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# DESIGN OF ABSORBER

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## 5.1 ABSORPTIONS

The removal of one or more component from the mixture of gases by using a suitable solvent is second major operation of Chemical Engineering that is based on mass transfer.

In gas absorption, soluble vapors are more or less absorbed in the solvent from its mixture with inert gas. The purpose of such gas scrubbing operations may be any of the following;

- a) For Separation of component having the economic value.
- b) As a stage in the preparation of some compound.
- c) For removing of undesired component (pollution).

## 5.2 TYPES OF ABSORPTION

- 1) Physical absorption,
- 2) Chemical Absorption.

### 5.2.1 Physical Absorption

In physical absorption mass transfer take place purely by diffusion and physical absorption is governed by the physical equilibria.

### 5.2.2 Chemical Absorption

In this type of absorption as soon as a particular component comes in contact with the absorbing liquid a chemical reaction take place. Then, by reducing the concentration of component in the liquid phase, which enhances the rate of diffusion.

### 5.3 TYPES OF ABSOR5RS

There are two major types of absorbers which are mainly used for absorption purposes:

- Packed column
- Plate column

### 5.4 **COMPARISON BETWEEN PACKED AND PLATE COLUMN**

- 1) The packed column provides continuous contact between vapors and liquid phases while the plate column brings the two phases into contact on stage wise basis.
- 2) **SCALE:** For column diameter of less than approximately 8 ft, it is more usual to employ packed towers because of high fabrication cost of small trays. But if the column is very large then the liquid distribution is problem and large volume of packing and its weight is problem.
- 3) **PRESSURE DROP:** Pressure drop in packed column is less than the plate column. In plate column there is additional friction generated as the vapor passes through the liquid on each tray. If there are large

number of Plates in the tower, this pressure drop may be quite high and the use of packed column could effect considerable saving.

- 4) **LIQUID HOLD UP:** Because of the liquid on each plate there may be a Urge quantity of the liquid in plate column, whereas in a packed tower the liquid flows as a thin film over the packing.
- 5) **SIZE AND COST:** For diameters of less than 3 ft. packed tower require lower fabrication and material costs than plate tower with regard to height, a packed column is usually shorter than the equivalent plate column.

From the above consideration packed column is selected as the absorber, because in our case the diameter of the column is approximately 2 meters. As the solubility is infinity so the liquid will absorb as much gases as it remain in contact with gases so packed tower provide more contact. It is easy to operate.

## **5.5 PACKING**

The packing is the most important component of the system. The packing provides sufficient area for intimate contact between phases. The efficiency of the packing with respect to both HTU and flow capacity determines to a significance extent the overall size of the tower. The economics of the installation is therefore tied up with packing choice.

The packings are divided into those types which are dumped at random into the tower and these which must be stacked by hand. Dumped packing consists of unit 1/4 lo 2 inches in major dimension and is used roost in the smaller columns. The units in stacked packing are 2 to about 8 inches in size; they are used only in the larger towers.

The Principal Requirement of a Tower packing are:

- 1) It must be chemically inert to the fluids in the tower.
- 2) It must be strong without excessive weight.
- 3) It must contain adequate passages for both streams without excessive liquid hold up or pressure drop.
- 4) It must provide good contact between liquid and gas.
- 5) It must be reasonable in cost.

Thus most packing is made of cheap, inert, fairly light materials such as clay, porcelain, or graphite. Thin-walled metal rings of steel or aluminum are some limes used.

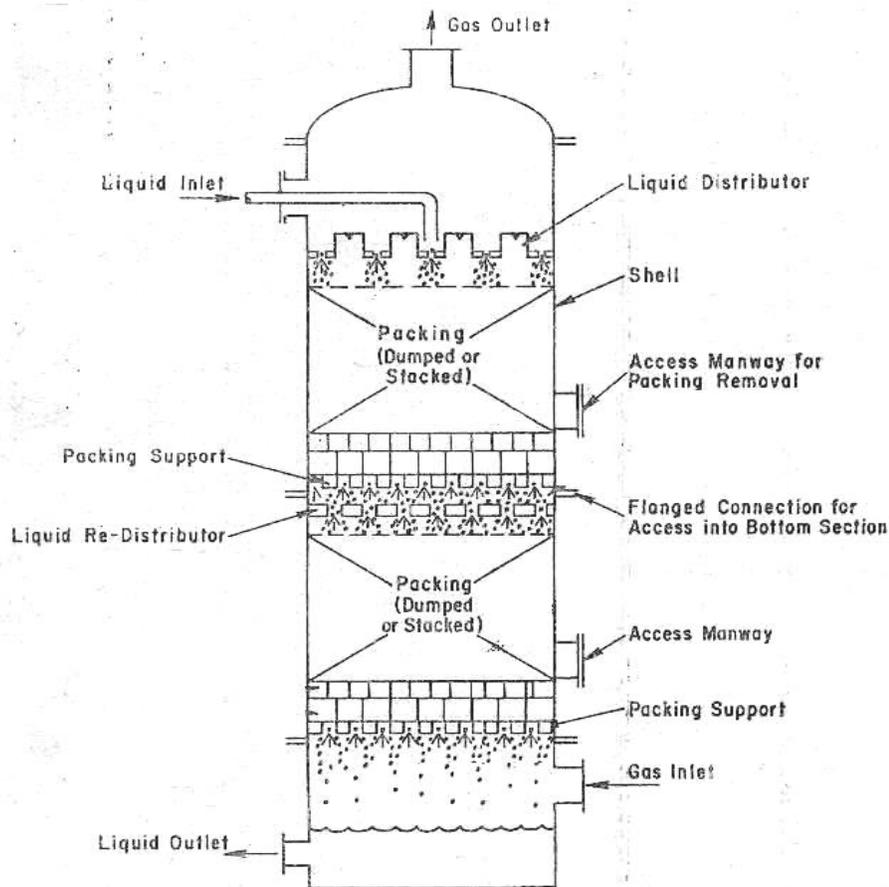
Common Packings are:

- a) Berl Saddle.
- b) Intalox Saddle.
- c) Rasching rings.
- d) Lessing rings.
- e) Cross-partition rings.
- f) Single spiral ring.
- g) Double - Spiral ring.
- h) Triple - Spiral ring.

## **5.6 DESIGNING STEPS FOR ABSORPTION COLUMN**

- Determining the approximate dia of the column
- Selection of column.
- Selection of packing and material
- Calculating the size of packing
- Calculating the actual dia of column

- Calculating the flooding velocity
  - a) Finding loading velocity with the knowledge the flooding velocity
  - b) Calculating actual dia of column
- Finding the no. of transfer units ( $N_{OG}$ )
- Determining the height of packing
- Determining the height of the column
- Determining the pressure drop.
- by equation  $\Delta P = \frac{a \times 10^{b\ell} g_F^2}{\rho_G}$  [in. water /ft of packing]



Cross-section of typical packed tower.

➔ **Design calculation of absorber**

We want to scrub Acetonitrile, HCN and acrylonitrile in the absorption tower using water stream .this is known as multi component absorption. The solubility data of these components shows that Acetonitrile and HCN Are infinitely soluble in water while acrylonitrile has limited solubility in water. Acrylonitrile is least soluble among three components, therefore we base our design of packed bed absorption tower on the solubility of acrylonitrile in water. The solvent used for this purpose is water.

**Basis: 1 hour operation**

**Input gaseous stream**

Components	Kg	Kgmol	Mol%
Acrylonitrile	9900	186.79	5.95
HCN	1687.4	62.50	1.99
Acetonitrile	996.46	24.30	0.77
CO	982	35.07	1.12
CO <sub>2</sub>	3394.8	77.16	2.46
Propane	1234.5	28.06	0.89
Propylene	736	17.54	0.56
Water	9847.8	547.10	17.43
Oxygen	2356.7	73.65	2.35
Nitrogen	58329	2083.19	66.37
Impurities	212.46	3.51	0.11
Total	89677.12	3138.87	100

Water entering in the absorber = 223607 Kg /hr

**Output liquid stream**

Components	Kg	Kgmol	Mol%
Acrylonitrile	9702	183.05	1.54
HCN	1654	61.25	0.512
Acetonitrile	976.5	23.817	0.199
Water	210109.3	11672.73	97.72
Impurities	212.5	3.507	.0293
Total	223654.3	11944.35	100

**→ Selection of Packing**

We have selected ceramic Intalox saddle.

Intalox saddle and pall rings are most popular choices. We have selected ceramic intalox saddle because they are most efficient. We have selected the ceramic material of packing because in our system oxygen and water are present and they can cause corrosion and ceramic material will prevent corrosion.

**→ Size of the Packing**

Now we will find the maximum size of intalox saddle which would be used for this particular dia of the column.

$$\text{Packing size} = \frac{1 \times D}{15} = \frac{1 \times 1}{15} = 0.0666 \text{ m} = 66 \text{ mm}$$

Although the efficiency of higher for small packing, it is generally accept that it is economical to use these small sizes in an attempt to improve the performance of a column. It is preferable to use the largest recommended size of a particular type of packing and to increase the packed height to compensate for small loss of efficiency.

### → Number of Transfer Units ( $N_{OG}$ )

In an absorption tower mol. fraction of acrylonitrile is  $X$ ,  $Y$  in liquid gas respectively. Then by acrylonitrile mass balance we have:

$$G_m(Y - Y_2) = L_m(X - X_2)$$

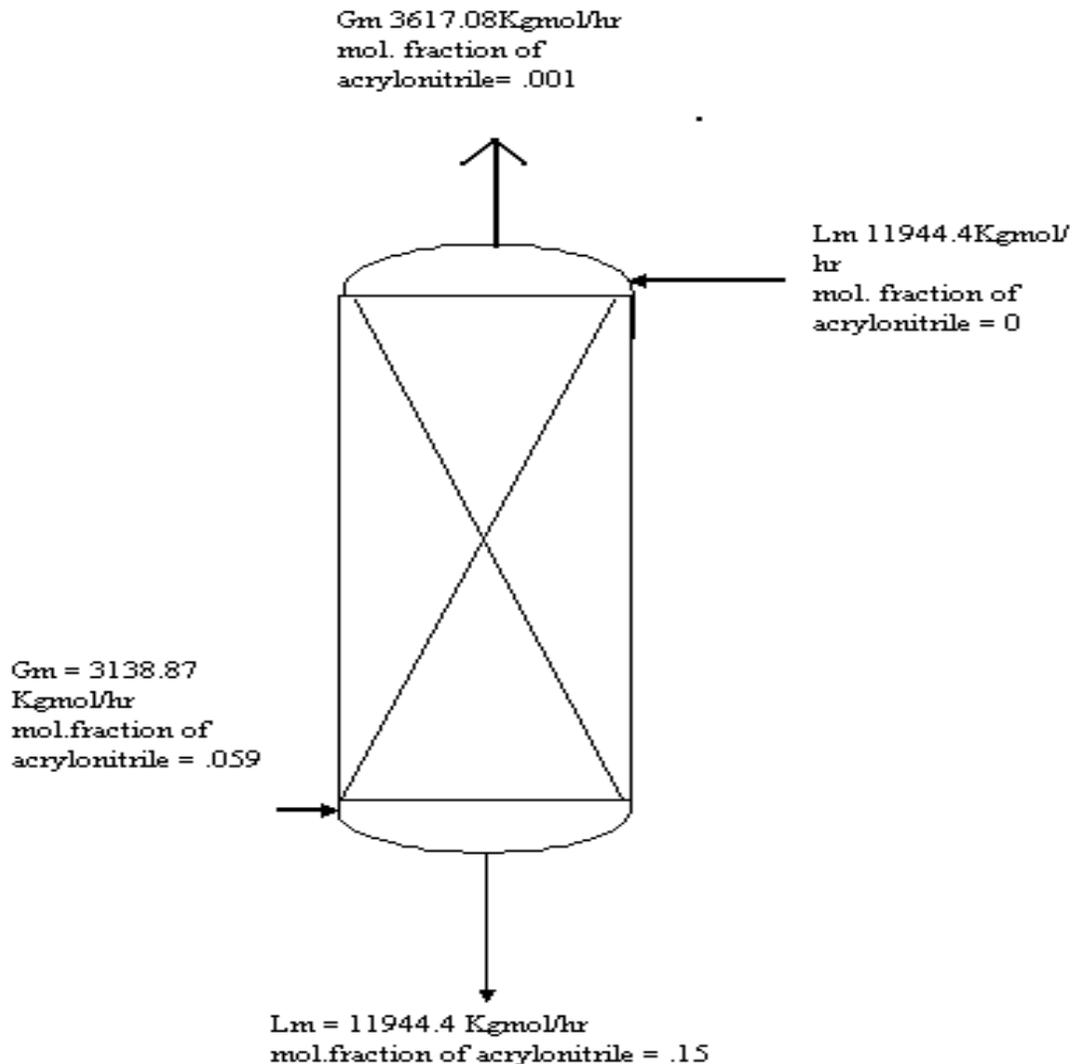
$$3138.87(Y - .001) = 11944.35(X - 0)$$

$$Y = 3.805X + .001 \rightarrow (1)$$

$$X = .26Y - .00026$$

$$X = .26(Y - .001) \rightarrow (2)$$

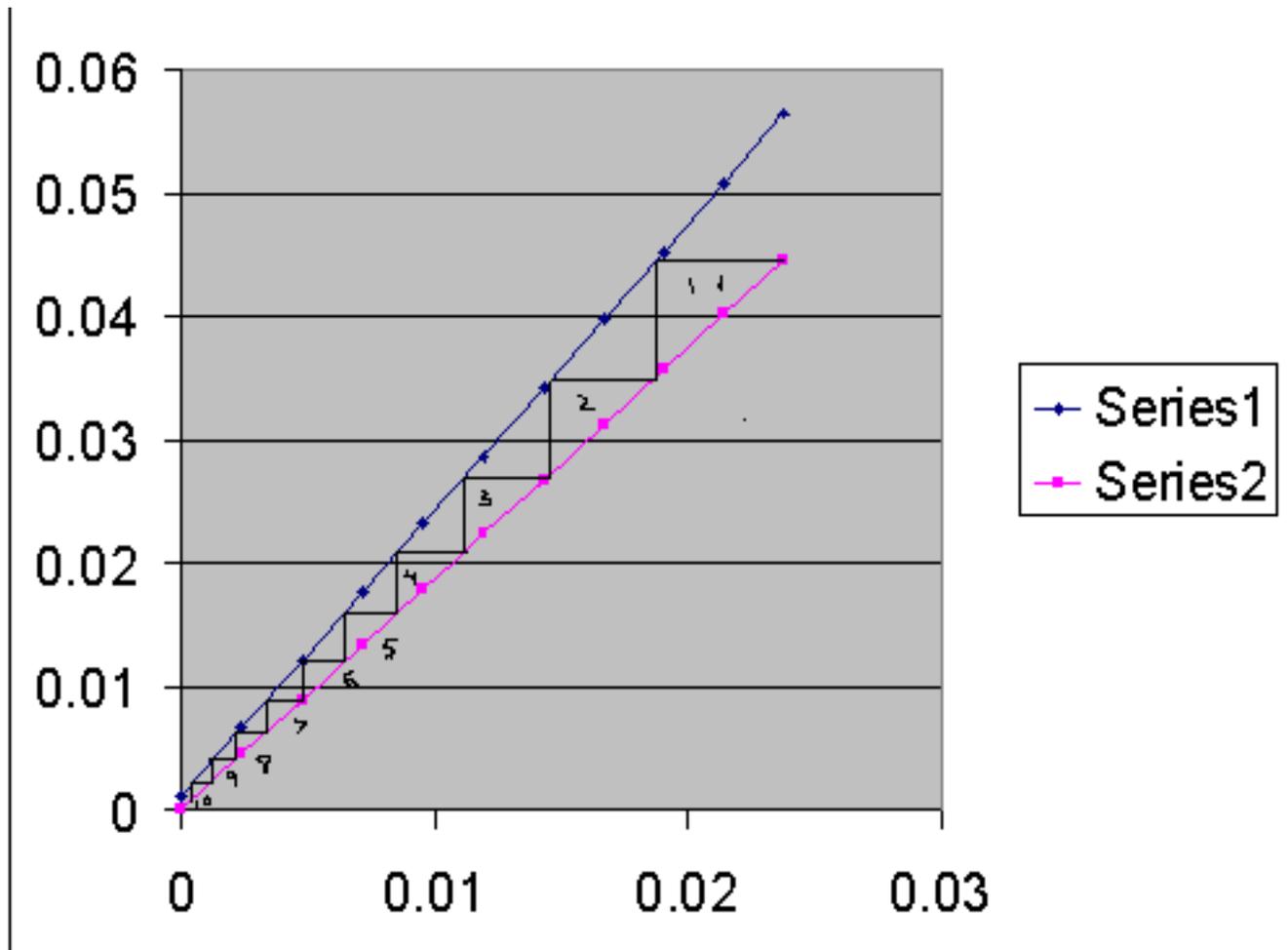
The equation no. 1 and 2 represents the operating line of absorption of acrylonitrile.



As the concentration of acrylonitrile is very small in stream, the equilibrium curve for the system will be a straight line with a slope of 1.7. Hence the equation of equilibrium curve for the acrylonitrile-water system is

$$Y^* = 1.7 X \rightarrow (3)$$

Now we assumed different values of  $Y$  and calculated their corresponding values of  $X$  &  $Y^*$  using equation (2) and (3). The graph is shown below:



Number of transfer unit after drawing graph between  $X$  Vs  $Y$  and  $X$  Vs  $Y^*$  we get using McCabe Thiele method:

$$N_{OG} = 10$$

### → Calculation of column dia:

Most methods for determining the size of randomly packed towers are derived from the Sherwood correlations which are used here to fine out diameter of absorber. The physical property of a gas can be taken as that of air at 45 °C and 170 KN/m<sup>2</sup> because concentration of acrylonitrile is very small in a gas mixture and average molecular weight of gas mixture is 28.56 Kg/Kgmol

Thus for abscisox =  $L/G \times \rho_v/\rho_l$

L = Flow rate of water = 223607 Kg/hr

G = Flow rate of a gas mixture = 89677.12 Kg/hr

$\rho_v$  = density of gas at 45 °C and 170 KN/m<sup>2</sup>

$\rho_l$  = density of water at 45 °C and 170 KN/m<sup>2</sup>

$\rho_v = PM/RT$  (where, R = 0.08205)

$$\rho_v = \frac{1.7 \times 28.56}{0.08205 \times 318} = 1.862 \text{ g/L} = 1.862 \text{ Kg/m}^3$$

$$\rho_l = 990 \text{ Kg/m}^3$$

Therefore

$$L/G \times \rho_v/\rho_l = 0.108$$

For our absorber we will design for 42mm water in packing.

Thus for 42mm of water / m of packing height

$$K_{4_} = 1.4, \quad F_p = 22.3$$

$\mu_L$  = viscosity of water at 45 °C = .63 Cp

$$G^* = [K_4 \rho_v (\rho_l - \rho_v) / 13.1 F_p (\mu_L / \rho_l)]^{1/2}$$

$$G^* = 6.06 \text{ Kg m}^2/\text{sec}.$$

Thus

$$A = \text{area of cross section} = G / G^*$$

$$A = 4.11 \text{ m}^2$$

$$\text{Diameter} = D = [4 \times A / \pi]^{1/2}$$

$$D = 2.3 \text{ m.}$$

### → Height of Packing(Z)

For ceramic intalox saddle:

$$H_{OG} = 1.14 \frac{(G_m)^{0.316}}{(L_m)^{0.315}}$$

Where

$G_m$  = gas flow rate, lb moles/hr. ft<sup>2</sup>

$L_m$  = liquid flow rate, lbmol/hr.ft<sup>2</sup>

We have,

$$G_m = 763.71 \text{ Kgmol/hr m}^2$$

Since cross-section area =  $A = 4.11 \text{ m}^2$

$$G_m = 113.3 / 0.502 \text{ Kg mol/m}^2\text{hr} = 0.212 \text{ Kgmol/m}^2\text{sec}$$

Similarly,

$$L_m = 11944.35 / 4.11 \text{ Kgmol/m}^2\text{hr} = .807 \text{ Kgmol/sec. m}^2$$

$$H_{OG} = 1.14 \times \frac{(0.212)^{0.316}}{(0.807)^{0.315}}$$

$$H_{OG} = 0.752 \text{ m}$$

Where

$H_{OG}$  = height of a transfer unit

$$Z = H_{OG} \times N_{OG}$$

$$Z = 0.752 \times 10 = 7.52 \text{ m}$$

$$Z = 7.52 \text{ m}$$

Where Z is the height of packing.

Allowance for liquid distribution = 1.00 m

Allowance for liquid redistribution = 1.00 m

Total height of column = 1.00+1.00+7.52

Total height of column = 9.5 m  $\approx$  10 m

### → Degree of wetting

$$LP = \frac{\text{Liquid rate}}{\text{Specific are of packing}}$$

Liquid flow rate = 62.11 Kg/sec

And

Specific area of packing = 11 m<sup>2</sup>/m<sup>3</sup>

$$L_p = 62.11/4.11 \times 11 \times 990 = 1.39 \times 10^{-3} \text{ m}^3/\text{msec}$$

### → Calculation of pressure drop at flooding region:

Pressure drop at flooding point = 3 in of water / ft of packing.

Therefore,

$$\Delta P_{\text{flood}} = 3 \text{ in of water / ft of packing} \times \text{height of packing}$$

$$\begin{aligned} \Delta P_{\text{flood}} &= 24.67 \times 3 \\ &= 74.01 \text{ in of water} \end{aligned}$$

$$\Delta P_{\text{flood}} = 18.43 \text{ KN/m}^2$$

### → Calculation of liquid hold up :

25% of packing weight can be taken as the liquid hold up for ceramic packing as the bulk density of 2 inch intalox saddles packing is

$$= 609 \text{ Kg/m}^3$$

$$\text{Volume of packing} = \pi/4 D^2 \times H_p$$

$$= \pi/4 (2.3)^2 \times 7.52$$

$$= 31.24 \text{ m}^3$$

$$\text{thus liquid hold up is} = 0.25 \times 31.24 \times 609$$

$$= 4756.29 \text{ kg of water.}$$

## **Module # 7**

### **PROCESS DESIGN OF MASS TRANSFER COLUMN: DESIGN OF DISTILLATION AND ABSORPTION COLUMN**

#### **1. Introduction**

#### **2. Plate contractors**

- 7.1. Definition of tray areas
- 7.2. Plate types
  - 7.2.1. Bubble cap plates
  - 7.2.2. Valve plates
  - 7.2.3. Sieve plate
  - 7.2.4. Selection of tray type
- 7.3. Effect of vapor flow conditions on tray design
  - 7.3.1. Flooding consideration
  - 7.3.2. Sieve tray weeping
  - 7.3.3. Liquid entrainment
- 7.4. Tray hydraulic parameters
- 7.5. Column sizing approximation
- 7.6. Provisional plate design
  - 7.6.1. Column diameter
  - 7.6.2. Hole diameter, hole pitch and plate thickness
  - 7.6.3. Weir height and weir length
  - 7.6.4. Calming zones

#### **3. Stepwise design tray procedure**

#### **4. Design problem**

#### **References**

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# Lecture 1: Introduction

## 1. Introduction

The typical gas-liquid contacting operations include distillation, absorption, stripping, leaching and humidification. Distillation and absorption are two most widely used mass transfer processes in chemical industries. Design of plate column for absorption and distillation involves many common steps of calculation such as determination of number of theoretical plates, column diameter, plate hydraulic design, etc. In absorption process, a soluble component is absorbed in a liquid (called solvent) from a gaseous mixture. The gas and liquid streams leaving the tray are in equilibrium under the ideal condition. The separation in distillation is based on the relative volatility of the components. Additional vapor phase is generated by the vaporization of more volatile components (called stripping) and by condensation of relatively less volatile components (called absorption) adds to the liquid phase.

**Selection of column type: Plate or Packed:** Packed towers (columns) are also used as the contacting devices for gas absorption, liquid-liquid extraction and distillation. The gaseous mixture is allowed to contact continuously with the liquid counter-currently in a packed column. The liquid flows downward over the packing surface, and the gaseous mixture flows upward through the space in the packing. The performance of the column strongly depends on the arrangement of the packings to provide good liquid and gas contact throughout the packed bed. The solute gas is absorbed by the fresh solvent (liquid) entering at the top of the tower where the lean gas leaves system. The liquid enriched with absorbed solute gas, leaves the column bottom through the exit port.

In a plate tower, the liquid and gas are contacted in stage-wise manner on the trays; while gas-liquid contact is continuous in a packed column. There are always some uncertainty to maintain good liquid distribution in a packed tower. For this reason, it is difficult to accurately estimate the packed tower efficiency. The course content is limited to design of plate column only and some typical criteria for the selection of column type are discussed below.

- Plate towers exhibit larger pressure drops and liquid holdup at higher gas flow rate. While, packed towers are not appropriate for very low liquid flow rates. Packed column is the preferred choice than a plate column to handle toxic and flammable liquids due to lower liquid holdup to keep the unit as small as possible for the sake of safety.
- Plate columns are normally suitable for fouling liquids or laden with solids. They are easier to clean and could handle substantial temperature variation during operation.
- Packed towers are more suitable for foaming and corrosive services.
- It is easier to make the provision for the installation of internal cooling coils or withdrawal of side streams from a plate column.

## **2. Plate contractors**

Plate contractors/ towers are vertical cylindrical columns in which a vertical stack of trays or plates are installed across the column height as shown in Figure 7.1. The liquid enters at the top of the column and flows across the tray and then through a downcomer (cross-flow mode) to the next tray below. The gas/vapor from the lower tray flows in the upward direction through the opening/holes in the tray to form a gas-liquid dispersion. In this way, the mass transfer between the phases (gas/vapor-liquid) takes place across the tray and through the column in a stage-wise manner.

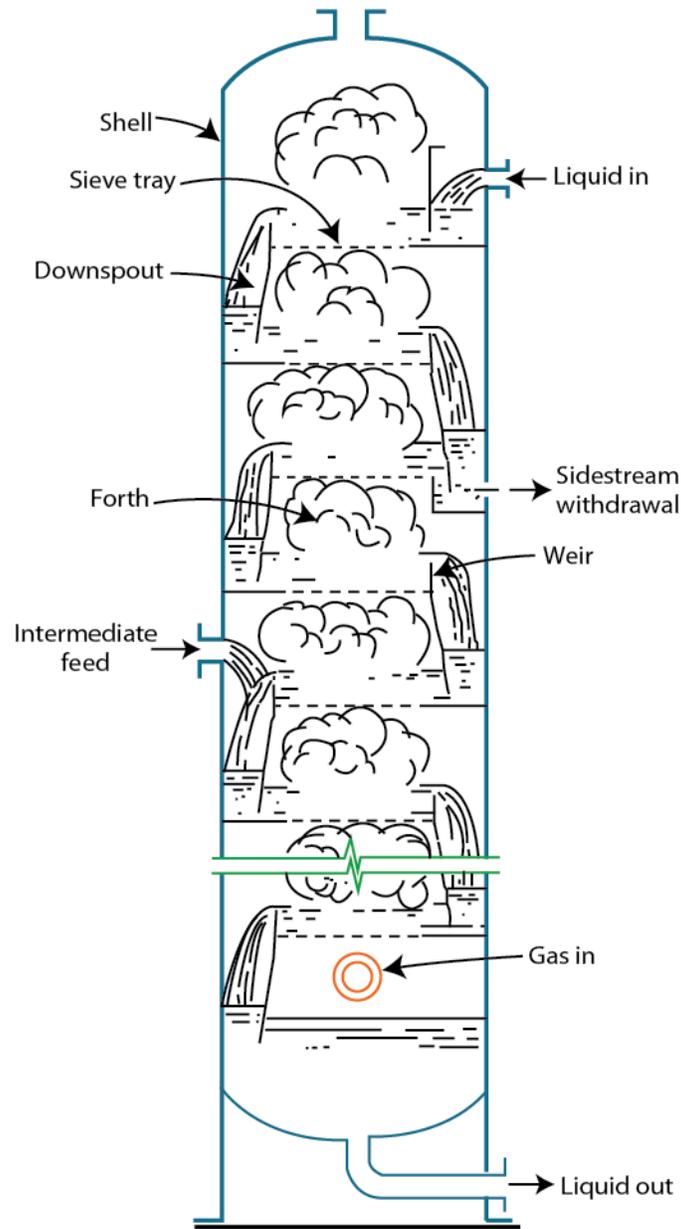


Figure 7.1. Schematic diagram of a plate contractor ([1] page 159).

**7.1. Definition of tray areas**

The definition of tray areas and its nomenclature illustrated in Figures 7.2 &7.3 are followed throughout the design procedure.

**Total tower cross-section area ( $A_T$ ):** The empty tower inside cross-sectional area without trays or downspouts.

**Net area ( $A_N$ ) (also called free area):**The total tower crosssectional area ( $A_T$ )minus the area at the top of the downcomer ( $A_{DT}$ ). The net area symbolizes the smallest area available for vapor flow in the inter-tray spacing.

**Bubbling area or active area ( $A_A$ ):** The total tower cross-sectional area minus sum of the downcomer top area( $A_{DT}$ ) and downcomer seal area ( $A_{DB}$ )and any other nonperforated areas on the tray. The bubbling area represents the area available for vapor flow just above the tray floor.

**Hole area ( $A_h$ ):** The total area of the perforations on the tray. The hole area is the smallest area available for vapor/gas passage.

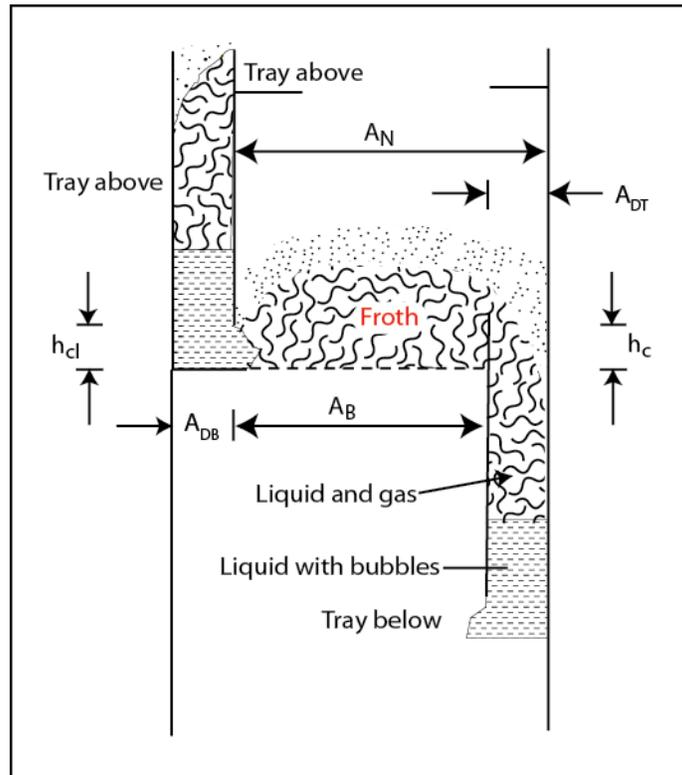


Figure 7.2.Schematic of a tray operating in the froth regime ([2] page 14-28).

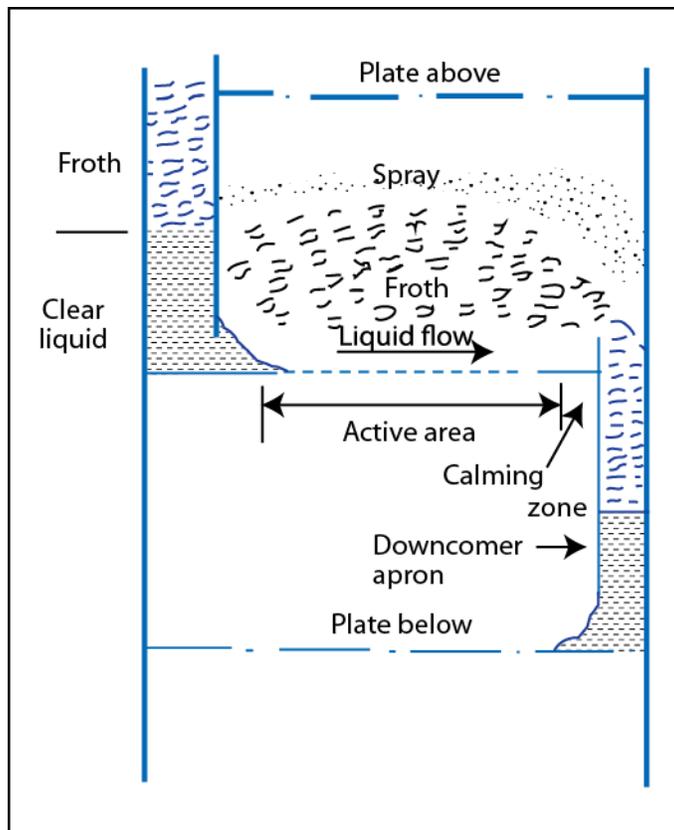


Figure 7.3. Typical cross-flow plate (sieve) ([3] page 557).

## 7.2. Plate types

Gas and liquid flow across the tray can either be by cross-flow or counter-flow manner (Figure 7.4). The cross-flow plates are most widely practiced and the three main types of cross flow plates are: bubble cap, valve and sieve trays with downcomer.

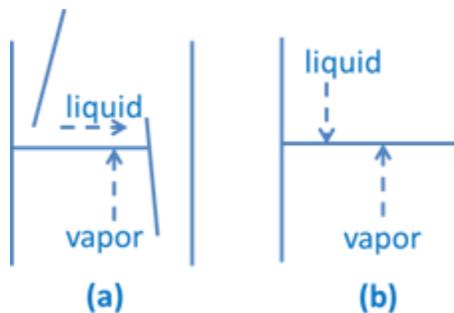


Figure 7.4. Classification of plate types based on flow mode- side view shown: (a) Cross-flow plate, (b): Counterflow plate.

### 7.2.1. Bubble cap plates

An enhanced gas-liquid contact can be achieved having bubble caps on the tray at very low liquid flow rates. A bubble cap consists of a riser (also called chimney) fixed to the tray through a hole and a cap is mounted over the riser (Figure 7.5). The gas flows up through the riser, directed downward by the cap through the annular space between riser and cap. Finally, the gas is dispersed into the liquid. A number of slots in the lower part of the cap help in gas bubble dispersion. Un-slotted types of cap designs are also common in application. Bubble caps are especially suitable for higher turndown ratio. Turndown ratio is the ratio of maximum operating vapor rate to the minimum allowable vapor rate, below which weeping starts.

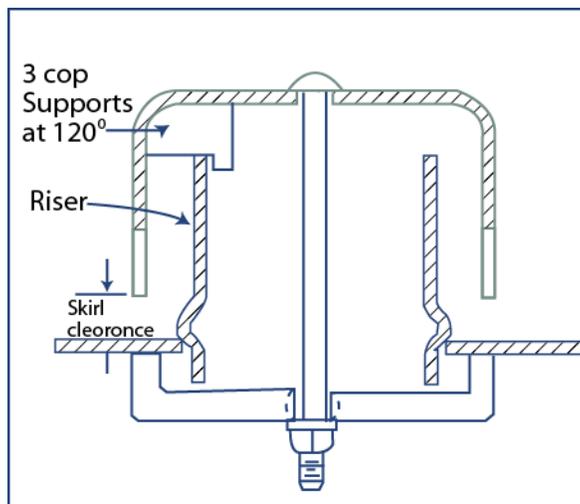


Figure 7.5. Bubble caps ([1] page 166).

### 7.2.2. Valve plates

Valve trays (or floating cap plate) are the modified design of sieve trays where relatively large plate perforations are covered by movable caps/valves (Figure 7.6). Valves cover may be round or rectangular. The very common hole diameter is 40 mm but upto 150 mm are also used. The valve lifts up as the vapor flow rate increases and the valve sits over the perforation at lower flow rate, thus stops the liquid from weeping. Valve trays provide good vapor-liquid contact at low flow rates (high turndown ratio).

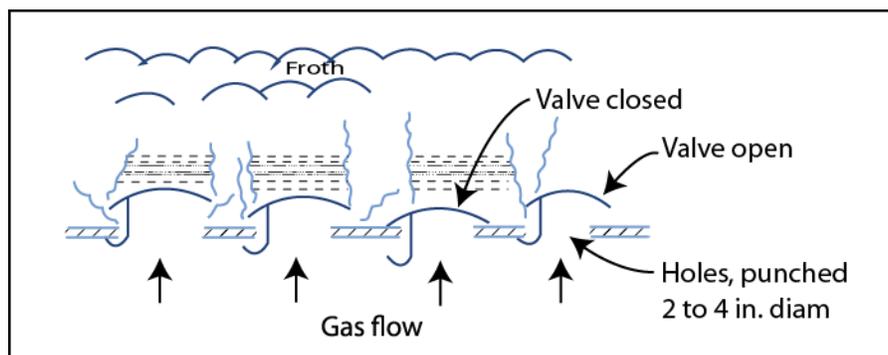


Figure 7.6. Valve tray ([4] page 14-25).

### 7.2.3. Sieve plate

The sieve tray (also known as perforated plate) is a flat perforated metal sheet (Figure 7.7). The hole diameter from 1.5 to 25 mm are very commonly used. The sieve tray layout is a typical square or equilateral triangular pitch holes. The gas/vapor flows upward through the perforation and disperses into the flowing liquid over the plate. There is no liquid seal in case of trays without downcomer and the liquid weeps (called weeping) through the holes at low flow rates, reducing the efficiency of plate. For this reason, sieve tray has the lowest turndown ratio. Sieve tray construction is simple and relatively cheap.

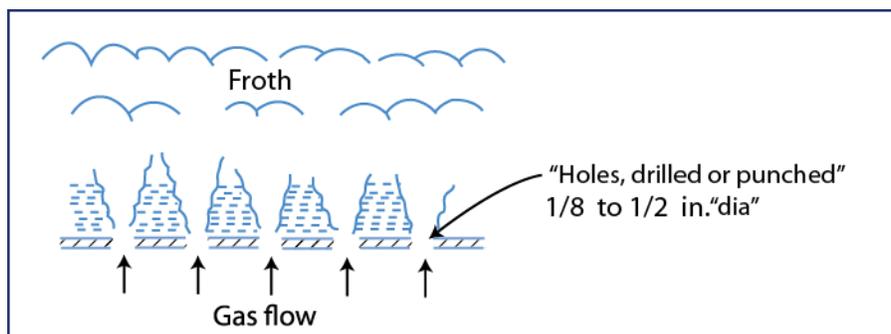


Figure 7.7. Sieve tray ([4] page 14-25).

### 7.2.4. Selection of tray type

The comparative performances of three common types of trays are summarized in Table 7.1.

The capacity, efficiency, pressure drop and entrainment of sieve and valve trays are almost same.

Bubble cap trays have lower capacity and efficiency and but higher pressure drop and entrainment compared to valve and sieve trays. The turndown ratio comes in the order of: bubble cap>valve>sieve. However, valve trays have the best turndown ratio in case of refinery applications. Sieve trays are the least expensive and suitable for almost all applications. Valve trays can be considered where higher turndown ratio is needed. Bubble cap trays should be used at very low liquid flow rate which is not achievable using sieve trays.

Table 7.1: Comparison of three types of cross-flow trays ([5] page 266).

Tray type	Capacity	Efficiency	Pressure drop	Entrainment	Turndown ratio	Cost
<b>Bubble cap</b>	Medium high	Medium high	High	~3 times than sieve tray	Excellent	100-200 % more than sieve tray
<b>Valve</b>	High to very high	High	Medium to high	Medium	4 to 10.1	20-50% more than sieve tray
<b>Sieve</b>	High	High	Medium	Medium	2.1	Cheapest of all types

## Lecture 2: Effect of Vapor Flow Conditions on Tray Design

### 7.3. Effect of vapor flow conditions on tray design

#### 7.3.1. Flooding consideration

Excessive liquid buildup inside the column leads to column flooding condition. The nature of flooding depends on the column operating pressure and the liquid to vapor flow ratio. It may be downcomer backup, spray entrainment or froth entrainment type floodings. Higher tray pressure drop due to excessive vapor flow rates holds up the liquid in the downcomer, increases the liquid level on the plate and leads to downcomer flooding situation. The column flooding conditions sets the upper limit of vapor velocity for steady operation.

Gas velocity through the net area at flooding condition can be estimated using Fair's correlation ([4], page 14-26):

$$U_{nf} = C_{sbf} \left( \frac{\sigma}{20} \right)^{0.2} \left( \frac{\rho_l - \rho_v}{\rho_v} \right)^{0.5} \quad [\text{m/s}] \quad (7.1)$$

$\rho_v$  = vapor density, kg/m<sup>3</sup>

$\rho_l$  = liquid density, kg/m<sup>3</sup>

$\sigma$  = liquid surface tension, mN/m (dyn/cm)

$C_{sbf}$  = capacity parameter (m/s) can be calculated ([4] page 14-27) in terms of plate

spacing and flow parameter  $F_{LG} = \frac{L}{V} \left( \frac{\rho_v}{\rho_l} \right)^{0.5}$

$$(7.2)$$

$L$  = liquid flow rate, kg/s

$V$  = vapor flow rate, kg/s

The design gas velocities ( $U_n$ ) is generally 80-85% of  $U_{nf}$  for non-foaming liquids and 75% or less for foaming liquids subject to acceptable entrainment and plate pressure drop.

### 7.3.2. Sieve tray weeping

Weeping occurs at low vapor/gas flow rates. The upward vapor flow through the plate perforations prevents the liquid from leaking through the tray perforation. At low vapor flow rates, liquid starts to leak/rain through the perforation (called weeping). When none of the liquid reaches the downcomer at extreme weeping condition at very low vapor flow rate, it is called dumping. The weeping tendency increases with increasing fractional hole area and liquid flow rates.

The vapor velocity at the weep point (where liquid leakage through holes starts) is the minimum value for stable operation. For a chosen hole area, the minimum operating vapor flow velocity ( $U_{min,op}$ ) at minimum flow rate for stable operation should be above weep point vapor velocity.

The minimum vapor velocity ( $U_{min}$ ) at the weep point ([3] page 569):

$$U_{min} = \frac{K_2 - 0.9(25.4 - d_h)}{\rho_v^{1/2}} \quad (7.3)$$

Where,  $d_h$  = hole diameter, mm,

$\rho_v$  = vapor density, kg/m<sup>3</sup> (maximum value of vapor density)

$K_2$  = constant ( $K_2$ ) of weep-point correlation depends on the depth of clear liquid (weir crest + weir height) on the plate ([3] page 571).

Weir crest ( $h_{wc}$ ) can be determined using the Francis' weir correlation ([3] page 571):

$$h_{wc} = 750 \left( \frac{L_{wc}}{L_W \rho_l} \right)^{2/3} \quad [\text{mm}] \quad (7.4)$$

$L_{WC}$  = weir length, m

$L_W$  = liquid flow rate over the crest, kg/s

$\rho_l$  = liquid density, kg/m<sup>3</sup>

Actual operating minimum vapor velocity:  $U_{min,op} = \frac{\text{minimum vapor flow rate}}{\text{hole area}} [\text{m/s}] \quad (7.5)$

To avoid weeping:  $U_{min,op} > U_{min}$ .

### 7.3.3. Liquid entrainment

Entrainment is the phenomena in which liquid droplets are carried by vapor/gas to the tray above. Therefore, the less volatile liquid components from bottom tray are mixed with liquid having relatively more volatile materials on the overhead tray. It counteracts the desired mass transfer operation and the plate efficiency decreases. Entrainment increases with vapor velocity. The fractional entrainment ( $\Psi = \frac{\text{kg}}{\text{kg gross liquid flow}}$ ) can be predicted using Fair's correlation in terms of the flow parameter [ $F_{LG} = \frac{L}{V} \left( \frac{\rho_v}{\rho_l} \right)^{0.5}$ ] and actual flooding velocity ([4] page 14-28).

Effect of  $\Psi$  on Murphree plate efficiency can be estimated using Colburn equation ([4] page 14-29):

$$E_a = \frac{E_{mv}}{1 + \frac{\Psi E_{mv}}{1 - \Psi}} \quad (7.6)$$

$E_{mv}$  =Murphree vapor efficiency

$E_a$ =Corrected Murphree vapor efficiency for liquid entrainment

### 7.4. Tray hydraulic parameters

#### Total plate pressure drop

All gas pressure drops ( $h_t$ ) are expressed as heads of the clear liquid and  $h_t$  is given by:

$$h_t = h_d + (h_{wc} + h_w) + h_r \quad (7.7)$$

Where,  $h_d$  =dry plate pressure drop, mm

$h_{wc}$  =height of liquid over weir (weir crest), mm

$h_w$  =weir height, mm

$h_r$  =residual head, mm

#### Dry plate pressure drop ( $h_d$ ):

Dry plate pressure drop occurs due to friction within dry short holes.  $h_d$  can be calculated using following expression derived for flow through orifices ([3] page 575).

$$h_d = 51 \left( \frac{U_{max}}{C_0} \right)^2 \frac{\rho_v}{\rho_l} \quad [\text{mm}] \quad (7.8)$$

$$\text{Maximum vapor velocity: } U_{max} = \frac{\text{Maximum volumetric vapor flow rate}}{A_H} \quad (7.9)$$

The orifice coefficient,  $C_0$  can be determined in terms of  $\frac{A_H}{A_p}$  and  $\frac{\text{plate thickness}}{\text{hole diameter}}$  ([3] page 576).

#### Residual gas pressure head ( $h_r$ ):

The residual pressure drop results mainly from the surface tension as the gas releases from a perforation. The following simple equation can be used to estimate  $h_r$  with reasonable accuracy ([3] page 575).

$$h_r = \frac{12.5 \times 10^3}{\rho_l} \quad (9.10)$$

#### Downcomer backup ( $h_b$ ) and downcomer residence time:

The liquid level and froth in the downcomer should be well below the top of the outlet weir on the tray above to avoid flooding ([3] page 576).

$$h_b = (h_{wc} + h_w) + h_t + h_{dc} \quad (7.11)$$

$$\text{Head loss in downcomer: } h_{dc} = 166 \left( \frac{L_{wd}}{\rho_l A_m} \right)^2 \quad (7.12)$$

$L_{wd}$  = Downcomer liquid flow rate, kg/s

$A_m$  = Smaller of clearance area under the downcomer apron ( $A_{ap}$ ) and downcomer area ( $A_D$ )

The average density of aerated liquid in the downcomer can be assumed as  $\frac{1}{2}$  of the clear liquid density. Therefore, half of the sum of the plate spacing and weir height should be greater than the downcomer backup.

$$\frac{1}{2} (\text{plate spacing} + \text{weir height}) > h_d \quad (7.13)$$

Downcomer residence time ( $t_{drt}$ ) should be sufficient for the disengagement of liquid and vapor in the downcomer to minimize entrained vapor. The value of  $t_{drt} > 3$  s is suggested. Downcomer residence time is given by ([3] page 578):

$$t_{drt} = \frac{A_D h_{bc} \rho_l}{L_{wd}} \text{ [s]}$$

(7.14)

$h_{bc}$  = clear liquid back up, mm

## Lecture 3: Plate Design

### 7.5. Column sizing approximation

The column sizing is a trial and error calculation procedure, starting with a tentative tray layout. The calculation is then revised until an acceptable design is obtained subject to satisfying the tray pressure drop, weeping, flooding and liquid entrainment limits. The column sizing is carried at the tray where the anticipated column loading is the highest and lowest for each section. However, the vapor flow rates have the highest impact on tower diameter. For an example, the sizing calculation is performed on the top tray for the above feed section and on the bottom tray for below feed section, for a single feed distillation column with one top and one bottom product. The tray spacing determines the column height. Lower tray spacing is desirable to minimize construction cost by checking against the column performance criteria. The suggested tray spacing ( $T_t$ ) with column diameter is appended below ([1] page 162). The detailed column sizing calculations are discussed in the solved example.

Tower diameter, m	Tray spacing, mm
1 or less	500 (150 mm is minimum)
1-3	600
3-4	750
4-8	900

### 7.6. Provisional plate design

#### 7.6.1. Column diameter

The column diameter is determined from the flooding correlation for a chosen plate spacing. The superficial vapor/gas velocity ( $U_{nf}$ ) at flooding through the net area relates to liquid and vapor densities according to Fair's correlation (**refer to section 7.3.1**).  $C_{sbf}$  is an empirical constant, depends on tray spacing and can be estimated against the flow parameter ( $F_{LG}$ ) based on mass flow rate of liquid ( $L$ ) and vapor ( $V$ ) ([3] page 567, [4] page 14-27).

Typically, the design velocity ( $U_n$ ) through the net area is about 80 to 85% of  $U_{nf}$  for non-foaming liquids and 75% or less for foaming liquid depending on allowable plate pressure drop and entrainment. It is a common practice to have uniform tower diameter in all sections of the column even though the vapor/gas and liquid loadings are expected to

be different to minimize the cost of construction. The uniformity in tower diameter may require selecting different plate spacing in different sections of the tower.

### 7.6.2. Hole diameter, hole pitch and plate thickness

The plate hole diameters ( $d_h$ ) from 3 to 12 mm are commonly used. The bigger sizes are susceptible to weeping. The holes may be drilled or punched and the plate is fabricated from stainless steel and other alloys than carbon steel. The centre to centre distance between two adjacent holes is called hole pitch ( $I_p$ ). Perforations can be arranged in square or equilateral triangular arrays with respect to the vapor/gas flow direction. The normal range of  $I_p$  is from 2.5 to 5 times of  $d_h$  ([1] page 168).

$$\text{For triangular pitch: } \frac{A_H}{A_P} = 0.907 \left( \frac{d_h}{I_p} \right)^2$$

(7.15)

Plate thickness ( $t_t$ ) typically varies from 0.2 to 1.2 times of the hole diameter and should be verified by checking the allowable plate pressure drop ([3] page 576).

### 7.6.3. Weir height and weir length

The depth of liquid on the tray is maintained by installing a vertical flat plate, called weir. Higher weir height ( $h_w$ ) increases the plate efficiency. But it increases plate pressure drop, entrainment rate and weeping tendency. Weir heights from 40 to 90 mm are common in applications for the columns operating above the atmospheric pressure. For vacuum operation,  $h_w=6$  to 12 mm are recommended. The weir length ( $L_w$ ) determines the downcomer area. A weir length of 60 to 80% of tower diameter is normally used with segmental downcomers. The dependency of  $L_w$  on downcomer area is calculated against the percentage value of  $\frac{A_D}{A_A}$  ([3] page 572).

### 7.6.4. Calming zones

Two blank areas called calming zone, are provided between the inlet downcomer or inlet weir and the perforation area, and also between the outlet weir and perforation area. Inlet calming zone helps in reducing excessive weeping in this area because of high vertical velocity of the entering liquid in the downward direction. Outlet calming zone allows disengagement of vapor before the liquid enters the downcomer area. A calming zone between 50 to 100mm is suggested.

### 3. Stepwise design tray procedure

Iterative tray design approach ([3] page 566) is listed below. The design is performed separately both above feed plate (top section) and below feed plate (bottom section) for single feed two product distillation column.

**Step #1:** Determine the number of theoretical plate and vapor and liquid flow-rates separately both in top and bottom sections.

**Step #2:** Obtain the physical properties of the system

**Step #3:** Select a trial plate spacing

**Step #4:** Estimate the column diameter based on flooding considerations

**Step #5:** Decide the liquid flow arrangement (reverse, single-pass, or multiple-pass). A guideline is provided in Figure 11.28 ([3] page 568).

**Step #6:** Make a provisional tray layout including downcomer area, active area, perforated area, hole area and size, weir height, weir length

**Step #7:** Check the weeping rate, if not satisfactory go back to step #6 and reselect tray layout

**Step #8:** Check the plate pressure drop, if too high return to step #6

**Step #9:** Check downcomer back-up, if too high go back to step #6 or #3

**Step #10:** Decide plate layout including calming zones and unperforated areas and check hole pitch, if unsatisfactory return to step #6

**Step #11:** Recalculate the percentage of flooding based upon selected tower diameter

**Step #12:** Check for entrainment, if too high then return to step #4

**Step #13:** Optimize design: repeat steps #3 to #9 to find smallest diameter and plate spacing acceptable to get the lowest cost for the specified application

**Step #14:** Finalize design: draw up the plate specification and sketch the layout

## Lecture 4: Hand on Design

### 4. Design problem

Design a continuous distillation column (plate) to recover acetone from a 50-50 mole % acetone-water mixture available at 30°C. The feed stream flow rate is 25,000 kg/h. The top product should contain at least 95 mole% acetone and the bottom product should contain <1 % acetone by mole. Consider reboiler as equivalent to one stage. This column is operated at atmospheric pressure (top tray). Column efficiency of 60% and pressure drop per plate of 1.25 kPa may be assumed. You can take the minimum liquid flow as 70% of the maximum rate both above and below the feed plate. The vapor liquid equilibrium (VLE) data for the acetone-water system at atmospheric pressure is provided in Table 7.2.

#### Data given:

Latent heat of water= 41,360 J/mol; latent heat of acetone= 28,410 J/mol

Specific heat of water=75.3 J/mol°C (mean); Specific heat of acetone 128 J/mol°C (mean)

Table 7.2: VLE data for the acetone-water system at 1 atm.

<i>x</i>	0.0	0.05	0.10	0.15	0.20	0.25	0.30	0.35	0.40	0.45	<b>0.50</b>	0.55	0.60	0.65	0.70	0.75	0.80	0.85	0.90	0.95
<i>y</i>	0.0	0.6381	0.7301	0.7716	0.7916	0.8034	0.8124	0.8201	0.8269	0.8376	<b>0.8387</b>	0.8455	0.8532	0.8615	0.8712	0.8817	0.8950	0.9118	0.9335	0.9627
BP, °C	100	74.8	68.53	65.26	63.59	62.6	61.87	61.26	60.75	60.35	<b>59.95</b>	59.54	59.12	58.71	58.29	57.9	57.49	57.08	56.68	56.3

*x*= Mole fraction of acetone in liquid; *y*= Mole fraction of acetone in vapor; BP: Bubble point

#### Step #1: Mass balance and determination of number of theoretical stage

Feed and products compositions:

<i>Component</i>	<i>Feed mole fraction</i>	<i>Top product mole fraction</i>	<i>Bottom product mole fraction</i>
Acetone	0.50	0.95	0.01
Water	0.50	0.05	0.99

Bubble point of feed (from the data shown in table) = 59.95°C

Latent heat of the feed = 28,410×0.5 + 41,360×(1 - 0.5) = 34,885 J/mol

Specific heat of the feed = (128×0.5) + 75.3× (1 - 0.5)= 101.75 J/mol °C

Heat required to vaporize 1 mole of the given feed =  $(59.95 - 30) \times 101.75 + 34,885 = 37933$  J

$$q = \frac{\text{Heat required to vaporize 1 mole of the given feed}}{\text{Latent heat of the feed}} = \frac{37933}{34885} = 1.09$$

$$\text{Slope of the } q\text{-line} = \frac{q}{q-1} = \frac{1.09}{1.09-1} = 12.44$$

Here, the top operating line just touches the equilibrium curve at the point of tangency of the rectifying section operating line at which the minimum reflux takes place.

From the Figure 7.8:  $\frac{x_D}{R_{\min} + 1} = 0.57$ ;  $R_{\min} = 0.67$  for  $x_D = 0.95$

Here, **reflux ratio**,  $R = 2.5 \times R_{\min} = 2.5 \times 0.67 = 1.675$  is taken for this design.

Average molecular wt. of feed =  $0.5 \times 58 + 0.5 \times 18 = 38$

Molar feed flow ( $F$ ) rate =  $25,000/38 = 657.9$  kmol/h

Acetone balance:  $D \times 0.95 = 657.9 \times 0.5 \Rightarrow D = 346.2$  kmol/h

Vapor flow ( $V$ ) rate above feed plate,  $V = D(1 + R) = 346.2(1 + 1.675) = 926.2$  kmol/h

(Assuming constant molar overflow)

Top section liquid flow rate,  $L = V - D = 580$  kmol/h

Bottom product:  $B = F - D = 657.9 - 346.2 = 311.7$  kmol/h

Mass balance below feed plate:  $L' = V' + B$

Slope of the bottom section operating line (Figure 7.8):  $\frac{L'}{V'} = 1.32$

$L'$  = Liquid flow rate below feed plate = 1285.7 kmol/h

$V'$  = Vapor flow rate below feed plate = 974 kmol/h

The construction of operating lines and number of theoretical stages are shown in this Figure 7.8.

**Total number of tray** = 6 (above feed) + 3 (below feed) = **9**

**Total number of real stages** =  $\frac{9-1}{0.6} \approx 14$  (60% column efficiency; reboiler was considered as equivalent to one theoretical tray)

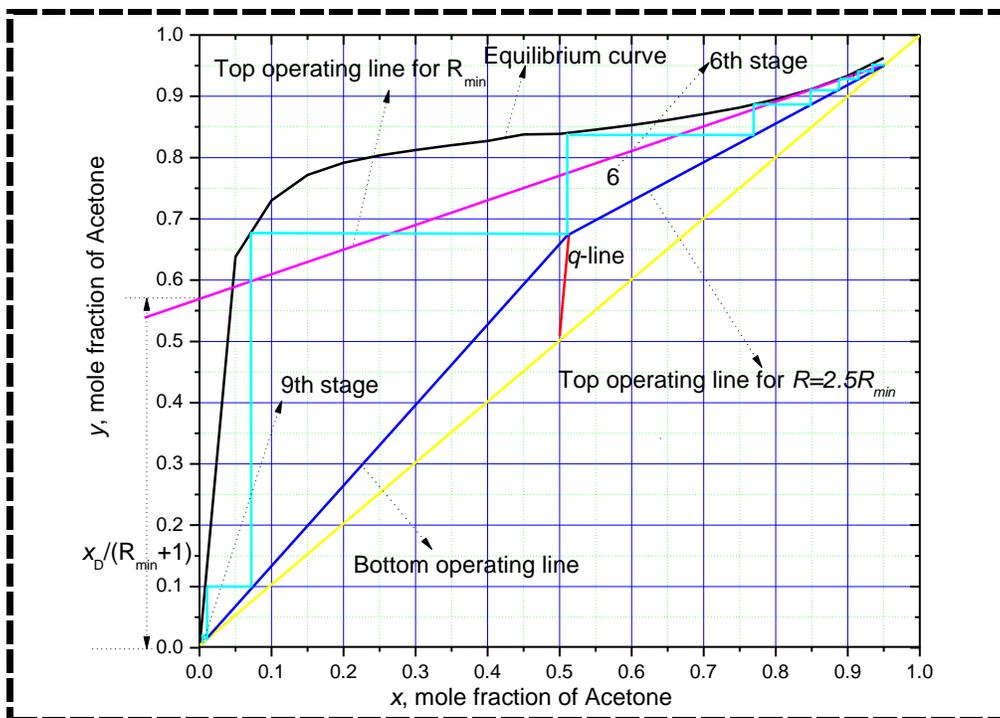


Figure 7.8: McCabe-Thiele construction.

### Step #2: Estimation of physical properties

Column top pressure= 101325 Pa (1 atm)

Column pressure drop= $1.25 \times 10^3 \times 14 = 16800$  kPa

Pressure drop of 1.25kPa per tray is specified

#### Top section:

Column top pressure= 101325 Pa (1.0147 bar) and temperature= 56.3 °C

$$\rho_v = \frac{PM}{RT} = \frac{101325 \times 56.5}{329.3 \times 8.314 \times 10^3} = 2.08 \text{ kg/m}^3$$

$\rho_l = 744 \text{ kg/m}^3$  (density of the mixture)

(water density= 985 and acetone density= 735  $\text{kg/m}^3$  at 56.3 °C)

Average molecular weight of vapor: M=56.5

Average molecular weight of liquid: M=56

Surface tension,  $\sigma = 20 \times 10^{-3} \text{ N/m}$

#### Bottom section:

Column bottom pressure=101325 +16800 = 118825 Pa (=1.19 bar)

Boiling point of water at 118825 Pa (1.19 bar)= 105 °C (bottom contains 99 mole % water)

From the steam table at 118825 Pa and 105 °C:  $\rho_v = 0.693$ ;  $\rho_l = 955 \text{ kg/m}^3$

Average molecular weight of vapor: M=40.58

Average molecular weight of liquid: M=18.4

	$\sigma = 58 \times 10^{-3} \text{ N/m}$
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## Lecture 5: Provisional Plate Design

### Step #3: Plate spacing

Plate spacing of 600 mm is considered for the first trial to calculate capacity parameter ( $C_{sbf}$ ) for the estimation of maximum allowable vapor velocity through the net plate area ([3] page 567, [4] page 14-27). The suggested plate spacing is 600 mm for column diameter > 1.5 m.

### Step #4: Column diameter (refer to sections 7.3.1 & 7.6.1; Eqs. 7.1 & 7.2)

**1<sup>st</sup> trial is started with the following considerations:**

- Design is performed for 80% flooding at maximum gas flow rate.
- Total downcomer top and bottom seal area is 10% of the net area.

Top section:	Bottom section:
Flow parameter ( $F_{LG}$ ) based on mass flow rate, $\frac{L}{V} \left( \frac{\rho_v}{\rho_l} \right)^{0.5} = \frac{580 \times 56}{926.2 \times 56.5} \left( \frac{2.08}{744} \right)^{0.5} = 0.033$	$F_{LG} = \frac{L'}{V'} \left( \frac{\rho_v}{\rho_l} \right)^{0.5}$ $= \frac{1285.7 \times 18.4}{974 \times 40.58} \left( \frac{0.693}{955} \right)^{0.5} = 0.016$
Capacity parameter ( $C_{sbf}$ ) = 0.12 m/s	$C_{sbf} = 0.11 \text{ m/s}$
Gas velocity through the net area at flooding: $U_{nf} = C_{sbf} \left( \frac{\sigma}{20} \right)^{0.2} \left( \frac{\rho_l - \rho_v}{\rho_v} \right)^{0.5} = 0.12 \times$ $\left( \frac{20}{20} \right)^{0.2} \left( \frac{744 - 2.08}{2.08} \right)^{0.5} = 2.26 \text{ m/s}$ [ $\sigma$ = liquid surface tension, mN/m]	$U_{nf} = C_{sbf} \left( \frac{\sigma}{20} \right)^{0.2} \left( \frac{\rho_l - \rho_v}{\rho_v} \right)^{0.5}$ $= 0.11 \times \left( \frac{58}{20} \right)^{0.2} \left( \frac{955 - 0.693}{0.693} \right)^{0.5}$ $= 5.05 \text{ m/s}$
The linear design gas velocity ( $U_n$ ) based on net area (80% flooding): $U_n = 0.8 \times 2.26 = 1.8 \text{ m/s}$	$U_n = 0.8 \times 5.05 = 4.04 \text{ m/s}$
The maximum volumetric vapor flow rate ( $Q_{max}$ ): $Q_{max} = \frac{V \times M}{\rho_v} = \frac{926.2 \times 56.5}{2.08}$ $= 25158.8 \text{ m}^3/\text{h} = 6.98 \text{ m}^3/\text{s}$	$Q_{max} = \frac{V \times M}{\rho_v} = \frac{974 \times 40.58}{0.693}$ $= 57034.5 \text{ m}^3/\text{h} = 15.84 \text{ m}^3/\text{s}$

<p>Net area required:</p> $\frac{Q_{max}}{U_n} = \frac{6.98}{1.8} = 3.88 \text{ m}^2$ <p>Totals tower cross-section area:</p> $\frac{3.88}{0.9} = 4.31 \text{ m}^2$ <p>(Total downcomer top and bottom seal area is 10% of the net area)</p> <p>Column (tower) diameter:</p> $\sqrt{\frac{4.31}{0.785}} = 2.34 \text{ m}$	$\frac{Q_{max}}{U_n} = \frac{15.84}{4.04} = 3.92 \text{ m}^2$ $= \frac{3.92}{0.9} = 4.36 \text{ m}^2$ <p>Column (tower) diameter:</p> $\sqrt{\frac{4.36}{0.785}} = 2.36 \text{ m}$
<p><b>Use the highvalue of the tower diameter for the uniformity between sections, if the difference is not greater than 20%. In this case, the bottom diameter is used both in top and bottomsections. Higher area than the design area (here top section) can be taken care by reducing the perforated area.</b></p> <p><b>The nearest recommended shell (nominal diameter 2400 mm) fabricated from carbon steel or stainless steel sheet in IS 2844-1964: ID 2403 mm with minimum wall thickness: 8 mm for carbon steel) and 6 mm for stainless steel.</b></p>	

#### Step #5: Selection of liquid-flow arrangement

Liquid volumetric flow rate in the top section =  $\frac{580 \times 56}{3600 \times 744} \approx 0.012 \text{ m}^3/\text{s}$

Liquid volumetric flow rate in the bottom section =  $\frac{1285.7 \times 18.4}{3600 \times 955} \approx 0.007 \text{ m}^3/\text{s}$

Therefore, single pass cross-flow sieve plate is chosen for this service ([3] page 568).

#### Step #6: Provisional plate design (refer to sections 7.6.2 & 7.6.3)

Column (tower) diameter (ID):  $D_T \approx 2.4 \text{ m}$

Column cross-section area:  $A_T = 0.785 \times D_T^2 = 4.52 \text{ m}^2$

Downcomer area:  $A_D = 0.1A_T = 0.452 \text{ m}^2$

Net area:  $A_N = A_T - A_D = 4.068 \text{ m}^2$

Weir Length ( $l_W$ ) =  $0.73 \times D_T = 1.752$  m ([3] page 573)

Weir height,  $h_w = 40$  mm is considered.

Active area:  $A_A = A_T - 2 \times A_D = 3.616$  m<sup>2</sup>

For the first trial, consider hole diameter:  $d_h = 12$  mm ( $\frac{1}{2}$  inch). The plate thickness = hole diameter is selected for the first trial.

**Step #7: Checking for weepage (refer to section 7.3.2; Eqs. 7.3 & 7.4)**

Top section	Bottom section
Maximum liquid flow rate ( $m_{max}$ ) = $\frac{580 \times 56}{3600} = 9.02$ kg/s	$m_{max} = \frac{1285.7 \times 18.4}{3600} = 6.571$ kg/s
Minimum liquid flow rate ( $m_{min}$ ) (70% of $m_{max}$ ) = $0.7 \times 9.02 = 6.32$ kg/s	$m_{min} = 0.7 \times 6.571 = 4.6$ kg/s
Maximum weir crest, $h_{wc} = 750 \left( \frac{L_{wc}}{L_W \rho_l} \right)^{2/3} = 750 \left( \frac{9.02}{1.752 \times 744} \right)^{2/3} = 27.2$ mm liquid height	$h_{wc} = 750 \left( \frac{6.571}{1.752 \times 955} \right)^{2/3} = 18.7$ mm liquid height
Minimum weir crest, $h_{wc} = 750 \left( \frac{L_{wc}}{L_W \rho_l} \right)^{2/3} = 750 \left( \frac{6.32}{1.752 \times 744} \right)^{2/3} = 21.5$ mm liquid height	$h_{wc} = 750 \left( \frac{4.6}{1.752 \times 955} \right)^{2/3} = 14.7$ mm liquid height
The constant ( $K_2$ ) of weep-point correlation = 30.3 at $h_{wc} + h_w = 40 + 21.5 = 61.5$ mm using minimum liquid flow rate ([3] page 571).	$K_2 = 30.2$ at $h_{wc} + h_w = 54.7$
The minimum vapor velocity ( $U_{min}$ ) at the weep point: $U_{min} = \frac{K_2 - 0.9(25.4 - d_h)}{\rho_v^{1/2}} = \frac{30.3 - 0.9(25.4 - 12)}{(2.08)^{1/2}} = 12.6$ m/s	$U_{min}$ at the weep point: $U_{min} = \frac{30.2 - 0.9(25.4 - 12)}{(0.693)^{1/2}} = 21.8$ m/s
Actual minimum vapor velocity at minimum vapor flow rate: $= \frac{\text{Actual vapor flow rate}}{A_H} = \frac{70 \% \text{ of } Q_{max}}{A_H} = \frac{0.7 \times 6.98}{0.489} = 10$ m/s	$= \frac{0.7 \times 15.84}{0.489} = 22.7$ m/s

**Therefore, the minimum operating velocity both in top and bottom sections is of above the weep point velocity.**

**Step #8: Plate pressure drop (refer to section 7.4; Eqs. 7.7, 7.8 & 7.10)**

Top section	Bottom section
Maximum vapor velocity: $U_{max} = \frac{Q_{max}}{A_H} = \frac{6.98}{0.489} = 14.3$ m/s	$U_{max} = \frac{Q_{max}}{A_H} = \frac{15.84}{0.489} = 32.4$ m/s
Maximum <b>dry plate pressure drop</b> : $h_d = 51 \left( \frac{U_{max}}{C_0} \right)^2 \frac{\rho_v}{\rho_l} = 51 \left( \frac{14.3}{0.88} \right)^2 \frac{2.08}{744} = 37.5$ mm liquid [The orifice coefficient, $C_0 = 0.88$ at $A_H/A_p = 15\%$ and $\frac{\text{plate thickness}}{\text{hole diameter}} = 1$ ([3] page 576).]	$h_d = 51 \left( \frac{32.4}{0.88} \right)^2 \frac{0.693}{955} = 50.2$ mm liquid
<b>Residual head</b> : $h_r = \frac{12.5 \times 10^3}{\rho_l} = \frac{12.5 \times 10^3}{744} = 16.8$ mm liquid	$h_r = \frac{12.5 \times 10^3}{\rho_l} = \frac{12.5 \times 10^3}{955} = 13$ mm liquid
Total plate pressure drop: $h_t = h_d + (h_{wc} + h_w) + h_r = 37.5 + (40 + 27.2) + 16.8 \approx 122$ mm liquid	$h_t = 50.2 + (40 + 18.7) + 13 \approx 122$ mm liquid
<b>The plate pressure drop of 1.25 kPa (=127 mm of water and 161 mm of acetone pressure) was assumed. The estimated value in the first trial is therefore acceptable.</b>	

**Step # 9: Downcomer backup liquid and downcomer residence time: (refer to section 7.4; Eqs. 7.11-7.13)**

Downcomer back: $h_b = (h_{wc} + h_w) + h_t + h_{dc}$ Head loss in downcomer: $h_{dc} = 166 \left( \frac{L_{wd}}{\rho_l A_m} \right)^2$ Downcomer liquid flow rate ( $L_{wd}$ ) = maximum liquid flow rate is taken $A_m$ is smaller of $A_{qp}$ and $A_D$ . $A_{qp} = h_{qp} l_w = 30 \times 10^{-3} \times 1.752 = 0.0525$ m <sup>2</sup> (typically $h_{qp} = h_w - 10$ ) Here, $A_{qp} < A_D = 0.452$ m <sup>2</sup>	
Top section	Bottom section
$h_{dc} = 166 \left( \frac{9.02}{744 \times 0.0525} \right)^2 = 8.8$ mm	$h_{dc} = 166 \left( \frac{6.571}{955 \times 0.0525} \right)^2 = 2.9$ mm
$h_b = (27.2 + 40) + 122 + 8.8 \approx 198$ mm	$h_b = (18.7 + 40) + 122 + 2.9 \approx 184$ mm

<p>Downcomer residence time:</p> $t_{dt} = \frac{A_D h_{bc} \rho_l}{L_{wd}} = \frac{0.452 \times (184 \times 10^{-3}) \times 955}{6.571} \approx 12 \text{ s}$ <p>&gt;3 s.</p> <p><math>h_{bc}</math> = clear liquid back up  (<math>h_{bc} \approx h_b</math> can be approximated)</p>	$t_{dt} = \frac{0.452 \times (184 \times 10^{-3}) \times 955}{6.571} \approx 12 \text{ s} > 3 \text{ s.}$
$\frac{1}{2}(\text{plate spacing} + \text{weir height}) = \frac{1}{2}(600 + 40) = 320 \text{ mm}$ $\frac{1}{2}(\text{plate spacing} + \text{weir height}) > h_d$ <p><b>Therefore, the plate spacing and downcomer residence time in both the sections meet the design required design criteria.</b></p>	

## Lecture 6: Provisional Plate Design

### Step #10: Calming zones and hole pitch (refer to sections 7.6.2 & 7.6.4; Eq. 7.15)

Perforated area ( $A_p$ ):  $A_p = A_A - A_{CZ} - A_{ES}$

Where,  $A_{CZ}$  = calming zone area (Figure 7.3).

$A_{ES}$  = area occupied by edge strip (Figure 7.9)

$$\frac{l_W}{D_T} = 0.73; \text{ now, } \theta_c = 95^\circ \text{ ([3] page 573)}$$

Angle subtended by the chord (edge plate),  $\alpha = 180^\circ - 95^\circ = 85^\circ$

The unperforated edge strip (edge plate) mean length from the geometry:

$$l_{ES} = (D_T - 50 \times 10^{-3}) \times \frac{\alpha \times \pi}{180} = (2.4 - 50 \times 10^{-3}) \times \frac{85 \times \pi}{180} = 3.49 \text{ m}$$

$$A_{ES} = 50 \times 10^{-3} \times l_{ES} = 0.175 \text{ m}^2$$

Use 50 mm wide calming zones. The approximate mean length of zones:

$l_{CZ}$  = Weir length ( $l_W$ ) + Width of unperforated edge strip

$$= 1.752 + 50 \times 10^{-3} = 1.802 \text{ m}$$

$$A_{CZ} = 2(50 \times 10^{-3} \times l_{CZ}) = 0.18 \text{ m}^2$$

Therefore, perforation area per tray ( $A_p$ ) =  $A_A - A_{CZ} - A_{ES} = 3.616 - 0.18 - 0.175 = 3.26 \text{ m}^2$

Take total hole area  $A_H = 0.15 A_A = 0.489 \text{ m}^2$

$$A_H = 0.785 \times d_h^2 \times n_h = 0.489 \text{ m}^2 \quad [\text{hole diameter } 12 \text{ mm}]$$

Number of holes ( $n_h$ ) = 4326

$$A_H / A_p = 0.15. \text{ For equilateral triangular pitch: } \frac{A_H}{A_p} = 0.907 \left( \frac{d_h}{l_p} \right)^2$$

This corresponds to hole-pitch to hole diameter ratio of  $(l_p / d_h) = 2.46$ . This is very close to the normal range of 2.5 to 4.0 times of hole diameter.

The estimated hole pitch ( $l_p$ ) = 29.5 mm.

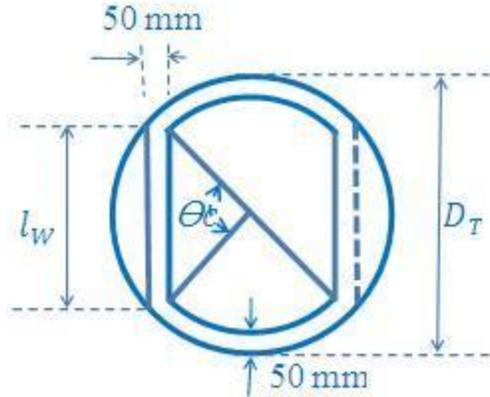


Figure 7.9. Angle subtended by the chord ([3] page 583).

**Steps # 11 and 12: Entrainment checking (refer to section 7.3.3; Eq. 7.6)**

Top section	Bottom section
<p>Actual vapor velocity (<math>U_v</math>) based on net area (<math>A_N</math>) selected provisionally:</p> $U_v = \frac{Q_{max}}{A_N} = \frac{6.98}{4.068} = 1.7 \text{ m/s}$ <p>% flooding = <math>\frac{U_v}{U_{nf}} \times 100 = \frac{1.7}{2.26} \times 100 = 76 \%</math></p> <p>The <b>fractional entrainment</b>, <math>\Psi = 0.09</math> at <math>F_{LG} = \frac{L}{V} \left( \frac{\rho_v}{\rho_l} \right)^{0.5} = 0.033</math> and actual flooding velocity of 76 % ([4] page 14-28).</p> <p>Effect of <math>\Psi</math> on Murphree plate efficiency can be estimated from ([4] page 14-29):</p> $E_a = \frac{E_{mv}}{1 + \frac{\Psi E_{mv}}{1 - \Psi}} = 0.57$ <p><math>E_{mv} = 0.6</math> (Murphree vapor efficiency 60%)  <math>E_a</math> = Murphree vapor efficiency, corrected for liquid entrainment</p>	<p><math>U_v = \frac{Q_{max}}{A_N} = \frac{15.84}{4.068} = 3.9 \text{ m/s}</math></p> <p>% flooding = <math>\frac{U_v}{U_{nf}} \times 100 = \frac{3.9}{5.05} \times 100 = 77\%</math></p> <p><math>\Psi = 0.18</math> at <math>F_{LG} = 0.016</math> and actual flooding velocity of 77 % ([4] page 14-28).</p> $E_a = 0.53$
<p><b>The actual flooding is below the design flooding value of 80%. Usually, <math>\Psi &lt; 0.1</math> is desirable. However, the optimum design value may be above this.</b></p>	

**Design problem: Absorption column**

An industrial gas stream is available @ 2 kg/s from a cracking operation of  $\text{NH}_3$  containing 72%  $\text{H}_2$ , 24%  $\text{N}_2$  and 4%  $\text{NH}_3$  by mole, at 202.65 kPa and 35°C. You have been asked to design a multistage countercurrent bubble cap absorber to remove  $\text{NH}_3$  from the above stream with water as the scrubbing liquid. The liquid mass flow rate is limited to be 2 to 3.5 times of gas mass rate.  $\text{NH}_3$  concentration should not be greater than 0.003 mg per  $\text{m}^3$  of the exit gas.

**Assumptions/ design considerations:**

- Lean water- $\text{NH}_3$  system follows Henry's law and the corresponding equilibrium relation:  $y^* = 0.85x$  @ 30°C
- Isothermal gas absorption at room temperature (~30°C)
- Optimum adsorption factor ( $A$ ) = 1.2 to 2
- Overall column efficiency = 70%
- Pressure drop per plate = 1 kPa
- Minimum liquid loading = 70% of expected maximum loading

**Hints:** The flow rates of liquid and gas entering and leaving the absorber is almost constant throughout the column if a small amount of the solute gas is absorbed. This is a typical case, also common in practice when the solute gas concentration in the feed stream is low (dilute gas absorption). For such operations, the variation of temperature between column top and bottom trays is insignificant (~isothermal operation). The pressure drop, if the column is not too tall, has minor effect on the physical properties of process fluids that could influence the column design. The section-wise determination of number trays and design are performed if the gas stream and/ or the solvent liquid (usually makeup solvent) are introduced at any intermediate point of the column. The number of theoretical trays can be estimated using either Kremser equation or graphical technique for multistage counter current lean gas absorption ([1] page 290). The vapor loading is the highest at the bottom tray even though its variation is not appreciable. For the safe side, the design is usually performed at the bottom tray.

The mole fractions ( $x, y$ ) between the phases are plotted in McCabe Thiele method of distillation calculation. In case of absorption, the mole ratios ( $X = \frac{x}{1-x}, Y = \frac{y}{1-y}$ ) are used for the determination of number of trays in graphical method instead of mole fractions.

### References

- [1]. Robert E. Treybal, Mass Transfer Operations, McGraw-Hill, Inc., 3<sup>rd</sup> ed. 1981.
- [2]. Perry's Chemical Engineers' Handbook, McGraw-Hill, Inc., 8<sup>th</sup> ed. 1997.
- [3]. R. K. Sinnott, Coulson & Richardson's Chemical Engineering: Chemical Engineering Design (vol. 6), Butterworth-Heinemann, 3<sup>rd</sup> ed. 1999.
- [4]. Perry's Chemical Engineers' Handbook, McGraw-Hill Companies, 7<sup>th</sup> ed. 1997.
- [5]. Henry Z. Kister, Distillation Design, McGraw-Hill, Inc., 1st ed. 1992.

# Process Flow sheet diagram

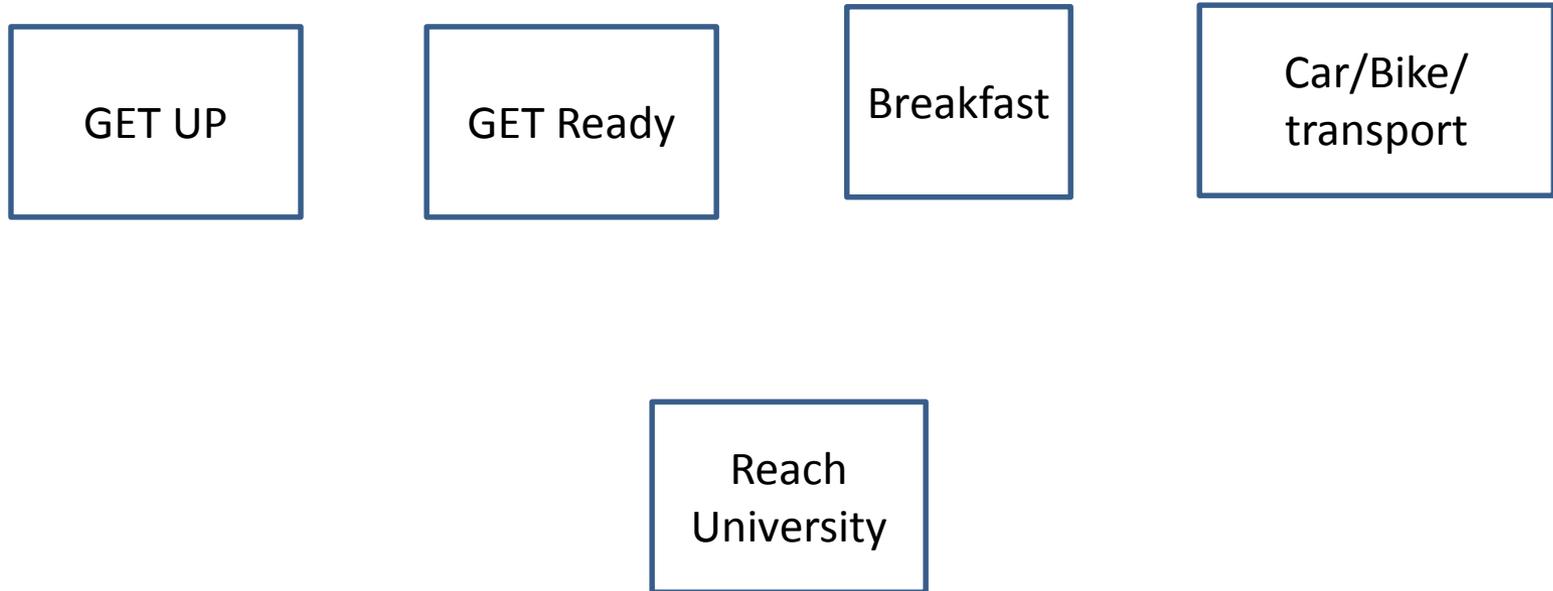


A picture / figure is  
better than 1000  
words

# Process Description

- Any physical or chemical process can be explained by the help of description or by use of blocks, or pictorially.
- Pictorial representation of a process makes it easier to understand, and presentation of process by pictures uses lesser sheets and words.

# A student coming to university can be represented by a diagram



- In a Similar way any process can be represented described by the help of diagram or picture

Try to describe any process with  
help of a picture/diagram

# Flow sheet

- The flow sheet is a key document in process design. it shows the arrangement of the equipment selected to carry out the process; the stream connections; stream flow-rate and compositions; and the operating conditions.
- It is a diagrammatic model of the process.
- This will include piping, instrumentation, and equipment design and plant layout

# Types of process flow diagrams

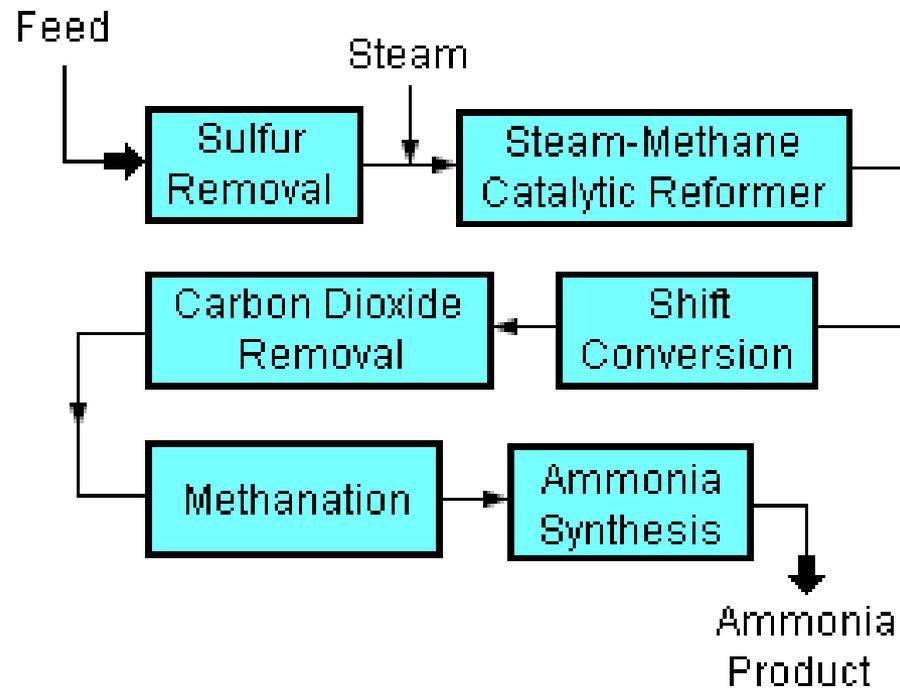
- 
- There are 3 types of process flow diagram in chemical industry.
- Block diagram.
- Process flow diagram(PFD)
- Process and instrumentation diagram(PID)

# Block diagrams

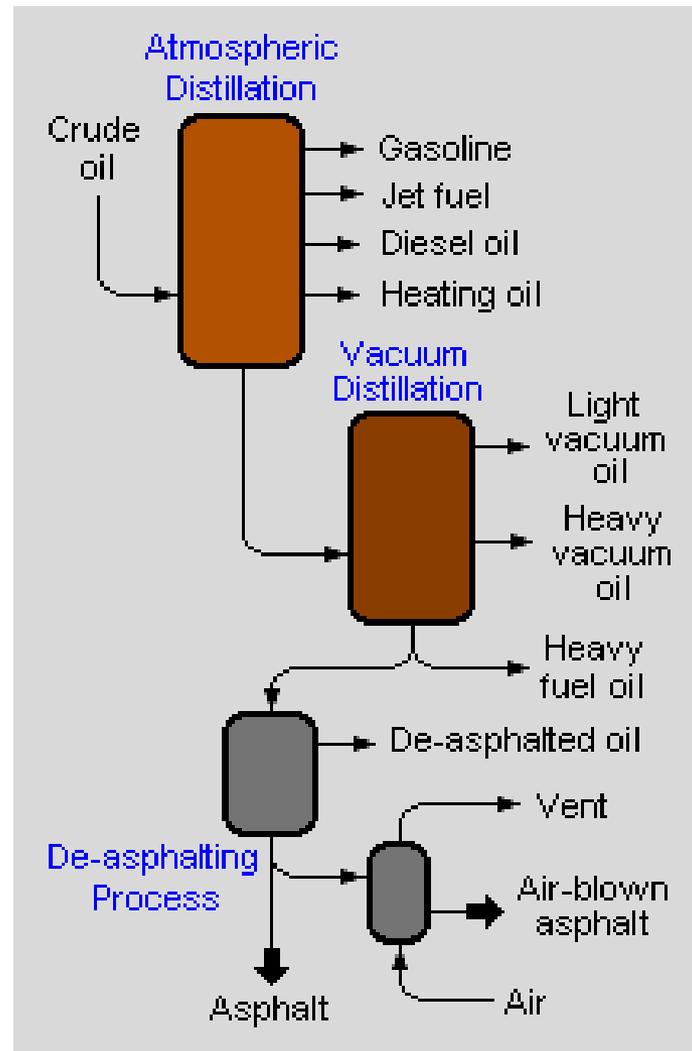
- It is a process description and flow direction depicted in the form of blocks. Every unit operation is represented as a block. In this you have no need to show any controls.
- Block diagrams are useful for representing a process in a simplified form in reports and textbooks, but have only a limited use as engineering documents.
- The block can be of any shape, but it is usually convenient to use a mixture of squares and circles, drawn with a template.

# Examples:

- Below are two examples of very simple process block flow diagrams. One is a diagram of the processes used to produce [ammonia](#) and the other is a diagram of a process for producing [petroleum asphalt](#):



Block flow diagram of steam-methane reforming and other processes used in producing liquid ammonia



Block flow diagram of asphalt production in a petroleum refinery

# Process flow diagram (PFD)

- In this u can simply replace the blocks in the Block Diagram and addition to it you have to show some major controls which affects the process more or without which plant can be prone to accidents. You have to mention Mass and Energy balance on the same sheet. You can mention stream no. and details of these streams on PFD.
- The PFD displays the relationship between major equipment of a plant facility and does not show minor details such as piping details and designations. Another commonly-used term for a PFD is a flow sheet

# A process flow sheet includes

- Process piping
- Major bypass and recirculation lines
- Major equipment symbols, names and identification numbers
- Flow directions
- Control loops that affect operation of the system
- Interconnection with other systems

# Process flow diagrams generally do not include:

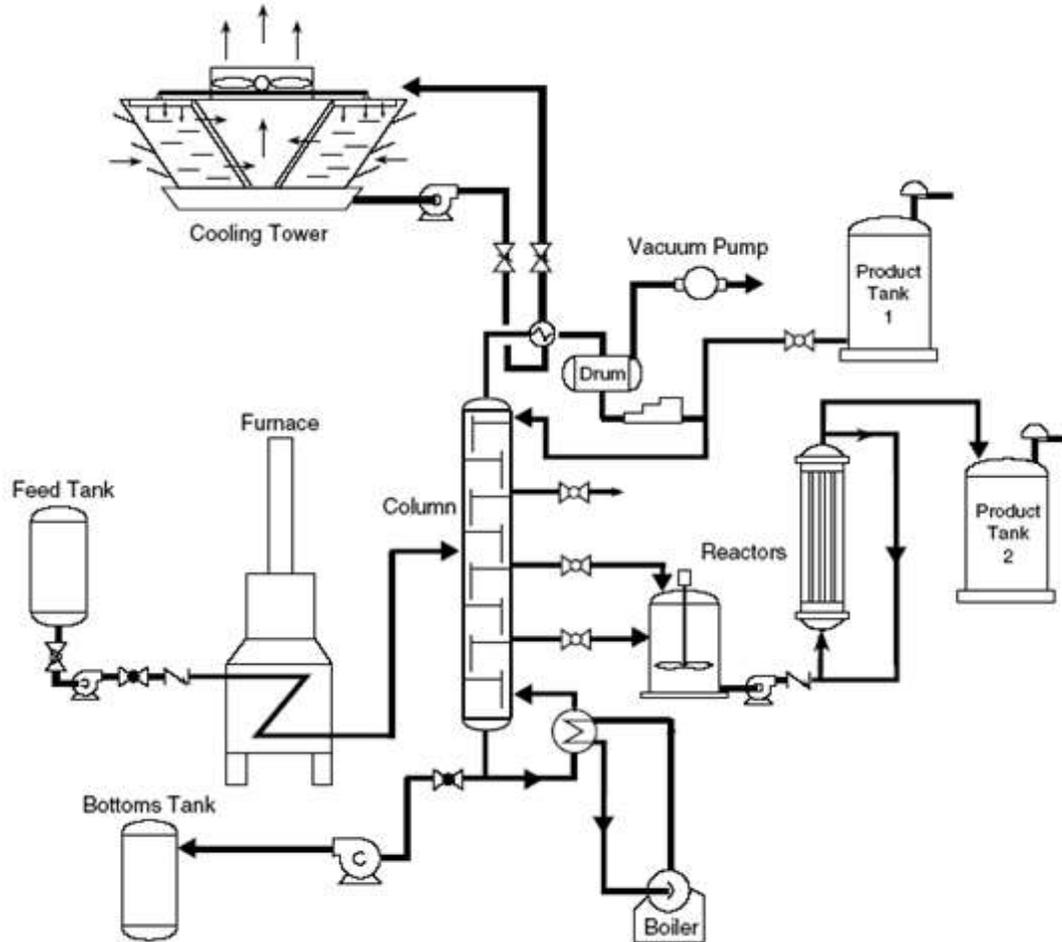
- 
- Pipe classes or piping line numbers
- Process control instrumentation (sensors and final elements)
- Minor bypass lines
- Isolation and shutoff valves
- Maintenance vents and drains
- Relief and safety valve
- Flanges

# Process Flow Diagram

*The goal is to present the most amount of information with the least amount of effort on the part of the reader.*

- The flowsheet should generally flow from left to right.*
- The flowsheet should not be cluttered - use multiple sheets.*
- The flowsheet should be in landscape with the bound edge at top.*
- The equipment should be drawn in approximately relative size, e.g. towers larger than drums, exchangers larger than pumps etc.*
- The major towers and reactors are generally on one, or nearly one, level.*
- The reader should be able to follow it with his or her eye.*
- The streams should have the minimum of direction changes.*
- The streams that enter across the battery limits should be on the left.*
- The streams that leave across the battery limits should be on the right.*
- The streams that move to the next sheet should leave on the right.*
- The streams that recycle to earlier sheets should leave on the left.*

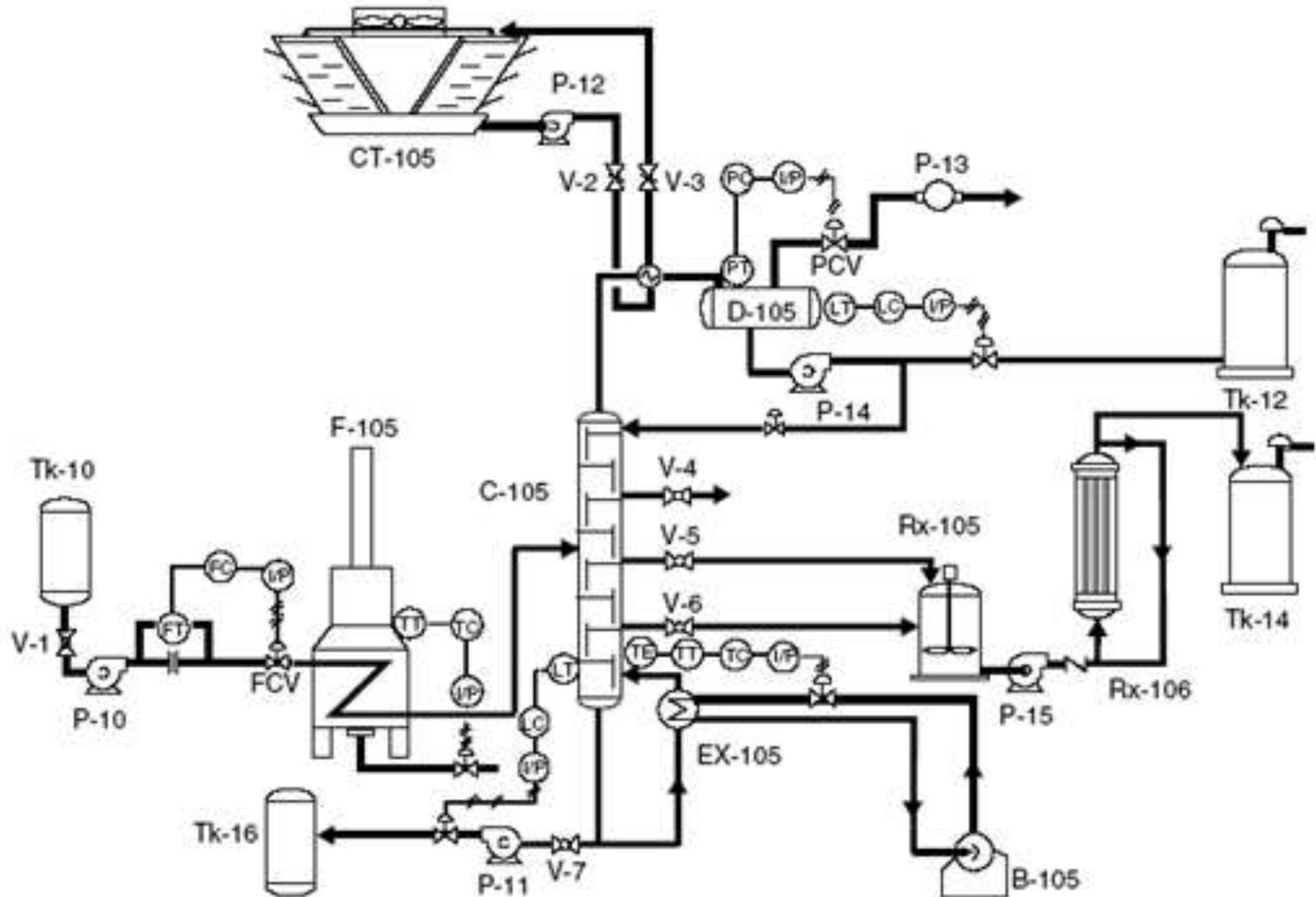
# Example of a refinery operation



# Process and instrumentations diagrams (PID)

- 
- This is the next step to PFD. Where you can show all the controls, details of equipments like size, design pressure/temp, insulation, nozzle details etc. You have to mention line sizes. It is the basic document and considered as basis for the construction of plant. it is often called the *Engineering Flow-sheet or Engineering Line Diagram*
- 
- The P & I diagram shows the arrangement of the process equipment, piping, pumps, instruments, valves and other fittings. It should include:
  - All process equipment identified by an equipment number. The equipment should be drawn roughly in proportion, and the location of nozzles shown.
  - All pipes, identified by a number. the pipe size and material of construction should be shown.
  - All valves control and block valves, with an identification number the type and the size should be shown. The type may be shown by the symbol used for the valve or included in the code used for valve number.

# PID of refinery



## **Plant Design CHEN 451**

Engineering design of new chemical and petrochemical plants and the expansion or revision of existing ones require the use of engineering principles and theories combined with a practical realization of the limits imposed by industrial conditions. A successful engineer needs more than a knowledge and understanding of the fundamental sciences and the related engineering subjects such as thermodynamics, reaction kinetics, and computer technology. The engineer must also have the ability to apply this knowledge to practical situations for the purpose of accomplishing something that will be beneficial to society. There are three parameters that must be defined namely:

**1-Design:** design is a creative activity and is defined as the synthesis, the putting together of ideas to achieve a desired purpose. Also it can be defined as the creation of manufacturing process to fulfill a particular need. The need may be public need or commercial opportunity.

**2-Process Design:** process design establishes the sequence of chemical and physical operations; operating conditions; the duties, major specifications, and materials of construction (where critical) of all process equipment (as distinguished from utilities and building auxiliaries); the general arrangement of equipment needed to ensure proper functioning of the plant; line sizes; and principal instrumentation. The process design is summarized by a process flowsheet.

Process design is intended to include:

1. Flowsheet development.
2. Process material and heat balances.
3. Auxiliary services material and heat balances (utilities requirements).
4. Chemical engineering performance design for specific items of equipments required for a flowsheet.
5. Instrumentation as related to process performance.
6. Preparation of specifications (specification sheets) in proper form for use by the project team as well as for the purchasing function.
7. Evaluation of bids and recommendation of qualified vendor.

**3-Plant Design:** includes items related directly to the complete plant, such as plant layout, general service facilities, and plant location.

## Design Development Stages:

The stages in the development of a design, from the initial identification of the objectives to the final design are shown in Fig.(1).

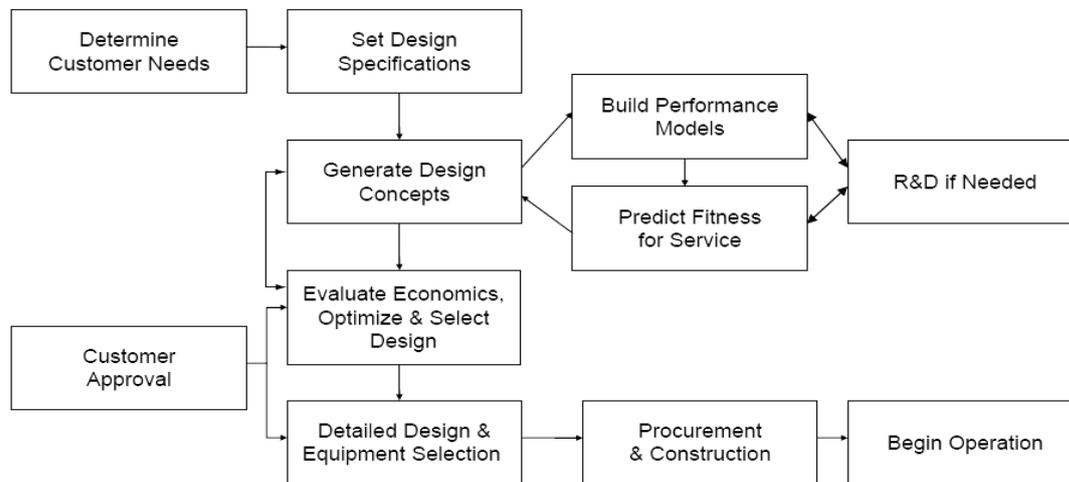


Fig.(1) The design process.

### (I) The Design Objectives (The Need)

Engineering projects can be divided into three types:

- A. New process development.
- B. New production capacity to meet growing sales.
- C. Modification and addition to existing plant.

In the design of a chemical process the need is the public need for the product, the commercial opportunity as foreseen by the sales and marketing organization.

### (II) Setting The Design Basis (Data Collection)

The most important step in starting a process design is translating the customer need into a design basis. The design basis is a more precise statement of the problem that is to be solved. It will normally include the production rate and purity specifications of the main product, together with information on constraints that will influence the design, such as:

1. Information on possible processes and the system of units to be used.
2. The national, local or company design codes that must be followed.
3. Details of raw materials that are available.
4. Information on potential sites where the plant might be located, including climate data, seismic conditions, and infrastructure availability.
5. Information on the conditions, availability, and price of utility services such as fuel (gas), steam, cooling water, process air, process water, and electricity, that will be needed to run the process.

### **(III) Generation of Possible Design Concepts (Solutions)**

It is the creative part of the design process. This part is concerned with the generation of possible solutions for analysis, evaluation, and selection (ways of meeting objective problems).

Source of solutions:

- a- Past experiences.
- b- Tried and tested methods.

### **(IV) Build Performance Model and Fitness Testing**

When design alternatives are suggested, they must be tested for fitness of purpose. In other words, the design engineer must determine how well each design concept meets the identified need. In the field of chemical engineering, it is usually prohibitively expensive to build several designs to find out which one works best (a practice known as “**prototyping**” which is common in other engineering disciplines). Instead, the design engineer builds a mathematical model of the process, usually in the form of computer simulations of the process, reactors, and other key equipment. In some cases, the performance model may include a pilot plant or other facility for predicting plant performance and collecting the necessary design data.

The design engineer must assemble all of the information needed to model the process so as to predict its performance against the identified objectives. For process design this will include information on possible processes, equipment performance, and physical property data

If the necessary design data or models do not exist, then research and development work is needed to collect the data and build new models. Once the data has been collected and a working model of the process has been established, then the design engineer can begin to determine equipment sizes and costs. At this stage it will become obvious that some designs are uneconomical and they can be rejected without further analysis. From this step a few candidate designs that meet the customer objective are identified.

### **(V) Economic Evaluation, Optimization, and Selection**

Once the designer has identified a few candidate designs that meet the customer objective, then the process of design selection can begin. The primary criterion for design selection is usually economic performance, although factors such as safety and environmental impact may also play a strong role. The economic evaluation usually entails analyzing the capital and operating costs of the process to determine the return on investment (R.O.I).

The economic analysis of the product or process can also be used to **optimize the design**. Every design will have several possible variants that make economic sense under certain conditions. For example, the extent of process heat recovery is a tradeoff between the cost of energy and the cost of heat exchangers (usually expressed as a cost of heat exchange area). In regions where energy costs

are high, designs that use a lot of heat exchange surface to maximize recovery of waste heat for reuse in the process will be attractive. In regions where energy costs are low, it may be more economical to burn more fuel and reduce the capital cost of the plant.

When all of the candidate designs have been optimized, the best design can be selected. Very often, the design engineer will find that several designs have very close economic performance, in which case the **safest design** or that which has the best commercial track record will be chosen. At the selection stage an experienced engineer will also look carefully at the candidate designs to make sure that they are safe, operable, and reliable, and to ensure that no significant costs have been overlooked.

### **(VI) Detailed Design and Equipment Selection**

Here the detailed specifications of equipment such as vessels, exchangers, pumps, and instruments are determined. During the detailed design stage there may still be some changes to the design, and there will certainly be ongoing optimization as a better idea of the project cost structure is developed. The detailed design decisions tend to focus mainly on equipment selection though, rather than on changes to the flowsheet. For example, the design engineer may need to decide whether to use a U-tube or a floating-head exchanger, or whether to use trays or packing for a distillation column.

### **(VII) Procurement, Construction, and Operation**

When the details of the design have been finalized, the equipment can be purchased and the plant can be built. Procurement and construction are usually carried out by an EPC firm (Engineering, Procurement, and Construction) unless the project is very small. Because they work on many different projects each year, the EPC firms are able to place bulk orders for items such as piping, wire, valves, etc., and can use their purchasing power to get discounts on most equipment. The EPC companies also have a great deal of experience in field construction, inspection, testing, and equipment installation. They can therefore normally contract to build a plant for a client cheaper (and usually also quicker) than the client could build it on its own. Finally, once the plant is built and readied for **startup**, it can begin operation. The design engineer will often then be called upon to help resolve any startup issues and teething problems with the new plant.

## Design Constraints

When considering possible ways of achieving the objective the designer will be constrained by many factors which are called the design constraints.

Design constraints are divided into two types Fig.(2):

- A. Internal constraints: over which the designer has some control.
- B. External constraints: fixed, invariable.

**What is meant by design constraints? What are the different types? Give examples?**

**What are the main items that should be included in the process design?**

**Draw a block diagram showing the main steps involved in the development of a design process?**

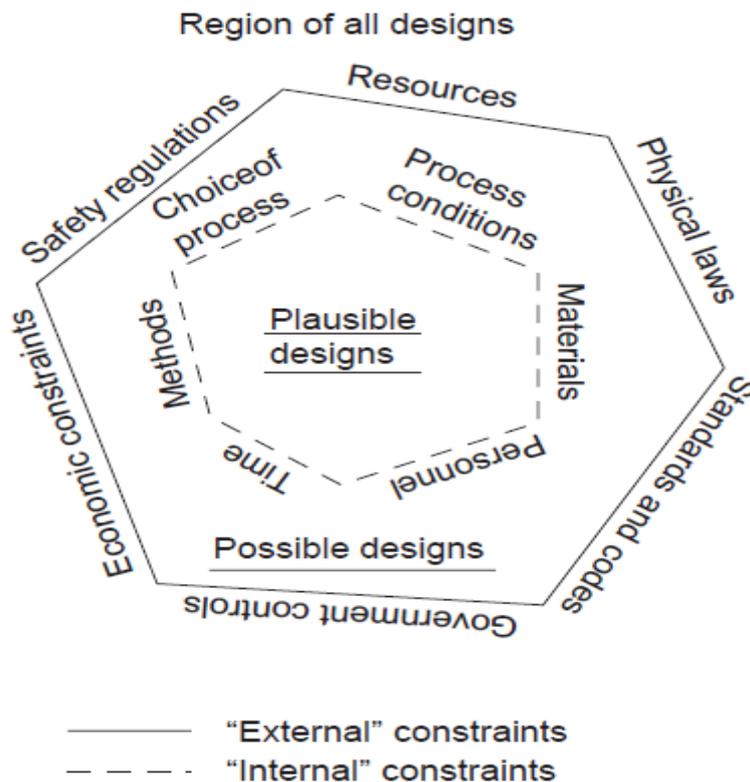


Fig.(2) Design constraints.

## Flow-sheeting (special language conveying information)

Process design normally starts with a process scheme (flowsheet). The flowsheet is the key document or road map in process design. It's a diagrammatic model of the process describe the process steps in a proper sequence using symbols to represent the various components (equipment, lines, and control instrumentation) that make up the unit.

### **The Flow-sheet Importance**

- Shows the arrangement of the equipment selected to carry out the process.
- Shows the streams concentrations, flow rates & compositions.
- Shows the operating conditions.
- During plant start up and subsequent operation, the flow sheet from a basis for comparison of operating performance with design. It's also used by operating personnel for the preparation of operating manual and operator training.

## Flowsheet Presentation

### **1- Block diagram**

- Represent the process in a simplified form.
- No details involved.
- Don't describe how a given step will be achieved.

### **When is it used?**

- In survey studies.
- Process proposal for packaged steps.
- Talk out a processing idea.

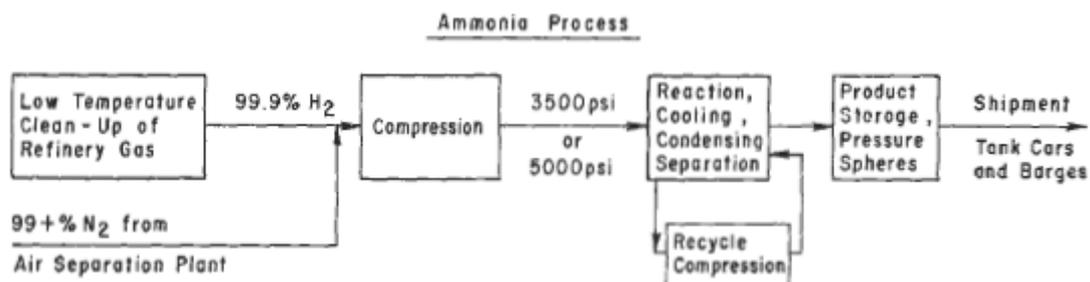


Fig.(3) Block diagram.

## 2- Pictorial Flow Sheet

The equipments are normally drawn in a stylized pictorial form. For tender documents or company brochures actual scale drawing of the equipment are sometimes used.

### Types of pictorial flow-sheets

#### a) Process Flow Diagram (PFD)

A PFD is a simplified flow diagram of a single process unit, a utility unit, a complete process module. The purpose of a PFD is to provide a preliminary understanding of the process system indicating only the main items of equipment, the main pipelines and the essential instruments, switches and control valves.

A PFD also indicates operating variables, such as mass flow, temperatures and pressures, which are tabulated at various points in the system.

The PFD is a document containing information on:

- Process conditions and physical data of the main process streams.
- Main process equipment with design data.
- Main Process lines.
- Mass (material) balance.
- Heat balance (if applicable).

**NOTE:** If the PFD doesn't contain any data about the flow rates, it is called a **qualitative flowsheet**, while if the flow rates are involved the PFD is called a **combined flowsheet** in which qualitative information and quantitative data are combined on the basis of one flowsheet.

#### b) Piping and Instrumentation Diagram (P & ID) (mechanical flow diagram)

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

- All process equipment identified by an equipment number.
- All pipes identified by a line size, material code and line number.
- All valves with an identified size and number.
- Fittings.
- All pumps identified by a suitable code number.
- All control loops and instruments.

#### c) Utility Flowsheet (Process Engineering Utility Flow Diagram (PEUFD))

Used to summarize and detail the interrelationship of utilities such as air, water (various types), steam (various types), heat transfer mediums, process vents and purges, safety relief blow-down,

etc., to the basic process. The amount of detail is often too great to combine on other sheets, so separate sheets are prepared.

The PEUFD is a document containing information on:

- Main distribution or arrangement of each individual utility system, except electrical systems.

**PEUFD Function:**

A typical process uses utilities such as water, air and electric power. Water may be used either in the process, or for cooling and/or production of steam. Air may also be used in the process or for instrument applications. Electric power of course is typically used at various points in the process and throughout the site.

It is always useful to develop diagrams that show the flow and utilization of each utility. An example of a water balance/utility diagram is shown below.

The PEUFD shall state characteristics and consumption figures of the particular utility concerned, cooling water, fire water, drinking water, steam, plant air, instrument air, fuel oil/gas, inert gas and similar utilities.

**d) Process Safeguarding Flow Diagram (PSFD)**

The PSFD is a document highlighting information on:

- Types and levels of protection offered by the devices installed and their inter relation to demonstrate the plant's safety.

The P&ID contains all information required for a PSFD; however, the PSFD highlights protection in case of extreme conditions and measures to be taken to safeguard personnel and environment.

**Note:** In general these schemes will only be made for complex installations like offshore process platforms. For simple applications the information shown on the P&ID is usually sufficient to highlight safety devices and aspects.

- ✓ **What is meant by the following identifications? PFD, P&ID, PEUFD and PSFD**
- ✓ **State the information you can get from the following schemes: PFD, P&ID, PEUFD and PSFD?**

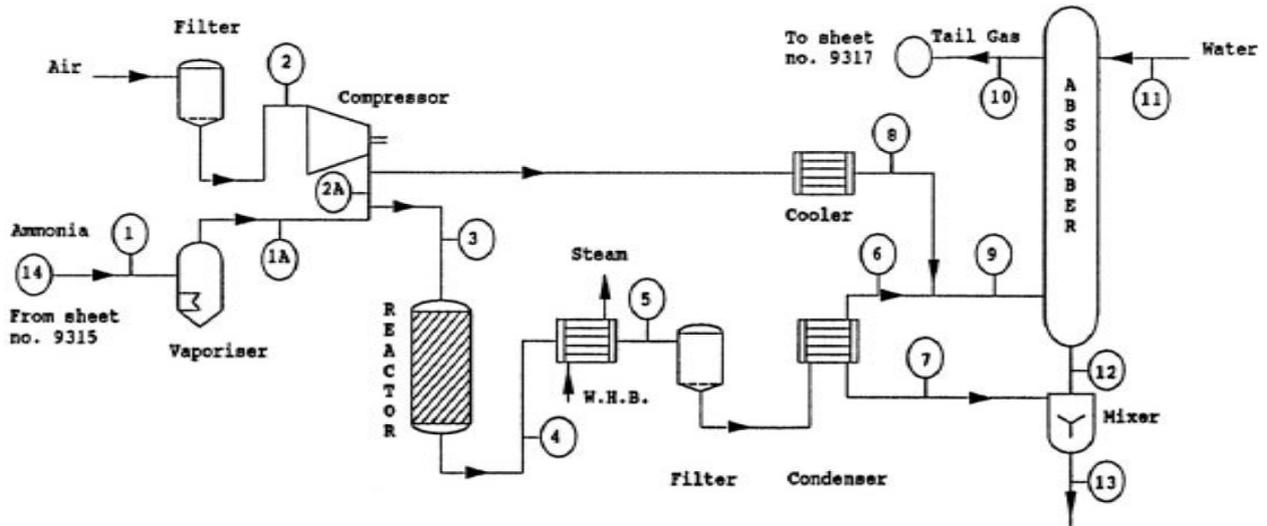


Fig.(4) PFD [Qualitative flow diagram for the manufacture of nitric acid by the ammonia-oxidation process].

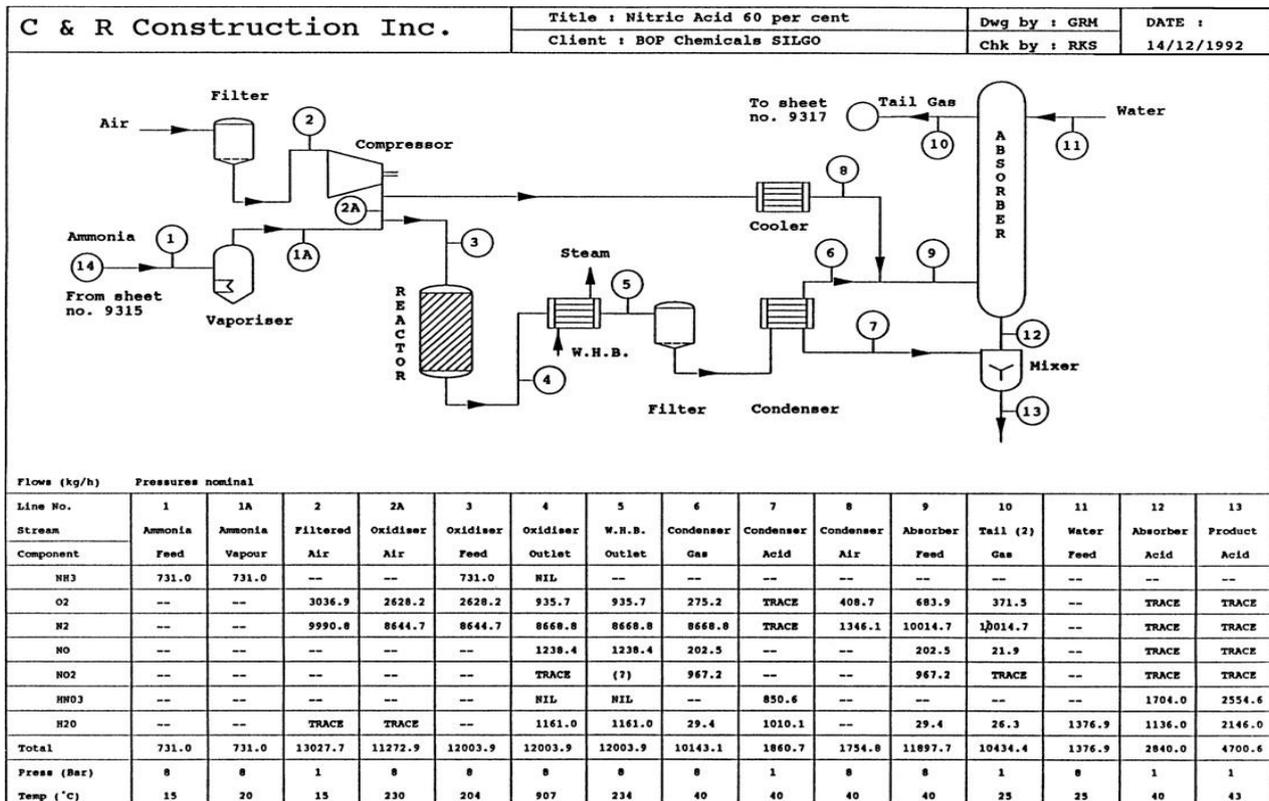


Fig.(5) PFD [Combined flow diagram for the manufacture of nitric acid by the ammonia-oxidation process].

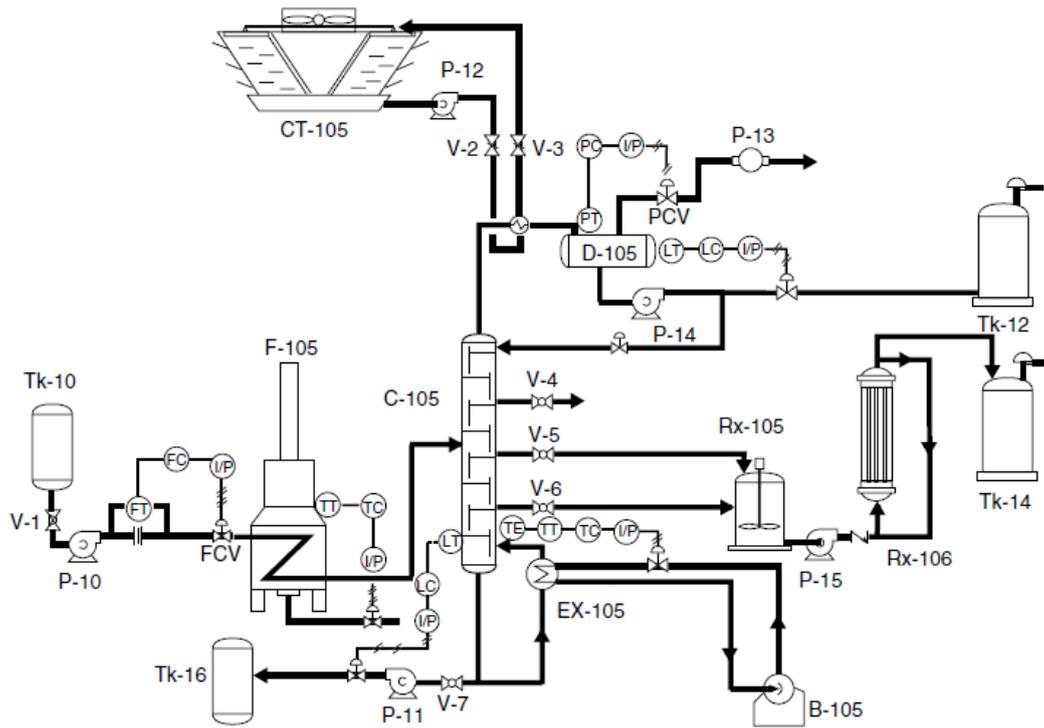


Fig.(6) Process and Instrument Diagram (P&ID)

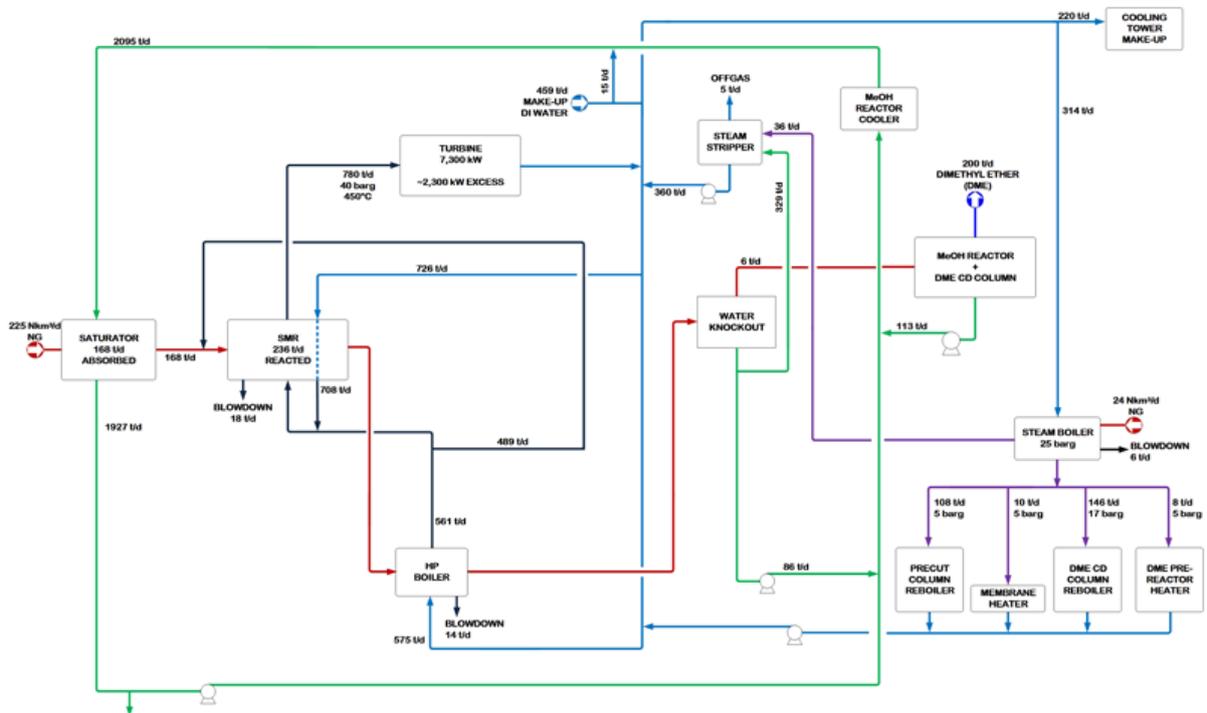


Fig.(7) Typical utility flow diagram.

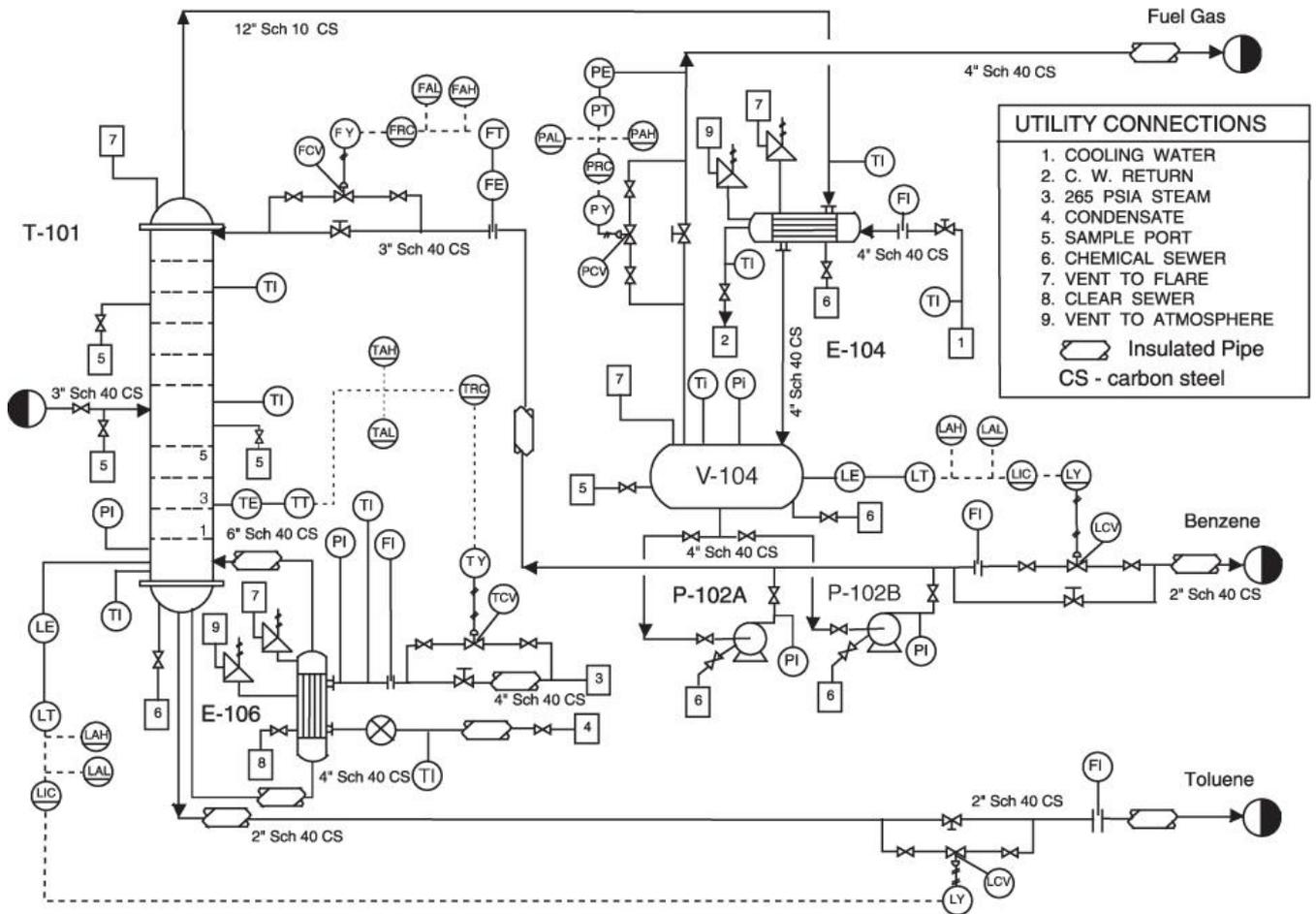


Fig.(8) Engineering P&ID flowsheet

## **Flowsheet Symbols**

To reduce detailed written descriptions on flowsheets, it is usual practice to develop or adopt a set of symbols and codes which suit the purpose. Many symbols are pictorial which is helpful in representing process as well as control and mechanical operations. See Fig.(9)

## **Line Symbols and Designation**

The two types of lines on a flowsheet are (1) those representing outlines and details of equipment, instruments, etc., and (2) those representing pipe carrying process or utility liquids, solids, or vapors and electrical or instrument connections. The latter must be distinguished among themselves as suggested by Figure (10).

The usual complete line designation contains the following: (1) line size (nominal); (2) material cod; (3) sequence number; and (4) materials of construction.

Examples: 2"-CL6-CS40

3"-CL6a-CS40

## **Equipment Designation**

Equipment code designations can be developed to suit the particular process, or as is customary a master coding can be established and followed for all projects. A suggested designation list (not all inclusive for all processes) for the usual process plant equipment is given in Table (1). The various items are usually numbered by type and in process flow order as set forth on the flowsheets. For example:

Item code	Designation
S-1	First separator in a process
S-2	Second separator in a process
C-1	First compressor in a process

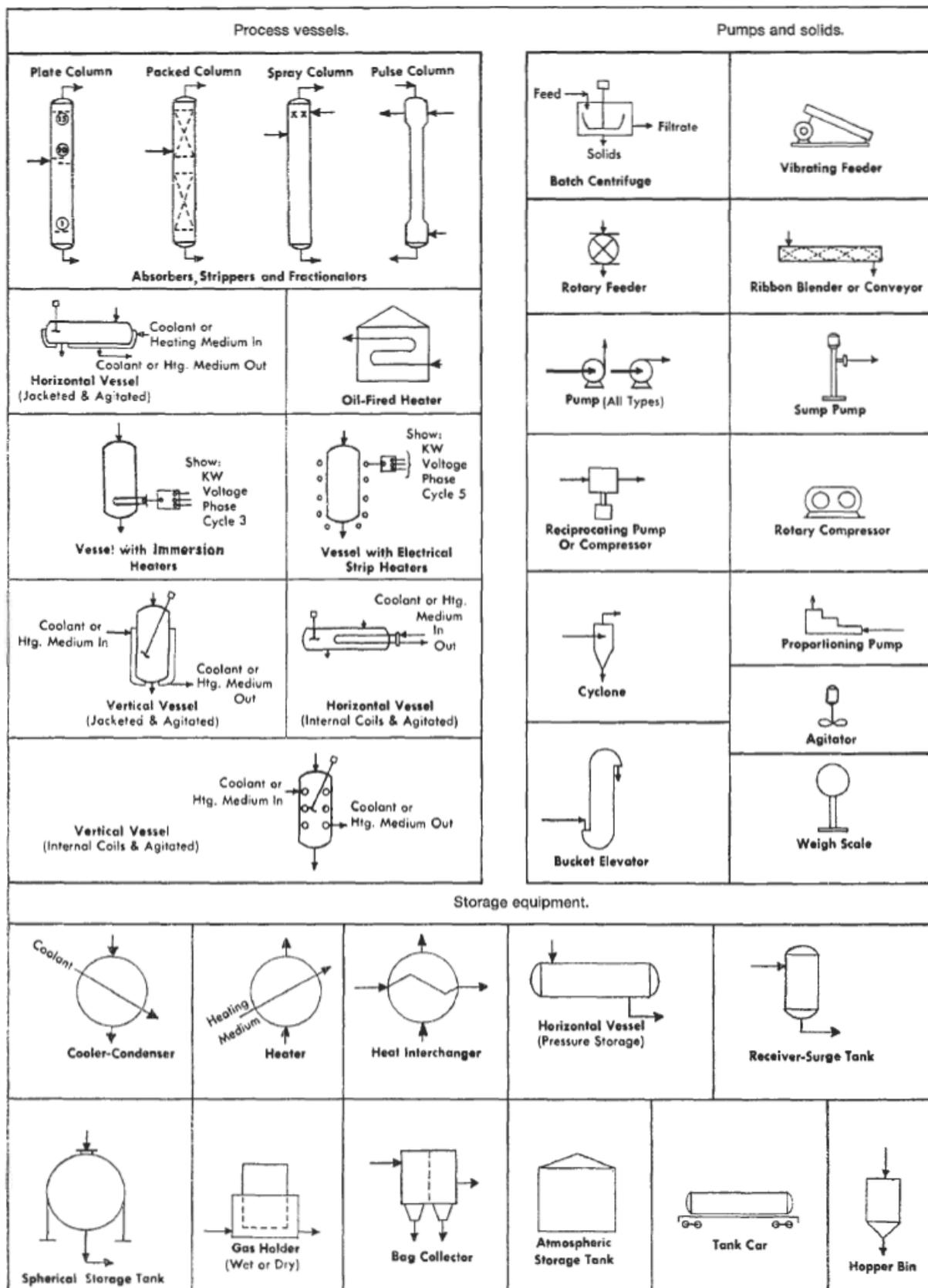


Fig. (9.a) Flowsheet symbols.

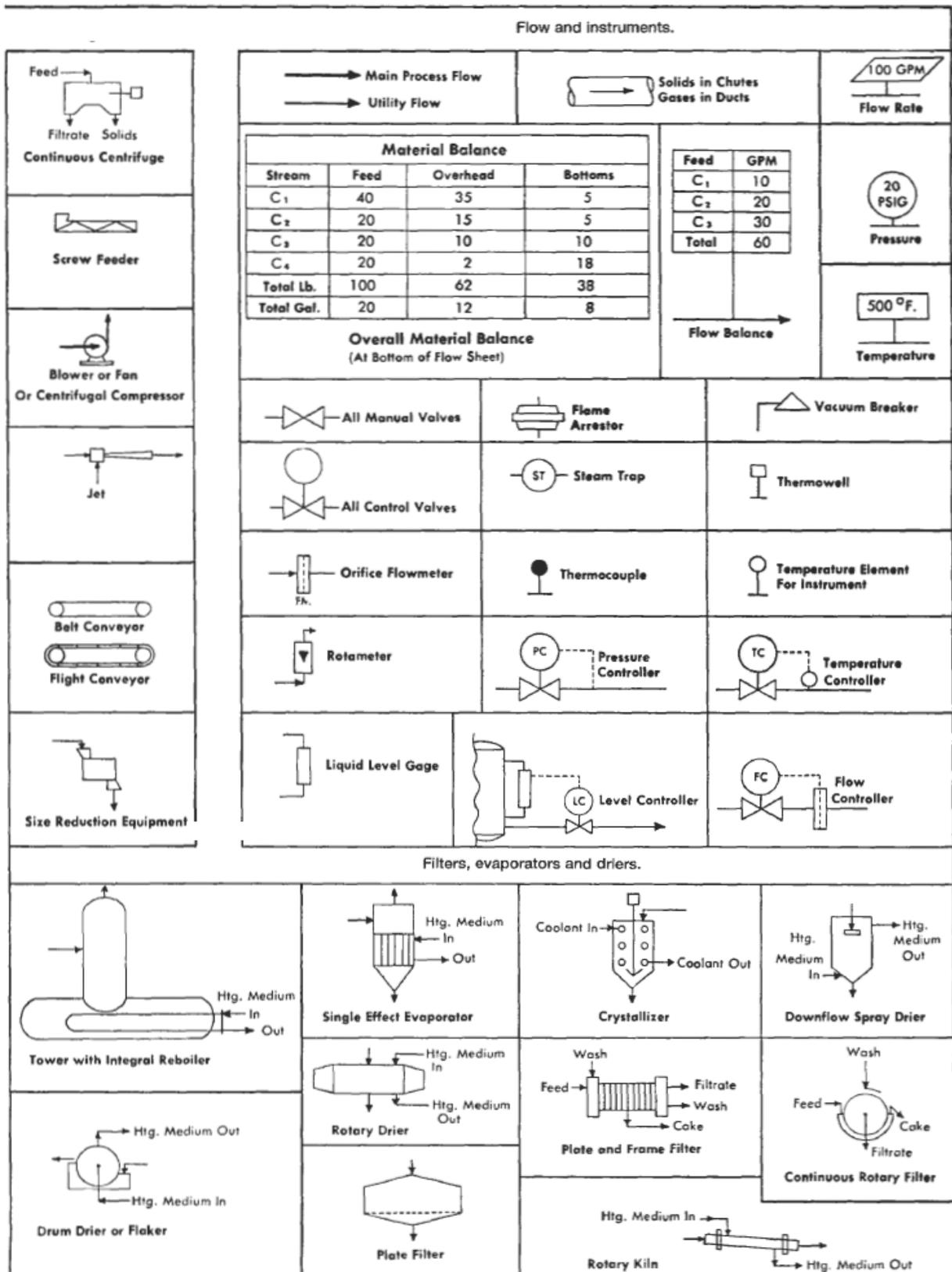


Fig. (9.b) Flowsheet symbols contd.

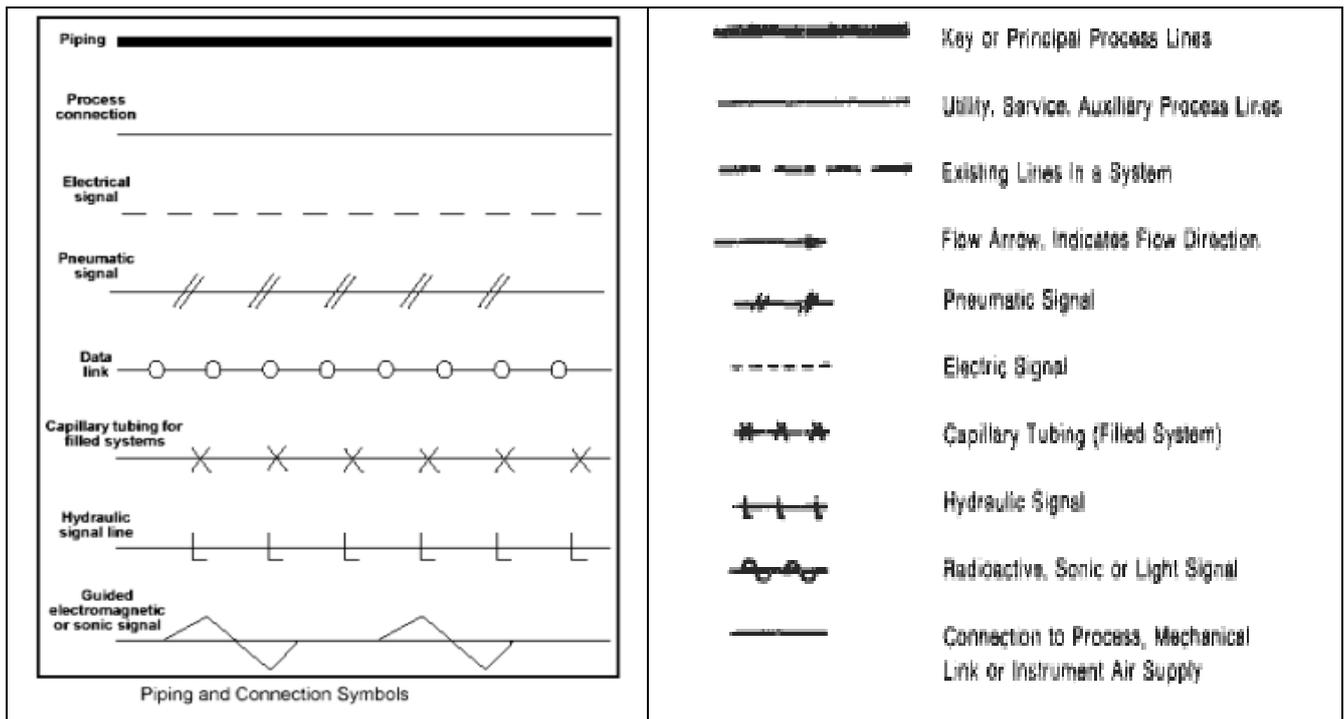


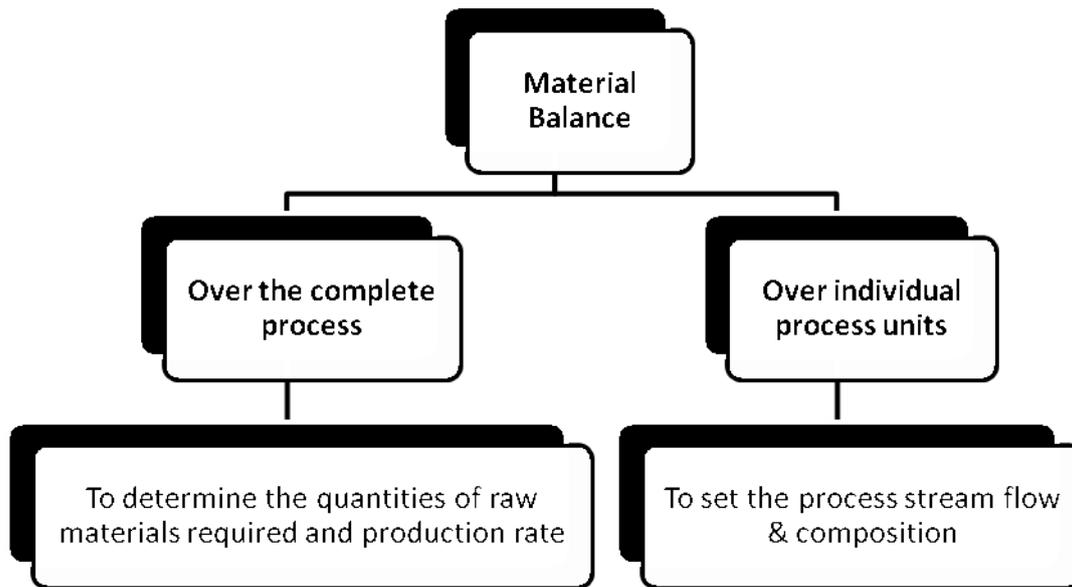
Fig.(10) piping and connection symbols from different sources(Line symbols)

Table 1

<b>A System of Equipment Designations</b>	
AD	→ Air Drier
AF	→ Air Filter
Ag	→ Agitator
B	→ Blower
BR	→ Barometric Refrigeration Unit
C	→ Compressor
CP	→ Car Puller
CT	→ Cooling Tower
CV	→ Conveyor
D	→ Drum or tank
DS	→ Desuperheater
E	→ Heat Exchanger, condenser, reboiler, etc.
Ej	→ Jet Ejector
Ex	→ Expansion Joint
F	→ Fan
FA	→ Flame Arrestor
Fi	→ Filter (line type, tank, centrifugal)
GT	→ Gas Turbine
MB	→ Motor for Blower
MC	→ Motor for Compressor
MF	→ Motor for Fan
MP	→ Motor for Pump
P	→ Pump
PH	→ Process Heater or Furnace
R	→ Reactor
S	→ Separator
St	→ Strainer
ST	→ Steam Turbine
Str	→ Steam trap
SV	→ Safety Valve
Tr	→ Trap
V	→ Valve
VRV	→ Vacuum Relief Valve

## Fundamentals of Material Balance

Material balances are the basis of process design



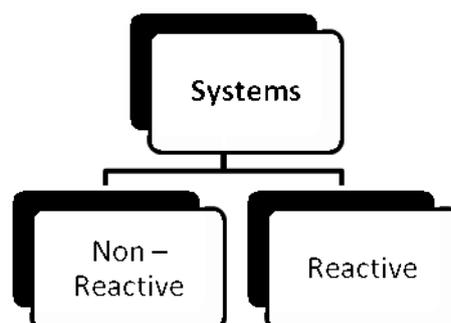
Material balance is also useful tool for the following:

1. The study of the plant operation & troubleshooting.
2. Check performance against design.
3. Check the instrument calibration.

## Conservation of Mass:

The general mass balance equation:

Input – output + generation – consumption = Accumulation



**For steady-state non reactive system:** Input = Output

[Number of equations = Number of components]

**For steady-state reactive system:** Input – output + generation – consumption = 0.0

## Some Important Parameters for Reactive System:

### ***Limiting Reactant:***

It's the reactant that would be completely consumed if the reaction proceeded to completion. As it disappears the reaction stops. It's also called the rate determining component since its concentration determines the reaction rate. All other reactants must either be fed in stoichiometric proportion to the limiting reactant (the feed rates are in the ratio of the stoichiometric coefficients) or in excess of the limiting reactant (in greater than stoichiometric proportion to it).

### ***Stoichiometry:***

It's used to balance chemical reaction equations. The stoichiometric equation for a chemical reaction states the number of molecules of the reactants and products that take part, from which the quantities can be calculated.

For simple reactions → can be done by inspection. بمجرد النظر

For complex reactions → take a base of 1 mole of one component and make an atomic balance on each element (It's better to choose one with many atoms as possible).

### ***Fractional Conversion:***

It's the ratio of amount reacted to amount fed. The fractional conversions of different reactants are generally different unless the reactants are fed in stoichiometric proportion.

$$\begin{aligned}\text{Fractional Conversion} &= \frac{\text{Amount reacted from the limiting reactant}}{\text{Amount fed from the limiting reactant}} \\ &= \frac{\text{Input of the limiting reactant} - \text{Output of the limiting reactant}}{\text{Input of the limiting reactant}}\end{aligned}$$

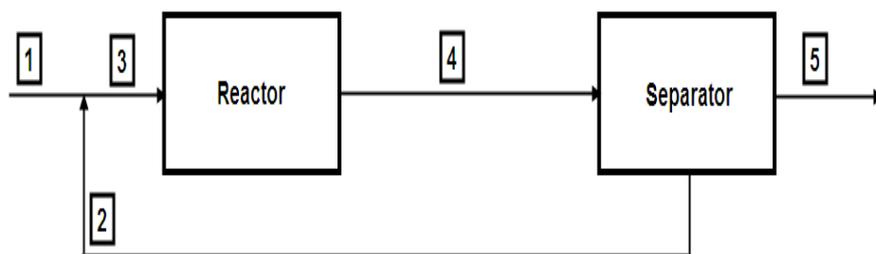
**Note:** In case of reaction process with recycling of unreacted reactants, there are 2 types of conversion:

#### **a- Single pass conversion**

$$= \frac{\text{Input to the reactor} - \text{Output from the reactor}}{\text{Input to the reactor}} = \frac{3-4}{3}$$

#### **b- Overall conversion**

$$= \frac{\text{Input to the process} - \text{Output from the process}}{\text{Input to the process}} = \frac{1-5}{1}$$



The overall conversion should be >>> single pass conversion [one of the recycle functions is to maximize conversion]

### **Selectivity:**

It's a measure of the efficiency of the reactor in converting reagent to the desired product. It is the fraction of the reacted material that was converted into the desired product. If no byproducts are formed, then the selectivity is 100%.

$$\text{Selectivity} = \frac{\text{moles of desired product actually formed}}{\text{moles of desired product theoretically could be formed}}$$

$$= \frac{\text{moles of desired product actually formed}}{\text{moles of reagent consumed} \times \text{stoichiometric factor}}$$

### **Yield:**

It's a measure of the performance of a reactor or a plant.

There are 2 types of yield:

#### **a- Reaction yield (Chemical yield)**

Reaction yield = Conversion  $\times$  Selectivity

$$= \frac{\text{Moles product produced}}{\text{Moles reagent converted}} \times \text{Stoichiometric factor}$$

Where moles reagent converted includes that consumed in both main & side reactions.

**Note:** Reaction yield = conversion when there is no side reactions take place.

#### **b- Plant Yield**

Plant yield is a measure of the overall performance of the plant and includes all chemical & physical losses (during separation process).

$$\text{Plant Yield} = \frac{\text{Moles product produced}}{\text{Moles reagent supplied to the process}} \times \text{Stoichiometric factor}$$

Stoichiometric factor = theoretical moles of reagent required per moles of product produced in the reaction balanced equation.

**Excess:**

A reagent may be supplied in excess to promote the desired reaction to:

- 1- Maximize the use of an expensive reagent.
- 2- Ensure complete reaction of a reagent, as in combustion.

$$\begin{aligned} \% \text{ Excess} &= \frac{\text{Actual Feed} - \text{Theoretical Feed}}{\text{Theoretical Feed}} \times 100 \\ &= \frac{\text{Actual Feed} - \text{Stoichiometric Feed}}{\text{Stoichiometric Feed}} \times 100 \end{aligned}$$

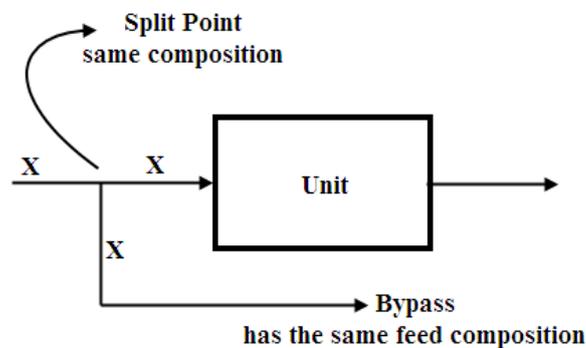
**Note:** Excess component actual feed = Theoretical feed  $\times$  (1 + fraction excess)

**Tie Component:**

If one component passes unchanged through a process unit (inert component), it can be used to tie the inlet & outlet compositions. Since its amount is the same in input & output so the total amount of input & output can be calculated if their compositions are known. Example: Nitrogen in combustion reactions.

**Bypass:**

A flow stream may be divided and some part diverted (bypassed) around some units. This procedure is often used to control stream composition or temperature.



**Recycle:**

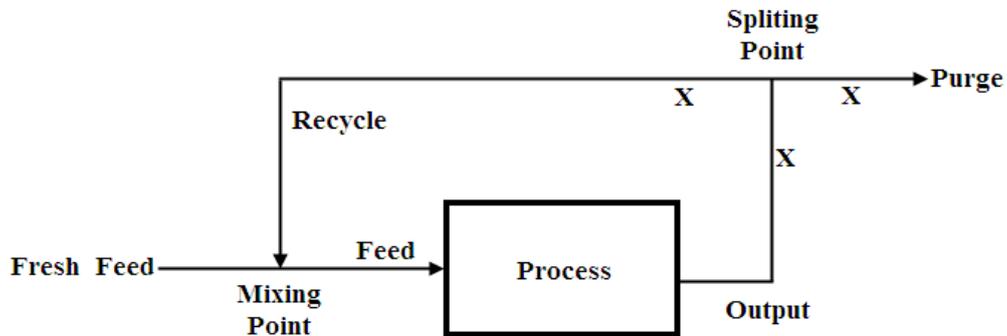
It's used to send unused raw materials emerging from a process unit back to the unit. Overall system balances are usually (but not always) convenient starting points for analyzing process with recycle.

**Purge:**

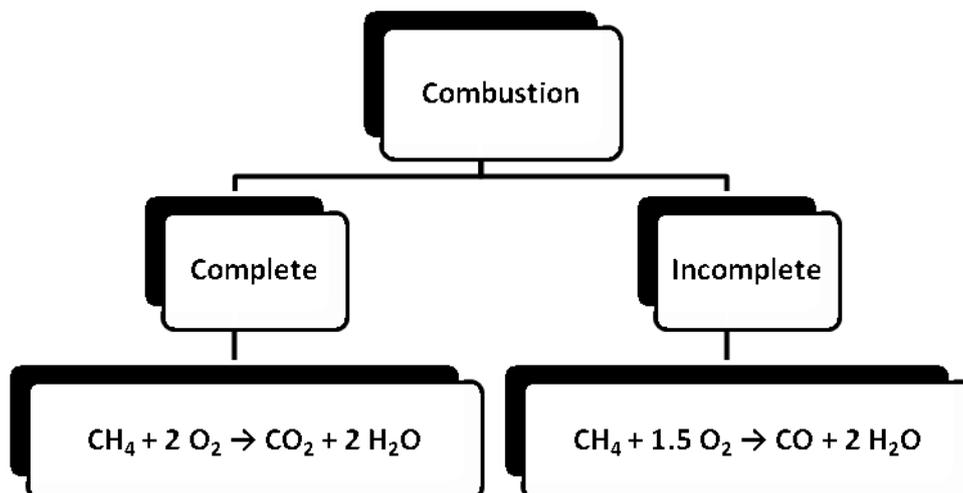
A stream that's withdrawn from a process when a species enters in the process feed and is completely recycled. If this species weren't removed in the purge, it would keep accumulating in the process system and eventually lead to shutdown.

**Purge stream used to:**

- Maintain the steady state conditions in the system
- Prevent the accumulation of inert or undesired materials [To rid the process of the undesired material]



**Combustion Reactions:**



Excess air is used in combustion to:

- 1- Ensure complete combustion.
- 2- Minimize Co & smoke formation.

**Note:** % excess air = % excess oxygen

Air = 79% N<sub>2</sub> + 21% O<sub>2</sub> by moles

Air = 77% N<sub>2</sub> + 23% O<sub>2</sub> by mass

The calculated amount of excess air doesn't depend on how much material is actually burned but what can be burned. Excess O<sub>2</sub> (air) is calculated from the complete combustion equation, i.e. based on conversion of all C → CO<sub>2</sub>, S → SO<sub>2</sub> & H → H<sub>2</sub>O.

**Important Note:** Using too much excess air means that more air will be heated (i.e. high energy losses). While if %excess air is less than 3, **flame impingement** to the furnace tubes.

### **Expressing the Composition of a Material Stream:**

- Mass fraction = component mass / total stream mass
- Mole fraction = component moles / total stream moles
- Volume fraction = component volume / total stream volume

**Note:** volume fraction = mole fraction for gases up to 25 bar.

For trace quantities:

ppm ► part per million

ppb ► part per billion

**Note:** UK billion =  $10^{12}$  while USA billion =  $10^9$

### **Choice of a basis for Calculations:**

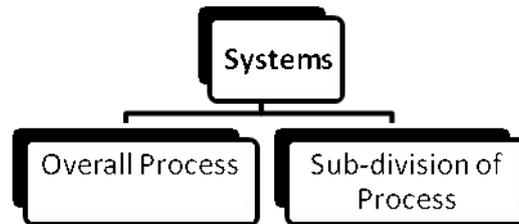
A basis of calculation for a process is an amount or flow rate of one of the process streams, preferably that stream with known composition. The basis may be a period of time for example, hours, or a given mass of material.

- 1- For continuous process (production or feed rate is given as kg/hr, ton/day, ..... etc) ► Basis is 1 hr or 1 operating day (unit of Time)
- 2- For batch process (production or feed rate is given as kg/batch, ton/batch, ... etc) ► Basis is 1 batch.
- 3- If the flow rates are not given ►
  - For composition given as mass fraction (in case of liquids or solids), basis is often 1 or 100  $lb_m$  or kg.
  - For composition given as mole fraction (in case of gases), basis is often 1 or 100  $lb_{mol}$  or kmol.

**Note:** It's important that your basis be indicated near the beginning of the problem.

### Choice of the system boundary

System boundary: The part of the process being considered.



The system chosen must have a degree of freedom = 0.0 in order to be able to be solved (i.e. No. of equations = No. of unknowns).

**Note:** For any reactive system, the following parameters should be defined:

- 1- Yield or conversion.
- 2- % Excess if present.
- 3- Concentrations (strength) of reactants & products streams.
- 4- Recovery.

$$\text{Recover} = \frac{\text{Amount of product finally out from the plant}}{\text{Amount of product produced from the reactor}}$$

### General Procedures for Material Balance Problems:

- 1- Draw a block diagram of the process.
- 2- List all the available data.
- 3- List all the information required from the balance.
- 4- Write out all the chemical reactions involved.
- 5- Decide the basis of your calculations.
- 6- Decide the system boundary.

## Fundamentals of Heat Balance

Heat balance is a special case of energy balance. Target of heat balance depends on the type of the system.

### 1. Reactive System:

Heat balance is important to:

- 1- Calculate the rate of heat addition or removal from the system to maintain the reaction mixture at the desired temperature to ensure certain degree of conversion.
- 2- Calculate the flow rates of heating or cooling utilities required.
- 3- Calculate the heat transfer area required for heating or cooling.

#### *a- Calculation of rate of heat removal / addition:*

##### **i- For continuous system:**

$$Q = \frac{n^{\circ}}{\gamma} \Delta \bar{H}$$

$n^{\circ}$ : Molar reaction rate of the limiting reactant (mol/ hr)

=  $n^{\circ}_{\text{feed}} \times \text{single pass conversion}$

$\Delta \bar{H}$ : Standard specific heat of reaction at 298K (kJ/mol)

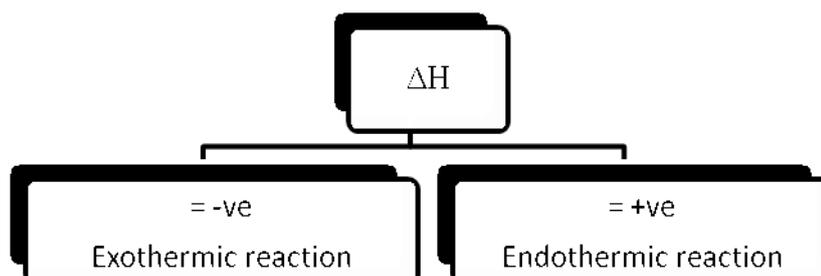
$$= \sum \gamma_i \Delta \bar{H}_f^{\circ} (\text{products}) - \sum \gamma_i \Delta \bar{H}_f^{\circ} (\text{reactants})$$

$\gamma_i$ : Component stoichiometric coefficient in the balanced equation.

$\Delta \bar{H}_f^{\circ}$ : Component standard heat of formation (kJ/mol).

**Note:** For a reaction  $A + 2 B \rightarrow C$   $\Delta \bar{H} = -240$  kJ/mol

That means that the heat released from the reaction of 1 mole of A with 2 mole of B to produce 1 mole of C = 240 kJ. If 100 mole of B reacts with 50 mole of A the heat produced =  $100/2 \times 240$  or =  $50/1 \times 240 = 1200$  kJ.



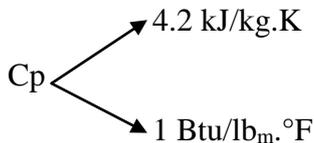
**For exothermic reaction:**

∴ Cooling is must

Calculate flow rate of cooling water:

$$Q = m^{\circ} C_p \Delta T$$

$m^{\circ}$ : cooling water mass flow rate



$\Delta T$ :  $T_{\text{out}} - T_{\text{in}} = 10 \rightarrow 15 \text{ °C}$  (if not given)

Cooling water pump capacity (pump volumetric flow rate) can be calculated:

Capacity of the cooling water pump =  $\frac{m^{\circ}}{\rho}$  ( $\text{m}^3/\text{hr}$ , Gpm or Lpm)

**For endothermic reaction:**

∴ Heating is must

Calculate the steam consumption:

- If steam enters as saturated vapor & exit as saturated liquid

$$Q = m_s^{\circ} \lambda_s$$

$m_s^{\circ}$ : Steam mass flow rate (kg/hr)

$\lambda_s$ : Latent heat of vaporization (kJ/kg) (from steam tables)

- If steam enters as super heated vapor & exit as sub-cooled liquid:

$$Q = m_s^{\circ} C_{p_v} \Delta T_1 + m_s^{\circ} \lambda_s + m_s^{\circ} C_{p_L} \Delta T_2$$

$\Delta T_1 = T_{\text{steam}} - T_{\text{boiling}}$        $C_{p_v}$ : Steam specific heat (kJ/kg.K)

$\Delta T_2 = T_{\text{boiling}} - T_{\text{out}}$        $C_{p_L}$ : water specific heat (kJ/kg.K)

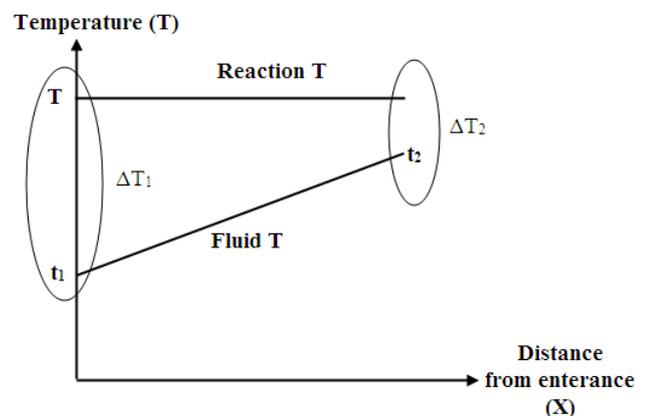
**Note:** you can calculate steam consumption regardless type of phase change:

$$Q = m_s [ \bar{H}_{\text{in}} - \bar{H}_{\text{out}} ]$$

- If heating is done by using oil: Mass flow rate of the heating oil:

$$Q = m^{\circ} C_p \Delta T_{\text{oil}}$$

$m^{\circ}$ : oil mass flow rate.



$$\Delta T_{oil} = T_{in} - T_{out}$$

ii- For batch system:

$$Q = \frac{n\Delta\bar{H}_r/\gamma}{\text{Reaction time}}$$

Reaction time = cycle time (if not given)

$\Delta\bar{H}_r$ : Specific heat of reaction

n: no of moles reacted from the limiting reactant per cycle

n = n fed per cycle  $\times$  conversion

$$n \text{ fed per cycle} = \frac{\text{No. of moles fed/day}}{\text{No. of cycles}}$$

$$\text{No. of cycles} = \frac{24 \text{ hr}}{\text{cycle time}}$$

**Important note:**

To calculate  $\Delta\bar{H}_r$  at any temperature:

$$\Delta\bar{H}_r = \Delta\bar{H}_r^\circ + R \int_{T_0}^T \frac{\Delta C_p}{R} dT$$

$$\int_{T_0}^T \frac{\Delta C_p}{R} dT = \Delta A T_0 [\tau - 1] + \frac{\Delta B}{2} T_0^2 [\tau^2 - 1] + \frac{\Delta C}{3} T_0^3 [\tau^3 - 1] + \frac{\Delta D}{T_0} \left[ \frac{\tau - 1}{\tau} \right]$$

$\tau = \frac{T}{T_0}$ , A, B, C & D are constants (from thermodynamic tables)

$\Delta A = \sum \gamma_i A_i (\text{products}) - \sum \gamma_i A_i (\text{reactants})$  & so for  $\Delta B$ ,  $\Delta C$  &  $\Delta D$

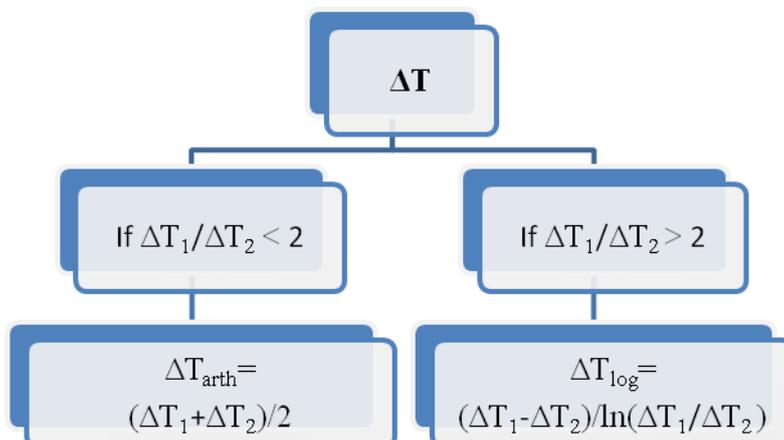
**b- Calculation of Heat Transfer Area Required for Heating or Cooling:**

$$Q = U A \Delta T_{lm}$$

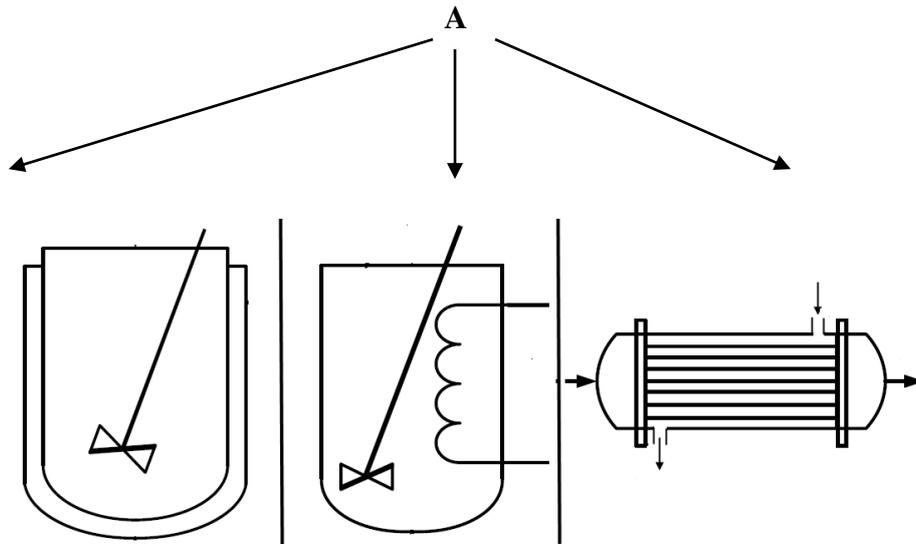
U: Overall heat transfer coefficient (W/m<sup>2</sup>.K).

A: Heat transfer area (m<sup>2</sup>).

$\Delta T_{lm}$ : Temperature difference (K or °C).



**Note:**  $\Delta T$  is calculated based on counter current flow unless otherwise is given.



**Reactor with jacket**

A= Jacket area

**Reactor with coil**

A= coil area

$$= \pi d_0 L$$

Coil length (L) can be calculated if  $d_0$  is known

**Heat Exchanger**

A= tubes area

$$= N \pi d_0 L$$

Number of tubes (N) can be calculated if the tube dims. are known (L &  $d_0$ )

**2. Non-Reactive System:**

This includes all the systems in which heat transfer takes place without chemical reaction.

Q in this case depends on the system.

**a. For Heaters/Coolers:**

$$Q = m^\circ C_p \Delta T \quad \text{Only sensible heat (no phase change)}$$

**b. For Vaporizers:**

General heat load equation:

$$Q = m^\circ C_{pL} (T_{\text{boiling}} - T_{\text{feed}}) + m^\circ \lambda + m^\circ C_{pV} (T_{\text{out}} - T_{\text{boiling}})$$

If the outlet is saturated vapor:

$$Q = m^\circ C_{pL} (T_{\text{boiling}} - T_{\text{feed}}) + m^\circ \lambda$$

If the feed is saturated liquid:

$$Q = m^\circ \lambda + m^\circ C_{pV} (T_{\text{out}} - T_{\text{boiling}})$$

If the feed is saturated liquid & the outlet is saturated vapor:

$$Q = m^{\circ} \lambda$$

*c. For Condenser:*

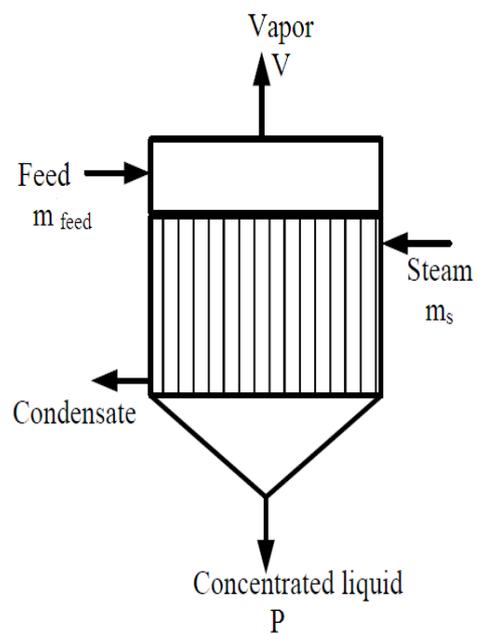
$$Q = m^{\circ} \lambda$$

*d. For Cooler-Condenser:*

$$Q = m^{\circ} C_p \Delta T + m^{\circ} \lambda$$

*e. For Evaporators:*

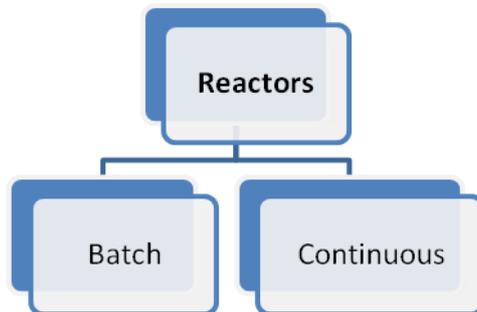
$$Q = m^{\circ}_{\text{feed}} C_{p \text{ feed}} (T_{\text{boiling}} - T_{\text{feed}}) + V \lambda$$



## Equipment Sizing

i.e. Calculation of reactors volume & dimensions.

Calculation of storage (holding) tank volumes.



### **1- Batch Reactor:**

$$= [V_{\text{reactants fed/cycle}} + V_{\text{catalyst (if exist)}}] * \frac{100}{\% \text{ Full}}$$

% Full = 75 if not given

$$V_{\text{reactants fed/cycle}} = \frac{V_{\text{reactants fed/cycle}}}{\text{No. of cycles}}$$

$$\text{No of cycles} = \frac{24}{t_{\text{cycle}} \text{ (hr)}}$$

After calculation of the reactor volume, its dimensions can be calculated:

$$V_{\text{reactor}} = \frac{\pi}{4} D^2 H$$

$H/D$  : Aspect ratio (If not given it can be assumed = 5)

### **2- Continuous Reactor:**

$$\text{Space time } \tau = \frac{V_{\text{reactor}}}{V^{\circ}_{\text{reactants}}}$$

$$\text{Space velocity } v = \tau^{-1} = \frac{V^{\circ}_{\text{reactants}}}{V_{\text{reactor}}}$$

For gas phase:

$$P V^{\circ}_{\text{reactants}} = n^{\circ} R T \text{ (assuming ideal gas behavior } P < 15 \text{ bar)}$$

$V^{\circ}_{\text{reactants}}$  : Feed reactants volumetric flow rate.

$n^{\circ}$  : Feed reactants molar flow rate.

For liquid phase:

$$V^{\circ}_{\text{reactants}} = \frac{m^{\circ}}{\rho}$$

**Note:** For mixed phase reaction the volume is calculated based on the gas volume.

### **General Procedure for Reactor Design:**

- 1) Collect together all the kinetic & thermodynamic data on the desired reaction and the possible side reactions (these data are obtained from literature or laboratory & pilot plant test).
- 2) Collect the physical properties required for the design 9 from literature, estimation by laboratory measurements).
- 3) Identify the controlling step [Kinetic or mass] and choose the suitable reactor type based on the experience with similar reactions or from the laboratory or pilot plant work.
- 4) Make initial selection of the reactor conditions to give the desired conversion & yield.
- 5) Size the reactor.
- 6) Select a suitable material of construction.
- 7) Make a preliminary mechanical design for the reactor.
- 8) Cost the proposed design (repeat steps 4→8 to optimize the design).

### **Storage (Holding) Tank:**

Storage tank is used for raw materials & final products.

Holding tanks is used in case of batch process.

$$V_{\text{Storage tank for a reactant}} = V_{\text{consumed per day}} \times \text{storage period in days}$$

$$V_{\text{Storage tank for a product}} = V_{\text{produced per day}} \times \text{storage period in days}$$

$$\text{Note: \% Tank full} = \frac{\text{Liquid height inside the tank}}{\text{Total tank height}}$$

## **Types of Designs**

The methods for carrying out a design project may be divided into the following classifications, depending on the accuracy and detail required:

### **1. Preliminary or quick-estimate designs**

Used as a basis for determining whether further work should be done on the proposed process. This type of design is based on approximate process methods, and rough cost estimates are prepared. Few details are included, and the time spent on calculations is kept at a minimum.

### **2. Detailed-estimate designs**

In this type of design, the cost and profit potential of an established process is determined by detailed analysis and calculations. However, exact specifications are not given for the equipment, and drafting-room work is minimized. The following factors should be established within narrow limits before a detailed-estimate design is developed:

- Manufacturing process
- Material and energy balances
- Temperature and pressure ranges
- Raw-material and product specifications
- Yields, reaction rates, and time cycles
- Materials of construction
- Utilities requirements
- Plant site

i.e the above factors should be determined after a preliminary design.

### **3. Firm process designs or detailed designs**

When the detailed-estimate design indicates that the proposed project should be a commercial success, the final step before developing construction plans for the plant is the preparation of a firm process design. In this type complete specifications are presented for all components of the plant, without any change in the process flowsheet and accurate costs based on quoted prices are obtained. The firm process design includes blueprints and sufficient information to permit immediate development of the final plans for constructing the plant.

## **Design Information** (literature survey)

General information and specific data required to the development of a design project can be obtained from many different sources such as:

### **A. Textbooks**

A large number of textbooks covering the various aspects of chemical engineering principles and design are available. In addition, many handbooks have been published giving physical properties and other basic data which are very useful to the design engineer. A primary source of information on all aspects of chemical engineering principles, design, costs, and applications is "The Chemical Engineers' Handbook" published by McGraw-Hill Book Company with R. H. Perry and D. W. Green as editors and Encyclopedia of Chemical Technology by Kirk Othmer.

### **B. Technical journals**

Regular features on design-related aspects of equipment, costs, materials of construction, and unit processes are published in Chemical Engineering. In addition to this publication, there are many other periodicals that publish articles of direct interest to the design engineer. The following periodicals are suggested as valuable sources of information for the chemical engineer who wishes to keep abreast of the latest developments in the field:

- American Institute of Chemical Engineers journal (AIChE)
- Chemical Engineering Progress
- Chemical and Engineering News
- Chemical Engineering Science
- Industrial and Engineering Chemistry Fundamentals
- Industrial and Engineering Chemistry Process Design and Development
- Journal of the American Chemical Society, Journal of Physical Chemistry
- Journal of the American Chemical Society
- Hydrocarbon Processing
- Oil and Gas Journal
- Engineering News-Record
- Canadian Journal of Chemical Engineering

### **C. Trade bulletins**

Trade bulletins are published regularly by most manufacturing concerns, and these bulletins give much information of direct interest to the chemical engineer preparing a design. Some of the trade-bulletin information is condensed in an excellent reference book on chemical engineering equipment, products, and manufacturers. This book is known as the "Chemical Engineering Catalog," and contains a large amount of valuable descriptive material. New information is constantly becoming available through publication in periodicals, books, trade bulletins, government reports, university bulletins, and many other sources. Many of the publications are devoted to shortcut methods for

estimating physical properties or making design calculations, while others present compilations of essential data in the form of nomographs or tables. The effective design engineer must make every attempt to keep an up-to-date knowledge of the advances in the field.

#### **D. Patents**

A patent is essentially a contract between an inventor and the public. In consideration of full disclosure of the invention to the public, the patentee is given exclusive rights to control the use and practice of the invention. A patent gives the holder the power to prevent others from using or practicing the invention for a period of 17 years from the date of granting. In contrast, trade-secrets and certain types of confidential disclosures can receive protection under common-law rights only as long as the secret information is not public knowledge. A new design should be examined to make certain no patent infringements are involved. If the investigation can uncover even one legally expired patent covering the details of the proposed process, the method can be used with no fear of patent difficulties.

#### **THE PRELIMINARY DESIGN**

In order to amplify the remarks made earlier concerning the design-project procedure, it is appropriate at this time to look more closely at a specific preliminary design. Only a brief presentation of the design will be attempted at this point. However, sufficient detail will be given to outline the important steps which are necessary to prepare such a preliminary design. The problem presented is a practical one of a type frequently encountered in the chemical industry; it involves both process design and economic considerations.

#### **Problem Statement**

A conservative petroleum company has recently been reorganized and the new management has decided that the company must diversify its operations into the petrochemical field if it wishes to remain competitive. The research division of the company has suggested that a very promising area in the petrochemical field would be in the development and manufacture of biodegradable synthetic detergents using some of the hydrocarbon intermediates presently available in the refinery. A survey by the market division has indicated that the company could hope to attain 2.5 percent of the detergent market if a plant with an annual production of 15 million pounds were to be built. To provide management with an investment comparison, the design group has been instructed to proceed first with a preliminary design and an updated cost estimate for a non biodegradable detergent producing facility similar to ones supplanted by recent biodegradable facilities.

## Literature Survey

A survey of the literature reveals that the majority of the non biodegradable detergents are alkyl benzene sulfonates (ABS). Theoretically, there are over 80,000 isomeric alkyl benzenes in the range of C<sub>10</sub> to C<sub>15</sub> for the alkyl side chain. Costs, however, generally favor the use of dodecene (propylene tetramer) as the starting material for ABS.

There are many different schemes in the manufacture of ABS. Most of the schemes are variations of the one shown in Fig. (11) for the production of sodium dodecylbenzene sulfonate. A brief description of the process is as follows:

This process involves:

- i. Reaction of dodecene with benzene in the presence of aluminum chloride catalyst (alkylation)
- ii. Fractionation of the resulting crude mixture to recover the desired boiling range of dodecylbenzene.
- iii. Sulfonation of the dodecylbenzene.
- iv. Neutralization of the sulfonic acid with caustic soda.
- v. Blending the resulting slurry with chemical “builders”; and drying.

## Process Description

Dodecene is charged into a reaction vessel containing benzene and aluminum chloride. The reaction mixture is agitated and cooled to maintain the reaction temperature of about **115°F** maximum. An excess of benzene is used to suppress the formation of by-products. Aluminum chloride requirement is **5 to 10 wt%** of dodecene. After removal of aluminum chloride sludge, the reaction mixture is fractionated to recover excess benzene (which is recycled to the reaction vessel), a light alkylaryl hydrocarbon, dodecylbenzene, and a heavy alkylaryl hydrocarbon.

Sulfonation of the dodecylbenzene may be carried out continuously or batch-wise under a variety of operating conditions using sulfuric acid (100 percent), oleum (usually 20 percent SO<sub>3</sub>), or anhydrous sulfur trioxide. The optimum sulfonation temperature is usually in the range of **100 to 140°F** depending on the strength of acid employed, mechanical design of the equipment, etc. Removal of the spent sulfuric acid from the sulfonic acid is facilitated by adding water to reduce the sulfuric acid strength to about **78 percent**. This dilution prior to neutralization results in a final neutralized slurry having approximately **85 percent** active agent based on the solids. The inert material in the final product is essentially Na<sub>2</sub>SO<sub>4</sub>.

The sulfonic acid is neutralized with **20 to 50 percent** caustic soda solution to a pH of 8 at a temperature of about **125°F**. Chemical “builders” such as trisodium phosphate, tetrasodium pyrophosphate, sodium silicate, sodium chloride, sodium sulfate, carboxymethyl cellulose, etc., are added to enhance the deterative, wetting, or other desired properties in the finished product. A flaked,

dried product is obtained by drum drying or a bead product is obtained by spray drying.

The basic reactions which occur in the process are the following.

Alkylation:



Sulfonation:



Neutralization:



A literature search indicates that yields of **85 to 95 percent** have been obtained in the alkylation step, while yields for the sulfonation process are substantially **100 percent**, and yields for the neutralization step are always **95 percent** or greater. All three steps are exothermic and require some form of jacketed cooling around the stirred reactor to maintain isothermal reaction temperatures.

Laboratory data for the sulfonation of dodecylbenzene, described in the literature, provide additional information useful for a rapid material balance.

This is summarized as follows:

1. Sulfonation is essentially complete if the ratio of 20 percent oleum to dodecylbenzene is maintained at 1.25.
2. Spent sulfuric acid removal is optimized with the addition of 0.244 lb of water to the settler for each 1.25 lb of 20 percent oleum added in the sulfonation step.
3. A 25 percent excess of 20 percent NaOH is suggested for the neutralization step.

Operating conditions for this process, as reported in the literature, vary somewhat depending upon the particular processing procedure chosen.

Required:

- 1- Making material and energy balances.
- 2- Equipment sizing and selection.

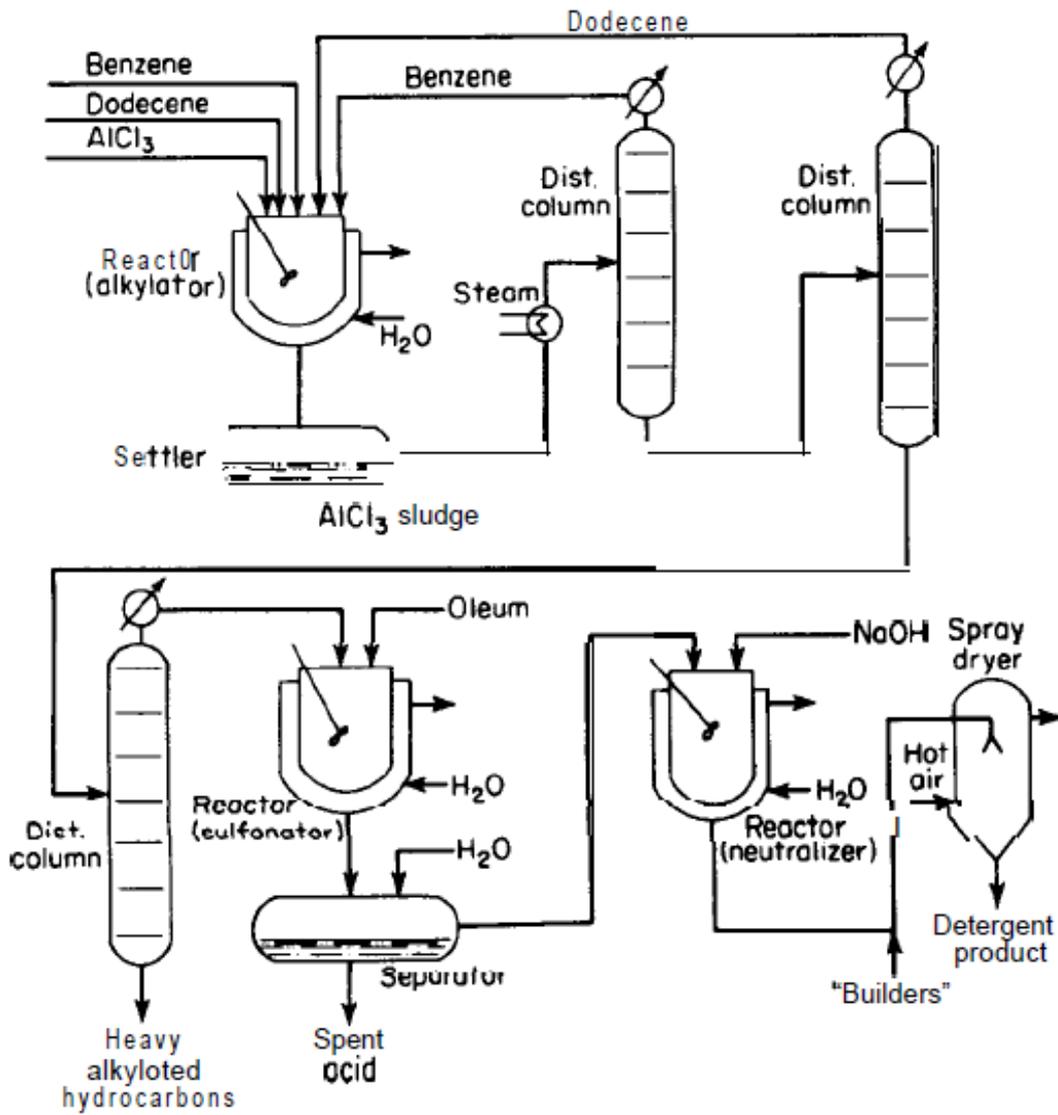


Fig. (11) Qualitative flow diagram for the manufacture of sodium dodecylbenzene sulfonate.

## **Economic Evaluation**

The design project can be economically evaluated through the following steps:

1. Estimation of the total capital investment.

Total capital investment = fixed capital + working capital

2. Estimation of the total annual product cost.

Total product cost = direct cost + indirect cost

3. Estimation of the expected annual profit.

Annual net profit = (total annual sales – total product cost) (1 – income taxes rate)

4. Using profitability estimation methods such as R.O.I to evaluate the attractiveness of the proposed project.

## **Safety Factors (Design Margins)**

**Definition:** These factors represent the amount of overdesign that would be used to account for the changes in the operating performance with time (fouling in H.X) and potential increases in capacity requirements.

Experienced designers include a degree of over-design known as a “design factor,” “design margin,” or “safety factor,” to ensure that the design that is built meets product specifications and operates safely.

Design factors are applied in process design to give some tolerance in the design. For example, the process stream average flows calculated from material balances are usually increased by a factor, typically 10%, to give some flexibility in process operation. This factor will set the maximum flows for equipment, instrumentation, and piping design. Where design factors are introduced to give some contingency in a process design, they should be agreed upon within the project organization and clearly stated in the project documents (drawings, calculation sheets, and manuals). If this is not done, there is a danger that each of the specialist design groups will add its own “factor of safety,” resulting in gross and unnecessary over-design. Companies often specify design factors in their design manuals. When selecting the design factor, a balance has to be made between the desire to make sure the design is adequate and the need to design to tight margins to remain competitive (economic consideration). Greater uncertainty in the design methods and data requires the use of bigger design factors.

In general design work, the magnitudes of safety factors are dictated by:

- Economic or market considerations,
- Accuracy of the design data and calculations,
- Potential changes in the operating performance,
- Background information available on the overall process.

Each safety factor must be chosen on basis of the existing conditions, and the chemical engineer should not hesitate to use a safety factor of zero if the situation warrants it. Some examples of recommended safety factors for equipment design are shown in Table (2).

**What is meant by design margins and what are the factors that control its values?**

### **Specification Sheets**

Standard specification sheets are normally used to transmit the information required for the detailed design, or purchase, of equipment items, such as heat exchangers, pumps, columns, pressure vessels, etc. As well as ensuring that the information is clearly and unambiguously presented, standard specification sheets serve as check lists to ensure that all the information required is included. A generalization for equipment design is that standard equipment should be selected whenever possible. If the equipment is standard, the manufacturer may have the desired size in stock. In any case, the manufacturer can usually quote a lower price and give better guarantees for standard equipment than for special equipment.

Before a manufacturer is contacted, the engineer should evaluate the design needs and prepare a preliminary specification sheet for the equipment. This preliminary specification sheet can be used by the engineer as a basis for the preparation of the final specifications, or it can be sent to a manufacturer with a request for suggestions and fabrication information. Preliminary specifications for equipment should show the following:

- 1) Identification
- 2) Function
- 3) Operation
- 4) Materials handled
- 5) Basic design data
- 6) Essential controls
- 7) Insulation requirements
- 8) Allowable tolerances
- 9) Special information and details pertinent to the particular equipment, such as materials of construction including gaskets, installation, necessary delivery date, supports, and special design details or comments. Figures 12 and 13 show typical types of specification sheets for equipment. These sheets apply for the normal type of equipment encountered by a chemical engineer in design work. The details of mechanical design, such as shell or head thicknesses, are not included, since they do not have a direct effect on the performance of the equipment. However, for certain types of equipment involving unusual or extreme operating conditions, the engineer may need to extend the specifications to include additional details of the mechanical design.

Table (2)

## Factors in equipment scale-up and design

Type of equipment	Is pilot plant usually necessary?	Major variables for operational design (other than flow rate)	Major variables characterizing size or capacity	Maximum scale-up ratio based on indicated characterizing variable	Approximate recommended safety or over-design factor, %
Agitated batch crystallizers	Yes	Solubility-temperature relationship	Flow rate Heat transfer area	>100:1	20
Batch reactors	Yes	Reaction rate Equilibrium state	Volume Residence time	>100:1	20
Centrifugal pumps	No	Discharge head	Flow rate Power input Impeller diameter	>100:1 >100:1 10:1	10
Continuous reactors	Yes	Reaction rate Equilibrium state	Flow rate Residence time	>100:1	20
Cooling towers	No	Air humidity Temperature decrease	Flow rate Volume	>100:1 10:1	15
Cyclones	No	Particle size	Flow rate Diameter of body	10:1 3:1	10
Evaporators	No	Latent heat of vaporization Temperatures	Flow rate Heat-transfer area	>100:1 >100:1	15
Hammer mills	Yes	Size reduction	Flow rate Power input	60:1 60:1	20
Mixers	No	Mechanism of operation System geometry	Flow rate Power input	>100:1 20:1	20
Nozzle-discharge centrifuges	Yes	Discharge method	Flow rate Power input	10:1 10:1	20 20

Factors in equipment scale-up and design (Continued)

Type of equipment	Is pilot plant usually necessary?	Major variables for operational design (other than flow rate)	Major variables characterizing size or capacity	Maximum scale-up ratio based on indicated characterizing variable	Approximate recommended safety or over-design factor, %
Packed columns	No	Equilibrium data Superficial vapor velocity	Flow rate Diameter Height to diameter ratio	>100:1 10:1	15
Plate columns	No	Equilibrium data Superficial vapor velocity	Flow rate Diameter	>100:1 10:1	15
Plate-and-frame filters	Yes	Cake resistance or permeability	Flow rate Filtration area	>100:1 >100:1	20
Reboilers	No	Temperatures Viscosities	Flow rate Heat-transfer area	>100:1 >100:1	15
Reciprocating compressors	No	Compression ratio	Flow rate Power input Piston displacement	>100:1 >100:1 >100:1	10
Rotary filters	Yes	Cake resistance or permeability	Flow rate Filtration area	>100:1 25:1	20
Screw conveyors	No	Bulk density	Flow rate Diameter Drive horsepower	90:1 8:1	20
Screw extruders	No	Shear rate	Flow rate Power input	100:1 100:1	20 10
Sedimentation centrifuges	No	Discharge method	Flow rate Power input	10:1 10:1	20 20
Settlers	No	Settling velocity	Volume Residence time	>100:1	15
Spray columns	No	Gas solubilities	Flow rate Power input	10:1	20

HEAT EXCHANGER	
<b>Identification:</b> Item Condenser Item No. <i>H-S</i> No. required <b>I</b>	Date <b>I-I-90</b> By <i>JRL</i>
Function: Condense overhead <i>vapors from methanol fractionation column</i>	
Operation: <i>Continuous</i>	
Type: <i>Horizontal</i> <i>Fixed tube sheet</i> <i>Expansion ring in shell</i> <b>Duty 3,400,000 Btu/h</b> <b>Outside area 470 sq ft</b>	
<b>Tube side:</b> Fluid handled <i>Cooling water</i> Flow rate <b>380 gpm</b> Pressure <b>20 psig</b> Temperature <b>15°C to 25°C</b> Head material <i>Carbon steel</i>	<b>Tubes:</b> <i>1 in. diam. 14 BWG</i> <b>1.25" Centers A Pattern</b> <b>225 Tubes each 8 ft long</b> <b>2 Passes</b> <b>Tube material</b> <i>Carbon steel</i>
<b>Shell side:</b> Fluid handled <i>Methanol vapor</i> Flow rate <b>7000 lb/h</b> Pressure <b>0 psig</b> Temperature <b>65°C to (constant temp.)</b>	<b>Shell:</b> <i>22 in. diam. I Passes</i> <b>(Transverse baffles Tube support Req'd)</b> <b>(Longitudinal baffles 0 Req'd)</b> <b>Shell material</b> <i>Carbon steel</i>
<b>Utilities:</b> <i>Untreated cooling water</i> <b>Controls:</b> <i>Cooling-water rare controlled by vapor temperature in vent line</i> <b>Insulation:</b> <i>2-in. rock cork or equivalent; weatherproofed</i> Tolerances: <i>Tubular Exchangers Manufacturers Association (TEMA) standards</i> Comments <b>and drawings:</b> <i>Location and sizes of inlets and outlets are shown on drawing</i>	

Fig. (12) Specification sheet for heat exchangers using U.S. customary units.

SIEVE-TRAY COLUMN				
Identification:	Item _____	Date _____		
	Item No. _____			
	No. required _____	BY _____		
Function:				
Operation:				
<b>Materials</b> handled:	<i>Feed</i>	<i>Overhead</i>	<b>Reflux</b>	<i>Bottoms</i>
Quantity	_____	_____	_____	_____
Composition	_____	_____	_____	_____
Temperature	_____	_____	_____	_____
Design data:	No. of trays _____		Reflux ratio _____	
	Pressure _____		Tray spacing _____	
	Functional height _____		Skirt height _____	
	Material of construction _____			
<i>Diameter:</i>	Liquid density _____ lb/ft <sup>3</sup> (- k g / m <sup>3</sup> )			
	Vapor density _____ lb/ft <sup>3</sup> (- k g / m <sup>3</sup> )			
	Maximum allowable vapor velocity (superficial) _____ ft/s (_____ m/s)			
	Maximum vapor flow rate _____ ft <sup>3</sup> /s (_____ m <sup>3</sup> /s)			
	Recommended inside diameter _____			
	Hole size and arrangement _____			
	Tray thickness _____			
Utilities:				
Controls:				
<b>Insulation:</b>				
<b>Tolerances:</b>				
Comments and <b>drawings:</b>				

Fig. (13) Specification sheet for sieve-tray distillation column.

