MASS TRANSFER AND GAS ABSORPTION EQUIPMENT

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TOPICS

- Review of heat transfer and heat exchangers
- Some fundamental aspects of mass transfer
- Analysis of a packed bed gas absorption tower

Imperial Flowsheet

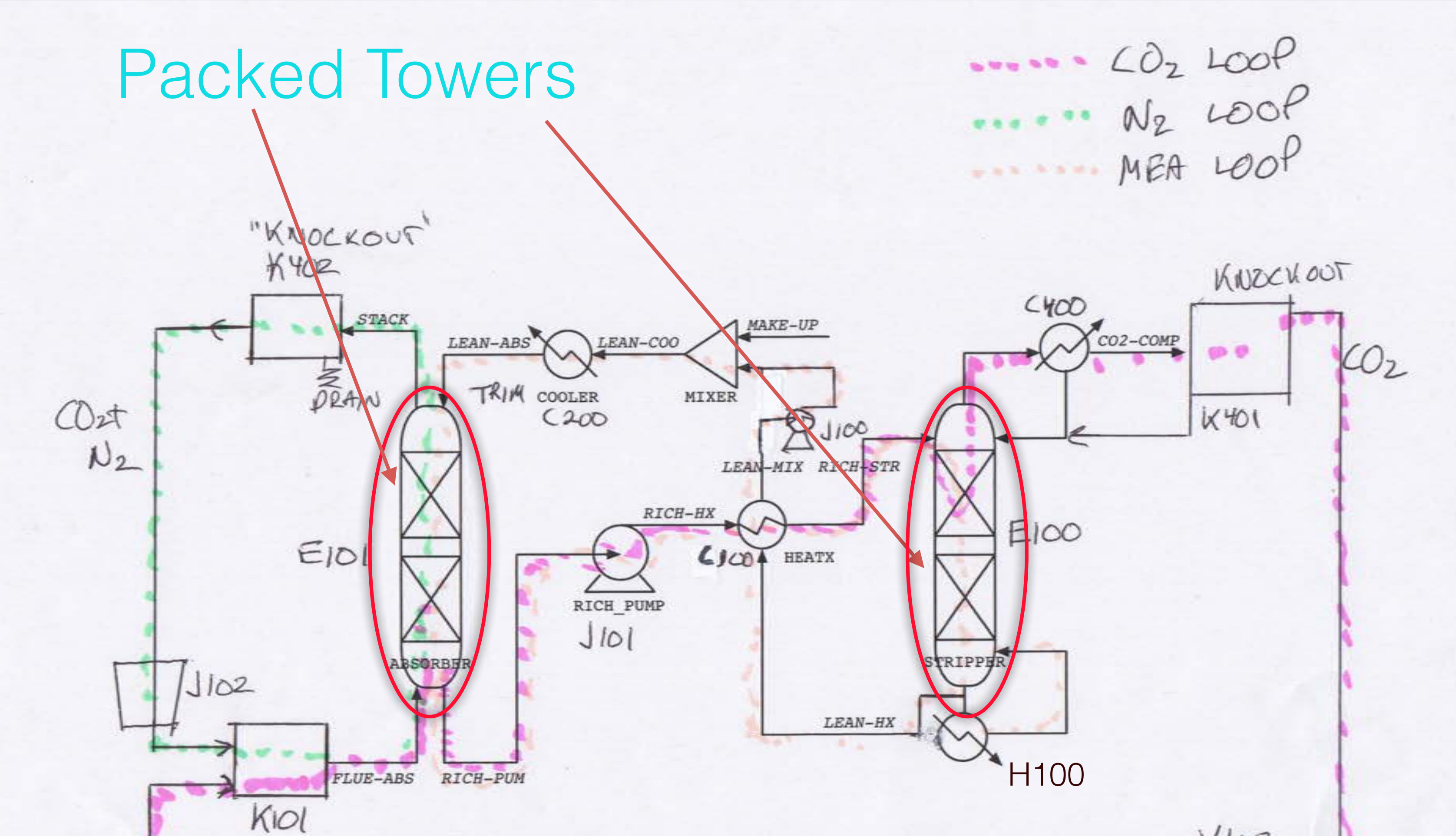


Figure 4.1: Base MEA absorption process flowsheet

GAS ABSORPTION/STRIPING

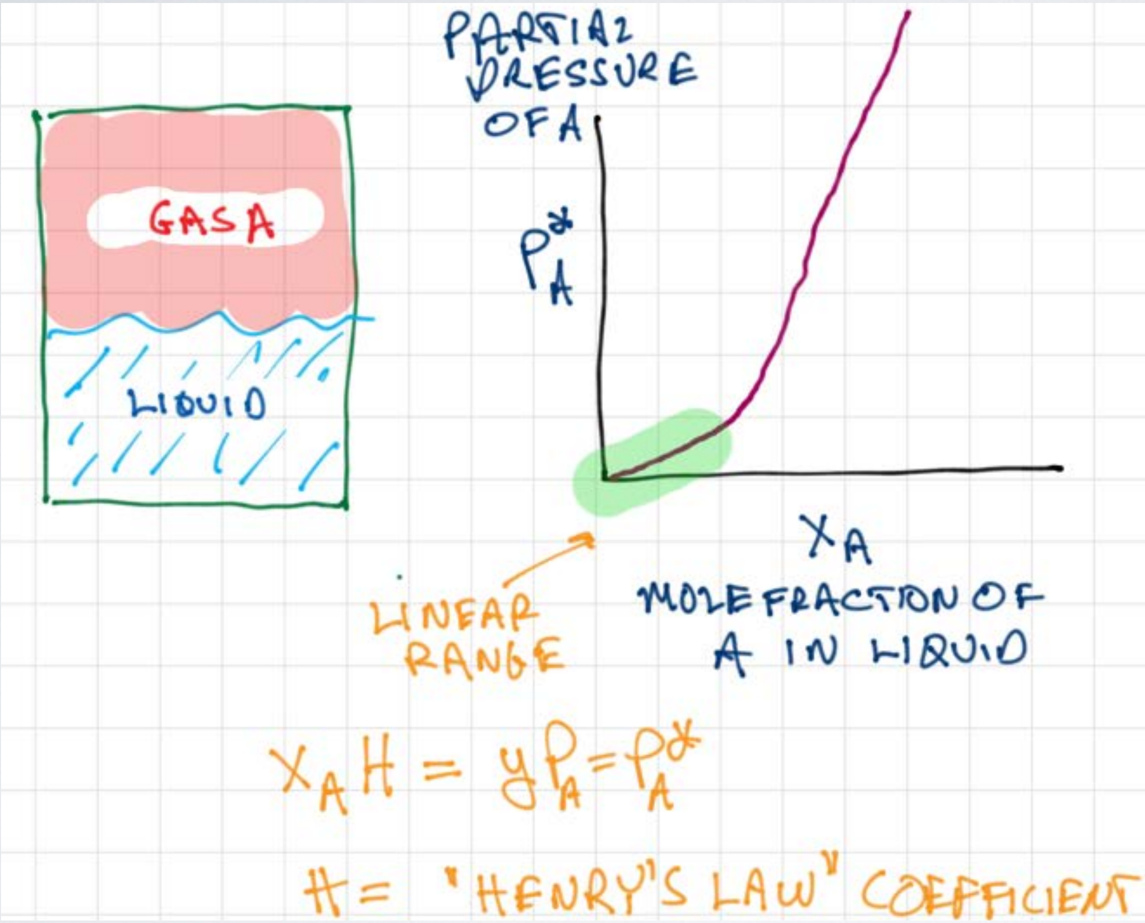
- Mass transfer is the main “business” of the process!
- The absorber is a column packed with a structured packing
  - Inlet gas mixture of N_2 and CO_2
  - Absorption liquid is monoethanolamine in water
    - reacts reversibly with CO_2, to selectively remove CO_2 from N_2
- The stripping column is packed with a random, metal packing
  - Steam flow to reboiler, boils the mixture, reversing the reaction and the steam that is generated helps to carry the CO_2 out of the column

ANALYSIS: TWO BASIC PRINCIPLES

- Conservation of mass
  - Keep track of chemical species in the two different flows
- Rate of interphase transfer equation
  - Analogous to Newton's Law of cooling
  - Driving force is a concentration difference:
    - Gas-liquid phase equilibria may include reaction

GAS-LIQUID PHASE EQUILIBRIA

- We need to make sure we understand the underlying thermodynamics



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Henry's Law is often a useful approximation for low solubility gases.

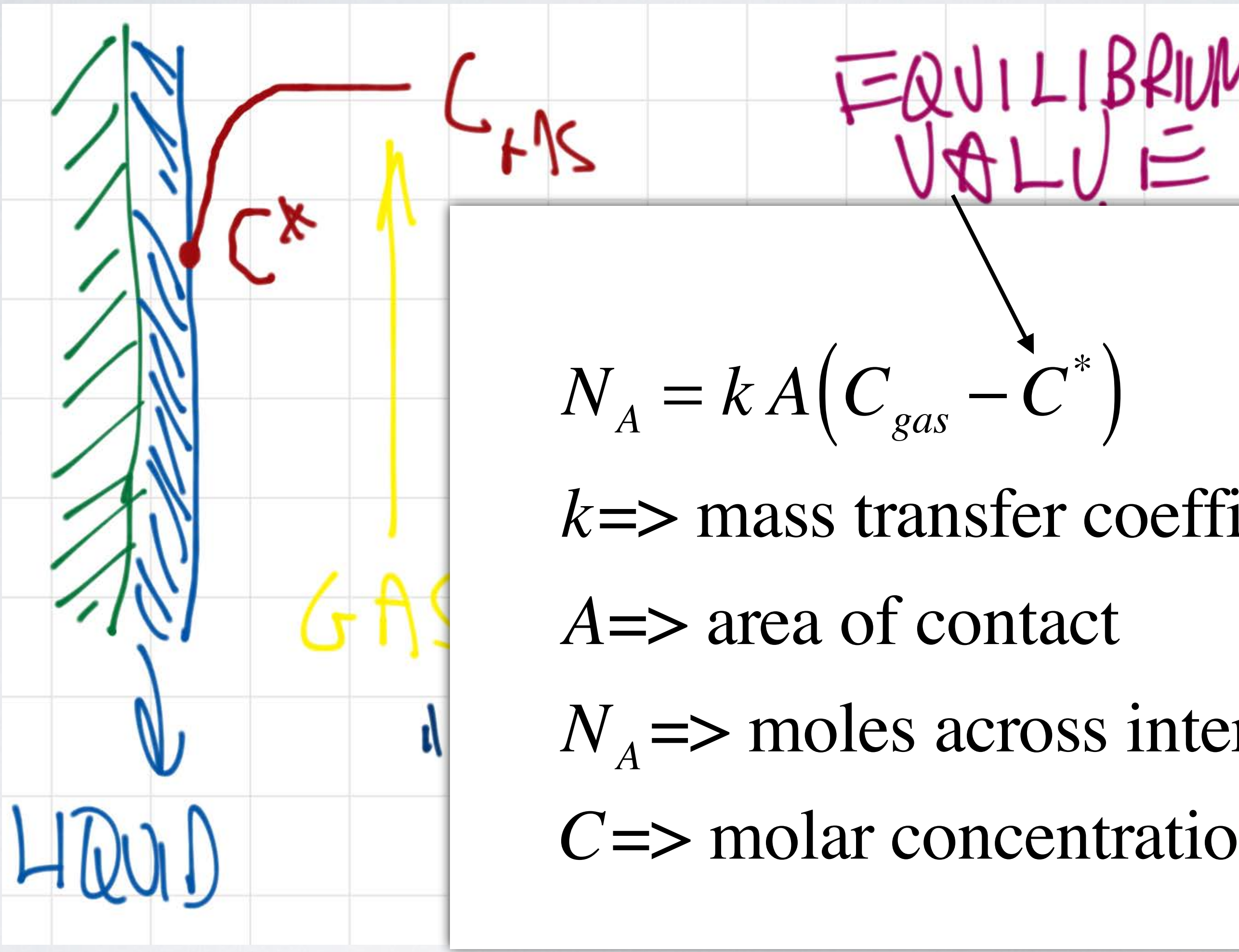
Note that as H increases, the solubility decreases.

Also, $H = H(T)$. In most cases, H will increase with T.

MECHANISMS OF MASS TRANSFER

- *Diffusion* (analogous to *Conduction* in heat transfer)
  - Transport by random molecular motion gases and liquids.
  - Fick's Law: $j = D\frac{\partial C}{\partial x}$ (same as Fourier's Law)
- *Convection* (essentially the same as *Convection* in heat transfer)
  - Transport by net motion of fluid. (molecular motion that is correlated, not random)
  - Describe using analog to Newton's law of cooling

MASS TRANSFER RATE EQUATION



$$N_A = k\,A\left(C_{gas} - C^*\right)$$

k=> mass transfer coefficient

A=> area of contact

N_A=> moles across interface/time

C=> molar concentration of CO_2 in gas

Flux = rate coefficient *(linear driving force)

MASS TRANSFER CORRELATIONS

- As with the heat transfer coefficient, we will get values for the mass transfer coefficient from correlations of dimensionless groups.
- The Sherwood number, *Sh*, is the group analogous to the Nusselt number. $Sh \equiv \frac{kd}{D}$ $D =$ MOLECULAR DIFFUSIVITY
- The Schmidt number, *Sc*, is the group analogous to the Prandtl number. $Sc \equiv \frac{\nu}{D}$
- The Colburn "j-factor" $j \equiv Sh/(Re\ Sc^{1/3})$ $(= f/2)$ is also used.
- For the same flow situations, the heat transfer and mass transfer correlations are exactly the same!

From R. Treybal, Mass Transfer Operations

Table 3.3 Mass transfer† for simple situations

Fluid motion	Range of conditions	Equation	Ref.
1. Inside circular pipes	Re = 4000–60 000 Sc = 0.6–3000	$j_D = 0.023\ \mathrm{Re}^{-0.17}$ $\mathrm{Sh} = 0.023\ \mathrm{Re}^{0.83}\ \mathrm{Sc}^{1/3}$	41, 52
	Re = 10 000 – 400 000 Sc > 100	$j_D = 0.0149\ \mathrm{Re}^{-0.12}$ $\mathrm{Sh} = 0.0149\ \mathrm{Re}^{0.88}\ \mathrm{Sc}^{1/3}$	44
2. Unconfined flow parallel to flat plates‡	Transfer begins at leading edge $\mathrm{Re}_x < 50\ 000$	$j_D = 0.664\ \mathrm{Re}_x^{-0.5}$	32
	$\mathrm{Re}_x = 5 \times 10^5 – 3 \times 10^7$ Pr = 0.7–380	$\mathrm{Nu} = 0.037\ \mathrm{Re}_x^{0.8}\ \mathrm{Pr}_0^{0.43}\left(\frac{\mathrm{Pr}_0}{\mathrm{Pr}_i}\right)^{0.25}$	65
	$\mathrm{Re}_x = 2 \times 10^4 – 5 \times 10^5$ Pr = 0.7–380	Between above and $\mathrm{Nu} = 0.0027\ \mathrm{Re}_x\ \mathrm{Pr}_0^{0.43}\left(\frac{\mathrm{Pr}_0}{\mathrm{Pr}_i}\right)^{0.25}$	
3. Confined gas flow parallel to a flat plate in a duct	$\mathrm{Re}_e = 2600–22\ 000$	$j_D = 0.11\ \mathrm{Re}_e^{-0.29}$	46
4. Liquid film in wetted-wall tower, transfer between liquid and gas	$\frac{4\Gamma}{\mu} = 0–1200$, ripples suppressed	Eqs. (3.18)–(3.22)	20, 37
	$\frac{4\Gamma}{\mu} = 1300–8300$	$\mathrm{Sh} = (1.76 \times 10^{-5})\left(\frac{4\Gamma}{\mu}\right)^{1.506} \mathrm{Sc}^{0.5}$	

5.	Perpendicular to single cylinders	Re = 400–25 000 Sc = 0.6–2.6	$\frac{k_G p_t}{G_M} Sc^{0.56} = 0.281\ Re'^{0.4}$	5
		$Re' = 0.1–10^5$ Pr = 0.7–1500	$Nu = (0.35 + 0.34\ Re'^{0.5} + 0.15\ Re'^{0.58})\ Pr^{0.3}$	16, 21, 42
6.	Past single spheres	Sc = 0.6–3200 $Re''\ Sc^{0.5}$ = 1.8–600 000	$Sh = Sh_0 + 0.347(Re''\ Sc^{0.5})^{0.62}$ $Sh_0 = \begin{cases} 2.0 + 0.569(Gr_D\ Sc)^{0.250} & Gr_D\ Sc < 10^8 \\ 2.0 + 0.0254(Gr_D\ Sc)^{0.333}\ Sc^{0.244} & Gr_D\ Sc > 10^8 \end{cases}$	55
7.	Through fixed beds of pellets§	Re″ = 90–4000 Sc = 0.6	$j_D = j_H = \frac{2.06}{\varepsilon} Re''^{-0.575}$	4, 23, 64
		Re″ = 5000–10 300 Sc = 0.6	$j_D = 0.95 j_H = \frac{20.4}{\varepsilon} Re''^{-0.815}$	
		Re″ = 0.0016–55 Sc = 168–70 600	$j_D = \frac{1.09}{\varepsilon} Re''^{-2/3}$	
		Re″ = 5–1500 Sc = 168–70 600	$j_D = \frac{0.250}{\varepsilon} Re''^{-0.31}$	

† Average mass-transfer coefficients throughout, for constant solute concentrations at the phase surface. Generally, fluid properties are evaluated at the average conditions between the phase surface and the bulk fluid. The heat-mass-transfer analogy is valid throughout.

‡ Mass-transfer data for this case scatter badly but are reasonably well represented by setting $j_D = j_H$.

§ For fixed beds, the relation between ε and d_p is $a = 6(1 - \varepsilon)/d_p$, where a is the specific solid surface, surface per volume of bed. For mixed sizes [58]

$$d_p = \frac{\sum_{i=1}^{n} n_i d_{pi}^3}{\sum_{i=1}^{n} n_i d_{pi}^2}$$

From R. Treybal, Mass Transfer Operations

THE TWO COLUMNS

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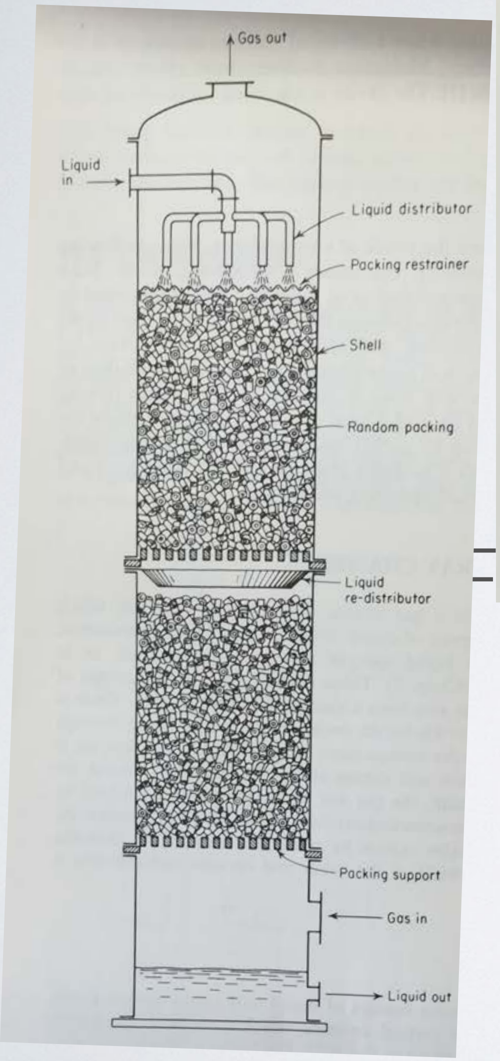
PACKED TOWER FOR GAS ABSORPTION



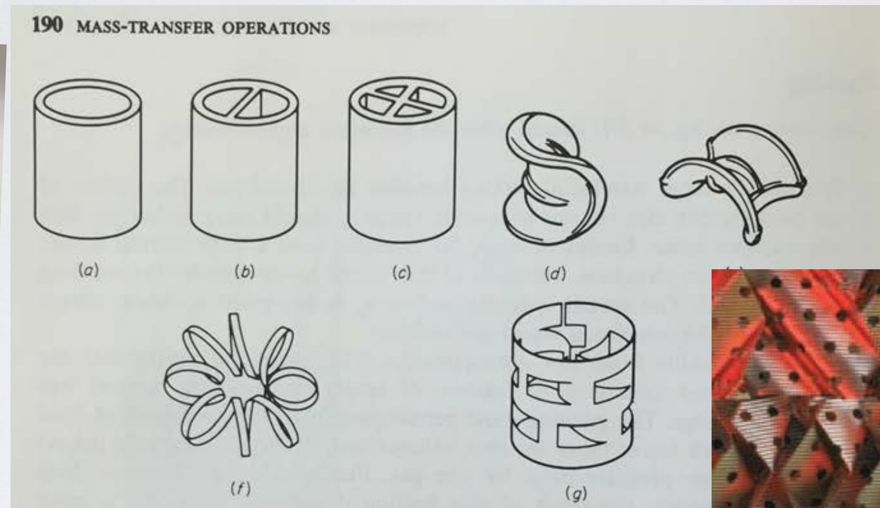


Figure 6.28 Some random tower packings: (a) Raschig rings, (b) Lessing ring, (c) … Berl saddle *(courtesy of Maurice A. Knight)*, (e) Intalox saddle *(Chemical Processing …* *Norton Co.)*, (f) Tellerette *(Ceilcote Company, Inc.)*, and (g) pall ring *(Chemical P… Division, Norton Co.)*.

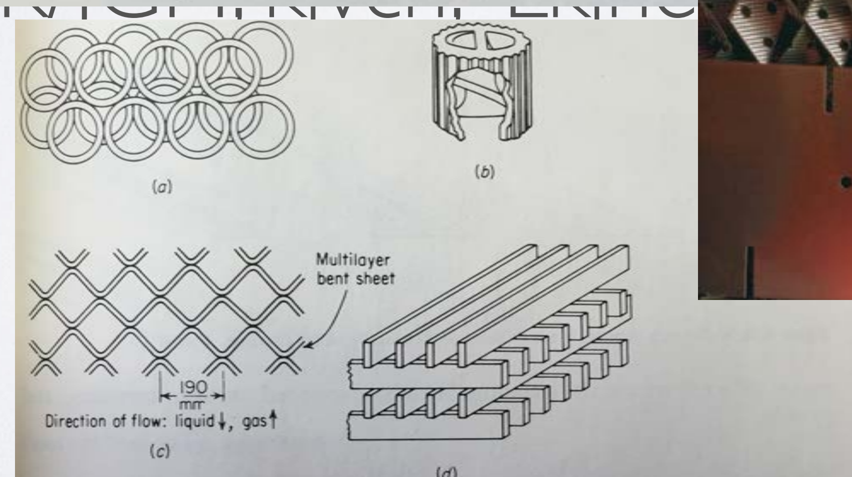
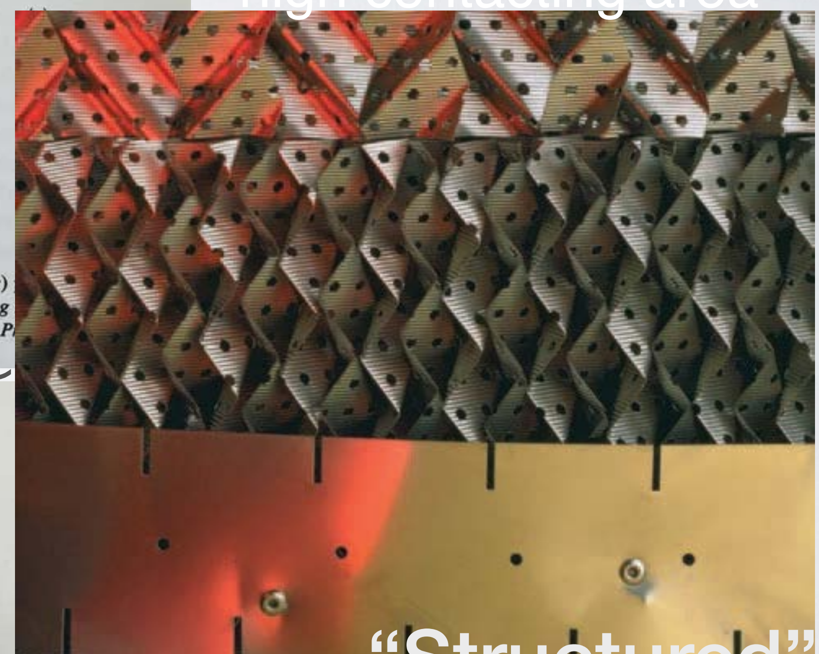


Figure 6.29 Regular, or stacked, packings: (a) Raschig rings, stacked staggered (top view), (b) double spiral ring *(Chemical Processing Products Division, Norton Co.)*, (c) section through expanded-metal-lath packing, (d) wood grids.

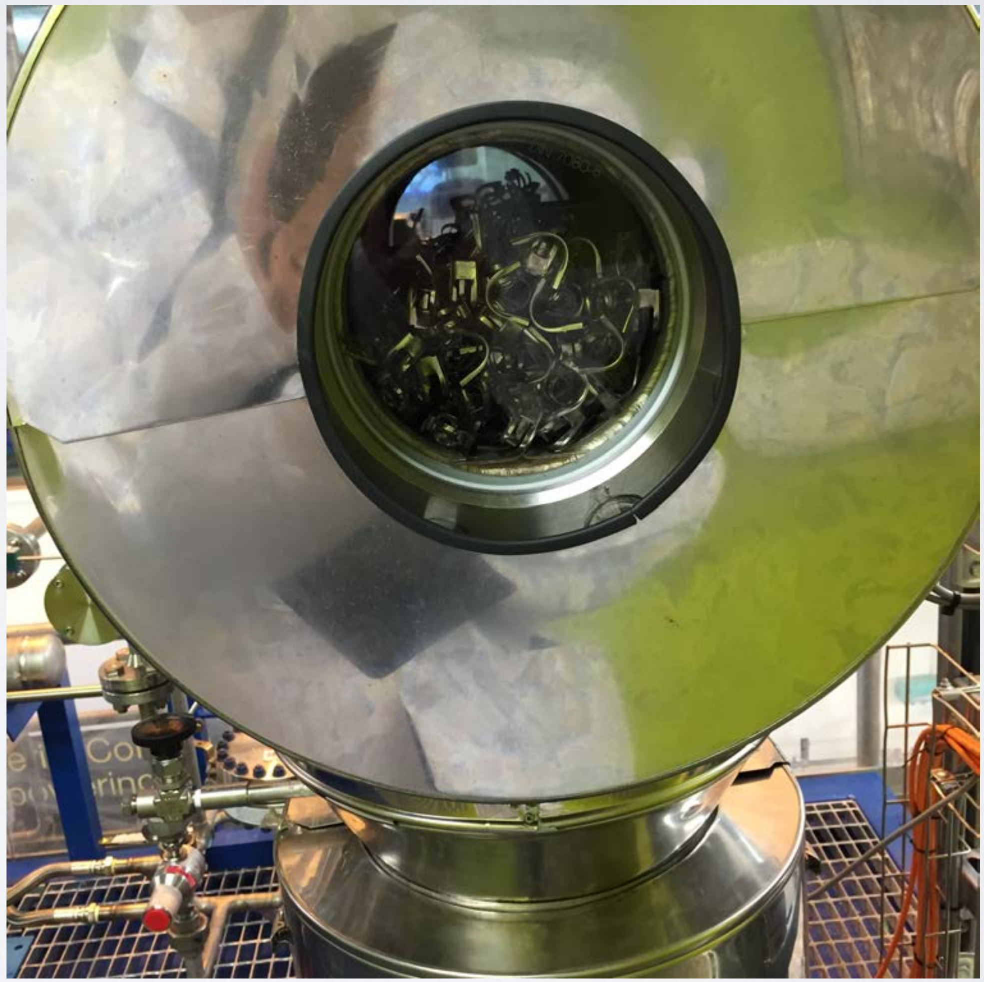
350 m^2/m^3

high contacting area



“Structured” packing

RANDOM PACKING IN STRIPPING COLUMN



PACKED TOWER

- **Countercurrent**
  - greater overall "driving force" (concentration difference) than if cocurrent (downward!)
  - (potentially) no limitation on amount of CO_2 removed
    - could contact lowest concentration exiting gas with "pure" solvent

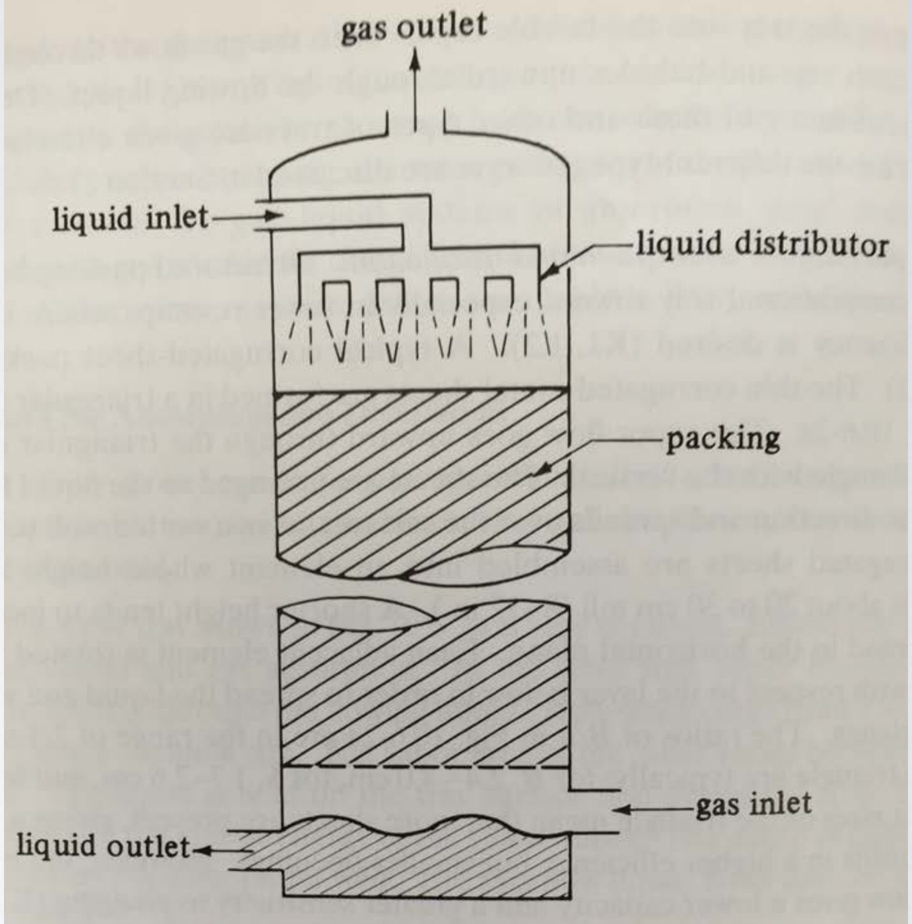


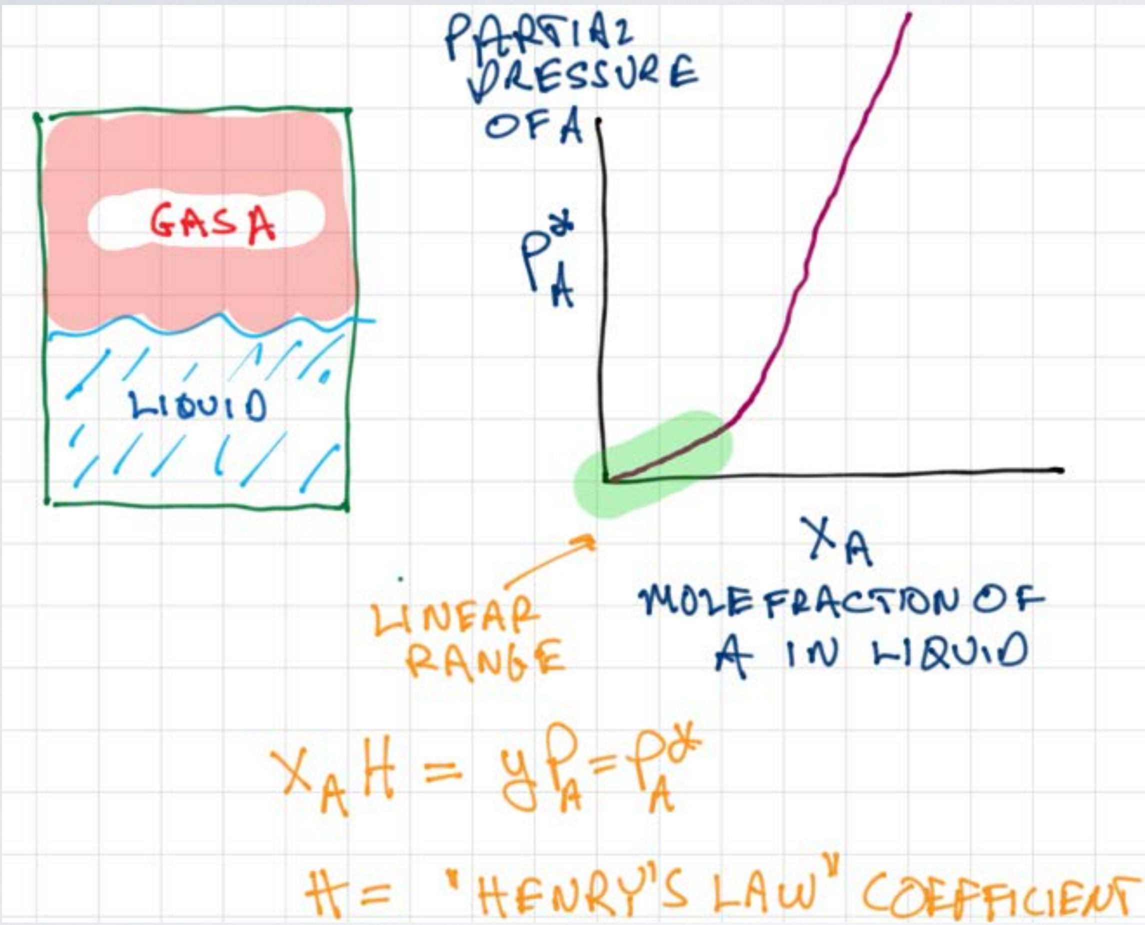
FIGURE 10.6-3. *Packed tower flows and characteristics for absorption.*

MASS TRANSFER ANALYSIS

- The analysis for gas absorption in a packed tower is very similar to the analysis we just did for a heat exchanger.
    - First you make sure that you have the mass balances correct to keep track of the total flows and the chemical species.
    - Then you need to match the change in a chemical species in a differential slice of column as represented by the mass balance, to the rate of interphase transfer from the mass transfer equation.

GAS-LIQUID PHASE EQUILIBRIA

- We need to make sure we understand the underlying thermodynamics



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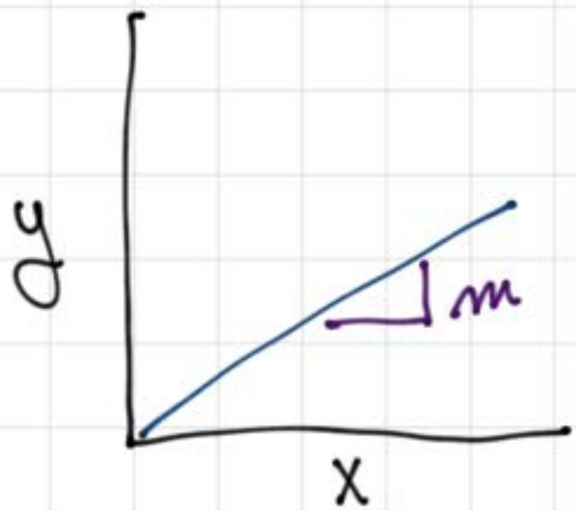
Note that as *H* increases, the solubility decreases.

Also, $H = H(T)$. In most cases, *H* will increase with *T*.

ANALYSIS FOR ABSORPTION OF A SLIGHTLY-SOLUBLE GAS

The phase equilibrium will be represented by Henry's Law, y = m x, for gas and liquid phase mole fractions, y and x.



$$y = m x = \frac{H}{P} x$$

$$m \equiv \frac{H}{P}$$

Overall mass balance for absorber. Note that L and G are "molar" flow rates for liquid and gas.

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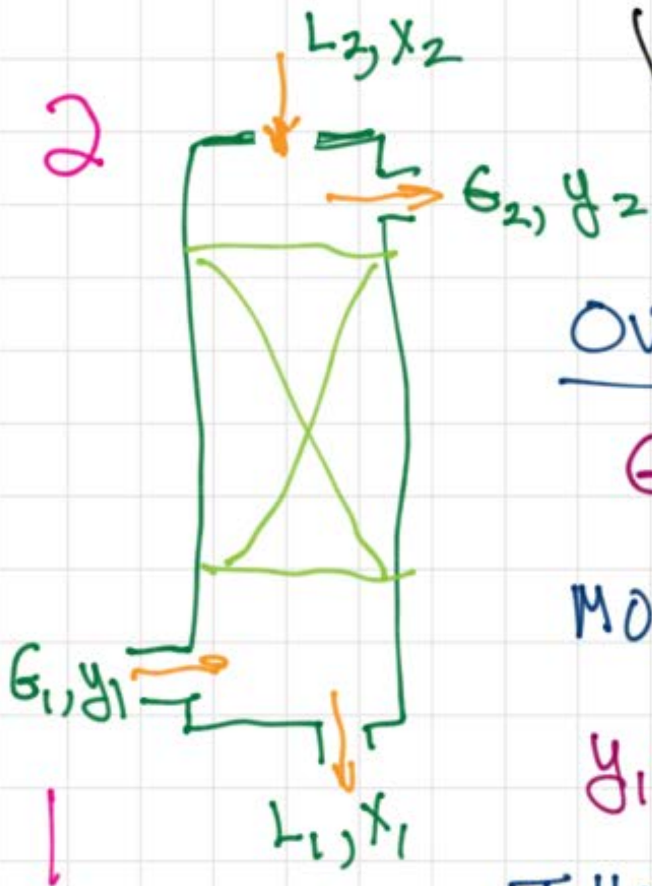


PACKED BED ABORBER

OVERALL MOLE BALANCE

$$G_1 + L_2 = G_2 + L_1$$

MOLE BALANCE FOR GAS A

$$y_1 G_1 + X_2 L_2 = y_2 G_2 + X_1 L_1$$

THIS CURVE IS CALLED THE "OPERATING LINE"
IT WILL BE A "LINE" IF
L, G = CONSTANTS

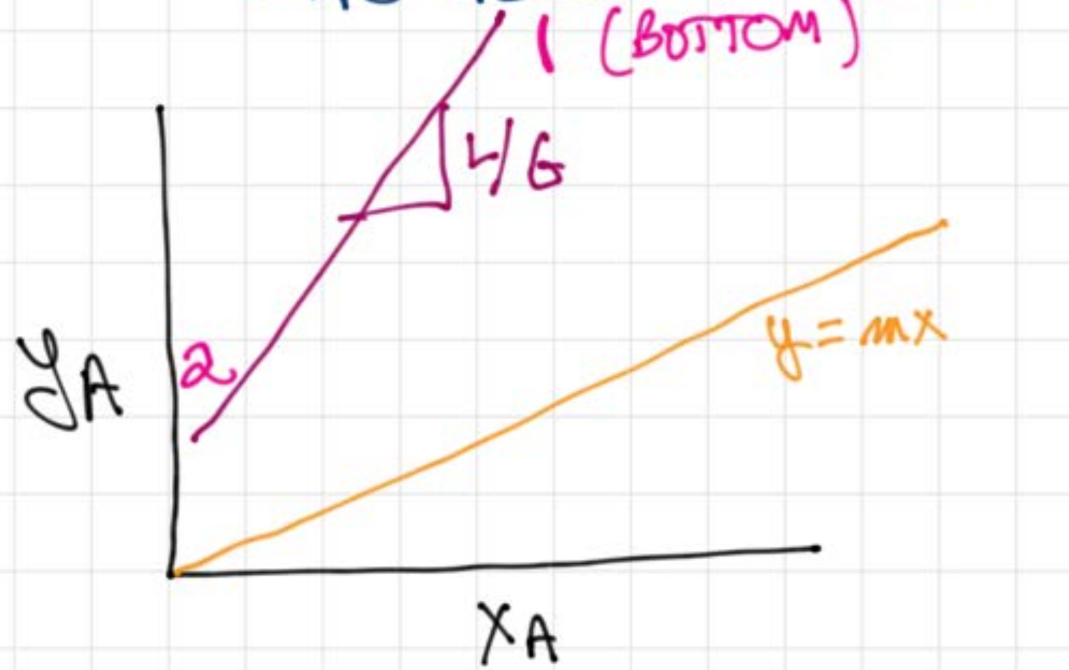
Side results..

$$(y_1 - y_2) G = (X_1 - X_2) L$$

$$\frac{y_1 - y_2}{X_1 - X_2} = \frac{L}{G}$$

$$y = \frac{L}{G} x + y_2 - X_2 \frac{L}{G}$$

GAS ABSORPTION

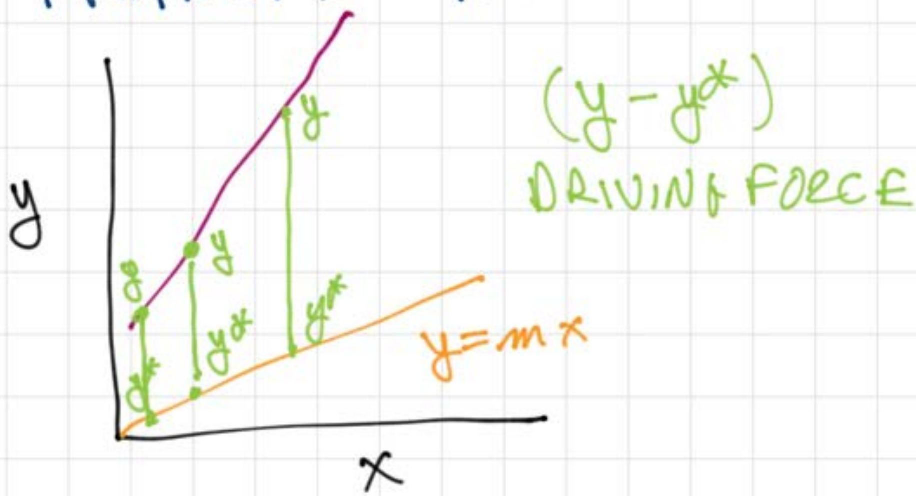


FOR DILUTE SYSTEMS, L = CONST
G = CONST

$$y = \frac{L}{G}x + y_2 - \frac{L}{G}x_2$$

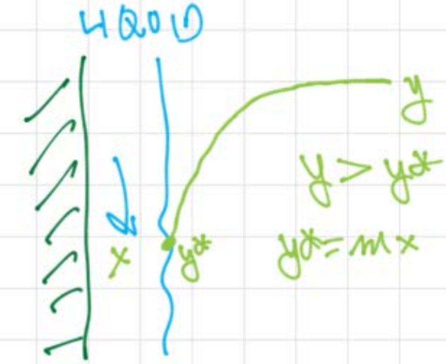
THIS TERM WOULD BE 0 FOR A PURE INLET SOLVENT

INTERFACIAL MASS TRANSFER ALONG COLUMN



ALL ALONG COLUMN $y > y^*$, THE MOLE FRACTION IN GAS IS HIGHER THAN EQUILIBRIUM SOLUBILITY

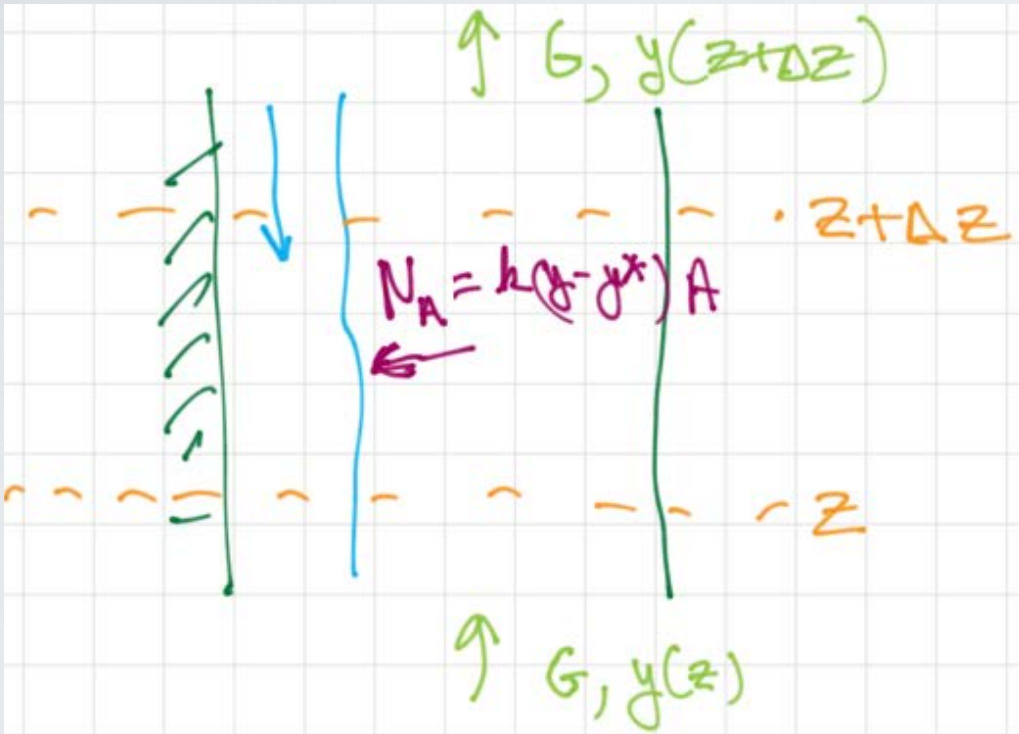
THUS: CO_2 IS ABSORBED.



$$\frac{\text{MOLES TRANSFERRED}}{\text{AREA TIME}} = \text{COEFFICIENT} \times \text{DRIVING FORCE}$$

$$N_{CO_2} = k(y - y^*)$$

As for the heat exchanger, the liquid and gas concentrations are changing along the device. We thus take the same approach of a "differential slice" through the column and equate the rate at which the gas concentration changes to the rate at which CO_2 crosses the interface.



$a \equiv \dfrac{\text{CONTACT AREA}}{\text{VOLUME OF TOWER}}$

$A_T \equiv$ CROSS-SECTION AREA OF COLUMN

$A_T \Delta z =$ VOLUME OF DIFFERENTIAL SLICE

$$G\left(y|_z - y|_{z+\Delta z}\right) = k\, a\, (y - y^*)\, A_T \Delta z$$

DESIRED RESULT: COLUMN HEIGHT, Z

SHRINK $\Delta z \rightarrow dz$

REARRANGE:

$$-\frac{G\left(y|_{z+\Delta z} - y|_{z}\right)}{\Delta z} = ka(y-y^*)A_T$$

$$-\frac{dy}{dz} = \frac{ka}{G}(y-y^*)A_T$$

THE DESIGN QUESTION IS THE HEIGHT OF COLUMN

$$\int_0^Z dz = -\frac{G/A_T}{ka}\int_{y_1}^{y_2}\frac{dy}{(y-y^*)}$$

Some conventional lingo

$$Z = -\left(\frac{G/A_T}{ka}\right)\left(\int_{y_1}^{y_2}\frac{dy}{(y-y^*)}\right)$$

Height of a transfer unit. Number of transfer units

COLUMN HEIGHT IN TERMS OF "HEIGHT OF A TRANSFER UNIT" AND NUMBER OF "TRANSFER UNITS"

$$Z = H_{OG} \; N_{OG}$$

H_{OG} WILL DECREASE AS THE MASS TRANSFER PERFORMANCE INCREASES, $ka \uparrow$

OR THE "NEED" DECREASES, $G \downarrow$

N_{OG} WILL DECREASE IF THE AVERAGE DRIVING FORCE, WHICH IS IN THE DENOMINATOR, INCREASES $(y - y^*) \uparrow$

SIMPLIFICATIONS

More simplifications

IF $y^* =$ CONST

$$-\int_{y_1}^{y_2} \frac{dy}{y-y^*} = \ln\left(\frac{y_1-y^*}{y_2-y^*}\right)$$

A CHEMICAL REACTION IN THE LIQUID CAN KEEP $y^* \approx 0$

$$-\int_{y_1}^{y_2} \frac{dy}{(y-y^*)} = \ln\left(\frac{y_1}{y_2}\right)$$

There would be some error in using this formula because the G changes by 5 - 10% in the column

The reaction of CO2 with monoethanolamine does keep the y* at essential 0 in the pilot plant gas absorber.

EVALUATE THE INTEGRAL FOR HENRY'S LAW

$$\int \frac{dy}{(y-y^*)} = \int \frac{dy}{y-mx}$$

WE NEED TO REPLACE X

USE OPERATING LINE,

$$y = \frac{L}{G}x + \left(y_1 - \frac{L}{G}x_2\right)$$

$$x = \frac{G}{L}(y-y_1) + x_2$$

$$N_{OG} = -\int_{y_1}^{y_2} \frac{dy}{y-y^*}$$

$$= \int_{y_1}^{y_2} \frac{dy}{y-mx}$$

$$= \int_{y_1}^{y_2} \frac{dy}{y-m\left(\frac{G}{L}(y-y_2)+x_2\right)}$$

CONTINUE THE EVALUATION

FOR $A \equiv \frac{L}{mG}$

$$= -\int_{y_1}^{y_2} \frac{dy}{\left(1-\frac{1}{A}\right)y + \frac{y_2}{A} - mx_2}$$

THIS CAN BE INTEGRATED...

$$= -\frac{1}{\frac{1}{A}-1} \ln\left[\frac{A(mx_2-y_1)+(y_1-y_2)}{A(mx_2-y_2)}\right]$$

$$= \frac{\ln\left[\frac{y_1-mx_2}{y_2-mx_2}\left(1-\frac{1}{A}\right)+\frac{1}{A}\right]}{1-\frac{1}{A}}$$

- Further simplification is possible for a pure solvent, $x_2 = 0$.
- Note that as *A* increases, the absorption process gets easier!

- NOG integral values from Treybal's book.

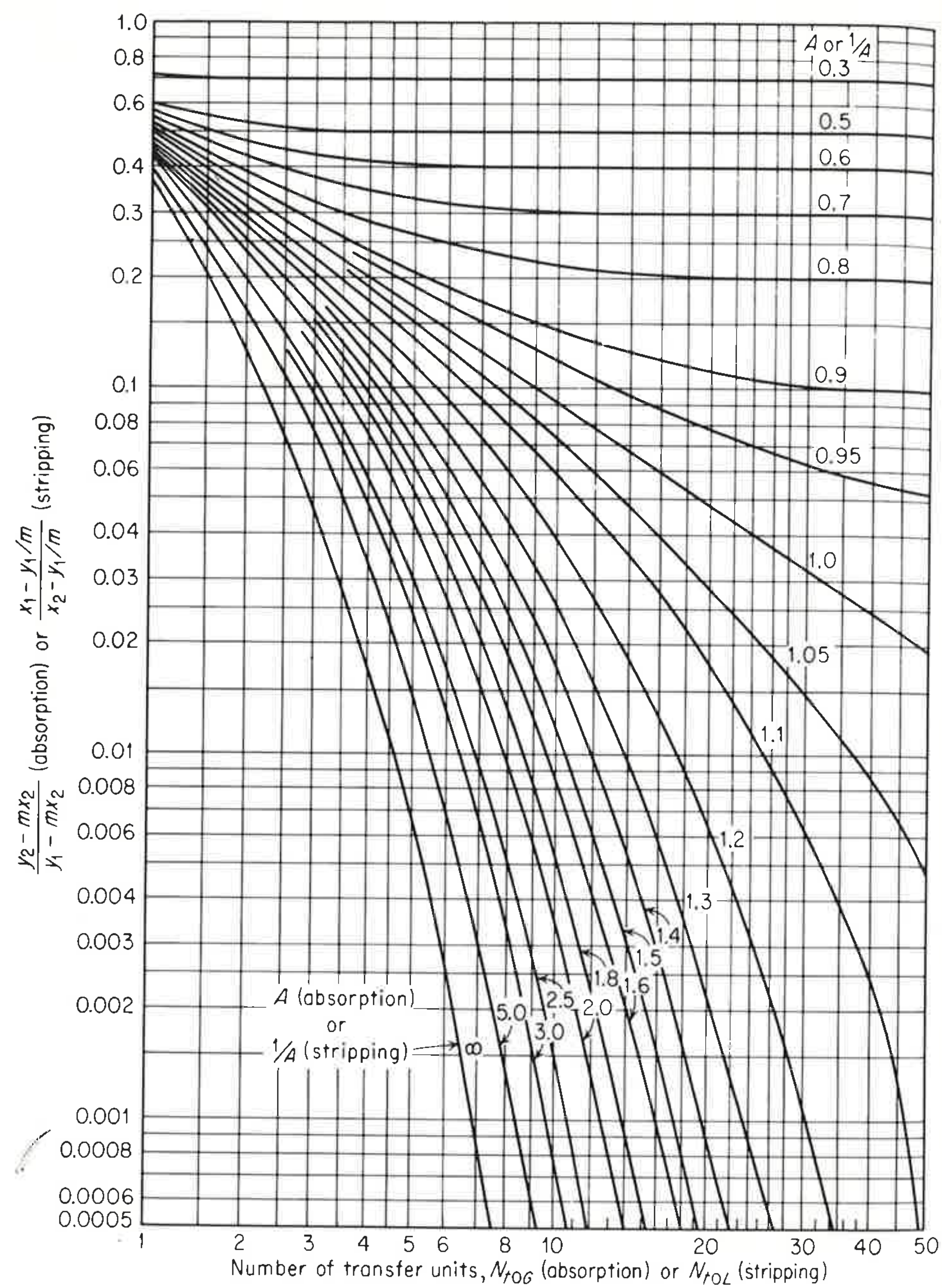


Figure 8.20 Number of transfer units for absorbers or strippers with constant absorption or stripping factor.

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A USEFUL NUMBER: CO2 SOLUBILITY IN WATER

IF SOLVENT IS WATER:

CO_2 INTO WATER

$$m = 1590 \exp\left(-2400\left(\frac{1}{T} - \frac{1}{298}\right)\right)$$

@ 298K, $m = 1590$

MASS TRANSFER COEFFICIENT FOR THE MEA ABSORBER

ANALYSIS TO GET A MASSTRANSFER COEFFICIENT FOR CO_2 ABSORPTION INTO MEA:

$$z = -\left(\frac{G/A_T}{ka}\right)\left(\int_{y_1}^{y_2} \frac{dy}{(y-y^*)}\right)$$

$$z = -\frac{G/A_T}{ka} \ln\left(\frac{y_1}{y_2}\right)$$

$$k = \frac{G/A_T}{a\,z} \ln\left(\frac{y_1}{y_2}\right)$$

- The mole fractions can be measured at the inlet and outlet. The packing properties and flow rate are known

- This would give a average "k" value for the entire column.

ALTERNATIVE (PREFERRED) MASS TRANSFER ANALYSIS

- There are many different definitions of "k". The one above gives dimensions of moles*length/time.
- The pilot plant allows measurement of the concentration at the entrance, exit and 4 intermediate points along the column, so we can get local values of the mass transfer coefficient.
- So, don't integrate the formula and also switch to the preferred notation for the pilot plant analysis…

"K" FOR PILOT PLANT

$$K_G = \frac{\dfrac{G}{P(y-y^*)}\,\dfrac{dY_a}{dz}}{a_v}$$

$$K_G [=] \frac{\text{KMOLES}}{m^2\, hr\ kPa}$$

$$G [=] \frac{\text{KMOLES}}{hr\ m^2} = \left(\frac{G}{A_T}\right)$$ ← OLD

P (=) TOTAL PRESSURE kPa

$$a_v = \frac{\text{CONTACT AREA}}{\text{VOLUM}} = 350\ \frac{m^2}{m^3}$$

$$\frac{dY_A}{dz} = \frac{\Delta Y_a}{\Delta z}$$ ← READ AT MULTIPLE Z LOCATIONS

$\Delta z = 1.37\ m$ BETWEEN MEASUREMENTS

$$Y = \frac{y}{1-y}$$ "MOLE RATIO"

$y^* = 0$ BECAUSE OF REACTION

"K" FOR PILOT PLANT

EXAMPLE NUMBERS ...

$\Delta Z = 1.37\,m$: $y = .05$, $y = .03$

$Y = .052$, $Y = .031$ → AVERAGE $.04$

$$K_G = \frac{\dfrac{\left(100\ \dfrac{\text{KMOLE}}{\text{hr}\ m^2}\right)}{101.3\ \text{kPa}\ (.04 - 0)}\left(\dfrac{.052 - .031}{1.37\ m}\right)}{350/m}$$

$$= .0011\ \frac{\text{kMOL}}{\text{hr kPa}\ m^2}$$

EXAMPLE PROBLEM

1. You wish to calculate the requisite "K_G" mass transfer coefficient for the absorber in the pilot plant. It has units of kmoles/(m^2 hr kPa). The following information is available.

FT304 (total gas flow rate) reads 131 kg/hour
AT400 (station 6) reads 6.31vol% CO_2
AT400 (station 5) reads 3.11 vol% CO_2
AT400 (station 4) reads 1.36 vol% CO_2
AT400 (station 3) reads 0.46 vol% CO_2
AT400 (station 2) reads 0.08 vol% CO_2
AT400 (station 1) reads 0.049 vol% CO_2
PT403[1] (pressure above the exit of absorber), reads 0.9 BAR (This is apparently gauge pressure. The "formula" for K_G uses absolute pressure.)
L' is 720 kg/hr.

The cross section area of the tower is A_T = 0.0415 m^2.

The area of packing per volume of packing is 350 m^2/m^3.

The distance up the column between sample ports is 1.37 m.

A. Find K_G between each station along the column. Explain possible reasons for any difference.
B. Plot the operating line on an *X*, *Y* plot. Assume that there is no CO_2 in the inlet MEA solution. A mass balance will determine how much CO2 is in the exit liquid.

PRESSURE DROP CORRELATION FOR PACKED BEDS: USED TO CHOOSE THE CROSS SECTIONAL AREA

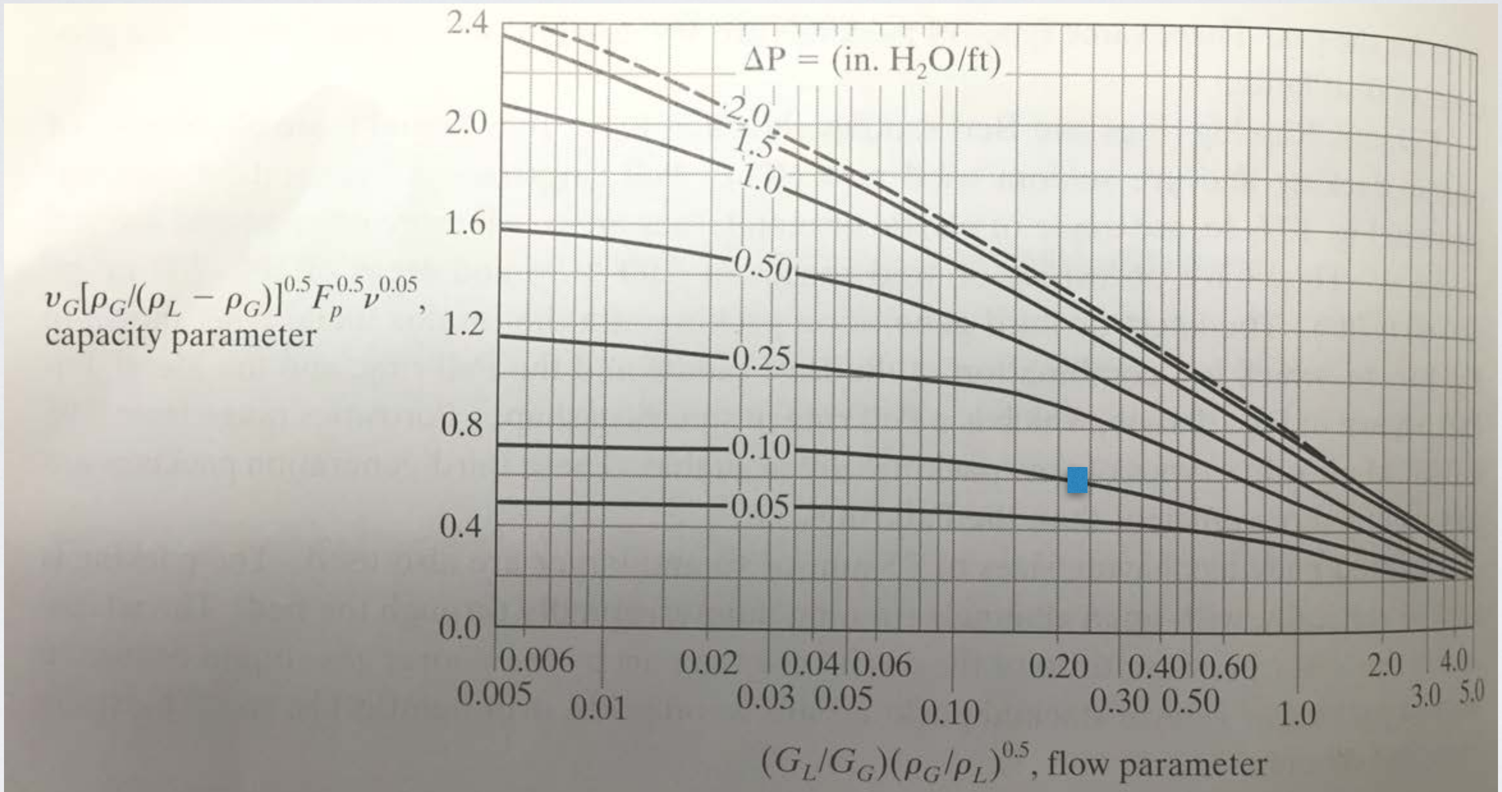

FIGURE 10.6-5. *Pressure-drop correlation for random packings by Strigle. (From R. F. Strigle, Jr.,* Random Packings and Packed Towers, *Houston: Gulf Publishing Company, 1987. With permission from Elsevier Science.)*

- Nominal operation point for Imperial absorber

ANALYSIS: TWO BASIC PRINCIPLES

- Conservation of mass
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  - Fick's Law: $j = D\frac{\partial C}{\partial x}$ (same as Fourier's Law)
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