbulent flow regime is given as:

$$Q = C N_1 F_R \sqrt{\frac{\Delta P_{actual}}{\rho_1 / \rho_o}} \quad (A.2)$$

The fundamental flow model for compressible fluids in the non-turbulent flow regime is given as:

$$W = C N_{27} F_R Y \sqrt{\frac{\Delta P (P_1 + P_2) M}{T_1}} \quad (A.3)$$

Table C.1 – Incompressible flow

Non-choked flow ($x_{F,actual} < x_{F,choked}$)	Choked flow ($x_{F,actual} \geq x_{F,choked}$)
$x_{F,choked}$ is to be calculated from Equation (4) with F_p and F_{LP} under non-choked flow conditions (see this table)	
$F_p = \sqrt{1 - \frac{\sum \zeta}{N_2} \left(\frac{C}{d^2} \right)^2}$	$F_p = \sqrt{\frac{1 - \frac{\Delta p}{p_1 - F_F \bullet p_v} \bullet \frac{\zeta_1 + \zeta_{B1}}{N_2} \left(\frac{C}{d^2} \right)^2}{1 + \frac{\Delta p}{p_1 - F_F \bullet p_v} \bullet \frac{1}{N_2} \left(\frac{\sum \zeta}{F_L^2} - (\zeta_1 + \zeta_{B1}) \right) \left(\frac{C}{d^2} \right)^2}}$
$F_{LP} = \frac{F_L}{\sqrt{1 + \frac{\zeta_1 + \zeta_{B1}}{N_2} \left(\frac{C}{d^2} \right) \frac{F_L^2}{F_p^2}}}$	$F_{LP} = F_L \bullet \sqrt{1 - \frac{\Delta P}{P_1 - F_F \bullet P_v} \bullet \frac{\zeta_1 + \zeta_{B1}}{N_2} \left(\frac{C}{d^2} \right)^2}$

Table C.2 – Compressible flow

Non-choked flow ($x_{actual} < x_{choked}$)	Choked flow ($x_{actual} \geq x_{choked}$)
x_{choked} is to be calculated from Equation (11) with F_p and x_{TP} under non-choked flow conditions (see this table)	
$F_p = \sqrt{1 - \frac{\sum \xi}{N_2} \left(\frac{C}{d^2} \right)^2}$	$F_p = \sqrt{\frac{1 - \frac{9}{4} \bullet \frac{\Delta p}{F_\gamma \bullet p_1} \bullet \gamma^2 \bullet \frac{\zeta_1 + \zeta_{B1}}{N_5} \left(\frac{C}{d^2} \right)^2}{1 + \frac{9}{4} \bullet \frac{\Delta p}{F_\gamma \bullet p_1} \bullet \gamma^2 \bullet \frac{1}{N_5} \left(\frac{\sum \zeta}{x_T} - (\zeta_1 + \zeta_{B1}) \right) \left(\frac{C}{d^2} \right)^2}}$
$x_{TP} = \frac{x_T}{F_p^2 + x_T \bullet \frac{\zeta_1 + \zeta_{B1}}{N_5} \left(\frac{C}{d^2} \right)^2}$	$x_{TP} = x_T \bullet \left(1 + \frac{\Delta p}{F_\gamma \bullet p_1} \bullet \gamma^2 \bullet \frac{\zeta_1 + \zeta_{B1}}{N_5} \left(\frac{C}{d^2} \right)^2 \right)$

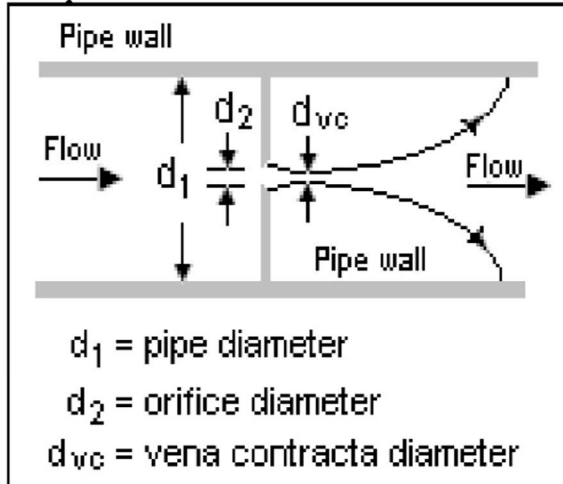
4.9 ORIFICE SIZING

Orifice Sizing Calculations as per ISO 5167:

Orifice plate

An **orifice plate** is a device used for measuring flow rate, for reducing pressure or for restricting flow (in the latter two cases it is often called a restriction plate). Either a volumetric or mass flow rate may be determined, depending on the calculation associated with the orifice plate. It uses the same principle as a Venturi nozzle, namely Bernoulli's principle which states that there is a relationship between the pressure of the fluid and the velocity of the fluid. When the velocity increases, the pressure decreases and vice versa.

Description



An orifice plate is a thin plate with a hole in it, which is usually placed in a pipe. When a fluid (whether liquid or gaseous) passes through the orifice, its pressure builds up slightly upstream of the orifice but as the fluid is forced to converge to pass through the hole, the velocity increases and the fluid pressure decreases. A little downstream of the orifice the flow reaches its point of maximum convergence, the vena contracta (see drawing to the right) where the velocity reaches its maximum and the pressure reaches its minimum. Beyond that, the flow expands, the velocity falls and the pressure increases. By measuring the difference in fluid pressure across tapings upstream and downstream of the plate, the flow rate can be obtained from Bernoulli's equation using coefficients established from extensive research.

Application

Orifice plates are most commonly used to measure flow rates in pipes, when the fluid is single-phase (rather than being a mixture of gases and liquids, or of liquids and solids) and well-mixed, the flow is continuous rather than pulsating, the fluid occupies the entire pipe (precluding silt or trapped gas), the flow profile is even and well-developed and the fluid and flow rate meet certain other conditions. Under these circumstances and when the orifice plate is constructed and installed according to appropriate standards, the flow rate can easily be determined using published formulae based on substantial research and published in industry, national and international standards.

Plates are commonly made with sharp-edged circular orifices and installed concentric with the pipe and with pressure tapings at one of three standard pairs of distances upstream and downstream of the plate; these types are covered by ISO 5167 and other major standards. There are many other possibilities. The edges may be rounded or conical, the plate may have an orifice the same size as the pipe except for a segment at top or bottom which is obstructed, the orifice may be installed eccentric to the pipe, and the pressure tapings may be at other positions. Variations on these possibilities are covered in various standards and handbooks. Each combination gives rise to different coefficients of discharge which can be predicted so long as various conditions are met, conditions which differ from one type to another.

Once the orifice plate is designed and installed, the flow rate can often be indicated with an acceptably low uncertainty simply by taking the square root of the differential pressure across the orifice's pressure tapings and applying an appropriate constant. Even compressible flows of gases that vary in pressure and temperature may be measured with acceptable uncertainty by merely taking the square roots of the absolute pressure and/or temperature, depending on the purpose of the measurement and the costs of ancillary instrumentation.

Orifice plates are also used to reduce pressure or restrict flow, in which case they are often called restriction plates.

Pressure tapings

There are three standard positions for pressure tapings (also called taps), commonly named as follows:

- Corner taps placed immediately upstream and downstream of the plate; convenient when the plate is provided with an orifice carrier incorporating tapings
- D and D/2 taps or radius taps placed one pipe diameter upstream and half a pipe diameter downstream of the plate; these can be installed by welding bosses to the pipe
- Flange taps placed 25.4 mm (1 inch) upstream and downstream of the plate, normally within specialised pipe flanges.

These types are covered by ISO 5167 and other major standards.

Plate

Standards and handbooks are mainly concerned with sharp-edged thin plates. In these, the leading edge is sharp and free of burrs and the cylindrical section of the orifice is short, either because the entire plate is thin or because the downstream edge of the plate is bevelled. Exceptions include the quarter-circle or quadrant-edge orifice, which has a fully rounded leading edge and no cylindrical section, and the conical inlet or conical entrance plate which has a bevelled leading edge and a very short cylindrical section. The orifices are normally concentric with the pipe (the eccentric orifice is a specific exception) and circular (except in the specific case of the segmental or chord orifice, in which the plate obstructs just a segment of the pipe). Standards and handbooks stipulate that the upstream surface of the plate is particularly flat and smooth. Sometimes a small drain or vent hole is drilled through the plate where it meets the pipe, to allow condensate or gas bubbles to pass along the pipe.

Pipe

Standards and handbooks stipulate a well-developed flow profile; velocities will be lower at the pipe wall than in the centre but not eccentric or jetting. Similarly the flow downstream of the plate must be unobstructed, otherwise the downstream pressure will be affected. To achieve this, the pipe must be acceptably circular, smooth and straight for stipulated distances. Sometimes when it is impossible to provide enough straight pipe, flow conditioners such as tube bundles or plates with multiple holes are inserted into the pipe to straighten and develop the flow profile, but even these require a further length of straight pipe before the orifice itself. Some standards and handbooks also provide for flows from or into large spaces rather than pipes, stipulating that the region before or after the plate is free of obstruction and abnormalities in the flow.

Computation

Flow rates through an orifice plate can be calculated without specifically calibrating the individual flowmeter so long as the construction and installation of the device complies with the stipulations of the relevant standard or handbook. The calculation takes account of the fluid and fluid conditions, the pipe size, the orifice size and the measured differential pressure; it also takes account of the coefficient of discharge of the orifice plate, which depends upon the orifice type and the positions of the pressure tapings. With local pressure tapings (corner, flange and D+D/2), sharp-edged orifices have coefficients around 0.6 to 0.63. While the coefficients for conical entrance plates are in the range 0.73 to 0.734 and for quarter-circle plates 0.77 to 0.85. The coefficients of sharp-edged orifices vary more with fluids and flow rates than the coefficients of conical-entrance and quarter-circle plates, especially at low flows and high viscosities.

For compressible flows such as flows of gases or steam, an expansibility factor or expansion factor is also calculated. This factor is primarily a function of the ratio of the measured differential pressure to the fluid pressure and so can vary significantly as the flow rate varies, especially at high differential pressures and low static pressures.

The equations provided in American and European national and industry standards and the various coefficients used to differ from each other even to the extent of using different combinations of correction factors, but many are now closely aligned and give identical results; in particular, they use the same Reader-Harris/Gallagher (1998) equation for the coefficient of discharge for sharp-edged orifice plates. The equations below largely follow the notation of the international standard ISO 5167 and use SI units.

Volume flow rate:

$$q_v = \frac{q_m}{\rho_1}$$

Mass flow rate:

$$q_m = \frac{C}{\sqrt{1-\beta^4}} \epsilon \frac{\pi}{4} d^2 \sqrt{2 \rho_1 \Delta p}$$

Coefficient of discharge for sharp-edged orifice plates with corner, flange or D and D/2 tappings and no drain or vent hole (Reader-Harris/Gallagher equation):

$$C = 0.5961 + 0.0261\beta^2 - 0.216\beta^{2.8} + 0.000521 \left(\frac{10^6 \beta}{Re_D} \right)^{0.7} - (0.0188 + 0.0063A) \beta^{3.5} \left(\frac{10^6}{Re_D} \right)^{0.3} + (0.043 + 0.080e^{-16L_1} - 0.123e^{-7L_1})(1 - 0.11A) \frac{\beta^4}{1 - \beta^4} - 0.031(M_2' - 0.8M_2'^{1.1})\beta^{-3}$$

and if $D < 71.2\text{mm}$ in which case this further term is added to C:

$$+ 0.011(0.75 - \beta) \left(2.8 - \frac{D}{0.0254} \right)$$

In the equation for C,

$$A = \left(\frac{19000\beta}{Re_D} \right)^{0.8}$$

$$M_2' = \frac{2L_2'}{1 - \beta}$$

and only the three following pairs of values for L_1 and L_2' are valid:

corner tappings:

$$L_1 = L_2' = 0$$

flange tappings:

$$L_1 = L_2' = \frac{0.0254}{D}$$

D and D/2 tappings:

$$L_1 = 1$$

$$L_2' = 0.47$$

Expansibility factor, also called expansion factor, for sharp-edged orifice plates with corner, flange or D and D/2 tappings:

if $p_2/p_1 > 0.75$ (at least - standards vary)

$$\epsilon = 1 - (0.351 + 0.256\beta^4 + 0.93\beta^8) \left[1 - \left(\frac{p_2}{p_1} \right)^{\frac{1}{\kappa}} \right]$$

but for incompressible fluids, including most liquids

$$\epsilon = 1$$

where:

C = coefficient of discharge, dimensionless

d = internal orifice diameter under operating conditions, m

D = internal pipe diameter under operating conditions, m

p_1 = fluid absolute static pressure in plane of upstream tapping, Pa

p_2 = fluid absolute static pressure in plane of downstream tapping, Pa

q_m = mass flow rate, kg/s

q_v = volume flow rate, m³/s

Re_D = pipe Reynolds number, $\frac{4q_m}{\pi\mu D}$, dimensionless

β = diameter ratio of orifice diameter to pipe diameter, $\frac{d}{D}$, dimensionless

Δp = differential pressure, Pa

ϵ = expansibility factor, also called expansion factor, dimensionless

κ = isentropic exponent, often approximated by specific heat ratio, dimensionless

μ = dynamic viscosity of the fluid, Pa.s

ρ_1 = fluid density in plane of upstream tapping, kg/m³

Overall pressure loss

The overall pressure loss caused by an orifice plate is less than the differential pressure measured across tappings near the plate. For sharp-edged plates such as corner, flange or D and D/2 tappings, it can be approximated by the equation

$$\frac{\Delta\bar{\omega}}{\Delta p} = 1 - \beta^{1.9}$$

or

$$\frac{\Delta\bar{\omega}}{\Delta p} = \frac{\sqrt{1 - \beta^4(1 - C^2)} - C\beta^2}{\sqrt{1 - \beta^4(1 - C^2)} + C\beta^2}$$

where

$\Delta\bar{\omega}$ = overall pressure loss, Pa

and other symbols are as above

where:

Q = volumetric flow rate (at any cross-section), m³/s

$\dot{m}$ = mass flow rate (at any cross-section), kg/s

C_d = coefficient of discharge, dimensionless

C = orifice flow coefficient, dimensionless

A_1 = cross-sectional area of the pipe, m²

A_2 = cross-sectional area of the orifice hole, m²

d_1 = diameter of the pipe, m

d_2 = diameter of the orifice hole, m

β = ratio of orifice hole diameter to pipe diameter, dimensionless

V_1 = upstream fluid velocity, m/s

V_2 = fluid velocity through the orifice hole, m/s

P_1 = fluid upstream pressure, Pa with dimensions of kg/(m·s²)

P_2 = fluid downstream pressure, Pa with dimensions of kg/(m·s²)

ρ = fluid density, kg/m³

Deriving the above equations used the cross-section of the orifice opening and is not as realistic as using the minimum cross-section at the vena contracta. In addition, frictional losses may not be negligible and viscosity and turbulence effects may be present. For that reason, the coefficient of discharge C_d is introduced. Methods exist for determining the coefficient of discharge as a function of the Reynolds number.

The parameter $\frac{1}{\sqrt{1 - \beta^4}}$ is often referred to as the velocity of approach factor and multiplying the coefficient of discharge by that parameter (as was done above) produces the flow coefficient C .

Methods also exist for determining the flow coefficient as a function of the beta function β and the location of the downstream pressure sensing tap. For rough approximations, the flow coefficient may be assumed to be between 0.60 and 0.75. For a first approximation, a flow coefficient of 0.62 can be used as this approximates to fully developed flow.

An orifice only works well when supplied with a fully developed flow profile. This is achieved by a long upstream length (20 to 40 pipe diameters, depending on Reynolds number) or the use of a flow conditioner. Orifice plates are small and inexpensive but do not recover the pressure drop as well as a venturi, nozzle, or venturi-nozzle does. Venturis also require much less straight pipe upstream. A venturi meter is more efficient, but usually more expensive and less accurate (unless calibrated in a laboratory) than an orifice plate.

4.10 PUMP SIZING

A pump is a mechanical device used to move fluids from one destination to another. Fluids can mean either liquids, gases or a combination of liquids and solids (slurries). For the purpose of this training manual, pumps are to be considered only for transporting liquids and slurries.

Types of Pumps

➤ Centrifugal Pump

A centrifugal pump is a roto-dynamic pump that uses a rotating impeller to increase the pressure and flow rate of a fluid. Centrifugal pumps are the most common type of pump used to move liquids through a piping system. The fluid enters the pump impeller along or near to the rotating axis and is accelerated by the impeller, flowing radially outward or axially into a diffuser or volute chamber, from where it exits into the downstream piping system. Centrifugal pumps are typically used for large discharge through smaller heads.

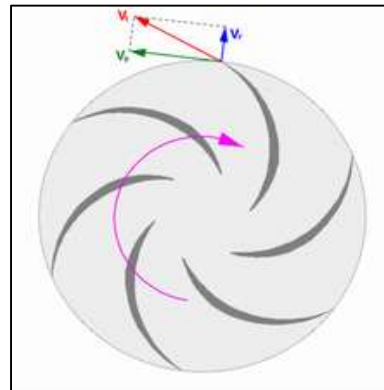
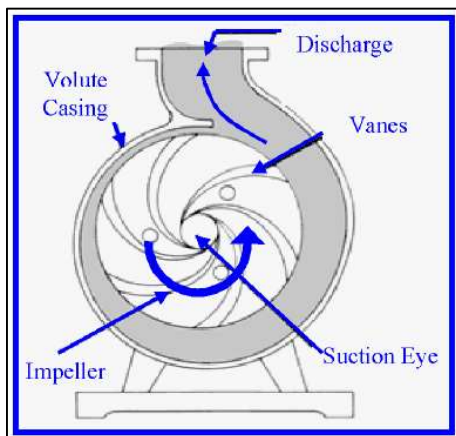


Fig 1.

➤ Positive Displacement (PD) Pump

A positive displacement pump causes a fluid to move by trapping a fixed amount of it and then forcing (displacing) that trapped volume into the discharge pipe.

Some positive displacement pumps work using an expanding cavity on the suction side and a decreasing cavity on the discharge side. Liquid flows into the pump as the cavity on the suction side expands and the liquid flows out of the discharge as the cavity collapses. The volume is constant given each cycle of operation.

➤ PD Pump behavior and safety

Positive displacement pumps, unlike centrifugal or roto-dynamic pumps, will in theory produce the same flow at a given speed (RPM) no matter what the discharge pressure. Thus, positive displacement pumps are "constant flow machines". However due to a slight increase in internal leakage as the pressure increases, a truly constant flow rate cannot be achieved.

A positive displacement pump must not be operated against a closed valve on the discharge side of the pump, because it has no shut-off head like centrifugal pumps. A positive displacement pump operating against a closed discharge valve will continue to produce flow and the pressure in the discharge line will increase, until the line bursts or the pump is severely damaged, or both.

A relief or safety valve on the discharge side of the positive displacement pump is therefore necessary. The relief valve can be internal or external. The pump manufacturer normally has the option to supply internal relief or safety valves. The internal valve should in general only be used as a safety precaution, an external relief valve installed in the discharge line with a return line back to the suction line or supply tank is recommended.

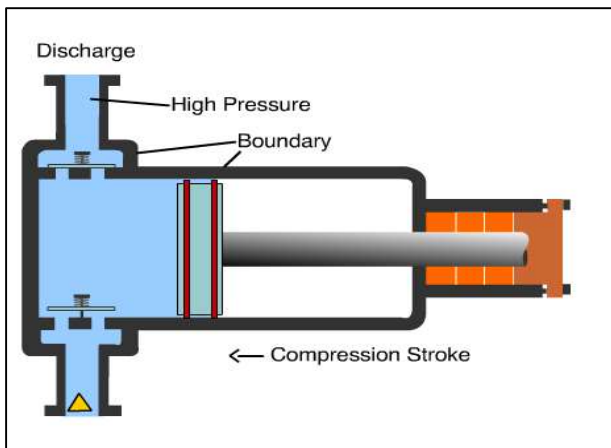


Fig 2.

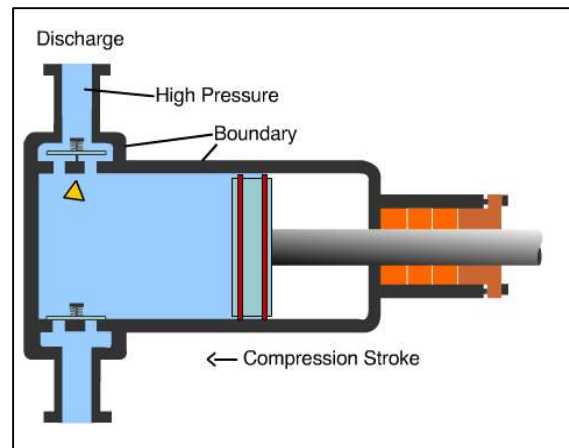


Fig 3.

➤ **PD pump types**

A positive displacement pump can be further classified according to the mechanism used to move the fluid:

Rotary-type positive displacement: internal gear, screw, shuttle block, flexible vane or sliding vane, circumferential piston, or liquid ring vacuum pumps

Reciprocating-type positive displacement: piston or diaphragm pumps.

For a detailed description of types of pumps refer the Wikipedia article on pumps at the link:

<http://en.wikipedia.org/wiki/pump>

Some illustrations for PD pumps are provided below:

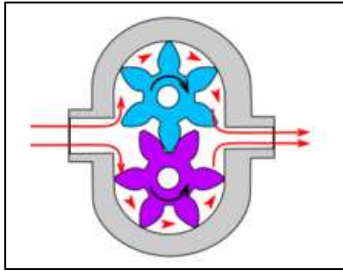


Fig 4. Gear Pump

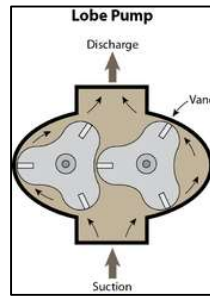


Fig 5. Lobe Pump

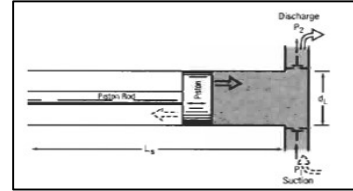


Fig 6. Piston Pump

Design Flow Rate

The pump design flow rate shall include a minimum 10% margin on the maximum flow rate except in the following cases:

- Intermittent service
- Minimum flow bypass using orifice plate, in which case pump design flow rate shall be the process flow rate (as per heat and material balance) plus the minimum flow bypass rate.

For reflux pumps, design flow rate shall be at least 120% of the operating flow rate.

Net Positive Suction Head Available (NPSHa)

It is defined as the pressure at the centerline of the pump datum level minus the fluid vapor pressure at the given temperature. It is expressed in meter or feet of fluid. Process designer shall provide NPSHa at reference datum level (e.g. grade level) such that pump vendor can select the correct pump by calculating NPSHa at his datum level.

The NPSHa must be greater than NPSHr (NPSH required, given by pump vendor). NPSHa should be at least 1 meter greater than the NPSHr throughout the range from minimum flow to rated capacity of the pump. NPSHr should be based on 3% head loss criteria (reference API STD 610).

However, a case-by-case study depending on the application shall be done to determine the NPSHa margin over the NPSHr. For example, more margin is required to be considered for liquids with dissolved gases, lower suction pressure (< 1 atm) etc.

Reciprocating pumps must consider the acceleration head when calculating the NPSHa.

Minimum Flow By-Pass

API STD 610 defines the following two minimum flows for centrifugal pumps:

- Minimum continuous stable flow: It is the lowest flow at which the pump can operate without exceeding acceptable vibration limits.

- Minimum continuous thermal flow: It is the lowest flow at which the pump can operate without its operation being impaired by the temperature rise of the pumped liquid.

Pump operations below these points can cause shaft vibration and reduce the mechanical seal life. To avoid such problems, high capacity pumps shall be provided with a minimum flow bypass with the flow controlled by a restriction orifice or a control valve. If a restriction orifice is used, the operating flow rate of the pump should be increased to account for the continuous bypass flow.

The sizing of the minimum bypass flow circuit shall be based on the higher of the two minimum flows.

If it is not possible to provide a minimum flow circulation line, then adequate instrument protection shall be provided to prevent pump operation below the minimum allowable flow rate.

General Guidelines for Proper Selection & Operation of Centrifugal Pumps

- The head / capacity characteristic curve shall continuously rise as flow is reduced to shut-off (or zero flow). The minimum recommended rise is 12% with reference to the rated point. For parallel operation the head rise shall be between minimum 15% to maximum 20%.
- The pump should be capable of a head increase at the design conditions by installing a larger impeller. The rated impeller diameter shall be between 85% and 90% of the maximum impeller diameter as per vendor data sheet.
- Pump(s) should not be operated below the minimum safe continuous flow as provided by the pump manufacturer.
- The installed power shall be adequate for maximum impeller diameter plus 120% of rated capacity.

General Guidelines for Proper Selection & Operation of Positive Displacement Reciprocating & Controlled Volume Pumps

Reciprocating/Controlled volume pumps are capable of handling very low flows ($<0.227 \text{ m}^3/\text{h}$) and are typically used for chemical injection duties in oil and gas installations.

- Reciprocating pumps should comply with API STD 674.
- Metering (or controlled volume) pumps should comply with API STD 675.
- Some selection guidelines for controlled volume pumps are as follows:
 - “Packed Plunger Pumps” are generally recommended for pressures above 340 barg
 - Diaphragm pumps are generally recommended for pressures up to 340 barg and temperatures up to 150°C.
 - Double Diaphragm Pumps of the sandwich type construction are generally recommended.

Double Diaphragm pumps should be used when:

- Dilution of the product flow by hydraulic oil is not acceptable.
- Pumped fluid is extremely hazardous and/or toxic.

- Process leakage to the atmosphere is unacceptable
- Process fluid must not come in contact with air
- Packed plunger pumps should not be used for the above conditions.

Capacity Adjustment for Metering Pumps (Reference API STD 675)

- Pump capacity shall be adjusted by changing the actual or effective stroke length or the pump stroking speed.
- The pump shall be capable of accepting manual or automatic capacity stroke control, either factory mounted or by field conversion.
- Integral pump devices used to vary capacity either manually or automatically shall be provided with visual indication of capacity setting, shown as a percentage of the nameplate rated flow. Manual control shall include a locking device to positively retain the capacity setting.
- The direction of movement to increase or decrease pump flow shall be clearly marked. All adjustment means and indicators shall be easily accessible with the pump installed.

Pressure Relief for Positive Displacement Pumps

Full flow pressure relief shall be provided at the pump discharge upstream of any isolating valve. Where the nature of the fluid would make relief valves unreliable, rupture discs should be used either alone or in conjunction with relief valves.

Pulsation Devices for Positive Displacement Pumps

When specified, the vendor shall furnish pulsation suppression devices to be located at the pump suction and/or discharge connections. The following are three basic types of pulsation suppression devices:

- Volume bottles without internals.
- Pulsation dampeners using a gas-filled chamber isolated from the pumped fluid by a piston or elastomeric diaphragm.
- Pulsation filters and attenuators, including proprietary commercial designs based on acoustical suppression techniques.

Reference is made to API STD 675 for the above.

Further Explanation on Key Terms Related to Pumps

NPSHa:

Net positive suction head available (NPSHa) is the suction head present at the pump suction over and above the vapor pressure of the liquid. NPSHa is a function of the suction system and is independent of the type of pump in the system. It should be calculated by the engineer or pump user, and supplied to the pump manufacturer as part of the application criteria or pump specification. The general formula for calculating NPSHa is:

$$\text{NPSHa} = P \pm H - H_f - H_{vp} - H_a \text{ ----- (1)}$$

Where:

P = absolute pressure on the surface of the liquid in the suction vessel, expressed in meter of liquid
 H = static distance from the surface of the liquid in the supply vessel to the centerline of the pump impeller, in meter; the term is positive if the pump has a static suction head, and negative if the pump has a static suction lift. For the purpose of NPSHa calculations, both the static suction head and the static suction lift should be considered at the “minimum operating liquid level” of the suction vessel. In other words, the NPSHa should be calculated with the minimum static head or the maximum static lift as the case maybe. The term “minimum operating liquid level” although is quite debatable since many engineering professionals as well as engineering companies differ on its definition. It would suffice to say that the minimum static head or maximum static lift may differ on a case-to-case and operating philosophy basis and the engineer performing the NPSHa calculations would require doing a careful analysis of this value before using it in his or her calculation.

H_f = friction loss in the suction line, including all piping, valves, fittings, filters, etc., expressed in meter of liquid; this term varies with flow, so NPSHa must be calculated based on a particular flow rate

H_{vp} = vapor pressure of the liquid at the pumping temperature, expressed in meter of liquid

H_a = Acceleration head. Acceleration head needs to be considered only for reciprocating pumps. For centrifugal pumps and rotary pumps the acceleration head is zero.

In a new pump application, NPSHa (and the static term H in the above formula) must be given to the manufacturer with reference to some known datum point such as the elevation of the pump mounting base. This is because the location of the pump impeller centerline elevation is generally not known when the NPSHa calculations are made. It is important that the datum point of reference be mentioned in the specification, as well as the calculated value of NPSHa.

New engineers often get confused when the suction vessel is not a vented vessel to atmosphere and there is absence of data about the maximum operating pressure in the vapor space of the vessel. In such a case, the set pressure if available of the relief device installed in the vapor space of the vessel may be used as the maximum vessel pressure for the purpose of NPSHa calculations.

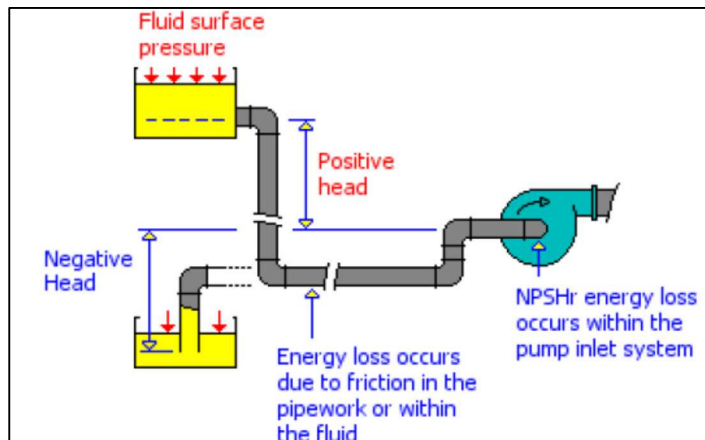


Fig 7.

NPSHr:

Net positive suction head required (NPSHr) is the suction head required at the impeller centerline over and above the vapor pressure of the liquid. NPSHr is strictly a function of the pump inlet design, and is independent of the suction piping system. The pump requires a pressure at the suction flange greater than the vapor pressure of the liquid because merely getting the liquid to the pump suction flange in a liquid state is not sufficient. The liquid experiences pressure losses when it first enters the pump, before it gets to the point on the impeller vane where pressure begins to increase. These losses are caused by frictional effects as the liquid passes through the pump suction nozzle, moves across the impeller inlet and changes direction to begin to flow along the impeller vanes.

NPSHr is established by the manufacturer using a special test, and the value of NPSHr is shown on the pump curve as a function of pump capacity.

It is important to note that the NPSHr increases with higher flow rate due to the increased amount of friction loss inside the pump inlet before the liquid reaches the pump impeller. In certain cases the NPSHr also increases with the flow remaining unchanged but the impeller diameter reduced.

For a pump to operate free of cavitation, NPSHa must be greater than NPSHr. In determining the acceptability of a particular pump operating in a particular system with regard to NPSH, the NPSHa for the system must be calculated by the engineer, and then the NPSHr for the pump to be used must be examined at the same flow rate by looking at this information on the pump curve. This comparison should be made at all possible operation points of the pump, with the worst case usually being at the maximum expected flow, also called the run out flow.

Suction Pressure or Suction Head:

Also called the minimum suction pressure it is nothing but the NPSHa plus the vapor pressure of the liquid. As an equation it can be represented as:

$$H_s = \text{NPSHa} + H_{vp} = P \pm H - H_f - H_a \text{ ----- (2)}$$

Where:

H_s = Suction Pressure or Suction Head in meter at the pump suction

Discharge Pressure or Discharge Head:

This is the pressure measured at the discharge connection of the pump. It can be measured either as pressure or head of liquid in meters. It can be represented in the following equation form:

$$H_{dis} = P_{dest} + H_{f(dis)} \pm H_{(dis)}$$

Where:

H_{dis} = Discharge pressure expressed in meter of liquid

P_{dest} = Destination pressure of destination vessel expressed in meter of liquid

Destination pressure estimation requires the knowledge of the inlet connection location from the pump to the destination vessel

If the inlet connection is located on top of the destination vessel / tank in the vapor space then the destination pressure would be the vapor pressure in the vapor space.

If the inlet connection is located at the bottom of the destination vessel in the liquid filled area then the destination pressure would be the vapor pressure in the vapor space plus the maximum operating liquid level (height) in the destination vessel.

$H_{f(dis)}$ = friction loss in the discharge line, including all piping, valves, fittings, equipment, instruments etc., expressed in meter of liquid; this term varies with flow, so discharge pressure must be calculated based on a particular flow rate.

Friction losses are given in terms of Equivalent length of pipe, which is covered briefly in below link.
<https://www.cheresources.com/invision/files/file/258-equivalent-length-calculations-for-pipes-and-valves/>

$H_{(dis)}$ = Static distance from pump discharge nozzle to the inlet connection at the destination receiver, expressed in meter of liquid; the term is positive if the inlet connection to the destination vessel is at a height above the pump discharge nozzle (datum for pump discharge) and negative if the inlet connection to the destination vessel is at a height below the pump discharge nozzle. The case where the inlet connection to destination vessel is below the pump discharge nozzle is not frequently encountered in pumping practices.

Differential Pressure or Differential head:

It is the difference between the discharge pressure and the suction pressure. Expressed as an equation it is:

$$P_d = H_{dis} - H_s$$

Where:

P_d = Differential pressure or differential head expressed in meter of liquid

At the beginning of the project due to uncertainties in suction and discharge calculations such as piping lengths, number of fittings, location of source and destination vessels and sometimes even in liquid properties, a margin may be provided on the calculated differential head which ensures

that the pump is adequately sized for the intended duty. The differential head after providing the margin is called the “rated differential head”. Typically the margin that is provided is a maximum 10 % over the calculated differential head. Thus expressed in equation form the rated differential head would be:

$$P_{d (rated)} = 1.1 \times P_d$$

Where:

$P_{d (rated)}$ = Rated differential pressure or differential head expressed in meter of liquid

Pump Motor Brake Horse Power (BHP) or Brake Kilowatt (BkW)

All pumps require a driver such as a motor or turbine to drive the pump. The BHP or BkW needs to be provided by the process engineer for the selection of the driver by the electrical engineer. Expressed in equation form the BHP or BkW would be:

$$BkW = Q \times P_{d (rated)} \times \rho \times g / (3600 \times \eta)$$

Where:

BkW = Brake Power in kW

Q = rated volume flow rate of the pump, m³/h

$P_{d (rated)}$ = Rated differential pressure, m

ρ = density of the liquid at the pumping temperature, kg/m³

g = acceleration due to gravity, 9.81 m/s²

η = pumping efficiency, expressed as a fraction

Converting Head to Pressure and Vice-Versa

An important task while performing pump hydraulic calculations is converting head in terms of meters of liquid to pressure in bar or vice-versa. The basic principle governing the conversion of head to pressure and pressure to head is in accordance to Pascal’s law which when expressed in terms of an equation is:

$$P = \rho \times g \times h$$

Where:

P = hydrostatic pressure, Pascal (N/m²)

ρ = fluid density, kg/m³

h = height of the liquid column, m

For pump hydraulic calculations where head is expressed in meters of liquid and pressure in bar the following equations can be used for conversion:

Converting Head in Meters to Pressure in Bar

$$P = h \times SG / 10.2$$

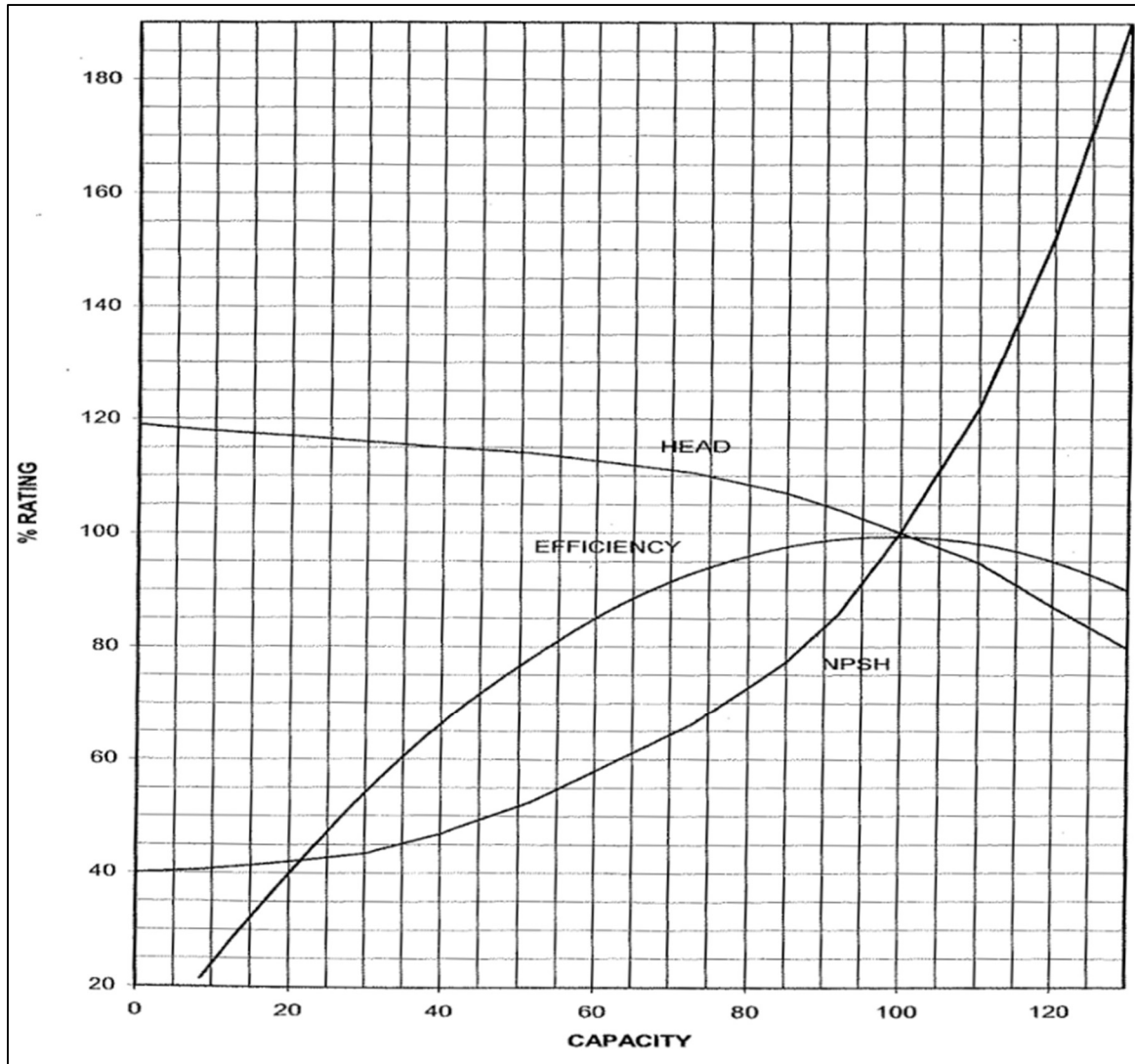
Converting Pressure in Bar to Head in Meters

$$h = P \times 10.2 / SG$$

Where:

SG = specific gravity of liquid

= density of liquid in $\text{kg/m}^3 / 1000$



Excel Programmable Formulas for Incompressible Gas Flow and Liquid Flow Pressure drop Calculations:

Nomenclature in formulas:

W = Liquid Flow Rate, kg/h

ρ = Liquid Density, kg/m³

μ = Liquid Viscosity, cP

ϵ = Absolute Pipe Roughness, mm

d = Inside Pipe Diameter, mm (Refer Note below)

Re = Reynolds number, dimensionless

f = Darcy friction factor, dimensionless

ΔP = Pressure Drop, kPa / 100m

v = Liquid velocity, m/s

Note: As per global industry practices, pipes are specified as nominal pipe size (NPS) identifying the outer diameter in inches and a pipe schedule which indicates the wall thickness. The pipe inside diameter needs to be calculated by subtracting 2-times the wall thickness from the outer diameter of the pipe. For a given NPS, but with different pipe schedules, the pipe inside diameter will change according to the pipe schedule. Refer the link below:

<http://www.dacapo.com/uploads/documents/35-document.pdf>

a) Formula for Initial Estimate of Pipe Diameter

$$d = ((125060 * W^2)/(\Delta P * \rho))^{1/5}$$

Use next largest diameter, in mm, for initial calculations

b) Formula for Reynolds Number

$$Re = (354 * W)/(\mu * d)$$

c) Formula for Friction factor

$$f = 0.25/(\log((\epsilon/3.7d) + (5.74/Re^{0.9})))^2$$

d) Formula for pressure drop

$$\Delta P = ((6253000 * f * W^2)/(d^5 * \rho))$$

e) Formula for velocity

$$v = ((353.7 * W)/(d^2 * \rho))$$

*PROCESS DESIGN ENGINEERING
MANUAL:
Part 3*

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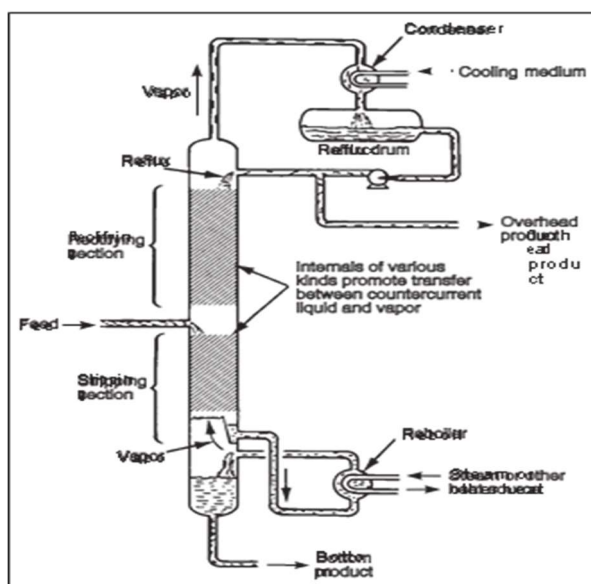
4.11 DISTILLATION COLUMN

Fractionation:

Fractionation is a unit operation utilized to separate mixtures into individual products. Fractionation involves separating components by relative volatility (α). The difficulty of a separation is directly related to the relative volatility of the components and the required purity of the product streams. Virtually all gas processing plants producing natural gas liquids require at least one fractionator to produce a liquid product which will meet sales specifications. The schematic of an example fractionator in Fig 19-2 shows the various components of the system. Heat is introduced to the reboiler to produce stripping vapors. The vapor rises through the column contacting the descending liquid. The vapor leaving the top of the column enters the condenser where heat is removed by some type of cooling medium. Liquid is returned to the column as re-flux to limit the loss of heavy components overhead.

Internals such as trays or packing promote the contact between the liquid and vapor streams in the column. Intimate contact of the vapor and liquid phases is required for efficient separation. Vapor entering a separation stage will be cooled which results in some condensation of heavier components. The liquid phase will be heated which results in some vaporization of the lighter components. Thus, the heavier components are concentrated in the liquid phase and eventually become the bottom product. The vapor phase is continually enriched in the light components which will make up the overhead product.

The vapor leaving the top of the column may be totally or partially condensed. In a total condenser, all vapor entering the condenser is condensed to liquid and the reflux returned to the column has the same composition as the distillate or overhead product. In a partial condenser, only a portion of the vapor entering the condenser is condensed to liquid. In most partial condensers only sufficient liquid will be condensed to serve as re-flux for the tower. In some cases, however, more liquid will be condensed than is required for reflux and there will actually be two overhead products, one a liquid having the same composition as the reflux and the other a vapor product which is in equilibrium with the liquid reflux.

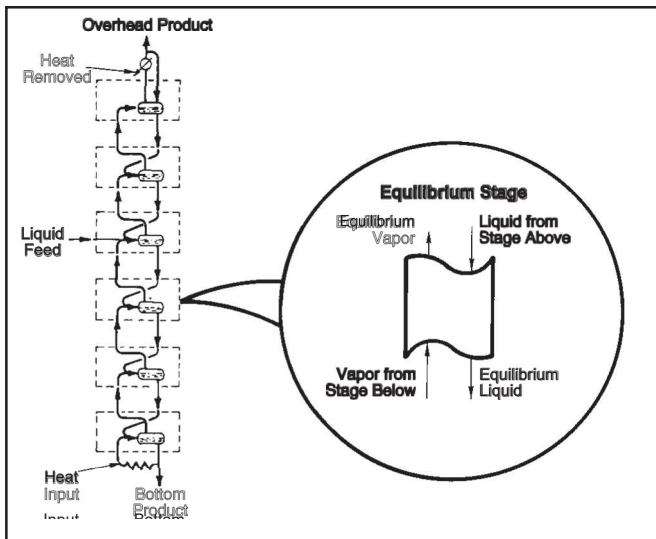


Fractionation Schematic Diagram

Fractionation Schematic Diagram

Equilibrium Stage Concept

All calculations are performed using theoretical (equilibrium) stages. A fractionation column may be considered as a series of equilibrium flashes with two feeds and two product streams (Fig. 19-3). The vapor enters the flash from the stage below at a higher temperature and the liquid stream enters from the stage above at a lower temperature. Heat and mass transfer occur in this stage such that the exiting streams are a bubble point liquid and dew point vapor at the same temperature and pressure. The compositions of these phases are related by the equilibrium relationship of $y_i = K_{ixi}$ (See Section 25). This relationship, along with heat and material balance considerations, is the basis for all fractionator design.



Product Specifications

A material balance around the column is the first step in fractionation calculations. In order to perform this balance, assumption of the product stream compositions must be made. There are three ways of specifying a desired product from a fractionator:

- A percentage recovery of a component in the overhead or bottom stream.
- A composition of one component in either product.
- A specific physical property, such as vapor pressure, for either product.

The recovery and composition specifications can be used directly in the material balance. However, property specifications are used indirectly. For instance if vapor pressure is a desired specification of a product, a material balance is performed with an assumed component split. The calculated vapor pressure of the resulting stream is then compared with the desired value and the material balance redone until reasonable agreement is reached.

In a multicomponent mixture, there are typically two components which are the “keys” to the separation. The light key component is defined as the lightest component in the bottom product in a significant amount. The heavy key component is the heaviest component in the overhead product in a significant amount. Normally, these two components are adjacent to each other in the volatility listing of the components. For hand calculations, it is normally assumed for material balance purposes that all components lighter than the light key are produced overhead and all components heavier than the heavy key are produced with the bottom product. By definition, the key components will be distributed between the product streams.

Example 19-1 — For the given feed stream, estimate the product stream compositions for 98% propane recovered in the overhead product with a maximum iso-butane content of the overhead stream of 1%.

Feed:	C ₂	2.4
	C ₃	162.8
	iC ₄	31.0
	nC ₄	76.7
	C ₅	76.5
		349.4 moles

Solution Steps

For Propane (light key):

$$\text{Moles in overhead} = (0.98) 162.8 = 159.5$$

$$\text{Moles in bottoms} = 162.8 - 159.5 = 3.3$$

For Ethane:

$$\text{Moles in overhead} = 2.4 \text{ (100\% to overhead)}$$

Since the isobutane (the heavy key) is 1% of the overhead stream, the sum of propane and ethane must be 99% (all n-C₄ and C₅+ are in the bottoms). Thus:

$$\text{Total Overhead Moles} = \frac{159.5 + 2.4}{0.99} = \frac{161.9}{0.99} = 163.5$$

$$\text{Moles of iC}_4 = 163.5 - 161.9 = 1.6$$

The overall balance is:					
Comp.	Feed	Overhead		Bottoms	
	Moles	Moles	Mole %	Moles	Mole %
C ₂	2.4	2.4	1.5	—	—
C ₃	162.8	159.5	97.5	3.3	1.8
iC ₄	31.0	1.6	1.0	29.4	15.8
nC ₄	76.7	—	—	76.7	41.2
C ₅	76.5	—	—	76.5	41.2
Total	349.4	163.5	100.0	185.9	100.0
In actual operation, the lighter than light key components and heavier than heavy key components will not be perfectly separated. For estimation purposes and hand calculations, perfect non-key separation is a useful simplifying assumption.					

Key Parameters

Two important considerations which affect the size and cost of a fractionation column are degree of separation and component volatility.

The degree of separation or product purity has a direct impact on the size of the column and the required utilities. Higher purity will require more trays, more reflux, larger diameter, and/or a reduced product quantity. One quantitative measure of the difficulty of a separation is the separation factor, SF, de- fined as:

$$S_F = \left(\frac{X_D}{X_B} \right)_{LK} \left(\frac{X_B}{X_D} \right)_{HK}$$

Note that above Equation defines the specification for the tower design.

Typically, for most fractionation problems this factor ranges from around 500 to 2000. However, for sharp separations, it can be in the 10,000 range. The number of trays will be roughly proportional to the log of the separation factor for a given system.

The volatility of the components is usually expressed as relative volatility, α . This quantity is computed as the ratio of the equilibrium K values of two components at a given temperature and pressure. For fractionation calculations the relative volatility of the key components is important. Therefore:

$$\alpha = K_{LK} / K_{HK}$$

This is a measure of the ease of separation. The larger α is, the easier is the separation.

Design Considerations

Operating Pressure

Before any design calculations can be made on a fractionation problem, a tower operating pressure must be determined. One of the primary considerations for operating pressure is the cooling medium available for the reflux condenser. The over- head product will be at bubble point conditions for a liquid product or at dew point conditions for a vapor product. The bubble point (or dew point) pressure is fixed by a desired component separation and the temperature of the cooling medium.

The cooling media typically used are air, water, and refrigerant. Air cooling is normally the least expensive cooling method. Practical exchanger design limits the process to a 10°C approach to the

ambient summer temperature. This translates to a process temperature of 46 to 52°C in most locations. With cooling water, process temperatures of 35 to 41°C are possible. Below about 35°C, mechanical refrigeration must be used to achieve the desired condensing temperature. This is the most expensive cooling method from both a capital and operating cost standpoint. Generally, it is desirable to operate at as low a pressure as possible to maximize the relative volatility between the key components of the separation. However, if reducing the pressure requires a change to a more expensive cooling method, this is usually not a desirable choice.

In some cases, the overhead from the column must be compressed to sales or another process unit. In this case a higher operating pressure may be desired to reduce compression power.

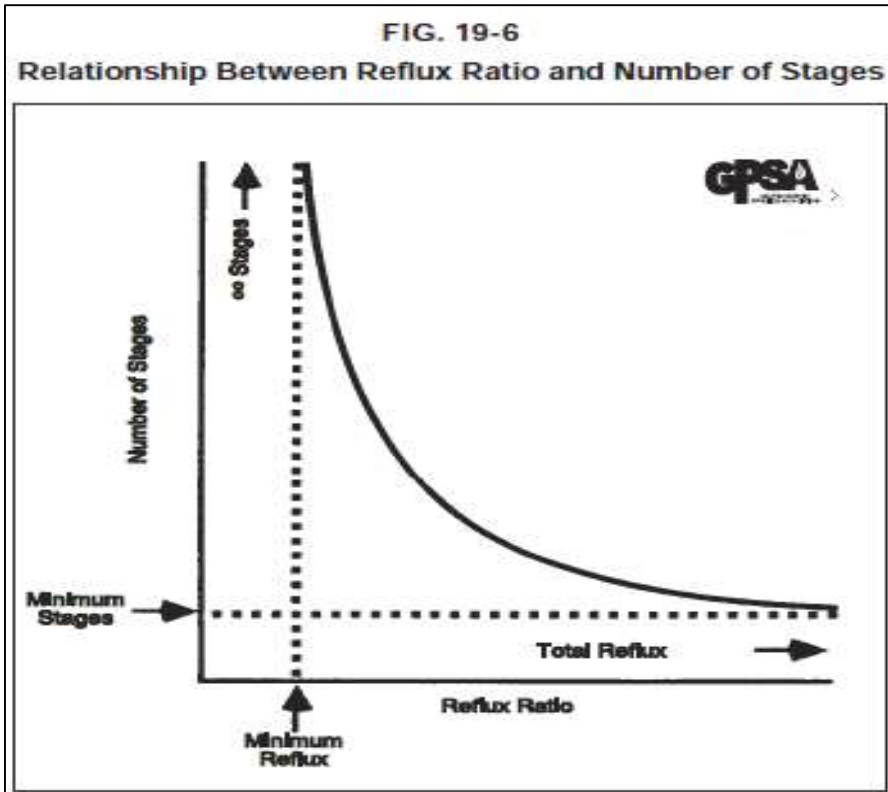
Other items must be considered which will limit pressure selection. If an operating pressure is too high, the critical temperature of the bottom product may be exceeded and the desired separation cannot be achieved. Additionally, the pressure cannot exceed the critical pressure of the desired overhead product.

The selection of a partial or total condenser is fixed by the disposition of the overhead product. A total condenser is used for a liquid product and a partial condenser for a vapor product. However, a liquid product can be produced as a vapor and subsequently cooled and/or compressed to produce a liquid product. There may be cases where this downstream liquefaction is economically attractive. In most cases, the fractionation system for a partial condenser will be cheaper and will have to be balanced against the cost of additional downstream equipment. Before a reliable economic comparison can be made, the column design must be made for each type condenser for a number of reflux ratios and operating pressures.

Reflux Ratio and Number of Stages

The design of a fractionation column is a capital cost versus energy cost trade-off problem. The primary parameters are the number of stages and the reflux ratio. Reflux ratio may be defined in several ways. For most calculations, reflux ratio is defined as the ratio of the molar rate of reflux liquid divided by the molar rate of net overhead product. The reboiler duty is a direct function of the reflux ratio as the fractionating column must maintain an overall heat and material balance for a given separation.

A fractionation column can only produce a desired separation between the limits of minimum reflux and minimum stages. Fig. 19-6 illustrates the relationship between reflux ratio and number of stages for a given separation. At minimum reflux an infinite number of stages is required. At total reflux a minimum number of stages is required. Neither of these situations represents actual operation but are the extreme limits of possible design configurations. Methods have been developed to calculate both these cases in a rigorous manner. However, these methods require a computer solution of tray by tray calculations. In order to begin a detailed design, estimates of minimum reflux ratio and minimum trays should be generated from simple methods using a binary key component analysis.



The minimum stages can be calculated for most multicomponent systems by the Fenske equation.

$$S_m = \frac{\log S_F}{\log(\alpha_{avg})}$$

Condenser if these items are used. The α_{avg} is the column average relative volatility of the key components in the separation. Various averaging techniques have been proposed such as square root averaging of the α at the top and bottom of the column. The most common approach is to use an arithmetic average:

$$\alpha_{avg} = \frac{\alpha_{top} + \alpha_{bottom}}{2}$$

If volatility varies widely, the approach of Winn is suggested, in which a modified volatility is used:

$$\beta_{ij} = K_{LK} / K_{HK}^b$$

Where the exponent b is obtained from K-value plots over the range of interest.

The minimum stage calculation is:

$$S_m = \frac{\log \left[\left(\frac{X_D}{X_B} \right)_{LK} \left(\frac{X_B}{X_D} \right)_{HK}^b \left(\frac{B}{D} \right)^{1-b} \right]}{\log \beta_{ij}}$$

Note that S_m includes the partial condenser and partial reboiler if they exist.

Minimum Reflux Ratio

The Underwood method is the most widely used of the methods for calculating minimum reflux ratio. Underwood assumed constant relative volatility and constant molal overflow in the development of this method. The first step is to θ evaluate by trial and error:

$$1 - q = \sum_{i=1}^n \frac{X_{Fi}}{(\alpha_i - \theta)/\alpha_i}$$

Once θ is determined, the minimum reflux ratio is:

$$(L_o/D)_m + 1 = R_m + 1 = \sum_{i=1}^n \frac{X_{Di}}{(\alpha_i - \theta)/\alpha_i}$$

Number of Stages

The number of theoretical stages required for a given separation at a reflux ratio between minimum and total reflux can be determined from empirical relationships. Erbar and Maddox made an extensive investigation of tray-by-tray fractionator calculations and developed the correlation. This correlation relates the ratio of minimum stages to theoretical stages to the minimum reflux ratio, R_m , and the operating reflux ratio, R (where $R = L_o/D$).

It can be used to determine an operating reflux for a given number of stages by entering the figure at the value of S_m/S , moving up to the line representing the value of $R_m/(R_m + 1)$ and reading a value of $R/(R + 1)$. The optimum operating reflux ratio has been found to be near the minimum reflux ratio. Values of 1.2 to 1.3 times the minimum are common. Thus for a given R , a value of S can be determined from.

This correlation is generated on the basis of bubble point feed. If the feed is between the bubble point and dew point then the operating reflux should be corrected. Erbar and Maddox proposed the following relationship to adjust the vapor rate from the top tray for non-bubble point feed.

$$V_{corr} = V_{calc} + \frac{\left(1 - \frac{D}{F}\right) [F(H_{VF} - H_{BP})]}{\left(\frac{Q_C}{L_o}\right)_{calc}}$$

The reflux rate can then be adjusted by material balance since:

$$L_o = V_i - D$$

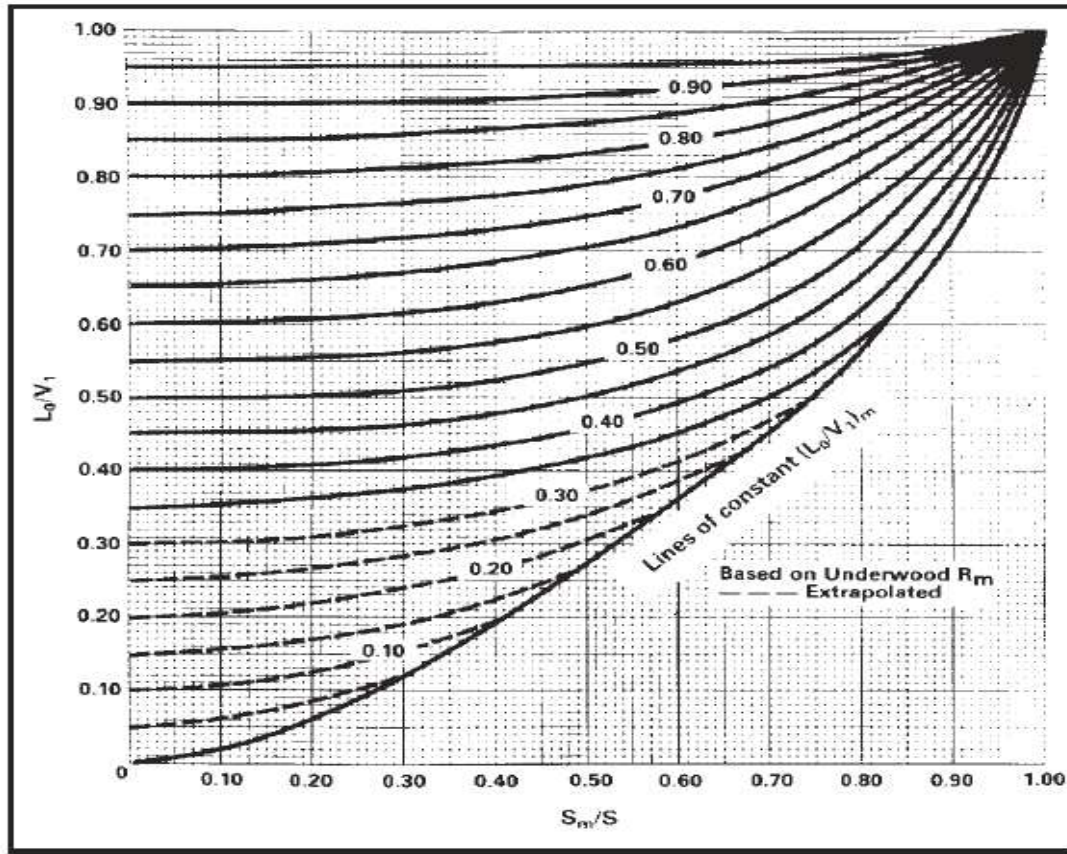
Computation Method

In order to determine the design parameters for a fractionation problem, the following method is recommended:

- Establish feed composition, flow rate, temperature, and pressure.
- Make product splits for the column and establish condenser temperature and column pressure. From column pressure, calculate the reboiler temperature.
- Calculate minimum number of theoretical stages from the Fenske equation (Equation 19-3).
- Calculate minimum reflux rate from the Underwood equations (Equation 19-7 and 19-8).

- Obtain theoretical stages/operating reflux relation from Fig. 19-7.
- Adjust actual reflux for feed vaporization if necessary (Equation 19-9, 19-10).

FIG. 19-7
Erbar-Maddox Correlation of Stages vs Reflux⁶



For the given feed stream, 1102 m³/day (bubble point feed).

Desire: 98% C₃ in the overhead (relative to the feed)

1% iC₄ in the overhead

Air cooling (49°C Condensing Temperature)

Feed Composition Mol % Moles/hr

Feed Composition	Mol %	Moles/hr
C ₂	2.07	21.5
C ₃	48.67	505.6
iC ₄	10.11	105.0
nC ₄	24.08	250.1
iC ₅	5.41	56.2
nC ₅	4.81	50.0
C ₆	4.85	50.4
	100.00	1038.8

Find the:

- Minimum trays required
- Minimum reflux ratio
- Actual trays at 1.3 times the minimum reflux ratio

Solution Steps

Estimate Product Splits from Material Balance:

	Overhead		Bottoms	
	Moles	Mol %	Moles	Mol %
C ₂	21.5	4.1	—	—
C ₃	495.4	94.9	10.2	2.0
iC ₄	5.2	1.0	99.8	19.3
nC ₄	—	—	250.1	48.4
iC ₅	—	—	56.2	10.9
nC ₅	—	—	50.0	9.7
C ₆	—	—	50.4	9.7
Totals	522.1	100.0	516.7	100.0

Get tower pressure at 49°C (bubble point calculation). Using K-values from Section 25:

	K	α
C ₂	2.80	
C ₃	0.93	2.067
iC ₄	0.45	

bubble point pressure = 1930 kPa

Estimate the bottom temperature using K-values at 280 psia (bubble point calculation) assuming negligible pressure drop:

	K	α
C ₃	2.30	1.643
iC ₄	1.40	
nC ₄	1.15	
iC ₅	0.68	
nC ₅	0.62	
C ₆	0.15	

bubble point temperature = 121°C

$$\alpha_{avg} = 1.855$$

Determine the minimum number of trays (Equation 19-3):

$$S_F = \left(\frac{94.9}{2.0} \right) \left(\frac{19.3}{1.0} \right) = 915.8$$

$$S_m = \frac{\log(915.8)}{\log(1.855)} = 11 \text{ trays}$$

Correct for change in relative volatility by using Equation 19-6:

$$K_{LH} = 0.93 = \beta_H (0.45)^b \text{ (condenser)}$$

$$K_{LH} = 2.30 = \beta_H (1.4)^b \text{ (reboiler)}$$

dividing gives $2.473 = (3.111)^b$

$$b = 0.798; \beta_H = 1.759$$

$$S_m = \frac{\log \left[\left(\frac{495.4}{10.2} \right) \left(\frac{99.8}{5.2} \right)^{0.798} \left(\frac{516.7}{522.1} \right)^{1-0.798} \right]}{\log(1.759)}$$

$$S_m = \frac{\log(512.1)}{\log(1.759)} = 11.05$$

Thus correcting for changing α did not affect the minimum tray calculation in this example.

Find the minimum reflux, R_m (Equation 19-7, 8):

	X_F	α_{avg} relative to C_6+	$\alpha X_F / (a - \theta)$		
			$\theta = 16$	$\theta = 15$	$\theta = 15.8$
C_2	0.0207	68.33	0.0270	0.0265	0.0269
C_3	0.4867	26.67	1.2165	1.1123	1.1941
iC_4	0.1011	13.83	-0.6443	-1.1951	-0.7098
nC_4	0.2408	11.00	-0.4506	-0.6622	-0.5518
iC_5	0.0541	5.83	-0.0310	-0.0344	-0.0316
nC_5	0.0481	5.00	-0.0219	-0.0241	-0.0223
C_6	0.0485	1.00	-0.0032	-0.0035	-0.0033
Total			0.0925	-0.7805	-0.0978

$$\theta = 15.9$$

$$R_m + 1 = \frac{68.33 (0.041)}{68.33 - 15.9} + \frac{26.67 (0.949)}{26.67 - 15.9} + \frac{13.83 (0.01)}{13.83 - 15.9}$$

$$R_m + 1 = 2.336$$

$$R_m = 1.336$$

Theoretical trays at $R = (1.3) R_m = 1.737$

$$L_0/V_1 = \frac{R}{R + 1} = \frac{1.737}{(1.737 + 1.0)} = 0.635$$

$$(L_0/V_1)_m = \frac{R_m}{R_m + 1} = \frac{1.336}{(1.336 + 1.0)} = 0.572$$

$$S_m/S = 0.54 \text{ (Fig. 19-7)}$$

$$S = 20.46 \text{ trays (use 21 trays)}$$

Tray Columns

Internals

Various types of trays are used in fractionation columns. Fig. 19-8 presents an isometric sketch of the top two trays in a fractionation column with bubble caps. The bubble caps, along with the weirs and down comers, maintain a liquid level on the trays. The liquid flows across the tray, into the down comer, and across the next tray in the opposite direction. The vapor flows up through the caps and through the slots mixing with the liquid.

Fig. 19-9 shows the vapor flow through bubble cap trays, sieve trays, and valve trays. Due to the riser in the bubble cap, it is the only tray which can be designed to prevent liquid from “weeping” through the vapor passage. Sieve or valve trays control weeping by vapor velocity. The bubble cap tray has the highest turndown ratio, with designs of 8:1 to 10:1 ratio being common. Bubble cap trays are commonly found in existing glycol dehydration columns but new designs typically use structured packing. Valve and sieve trays are popular due to the lower cost and increased capacity over bubble cap trays for a given tower diameter.

Fig. 19-10 shows two valve designs. The upper drawing shows a floating valve free to open and close with varying vapor flow rates. The lower drawing shows a “caged” valve which prevents valve loss due to erosion of the tray. Various other designs are common such as using multiple disks and rectangular valves. Valves of assorted weights have also been used to increase flexibility. The sieve or perforated tray is the simplest construction of the three general types and thus is the least expensive option. The sieve tray is simply a plate with holes for vapor passage. Although the sieve tray generally has lower pressure drop than other tray types, its main disadvantage is that sieve trays will be susceptible to “weeping” or “dumping” of the liquid through the holes at low vapor rates and its turndown capacity is limited. Vendor-proprietary fixed valves and mini-valves are also available for increased capacity. Fixed valves can provide turndown capacity between that of sieve trays and standard (movable) valves. Larger fixed valves are used as alternatives to sieve trays in fouling services.

Examples include

Koch-Glitsch PROVALVE®, VG-10 (fixed valves)

VG-0 (fixed mini-valve)

MV-1 (movable mini-valve)

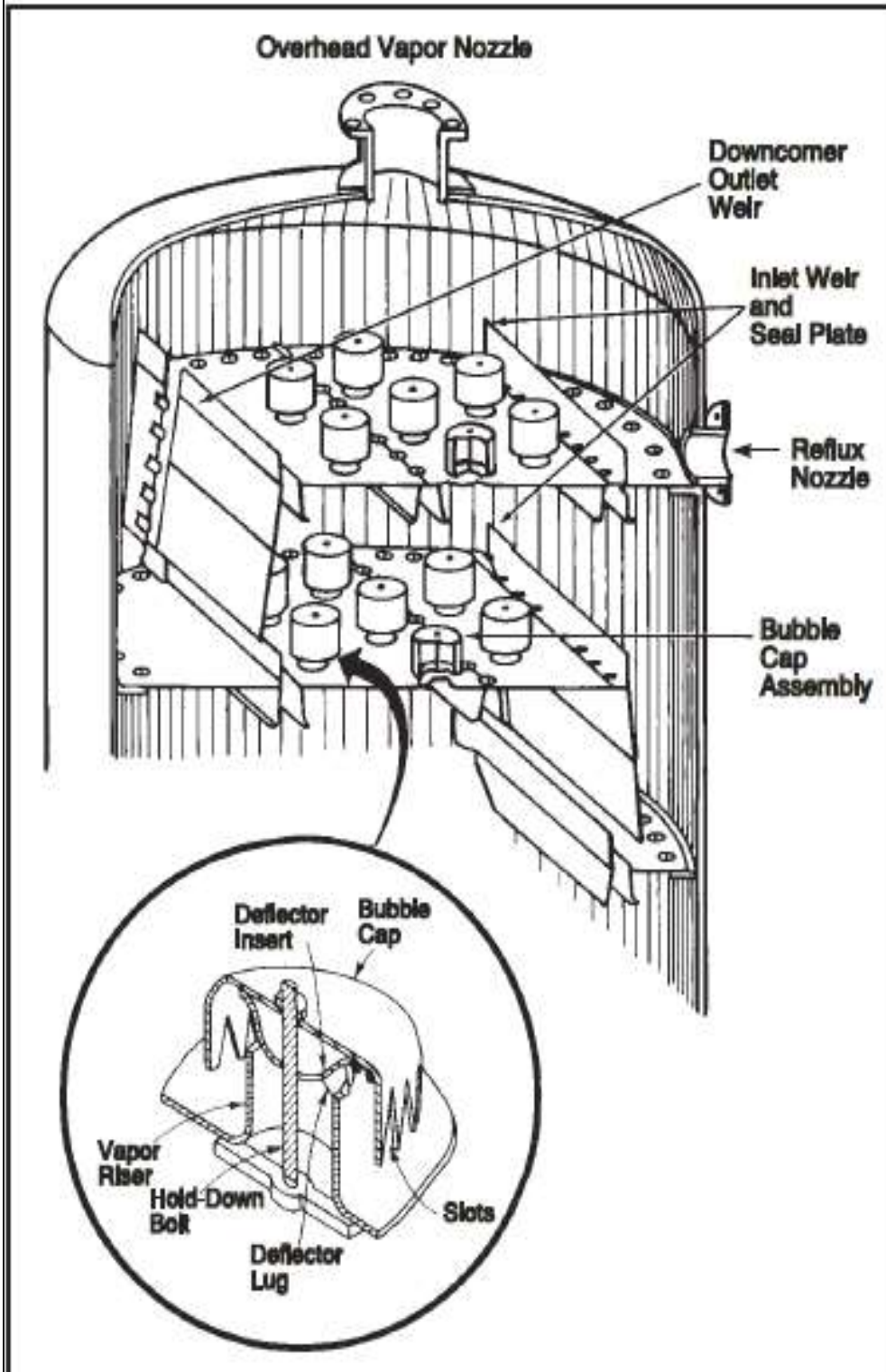
Sulzer SVG (fixed valve)

MVG (fixed mini-valve)

Trayed columns generally provide satisfactory operation

Over a wide range of vapor and liquid loadings. Fig. 19-11 shows

FIG. 19-8
Top Two Trays of a Bubble-cap Column²⁸



operating characteristics for a representative system. The vapor and liquid rates can vary independently over a broad range and the column will operate satisfactorily. At low vapor rates unsatisfactory tray dynamics may be characterized by vapor pulsation, dumping of liquid, or uneven distribution. At high vapor rates, the tower will eventually flood as liquid is entrained to the tray above or backed-up in the downcomers. At low liquid

FIG. 19-9
Flow Through Vapor Passages²⁸

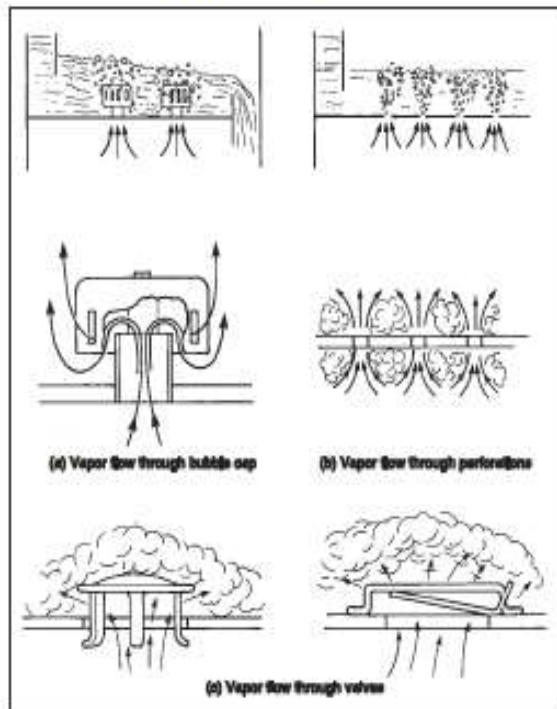
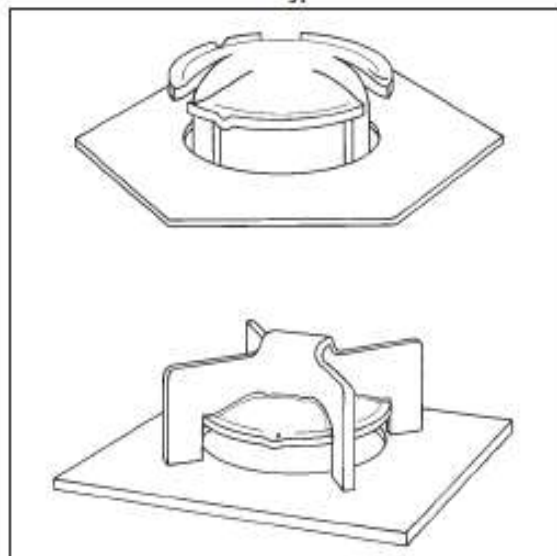


FIG. 19-10
Valve Types²⁸



rates, poor vapor-liquid contact can result. High liquid rates can cause flooding and dumping as the liquid capacity of the downcomers is exceeded.

In order to handle higher liquid rates, more downcomer area is required. This is often achieved by using multiple pass trays. Multipass trays increase liquid handling capacity for a given diameter due to increased weir length and reductions in the weir crest. Fig. 19-12 shows various configurations beyond a one pass tray where the liquid phase is split into two to four flow paths to increase liquid handling capacity.

Sizing

"C" Factor Method — Many design methods for sizing trayed fractionators have been used. Generally these methods are oriented toward liquid entrainment limitations or correlations for flooding limits. A simple method called the Souders and Brown equation⁹ involves using a Stokes' Law type formula:

$$V_{max} = (0.3048) C \sqrt{\frac{\rho_L - \rho_v}{\rho_v}} \quad \text{Eq 19-11}$$

Note that ρ_L and ρ_v are at flowing temperature and pressure.

The value of C can be found from Fig. 19-13 based on tray spacing and liquid surface tension. The column diameter is:

$$D_T = \sqrt{\frac{V_{max}}{V_{max} (0.7854)}} \quad \text{Eq 19-12}$$

This method was originally developed for bubble cap trays and gives a conservative diameter, especially for other types of trays.

Nomograph Method — Manufacturers of valve trays have developed design methods for their trays. Design procedures are made available^{10, 11} for preliminary studies. One such procedure starts with the nomograph in Fig. 19-14.¹⁰ This is a simple relationship of liquid rate (GPM) and a quantity V_{load} defined as:

$$V_{load} = CFS \sqrt{\frac{\rho_v}{(\rho_L - \rho_v)}} \quad \text{Eq 19-13}$$

FIG. 19-11
Limits of Satisfactory Tray Operation for a
Specific Set of Tray Fluid Properties⁹

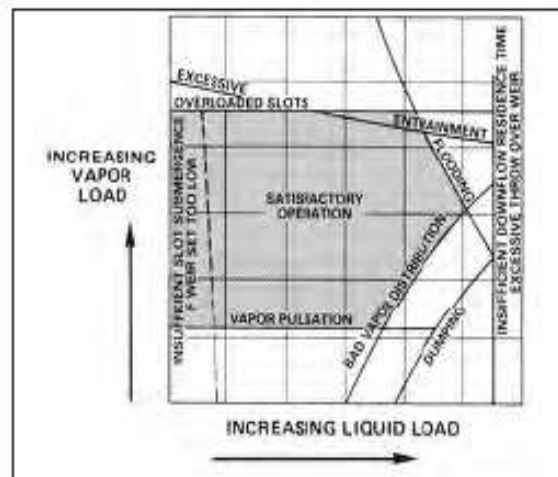
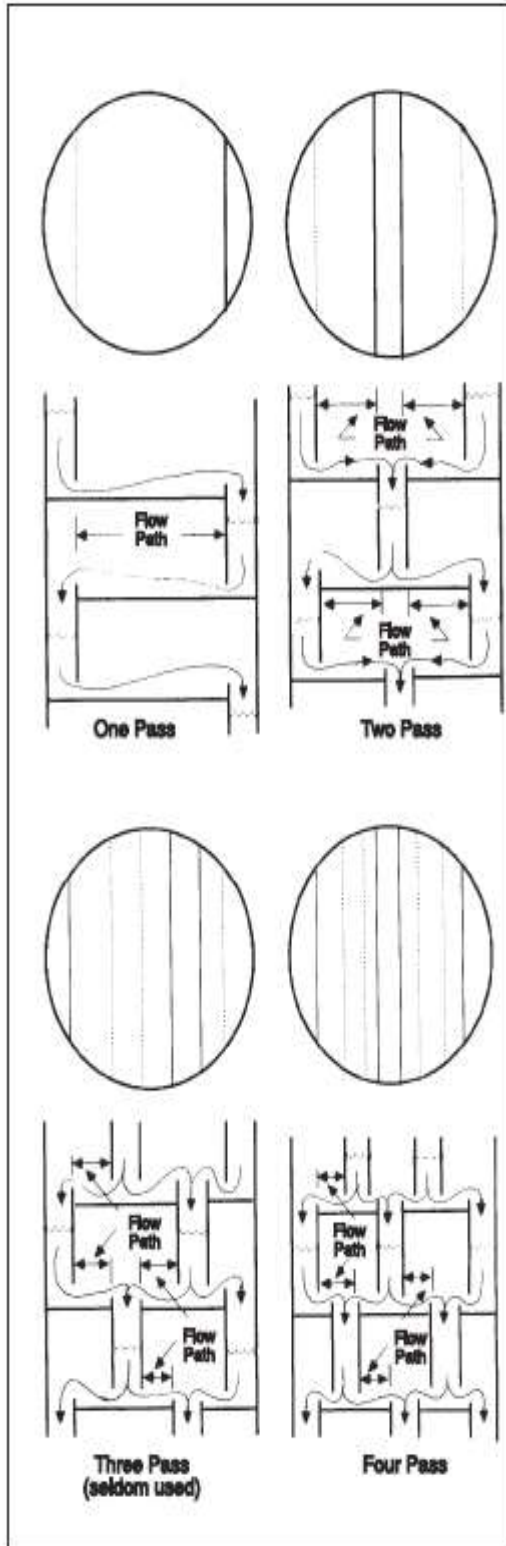


FIG. 19-12
Alternative Liquid Flow Paths



Simplified Hand Method — Tray vendors today provide computer programs to users to size both trayed and packed columns. These vendors should be contacted for copies of their programs for their products. The method included here is a hand method that can be used for preliminary sizing of trayed columns and to understand the key parameters that affect column sizing.

Fig. 19-14 is an approximation only and does not take into account foaming which is a major consideration in many systems. In order to compensate for foaming, a System Factor is used to adjust the vapor and liquid capacities (Fig. 19-15).

The downcomer velocity VD_{sq} is found from Fig. 19-16. VD_{sq} is corrected by the System Factor:

$$VD_{sq} = VD_{sq}^* (\text{System Factor}) \quad \text{Eq. 19-14}$$

The other factor required for this design method is the vapor capacity factor CAF.

$$CAF = CAF_0 (\text{System Factor}) \quad \text{Eq. 19-15}$$

CAF_0 is read from Fig. 19-17. In order to compute the column cross sectional area, three quantities are needed.

FIG. 19-13
Souders-Brown Correlation for Approximate Tower Sizing¹¹

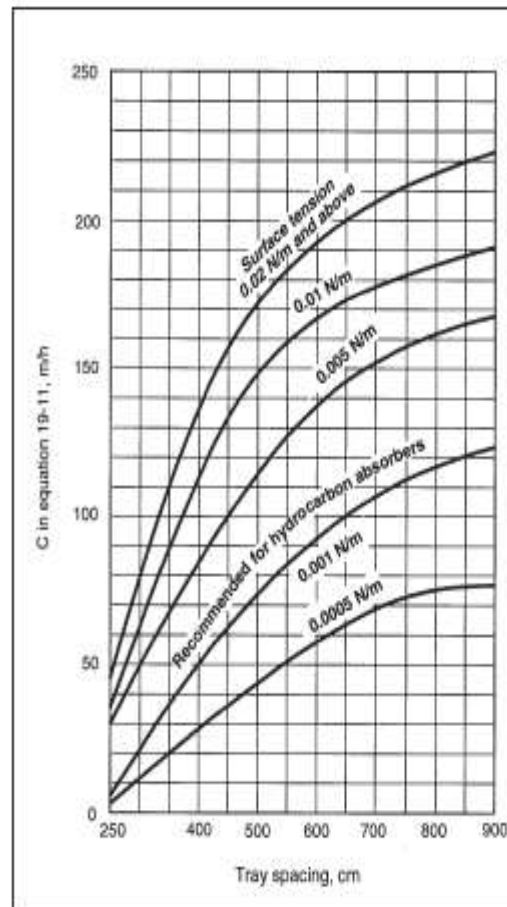
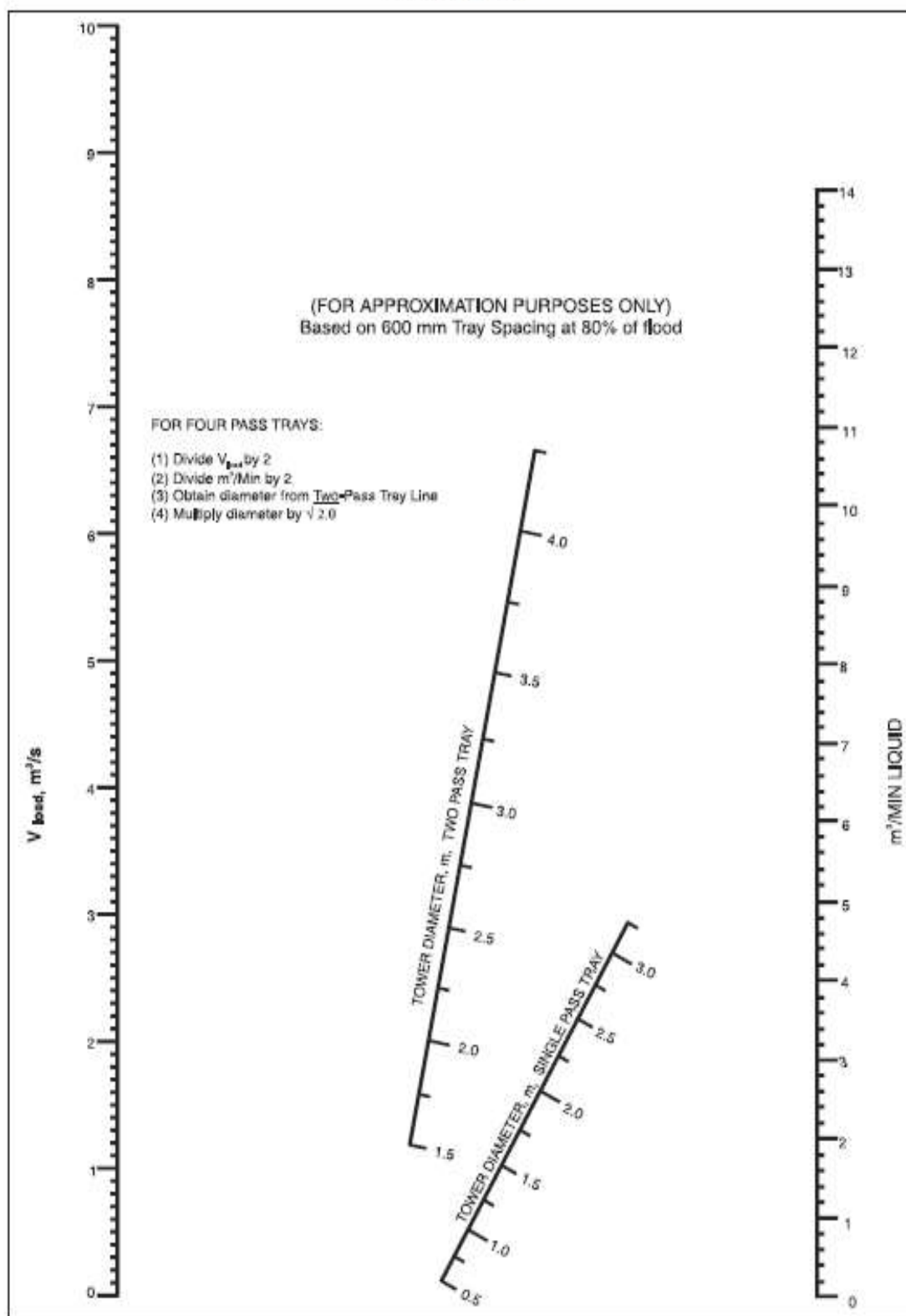


FIG. 19-14
Valve Tray Diameter¹⁰



The flow path length, FPL:

$$FPL = 9 D_T / NP \quad \text{Eq 19-16}$$

D_T and NP are found from Fig. 19-14.

The active area, AAM:

$$AAM = \frac{V_{load} + [(GPM (FPL/530))]}{CAF \cdot FF} \quad \text{Eq 19-17}$$

FF, the flooding factor commonly used is 0.82 for most systems.

The downcomer area, ADM:

$$ADM = GPM / (VD_{avg} \cdot FF) \quad \text{Eq 19-18}$$

If ADM is less than 11% of AAM, use 11% of AAM or double ADM, whichever is smaller.

The tower cross sectional area is then:

$$ATM = AAM + 2 (ADM) \quad \text{Eq 19-19}$$

or

$$ATM = \frac{V_{load}}{0.78 \cdot CAF \cdot FF} \quad \text{Eq 19-20}$$

The larger of these two values is used. Then:

$$D_T = \sqrt{ATM / 0.7854} \quad \text{Eq 19-21}$$

Example 19-3 — Determine the diameter of a depropanizer with the following data:

vapor rate: 1994 m³/h
vapor density: 48 kg/m³
liquid rate: 4.5 m³/min
liquid density: 461 kg/m³
liquid surface tension: 0.0033 N/m
tray spacing: 61cm

"C" Factor Method

From Fig. 19-13: C = 430

$$v_{max} = 430 (0.3048) \sqrt{\frac{461 - 48}{48}} = 384.4 \text{ m/h} \quad \text{from Equation 19-11}$$

$$D_T = \sqrt{\frac{1944}{384.4 (0.7854)}} = 2.54 \text{ m} = 2540 \text{ mm} \quad \text{from Equation 19-12}$$

Nomograph (Fig. 19-14)

$$V_{load} = \frac{1994}{3600} \sqrt{\frac{48}{461 - 48}} = 0.1889 \text{ m}^3/\text{s} \quad \text{from Equation 19-13}$$

then from Fig. 19-14 @ GPM = 4.5 m³/s

$$D_T = 290 \text{ cm for a 1 pass tray} = 2900 \text{ mm} \\ 229 \text{ cm for a 2 pass tray} = 2290 \text{ mm}$$

Detailed Method

$$VD_{avg} = 7.6 \text{ m}^3/\text{min}/\text{m}^2 \text{ at } \rho_v - \rho_L = 413$$

$$\text{System Factor} = \frac{2.93}{\left(\frac{413}{48}\right)^{0.35}} = 0.85 \quad (\text{Fig. 19-15})$$

$$VD_{avg} = 7.6 (0.85) \\ = 6.46 \text{ m}^3/\text{min}/\text{m}^2$$

From Fig. 19-17:

$$CAF_v = 0.126 \text{ m/s}$$

$$CAF = (0.126) (0.85) = 0.107 \text{ m/s}$$

$$FPL = \frac{9 (2.29)}{2} = 10.3 \text{ m} \quad \text{from Equation 19-16}$$

$$AAM = \frac{0.1889 + [4.5 (10.3/530)]}{(0.107) (0.82)} \\ = 3.15 \text{ m}^2 \quad \text{from Equation 19-17}$$

$$ADM = \frac{4.5}{(6.46) (0.82)} = 0.85 \text{ m}^2 \quad \text{from Equation 19-18}$$

$$ATM = 3.15 + 2 (0.849) = 4.85 \text{ m}^2 \quad \text{from Equation 19-19}$$

$$ATM = \frac{0.1889}{(0.78) (0.107) (0.82)} = 2.76 \text{ m}^2 \quad \text{from Equation 19-20}$$

$$D_T = \sqrt{\frac{4.85}{0.7854}} = 2.48 \text{ m} = 2480 \text{ mm} \quad \text{from Equation 19-21}$$

A comparison of the methods (rounded to the nearest 50 mm)

C Factor	2550 mm
Nomograph	2300 mm (2900 mm for single pass)
Detailed Method*	2500 mm

FIG. 19-15
System Factors⁸

Systems with foaming tendencies are taken into account by using a factor to derate the capacity of a given tray design. A list of the more common foaming systems and their recommended factor is below.

System	Factor
Absorbers (over -18°C)	0.85
Absorbers (below -18°C)	0.80
Amine Contactor	0.80
Vacuum Towers	0.85
Amine Stills (Amine Regenerator)	0.85
H ₂ S Stripper	0.85
Purifical Fractionator	0.85
Glycol Contactors	0.50
Glycol Stills	0.65
CO ₂ Absorber	0.80
CO ₂ Regenerator	0.85
Caustic Wash	0.65
Caustic Regenerator, Foul Water, Sour Water	0.60
Stripper	
Alcohol Synthesis Absorber	0.35
Hot Carbonate Contactor	0.85
Hot Carbonate Regenerator	0.90
Oil Reclaimer	0.70

The capacity of a given tray design used in *high pressure fractionation services* with a vapor density of 28.8 kg/m³ and higher should be derated by a system factor calculated by the following formula:

$$\text{System factor} = \frac{2.93}{(\rho_v)^{0.35}}$$

FIG. 19-16
Downcomer Design Velocity¹⁰

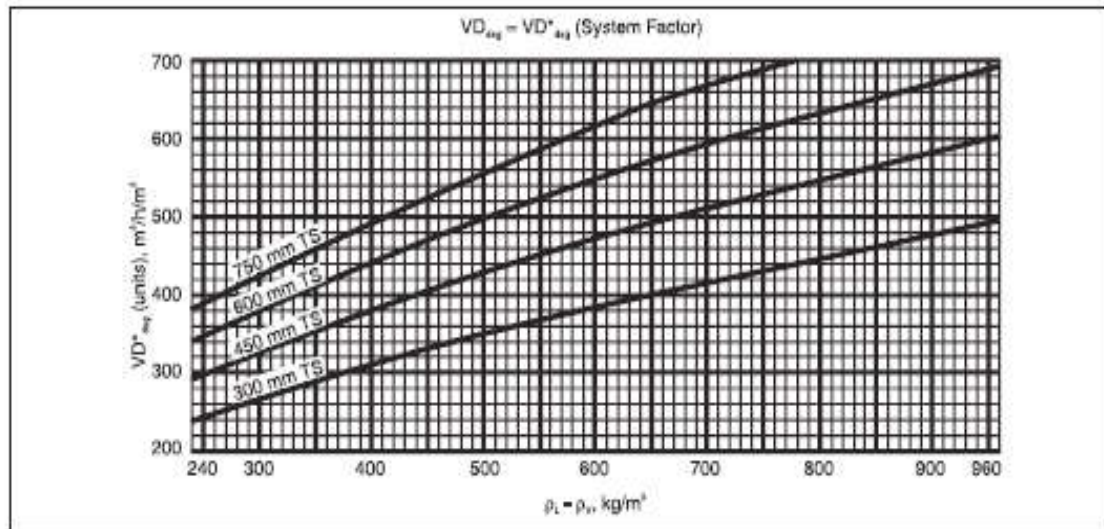
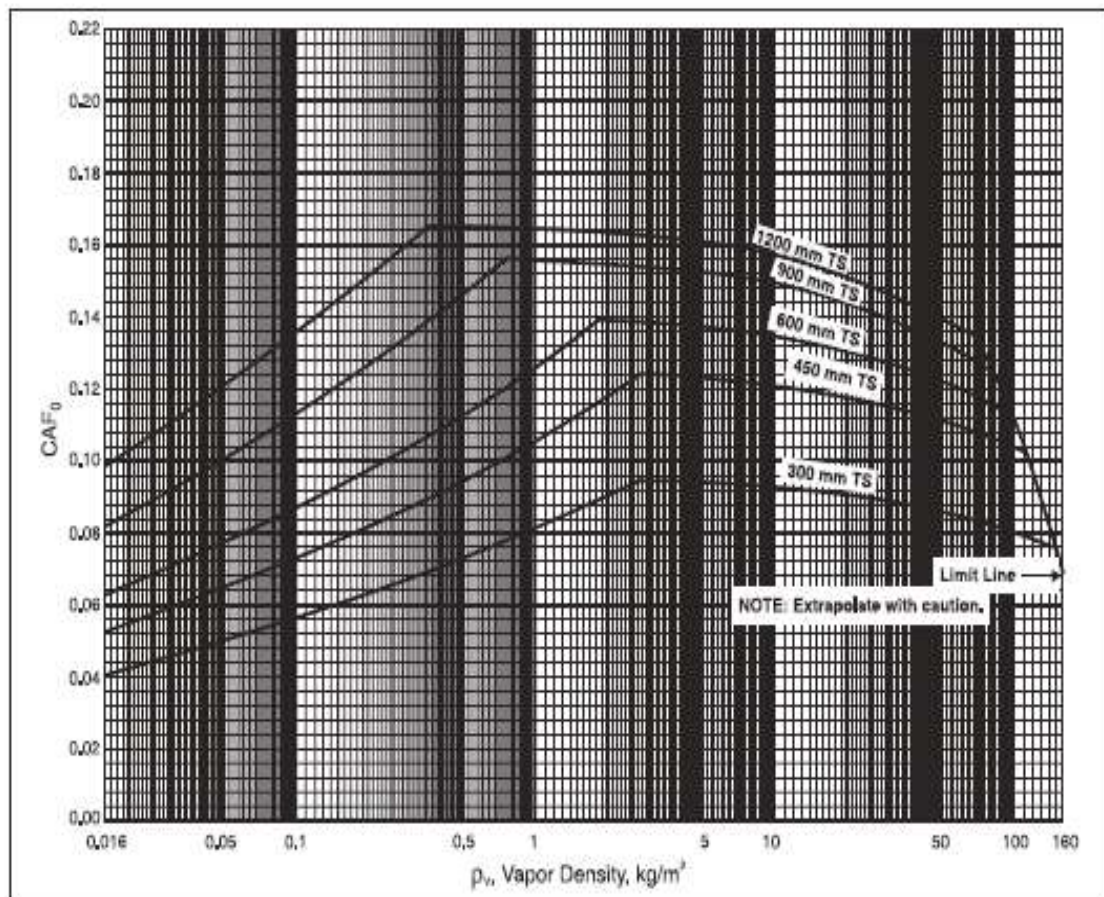


FIG. 19-17
Approximate Flood Capacity of Valve Trays¹⁰



Tray Efficiency

All column design work is performed using theoretical trays. An actual tray will not achieve equilibrium due to limitations of vapor-liquid contact time. In an actual column, more trays are required to obtain the desired separation. This determination is usually accomplished by the use of an overall tray efficiency defined as:

$$\varepsilon = \frac{\text{theoretical trays}}{\text{actual trays}} \quad \text{Eq 19-22}$$

The determination of tray efficiencies from theoretical parameters is the topic of numerous technical articles.^{12, 13, 14} A detailed discussion of this subject is beyond the scope of this book.

O'Connell¹⁵ correlated the tray efficiencies of fractionators and absorbers. For fractionators, this correlation considered thirty-eight systems of which 27 are hydrocarbon fractionators. The correlation, shown in Fig. 19-18, relates overall tray efficiency to relative volatility computed at average column conditions and the feed viscosity at average column conditions.

Example 19-4 — Evaluate the tray efficiency for the system in Example 19-2.

Average column temperature = 85°C

Feed viscosity @ 85°C = 0.076 mPa • s

Average α = 1.854

Solution Steps

$$(\alpha)(\mu) = 0.141$$

From Fig. 19-18, $\varepsilon \cong 80\%$

The system in Example 19-2 required 21 theoretical stages including the reboiler. The total actual trays is:

$$\frac{21-1}{0.80} \cong 25 \text{ trays}$$

Typically an extra tray is added to the tray count for each feed tray and each side exchanger. Using this criteria, this column should have 26 trays.

Typical operating pressures, tray counts, reflux ratio, and tray efficiencies for various gas processing systems are shown in Fig. 19-19. These are not design values; rather guidelines for typical values in previous applications. The actual selection depends on many factors such as feed composition, energy cost, and capital cost.

FIG. 19-18

Effect of Relative Volatility and Viscosity on Plate Efficiency of Fractionating Columns

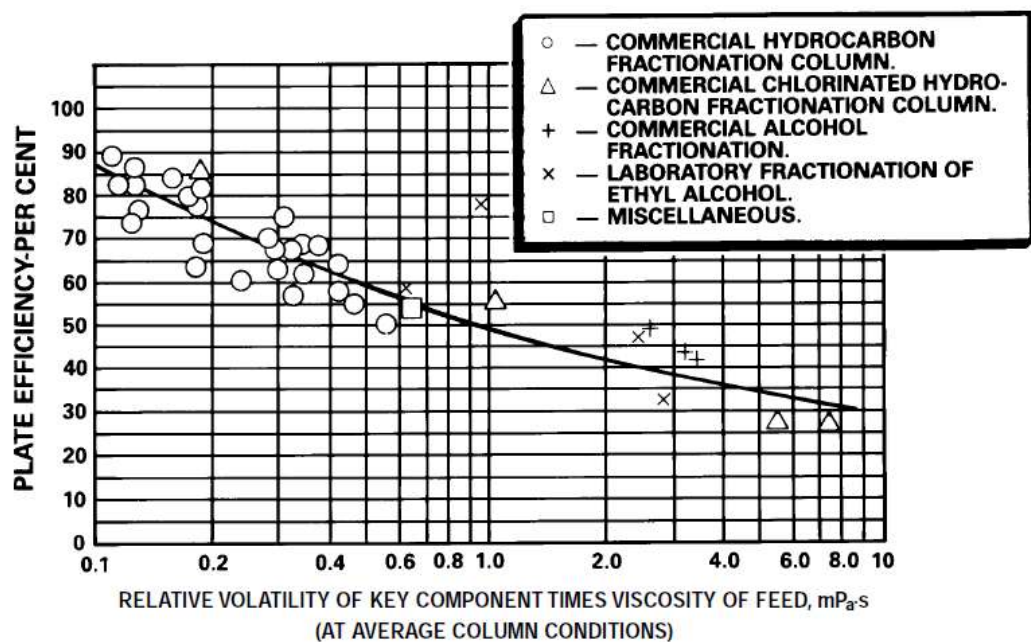


FIG. 19-19

Typical Fractionator Parameters

	Operating Pressure, kPa (ga)	Number of Actual Trays	Reflux ¹ Ratio	Reflux ² Ratio	Tray Efficiency, %
Demethanizer	1380 – 2750	18 – 26	Top Feed	Top Feed	45 – 60
Deethanizer	2590 – 3100	25 – 35	0.9 – 2.0	0.6 – 1.0	60 – 75
Depropanizer	1650 – 1860	30 – 40	1.8 – 3.5	0.9 – 1.1	80 – 90
Debutanizer	480 – 620	25 – 35	1.2 – 1.5	0.8 – 0.9	85 – 95
Butane Splitter	550 – 690	60 – 80	6.0 – 14.0	3.0 – 3.5	90 – 100
Rich Oil Fractionator (Still)	900 – 1100	20 – 30	1.75 – 2.0	0.35 – 0.40	Top 67 Bottom 50
Rich Oil Deethanizer	1380 – 1725	40	–	–	Top 25 – 40 Bottom 40 – 60
Condensate Stabilizer	690 – 2750	16 – 24	Top Feed	Top Feed	50 – 75

¹Reflux ratio relative to overhead product, mol/mol²Reflux ratio relative to feed, m³/m³

High Capacity Trays

The 90s saw the proliferation of High Capacity Trays by distillation equipment vendors and users. These trays employ unconventional down comer and deck configurations to effect vapour and/or liquid handling capability increases, when used to revamp distillation columns. High Capacity Trays have been particularly effective in demethanizers, deethanizers, depropanizers and butane columns.

One such High Capacity Tray, the NYE TRAY®, is shown in Fig. 19-20. This tray increases vapor capacity by raising the receiving pan and increasing the area available for vapor flow. This, and similar trays, employ a cross flow arrangement with liquid traveling horizontally across the decks and vapor bubbling up through the liquid, creating a froth where the mass transfer occurs. Proprietary down comer designs and deck enhancements such as fixed or mini-valves are often employed. Examples of other High Capacity cross flow trays include the following:

NYE Tray (Raschig-Jaeger)
SUPERFRAC® (Koch-Glitsch)
VGPlus (Sulzer ChemTech)

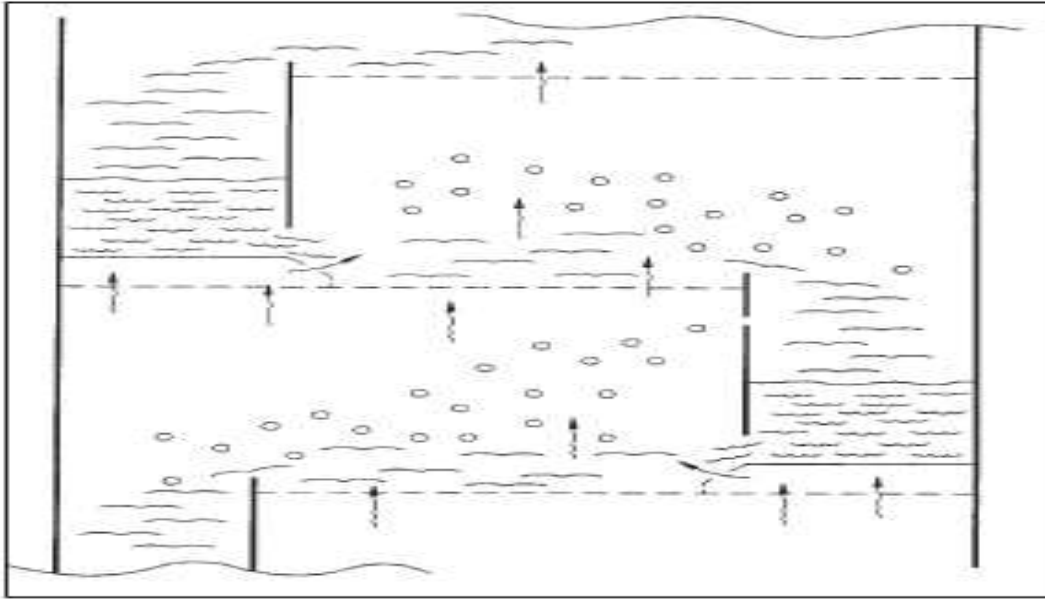
Another style of High Capacity Trays employs multiple down comers to increase weir length and liquid handling capacity beyond what conventional multi-pass trays can provide. Examples of such trays include the following:

MD™ (UOP)
ECMD™ (UOP)
Hi-Fi (Shell Global Solutions)

Efficiencies of these counter flow trays are often lower than those of cross flow trays due to the reduced contact time between the phases. Capacities can be quite high, and tray spacings quite small, due to the very long outlet weir that these trays are capable of providing.

Still another tray configuration, called “cocurrent flow,” is expected to gain greater acceptance in the future. With concurrent flow, the vapor and liquid phases are allowed to flow together, unidirectionally, for a while inside contacting elements. Examples of such trays are ULTRA-FRAC® (Koch-Glitsch), Simulflow (UOP) and ConSep (Shell Global Solutions). Some such trays are only capable of functioning at low liquid rates such as those that are encountered in glycol dehydration columns. Little information is publicly available regarding the efficiency of cocurrent flow trays. Sizing for these, and in fact most High Capacity Trays, is regarded as proprietary by their vendors. Fractionation Research Inc. (FRI) has performed extensive studies of both trays and packing in many configurations. Companies that are members of Fractionation Research Inc. (FRI) have access to this extensive set of test data and correlations to predict capacity and efficiency.

**FIG. 19-20
NYE TRAY Schematic**



Packed Columns

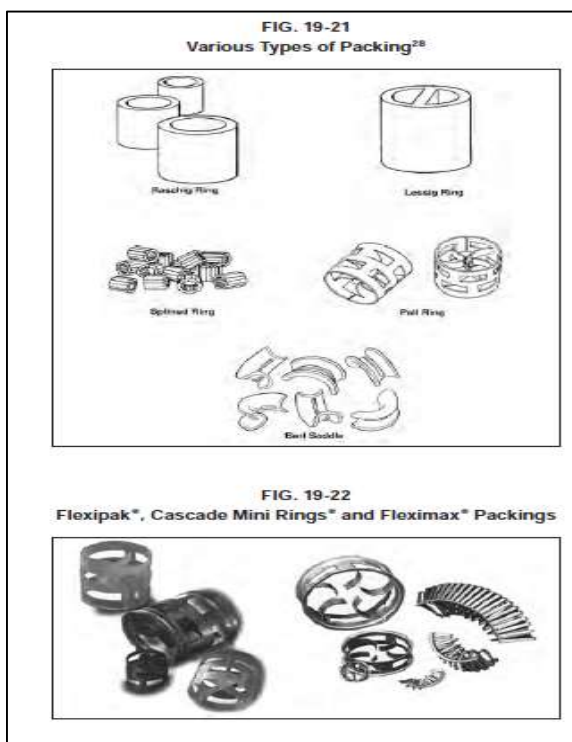
Traditionally the majority of fractionation columns in gas processing plants were equipped with trays. However, an option to trayed columns is to use packing. With packed columns, contact between the vapor and liquid phases is achieved throughout the column rather than at specific levels.

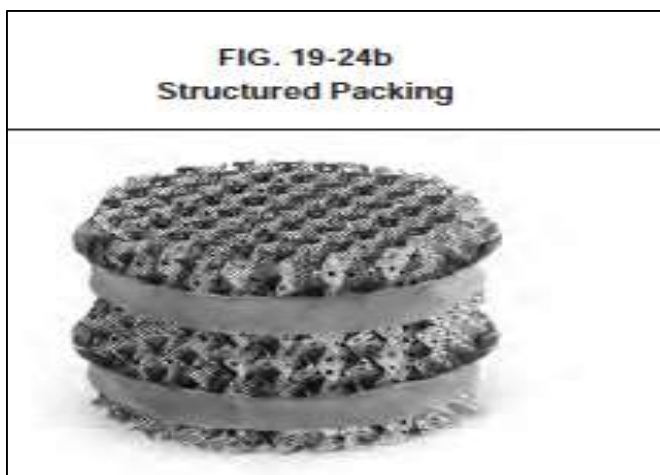
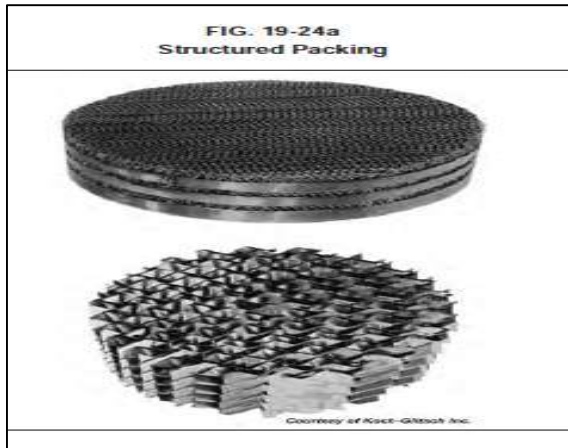
There are generally three types of packed columns:

- Random packing wherein discrete pieces of packing are dumped in a random manner into a column shell. These packings are of a variety of designs. Each design has particular Surface area, pressure drop and efficiency characteristics. Examples of various packing types are shown in Fig. 19-21. Random packing have gone through various Development phases from the first generation packing's which were two basic shapes, the Raschig Ring and the Berl Saddle. Second generation packings include the Pall Ring and the Intalox Saddle which are still used extensively today. Third generation packings come in a multitude of geometries most of which evolved from the Pall Ring and The Intalox Saddle. (Fig. 19-22 and 19-23). Popular third generation packings include IMTP and Nutter Rings. More recent developments include Raschig Super Rings and Koch-Glitsch's Intalox Ultra.
- Structured packing where a specific geometric configuration is achieved. These types of packing can either be the knitted-type mesh packing or sectionalized beds made of Corrugated sheets (Figs. 19-24a and 19-24b). There are a number of commercially available packings which differ in the angle of the crimps, the surface grooves and the use of perforations.
- Grids which are systematically arranged packing which use an open lattice structure. These types of packings have found application in vacuum operation and low pressure drop applications. Little use of these types of packings are seen in high pressure services.

Structured packing has found application in low liquid loading applications which are below $49 \text{ m}^3/\text{h}/\text{m}^2$. Structured packing has performed very well in extremely low liquid loading applications such as glycol dehydration (See Section 20). The high surface tension in glycol dehydrators also helps the structured packing to perform well. Care is needed in the design to ensure sufficient liquid to wet the entire packing surface.

Above $49 \text{ m}^3/\text{h}/\text{m}^2$, random packings are more advantageous. Structured packings have been tried in fractionators with little success. Numerous case of structured packing failures have been experienced in high pressure and/or high-liquid rate services. Predicting structured packing efficiency has been problematic and unreliable in high pressure systems. Structured packings generally have lower pressure drop per theoretical stage than random packings. This can be important in low pressure applications but not for high pressure NGL fractionators.





Column Sizing

The Eckert generalized pressure drop correlation (GPDC) is often used for sizing randomly packed columns. The chart in Fig. 19-26, which is a modified correlation, can be used to predict pressure drop for a given loading and column diameter. Alternatively, for a given pressure drop the diameter can be determined.

Most packed columns are designed for pressure drop of between 5 and 15 mm of water per foot of packed depth with 25 mm of water being the maximum.

The packing factors for various packings are shown in Fig. 19-25. Broadly speaking, packings smaller than 25 mm size are intended for towers 300 mm or smaller in diameter, packings 25 mm or 38 mm in size for towers over 300 mm to 900 mm in diameters, and 50 or 75 mm inch packings are used for towers three or more feet in diameter. This results from trade-offs of capacity and efficiency. The designer should select the proper size of packing, and therefore the proper packing factor for calculations.

The packing factors in Fig. 19-25 are average values which are sufficient for preliminary sizing but specific packing vendors should be contacted for design applications.

The GPDC has limitations in describing the performance of packings. Efforts to improve the correlation for specific packing geometries have led to the development and publication of charts for each packing which strive to correlate packing performance information with the same abscissa and ordinate as the GPDC chart. As discussed previously, FRI member companies have access to an extensive set of test data and correlations to predict capacity and efficiency for many packings.

Example 19-5 — Determine the packed column diameter for example 19-3 using 50 mm plastic Pall rings.

Given: $\mu = 7.6 (10^{-5}) \text{ Pa} \cdot \text{s}$

$\Delta P = 42 \text{ mm water/m of packing}$

Solution Steps

$$M_L = (4.5) (60) (461) = 124\,470 \text{ kg/h}$$

$$M_G = (1994) (48) = 95\,712 \text{ kg/h}$$

$$\frac{L_p}{G_p} \sqrt{\frac{\rho_v}{\rho_L}} = \frac{M_L}{M_G} \sqrt{\frac{\rho_v}{\rho_L}} = \frac{124\,470}{95\,712} \sqrt{\frac{48}{461}} = 0.42$$

From Fig. 19-26 at $\Delta P = 42 \text{ mm water/m of packing}$:

$$\frac{G_p^2 \mu_L^{0.1} F_p (\rho_w/\rho_L)^{0.1}}{\rho_v (\rho_L - \rho_v)} = 0.024$$

From Fig. 19-25, $F_p = 26$, then:

$$G_p^2 = \frac{(0.024) (461 - 48) (48)}{(0.000076)^{0.1} (26) (1000/461)^{0.1}} = 43.73$$

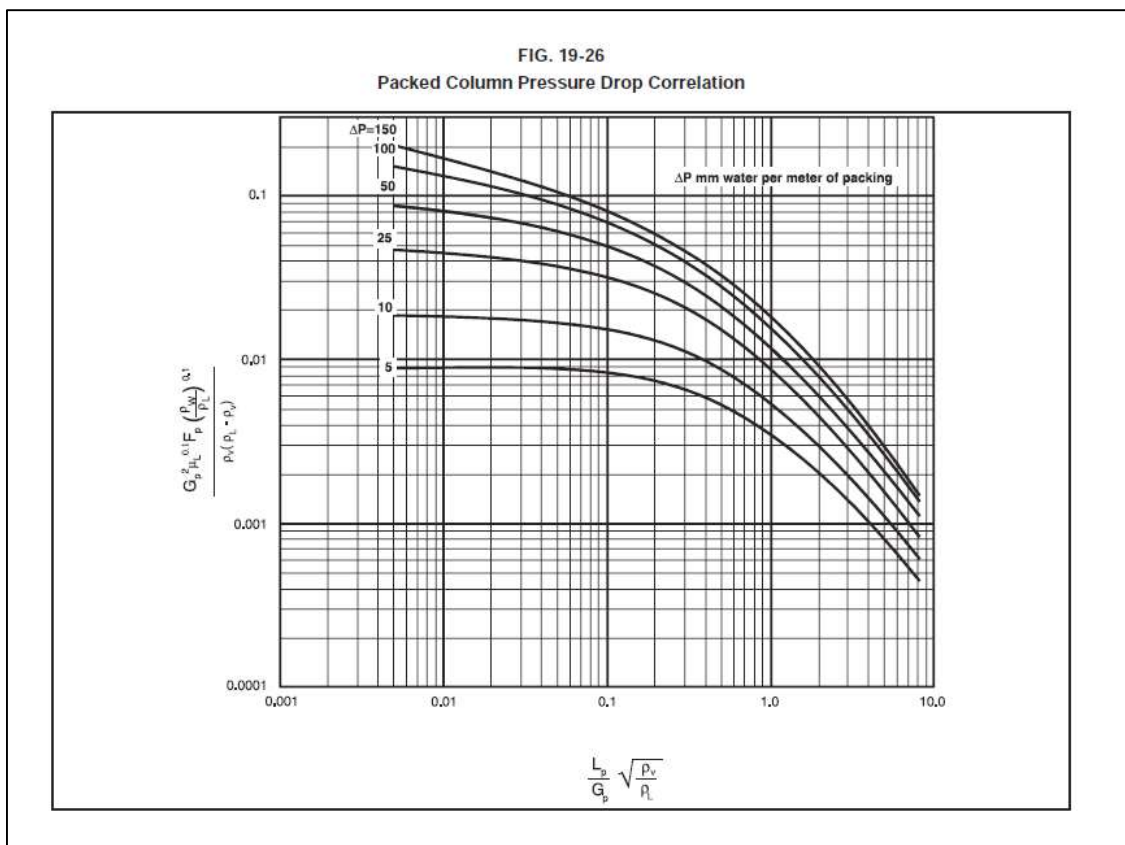
$$G_p = 6.61 \text{ kg}/(\text{m}^2 \cdot \text{s})$$

$$A_c = \frac{95\,712}{(6.61) (3600)} = 4.02 \text{ m}^2$$

$$D_T = 2.26 \text{ m}$$

FIG. 19-25
Packing Factors (F_p) (Dumped Packing)

Packing Type	Material	Nominal Packing Size (mm)										
		6	9	12	15	18	25	31	37	50	75	87
IMTP®	Metal				51		40		24	18	12	
Hy-Pak™	Metal						45		29	26		16
Super Intalox Saddles®	Ceramic						60			30		
Super Intalox Saddles®	Plastic						40			28		18
Pall Rings	Plastic				75		55		40	26		17
Pall rings	Metal				70		56		40	27		18
Intalox Saddles®	Ceramic	725	330	200		145	92		52	40	22	
Raschig Rings	Ceramic	1600	1000	580	380	255	155	125	95	65	37	
Raschig Rings	1/32" Metal	700	390	300	170	155	115					
Raschig Rings	1/16" Metal			410	300	220	144	110	93	62	32	
Berl Saddles	Ceramic	900		240		170	110		65	45		
Flexiring	Metal						49			23		16
Fleximax	Metal						35		26	17	11	
Cascade Mini Rings	Metal						39	36	33	26	18	
Cascade Mini Rings	Plastic						44		33	20	17	
NOTE: Values in this table are average values for the packing factor (F_p). F_p is actually a function of loading. Specific correlations for each packing from the vendors should be used for design purposes.												



Packing Height

In order to determine the height of a packed column bed, the height of a theoretical plate, HETP, is required. HETP times the number of theoretical stages gives the height of the packing. Generally HETPs range from 300 to 900 mm but can be as high as 1500 mm. Packed columns have found wide usage in cryogenic plant demethanizers. Typical HETP's for demethanizers are 91 cm for the upper section and 76 cm for the lower section.

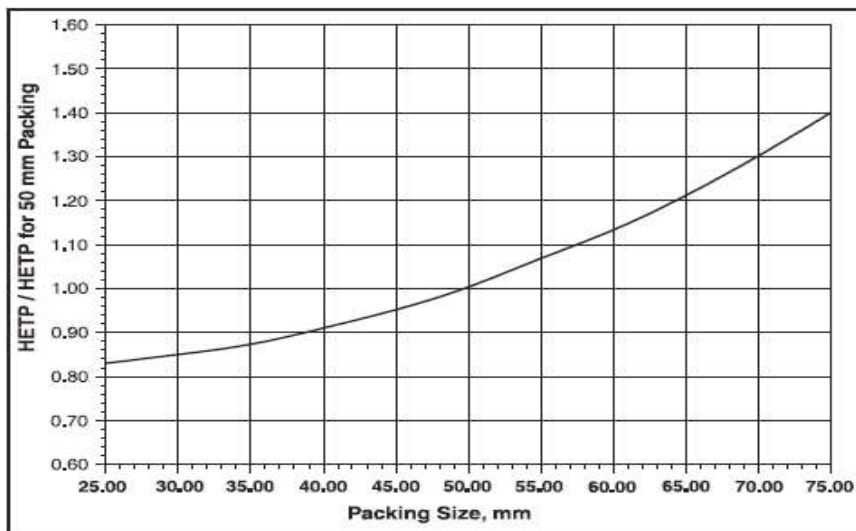
The prediction of the HETP from theory or empirical relations is a complex subject. Recent research by Fractionation Research Inc. has underscored the sensitivity of HETP with a number of variables. HETP is a function of flow rates and properties of the system as well as the specific geometric and mechanical factors. In order to determine packing requirements, a packing manufacturer should be consulted.

Fig. 19-27 provides some example HETPs for hydrocarbon systems in the gas processing industry. HETP's are also a function of the packing size. In general, the smaller packings have lower HETP values. Fig. 19-28 shows an example trend of packing HETP's for one type of packing.

FIG. 19-27
Typical Packing Depths¹⁸

System	L/G, kg/(h · m ²)	Diam. cm	Packing			Bed depth, m	HETP, m	HTU, m	System press., kPa (abs)	ΔP, in mm H ₂ O/m pkg	% Overhead
			Type	Size, in.	Size, mm						
Absorber	350 / 470	91	Pall rings	2	50	7	0.85	—	5964	46	—
L.O.-Top fractionator	150 / 200	91	Pall rings	2	50	5	0.76	—	1082	10	—
L.O.-Bottom fractionator	450 / 240	122	Pall rings	2	50	5	0.85	—	1082	25	—
Deethanizer top	470 / 190	46	Pall rings	1½	37	6	0.88	—	2068	17	—
Deethanizer bottom	800 / 190	76	Pall rings	2	37	55	1.00	—	2068	25	—
Depropanizer top	240 / 175	59	Pall rings	1½	37	4.9	0.98	—	1862	25	—
Depropanizer bottom	240 / 175	59	Pall rings	1½	37	7.3	0.73	—	1862	25	—
Debutanizer top	80 / 130	50	Pall rings	1½	37	3.7	0.73	—	621	10	—
Debutanizer bottom	80 / 130	50	Pall rings	1½	37	5.5	0.61	—	621	10	—
Pentane-iso-pentane	90 / 80	46	Pall rings	1	25	2.7/2.3	0.46	—	Atmos.	33	—
Light and heavy naphtha	30 / 50	38	Pall rings	1	25	3	0.61	0.62	100 mm. Hg	92	95.0
	13 / 25	38	Pall rings	1	25	3	0.99	0.76	100 mm. Hg	17	95.0
	26 / 60	38	Pall rings	1	25	3	0.44	0.38	100 mm. Hg	146	97.5
	10 / 27	38	Pall rings	1	25	3	0.44	0.40	100 mm. Hg	17	97.5
	16 / 36	38	Intalox	1	25	3	0.70	0.60	100 mm. Hg	67	93.0
	9 / 21	38	Intalox	1	25	3	0.82	0.64	100 mm. Hg	18	99.0
	14 / 34	38	Raschig rings	1	25	3	0.59	0.43	100 mm. Hg	92	91.6
	9 / 21	38	Raschig rings	1	25	3	0.82	0.60	100 mm. Hg	33	96.5
Iso-octane Toluene	83 / 97	38	Pall rings	1	25	3	0.41	0.41	Atmos.	58	82.0
	40 / 90	38	Pall rings	1	25	3	0.58	0.66	Atmos.	8	76.0
	90 / 110	38	Pall rings	1	25	3	0.24	0.31	Atmos.	142	84.0
	40 / 50	38	Pall rings	1	25	3	0.47	0.43	Atmos.	12	74.0
	46 / 55	38	Pall rings	1	25	3	0.41	0.39	100 mm. Hg	90	92.5
	21 / 25	38	Pall rings	1	25	3	0.57	0.55	100 mm. Hg	17	87.0
	43 / 55	38	Pall rings	1	25	3	0.51	0.49	100 mm. Hg	95	92.0
	20 / 55	38	Pall rings	1	25	3	0.63	0.61	100 mm. Hg	17	89.0
Gas plant asorber	200 / 250	122	Pall rings	2	50	7	0.88	—	6205	9	92.00 propane absorbed

FIG. 19-28
Example Effect of Packing Size on HETP



Packed Column Internals

A critical consideration in packed columns is the control of the vapor and liquid phases. Fig. 19-29 shows a cross section of a packed tower with various internals. Each section of packing is supported by a support plate or grid whose function is to carry the weight of the bed with minimum pressure drop. Hold-down grids are used at the top to prevent lifting of the bed by the vapor phase. Liquid distribution is a critical consideration in packed columns. Poor liquid distribution causes dramatic loss of efficiency. Various designs have been used to distribute liquid feeds and to collect and redistribute the liquid at various points in the tower. Generally the liquid should be redistributed every 6 m of packing height or every 10 column diameters, whichever is smaller. Each of the packing vendors has several types of distributors. The design and configuration of the distributor varies according to tower vapor and liquid loading.

Dumped Packing versus Trays

Packed columns have been used extensively in the chemical industry for many years. Packings are selected instead of trays for several reasons:

- Pressure drop — packed towers usually yield a lower pressure drop per theoretical stage. This can be important for low pressure operations. However, at the elevated pressures encountered in natural gas processing, column pressure drop is usually not a major issue.
- High liquid loading — for high liquid-to-vapor ratio systems, a packed column will have more capacity for a given diameter. Some fractionation applications are characterized by low liquid/vapor ratios and packing has less of an advantage for these designs.
- Foaming — in systems prone to foaming, packing is preferred over trays because packing does not promote foam. Packing is less susceptible to foaming so credit may be taken in the capacity correlation itself or by applying a less severe system factor (than trays in the same service) to the existing correlation. Designers should consult with their company fractionation specialist and/or the packing vendor to confirm the appropriate system factor to use.
- Corrosion — for corrosive systems, packing can be fabricated from ceramics or plastics. Trays may have to be fabricated from expensive alloy materials.

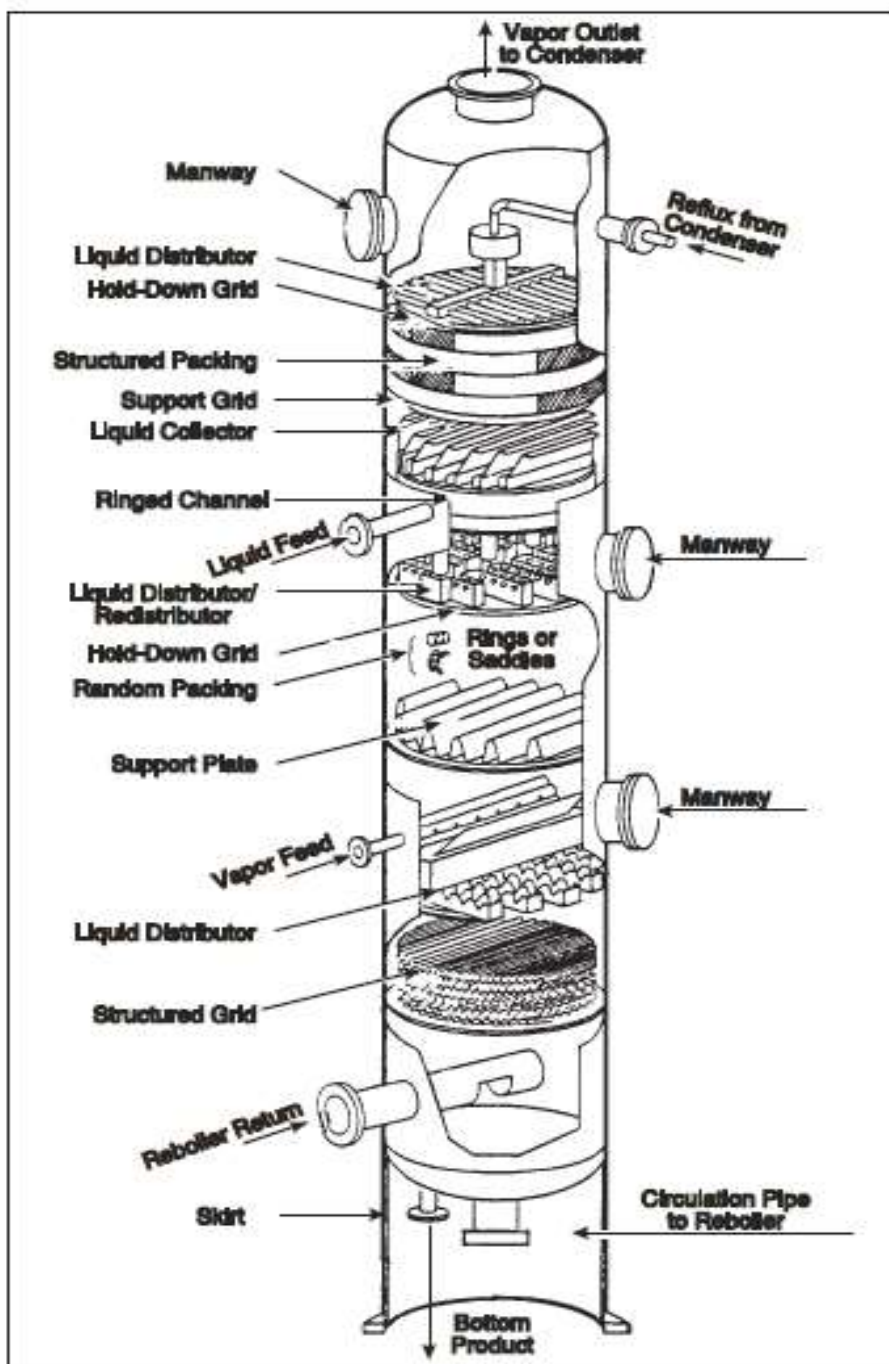
Packed columns also have several disadvantages which must be taken into account in a fractionation design:

- Liquid distribution — In trayed columns, the liquid phase is forced to flow across a tray surface. With gas bubbling through the liquid, contact is assured. In packed towers, the liquid and vapor are free to seek their own flow paths, and channelling can occur. It is critical that the liquid phase be properly distributed at the top of the column and be redistributed at 6 m intervals or every 10 column diameters, whichever is smaller.
- Turndown — Turndown in a packed tower is usually limited by the liquid distributor design. Typical distributor designs can achieve 50% turndown, but special designs can provide a wide operating range, e.g. 10:1. Note that the high turndown distributor often occupies more tower height than a standard design. Turndown for a tray varies with tray type and design details. Moveable valve trays typically can be operated at 20- 30% of full load, and special valve tray

designs can operate at even lower turndown. This can be important in situations where gas production is phased in and throughput rates build up over time.

- Plugging — Packed towers will be more susceptible to plugging from dirt and other foreign materials.
- Packing Height — The HETP for a packed column is an uncertain matter for new services. Often they must be determined by testing or field applications. HETP's can vary from a few inches to several feet.
- Inspection - It is difficult to inspect internals without removing all the contents of a column.

FIG. 19-29
Example Packed Column Internals²⁹



4.12 SEPARATOR

Separator is a pressure vessel used for separating well fluids from oil & gas wells into gaseous and liquid components. A separator for petroleum production is a large vessel designed to separate production fluids into the components of oil, gas and water.

Types of Separator

➤ Flash Vessel

A vessel used to separate the gas evolved from liquid flashed from a higher pressure to a lower pressure.

➤ Gas – Liquid- Soli Separator (Filter Separator)

A filter separator usually has two compartments. The first compartment contains filter-coalescing elements. The liquid particles coalesce into large droplet and when the droplets reach sufficient size, the gas flow causes them to flow out of the filter elements into the center core. The particles are then carried into the second compartment of the vessel (containing a vane-type or knitted wire mesh mist extractor) where the larger droplets are removed.

➤ Scrubber (or) Knock out drum (Gas-Liquid Separator)

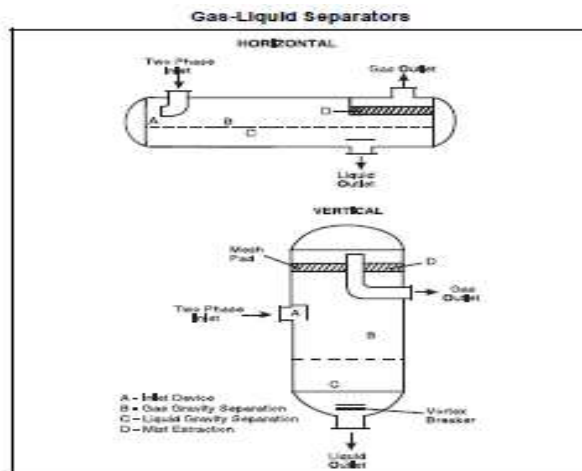
A vessel designed to handle streams with high gas to liquid ratios (GLR). The liquid is generally entrained as mist in the gas or is free flowing along the pipe wall. These vessels usually have a small liquid collection section.

➤ Two Phase Separator (Gas-Liquid Separator)

A vessel that separates the well fluid into gas and liquid. A two-phase separator can be horizontal, vertical or spherical. The liquid leaves the vessel at the bottom through a level control or dump valve. The gas leaves the vessel at the top passing through a mist extractor to remove the entrained liquid droplets in the gas.

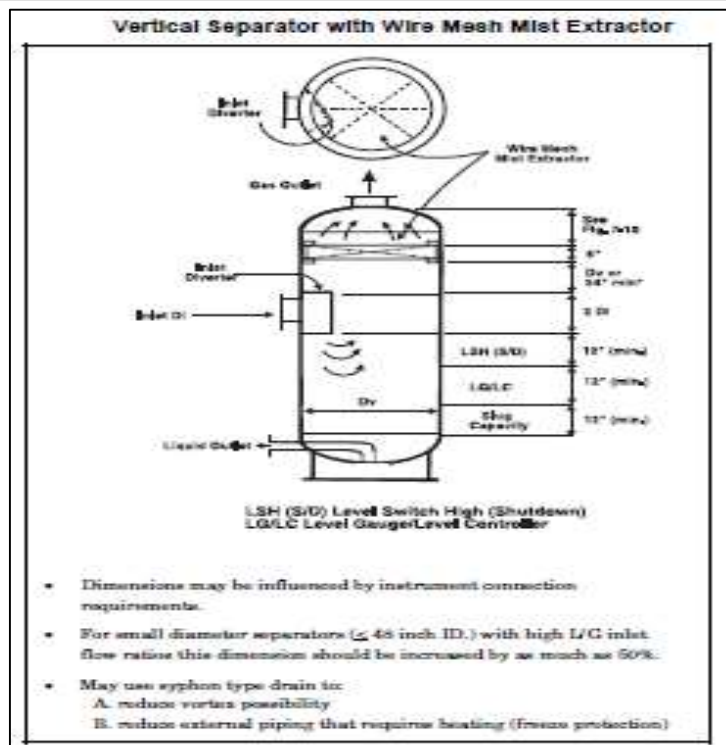
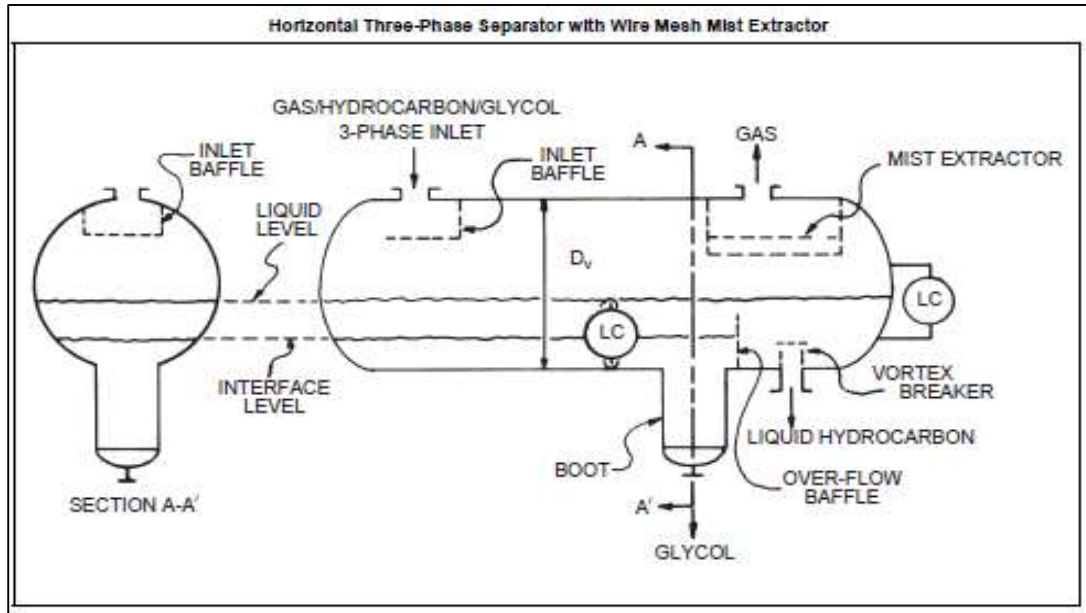
➤ Liquid-Liquid Separator

Two immiscible liquid phases can be separated using the same principles as gas-liquid separators except that they must be designed for much lower feed velocities because of the difference in density between two liquids is less and the separation become more difficult.



➤ **Gas –Liquid (Oil)-Liquid (Water) Separator (Three Phase Separator)**

A vessel that separates the well fluids into gas and two types of liquid oil & water. A three phase separator can be horizontal, vertical or spherical. This type of separator is commonly called a free water knock because its main purpose is to remove free water that can cause problems such as corrosion and formation of hydrates.



Based on the purpose oil & gas separators are further classified as below

Classification by operating configuration

Sr No.	Type of Equipment	Size of the Equipment
1	Horizontal	0.25m – 0.3048m ϕ and 1.22m – 1.5m (T to T) 4.57m - 4.88m ϕ and 18.29m – 21.34m (T to T)
2	Vertical	0.25m – 0.3048m ϕ and 1.22m – 1.5m (T to T) 0.25m – 0.3048m ϕ and 3.66m – 4.57m (T to T)
3	Sphere	7.32m - 9.14 m ϕ to 1.67- 1.83 m ϕ

Horizontal oil & gas separators are manufactured with monotube and dual tube shells. Monotube units have one cylindrical shell and dual tube units have two cylindrical parallel shells with one above the other. Both types of units can be used for two-phase and three-phase service. A monotube horizontal oil & gas separator is usually preferred over a dual-tube separator.

Note: **Mono Tube:** One cylindrical shell with greater area for gas flow as well as a greater oil/gas interface area. It will usually provide longer retention time.

Classification by Application

Oil and gas separators are further classified according to the application as below:

➤ Test Separator

A test separator is used to separate and to meter the well fluids. The test separator can be referred to as a well tester or well checker. They can be permanently installed or portable. Test separators can be equipped with various types of measuring devices for measuring the Gas oil ratio (GOR), Water cut, Basic sediments & Water (BS & W).

➤ Production Separator

Production separator is used to separate the produced well fluid from a well, group of wells, or a lease on a daily or continuous basis.

➤ Separator Basic Design Criteria

- A vessel that includes a primary separation device or section, secondary gravity settling section, mist extractor to remove small liquid particles from the gas, gas outlet, liquid settling section to remove gas from liquids, to separate water from oil and to separate solids from the liquids, to separate water from oil and to separate solids from the liquids, and oil and water outlet.
- Adequate volumetric liquid capacity to handle liquid surges from well or flow lines.
- Adequate vessel diameter and height to allow most of the liquid to separate from the gas so that mist eliminator will not be flooded.
- Controlling of oil/water levels in the separator, which usually includes a liquid-level controller and outlet control valves for oil & water.

- A backpressure valve on the gas outlet line to control the pressure in the separator.
- Pressure relief device to safeguard the separator.

Function & Separation Device Combinations

Sr.No	Function	Method of Separation	Internal Devices and Separation Aids
1	Separation of Oil from gas	Gravity	Mist extractors, impingement type, change of direction type, cyclone filter cartridges and washing.
2	Separation of gas from Oil	Gravity	Cyclone, plate packs, matrix packs, agitation and heat.
3	Separation of water from Oil	Gravity	Cyclone, plate packs, matrix packs, agitation and heat.
4	Separation of Oil from water	Gravity	Cyclone, plate packs, matrix packs, agitation and flotation.
5	Separation of Solids from Liquid	Gravity	Cyclone, agitation, heat and flotation.

Separator Orientation

Sr No	Operation	Horizontal	Vertical	Discussion
1	Separation	Better	--	In a horizontal vessel, the drops or bubbles do not have to settle or rise through a counter current flow.
2	Solids	--	Better	Solids are more easily removed from the bottom of the vertical vessel
3	Foaming	Better	--	Horizontal vessel provides more surface area for bubbles to escape
4	Surges	--	Better	A change in liquid level does not affect the gas capacity of the vessel

For detail description of types of separator refer GPSA handbook separator section 7.1 and Wikipedia article on pumps at the link:
http://en.wikipedia.org/wiki/Oil_and_gas_separator

Sizing of the feed and Outlet Nozzles

The sizing of the nozzles shall be based on the maximum flow rates, including the appropriate design margin.

Feed Inlet Nozzle

The internal nozzle diameter, d_1 , may be taken equal to that of the feed pipe, but also a momentum criterion (dependent on the inlet device, if any) shall be satisfied:

If no inlet device is used:

$$\rho_m v_m^2 \leq 2100 \text{ [Pa]}$$

Where:

ρ_m is the mean density of the mixture in the feed pipe

$$\rho_m = (M_G + M_L) / (Q_G + Q_L) \text{ [kg/m}^3\text{]}$$

M_G = Gas Mass flow rate, kg/s

M_L = Liquid Mass flow rate, kg/s

Q_G = Gas Volume flow rate, m³/s

Q_L = Liquid Volume flow rate, m³/s

$V_{m,in}$ is the velocity of the mixture in the inlet nozzle

$$V_{m,in} = (Q_G + Q_L) / (\pi d_1^2 / 4) \text{ [m/s]}$$

d_1 = inside diameter of feed nozzle, m

If a **half-open** pipe is used as inlet device:

$$\rho_m v_{m,in}^2 \leq 2100 \text{ [Pa]}$$

If a **Schoepentoeter** is used as inlet device:

$$\rho_m v_{m,in}^2 \leq 8000 \text{ [Pa]}$$

For vessels with normally no liquid, sizing the inlet nozzle for an inlet momentum of 8000 Pa at normal operating conditions is allowed. The design margin can be omitted for these cases. This minimizes the need for reducers.

When rating the capacity of existing vessels, the inlet momentum regularly exceeds the design inlet momentum. The consequence of a higher than design inlet momentum is that the droplets in the feed to the vessel typically get smaller, resulting in a lower separation efficiency of the inlet device and the separator. Shell DEP.31.20.20.31-Gen. provides guidelines for the maximum operating load over the feed nozzle for meeting the mechanical integrity requirements of the Schoepentoeter.

The pressure drop across the Schoepentoeter approximates to:

$$\Delta P_{sch} = 0.08 * \rho_m v_{m,in}^2 \text{ [Pa]}$$

In units or separators where the inlet velocity can be very high because of the low gas density, the use of a Schoepentoeter as feed inlet device is mandatory. The following velocity limits shall also be satisfied:

- To prevent erosion:

$$v_{G,in} \leq 70 \text{ [m/s]}$$

- To prevent choking or damage due to vibrations:

$$v_{G,in} \leq 0.8 v_{sonic,G}$$

Where $v_{sonic,G}$ is the sonic velocity if only gas is present.

$$v_{sonic,G} = \sqrt{\frac{\kappa R T}{MW_G}}$$

R is the gas constant (8314 J/kmol/K)

T is the absolute temperature (in Kelvin)

κ is the ratio of the specific heats (C_p/C_v)

MW_G is the mean molecular weight of the gas phase (in kg/kmol)

Gas Outlet Nozzle

The diameter of the gas outlet nozzle, d_2 , should normally be taken equal to that of the outlet pipe, but the following criterion shall be satisfied:

$$\rho_G v_{G,out}^2 \leq 4500 \text{ [Pa]}$$

In High Vacuum Units this criterion may result in a high outlet velocity, leading to a pressure drop which is too high. In that case the gas outlet nozzle shall be sized such that the pressure drop requirements between column and downstream system are met.

Liquid Outlet Nozzle

The diameter of the liquid outlet nozzle, d_3 , shall be chosen such that the liquid velocity does not exceed 1 m/s. The minimum diameter is 0.05 m (2 in). The nozzle shall be equipped with a vortex breaker.

When rating the capacity of existing vessels a nozzle liquid velocity up to 2 - 2.5 m/s is regularly experienced. This higher nozzle liquid velocity can be accepted as long as the available minimum liquid static head in the vessel, determined by the level control settings, can provide the higher liquid velocity, as estimated from the equation: $h_L = 100 V_L^2$

Where

h_L = Liquid static head above the nozzle in mm

V_L = Nozzle Liquid velocity in m/s

Separator Internals

Inlet Devices

The main function of the inlet devices are:

- Reduce the momentum of the inlet stream and enhance flow distribution of the gas and liquid phase.
- Efficient separation of the bulk liquid phase.
- Prevent droplet shattering and re-entrainment of bulk liquid phase.
- These are the commonly used separator inlet devices:
- No inlet device.
- Diverter plate.
- Half-pipe.
- Vane-type.
- Cyclonic type.
- Schoepentoeter

In addition to the inlet device, inlet piping configuration is also important for proper separation. Generally vane-type and cyclonic inlet device provided improved separation compared to others.

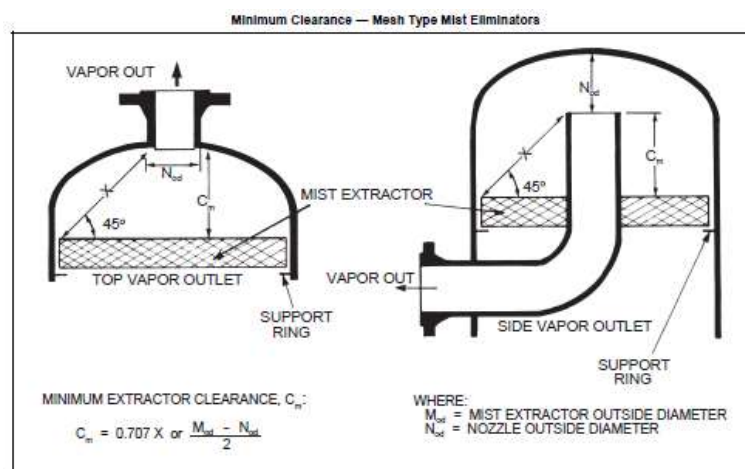
Mist Extraction Equipment

Mist extractor are used to separate the small liquid droplets from the gas phase that were not removed by the inlet device or gas settling section of the separator and the droplets are less than 150-500 microns in size. It is not economic to separate these small droplets by gravity settling.

Wire Mesh

Wire-mesh mist extractors, or pads, are made by knitting wire, metal or plastic, into tightly packed layers which are then crimped and stacked to achieve the required pad thickness. If removal of very small droplets, i.e. less than 10 microns is required, much finer fibers may be interwoven with the primary mesh to produce a co-knit pad. Performance will affect adversely if the pad is tilted more than 30 degrees from the horizontal.

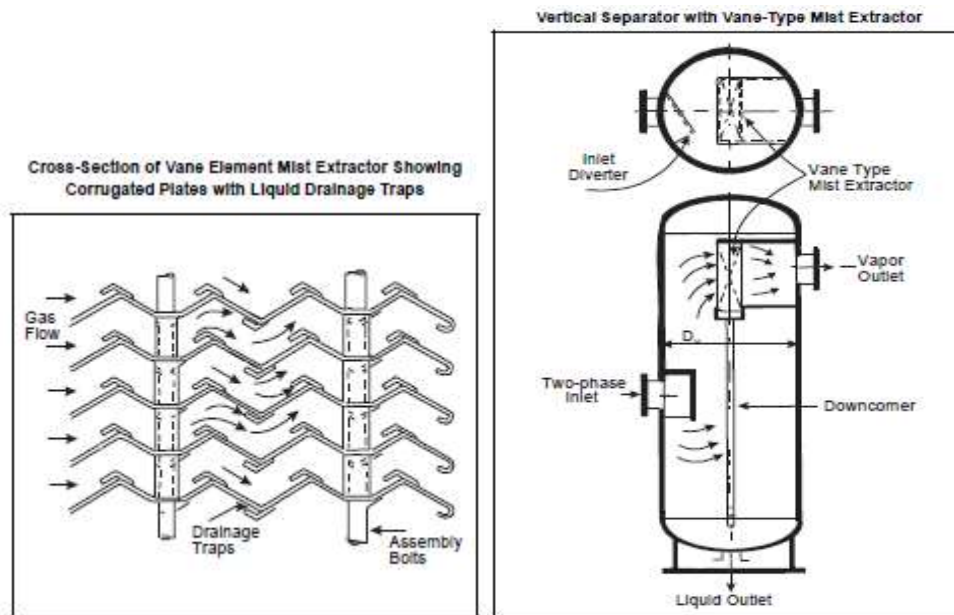
Most of the mesh pad installations will use 150mm thick pad with 144 – 192 kg/m³ bulk density. Minimum recommended pad thickness is 100mm. Manufacturer has to confirm the mesh pad thickness based on the requirement. Typical mesh pad installation is shown below.



Vane

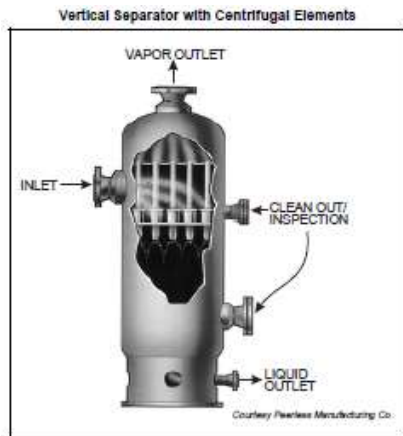
Vane or chevron-type mist extractors (vane-pack) use relatively closely spaced blades arranged to provide sinusoidal or zigzag flow paths. The changes in gas flow direction combined with inertia of the entrained liquid droplets, cause impingement of the droplets onto the plate surface, followed by coalescence and drainage of the liquid to the liquid collection section of the separator. Vane packs may be installed in either horizontal or vertical orientations, though capacity is typically reduced significantly for vertical up flow applications.

A number of different vane pack design are available. Pack thicknesses are generally in the range of 6-12 inches. Vanes are usually arranged in a zigzag or sinusoidal pattern, with vane spacing's of 1-1.5 inches typical.



Cyclonic Separator

Cyclonic separation is a method of removing particulates from an air, gas or liquid stream, without the use of filters, through vortex separation. Rotational effects and gravity are used to separate mixtures of solids and fluids. The method can also be used to separate fine droplets of liquid from a gaseous stream. A high speed rotating (air) flow is established within a cylindrical or conical container called a cyclone.



Souders-Brown Equation

The Souders–Brown equation has been the tool for obtaining the maximum allowable vapor velocity in vapor–liquid separation vessels (variously called flash drums, knockout drums, knockout pots, compressor suction drums and compressor inlet drums). It has also been used for the same purpose in designing tray fractionating columns, tray absorption columns and other vapor–liquid contacting columns. A vapor–liquid separator drum is a vertical vessel into which a liquid and vapor mixture (or a flashing liquid) is fed and wherein the liquid is separated by gravity, falls to the bottom of the vessel, and is withdrawn. The vapor travels upward at a design velocity which minimizes the entrainment of any liquid droplets in the vapor as it exits the top of the vessel.

The diameter of a vapor–liquid separator drum is dictated by the expected volumetric flow rate of vapor and liquid from the drum. The following sizing methodology is based on the assumption that those flow rates are known. Use a vertical pressure vessel with a length–diameter ratio of about 3 to 4, and size the vessel to provide about 5 minutes of liquid inventory between the normal liquid level and the bottom of the vessel (with the normal liquid level being somewhat below the feed inlet).

Calculate the maximum allowable vapor velocity in the vessel by using the Souders–Brown equation:

$$V = (k) \sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

Where:

V	=	maximum allowable vapor velocity, m/s	
ρ_L	=	liquid density, kg/m ³	
ρ_V	=	vapor density, kg/m ³	
K	=	0.107 m/s (when the drum includes a de-entraining mesh pad)	

Then the cross-sectional area of the drum (A) is obtained from:

$$\text{Area } A = \text{Vapor volume flow rate} / \text{Vapor velocity}$$

And the drum diameter (D) is:

$$\text{Diameter } D = (1.273 * A)^{1/2}$$

The drum should have a vapor outlet at the top, liquid outlet at the bottom, and feed inlet at about the half-full level. At the vapor outlet, provide a de-entraining mesh pad within the drum such that the vapor must pass through that mesh before it can leave the drum. Depending upon how much liquid flow is expected, the liquid outlet line should probably have a liquid level control valve.

As for the mechanical design of the drum (materials of construction, wall thickness, corrosion allowance, etc.), use the same criteria as for any pressure vessel.

Recommended values of k

The GPSA Engineering Data Book recommends the following k values for vertical drums with horizontal mesh pads (at the denoted operating pressures):

- At a gauge pressure of 0 bar: 0.107 m/s
- At a gauge pressure of 7 bar: 0.107 m/s
- At a gauge pressure of 21 bar: 0.101 m/s
- At a gauge pressure of 42 bar: 0.092 m/s
- At a gauge pressure of 63 bar: 0.083 m/s
- At a gauge pressure of 105 bar: 0.065 m/s

GPSA Notes:

- $k = 0.107$ at a gauge pressure of 7 bar. Subtract 0.003 for every 7 bar above a gauge pressure of 7 bar.
- For glycol or amine solutions, multiply above k values by 0.6 – 0.8
- Typically use one-half of the above k values for approximate sizing of vertical separators without mesh pads
- For compressor suction scrubbers and expander inlet separators, multiply k by 0.7 – 0.8

4.13 HEAT EXCHANGERS

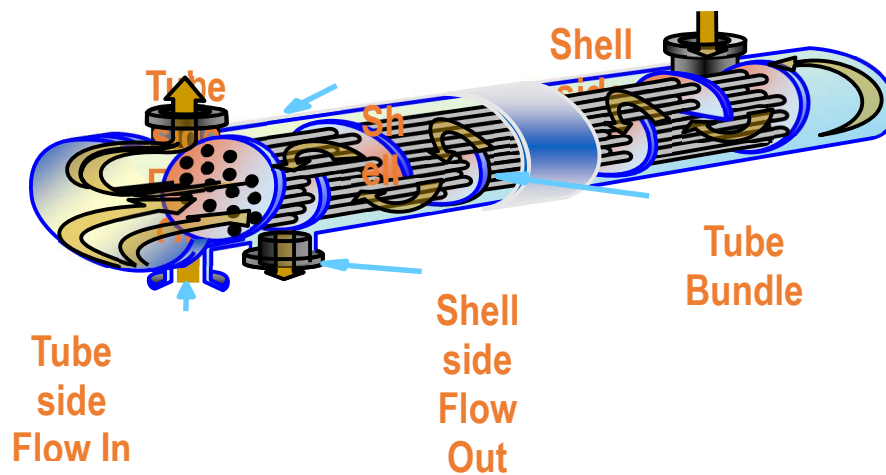
Introduction

The shell-and-tube heat exchanger is by far the most common type of heat exchanger used in industry. It can be fabricated from a wide range of materials both metallic and non-metallic. Design pressures range from full vacuum to 6,000 psig. Design temperatures range from -250°C to 800°C.

The shell-and-tube design is more rugged than other types of heat exchangers. It can stand more (physical and process) abuse.

Shell & Tube Heat Exchangers can be used as condensers, reboilers, process heaters and coolers.

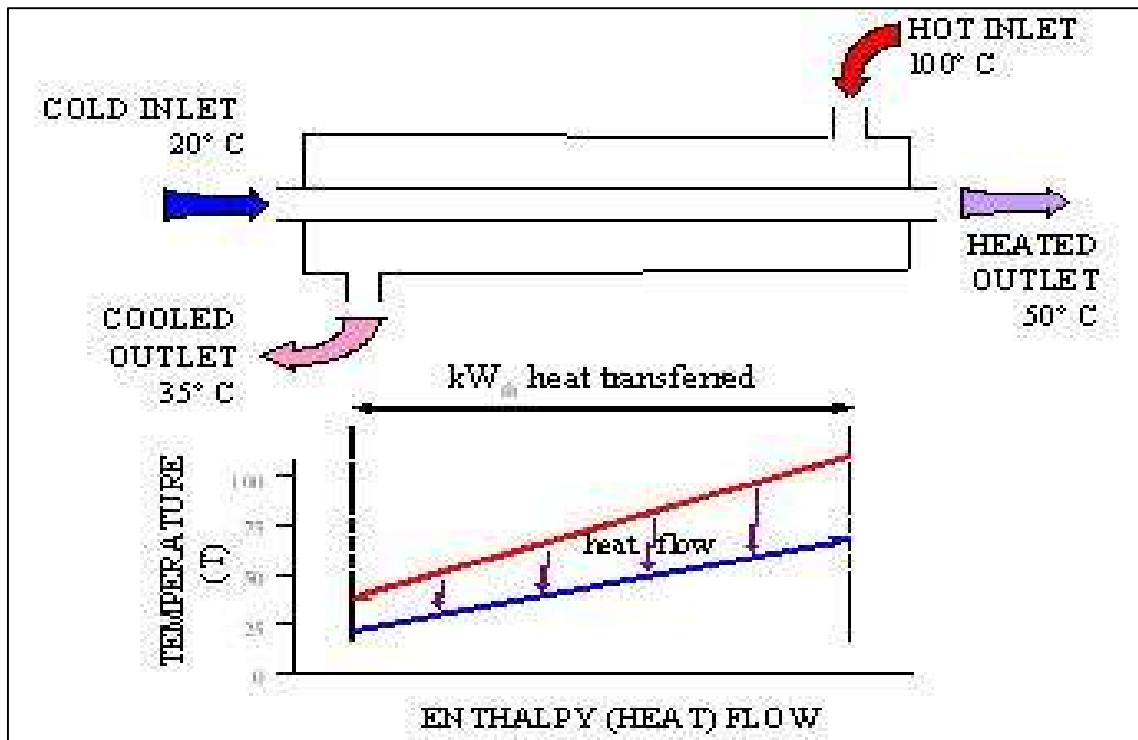
Low Pattern in a Simple Shell & Tube Heat Exchanger



Working Principle / Governing Equation

Simplified heat exchanger concepts:

Heat exchangers work because heat naturally flows from higher temperature to lower temperatures. Therefore if a hot fluid and a cold fluid are separated by a heat conducting surface, heat can be transferred from the hot fluid to the cold fluid.



The rate of heat flow at any point (kW/m^2 of transfer surface) depends on:

- Heat transfer coefficient (u), itself a function of the properties of the fluids involved, fluid velocity, materials of construction, geometry and cleanliness of the exchanger
- Temperature difference between hot and cold streams

Total heat transferred (q) depends on:

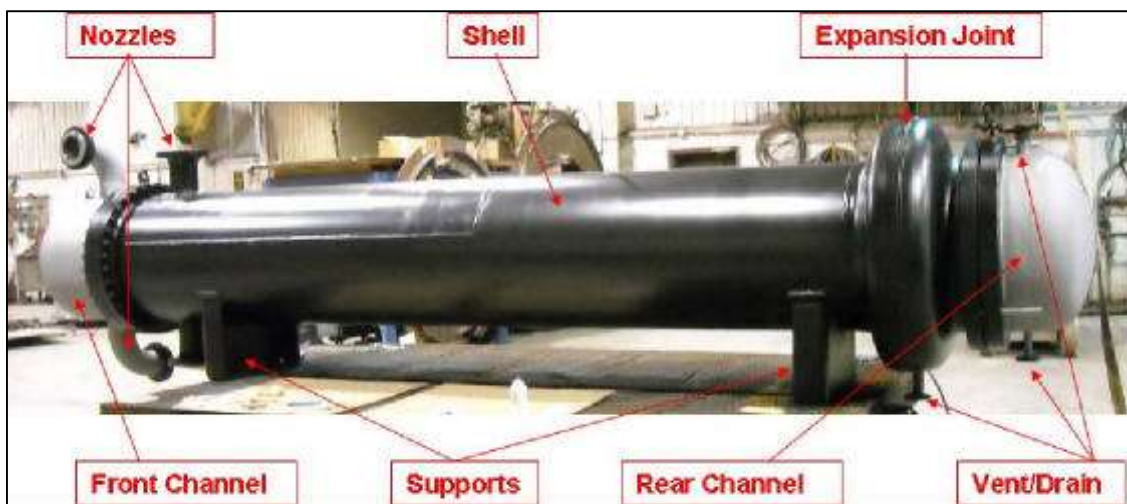
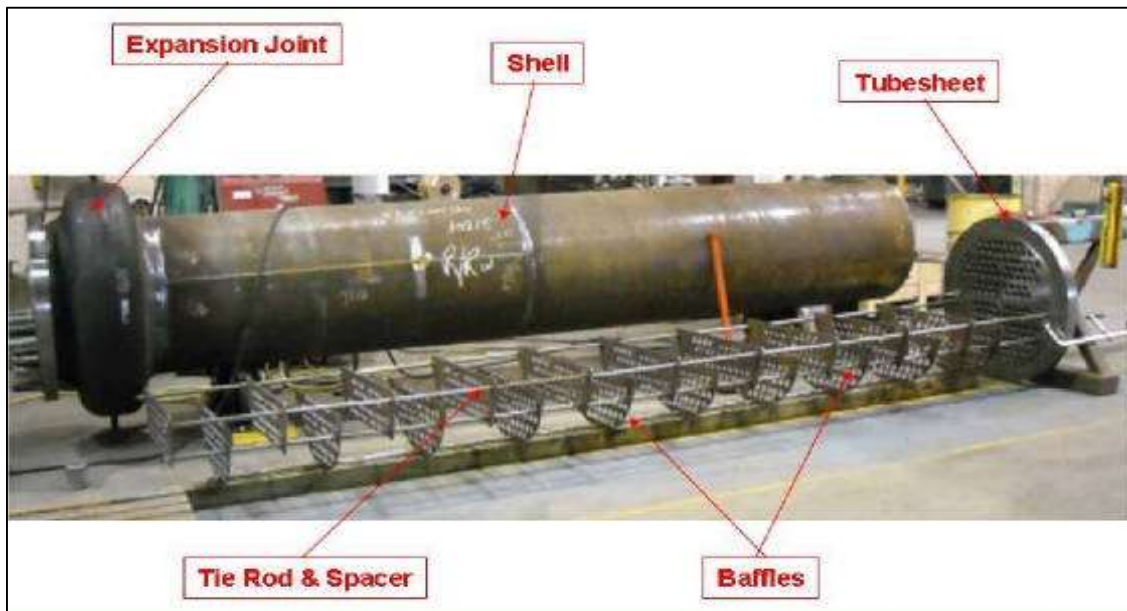
- heat transfer surface area (a)
- heat transfer coefficient (u)
- average temperature difference between the streams, strictly the log mean (Δt_{lm})

Thus total heat transferred:

But the larger the area the greater the cost of the exchanger

Therefore there is a trade-off between the amount of heat transferred and the exchanger cost

Heat Exchanger: Pictorial View

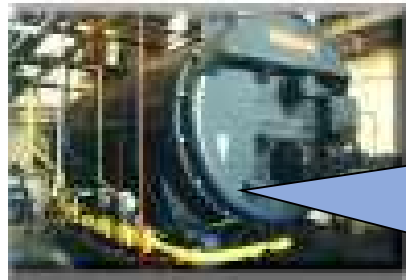


Other Types of Heat Exchangers

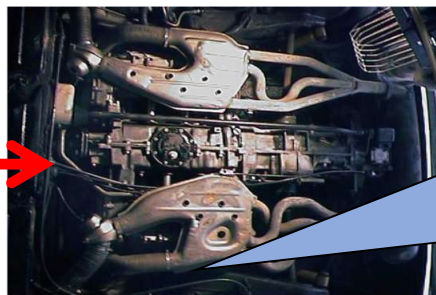
- Gasketed / welded plate and frame heat exchanger
- Spiral heat exchanger
- Tubular heat exchanger
- Plate coils heat exchanger
- Scraped surface heat exchanger



Applications of Heat Exchangers



Heat exchangers are used in Industry for heat transfer



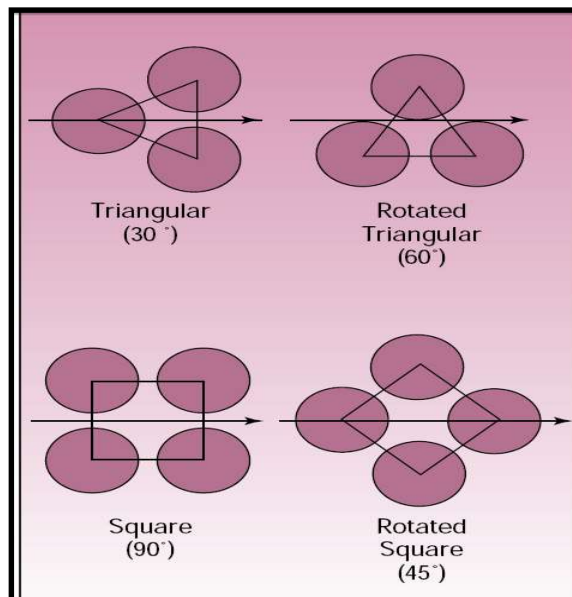
Heat Exchangers prevent car engine overheating



Heat exchangers are used in AC and furnaces

Different Parts of a Typical Heat Exchanger

- Shell – The Shell is simply a container for the shell side fluid. The shell cylinder can be fabricated from rolled plate or from piping (up to 24 inch diameters).
- Tubes / Tube Bundle -
 - ✓ Tubing may be seamless or welded.
 - ✓ Normal tube diameters are 5/8 inch, 3/4 inch and 1 inch.
 - ✓ The normal tube wall thickness ranges from 12 to 16 BWG (from 0.109 inches to 0.065 inches thick).
 - ✓ Tubing may be finned to provide more heat transfer surface.
- Tubesheets -
 - ✓ Tubesheets are plates or forgings drilled to provide holes through which tubes are inserted.
 - ✓ Tubes are appropriately secured to the tube sheet so that the fluid on the shell side is prevented from mixing with the fluid on the tube side.
 - ✓ The distance between the centers of the tube hole is called the tube pitch; normally the tube pitch is 1.25 times the outside diameter of the tubes.



Square Pitch – When the shell side fluid is fouling and mechanical cleaning is required.

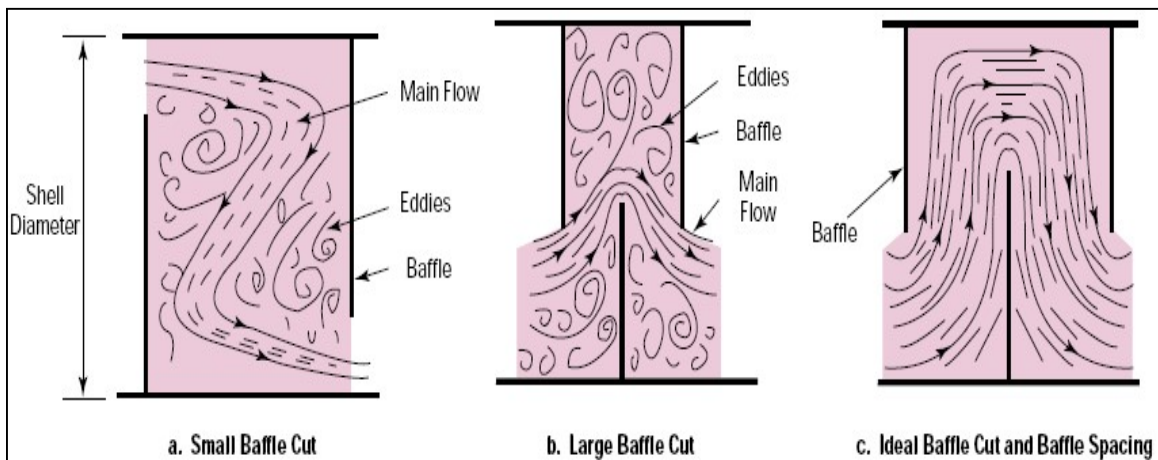
Triangular Pitch – When the shell side fluid is clean.

Rotated Triangular Pitch – Seldom used. Offers no advantage over triangular pitch.

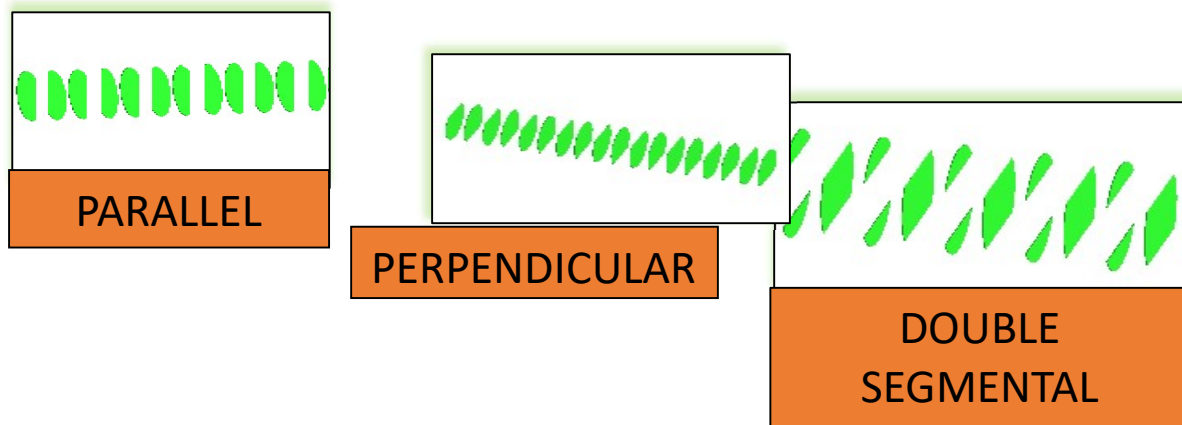
Rotated Square Pitch – When higher turbulence is required i.e. when Reynolds Number is low (< 2000)

- Baffles –
 - ✓ Baffles serve three functions:
 - ✓ Support the tube.
 - ✓ Maintain the tube spacing.
 - ✓ Direct the flow of fluid in the desired pattern through the shell side.
 - ✓ A segment, called the baffle cut, is cut away to permit the fluid to flow parallel to the tube axis as it flows from one baffle space to another.
 - ✓ The spacing between segmental baffles is called the baffle pitch.
 - ✓ When the shell side heat transfer is sensible heating or cooling with no phase change, the baffle cut should be horizontal. This causes the fluid to follow an up-and-down path and prevents stratification with warmer fluid at the top of the shell and cooler fluid at the bottom of the shell.
 - ✓ For shell side condensation, the baffle cut is vertical to allow the condensate to flow towards the outlet without significant liquid holdup by the baffle. For shell side boiling, the baffle cut may be either vertical or horizontal.

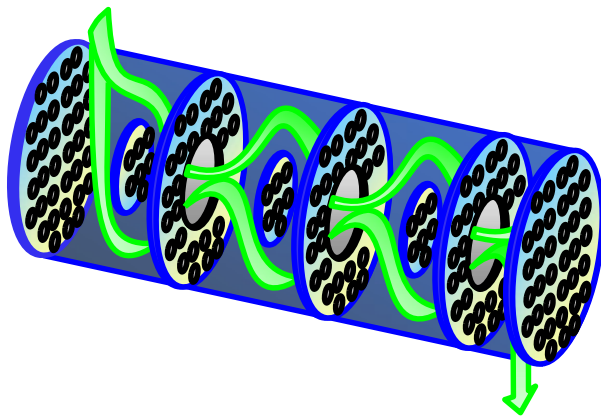
Effects of Baffle Cut



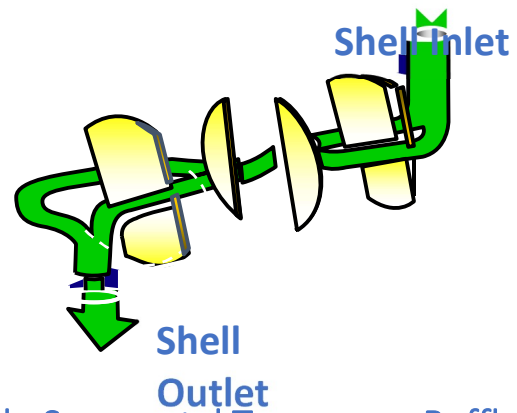
Types of Baffle Arrangement



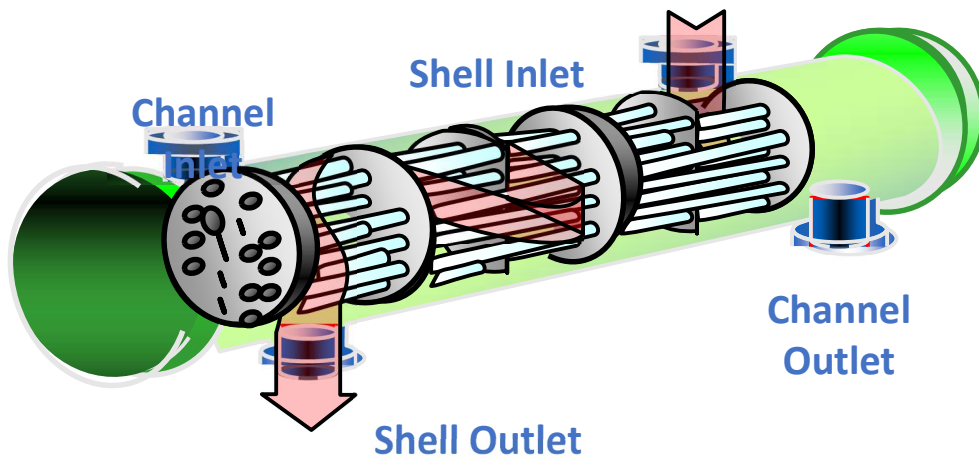
Flow Pattern in Different Types of Baffles



Doughnut and Disc
Type Baffles

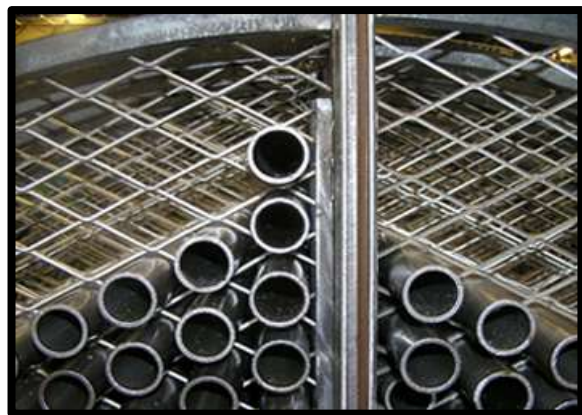
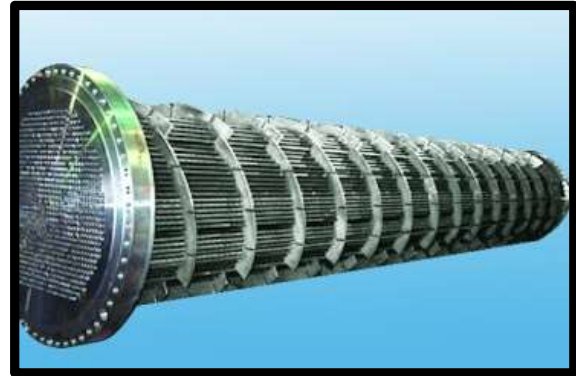


Double Segmental Transverse Baffles

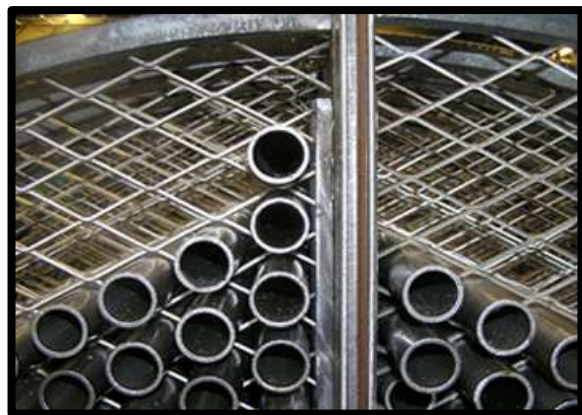
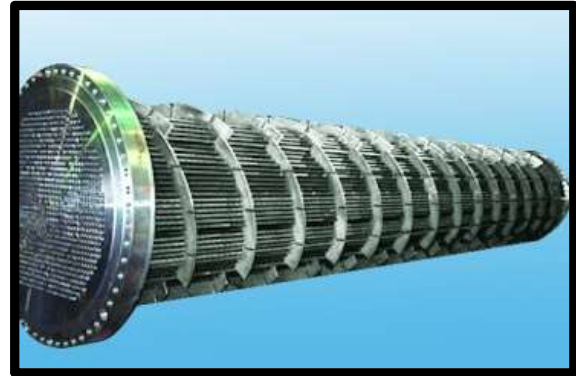


Single Segmental Transverse Baffles

Types of Baffles



Types of Baffles



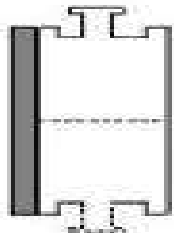
- Tie Rods & Spacers - Tie rods and spacers are used for two reasons:
 - ✓ Hold the baffle assembly together
 - ✓ Maintain the selected baffle spacing.
- Channels (Heads)
 - ✓ Channels or heads are required for shell-and-tube heat exchangers to contain the tube side fluid
 - ✓ To provide the desired flow path.
 - ✓ The three (3) letters TEMA (Tubular Exchanger Manufacturers Association) designation is the standard method for identifying the type of channels and the type of shell of shell-and-tube heat exchangers.
 - ✓ The channel type is selected based on the application.
 - ✓ The most commonly used channel type is the bonnet. It is used for services which do not require frequent removal of the channel for inspection or cleaning.
 - ✓ Removable cover channels are provided when frequent cleaning is required.

TEMA Class for Channels

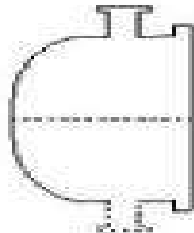
TEMA CLASS FOR CHANNELS	APPLICATION
R	Severe requirements of petroleum and related process applications
C	Moderate requirements of commercial and general process applications
B	Chemical process service

TEMA Front & Rear Heads

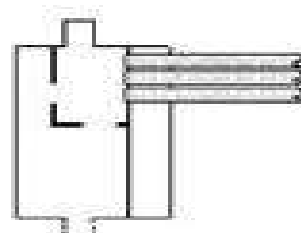
TEMA Front Heads



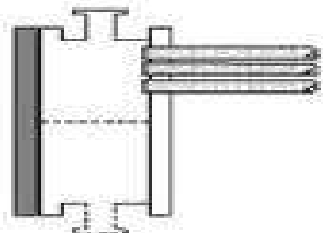
A
**Channel and
Removable Cover**



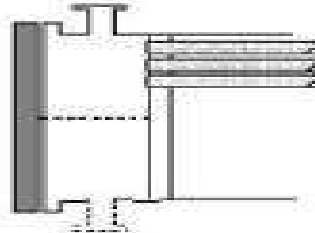
B
**Bonnet
Integral Cover**



D
**Special High
Pressure Closure**

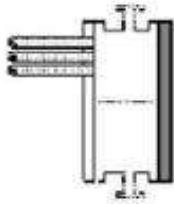


C

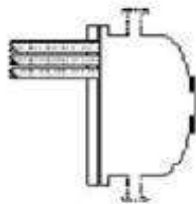


N

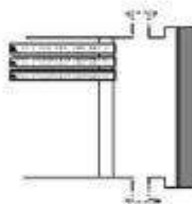
TEMA Rear Heads



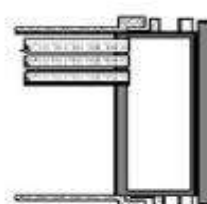
L
Fixed Tubesheet
Like "A" Stationary Head



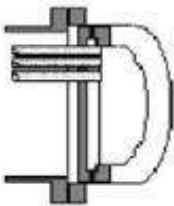
M
Fixed Tubesheet
Like "B" Stationary Head



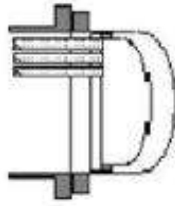
N
Fixed Tubesheet
Like "N" Stationary Head



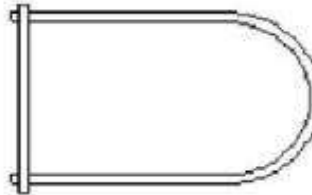
P
Outside Packed
Floating Head



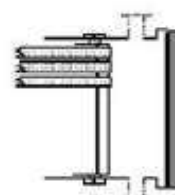
S
Floating Head with
Packing Device



T
Pull Through
Floating Head



U
U Tube Bundle



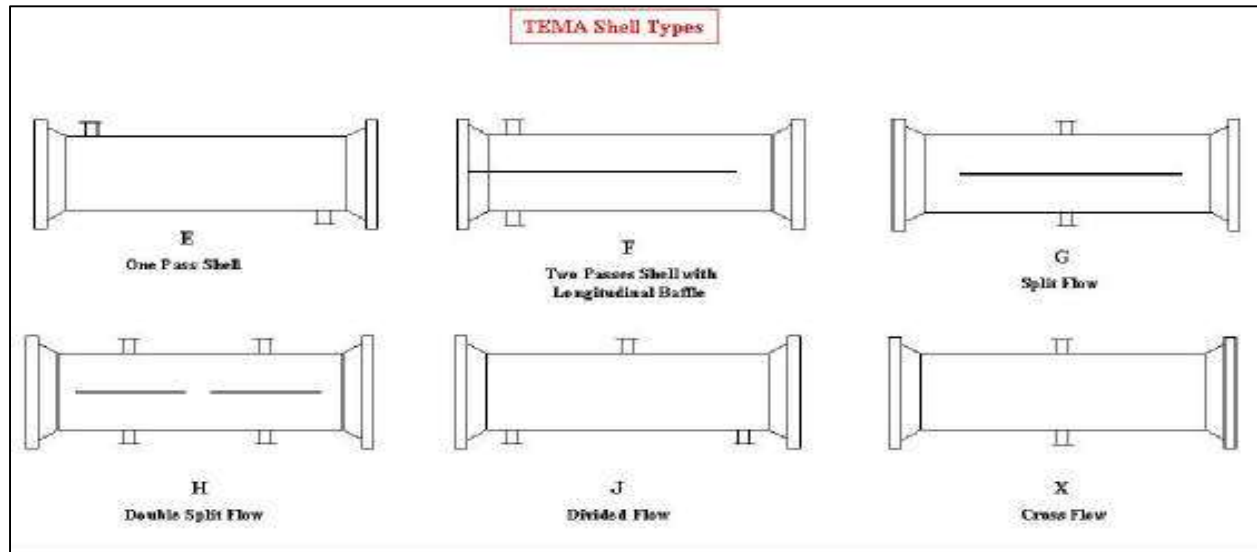
W
Externally Sealed
Floating Head

Selection of Heads

TEMA Front Head Selection	
A	Easy to open for tube side access. For low pressure applications.
B	For Higher-pressure applications, preferred with clean tube side fluid. It is less expensive than Type A.
C	Tube side is corrosive, toxic or hazardous and when removable tube bundle is required. It is normally used for low-pressure operations.
N	For application where tube side is corrosive, toxic or hazardous and shell side fluid is clean and any leakage possibility is to be eliminated.
D	Very High-pressure applications.
TEMA Rear Head Selection	
L,M,N	Fixed Head, Should be used when thermal differential expansion of the shell and tubes is low and shell side fluid is clean.
L	For low pressure applications.
M	For Higher-pressure applications, not requiring frequent maintenance.
N	For application where tube side is Corrosive, toxic or hazardous and where leakage of shell to tube side fluid and vice versa, is to be eliminated.
U	For thermal differential expansion of the shell and tubes is higher and tube side fluid is clean. For high-pressure applications or, with hazardous/ toxic fluid on shell side.
P,S,T,W	Should be used when shell side fluid or both shell and tube side fluid are Dirty.
P	Pressure is low and shell side fluid is not toxic or hazardous. Where risk of internal flange leakage is to be avoided.
S	Normal Pressure requirements, relatively lesser maintenance requirements.
T	High-pressure requirements, frequent need to takeout the tube bundle.
W	For low-pressure application.

TEMA Shell Types

SELECTION OF SHELL



Common TEMA type of Shell & Tube Heat Exchangers:

- ✓ AES
- ✓ BEU
- ✓ BHU
- ✓ BXU
- ✓ BEM
- ✓ AKU
- ✓ AET
- ✓ AEL

Selection of Shell

TEMA Shell Selection	
E	The E shell is the most common as it is inexpensive and simple.
F	F shell is rarely used in practice because there are many problems associated with the design. It is difficult to remove/replace the tube bundle, problems of fabrication and maintenance, internal leakage, unbalanced thermal expansion in case of large temperature difference between inlet & outlet.
G & H	To accommodate high inlet velocities. They are used as horizontal thermosiphon reboilers, condensers, and other phase-change applications.
J	Used for low pressure drop applications such as a condenser in vacuum.
K	The K shell is used for partially vaporizing the shell fluid. It is used as a kettle reboiler in the process industry and as a flooded chiller in the refrigeration industry. They are used when essentially 100% vaporization is required.
X	It is used for gas heating and cooling and for vacuum condensation. It is also used when shell flows are large.

Allocation of Streams

- **High Pressure**

If one of the stream is at a high pressure, it is desirable to put that stream inside the tubes. In this case, only the tubes and tube side fittings need to be designed to withstand high pressure, whereas the shell may be made of lighter weight metal.

- **Corrosion**

Corrosion generally dictates the choice of material of construction, rather than exchanger design. However, since most corrosion resistant alloys are more expensive than the ordinary materials of construction, the corrosive fluid will normally be placed inside the tubes to reduce the cost by avoiding use of corrosion resistant alloys for the shell side.

- **Viscosity**

Highly viscous fluid is placed at shell side so that turbulence can be induced by introducing baffles in shell side.

- **Fouling**
Fouling enters into the design of almost every process exchanger to a measurable extent, but certain streams foul so badly that the entire design is dominated by features which seek a) to minimize fouling or b) to facilitate cleaning.
- **Low Heat Transfer Coefficient**
If one stream has an inherently low heat transfer coefficient (such as low pressure gases or viscous liquids), this stream is preferentially put on the shell side.
- **Flow Rate**
Generally smaller flow rate fluid is placed on the shell side. This facilitates provision of adequate turbulence by increasing number of baffles.
- **Phase Of Fluid**
Two phase fluid should be placed in **shell side**.

Major Practical Problems Encountered In Heat Exchangers

Fouling

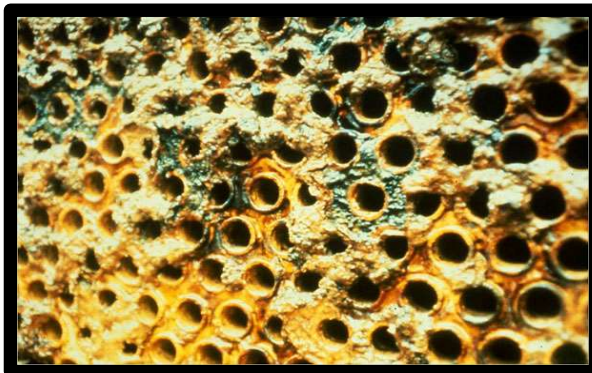
Fouling is generally defined as the accumulation of unwanted materials on the surfaces of processing equipment. It has been recognized as a nearly universal problem in design and operation and affects the operation of equipment in two ways:

- The fouling layer has a low thermal conductivity. This increases the resistance to heat transfer and reduces the effectiveness of heat exchangers – increasing temperature.
- As deposition occurs, the cross-sectional area is reduced, which causes an increase in pressure drop across the exchanger.

Fouling can be caused by:

- Frequent use of the heat exchanger
- Not cleaning the heat exchanger regularly
- Reducing the velocity of the fluids moving through the heat exchanger
- Over-sizing of the heat exchanger

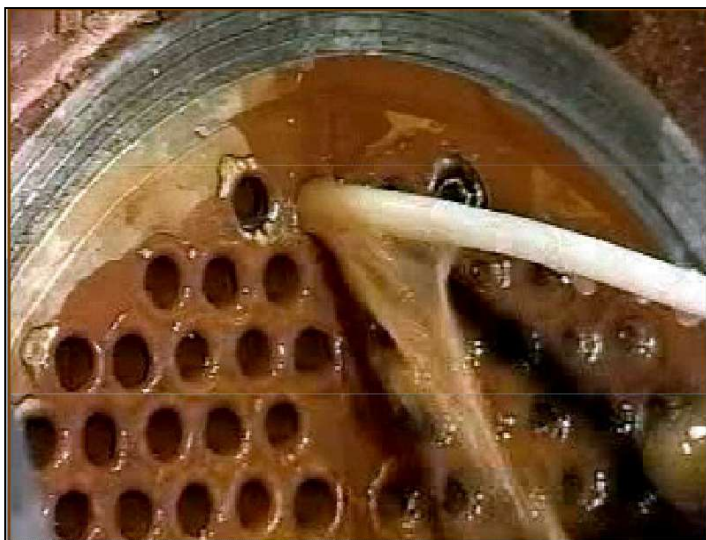
Fouled Heat Exchanger Tubes

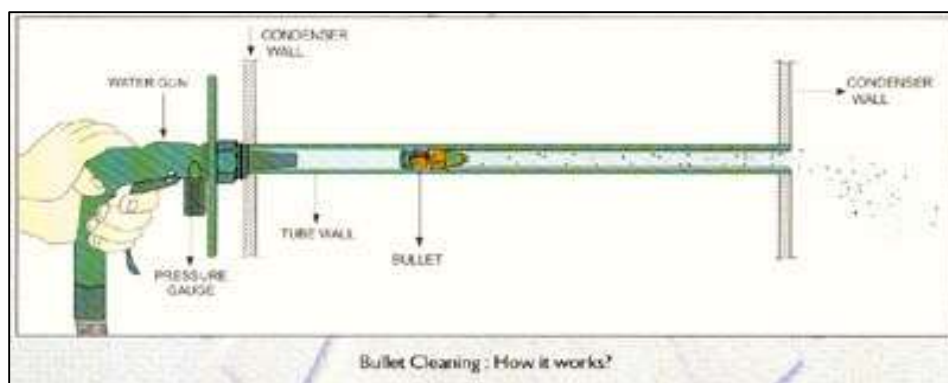


Fouling Treatment

Tubular heat exchangers can be cleaned by such methods as chemical cleaning, sandblasting, high pressure water jet, bullet cleaning, or drill rods.

In large-scale cooling water systems for heat exchangers, water treatment such as purification, addition of chemicals, and testing is used to minimize fouling of the heat exchanger.





Fouling Resistances for Industrial Fluids

Oils:		
Fuel Oil #2		0.002
Fuel Oil #6		0.005
Transformer Oil		0.001
Engine Lube Oil		0.001
Quench Oil		0.004
Gases And Vapors:		
Manufactured Gas		0.010
Engine Exhaust Gas		0.010
Steam (Non-Oil Bearing)		0.0005
Exhaust Steam (Oil Bearing)		0.0015-0.002
Refrigerant Vapors (Oil Bearing)		0.002
Compressed Air		0.001
Ammonia Vapor		0.001
CO ₂ Vapor		0.001
Chlorine Vapor		0.002
Coal Flue Gas		0.010
Natural Gas Flue Gas		0.005
Liquids:		
Molten Heat Transfer Salts		0.0005
Refrigerant Liquids		0.001
Hydraulic Fluid		0.001
Industrial Organic Heat Transfer Media		0.002
Ammonia Liquid		0.001
Ammonia Liquid (Oil Bearing)		0.003
Calcium Chloride Solutions		0.003
Sodium Chloride Solutions		0.003
CO ₂ Liquid		0.001
Chlorine Liquid		0.002
Methanol Solutions		0.002
Ethanol Solutions		0.002
Ethylene Glycol Solutions		0.002

Fouling Resistances for Chemical Processing Streams

Gases And Vapors:	
Acid Gases	0.002-0.003
Solvent Vapors	0.001
Stable Overhead Products	0.001
Liquids:	
MEA And DEA Solutions	0.002
DEG And TEG Solutions	0.002
Stable Side Draw And Bottom Product	0.001-0.002
Caustic Solutions	0.002
Vegetable Oils	0.003

Fouling Resistances For Natural Gas-Gasoline Processing Streams	
Gases And Vapors:	
Natural Gas	0.001-0.002
Overhead Products	0.001-0.002
Liquids:	
Lean Oil	0.002
Rich Oil	0.001-0.002
Natural Gasoline And Liquified Petroleum Gases	0.001-0.002

Fouling Resistances for Oil Refinery Streams

Atmospheric Tower Overhead Vapors							0.001
Light Naphthas							0.001
Vacuum Overhead Vapors							0.002
Crude And Vacuum Liquids:							
Crude Oil							
0 to 250 ° F VELOCITY FT/SEC				250 to 350 ° F VELOCITY FT/SEC			
	<2	2-4	>4		<2	2-4	>4
DRY	0.003	0.002	0.002	0.003	0.002	0.002	0.002
SALT*	0.003	0.002	0.002	0.005	0.004	0.004	0.004
350 to 450 ° F VELOCITY FT/SEC				450 ° F and over VELOCITY FT/SEC			
	<2	2-4	>4		<2	2-4	>4
DRY	0.004	0.003	0.003	0.005	0.004	0.004	0.004
SALT*	0.006	0.005	0.005	0.007	0.006	0.006	0.006
*Assumes desalting @ approx. 250 ° F							
Gasoline							0.002
Naphtha And Light Distillates							0.002-0.003
Kerosene							0.002-0.003
Light Gas Oil							0.002-0.003
Heavy Gas Oil							0.003-0.005
Heavy Fuel Oils							0.005-0.007
Asphalt And Residuum:							
Vacuum Tower Bottoms							0.010
Atmosphere Tower Bottoms							0.007
Cracking And Coking Unit Streams:							
Overhead Vapors							0.002
Light Cycle Oil							0.002-0.003
Heavy Cycle Oil							0.003-0.004
Light Coker Gas Oil							0.003-0.004
Heavy Coker Gas Oil							0.004-0.005
Bottoms Slurry Oil (4.5 Ft/Sec Minimum)							0.003
Light Liquid Products							0.002

Catalytic Reforming, Hydrocracking And Hydrodesulfurization Streams:	
Reformer Charge	0.0015
Reformer Effluent	0.0015
Hydrocracker Charge And Effluent*	0.002
Recycle Gas	0.001
Hydrodesulfurization Charge And Effluent*	0.002
Overhead Vapors	0.001
Liquid Product Over 50 ° A.P.I.	0.001
Liquid Product 30 - 50 ° A.P.I.	0.002
*Depending on charge, characteristics and storage history, charge resistance may be many times this value.	
Light Ends Processing Streams:	
Overhead Vapors And Gases	0.001
Liquid Products	0.001
Absorption Oils	0.002-0.003
Alkylation Trace Acid Streams	0.002
Reboiler Streams	0.002-0.003
Lube Oil Processing Streams:	
Feed Stock	0.002
Solvent Feed Mix	0.002
Solvent	0.001
Extract*	0.003
Raffinate	0.001
Asphalt	0.005
Wax Slurries*	0.003
Refined Lube Oil	0.001
*Precautions must be taken to prevent wax deposition on cold tube walls.	
Visbreaker:	
Overhead Vapor	0.003
Visbreaker Bottoms	0.010
Naphtha Hydrotreater:	
Feed	0.003
Effluent	0.002
Naphthas	0.002
Overhead Vapors	0.0015

Fouling Resistance for Water

Temperature Of Heating Medium	Up To 240° F		240 to 400° F	
Temperature Of Water	125 ° F		Over 125° F	
	Water Velocity Ft/Sec		Water Velocity Ft/Sec	
	3 and Less	Over 3	3 and Less	Over 3
Sea Water	0.0005	0.0005	0.001	0.001
Brackish Water	0.002	0.001	0.003	0.002
Cooling Tower And Artificial Spray Pond:				
Treated Make Up	0.001	0.001	0.002	0.002
Untreated	0.003	0.003	0.005	0.004
City Or Well Water	0.001	0.001	0.002	0.002
River Water:				
Minimum	0.002	0.001	0.003	0.002
Average	0.003	0.002	0.004	0.003
Muddy Or Silty	0.003	0.002	0.004	0.003
Hard (Over 15 Grains/Gal.)	0.003	0.003	0.005	0.005
Engine Jacket	0.001	0.001	0.001	0.001
Distilled Or Closed Cycle				
Condensate	0.0005	0.0005	0.0005	0.0005
Treated Boiler Feedwater	0.001	0.0005	0.001	0.001
Boiler Blowdown	0.002	0.002	0.002	0.002

Vibrations

Fluid flowing through a heat exchanger can cause the heat exchanger tubes to vibrate. Different types of vibration mechanisms are as follows:

✓ Fluid Elastic Instability:

Fluid elastic instability is important for both gases & liquids. This occurs above a critical flow velocity. There are different methods to avoid fluid elastic instability such as:

- Decreasing the span lengths
- Increasing the tube diameter
- Reducing clearance between tube & baffle
- Increasing tube pitch also helps in minimizing tube vibrations
- Tubes in the window region can be removed so that all tubes are supported.

✓ Vortex Shedding:

Vortex shedding is caused by the periodic shedding of the vortices from the tubes and can lead to damage of tubes if vibrations coincide with the tube natural frequency. Some measures such as changing span lengths can be taken to avoid vibrations.

✓ Acoustic Resonance:

Acoustic resonance is very important in case of gases. It occurs when the frequency of an acoustic wave in the heat exchanger coincides with tube natural frequency. Even if acoustic wave does not cause any vibrations, it can lead to intolerable noise. It can be avoided by changing span lengths. Generally deresonating baffles are placed in all cross passes of the heat exchanger parallel to both the direction of cross flow and the centerline of the tubes, which increases the acoustic frequency of the acoustic wave thus, eliminating acoustic vibration problem.

✓ Turbulent Buffeting:

Turbulent buffeting mechanism is very important in case of two-phase flow. The turbulence in the flowing fluid contains a broad range of frequencies and can coincide with the tube natural frequency to cause tube vibrations.

✓ Flow pulsation:

Flow pulsation is because of periodic variations in the flow. This can become very important in case of two-phase flow.

Table for Selection

Shell and Tube Exchanger Selection Guide (Cost Increases from Left to Right)

Type of Design	"U" Tube	Fixed Tubesheet	Floating Head Outside Packed	Floating Head Split Backing Ring	Floating Head Pull-Through Bundle
Provision for differential expansion	individual tubes free to expand	expansion joint in shell	floating head	floating head	floating head
Removeable bundle	yes	no	yes	yes	yes
Replacement bundle possible	yes	not practical	yes	yes	yes
Individual tubes replaceable	only those in outside row	yes	yes	yes	yes
Tube interiors cleanable	difficult to do mechanically, can do chemically	yes, mechanically or chemically	yes, mechanically or chemically	yes, mechanically or chemically	yes, mechanically or chemically
Tube exteriors with triangular pitch cleanable	chemically only	chemically only	chemically only	chemically only	chemically only
Tube exteriors with square pitch cleanable	yes, mechanically or chemically	chemically only	yes, mechanically or chemically	yes, mechanically or chemically	yes, mechanically or chemically
Number of tube passes	any practical even number possible	normally no limitations	normally no limitations	normally no limitations	normally no limitations
Internal gaskets eliminated	yes	yes	yes	no	no

Thermal Designing & Rating of Heat Exchangers

Designing & Rating of Heat Exchangers is carried out by using following software:

- HTRI Exchanger Suite: HTRI is developed by Heat Transfer Research Inc.
- EDR: Exchanger Design Rating from Aspen

Inputs Required

Before proceeding in either designing or rating of Heat Exchangers, there are some essential inputs required from the client. These are:

Complete and latest Process datasheet. Also, old Process datasheet in case of rating of Heat Exchanger.

Essential data includes:

- Total Heat Duty
- Flow rates, temperatures, pressures for both sides
- Fluid properties including: density, specific heat, thermal conductivity, and viscosity for at least two points.
- For condensers and evaporators, data such as a condensing curve, boiling point elevation, and/or other parameters may be required.
- Process conditions and limitations such as fouling resistance, pressure drop limitations, MOC etc.

Specifications and other special requirements/limitations.

Exchanger drawings & old TEMA datasheet in case of rating of heat exchanger.

Outputs Delivered

When an Exchanger is designed following are the deliverables to the client:

- TEMA Datasheet
- Rating Report stating the adequacy of the Heat Exchanger

HTRI OUTPUT / RESULTS INTERPRETATION

- Output of HTRI is interpreted in various ways depending upon the type of problem i.e. type of service, type of exchanger, any special guideline/limitation from the client, etc. Normally, an output is studied for the following parameters :

- Heat duty
- Allowable pressure drop
- B & F flow fractions
- Vibrations
- Shell side, Tube side, Cross flow, Wind flow velocities
- Overdesign percentage
- Runtime messages (if any)
- Rho-V2 values

Heat Exchanger design:

1. Initially calculate heat load

$$Q = M_s C_{p_s} (T_2 - T_1) = M_t C_{p_t} (t_2 - t_1) = U_D A_D LMTD$$

Where M_s = Shell side mass flowrate kg/hr

C_{p_s} : Specific heat of fluid on shell side kcal/kgC

T_1, T_2 : Shell side inlet/outlet temperature.

M_t : Tube side mass flowrate kg/hr

C_{p_t} : Specific heat of fluid on tube side kcal/kgC

t_1, t_2 : Tube side inlet/outlet temperature.

Q : Heat load Kcal/hr

U_D : Overall Design heat transfer coefficient Kcal/m²hrC

A_D : Design surface area M²

LMTD: Log mean temperature difference C

2. Based on the co-current/counter-current flow, calculate LMTD

3. Decide Correction factor F_t for LMTD based on no of passes

R : Shell side temperature difference/Tube side temperature difference

S : Tube side temperature difference/ Shell inlet temperature- tube inlet temperature

Get the correction factor from R, S values for the selected no of passes

LMTD corrected: LMTD true $\times$ Correction factor

Log Mean Temperature Difference

The LMTD is the temperature difference at one end of the heat exchanger less the temperature.

The hot and cold fluids are assumed to be flowing countercurrent to each other.

$$LMTD = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \left[\frac{(T_1 - t_2)}{(T_2 - t_1)} \right]}$$

CLMTD = $F \times$ LMTD

4. Assume U_D

5. Calculated the surface area $A_D = \frac{Q}{U_D} \times LMTD_{corrected}$

6. Select the tube configuration viz Tube dia, thk, length, and layout

Calculate No of tubes, no of passes, tubes/ pass

7. Next determine Inside & Outside heat transfer coefficients

Inside heat transfer coefficients h_{io} : $h_i \times d_i / d_o$

Outside heat transfer coefficient h_o

8. Then calculate Tube metal wall temperature t_w

When the hot fluid is inside the shell/ cold fluid in tube

$$t_w = t_c + (h_o / (h_{io} + h_o)) (T_c - t_c)$$

$$t_w = T_c - (h_{io} / (h_{io} + h_o)) (T_c - t_c)$$

When the hot fluid is inside the tube/ cold fluid in shell

$$t_w = t_c + (h_o / (h_{io} + h_o)) (T_c - t_c)$$

$$t_w = T_c - (h_{io} / (h_{io} + h_o)) (T_c - t_c)$$

Where T_c , T_c Is Calorific or Average Fluid Temperature on Shell/ Tube Side

$$T_c = (T_1 + T_2) / 2, t_c = (t_1 + t_2) / 2$$

9. Get the viscosity at tube metal wall temperature, μ_w

Calculate viscosity correction factor $(\mu / \mu_w)^{0.14}$

10. Calculate the corrected inside and outside heat transfer coefficient

11. Calculate clean overall heat transfer coefficient U_c

$$U_c = h_{io} \times h_o / (h_{io} + h_o)$$

12. Then check the dirt factor R_d

$$h_d = U_c U_D / (U_c - U_D)$$

$$R_d = 4.88 \times (U_c - U_D) / U_c U_D$$

Max h_d 1625 Kcal/M²HrC

$$R_d \leq 0.003 \text{ Btu/Ft}^2 \text{ Ht}^\circ \text{F}$$

If dirt coefficient within limits, then assumed U_D is ok.

Then surface area is sufficient for the duty

13. Finally calculate tube & shell side pressure drop

Shell side frictional pressure drop: $f G_s^2 D_s (N+1) / 2 G \rho D_e$ in kg/m²

D_s : Shell inside diameter m

G_s : Shell side flow kg/m²hr

D_e : Equivalent diameter m

ρ : Density kg/m³

$N + 1$: L / B

L: Shell length mm

B: Baffle spacing mm

Friction factor $f : \text{Exp}(0.576 - (0.19 L_n R_e))$

14. Tube side frictional pressure drop: $4 f (L / D_i) (\rho u^2 / 2 g)$ kg/m²

D_t : Inside tube diameter m

L: Total tube length M= Tube length $\times$ No of passes

Return losses through tube passes: $n4f(4\rho u^2/2g)$

Where n: no of passes

ρ : Density kg/m³

g: 9.81 m/s²

u: Velocity m/s

Total tube side pressure drop: pressure drop due to frictional + return losses

Allowable pressure drops: Shell side: 2 psi, Tube: 10 psi

4.14 FLARE SYSTEM

Flare

The flare is a last line of defense in the safe emergency release system in a refinery or chemical plant. It uses unrecoverable gases emerging with oil from oil wells, vented gases and gaseous water from gas processing units. The flare provides a means of safe disposal of the vapor streams from its facilities, by burning them under controlled conditions such that adjacent equipment or personnel are not exposed to hazards, and at the same time obeying the environmental regulation of pollution control and public relation requirements.

The chemical process used for flaring is a high temperature oxidation reaction to burn combustible components, mostly hydrocarbons, or waste gases from industrial operations. In combustion, the gaseous hydrocarbon reacts with atmospheric oxygen to form carbon dioxide and water. Several by-products formed will be carbon monoxide, hydrogen and others dependent upon what is being burned. Efficiency of hydrocarbon conversion is generally 98%.

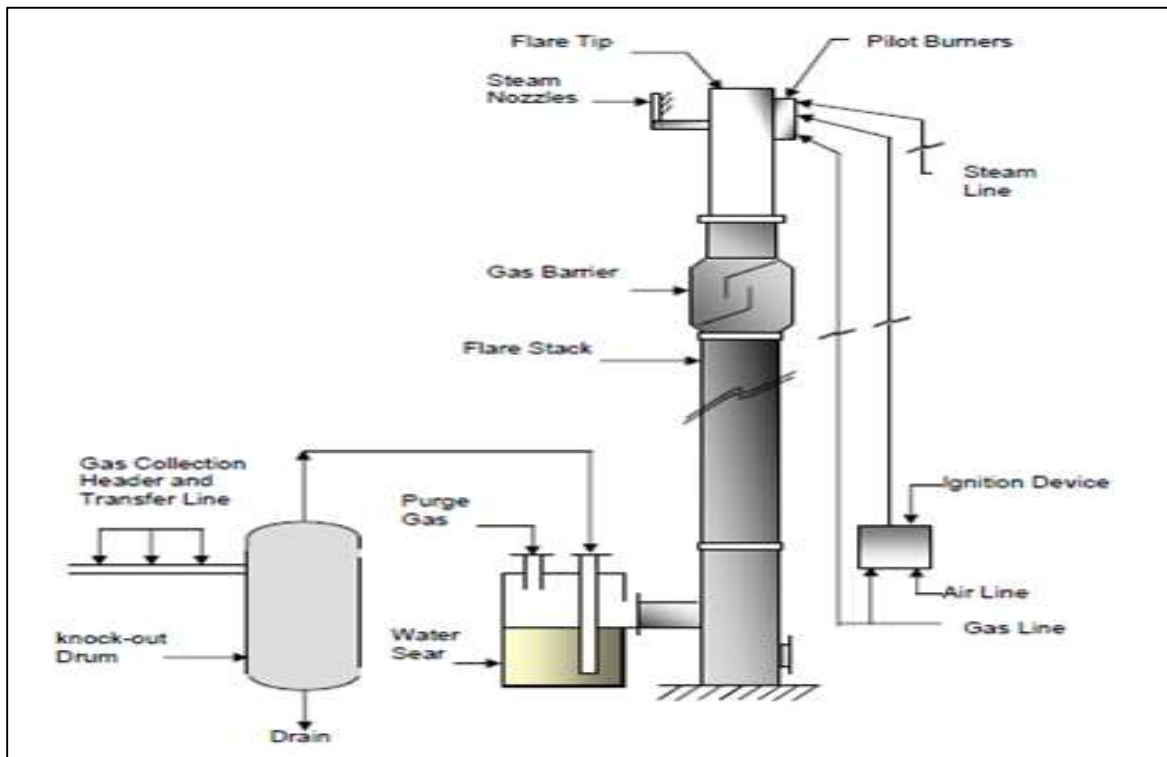
Flare Type

In industry commonly used flare systems are elevated flares and ground flares. Selection of the flare depends on parameters such as space availability, characteristics of the flare gas, economics, investment, operating costs and government regulations.

Elevated Flare

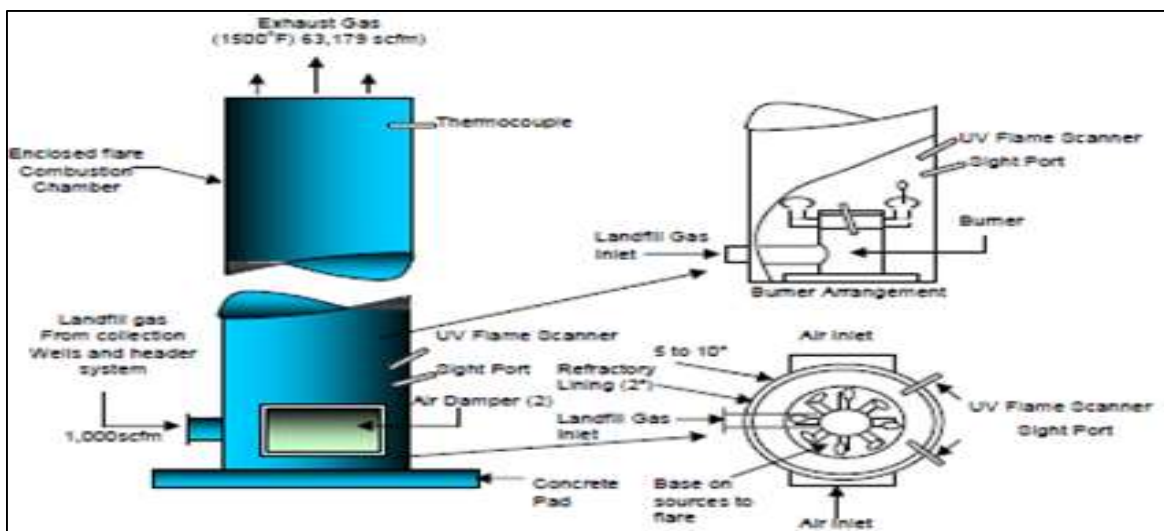
Elevated flare is the most commonly used type in refineries and chemical plants. These have larger capacities than ground flares. The waste gas stream is fed through a stack from 10 m to 100 m tall and is combustible at the tip of the stack.

The elevated flare, can be steam assisted, air assisted or non-assisted. Elevated can utilize steam injection/air injection to make smokeless burning and with low luminosity up to about 20% of maximum flaring load. Disadvantage of steam/air injection will be source of noise and cause noise pollution. By increasing the elevation of stack, flare has the best dispersion characteristics for toxic combustion products. Capital costs are relatively high and an appreciable plant area may be rendered unavailable for plant equipment, because of radiant heat considerations.



Ground Flare

A ground flare is where the combustion takes place at ground level. It varies in complexity, and may consist either of conventional flare burners discharging horizontally with no enclosure or of multiple burners in refractory-lined steel enclosures. Compare to elevated flare, ground flare can be achieved smokeless operation as well, but with essentially no noise or luminosity problems, provided that the design gas rate to the flare is not exceeded.



Flare System Components

Typical flare system consists of

- Gas collection header and piping for collecting gases from processing units.
- A Knockout drum to remove and store condensable and entrained liquids,
- A proprietary seal, water seal, or purge gas supply to prevent flash-back
- A single or multiple burner unit and a flare stack,
- Gas pilots and an igniter to ignite the mixture of waste gas and air and
- A provision for external momentum force (Steam injection or forced air) for smokeless flaring.

Design Factors

Following points are to be considered for designing the flare system design:

Based on the Gas Stream released

- Gas Flow Rate
- Gas Composition
- Gas temperature
- Gas Pressure available
- Utility Cost and availability

Based on the Regulatory Requirement

- Safety Requirements.
- Environmental Requirements.
- Social Requirements.

Design Factors

When designing the flare system, several important parameters have to be considered, these are, flare head design, flare exit velocity, VOC heating value, and whether the flame is assisted by steam or air.

Besides that, the design should be based on consideration below:

- Flare Spacing, Location, and Height
- Radiant Heat
- Burning Liquid Fall out
- Pollution limitations

Flare Spacing, Location, and Height

Flare design capacity is design to handle largest vapor release from pressure relief valve, vapor blow down and other emergency system.

Flashback Seals-flashback protection, which prevents a flame front from travelling back to the upstream piping and equipment. Sizing of flare systems is a function of maximum allowable back pressure on pressure relief valves and other sources of release into the emergency system.

Flare Glossary

- Back Pressure - Back pressure is the sum of the superimposed and build-up back pressures. The pressure that exists at the outlet of a pressure relief device is as a result of the pressure in the discharge system.
- Gas Blower - Device for blowing air to flare system.
- Blow down - The difference between the set pressure and the closing pressure of a pressure relief valve, expressed as a % of the set pressure in pressure units.
- Closed Disposal System - Disposal system which is capable of containing pressure that is different from atmospheric pressure.
- Flare System – A system that safely disposing of waste gases through the use of combustion.
- Flare Stack – It is an elevated vertical stack found on oil wells or oil rigs, and in refineries, chemical plants and landfills used for burning off unusable waste gas or flammable gas and liquids released by pressure relief valves during unplanned over-pressuring of plant equipment.
- Flame Arrestors - A crimped ribbon aluminum or stainless steel flame cell to protect against rapid burn backs in low-pressure situations. These passive safety devices are guaranteed to prevent flame fronts from propagating back through lines, destroying facilities, and causing injuries.
- Flare Tips - Structure at top of the flare play the role to keep an optimum burn and control over all flow rates, which results in a cleaner combustion. The design of the tip makes sure that the tip does not come into contacting with the flame making the tips reliable and long lasting.
- Ignitions system – Is a system use to ignite the flare of flare systems. Normally this system designed to ignite the flare quickly the first time, maintain combustion and re-ignite rapidly to prevent industrial hazards and personal injury while protecting the environment.
- Knockout Drum – Is a drum installed near the flare base, and serves to recover liquid hydrocarbons, prevent liquid slugs, and remove large liquid particles from the gas streams released from relief system.
- Open Disposal System - A disposal system that discharges directly from relief system to atmosphere without other devices.
- Overpressure - Pressure value increase more that the set point pressure of the relieving device, expressed in percent.

- **Pressure Relieving System** - An arrangement of a pressure-relieving device, piping and a means of disposal intended for the safe relief, conveyance, and disposal of fluids in a vapor, liquid, or gaseous phase. It can consist of only one pressure relief valve or rupture disk, either with or without discharge pipe, on a single vessel or line.
- **Rupture Disk Device** - A non-reclosing differential pressure relief device actuated by inlet static pressure and designed to function by bursting the pressure-containing rupture disk. A rupture disk device includes a rupture disk and a rupture disk holder.
- **Relief Valve** – A spring-loaded pressure relief valve is actuated by the static pressure upstream of the valve. The valve opens normally in proportion to the pressure increase over the operating pressure. A relief valve is used primarily with incompressible fluids.
- **Support Structure** – Structure which is designed to withstand local wind conditions for flares. Three types are available: self-supported, Guy-wire supported and Derrick supported.
- **Windbreaker** - A windbreaker prevents the wind from extinguishing the flames at flare tip. It serves also to hide the flames.

4.15 SOFTWARE TOOLS (COMMONLY USED)

Aspen HYSYS

A general process simulation and modelling software very commonly used in the upstream oil and gas industry is HYSYS. HYSYS is extensively used for developing PFD and mass and heat balance.

Aspen Flare System Analyser

It is extensively used for modeling of flare or cold vent networks commonly required in upstream oil and gas. It allows to determine type of relief device and sizing of connected piping including individual tail pipes, flare sub-header and flare main header.

Aspen Exchanger Design and Rating (EDR)

Aspen Exchanger Design and Rating (EDR) enables you to easily design and simulate the heat exchangers. Deliver the optimum heat exchanger size and rating when both cost and performance must be balanced with rigorous up-to-date integrated modeling capabilities.

Bentley WaterGEMS

It is used for analysis and design of water networks. Create and manage customized reports that automatically combine graphs, data tables, color-coded and annotated plan views, and more into a single report. Fire flow, hydraulics, operations, criticality, and water quality can be modeled.

Bentley HAMMER

HAMMER determines appropriate surge control strategies and reduces transients. It uses the Method of Characteristics, the benchmark standard for hydraulic transient analyses. We can run both transient and steady-state analyses in HAMMER.

PIPESIM from SCHLUMBERGER

A simulation and modeling software used for steady-state modeling of “pipelines” both for single-phase, two-phase and three-phase flow.

UniSim

UniSim Design Suite is an intuitive process modeling software that helps to create steady-state and dynamic models for plant design, performance monitoring, troubleshooting, business planning, and asset management. Its major use:

- Process flowsheet development
- Utilizing case scenarios tool to optimize designs against business criteria
- Equipment rating across a broad range of operating conditions
- Evaluating the effect of feed changes, upsets and alternate operations on process safety, reliability and profitability
- Accurately size and select the appropriate material for blowdown systems
- Monitoring equipment performance against operating objectives.

PIPENET

PIPENET is the leader software for rapid fluid flow analysis of pipe and duct networks. Three modules (Standard Module, Spray/Sprinkler Module, Transient Module) ensure that, no matter how complex your network, PIPENET software will perform flow simulation fast and accurately.