

# Distillation Operations: Methods, Operational and Design Issues

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## Abstract:

*Distillation is one of the oldest and most common methods for the purification of liquids. It has been used for centuries to concentrate dilute alcoholic beverages and to obtain perfumes from fruits and flowers. In this paper various methods of distillation, operating principles, equipment for distillation are discussed in detail. Various design and operating policies for binary distillation columns are discussed with objective of maximizing production rate. Significant increasing in capacity can be realized by optimum design of column- i.e. number of trays, trays hold up, reflux drum hold up etc. and optimum operation- i.e. reflux ratio and start up procedure. Equipment and column size issues discussed here may be useful to practicing engineers in early stage of distillation design. Some issues in the design and operation of a batch and continuous distillation column are addressed.*

**Key words: distillation operations, methods, distillation equipment, operational and design Issues**

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## 1. Introduction

Distillation is by far the most predominant separation technique used in the chemical process industry (CPI). Distillation is a cascade of equilibrium stages at which the vapor mixture is in equilibrium with the liquid mixture. An in-depth understanding of vapor-liquid equilibrium (VLE) is therefore essential to the understanding and design of distillation processes.

## 2. Rules of Thumb for Feasible Distillation Operation

**a) Azeotropes:** If the mixture is azeotropic, then more advanced types of separation must be considered. [1]

**b) Solids:** If the material to be separated is high in solids, or contains tars or resins that could plug or foul a continuous unit, then a batch separation should be considered ([2]

### c) Optimum Pressure

(i) *Vacuum operation:* Use of a vacuum should be considered for heat sensitive compounds or polymerizable materials. Vacuum is usually not used unless required [2,3] e.g. a low bottom

temperature is needed to avoid thermal decomposition.

(ii) *Distillation column is above atmospheric pressure:* Column shell should be thicker to withstand pressure of the column.

(iii) If the column pressure required accomplishing overhead condensation with cooling water is less than 250 lb/in<sup>2</sup>, then the column pressure should give an average temperature driving force of 5-15°C in the overhead condenser.

(iv) If the column pressure required accomplishing overhead condensation with cooling water is greater than 250 lb/in<sup>2</sup>, then consider an alternative of using a refrigerant on the overhead and running the column at a lower pressure.

### d) Optimum Temperature Differences in Reboiler and Condensers

(i) Reboiler temperatures should be kept low enough to avoid bottoms degradation and/or fouling. (ii) Common temperature differences used for heat exchange across reboiler and condensers are shown in the following Table 1.

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Table 1 Common temperature differences for Heat exchange [2,3]

	Temp, K
<b>Condenser:</b>	
Refrigeration	3-10
Cooling Water	6-20
Pressurized	10-20
Fluid	20-40
Boiling Water	20-50
Air	20-50
<b>Reboiler:</b>	
Process Fluid	10-20
Steam	10-60
Hot Oil	20-60

### 3. Basic Distillation Equipment and Operation

#### 3.1 Main Components of Distillation Columns

Distillation columns are made up of several components, each of which is used either to transfer heat energy or enhance material transfer. A typical distillation unit contains several major components:

- A vertical shell where the separation of liquid components is carried out.
- Column internals such as trays/plates and/or pickings, which are used to enhance component separations.
- A reboiler to provide the necessary vaporization for the distillation process.
- A condenser to cool and condense the vapour leaving the top of the column.
- A reflux drum to hold the condensed vapour from the top of the column, so that liquid (reflux) can be recycled back to the column.

The vertical shell houses the column internals and together with the condenser and reboiler, constitute a distillation column. A schematic of a typical distillation unit with a single feed and two product streams is shown Fig. 1

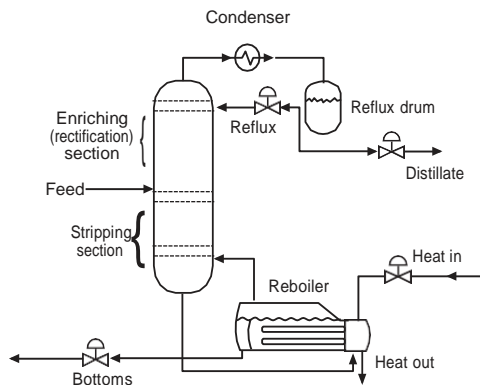


Fig. 1 A typical Distillation column

#### 3.2 Basic Operation and Terminology

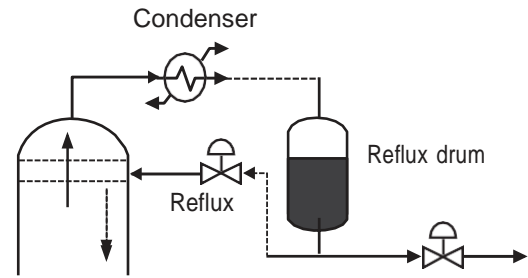


Fig. 2 Batch Distillation with Rectifying Column

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

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

The vapour moves up the column, and as it exits the top of the unit, a condenser cools it. The condensed liquid is stored in a holding vessel known as the reflux drum. Some of this liquid is recycled back to the top of the column and this is called the reflux. The condensed liquid that is removed from the system is known as the distillate or top product [2]

#### 3.3 Batch Distillation with rectifying column

Batch distillation is an unsteady state operation and carried out in a batch still to which a column equivalent to a number of equilibrium stages is attached. The Fig. 3 shows a typical set-up for batch distillation with rectifying column.

During distillation, the vapour passes upward through the column. The whole column is an enriching section. The vapour is condensed into liquid at the top of the column. Part of the liquid is returned to the column as reflux, and the remainder withdrawn as distillate. Nothing is added or withdrawn from the still until the run is completed. As distillation progress, the MVC in the batch still decreases.

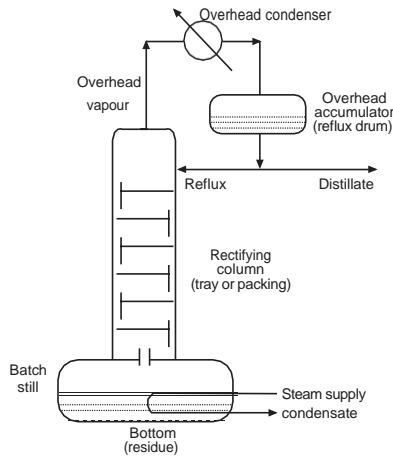


Fig. 3 Batch Distillation with Rectifying Column

In principle, the alternative setup as shown in Fig. 4 is also feasible. Here the vessel is located at the top of the column that is operated as a stripping column. Such a set-up is often known as the inverted batch distillation or the batch stripper. During operation, the LVC are primarily separated from the system. The liquid in the still depletes in LVC and enriches in MVC. [7]

In a batch process, the main steps are operated discontinuously. In contrast with a continuous process, a batch process does not deliver its product continuously but in discrete manner. In practice, most batch processes are made up of a series of batch and semi-continuous steps. A semi-continuous step runs continuously with periodic start-ups and shutdowns. [6]

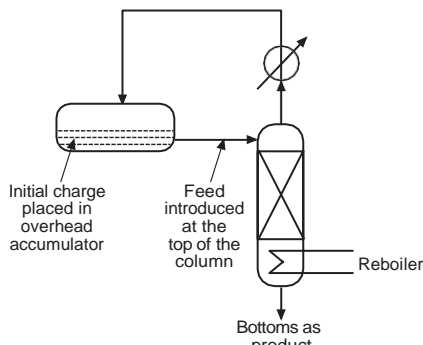


Fig. 4 Alternative Set-up batch Distillation with Rectifying Column

#### 4. Continuous Distillation Processes

The following types of distillation processes are briefly discussed here:

##### 4.1 Steam Distillation

Steam distillation refers to a process in which live steam is in direct contact with the distilling system in either batch or continuous operation. The basis of steam distillation rests on the fact that water forms immiscible mixtures with most organic

substances, and these mixtures will boil at a temperature below that of either water or the other materials. This is a very desirable feature, especially if the organic compound has a high boiling point at which it may be unstable or decompose. [10]

Steam distillation is commonly used in the following situations: [12]

- To separate relatively small amounts of volatile impurity from a large amount of material.
- To separate appreciable quantities of higher-boiling materials.
- Recover high-boiling materials from small amounts of impurity, which have a higher boiling point.
- Where the material to be distilled is thermally unstable or reacts with other components associated with it at the boiling temperature.
- Where the material cannot be distilled by indirect heating even under low pressure because of the high boiling temperature.
- Where direct-fired heaters cannot be used because of fire hazards.

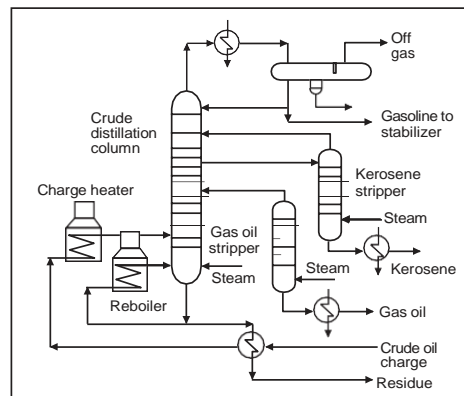


Fig.5 Typical Crude Distillation with Steam Strippers

Fig. 5 a typical crude distillation unit with steam strippers for its side draws products, namely kerosene and gas oil. Steam distillation is most commonly employed in petroleum refining operation, usually in combination with vacuum distillation.

##### 4.2 Vacuum Distillation

This method of distillation similar to conventional distillation except that it operates at very low (near vacuum) pressure. This method of distillation allows the mixture to boil at a lower temperature and thus avoids the thermal degradation problem. [3]

In addition, to handle the larger volume of vapors that result from distillation at reduced

pressure, a vacuum distillation column usually have fairly large diameter and tray spacing.

An example flow sheet for a crude distillation column (using steam distillation) and a vacuum distillation column is shown in Fig. 6.

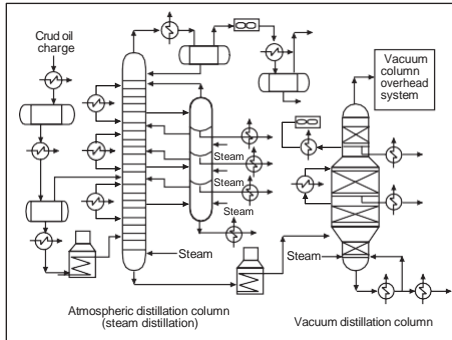


Fig. 6 Vacuum Distillation Column

### 4.3 Extractive Distillation

This process is very similar to azeotropic distillation. Extractive distillation refers to those processes in which a high-boiling solvent is added alter the relative volatilities of components in the feed. [5]

#### 4.3.1 Comparison between Azeotropic and Extractive Distillation:

Using the example of ethanol-water, benzene is the entrainer. Ethanol is removed as the bottoms product from the column. Benzene is too hazardous for various reasons ranging from workplace to product to environmental safety.

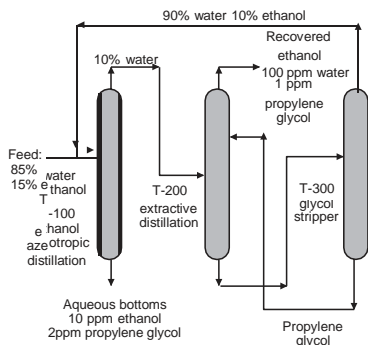


Fig. 7 Extractive Distillation

An alternative to recover ethanol is to use extractive distillation. The solvent used is Propylene Glycol. Recall also that ethanol forms a minimum-boiling azeotrope with water at approximately 89.4 mole% (96 wt %) ethanol. A process schematic for the process is shown in Fig.7

For this separation, propylene glycol meets all the requirements of an ideal extractive solvent

- It has a higher boiling point than water (187 deg C at 1 atm).
- It does not form an azeotrope with water.
- It has a molecular affinity for water (the hydroxyl -OH group forms a weak bond with water molecule).
- It is a relatively safe workplace material.

In the above system, the first column is the ordinary azeotropic distillation that produces an ethanol-water azeotrope as the distillate and nearly pure water as the bottoms. The distillate is fed to the second column for extractive distillation, where propylene glycol is added. Ethanol is produced as the distillate, leaving the top of the column.

This column can be conceptually divided into three sections. The middle section is the rectifying section where ethanol is purified by the removal of water. Bonding of the water molecules with glycol raises ethanol's relative volatility with respect to water, thus facilitating separation.

The top section reduces the concentration of propylene glycol in the ethanol distillate to negligible level. The bottom section strips ethanol from water. The bottoms from the second column are sent to the third column, a glycol stripper, where the glycol is recovered. The propylene glycol leaves the stripper as a bottoms product and is recycled back to the extraction column as the source of solvent. The overhead from the glycol stripper (containing mainly water and some ethanol) is sent back to the first column where it combined with fresh feed.

#### 4.3.2 Selection of a Solvent

The choice of solvent determines which of the two components in the original feed is removed predominantly in the distillation. For example, if the fresh feed to the distillation is a mixture of 83 mole% ethanol and 17 mole% water, and ethylene glycol (boiling point 197.35 C) is the solvent; the volatility of ethanol is increased more than that of water. Therefore, ethanol is removed as the distillate from the extractive distillation column, and water is separated in the solvent recovery column. [12]

The separation of toluene (boiling point 110.8°C) from isooctane (boiling point 99.3°C) is difficult using conventional distillation. Addition of phenol (boiling point 181.4°C) results in the formation of phenol-toluene mixture that leaves the extractive distillation column as bottoms, while relatively isooctane is recovered as overhead product. The phenol-toluene mixture is further separated in a second column (solvent recovery column) whereby toluene appears as

distillate and the bottoms product, phenol, is recycled back to the first column. Fig. 8 shows process flow sheet for separation of toluene and isooctane using phenol as the solvent.

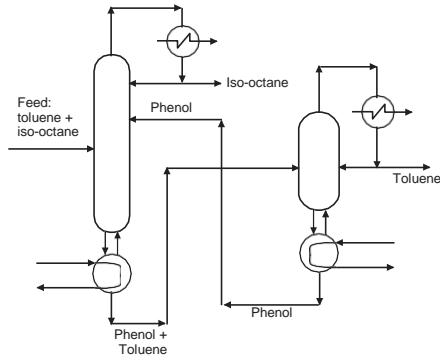


Fig. 8 Separation of Toluene and Isooctane using Phenol as the Solvent

#### 4.4 Reactive (Catalytic) Distillation

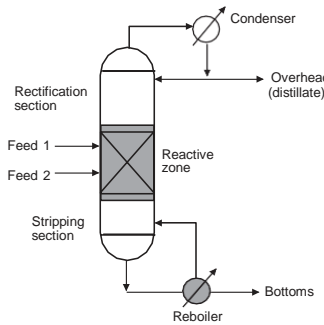


Fig. 9 Reactive Distillation Process

This technique embodies simultaneous reaction and separation. Catalysts are enclosed in special packing inside a distillation column. This method uses distillation to remove products as they are formed in order to shift the chemical equilibrium of a reaction towards 100% conversion shown in Fig. 9[4].

In reactive distillation (RD), one or more feeds can be used, and the reactive zone can be anywhere in the column and depending on the product, it may exit from either the overhead or bottom. Enhancement in conversion, selectivity, energy utilization, reduction in equipment cost and separation of otherwise difficult to – separate mixtures are some of the motives behind its several applications. [4, 8]]

Today, RD is used mainly for etherification and Esterifications reactions, but it has also been successfully applied to alkylation and nitration reactions. Most early applications involved liquid-phase reactions; and if a catalyst was utilized, it was in solution. In more recent applications, solid catalysts had been used. An example is the production of methyl-tert-butyl-ether (MTBE), an

octane booster for gasoline. The set-up is shown in Fig. 10.

The pre-reactor provides reaction equilibrium of over 90% MTBE, with the RD pushing the reaction another 5-10% to almost total conversion of the reactants. In the reactive zone, bales containing beads of catalyst resins are stacked on top each other. [8]

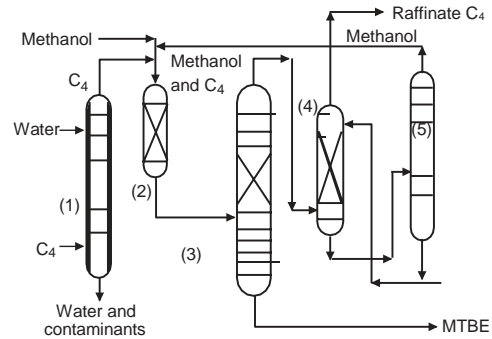


Fig.10 Production of MTBE by RD

The advantage of reactive distillation is the elimination of complicated product recovery and separation and recycling of unconverted reactants, all of which lead to savings in equipment and energy costs.

#### 4.5 Pressure-Swing Distillation

As mentioned earlier, some binary azeotrope mixtures lose azeotropic behavior when the system pressure is changed. In this case, separation can be achieved without using an additional entrainer. For example, in the case of ethanol-water mixture (azeotropic composition 89.4 mole% at 101.325 kPa), an azeotrope does not form below a pressure of approximately 9.2 kPa. [2, 12]

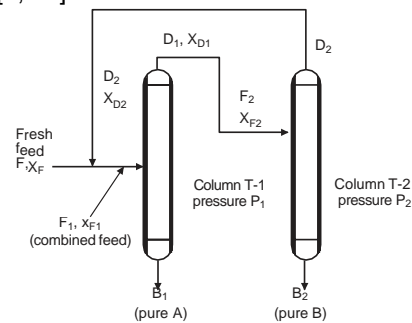


Fig. 11 Pressure-Swing Distillation Process

Pressure-swing distillation is a method for separating a pressure-sensitive azeotrope that utilizes two columns operated in sequence at two different pressures.

As the pressure is reduced to  $P_1$ , the azeotropic concentration moves toward a smaller

percentage of A ( $x_{P2}$  to  $x_{P1}$ ). The set-up for a pressure-swing distillation is shown in Fig. 11

The total feed  $F_1$  to Column  $T_1$  operating at the lower pressure  $P_1$  is the sum of the fresh feed  $F$  at mole fraction  $x_F$ , and the recycled distillate  $D_2$  at mole fraction  $x_{D2}$  (close to azeotropic mole fraction  $x_{P2}$ ). The mole fraction of the total feed  $F_1$  is  $x_{F1}$ , and it is richer in A than the azeotropic mole fraction of  $x_{P1}$  at pressure  $P_1$ . The bottoms leaving column  $T_1$  is almost pure A. The distillate leaving  $T_1$  is  $D_1$ , having the mole fraction  $x_{D1}$ . This mole fraction is richer in A than the azeotropic mole fraction of  $x_{P1}$ , but less rich in A than the azeotrope mole fraction of  $x_{P2}$ .

Distillate  $D_1$  is sent to Column  $T_2$  as feed  $F_2$ . Nearly pure B is obtained from bottom of  $T_2$ , and the distillate  $D_2$  is recycled to Column  $T_1$ . The change on the phase diagram is shown in Fig. 13

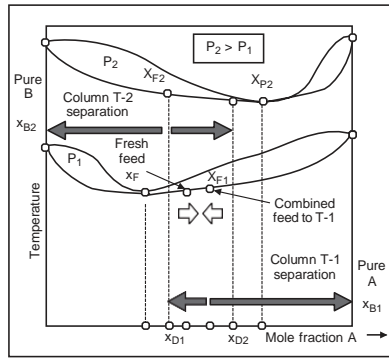


Fig.12 Changes on the Phase Diagram for Pressure-Swing Distillation

## 5. Equipment and Column Sizing

In order to have stable operation in a distillation column, the vapour and liquid flows must be managed. Requirements are: [12]

- Vapour should flow only through the open regions of the tray between the downcomers.
- Liquid should flow only through the downcomers.
- Liquid should not weep through tray perforations.
- Liquid should not be carried up the column entrained in the vapour.
- Vapour should not be carried down the column in the liquid.
- Vapour should not bubble up through the downcomers.

### 5.1 Tray Constructions and Hydraulics

Three main types of trays are: Bubble Cap Trays, Sieve Trays and Valve Trays. Typically, a weir on each tray governs the liquid flow between

trays. The flow depends on the length of the weir and how high the liquid level on the tray is above the weir. [11]

### 5.2 Tray Efficiency

Ideally, tray efficiencies are determined by measurements of the performance of actual trays separating the materials of interest; however, this is usually not practical in the early phases of a design. Estimates can be based on theory or on data collected from other columns.

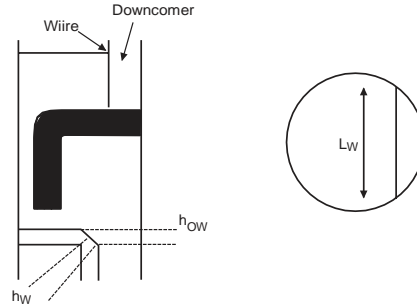


Fig. 13 Tray Efficiency for Distillation Column

The O'Connell correlation is based on data collected from actual columns. It is based on bubble cap trays and is conservative for sieve and valve trays. [9] It correlates the overall efficiency of the column with the product of the feed viscosity and the relative volatility of the key component in the mixture. These properties should be determined at the arithmetic mean of the column top and bottom temperatures. A fit of the data has been determined:

$$\eta_{oa} = 0.492(\mu_F a_{HK})^{-0.245} \quad [1]$$

This, or a similar data set, can be used to get preliminary estimates of efficiency numbers.

### 5.3 Column Diameter

Column diameter is determined based on the constraints imposed by *flooding*. Before beginning a diameter calculation, we want to know the vapour and liquid rates throughout the column. We then do a diameter calculation for each point where the loading might be an extreme: the top and bottom trays; above and below feeds, side draws, or heat addition or removal; and any other places where we suspect peak loads. [9]

Once we have calculated these diameters, we select one to use for the column, and then check it to make sure it will work.

### 5.4 Flooding

Downcomer flooding occurs when liquid backs up on a tray because the downcomer area is too small. This is not usually a problem. More worrisome is *entrainment flooding*, caused by too

much liquid being carried up the column by the vapour stream.

A number of correlations and techniques exist for calculating the *flooding velocity*; from this, the active area of the column is calculated, so that the actual velocity can be kept to no more than 80-85% of flood; values down to 60% are sometimes used.

A force balance can be made on droplets entrained by the vapour stream (which can lead to entrainment flooding). This balance yields an expression relating the vapour and liquid densities and a capacity factor (C, with velocity units) to the flooding velocity:

$$V_{\text{flood}} = C \sqrt{\frac{p_L - p_V}{p_V}} \quad [2]$$

### 5.5 Capacity Factors

The capacity factor can be determined from theory (it depends on droplet diameter, drag coefficient etc.), but is usually obtained from correlations based on experimental data from distillation tray tests. Depending on the correlation used, C may include the effects of surface tension, tendency to foam, and other parameters.

A common correlation proposed by Fair in the late 50s - early 60s for sieve trays is available in a wide range of sources. The correlation takes the form of a plot of a capacity factor (which must be corrected for surface tension) versus a functional group based on the liquid to vapour mass ratio:

$$F_{LV} = \frac{L}{V} \sqrt{\frac{p_V}{p_L}} \quad [3]$$

Enter the plot from the bottom with this

number, and then read the capacity factor from the left. This capacity factor applies to non-foaming systems and trays meeting certain hole and weir size restrictions. It will need to be corrected for surface tension:

$$C = \left( \frac{\sigma}{20} \right)^{0.2} \quad [4]$$

Where,  $\sigma$  the surface tension is in dynes/cm.

### 5.6 Diameter

Once we have the capacity factor, we can readily solve for the flooding velocity:

$$V_{\text{Flood}} = C_{\text{Fair}} \left( \frac{\sigma}{20} \right)^{0.2} \sqrt{\frac{p_L - p_V}{p_V}} \quad [5]$$

(This solution is for the Fair correlation, and adds the surface tension correction).

We know that flow = velocity  $\times$  area, so we can calculate the flow area from the known vapour flow rate and the desired velocity (a fraction of flood). This area needs to be increased to account

for the downcomer area, which is unavailable for mass transfer. The resulting tray area can be then used to calculate the column diameter. So, with everything lumped together, we have:

$$D_T = \sqrt{\frac{4V_{\text{Mass}}}{V_{\text{Flood}} \cdot F_{\text{Design}} \cdot \pi \pi p \left( 1 - \frac{A_d}{A} \right)}} \quad [6]$$

Trays probably aren't a good idea for columns less than about 1.5 ft in diameter (we can't work on them) - these are normally packed. Packing is less desirable for large diameter columns (over about 5 ft in diameter).

### 5.7 Pressure Drop

This gradient is normally expressed in terms of a pressure drop per tray, usually on the order of 0.10 psi. Detailed calculations are possible, but these depend so much on the actual tray specifications that final values are usually obtained from experts, but approximate methods can be used to get values to put in your design basis.

There are two main components to the pressure drop: the "dry tray" drop caused by restrictions to vapour flow imposed by the holes and slots in the trays and the head of the liquid that the vapour must flow through.

$$H_{\text{tray}} = h_d + h_l \quad [7]$$

### 5.8 Dry Tray Losses

The dry tray head loss can be related to an orifice flow equation:

$$h_d = \left( \frac{0.186}{C_0} \right)^2 \left( \frac{p_V}{p_L} \right) \cdot V_{\text{hole}}^2 \quad [8]$$

This equation determines the dry tray drop in inches of fluid. The constant 0.186 takes care of the units and is appropriate for sieve trays. The orifice size coefficient  $C_0$  depends on the tray configuration and will usually fall between 0.65 and 0.85. The hole velocity can be obtained by dividing the vapour flow rate by the total hole area of the tray.

### 5.9 Liquid Losses

The liquid head pressure drop includes the effects of surface tension and of the frothing on the tray. It is typically represented as the product of an aeration factor and the height of liquid on the tray:

$$h_l = \beta(h_w + h_{ow}) \quad [9]$$

Correlations are available for the aeration factor ( $\beta$ ); a value of 0.6 is good for a wide variety of situations. The height of liquid on the tray is the sum of the weir height and the height of liquid over the weir. The total height can be

calculated directly from the volume of liquid on the tray and its active area. Another approach is to back the height out of a version of the Francis weir equation (which relates flow-off a tray to liquid height and weir length). One version, for a straight weir, in units of inches and gal/min is:

$$h_{ow} = 0.48 \left( \frac{L}{\text{weir}} \right)^{2/3} \quad [10]$$

Note that these equations depend on the size and shape of the weir.

### 5.10 Column Height:

The height of a trayed column is calculated by multiplying the number of (actual) stages by the tray separation. Tray spacing can be determined as a cost optimum, but is usually set by mechanical factors. The most common tray spacing is 24 inches – it allows enough space to work on the trays whenever the column is big enough around (> 5 ft diameter) that workers must crawl inside. Smaller diameter columns may be able to get by with 18 inch tray spacing. [1,2,10].

In addition to the space occupied by the trays, height is needed at the top and bottom of the column. Space at the top - typically an additional 5 to 10 ft - is needed to allow for disengaging space. The bottom of the tower must be tall enough to serve as a liquid reservoir.

## 6. Factors Affecting Distillation Column Operation

The performance of a distillation column is determined by many factors as given below: [10]

- Feed conditions.
- State of feed.
- Composition of feed.
- Trace elements that can severely affect the VLE of liquid mixtures.
- Internal liquid and fluid flow conditions.
- State of trays (packings).
- Weather conditions.

### 6.1 Feed Conditions

The state of the feed mixture and feed composition affects the operating lines and hence the number of stages required for separation. It also affects the location of feed tray. During operation, if the deviations from design specifications are excessive, then the column may no longer be able handle the separation task.

### 6.2 Reflux Conditions:

As the reflux ratio is increased, the gradient of operating line for the rectification section moves towards a maximum value of 1. Minimum trays are required under total reflux conditions, i.e. there is no withdrawal of distillate. Most columns are designed to operate between 1.2 to 1.5 times the

minimum reflux ratios because this is approximately the region of minimum operating costs (more reflux means higher reboiler duty).

### 6.3 Vapour Flow Conditions

Adverse vapour flow conditions can cause: Foaming, Entrainment, weeping/dumping and Flooding.

#### 6.4.1 Foaming

Foaming refers to the expansion of liquid due to passage of vapour or gas. Although it provides high interfacial liquid-vapour contact, excessive foaming often leads to liquid buildup on trays. In some cases, foaming may be so bad that the foam mixes with liquid on the tray above.

#### 6.4.2 Entrainment

Entrainment refers to the liquid carried by vapour up to the tray above and is again caused by high vapour flow rates. It is detrimental because tray efficiency is reduced: lower volatile material is carried to a plate holding liquid of higher volatility. It could also contaminate high purity distillate. Excessive entrainment can lead to flooding.

#### 6.4.3 Weeping/Dumping:

This phenomenon is caused by low vapour flow. The pressure exerted by the vapour is insufficient to hold up the liquid on the tray. Therefore, liquid starts to leak through perforations. Excessive weeping will lead to dumping. A sharp pressure drops in the column and reduced separation efficiency indicates weeping.

#### 6.4.4 Flooding

Flooding is due to excessive vapour flow, causing liquid to be entrained in the vapour up the column. The increased pressure from excessive vapour also backs up the liquid in the Downcomer, causing an increase in liquid holdup on the plate above.

### 6.4 Column Diameter

Most of the above factors that affect column operation are due to vapour flow conditions: either excessive or too low. [5,9,11] Vapour flow velocity is dependent on column diameter. Weeping determines the minimum vapour flow required while flooding determines the maximum vapour flow allowed, hence column capacity.

### 6.5 State of Trays and Packings

The actual number of trays required for a particular separation duty is determined by the efficiency of the plate, and the packings if packings are used. [11] Thus, any factors that cause a decrease in tray efficiency will also change the performance of the column. Fouling,



wear and tear and corrosion, and the rates at which these occur depends on the properties of the liquids being processed affect tray efficiencies.

### 6.6 Weather Conditions

Most distillation columns are open to the atmosphere. Although many of the columns are insulated, changing weather conditions can still affect column operation. Thus the reboiler must be appropriately sized to ensure that enough vapour can be generated during cold and windy spells and that it can be turned down sufficiently during hot seasons.

#### Conclusion:

In this paper, various methods of distillation along with their operating principles are discussed in detail. The distillation equipments and operation are presented with industrial case studies. Some issues in the design and operation of a batch and continuous distillation column are addressed. Equipment and column size issues presented here may be useful to practicing engineers in early stage of distillation column design. Finally various factors affecting on industrial scale distillation operations are discussed with some guidelines.

#### Acknowledgement:

We are thankful to MAEER's MIT, Pune for providing necessary library facilities and Prof. Michel D'Souza for proof reading of the manuscript.

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