PACKED COLUMNS

① ② ③

Packed columns are used for distillation, gas absorption, and liquid-liquid extraction; only distillation and absorption will be considered in this section. Stripping (desorption) is the reverse of absorption and the same design methods will apply.

The gas liquid contact in a packed bed column is continuous, not stage-wise, as in a plate column. The liquid flows down the column over the packing surface and the gas or vapour, counter-currently, up the column. In some gas-absorption columns co-current flow is used. 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.

A schematic diagram, showing the main features of a packed absorption column, is given in Figure 11.36. A packed distillation column will be similar to the plate columns shown in Figure 11.1, with the plates replaced by packed sections.

Choice of plates or packing

The choice between a plate or packed column for a particular application can only be made with complete assurance by costing each design. However, this will not always be worthwhile, or necessary, and the choice can usually be made, on the basis of experience by considering main advantages and disadvantages of each type; which are listed below: ← Plates or packing ?

1. Plate 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 plate can be predicted with more certainty than the equivalent term for packing (HETP or HTU).
4. Plate 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 plate column; coils can be installed on the plates.
6. It is easier to make provision for the withdrawal of side-streams from plate columns.
7. If the liquid causes fouling, or contains solids, it is easier to make provision for cleaning in a plate column; manways can be installed on the plates. With small-diameter columns it may be cheaper to use packing and replace the packing when it becomes fouled.
8. For corrosive liquids a packed column will usually be cheaper than the equivalent plate column.
9. The liquid hold-up is appreciably lower in a packed column than a plate 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 plates; 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 plates would be difficult to install, and expensive.

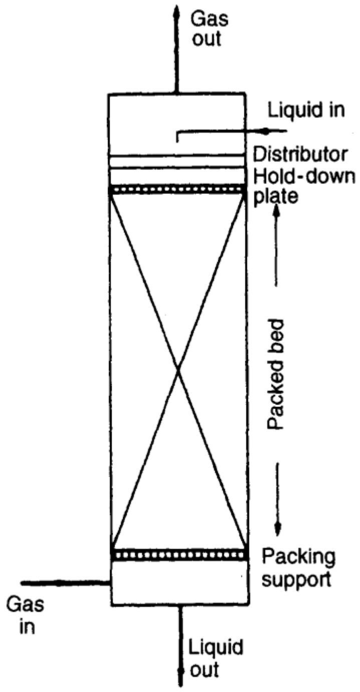


Figure 11.36. Packed absorption column

Note: Page 1
HETP is height equivalent of a theoretical plate.
4- about heat evolution, Cooling is easier in a plate column.
5- about fouling, for a big diameter use plate and for a small diameter use packed bed.
Small size packing provides high surface area.

Packed-column design procedures

The design of a packed column will involve the following steps:

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

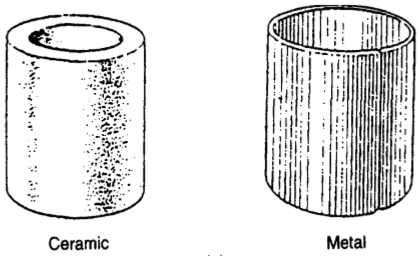
The principal requirements of a packing are that it should:

① – Provide a large surface area: a high interfacial area between the gas and liquid.
② – Have an open structure: low resistance to gas flow.
③ – Promote uniform liquid distribution on the packing surface.

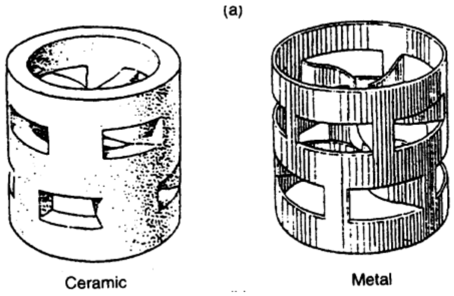
– Promote uniform vapor gas flow across the column cross-section.

Random packing

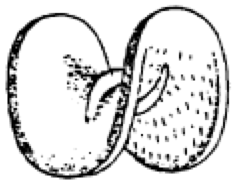
1. Rashing ring



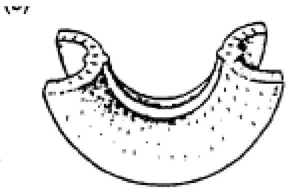
2. Pall ring



3. Saddle

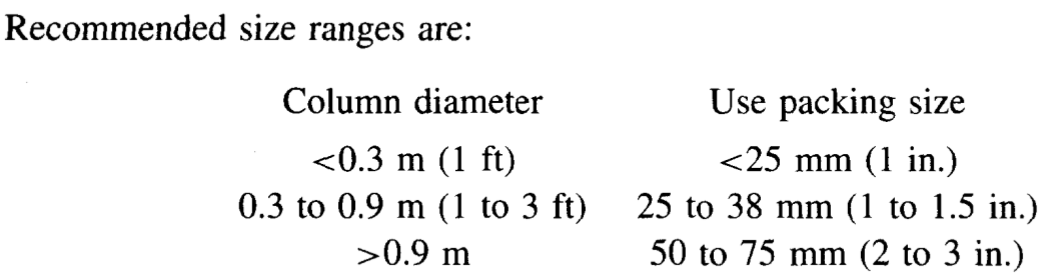


Berl saddle rings



Intalokx saddle rings

Packing size



Recommended size ranges are:

Column diameter	Use packing size
<0.3 m (1 ft)	<25 mm (1 in.)
0.3 to 0.9 m (1 to 3 ft)	25 to 38 mm (1 to 1.5 in.)
>0.9 m	50 to 75 mm (2 to 3 in.)

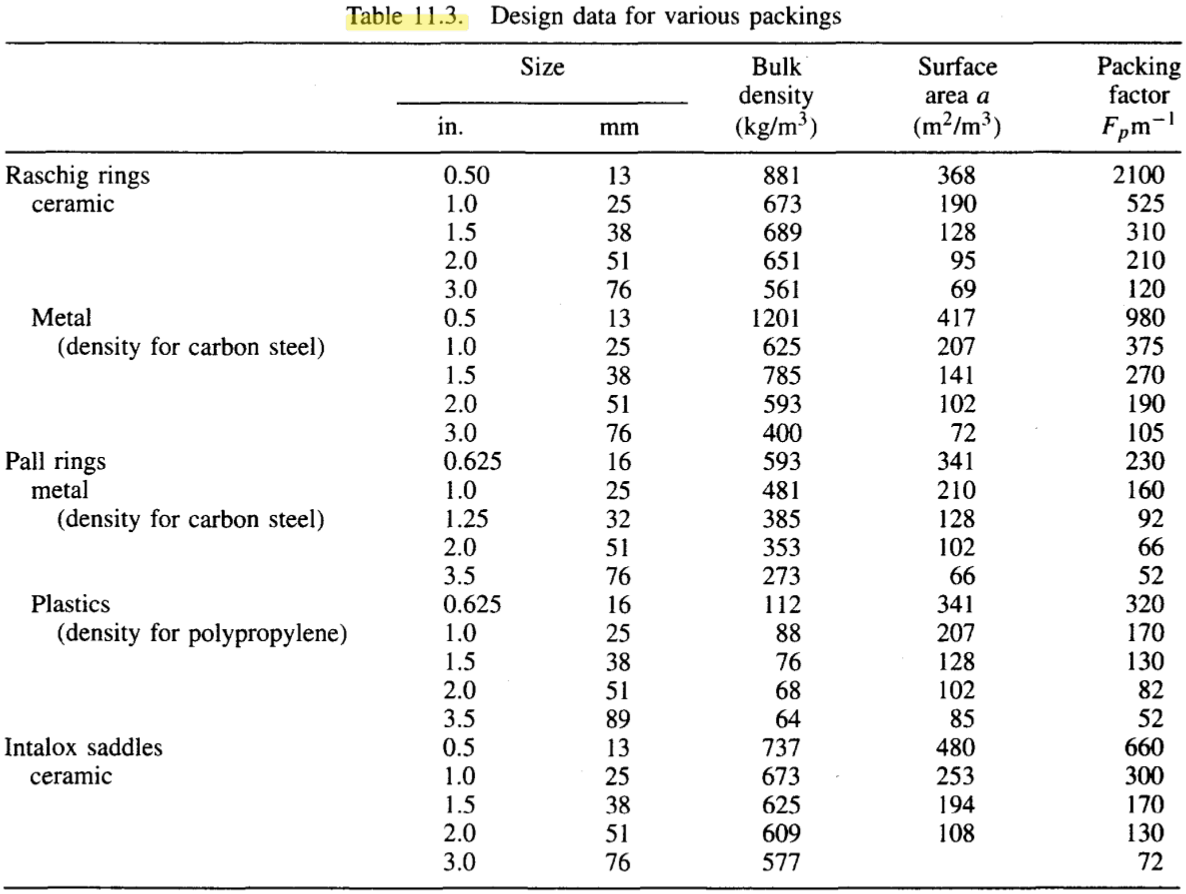


Table 11.3. Design data for various packings

	Size		Bulk density (kg/m^3)	Surface area a (m^2/m^3)	Packing factor F_pm^{-1}
	in.	mm			
Raschig rings	0.50	13	881	368	2100
ceramic	1.0	25	673	190	525
	1.5	38	689	128	310
	2.0	51	651	95	210
	3.0	76	561	69	120
Metal	0.5	13	1201	417	980
(density for carbon steel)	1.0	25	625	207	375
	1.5	38	785	141	270
	2.0	51	593	102	190
	3.0	76	400	72	105
Pall rings	0.625	16	593	341	230
metal	1.0	25	481	210	160
(density for carbon steel)	1.25	32	385	128	92
	2.0	51	353	102	66
	3.5	76	273	66	52
Plastics	0.625	16	112	341	320
(density for polypropylene)	1.0	25	88	207	170
	1.5	38	76	128	130
	2.0	51	68	102	82
	3.5	89	64	85	52
Intalox saddles	0.5	13	737	480	660
ceramic	1.0	25	673	253	300
	1.5	38	625	194	170
	2.0	51	609	108	130
	3.0	76	577		72

Packed-bed height for absorption

For packed absorption and stripping columns where the concentration of the solute is small, say less than 10 per cent, the flow of gas and liquid will be essentially constant throughout the column, and the height of packing required, Z, is given by:

Note: Absorption is transfer from gas phase to liquid phase.

1-Solute concentration y1 is higher than ye. Then solute may be transfer to the liquid phase until it reaches equilibrium ($Y_1 > Y_e$).
2-Solute concentration in the liquid phase is less than equilibrium. Then solute concentration may be increased until it reaches equilibrium ($X_e > X$).
3- $X_1 > X_2$.
4-Stripping is transfer from liquid phase to liquid phase.

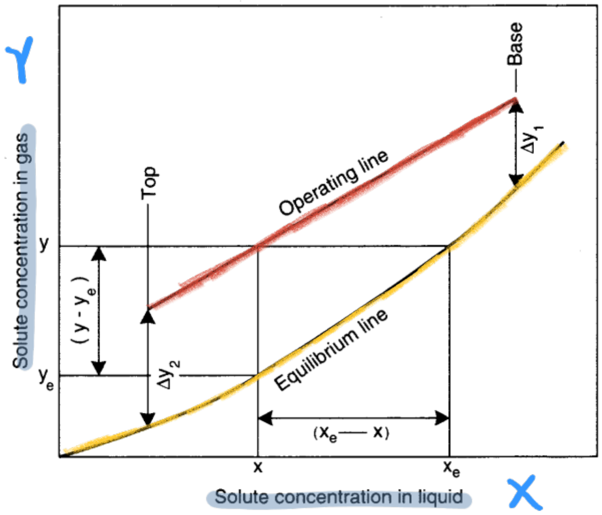


Figure 11.39. Gas absorption concentration relationships

$$\boxed{\begin{aligned} Z &= \mathbf{H}_{OG}\mathbf{N}_{OG} \\ Z &= \mathbf{H}_{OL}\mathbf{N}_{OL} \end{aligned}}$$

$$Z = \frac{G_m}{K_G a P} \int_{y_2}^{y_1} \frac{dy}{y - y_e} \qquad \text{----------(1)}$$

in terms of the overall gas phase mass transfer coefficient K_G and the gas composition. Or,

$$Z = \frac{L_m}{K_L a C_t} \int_{x_2}^{x_1} \frac{dx}{x_e - x} \qquad \text{---------------(2)}$$

in terms of the overall liquid-phase mass-transfer coefficient *KL* and the liquid composition,

where G_m = molar gas flow-rate per unit cross-sectional area,
L_m = molar liquid flow-rate per unit cross-sectional area,
a = interfacial surface area per unit volume,
P = total pressure,
C_t = total molar concentration,
y_1 and y_2 = the mol fractions of the solute in the gas at the bottom and top of the column, respectively,
x_1 and x_2 = the mol fractions of the solute in the liquid at the bottom and top of the column, respectively,
x_e = the concentration in the liquid that would be in equilibrium with the gas concentration at any point,
y_e = the concentration in the gas that would be in equilibrium with the liquid concentration at any point.

where $\mathbf{H}_{OG}$ is the height of an overall gas-phase transfer unit

$$\boxed{H_{OG} = \frac{G_m}{K_G a P}} \qquad \text{----(3)}$$

$\mathbf{H}_{OL}$ is the height of an overall liquid-phase transfer unit

$$\boxed{H_{OL} = \frac{L_m}{K_L a C_t}} \qquad \text{------(4)}$$

The number of overall gas-phase transfer units is often more conveniently expressed in terms of the partial pressure of the solute gas.

$$\mathbf{N}_{OG} = \int_{p_1}^{p_2} \frac{\mathrm{d}p}{p - p_e} \qquad \text{---------------(5)}$$

The relationship between the overall height of a transfer unit and the individual film transfer units $\mathbf{H}_L$ and $\mathbf{H}_G$, which are based on the concentration driving force across the liquid and gas films, is given by:

$\underline{G\Delta Y = L\Delta X}$
$\underline{\Delta Y/\Delta X = L/G}$

$$\mathbf{H}_{OG} = \mathbf{H}_G + m\frac{G_m}{L_m}\mathbf{H}_L \quad \text{--------(8)}$$

$$\mathbf{H}_{OL} = \mathbf{H}_L + \frac{L_m}{mG_m}\mathbf{H}_G \quad \text{------(9)}$$

where *m* is the slope of the equilibrium line and *Lm/Gm* the slope of the operating line.
The number of transfer units is obtained by graphical or numerical integration
Special cases:

Where the operating and equilibrium lines are straight, and they can usually be considered to be so for dilute systems, the number of transfer units is given by:

$$\mathbf{N}_{OG} = \frac{y_1 - y_2}{\Delta y_{\text{lm}}} \quad \text{------(10)}$$

where Δy_{lm} is the log mean driving force, given by:

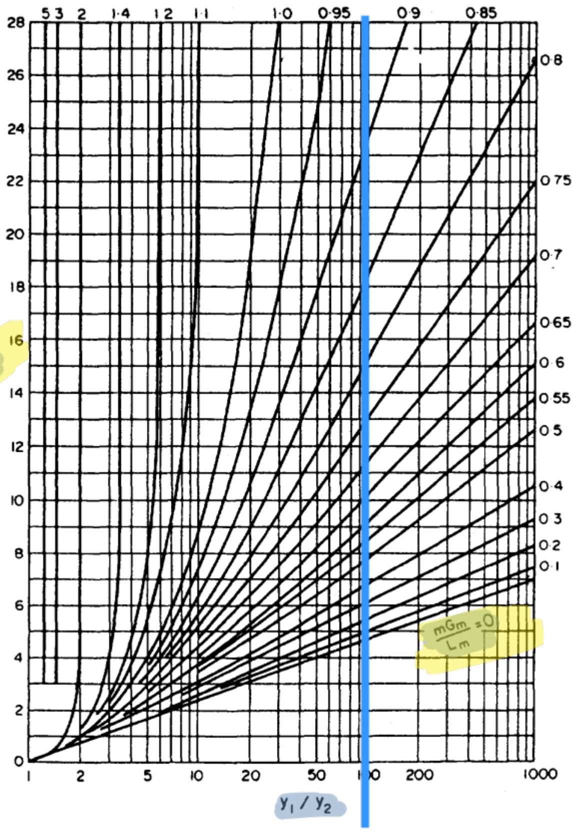
$$y_{\text{lm}} = \frac{\Delta y_1 - \Delta y_2}{\ln\left(\frac{\Delta y_1}{\Delta y_2}\right)} \quad \text{----(11)}$$

where $\Delta y_1 = y_1 - y_e$,
$\Delta y_2 = y_2 - y_e$.

If the equilibrium curve and operating lines can be taken as straight and the solvent feed essentially solute free, the number of transfer units is given by:

$$\mathbf{N}_{OG} = \frac{1}{1 - \left(\frac{mG_m}{L_m}\right)} \ln\left[\left(1 - \frac{mG_m}{L_m}\right)\frac{y_1}{y_2} + \frac{mG_m}{L_m}\right] \quad \text{----(12)}$$

It can be seen from Figure 11.40 that the number of stages required for a given separation is very dependent on the flow rate *Lm,* If the solvent rate is not set by other process considerations, Figure 11.40 can be used to make quick estimates of the column height at different flow rates to find the most economic value. It is suggested that the optimum value for the term *mGm/Lm* will lie between 0.7 to 0.8.



To find Nog →

Figure 11.40. Number of transfer units N_{OG} as a function of y_1/y_2 with mG_m/L_m as parameter

Column diameter (capacity)

The capacity of a packed column is determined by its cross-sectional area. Normally, the column will be designed to operate at the highest economical pressure drop, to ensure good liquid and gas distribution. For random packing the pressure drop will not normally exceed 80 mm of water per meter of packing height. At this value the gas velocity will be about 80 per cent of the flooding velocity. Recommended design values, mm water per m packing, are:

Absorbers and strippers	15 to 50
Distillation, atmospheric and moderate pressure	40 to 80

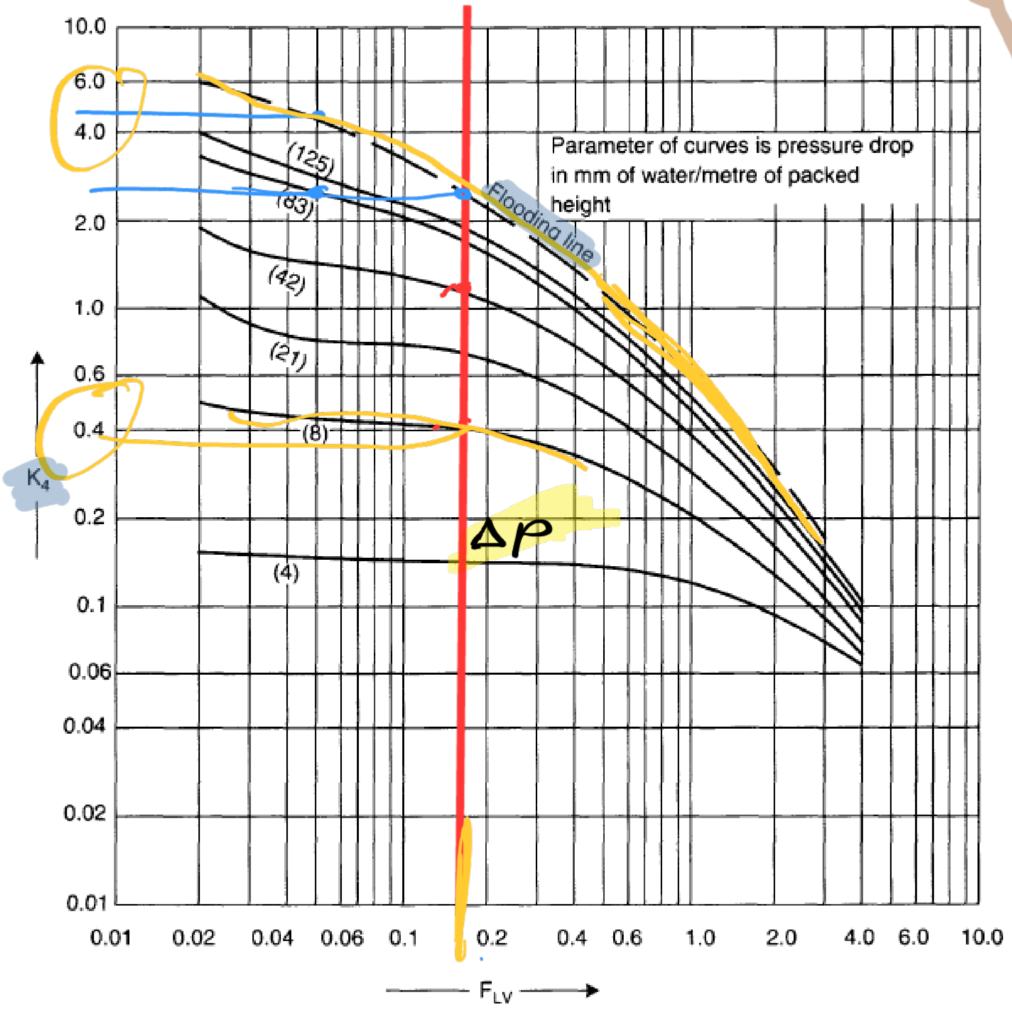
The column cross-sectional area and diameter for the selected pressure drop can be determined from the generalized pressure-drop correlation given in Figure 11.44. The figure correlates the liquid and vapor flow rates, system physical properties and packing characteristics, with the gas mass flow-rate per unit cross-sectional area; with lines of constant pressure drop as a parameter.

The term K_4 on Figure 11.44 is the function:

$$K_4 = \frac{13.1(V_w^*)^2 F_p \left(\frac{\mu_L}{\rho_L}\right)^{0.1}}{\rho_v(\rho_L - \rho_v)}$$

where V_w^* = gas mass flow-rate per unit column cross-sectional area, kg/m^{2}s
F_p = packing factor, characteristic of the size and type of packing, see Table 11.3, m^{-1}.
μ_L = liquid viscosity, Ns/m^2
ρ_L, ρ_v = liquid and vapour densities, kg/m^3

V_w^* → * means per area. and w means mass flow rate.



$$F_{LV} = \frac{L_W^*}{V_W^*}\sqrt{\frac{\rho_V}{\rho_L}}$$

Figure 11.44. Generalised pressure drop correlation, adapted from a figure by the Norton Co. with permission

$$\rho_v = \frac{Mwt}{22.4} \times \frac{273}{T_2} \times \frac{P_2}{1}$$

The values of the flow factor F_{LV} given in Figure 11.44 covers the range that will generally give satisfactory column performance.

Cornell's method

$$\mathbf{H}_G = 0.011\psi_h (Sc)_v^{0.5} \left(\frac{D_c}{0.305}\right)^{1.11} \left(\frac{Z}{3.05}\right)^{0.33} \Big/ (L_w^* f_1 f_2 f_3)^{0.5}$$

$$\mathbf{H}_L = 0.305\phi_h (Sc)_L^{0.5} K_3 \left(\frac{Z}{3.05}\right)^{0.15}$$

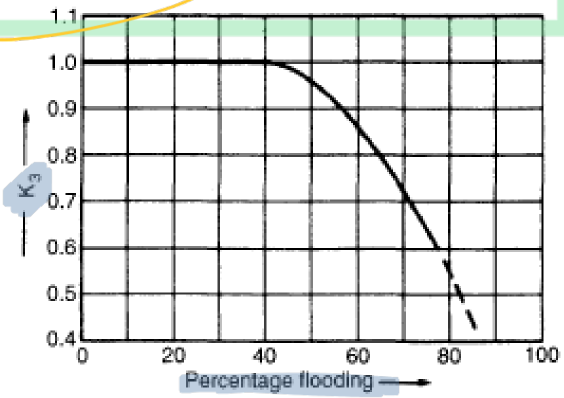


Figure 11.41. Percentage flooding correction factor

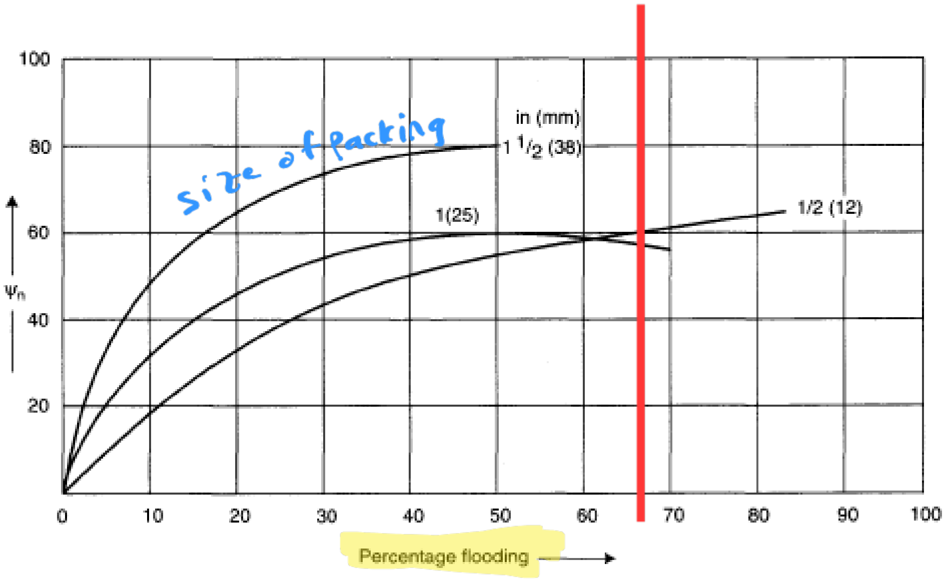


Figure 11.42. Factor for $\mathbf{H}_G$ for Berl saddles

where $\mathbf{H}_G$ = height of a gas-phase transfer unit, m,
$\mathbf{H}_L$ = height of a liquid-phase transfer unit, m,
$(Sc)_v$ = gas Schmidt number = $(\mu_v/\rho_v D_v)$,
$(Sc)_L$ = liquid Schmidt number = $(\mu_L/\rho_L D_L)$,
D_c = column diameter, m,
Z = column height, m,
K_3 = percentage flooding correction factor, from Figure 11.41,
ψ_h = $\mathbf{H}_G$ factor from Figure 11.42,
ϕ_h = $\mathbf{H}_L$ factor from Figure 11.43,
L_w^* = liquid mass flow-rate per unit area column cross-sectional area, kg/m^{2}s,
f_1 = liquid viscosity correction factor = $(\mu_L/\mu_w)^{0.16}$,

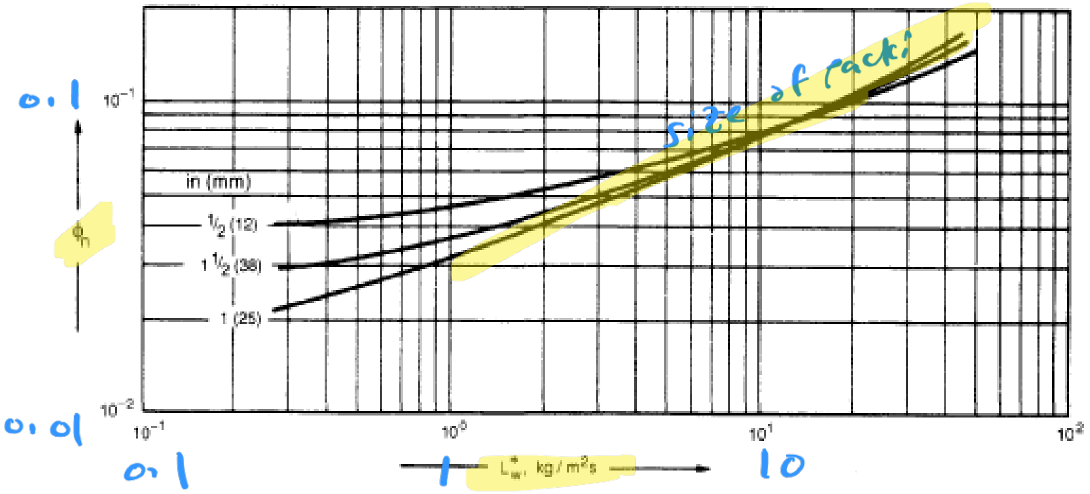


Figure 11.43. Factor for $\mathbf{H}_L$ for Berl saddles

f_2 = liquid density correction factor = $(\rho_w/\rho_L)^{1.25}$,
f_3 = surface tension correction factor = $(\sigma_w/\sigma_L)^{0.8}$,

where the suffix w refers to the physical properties of water at 20°C; all other physical properties are evaluated at the column conditions.

Design procedures:
1-Calculate Y_1/Y_2 from a given recovery.
2-Choose optimum $\dot{m}G_m/L_m$.
3-Evaluate L_m.
4-Use Figure 11.4 to find NOG at $\dot{m}G_m/L_m$ and Y_1/Y_2.
5-Choose suitable ΔP.
6- Calculate F_{LV}.
7-Find K_4 from Figure 11.44.
8-Calculate percentage flooding.
9-Calculate V_W^* from equation.
10-Calculate D_C from:
Area = V_W/V_W^*
$A = \pi/4(D_C)^2$
11-Calculate H_G, H_L and HOG from equation.
12-Calculate Z from : Z = NOG.HOG

Note: If there is available data about other packing we may use their information in the design of packed absorber.
Note: Increasing $\dot{m}G_m/L_m$ → decreasing L_m → decreasing wetting rate → decreasing mass transfer → increasing NOG → increasing no. of stages → increasing Z.
Note: If a solution is diluted as in the example, consider it as air-water mixture.

Example 11.14

L

G

Sulphur dioxide produced by the combustion of sulphur in air is absorbed in water. Pure SO_2 is then recovered from the solution by steam stripping. Make a preliminary design for the absorption column. The feed will be 5000 kg/h of gas containing 8 per cent v/v SO_2. The gas will be cooled to 20°C. A 95 per cent recovery of the sulphur dioxide is required.

Solution

As the solubility of SO_2 in water is high, operation at atmospheric pressure should be satisfactory. The feed-water temperature will be taken as 20°C, a reasonable design value.

Solubility data

SO_2	per cent w/w solution	0.05	0.1	0.15	0.2	0.3	0.5	0.7	1.0	1.5
	Partial press. gas mmHg	1.2	3.2	5.8	8.5	14.1	26	39	59	92

Partial pressure of SO_2 in the feed = (8/100) × 760 = 60.8 mm Hg
These figures are plotted in Figure (*d*).

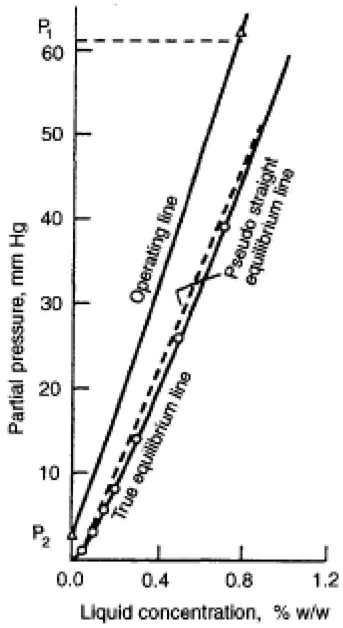


Figure (*d*). SO_2 absorber design (Example 11.14)

Number of stages

Partial pressure in the exit gas at 95 per cent recovery $= 60.8 \times 0.05 = 3.04$ mm Hg

Over this range of partial pressure the equilibrium line is essentially straight so Figure 11.40 can be used to estimate the number of stages needed.

The use of Figure 11.40 will slightly overestimate the number of stages and a more accurate estimate would be made by graphical integration of equation 11.104; but this is not justified in view of the uncertainty in the prediction of the transfer unit height.

Molecular weights: $SO_2 = 64$, $H_2O = 18$, air $= 29$

Slope of equilibrium line

From the data: partial pressure at 1.0% w/w $SO_2 = 59$ mm Hg.

$$\text{Mol. fraction in vapour} = \frac{59}{760} = 0.0776$$

$$\text{Mol. fraction in liquid} = \frac{\frac{1}{64}}{\frac{1}{64} + \frac{99}{18}} = 0.0028$$

$$m = \frac{0.0776}{0.0028} = 27.4$$

To decide the most economic water flow-rate, the stripper design should be considered together with the absorption design, but for the purpose of this example the absorption design will be considered alone. Using Figure 11.40 the number of stages required at different water rates will be determined and the "optimum" rate chosen:

$$\frac{y_1}{y_2} = \frac{p_1}{p_2} = \frac{60.8}{3.04} = 20$$

$m\dfrac{G_m}{L_m}$	0.5	0.6	0.7	0.8	0.9	1.0
N_{OG}	3.7	4.1	6.3	8	10.8	19.0

It can be seen that the "optimum" will be between $mG_m/L_m = 0.6$ to 0.8, as would be expected. Below 0.6 there is only a small decrease in the number of stages required with increasing liquid rate; and above 0.8 the number of stages increases rapidly with decreasing liquid rate.

Check the liquid outlet composition at 0.6 and 0.8:

$$\text{Material balance } L_m x_1 = G_m(y_1 - y_2)$$

$$\text{so } x_1 = \frac{G_m}{L_m}(0.08 \times 0.95) = \frac{m}{27.4}\frac{G_m}{L_m}(0.076)$$

$$\text{at } \frac{mG_m}{L_m} = 0.6, x_1 = 1.66 \times 10^{-3} \text{ mol fraction}$$

$$\text{at } \frac{mG_m}{L_m} = 0.8, x_1 = 2.22 \times 10^{-3} \text{ mol fraction}$$

Use 0.8, as the higher concentration will favour the stripper design and operati without significantly increasing the number of stages needed in the absorber.

$$N_{OG} = \underline{\underline{8}}$$

Column diameter

The physical properties of the gas can be taken as those for air, as the concentration SO_2 is low.

$$\text{Gas flow-rate} = \frac{5000}{3600} = 1.39 \text{ kg/s}, = \frac{1.39}{29} = 0.048 \text{ kmol/s}$$

$$\text{Liquid flow-rate} = \frac{27.4}{0.8} \times 0.048 = 1.64 \text{ kmol/s}$$

$$= 29.5 \text{ kg/s.}$$

Select 38 mm ($1\frac{1}{2}$ in.) ceramic Intalox saddles.

From Table 11.3, $F_p = 170 \text{ m}^{-1}$

$$\text{Gas density at } 20^\circ\text{C} = \frac{29}{22.4} \times \frac{273}{293} = 1.21 \text{ kg/m}^3$$

$$\text{Liquid density} \simeq 1000 \text{ kg/m}^3$$

$$\text{Liquid viscosity} = 10^{-3} \text{ Ns/m}^2$$

$$\frac{L_W^*}{V_W^*}\sqrt{\frac{\rho_v}{\rho_L}} = \frac{29.5}{1.39}\sqrt{\frac{1.21}{10^3}} = 0.74$$

Design for a pressure drop of 20 mm H_2O/m packing

From Figure 11.44,

$$K_4 = 0.35$$

$$\text{At flooding } K_4 = 0.8$$

$$\text{Percentage flooding} = \left[\frac{K_4 \text{ at design pressure drop}}{K_4 \text{ at flooding}}\right]^{1/2}$$

$$\text{Percentage flooding} = \sqrt{\frac{0.35}{0.8}} \times 100 = 66 \text{ per cent, satisfactory.}$$

From equation 11.118

$$V_W^* = \left[\frac{K_4 \rho_V (\rho_L - \rho_v)}{13.1 F_p (\mu_L/\rho_L)^{0.1}}\right]^{1/2}$$

$$= \left[\frac{0.35 \times 1.21(1000 - 1.21)}{13.1 \times 170(10^{-3}/10^3)^{0.1}}\right]^{1/2} = 0.87 \text{ kg/m}^2\text{s}$$

$$\boxed{K_4 = \frac{13.1 (V_w^*)^2 F_p \left(\dfrac{\mu_L}{\rho_L}\right)^{0.1}}{\rho_v(\rho_L - \rho_v)} \qquad (11.118)}$$

$$\text{Column area required} = \frac{1.39}{0.87} = 1.6 \text{ m}^2$$

$$\text{Diameter} = \sqrt{\frac{4}{\pi} \times 1.6} = 1.43 \text{ m}$$

Round off to $\underline{\underline{1.50 \text{ m}}}$

$$\text{Column area} = \frac{\pi}{4} \times 1.5^2 = 1.77 \text{ m}^2$$

$$\text{Packing size to column diameter ratio} = \frac{1.5}{38 \times 10^{-3}} = 39,$$

A larger packing size could be considered.
Percentage flooding at selected diameter

$$= 66 \times \frac{1.6}{1.77} = 60 \text{ per cent,}$$

Could consider reducing column diameter.

Estimation of H_{OG}

Cornell's method

$$D_L = 1.7 \times 10^{-9}\ \text{m}^2/\text{s}$$

$$D_v = 1.45 \times 10^{-5}\ \text{m}^2/\text{s}$$

$$\mu_v = 0.018 \times 10^{-3}\ \text{Ns/m}^2$$

$$(Sc)_v = \frac{0.018 \times 10^{-3}}{1.21 \times 1.45 \times 10^{-5}} = 1.04$$

$$(Sc)_L = \frac{10^{-3}}{1000 \times 1.7 \times 10^{-9}} = 588$$

$$L_W^* = \frac{29.5}{1.77} = 16.7\ \text{kg/s m}^2$$

From Figure 11.41, at 60 per cent flooding, $K_3 = 0.85$.
From Figure 11.42, at 60 per cent flooding, $\psi_h = 80$.
From Figure 11.43, at $L_W^* = 16.7$, $\phi_h = 0.1$.

$\mathbf{H}_{OG}$ can be expected to be around 1 m, so as a first estimate Z can be taken as 8 m. The column diameter is greater than 0.6 m so the diameter correction term will be taken as 2.3.

For design purposes the diameter correction term should be taken as a fixed value of 2.3 for columns above 0.6 m.

$$\mathbf{H}_L = 0.305 \times 0.1(588)^{0.5} \times 0.85\left(\frac{8}{3.05}\right)^{0.15} = 0.7\ \text{m}$$

As the liquid temperature has been taken as 20°C, and the liquid is water,

$$f_1 = f_2 = f_3 = 1$$

$$\mathbf{H}_G = 0.011 \times 80(1.04)^{0.5}(2.3)\left(\frac{8}{3.05}\right)^{0.33} \Big/ (16.7)^{0.5} = 0.7\ \text{m}$$

$$\mathbf{H}_{OG} = 0.7 + 0.8 \times 0.7 = 1.3\ \text{m}$$

$Z = 8 \times 1.3 = 10.4$ m, close enough to the estimated value.

round to 11m.

Sheet #4

Design an absorption column for this duty.

Q.1

An acetone–air mixture containing 0.015 mole fraction of acetone has the mole fraction reduced to 5 per cent of this value by countercurrent absorption with water in a packed tower. The gas flow rate *G* is 1 kg/s of air and the water flow rate entering is 1.6 kg/s. For this system, Henry's law holds and *ye* = 1.75*x*, where *ye* is the mole fraction of acetone in the vapor in equilibrium with a mole fraction *x* in the liquid.

Q.2

Ammonia is removed from a 10 per cent ammonia–air mixture by scrubbing with water in a packed tower, so that 99 per cent of the ammonia is removed. The gas enters at a rate of 6000kg/hr at 30°C and 150psig. Diffusivity (NH_3 -water) =6.83E-5ft^2/hr, Diffusivity (NH_3 -air) 0.065ft^2/hr.

Equilibrium data

y	0.01	0.02	0.035	0.06
x	0.04	0.06	0.08	0.1

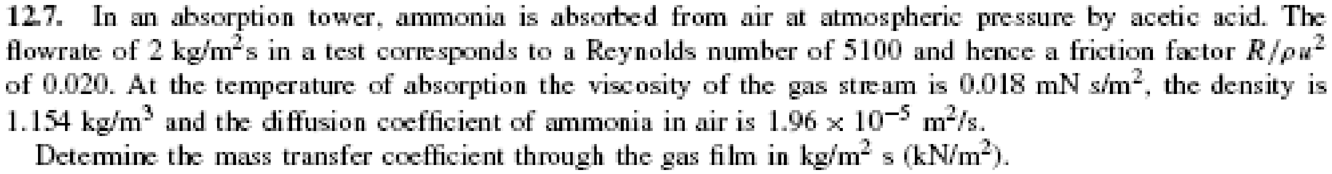
Q.3 Repeat Q.2

- For 25mm packing.

- For 97% recovery.

Problem / class B. final date 5/4/2009

A process stream containing nitrogen and carbon dioxide is to be scrubbed with water for CO_2 removal. The inlet gas contains 40.6mol/ hr of CO_2 and 365.4 mol/hr N_2. the absorber is to operate at 32°C and 3 atm. Design a suitable absorber for 95% recovery.



12.7. In an absorption tower, ammonia is absorbed from air at atmospheric pressure by acetic acid. The flowrate of 2 kg/m^2s in a test corresponds to a Reynolds number of 5100 and hence a friction factor $R/\rho u^2$ of 0.020. At the temperature of absorption the viscosity of the gas stream is 0.018 mN s/m^2, the density is 1.154 kg/m^3 and the diffusion coefficient of ammonia in air is 1.96×10^{-5} m^2/s.

Determine the mass transfer coefficient through the gas film in kg/m^2 s (kN/m^2).