Design of Distillation Column

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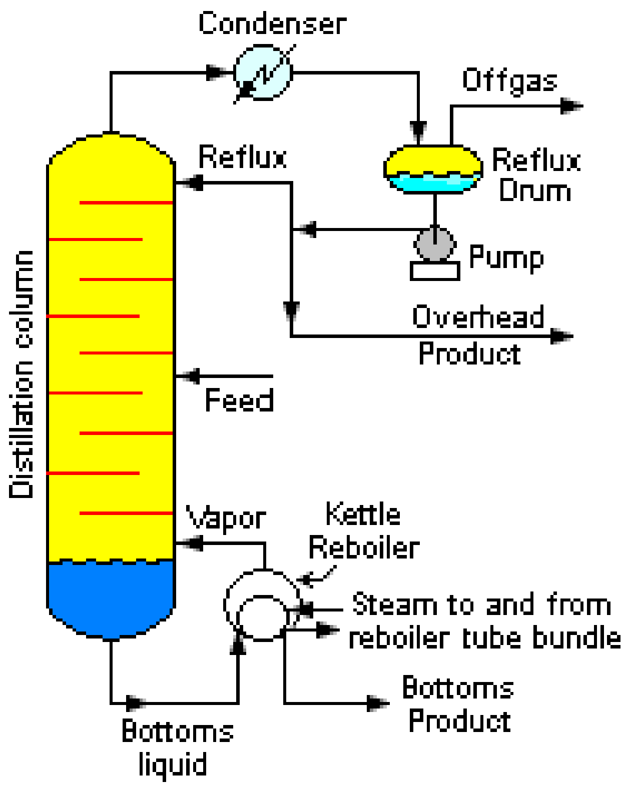
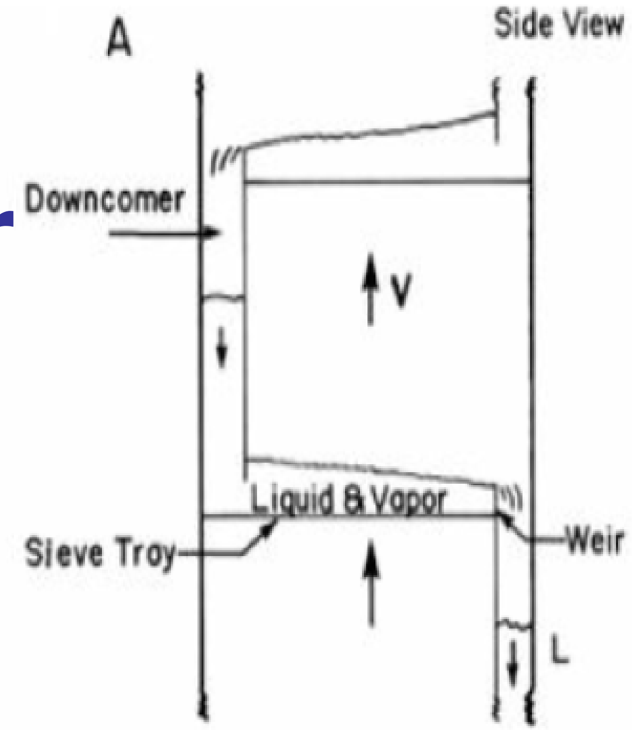
Distillation Equipment

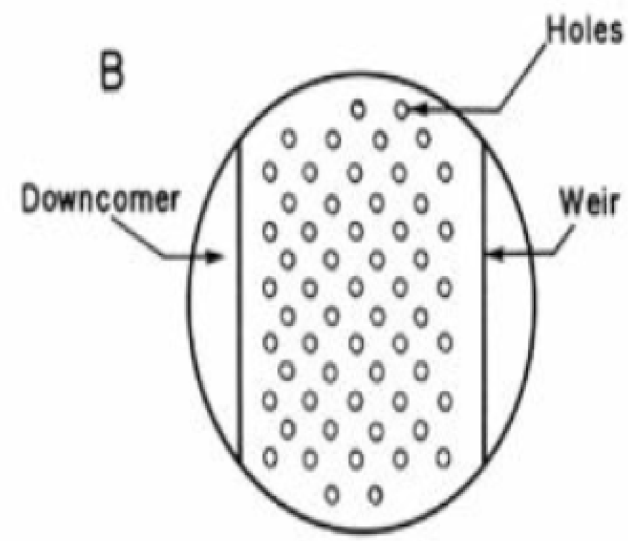
- Columns are built in metal and have circular cross- sections
- Trays (stages) are built so that liquid-vapor contact occurs
- Sieve trays are sheets of metal with holes punched into them to allow vapor to pass through



Downcomer and Weir

- Liquid flow down from tray above in a downcomer
- Liquid is contacted with vapor as it flows across the sieve tray
- Rising vapor prevents liquid from dripping downward
- Metal weir allows a sufficient liquid level on each tray
- The metal weir acts as a dam to keep a sufficient level of liquid on the plate (tray).





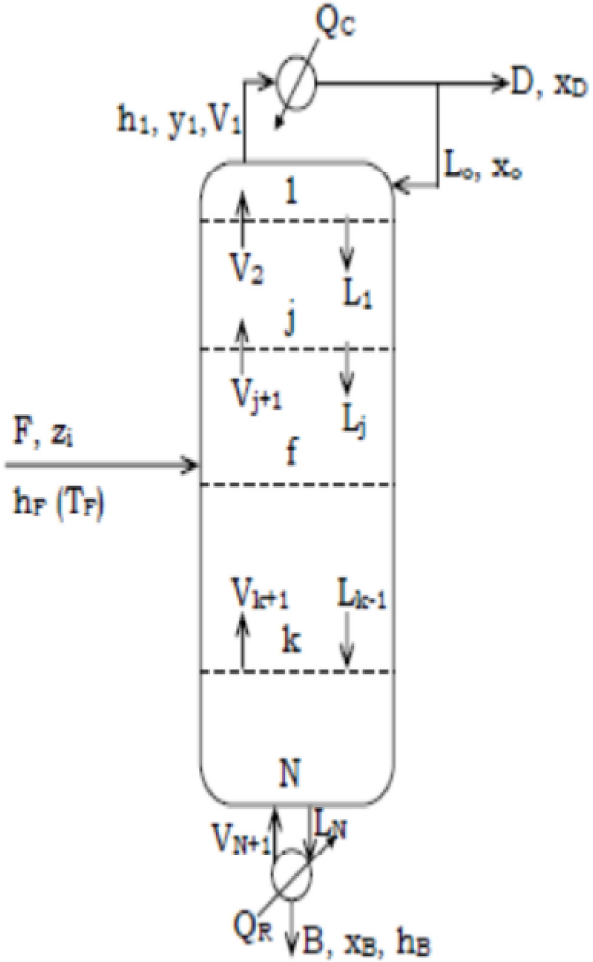
External Column Balances

Overall mass balance:

$$F = B + D$$

$$D = \frac{Z - X_B}{X_D - X_B} F$$

Mass balance on the more volatile component:



$$Fz = BX_B + DX_D$$

$$B = \frac{X_D - Z}{X_D - X_B} F$$

$$D = \left(\frac{Z - X_B}{X_D - X_B} \right) F$$

$$B = \left(\frac{X_D - Z}{X_D - X_B} \right) F$$

External Column Energy Balances

$$Fh_F + Q_R = Dh_D + Bh_B + Q_C$$

$$V_1 H_1 = V_1 h_D + Q_C \Rightarrow Q_C = V_1 \left(H_1 - h_D \right) = \lambda V_1$$

$$Q_C = \left(1 + \frac{L_0}{D} \right) \left(\frac{Z - X_B}{X_D - X_B} \right) F\lambda$$

$$D = \left(\frac{Z - X_B}{X_D - X_B} \right) F \qquad B = \left(\frac{X_D - Z}{X_D - X_B} \right) F$$

$$Q_R = Dh_D + Bh_B - Fh_F + \left(1 + \frac{L_0}{D} \right) D\lambda$$

$$V_1 = (D + L_0) = (1 + L_0 / D) D$$

$$Q_R = \left(\frac{Z - X_B}{X_D - X_B} \right) Fh_D + \left(\frac{X_D - Z}{X_D - X_B} \right) Fh_B - Fh_F + \left(1 + \frac{L_0}{D} \right) \left(\frac{Z - X_B}{X_D - X_B} \right) F\lambda$$

In Condenser: energy is lost from process stream to utility stream
In Reboiler: energy is supplied from utility stream to process stream

Points to Consider

- Enthalpies (h_i) is a state function that obeys Gibbs phase rule [for binary system (C=2), one phase (P=1), then 3 degrees of freedom (F = 3): T, P, and composition].

  $$F=C-P+2$$
- Stream V_1 enters the condenser and experiences ONLY a phase change => composition remains unchanged ($y_1=x_D=x_o$)
- D and L_o are both liquids (total condenser) with same composition => their enthalpies are also the same ($h_D = h_o$)

Stagewise Balances and Relations

If partial condenser is used,
it acts as an equilibrium stage

- Overall balance: $V_{j+1} = L_j + D$

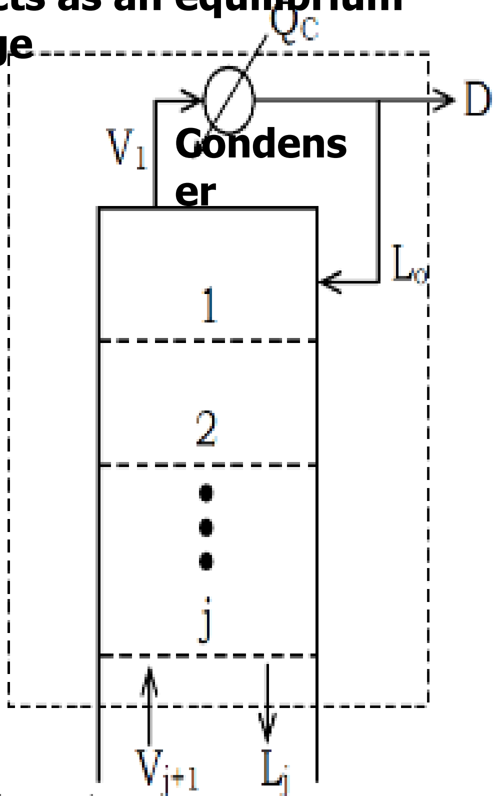
- Species balance (for *more volatile* component): $y_{j+1}V_{j+1} = x_j L_j + x_D D$

- Energy balance:

$$H_{j+1}V_{j+1} + Q_C = h_j L_j + h_D D$$

- Relations

$$h_j = h_j(x_j);\ H_{j+1} = H_{j+1}(y_{j+1});\quad x_j = x_j(y_j)$$

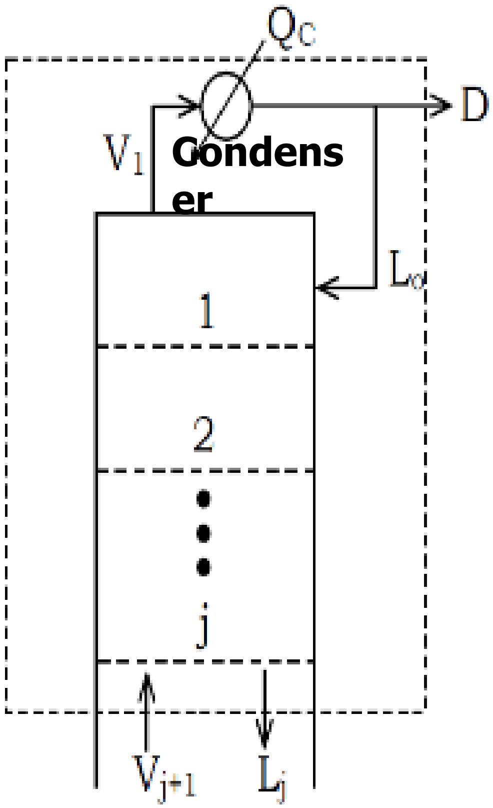


Rectifying Section:
Stage J Solution

$$V_{j+1} = L_j + D; \qquad y_{j+1}V_{j+1} = x_j L_j + x_D D$$

$$H_{j+1}V_{j+1} + Q_C = h_j L_j + h_D D$$

If partial condenser is used, it acts as an equilibrium stage



There are:

6 unknowns (L_j, V_{j+1}, x_j, y_{j+1}, H_{j+1}, h_j)

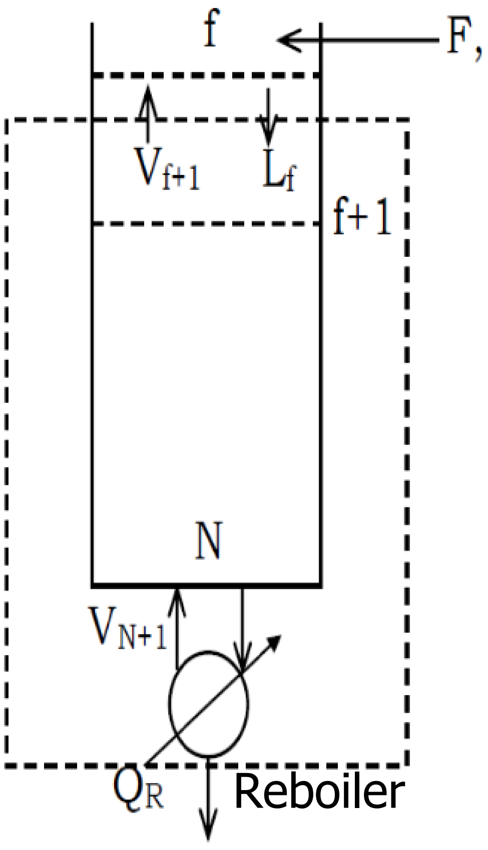
To find the unknowns, we need:

6 equations

Solve the equations to determine the unknowns

Stripping Section:
Stage f+1 Balances and Relations

- **Feed stage is f; the stage just below the feed stage is f+1**



A partial reboiler acts as an equilibrium stage

Overall balance: $\bar{V}_{f+1} = \bar{L}_f - B$

Species balance (for *more volatile* component):

$$y_{f+1}\bar{V}_{f+1} = x_f \bar{L}_f - x_B B$$

Energy balance:

$$h_f \bar{L}_f + Q_R = y_{f+1}\bar{V}_{f+1} + h_B B$$

Relations: $h_f = h_f(x_f);\; H_{f+1} = H_{f+1}(y_{f+1});\quad x_f = x_f(y_f)$

Stripping Section:
Stage f+1 Solution

There are:

4 knowns

x_B specified

P specified

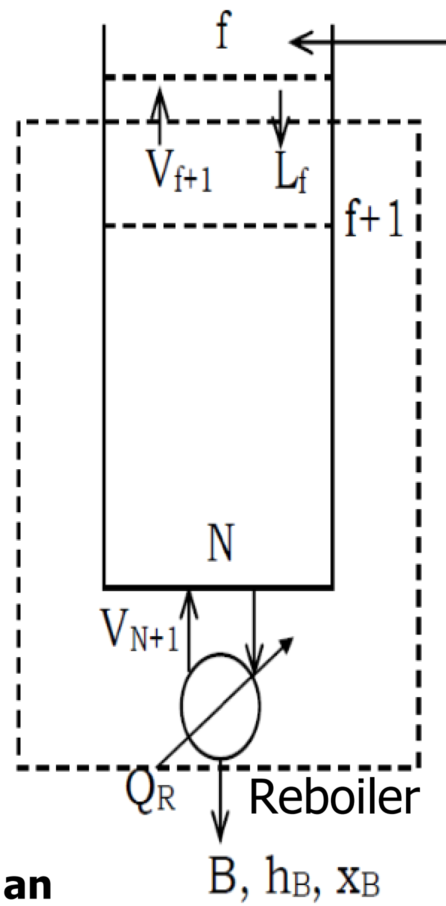
B calculated from external mass balance

Q_R calculated from external energy balance

6 unknowns (L_f, V_{f+1}, x_f, y_{f+1}, H_{f+1}, h_f)

Need: 6 equations

Solve the equations to determine the unknowns



A partial reboiler acts as an equilibrium stage

Tray Efficiency

- Overall column efficiency: $E_O = \dfrac{N_{equil}}{N_{actual}}$

- Murphree vapor efficiency:

$$E_{MV} = \frac{y_{out} - y_{in}}{y^*_{out} - y_{in}} = \frac{\text{actual change in vapor}}{\text{change in vapor at equilibrium}}$$

- Murphree liquid efficiency:

$$E_{MV} = \frac{x_{out} - x_{in}}{x^*_{out} - x_{in}} = \frac{\text{actual change in liquid}}{\text{change in liquid at equilibrium}}$$

Tray efficiency

Relationship between Murfree and Overall Efficiency

$$E_O = \frac{\log\left[1 + E_{MV}\left(\frac{mV}{L} - 1\right)\right]}{\log\left(\frac{mV}{L}\right)}$$

$$y^*_{out} = mx_{out} + b$$

Tray Efficiency Versus Vapor Flow rate

High vapor flow rates

↓

Excessive entrainment

↓

Efficiency decreases

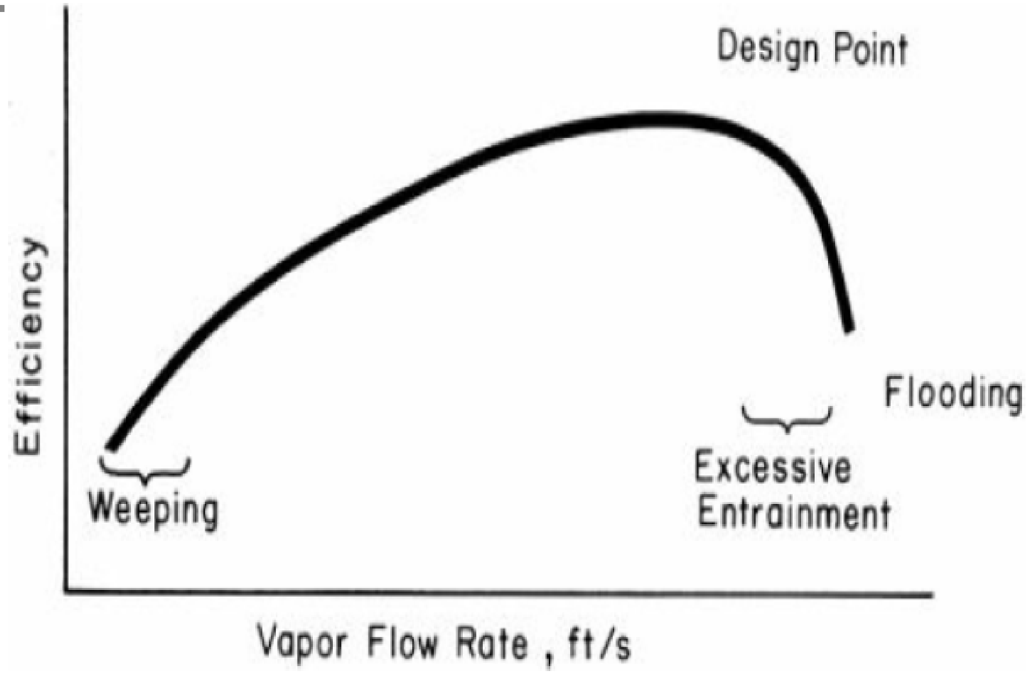
Low vapor flow rates

↓

Weeping

↓

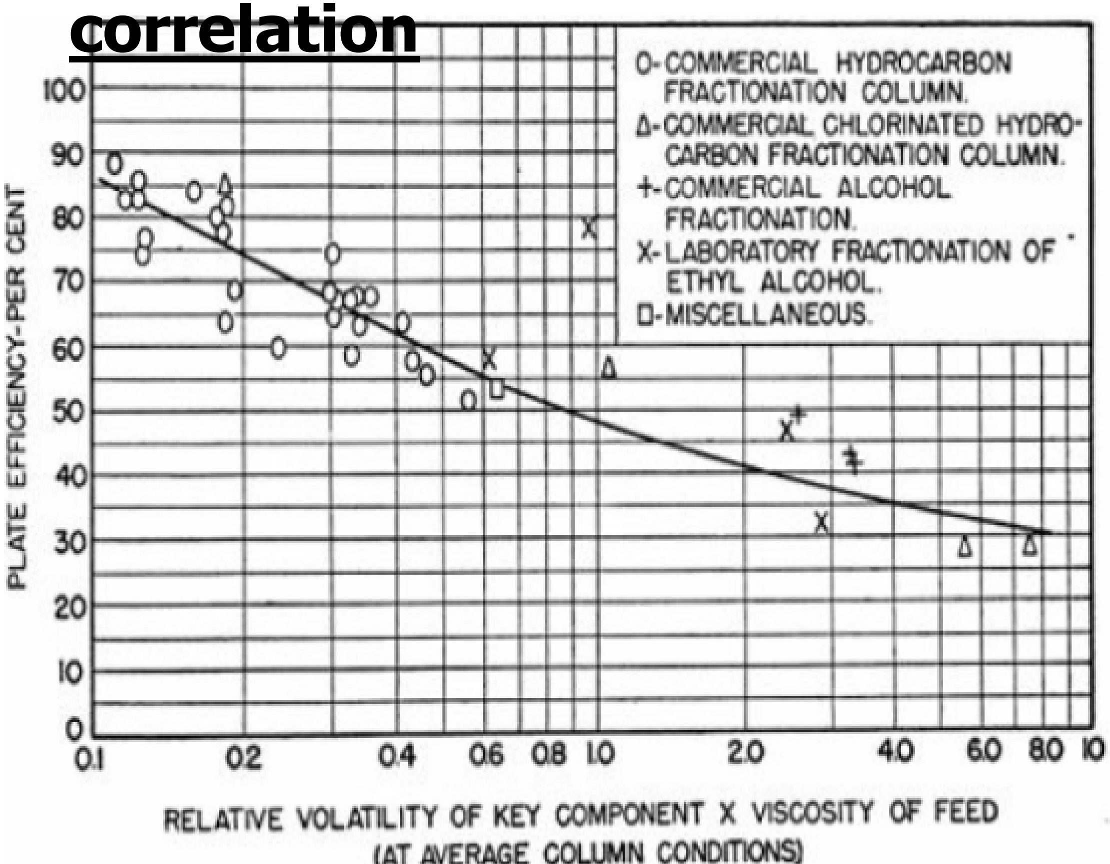
Efficiency decreases



- As one lowers vapor flow rate: mass transfer becomes less efficient
- Trays with good turndown characteristics have a wide maximum, so there is little loss in efficiency when vapor velocity decreases

Determination of Efficiency using Correlations

O'Connell correlation



$$E_o = 0.52782 - 0.27511 \log(\alpha\mu) + 0.044923 [\log(\alpha\mu)]^2$$

μ (in cP): viscosity at feed composition

α : relative volatility of key components

Both μ and α are determined at average T and P of the column

- If μ increases then efficiency decreases (mass transfer rates are lower)

Minimum Number of Stages (Fenske's equation)

$$N_{min} = \log \frac{\left[\left(\frac{d_{LK}}{b_{LK}}\right)\left(\frac{b_{HK}}{d_{HK}}\right)\right]}{\log(\alpha_{LK,HK})}$$

where d and b are the flow rates in the distillate and bottoms of the light (LK) and heavy (HK), and $\alpha_{LK,HK}$ is the relative volatility of the light to heavy component and is equal to K_{LK}/K_{HK}, taken as constant value throughout the column.

Column Diameter Calculations

There are several procedures for calculating column diameter.
The most widely used is the Fair's procedure:

- ❖ Determine vapor velocity that will cause flooding (u_{flood})
- ❖ Determine the operating velocity (u_{op})
- ❖ Determine column diameter

Flooding Vapor Velocity

in ft/s

surface tension (in dynes/ cm)

$$u_{flood} = C_{sb,\,f}\left(\frac{\sigma}{20}\right)^{0.2}\sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

capacity factor (constant)

average liquid density

average vapor density

Determination of the Capacity Factor

Flow parameter

$$\log\left(C_{sb,f}\right) = -1.1977 - 0.53143\log\left(F_{lv}\right) - 0.18760\left(\log\left(F_{lv}\right)\right)^2$$

<u>**This correlation is for 6-inch try spacing (see the textbook, pages 371-372, for other tray spacing correlations)**</u>

Mass flow rate of liquid

Volumetric flow rate of liquid

$$F_{lv} = \frac{W_L}{W_V}\sqrt{\frac{\rho_V}{\rho_L}} = \frac{Q_L}{Q_V}\sqrt{\frac{\rho_L}{\rho_V}}$$

Mass flow rate of vapor

Volumetric flow rate of vapor

Determination of the Capacity Factor – Plate Spacing

This chart will provide conservative estimate for the flooding velocity value

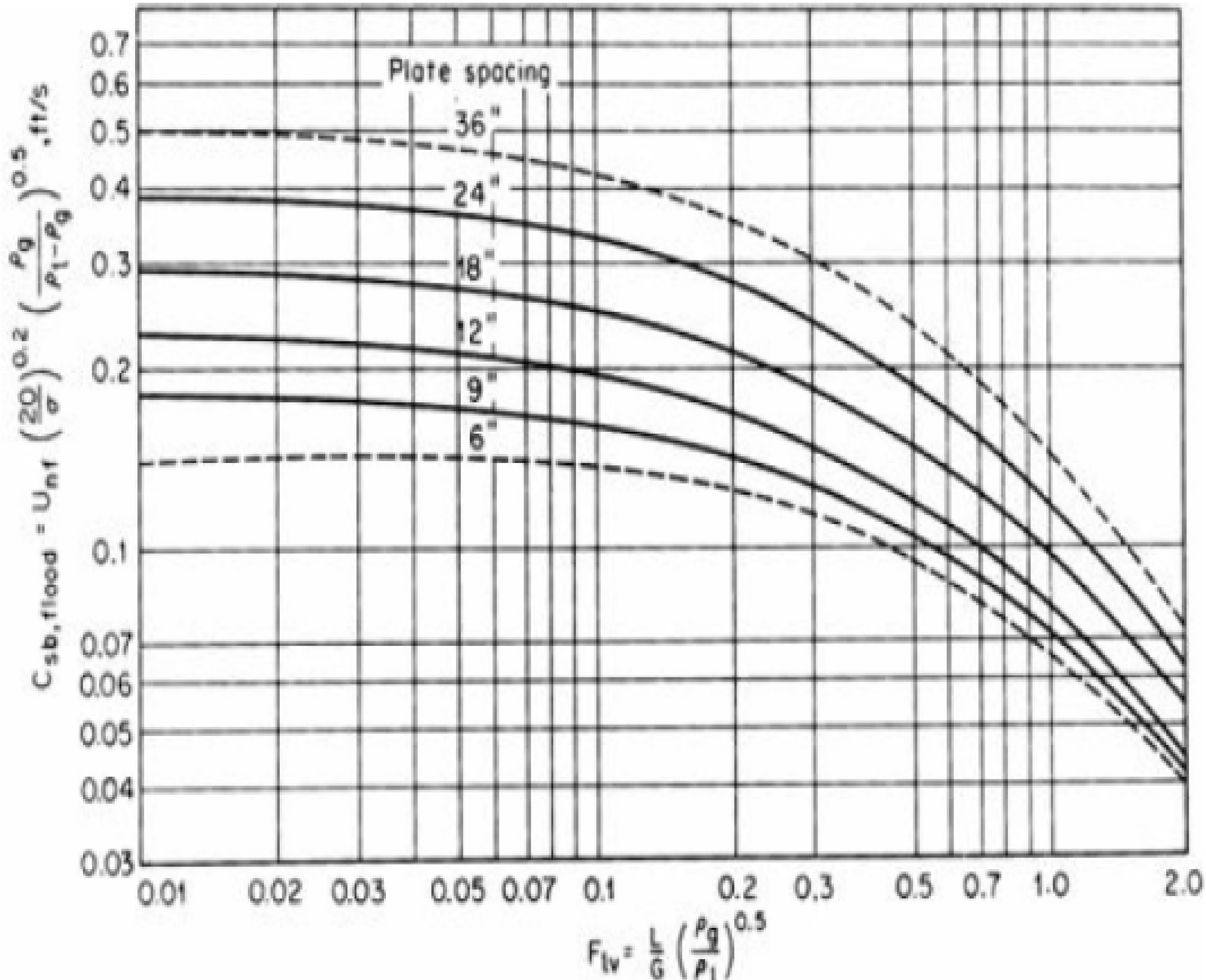


Plate spacing is the distance between the adjacent trays

$$W_L = L$$

$$W_V = G$$

$$\rho_V = \rho_g$$

$$\rho_L = \rho_l$$

Ratio of Holes Area (Ah) to Active Area (Aa)

$$\beta = \frac{A_h}{A_a}$$

This correlation $u_{flood} = C_{sb,f}\left(\frac{\sigma}{20}\right)^{0.2}\sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$ is based on $\beta \geq 0.1$

- If $\beta = 0.08$, multiply u_{flood} by 0.9
- If $\beta = 0.06$, multiply u_{flood} by 0.8

The relationship between β and the correction factor of u_{flood} is linear

Sieve Tray Spacing

- Usually selected according to maintenance requirements because it has little effect on tray efficiency
- Sieve trays are spaced 6 to 36 inches apart with 12 to 16 inches a common range for smaller (less than 5 feet) towers.
- Tray spacing is usually greater in large-diameter columns
- The minimum tray spacing is 0.4572 m (24-inch)
- Typical tray spacing 0.6096 m (24-inch) typical to allow workers to crawl through the column for inspection and maintenance.

Operating Vapor Velocity and Column Diameter

$$u_{op} = (fraction)\, u_{flood}$$

- Fraction is generally between 0.65 and 0.9
- You can always use 0.75

Molar vapor flow rate (mol/h)

$$u_{op}(ft/s) = \frac{V\,\overline{MW}_V}{3600\,\rho_V A_{net}} \Rightarrow A_{net}(ft^2) = \frac{\pi (Dia)^2}{4}\eta$$

$$\eta = \frac{\text{cross-sectional area available for vapor flow}}{\text{total cross-sectional area of the column}}$$

$$1-\eta = \text{fraction of the column area taken up by downcomer}$$

Column Diameter

$$Dia(ft) = \sqrt{\frac{4V\,\overline{MW}_V}{3600\ \pi\eta\rho_V(fraction)u_{flood}}}$$

If the vapor behaves as ideal gas: $\rho_V = \dfrac{P\overline{MW}_V}{RT}$

$$Dia(ft) = \sqrt{\frac{4VRT}{3600\ \pi\eta P(fraction)u_{flood}}}$$

Ratio of Column Diameter at Top and Bottom

For columns operating at or above atmospheric pressure, the pressure is essentially constant in the column:

$$\frac{Dia_{top}}{Dia_{bot}} = \left(\frac{V_{top}}{V_{bot}}\right)^{0.5} \left(\frac{MW_{V,top}}{\overline{MW}_{V,bot}}\right)^{0.25} \left(\frac{\rho_{L,bot}}{\rho_{L,top}}\right)^{0.25} \left(\frac{T_{top}}{T_{bot}}\right)^{0.25} \left(\frac{\sigma_{bot}}{\sigma_{top}}\right)^{0.1} \left(\frac{C_{sb,f,bot}}{C_{sb,f,top}}\right)^{0.5}$$

- If the calculation is done at different locations, different diameters will be obtained
- The largest diameter should be used and rounded off to the next highest ½-foot increment
- Columns with diameters less than 2.5 feet are usually constructed as packed columns
- If a column with a single diameter is constructed, the efficiencies in different parts of the column may vary considerably

Relationship between Murphree Tray Efficiency and Fractional Entrainment

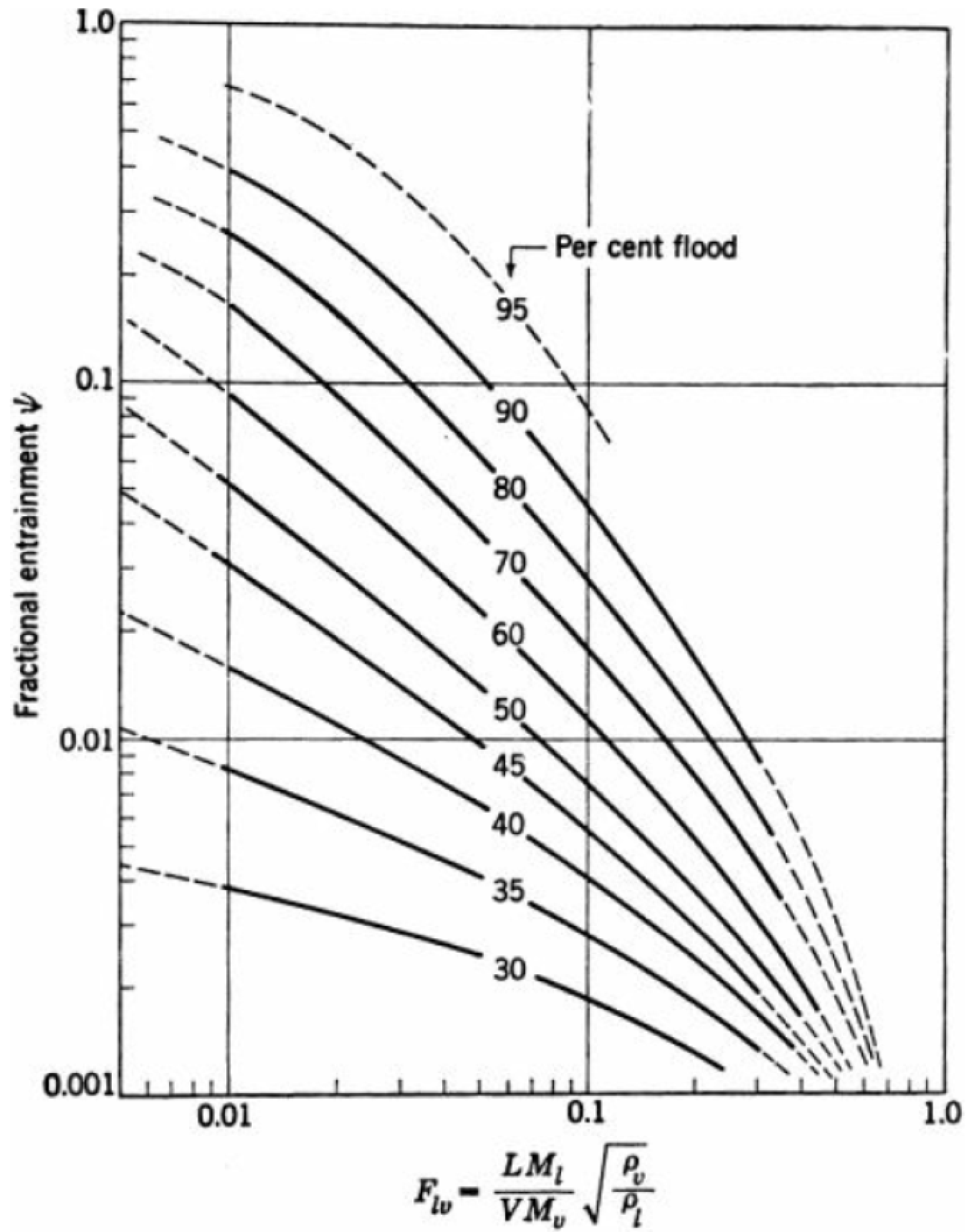
$$E_{MV,entraimmnet} = E_{MV}\left(\frac{1}{1+\frac{E_{MV}\psi}{1-\psi}}\right)$$

Fractional entrainment

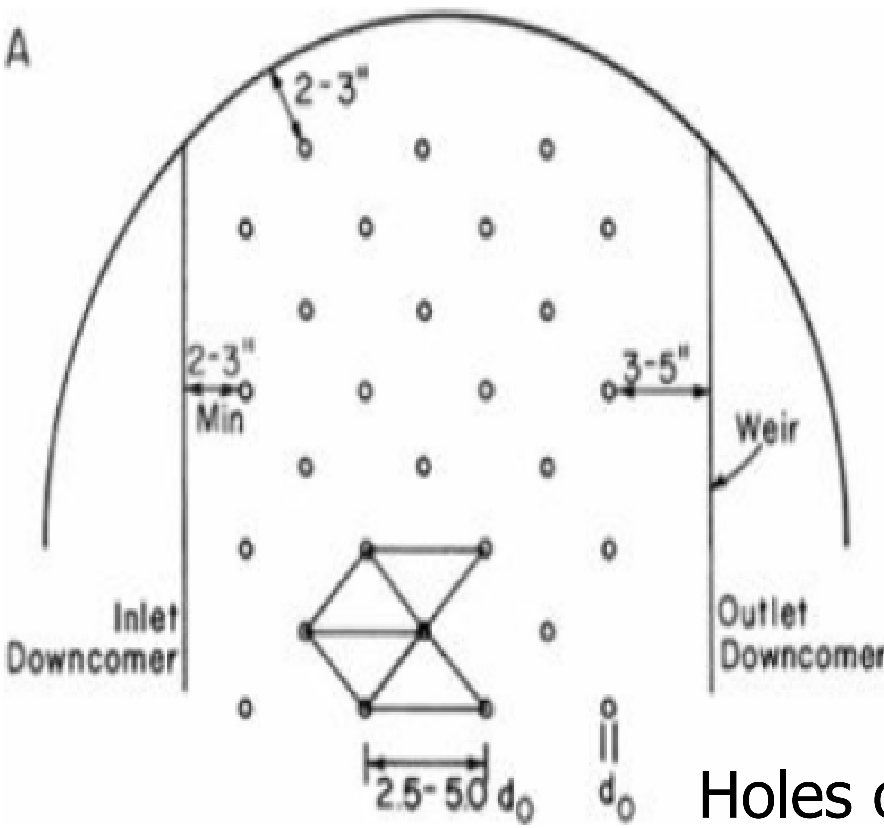
$$\psi = \frac{e}{L+e} = \frac{\text{absolute entrainmnet}}{\text{total liquid flow rate}} \quad \Rightarrow e = \frac{\psi L}{1-\psi}$$

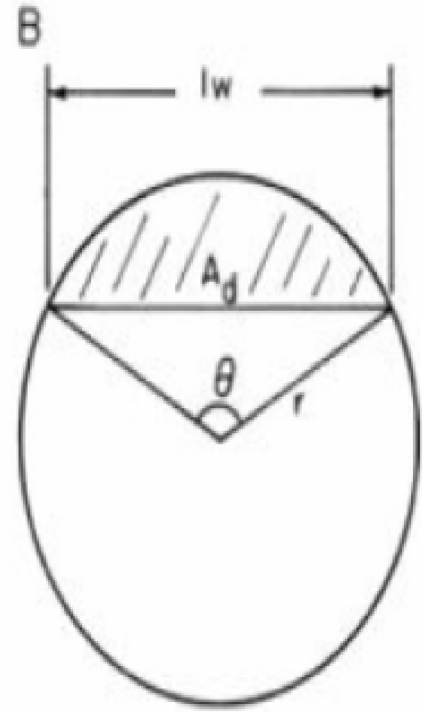
- Usually, entrainment is not a problem until $\psi > 0.1$
- $\psi > 0.1$ when $u_{op} = 85-100\%\ u_{flood}$
- When $u_{op} = 75\%\ u_{flood}$, entrainment will not be significant

Entrainment Correlation



Sieve Tray Layout





Holes on sieve plate have pattern to ensure even flow of vapor and liquid on
the tray

Sieve Tray Layout

- d_o ~1/8 to 1 inch
- Efficiency is the highest (and it is constant) for d_o ~1/2 to 1 inch
- Holes are 2.5 d_o – 5.0 d_o apart (take average of 3.8 d_o)
- Distance between the nearest holes to the column shell/inlet downcomer is 2 – 3 inches
- Distance between of the nearest holes to weir (outlet downcomer) is 3 – 5 inches
- A_h is typically 4 to 15% of A_{column}
- The ratio of A_h to the active area A_a is typically 6 to 25% (10% is a reasonable first guess)

Sieve Tray Layout

Vapor velocity through holes: $$v_o = \frac{V\,\overline{MW}_V}{3600\,\rho_V\,A_h}$$

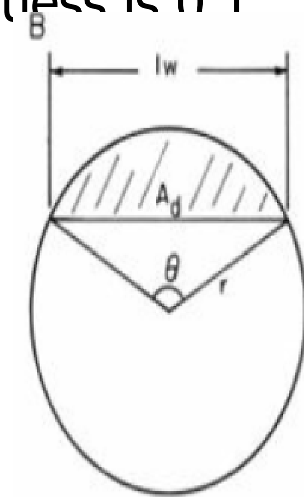
v_o should be above weeping and below flooding points

Area of holes: $$A_h = (\#\ of\ holes)\frac{\pi\ d_o}{4} = \beta\ A_a$$

Reasonable first guess is 0.1

Active area: $$A_a = A_{column}(1-2(1-\eta)) = A_{column}(2\eta - 1)$$

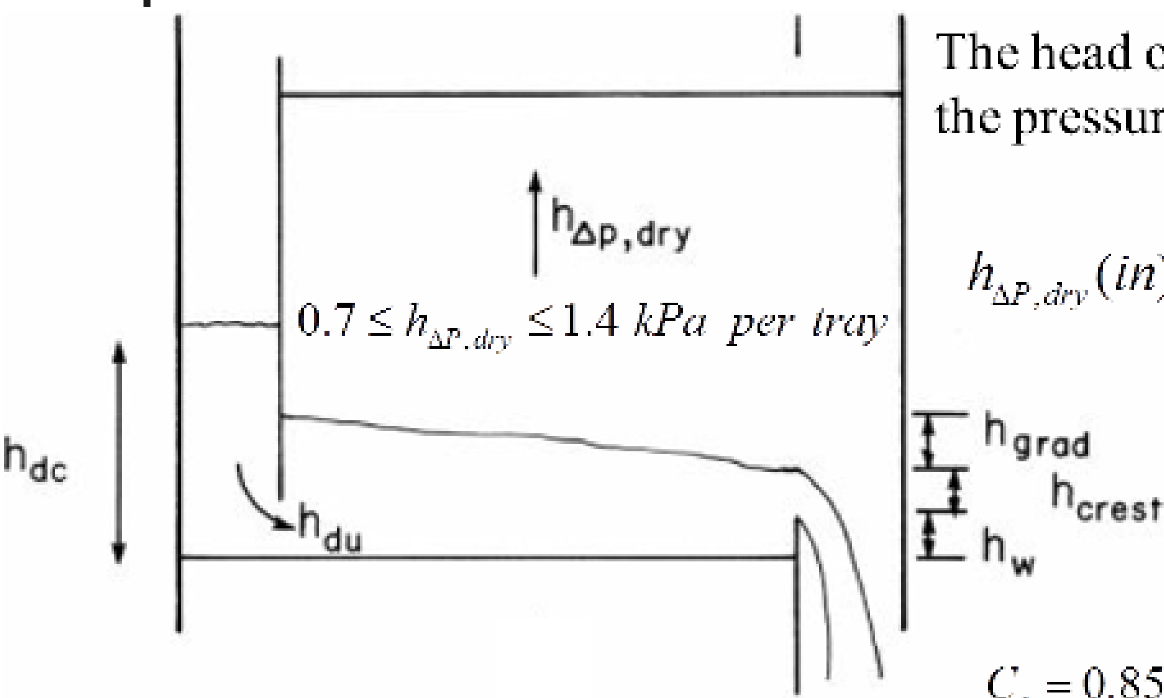
Downcomer area: $$A_d = A_{column}(1-\eta)$$



η	0.8	0.825	0.85	0.875	0.900	0.925	0.95	0.975
l_{weir}/Dia	0.871	0.843	0.811	0.773	0.726	0.669	0.593	0.478

l_{weir}: is length of the weir

Sieve Tray Hydraulics - 1



The head of liquid required to overcome the pressure drop of gas on dry tray ($h_{\Delta P,dry}$):

$$h_{\Delta P,dry}(in) = 0.003\,v_o^2 \rho_V \left(\frac{\rho_{water}}{\rho_L}\right)\left(\frac{1-\beta^2}{C_o^2}\right)$$

Vapor velocity through holes (ft/s)

Orifice coefficient

$$C_o = 0.85032 - 0.04231\frac{d_o}{t_{tray}} + 0.0017954\left(\frac{d_o}{t_{tray}}\right)^2$$

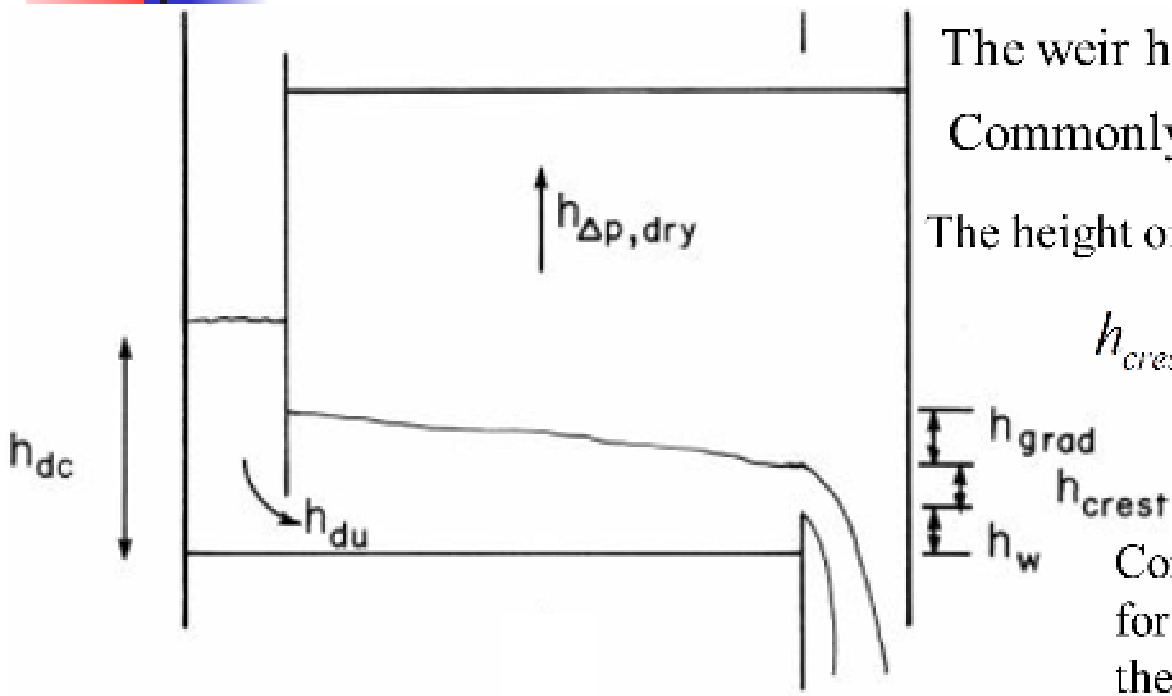
Tray thickness

$$\frac{d_o}{t_{tray}}\ must\ be \ge 1$$

Downcomer pressure head of clear liquid (h_{dc}):

$$h_{dc} = h_{\Delta P,dry} + h_{weir} + h_{crest} + h_{grad} + h_{du}$$

Sieve Tray Hydraulics - 2



The weir height (h_{weir}): $h_{weir} \geq 0.5\ in$

Commonly: $2 \leq h_{weir} \leq 4\ in$

The height of the liquid crest over the weir (h_{crest}):

$$h_{crest}(in) = 0.092\, F_{weir}\, (L_g / l_{weir})^2$$

Correction factor to account for column wall curvature in the downcomer (see next slide)

Total liquid flow rate (gal/min)

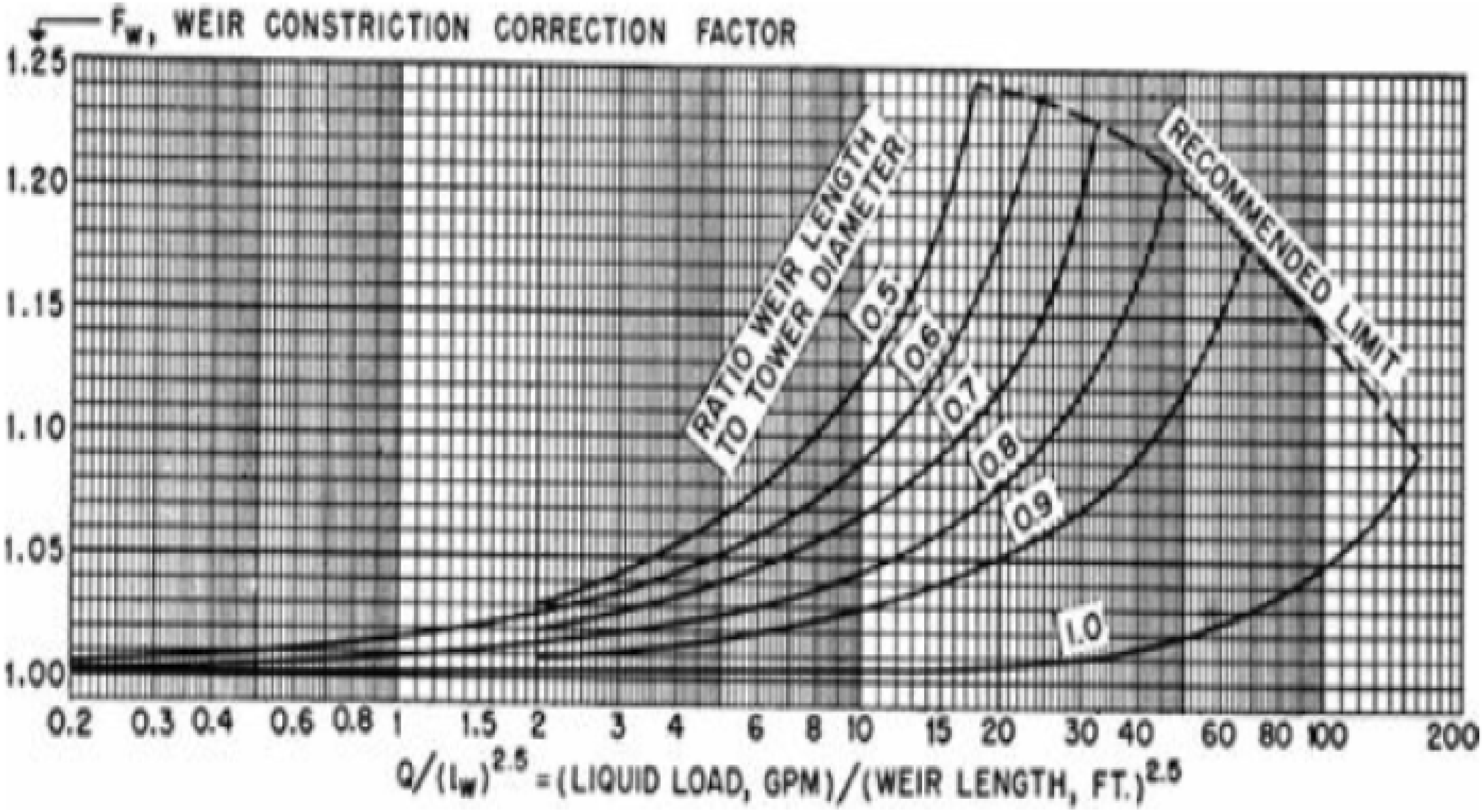
The length of weir (ft)

Downcomer pressure head of clear liquid (h_{dc}):

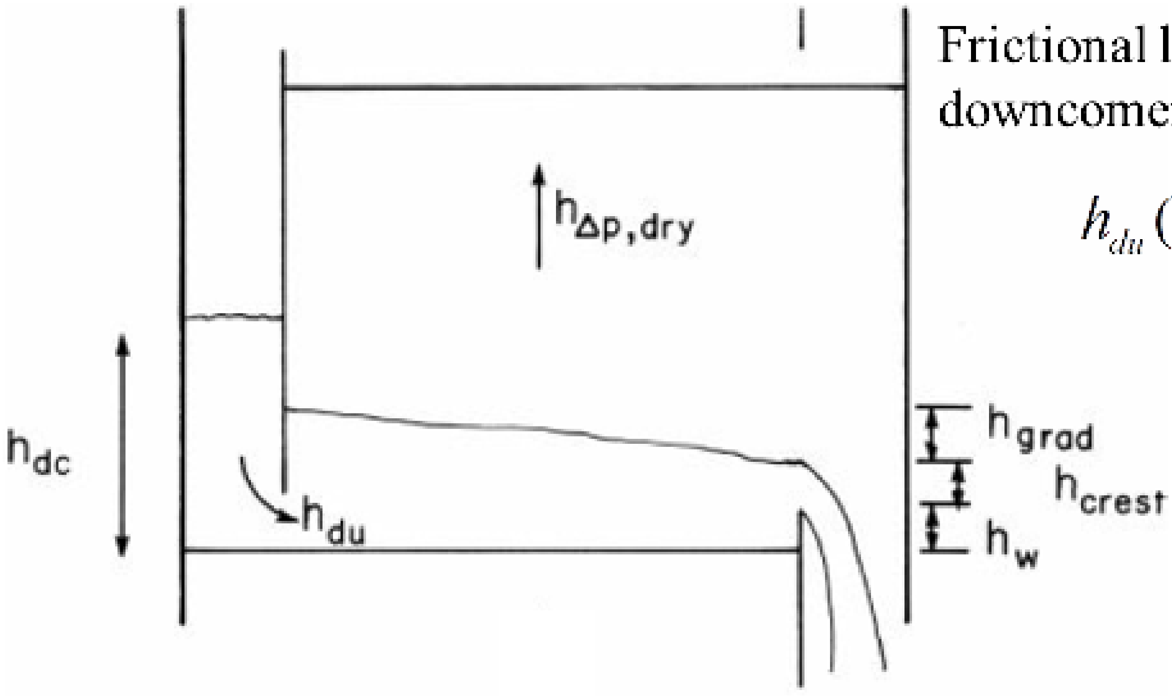
$$h_{dc} = h_{\Delta P,dry} + h_{weir} + h_{crest} + h_{grad} + h_{du}$$

Gradient of liquid height on trays. It is negligible for sieve trays

Weir Correction Factor (Fweir)



Sieve Tray Hydraulics - 3



Frictional loss due to flow in and under downcomer (h_{du}):

$$h_{du}(in) = 0.56\left(\frac{L_g}{449\ A_{du}}\right)^2$$

Flow area under downcomer apron (in ft^2)

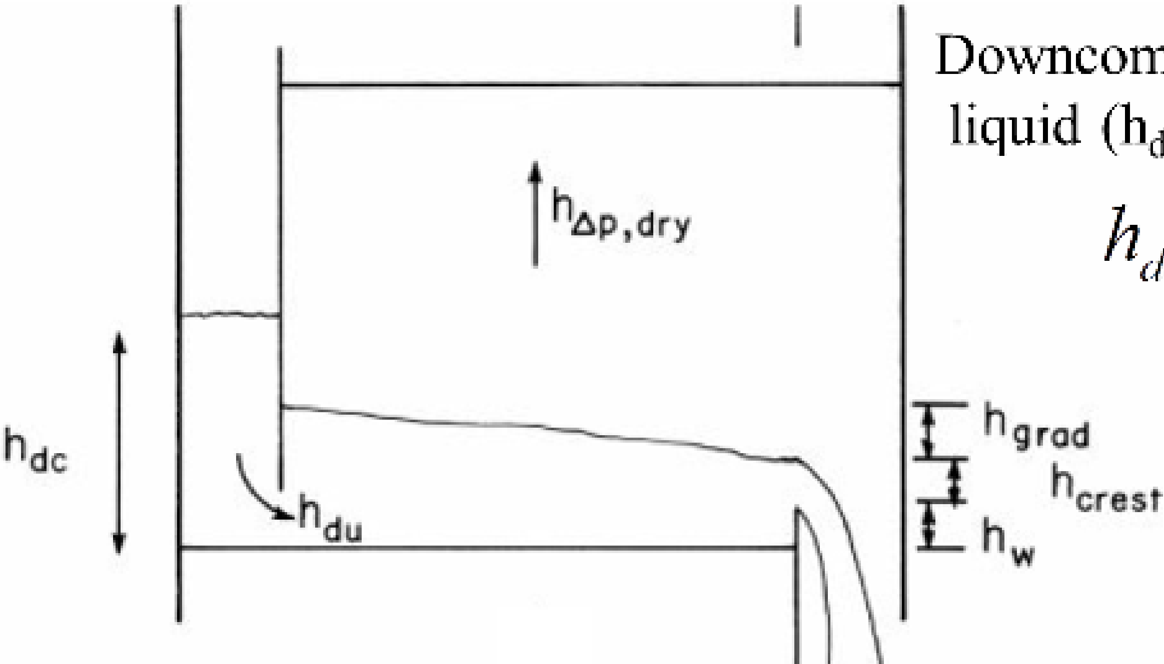
$$A_{du} = (gap)\,l_{weir}$$

Usually 1 in

Downcomer pressure head of clear liquid (h_{dc}):

$$h_{dc} = h_{\Delta P, dry} + h_{weir} + h_{crest} + h_{grad} + h_{du}$$

Sieve Tray Hydraulics - 4



Downcomer pressure head of aerated liquid ($h_{dc,\ aerated}$):

$$h_{dc,aerated} = h_{dc} / \phi_{dc}$$

Relative froth density

For normal operations: $\phi_{dc} = 0.5$

To avoid downcomer flooding, tray spacing must be > $h_{dc,aerated}$

In normal operations, the tray spacing is > $2h_{dc}$

Condition for Avoiding Weeping

$$h_{\Delta P,dry} + h_\sigma \geq 0.10392 + 0.25119x - 0.021675x^2$$

$$h_\sigma(in) = \frac{0.04\sigma}{\rho_L d_o}$$

in lb/ft^3

in dynes/cm

in in

$$x = h_{weir} + h_{crest} + h_{grad}$$

This equation is valid for $0.06 \leq \beta \leq 0.14$

END