

Distillation

To reduce load times, this material is divided into seven files, corresponding to the numbered points below. The present file (distill7.html) contains point 7 only.

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

- vapor 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 vapor
- vapor should not be carried down the column in the liquid
- vapor should not bubble up through the downcomers

These requirements can be met if the column is properly sized and the tray layouts correctly determined.

Tray layout and column internal design is quite specialized, so final designs are usually done by specialists; however, it is common for preliminary designs to be done by ordinarily superhuman process engineers. These notes are intended to give you an overview of how this can be done, so that it won't be a complete mystery when you have to do it for your design project.

Basically in order to get a preliminary sizing for you column, you need to obtain values for

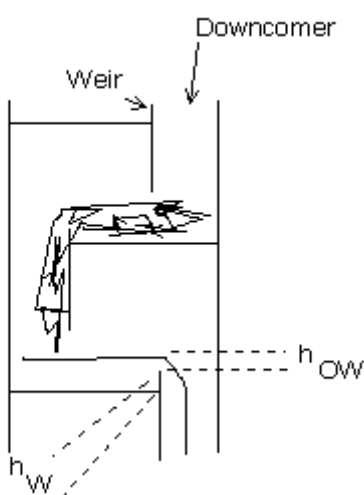
- the tray efficiency
- the column diameter
- the pressure drop
- the column height

Tray Construction & Hydraulics

Three main types of trays are to be discussed:

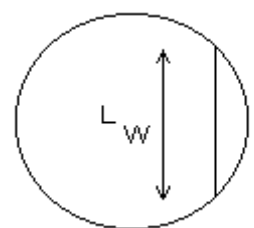
- Bubble Cap Trays
- Sieve Trays
- Valve Trays

Typically, the liquid flow between trays is governed by a weir on each tray. The flow depends on the length of the weir and how high the liquid level on the tray is above the weir. The Francis weir equation is one example of how the flow off a tray may be modeled.



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. Consequently, some form of estimation is required. Estimates can be based on theory or on data collected from other columns.



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. 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:

$$\eta_{oa} = 0.492(\mu_F \alpha_{HK})^{-0.245}$$

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

Column Diameter

Column diameter is found based on the constraints imposed by *flooding*. The number of ideal stages isn't needed to find the diameter -- only the vapor and liquid loads. You do need the number of actual stages to get the column height.

Before beginning a diameter calculation, you want to know the vapor and liquid rates throughout the column. You then do a diameter calculation for each point where the loading might be an extreme: the top and bottom trays; above and below feeds, sidedraws, or heat addition or removal; and any other places where you suspect peak loads.

Once you've calculated these diameters, you select one to use for the column, then check it to make sure it will work. Some columns will have two sections with different diameters -- consider this possibility if you end up with regions where the estimated diameter varies by 20% or more, but realize it will be more expensive than a column that is the same all the way up.

One issue that ought to be considered is the validity of your design numbers. If you are following the "traditional" approach, you've probably designed your column for reflux rates in the range of 1.1 to 1.2 times the minimum. This may not give you a column that can handle "upsets" well, so you may want to design for a capacity slightly greater than that -- increasing the flows by about 20% might be wise.

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 vapor 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 vapor stream (which can lead to entrainment flooding). This balance yields an expression relating the vapor and liquid densities and a capacity factor (C, with velocity units) to the flooding velocity:

$$v_{flood} = C \sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

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 is one proposed by Fair in the late 50s - early 60s. The version for sieve trays is available in a wide range of sources (including Figure 21.28 of MSH). The correlation takes the form of a plot of a capacity factor (which must be corrected for surface tension) vs. a functional group based on the liquid to vapor mass ratio:

$$F_{LV} = \frac{L_{mass}}{V_{mass}} \sqrt{\frac{\rho_V}{\rho_L}}$$

Enter the plot from the bottom with this number, and then read the capacity factor from the left. This capacity factor applies to nonfoaming systems and trays meeting certain hole and weir size restrictions. It will need to

be corrected for surface tension:

$$C = C_{Fair} \left(\frac{\sigma}{20} \right)^{0.2}$$

where the surface tension is in dynes/cm.

Other correlations for the capacity factor are also available. Several are based on more recent information, and may well be more accurate than the Fair plot; however, they also tend to be less broadly known and often require more *a priori* information on the system. You should use a correlation that is acceptable for your problem.

Diameter

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

$$v_{flood} = C_{Fair} \left(\frac{\sigma}{20} \right)^{0.20} \sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

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

We know that flow=velocity*area, so we can calculate the flow area from the known vapor 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 then be used to calculate the column diameter. So, with everything lumped together, we have:

$$D_T = \sqrt{\frac{4V_{mass}}{v_{flood} f_{design} \pi \rho_V \left(1 - A_d/A \right)}}$$

The only "new" term is the ratio of downcomer area to tray area. This should probably never be less than 0.1, and probably seldom will be greater than 0.2.

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

Pressure Drop

There is a pressure gradient through the column -- otherwise the vapor wouldn't flow. This gradient is normally expressed in terms of a pressure drop per tray, usually on the order of 0.10 psi.

The best source of pressure drop information is to measure the actual drop between trays, but this isn't always feasible at the beginning of a design. 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 vapor flow imposed by the holes and slots in the trays and the head of the liquid that the vapor must flow through.

$$h_{tray} = h_d + h_l$$

Dry Tray Losses

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

$$h_d = \left(\frac{0.186}{C_o^2} \right) \left(\frac{\rho_v}{\rho_l} \right) v_{hole}^2$$

This equation determines the dry tray drop in inches of fluid (your text has a similar equation in SI units). The constant 0.186 takes care of the units and is appropriate for sieve trays. The orifice size coefficient C_o depends on the tray configuration and will usually fall between 0.65 and 0.85. The hole velocity can be obtained by dividing the vapor flow rate by the total hole area of the tray.

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:

$$h_l = \beta(h_w + h_{ow})$$

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:

$$h_{ow} = 0.48 \left(\frac{L}{l_{weir}} \right)^{2/3}$$

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

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 spacings.

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. Depending on your boss's feelings about keeping inventory in the column, you will probably design the base for about 5 minutes of holdup, so that the total material entering the base can be contained for at least 5 minutes before reaching the bottom tray.

The total of height added to the top and bottom will usually amount to about 15% or so added to that required by the trays.

You rarely will see a real tower that is more than about 175 ft. tall. Tall, skinny towers are not a good idea, so watch the height/diameter ratio. You generally want to keep it less than 20 or 30. If your tower ends up exceeding these values, you probably want to look at a redesign, maybe by reducing the tray spacing, or splitting the tower into two parts.

[Quattro Pro 6.0 Example -- Distillation Sizing](#)

Additional References:

Primary references

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 5. Seader, J.D. and Ernest J. Henley, *Separation Process Principles*, John Wiley, 1998, pp. 305-312.
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R.M. Price

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