References

1. Sinnott R.K and Towler C., "***Chemical Engineering Design***," 5th edition Butterworth-Heinemann, (**2009**).

2. Coke A.K., "***Ludwig Applied Process Design of Chemical and Petrochemical Plant***," vol. 1, 4th edition Gulf professional Publisher, (**2007**).

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5. Peters M, Timmerhause K.D and West R, "***Plant Design and Economics for Chemical Engineers***," 5th edition McGraw-Hill, (**2003**).

6. Perry R and Green D; "***Perry's Chemical Engineers Handbook***," 7th edition McGraw –Hill, (**1997**).

7. Sinnott R.K., "***Chemical Engineering Design***," vol. 6, 4^{th} edition, Coulson & Richardson's Chemical Engineering Series, Elsevier Butterworth-Heinemann, (**2005**).

Courses Requirements:

- Equipment design has two courses each course includes 45 hours per course. All students have to attend course lectures, doing daily and monthly exams, otherwise, unexcused absent for 10% (5 hours) of the total course hours will cause failure in this course. The optional activities for equipment design are participating to solve industrial problems and challenge design day. Feel free to contact me at office hours or in emergency cases at the following emails: em2ihsan@mu.edu.iq; em2ihsan@yahoo.com

Chapter One
Basic Concepts

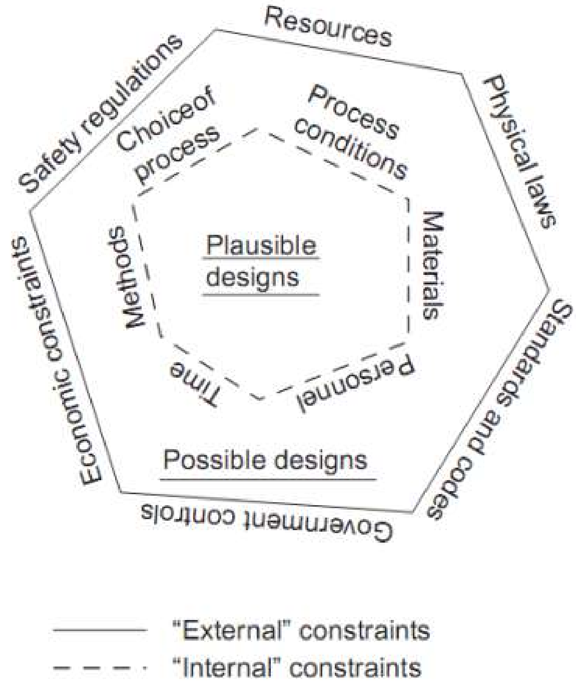
1.1 Introduction

Design is a creative activity, and as such can be one of the most rewarding and satisfying activities undertaken by an engineer. The designer begins with a specific objective or customer need in mind and, by developing and evaluating possible designs, arrives at the best way of achieving that objective for the chemical engineer, a new chemical product or production process.

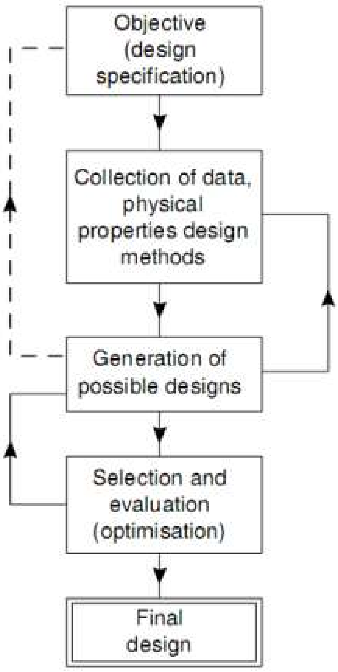
When considering possible ways of achieving the objective, the designer will be constrained by many factors, which will narrow down the number of possible designs. There will rarely be just one possible solution to the problem, just one design. Several alternative ways of meeting the objective will normally be possible, even several best designs, depending on the nature of the constraints.

1- Economic considerations are obviously a major constraint on any engineering design: plants must make a profit.
2- Time will also be a constraint. The time available for completion of a design will usually limit the number of alternative designs that can be.
3- Physical laws.
4- Resources.
5- Safety Regulations.
6- Standard and codes.

These set the outer boundary of possible designs, as shown in figure below. Within this boundary, there will be a number of plausible designs bounded by the other. Economic considerations are obviously a major constraint on any engineering design. Time will also be a constraint. The time available for completion of a design will usually limit the number of alternative designs that can be considered.



The stages in the development of a design, from the initial identification of the objective to the final design, are shown diagrammatically in figure below. This figure shows design as an iterative procedure; as the design develops, the designer will be aware of more possibilities and more constraints, and will be constantly seeking new data and ideas, and evaluating possible design solutions.



1.2 Generation of Possible Design Concepts

Chemical engineering projects can be divided into three types, depending on the novelty involved:

1. Modifications, and additions, to existing plant; usually carried out by the plant design group.
2. New production capacity to meet growing sales demand, and the sale of established processes by contractors. Repetition of existing designs with only minor design changes.
3. New processes, developed from laboratory research, through pilot plant, to a commercial process. Even here, most of the unit operations and process equipment will use established designs.

1.3 Data Collection

Process design will include information on possible processes, equipment performance, and physical property data. This stage can be one of the most time consuming. Most organizations will have design manuals covering preferred methods and data for the more frequently used, routine, design procedures.

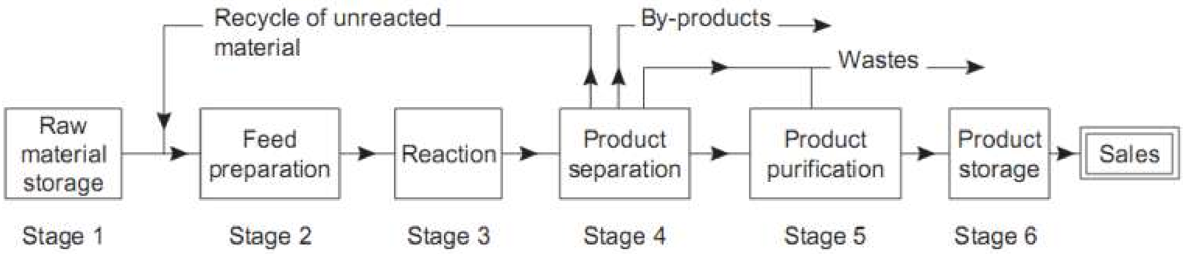
Setting the Design Basis

The most important step in starting a process design is translating the customer need into a design basis. It will normally include the production rate of the main product together with the information on constraints that will influence the design such as:

1- The system of the units to be used.
2- The national, local or company design codes that must be followed.
3- Details of raw materials that available.
4- Information on potential sites where the plant might be located.
5- Information on the condition, availability and prices of utility services.

1.4 The Anatomy of A chemical Manufacturing Process

Chemical engineering design is concerned with the selection and arrangement of the stages and the selection, specification, and design of the equipment required to perform the function of each stage.



Stage 1. Raw material storage

Unless the raw materials (also called essential materials, or feed stocks) are supplied as intermediate products (intermediates) from a neighboring plant, some provision will have to be made to hold several days, or weeks, storage to smooth out fluctuations and interruptions in supply.

Stage 2. Feed preparation

Some purification, and preparation, of the raw materials will usually be necessary before they are sufficiently pure, or in the right form, to be fed to the reaction stage.

Stage 3. Reactor

The reaction stage is the heart of a chemical manufacturing process. In the reactor the raw materials are brought together under conditions that promote the production of the desired product; invariably, by-products and unwanted compounds (impurities) will also be formed.

Stage 4. Product separation

In this first stage after the reactor, the products and by-products are separated from any unreacted material. If in sufficient quantity, the unreacted material will be recycled to the reactor. They may be returned directly to the reactor, or to the feed purification and preparation stage. The by-products may also be separated from the products at this stage.

Stage 5. Purification

Before sale, the main product will usually need purification to meet the product specification. If produced in economic quantities, the by-products may also be purified for sale.

Stage 6. Product storage

Some inventory of finished product must be held to match production with sales. Provision for product packaging and transport will also be needed, depending on the nature of the product. Liquids will normally be dispatched in drams and in bulk tankers (road, rail and sea), solids in sacks, cartons or bales. The stock held will depend on the nature of the product and the market.

Ancillary Processes

In addition to the main process stages, provision will have to be made for the supply of the services (utilities) needed; such as, process water, cooling water, compressed air, steam. Facilities will also be needed for maintenance, firefighting, offices and other accommodation, and laboratories.

1.5 Continuous and Batch Processes:

Continuous processes are designed to operate 24 hours a day, 7 days a week, throughout the year. Some down time will be allowed for maintenance and, for some processes, catalyst regeneration. The plant attainment; that is, the percentage of the available hours in a year that the plant operates, will usually be 90 to 95%. hours operated ,Continuous processes will usually be more economical for large scale production.

Batch processes are designed to operate intermittently. Some, or all, the process units being frequently shut down and started up. Batch processes are used where some flexibility is wanted in production rate or product specification.

Choice of Continuous Versus Batch Production

The choice between batch and continuous operation will not be clear cut, but the following rules can be used as a guide.

Continuous

1. Production rate greater than 5×10^6 kg/h
2. Single product
3. No severe fouling
4. Good catalyst life
5. Proven processes design
6. Established market

Batch

1. Production rate less than 5×10^6 kg/h
2. A range of products or product specifications
3. Severe fouling
4. Short catalyst life
5. New product
6. Uncertain design

Given the higher fixed costs and lower plant utilization of batch processes, batch processing usually makes sense only for products that have high value and are produced in small quantities. Batch plants are commonly used for

- Food products.
- Pharmaceutical products such as drugs, vaccines, and hormones.
- Personal care products.
- Specialty chemicals.

Even in these sectors, continuous production is favored if the process is well understood, the production volume is large, and the market is competitive.

1.6 FACTORS OF SAFETY (DESIGN FACTORS)

Design is an inexact art; errors and uncertainties will arise from uncertainties in the design data available and in the approximations necessary in design calculations. To ensure that the design specification is met, factors are included to give a margin of safety in the design; safety in the sense that the equipment will not fail to perform satisfactorily, and that it will operate safely: will not cause a hazard. "Design factor" is a better term to use, as it does not confuse safety and performance factors.

1.7 The Flow-Sheet

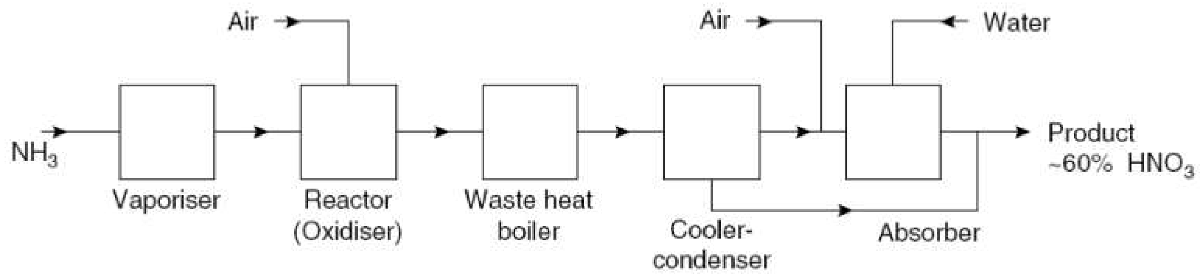
The flow-sheet, also known as the process flow diagram (PFD), is the key document in process design. It shows the arrangement of the equipment selected to carry out the process; the stream connections; stream flow-rates and compositions; and the operating conditions. It is a diagrammatic model of the process. The flow sheet will be used by the specialist design groups as the basis for their designs. This will include piping, instrumentation, equipment design and plant layout. It will also be used by operating personnel for the preparation of operating manuals and operator training. The flow sheet is drawn up from material balances made over the complete process and each individual unit. Energy balances are also made to determine the energy flows and the service requirements.

1.7.1 Flow-Sheet Presentation

As the process flow-sheet is the definitive document on the process, the presentation must be clear, comprehensive, accurate, and complete. The various types of flow-sheets are discussed in the following sections. There are various types of flow-sheet.

1. **Block Diagrams**

A block diagram is the simplest form of presentation. Each block can represent a single piece of equipment or a complete stage in the process. It is useful for showing simple processes. With complex processes, their use is limited to showing the overall process, broken down into its principal stages as nitric acid production:

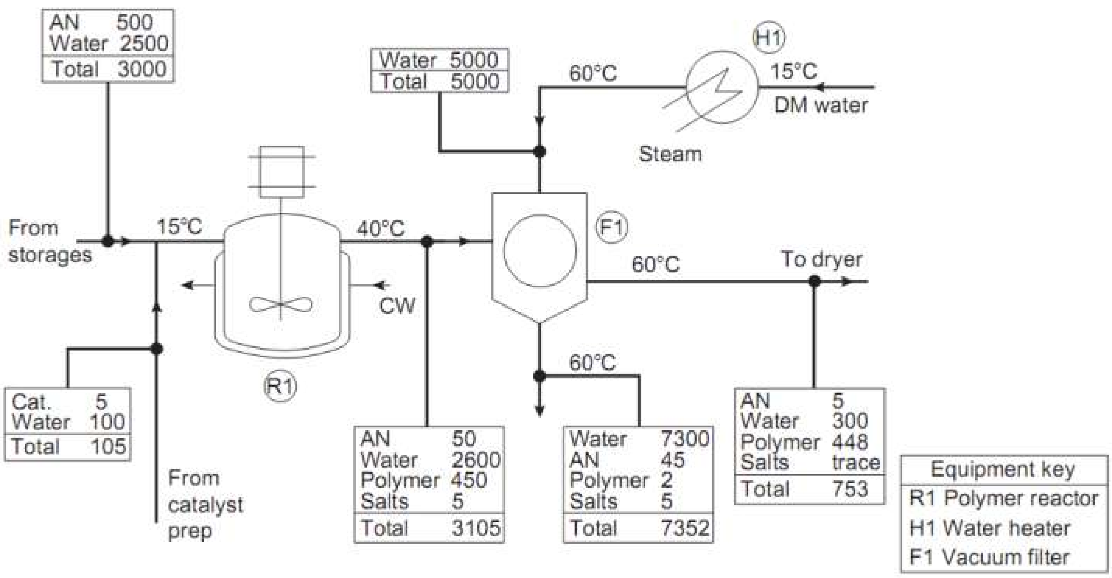


Block diagram: Production of nitric acid by oxidation of ammonia

In that example each block represented the equipment for a complete reaction stage: the vaporizer, reactor, boiler, cooler and absorber column. Block diagrams are useful for representing a process in a simplified form in reports and textbooks, but have only a limited use as engineering documents. The blocks can be of any shape, but it is usually convenient to use a mixture of squares and circles, drawn with a template.

2. Pictorial Representation

A better method for the presentation of data on flow-sheets is shown in this method. In this method each stream line is numbered and the data tabulated at the bottom of the sheet. Alterations and additions can be easily made. This is the method generally used by professional design offices.



Flow-sheet: polymer production

Process Flow Diagram (PFD)

A. Used to present the heat and mass balances of the process.

B. Show all production steps, starting from raw material to final products.

C. Show the operating conditions of each production step in the process (operating conditions are: Temperature, Pressure, Flow-rate, -----).

D. Show the type and quantities of utilities that required for the process (such as: water, steam, O_2, H_2, -----)

E. Shows the process equipment and their connection pipes.

F. Gives some details for the main process equipment (such as: distillation column: its diameter, height, type of internals (trays or packing types), material of construction, thickness, ---)

G. Show the main control system (type of instruments) required for the process.

H. Used as data base for Piping and Instrumentation Diagram (**PID**) and for equipment schedule (equipment summary).

1.7.2 Information to be Included

The amount of information shown on a flow-sheet will depend on the custom and practice of the particular design office. The list given below has therefore been divided into essential items and optional items. The essential items must always be shown, the optional items add to the usefulness of the flow-sheet but are not always included.

Essential information

1. Stream composition, either:

(i) The flow-rate of each individual component, kg/h, which is preferred, or

(ii) The stream composition as a weight fraction.

2. Total stream flow-rate, kg/h.

3. Stream temperature, degrees Celsius preferred.

4. Nominal operating pressure (the required operating pressure).

Optional information

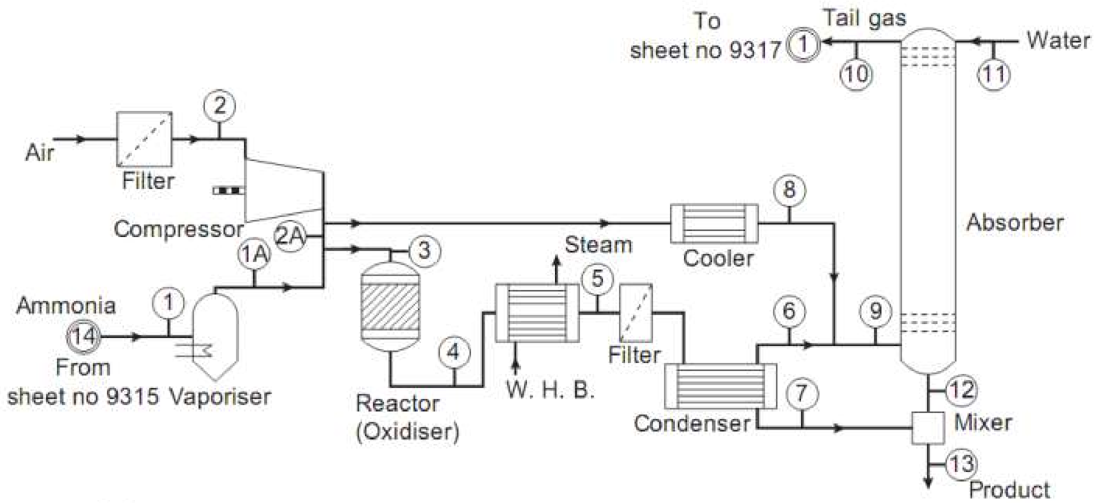
1. Molar percentages composition.

2. Physical property data, mean values for the stream, such as:

(i) Density, kg/m^3.

(ii) Viscosity, $mN\ s/m^2$.

3. Stream name, a brief, one or two-word, description of the nature of the stream.



Flows kg/h pressures nominal

Line no. Stream component	1 Ammonia feed	1A Ammonia vapor	2 Filtered air	2A Oxidiser air	3 Oxidiser feed	4 Oxidiser outlet	5 W.H.B. outlet	6 Condenser gas	7 Condenser acid	8 Secondary air	9 Absorber feed	10 Tail(2) gas	11 Water feed	12 Absorber acid	13 Product acid
NH_3	731.0	731.0	—	—	731.0	Nil	—	—	—	—	—	—	—	—	—
O_2	—	—	3036.9	2628.2	2628.2	935.7	(935.7)[1]	275.2	Trace	408.7	683.9	371.5	—	Trace	Trace
N_2	—	—	9990.8	8644.7	8644.7	8668.8	8668.8	8668.8	Trace	1346.1	10,014.7	10,014.7	—	Trace	Trace
NO	—	—	—	—	—	1238.4	(1238.4)[1]	202.5	—	—	202.5	21.9	—	Trace	Trace
NO_2	—	—	—	—	—	Trace	(?)[1]	967.2	—	—	967.2	(Trace)[1]	—	Trace	Trace
HNO_3	—	—	—	—	—	Nil	Nil	—	850.6	—	—	—	—	1704.0	2554.6
H_2O	—	—	Trace	—	—	1161.0	1161.0	29.4	1010.1	—	29.4	26.3	1376.9	1136.0	2146.0
Total	731.0	731.0	13,027.7	11,272.9	12,003.9	12,003.9	12,003.9	10,143.1	1860.7	1754.8	11,897.7	10,434.4	1376.9	2840.0	4700.6
Press bar	8	8	1	8	8	8	8	8	1	8	8	1	8	1	1
Temp. °C	15	20	15	230	204	907	234	40	40	40	40	25	25	40	43

Flow-sheet: simplified nitric acid process

1.8 Basis of the Calculation

It is good practice to show on the flow-sheet the basis used for the flow-sheet calculations. This would include: the operating hours per year; the reaction and physical yields; and the datum temperature used for energy balances. It is also helpful to include a list of the principal assumptions used in the calculations. This alerts the user to any limitations that may have to be placed on the flow-sheet information.

1.9 Services (Utilities)

To avoid cluttering up the flow-sheet, it is not normal practice to show the service headers and lines on the process flow-sheet. The service connections required on each piece of equipment should be shown and labeled. The service requirements for each piece of equipment can be tabulated on the flow-sheet.

1.10 Equipment Identification

Each piece of equipment shown on the flow-sheet must be identified with a code number and name. The identification number (usually a letter and some digits) will normally be that assigned to a particular piece of equipment as part of the general project control procedures, and will be used to identify it in all the project documents. If the flow-sheet is not part of the documentation for a project, then a simple, but consistent, identification code should be devised. The easiest code is to use an initial letter to identify the type of equipment, followed by digits to identify the particular piece. For example,[H] heat exchangers, [C] columns, [R] reactors. The key to the code should be shown on the flow-sheet.

Chapter Two
PIPING AND INTRUMENT DIAGRAM

2-1 Piping and Instrument Diagram (PID)

The P and I diagram shows the arrangement of the process equipment, piping, pumps, instruments, valves, and other fittings. It should include:

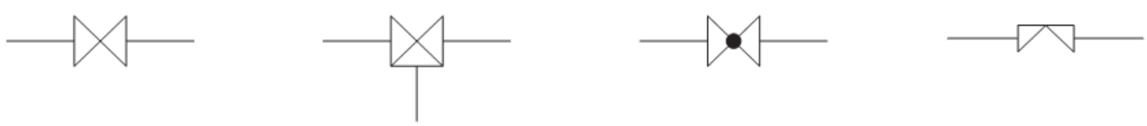
1- Prepared by chemical engineer with the aid of mechanical and control Engineers.
2- All process equipment, identified by an equipment number. The equipment should be drawn roughly in proportion and the location of nozzles shown.
3- All pipes, identified by a line number. The pipe size and material of construction should be shown. The material may be included as part of the line identification number.
4- All valves, control and block valves, with an identification number. The type and size should be shown.
5- Ancillary fittings that are part of the piping system, such as inline sight-glasses, strainers, and steam traps, with an identification number.
6- Pumps, identified by a suitable code number.
7- All control loops and instruments, with an identification number.

2.2 Basic Symbols

The symbols of control valves and instruments will be illustrated:

2.2.1 Control Valves

There are different types of valves:



General Three-way Globe Diaphragm

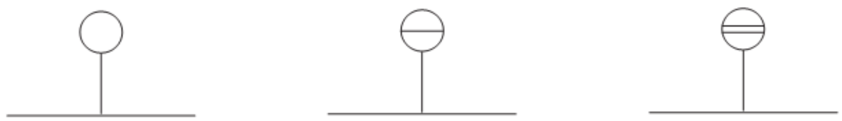
2.2.2 Failure Mode

The direction of the arrow shows the position of the valve on failure of the power supply.

Fails open Fails closed Fails locked in current position Failure mode indeterminate

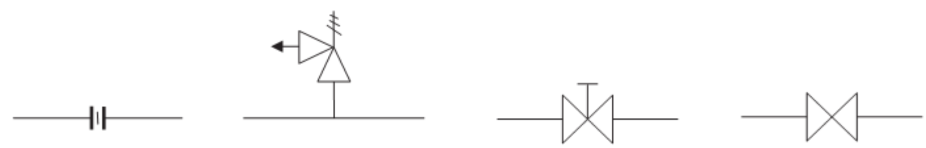
2.2.3 General Instrument and Controller Symbols

Locally mounted means that the controller and display are located out on the plant near to the sensing instrument location. Main panel means that they are located on a panel in the control room. Except on small plants, most controllers would be mounted in the control room.



Field mounted | Panel mounted in primary location | Panel mounted in auxiliary location (local panel)

2.2.4 Other Common Symbols



Restriction orifice | Pressure relief or safety valve | Hand control valve | Gate valve or isolation valve

Self-contained backpressure regulator | Stop check (non-return) valve

2.3 Type of Instrument

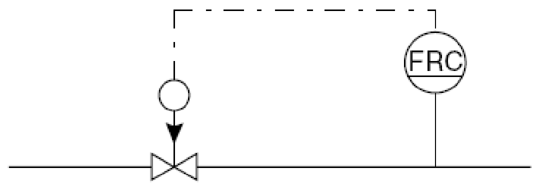
This is indicated on the circle representing the instrument-controller by a letter code as table below:

Initiating or Measured Variable	First Letter	Indicating Only	Controllers			Transmitters	Final Control Element
			Recording	Indicating	Blind		
Analysis (composition)	A	AI	ARC	AIC	AC	AT	AV
Flow rate	F	FI	FRC	FIC	FC	FT	FV
Flow ratio	FF	FFI	FFRC	FFIC	FFC	FFT	FFV
Power	J	JI	JRC	JIC		JT	JV
Level	L	LI	LRC	LIC	LC	LT	LV
Pressure, vacuum	P	PI	PRC	PIC	PC	PT	PV
Pressure differential	PD	PDI	PDRC	PDIC	PDC	PDT	PDV
Quantity	Q	QI	QRC	QIC		QT	QZ
Radiation	R	RI	RRC	RIC	RC	RT	RZ
Temperature	T	TI	TRC	TIC	TC	TT	TV
Temperature differential	TD	TDI	TDRC	TDIC	TDC	TDT	TDV
Weight	W	WI	WRC	WIC	WC	WT	WZ

The first letter indicates the property measured; for example, F = flow. Subsequent letters indicate the function; for example,

I = indicating

RC = recorder controller



A typical control loop

2.4 Valve Selection

The valves used for a chemical process plant can be divided into two broad classes, depending on their primary function:

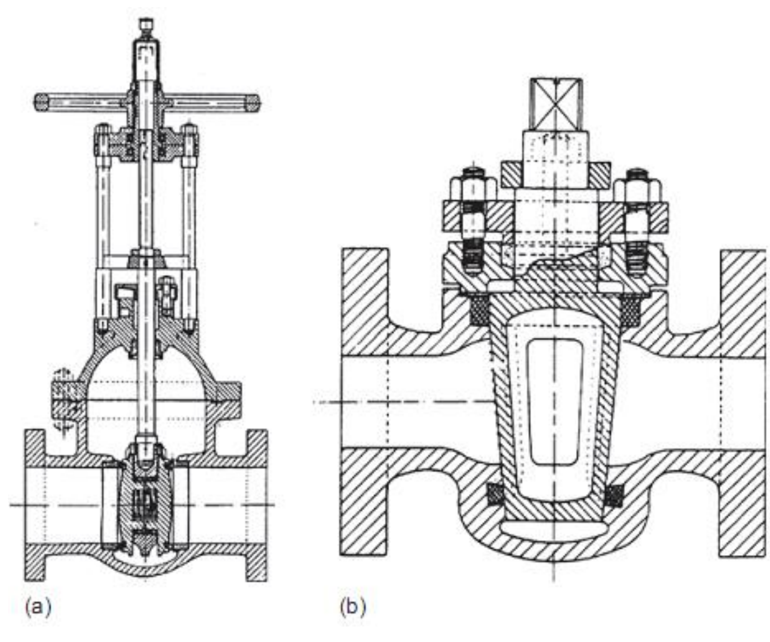
1. Shut-off valves (block valves or isolation valves), whose purpose is to close off the flow.

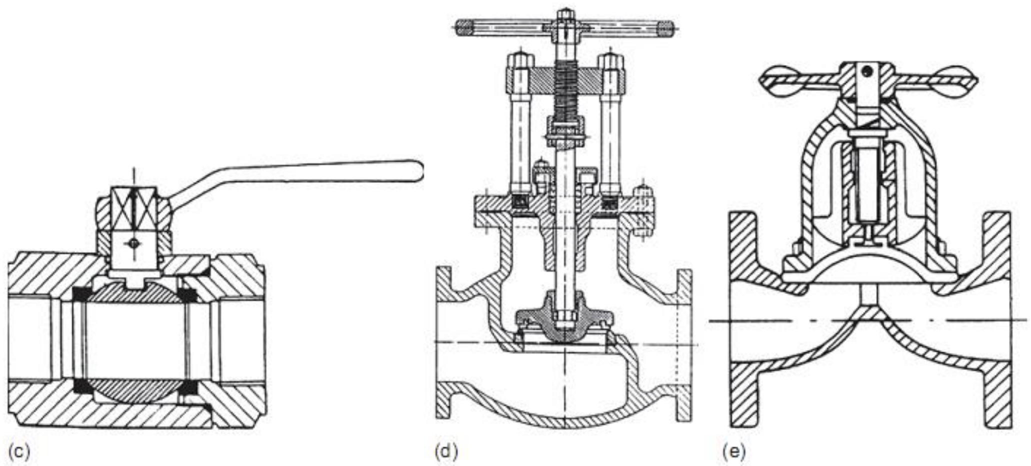
2. Control valves, both manual and automatic, used to regulate flow.

The main types of valves used are:

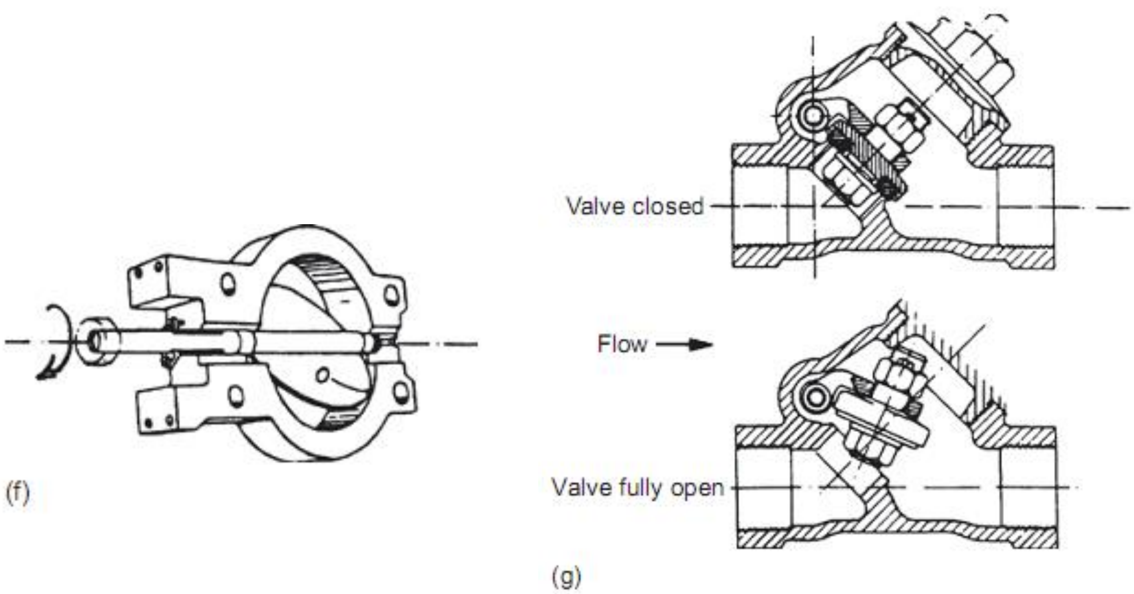
1-Gate valve **2- Plug valve** **3- Ball valve** **4- Globe valve**

5- **Diaphragm valve** **6- Butterfly valve** **7- Non-return valve**





(a) Gate valve (slide valve). (b) Plug valve. (c) Ball valve. (d) Globe valve.
(e) Diaphragm valve.



(f) Butterfly valve. (g) Non-return valve, check valve.

Characteristics of selected Valves:

(1) **Gate valve**: primarily designed to serve as isolation valves. These valves generally are fully open or fully closed.

(2) **Globe valve:** It used to regulate the flow of liquids and vapors. These valves install at By-pass to regulate the flow rate when the (electrical or air) signal was cut-off on the control valve. Also, it install on steam lines for steam turbine.

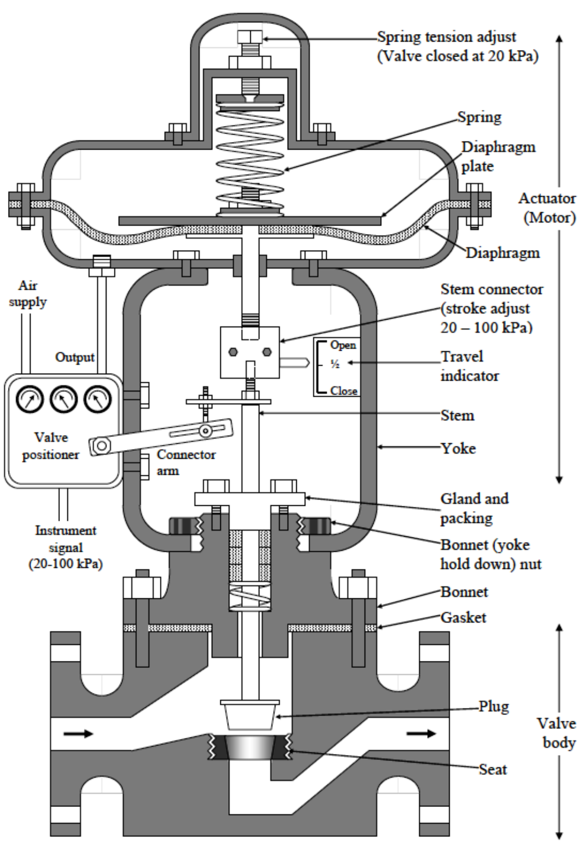
(3) **Ball valve**: It used for more accurate flow of liquids and gases. Also, it is fast open or close by moving arm.

(4) **Butterfly valve**: It used to control and regulate or throttle the flow. It required only a quarter-turn from closed to fully-open position. Butterfly valves are often used for the control of gas and vapor flows.

(5) **Non-return valves**: It used to prevent backflow of fluid in a process line. The valve is kept open by forward flow of fluid and quickly closed by reverse flow.

The careful selection and design of control valves is important; good flow control must be achieved, while keeping the pressure drop as low as possible. The valve must also be sized to avoid the flashing of hot liquids and the supercritical flow of gases and vapors.

Control valves have basically four interactive components : (1) valve body, (2) actuating device (usually a spring diaphragm type), (3) valve positioner (an instrument that converts an electric control signal into air signal to control the position of the valve)m and (4) an airset to supply air pressure to the positioner.



Reverse acting (air to open) pneumatic control valve

2.5 Mechanical Design of Piping Systems

2.5.1. Wall thickness, pipe schedule:

The pipe wall thickness is selected to resist the internal pressure, with an allowance for corrosion. Processes pipes can normally be considered as thin cylinders; only high pressure pipes, such as high-pressure steam lines, are likely to be classified as thick cylinders and must be given special consideration, the following formula used for calculate pipe thickness:

$$t = \frac{Pd}{20\sigma_d + P} \qquad \ldots(2.1)$$

where P = internal pressure, bar.

d = pipe O.D, mm.

σ_d = design stress at working temperature, N/mm.

Pipes are often specified by a schedule number which is defined by:

$$\text{Schedule number} = \frac{P_s \times 1000}{\sigma_s} \quad \ldots(2.2)$$

where P_s = safe working pressure, lb/in^2 (or N /mm^2).

σ_s = safe working stress, lb/in^2 (or N /mm^2).

Schedule 40 pipe is commonly used for general purposes.

Example 2.1

Estimate the safe working pressure for a 4 in. (100 mm) dia., schedule 40 pipe, carbon steel, butt welded, working temperature 100 ^oC. The safe working stress for butt welded steel pipe up to 120 ^oC is 6000 Ib/in^2 (41.4 N/mm^2).

Solution:

$$P_s = \frac{(\text{schedule no.}) \times \sigma_s}{1000} = \frac{40 \times 6000}{1000} = \underline{\underline{240 \text{ lb/in}^2}} = \underline{\underline{1656 \text{ kN/m}^2}}$$

2.6 Economic Pipe Diameter

The capital cost of a pipe run increases with diameter, whereas the pumping costs decrease with increasing diameter. The most economic pipe diameter will be the one which gives the lowest annual operating cost. The following equations for the optimum diameter, for turbulent flow:

Carbon steel pipe:

$$d\text{, optimum} = 293\ G^{0.53}\rho^{-0.37} \qquad \ldots (2.3)$$

Stainless steel pipe:

$$d\text{, optimum} = 260\ G^{0.52}\rho^{-0.37} \qquad \ldots (2.4)$$

where d = pipe inside diameter, mm.
G = mass flow rate, kg/s.
ρ = liquid density, kg/m^3.

Equations (2.3) and (2.4) can be used to make an approximate estimate of the economic pipe diameter for normal pipe runs for turbulent flow

Example 2.2

Estimate the optimum pipe diameter for a water flow rate of 10 kg/s, at 20 ^oC. Carbon steel pipe will be used. Density of water 1000 kg/m.

Solution:

$$d\text{, optimum} = 293 \times (10)^{0.53}\ 1000^{-0.37}$$
$$= \underline{\underline{77.1\ \text{mm}}}$$

use 80-mm pipe.

Viscosity of water at 20°C = 1.1×10^{-3} Ns/m^2,

$$Re = \frac{4G}{\pi\mu d} = \frac{4 \times 10}{\pi \times 1.1 \times 10^{-3} \times 80 \times 10^{-3}} = 1.45 \times 10^5$$

>4000, so flow is turbulent.

Example 2.3

Estimate the optimum pipe diameter for a flow of gaseous HCl of 7000 kg/h a t 5 bar, 15°C, stainless steel pipe. Molar volume 22.4 m^3/kmol, at 1 bar and 0°C.

Solution:

Molecular weight HCl = 36.5.

$$\text{Density at operating conditions} = \frac{36.5}{22.4} \times \frac{5}{1} \times \frac{273}{288} = \underline{\underline{7.72\ \text{kg/m}^3}}$$

$$\text{Optimum diameter} = 260\left(\frac{7000}{3600}\right)^{0.52} 7.72^{-0.37}$$
$$= \underline{\underline{172.4\ \text{mm}}}$$

use 180-mm pipe.

Viscosity of HCl 0.013 m Ns/m^2

$$Re = \frac{4}{\pi} \times \frac{7000}{3600} \times \frac{1}{0.013 \times 10^{-3} \times 180 \times 10^{-3}} = \underline{\underline{1.06 \times 10^6}}, \text{turbulent}$$

2.7 Flanged Joints

Flanged joints are used for connecting pipes and instruments to vessels, for manhole covers, and for removable vessel heads when ease of access is required. Flanges may also be used on the vessel body, when it is necessary to divide the vessel into sections for transport or maintenance. Flanged joints are also used to connect pipes to other equipment, such as pumps and valves. Flanges range in size from a few millimeters diameter for small pipes, to several meters diameter for those used as body or head flanges on vessels.

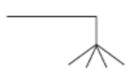
Graphical Symbols for Piping Systems and Plant

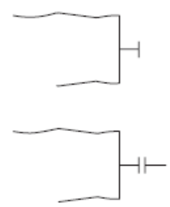
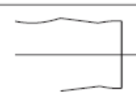
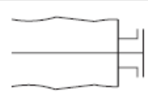
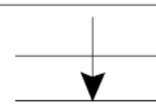
BASED ON BS 1553: PART 1: 1977

Scope

This part of BS 1553 specifies graphical symbols for use in flow and piping diagrams for process plant.

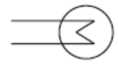
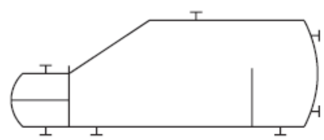
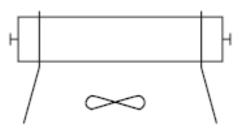
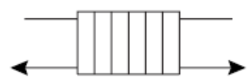
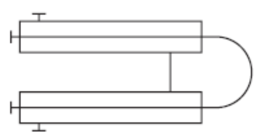
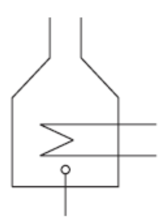
Symbols (or elements of symbols) for use in conjunction with other symbols

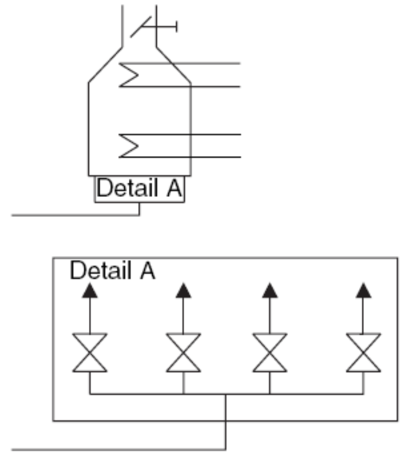
Mechanical linkage	
Weight device	
Electrical device	
Vibratory or loading device (any type)	
Spray device	
Rotary movement	
Stirring device	
Fan	

Access point	
Equipment branch: general symbol Note. The upper representation does not necessarily imply a flange, merely the termination point. Where a breakable connection is required the branch/pipe would be as shown in the lower symbol	
Equipment penetration (fixed)	
Equipment penetration (removable)	
Boundary line	
Point of change	
Discharge to atmosphere	

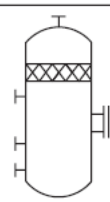
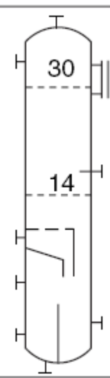
Basic and developed symbols for plant and equipment

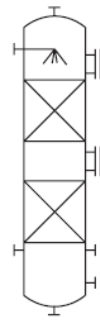
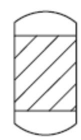
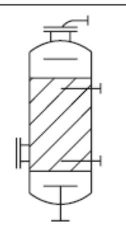
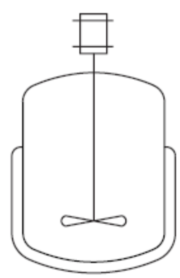
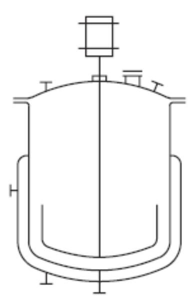
Heat transfer equipment

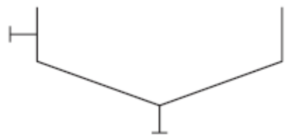
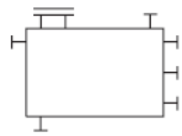
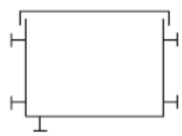
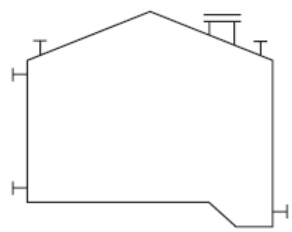
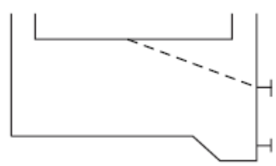
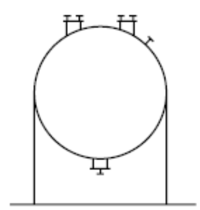
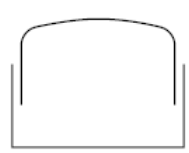
Heat exchanger (basic symbols) Alternative:	 
Shell and tube: fixed tube sheet	
Shell and tube: U tube or floating head	
Shell and tube: kettle reboiler	
Air - blown cooler	
Plate type	
Double pipe type	
Heating/cooling coil (basic symbol)	
Fired heater/boiler (basic symbol)	

Upshot heater Where complex burners are employed the "burner block" may be detailed elsewhere on the drawing, thus	

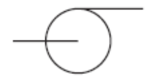
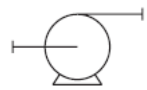
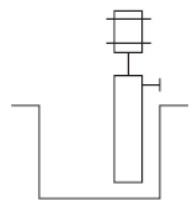
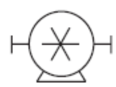
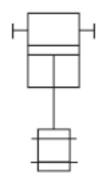
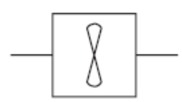
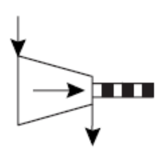
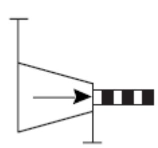
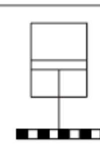
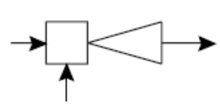
Vessels and tanks

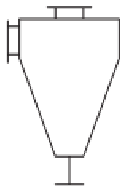
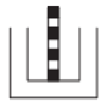
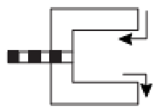
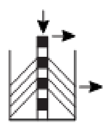
Drum or simple pressure vessel (basic symbol)	
Knock-out drum (with demister pad)	
Tray column (basic symbol)	
Tray column Trays should be numbered from the bottom; at least the first and the last should be shown. Intermediate trays should be included and numbered where they are significant.	

Fluid contacting vessel (basic symbol)	
Fluid contacting vessel Support grids and distribution details may be shown	
Reaction or absorption vessel (basic symbol)	
Reaction or absorption vessel Where it is necessary to show more than one layer of material alternative hatching should be used	
Autoclave (basic symbol)	
Autoclave	

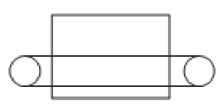
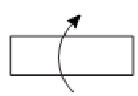
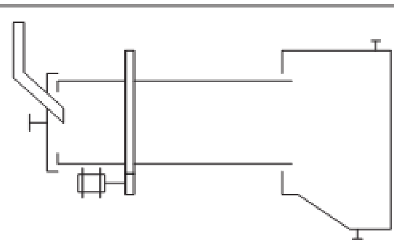
Open tank (basic symbol)	
Open tank	
Clarifier or settling tank	
Sealed tank	
Covered tank	
Tank with fixed roof (with draw-off sump)	
Tank with floating roof (with roof drain)	
Storage sphere	
Gas holder (basic symbol for all types)	

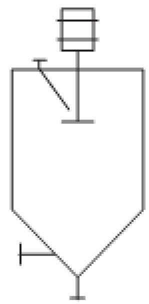
Pumps and compressors

Rotary pump, fan or simple compressor (basic symbol)	
Centrifugal pump or centrifugal fan	
Centrifugal pump (submerged suction)	
Positive displacement rotary pump or rotary compressor	
Positive displacement pump (reciprocating)	
Axial flow fan	
Compressor: centrifugal / axial flow (basic symbol)	
Compressor: centrifugal / axial flow	
Compressor: reciprocating (basic symbol)	
Ejector / injector (basic symbol)	

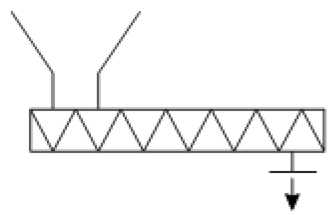
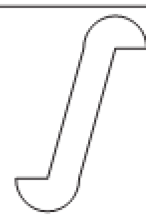
Cyclone and hydroclone (basic symbol)	
Cyclone and hydroclone	
Centrifuge (basic symbol)	
Centrifuge: horizontal peeler type	
Centrifuge: disc bowl type	

Drying

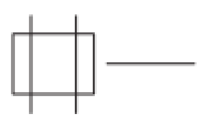
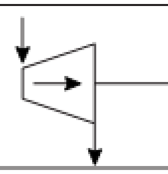
Drying oven	
Belt drier (basic symbol)	
Rotary drier (basic symbol)	
Rotary kiln	

Spray drier	

Materials handling

Belt conveyor	
Screw conveyor	
Elevator (basic symbol)	

Prime movers

Electric motor (basic symbol)	
Turbine (basic symbol)	

Chapter Two
PIPING AND INTRUMENT DIAGRAM

2-1 Piping and Instrument Diagram (PID)

The P and I diagram shows the arrangement of the process equipment, piping, pumps, instruments, valves, and other fittings. It should include:

1- Prepared by chemical engineer with the aid of mechanical and control Engineers.
2- All process equipment, identified by an equipment number. The equipment should be drawn roughly in proportion and the location of nozzles shown.
3- All pipes, identified by a line number. The pipe size and material of construction should be shown. The material may be included as part of the line identification number.
4- All valves, control and block valves, with an identification number. The type and size should be shown.
5- Ancillary fittings that are part of the piping system, such as inline sight-glasses, strainers, and steam traps, with an identification number.
6- Pumps, identified by a suitable code number.
7- All control loops and instruments, with an identification number.

2.2 Basic Symbols

The symbols of control valves and instruments will be illustrated:

2.2.1 Control Valves

There are different types of valves:

General Three-way Globe Diaphragm

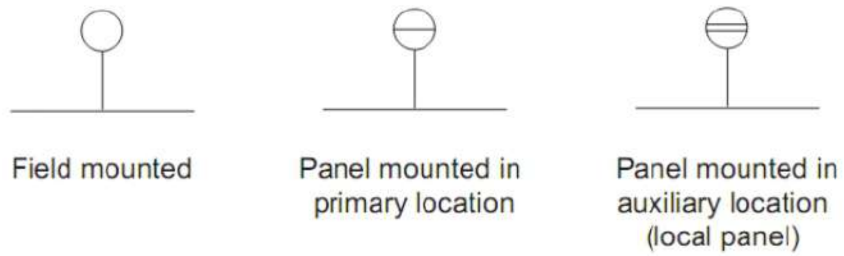
2.2.2 Failure Mode

The direction of the arrow shows the position of the valve on failure of the power supply.

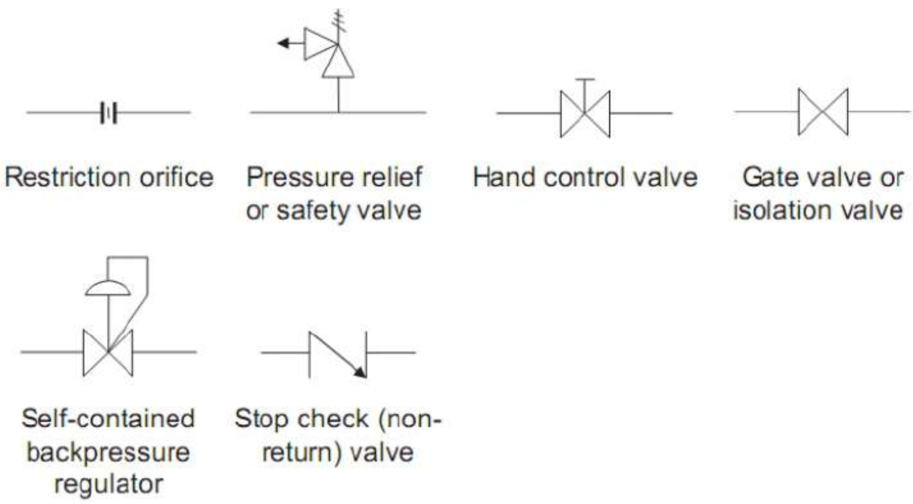
Fails open Fails closed Fails locked in current position Failure mode indeterminate

2.2.3 General Instrument and Controller Symbols

Locally mounted means that the controller and display are located out on the plant near to the sensing instrument location. Main panel means that they are located on a panel in the control room. Except on small plants, most controllers would be mounted in the control room.



2.2.4 Other Common Symbols



2.3 Type of Instrument

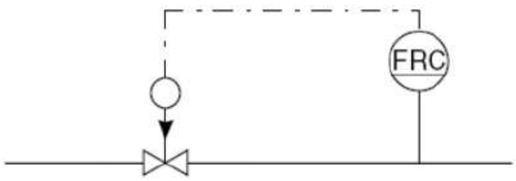
This is indicated on the circle representing the instrument-controller by a letter code as table below:

Initiating or Measured Variable	First Letter	Indicating Only	Controllers			Transmitters	Final Control Element
			Recording	Indicating	Blind		
Analysis (composition)	A	AI	ARC	AIC	AC	AT	AV
Flow rate	F	FI	FRC	FIC	FC	FT	FV
Flow ratio	FF	FFI	FFRC	FFIC	FFC	FFT	FFV
Power	J	JI	JRC	JIC		JT	JV
Level	L	LI	LRC	LIC	LC	LT	LV
Pressure, vacuum	P	PI	PRC	PIC	PC	PT	PV
Pressure differential	PD	PDI	PDRC	PDIC	PDC	PDT	PDV
Quantity	Q	QI	QRC	QIC		QT	QZ
Radiation	R	RI	RRC	RIC	RC	RT	RZ
Temperature	T	TI	TRC	TIC	TC	TT	TV
Temperature differential	TD	TDI	TDRC	TDIC	TDC	TDT	TDV
Weight	W	WI	WRC	WIC	WC	WT	WZ

The first letter indicates the property measured; for example, F = flow. Subsequent letters indicate the function; for example,

I = indicating

RC = recorder controller



A typical control loop

2.4 Valve Selection

The valves used for a chemical process plant can be divided into two broad classes, depending on their primary function:

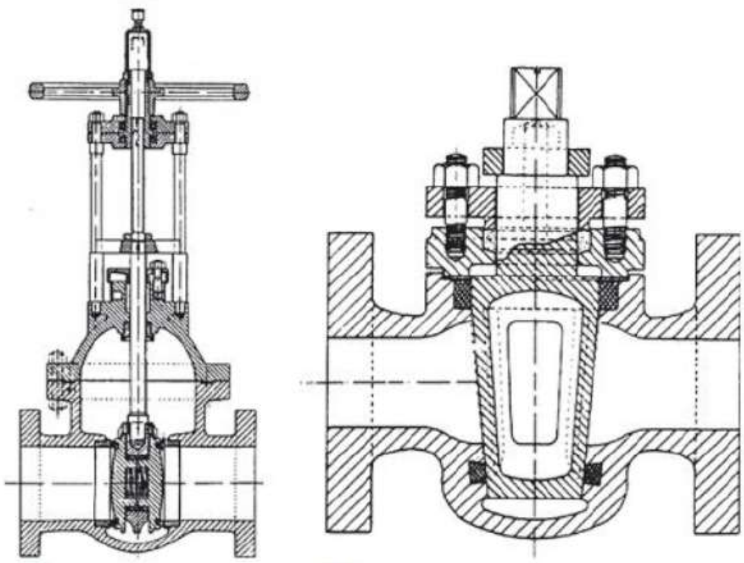
1. Shut-off valves (block valves or isolation valves), whose purpose is to close off the flow.

2. Control valves, both manual and automatic, used to regulate flow.

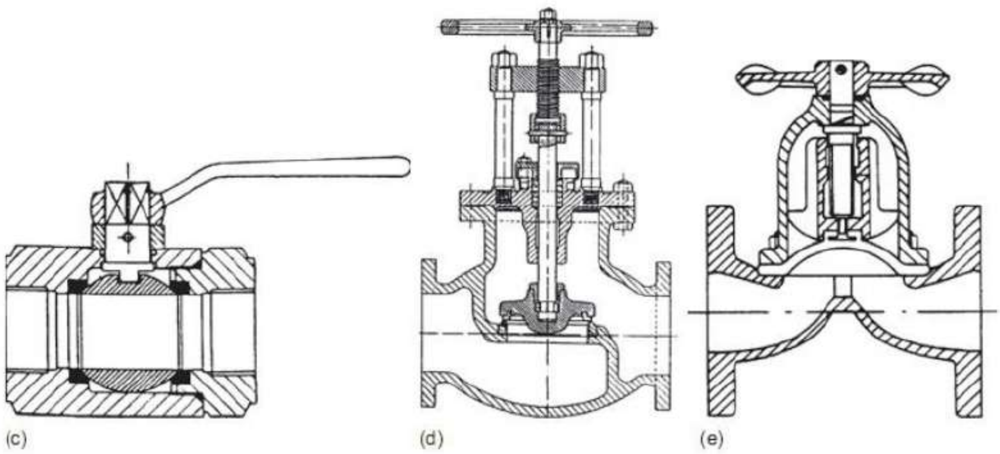
The main types of valves used are:

1-Gate valve **2- Plug valve** **3- Ball valve** **4- Globe valve**

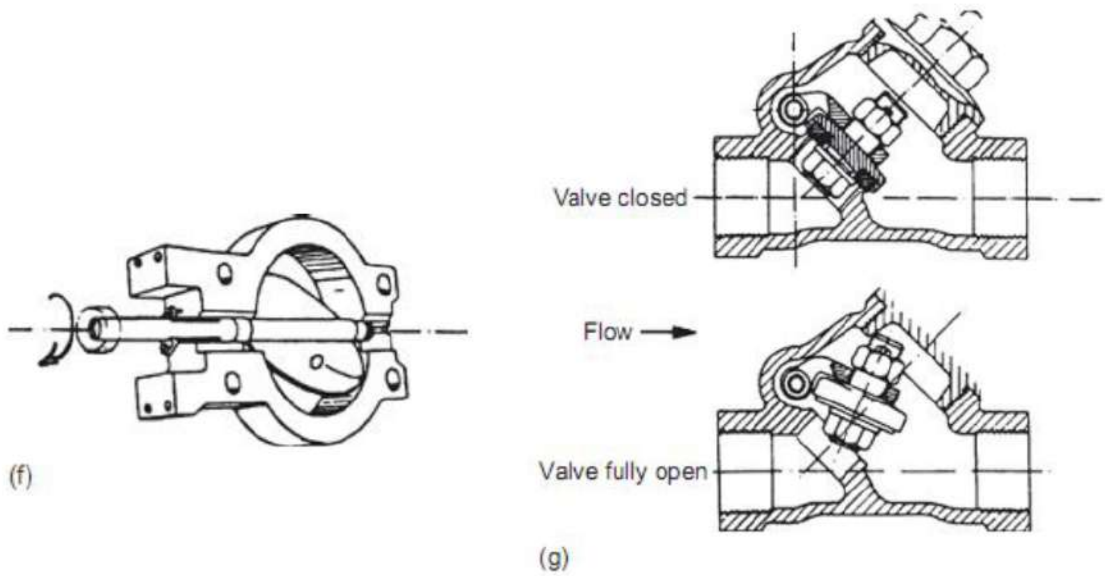
5- **Diaphragm valve** **6- Butterfly valve** **7- Non-return valve**



(a) (b)



(a) Gate valve (slide valve). (b) Plug valve. (c) Ball valve. (d) Globe valve. (e) Diaphragm valve.



(f) Butterfly valve. (g) Non-return valve, check valve.

Characteristics of selected Valves:

(1) **Gate valve**: primarily designed to serve as isolation valves. These valves generally are fully open or fully closed.

(2) **Globe valve:** It used to regulate the flow of liquids and vapors. These valves install at By-pass to regulate the flow rate when the (electrical or air) signal was cut-off on the control valve. Also, it install on steam lines for steam turbine.

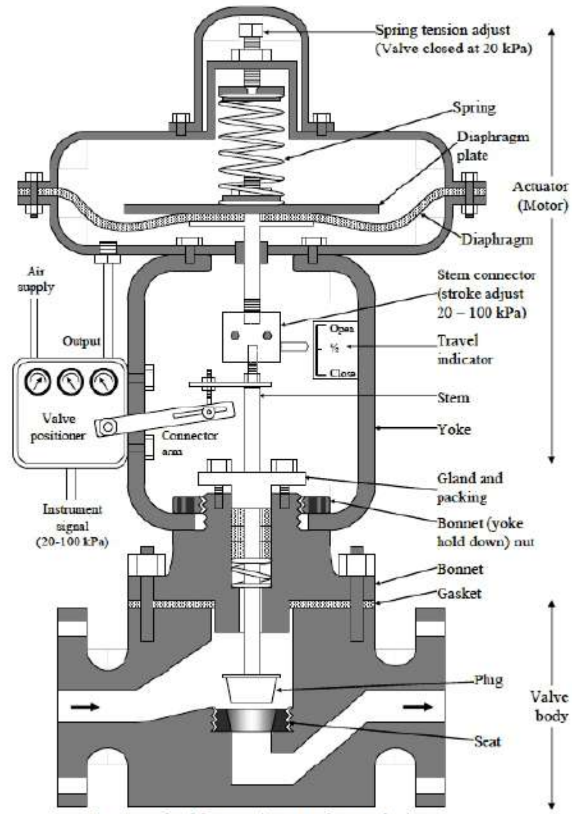
(3) **Ball valve**: It used for more accurate flow of liquids and gases. Also, it is fast open or close by moving arm.

(4) **Butterfly valve**: It used to control and regulate or throttle the flow. It required only a quarter-turn from closed to fully-open position. Butterfly valves are often used for the control of gas and vapor flows.

(5) **Non-return valves**: It used to prevent backflow of fluid in a process line. The valve is kept open by forward flow of fluid and quickly closed by reverse flow.

The careful selection and design of control valves is important; good flow control must be achieved, while keeping the pressure drop as low as possible. The valve must also be sized to avoid the flashing of hot liquids and the supercritical flow of gases and vapors.

Control valves have basically four interactive components : (1) valve body, (2) actuating device (usually a spring diaphragm type), (3) valve positioner (an instrument that converts an electric control signal into air signal to control the position of the valve)m and (4) an airset to supply air pressure to the positioner.



Reverse acting (air to open) pneumatic control valve

2.5 Mechanical Design of Piping Systems

2.5.1. Wall thickness, pipe schedule:

The pipe wall thickness is selected to resist the internal pressure, with an allowance for corrosion. Processes pipes can normally be considered as thin cylinders; only high pressure pipes, such as high-pressure steam lines, are likely to be classified as thick cylinders and must be given special consideration, the following formula used for calculate pipe thickness:

$$t = \frac{Pd}{20\sigma_d + P} \qquad \ldots(2.1)$$

where P = internal pressure, bar.
d = pipe O.D, mm.
σ_d = design stress at working temperature, N/mm.

Pipes are often specified by a schedule number which is defined by:

$$\text{Schedule number} = \frac{P_s \times 1000}{\sigma_s} \qquad \ldots(2.2)$$

where P_s = safe working pressure, lb/in^2 (or N /mm^2).
σ_s = safe working stress, lb/in^2 (or N /mm^2).

Schedule 40 pipe is commonly used for general purposes.

Example 2.1

Estimate the safe working pressure for a 4 in. (100 mm) dia., schedule 40 pipe, carbon steel, butt welded, working temperature 100 ^oC. The safe working stress for butt welded steel pipe up to 120 ^oC is 6000 Ib/in^2 (41.4 N/mm^2).

Solution:

$$P_s = \frac{(\text{schedule no.}) \times \sigma_s}{1000} = \frac{40 \times 6000}{1000} = \underline{\underline{240\ \text{lb/in}^2}} = \underline{\underline{1656\ \text{kN/m}^2}}$$

2.6 Economic Pipe Diameter

The capital cost of a pipe run increases with diameter, whereas the pumping costs decrease with increasing diameter. The most economic pipe diameter will be the one which gives the lowest annual operating cost. The following equations for the optimum diameter, for turbulent flow:

Carbon steel pipe:

$$d\text{, optimum} = 293\ G^{0.53}\rho^{-0.37} \qquad \ldots (2.3)$$

Stainless steel pipe:

$$d\text{, optimum} = 260\ G^{0.52}\rho^{-0.37} \qquad \ldots (2.4)$$

where d = pipe inside diameter, mm.
G = mass flow rate, kg/s.
ρ = liquid density, kg/m^3.

Equations (2.3) and (2.4) can be used to make an approximate estimate of the economic pipe diameter for normal pipe runs for turbulent flow

Example 2.2

Estimate the optimum pipe diameter for a water flow rate of 10 kg/s, at 20 °C. Carbon steel pipe will be used. Density of water 1000 kg/m.

Solution:

$$d\text{, optimum} = 293 \times (10)^{0.53}\ 1000^{-0.37}$$
$$= \underline{\underline{77.1\text{ mm}}}$$

use 80-mm pipe.

Viscosity of water at 20°C = 1.1×10^{-3} Ns/m^2,

$$Re = \frac{4G}{\pi\mu d} = \frac{4 \times 10}{\pi \times 1.1 \times 10^{-3} \times 80 \times 10^{-3}} = 1.45 \times 10^5$$

>4000, so flow is turbulent.

Example 2.3

Estimate the optimum pipe diameter for a flow of gaseous HCl of 7000 kg/h a t 5 bar, 15°C, stainless steel pipe. Molar volume 22.4 m^3/kmol, at 1 bar and 0°C.

Solution:

Molecular weight HCl = 36.5.

$$\text{Density at operating conditions} = \frac{36.5}{22.4} \times \frac{5}{1} \times \frac{273}{288} = \underline{\underline{7.72 \text{ kg/m}^3}}$$

$$\text{Optimum diameter} = 260 \left(\frac{7000}{3600}\right)^{0.52} 7.72^{-0.37}$$

$$= \underline{\underline{172.4 \text{ mm}}}$$

use 180-mm pipe.

Viscosity of HCl 0.013 m Ns/m^2

$$Re = \frac{4}{\pi} \times \frac{7000}{3600} \times \frac{1}{0.013 \times 10^{-3} \times 180 \times 10^{-3}} = \underline{\underline{1.06 \times 10^6}}, \text{ turbulent}$$

2.7 Flanged Joints

Flanged joints are used for connecting pipes and instruments to vessels, for manhole covers, and for removable vessel heads when ease of access is required. Flanges may also be used on the vessel body, when it is necessary to divide the vessel into sections for transport or maintenance. Flanged joints are also used to connect pipes to other equipment, such as pumps and valves. Flanges range in size from a few millimeters diameter for small pipes, to several meters diameter for those used as body or head flanges on vessels.

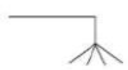
Graphical Symbols for Piping Systems and Plant

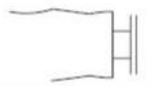
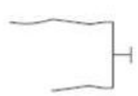
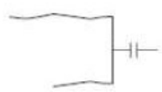
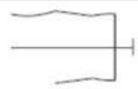
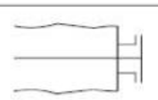
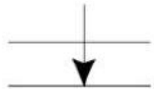
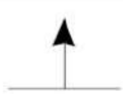
BASED ON BS 1553: PART 1: 1977

Scope

This part of BS 1553 specifies graphical symbols for use in flow and piping diagrams for process plant.

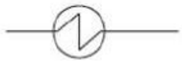
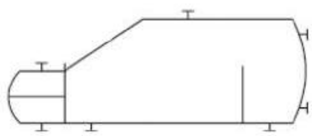
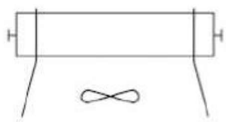
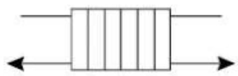
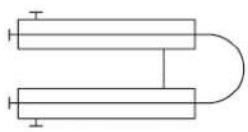
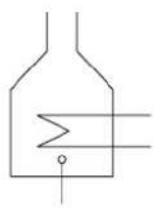
Symbols (or elements of symbols) for use in conjunction with other symbols

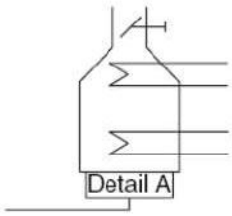
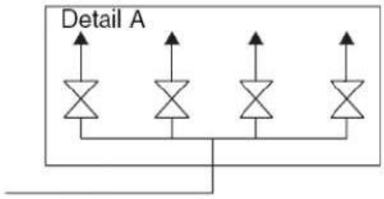
Mechanical linkage	
Weight device	
Electrical device	
Vibratory or loading device (any type)	
Spray device	
Rotary movement	
Stirring device	
Fan	

Access point	
Equipment branch: general symbol Note. The upper representation does not necessarily imply a flange, merely the termination point. Where a breakable connection is required the branch/pipe would be as shown in the lower symbol	 
Equipment penetration (fixed)	
Equipment penetration (removable)	
Boundary line	
Point of change	
Discharge to atmosphere	

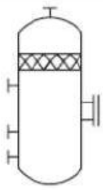
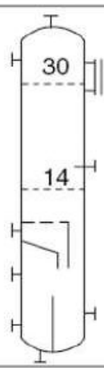
Basic and developed symbols for plant and equipment

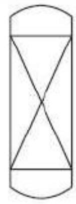
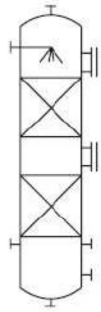
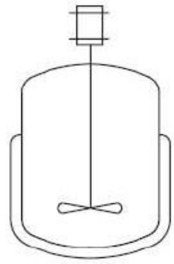
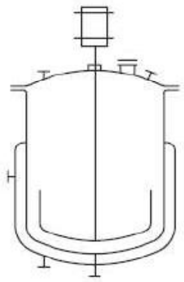
Heat transfer equipment

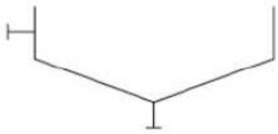
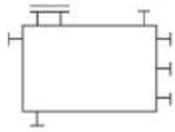
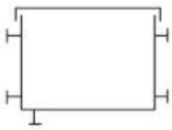
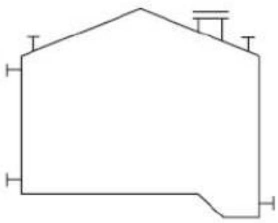
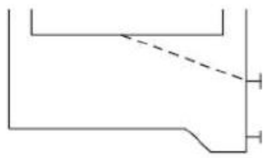
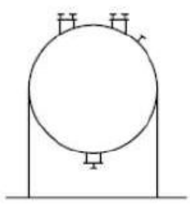
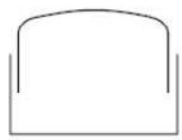
Heat exchanger (basic symbols) Alternative:	 
Shell and tube: fixed tube sheet	
Shell and tube: U tube or floating head	
Shell and tube: kettle reboiler	
Air - blown cooler	
Plate type	
Double pipe type	
Heating/cooling coil (basic symbol)	
Fired heater/boiler (basic symbol)	

Upshot heater	
Where complex burners are employed the "burner block" may be detailed elsewhere on the drawing, thus	

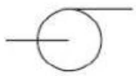
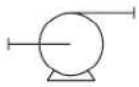
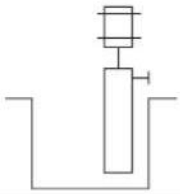
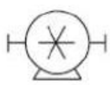
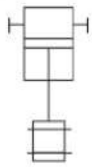
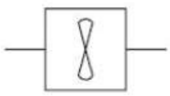
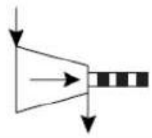
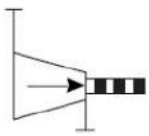
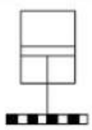
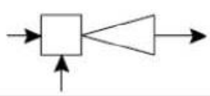
Vessels and tanks

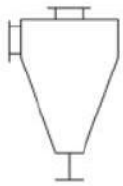
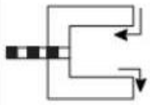
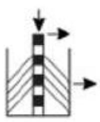
Drum or simple pressure vessel (basic symbol)	
Knock-out drum (with demister pad)	
Tray column (basic symbol)	
Tray column Trays should be numbered from the bottom; at least the first and the last should be shown. Intermediate trays should be included and numbered where they are significant.	

Fluid contacting vessel (basic symbol)	
Fluid contacting vessel Support grids and distribution details may be shown	
Reaction or absorption vessel (basic symbol)	
Reaction or absorption vessel Where it is necessary to show more than one layer of material alternative hatching should be used	
Autoclave (basic symbol)	
Autoclave	

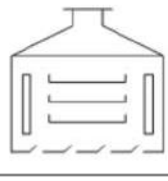
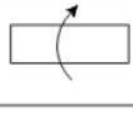
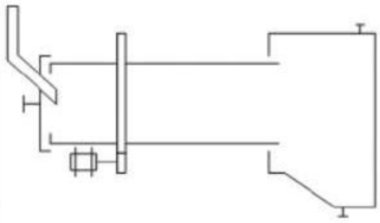
Open tank (basic symbol)	
Open tank	
Clarifier or settling tank	
Sealed tank	
Covered tank	
Tank with fixed roof (with draw-off sump)	
Tank with floating roof (with roof drain)	
Storage sphere	
Gas holder (basic symbol for all types)	

Pumps and compressors

Rotary pump, fan or simple compressor (basic symbol)	
Centrifugal pump or centrifugal fan	
Centrifugal pump (submerged suction)	
Positive displacement rotary pump or rotary compressor	
Positive displacement pump (reciprocating)	
Axial flow fan	
Compressor: centrifugal / axial flow (basic symbol)	
Compressor: centrifugal / axial flow	
Compressor: reciprocating (basic symbol)	
Ejector / injector (basic symbol)	

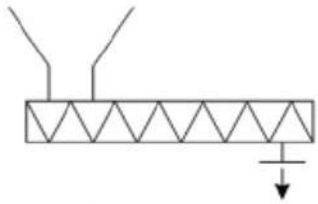
Cyclone and hydroclone (basic symbol)	
Cyclone and hydroclone	
Centrifuge (basic symbol)	
Centrifuge: horizontal peeler type	
Centrifuge: disc bowl type	

Drying

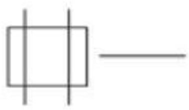
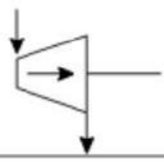
Drying oven	
Belt drier (basic symbol)	
Rotary drier (basic symbol)	
Rotary kiln	

Spray drier	

Materials handling

Belt conveyor	
Screw conveyor	
Elevator (basic symbol)	

Prime movers

Electric motor (basic symbol)	
Turbine (basic symbol)	

Chapter Three
COST ESTIMATIONS

3.1 Costing and Project Evaluation

The design engineer needs to be able to make quick, rough, cost estimates to decide between alternative designs and for project evaluation. Chemical plants are built to make a profit, and an estimate of the investment required and the cost of production are needed before the profitability of a project can be assessed.

3.2 Accuracy and Purpose of Capital Cost Estimates

The accuracy of an estimate depends on the amount of design detail available. The accuracy of the cost data available and the time spent on preparing the estimate. Capital cost estimates can be broadly classified into three types according to their accuracy and purpose:

1. Preliminary (approximate) estimates, accuracy typically ±30 % which are used in initial feasibility studies and to make coarse choices between design alternatives.
2. Authorization (Budgeting) estimates, accuracy typically ±10 - 15 % which are used for the authorization of funds.
3. Detailed (Quotation) estimates, accuracy ±5 - 10 % which are used for project cost control and estimates for fixed price contracts.

3.3 Fixed and Working Capital

Fixed capital is the total cost of the plant ready for start-up. It is the cost paid to the contractors. It includes the cost of:

1. Design, and other engineering and construction supervision.
2. All items of equipment and their installation.
3. All piping, instrumentation and control systems.
4. Buildings and structures.
5. Auxiliary facilities, such as utilities, land and civil engineering work.

Working capital is the additional investment needed, over and above the fixed capital, to start the plant up and operate it to the point when income is earned. It includes the cost of:

1. Start-up.
2. Initial catalyst charges.
3. Raw materials and intermediates in the process.

4. Finished product inventories.
5. Funds to cover outstanding accounts from customers.

Most of the working capital is recovered at the end of the project. The total investment needed for a project is the sum of the fixed and working capital.

3.4 Cost Inflation

All cost-estimating methods use historical data, and are themselves forecasts of future costs. Some method has to be used to update old cost data for use in estimating at the design stage, and to forecast the future construction cost of the plant.

$$\text{Cost in year A} = \text{Cost in year B} \times \frac{\text{Cost index in year A}}{\text{Cost index in year B}}$$

Many methods had been adopted for estimating the values of the cost index. One is the process engineering index.

To estimate the future cost of a plant some prediction has to be made of the future annual rate of inflation. This can be based on the extrapolation of one of the published indices, tempered by the engineer's own assessment of what the future may hold.

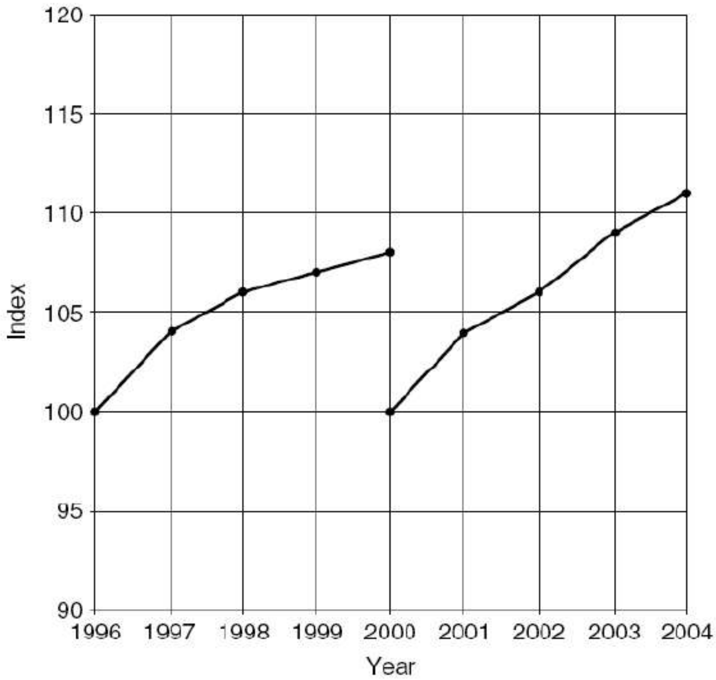


Fig. (3.1) Process Engineering Index

A composite index for the United States process plant industry is published monthly in the journal *Chemical Engineering*, the CPE plant cost index. The longer the period over which the correlation is made the more unreliable the estimate. Between 1970 and 1990 prices rose dramatically. Since then the annual rise has slowed down and is now averaging around 2 - 3 % per year.

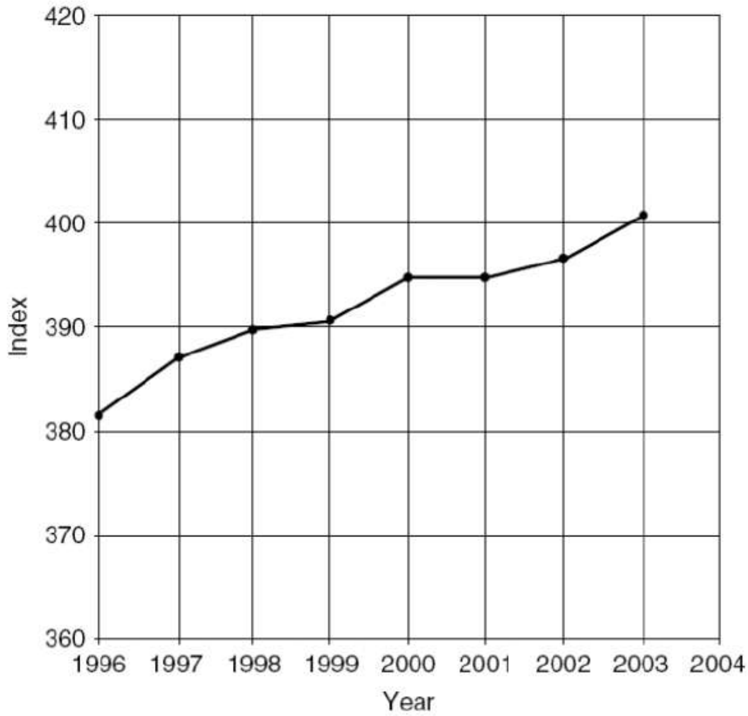


Fig. (3.2) CPE index

Example 3.1:

The purchased cost of a shell and tube heat exchanger, carbon shell, stainless steel tubes, heat transfer area 500 m^2, was £7600 in January 1998; estimate the cost in January 2006. Use the *Process Engineering* plant index.

Solution

From Process Engineering Index figure (3.1) :

Index in 1998 = 106

2000 = 108, 100 (change of base)

2004 = 111

So, estimated cost in January 2000 = 7600 × 108/106 = £7743

and in 2004 = 7743 × 111/100 = £8595

From Process Engineering Index figure (3.1):

The average increase in costs = (111-100)/4= 2.75 per year.

Use this value to predict the exchanger cost in 2006.
The cost index in 2006 = 2 × 2.75 + 111 = 116.5
Cost in 2006 = 8595 × 116.5/111 = £9021 say £9000.

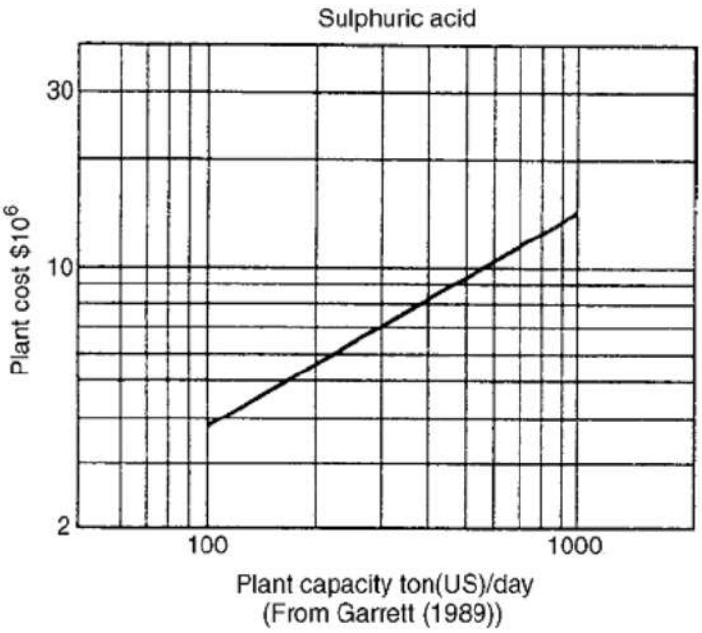
3.5 Historical Costs

An approximate estimate of the capital cost of a project can be obtained from knowledge of the cost of earlier projects using the same manufacturing process. The capital cost of a project is related to capacity by the equation

$$C_2 = C_1 \left(\frac{S_2}{S_1}\right)^n$$

Where:
C_2 = capital cost of the project with capacity, S_2
C_1 = capital cost of the project with capacity, S_1

The value of the index n is traditionally taken as 0.6; the well-known six-tenths rule. This value can be used to get a rough estimate of the capital cost if there are not sufficient data available to calculate the index for the particular process. This equation is only an approximation, and if sufficient data are available the relationship is best represented on a log-log plot. Garrett (1989) has published capital cost-plant capacity curves for over 250 processes.



(From Garrett (1989))

3.6 Estimating Equipment Costs by Scaling

The six-tenths rule should only be used for heat exchangers in the absence of other information. In general, the cost-capacity concept should not be used beyond a tenfold range of capacity, and care must be taken to make certain the two pieces of equipment are similar with regard to type of construction, materials of construction, temperature and pressure operating range, and other pertinent variables. Table 1 contains values for other units:

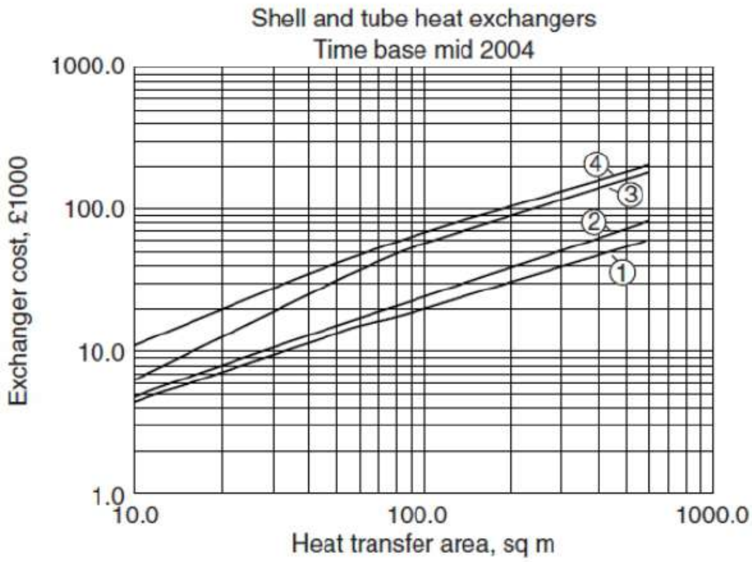
Table (3.1): Typical exponents for equipment cost vs. capacity

Equipment	Siie range	Exponent
Blender, double cone rotary, c.s.	SO-250 ft^3	0.49
Blower, centrifugal	10^3–10^4 ft^3/min	0.59
Centrifuge, solid bowl, c.s.	10–10^2 hp drive	0.67
Crystallizer, vacuum batch, c.s.	500-7000 ft^3	0.37
Compressor, reciprocating, air cooled, two-stage, 150 psi discharge	10–400 ft^3/min	0.69
Compressor, rotary, single-stage, sliding vane, 150 psi discharge	10^2–10^3 ft^3/min	0.79
Dryer, drum, single vacuum	10–10^2 ft^2	0.76
Dryer, drum, single atmospheric	10–10^2 ft^2	0.40
Evaporator (installed), horizontal tank	10^2–10^4 ft^2	0.54
Fan, centrifugal	10^3–10^4 ft^3/min	0.44
Fan, centrifugal	2×10^4–7×10^4 ft^3/min	1.17
Heat exchanger, shell and tube, floating head, c.s.	100-400 ft^2	0.60
Heat exchanger, shell and tube, fixed sheet, c.s.	100-400 ft^2	0.44
Kettle, cast iron, jacketed	250-800 gal	0.27
Kettle, glass lined, jacketed	200-800 gal	0.31
Motor, squirrel cage, induction, 440 volts, explosion proof	5-20 hp	0.69
Motor, squirrel cage, induction, 440 volts, explosion proof	20-200 hp	0.99
Pump, reciprocating, horizontal cast iron (includes motor)	2–100 gpm	0.34
Pump, centrifugal, horizontal, cast steel (includes motor)	10^4–10^5 gpm $\times$ psi	0.33
Reactor, glass lined, jacketed (without drive)	50–600 gal	0.54
Reactor, s.s, 300 psi	10^2–10^3 gal	0.56
Separator, centrifugal, c.s.	50-250 ft^3	0.49
Tank, flat head, c.s.	10^2–10^4 gal	0.57
Tank, c.s., glass lined	10^2–10^3 gal	0.49
Tower, c.s.	10^3–2×10^6 lb	0.62
Tray, bubble cup, c.s.	3-10 ft diameter	1.20
Tray, sieve, c.s.	3-10 ft diameter	0.86

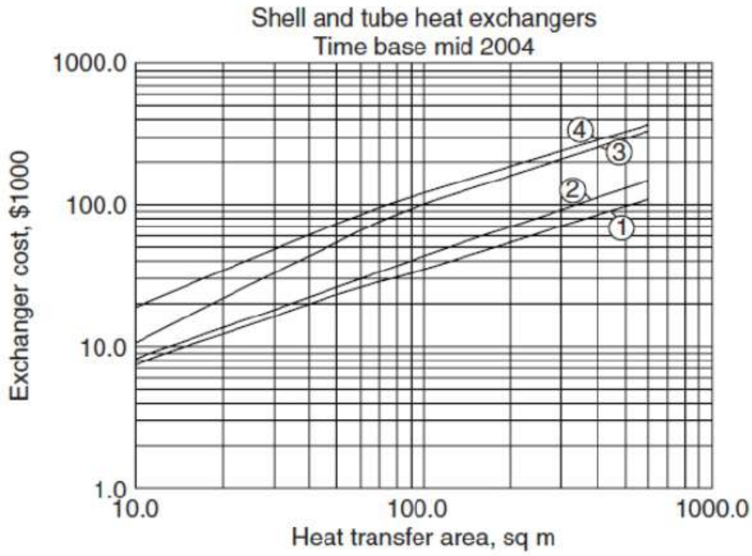
3.7 Estimation of Equipment Costs:

The cost data given in Figures (3.3) to (3.7), and Table (3.2) have been compiled from various sources. They can be used to make preliminary estimates in pounds sterling and US dollars. The base date is mid-2004, and the prices are thought to be accurate to within ±25 %.

The cost of specialized equipment, which cannot be found in the literature, can usually be estimated from the cost of the components that make up the equipment. For example, a reactor design is usually unique for a particular process but the design can be broken down into standard components (vessels, heat-exchange surfaces, agitators) the cost of which can be found in the literature and used to build up an estimate of the reactor cost.



(*a*) Pounds sterling



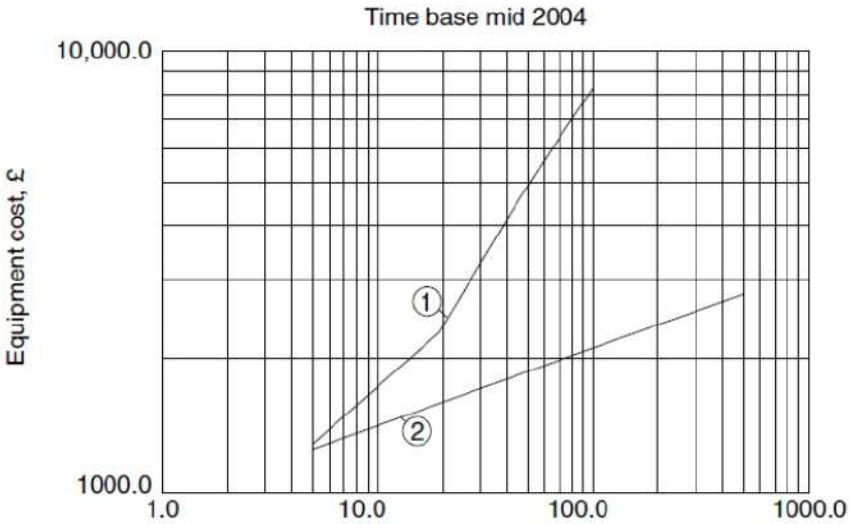
(*b*) US dollars

Materials	
Shell	Tubes
① Carbon steel	Carbon steel
② C.S.	Brass
③ C.S.	Stainless steel
④ S.S.	S.S.

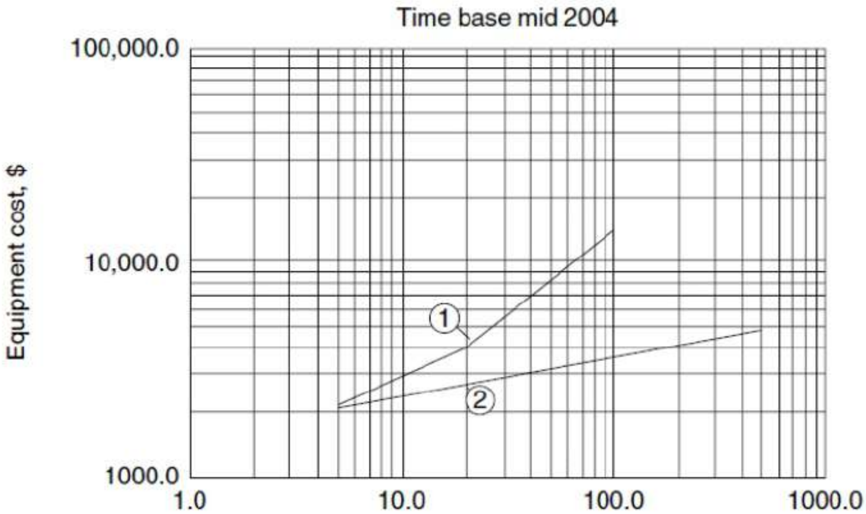
Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

Type factors	
Floating head	× 1.0
Fixed tube sheet	× 0.8
U tube	× 0.85
Kettle	× 1.3

Figure (3.3): Shell and tube heat exchangers. Time base mid-2004
Purchased cost = (bare cost from figure) × Type factor × Pressure factor



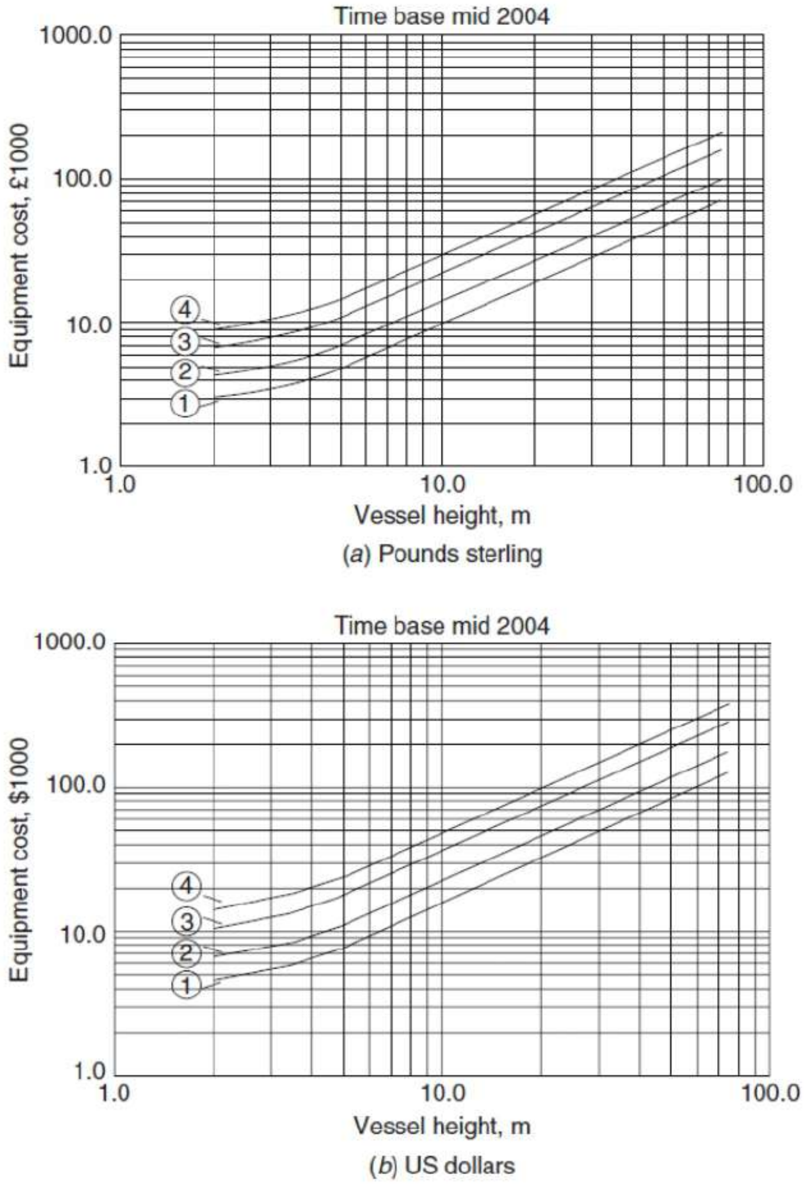
(*a*) Pounds sterling



(*b*) US dollars

Type	Area scale	Material
(1) Gasketed plate	m^2	Stainless steel
(2) Double pipe	$m^2 \times 10$	Carbon steel

Figure (3.4): Gasketed plate and frame and double pipe heat exchangers, Time base mid-2004



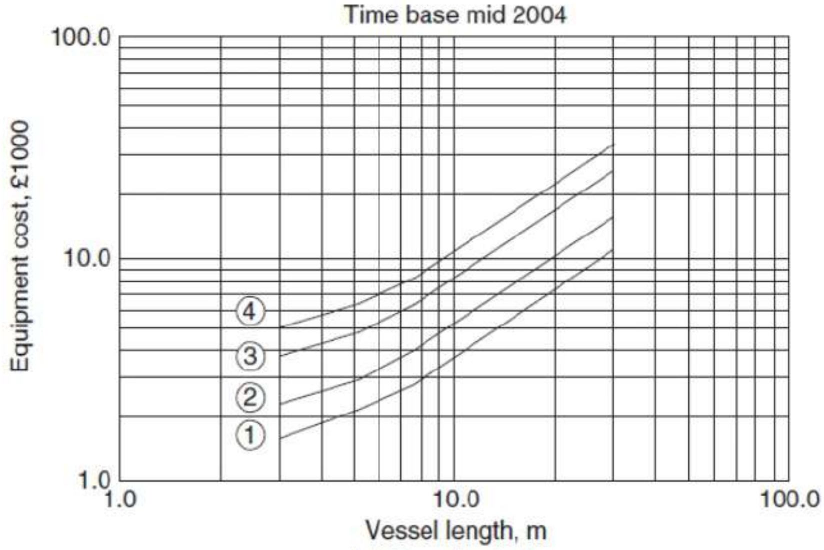
Diameter, m	
①—0.5	③—2.0
②—1.0	④—3.0

Material factors	
C.S.	× 1.0
S.S.	× 2.0
Monel	× 3.4
S.S. clad	× 1.5
Monel clad	× 2.1

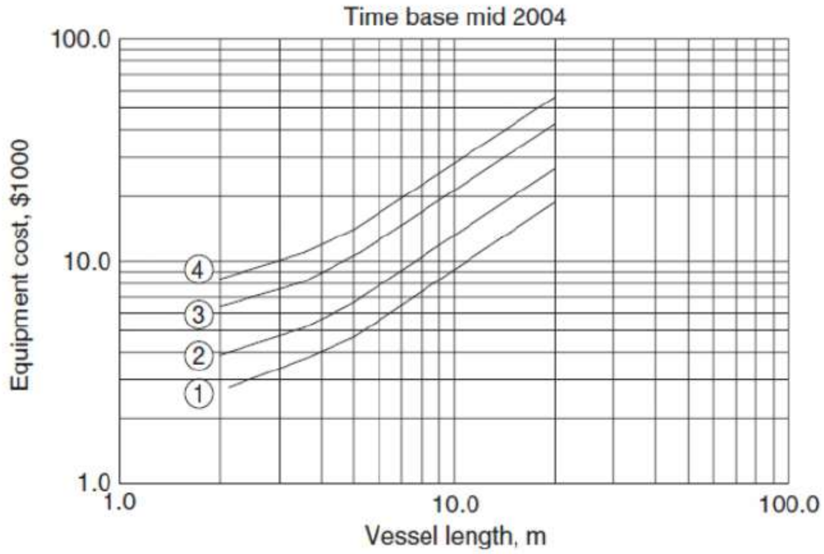
Pressure factors	
1–5 bar	× 1.0
5–10	× 1.1
10–20	× 1.2
20–30	× 1.4
30–40	× 1.6
40–50	× 1.8
50–60	× 2.2

Temperature up to 300°C

Figure (3.5): Vertical pressure vessels. Time base mid-2004.
Purchased cost = (bare cost from figure) × Material factor × Pressure factor



(*a*) Pounds sterling



(*b*) US dollars

Diameter, m	
①—0.5	③—2.0
②—1.0	④—3.0

Material factors	
C.S.	× 1.0
S.S.	× 2.0
Monel	× 3.4
S.S. clad	× 1.5
Monel clad	× 2.1

Pressure factors	
1–5 bar	× 1.0
5–10	× 1.1
10–20	× 1.2
20–30	× 1.4
30–40	× 1.6
40–50	× 1.8
50–60	× 2.2

Temperature up to 300°C

Figure (3.6): Horizontal pressure vessels. Time base mid-2004.
Purchase cost = (bare cost from figure) × Material factor × Pressure factor

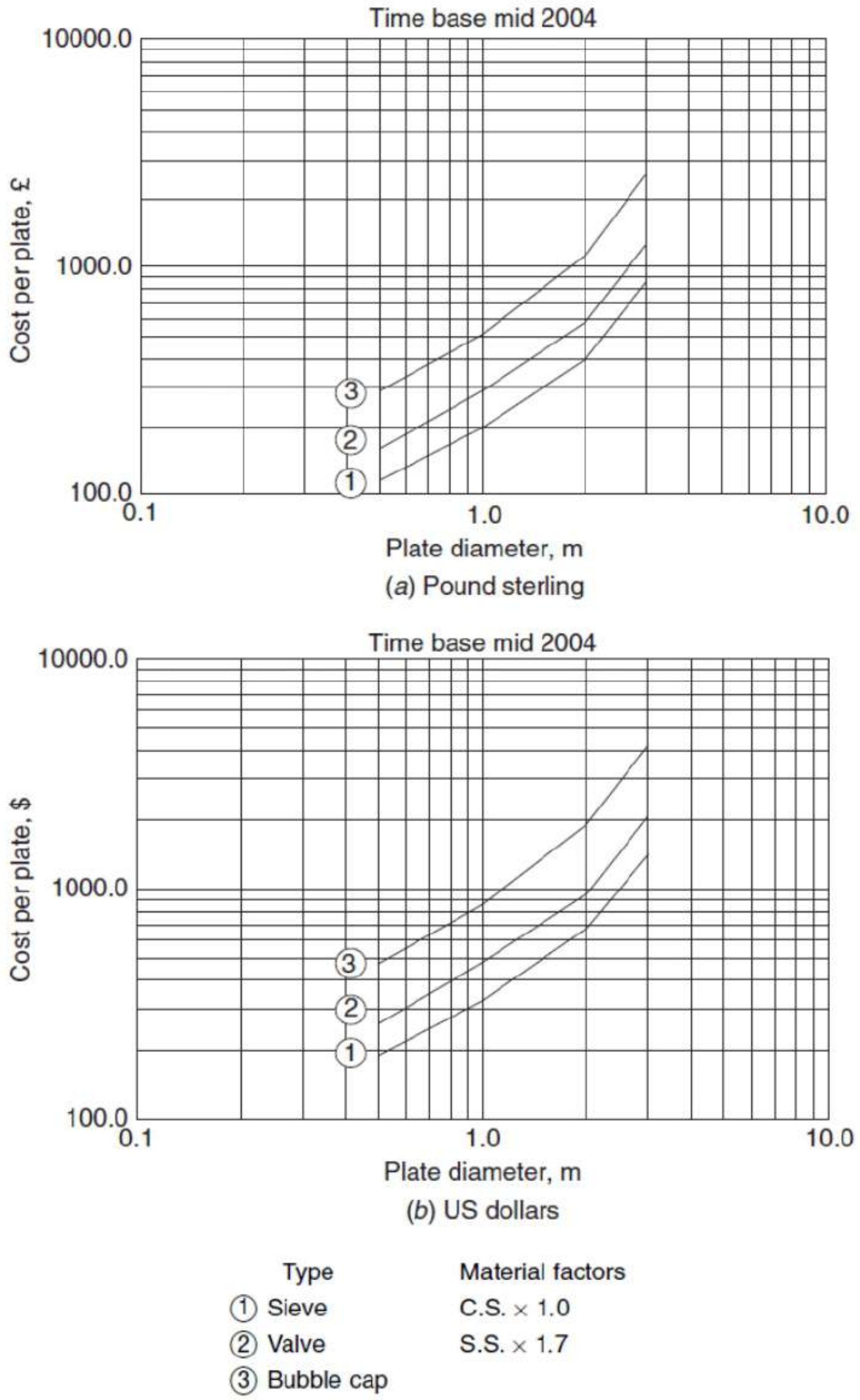


Figure (3.7): Column plates. Time base mid-2004 (for column costs see Figure 3)
Installed cost = (cost from figure) × Material factor

Table (3.2) : Cost of equipment. Cost basis mid 2004

Equipment	Size unit, S	Size range	Constant C,£	C,$	Index n	Comment
Agitators						
Propeller	driver	5–75	1200	1900	0.5	
Turbine	power, kW		1800	3000	0.5	
Boilers						
Packaged						oil or gas fired
up to 10 bar	kg/h steam	$(5-50)\times10^3$	70	120	0.8	
10 to 60 bar			60	100	0.8	
Centrifuges						
Horizontal basket	dia., m	0.5–1.0	35,000	58,000	1.3	carbon steel
Vertical basket			35,000	58,000	1.0	×1.7 for ss
Compressors						
Centrifugal	driver power, kW	20–500	1160	1920	0.8	electric, max. press.
Reciprocating			1600	2700	0.8	50 bar
Conveyors						
Belt	length, m	2–40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20–200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m^2	5–30	21,000	35,000	0.45	direct
Pan		2–10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m^2	10–100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m^2	5–50	5400	8800	0.6	cast iron
Vacuum drum		1–10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	10^3-10^4	330	540	0.77	carbon steel
Box		10^3-10^5	340	560	0.77	×2.0 ss
Reactors						
Jacketed,	capacity, m^3	3–30	9300	15,000	0.40	carbon steel
agitated			18,500	31,000	0.45	glass lined
Tanks						
Process	capacity, m^3					
vertical		1–50	1450	2400	0.6	atmos. press.
horizontal		10–100	1750	2900	0.6	carbon steel
Storage						
floating roof		50–8000	2500	4350	0.55	×2 for
cone roof		50–8000	1400	2300	0.55	stainless

Table (3.3): Cost of column packing. Cost basis mid 2004

	Cost	£/m^3 ($/$m^3$)	
Size, mm	25	38	50
Saddles, stoneware	840 (1400)	620 (1020)	580 (960)
Pall rings, polypropylene	650 (1080)	400 (650)	250 (400)
Pall rings, stainless steel	1500 (2500)	1500 (2500)	830 (1360)

To use Table (3.2), substitute the values given for the particular type of equipment into the equation:

$$Ce = C\,S^n$$

where

Ce = equipment cost.

S = characteristic size parameter, in the units given in Table 2.

C = cost constant from Table (3.2).

n = index for that type of equipment.

3.8 Estimation of Total Product Cost

The cost of producing a chemical product will include the items listed below. They are divided into two groups.

Fixed costs

1. Maintenance.
2. Operating labor.
3. Laboratory costs.
4. Supervision.
5. Plant overheads.
6. Capital charges.
7. Insurance.
8. Local taxes.
9. License fees and royalty payments.

Variable costs

1. Raw materials. The prices of a selected range of chemicals are given in Table (3.4).
2. plant supplies.
3. Utilities (Services),Table (3.5), can be used to make preliminary estimates.
4. Shipping and packaging. Depend on the nature of the product.

Table (3.4): Raw material and product costs

Chemical, and state	Cost unit	Cost £/unit	Cost $/unit
Acetaldehyde, 99%	kg	0.53	0.48
Acetic acid	kg	0.60	1.10
Acetic anhydride	kg	0.70	1.15
Acetone	kg	0.63	1.03
Acrylonitrile	kg	1.20	1.90
Ally alcohol	kg	1.40	2.30
Ammonia, anhydrous	t	180	280
Ammonium nitrate, bulk	t	100	170
Ammonium sulphate, bulk	t	90	150
Amyl alcohol, mixed isomers	kg	0.67	1.20
Aniline	kg	0.52	0.84
Benzaldehyde, drums	kg	1.95	3.21
Benzene	kg	0.20	0.33
Benzoic acid, drums	kg	2.20	3.60
Butene-1	kg	0.30	0.40
n-Butyl alcohol	kg	0.75	1.30
n-Butyl ether, drums	kg	1.95	3.20
Calcium carbide, bulk	t	320	530
Calcium carbonate, bulk, coarse	t	105	145
Calcium chloride, bulk	t	200	275
Calcium hydroxide (lime), bulk	t	55	90
Carbon disulphide	t	370	500
Carbon tetrachloride, drums	kg	0.50	0.83
Chlorine	t	140	200
Chloroform	kg	0.45	0.70
Cupric chloride, anhydrous	kg	3.30	5.5
Dichlorobenzene	kg	0.95	1.54
Diethanolamine	kg	1.20	1.70
Ethanol, 90%	kg	4.20	6.50
Ethyl ether	kg	0.80	1.35
Ethylene, contract	kg	0.46	0.70
Ethylene glycol	kg	0.56	0.83
Ethylene oxide	kg	0.60	0.90
Formaldehyde, 37% w/w	kg	0.31	0.46
Formic acid, 94% w/w, drums	kg	0.63	1.05
Glycerine, 99.7%	kg	1.30	1.70
Heptane	kg	0.30	0.40
Hexane	kg	0.20	0.33
Hydrochloric acid, anhyd.	kg	1.00	1.70
Hydrochloric acid, 30% w/w	t	60	90
Hydrogen fluoride, anhydrous	kg	0.90	1.40
Hydrogen peroxide, 50% w/w	kg	0.50	0.80
Isobutanol, alcohol	kg	0.75	1.1
Isopropanol alcohol	kg	0.73	1.12
Maleic anhydride, drums	kg	1.80	2.90
Methanol	kg	0.63	1.00
Methyl ethyl ketone	kg	0.64	1.06
Monoethanolamine	kg	1.02	1.54
Methylstyrene	kg	0.70	1.15
Nitric acid, 50% w/w	t	130	220
98% w/w	t	220	370
Nitrobenzene	kg	0.47	0.78

Chemical, and state	Cost unit	Cost £/unit	Cost $/unit
Oxalic acid, sacks	kg	0.58	0.96
Phenol	kg	0.90	1.53
Phosgene, cyl.	kg	1.09	1.62
Phosphoric acid 75% w/w	kg	0.47	0.78
Potassium bicarbonate, sacks	kg	0.45	0.75
Potassium carbonate, sacks	kg	0.56	0.92
Potassium chloride	t	70	110
Potassium chromate, sacks	kg	0.80	1.30
Potassium hydroxide	kg	2.00	3.70
Potassium nitrate, bulk	t	350	570
Propylene	kg	0.43	0.64
Propylene oxide	kg	1.00	1.60
n-Propanol	kg	0.93	1.438
Sodium carbonate, sacks	kg	0.35	0.58
Sodium chloride, drums	kg	0.40	0.65
Sodium hydroxide, drums	kg	1.60	2.60
Sodium sulphate, bulk	t	72	120
Sodium thiosulphate	kg	0.38	0.57
Sulphur, crude, 99.5%, sacks	t	85	140
Sulphuric acid, 98% w/w	t	40	65
Titanium dioxide, sacks	kg	1.50	2.50
Toluene	kg	0.32	0.47
Toluene diisocyanate	kg	2.20	3.20
Trichloroethane	kg	0.56	0.94
Trichloroethylene	kg	0.84	1.40
Urea, 46% nitrogen, bulk	t	120	160
Vinyl acetate	kg	0.65	1.08
Vinyl chloride	kg	0.44	0.66
Xylenes	kg	0.29	0.43

Anhyd. = anhydrous, cyl. = cylinder, refin. = refined

Table (3.5) Cost of utilities, typical figures mid-2004

Utility	UK	USA
Mains water (process water)	60 p/t	50 c/t
Natural gas	0.4 p/MJ	0.7 c/MJ
Electricity	1.0 p/MJ	1.5 c/MJ
Fuel oil	65 £/t	100 $/t
Cooling water (cooling towers)	1.5 p/t	1 c/t
Chilled water	5 p/t	8 c/t
Demineralised water	90 p/t	90 c/t
Steam (from direct fired boilers)	7 £/t	12 $/t
Compressed air (9 bar)	0.4 p/m^3 (Stp)	0.6 c/m^3
Instrument air (9 bar) (dry)	0.6 p/m^3 (Stp)	1 c/m^3
Refrigeration	1.0 p/MJ	1.5 c/MJ
Nitrogen	6 p/m^3 (Stp)	8 c/m^3

Note: £1 = 100p, 1$ = 100c, 1 t = 1000 kg = 2200 ib, stp = 1 atm, 0°C

Table 6.6. Summary of production costs

Variable costs	*Typical values*
1. Raw materials	from flow-sheets
2. Miscellaneous materials	10 per cent of item (5)
3. Utilities	from flow-sheet
4. Shipping and packaging	usually negligible
Sub-total A	....
Fixed costs	
5. Maintenance	5–10 per cent of fixed capital
6. Operating labour	from manning estimates
7. Laboratory costs	20–23 per cent of 6
8. Supervision	20 per cent of item (6)
9. Plant overheads	50 per cent of item (6)
10. Capital charges	10 per cent of the fixed capital
11. Insurance	1 per cent of the fixed capital
12. Local taxes	2 per cent of the fixed capital
13. Royalties	1 per cent of the fixed capital
Sub-total B	....
Direct production costs A + B	....
13. Sales expense	20–30 per cent of the direct production cost
14. General overheads	
15. Research and development	
Sub-total C	....
Annual production cost = A + B + C =	....

$$\text{Production cost £/kg} = \frac{\text{Annual production cost}}{\text{Annual production rate}}$$

Example 3.2:

Preliminary design work has been done on a process to recover a valuable product from an effluent gas stream. The gas will be scrubbed with a solvent in a packed column; the recovered product and solvent separated by distillation; and the solvent cooled and recycled. The major items of equipment that will be required are detailed below.

1. Absorption column : diameter 1 m, vessel overall height 15 m, packed height 12m, packing 25mm ceramic intalox saddles, vessel carbon steel, operating pressure 5 bar.

2. Recovery column : diameter 1 m, vessel overall height 20 m, 35 sieve plates, vessel and plates stainless steel, operating pressure 1 bar.

3. Reboiler : forced convection type, fixed tube sheets, area 18.6 m^2, carbon steel shell, stainless-steel tubes, operating pressure 1 bar.

4. Condenser : fixed tube sheets, area 25.3 m^2, carbon steel shell and tubes, operating pressure 1 bar.
5. Recycle solvent cooler : U-tubes, area 10.1 m^2, carbon steel shell and tubes, operating pressure 5 bar.
6. Solvent storage tanks : cone roof, capacity 100 m^3, carbon steel.

Solution

Absorption column

Bare vessel cost (Figure 3.5a) £21,000; material factor 1.0, pressure factor 1.1

Vessel cost = 21,000 × 1.0 × 1.1 = £23,000

Packing cost (Table 3.3) £840/m^3

Volume of packing = $(\pi/4)\ d^2 \times L = (\pi/4) \times 12 = 9.4\ m^3$

Cost of column packing = 9.4 × 840 = £7896

Total cost of column 23,000 + 7896 = 30,896 say £31,000

Recovery column

Bare vessel cost (Figure 3.5a) £26,000; material factor 2.0, pressure factor 1.0

Vessel cost 26,000 × 2.0 × 1.0 = £52,000

Cost of a plate (Figure 3.7a), material factor 1.7 = 200 × 1.7 = £340

Total cost of plates = 35 × 340 = £11,900

Total cost of column = 52,000 + 11,900 = 63,900 say £64,000

Reboiler

Bare cost (Figure 3.3a) £11,000; type factor 0.8, pressure factor 1.0

Purchased cost = 11,000 × 0.8 × 1.0 = £8800

Condenser

Bare cost (Figure 3.3a) £8500; type factor 0.8, pressure factor 1.0

Purchased cost = 8500 × 0.8 × 1.0 = £6800

Cooler

Bare cost (Figure 3.3a) £4300; type factor 0.85, pressure factor 1.0

Purchased cost = 4300 × 0.85 × 1.0 = £3700

Solvent tank

Purchase cost (Table 3.2) = $1400 \times (100)^{0.55} = £17625$

Total purchase cost of major equipment items (PCE)

Absorption column	31,000
Recovery column	64,000
Reboiler	8,800
Condenser	6,800
Cooler	3,700
Solvent tank	17,700
Total	**£132,000**

4. Finished product inventories.
5. Funds to cover outstanding accounts from customers.

Most of the working capital is recovered at the end of the project. The total investment needed for a project is the sum of the fixed and working capital.

3.4 Cost Inflation

All cost-estimating methods use historical data, and are themselves forecasts of future costs. Some method has to be used to update old cost data for use in estimating at the design stage, and to forecast the future construction cost of the plant.

$$\text{Cost in year A} = \text{Cost in year B} \times \frac{\text{Cost index in year A}}{\text{Cost index in year B}}$$

Many methods had been adopted for estimating the values of the cost index. One is the process engineering index.

To estimate the future cost of a plant some prediction has to be made of the future annual rate of inflation. This can be based on the extrapolation of one of the published indices, tempered by the engineer's own assessment of what the future may hold.

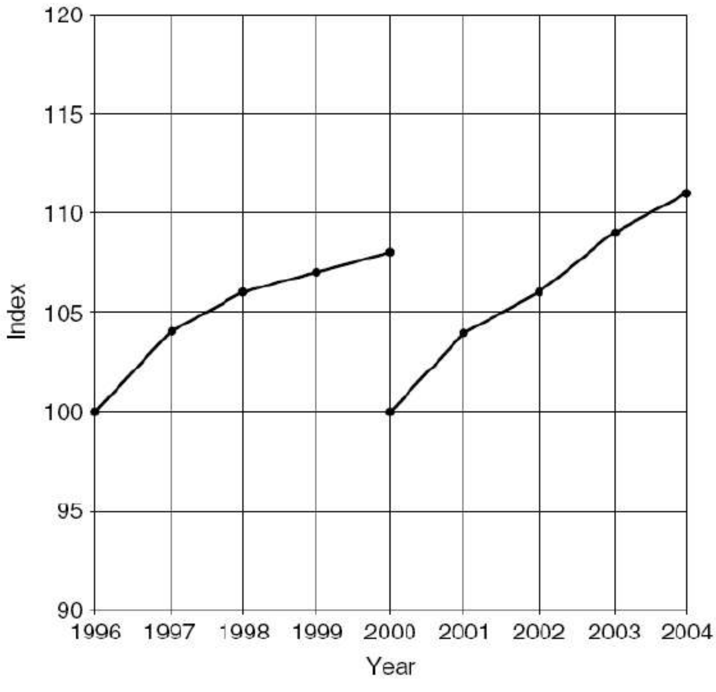


Fig. (3.1) Process Engineering Index

A composite index for the United States process plant industry is published monthly in the journal *Chemical Engineering*, the CPE plant cost index. The longer the period over which the correlation is made the more unreliable the estimate. Between 1970 and 1990 prices rose dramatically. Since then the annual rise has slowed down and is now averaging around 2 - 3 % per year.

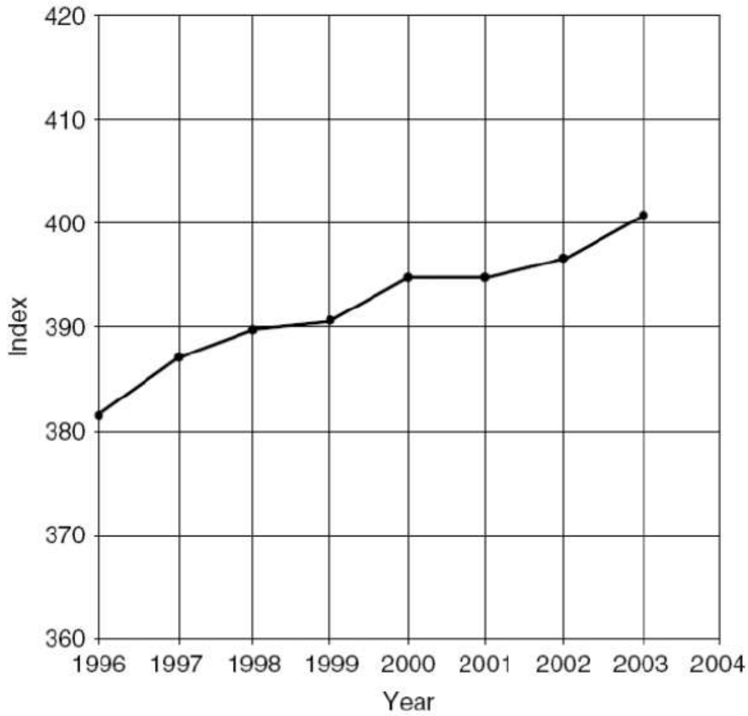


Fig. (3.2) CPE index

Example 3.1:

The purchased cost of a shell and tube heat exchanger, carbon shell, stainless steel tubes, heat transfer area 500 m^2, was £7600 in January 1998; estimate the cost in January 2006. Use the *Process Engineering* plant index.

Solution

From Process Engineering Index figure (3.1) :

Index in 1998 = 106

2000 = 108, 100 (change of base)

2004 = 111

So, estimated cost in January 2000 = 7600 × 108/106 = £7743

and in 2004 = 7743 × 111/100 = £8595

From Process Engineering Index figure (3.1):

The average increase in costs = (111-100)/4= 2.75 per year.

Use this value to predict the exchanger cost in 2006.
The cost index in 2006 = 2 × 2.75 + 111 = 116.5
Cost in 2006 = 8595 × 116.5/111 = £9021 say £9000.

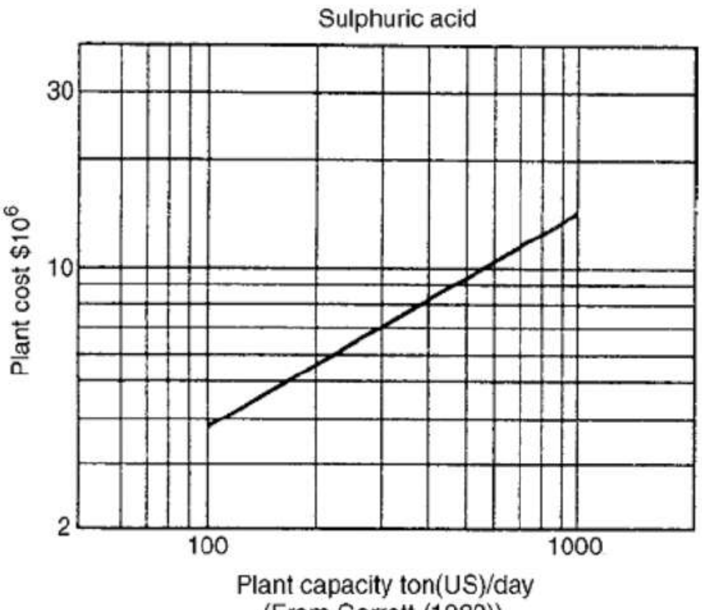
3.5 Historical Costs

An approximate estimate of the capital cost of a project can be obtained from knowledge of the cost of earlier projects using the same manufacturing process. The capital cost of a project is related to capacity by the equation

$$C_2 = C_1 \left(\frac{S_2}{S_1}\right)^n$$

Where:
C_2 = capital cost of the project with capacity, S_2
C_1 = capital cost of the project with capacity, S_1

The value of the index n is traditionally taken as 0.6; the well-known six-tenths rule. This value can be used to get a rough estimate of the capital cost if there are not sufficient data available to calculate the index for the particular process. This equation is only an approximation, and if sufficient data are available the relationship is best represented on a log-log plot. Garrett (1989) has published capital cost-plant capacity curves for over 250 processes.



(From Garrett (1989))

3.6 Estimating Equipment Costs by Scaling

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Table (3.1): Typical exponents for equipment cost vs. capacity

Equipment	Siie range	Exponent
Blender, double cone rotary, c.s.	SO-250 ft^3	0.49
Blower, centrifugal	10^3–10^4 ft^3/min	0.59
Centrifuge, solid bowl, c.s.	10–10^2 hp drive	0.67
Crystallizer, vacuum batch, c.s.	500-7000 ft^3	0.37
Compressor, reciprocating, air cooled, two-stage, 150 psi discharge	10–400 ft^3/min	0.69
Compressor, rotary, single-stage, sliding vane, 150 psi discharge	10^2–10^3 ft^3/min	0.79
Dryer, drum, single vacuum	10–10^2 ft^2	0.76
Dryer, drum, single atmospheric	10–10^2 ft^2	0.40
Evaporator (installed), horizontal tank	10^2–10^4 ft^2	0.54
Fan, centrifugal	10^3–10^4 ft^3/min	0.44
Fan, centrifugal	2×10^4–7×10^4 ft^3/min	1.17
Heat exchanger, shell and tube, floating head, c.s.	100-400 ft^2	0.60
Heat exchanger, shell and tube, fixed sheet, c.s.	100-400 ft^2	0.44
Kettle, cast iron, jacketed	250-800 gal	0.27
Kettle, glass lined, jacketed	200-800 gal	0.31
Motor, squirrel cage, induction, 440 volts, explosion proof	5-20 hp	0.69
Motor, squirrel cage, induction, 440 volts, explosion proof	20-200 hp	0.99
Pump, reciprocating, horizontal cast iron (includes motor)	2–100 gpm	0.34
Pump, centrifugal, horizontal, cast steel (includes motor)	10^4–10^5 gpm $\times$ psi	0.33
Reactor, glass lined, jacketed (without drive)	50–600 gal	0.54
Reactor, s.s, 300 psi	10^2–10^3 gal	0.56
Separator, centrifugal, c.s.	50-250 ft^3	0.49
Tank, flat head, c.s.	10^2–10^4 gal	0.57
Tank, c.s., glass lined	10^2–10^3 gal	0.49
Tower, c.s.	10^3–2×10^6 lb	0.62
Tray, bubble cup, c.s.	3-10 ft diameter	1.20
Tray, sieve, c.s.	3-10 ft diameter	0.86

3.8 Plant Layout

plant layout a method of organizing the elements of a production process, in which similar processes and functions are grouped together. The basic objective of layout design is to facilitate a smooth flow of work, material, and information through the system. Plant Layout is the physical arrangement of equipment and facilities within a plant to improve productivity, safety and quality of products.

3.9 Plant Location and Site Selection

The location of the plant can have a crucial effect on the profitability of a project and the scope for future expansion. Many factors must be considered when selecting a suitable site. The principal factors to consider are

1. Location, with respect to the marketing area;
2. Raw material supply;
3. Transport facilities;
4. Availability of labor;
5. Availability of utilities: water, fuel, power;
6. Environmental impact, including effluent disposal;
7. Local community considerations;
8. Availability of suitable land
9. Climate;
10. Political and strategic considerations

3.9.1 Marketing Area

For materials that are produced in bulk quantities, such as cement, mineral acids, and fertilizers, where the cost of the product per metric ton is relatively low and the cost of transport is a significant fraction of the sales price, the plant should be located close to the primary market. This consideration is much less important for low-volume production and high-priced products, such as pharmaceuticals.

3.9.2 Raw Materials

The availability and price of suitable raw materials will often determine the site location. Plants that produce bulk chemicals are best located close to the source of the major raw material, as long as the costs of shipping product are not greater than the cost of shipping feed. For example, at the time of writing much of the new ethylene capacity that is being added worldwide is being built in the Middle East, close to supplies of cheap ethane from natural gas. Oil refineries, on the other hand, tend to be located close to major population centers, as an oil refinery produces many grades of fuel, which are expensive to ship separately.

3.9.3 Transport

The transport of materials and products to and from the plant can be an overriding consideration in site selection. If practicable, a site should be selected that is close to at least two major forms of transport: road, rail, waterway (canal or river), or a sea port. Road transport is increasingly used and is suitable for local distribution from a central warehouse. Rail transport is usually cheaper for the long-distance transport of bulk chemicals. Air transport is convenient and efficient for the movement of personnel and essential equipment and supplies.

3.9.4 Availability of Labor

Labor will be needed for construction of the plant and its operation. Skilled construction workers are usually brought in from outside the site area, but there should be an adequate pool of unskilled labor available locally, and labor suitable for training to operate the plant. Skilled craft workers such as electricians, welders, and pipe fitters will be needed for plant maintenance.

3.9.5 Utilities (Services)

The word utilities is used for the ancillary services needed in the operation of any production process. These services are normally supplied from a central site facility and include

1. Electricity;
2. Steam;
3. Cooling water;
4. Water for general use;
5. Demineralized water;
6. Compressed air;
7. Inert-gas supplies;
8. Refrigeration;
9. Effluent disposal facilities.

3.9.6 Environmental Impact and Effluent Disposal

All industrial processes produce waste products, and full consideration must be given to the difficulties and cost of their disposal. The disposal of toxic and harmful effluents will be covered by local regulations, and the appropriate authorities must be consulted during the initial site survey to determine the standards that must be met. An environmental impact assessment should be made for each new project or major modification or addition to an existing process.

3.9.7 Local Community Considerations

The proposed plant must fit in with and be acceptable to the local community. Full consideration must be given to the safe location of the plant so that it does not impose a significant additional risk to the local population. Plants should generally be sited so as not to be upwind of residential areas under the prevailing wind. Some communities welcome new plant construction as a source of new jobs and economic prosperity. More affluent communities generally do less to encourage the building of new manufacturing plants and in some cases may actively discourage chemical plant construction.

3.9.8 Land (Site Considerations)

Sufficient suitable land must be available for the proposed plant and for future expansion. The land should ideally be flat, well drained, and have suitable load-bearing characteristics. A full site evaluation should be made to determine the need for piling or other special foundations.

3.9.9 Climate

Adverse climatic conditions at a site will increase costs. Abnormally low temperatures require the provision of additional insulation and special heating for equipment and pipe runs. Stronger structures are needed at locations subject to high winds or earthquakes. Political and Strategic Considerations Capital grants, tax concessions, and other inducements are often given by governments to direct new investment to preferred locations, such as areas of high unemployment. The availability of such grants can be the overriding consideration in site selection.

3.10 Design of Plant Layout

The economic construction and efficient operation of a process unit will depend on how well the plant and equipment specified on the process flow-sheet is laid out. Figs. (3.8) and (3.9) show the flow-sheet and typical plant layout of crude oil refinery.

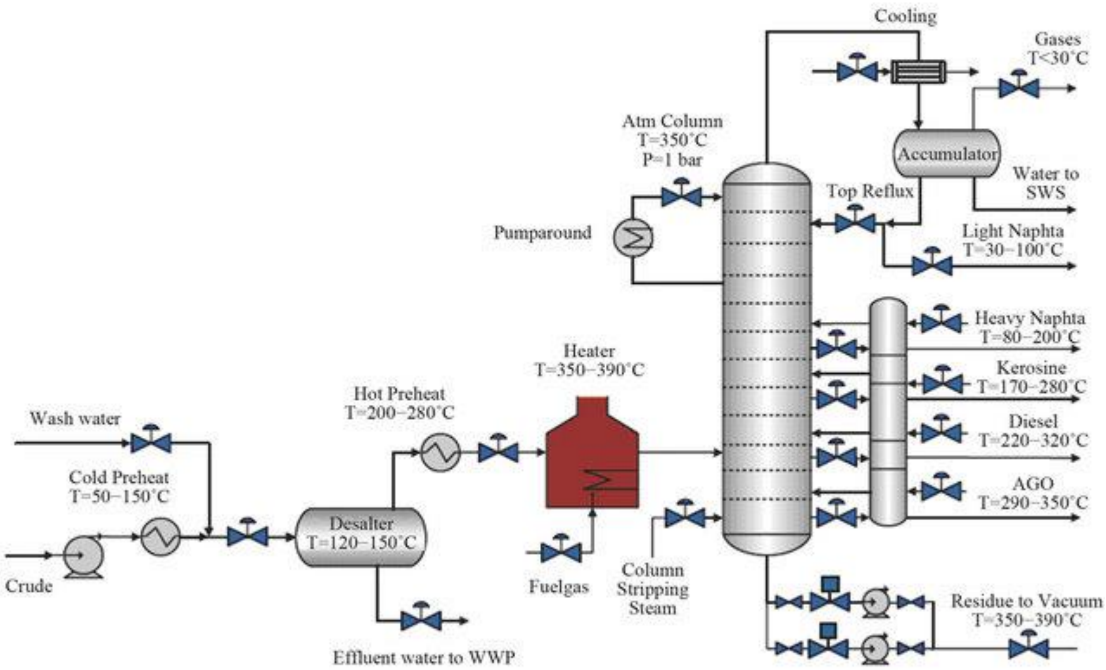


Figure (3.8) Flow-sheet of Crude Oil Refinery.

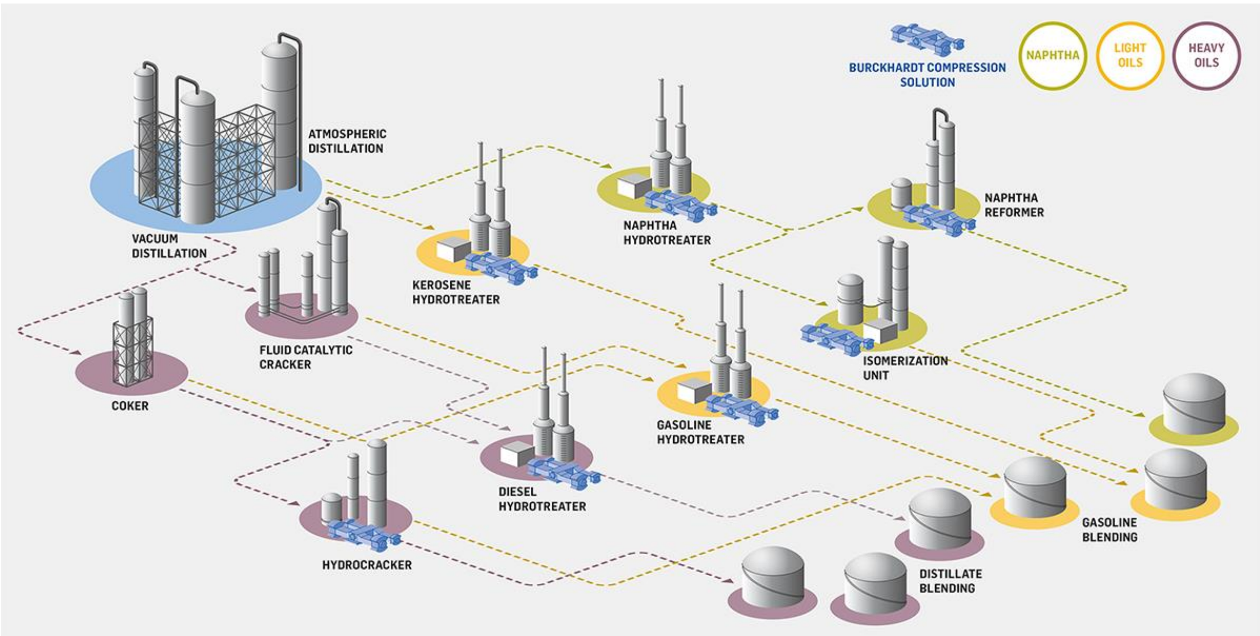


Figure (3.9) Typical plant layout of crude oil refinery.

The main storage areas should be placed between the loading and unloading facilities and the process units they serve as shown in Fig. (3.10) and Fig. (3.11).

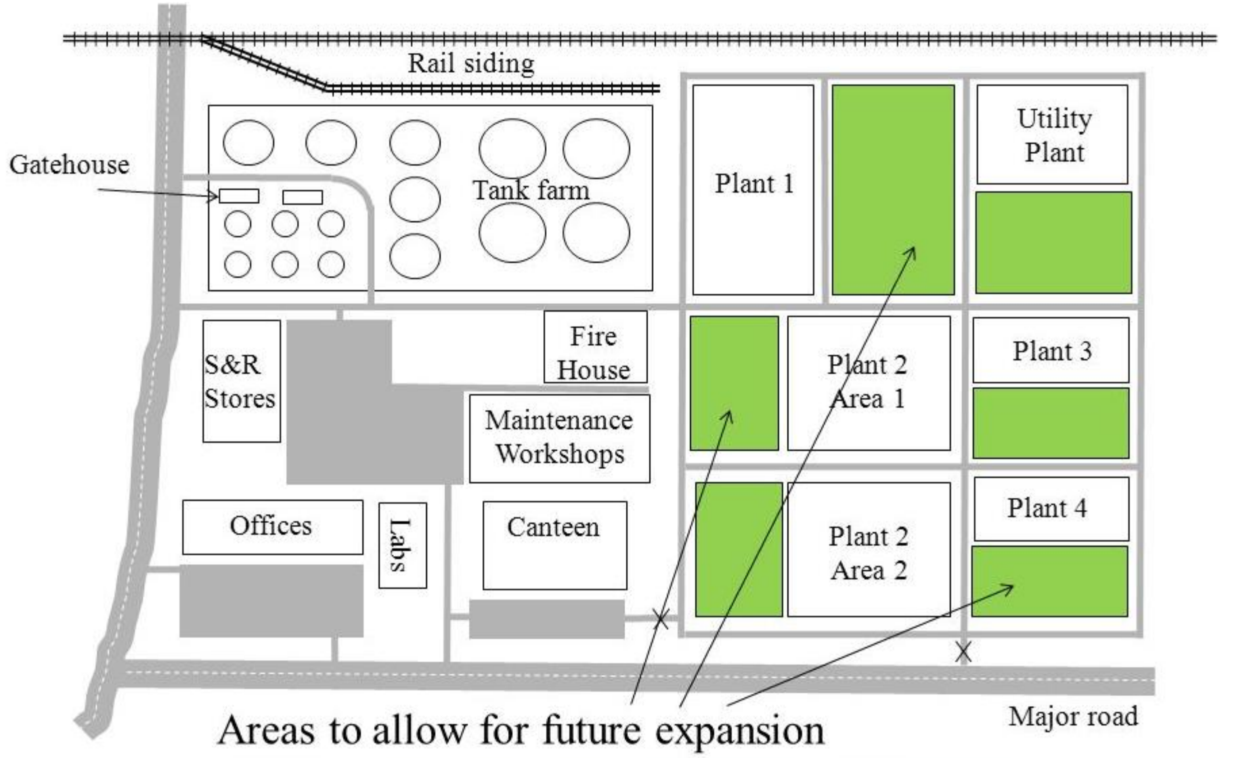


Figure (3.10) Typical site plant.

Figure (3.11) Typical site plant

Fig. (3.12) shows typical plant site for general plant.

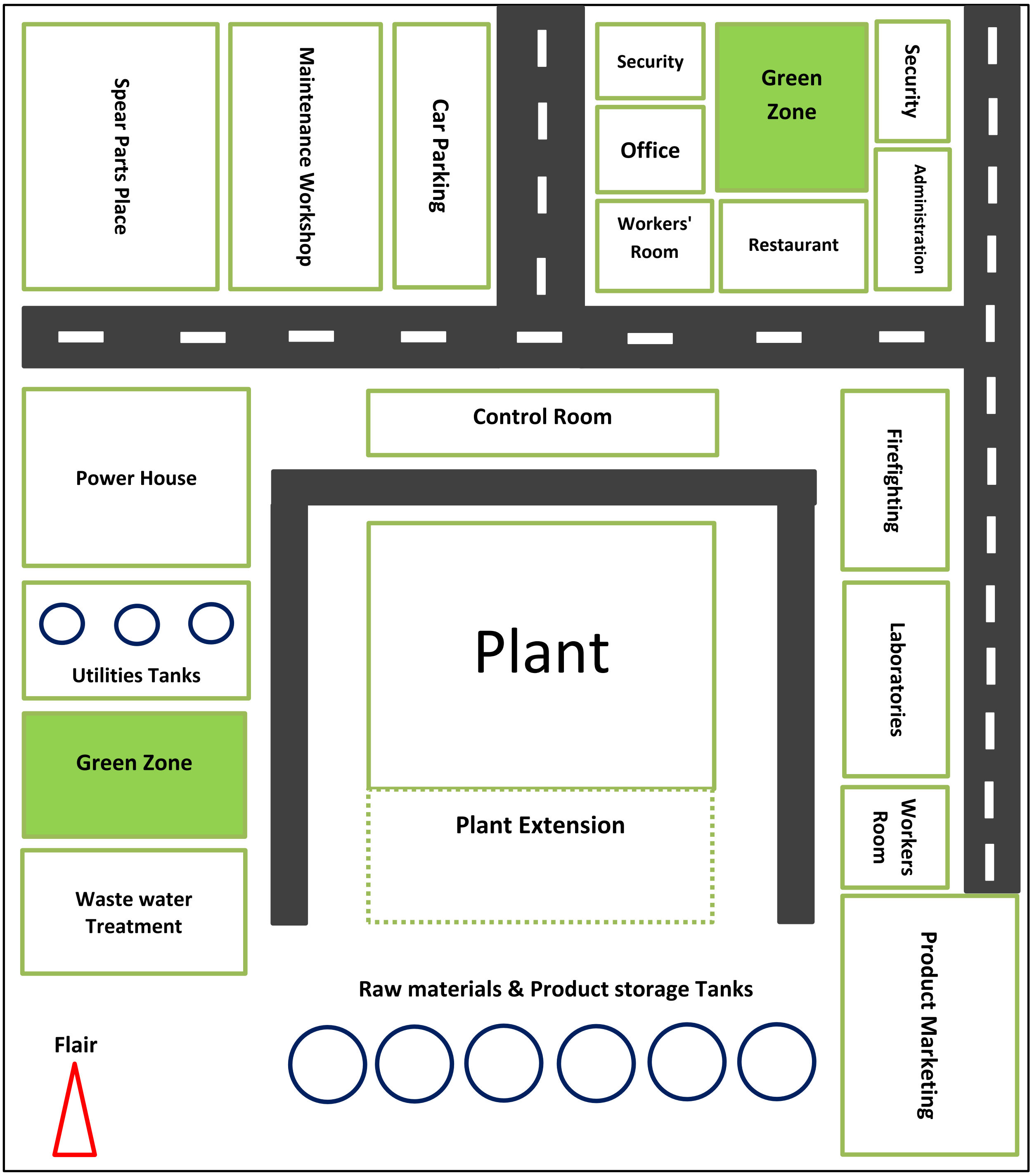


Figure (3.12) Typical plant site.

CHAPTER FOUR
EQUIPMENT SELECTION

4.1 Specification and Design

The chemical engineer's part in the design of equipment is usually limited to selecting and sizing the equipment. For example, in the design of a distillation column his work will typically be to determine the number of plates; the type and design of plate; diameter of the column; and the position of the inlet, outlet and instrument nozzles. This information would then be transmitted, in the form of sketches and specification sheets, to the specialist mechanical design group for detailed design.

4.2 Process Vessels

4.2.1 Liquid – Liquid Separation

Separation of two liquid phases, immiscible or partially miscible liquids, is a common requirement in the process industries. For example, in the unit operation of liquid-liquid extraction the liquid contacting step must be followed by a separation stage. The simplest form of equipment used to separate liquid phases is the gravity settling tank, the decanter.

4.2.1.1 Decanters (settlers)

Decanters are used to separate liquids where there is a sufficient difference in density between the liquids for the droplets to settle readily. In an operating decanter there will be three distinct zones or bands: clear heavy liquid; separating dispersed liquid; and clear light liquid.

Typical designs are shown in figures (4.1 a, b). The position of the interface can be controlled, with or without the use of instruments. The height of the take-off can be determined by making a pressure balance. Neglecting friction loss in the pipes, the pressure exerted by the combined height of the heavy and light liquid in the vessel must be balanced by the height of the heavy liquid in the take-off leg.

$$(z_1 - z_3)\rho_1 g + z_3 \rho_2 g = z_2 \rho_2 g$$

$$z_2 = \frac{(z_1 - z_3)\rho_1}{\rho_2} + z_3$$

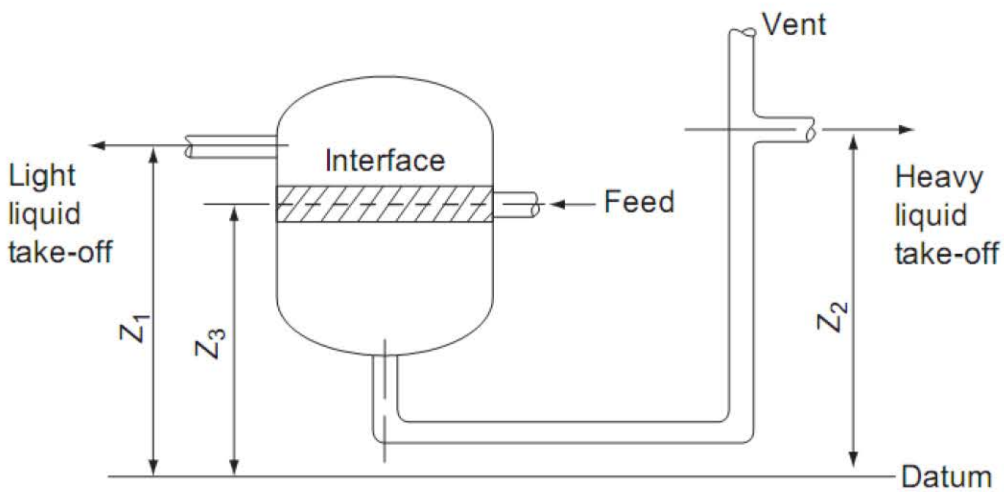


Figure (4.1a): Vertical decanter

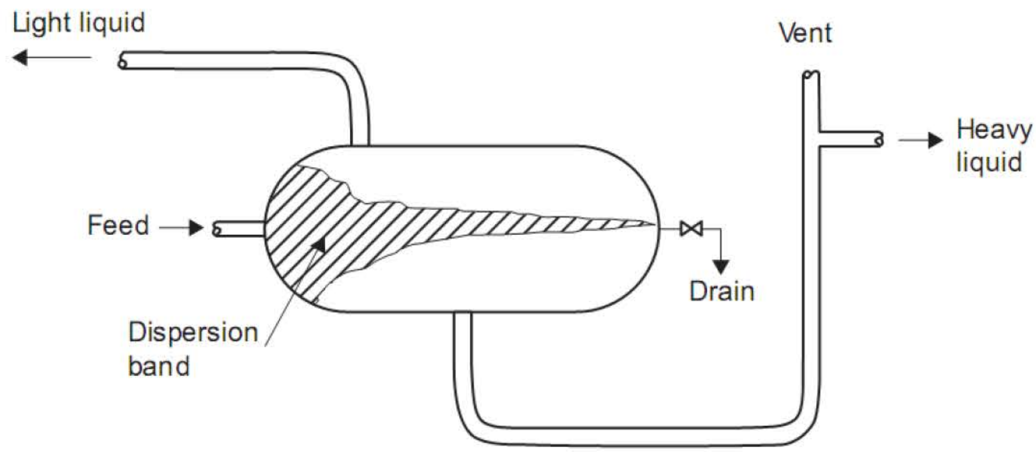


Figure (4.1b): Horizontal decanter

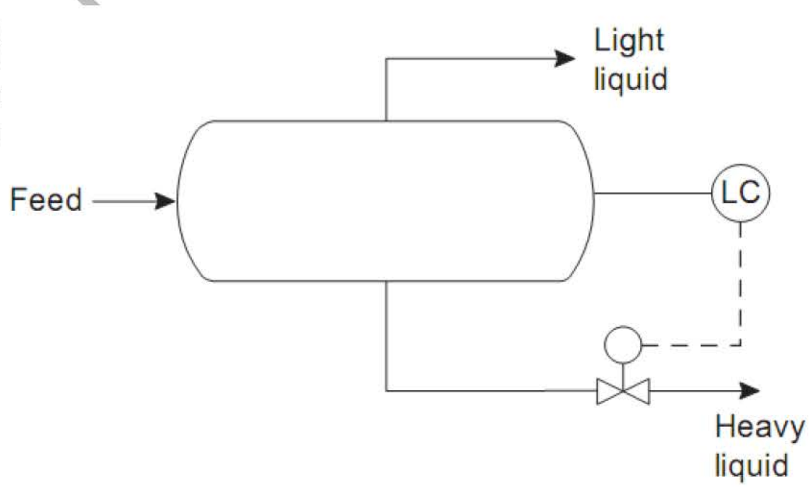


Figure (4.2): Automatic control, level controller detecting interface.

where

ρ_1 = density of the light liquid, kg/m^3.

ρ_2 =density of the heavy liquid, kg/m^3.

z_1 = height from datum to light liquid overflow, m.

z_2 = height from datum to heavy liquid overflow, m.

z_3 = height from datum to the interface, m.

The height of the liquid interface should be measured accurately when the liquid densities are close, when one component is present only in small quantities, or when the throughput is very small. Fig. (4.2) a typical scheme for the automatic control of the interface, using a level instrument that can detect the position of the interface

4.2.1.2 Decanter Design

A rough estimate of the decanter volume required can be made by taking a hold-up time of 2 to 10 min, which is usually sufficient where emulsions are not likely to form. The decanter vessel is sized on the basis that the velocity of the continuous phase must be less than settling velocity of the droplets of the dispersed phase. Plug flow is assumed, and the velocity of the continuous phase calculated using the area of the interface:

$$u_c = \frac{L_c}{A_i} < u_d$$

where

u_d = settling velocity of the dispersed phase droplets, m/s.

u_c = velocity of the continuous phase, m/s.

L_c = continuous phase volumetric flow rate, m^3/s.

A_i = area of the interface, m^2.

Stokes' law is used to determine the settling velocity of the droplets:

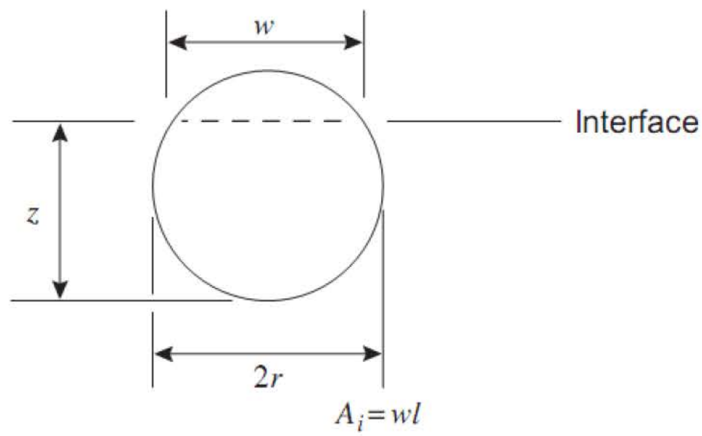
$$u_d = \frac{d_d^2 g(\rho_d - \rho_c)}{18\mu_c}$$

where

d_d = droplet diameter, m,

u_d = settling (terminal) velocity of the dispersed phase droplets, m/s.
ρ_c = density of the continuous phase, kg/m^3.
ρ_d = density of the dispersed phase, kg/m^3.
μ_c = viscosity of the continuous phase, N s/m^2.
g = gravitational acceleration, 9.81 m/s^2.

This equation is used to calculate the settling velocity with an assumed droplet size of 150 μm, which is well below the droplet sizes normally found in decanter feeds. If the calculated settling velocity is greater than 4×10^{-3} m/s, then a figure of 4×10^{-3} m/s is used. For a horizontal, cylindrical, decanter vessel, the interfacial area will depend on the position of the interface.



$$w = 2(2rz - z^2)^{1/2}$$

where
w = width of the interface, m.
z = height of the interface from the base of the vessel, m.
l = length of the cylinder, m/
r = radius of the cylinder, m.

For a vertical, cylindrical decanter: $\boldsymbol{A_i = \pi r^2}$

The depth of the dispersion band is a function of the liquid flow rate and the interfacial area. A value of 10 % of the decanter height is usually taken for design purposes.

Example 4.1

Design a decanter to separate a light oil from water.

The oil is the dispersed phase.

Oil, flow rate 1000 kg/h, density 900 kg/m^3, viscosity 3 mN s/m^2.

Water, flow rate 5000 kg/h, density 1000 kg/m^3, viscosity 1 mN s/m^2.

Solution

Take $d_d = 150$ µm

$$u_d = \frac{(150 \times 10^{-6})^2\, 9.81(900 - 1000)}{18 \times 1 \times 10^{-3}}$$

= - 0.0012 m/s, - 1.2 mm/s (rising)

$$L_c = \frac{5000}{1000} \times \frac{1}{3600} = 1.39 \times 10^{-3}\ \text{m}^3/\text{s}$$

$$u_c \ngtr u_d, \text{ and } u_c = \frac{L_c}{A_i}$$

$$A_i = \frac{1.39 \times 10^{-3}}{0.0012} = 1.16\ \text{m}^2$$

$$r = \sqrt{\frac{1.16}{\pi}} = 0.61\ \text{m}$$

$$\text{diameter} = \underline{\underline{1.2\ \text{m}}}$$

Take the height as twice the diameter, height = 2.4 m

Take the dispersion band as 10 % of the height = 0.24 m

Check the residence time of the droplets in the dispersion band

$$= \frac{0.24}{u_d} = \frac{0.24}{0.0012} = 200\ \text{s}\ (\sim 3\ \text{min})$$

This is satisfactory, a time of 2 to 5 min is normally recommended.

Piping arrangement

To minimize entrainment by the jet of liquid entering the vessel, the inlet velocity for a decanter should keep below 1 m/s.

$$\text{Flow rate} = \left[\frac{1000}{900} + \frac{5000}{1000}\right] \frac{1}{3600} = 1.7 \times 10^{-3}\ \text{m}^3/\text{s}$$

$$\text{Area of pipe} = \frac{1.7 \times 10^{-3}}{1} = 1.7 \times 10^{-3}\ \text{m}^2$$

$$\text{Pipe diameter} = \sqrt{\frac{1.7 \times 10^{-3} \times 4}{\pi}} = 0.047\ \text{m, say } \underline{\underline{50\ \text{mm}}}$$

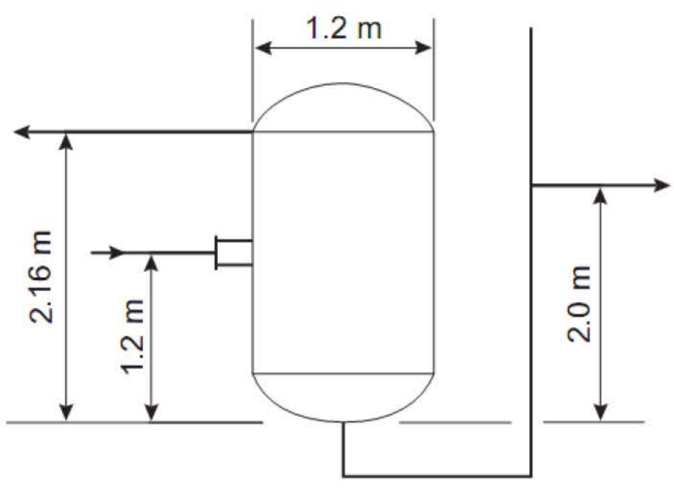
Take the position of the interface as half-way up the vessel and the light liquid off-take as at 90 % of the vessel height:

$$z_1 = 0.9 \times 2.4 = 2.16\ \text{m}$$

$$z_3 = 0.5 \times 2.4 = 1.2\ \text{m}$$

$$z_2 = \frac{(2.16 - 1.2)}{1000} \times 900 + 1.2 = \underline{\underline{2.06\ \text{m}}},\ \ \text{say } \underline{\underline{2.0\ \text{m}}}$$

Proposed design



Drain valves should be fitted at the interface so that any tendency for an emulsion to form can be checked; and the emulsion accumulating at the interface drained off periodically as necessary.

4.2.2 Gas – Liquid Separators

The separation of liquid droplets from gas phase streams by gravity settling in a vertical or horizontal separating vessel which called knockout drums. Knitted mesh demisting pads are frequently used to improve the performance of separating vessels where the droplets are likely to be small, down to 1 μm, and where high separating efficiencies are required. Demister pads are available in a wide range of materials such as metals and plastics. Separating efficiencies above 99% can be obtained with low pressure drop.

The following equation can be used to estimate the settling velocity of the liquid droplets, for the design of separating vessels.

$$u_t = 0.07\left[\frac{(\rho_l - \rho_v)}{\rho_v}\right]^{1/2}$$

where u_t = settling velocity, m/s.
ρ_l = liquid density, kg/m^3.
ρ_v = vapor density, kg/m^3.

If a demister pad is not used, the value of u_t obtained from equation above should be multiplied by a factor of 0.15 to provide a margin of safety and to allow for flow surges. This factor may use in case of vapors.

4.2.2.1 Vertical Knockout Drums

The layout and typical proportions of a vertical liquid gas separator are shown in Fig. (4.3). The diameter of the vessel must be large enough to slow the gas down to below the velocity at which the particles will settle out. So the minimum allowable diameter will be given by:

$$D_v = \sqrt{\frac{4\,V_v}{\pi\,u_s}}$$

where D_v = minimum vessel diameter, m.
V_v = gas, or vapor volumetric flow-rate, m^3/s.
$u_s = u_t$, if a demister pad is used, and 0.15 u_t for a separator without a demister pad, m/s.

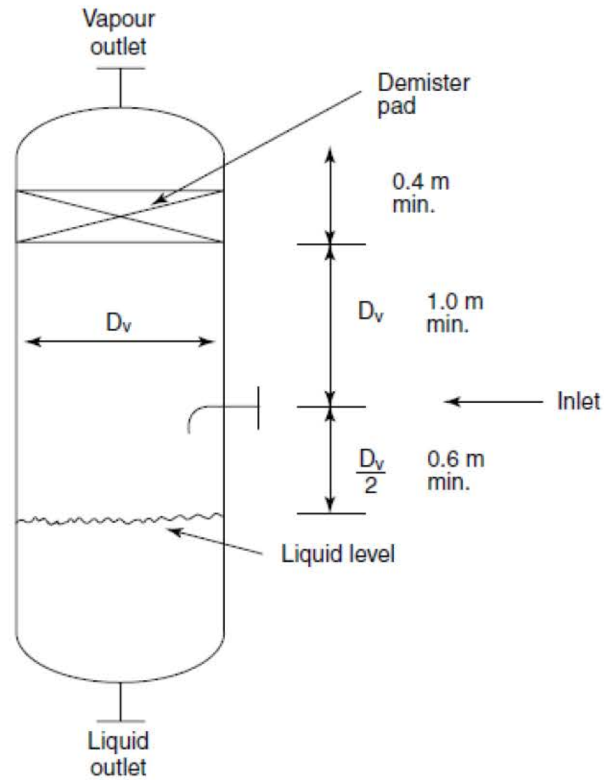


Figure (4.3): Vertical liquid-vapor separator

The height of the vessel outlet above the gas inlet should be sufficient to allow for disengagement of the liquid drops. A height equal to the diameter of the vessel should be used or 1 m if the diameter is less than 1m.

The liquid level will depend on the hold-up time necessary for smooth operation and control; typically 10 minutes would be allowed.

Example 4.2

Make a preliminary design for a separator to separate a mixture of steam and water; flow-rates: steam 2000 kg/h, water 1000 kg/h; operating pressure 4 bar.

Solution

From steam tables, at 4 bar: saturation temperature 143.6°C

liquid density 926.4 kg/m^3, vapor density 2.16 kg/m^3.

$u_t = 0.07[(926.4 - 2.16) /2.16]^{1/2} = 1.45$ m/s

As the separation of condensate from steam is unlikely to be critical, a demister pad will not be specified.

So, $u_s = 0.15 \times 1.45 = 0.218$ m/s

$$\text{Vapour volumetric flow-rate} = \frac{2000}{3600 \times 2.16} = 0.257\ \text{m}^3/\text{s}$$

$$D_v = \sqrt{[(4 \times 0.257)/(\pi \times 0.218)]} = 1.23\ \text{m, round to } 1.25\ \text{m (4 ft).}$$

$$\text{Liquid volumetric flow-rate} = \frac{1000}{3600 \times 926.14} = 3.0 \times 10^{-4}\ \text{m}^3/\text{s}$$

Allow a minimum of 10 minutes hold-up.

$$\text{Volume held in vessel} = 3.0 \times 10^{-4} \times (10 \times 60) = 0.18\ \text{m}^3$$

$$\text{Liquid depth required, } h_v = \frac{\text{volume held-up}}{\text{vessel cross-sectional area}}$$

$$= \frac{0.18}{(\pi \times 1.25^2/4)} = 0.15\ \text{m}$$

Increase to 0.3 m to allow space for positioning the level controller.

4.2.2.2 Horizontal separators

The layout of a typical horizontal separator is shown in the Fig. (4.4). A horizontal separator would be selected when a long liquid hold-up time is required.

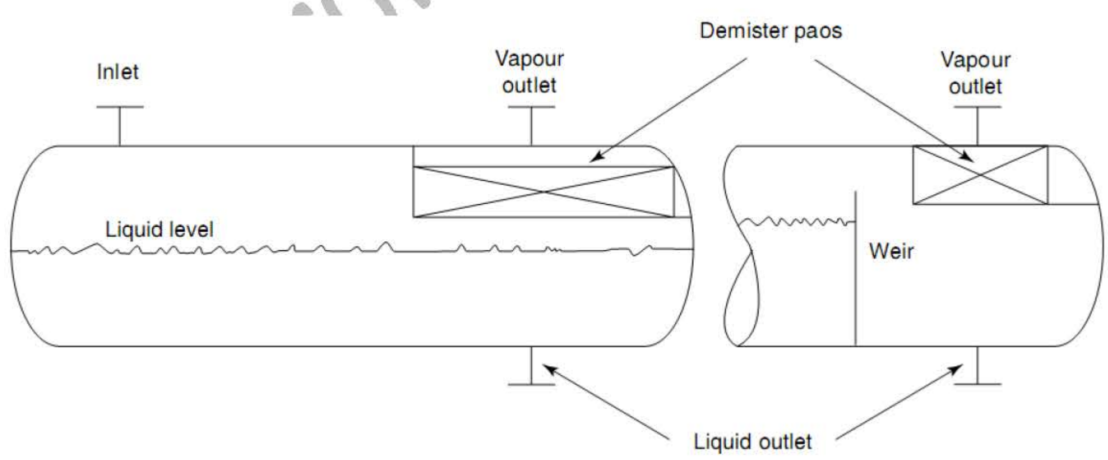


Figure (4.4): Horizontal liquid vapor separator

In the design of a horizontal separator the vessel diameter cannot be determined independently of its length, unlike for a vertical separator. The diameter and length, and the liquid level, must be chosen to give sufficient vapor residence

As the separation of condensate from steam is unlikely to be critical, a demister pad will not be specified.

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Increase to 0.3 m to allow space for positioning the level controller.

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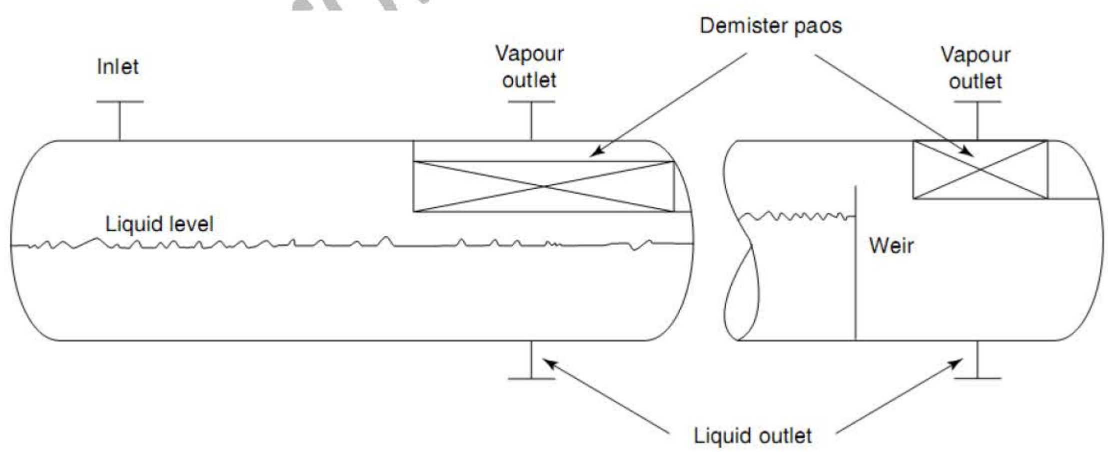


Figure (4.4): Horizontal liquid vapor separator

In the design of a horizontal separator the vessel diameter cannot be determined independently of its length, unlike for a vertical separator. The diameter and length, and the liquid level, must be chosen to give sufficient vapor residence time for the liquid droplets to settle out, and for the required liquid hold-up time to be met. The most economical length to diameter ratio will depend on the operating pressure. As a general guide the following values can be used:

Operating pressure, bar	Length: diameter, L_v/D_v
0–20	3
20–35	4
>35	5

The liquid height can set at the half of the vessel diameter and the height above the liquid level, h_v , can be calculated from the following equation:

$$h_v = D_v/2 \quad \text{and} \quad f_v = 0.5$$

where f_v is the fraction of the total cross-sectional area occupied by the vapor.

Example 4.3

Design a horizontal separator to separate 10,000 kg/h of liquid, density 962 kg/m^3, from 12,500 kg/h of vapor, density 23.6 kg/m^3. The vessel operating pressure will be 21 bar.

Solution

$$u_t = 0.07[(962.0 - 23.6)/23.6]^{1/2} = 0.44 \text{ m/s}$$

Try a separator without a demister pad.

$$u_S = 0.15 \times 0.44 = 0.066 \text{ m/s}$$

$$\text{Vapour volumetric flow-rate} = \frac{12{,}500}{3600 \times 23.6} = 0.147 \text{ m}^3/\text{s}$$

Take $h_v = 0.5D_v$ and $L_v/D_v = 4$

$$\text{Cross-sectional area for vapour flow} = \frac{\pi D_v^2}{4} \times 0.5 = 0.393Dv^2$$

$$\text{Vapour velocity, } u_v = \frac{0.147}{0.393Dv^2} = 0.374D_v^{-2}$$

Vapour residence time required for the droplets to settle to liquid surface

$$= h_v/u_s = 0.5D_v/0.066 = 7.58D_v$$

Actual residence time = vessel length/vapour velocity

$$= L_v/u_v = \frac{4D_v}{0.374\ Dv^{-2}} = 10.70D_v^3$$

For satisfactory separation required residence time = actual.

So, $7.58D_v = 10.70D_v^3$

$D_v = 0.84$ m, say 0.92 m (3 ft, standard pipe size)

Liquid hold-up time,

$$\text{liquid volumetric flow-rate} = \frac{10{,}000}{3600 \times 962.0} = 0.00289\ \text{m}^3/\text{s}$$

$$\text{liquid cross-sectional area} = \frac{\pi \times 0.92^2}{4} \times 0.5 = 0.332\ \text{m}^2$$

Length, $L_v = 4 \times 0.92 = 3.7$ m

Hold-up volume $= 0.332 \times 3.7 = 1.23\ \text{m}^3$

Hold-up time = liquid volume/liquid flow-rate

$= 1.23/0.00289 = 426$ s = 7 minutes.

This is unsatisfactory, 10 minutes minimum required.

Need to increase the liquid volume. This is best done by increasing the vessel diameter. If the liquid height is kept at half the vessel diameter, the diameter must be increased by a factor of roughly $(10/7)^{0.5} = 1.2$.

New $D_v = 0.92 \times 1.2 = 1.1$ m

Check liquid residence time,

$$\text{new liquid volume} = \frac{\pi \times 1.1^2}{4} \times 0.5 \times (4 \times 1.1) = 2.09\ \text{m}^3$$

new residence time $= 2.09/0.00289 = 723$ s = 12 minutes, satisfactory

Increasing the vessel diameter will have also changed the vapor velocity and the height above the liquid surface.

4.2.3 Gas-Solids Separations (Gas Cleaning)

Process gas streams must be cleaned up to prevent contamination of catalysts or products, and to avoid damage to equipment, such as compressors. Gas cleaning equipment can be classified according to the mechanism employed to separate the particles: gravity settling, impingement, centrifugal force, filtering, washing and electrostatic precipitation.

A variety of equipment has been developed for gas cleaning. The principal types used in the process industries are listed in Table (4.1) which shows the general field of application of each type in terms of the particle size separated, the expected separation efficiency, and the throughput.

Cyclones are the principal type of gas-solids separator employing centrifugal force, and are widely used. Cyclones are suitable for separating particles above about 10μm diameter; smaller particles, down to about 0.5μm, can be separated where agglomeration occurs.

The most commonly used design is the reverse-flow cyclone, Fig. (4.5). In a reverse-flow cyclone the gas enters the top chamber tangentially and spirals down to the apex of the conical section; it then moves upward in a second, smaller diameter, spiral, and exits at the top through a central vertical pipe. The solids move radially to the walls, slide down the walls, and are collected at the bottom.

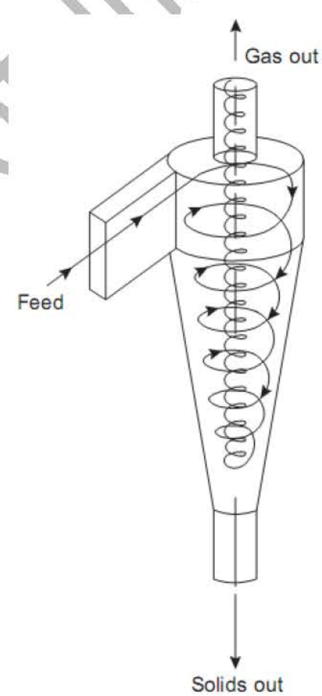


Figure (4.5): Reverse-flow cyclone

Table (4.1): Gas cleaning equipment

Type of equipment	Minimum particle size (μm)	Minimum loading (mg/m^3)	Approx. efficiency (%)	Typical gas velocity (m/s)	Maximum capacity (m^3/s)	Gas pressure drop (mm H_2O)	Liquid rate (m^3/10^3 m^3 gas)	Space required (relative)
Dry collectors								
Settling chamber	50	12,000	50	1.5–3	none	5	—	Large
Baffle chamber	50	12,000	50	5–10	none	3–12	—	Medium
Louver	20	2500	80	10–20	15	10–50	—	Small
Cyclone	10	2500	85	10–20	25	10–70	—	Medium
Multiple cyclone	5	2500	95	10–20	100	50–150	—	Small
Impingement	10	2500	90	15–30	none	25–50	—	Small
Wet scrubbers								
Gravity spray	10	2500	70	0.5–1	50	25	0.05–0.3	Medium
Centrifugal	5	2500	90	10–20	50	50–150	0.1–1.0	Medium
Impingement	5	2500	95	15–30	50	50–200	0.1–0.7	Medium
Packed	5	250	90	0.5–1	25	25–250	0.7–2.0	Medium
Jet	0.5 to 5 (range)	250	90	10–100	50	none	7–14	Small
Venturi	0.5	250	99	50–200	50	250–750	0.4–1.4	Small
Others								
Fabric filters	0.2	250	99	0.01–0.1	100	50–150	—	Large
Electrostatic precipitators	2	250	99	5–30	1000	5–25	—	Large

4.2.3.1 Cyclone Design

Stairmand developed two standard designs for gas-solid cyclones: a high-efficiency cyclone, Fig. (4.6 a), and a high throughput design, Fig. (4.6 b). The performance curves for these designs, obtained experimentally under standard test conditions. These curves can be transformed to other cyclone sizes and operating conditions by use of the following scaling equation, for a given separating efficiency:

$$d_2 = d_1 \left[\left(\frac{D_{c_2}}{D_{c_1}} \right)^3 \times \frac{Q_1}{Q_2} \times \frac{\Delta\rho_1}{\Delta\rho_2} \times \frac{\mu_2}{\mu_1} \right]^{1/2} \quad \ldots(4.1)$$

where d_1 = mean diameter of particle separated at the standard conditions, at the chosen separating efficiency, curve (a) or (b).

d_2 = mean diameter of the particle separated in the proposed design, at the same separating efficiency.

Dc_1 = diameter of the standard cyclone = 8 inches (203 mm).

Dc_2 = diameter of proposed cyclone, mm.

Q_1 = standard flow rate:

for high efficiency design=223 m^3/h.

for high throughput design = 669 m^3/h.

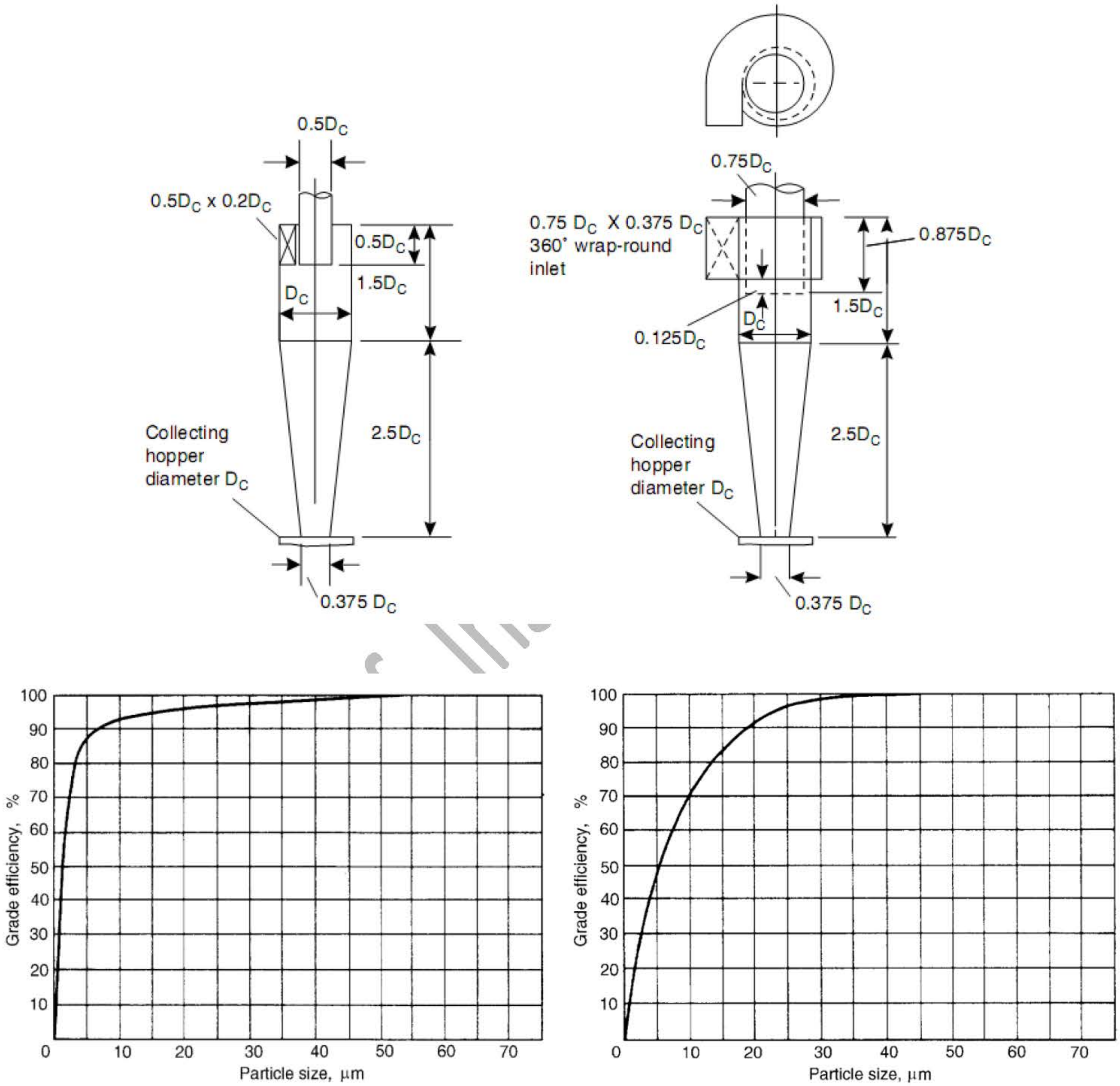
Q_2 = proposed flow rate, m^3/h.

$\Delta\rho_1$ = solid-fluid density difference in standard conditions = 2000 kg/m^3.

$\Delta\rho_2$ = density difference, proposed design.

μ_1 = test fluid viscosity (air at 1 atm, 20°C) = 0.018 mN s/m^2.

μ_2 = viscosity, proposed fluid.



(a) High efficiency cyclone **(b) High gas rate cyclone**

Fig. (4.6): Standard cyclone dimension and performance curves.

Pressure Drop

The pressure drop in a cyclone will be due to the entry and exit losses, and friction and kinetic energy losses in the cyclone. Empirical equation (4.2) can be used to estimate the pressure drop:

$$\Delta P = \frac{\rho_f}{203}\left\{u_1^2\left[1 + 2\phi^2\left(\frac{2r_t}{r_e} - 1\right)\right] + 2u_2^2\right\} \quad \ldots(4.2)$$

where : ΔP = cyclone pressure drop, millibars. (1 mbar = 10.47 mmH_2O)

ρ_f = gas density, kg/m^3

u_1 = inlet duct velocity, m/s.

u_2 = exit duct velocity, m/s.

r_t = radius of circle to which the center line of the inlet is tangential, m.

r_e = radius of exit pipe, m.

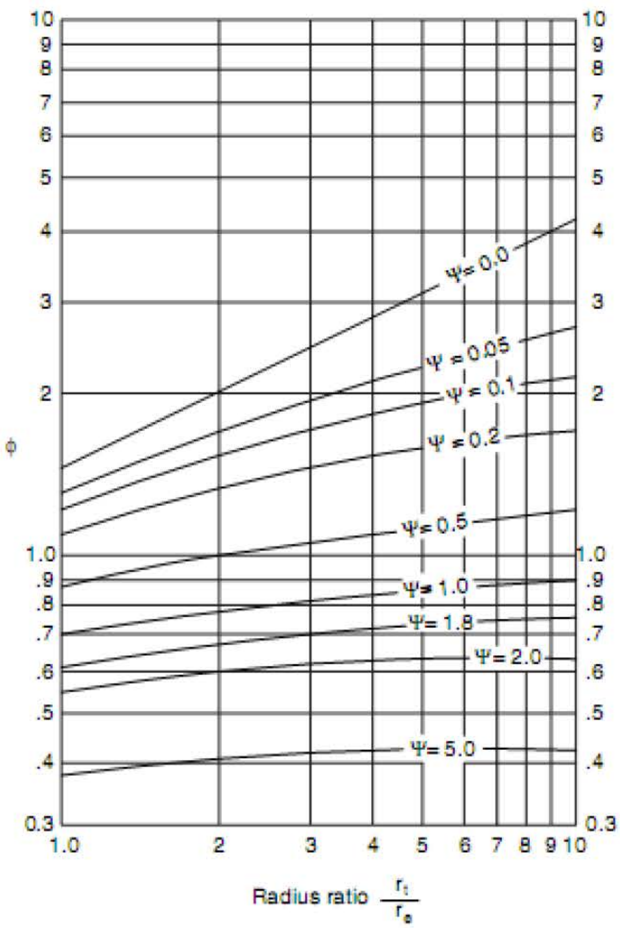


Figure (4.7) Cyclone pressure drop factor ø = factor from Figure (4.7).

Ψ = parameter in Figure (4.7), given by:

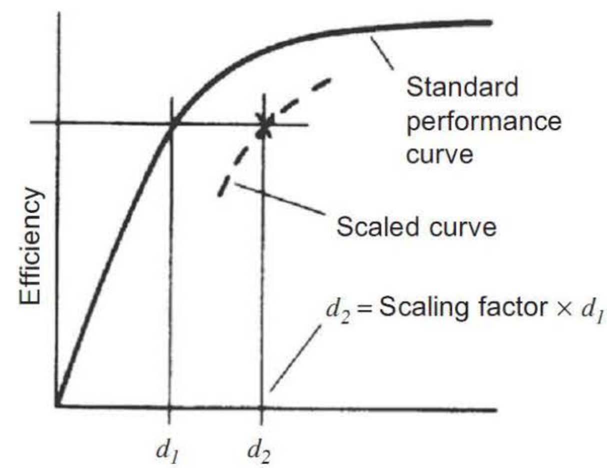
$$\psi = f_c \frac{A_s}{A_1} \quad \dots(4.3)$$

f_c = friction factor, taken as 0.005 for gases.

A_s = surface area of cyclone exposed to the spinning fluid, m^2

For design purposes this can be taken as equal to the surface area of a cylinder with the same diameter as the cyclone and length equal to the total height of the cyclone (barrel plus cone).

A_1 = area of inlet duct, m^2.



Scaled performance curve.

General design procedure

1. Select either the high-efficiency or high-throughput design, depending on the performance required.
2. Estimate the number of cyclones needed in parallel.
3. Obtain an estimate of the particle size distribution of the solids in the stream to be treated.
4. Calculate the cyclone diameter for an inlet velocity of 15 m/s. Scale the other cyclone dimensions from Figures (4.6 a) or (4.6 b).
5. Calculate the scale-up factor for the transposition of Figures (4.6a) or (4.6 b).
6. Calculate the cyclone performance and overall efficiency (recovery of solids). If unsatisfactory try a smaller diameter.
7. Calculate the cyclone pressure drop.

Example 4.4

Design a cyclone to recover solids from a process gas stream. The anticipated particle size distribution in the inlet gas is given below. The density of the particles is 2500 kg/m^3, and the gas is essentially nitrogen at 150°C. The stream volumetric flow-rate is 4000 m^3/h, and the operation is at atmospheric pressure. An 80 percent recovery of the solids is required.

Particle size (μm)	50	40	30	20	10	5	2
Percentage by weight less than	90	75	65	55	30	10	4

Solution

As 30 per cent of the particles are below 10 μm the high-efficiency design will be required to give the specified recovery.

$$\text{Flow-rate} = \frac{4000}{3600} = 1.11\ \text{m}^3/\text{s}$$

$$\text{Area of inlet duct, at 15 m/s} = \frac{1.11}{15} = 0.07\ \text{m}^2$$

From Figure 4 .6 *a*, duct area $= 0.5\ D_c \times 0.2\ D_c$

so, $D_c = 0.84$

This is clearly too large compared with the standard design diameter of 0.203 m.

Many units must be used,

A rough estimation of the number of units is done as:

Number of units = 0.84/0.203 = 4.14 say 4 cyclones

Cross sectional area of the new cyclone = $(\frac{\pi}{4} \times 0.84^2)/4 = 0.1385\ \text{m}^2$

The new $D_c = \sqrt{\frac{0.1385}{(\pi/4)}} = 0.42$ m

Try four cyclones in parallel, $D_c = 0.42$ m.

$$\text{Flow-rate per cyclone} = 1000\ \text{m}^3/\text{h}$$

$$\text{Density of gas at } 150^\circ\text{C} = \frac{28}{22.4} \times \frac{273}{423} = 0.81\ \text{kg/m}^2,$$

negligible compared with the solids density

$$\text{Viscosity of } N_2 \text{ at } 150^\circ\text{C} = 0.023\ \text{cp(mN s/m}^2)$$

From equation 4 .1,

$$\text{scaling factor} = \left[\left(\frac{0.42}{0.203}\right)^3 \times \frac{223}{1000} \times \frac{2000}{2500} \times \frac{0.023}{0.018}\right]^{1/2} = \underline{\underline{1.42}}$$

The performance calculations, using this scaling factor and Figure 4 .6 *a*, are set out in the table below:

Calculated performance of cyclone design

1	2	3	4	5	6	7
Particle size (μm)	Per cent in range	Mean particle size ÷ scaling factor	Efficiency at scaled size % (Figure 10.46a)	Collected $\frac{(2) \times (4)}{100}$	Grading at exit (2)–(5)	Per cent at exit
>50	10	35	98	9.8	0.2	1.8
50–40	15	32	97	14.6	0.4	3.5
40–30	10	25	96	9.6	0.4	3.5
30–20	10	18	95	9.5	0.5	4.4
20–10	25	11	93	23.3	1.7	15.1
10–5	20	5	86	17.2	2.8	24.8
5–2	6	3	72	4.3	1.7	15.1
2–0	4	1	10	0.4	3.6	31.8
	100		Overall collection efficiency	88.7	11.3	100.0

The collection efficiencies shown in column 4 of the table were read from Figure 4.6 *a* at the scaled particle size, column 3. The overall collection efficiency satisfies the specified solids recovery. The proposed design with dimension in the proportions given in Figure 4.6 *a* is shown in Figure 4.8.

Pressure-drop calculation

$$\text{Area of inlet duct, } A_1, = 210 \times 80 = 16{,}800 \text{ mm}^2$$

$$\text{Cyclone surface area, } A_s = \pi 420 \times (630 + 1050)$$

$$= 2.218 \times 10^6 \text{ mm}^2$$

f_c taken as 0.005

$$\psi = \frac{f_c, A_s}{A_1} = \frac{0.005 \times 2.218 \times 10^6}{16{,}800} = 0.66$$

$$\frac{r_t}{r_e} = \frac{(420 - (80/2))}{210} = 1.81$$

From Figure 4.7, $\phi = 0.9$.

$$u_1 = \frac{1000}{3600} \times \frac{10^6}{16{,}800} = 16.5 \text{ m/s}$$

Area of exit pipe $= \dfrac{\pi \times 210^2}{4} = 34{,}636 \text{ mm}^2$

$$u_2 = \frac{1000}{3600} \times \frac{10^6}{34{,}636} = 8.0 \text{ m/s}$$

From equation 4.2

$$\Delta P = \frac{0.81}{203}[16.5^2[1 + 2 \times 0.9^2(2 \times 1.81 - 1)] + 2 \times 8.0^2]$$
$$= \underline{\underline{6.4 \text{ millibar}}} \ (67 \text{ mm } H_2O)$$

This pressure drop looks reasonable.

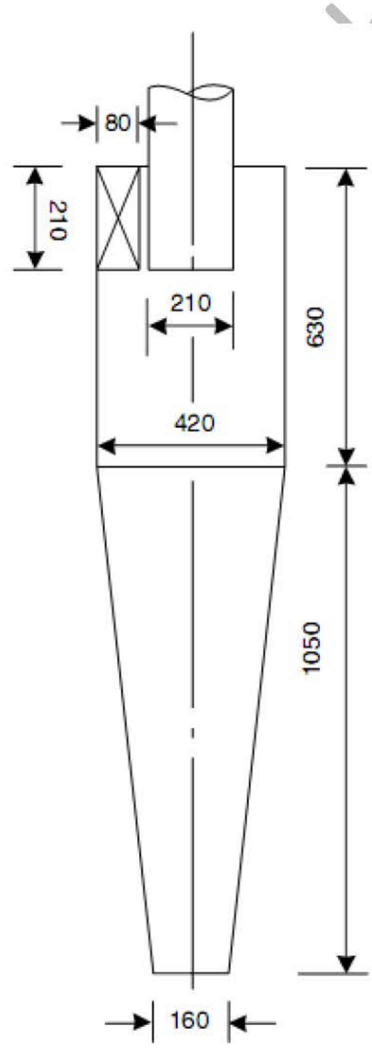


Figure 4.8. Proposed cyclone design, all dimensions mm

4.3 Storage Tanks

Liquids are usually stored in bulk in vertical cylindrical steel tanks. Fixed and floating-roof tanks are used. In a floating-roof tank a movable piston floats on the surface of the liquid and is sealed to the tank walls. Floating-roof tanks are used to eliminate evaporation losses and, for flammable liquids, to obviate the need for inert gas blanketing to prevent an explosive mixture forming above the liquid, as would be the situation with a fixed-roof tank. Horizontal cylindrical tanks and rectangular tanks are also used for storing liquids, usually for relatively small quantities.

4.3.1 Horizontal tanks: Above ground they are limited to ***35,000*** gal (***160*** m^3). Normally they are supported on steel structures or concrete saddles at elevations of ***6 to 10 ft***. The minimum thickness of shell and heads is ***3/16 in*** with diameters of ***(48-72) in*** and ***1/4 in*** with diameters of ***(73-132) in***.

4.3.2 Vertical tanks: Vertical cylindrical tanks, with flat bases and conical roofs, are universally used for the bulk storage of liquids at atmospheric pressure. Tank size vary from tens of cubic meters to several hundred cubic meters as shown in Fig. (4.9).

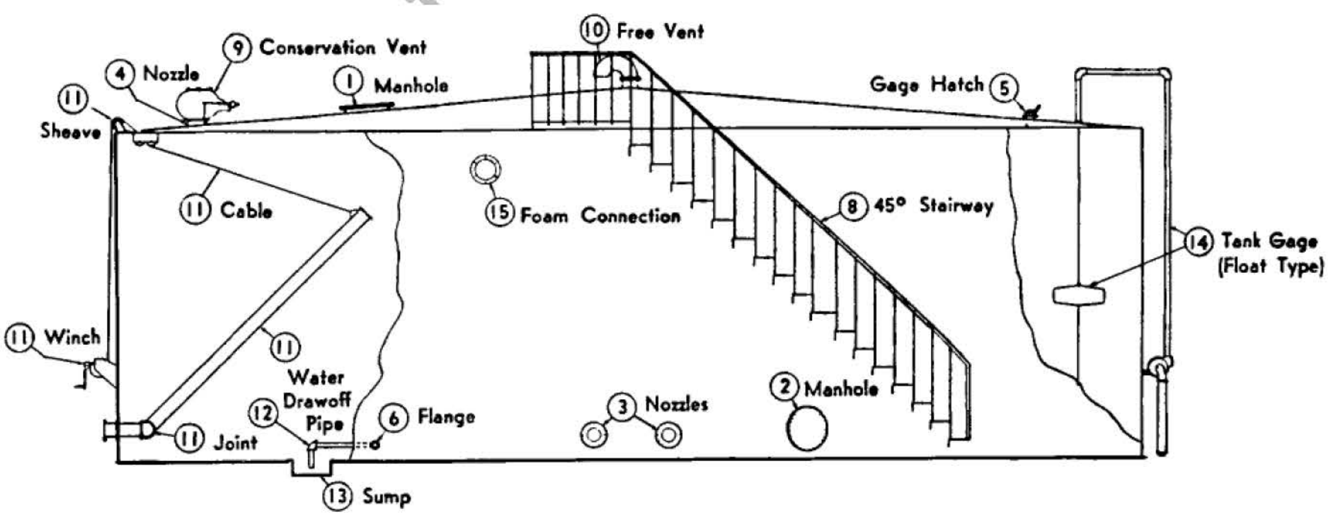


Figure (4.9) A large tank and its appurtenance

The main load to be considered in the design of these tanks is the hydrostatic pressure of the liquid, moreover the tanks must also be designed to withstand wind loading and, for some location, the weight of snow on the tank roof.

The minimum wall thickness required to resist the hydrostatic pressure can be calculated from the following equation:

$$t = \frac{\rho_l \, g \, H_l \quad D}{2\sigma \, J \quad\quad 10^3}$$

where t = tank thickness required at depth H_l , mm.

H_l = liquid depth, m.

ρ_l = liquid density, kg/m^3.

J = welded joint factor = 1 for virgin plates and less for welded plates.

g = gravitational acceleration, 9.81 m/s^2.

σ = design stress for tank material, N/mm^2.

D = tank diameter, m.

the joint factor (J) can take as:

Radiography = 1 (virgin plates)

Spot radiography = 0.8

No radiography = 0.7

For small tanks, a constant wall thickness would normally be used, calculated at maximum liquid depth. With large tanks, it is economical to take account of the variation in hydrostatic pressure with depth, by increasing the plate thickness progressively from the top to the bottom of the tank.

Liquid stored at near atmospheric pressure, as the tank cools during the night air is down in, then the vaporization occurs to saturation, and the vapor mixture is expelled as the tank warms up during the day. Volatile liquids such as gasoline consequently suffer a material loss and also a change in composition because of the selective loss of lighter constituents.

In order to minimize such effects, a floating roof is a pad which floats on the surface of the stored liquid with a diameter of about a foot less than that of the tank. The annular space between the float and the shell may be sealed by the special fabric seal as shown in Fig. (4.10).

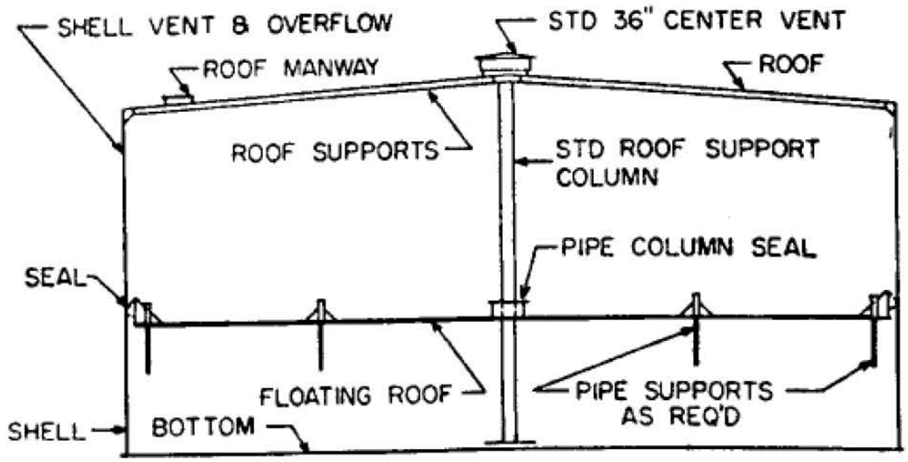


Figure (4.10) Floating roof tank

Example 4.5

A storage tank at atmospheric pressure used to store sulfuric acid (density = 1520 kg/m^3) is to have an inside diameter of 4 m and a height of 12 m. The maximum liquid level in the tank will be at 11 m. Estimate the virgin plate thickness required of the tank. Take the allowable design stress as 90 N/mm^2.

Solution

$$t = \frac{\rho_l\, g\, H_l}{2\sigma J}\frac{D}{10^3} = \frac{\mathbf{1520 * 9.81 * 11 * 4}}{\mathbf{2 * 90 * 1 * 10^3}} = \mathbf{3.64\ mm}$$

CHAPTER FIVE
PRESSURE VESSELS

5.1 Introduction

A pressure vessel is defined as a container with a pressure differential between inside and outside. The inside pressure is usually higher than the outside, except for some isolated situations. The fluid inside the vessel may undergo a change in state as in the case of steam boilers, or may combine with other reagents as in the case of a chemical reactor.

5.2 Design Considerations

5.2.1 Design Pressure

A vessel must be designed to withstand the maximum pressure to which it is likely to be subjected in operation. For vessels under internal pressure, the design pressure is normally taken as the pressure at which the relief device is set. This will normally be 5 - 10 % above the normal working pressure, to avoid spurious operation during minor process upsets. When deciding the design pressure, the hydrostatic pressure in the base of the column should be added to the operating pressure, if significant.

5.2.2 Design Temperature

The strength of metals decreases with increasing temperature so the maximum allowable design stress will depend on the material temperature. The design temperature at which the design stress is evaluated should be taken as the maximum working temperature of the material.

5.2.3 Materials

Pressure vessels are constructed from plain carbon steels, stainless steels, clad plate, and reinforced plastics. Selection of a suitable material must take into account the suitability of the material for fabrication (particularly welding) as well as the compatibility of the material with the process environment.

5.2.4 Design Stress

For design purposes it is necessary to decide a value for the maximum allowable stress that can be accepted in the material of construction. Typical design stress factors for pressure components are shown in Table (5.1).

5.2.5 Welded Joint Efficiency

The strength of a welded joint will depend on the type of joint and the quality of the welding checked by radiography. Taking the factor as 1 implies that the joint is equally as strong as the virgin plate or welds are fully radiographed. For other types of joint, the joint factor can take as:

Radiography = 1

Spot radiography = 0.8

No radiography = 0.7

5.2.6 Corrosion Allowance

The "corrosion allowance" is the additional thickness of metal added to allow for material lost by corrosion and erosion, or scaling. Usually, a minimum allowance of 2 mm should be used.

5.2.7 Minimum Practical Wall Thickness

There will be a minimum wall thickness required to ensure that any vessel is sufficiently rigid to withstand its own weight, and any incidental loads. Wall thickness of any vessel should not be less than the values given in Table (5.2); the values include a corrosion allowance of 2 mm.

Table (5.1) Typical design stresses for plate

Material	Tensile strength (N/mm^2)	Design stress at temperature °C (N/mm^2)									
		0 to 50	100	150	200	250	300	350	400	450	500
Carbon steel (semi-killed or silicon killed)	360	135	125	115	105	95	85	80	70		
Carbon-manganese steel (semi-killed or silicon killed)	460	180	170	150	140	130	115	105	100		
Carbon-molybdenum steel, 0.5 per cent Mo	450	180	170	145	140	130	120	110	110		
Low alloy steel (Ni, Cr, Mo, V)	550	240	240	240	240	240	235	230	220	190	170
Stainless steel 18Cr/8Ni unstabilised (304)	510	165	145	130	115	110	105	100	100	95	90
Stainless steel 18Cr/8Ni Ti stabilised (321)	540	165	150	140	135	130	130	125	120	120	115
Stainless steel 18Cr/8Ni Mo $2\frac{1}{2}$ per cent (316)	520	175	150	135	120	115	110	105	105	100	95

Table (5.2) Minimum Practical Wall Thickness

Vessel diameter (m)	Minimum thickness (mm)
1	5
1 to 2	7
2 to 2.5	9
2.5 to 3.0	10
3.0 to 3.5	12

5.3 The Design of Thin-Walled Vessels Under Internal Pressure

5.3.1 Cylinders and Spherical Shells

For a cylindrical and spherical shell, the minimum thickness required to resist internal pressure can be determined from equation.

$$t = \frac{P_i D_i}{2Jf - P_i} \qquad \text{(for cylindrical shell)}$$

$$t = \frac{P_i D_i}{4Jf - 1.2P_i} \qquad \text{(for spherical shell)}$$

where t = wall thickness.
P_i = Internal pressure.
D_i = Inside diameter.
f = design stress (from Table 5.1).
J = the joint factor.

Any consistent set of units can be used for equations above.

5.4 Heads and Ends (Closures)

The ends of a cylindrical vessel are closed by heads of various shapes. The principal types used are:

1. Flat plates and formed flat heads(flanged plate, welded plate, bolted plate).
2. Domed head (Torispherical heads, Ellipsoidal heads, Hemispherical heads).

Domed are formed by pressing or spinning; large diameters are fabricated from formed sections. Torispherical heads are often referred to as dished ends.

5.4.1 Choice of Closure

1- Flat plates are used as covers for manways, and as the channel covers of heat exchangers. and for low-pressure vessels.

2- Torispherical heads (dished ends) are the most commonly used end closure for vessels up to operating pressures of 15 bar. They can be used for higher pressures, but above 10 bar their cost should be compared with that of an equivalent ellipsoidal head.

3- Above 15 bar an ellipsoidal head will usually prove to be the most economical closure to use.

4- A hemispherical head is the strongest shape; capable of resisting about twice the pressure of a torispherical head of the same thickness. The cost of forming a hemispherical head will be higher than that for a shallow torispherical head. Hemispherical heads are used for high pressures.

5.4.2 Design of Flat Ends

Though the fabrication cost is low, flat ends are not a structurally efficient form, and very thick plates would be required for high pressures or large diameters. The design equations used to determine the thickness of flat and domed ends are listed in Table (5.3) with the degree of constraint at the plate periphery. The minimum thickness required is given by:

$$t = C_p D_e \sqrt{\frac{P_i}{f}}$$

where C_p= a design constant, dependent on the edge constraint,
D_e = nominal plate diameter,
f = design stress.

Any consistent set of units can be used.

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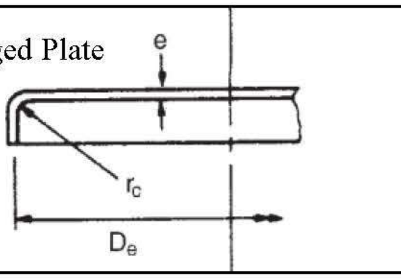
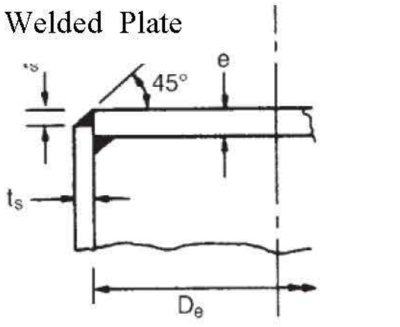
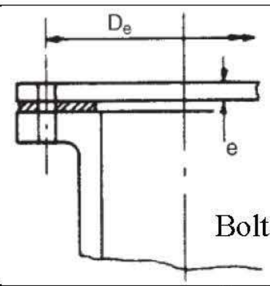
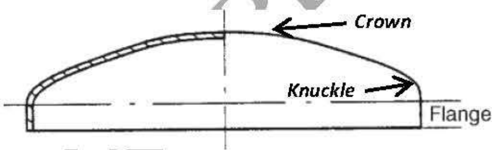
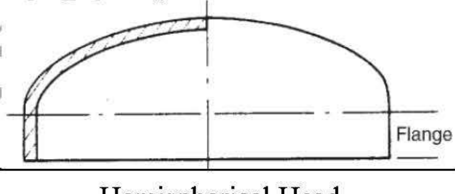
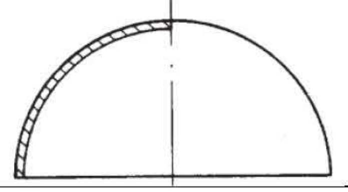
$$t = C_p D_e \sqrt{\frac{P_i}{f}}$$

where C_p= a design constant, dependent on the edge constraint,
D_e = nominal plate diameter,
f = design stress.

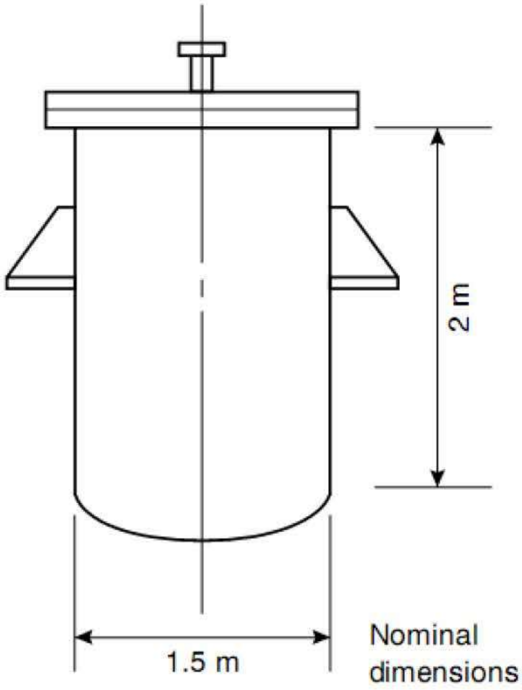
Any consistent set of units can be used.

Table (5.3) Design of Heads and Ends

Nomenclature		Minimum Thickness	Constrains
Flat-ended closer	Flanged Plate 	$t = C_p D_e \sqrt{\frac{P_i}{f}}$	*For* $D_i < 0.6$ *and* $r_{corner} = 0.25\,t$ $C_p = 0.45$ $D_e = D_i$
	Welded Plate 		$C_p = 0.55$ *and* $D_e = D_i$
	 Bolted Cover		$C_p = 0.4$ *and* D_e *=bolt circle diameter*
Domed Heads	Torispherical Head 	$t = \frac{P_i R_c C_s}{2fJ + P_i(C_s - 0.2)}$ $C_s = \frac{1}{4}(3 + \sqrt{R_c/R_k})$ Rc = crown radius Rk = knuckle raduis	$R_c \leq D_i$ *and* $R_k/R_c \geq 0.06$ *for formed head (no joints in the head) the joint factor is taken as 1*
	Ellipsodial Head 	$t = \frac{P_i D_i}{2Jf - 0.2P_i}$	*Major axis :minor axis 2:1 formed head (no joints in the head) the joint factor is taken 1*
	Hemispherical Head 	*t = 0.6 t cylindrical shell*	*For formed head (no joints in the head) the joint factor is taken as 1*

Example 5.1

Estimate the thickness and heads required for the component parts of the vessel shown in the diagram. The vessel is to operate at a pressure of 14 bar (absolute) and temperature of 300°C. The material of construction will be plain carbon steel. Welds will be fully radiographed. A corrosion allowance of 2 mm should be used.



Solution

Design pressure, take as 10 per cent above operating pressure,

$$= (14 - 1) \times 1.1$$

$$= 14.3 \text{ bar}$$

$$= 1.43 \text{ N/mm}^2$$

Design temperature 300°C.
From Table 1, typical design stress = 85 N/mm^2.

Cylindrical section

$$t = \frac{1.43 \times 1.5 \times 10^3}{2 \times 85 - 1.43} = 12.7 \text{ mm}$$

add corrosion allowance $12.7 + 2 = 14.7$

say $\underline{\underline{15 \text{ mm plate}}}$

Domed head

(i) Try a standard dished head (torisphere):

$$\text{crown radius } R_c = D_i = 1.5 \text{ m}$$

$$\text{knuckle radius} = 6 \text{ per cent } R_c = 0.09 \text{ m}$$

A head of this size would be formed by pressing: no joints, so $J = 1$.

$$C_s = \frac{1}{4}\left(3 + \sqrt{\frac{R_c}{R_k}}\right) = \frac{1}{4}\left(3 + \sqrt{\frac{1.5}{0.09}}\right) = 1.77$$

$$t = \frac{1.43 \times 1.5 \times 10^3 \times 1.77}{2 \times 85 + 1.43(1.77 - 0.2)} = \underline{\underline{22.0 \text{ mm}}}$$

(ii) Try a "standard" ellipsoidal head, ratio major : minor axes = 2 : 1

$$t = \frac{1.43 \times 1.5 \times 10^3}{2 \times 85 - 0.2 \times 1.43}$$
$$= \underline{\underline{12.7 \text{ mm}}}$$

So an ellipsoidal head would probably be the most economical. Take as same thickness as wall 15 mm.

Flat head

Use a full face gasket $C_p = 0.4$

D_e = bolt circle diameter, take as approx. 1.7 m.

$$t = 0.4 \times 1.7 \times 10^3 \sqrt{\frac{1.43}{85}} = \underline{88.4 \text{ mm}}$$

Add corrosion allowance and round-off to $\underline{\underline{90 \text{ mm}}}$.

This shows the inefficiency of a flat cover. It would be better to use a flanged domed head.