# PEM

PROCESS ENGINEERING MANUAL
VOLUME – V

# PROCESS EQUIPMENT DESIGN

# Preface

This Process Engineering Manual (Vol V) on ``**Process Equipment Design**" has been prepared with a view to enable TPPL Process Engineers to size and specify the more common types of equipment encountered in the design of process plants. The information provided by the process engineer (in the form of a *Process Datasheet*) is then used by the relevant detailed engineering group to do the detailed design and prepare the purchase specifications.

The present manual covers **Stationary Equipment** as well as **Rotary Equipment** in **Part I** and **Part II** respectively.

The document is a support document for Process Design and Engineering and is intended for internal use only.

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# PART- 1

# STATIONARY EQUIPMENT

# CHAPTER - 1

# COLUMN INTERNALS

## 1.1 Introduction

All separation process involving liquid-gas contacting e.g. distillation, flashing, stripping, absorption etc. require column and internals for mass transfer operation. The various column internals include trays, grid and random packing, demisters, weirs, downcorners and spray nozzles etc. To enable the mass transfer specialist to make detailed specifications for the column internals to be issued to the fabricator a process engineer is required to furnish the necessary information to in the form of a *Tray Date Sheet.* This tray data sheet should be accompanied by the *Column Data Sheet* showing the relevant nozzles, draw-offs etc. Following guidelines should help the process engineer to understand the important features of mass transfer equipment and prepare the process data sheet for trays.

## 1.2 Liquid-Gas Contacting Systems

Liquid-gas contacting system are utilised for the separation of materials based on phase-equilibrium relationship, with the rate of transfer controlled by molecular and eddy-diffusion mechanisms. Such system may include:

- **Distillation**

It is a separation of the constituents of a liquid mixture via partial vaporisation of the mixture and separate recovery of vapour and residue.

- **Flashing**

It is a distillation process wherein the total vapour removed approaches phase equilibrium with the residue liquid.

- **Stripping**

It is the transfer of gas dissolved in the liquid into a gas stream. The term is also applied to the section of a fractionating tower below the feed plate.

- **Absorption**

It is the transfer of a soluble component in a gas-phase mixture into a liquid absorbent whose volatility is low under the desired process conditions.

## 1.3 Column Internals

From the viewpoint of a process engineer various internals in a column may be divided into two groups:

- Liquid-vapour contacting devices e.g. Trays or plates, random packings, grid packing etc.
- Mist and entrainment eliminators, spray nozzles, draw-off pans etc.

## 1.4 Packed Columns

In a packed column, random packings of ceramic, steel or polypropylene may be employed. The latter is increasingly preferred because of its low weight, easy handling and resistance to corrosion. Packing grids of a variety of proprietary designs are being increasingly used these days. Some typical proprietary designs are Glitsch grids, Sulzer packing etc. The mode of flow in packed columns is counter-current or co-current, the continuous phase being generally gas and used widely in absorption stripping and distillation operations.

## 1.5 Plate Columns

The mode of flow here is generally cross flow or counter-current, and used in distillation, absorption, stripping, gas cleaning etc. Plate or trays may be the following popular types:

- Bubble-cap Trays
- Sieve or perforated Trays
- Valve Trays
- Other types like channel. Discs and dough-nut etc.

The comparison between different types of trays is described in Section 1.7.

## 1.6 Packed Columns Vs Plate Columns

Packed or plate towers may be used for many gas-liquid contacting operations. A process engineer can use the following guidelines in making a preliminary selection on whether to use a packed column or a plate column for a particular application:

- If the operation involves liquid that contains dispersed solids, use of plate tower is preferred because the plates are more accessible and convenient for cleaning.
- Plate towers can be designed to handle wide range of liquid rates without flooding.
- Plate towers are preferred if interstage cooling is required to remove heats of reaction or solution. This is because cooling coils can be installed on the plates or the liquid delivery line from plate to plate can be passed through an external cooler.
- Plate columns are preferred if side streams are to be produced like refinery crude distillation column.
- The total weight of a dry plate tower is usually less than that of a packed tower designed for the same duty. However, if liquid hold up during operation is taken into account, both types of towers have about the same weight.
- When large temperature changes are involved, as in distillation operations, plate towers are often preferred because thermal expansion or contraction of the equipment components may crush the packing.
- Design information for plate towers is generally more readily available and more that for packed columns.
- Random-packed towers are seldom designed with diameters larger than 4 feet and diameters of commercial plate towers are seldom less than 2 feet. For columns less than 2 feet in diameters, packing are usually cheaper than plates, unless alloy metals are used. Larger diameter which of the two internals would be more economical . Larger diameter packed towers with packing grids are being increasingly used these days.
- Acids and may other corrosive materials can be easily handled in packed columns because construction can be of ceramic, carbon, polymers or other corrosion resistant materials.
- Packing often exhibit more desirable efficiency-pressure drop characteristics. Pressure drop through packed tower is usually less than the pressure drop through a plate tower for the same duty. This advantage, plus the fact that the packings serve to lesson the possibilities of tower –wall collapse, makes packed tower particularly desirable for vacuum operation.
- Packed towers are usually preferred if the liquids have a large tendency to foam.
- The amount of liquid hold up is considerably less in packed towers, an advantage when the liquid is thermally sensitive.

## 1.7 Comparison Of Tray Types

Important features of different types of trays are described below. This should help the process engineer to select the type of trays for the services under question.

### 1.7.1 Bubble Cap

| | |
|---|---|
| Capacity | : Moderately high, maintains efficiency. |
| Efficiency | : As high as other tray designs. |
| Entertainment | : About three times that of perforated type plate or sieve tray. |
| Flexibility | : Most flexible of tray designs for high and low vapour and liquid rates. Allows positive drain of liquid from tray. Liquid heads maintained by weirs. |
| Application | : All services except extremely cooking, polymer formation or other high fouling conditions. Used for extremely low flow conditions where tray must remain wet and maintain a vapour seal. |
| Tray spacing | : 18" average, 24 to 36" for vacuum conditions. |

### 1.7.2 Sieve Tray or Perforated Tray With Downcomers

Vapour rises through small holes (1/8"to 1") in tray floor and bubbles through liquid on tray below.

| | |
|---|---|
| Capacity | : As high or higher than bubble cap at design or down to 60% of design rates with good efficiency. At lower throughputs, performance drops as efficiency falls of rapidly. |
| Efficiency | : As high as bubble caps in region of design, but falls to unacceptable values when capacity reduced below 60% (approx.). |
| Entertainment | : Only about $\frac{1}{3}$rd that of bubble cap trays. |
| Flexibility | : Not generally suitable for columns operating under variable load, falling below 60% of design. Tray weeps liquid at low vapour rates. |
| Application | : System where high capacity near design to be maintained in continuos service. Because of small holes required to minimise weeping, these trays are not very suitable when solids and corrosive materials are to be handled which eventually lead to plugging of these holes. |
| Tray spacing | : Can be lesser than bubble cap due to improved entertainment. 15" is average, 9", 10" and 12" are acceptable with 20" to 30" for vacuum. |

### 1.7.3 Perforated Plate Without Downcomers

These are of perforated, turbogrid or kettle type trays. The vapour-liquid traffic is through the same opening and does not require a separate downcomer for liquid. Vapour rises through holes (3/16" to 1") in tray floor and bubbles through liquid. At the same time, liquid head forces the liquid counter-currently through these holes and onto tray below.

| | |
|---|---|
| Capacity | : Quite similar to sieve tray, as high or higher than bubble cap tray from 50% upto 100% design rate (varies with system and design criteria). Performance falls off at lower rates. |
| Efficiency | : Usually not a high as bubble caps in region of design but falls to unacceptable values below 60% design rate. |
| Entertainment | : Only about $\frac{1}{3}$rd that of bubble cap tray. |
| Application | : System where high capacity near design rates to be maintained in continuous service. Good in Vacuum or low pressure drop design. |
| Tray spacing | **:** Can be closer than bubble cap due to improved entertainment, 12" is average, 9" to 18" acceptable, 18" to 30" for vacuum. |

### 1.7.4 Valve Trays

Valve trays are similar to sieve trays with one important exception, the perforations are covered with liftable lids or "valves" which rise and fall with variations in vapour flow. The lids thus act as check valves to limit liquid weeping or dumping at low vapour rates. The main advantage of valve trays is that high efficiency can be maintained over a wide range of operating throughputs. Design turndown ratio can be as high as 10%. Valve trays provide more operating flexibility than sieve trays because of their variable orifice characteristics. Since they are more complex mechanically, their fabrication is somewhat more expensive than sieve trays, but much cheaper than bubble cap trays. Other features, such as entertainment, flooding, resistance to fouling etc. are not likely to be greatly different between valve trays and sieve trays. These trays match bubble cap trays in turndown requirement and have capacity similar to sieve trays and sometimes better. Tray spacing is around 18" –24" on an average.

### 1.7.5 Other Trays

Baffle trays of the type of discs, doughnuts and chimney trays etc. are also used in some services. Discs are solid horizontal plates, axially mounted and with diameter less than that of the column. The doughnuts are horizontal annular things attached to the walls of the column. These types are used in dirty services e.g. quench towers, absorbers, bottom section of fractionates in cooking units, FCC and visbreaker fractionator bottom and highly polymerising services. Chimney trays are used in distillation towers at locations where the total liquid is withdrawn.

## 1.8 Tray Performance

Tray performance means, the study of capacity and efficiency of the tray at various liquid and vapour rates. A tray gets limited in capacity either because of flooding in down comers or as a result of massive entertainment of liquid with the vapour to the tray above. The following description may be of help in understanding the various factors affecting tray performance.

### 1.8.1 Flooding

The downcomer then contains a mixture of liquid and foam, its capacity of handling liquid gets limited and the level rises in the downcomer. The level finally extends on to the tray above and eventually may propagate to the point of filling the tower. Flooding is generally accompanied with high liquid loads. The phenomena is more severe if the liquid is of foaming type. This results in loss of tray efficiency and excessively high tray pressure drops.

### 1.8.2 Blowing and Entertainment

The capacity of a tray may also be limited due to too high a vapour rate. This phenomenon is called entertainment. At very high vapour rates and relatively low liquid /gas ratio, the efficiency of the tray may drop markedly because of blowing, wherein the tray is blown clear of liquid in the immediate vicinity of the vapour distributors.

### 1.8.3 Weeping

When liquid to gas ration is high, the quantity of liquid flow across the plate may require a very high liquid gradient in order to drive the flow. This gives tendencies towards flooding

or too high a pressure drop. Another result of high liquid gradient can be phase maldistribution, wherein the vapour flows preferentially through the perforations in the plate near the liquid flows preferentially through the perforations in the plate near the liquid outlet and the liquid flows in part downward through the perforations near the liquid inlet where the liquid depth is greatest. This flow of liquid downward through the perforations rather than through the downcomer is known as weeping. The problem of a high liquid gradient is particularly severe for plate columns of large diameters. One way to prevent a large liquid gradient is to use split flow trays. Two, four or more pass trays are generally used for large diameter columns. Refer Figure 1.1 for sketches of trays with split flows.

### 1.8.4 Tray Flexibility

A tray is flexible when it operates with acceptable efficiency under condition which deviate significantly from those established for design. The usual changes affecting flexibility are vapour and liquid loadings. A tray may operate down to 50% and up to 120% of vapour load, and down to 15% and down to 130% of liquid load and still be efficient. Beyond these points the efficiency usually falls.

### 1.8.5 Tray stability

A tray is stable when it can operate with acceptable efficiencies under conditions which fluctuate on surge, developing unstable conditions. This type of operation is difficult to anticipate in design and most trays will not operate long without sharing loss of efficiency.

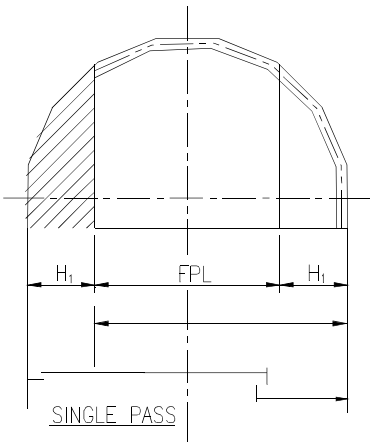
## 1.9 Tray Data Sheet

Process Data Sheet for tray should be filled in by the process engineer to give all the pertinent information asked for in the data sheet. Liquid and vapour rates should be specified for critical trays. In general, for pressure services, step load will occur on the top tray if cold reflux is being fed. Similarly, pump around sections will give rise to step changes in loadings on such trays. In vacuum distillation because of steep variation in density from bottom most to top trays, it is essential to give loadings on a number of trays. In case of any doubt, complete vapour and liquid profile along the tower should be supplied.

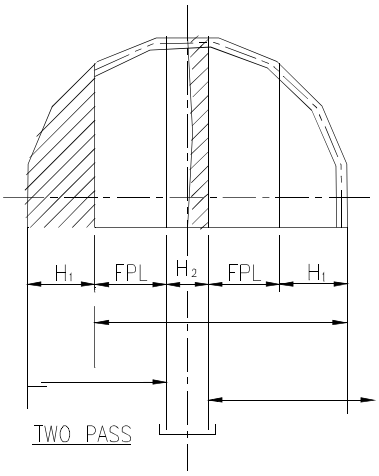
System frothing characteristics and properties like viscosity and surface tension should always be indicated. The design of down comers is greatly effected by the foam forming properties of the system. Refer Table 1.1 for system factors. Specify the allowable pressure drop judiciously. The pressure drop should not be unnecessarily conservative. Refer Section 1.12 for pressure drops in columns.

The column/vessel sketch showing relevant nozzle locations, orientation and location of manholes, draw – off sumps etc. should always be supplied along with the *Tray data sheet.* This sketch should be on the *Column data sheet.*

FIGURE 1.1: SKETCHES FOR TRAY TYPE.




SINGLE PASS

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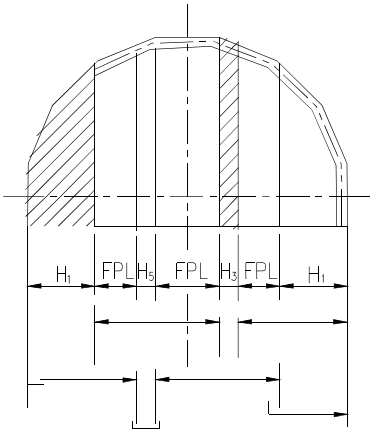
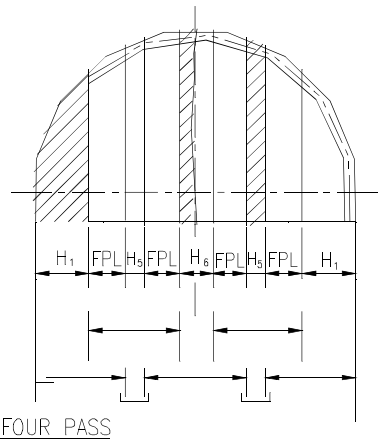
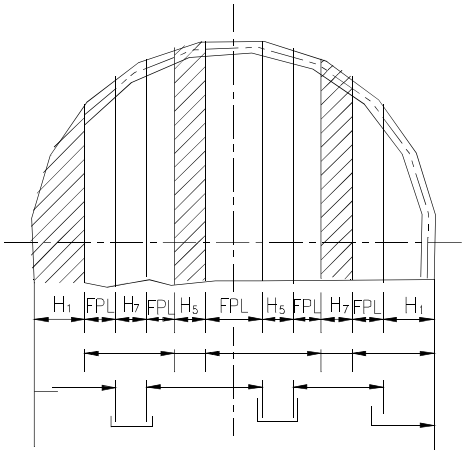



THREE PASS




FOUR PASS




FIVE PASS

## 1.10 Preliminary Sizing Of Trays / Packings

Once the column internal traffic data is generated, the next step is to carry out the sizing of the column. While the detailed sizing and design of the internals are carried out by the mass transfer specialist, it is necessary for the process engineer to do a reasonably accurate calculation to tower diameter and height using the methods outlined in the following pages. This is required for initiating the initial design without waiting for the accurate design to be done by the mass transfer specialist.

Table1.1: Suggested System for Factors

| System | 'G' or $C_F$ |
|---|---|
| Non-Foaming<br>Most clean distillation applications | 1.0 |
| Mild or Slight Foaming<br>Gasoline Stabilizers<br>Crude Distillation<br>Vacuum Crude Distillation | 0.9<br>0.8-1.0<br>0.85 |
| Moderate Foaming<br>Debutanizer, Depropanizer<br>Amine Absorbers<br>Glycol Regenerators | 0.85<br>0.8<br>0.85<br>0.8 |
| High Foaming<br>Deethanizer, Deethanizer<br>Amine Absorbers<br>Glycol Absorbers | 0.7-0.75<br>0.65-0.7<br>0.7 |
| Severe Foaming<br>M.E.K. Units | 0.6 |
| Foam Stable<br>Caustic Regenerators | 0.3 |

Table 1.2: $K_1$ and $K_2$ Values for Different Tray Spacings

| Tray Spacing | $K_1$ | $K_2$ |
|---|---|---|
| mm | | |
| | | |
| 300 | 180 | 130 |

| | | |
|---|---|---|
| 450 | 240 | 155 |
| 600 | 300 | 170 |
| 900 | 360 | 190 |

### 1.10.1 Valve Tray Sizing

(a) Preliminary tray diameter can be calculated by the following equation or by using figure 1.2. While using the equation, proceed as below :

$$A_T, \text{Tower area } m^2 = \frac{1}{C_F}\left(\frac{V}{K_1} + \frac{Q_L}{K_2}\right)$$

Where,

$V$ = Vapour flow $(m^3/hr) * \left(\frac{D_V}{D_L - D_V}\right)^{0.5}$

$Q_L$ = Liquid flow $(m^3/hr)$

$D_V$ = Vapour density $(kg/m^3)$

$D_L$ = Liquid density $(kg/m^3)$

$A_F$, Active area $m^2$ = $\frac{V}{K_1} * \frac{1}{C_F}$

Downcomer area = $\frac{Q_L}{K_2} * \frac{0.5}{C_F}$

Reference is made to Tables 1.1 and 1.2 to find out the values of the various constants listed above. Tray spacing can be based on the guidelines given in Table 1.3.

(b) For using Figure 1.2, proceed as below :

- Use Figure 1.3 to calculate the flood capacity factor CAF.
- Calculate $V_{load}$ as follows:

$$V_{load} = v\left(\frac{D_V}{D_L - D_V}\right)^{0.5}$$

Where,

$V$ = Vapour rate in $m^3/hr$

$D_v$ = Vaopur density in $kg/m^3$

$D_l$ = Liquid density in $kg/m^3$

Determine $Q_L$ in $m^3/hr$ where $Q_L$ = Liquid flow rate from the tray.

Use Figure 1.2 to read approximate tower diameter, *Dr*.

Table 1.3: Guidelines for Selection of Tray Spacing

| Service | Tray Spacing (mm) |
|---|---|
| Atmospheric Distillation tower, Non-foaming applications, clean Hydrocarbon system | 450-600 |
| Pump-around section in Atmospheric or vacuum columns, Foaming services | 600-900 |
| Pressure towers (> 5 kg/cm$^2$ abs) In non-foaming service | 300-450 |
| Tray exceeding 2500 mm diameter should preferably be on 600mm spacing. | |

(c) Confirm the approximate area as below:

- Flow path length $FPL = \dfrac{0.75 * D_L}{N_P}$ where $N_P$ is the number of passes assumed
- Calculate jet flood percentage as follows:

$$Jet\ flood\ \% = \left(V_{load}/102 + Q_L * FPL/74000\right) * \frac{100}{A_F * CAF * 10.5}$$

Where,

$A_F$ = Active area of tray deck
= $A_T$ Downcorner area
$A_T$ = tower area

In case downcorner area is not available, assume $A_F = 0.85\ A_T$.

- If G is the suggested system factor (Refer Table 1.1), then

  Corrected jet flood = $\frac{\textit{Jet Flood as calculated above}}{G}$

This value should be less than the assumed system flooding factor. Consider the following flooding factors:

| | |
|---|---|
| In General | 0.82 |
| Vacuum Towers | 0.77 |
| For towers less than 900 mm diameter | 0.65 |

(d) Downcorner Area Calculation

Downcorner area required for a service can be obtained form the downcorner design velocity. Downcorner velocity can be read from Figure 1.4 or calculated as follows:

$V_D$ = 150*$G$ (where G is the system factor)

$V_D$ = $25\left(D_L - D_V\right)^{\frac{1}{2}} G$

$V_D$ = $4.5\left(TS\right)^{\frac{1}{2}}\left(D_L - D_V\right)^{\frac{1}{2}} G$

$T_S$ = Tray spacing, inches

$D_L$ = Liquid density in lb/ft$^3$

$Dv$ = Vapour density in lb/ft$^3$

The above equations give $V_D$ in gpm/sq.ft.

$$V_D = 2.5 * \frac{gpm}{sq.ft} m^3 / m^2 hr$$

The minimum of the values calculated above is selected. Then,

$$\text{Downcomer area} = \frac{\text{Liquid flow (m}^3\text{ / hr)}}{V_D}$$

### 1.10.2 Sieve Tray Sizing

The active area is estimated by calculated the flooding velocity through the column . The following correlations are used:

$$F_{LV} = \frac{L}{G}\left(\frac{D_V}{D_L}\right)^{0.5}$$

$$C_{abf} = U_{nf}\left(\frac{D_V}{D_L - D_V}\right)^{0.5}$$

Where $L$ and $G$ are mass flow velocities of liquid and vapour per unit tower area respectively in $lb/hr\ ft^2$. $D_L$ and $D_V$ are liquid and vapour densities in lb/ft$^3$. $U_{nf}$ is the flooding vapour velocity in ft/sec. Figure 1.5 indicates the parameters for various tray spacing for liquid surface tension of 20 dynes/cm. The steps are as follows:

*(i)* Calculate $F_{LV}$

*(ii)* Assume tray spacing and use Figure 1.5 to read $C_{abf}$

*(iii)* Use the value of $C_{abf}$ to calculate $U_{nf}$

(iv) Correct $U_{nf}$ for hole/active area ratio using the following guideline.

| Hole/active area Ratio | Multiply $U_{nf}$ calculated in step (iii) by |
|---|---|
| 0.10 | 1.0 |
| 0.08 | 0.9 |
| 0.06 | 0.8 |

(v) For tower design, use 80% of $U_{nf}$ calculated in step (iv)

*(vi)* Calculate tower area = *Vapour flow / $U_{nf}$ calculated in step (v)*

### 1.10.3 Bubble Cap Tray Design

For the bubble cap trays, use the following correlations:

$$F_{LV} = \frac{L}{G}\left(\frac{D_V}{D_L}\right)^{0.5}$$

$$C_{ab} = 1.3\ U_N \left(\frac{D_V}{D_L - D_V}\right)^{0.5}$$

Where, L and G are mass flow rates of liquid and vapour respectively in *lb/hr*. $D_L$ and $D_V$ are liquid and vapour densities in $lb/ft^3$. $U_N$ is the vapour velocity based on the active area above the tray. Figure 1.6 indicates the parameters for various tray spacing for liquid surface tension of *20 dynes/cm*. The steps are as follows:

(i) Calculate liquid and vapour flow rates in $m^3/hr$.

(ii) Corrected vapour flow = Vapour flow $(m^3/hr)\left(\frac{D_V}{D_L - D_V}\right)^{0.5}$

(iii) Calculate downcorner velocity as explained for valve tray sizing.

(iv) Use downcorner velocity and liquid flow rate to find downcorner area.

*(v)* Calculate $F_{LV}$

(vi) Use figure 1.6 to read $C_{ab}$

(vii) Use the value of $C_{ab}$ to calculate $U_N$

(viii) Calculate tower active area = (Vapour flow rate)/ ($U_N$* flooding factor *system factor)

(ix) Total area = Active area + downcorner area.

(x) From tower area calculate tower diameter. This is the I.D. of the tower. Round this to the nearest standard O.D. after accounting for shell thickness.

### 1.10.4 Design of Packed Columns

The foregoing sections give the guidelines for sizing of columns, having trays. The use of packings as a substitute for trays in mass transfer as well as hear transfer service has gained widespread acceptance because of the advantages of increased capacity. Lower pressure drops improved quality of products like vacuum gas oil and lower energy requirements.

For sizing of a packed tower reference is made to Figure 1.7 This figure gives the capacity factor for glitsch grid as a function as liquid rate. The procedure for sizing the column is as follows:

(i) Calculate liquid load in GPM

(ii) Calculate vapour load in $ft^3$/sec.

(iii) Corrected vapour load = Vapour load ($m^3/hr$)$\left(\frac{D_V}{D_L - D_V}\right)^{0.5}$ where $D_L$ and $D_V$ are the densities of liquid and vapour in lb/ft$^3$

(iv) Assume tower diameter and calculate tower area in ft$^2$

(v) Calculate liquid rate in ft$^2$

(vi) Using the liquid rate calculated in step (v) read off the vapour capacity factor from Figure 1.7. This gives the maximum capacity factor corresponding to flooding.

(vii) Use the corrected vapour load and assumed tower area to calculate vapour capacity factor. This value should be about 80% of the value read off from Figure 1.7 in step (vi).

In case the value is higher, select a higher tower diameter and repeat the procedure.

The above calculations are based on using glitsch grid as the tower packing. It is recommended to design new columns at not more than 80% of the flood point. A C factor of about 0.4 to 0.45 may be used for most services. If pressure drop is critical, a design of somewhat lower values may become necessary. In case other packings such as pall rings, IMTP, Norton Hypak etc. are used, the C factor considered for preliminary sizing of the tower should be around 0.35.

The volume (and height) of the packings are calculated by the mass transfer specialist on the basis of heat transfer across the packing, HETP (Height Equivalent to Theoretical Plate) considerations and pressure drop across the packing.

## 1.11 Tower Height and Diameter

The foregoing section describes how the process engineer can estimate the diameter of a tower, using the internal traffic data generated during the heat and material balance calculations. In order to estimate the tower height, the following guidelines would be helpful.

- Provide at least 150 mm space between low liquid level and bottom tangent line.

- Provide 5 minutes hold up between LLL and HLL. However, for equipment like vacuum distillation, FCC and Visbreaker Fractionators etc. where coking is expected, the residence time should be kept low to reduce coking.
- The space between HLL and the first bottom tray should be equal to 300 mm + Tray spacing (TS) + dia of steam inlet nozzle if any.

- Allow adequate space between the feet nozzle and first tray above it in vacuum towers to prevent carry over of asphaltenes etc. to the trays above.
- In case of a change in tower diameter allow a minimum of twice the normal tray spacing but not less than 1200 *mm* between the trays.
- Allow a minimum of TS + 150 mm for the downcorner if a scalpan is to be used underneath.
- Allow a minimum of (1.5*TS) or 750 *mm* in case of a transition due to change in number of flow paths between two trays.
- Allow a minimum of (TS * 1.5) or 750 *mm* whichever is more between trays with an intermediate feed.
- Allow a minimum of (TS * 1.5) or 750 *mm* whichever is more between trays accommodating draw off pans.
- For calculating spacing between a chimney tray and the tray above, calculate the liquid height between LLL and HLL on the basis of 2-3 *minutes* hold up when liquid is withdrawn under level control. Provide a minimum of 150 *mm* between LLL and tray deck and between HLL and top of chimney. The total spacing would normally be larger than 1500 mm.
- At the top of the tower, provide a minimum of (1.5 * TS), or 750 *mm* whichever is more, for vapour disengaging space between the top tray and top tangent line.
- Add the above margins whichever are applicable to the height calculated based on the number of trays provided and tray spacing. The height so estimated should be specified in the process data sheet. This should however, be reviewed by the mass transfer specialist from mechanical and other aspects and corrected, if necessary.

## 1.12 Pressure Drop in Towers

The following guidelines can be used while specifying pressure drops across various tower internals.

| Service | Pressure Drop |
|---|---|
| Trays in Atmospheric Distillation towers, stabilizers, absorbers | 0.01 $kg/cm^2$ per tray |
| Trays in vacuum towers | 3 mm Hg max per tray |
| Chimney trays | 0.5-1 mm Hg per tray |
| Demisters | 0.5 mm Hg |
| Grid of Packings | 1 mm Hg per metre of bed |
| Random packings (e.g. Pall rings) | 1-1.5 mm Hg per metre of bed |

The above are indicative only. Normally a vacuum column having column having packing in all the sections is designed for a total pressure drop of 15 mm Hg. The diameter and the distribution of pressure drops in various sections are calculated by the mass transfer specialist during detailed design.

## 1.13 Other Internals

The foregoing sections have concentrated on the internals which are commonly used for liquid vapour contacting. Additionally, one can consider the following internals which are of column.

### 1.13.1 Mist Eliminators

These are used to prevent carry-over of entrained liquid in the vapours leaving a particular section of the column. Thus, they are provided above the wash some packing in a vacuum column to prevent carry-over entrained asphaltenes into the vacuum gas oil. Provision of the demisters in the top section of many columns prevents the carry over of liquid hydrocarbons to overhead system connected to an ejector, compressor, etc. Such service may include vacuum towers, absorbers and columns in foaming services. While giving the specifications for the demisters, the operating conditions of requirements (droplet size to be retained), allowable pressure drop and material of construction should be clearly spelt out. The pertinent data for demisters should be filled in the standard data sheet.

### 1.13.2 Spray Nozzles

These are very vital to ensure the performance of the packing in the columns. Gravity flow through a downcorner will not distribute over the packed uniformly. It is, therefore, essential that the liquid be sprayed over the packing in a pattern which ensures uniform distribution over the cross section. The liquid being sprayed through the nozzles should be clear of particulate matter to avoid erosion and choking of the spray nozzles. For this purpose, appropriate strainers are generally used.

A pressure drop of 2-3 $kg/cm^2$ through the spray nozzles should normally be considered when pumping requirements are calculated.

### 1.13.3 Chimney Trays

These are normally provided in columns when the liquid to the section below the chimney tray has to be supplied under pressure e.g. through a spray nozzle. The liquid is generally withdrawn under level control of the tray and sufficient hold up should be provided to ensure that the associated pumps do not lose suction or to take care of eventualities like switching over to the spare pump in case of pump failure. A minimum hold up of 2-3 minutes should be provided.

## 1.14 Material of Construction

A variety of materials are available for trays and other internals of columns. The exact selection will depend on the service under questions. The process engineer should get the recommendations of the material specialist for the material selection and specify the same in the process data sheet. The corrosion allowance for tray should be specified as total corrosion allowance as the tray is subjected to corrosive atmosphere only from inside. Corrosion allowances for trays of alloy materials is generally nil and for C.S. is 3.0 mm

## 1.15 Draw-of Nozzle Sizing

Table 1.4 is intended for use when sizing lines and nozzles under self venting flow conditions, particularly draw-off lines from column trays etc.

Table 1.4: Draw –off Nozzle Sizing

| Size, mm | Hot Liquid flow $m^3/hr$ | Min. Liquid Head Above Nozzle, mm |
|---|---|---|

| | | |
|---|---|---|
| 38 | Upto 1.0 | 20.0 |
| 50 | 1.1-2.0 | 20.0 |
| 80 | 2.1-5.0 | 20.0 |
| 100 | 5.1-10.0 | 25.0 |
| 150 | 10.1-30.0 | 40.0 |
| 200 | 31.0-60.0 | 50.0 |
| 250 | 61.0-100.0 | 80.0 |
| 300* | 101.0-200.0 | 80.0 |
| 350* | 201.0-250.0 | 100.0 |
| 400* | 251.0-300.0 | 100.0 |
| 450* | 301.0-400.0 | 100.0 |
| 500* | 401.0-500.0 | 150.0 |

Notes :

* It is preferred to use more than one draw off connection if draw off nozzle size exceeds 300mm.

# CHAPTER - 2

# PROCESS VESSELS

## 2.1 General Guidelines

Design and specification of process vessels is one of the most important jobs of a process engineer. Vessels constitute a major cost component in any process plant. Their contributions to the total plant investment may vary between 10-30%. Moreover, smooth operation of several critical equipment in a process plant largely depends upon the performance and adequacy of process vessels. Therefore, it becomes necessary that extreme care be taken while designing and specifying process vessels.

### 2.1.1 Classification of Process Vessels

Process vessels can be classified into several categories based upon their function and location in the plant. However, fundamentally, process vessels can be divided into following categories:

(i) Two phase separators
- Vapour- liquid separator
- Liquid-Liquid separator
- Vapour- solid separators

(ii) Three phase separators
- Vapour-liquid-liquid separators

(iii) Hold-up drums

The above categories would include following types of process vessels based on their popular nomenclature:

**Two Phase Vapour-Liquid Separators**

- Compressor Interstage knock-out drums (Single liquid phase)
- Reflux drums for partial/compound condensers
- Refrigeration system interstage drums
- Flare knock-out drums
- Steam drums.
- Fuel Gas drums
- Condensate flash drums

**Two Phase Liquid-Liquid Separators**

- Immiscible liquid-liquid separators
- Coalescers
- Water boots

**Three Phase Vapour-liquid-liquid Seprators**

- Compressor interstage knock-out drum (two liquid phases)
- Reflux drums (two liquid phases)

**Two Phase Vapour-Solid Separators**

- Decoking drum

**Hold-up drums**

- Feed/Product Surge drums
- Reflux drums (total condenser, single phase)
- Hot wells for steam jet ejectors
- Refrigerant accumulators
- Chemicals day/dosing tanks

### 2.1.2 Design Concepts

Design philosophy for process vessels depends upon the function they are required to perform. As clear from their classification, prime functions that a vessel may perform are:

- Separation of vapours from liquid
- Separation of two immiscible liquid phases.
- Separation of solid particles from liquid phase.
- Hold up volume for liquids.

The final design of the vessel should be able to perform one or more above listed functions as per the need of the process. Design criteria and general guidelines for sizing vessels for above mentioned functions are discussed below.

**Separation of vapour from Liquid**

Separation of vapour and liquid in a vessel is accomplished by the virtue of density difference aided by gravity force. Gravity accelerates the falling of a particle until its force if offset by drag force. From then on, the particle falls at constant velocity known as *Terminal Velocity or Critical Velocity*.

For spherical particle, critical velocity is given by

$$V_c^2 = \frac{4gd(S_L - S_v)}{3S_V C} \qquad (2.1)$$

where,

$Vc$ = Critical velocity m/sec.

$d$ = Particle diameter m

$S_L$ = Liquid density kg/m$^3$

$S_V$ = Vapour density kg/m$^3$

$C$ = Drag Coefficient

The above equation can be rewritten in following form:

$$V_c = K\left(\frac{S_L - S_V}{S_V}\right)^{0.5} \qquad (2.2)$$

Where, $K$ is constant with units of velocity and depends upon system characteristics like surface tension, droplet size, viscosity etc. For most industrial system k lies between 0.03 and 0.10 in *m/s* units. A typical value of 0.048 is recommended for designing process liquid vapour separators. For specific system, if detailed analysis becomes necessary, $K$value can be estimated from Figure 2.1.

For designing a liquid-vapour separator actual vapour velocity or maximum allowable vapour velocity should always be less than the critical velocity calculated above. However, if special internals (like demister pad etc.) are provided , vapour velocity higher than critical may also be acceptable.

**Separation of Two Immiscible Liquid Phases**

The separation of two immiscible liquid phases occurs under the influence of gravity, owing to the difference in the density of two liquids. The rate of settling depends upon viscosity, droplet size and density difference. Stoke's law is used to determine the settling velocity. The correlation for velocity changes with turbulence encountered in the system. The indicating parameter for turbulence is the Reynold's number $N_{Re}$ defined as

$$N_{re} = \frac{1.667dvs}{\mu} \qquad (2.3)$$

For Reynold's number less than 2 (Laminar region), settling velocity is calculated as

$$V = \frac{3.268x10^5 xd^2(S_1 - S_2)}{\mu} \qquad (2.4)$$

For Reynols's number between 2 to 500 (Transition region), settling velocity is

$$V = \frac{8000xd^{1.14}S_1^{0.71}}{S_2^{0.29}\mu^{0.43}} \qquad (2.5)$$

For Reynold's number higher than 500 (Turbulent Region) settling velocity is calculated using Newton's law and is given by:

$$V = 3250\left[\frac{d(S_1 - S_2)}{S_2}\right]^{0.5} \qquad (2.6)$$

where

| | | |
|---|---|---|
| $V$ | = Settling velocity | *cm/min* |
| $d$ | = droplet diameter | *cm* |
| $S_1$ | = density of heavy phase | $9/cm^3$ |
| $S_2$ | = density of light phase | $g/cm^3$ |
| $\mu$ | = viscosity of continuous phase | cp |

For calculation of setting velocity and Reynold's number of heavy phase droplets in continuous light phase, viscosity and density of light phase should be used. On the other hand when rising velocity of light phase droplet in heavy phase is to be calculated, viscosity and density of heavy phase should be used. For calculating settling velocity, assume the region calculate velocity and check back the Reynold's Number. For designing a settling drum, dimensions and volume of vessel should be adequate to allow the separation of light and heavy phases.

**Separation of solid particles from Vapour Phase**

Separation of solid particles from vapour phase occurs under the influence of gravity. Forces which affect the rate of settling of solid particles are:

(i) gravity force

(ii) drag force due to relative motions

(iii) buoyancy force

Setting velocity is determined by the same equation as for vapour-liquid separation (equation 2.2) except that density of liquid phase ($S_L$) is replaced by the density of solid particle.

**Liquid Hold-Up**

Quite often, the function of the process vessel is to provide a hold-up or surge volume for liquid stream. Under such circumstances, it is necessary to establish the hold-up time and volume of liquid. Once this data is established, the vessel can be sized for a reasonable length to diameter ratio without any consideration to vapour and/or liquid velocity.

**Multiple Function Process Vessels**

Most of the vessels in Process industries have to perform more than one function like separation of two or three phases along with h old-up of liquid phase. Under such circumstances, the process vessel should be designed to satisfy all functional requirements. For example, a compressor knock out drum may have to separate vapour, hydrocarbon liquid and water. It may also be necessary to provide hold-up volume for the liquid phases. This means that the vessel dimensions should satisfy following criteria:

1. Vapour velocity as per vapour-liquid separation
2. Liquid droplet velocity as per liquid-liquid separation
3. Liquid hold-up as required

Detailed design procedure for such vessels as well as for single function vessels are given in subsequent sections.

### 2.1.3 General Guidelines For Process Vessel Design

**Types of Vessel**

A process vessel could be horizontal or vertical. Spherical vessels may also be used for high pressure and high liquid hold-up systems like storage of light hydrocarbons etc. However, design of spherical vessels is not included in this manual. The choice between horizontal or vertical type of vessel primarily depends upon following process requirements:

- Relative liquid and vapour load
- Availability of plot area
- Economics
- Special considerations

From process point of view and relative vapour and liquid load consideration, general guidelines given in Table 2.1 may be used to decide the type of vessel.

Table 2.1 : Selection Guidelines for Type of vessel

| System Characteristics | Type Of Vessel |
|---|---|
| Large vapour, less liquid Load (by volume) | Vertical |
| Large liquid, less vapour Load (by volume) | Horizontal |
| Large vapour, large liquid Load (by volume) | Horizontal with split flow (*) |
| Liquid-liquid separation | Horizontal |
| Liquid-solid separation | Vertical |

*Split flow vessel has one entry in the centre and two vapour outlets (one at each end)

Sometimes, the choice between horizontal and vertical vessel may not be obvious. Under these conditions, it is recommended that design be performed for both types and cheaper one should be selected. It may be noted that cost of the vessel would depend upon its dimensions (height and diameter), wall thickness case of fabrication etc. A simple comparison of height and diameter may not always give the true picture and a detailed analysis may be necessary.

**Height/Length to Diameter Ratio**

Whenever liquid hold-up controls the size of the vessel ( large liquid rate and less vapour rate), it is possible to change the height/length (between bottom and top tangent line) to diameter ratio of the vessel and still meet the vapour-liquid separation requirement. In such cases, a reasonable value of this ratio should be selected based on following considerations:

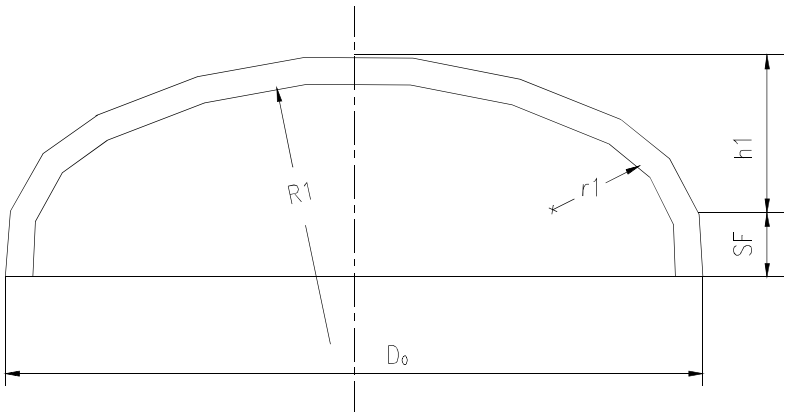
(i) Vessel diameter should always be as per standard dished end sizes available (given in Figure 2.2). Increasing diameter will reduce length/height. On the whole, surface area of the vessel will reduce.
(ii) Wall thickness will increase with increasing diameter.
(iii) Plates are available in standard thickness and change occurs in steps (given in Table 2.2). Therefore, as long as in rease in diameter does not result in step change in thickness, diameter should be increased.
(iv) Plate sizes are also standard (Given in Table 2.2). It is difficult to effectively use the small cutouts from plates. Therefore, adjusting height/length and diameter within the standard plate size will also reduce cost. Based on above considerations, it has been found that a ratio of about 2.5 to 3.5 is optimum. It is to be noted that selection of this ratio has to be made within the constraints of process requirements i.e. critical velocity of vapours, settling velocity of liquid etc.

**Liquid Hold-up Volume**

Liquid hold-up volume in vessel is based on downstream process requirements. General guidelines for this are given in Table 2.3.

Figure 2.2: Dimensions for Dished Ends

## DIMENSIONS FOR DISHED ENDS.




| Outside diameter ($D_0$) | Crown Radius ($R_1$) | Intside Knuckle Radius ($r_1$) | Outside diameter ($D_0$) | Crown Radius ($R_1$) | Intside Knuckle Radius ($r_1$) |
|---|---|---|---|---|---|
| 300 | 240 | 50 | 2000 | 1600 | 300 |
| (350) | 240 | 50 | (2100) | 1680 | 350 |
| 400 | 280 | 60 | 2200 | 1760 | 350 |
| 500 | 320 | 60 | (2300) | 1840 | 350 |
| 600 | 400 | 80 | 2400 | 1920 | 400 |
| 700 | 560 | 125 | 2600 | 2080 | 400 |
| 800 | 640 | 125 | 2800 | 2240 | 450 |
| 900 | 720 | 150 | 3000 | 2400 | 450 |
| 1000 | 800 | 150 | 3200 | 2560 | 500 |
| 1100 | 880 | 175 | 3400 | 2640 | 600 |
| 1200 | 960 | 200 | 3600 | 2880 | 600 |
| (1300) | 1040 | 200 | (3800) | 3040 | 600 |
| 1400 | 1120 | 230 | 4000 | 3200 | 600 |
| (1500) | 1200 | 250 | 4250 | 3400 | 700 |
| 1600 | 1280 | 250 | 4500 | 3600 | 700 |
| (1700) | 1360 | 300 | 4750 | 3800 | 800 |
| 1800 | 1440 | 300 | 5000 | 4000 | 800 |
| (1900) | 1520 | 300 | | | |

Notes:

1. Values in brackets are second preference values and shall be avoided where possible.
2. The length of the straight flange shall not be less then three times the end thickness with a minimum of 20 mm unless otherwise agreed between the manufacturer and the purchaser.
3. For elliptical heads, take $D_0 / R_1 = 2$.
4. For hemispherical heads, take $D_0 / R_1 = 4$.

Table 2.2: Standard Plate Thickness

A. for Carbon Steel Plates (mm)

| | | | | | |
|---|---|---|---|---|---|
| 5 | 6 | 8 | 10 | 12 | 14 |
| 16 | 18 | 20 | 22 | 25 | 28 |
| 30 | 32 | 35 | 37 | 38 | 40 |

B. For Stainless Steel Plates

| | | | | | | | |
|---|---|---|---|---|---|---|---|
| 2 | 3 | 4 | 5 | 6 | 8 | 10 | 12 |

C. For Carbon Steel Plates

(i) 2.5 meter x 6.5 meters

(ii) 3.0 meters x 6.5 meters

## 2.2 Detailed Procedures For Process Vessels Design

Detailed design procedures for various types of process vessels is discussed in this chapter as per their classification made in Section 2.1. Special considerations given in Section 2.2.9 should be taken care of during design of process vessels.

### 2.2.1 Two Phase V-L Separator

Two phase V-L separators will include compressor interstage knock out drum, reflux drum, refrigeration drum etc. as listed in Section 2.1. Stepwise procedure for sizing these vessels is given below:

Table 2.3 : Normal Liquid Hold-up in Process Vessels.

| Service | Liquid Residence Time (Minutes) |
|---|---|
| Reflux to Tower | 5 |
| Distillate to Storage | 2 |
| Distillate to Surge drum of other unit | 2 |
| Distillate/liquid feed to subsequent Tower or Furnace | |
| Drum diameter less than 1.2 meters | 20 |
| Drum diameter 1.2 to 2 meters | 15 |
| Drum diameter above 2 meters | 10 |
| Compressor Interstage knockout drums | 5 |

| | |
|---|---|
| Refrigeration Systems | |
| Intermediate Stage drums | 5 |
| Main accumulator | 15 |
| Feed surge drum | 10-30 |
| Hot Oils Surge Drum | 10 |
| Very low liquid rate (Manual intermittent draining) | 8-24 hours |
| Other general services | 3-10 |

**Notes:**

1. When large hold-up is required for reflux and distillate, use larger volume as distillate hold-up rater tan sum of the two.

1. Select type of vessel - horizontal or vertical.
2. Calculate critical velocity $V_c$ using equation 2.2 in Section 2.1 using K = 0.048 (in MKS Units).
3. Calculate maximum allowable vapour velocity $V_{max}$ as follows :

   Without mist eliminator : $V_{max} = 0.8\ V_c$

   With mist eliminator : In general $V_{max} = 1.7\ V_c$

   If specific information from the vendor of mist eliminator s available use that to find out $V_{max}$.
4. Calculate normal liquid hold-up volume as per Table 2.4
5. For a given allowable velocity and liquid hold-up, dimensions for a vertical and horizontal vessel would be different and shall be calculated as follows :

## 5.1 Vertical Vessel

5.1.1 Minimum drum diameter, $D_{min}$

$$D_{min} = 1.129 \left(\frac{\text{Vapour flow rate}}{V_{max}}\right)^{0.5}$$

Vapour flow rate : $m^3/sec$
$V_{max}$ : m/sec
$D_{min}$ : m/sec

5.1.2 Equivalent drum height between LLL and HLL ($h_2$ meters)

$$h_2 = 1.273 * \text{Normal hold-up volume } (m^3)/D_{min}^{\ 2}$$

5.1.3 Establish total vessel dimensions using Table 2.4. This will require calculation for inlet nozzle size, which can be taken equivalent to pipe size. Figure 2.3 should be used for listing all elevations in a vertical vessel.

5.1.4 Check that L/D is between 2.5 to 3.5 If not, increase the diameter and repeat the calculations.

## 5.2 Horizontal Vessel

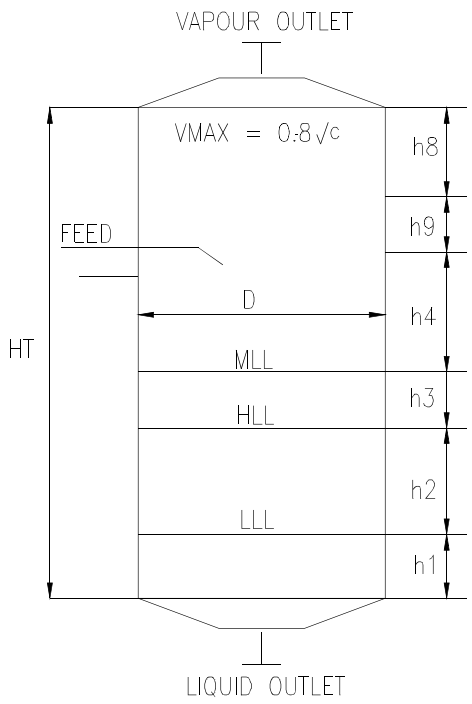
In a horizontal vessel, the vertical cross section is divided between vapour and liquid phases. Trial procedures becomes necessary in this case.

Table 2.4 : Liquid Levels and Relative Locations of Nozzles, Mist Eliminators etc.

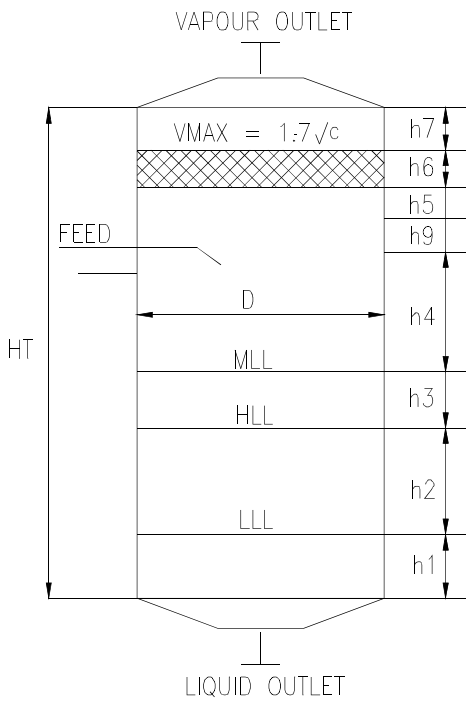
| Particulars | Notation | Height (under level control) |
|---|---|---|
| Height of LLL from lower tangent line | $h_1$ | 150 mm (Note-1) |
| Height of HLL from LLL | $h_2$ | Depends on hold-up |
| Height of maximum liquid level (MLL) from HLL | $h_3$ | 350 mm |
| Height of bottom of Inlet nozzle from MLL | $h_4$ | 150 mm |
| Top of inlet nozzle to bottom of mist eliminator | | |
| D less than 900 mm | $h_5$ | 300 mm |
| D more than 900 mm | $h_5$ | 450 mm |
| Thickness of mist eliminator | | |
| D less than 900 mm | $h_6$ | 150 mm |
| D more than 900 mm | | |
| Non-coking service | $h_6$ | 100 mm |
| Coking service | $h_6$ | 150 mm |
| Distance from top of mist eliminator to vapour outlet for | | |
| D less than 1200 mm | $h_7$ | 700 mm |
| D more than 1200 mm | $h_7$ | 900 mm |
| Inlet nozzle to vapour outlet nozzle (for drum with no mist eliminator) | $h_8$ | Max. of 900 mm or 35% of D |

Notes :1. If liquid draw-off is on manual control then $h_1 = 0$ and $h_2 = 200$ mm.

FIGURE 2.3: LIQUID LEVELS AND NOZZLE ELEVATION FOR VERTICAL VESSELS.




WITH OUT MIST ELIMINATOR




WITH MIST ELIMINATOR

h1 = 150 M.M.
h3 = 350 M.M.
h4 = 150 M.M.
h8 = MIN. 900 M.M. OR 0.35D
h9 = DIAMETER OF FEED NOZZLE M.M.

LLL = LOW LIQUID LEVEL
HLL = HIGH LIQUID LEVEL
MLL = MAXIMUM LIQUID LEVEL

5.2.1 Assume vessel volume 1.2 * normal liquid hold-up.

5.2.2 Calculate vessel length (L) and diameter (D) using L/D = 3 and neglecting volume of dished ends.

5.2.3 Vertical cross section Area Ar =0.785 $D_2$

5.2.4 Total drum volume VT (including dished end) is calculated from Figures 2.6 to 2.10. Horizontal axis has L/D ratio. Vessel volume is plotted on vertical axis for various diameters Graphs are available for both elliptical ends as well as hemispherical ends.

5.2.5 Select Low liquid level (LLL);

5.2.6 $h_1$ = 150mm and find volume of liquid ($V_1$) for height ($h_1$) from Figure 2.11. Horizontal axis is h/D and vertical axis is v/V where,

h = Liquid height

d = Drum diameter

v = Volume of liquid for h height

V = Total vessel volume (inclusive of dished ends)

5.2.7 For establishing the high liquid level HLL ($h_2$), the distance of HLL from the top of the drum should be 20% of the drum diameter or 300 mm whichever is greater. However, special consideration should be given when the gas product discharges into a refinery gas system or other system where liquid carry-over would constitute a potential hazard. In such a case make $h_2$ = d/3 or 300 mm whichever is greater.

5.2.8 Calculate volume $V_2$ for height $h_2$ using Figure 2.11.

5.2.9 Hold-up volume available $V_3$ where

$V_T$ = Total volume of drum

$V_1$ = Volume below LLL

$V_2$ = Volume above HLL

5.2.10 Check that the available hold-up volume $V_3$ is slightly higher than the normal hold-up volume calculated in step 4. If not, assume a larger diameter D and repeat the procedure from steps 5.2 through 5.9. if the available hold-up volume is more, a reduction in drum size could be achieved by assuming a smaller diameter and repeating the procedure from steps 5.2 through 5.9

5.2.11 $A_2/A_T$ : Knowing $h_2$ /D the ratio of segmental cross-sectional area above HLL ($A_2$) to the total area of cross section ($A_T$ ) can be read from Figures 2.12 to 2.14.

$$A_2 = \frac{A_2}{A_T} * A_T$$

5.2.12 For calculating vapour velocity in drum, divide vapour flow rate ($m^3$/sec) by vapour space cross-sectional area $A_2$ ($m_2$).

Check that the vapour velocity calculated in step 5.11 does not exceed the maximum allowable vapour vapour velocity, calculated in step 3. If it does, the vapour space in the drum should be increased. For this, assume a large diameter D and repeat the procedure from steps 5.3 to 5.11 by increasing the vapour space height.($h_2$).

After vapour velocity and hold-up criteria are met, final diameter and length should be decided as per Figure 2.4 and rounded off to standard sizes available.

### 2.2.2. Two Phase L-L Separator

Two phase liquid separators are used to separate immiscible liquid phases. Most of the time, whenever no vapour phase is involved, these vessels are horizontal. Following step-wise procedure should be used for design of horizontal L-L Separators.

1. Calculate hold-up volume of both liquid phases or residence time required, if any, for either phase for controlling process, or for start-up, make-up etc. Use table 2.3 if necessary. Nomenclature for hold-up volume is :
   $v_1$ = Hold-up volume for heavy phase
   $v_2$ = Hold-up volume for light phase
2. Minimum droplet diameter (d)
   Take d = 0.009 cm for both light and heavy phases. When separation of heavy phase from light phase is important (light phase is continuous), follow step 3 through 18.
3. Calculate settling velocity of heavy phase droplets in light phase using equations given in Section 2.1.2
4. Design settling velocity of heavy phase droplets: If the calculated settling velocity is higher than 25 cm/ min take the design settling velocity as 25 cm / min.
5. Diameter of drum: Assume that heavy phase occupies 1/3 of the total volume of the vessel.
6. Assume L / D = 4.

FIGURE 2.4: LIQUID LEVELS AND NOZZLE ELEVATION FOR HORIZONTAL VESSELS.

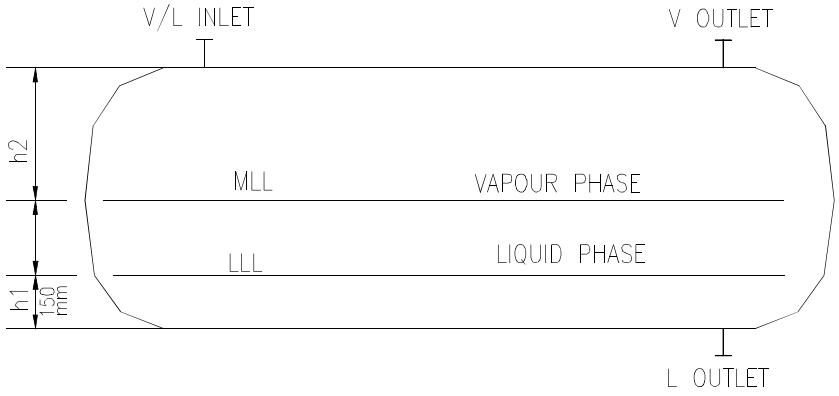



FIG. A V–L SEPARATOR (HORIZONTAL)

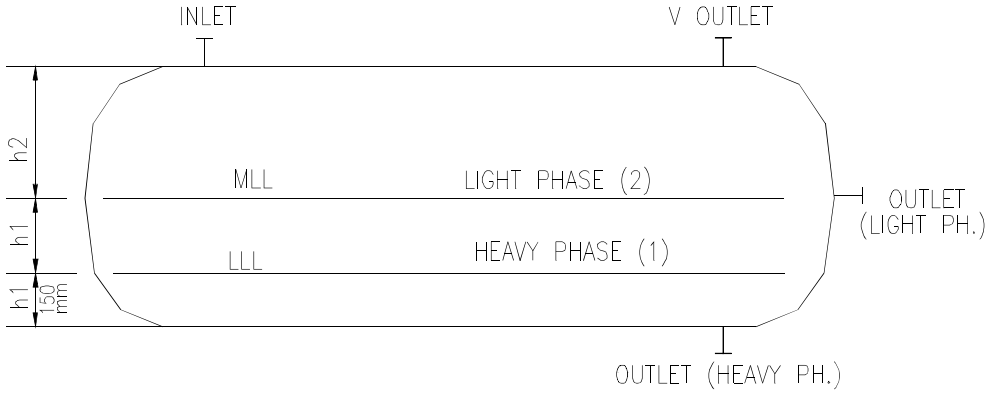



FIG. B L–L SEPARATOR (HORIZONTAL)

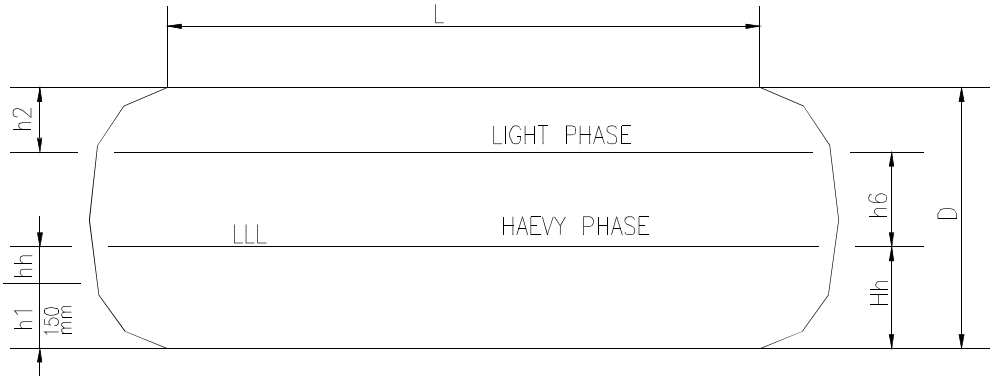



FIG. C V–L–L SEPARATOR (HORIZONTAL)

7. Based on the assumptions in steps 5 and 6, calculate vessel length and diameter excluding volume of dished ends.
8. Total volume of drum (VT): For a given diameter (meters) and length (meters), the total volume (m3) of a drum including the elliptical ends, is read from Figures 2.6 to 2.10.
9. $V_2/V_T$ = Hold-up volume of light phase (m3)/Total volume of drum ($m^3$)
10. h1 /D : knowing v2/VT the ratio h2/D can be read from Figure 2.11.

    $h_1$ = distance of the interface from low liquid level in the vessel.

    $h_2$ = distance of the interface from top of the drum (mm).

    Calculate h1 and h2
11. Calculate $h_1$ /D.

    Knowing $h_1$/D, the ratio $V_1/ V_T$ can be found from Figure 2.11.
12. Check that the calculated volume V1 is equal to or higher than the required hold-up volume of the heavy phase liquid. If not, assume a larger diameter D and repeat the procedure from steps 5 to 12.
13. The flow path of heavy phase droplets is height of light liquid phase (h2). Settling time for heavy phase droplets is calculated as follows.

    Settling Time (minutes) = $h_2$/Settling velocity (cm/min)

    Where, $h_2$ is the settling path length in cm.
14. Flow path length for light phase will be less than the length of drum due to nozzle arrangement. For estimating purposes, calculate flow path length as follows:

    **Flow path length (mm) = L – (di + do + 300)**

    Where,

    L = Length of drum (mm)

    $d_1$ = inlet nozzle o.d. (mm)

    $d_o$ = light phase outlet nozzle o.d. (mm)

    The nozzle sizes have to be estimated using the standard procedure given in subsequent sections.
15. $A_2/A_T$ : Knowing $h_2$ / D, the ratio of segmental area above interface to the total area of cross section of drum ($A_2/A_T$) can be found from Figure 2.12 to 2.14. Calculate $A_2$ (area of cross section occupied by light phase).
16. Calculate longitudinal velocity of light phase liquid in mm / min units as follows:

    **$V_L$ = Flow rate of light phase/$A_2$**
17. Residence time of light phase (min) is obtained by dividing the flow path length (mm) by the longitudinal velocity (mm / min).
18. Check that the residence time of the light phase fluid is slightly higher than the settling time of the heavy phase droplets. If not, a new diameter D must be assumed for the drum and the calculation repeated from steps 5 to 18.

When the separation of the light phase from the heavy phase is of equal importance as the separation of heavy phase from the light phase, a rate must be calculated for the rise of the light droplets through the heavy phase in essentially the same manner as discussed above for the light phase.

19. Calculate rising velocity of light phase droplets in heavy continuous phase using equations given in Section 2.1.2.
20. Design velocity of rise of light phase droplets : if the calculated rising velocity is higher than 25 cm / min, take the design velocity of rise as 25 cm / min.
21. Time for size of light phase droplets to the interface is obtained by dividing h2 (distance from the bottom of the drum to the interface) by the velocity of rise.
22. **Flow path length for heavy phase fluid = L – (d1+do+300)**

    Where,

    L = length of drum (mm)

    D1 = inlet nozzle o.d. (mm)

    Do = heavy phase outlet nozzle o.d. (mm)
23. Calculate area of cross section occupied by heavy phase

$$A_1 = A_T - A_2$$

24. Calculate longitudinal velocity of heavy phase fluid 40 mm/min units

$$V_H = \text{Flow rate of light phase} / A_1$$

25. Residence time of heavy phase (minutes) is obtained by dividing flow Path length by longitudinal velocity.
26. Check that the residence time of the heavy phase fluid is slightly higher than the rising phase droplets. If not a new diameter must be assumed and calculations should be repeated from steps 5 to 26.

### 2.2.3. Water Boot Sizing

When small quantity of water is settled, a draw off pot is provided to make separation of water easier. Draw off pot (water boot) sizing procedure is a as follows:

1. Water boot diameter ($d_B$)

Water boot diameter should be in between $d_{min}$ and $d_{max}$ defined as defined below:

| Drum Diameter (D) | Dmin | **Dmax** |
|---|---|---|

| | | |
|---|---|---|
| Less than 1500 mm | 300 mm | 1/2 D |
| 1500 – 2400 mm | 450 mm | 1/3 D |
| More than 2400 mm | 500 mm | 1/3 D |

2. Water Hold-Up ($V_b$)
3. Calculate height of water corresponding to its hold-up.
4. Total boot height = 150 mm + height corresponding to water hold-up.
5. The maximum boot height should be 1500 mm. If total boot height as calculated above is more than 1500 mm, assume boot height as 1500 mm and proceed from steps 6 to 10.
6. Calculate boot diameter for desired hold-up taking height as 1500 mm, if it is more than $d_{max}$ assume high interface level (HLL) within main drum, say 75 or 100 mm form bottom.
7. Determine the volume of drum corresponding to HLL and subtract from the required hold-up volume for water to obtain the hold-up volume to be provided in the boot.
8. Calculate the boot diameter based on revised hold-up criteria.
9. If height of boot is still more than 1500 mm provide automatic interface level control in the boot.
10. Boot size for automatic water drain:
    LLL : 150 mm from the lower tangent line of boot.
    HLL : 250 mm from bottom of the main drum.
    Distance between HLL and LLL is 350 mm or 5min hold-up on water whichever is higher.
11. Total boot height = Between 900 mm and 1500 mm.

### 2.2.4. Coalescer Design

When water is present in hydrocarbon stream in finely dispersed form, the size of water droplets is too small to achieve gravity separation. In such situation, it is necessary to have coalescence effect so that smaller droplets join together to make bigger droplets which can be separated easily.

A coalescer vessel is provided with a stainless steel fine wire mesh element. The wire mesh has an affinity for fine water droplets.

Coalescers are normally vendor supplied items. Therefore, their design aspects are not covered in this manual.

### 2.2.5. Three Phase V-L-L Separator

Typical three phase vapour-liquid-liquid separators are compressor interstage drums with liquid phases (hydrocarbon and water), reflux drums with two liquid phase with vapour product. Design of these vessels should take care of vapour-liquid separation as well as liquid-liquid separation. Detailed procedure for sizing two phase V-L separator and two phase L-L separator described earlier in Sections 2.2.1 and 2.2.2 is applicable to three phase V-L-L separators also. Major steps are repeated below:

1. Select type of vessel – horizontal or vertical (Refer Section 2.1.3).
2. Calculate critical velocity, maximum allowable vapour velocity and liquid hold up volume as described in Section 2.2.1 for vapour liquid separator. Here, average density of two liquid phases should be used.
3. For given allowable velocity and liquid hold-up, dimensions for vertical or horizontal vessel would be different.

#### 3.1 Vertical Vessel

3.1.1 Establish minimum diameter for vertical vessel as per procedure given in Section 2.2.1

3.1.2 Calculate hold-up of individual liquid phases and determine corresponding vessel height for both phases. This will establish normal interface elevation. Say height of heavy liquid phase and light liquid phase in $h_H$ and $h_L$ respectively.

3.1.3 Calculate settling velocity of heavy liquid droplets in light liquid phase and rising velocity of light liquid droplets in heavy liquid phase as per procedure given in Section 2.1.2 say velocity for heavy and light liquid phase is $v_H$ and $v_L$ respectively.

3.1.4 Calculate settling. Time of heavy liquid through light phase and rising time of light phase droplets through heavy phase as follows:

$T_S$ = Settling time for heavy phase

$T_R$ = Rising time for light phase

$T_S$ = Height of Light phase ($h_L$) /Settling velocity of heavy phase $v_H$

$T_R$ = Height of Heavy Phase ($h_H$)/Rising velocity of light phase ($v_L$)

3.1.5 Check hold-up time is higher than TS and TR. If not increase diameter and repeat steps 4.1 to 4.5 here it is to be noted that by changing holdup volume, criteria of step 4.5 cannot be met. Changing hold-up volume for a fixed diameter will change height

of both phases and hence TR and TS therefore, effort should be to keep same liquid hold-up but decrease $h_H$ and $h_L$ increasing diameter. However, minimum height of each phase should be 150 mm.

3.1.6 Establish total vessel dimensions using Table 2.5 and Figure 2.3.

### 3.2 Horizontal Vessel

In a horizontal V-L-L separator, vertical cross section of the drum is divided into three parts (i) Vapour phase (ii)light liquid phase and (iii) heavy liquid phase. Therefore, trial procedure becomes necessary and is outlined below:

3.2.1 Establish dimensions of horizontal vessel on the basis of V-L separation only as per procedure given in Section 2.2.1.

3.2.2 Calculate hold-up volume of both liquid phases or specific residence tome required, if any, for either phase for controlling process or for, start-up make-up etc. use Table 2.4 if necessary. Nomenclature for hold-up volume is

$V_l$ = Hold-up volume for light phase

$V_h$ = Hold-up volume for heavy phase

3.2.3 Take minimum droplet dia d= 0.009 cm for both phases. When separation of heavy phase from light phase is important, (light phase continuous) follow steps 4.4 through .416

3.2.4 Calculate settling velocity of heavy phase droplets in light phase using equations given in Section 2.1.2.

3.2.5 Use design settling velocity for heavy phase droplets as 25 cm/min if calculated value is higher than 25 cm/min.

3.2.6 $V_A/V_T$ = Hold-up volume of heavy phase ($m^3$)/Total volume of drum VT ($m^3$)

3.2.7 Read $h_h/D$ against $V_h/V_t$ from Figure 2.11 and hence calculate $h_h$ (height of heavy phase from LLL). Bottom 150 *mm* height is not considered for hold-up purpose. Thus, total heavy phase height = interphase level from bottom = $h_h$+150 *mm*.

3.2.8 $(V_h + V_l)/V_T$ = (Hold-up volume for heavy phase + light phase)/Total volume of drum VT ($m^3$)

3.2.9 Read $(h_h + hi)/D$ against $(Vh + Vi)/Vt$ from Figure 2.11 and calculate $h_1$.

3.2.10 The flow path of heavy phase droplet is height of light phase($h_1$).

Settling Time *Ts* (min) = (Height of light phase $h_1$ *cm*)/ (Design settling velocity, cm/min)

3.2.11 Flow path length of light phase will be less than the length of drum due to nozzle Arrangements. For estimating purpose, calculate flow path length as follows:

Flow path length (mm) = L – (di +do+300)

Where,

L = length of drum (mm)

$d_i$ = inlet nozzle o.d. (mm)

$d_o$ = light phase outlet nozzle 0.d. (mm)

3.2.12 $A_h$ (Area of cross section occupied by heavy phase): Read $A_h/A_T$ against h/D from Figure 2.12 to 2.14 and hence calculate $A_l$.

3.2.13 $A_l$ (Area of cross section occupied by light phase): Read $(A_h+Al)/A_T$ against $h_H + h_l /D$ from Figure 2.12 to 2.14 and hence calculate $A_l$.

3.2.14 Calculate longitudinal velocity of light phase liquid in mm/min units as follows :

$V_l$ = (Volumetric flow rate of light phase)/$A_l$

3.2.15 Residence time of light phase (min) is obtained by dividing the flow path length (*mm*) by the longitudinal velocity (*mm/min*).

3.2.16 Check that the residence of the light liquid phase is slightly higher than the settling time of heavy phase droplets. If not, a new diameter D must be assumed for the drum and the calculation repeated from steps 4.1 to 4.16.

When separation of light phase from heavy phase is of equal importance as the separation of heavy phase from light phase, rate of rise of light droplet through continuous heavy phase, rate of rise of light droplet through continuous heavy phase should be calculated.

3.2.17 Calculate rising velocity of light phase droplet in heavy continuous phase using equations given in section 2.1.2.

3.2.18 Use design velocity of light phase droplets as 25 cm/min in calculated value is higher.

3.2.19 The flow path of light phase droplet is height of heavy phase ($h_h$).

Rising Time $T_R$ (*min*) = Hight of heavy phase $h_h$ / Desing Rising Velocity (*cm/min*)

3.2.20 Flow path length for heavy phase fluid is calculated as follows :

Flow path length (*mm*) = L – (di + do + 300)

Where,

L = length of drum (mm)

$d_i$ = inlet nozzle o.d. (mm)

$d_o$ = heavy phase outlet nozzle o.d. (mm)

3.2.21 Calculate longitudinal velocity of heavy phase liquid in mm/min units as follows:

$V_l$ = (Volumetric flow rate of heavy phase)/ $A_h$

3.2.22 Residence time of heavy phase (*min*) is obtained by dividing the flow path length (*mm*) by the longitudinal velocity (*mm/min*)

3.2.23 Check that the residence time of heavy phase fluid is slightly higher than the rising time of light phase droplets. If not, a new diameter should be assumed and calculations should be repeated from steps 4.1 to 4.23.

### 2.2.6 Two Phase V-S Separators

The most commonly encountered two phase V-S separator in process industry is the decoking drum. A Decoking drum is provided for decoking of furnace tubes. Inlet stream to decoking drum may consist of steam, air and coke particles. Separation of coke particles from gases is essential to prevent coke particles being discharged to atmosphere or heater stack. These are vertical vessels and procedure for sizing is same as any other vessel except the points given below:

1. Calculate critical vapour velocity and maximum allowable vapour velocity using the following equations :

$$V_c = 0.107\left(\frac{S_s - S_v}{S_v}\right)^{0.5}$$

$$V_{\max} = 0.7V_c$$

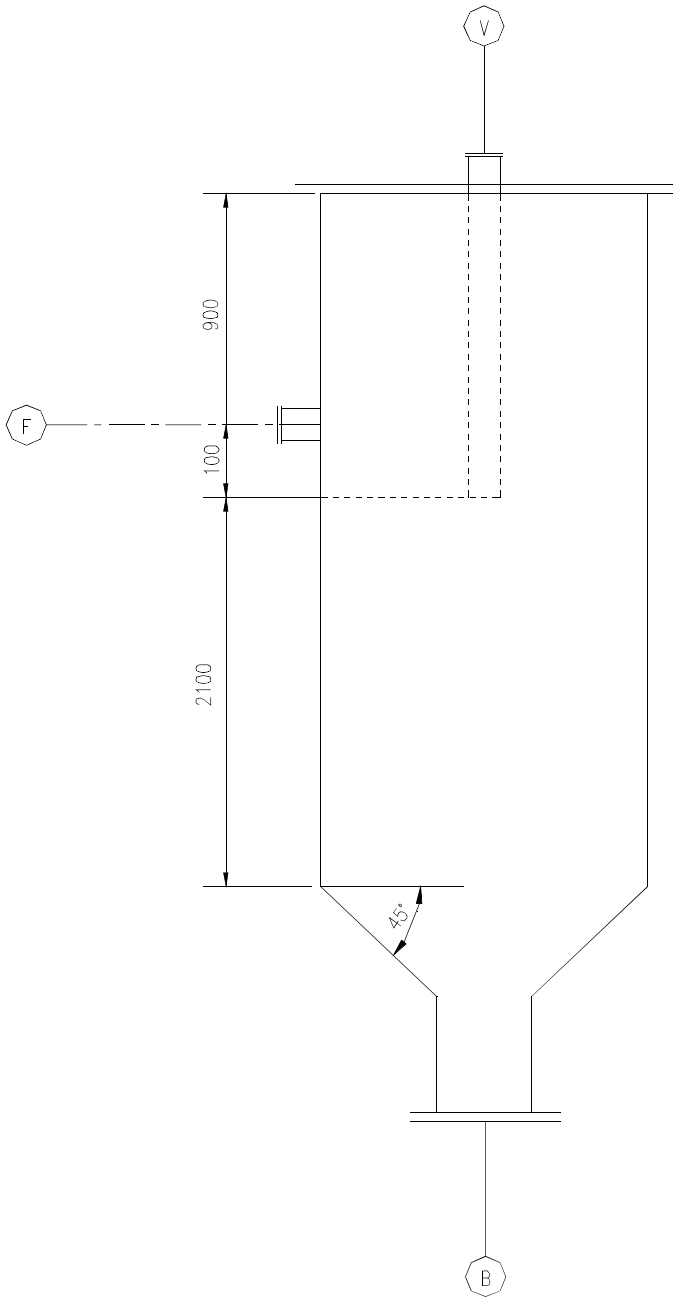
2. As solid particles settle at the bottom of the drum, provision is to be kept to remove them intermittently. To facilitate this, the bottom of the drum is made conical with bottom flange opening. Cone angle should preferably be based on angle of repose for solid particles under consideration. If it is not possible to determine angle of repose, a value of 45 deg may be used.

3. Typical configuration of decoking pot is given in Figure 2.5.

### 2.2.7 Hold-up Drums

There is no specific requirement for vapour space in hold-up drums. However, these should be designed for 80% liquid full (between HLL and LLL).

FIGURE 2.5: TYPICAL CONFIGURATION FOR DECOKING POT.



### 2.2.8 Nozzle Sizing for Process Vessels

- **Standard Nozzle Size**
  Select nozzle diameter equal to connecting pipe size. However, minimum nozzle should be equal to 1-1/2".

- **Miscellaneous Nozzles**
  Sizes of miscellaneous nozzles like instrument connections, vessel vent, drain, pumpout etc. should be as per Table 2.5

- **Minimum Distance of Nozzles from Tangent Line**
  Minimum distance of different sizes of nozzles from tangent line should be as per Table 2.6.

### 2.2.9 Special Design Considerations

- **Liquid Surge in Distillate Drums**

When the gas product discharges into a refinery gas system or other system where liquid carry over would constitute a potential hazard.

(i) Design the vapour disengaging space for surges of 150 – 200 % of normal design gas rate without exceeding the allowable vapour velocity.

(ii) Design the volume of the vapour space above high liquid level for not less than 10 minutes hold-up on liquid product rate.

- **Settling Drums**

Special requirements may make it necessary for deign a settling drum larger than size based on settling rates alone. Typical examples are:

(i) In recirculating caustic wash drum, the volume of the caustic layer is made at least equal to the caustic make up requirement for one shift. Thus, caustic make-up is required once a shift only.

(ii) For settlers handling emulsions, as in alkylation plants, time must be provided for breaking the emulsion in addition to that required for subsequent settling.

Table2.5: Standards for Nozzles in Process Vessels

| Service | Size | Type |
|---|---|---|
| Inlet/Outlet | Calculated size or line size, whichever is larger (2" minimum) | Flanged |
| Vent | Depends on drum volume as follows: | Flanged |

| | | | |
|---|---|---|---|
| | Drum Vol. | Nozzle Size | |
| | Up to 6 $m^3$ | 1-1/2” | |
| | More than 6 $m^3$ | 2” | |
| Drain | Depends on drum volume as follows: | | Flanged |
| | Drum Volume | Nozzle Size | |
| | Up to 6 $m^3$ | 1-1/2” | |
| | 6.1 to 15 m3 | 2” | |
| | More than 15 m3 | 3” | |
| Pumpout | Depends on drum volume as follows: | | Flanged |
| | Drum Volume | Nozzle Size | |
| | Up to 6 $m^3$ | 1-1/2” | |
| | 6.1 to 15 m3 | 2” | |
| | More than 15 m3 | 3” | |
| Safety Valve | Same as inlet nozzle size of valve (1” minimum) | | Flanged |
| Manhole | 18” or 20” | | Flanged |
| Blowdown | Depends on drum volume as follows: | | Flanged |
| | Drum Volume | Nozzle Size | |
| | Up to 1.5 $m^3$ | 2” | |
| | 1.5 to 15 m3 | 3” | |
| | More than 15 m3 | 4” | |
| Steamout | Depends on drum volume as follows: | | Flanged |
| | Drum Volume | Nozzle Size | |
| | Up to 15 $m^3$ | 1-1/2” | |
| | 1.5 to 15 m3 | 2” | |
| | More than 15 m3 | 3” | |

Notes:

1. No single nozzle cut out should be more than 50% of vessel diameter.

Table2.6: Minimum Distance of Nozzles from Tangent Line

| Nozzle Size (*inch*) | Minimum Distance Between Centre of the Nozzle Tangent line (*mm*) |
|---|---|
| 1-1/2 | 150 |
| 2 | 150 |
| 3 | 225 |
| 4 | 260 |
| 6 | 330 |

| | |
|---|---|
| 8 | 400 |
| 10 | 450 |
| 12 | 500 |
| 18 | 775 |
| 20 | 850 |
| 24 | 925 |

Notes:

1. Except for 1-1/2" and 2" sizes, minimum distance includes clearance for reinforcement pad.

- **Compressor Suction Knockout Drum**

In some cases, it is economical to combine the compressor knock-out service with another drum service, for example, the primary fractionator distillation drum in a catalytic cracking unit. In such cases, the emergency liquid surge requirements for compressor suction knockout are added to the other service requirements. Horizontal drums are common in this type of combination.

- **Steam Drums**

In addition to limiting steam velocity, it is a common practice to also limit steam release in steam drums to the following rates:

| Pressure kg/cm2 | Maximum Steam Rate (kg/hr)/cu.m. of steam space |
|---|---|
| 8.0 | 4600 |
| 11.5 | 5400 |
| 15.0 | 6200 |
| 18.5 | 6900 |
| 22.0 | 7300 |
| 26.0 | 7800 |

The steam space is considered to be the entire volume of the drum above high liquid level.

- **Flare Knockout Drums**

If there is any probability of liquid discharge into the flare system, a knockout drum should be provided immediately adjacent to each process unit or group of units. Facilities for automatically transferring any accumulated liquid to storage or other disposal are highly desirable. In addition to any such drum provided at the battery limit, a knockout pot is needed adjacent to the base of the stack to trap any condensed liquid. The flare header should be loped to drain to the knockout drum.

Special considerations for flare knockout drums are;

- **Droplet size**

The he flare can handle small size liquid droplets so the allowable vertical velocity in the rum may be based upon that necessary to separate droplets of 400 microns (0.04 cm) or greater. The flare knockout drums should be designed as per API-RP-521 "Guide for pressure relief and depressurising system".

**(i) Holdup Volume**

A liquid holdup time for 10 minutes is provided based on maximum condensation at design flaring load.

**(ii) Expansion Vessels**

Expansion vessels in heating oil circuit should be designed to have liquid holdup corresponding to:

~ Expansion of hot oil from ambient temperature to thermal operating temperature.

~ Total liquid inventory (estimated volume of equipment and piping), if drain/storage tank to take this volume is not available.

### 2.2.10 Design Considerations for Mist Eliminator

A mist eliminator is a device in which a gas stream having entrained liquid is passed around or through fixed elements interposed in the gas stream. The greater inertia of entrained liquid particles results in a tendency for the particles to deposit on this element rather than to follow the gas stream.

- **Types and selction of Mist Eliminator**

Several types of mist eliminators are used in process industries: These are:

(i) Knitted wire mesh type

(ii) Staggered baffles or channels

(iii)Perforated plates

Out of these, knitted wire mesh type separators are most popular in the industry. They have the following advantages over other types:

(a) High removal efficiency

(b) By using pads of different thickness, different types of entrainments encountered in process industries can be handled.

(c) Low initial and maintenance costs.

(d) Installation does not need special construction except for supports.

(e) Pressure drop is low.

(f) The separation efficiency is maintained over a wide range of operating velocity.

A disadvantage of knitted wire mesh is that if solid particles are present in the system or the fluid handled has fouling tendency, danger of plugging exists. In such cases, perforated plates or staggered baffles are used because they are less susceptible to plugging.

- **Construction of knitted wire mesh**

This type of mist eliminator consists of layers of knitted wire mesh made of fine wire, plastic or fibrous material. It is assembled in multiple layer sections which are usually sandwiched between light weight support grids to hold it in proper position. It may be made in serious for case of installation. The final product is self-contained integral unit with high free volume and large impingement area. The free voids are about 97% to 99%. Normally, 4" to 6" thick pads are used in process industries.

- **Material of Construction**

Wire mesh mist eliminators are manufactured in several materials for variety of process services.

- **Allowable vapour velocity**

Allowable vapour velocity for effective vapour-liquid separation with mist eliminator is higher than for vessels without mist eliminator. As a normal practice,

$$V_{allowable} = 1.7 * V_{without\ mist\ eliminator}$$

Specific data for a particular type of mist eliminator is normally supplied by vendors and same should be used for accurate calculations.

- **Pressure Drop in Mist Eliminator**

The pressure drop in mist eliminator is a function of the entrainment load, the mist eliminator design and gas velocity. Normally pressure drop for wire mesh type mist eliminator is less than 1" of water. Because of this small pressure drop, the wire mesh elements do not have to be "held down" and normally only wired to the support. A mist eliminator should be accessible inside the vessel.

## 2.3 Deign Specifications for Process Vessels

### 2.3.1 Design Pressure

(i) Vessels protected by safety relief valves shall be deigned for an internal pressure that exceeds maximum specified operating pressure by 10% or 2 $kg/cm^2$, which ever is more.

(ii) Vessels protected by safety relief valves shall be designed for minimum of 4.5 kg/cm2a pressure.

(iii) Vessels without pressure reliving devices shall be provided with an outlet which cannot be completely blocked off. The minimum outlet opening shall be sized so that the maximum pressure which can be developed in the vessel is not greater than the design pressure.

(iv) All vessels containing liquids with a vapour pressure at minimum ambient temperature lower than atmospheric pressure shall be deigned for full vacuum.

(v) Vessels open to atmosphere shall be deigned for pressure due to hydrostatic head only.

### 2.3.2 Design Temperature

(i) The design temperature for vessels shall be equal to the maximum operating temperature of fluid in the vessels plus 15 $^oC$ subject to a minimum of 65 $^oC$. For vessels operating at temperature below 29 $^oC$, the design temperature shall be the lowest operating temperature.

(ii) Vessels provided with steam flushing connection shall be designed for a minimum temperature equivalent to flushing steam operating temperature.

(iii) The vessels provided with drying air/gas connection shall be designed for operating temperature of drying air/gas plus 15 oC.

(iv) For pressure vessels storing liquefied hydrocarbons at ambient temperature, the lowest side design temperature (based on complete depressurisation) shall correspond to the coincident design pressure when this lowest design temperature shall be reached. This coincident design pressure can be different from the maximum design pressure specified for the vessels depending upon the system under consideration.

The design conditions (pressure and temperature) for all possible operations as specified above should be mentioned in the process data sheet.

### 2.3.3 Test Pressure

(i) Al pressure vessels shall be shop hydrostatically tested for 1.5 times of the design pressure.

(ii) Vessels that are open to atmosphere shall be tested by filling with water/liquid (whichever is heavier) to the top.

### 2.3.4 Corrosion Allowance

(i) The corrosion allowance for each vessel shall be determined by its intended service and shall be added to all pressure parts (except internal pressure piping) and non-removable internal parts on all surfaces exposed to the flowing medium. Parts of surfaces fabricated of, or surface protected with, corrosion resistant material shall not require corrosion allowance.

(ii) The minimum corrosion allowance for vessels not protected with corrosion resistant surface shall be 3.0 mm.

(iii) When the service conditions are such that it is not practicable to provide corrosion allowance by added base metal, a corrosion resistant lining with 2 mm minimum thickness shall be provided for the pressure parts.

### 2.3.5 Minimum Thickness

The minimum shell and head thickness exclusive of corrosion allowance for carbon and low ally steel vessels shall be 5.0 mm and for high alloy steel vessels shall be 2.5 mm. Standard plat thickness available are given in Table 2.2.

### 2.3.6 Special Provisions

**(i) Impingement Plate**

In general, if the velocity of the flow through a nozzle into a vessel exceeds 4.5 m/sec for liquid streams or 15 m/sec for vapour or vapour/liquid streams and can impinge directly on the shell wall, the vessel should be provided with an adequate removable impingement plate, depending on the diameter of the vessel.

**(ii) Baffles**

Internal baffles should be provided in the vessel for overflowing of light phase separated from heavy phase.

**(iii) Riser Nipple**

When there is a possibility of fouling or settling of fine particles, a riser nipple of 150 mm should be provided. However, in a drain nozzle no riser nipple should be provided.

**(iv) Drainage Time**

Whenever it is important to drain the vessel in fixed time, drain nozzle should be designed as per relevant standard.

## 2.4 Engineering Aspects of Process Vessel Design

### 2.4.1 Codes and Standards

All pressure vessels except those listed below shall be designed in accordance with Section VIII, Division I of the ASME boiler and pressure vessel code.

(i) Manufacturer's standard air receivers and pressure vessels furnished as part of proprietary and standardised equipment shall confirm to the applicable code requirements and the manufacturer's standards for the design conditions.

(ii) Steam generating equipment shall be designed in accordance with IBR Code.

(iii) Vessels designed for pressure above 3000 psig (210 kg/cm2g) or those of special (proprietary) design or construction shall confirm to applicable ASME Code requirements and to the manufacturer's proprietary designs and construction practices for such equipment.

### 2.4.2 Manholes, Handholes and Inspection Openings

(i) Vessels shall be provided with manholes and inspection openings in each pressure compartment. These openings shall be provided as follows:

| Vessel inside diameter | Required Opening |
|---|---|
| Less than 900 mm (36 inch) | Two handholes, 250 mm (8 inch) NPS or ½ the vessel I.D., whichever is less. |
| 900 mm (36 inch) and larger | 450 mm (18") NB manhole. |

(ii) Manholes shall be located as follows:

- For horizontal drums: in the heads unless the arrangement of access platforms makes this impracticable.
- For vertical vessels: In the shell.

(iii) Manhole opening in the vertical plane shall be furnished with handgrips inside the vessel.

### 2.4.3 Guidelines for Review of Detailed Engineering Documents on Process Vessels (Squad Check Procedure)

Following engineering documents are to be reviewed by the process department:

- Engineering drawing
- Nozzle orientation

It is very difficult to list out all the points which should be checked in vessel drawings. General guidelines are given below which should be followed with common sense:

(a) Check that elevation of vessel from ground level as per data sheet.

(b) Check that all instruments are conveniently located and approachable.

(c) Check that utility connection and sample points are conveniently located and approachable.

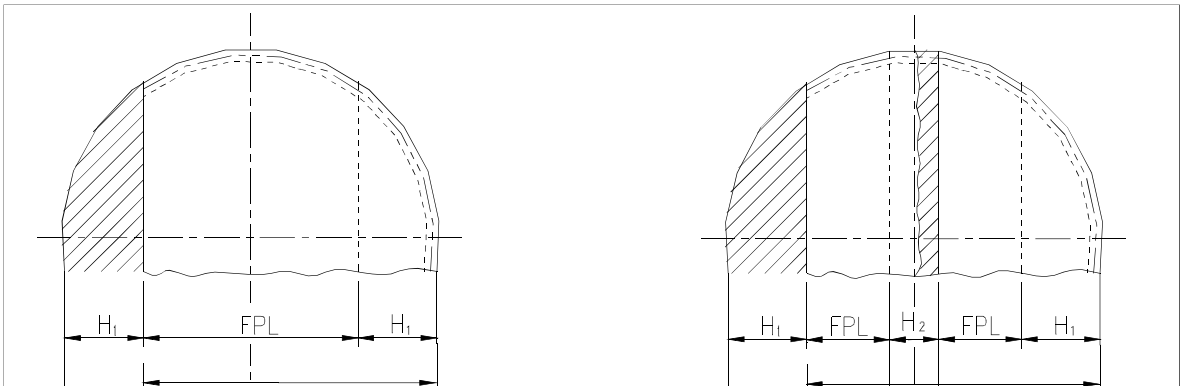
(d) Check that drain is provided at the lowest point and vent is provided at highest point.

(e) Check that all valves are conveniently located and approachable.

(f) Check level instrument nozzle elevation, especially for vessels with more than one phase.

(g) Check that instrument nozzles are away from feed nozzle where turbulence is expected.

Figure 2.11: Volume of segment for Horizontal Vessel



### 2.4.4 Sizing of Heating and Cooling Coils

The heat transfer across the coils can be characterized as an **Unsteady-State Batch** Operation. Hence the commonly used equation Q = U. A. LMTD is not valid.

The following steps are listed to help one perform the calculations.

#### (i) Basic Assumption

(a) U is constant for the process and cover the entire surface

(b) Specific heats are constant for the process

(c) Heat losses are negligible

(d) No partial phase changes occur

(e) Batch fluid temperature is uniform

#### (ii) mbols Used

| | |
|---|---|
| $T_1$ | Initial batch temperature |
| $T_2$ | Final batch temperature |
| $t_1$ | Initial medium temperature |
| $t_2$ | Final medium temperature |
| $\Delta T$ | Change in temperature of the batch liquid |
| $\Delta t$ | Change in temperature of the medium |
| M | Mass of the liquid in the batch |
| w | Mass flow rate of medium |
| c | Specific heat of medium |
| C | Specific heat of batch liquid |
| time | Time for the batch heating/cooling |
| Q' | Amount of heat transferred to/from the batch |
| Q | heat transferred per unit time |
| A | heat transfer area of the coil |
| U | Overall heat transfer coefficient |

#### (iii) Basic Steps

##### (a) Determination of heat load

The amount of heat to be added or removed form the batch is given by:

$$Q' = M.C.\Delta T$$

The amount of heat transferred per unit time is given by:

$$Q = Q'/time$$

**(b) Calculation of coil area using the following formulae (as applicable in different cases)**

1. COIL-IN-TANK, ISOTHERMAL HEATING MEDIUM

$$\ln\frac{t_1 - T_1}{t_1 - T_2} = \frac{U.A.time}{M.C}$$

2. COIL-IN-TANK, ISOTHERMAL COOLING MEDIUM

$$\ln\frac{T_1 - t_1}{T_2 - t_1} = \frac{U.A.time}{M.C}$$

3. COIL-IN-TANK, NON-ISOTHERMAL HEATING MEDIUM

$$\ln\frac{t_1 - T_1}{t_1 - T_2} = \frac{w.c\left(k_1 - \frac{1}{k_1}\right)}{M.C}$$

Where,

$$k_1 = e^{\frac{U.A}{w.c}}$$

4. COIL-IN-TANK, NON-ISOTHERMAL HEATING MEDIUM

$$\ln\frac{T_1 - t_1}{T_2 - t_1} = \frac{w.c\left(k_2 - \frac{1}{k_2}\right)}{M.C}$$

Where,

$$k_2 = e^{\frac{U.A}{w.c}}$$

**(c) Determination of U**

For the determination of U, for and agitated tank, several well-defined correlations are available in literature. However, for vessels without agitation there is a dearth of literature. The following values can be used in the calculations for the vessels without agitation:

| Case Description | U (Btu/hr.ft$^2$.$^o$F) | U (Kcal/hr.m$^2$.$^o$C) |
|---|---|---|
| 1. Steam condensing in coil and water boiling in copper vessel | 250 | 1220 |
| 2. Steam condensing in coil and water boiling in steel vessel | 175 | 854 |

| | | |
|---|---|---|
| 3. Water to water heating/cooling | 100 | 488 |
| 4. Aqueous solution heating | 75-80 | 366-390 |
| 5. Non-viscous hydrocarbons heating /cooling | 50 | 244 |
| 6. Medium organics heating/cooling | 10-20 | 48-96 |