SKF4153- PLANT DESIGN

EQUIPMENT SIZING

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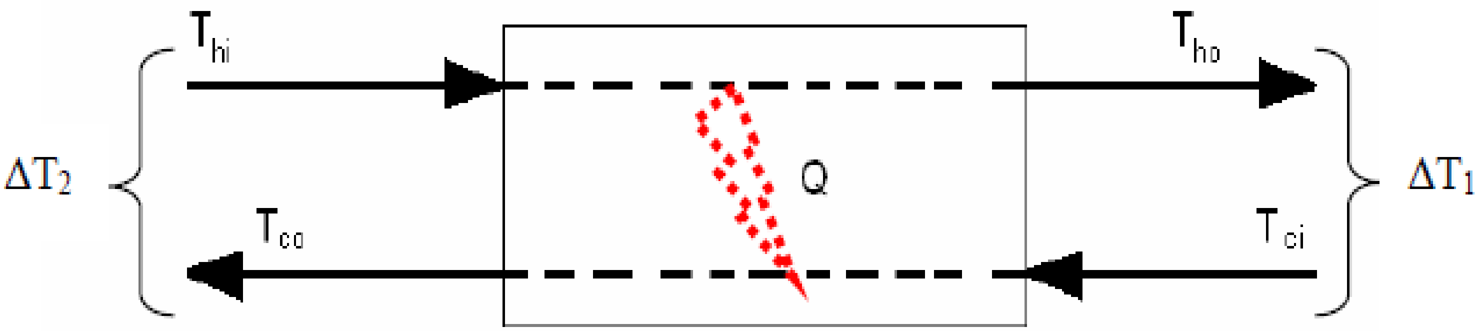
Heat Exchangers

- What to design ??
- Consider the following block diagram.



$$Q = m_h \cdot c_{p(h)} \cdot (T_{hi} - T_{ho}) = m_c \cdot c_{p(c)} \cdot (T_{co} - T_{ci}) = U \cdot A \cdot \Delta T_{LM}$$

$$\Delta T_{LM} = \frac{\Delta T_1 - \Delta T_2}{\ln\left(\frac{\Delta T_1}{\Delta T_2}\right)}$$

Shell-and-Tube Heat Exchangers

✓ For shell-and-tube heat exchangers, tubes are typically ¾-in. O.D., 16 ft long, and on 1-in. triangular spacing. A 1-ft I.D. shell can accommodate 300 ft^2 tubes outside area, 2-ft: 1330 ft^2, 3-ft: 3200 ft^2.

✓ The tube side is for corrosive, fouling, scaling, hazardous, high T & P, and expensive fluids. The shell side is for more viscous, cleaner, lower flow rate, evaporating and condensing fluids.

Reference: W.D. Seider, J.D. Seider, D.R. Lewin, Product and Process Design Principles: Synthesis, Analysis and Evaluation, John Wiley and Sons, Inc., 2010.

Pump

- Pumping action- cause increase the elevation, velocity and pressure of a fluid.
- Main purpose to provide energy to move liquids from one place to another.
- Common application is to increase the pressure of liquid.

Power Requirement

$$\dot{W} = Fv(\Delta P)$$

- F is molar flow rate; v is molar volume; P is pressure
- Due to smaller liquid molar volume, pump requires less power than compressor for the same molar flow rate and increase in P.
- Normally the outlet T of liquid increase only slightly.
- To increase a pressure of a stream, pump a liquid rather than compress a gas, unless refrigeration is needed.

  (Note: to condense a gas through refrigeration and then pump the condensate are expensive)

Pump Characteristics

- Capacity (Q) in gal/min (gpm) [conversion, 1ft^3= 7.48 gal]
- Pump head (H) in ft or m.

$$H = \left(\frac{V_d^2}{2g} + z_d + \frac{P_d}{\rho_d g} \right) - \left(\frac{V_s^2}{2g} + z_s + \frac{P_s}{\rho_s g} \right)$$

for negligible ΔV and Δz and constant ρ,

$$H = \frac{\Delta P}{\rho g}$$

Subscripts d and s refer to discharge and suction, respectively.

Heuristics

- ✓ For heads up to 3200 ft (multiple stages) and flow rates in the range 10 to 5000 gpm, use centrifugal pump.
- ✓ For high heads up to 20000 ft and flow rate up to 500 gpm, use reciprocating pump.
- ✓ Less common are axial pumps for heads up to 40 ft for flow rates in the range of 20 to 100000 gpm and rotary pumps for heads up to 3000 ft for flow rate in the range 1 to 1500 gpm.
- ✓ For liquid water,
  - ❖ Head of 3000 ft correspond to ΔP of 1300 psi,
  - ❖ Head of 20000 ft correspond to ΔP of 8680 psi.

Heuristics (cont'd)

- For liquid flow, we need to include the following when determining the required pumping head.
    - A pipeline pressure drop of 2 psi/100ft of pipe
    - A control valve pressure drop of at least 10 psi
    - A pressure drop of 4 psi per 10-ft rise in elevation
- Estimate the theoretical horsepower (THp) for pumping liquid using,

$$THp = (gpm)\left(pressure\ increase, psi\right)/1714$$

Reference: W.D. Seider, J.D. Seider, D.R. Lewin, Product and Process Design Principles: Synthesis, Analysis and Evaluation, John Wiley and Sons, Inc., 2010.

Compressor and Expander

- Compressor: to increase the velocity and/or pressure of gases.
  - Presence of liquid can damage the compressor blades.
  - Centrifugal, positive displacement and momentum transfer.
  - If exit T exceeds 375 °F, a multistage compressor with intercoolers must be employed.
- Expander or expansion turbine: used in place of valve to recover power as pressure is decreased.
  - The gas T is reduced, check for possible condensation avoid impeller erosion.
  - Often chilling of gas is more important than the power recovery.
  - For high P liquid, the power recovery through turbine is not economical.

$$Thp = SCFM\left(\frac{T_1}{8130a}\right)\left[\left(\frac{P_2}{P_1}\right)^a - 1\right]$$

- THp (adiabatic) for compressing a gas.
- SCFM= std cubic ft per min at 60°F and 1 atm
- T_1 inlet gas T in °R
- P_1 and P_2 are inlet and outlet P (absolute)
- a=(k-1)/k where k=C_p/C_v is specific heat ratio
- T_2 is estimated as,

$$T_2 = T_1\left(\frac{P_2}{P_1}\right)^a$$

Separation Tower

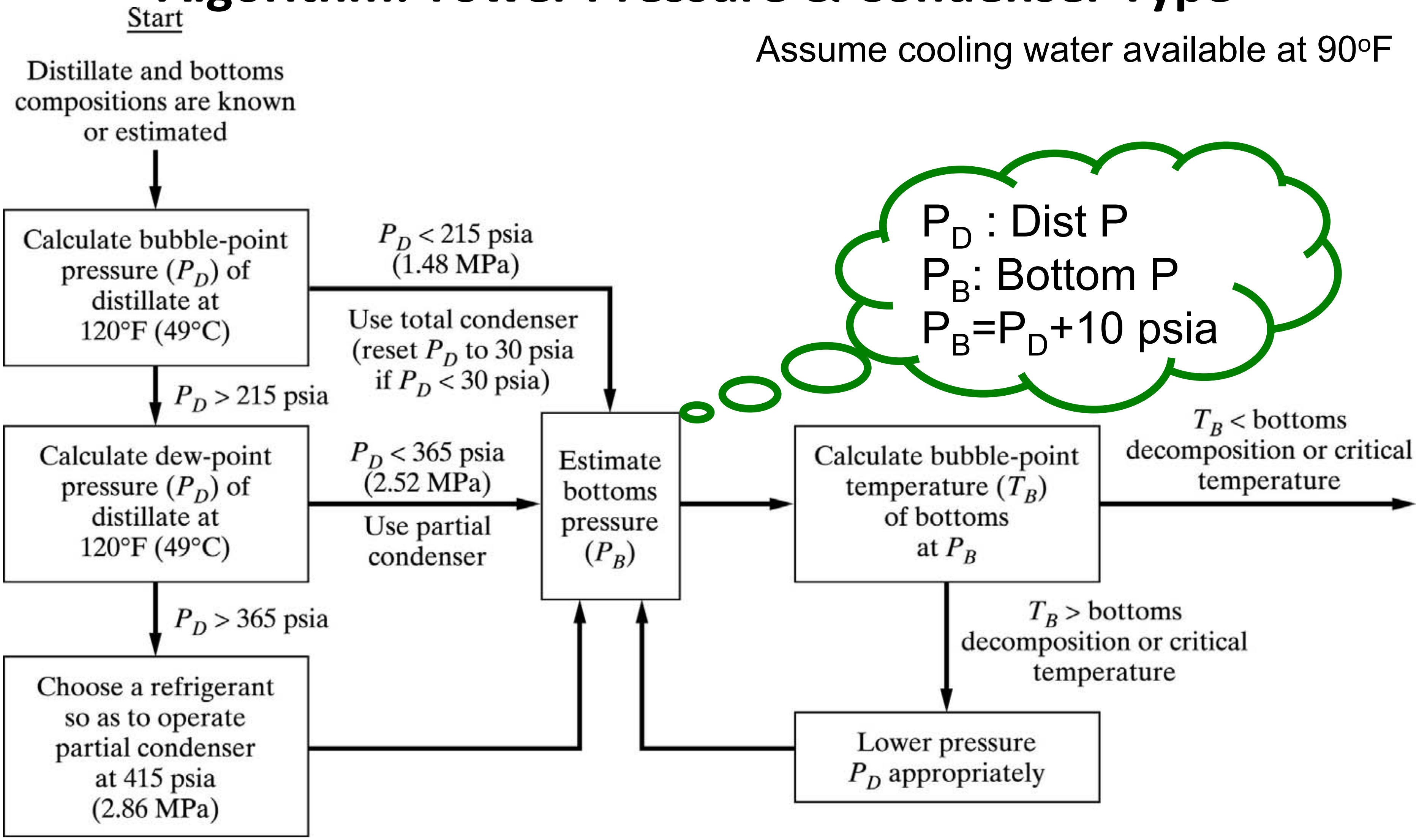
- Determine the tower operating conditions (T,P), and the type of condenser.
- Determine the equilibrium number of stages and reflux required.
- Select an appropriate contacting method (plates or packing).
- Determine the number of actual plates or packing height required, as well as the locations of feed and product.
- Determine the tower diameter.
- Determine other factors that may influence tower operation.

Distillation Tower Operating Conditions

- Proximity of critical conditions should be avoided.
- Typical operating P is 1 to 415 psia (29 bar).
- For vacuum operation P > 5 mmHg.
- Normally total condenser is used (except for low boiling components and where vapor distillate is desired).
- Preliminary material balance to estimate the distillate and bottom product compositions.

Algorithm: Tower Pressure & Condenser Type

Assume cooling water available at 90°F



Start

Distillate and bottoms compositions are known or estimated

Calculate bubble-point pressure (P_D) of distillate at 120°F (49°C)

$P_D < 215$ psia (1.48 MPa)

Use total condenser (reset P_D to 30 psia if $P_D < 30$ psia)

$P_D > 215$ psia

Calculate dew-point pressure (P_D) of distillate at 120°F (49°C)

$P_D < 365$ psia (2.52 MPa)

Use partial condenser

$P_D > 365$ psia

Choose a refrigerant so as to operate partial condenser at 415 psia (2.86 MPa)

Estimate bottoms pressure (P_B)

Calculate bubble-point temperature (T_B) of bottoms at P_B

$T_B <$ bottoms decomposition or critical temperature

$T_B >$ bottoms decomposition or critical temperature

Lower pressure P_D appropriately

P_D : Dist P
P_B: Bottom P
$P_B = P_D + 10$ psia

Reference: W.D. Seider, J.D. Seider, D.R. Lewin, Product and Process Design Principles: Synthesis, Analysis and Evaluation, John Wiley and Sons, Inc., 2010.

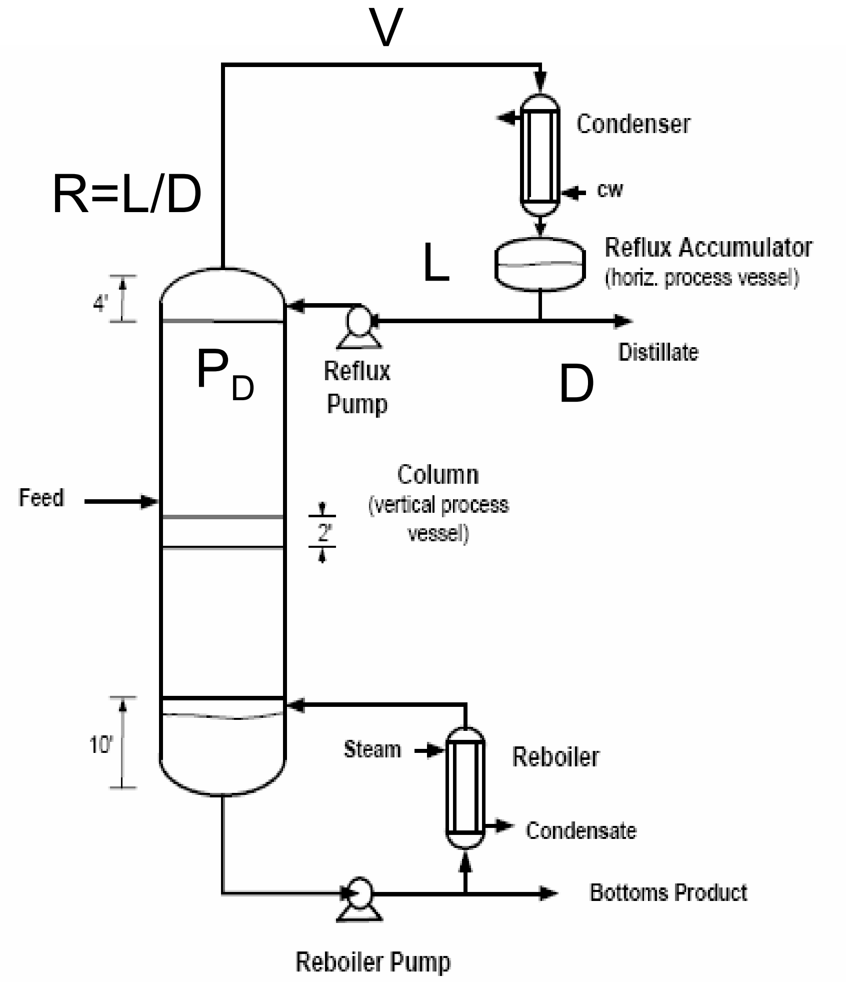
Other Things to Consider

- If top stream contains both condensable and non-condensable components, the condenser is designed to produce both vapor distillate and liquid distillate.
- The P_D is calculated at 120°F for the condensable components in liquid distillate.
- For vacuum operation, the vapor distillate is sent to vacuum pump.
- If refrigerant is used, always consider placing water-cooled partial condenser ahead of it (to reduce coolant requirement).

Design of Distillation Tower

- ❖ What to design ??
- ❖ Consider the following scenarios:
  - ➢ At total reflux ratio
  - ✓ *Minimum no. of stages*
  - ✓ *High utility cost*

  - ➢ At minimum reflux ratio
  - ✓ *Infinite no. of stages*
  - ✓ *Low utility cost*



Fenske-Underwood-Gilliland (FUG) Method

- Valid for single feed, distillate and bottoms, i.e. Ordinary Distillation.
- To estimate:
  - Reflux ratio
  - No. of equilibrium stages & feed location.
- Quite accurate for ideal mixtures of narrow boiling range.
- Not for non-ideal mixtures, azeotropes and mixtures of wide-boiling range (need to use rigorous model).

- ✓ **Step 1**: Use Fenske Equation to determine minimum number of equilibrium stages (i.e. at total reflux, D=0, R=∞)

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

d and b are component molar flowrates at distillate and bottom respectively. HK (heavy key), LK (light key), α is relative volatility

- ✓ **Step 2**: Also use Fenske Equation to determine the distribution of non-key component between distillate and bottom (d/b) streams at total reflux. Good estimate for the distribution (d/b) at finite reflux conditions.

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

- **Step 3**: Using Underwood Equation to determine minimum reflux ratio (R_{min}) that correspond to infinite number of equilibrium stages (N=∞).

$$1-q=\sum\frac{\alpha_i x_{iF}}{\alpha_i-\theta}$$

Solve ϴ by trial and error. Note: α_{HK} (equals 1) < ϴ < α_{LK}.

$$R_{\min}+1=\sum\frac{\alpha_i x_{iD}}{\alpha_i-\theta}$$

Refer C.J. Geankoplis textbook for detail calculations.

- **Step 4**: Using Gilliland correlation to estimate the actual number of equilibrium stages (N) at a specified ratio of R/R_{min}. Note: R = 1.1~1.5 R_{min}.

- **Step 5**: Estimate the feed location by using Fenske Equation.

  Calculate $N_{R,min}$ for rectifying section (between feed and distillate) and $N_{S,min}$ for stripping section (between feed and bottom).

  Assume that, $N_{R,min}/N_{S,min} = N_R/N_S$; also $N = N_R + N_S$.

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

- Alternatively, use Erbar-Maddox correlation and Kirkbride equation (refer to Geankoplis text book).

Kremser Shortcut Method for Absorption and Stripping

- For column with one feed, one absorbent or stripping agent, and two product streams.
- To estimate minimum absorbent (L_{min}) or stripping agent (V_{min}) flow rate, and the number of equilibrium stages N.
- Instead of relative volatility α, this method uses $A_e = L/KV$ for absorption and $S_e = KV/L$ for stripping.

Absorption Tower

- Minimum absorbent molar flow rate

$$L_{\min} = K_K V_{in}(1-\phi_{A_K})$$

K_K is K-value of key component computed at average T and P,

$(1-\phi_{A_K})$ is fraction of key component in gas to be absorbed

Typical actual absorbent rate L,

$$L = 1.5 L_{\min}$$

To calculate number of equilibrium stages N, use

$$\phi_{A_K} = \frac{A_{e_K} - 1}{A_{e_K}^{N+1} - 1}$$ Where, $$A_{e_K} = \frac{L}{K_K V}$$

Stripping Tower

- Min stripping agent molar flow rate

$$V_{\min} = \frac{L_{in}}{K_K}(1-\phi_{S_K})$$

K_K is K - value of key component computed at average T and P

$(1-\phi_{S_K})$ is fraction of key component in the feed liquid to be stripped

Typical actual stripping agent rate V,

$$V = 1.5V_{\min}$$

To calculate number of equilibrium stages N, use

$$\phi_{S_K} = \frac{S_{e_K} - 1}{S_{e_K}^{N+1} - 1} \qquad \text{Where,} \quad S_{e_K} = \frac{K_K V}{L}$$

Plate Efficiency and HETP

- Plate efficiency (E_o) to convert $N_{equilibrium}$ (equilibrium stages) to actual trays (N_{actual}).

$$N_{actual} = \frac{N_{equilibrium}}{E_o}$$

- Height equivalent to a theoretical plate (HETP) Height equivalent to a theoretical plate (HETP) to convert $N_{equilibrium}$ to packed height.

$$\text{Packed Height} = N_{equilibrium}(HETP)$$

Height of Tray Tower

Column Height (ft) =

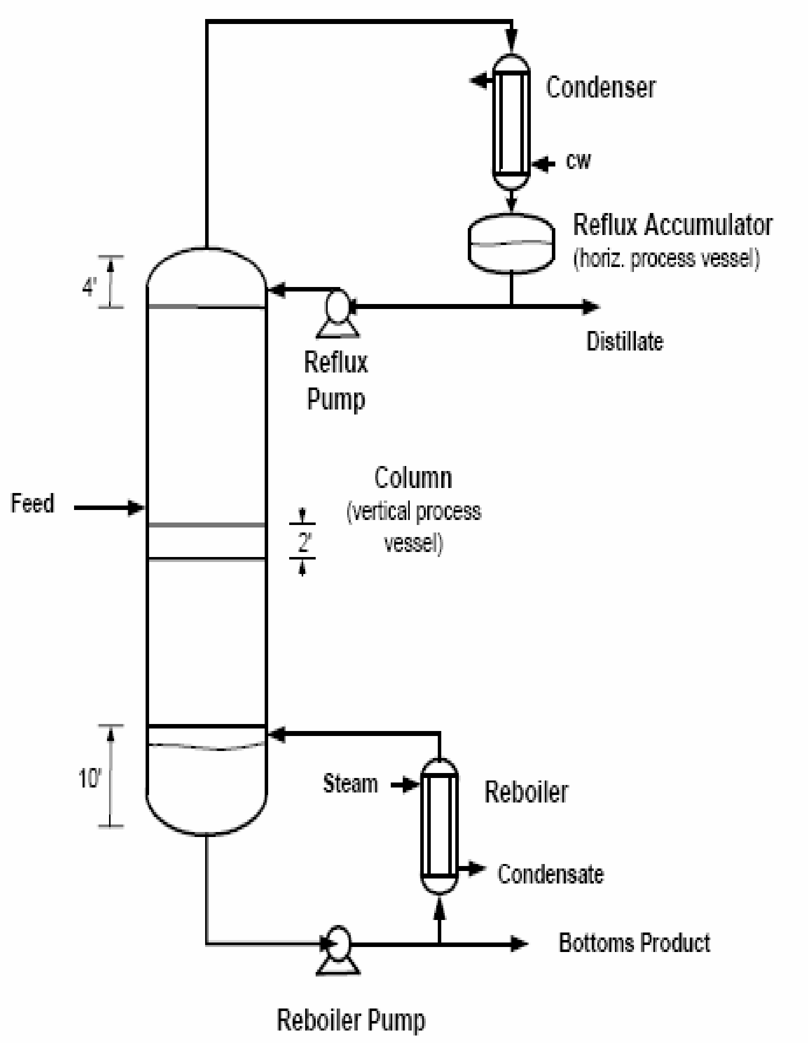
(N_{actual} - 1) x (Tray Spacing)

+ Height of sump below bottom tray

+ Disengagement height above top tray

Note:

For structural reasons, tower height must not exceed 200 ft. If calculated height exceeding 200 ft, consider using tower in series.



Height equivalent to a theoretical plate (HETP)

- Values of HETP are usually derived from experimental data for a particular type and size of packing.
- Packing vendors/manufacturers can provide HETP values.
- Typical values of :
  - ✓ For modern random packing: 2 ft
  - ✓ For structured packing: 1 ft
- HETP as a function of <u>nominal diameter</u> of random packings, and <u>specific surface area</u> as recommended by Kister (1992).

Why Tower Diameter ?

- Tower diameter is calculated to avoid flooding (i.e. liquid began to fill the tower and leave with vapor at top).
- The diameter depends on,
  - ✓ Flowrates of vapor and liquid.
  - ✓ Properties of vapor and liquid.

Diameter of Tray Tower

Tower inside diameter,

$$D_T = \left[\frac{4G}{(fU_f)\pi\left(1 - \frac{A_d}{A_T}\right)\rho_G} \right]^{\frac{1}{2}}$$

Flooding velocity,

$$U_f = C\left(\frac{\rho_L - \rho_G}{\rho_G} \right)^{\frac{1}{2}}$$

G = mass flow rate of gas L = mass flow rate of liquid

f = fraction of flooding velocity = 0.75 to 0.85

A_d = downcomer area A_T = tower inside cross-sectional area

ρ_G = density of gas ρ_L = density of liquid

F_{LG} = flow ratio parameter = $(\mathrm{L/G})(\rho_G / \rho_L)^{0.5}$

Also,

$$\frac{A_d}{A_T} = 0.1 \text{ if } F_{LG} \le 0.1$$

$$= 0.1 + \frac{(F_{LG} - 0.1)}{9} \text{ if } 0.1 \le F_{LG} \le 1.0$$

$$= 0.2 \text{ if } F_{LG} \ge 1.0$$

C = capacity parameter = $C_{SB} F_{ST} F_F F_{HA}$

C_{SB} = capacity parameter for perforated plate (see figure 19.4)

F_{ST} = surface tension factor = $(\sigma/20)^{0.2}$ σ is surface tension in dyne/cm

F_F = foaming factor

= 1 for non-foaming system (typical distillation)

= 0.5 - 0.75 for foaming system (typical absorption)

F_{HA} = hole area factor

= 1 for valve and bubble cap trays

= 1 for sieve tray with $(A_h / A_a) \geq 0.1$

= $[5(A_h / A_a) + 0.5]$ for sieve tray with $0.06 \leq (A_h / A_a) \leq 0.1$

A_a = active area of the tray = $(A_T - 2A_d)$ A_h = total hole area on the tray

Diameter of Packed Tower

Tower inside diameter,

$$D_T = \left[\frac{4G}{(fU_f)\pi\rho_G}\right]^{\frac{1}{2}}$$

For flooding velocity, use

$$Y = \frac{U_f^2 F_P}{g}\left(\frac{\rho_G}{\rho_{H2O(l)}}\right) f\{\rho_L\} f\{\mu_L\}$$

F_P = packing factor (See Table 19.1) g = 32.2ft/s^2

f = fraction of flooding velocity = 0.70

ρ_G = density of vapor ρ_L = density of liquid

Where,

$$Y = \exp\left[-3.7121 - 1.0371(\ln F_{LG}) - 0.1501(\ln F_{LG})^2 - 0.007544(\ln F_{LG})^3\right]$$

valid for Y = 0.01 to 10; F_{LG} = flow ratio parameter (as in Tray Tower)

Density function,

$$f\{\rho_L\} = -0.8787 + 2.6776\left(\frac{\rho_{H2O(L)}}{\rho_L}\right) - 0.6313\left(\frac{\rho_{H2O(L)}}{\rho_L}\right)^2$$

valid for density ratios from 0.65 to 1.4

Viscosity function (for random packing nominal diameter 1 in. or greater),

$$f\{\mu_L\} = 0.96\mu_L^{0.19}$$

valid for liquid viscosities from 0.3 to 20 cP

References

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