PROCESS EQUIPMENT DESIGN COURSE

FRACTIONATORS & ABSORBERS

Prepared and Presented by :

A. Kiashemshaki

Introduction

- Distillation is probably the most widely used separation process in the chemical and allied industries; its applications ranging from the rectification of alcohol, to the fractionation of crude oil.
- A good understanding of methods used for correlating vapor-liquid equilibrium data is essential to the understanding of distillation and other equilibrium-staged processes.
- Fractionation is a unit operation utilized to separate mixtures into individual products. Fractionation involves separating components by relative volatility (α).

Introduction

Distillation column design

- The design of a distillation column can be divided into the following steps:

  1. Specify the degree of separation required: set product specifications.

  2. Select the operating conditions: batch or continuous; operating pressure.

  3. Select the type of contacting device: trays or packing.

  4. Determine the stage and reflux requirements: the number of equilibrium stages.

  5. Size the column: diameter, number of real stages.

  6. Design the column internals: trays, distributors, packing supports.

  7. Mechanical design: vessel and internal fittings.

Introduction

Distillation column design

- The principal step will be to determine the stage and reflux requirements.
- This is a relatively simple procedure when the feed is a binary mixture, but a complex and difficult task when the feed contains more than two components (multi-component systems)

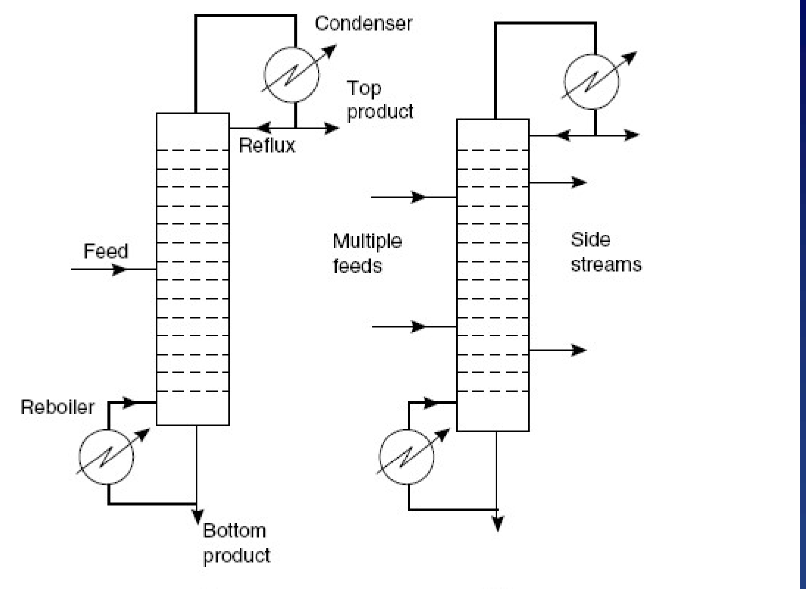


Figure 11.1. Distillation column (*a*) Basic column (*b*) Multiple feeds and side streams

Introduction

Distillation column design

- In the section below the feed, the more volatile components are stripped from the liquid and this is known as the *stripping section.*
- Above the feed, the concentration of the more volatile components is increased and this is called the enrichment, or more commonly, the *rectifying section.*
- The reflux ratio, R, is normally defined as:

$$R = \frac{\text{flow returned as reflux}}{\text{flow of top product taken off}}$$

- Total reflux is the condition when all the condensate is returned to the column as reflux: no product is taken off and there is no feed.

Introduction

Distillation column design

- As the reflux ratio is reduced a *pinch point will occur at which the separation can only be* achieved with an infinite number of stages. This sets the minimum possible reflux ratio for the specified separation.
- Practical reflux ratios will lie somewhere between the minimum for the specified separation and total reflux. The designer must select a value at which the specified separation is achieved at minimum cost.
- for many systems the optimum will lie between 1.2 to 1.5 times the minimum reflux ratio.

Introduction

Distillation column design

- The precise location of the feed point will affect the number of stages required for a specified separation and the subsequent operation of the column.
- As a general rule, the feed should enter the column at the point that gives the best match between the feed composition (vapor and liquid if two phases) and the vapor and liquid streams in the column.

Introduction

Selection of column pressure

- The before any design calculations can be made on a fractionation problem, a tower operating pressure must be determined.
- Since alpha's for hydrocarbons increase with decreasing pressure, the lowest pressure gives the lowest utilities.
- Optimum pressure is considered to be the lowest pressure obtainable for a reasonable temperature difference between process (overhead product bubble point, or receiver temperature) and conventional cooling sources (air or water).

Introduction

Selection of column pressure

- Exceptions to the practice of using lowest possible pressure are:

1. Vacuum columns are usually only employed where a temperature degradation problem exists. Disadvantages of vacuum operation are low tray efficiency and low turndown, high capital cost due to large diameter column and piping which results from low vapor density and process oxygen contamination possibility.
2. Routing of gas product to destination without compression.
3. Minimizing loss of valuable components in gas products sent to fuel gas or disposal.
4. increasing pressure to where the temperature increase allows column heat removal Via external eXChange against process or utility streams can have an attractive payout. Total column heat input increases andcapital costs are usually significantly higher.

Introduction

Selection of column pressure

- One of the primary considerations for operating pressure is the cooling medium available for the reflux condenser.
- The overhead product will be at bubble point conditions for a liquid product or at dew point conditions for a vapor product. The bubble point (or dew point) pressure is fixed by a desired component separation and the temperature of the cooling medium.
- The cooling media typically used are air, water, and refrigerant.
- In some cases, the gas produced as overhead from a column must be compressed to sales gas export pressure or to feed pressure of another process unit. Therefore a higher operating pressure may be desired to reduce compression horse power.

Introduction

Selection of column pressure

- However other items must be considered which will limit pressure selection. If an operating pressure is too high, the critical temperature of the bottom product may be exceeded and the desired separation cannot be achieved. Additionally, the pressure cannot exceed the critical pressure of the desired overhead product.
- Air cooling is normally the least expensive cooling method. Practical exchanger design limits the process to a 10°C approach to the ambient summer temperature.
- With cooling water, process temperatures of 35 to 41°C are possible.
- Below about 35°C, mechanical refrigeration must be used to achieve the desired condensing temperature. This is the most expensive cooling method from both a capital and operating cost standpoint.

Introduction

Selection of column pressure

- Order of magnitude of operating pressure in the reboiler is given by following relation:

- operating pressure in reboiler = operating pressure in reflux drum,
  + pressure drop in condenser,
  + pressure drop in the actual trays (or packing) of the column,

- Order of magnitude of pressure drop in condenser = 0.1 bar to 0.7 bar,
- pressure drop in the trays of the column = n x (ΔP) t

where: (ΔP)t = pressure drop in an actual tray

- For normal operations, pressure drop = 0.5 to 1.3 kPa/tray
- In a first approach adopt a value of 1 kPa/tray.

Basic Principles

- Material and energy balance equations can be written for any stage in a multistage process:

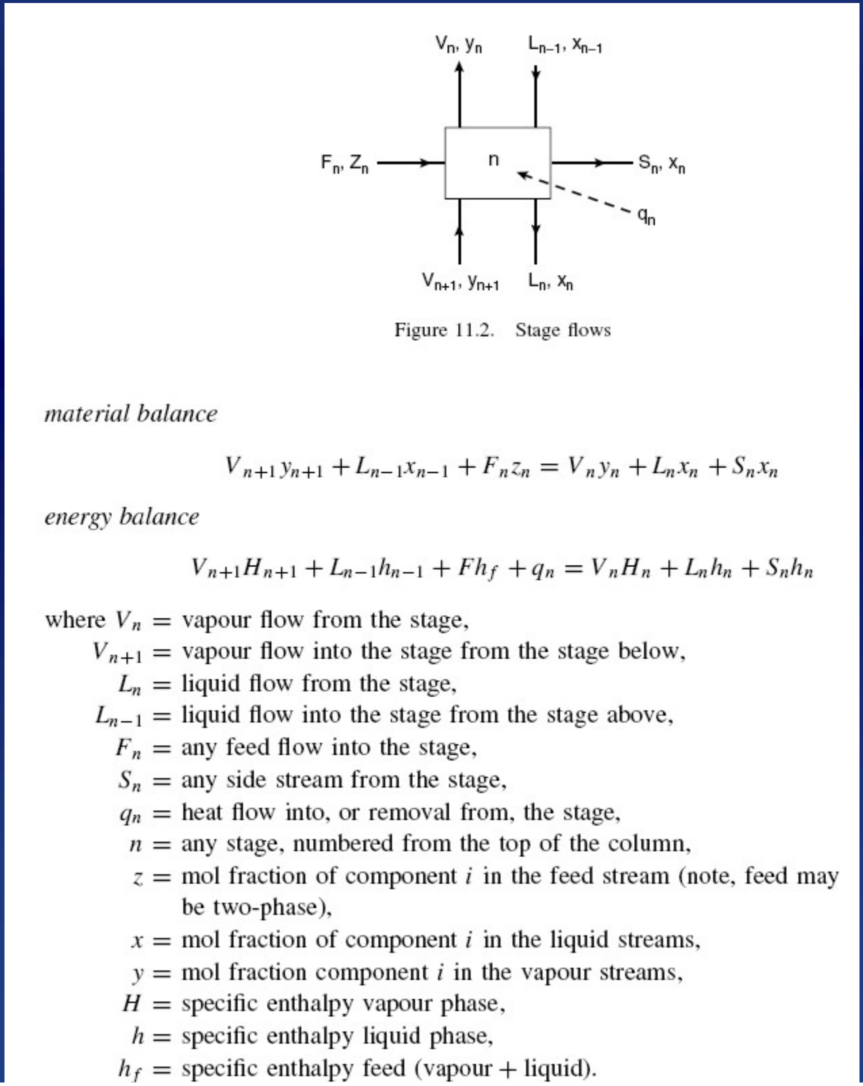


Figure 11.2. Stage flows

material balance

$$V_{n+1}y_{n+1} + L_{n-1}x_{n-1} + F_n z_n = V_n y_n + L_n x_n + S_n x_n$$

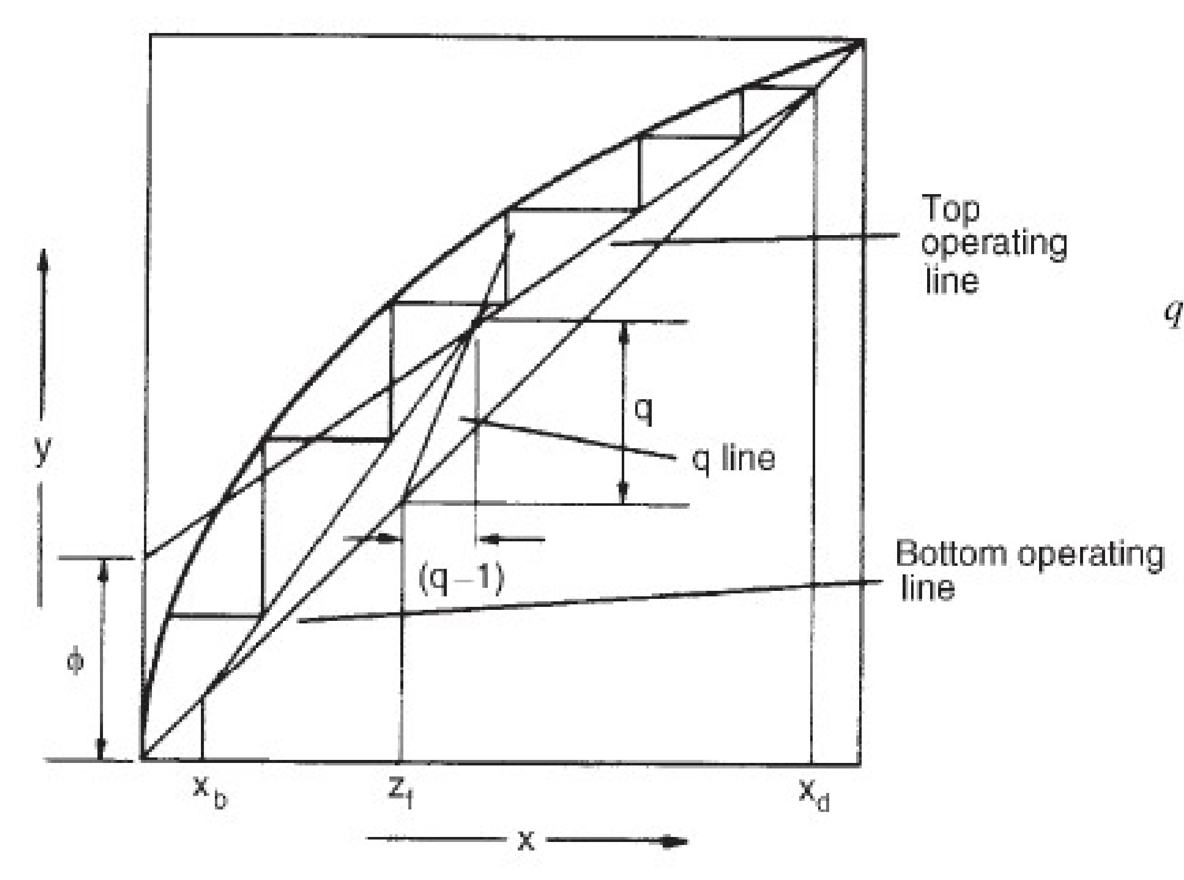
energy balance

$$V_{n+1}H_{n+1} + L_{n-1}h_{n-1} + Fh_f + q_n = V_n H_n + L_n h_n + S_n h_n$$

where V_n = vapour flow from the stage,
V_{n+1} = vapour flow into the stage from the stage below,
L_n = liquid flow from the stage,
L_{n-1} = liquid flow into the stage from the stage above,
F_n = any feed flow into the stage,
S_n = any side stream from the stage,
q_n = heat flow into, or removal from, the stage,
n = any stage, numbered from the top of the column,
z = mol fraction of component i in the feed stream (note, feed may be two-phase),
x = mol fraction of component i in the liquid streams,
y = mol fraction component i in the vapour streams,
H = specific enthalpy vapour phase,
h = specific enthalpy liquid phase,
h_f = specific enthalpy feed (vapour + liquid).

Basic Principles

McCabe-Thiele Diagram



$$q = \frac{\text{heat to vaporise 1 mol of feed}}{\text{molar latent heat of feed}}$$

$$\text{slope} = q/(q-1)$$

$$\phi = \frac{x_d}{1+R}$$

for B.P. feed, q = 1.0
for D.P. feed, q = 0
for 2 phase feed, 0 < q < 1.0

Basic Principles

Multi-Component Distillation : General Consideration

- The problem of determining the stage and reflux requirements for multi-component distillations is much more complex than for binary mixtures.
- With a multi-component mixture, fixing one component composition does not uniquely determine the other component compositions and the stage temperature.
- The separation between the top and bottom products is specified by setting limits on two “key” components, between which it is desired to make the separation.
- The light key will be the component that it is desired to keep out of the bottom product, and the heavy key the component to be kept out of the top product.

Basic Principles

<u>Multi-Component Distillation : Key Parameters</u>

- If any uncertainty exists, trial calculations should be made using different components as the keys to determine the pair that requires the largest number of stages for separation (the worst case).

- Two important considerations which affect the size and cost of a fractionation column are degree of separation and component volatility. The degree of separation or product purity has a direct impact on the size of the column and the required utilities.

- Higher purity will require more trays, more reflux, larger diameter, and/or a reduced product quantity. One quantitative measure of the difficulty of a separation is the separation factor, SF, defined as:

$$S_F = \left(\frac{X_D}{X_B}\right)_{LK} \left(\frac{X_B}{X_D}\right)_{HK}$$

Basic Principles

Multi-Component Distillation : Key Parameters

- Typically, for most fractionation problems this factor ranges from around 500 to 2000. However, for sharp separations, it can be in the 10,000 range. The number of trays will be roughly proportional to the log of the separation factor for a given system.
- The volatility of the components is usually expressed as relative volatility, α. This quantity is computed as the ratio of the equilibrium K-values of two components at a given temperature and pressure. For fractionation calculations the α of the key components is important

$$\boxed{\alpha = K_{LK}/K_{HK}}$$

- The larger α is, the easier is the separation.

Basic Principles

Multi-Component Distillation : Minimum Stages

- The minimum stages can be calculated for most multicomponent systems by the **Fenske** equation:

$$S_m = \frac{\log S_F}{\log (\alpha_{avg})}$$

- S_m in this equation includes a partial reboiler and a partial condenser if these items are used.

- The α_{avg} is the column average relative volatility of the key components in the separation.

$$\alpha_{avg} = \frac{\alpha_{top} + \alpha_{bottom}}{2}$$

Basic Principles

Multi-Component Distillation : Minimum Stages

- If volatility varies widely, the approach of Winn is suggested, in which a modified volatility is used:

$$\beta_{ij} = K_{LK}/K_{HK}^{b}$$

- where the exponent b is obtained from K value plots over the range of interest.

- The minimum stage calculation is:

$$S_m = \frac{\log\left[\left(\frac{X_D}{X_B}\right)_{LK}\left(\frac{X_B}{X_D}\right)_{HK}^{b}\left(\frac{B}{D}\right)^{1-b}\right]}{\log \beta_{ij}}$$

Basic Principles

Multi-Component Distillation : Minimum Reflux Ratio

- The **Underwood** method is the most widely used of the methods for calculating minimum reflux ratio. Underwood assumed constant relative volatility and constant molal overflow in the development of this method. The first step is to evaluate θ by trial and error:

$$1 - q = \sum_{i=1}^{n} \frac{X_{Fi}}{(\alpha_i - \theta)/\alpha_i}$$

- q is moles of saturated liquid in the feed per mole of feed
- Once θ is determined, the minimum reflux ratio is:

$$(L_o/D)_m + 1 = R_m + 1 = \sum_{i=1}^{n} \frac{X_{Di}}{(\alpha_i - \theta)/\alpha_i}$$

Basic Principles

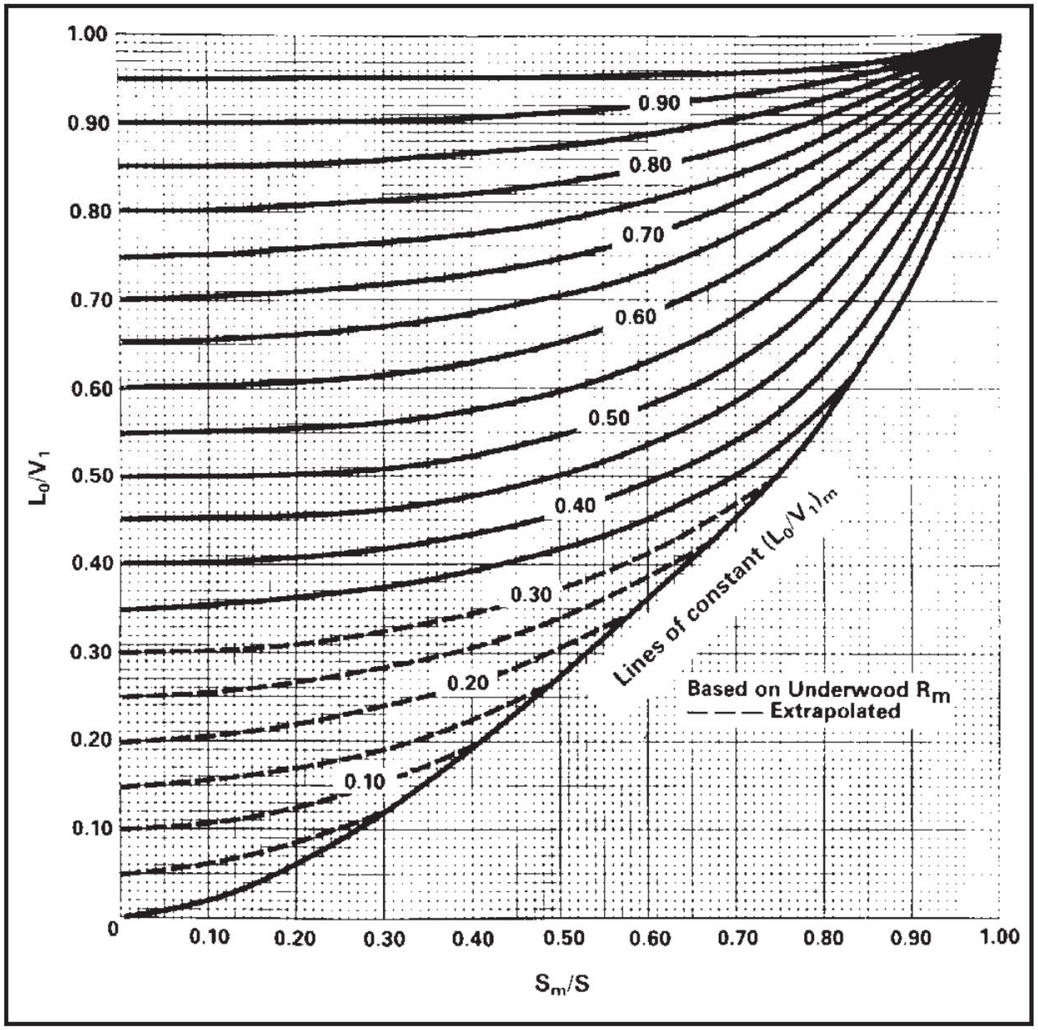
Multi-Component Distillation : Number of Stages

- The number of theoretical stages required for a given separation at a reflux ratio between minimum and total reflux can be determined from empirical relationships.
- **Erbar and Maddox** made an extensive investigation of tray by tray fractionator calculation.
- Next slide figure be used to determine an operating reflux for a given number of stages by entering the figure at the value of Sm/S, moving up to the line representing the value of Rm/(Rm + 1) and reading a value of R/(R + 1).
- The optimum operating reflux ratio has been found to be near the minimum reflux ratio. Values of 1.2 to 1.5 times the minimum are common.

Basic Principles

Multi-Component Distillation : Number of Stages

Erbar-Maddox Correlation of Stages vs Reflux[6]



Basic Principles

Multi-Component Distillation : Number of Stages

- This correlation is generated on the basis of bubble point feed. If the feed is between the bubble point and dew point then the operating reflux should be corrected. Erbar and Maddox proposed the following relationship to adjust the vapor rate from the top tray for nonbubble point feed.

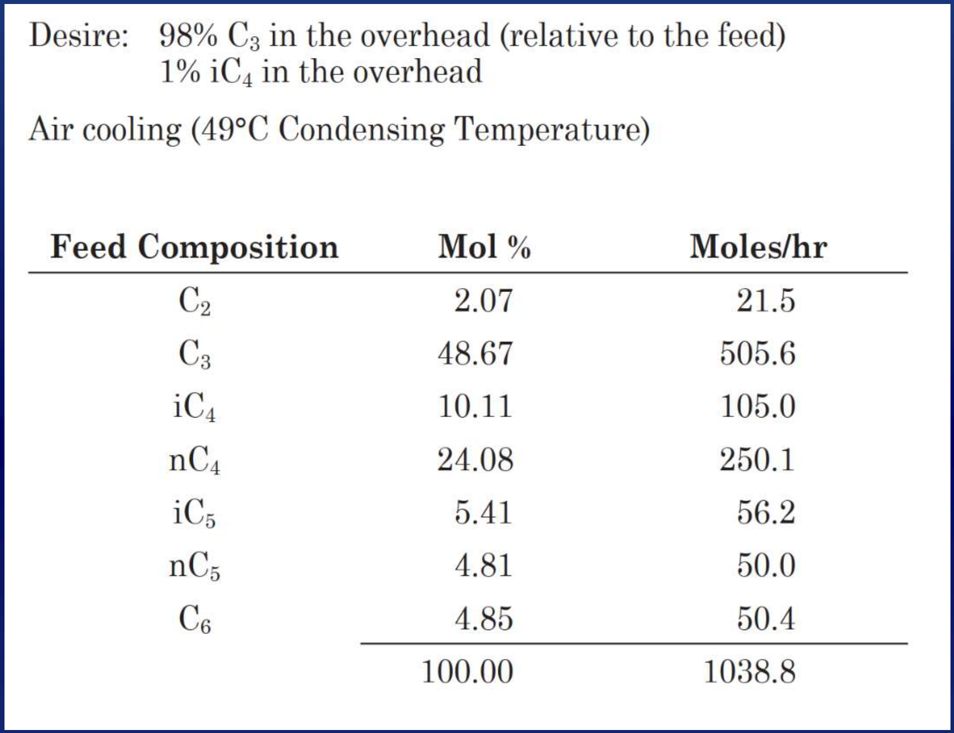
$$V_{corr} = V_{calc} + \frac{\left(1 - \frac{D}{F}\right)[F(H_{VF} - H_{BP})]}{\left(\frac{Q_C}{L_o}\right)_{calc}}$$

- The reflux rate can then be adjusted by material balance since:

$$L_o = V_1 - D$$

Basic Principles

Multi-Component Distillation : Example



Desire: 98% C_3 in the overhead (relative to the feed)
1% iC_4 in the overhead

Air cooling (49°C Condensing Temperature)

Feed Composition	Mol %	Moles/hr
C_2	2.07	21.5
C_3	48.67	505.6
iC_4	10.11	105.0
nC_4	24.08	250.1
iC_5	5.41	56.2
nC_5	4.81	50.0
C_6	4.85	50.4
	100.00	1038.8

- Find the:

  Minimum trays required

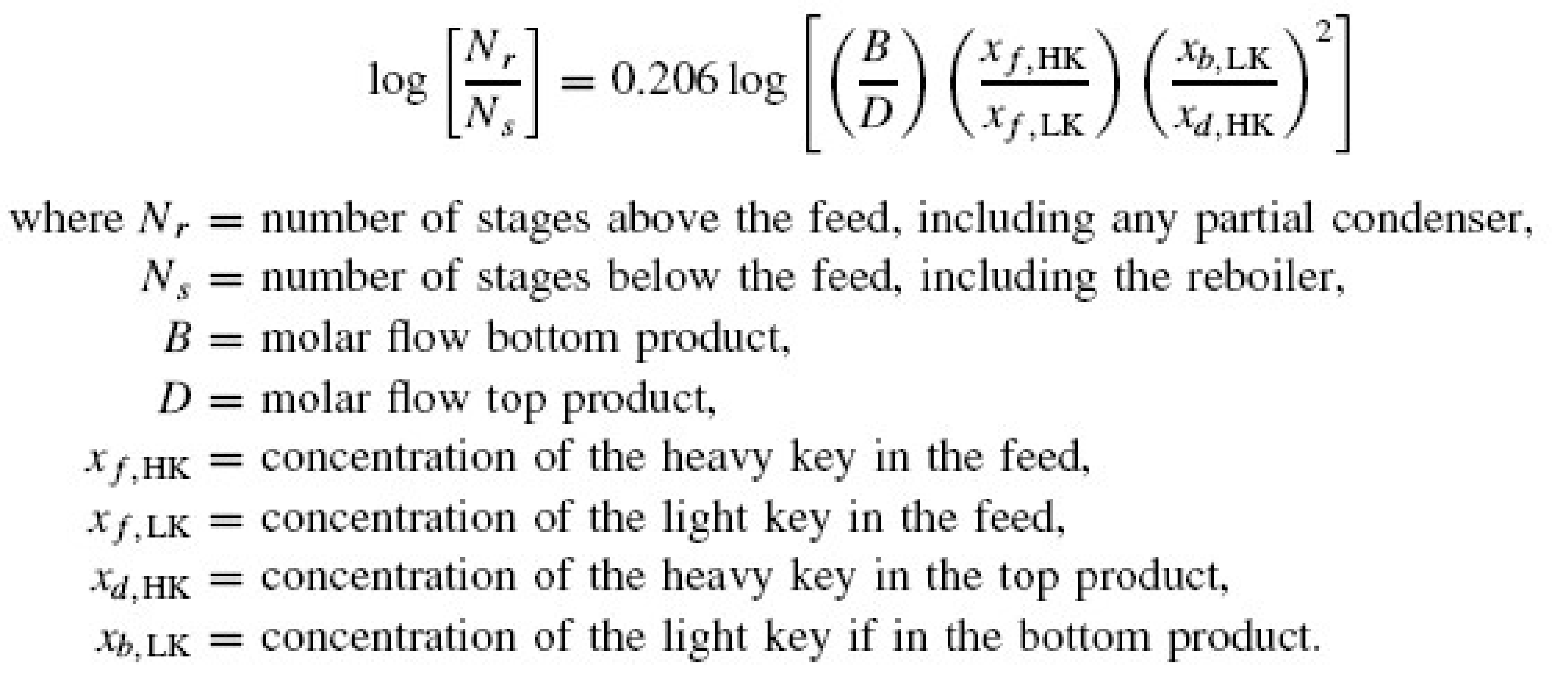
  Minimum reflux ratio

  Actual trays at $R=1.3\ R_m$ and feed is at its boiling point.

Basic Principles

Multi-Component Distillation : Empirical Equations

- Feed-Point Location by Kirkbride (1944):
- The feed tray location should be determined so that feed enters at a temperature so that no sharp change occurs in the column temperature gradient at the feed tray.



$$\log\left[\frac{N_r}{N_s}\right] = 0.206\log\left[\left(\frac{B}{D}\right)\left(\frac{x_{f,\mathrm{HK}}}{x_{f,\mathrm{LK}}}\right)\left(\frac{x_{b,\mathrm{LK}}}{x_{d,\mathrm{HK}}}\right)^2\right]$$

where N_r = number of stages above the feed, including any partial condenser,

N_s = number of stages below the feed, including the reboiler,

B = molar flow bottom product,

D = molar flow top product,

$x_{f,\mathrm{HK}}$ = concentration of the heavy key in the feed,

$x_{f,\mathrm{LK}}$ = concentration of the light key in the feed,

$x_{d,\mathrm{HK}}$ = concentration of the heavy key in the top product,

$x_{b,\mathrm{LK}}$ = concentration of the light key if in the bottom product.

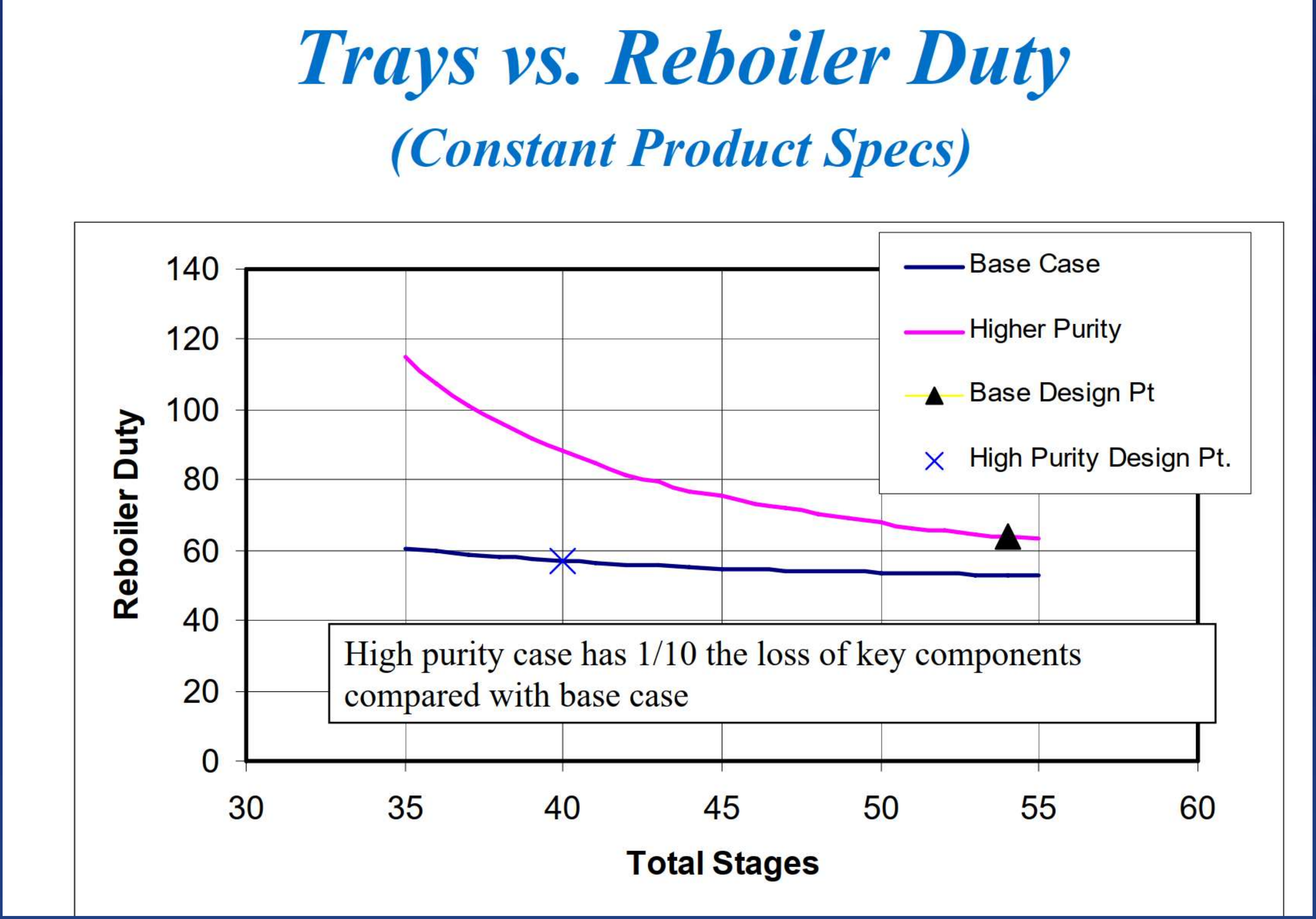
Basic Principles

Optimize Column Design

- Best Trays vs Reboiler
  - More trays – smaller reboiler and condenser
  - More capital – smaller utility cost
  - Economic analysis with help from simple guidelines
- Best Feed Tray Location

Basic Principles

Optimize Column Design



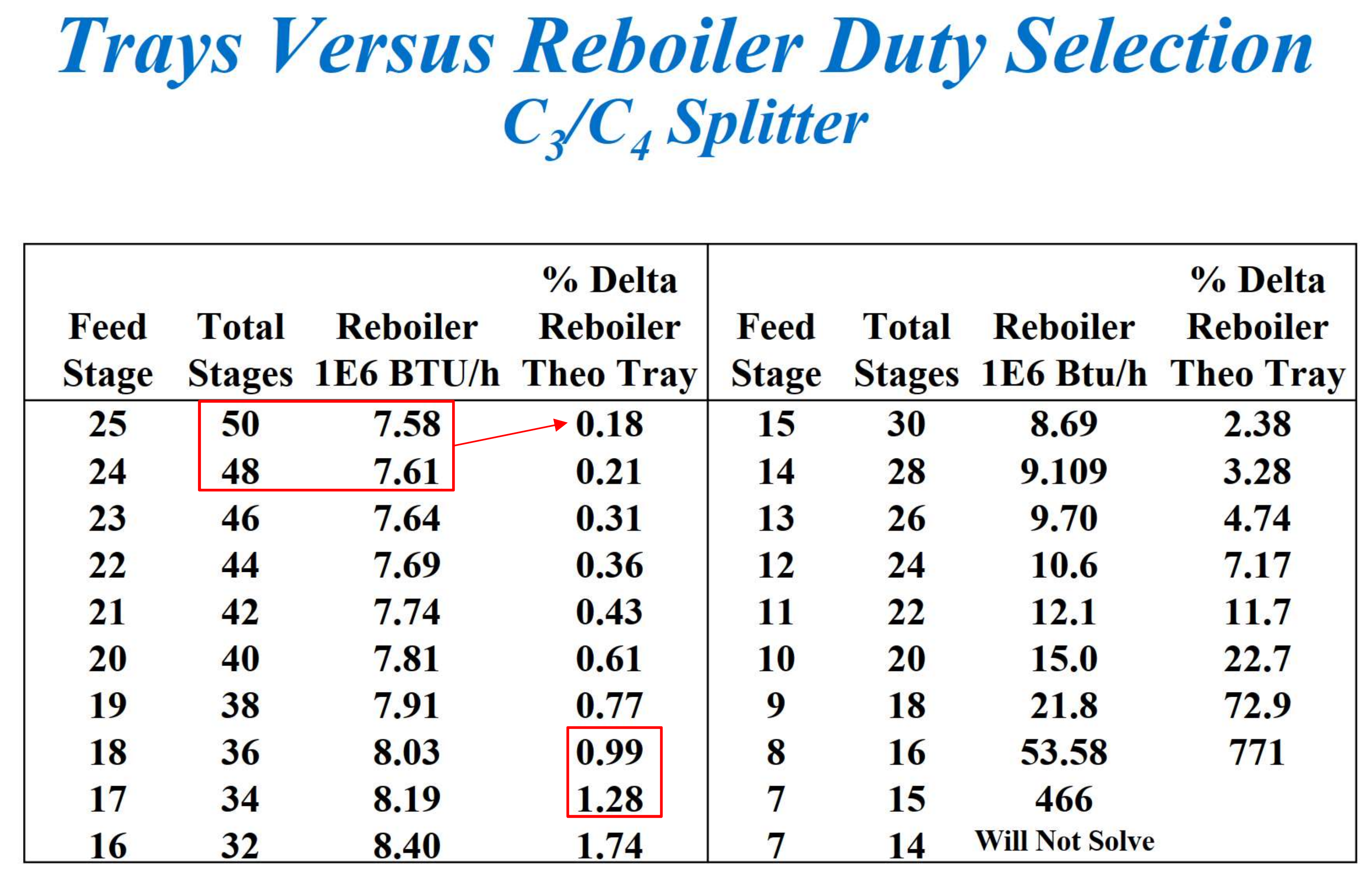
Trays vs. Reboiler Duty

(Constant Product Specs)

High purity case has 1/10 the loss of key components compared with base case

Basic Principles

Optimize Column Design



Trays Versus Reboiler Duty Selection

C_3/C_4 Splitter

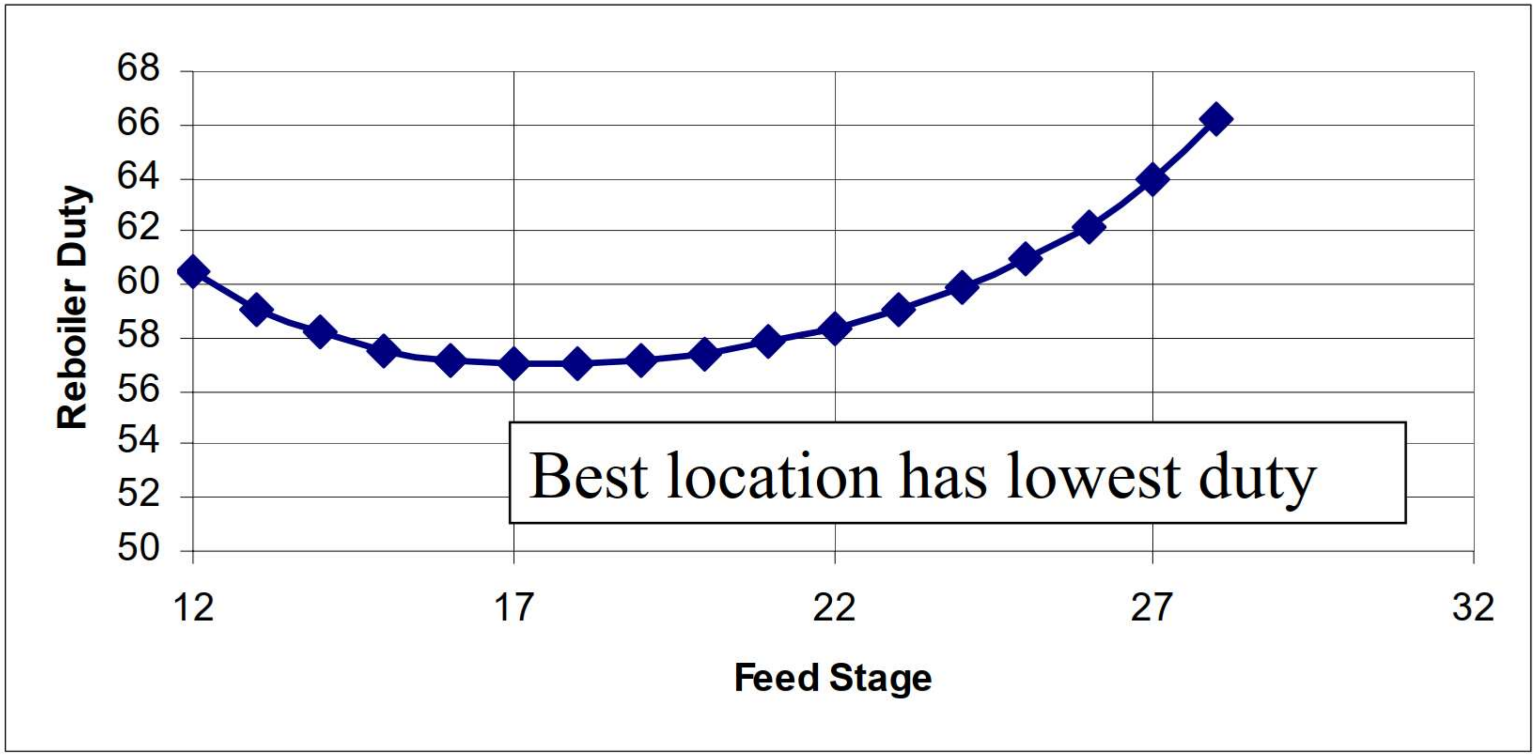
Feed Stage	Total Stages	Reboiler 1E6 BTU/h	% Delta Reboiler Theo Tray	Feed Stage	Total Stages	Reboiler 1E6 Btu/h	% Delta Reboiler Theo Tray
25	50	7.58	0.18	15	30	8.69	2.38
24	48	7.61	0.21	14	28	9.109	3.28
23	46	7.64	0.31	13	26	9.70	4.74
22	44	7.69	0.36	12	24	10.6	7.17
21	42	7.74	0.43	11	22	12.1	11.7
20	40	7.81	0.61	10	20	15.0	22.7
19	38	7.91	0.77	9	18	21.8	72.9
18	36	8.03	0.99	8	16	53.58	771
17	34	8.19	1.28	7	15	466	
16	32	8.40	1.74	7	14	Will Not Solve	

Basic Principles

Optimize Column Design

Feed Tray Location vs. Reboiler Duty

(Constant Product Specs)

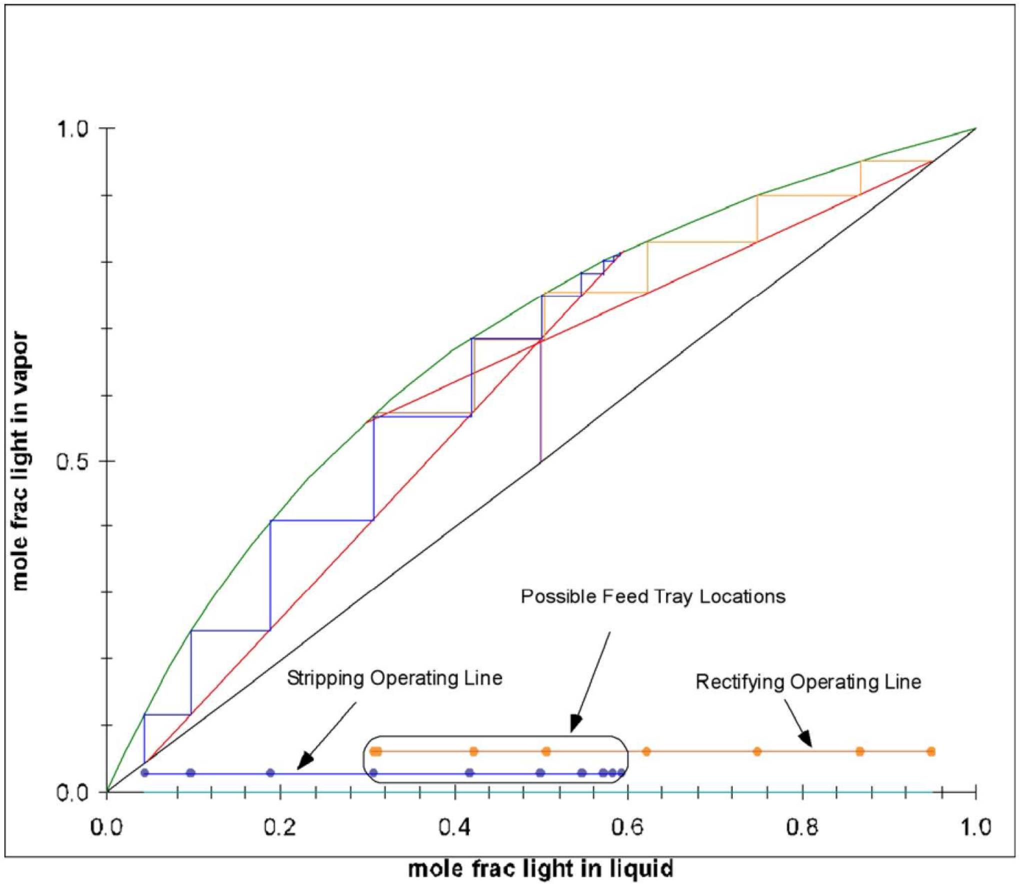


Best location has lowest duty

Basic Principles

Optimize Column Design

Feed Tray Location



- **Rooks, R.E., Chemical Processing, May 2006**

Basic Principles

Multi-Component Distillation : Rigorous Calculation

- The basic steps in any rigorous solution procedure will be:

1. Specification of the problem; complete specification is essential for computer methods.

2. Selection of values for the iteration variables; for example, estimated stage temperatures, and liquid and vapor flows (the column temperature and flow profiles).

3. A calculation procedure for the solution of the stage equations.

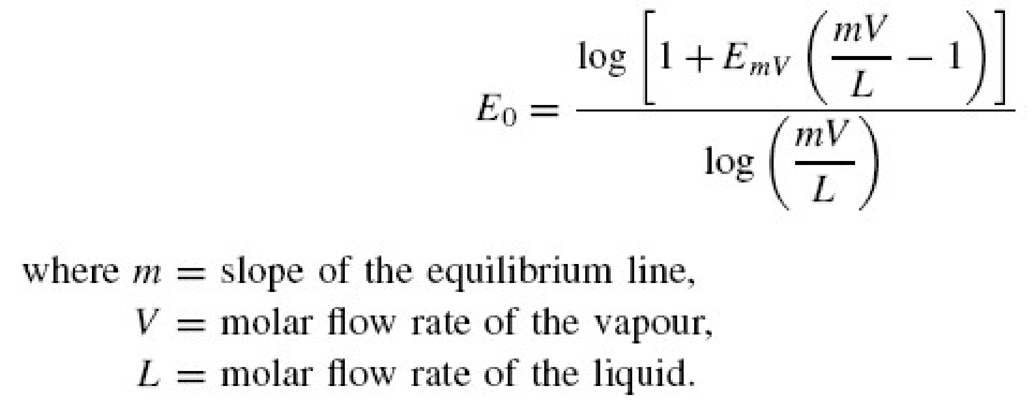
4. A procedure for the selection of new values for the iteration variables for each set of trial calculations.

5. A procedure to test for convergence; to check if a satisfactory solution has been achieved.

Basic Principles

Tray Efficiency

- Murphree tray efficiency $E_{mV} = \frac{y_n - y_{n-1}}{y_e - y_{n-1}}$
- For the idealised situation where the operating and equilibrium lines are straight, the overall column efficiency and the Murphree tray efficiency are related by an equation derived by Lewis (1936):



$$E_0 = \frac{\log\left[1 + E_{mV}\left(\frac{mV}{L} - 1\right)\right]}{\log\left(\frac{mV}{L}\right)}$$

where m = slope of the equilibrium line,
V = molar flow rate of the vapour,
L = molar flow rate of the liquid.

- Whenever possible the tray efficiencies used in design should be based on experimental values for similar systems, obtained on full-sized columns.

Basic Principles

Tray Efficiency

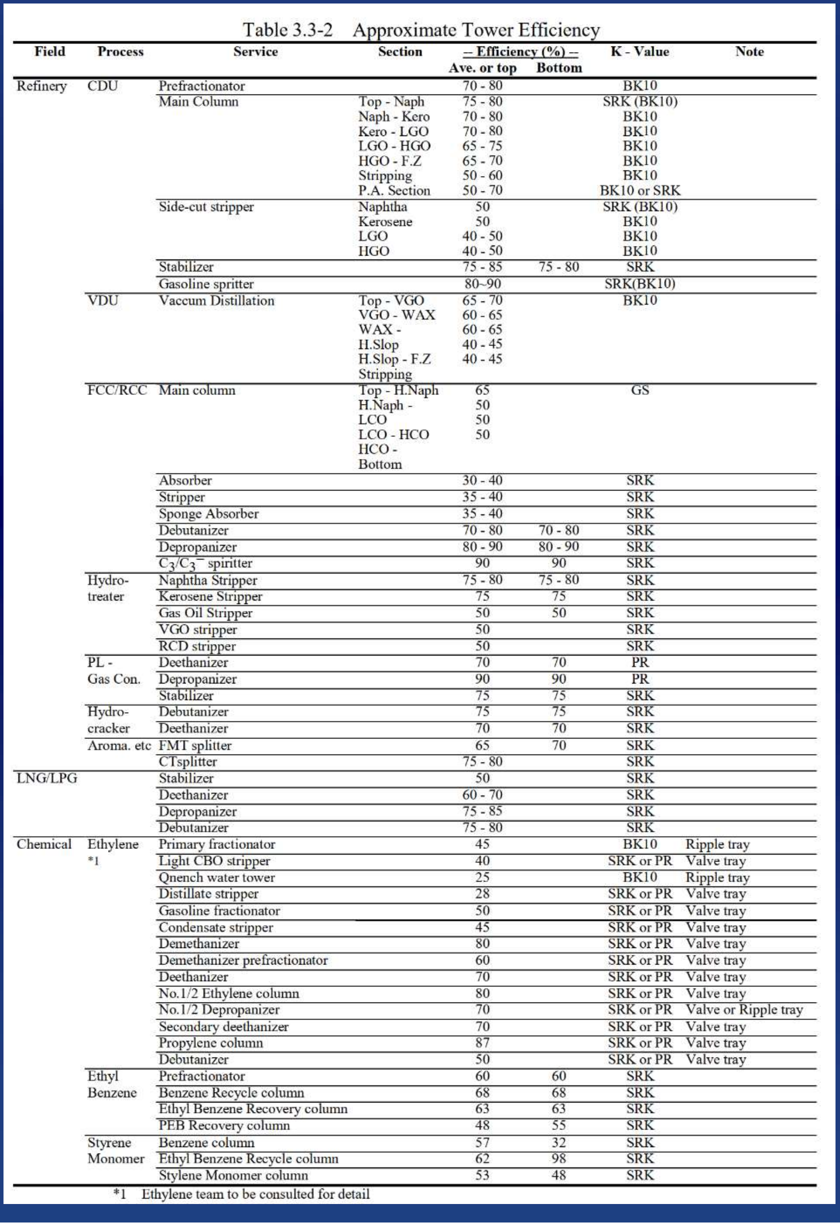


Table 3.3-2 Approximate Tower Efficiency

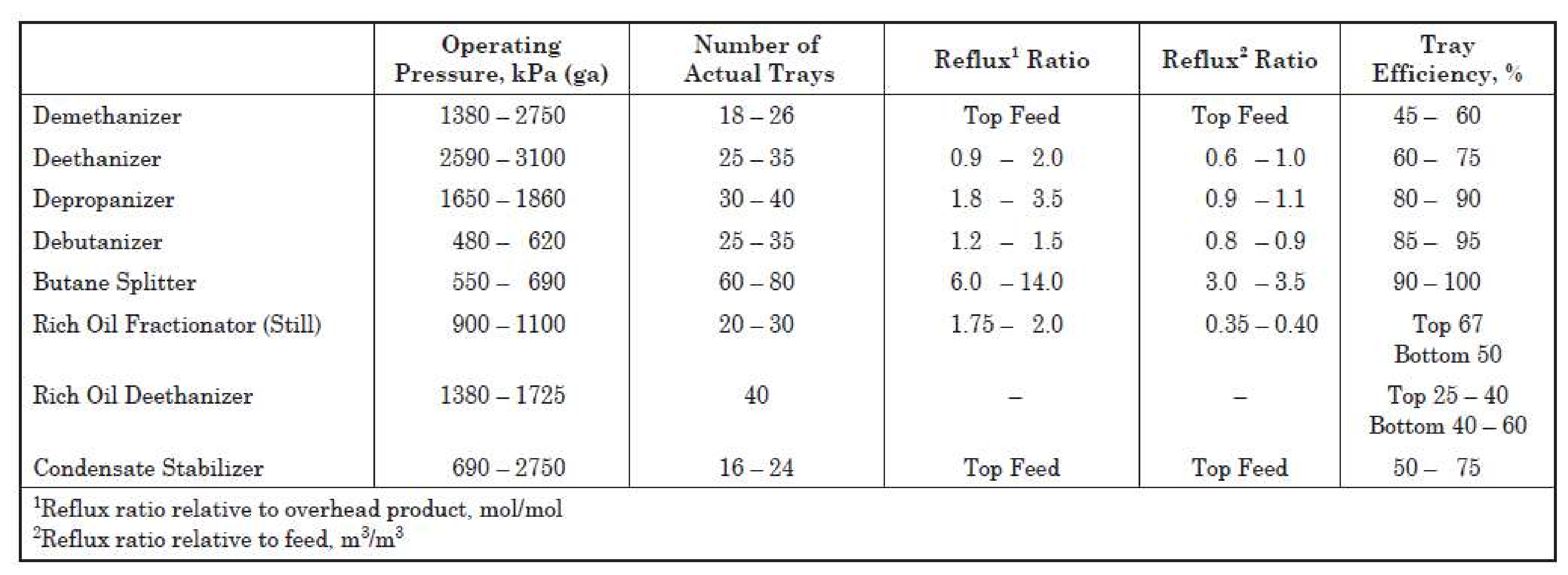
Field	Process	Service	Section	– Efficiency (%) –		K - Value	Note
				Ave. or top	Bottom		
Refinery	CDU	Prefractionator		70 - 80		BK10	
		Main Column	Top - Naph	75 - 80		SRK (BK10)	
			Naph - Kero	70 - 80		BK10	
			Kero - LGO	70 - 80		BK10	
			LGO - HGO	65 - 75		BK10	
			HGO - F.Z	65 - 70		BK10	
			Stripping	50 - 60		BK10	
			P.A. Section	50 - 70		BK10 or SRK	
		Side-cut stripper	Naphtha	50		SRK (BK10)	
			Kerosene	50		BK10	
			LGO	40 - 50		BK10	
			HGO	40 - 50		BK10	
		Stabilizer		75 - 85	75 - 80	SRK	
		Gasoline spritter		80~90		SRK(BK10)	
	VDU	Vaccum Distillation	Top - VGO	65 - 70		BK10	
			VGO - WAX	60 - 65			
			WAX -	60 - 65			
			H.Slop	40 - 45			
			H.Slop - F.Z	40 - 45			
			Stripping				
	FCC/RCC	Main column	Top - H.Naph	65		GS	
			H.Naph -	50			
			LCO	50			
			LCO - HCO	50			
			HCO -				
			Bottom				
		Absorber		30 - 40		SRK	
		Stripper		35 - 40		SRK	
		Sponge Absorber		35 - 40		SRK	
		Debutanizer		70 - 80	70 - 80	SRK	
		Depropanizer		80 - 90	80 - 90	SRK	
		C_3/C_3^- spiritter		90	90	SRK	
	Hydro-treater	Naphtha Stripper		75 - 80	75 - 80	SRK	
		Kerosene Stripper		75	75	SRK	
		Gas Oil Stripper		50	50	SRK	
		VGO stripper		50		SRK	
		RCD stripper		50		SRK	
	PL - Gas Con.	Deethanizer		70	70	PR	
		Depropanizer		90	90	PR	
		Stabilizer		75	75	SRK	
	Hydro-cracker	Debutanizer		75	75	SRK	
		Deethanizer		70	70	SRK	
	Aroma. etc	FMT splitter		65	70	SRK	
		CTsplitter		75 - 80		SRK	
LNG/LPG		Stabilizer		50		SRK	
		Deethanizer		60 - 70		SRK	
		Depropanizer		75 - 85		SRK	
		Debutanizer		75 - 80		SRK	
Chemical	Ethylene *1	Primary fractionator		45		BK10	Ripple tray
		Light CBO stripper		40		SRK or PR	Valve tray
		Qnench water tower		25		BK10	Ripple tray
		Distillate stripper		28		SRK or PR	Valve tray
		Gasoline fractionator		50		SRK or PR	Valve tray
		Condensate stripper		45		SRK or PR	Valve tray
		Demethanizer		80		SRK or PR	Valve tray
		Demethanizer prefractionator		60		SRK or PR	Valve tray
		Deethanizer		70		SRK or PR	Valve tray
		No.1/2 Ethylene column		80		SRK or PR	Valve tray
		No.1/2 Depropanizer		70		SRK or PR	Valve or Ripple tray
		Secondary deethanizer		70		SRK or PR	Valve tray
		Propylene column		87		SRK or PR	Valve tray
		Debutanizer		50		SRK or PR	Valve tray
	Ethyl Benzene	Prefractionator		60	60	SRK	
		Benzene Recycle column		68	68	SRK	
		Ethyl Benzene Recovery column		63	63	SRK	
		PEB Recovery column		48	55	SRK	
	Styrene Monomer	Benzene column		57	32	SRK	
		Ethyl Benzene Recycle column		62	98	SRK	
		Stylene Monomer column		53	48	SRK	

*1 Ethylene team to be consulted for detail

Basic Principles

Tray Efficiency

- GPSA



	Operating Pressure, kPa (ga)	Number of Actual Trays	Reflux[1] Ratio	Reflux[2] Ratio	Tray Efficiency, %
Demethanizer	1380 – 2750	18 – 26	Top Feed	Top Feed	45 – 60
Deethanizer	2590 – 3100	25 – 35	0.9 – 2.0	0.6 – 1.0	60 – 75
Depropanizer	1650 – 1860	30 – 40	1.8 – 3.5	0.9 – 1.1	80 – 90
Debutanizer	480 – 620	25 – 35	1.2 – 1.5	0.8 – 0.9	85 – 95
Butane Splitter	550 – 690	60 – 80	6.0 – 14.0	3.0 – 3.5	90 – 100
Rich Oil Fractionator (Still)	900 – 1100	20 – 30	1.75 – 2.0	0.35 – 0.40	Top 67 Bottom 50
Rich Oil Deethanizer	1380 – 1725	40	–	–	Top 25 – 40 Bottom 40 – 60
Condensate Stabilizer	690 – 2750	16 – 24	Top Feed	Top Feed	50 – 75

[1] Reflux ratio relative to overhead product, mol/mol

[2] Reflux ratio relative to feed, m^3/m^3

Basic Principles

Tray Efficiency

- Tray, and overall column, efficiencies will normally be between 30 % and 70 %, and as a rough guide a figure of 50 % can be assumed for preliminary designs.

- Correlations have been proposed to define the overall efficiency empirically. O'Connell correlation (1946) was developed for bubble cap trays and its application to other tray technologies is conservative:

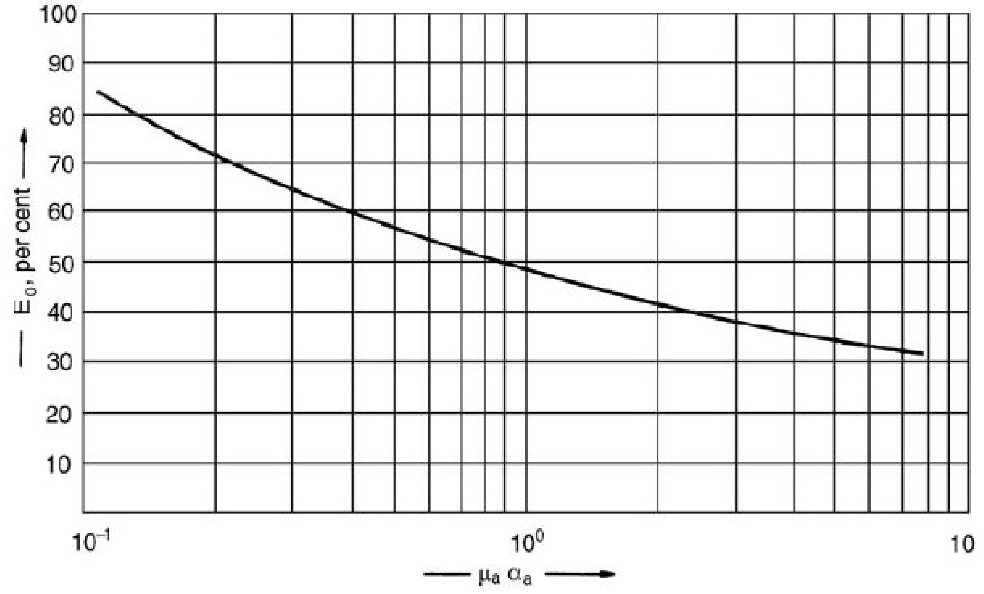


Figure 11.13. Distillation column efficiencies (bubble-caps) (after O'Connell, 1946)

$$E_{overall} = 0.492\,(\mu_L\,\alpha)^{-0.245}$$

with:

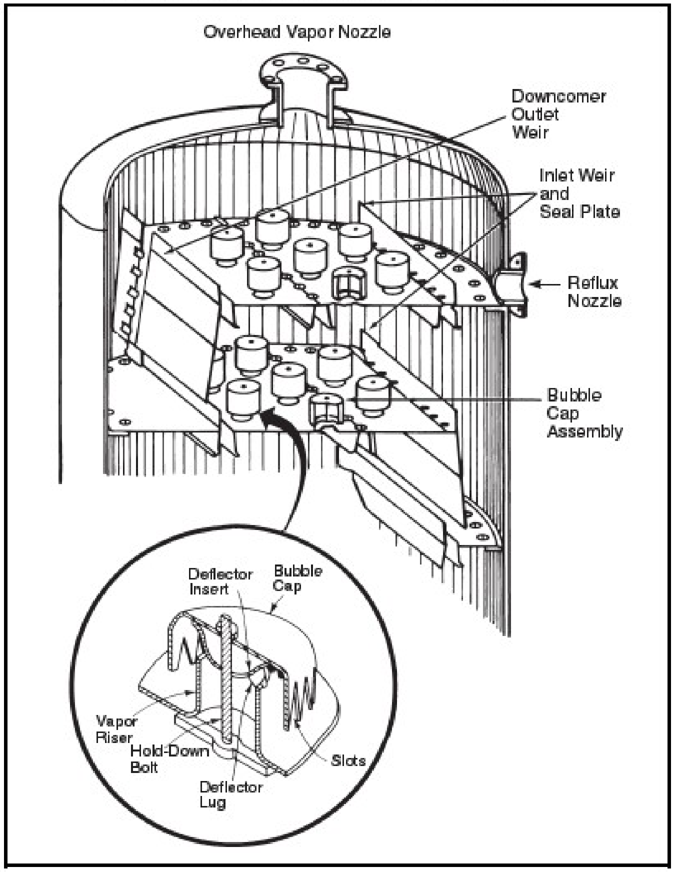
μ_L viscosity of the liquid (mPa.s)
α relative volatility of the two key components

Column Sizing

An approximate estimate of the overall column size can be made once the number of real stages required for the separation is known. This is often needed to make a rough estimate of the capital cost for project evaluation.

Internals

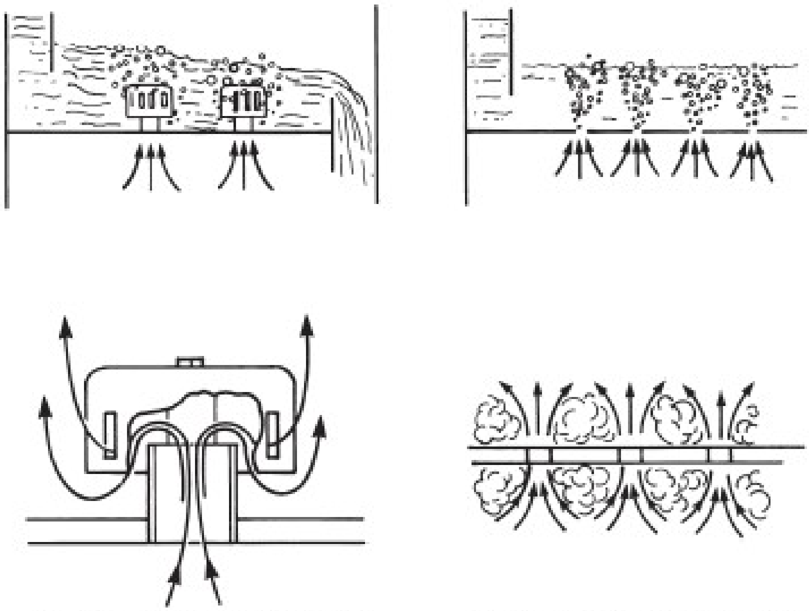
Top Two Trays of a Bubble-cap Column[28]



Column Sizing

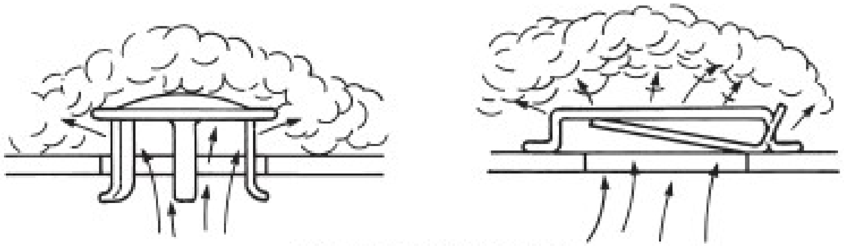
Internals

Flow Through Vapor Passages[28]



(a) Vapor flow through bubble cap

(b) Vapor flow through perforations



(c) Vapor flow through valves

Column Sizing

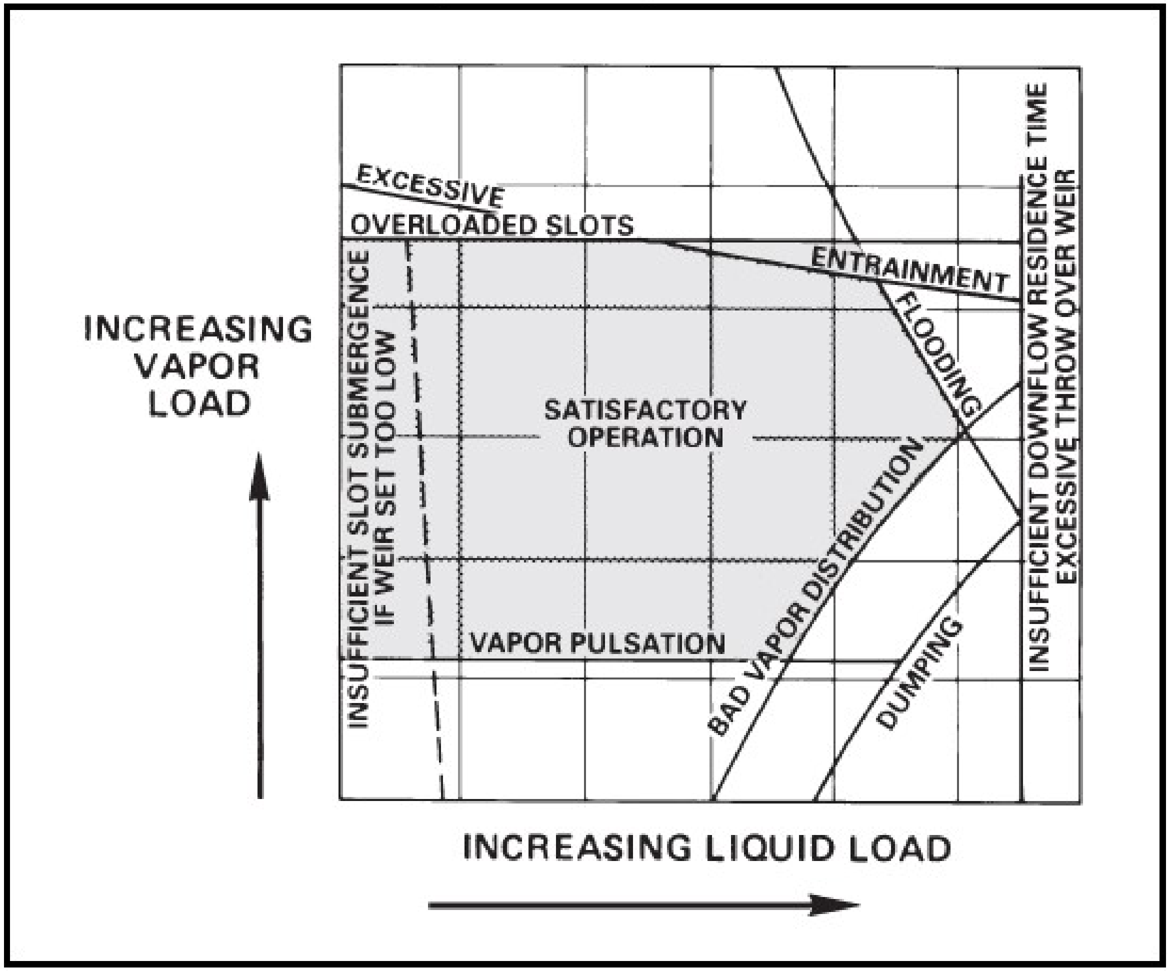
Internals

- Due to the riser in the bubble cap, it is the only tray which can be designed to prevent liquid from "weeping" through the vapor passage.
- Sieve or valve trays control weeping by vapor velocity.
- The bubble cap tray has the highest turndown ratio, with designs of 8:1 to 10:1 ratio being common. Bubble cap trays are almost always used in glycol dehydration columns.
- Valve and sieve trays are popular due to the lower cost and increased capacity over bubble cap trays for a given tower diameter.
- Although the sieve tray generally has higher capacity, its main disadvantage is that sieve trays will be susceptible to "weeping" or "dumping" of the liquid through the holes at low vapor rates and its turndown capacity is limited.

Column Sizing

Internals

Limits of Satisfactory Tray Operation for a Specific Set of Tray Fluid Properties[8]



Column Sizing

Internals

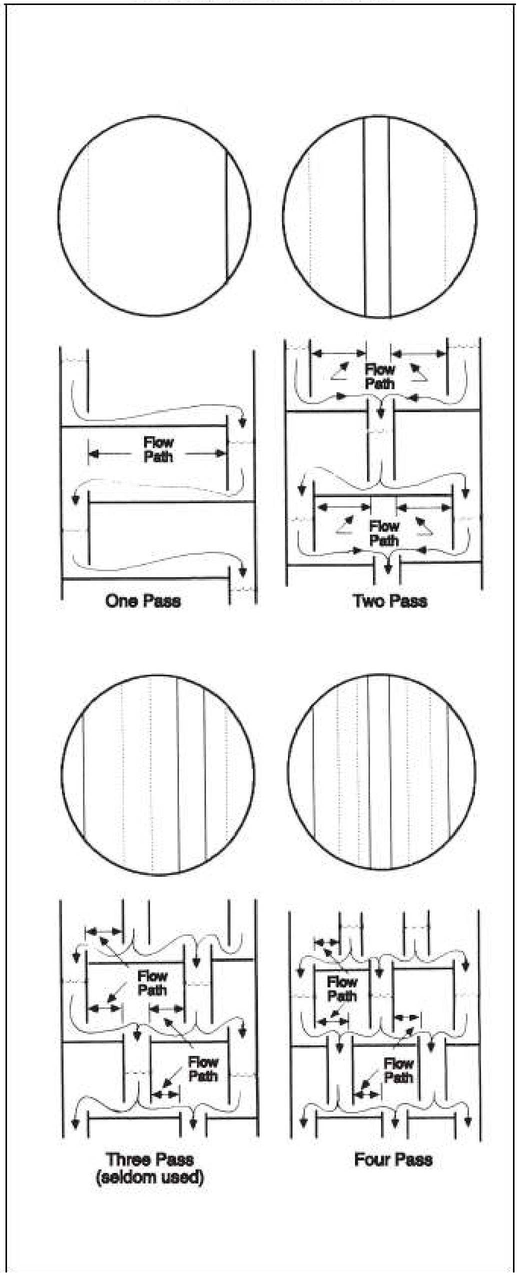
Flooding which occurs for both vapor and liquid:

- **Blowing.** occurs when vapor flow rates are too high in relation to the liquid flow. A finely dispersed mist is formed above the contact zone. This results in a poor liquid-vapor transfer with the formation of froth and fine droplets entrained to the trays overhead.
- **Jet flooding.** The simultaneous increase in liquid and vapor flow rates, common when maximum capacity is desired, is similar to the preceding case. More liquid is then entrained to the tray above due to the increased liquid height on the tray, thereby reducing the vapor disengagement height. Operation becomes unstable.
- **Flooding by excessive liquid flow.** The aerated liquid flow coming from the active area and passing into the downcomer is too high, causing flooding. This may be due to improper vapor disengagement (foaming system or excessively small downcomer area), too much vapor pressure drop, too little space under the downcome; inadequate tray spacing.

Column Sizing

Passes

Alternative Liquid Flow Paths



One Pass

Two Pass

Three Pass
(seldom used)

Four Pass

Column Sizing

Selection of Tray Type

- The principal factors to consider when comparing the performance of bubble-cap, sieve and valve trays are: cost, capacity, operating range, efficiency and pressure drop.
- *Cost:* Bubble-cap tray are appreciably more expensive than sieve or valve trays. The relative cost will depend on the material of construction used; for mild steel the ratios, bubble-cap : valve : sieve, are approximately 3.0 : 1.5 : 1.0.
- *Capacity:* There is little difference in the capacity rating of the three types (the diameter of the column required for a given flow-rate); the ranking is sieve, valve, bubble-cap.

Column Sizing

Selection of Tray Type

- *Operating range. This is the most significant factor. By operating range is meant the* range of vapor and liquid rates over which the tray will operate satisfactorily (the stable operating range).

  Some flexibility will always be required in an operating plant to allow for changes in production rate, and to cover start-up and shut-down conditions.

  The ratio of the highest to the lowest flow rates is often referred to as the “turn-down” ratio. Bubble-cap trays have a positive liquid seal and can therefore operate efficiently at very low vapor rates.

  Sieve trays rely on the flow of vapor through the holes to hold the liquid on the tray, and cannot operate at very low vapor rates. But, with good design, sieve trays can be designed to give a satisfactory operating range; typically, from 50 per cent to 120 percent of design capacity.

Column Sizing

Selection of Tray Type

- *Pressure drop.* The pressure drop over the trays can be an important design consideration, particularly for vacuum columns.

  The tray pressure drop will depend on the detailed design of the tray but, in general, sieve trays give the lowest pressure drop, followed by valves, with bubble-caps giving the highest.

- *Summary.* Sieve trays are the cheapest and are satisfactory for most applications.

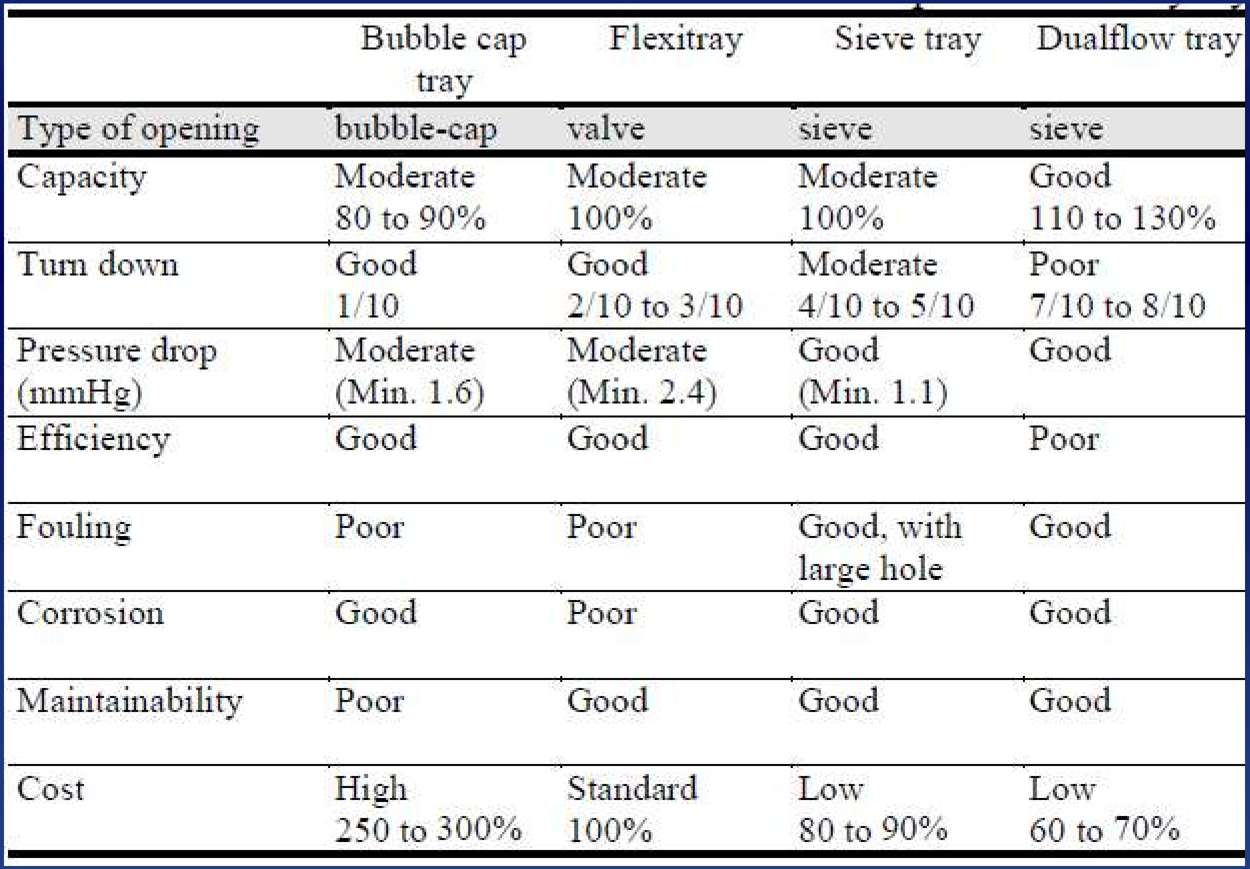
  Valve trays should be considered if the specified turn-down ratio cannot be met with sieve trays.

  Bubble-caps should only be used where very low vapor (gas) rates have to be handled and a positive liquid seal is essential at all flow-rates.

Column Sizing

Selection of Tray Type

- *Summary*



	Bubble cap tray	Flexitray	Sieve tray	Dualflow tray
Type of opening	bubble-cap	valve	sieve	sieve
Capacity	Moderate 80 to 90%	Moderate 100%	Moderate 100%	Good 110 to 130%
Turn down	Good 1/10	Good 2/10 to 3/10	Moderate 4/10 to 5/10	Poor 7/10 to 8/10
Pressure drop (mmHg)	Moderate (Min. 1.6)	Moderate (Min. 2.4)	Good (Min. 1.1)	Good
Efficiency	Good	Good	Good	Poor
Fouling	Poor	Poor	Good, with large hole	Good
Corrosion	Good	Poor	Good	Good
Maintainability	Poor	Good	Good	Good
Cost	High 250 to 300%	Standard 100%	Low 80 to 90%	Low 60 to 70%

Column Sizing

Tray Spacing

- Spacing is generally assumed to be 45 cm for diameters of less than 1.2 m and 60 or 75 cm for more than 1.2 m. Spacing of 90 cm is only seldom necessary in zones where there is a high liquid flow rate (circulating reflux zone, for example).

- The spacing choice can be modified locally due to the presence of manholes required for inspection.

Column Sizing

Downcomers

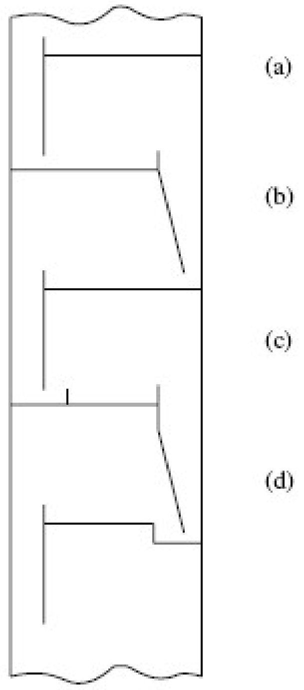


Figure 11.24. Segment (chord) downcomer designs. (*a*) Vertical apron (*b*) Inclined apron (*c*) Inlet weir (*d*) Recessed well

- The segmental, or chord downcomer is the simplest and cheapest form of construction and is satisfactory for most purposes.

Column Sizing

Downcomers

- The apron is usually vertical, but may be sloped to increase the tray area available for perforation.
- If a more positive seal is required at the downcomer at the outlet, an inlet weir can be fitted or a recessed seal pan used. Circular downcomers (pipes) are sometimes used for small liquid flow-rates.

Column Sizing

Tray Column Sizing

1. "C" factor method

- This very simple method uses the Souders and Brown equation, which gives the maximum acceptable vapor velocity below one tray to prevent excessive entrainment of liquid from this tray to the tray above

$$\vartheta_m = C\sqrt{\frac{\rho_l - \rho_v}{\rho_v}}$$

Vm = maximum acceptable vapor velocity in the space below one tray, (m/h),

ρl = liquid density at operating temp. and pressure of the tray (kg/m3),

ρv = vapor density at operating temp. and pressure of the tray (kg/m3),

C = Souders-Brown factor given by figure 12, in m/h versus tray spacing in cm and liquid surface tension in N/m,

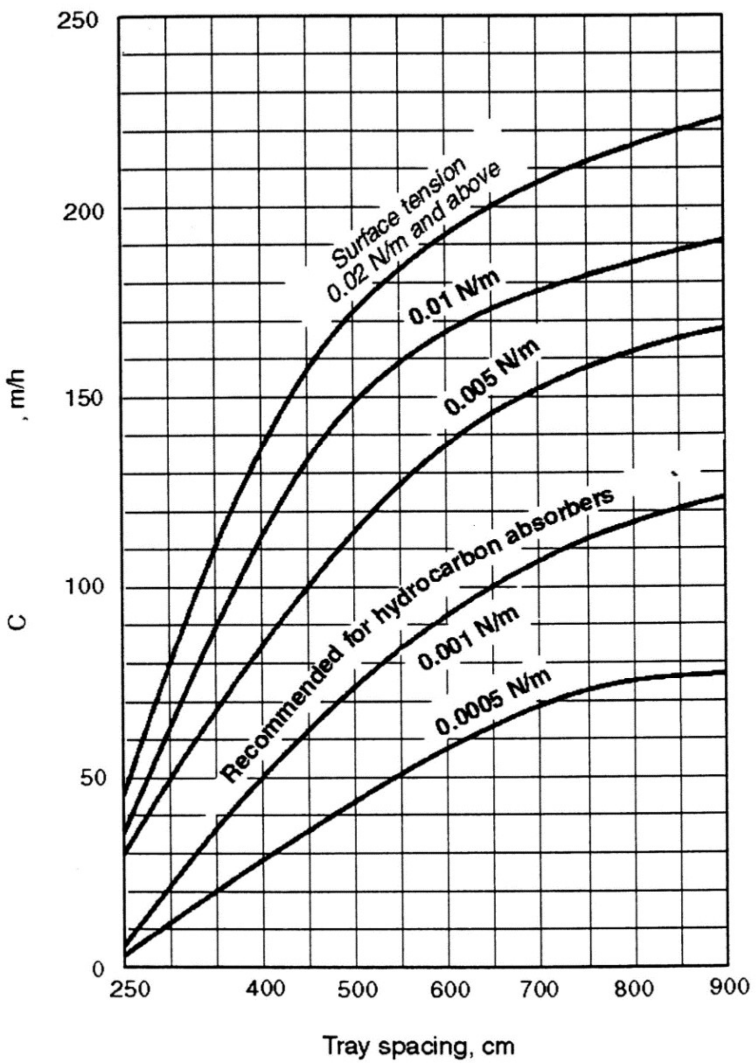
- Usual tray spacing: 18 in (46 cm), 24 in (61 cm), 30 in (76 cm). 24 in (61 cm) is the most frequent particularly for glycol and amine absorbers. Actually, this spacing also depends on downcomer design.

Column Sizing

Tray Column Sizing

1. "C" factor method

Figure 12 - SOUDERS-BROWN FACTOR FOR APPROXIMATE COLUMN SIZING

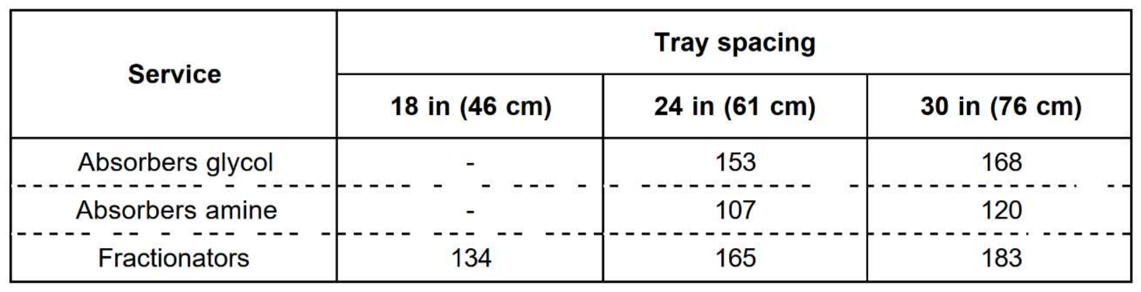


Column Sizing

Tray Column Sizing

1. "C" factor method

- Instead of figure 12, "C" factor values given by table as follows (from Campbell – gas Conditioning and Processing) can also be used.



Service	Tray spacing		
	18 in (46 cm)	**24 in (61 cm)**	**30 in (76 cm)**
Absorbers glycol	-	153	168
Absorbers amine	-	107	120
Fractionators	134	165	183

- Manufacturers provide specific capacity factors for each proprietary use.
- The column diameter of the column is given by equation as follows:

$$D = \sqrt{\frac{Q}{(0.7854)\vartheta_m}}$$

D = inside diameter of the column in meters,

Q = vapor flowrate at actual tray conditions (m3/h),

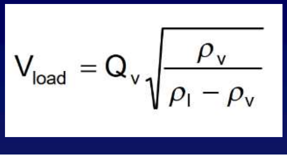
- This method was originally developed for bubble cap trays and gives a rough diameter value, especially for other types of tray.

Column Sizing

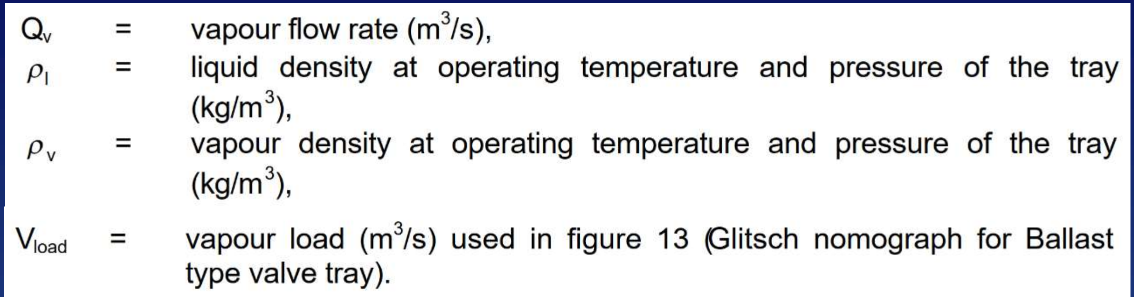
Tray Column Sizing

2. Nomograph method for valve trays

- Manufacturers of valve trays have developed design methods for their trays. Design procedures are made available for preliminary studies. Figure 13 is an example of such nomograph method which gives by simple reading tray diameter with number of pass by tray. It requires the knowledge for each tray of the liquid flow rate in m3/min and the vapour load determined with equation as follows:



$$V_{load} = Q_v \sqrt{\frac{\rho_v}{\rho_l - \rho_v}}$$



Q_v = vapour flow rate (m^3/s),

ρ_l = liquid density at operating temperature and pressure of the tray (kg/m^3),

ρ_v = vapour density at operating temperature and pressure of the tray (kg/m^3),

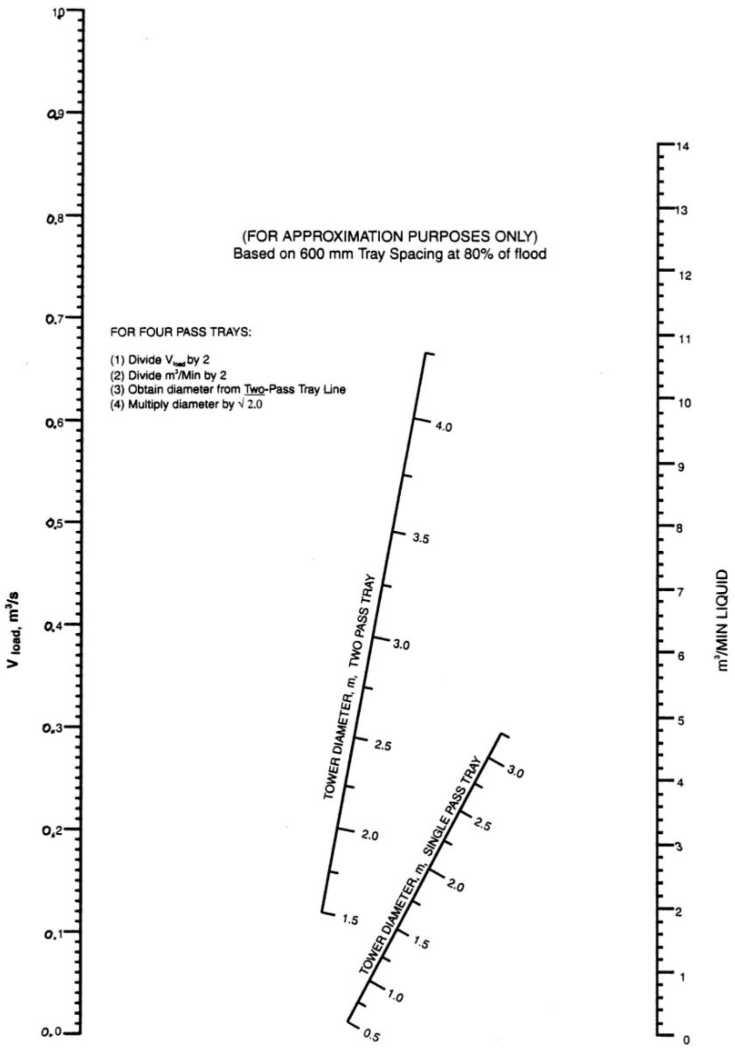
V_{load} = vapour load (m^3/s) used in figure 13 (Glitsch nomograph for Ballast type valve tray).

Column Sizing

Tray Column Sizing

2. Nomograph method for valve trays

Figure 13 - VALVE TRAY DIAMETER BY NOMOGRAPH METHOD



(FOR APPROXIMATION PURPOSES ONLY)
Based on 600 mm Tray Spacing at 80% of flood

FOR FOUR PASS TRAYS:

(1) Divide V_{load} by 2
(2) Divide m^3/Min by 2
(3) Obtain diameter from Two-Pass Tray Line
(4) Multiply diameter by $\sqrt{2.0}$

Column Sizing

Tray Column Sizing

3. Detailed method for valve trays

- Previous nomograph method corresponds to a first approach. It does not take account of foaming which is the source of major problems in many systems.

- The Glitsch manual method for Ballast type valve tray gives results with the following steps:

1st step: Determination of the flow path length (FPL)

- An approximate flow path length is useful for establishing the minimum column diameter.
- With the values of the diameter (DT) and the number of pass (NP) determined with the nomograph method (Figure 13), calculate the flow path length (FPL) with equation as follows:

$$(FPL) = 0.75\,\frac{(DT)}{(NP)}$$

(FPL) = flow path length (m),
(DT) = internal diameter of the column (m),
(NP) = number of pass

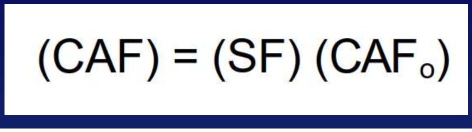
Column Sizing

Tray Column Sizing

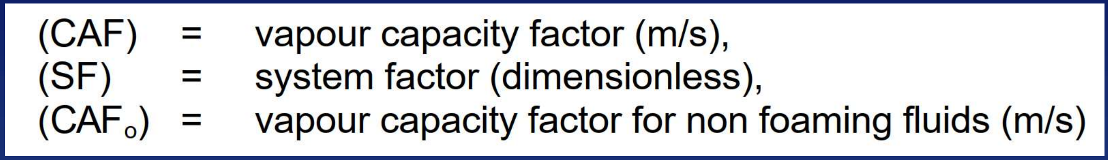
3. Detailed method for valve trays

2nd step: Determination of Vapor Capacity Factor (CAF)

- Figure 14 allows to determine the vapour capacity factor (CAFo) in meter per second, versus vapor density and tray spacing for non-foaming fluids.
- For foaming fluids this vapour capacity factor must be corrected by the system factor value indicated in the table of figure 15.



$$(CAF) = (SF)\,(CAF_o)$$



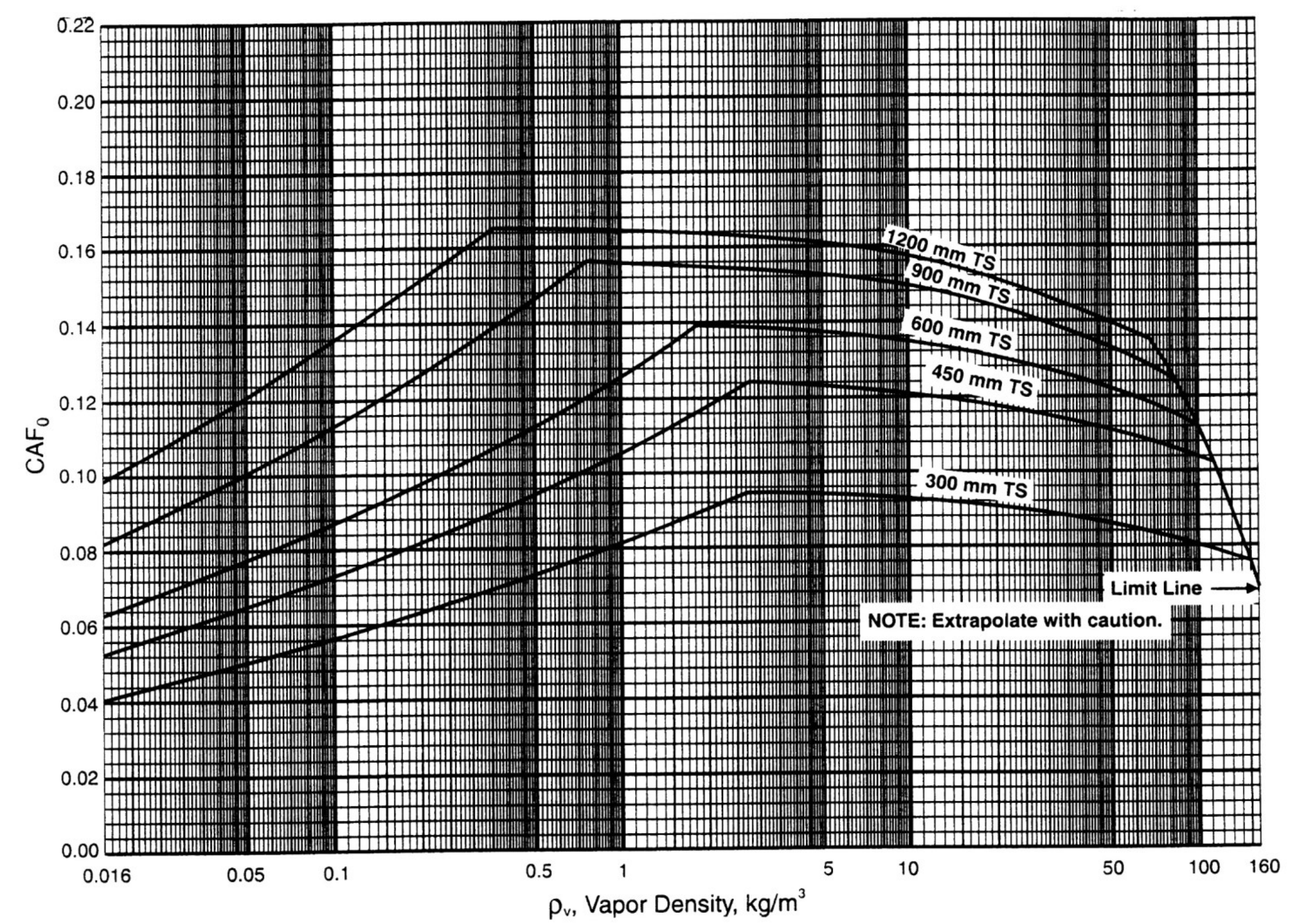
(CAF) = vapour capacity factor (m/s),
(SF) = system factor (dimensionless),
(CAF_o) = vapour capacity factor for non foaming fluids (m/s)

Column Sizing

Tray Column Sizing

3. Detailed method for valve trays

Figure 14 - APPROXIMATE FLOOD CAPACITY OF VALVE TRAYS

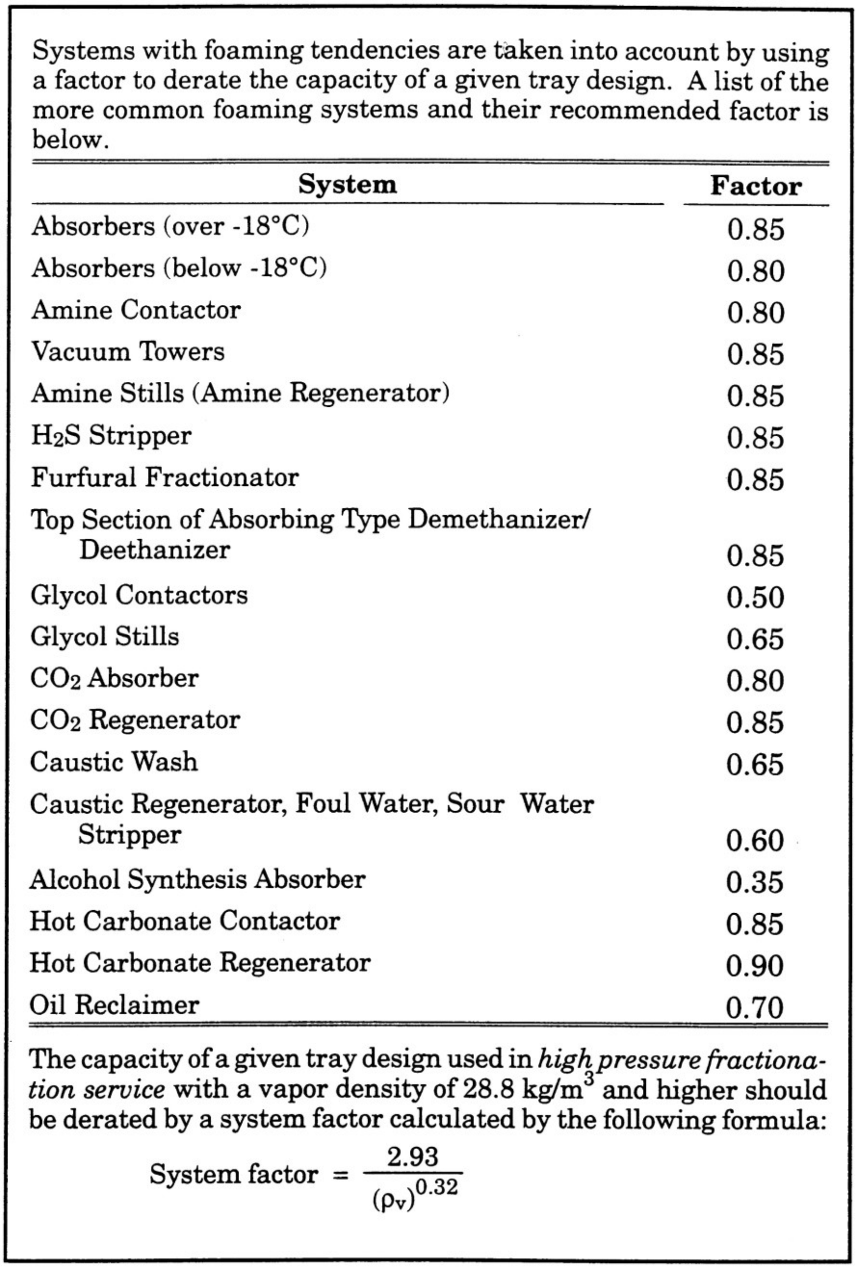


Column Sizing

Tray Column Sizing

3. Detailed method for valve trays

Figure 15 - SYSTEM FACTORS (SF)



Systems with foaming tendencies are taken into account by using a factor to derate the capacity of a given tray design. A list of the more common foaming systems and their recommended factor is below.

System	Factor
Absorbers (over -18°C)	0.85
Absorbers (below -18°C)	0.80
Amine Contactor	0.80
Vacuum Towers	0.85
Amine Stills (Amine Regenerator)	0.85
H_2S Stripper	0.85
Furfural Fractionator	0.85
Top Section of Absorbing Type Demethanizer/ Deethanizer	0.85
Glycol Contactors	0.50
Glycol Stills	0.65
CO_2 Absorber	0.80
CO_2 Regenerator	0.85
Caustic Wash	0.65
Caustic Regenerator, Foul Water, Sour Water Stripper	0.60
Alcohol Synthesis Absorber	0.35
Hot Carbonate Contactor	0.85
Hot Carbonate Regenerator	0.90
Oil Reclaimer	0.70

The capacity of a given tray design used in *high pressure fractionation service* with a vapor density of 28.8 kg/m^3 and higher should be derated by a system factor calculated by the following formula:

$$\text{System factor} = \frac{2.93}{(\rho_v)^{0.32}}$$

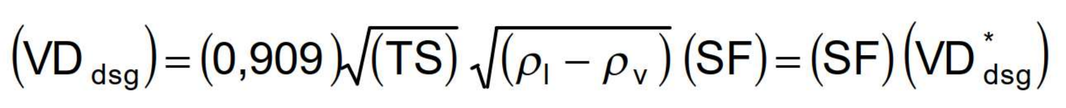
Column Sizing

Tray Column Sizing

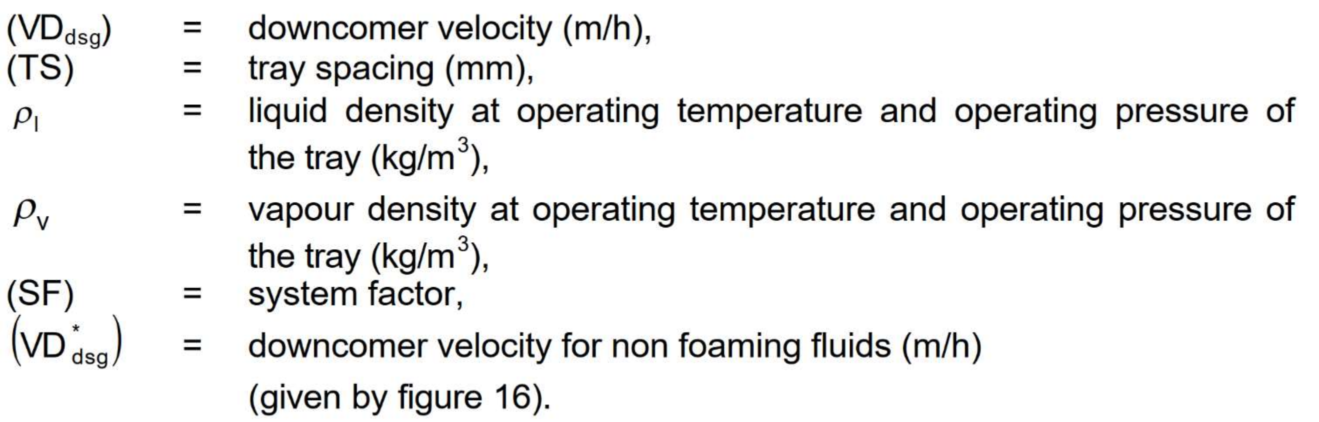
3. Detailed method for valve trays

3rd step: Determination of the Downcomer Velocity (VDdsg)

- The procedure used in this method for establishing downcomer area is based on a "design“ velocity in meter per hour given by figure 16 for non foaming fluid or by equation as follows:



$$\left(VD_{dsg}\right) = (0{,}909)\sqrt{(TS)}\sqrt{(\rho_l - \rho_v)}\,(SF) = (SF)\left(VD^{*}_{dsg}\right)$$



(VD_{dsg}) = downcomer velocity (m/h),

(TS) = tray spacing (mm),

ρ_l = liquid density at operating temperature and operating pressure of the tray (kg/m^3),

ρ_v = vapour density at operating temperature and operating pressure of the tray (kg/m^3),

(SF) = system factor,

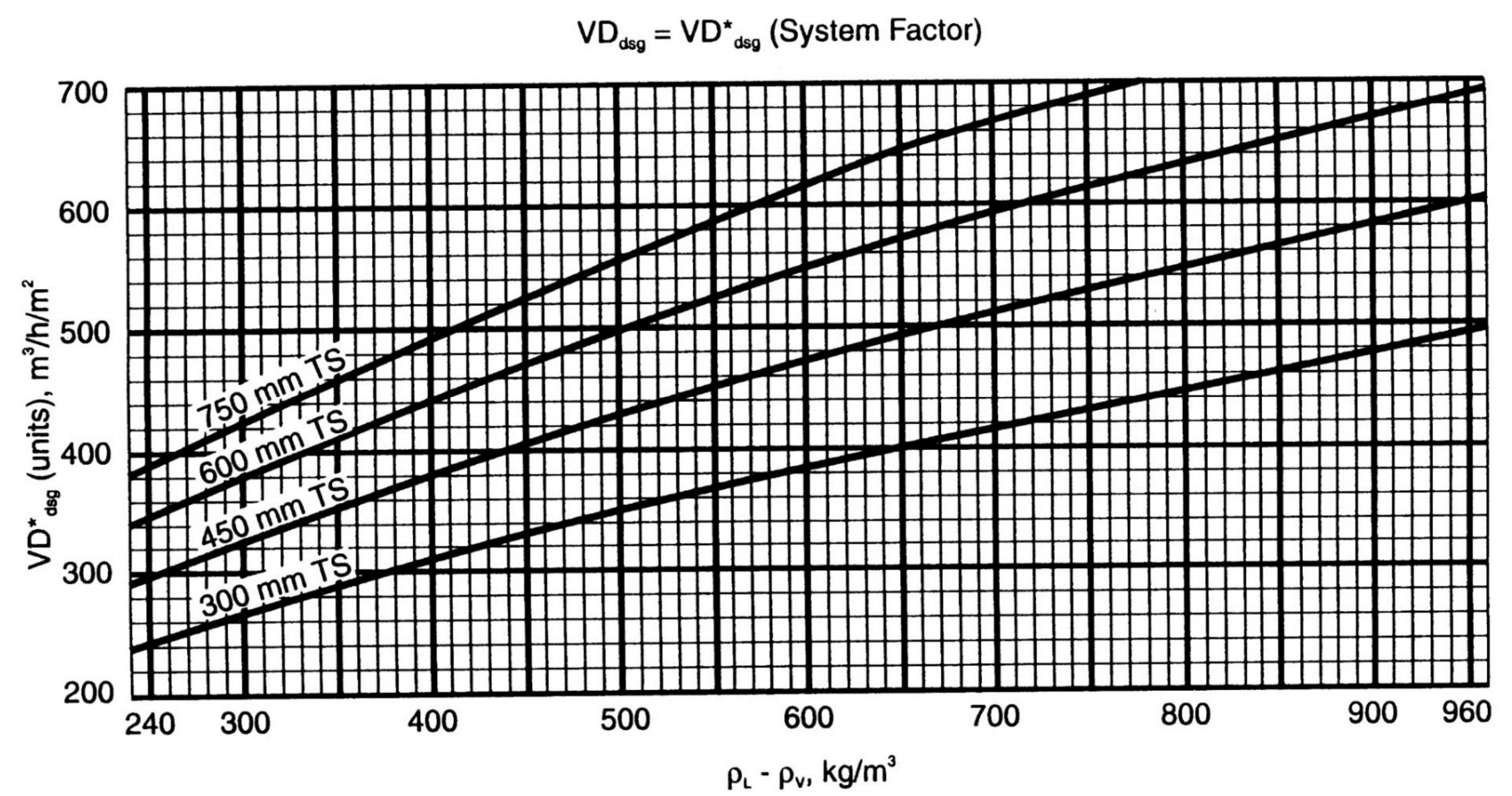
$\left(VD^{*}_{dsg}\right)$ = downcomer velocity for non foaming fluids (m/h) (given by figure 16).

Column Sizing

Tray Column Sizing

3. Detailed method for valve trays

Figure 16 - DOWNCOMER DESIGN VELOCITY



Column Sizing

Tray Column Sizing

3. Detailed method for valve trays

4th step: Determination of Active Area (AAM)

- The minimum active area is a function of vapor and liquid loads, system properties, flood factor and flow path length.
- The flood factor (FF) is used in certain equations for purpose of estimating column size. It is the "design percent of flood" expressed as a fraction.
- A value of not more than 0.77 is normally used for vacuum columns and a value not more than 0.82 is used for other services.
- For demethanisers and near critical point values, it is recommended to adopt a value in the range 0.6 to 0.7.
- These values are intented to give not more than approximately 10 % entrainment.
- Higher flood factorsmay result in excessive entrainment and/or a column sized too small for effective operation.

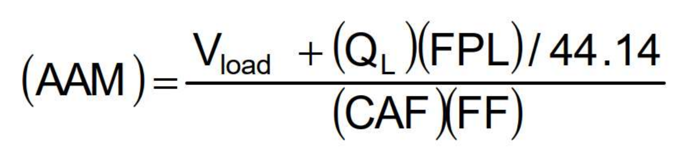
Column Sizing

Tray Column Sizing

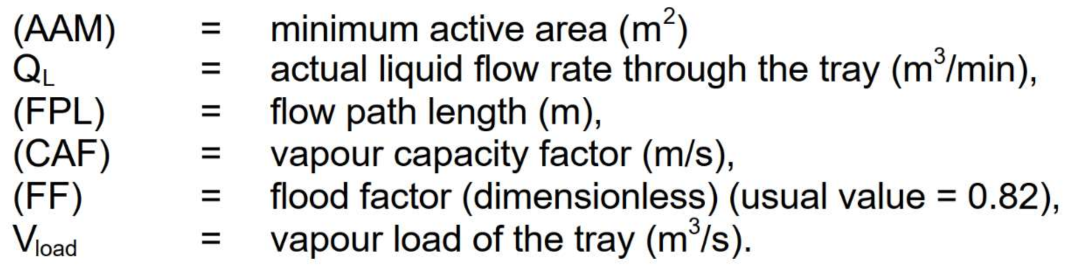
3. Detailed method for valve trays

4th step: Determination of Active Area (AAM)

- A flood factor of 0.65 to 0.75 should be used for column diameters under 36" (90 cm).
- The minimum active area is determined with equation as follows:



$$(AAM) = \frac{V_{load} + (Q_L)(FPL)/44.14}{(CAF)(FF)}$$



(AAM) = minimum active area (m^2)
Q_L = actual liquid flow rate through the tray (m^3/min),
(FPL) = flow path length (m),
(CAF) = vapour capacity factor (m/s),
(FF) = flood factor (dimensionless) (usual value = 0.82),
V_{load} = vapour load of the tray (m^3/s).

Column Sizing

Tray Column Sizing

3. Detailed method for valve trays

4th step: Determination of Active Area (AAM)

$$V_{load} = Q_V \sqrt{\frac{\rho_v}{\rho_l - \rho_v}}$$

Q_V = actual vapour flow rate through the tray (m^3/s),

ρ_l = liquid density at operating temperature and operating pressure of the tray (kg/m^3),

ρ_v = vapour density at operating temperature and operating pressure of the tray (kg/m^3),

Column Sizing

Tray Column Sizing

3. Detailed method for valve trays

5th step: Determination of the Downcomer area (ADM)

- The minimum downcomer area is a function of liquid rate, downcomer design velocity and flood factor.

$$(ADM_m) = \frac{60 \times Q_L}{(VD_{dsg})(FF)}$$

Q_L = actual liquid flow rate through the tray (m^3/min),
VD_{dsg} = downcomer velocity calculate in 3th step (m/h),
FF = flood factor defined in 4th step (dimensionless),
(ADM_m) = minimum downcomer area (m^2).

- If the downcomer area calculated by this equation is less than 11 % of the active area (AAM) adopt for (ADM) the smaller value of relations as follows:

$$(ADM) = (0.11)\,(AAM)$$

$$\text{or } (ADM) = 2\,(ADM_m)$$

Column Sizing

Tray Column Sizing

3. Detailed method for valve trays

6th step: Determination of the minimum inside diameter (DC) of the column

- The approximate column cross sectional area is calculated by equations as follows:

$$(\mathrm{ATM}) = (\mathrm{AAM}) + 2\,(\mathrm{ADM})$$

$$\text{or } (\mathrm{ATM}) = \frac{V_{\mathrm{load}}}{(0.78)(\mathrm{CAF})(\mathrm{FF})}$$

- The higher value is adopted.
- Minimum inside diameter of the column (DC) in meters is calculated with relation:

$$(\mathrm{DC}) = \sqrt{\frac{(\mathrm{ATM})}{0.7854}}$$

Column Sizing

Column (or tower) height estimation

- The height of a trayed column takes account of:

1. the number of actual trays,
2. the space required for the disengagement of the vapour at the top of the column to minimise the liquid carry-over,
3. the retention liquid volume required at the bottom of the column to control the liquid discharge and to prevent the gas carry-through in the liquid.

- Top part (vapour disengagement) and bottom part (retention liquid volume) are designed with the same criterions used for the sizing of vertical separators (see Vessels – Vapor-liquid separators).

Column Sizing

Column (or tower) height estimation

- Total height = height of the top part (distance between the tangent line of the shell top and the first tray in the column);

  \+ distance between the upper tray and the lower tray in the column

  (= (number of trays - 1) x tray spacing);

  \+ distance between the lower tray and the high liquid level (HLL)

  (= 2 x tray spacing);

  \+ height of the bottom part

  (distance between the high liquid level and the tangent line of the shell bottom);

  \+ height of the transition space if the diameter of the column changes (one time)

  along the total height of the column

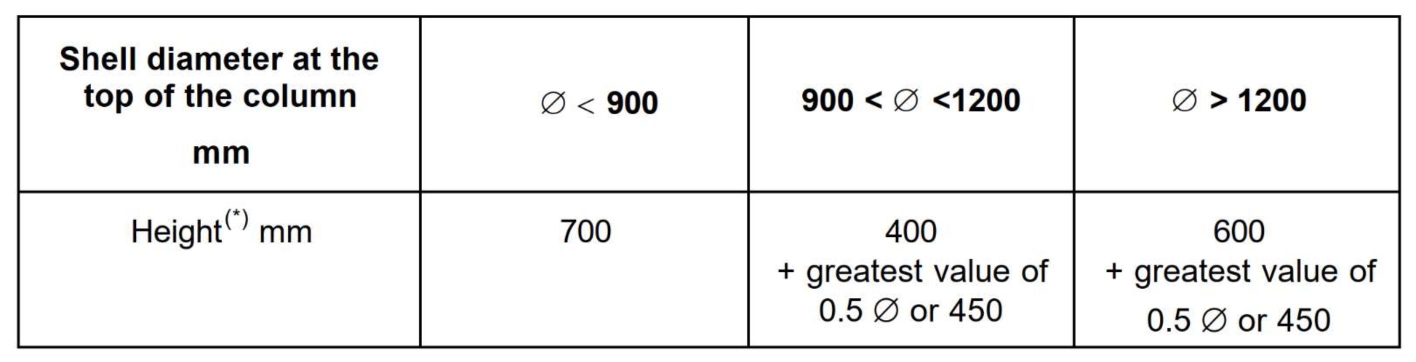
  $$\left(=\frac{\sqrt{3}}{2}(\Delta D)\text{ where }\Delta D=\text{difference of the two diameter values}\right);$$

  \+ feed tray space when special distribution is required.

Column Sizing

Column (or tower) height estimation

- Usual tray spacing: 18 in (46 cm), 24 in (61 cm), 30 in (76 cm) (24 in is the most frequent).
- Height of the top part of the column is given in table as follows.



Shell diameter at the top of the column mm	$\varnothing < 900$	$900 < \varnothing < 1200$	$\varnothing > 1200$
Height$^{(*)}$ mm	700	400 + greatest value of 0.5 $\varnothing$ or 450	600 + greatest value of 0.5 $\varnothing$ or 450

*Add 100~150 mm to this value if a mesh pad is installed.

Column Sizing

Column (or tower) height estimation

- Height determination of the bottom part of the column

= height of the liquid volume corresponding to a retention time of:

- 15 minutes for a feed to a downstream column,
- 8 minutes for a feed to a downstream furnace,
- 5 minutes for a feed to storage or an other unit with pump,
- 3 minutes for a feed to a storage or an other unit without pump,

(minimum value = 300 mm).

+ height between the low liquid level (LLL) and the low level alarm (LLA)

(= height of the liquid volume corresponding to a retention time of 1 to 2 minutes with a minimum value of 150 mm).

+ height between the low level alarm (LLA) and the tangent line of the shell bottom (300 mm).

Column Sizing

Process Limitations for Trayed

1. ***Liquid Flow Paths Versus Liquid Flowrate***

- The weir loading is generally used to determine the liquid loading of the tray. Although there is no specific design limit for weir loading, increased number of flow passes should be considered when the weir loading is greater than 100 – 120 gpm/ft (74.5 – 89.4 m3/h/m) in order to provide increased tray capacity.
- A minimum weir loading of 5 gpm/ft (3.7 m3/h/m) is recommended at turndown. Picket fence outlet weirs (or "weir blocks") can be added to increase the weir loading by specifying an override to the calculated outlet weir length or specifying a percentage of the outlet weir to be blocked.

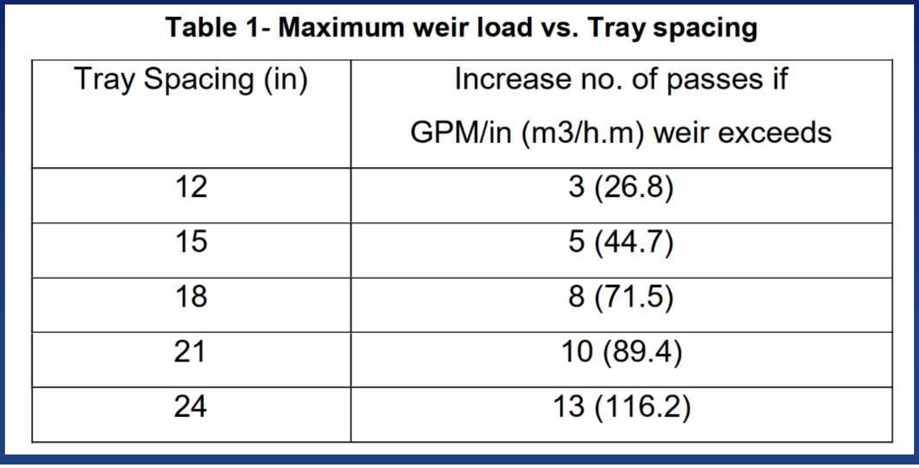


Table 1- Maximum weir load vs. Tray spacing

Tray Spacing (in)	Increase no. of passes if GPM/in (m3/h.m) weir exceeds
12	3 (26.8)
15	5 (44.7)
18	8 (71.5)
21	10 (89.4)
24	13 (116.2)

Column Sizing

Process Limitations for Trayed

2. Jet Flood

- Jet flood rating should be limited to 85% or other value limited in datasheet, of flood to avoid the possibility of flooding and/or inefficient operation. Increasing tower diameter, active area and/or tray spacing can be used to reduce the jet flood rating.

3. Downcomer Flood

- Downcomer rating should be limited to 85% of flood. The downcomer rating is generally set by the size of the downcomer area at the top. For a well balanced tray design, the downcomer and jet flood ratings should be fairly equal.

Column Sizing

Process Limitations for Trayed

4. Downcomer Backup

- The allowable downcomer backup is measured as the percentage of the tray spacing that the liquid level in the downcomer is allowed to reach. The downcomer backup should not exceed 40% of the tray spacing for high vapor density systems (greater than 3 lb/ft3 (48 kg/m3)), 50% for normal densities (between 16 and 48 kg/m3), and 60% for densities less than 1 lb/ft3 (16 kg/m3).
- The downcomer backup is dependent on the tray pressure drop and the clearance under the downcomer.

5. DC Exit Velocity

- The downcomer exit velocity is the liquid velocity as it flows horizontally through the downcomer clearance. This value should be limited to 1.5 ft/sec (0.46 m/s) for conventional valve tray designs. The downcomer exit velocity is most easily adjusted by changing the downcomer clearance.

Column Sizing

Process Limitations for Trayed

6. Dry Tray DP

- The dry tray pressure drop is an intermediate term in calculating the total tray pressure drop that does not include the effect of the liquid head. It can be used to provide a relative indication of vapor velocity through the valves.
- A good starting point for many tray designs is a dry tray pressure drop of around 2 inches of hot liquid.
- As a rule of thumb, the dry tray pressure drop should be limited to 15% of the tray spacing when possible.

7. Total Tray DP

- Many tray design methods do not set specific limits on the tray pressure drop; however, a tray typically reaches flood at a pressure drop of around 8-10 mm Hg per tray.
- The tray pressure drop is also a very key component to downcomer hydraulics due to its impact on downcomer backup.

Column Sizing

Process Limitations for Trayed

8. ***Head Loss Under DC***

- The head loss under the downcomer is based on the downcomer clearance and the shape of the downcomer edge.
- Typically, the head loss should be designed somewhere between 0.06 to 1.0 inches (1.5 to 25 mm).
- The head loss can be adjusted bychanging the downcomer clearance or utilizing radius-tip downcomers.

9. ***DC Residence Time***

- The calculated residence time in the limiting downcomer is based on the liquid flow rate and the available volume of the downcomer.
- Downcomer residence time is not typically used by Koch-Glitsch to determine proper downcomer sizing.
- This parameter is used by some tray designers to size downcomers in foaming systems.

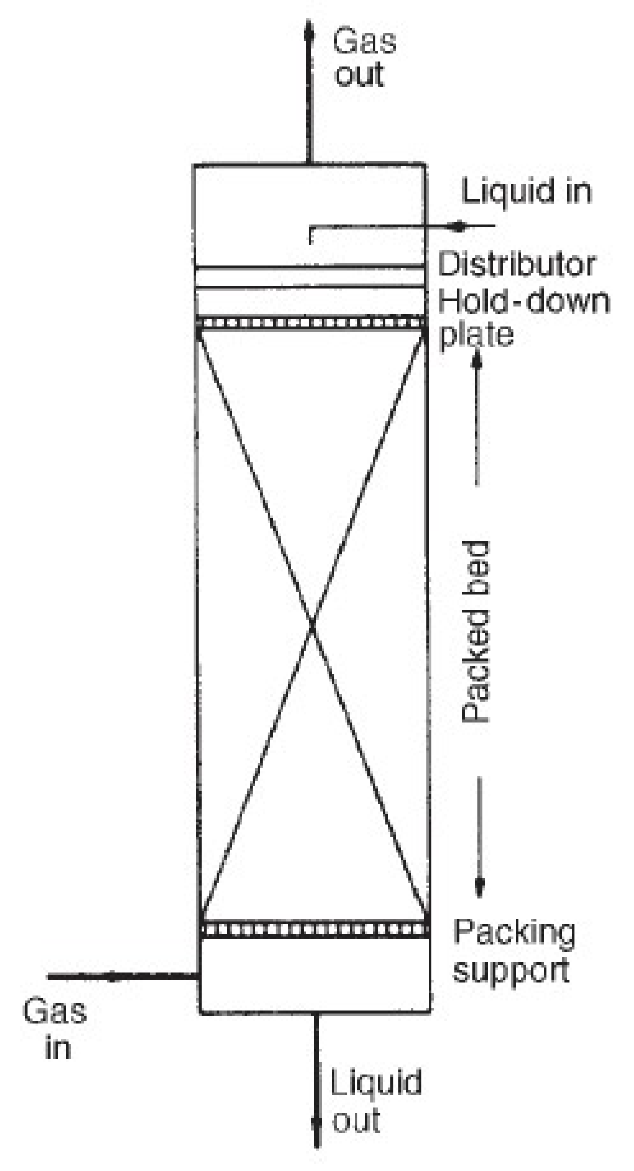
Packed Column

- Packed columns are used for distillation, gas absorption, and liquid-liquid extraction

- The gas liquid contact in a packed bed column is continuous, not stage-wise, as in a tray column. The liquid flows down the column over the packing surface and the gas or vapor, counter-currently, up the column.

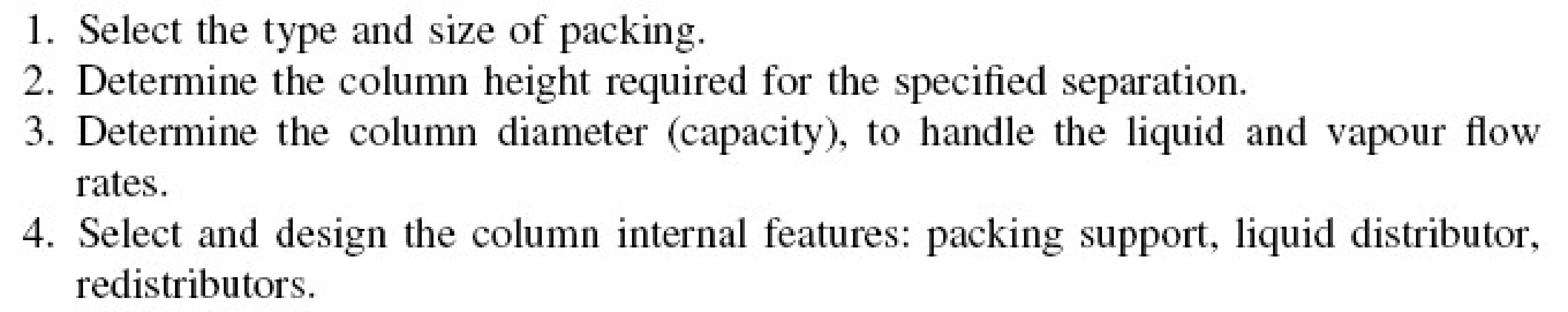
- The performance of a packed column is very dependent on the maintenance of good liquid and gas distribution throughout the packed bed, and this is an important consideration in packed-column design.

Packed Column



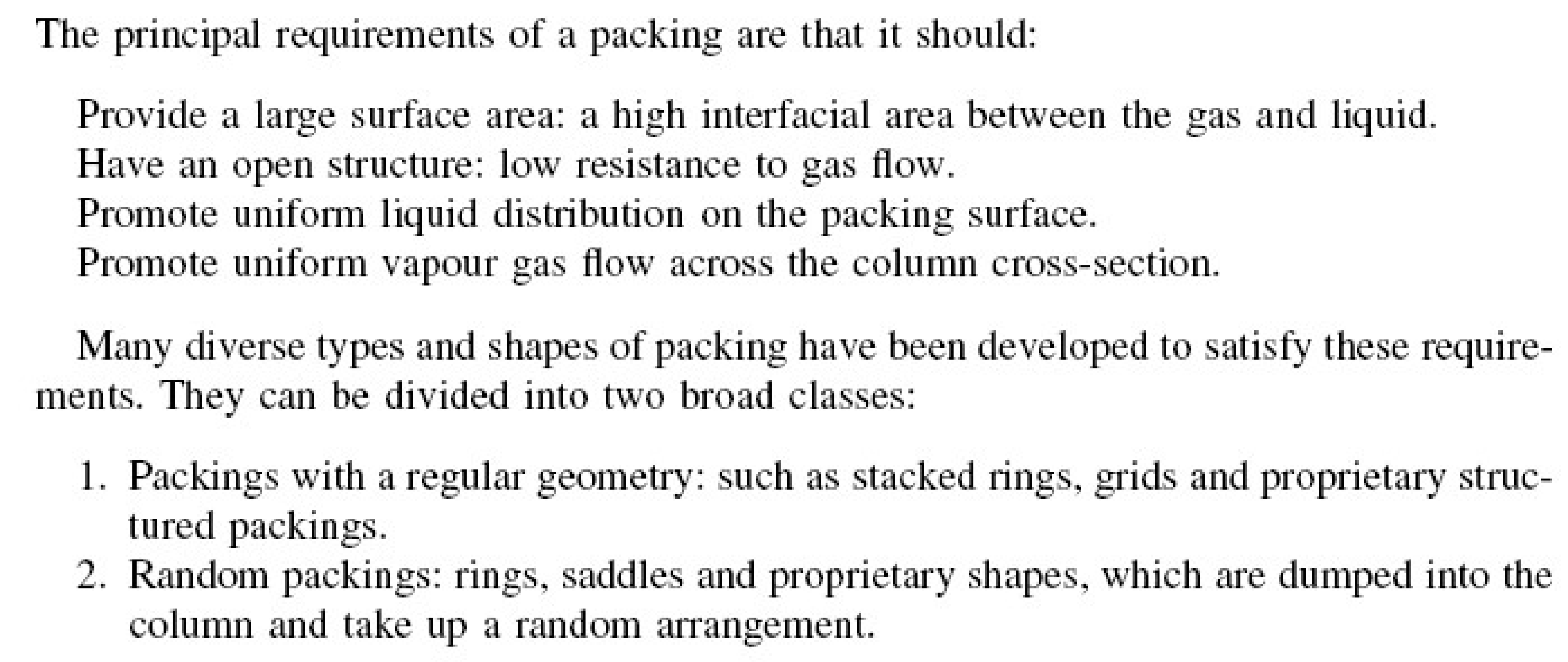
Packed Column

Packed-Column Design Procedures



1. Select the type and size of packing.
2. Determine the column height required for the specified separation.
3. Determine the column diameter (capacity), to handle the liquid and vapour flow rates.
4. Select and design the column internal features: packing support, liquid distributor, redistributors.

Types of Packing



The principal requirements of a packing are that it should:

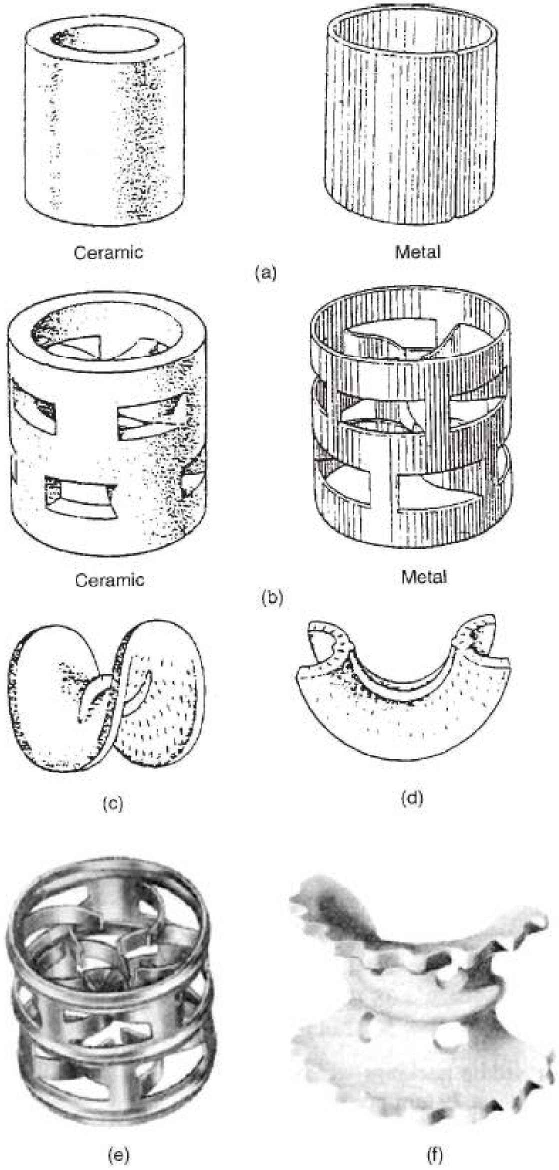
Provide a large surface area: a high interfacial area between the gas and liquid.
Have an open structure: low resistance to gas flow.
Promote uniform liquid distribution on the packing surface.
Promote uniform vapour gas flow across the column cross-section.

Many diverse types and shapes of packing have been developed to satisfy these requirements. They can be divided into two broad classes:

1. Packings with a regular geometry: such as stacked rings, grids and proprietary structured packings.
2. Random packings: rings, saddles and proprietary shapes, which are dumped into the column and take up a random arrangement.

Packed Column

Types of Packing



gure 11.37. Types of packing (Norton Co.). (*a*) Raschig rings (*b*) Pall rings (*c*) Berl saddle ceramic (*d*) Intalox saddle ceramic (*e*) Metal Hypac (*f*) Ceramic, super Intalox

Packed Column

Structured Packing

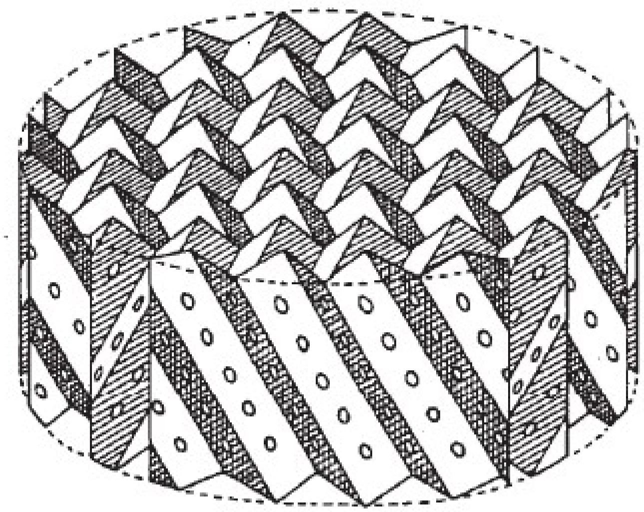
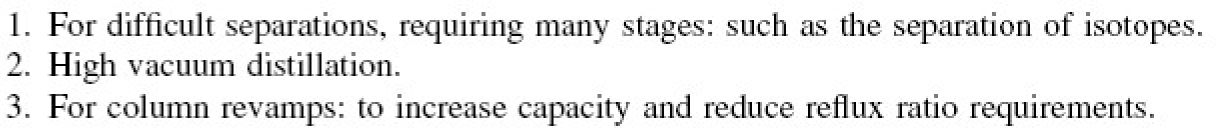


Figure 11.38. Make-up of structured packing. (Reproduced from Butcher (1988) with permission.)



1. For difficult separations, requiring many stages: such as the separation of isotopes.
2. High vacuum distillation.
3. For column revamps: to increase capacity and reduce reflux ratio requirements.

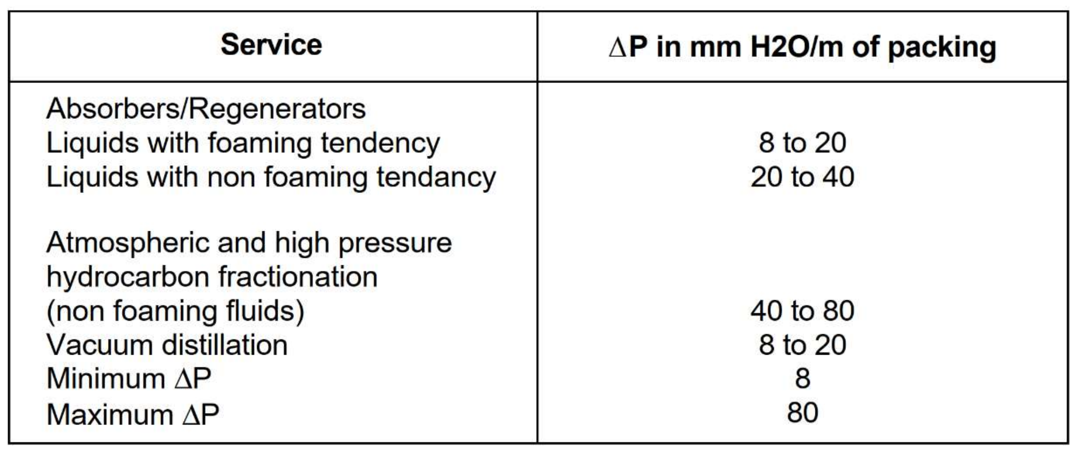
- The advantage of structured packings over random packing is their low HETP (typically less than 0.5 m) and low pressure drop (around 100 Pa/m).

Column Sizing

RANDOM PACKED COLUMN SIZING

Diameter sizing

- Correlations have been developed to predict the pressure drop in a random packed column for a given loading and column diameter at actual gas velocity lower than the gas flooding velocity which corresponds to the maximum capacity condition for a packed column.
- Alternatively these correlations can be used to determine the column diameter for a given pressure drop.
- The design pressure drop depends on the service. The following values may serve as a guide.



Service	ΔP in mm H2O/m of packing
Absorbers/Regenerators	
Liquids with foaming tendency	8 to 20
Liquids with non foaming tendancy	20 to 40
Atmospheric and high pressure hydrocarbon fractionation (non foaming fluids)	40 to 80
Vacuum distillation	8 to 20
Minimum ΔP	8
Maximum ΔP	80

Column Sizing

RANDOM PACKED COLUMN SIZING

Diameter sizing

- Chart of the figure 17 – Packed column pressure drop correlation, gives for a given pressure drop and the flow parameter value in abscissa the value of the capacity term.

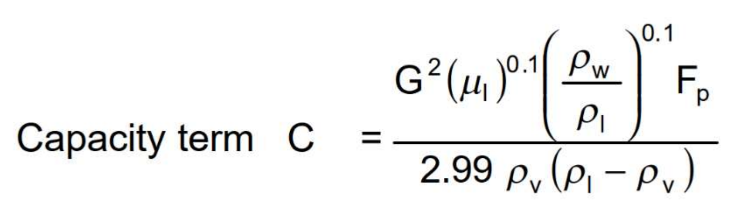
$$\text{flow parameter} = \frac{L_m}{G_m}\sqrt{\frac{\rho_v}{\rho_l}} = \frac{L}{G}\sqrt{\frac{\rho_v}{\rho_l}}$$

L_m = liquid mass flow rate (kg/s),
L = liquid mass velocity (kg/(m^2.s)),
G_m = gas (or vapour) mass flow rate (kg/s),
G = gas (or vapour) mass velocity (kg/(m^2.s)),
ρ_v = gas (or vapour) density at operating temperature and operating pressure of the contact section (kg/m^3),
ρ_l = liquid density at operating temperature and operating pressure of the contact section (kg/m^3).

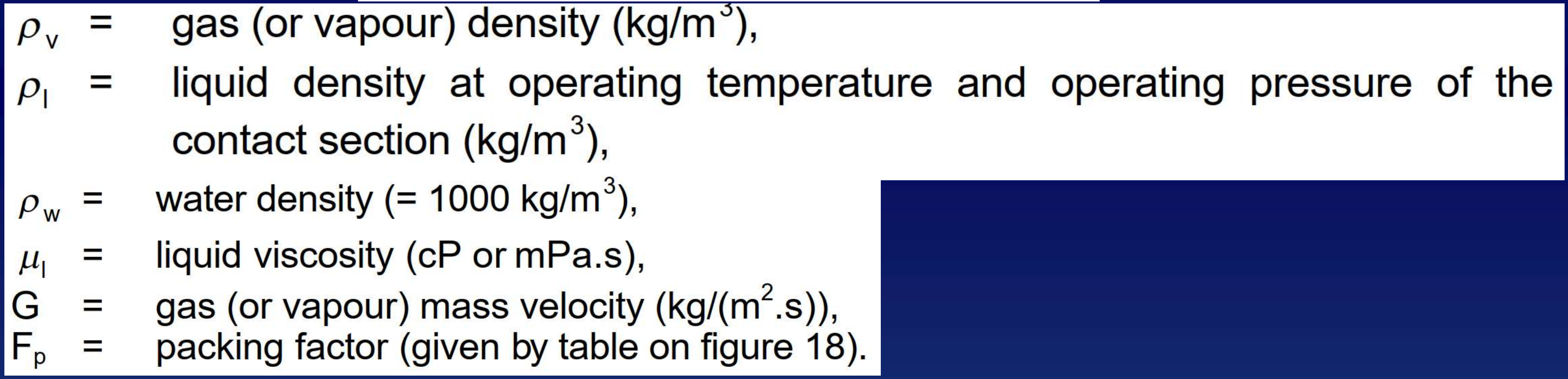
Column Sizing

RANDOM PACKED COLUMN SIZING

Diameter sizing



$$\text{Capacity term} \quad C = \frac{G^2 (\mu_l)^{0.1} \left(\frac{\rho_w}{\rho_l} \right)^{0.1} F_p}{2.99\, \rho_v (\rho_l - \rho_v)}$$



ρ_v = gas (or vapour) density (kg/m^3),

ρ_l = liquid density at operating temperature and operating pressure of the contact section (kg/m^3),

ρ_w = water density (= 1000 kg/m^3),

μ_l = liquid viscosity (cP or mPa.s),

G = gas (or vapour) mass velocity (kg/(m^2.s)),

F_p = packing factor (given by table on figure 18).

- The packing factor is determined by experiments. Packing factor values may be obtained from the packing manufacturer.

Column Sizing

RANDOM PACKED COLUMN SIZING

Diameter sizing

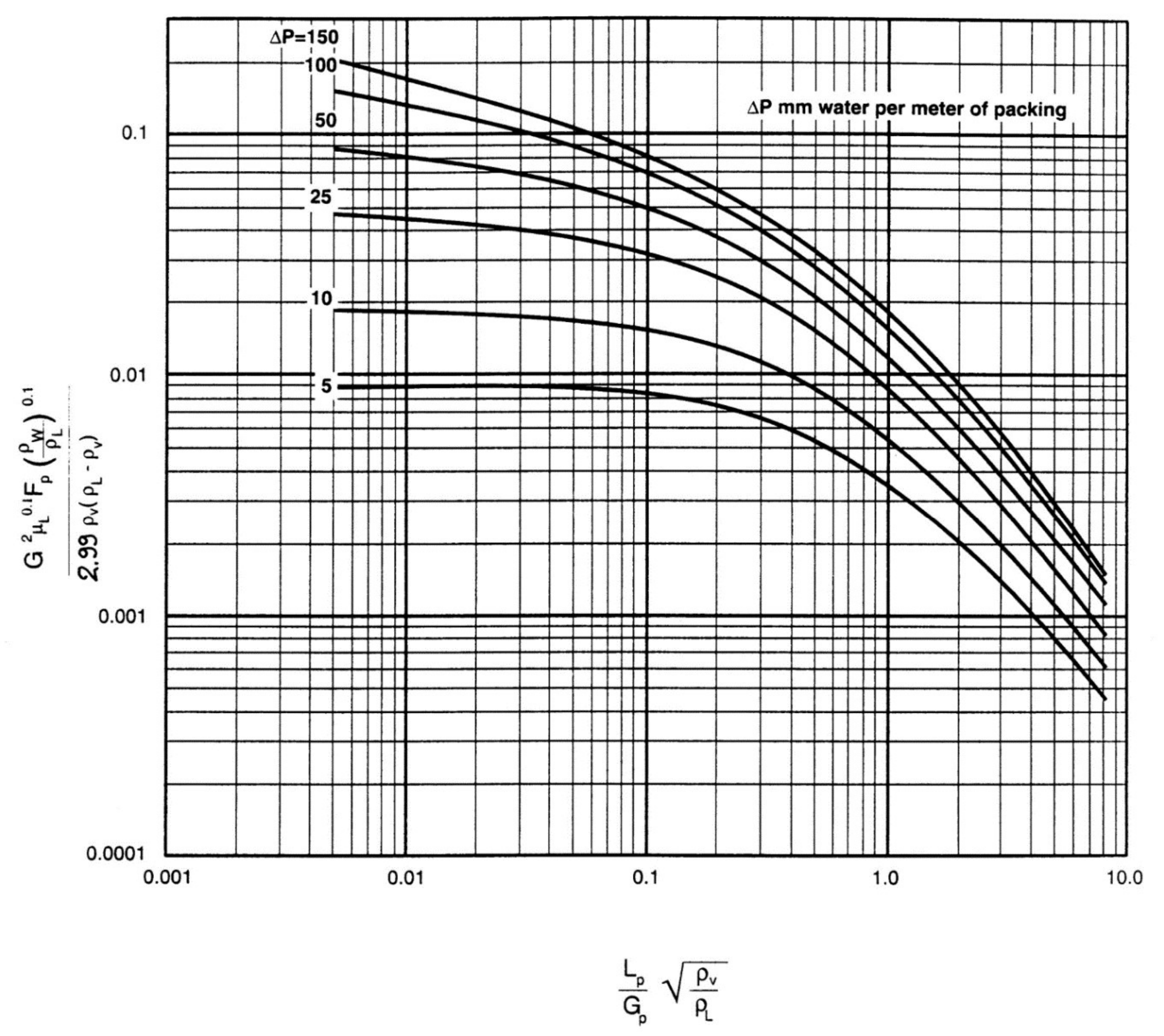
- The packing factors for various packings are shown in figure 18. Usually packings smaller than 25 mm (1 in) size are intended for column diameters of 300 mm or smaller, packings of 25 mm to 37 mm (1 in to 1½ in) in size for column diameters from 300 mm to 900 mm, and packings from 50 mm to 75 mm (2 in to 3 in) in size for column diameters of 900 mm and more.

- The designer has to select the proper size of packing, and therefore the proper packing factor for calculations.

Column Sizing

RANDOM PACKED COLUMN SIZING

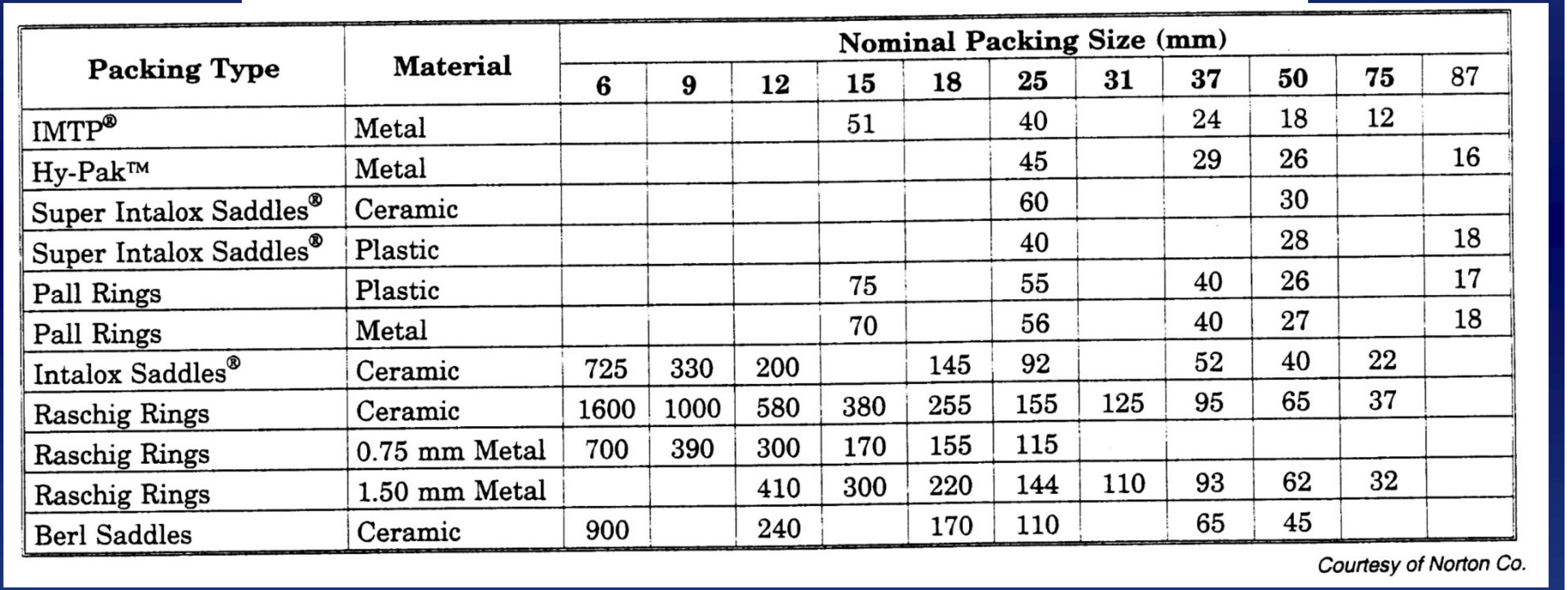
Figure 17 - PACKED COLUMN PRESSURE DROP CORRELATION



Column Sizing

RANDOM PACKED COLUMN SIZING

Figure 18 - PACKING FACTOR (Fp) FOR DUMPED PACKING



Packing Type	Material	Nominal Packing Size (mm)										
		6	9	12	15	18	25	31	37	50	75	87
IMTP®	Metal				51		40		24	18	12	
Hy-Pak™	Metal						45		29	26		16
Super Intalox Saddles®	Ceramic						60			30		
Super Intalox Saddles®	Plastic						40			28		18
Pall Rings	Plastic				75		55		40	26		17
Pall Rings	Metal				70		56		40	27		18
Intalox Saddles®	Ceramic	725	330	200		145	92		52	40	22	
Raschig Rings	Ceramic	1600	1000	580	380	255	155	125	95	65	37	
Raschig Rings	0.75 mm Metal	700	390	300	170	155	115					
Raschig Rings	1.50 mm Metal			410	300	220	144	110	93	62	32	
Berl Saddles	Ceramic	900		240		170	110		65	45		

Courtesy of Norton Co.

Column Sizing

RANDOM PACKED COLUMN SIZING

1st step: Determine the capacity term value

- Choice a ΔP value and calculate the flow parameter $\frac{L}{G}\sqrt{\frac{\rho_v}{\rho_l}}$.
- With the chart of figure 17 determine in ordinate the capacity term C value.

2nd step: Determine the gas (or vapour) mass velocity (G)

- Choice a packing type and determine the packing factor (Fp) with the table of figure 18.
- Calculate the gas (or vapour) mass velocity G with the capacity term C value determined in previous step.

$$G^2 = \frac{2.99\rho_v(\rho_l - \rho_v)C}{(\mu_l)^{0.1}\left(\frac{\rho_w}{\rho_l}\right)^{0.1} F_p}$$

Column Sizing

RANDOM PACKED COLUMN SIZING

3th step: Determine the minimum inside diameter (D) of the column

$$D = \sqrt{\frac{G_m}{0.7854\ G}}$$

D = minimum diameter (m),

G_m = gas (or vapour) mass flowrate (kg/s),

G = gas (or vapour) mass velocity (kg/(m^2.s)).

Column Sizing

RANDOM PACKED COLUMN SIZING

Determination of the packing height

Height Equivalent to a Theoritical Plate (HETP) definition

- The optimum number of equilibrium stages (or theoritical plates or trays) to perform a required separation is determined in many cases with a thermodynamical simulation program.

- To determine the height of a packed column bed, when the number of equilibrium stages is fixed, the height of packing to achieve the same separation as one equilibrium stage is required. This height is called HETP (Height Equivalent to a Theoritical Plate).

$$h = (HETP)\ (n)$$

h = packing height,
n = optimum number of equilibrium stages,
(HETP) = height equivalent to a theoritical plate.

Column Sizing

RANDOM PACKED COLUMN SIZING

HETP determination

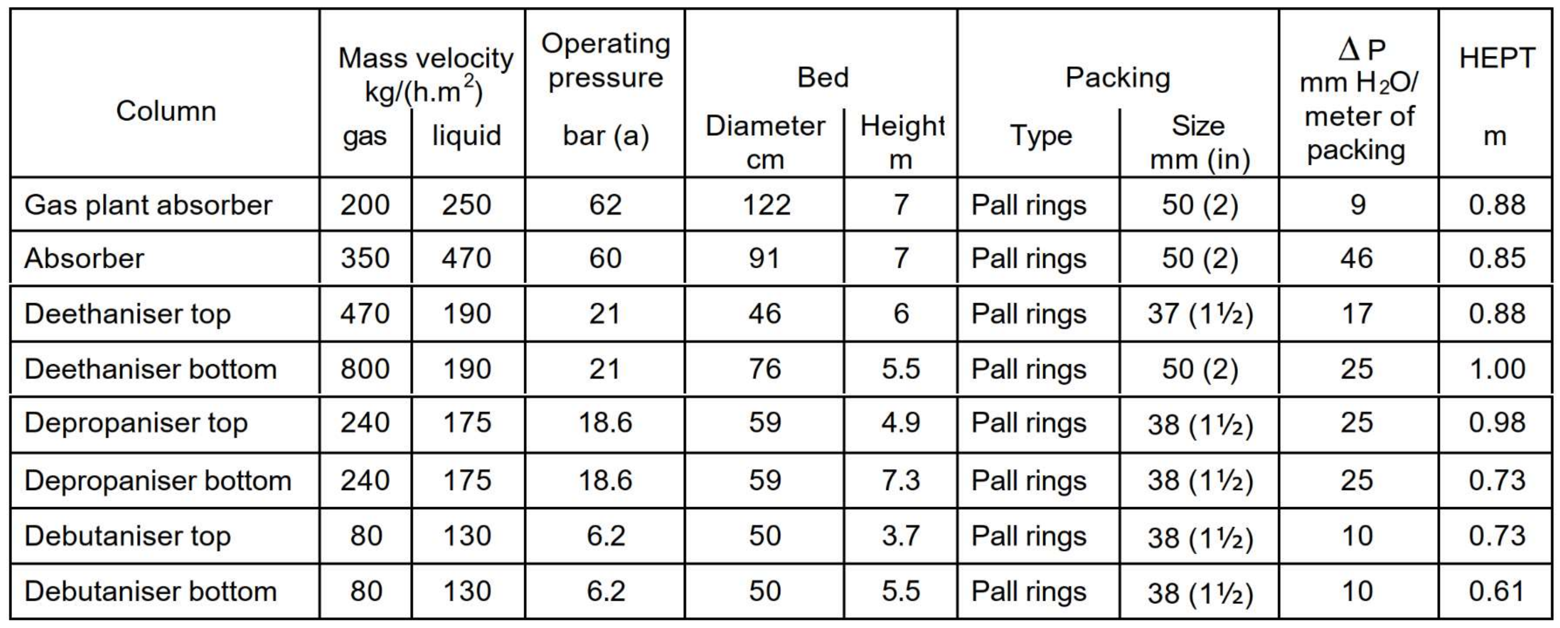
- The HETP is determined experimentally in laboratory or pilot plant tests. It is a function, among others, of packing type, vapor and liquid densities, liquid viscosity and surface tension, vapor and liquid loading.
- Few generalised methods for calculating HETP are available in published literature. Consult a packing manufacturer to obtain reasonable estimates of packing requirements and HETP for a particular service.
- Generally HETP values range from 300 to 900 mm but can be as high as 1500 mm (e.g. glycol stripper).

Column Sizing

RANDOM PACKED COLUMN SIZING

HETP determination

- Table as follows provides some examples of HETP values for hydrocarbon systems in gas processing industry.



Column	Mass velocity kg/(h.m^2)		Operating pressure	Bed		Packing		Δ P mm H_2O/ meter of packing	HEPT
	gas	liquid	bar (a)	Diameter cm	Height m	Type	Size mm (in)		m
Gas plant absorber	200	250	62	122	7	Pall rings	50 (2)	9	0.88
Absorber	350	470	60	91	7	Pall rings	50 (2)	46	0.85
Deethaniser top	470	190	21	46	6	Pall rings	37 (1½)	17	0.88
Deethaniser bottom	800	190	21	76	5.5	Pall rings	50 (2)	25	1.00
Depropaniser top	240	175	18.6	59	4.9	Pall rings	38 (1½)	25	0.98
Depropaniser bottom	240	175	18.6	59	7.3	Pall rings	38 (1½)	25	0.73
Debutaniser top	80	130	6.2	50	3.7	Pall rings	38 (1½)	10	0.73
Debutaniser bottom	80	130	6.2	50	5.5	Pall rings	38 (1½)	10	0.61

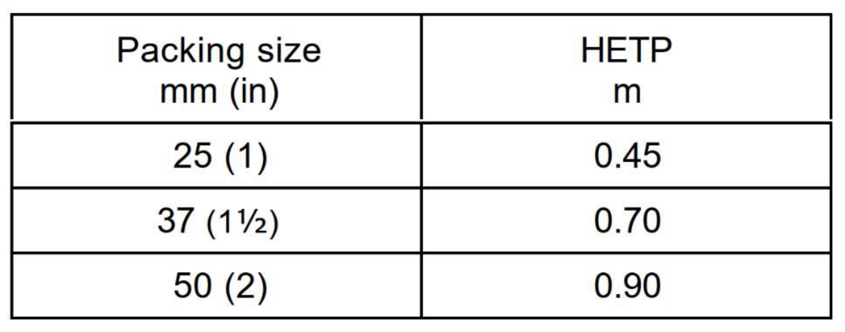
- In cryogenic plants some demethanisers run with a HEPT value from 450 to 600 m.

Column Sizing

RANDOM PACKED COLUMN SIZING

HETP determination

- For preliminary sizing calculations the following values could be adopted.



Packing size mm (in)	HETP m
25 (1)	0.45
37 (1½)	0.70
50 (2)	0.90

Column Sizing

RANDOM PACKED COLUMN SIZING

Column height estimation

- The height of a packed column takes account of:

1. the packing height,
2. the space required for the disengagement of the vapor at the top of the column to minimise the liquid carry-over,
3. the retention liquid volume required at the bottom of the column to control the liquid discharge and to prevent the gas carry-through in the liquid,
4. the space required for the feed stream, when the column has a stripping and a rectifying section,
5. the space for location of liquid distributors and redistributors.

Column Sizing

RANDOM PACKED COLUMN SIZING

Column height estimation

- Packing height determination is described in previous paragraphs.
- Height determination of the bottom part is identical to the height determination of the bottom part of a trayed column.
- Height determination of the top part and of the feed part of the column follows the same rules described for trayed columns. However this value must be increased by the space required for liquid distributors location (15 to 45 cm above the packing).

Column

Choice of Trays or Packing

1. Tray columns can be designed to handle a wider range of liquid and gas flow-rates than packed columns.
2. Packed columns are not suitable for very low liquid rates.
3. The efficiency of a tray can be predicted with more certainty than the equivalent term for packing (HETP or HTU).
4. Tray columns can be designed with more assurance than packed columns. There is always some doubt that good liquid distribution can be maintained throughout a packed column under all operating conditions, particularly in large columns.
5. It is easier to make provision for cooling in a tray column; coils can be installed on the trays.
6. It is easier to make provision for the withdrawal of side-streams from tray columns.
7. If the liquid causes fouling, or contains solids, it is easier to make provision for cleaning in a tray column; manways can be installed on the trays. With small diameter columns it may be cheaper to use packing and replace the packing when it becomes fouled.

Column

Choice of Trays or Packing

8. For corrosive liquids a packed column will usually be cheaper than the equivalent tray column.
9. The liquid hold-up is appreciably lower in a packed column than a tray column. This can be important when the inventory of toxic or flammable liquids needs to be kept as small as possible for safety reasons.
10. Packed columns are more suitable for handling foaming systems.
11. The pressure drop per equilibrium stage (HETP) can be lower for packing than trays; and packing should be considered for vacuum columns.
12. Packing should always be considered for small diameter columns, say less than 0.6 m, where trays would be difficult to install, and expensive.