CE407 SEPARATIONS

Lecture 21

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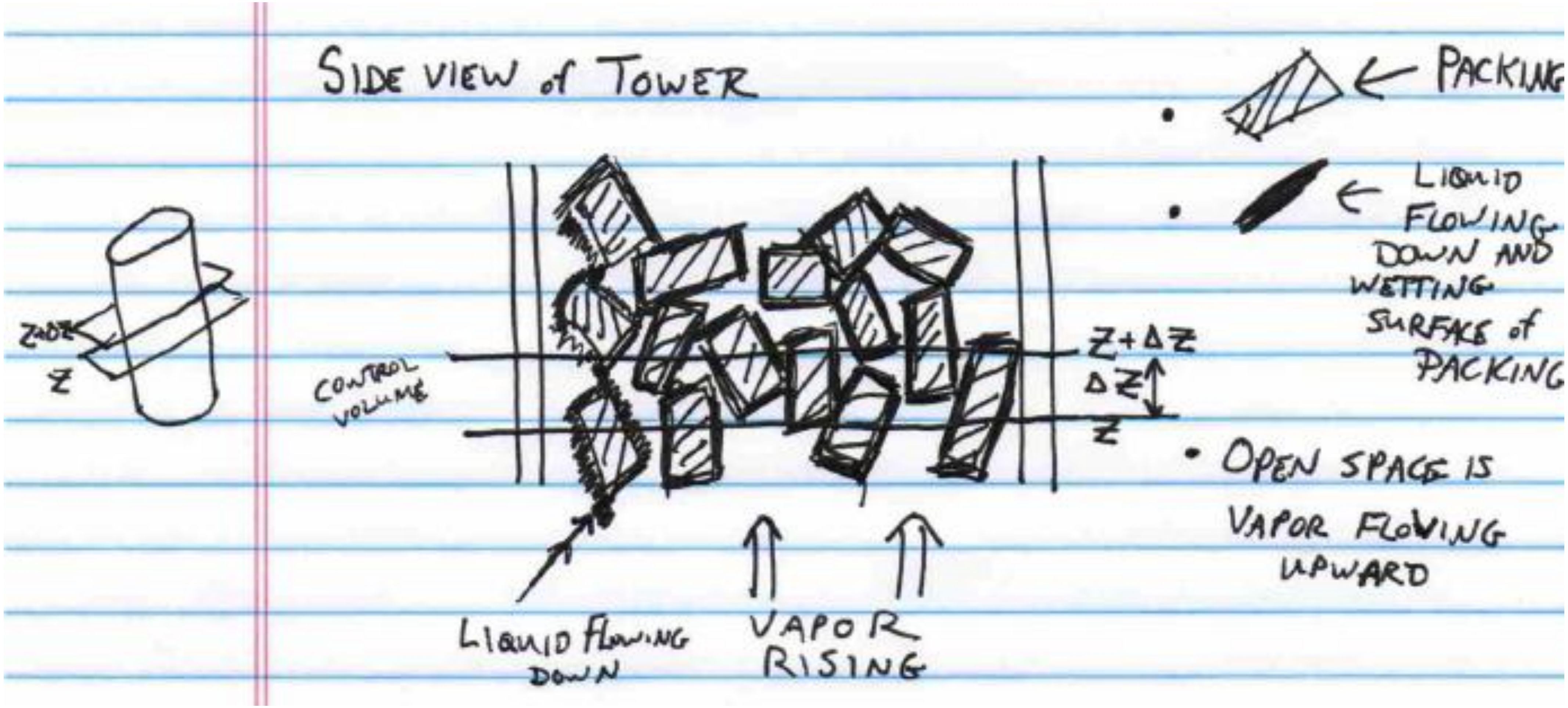
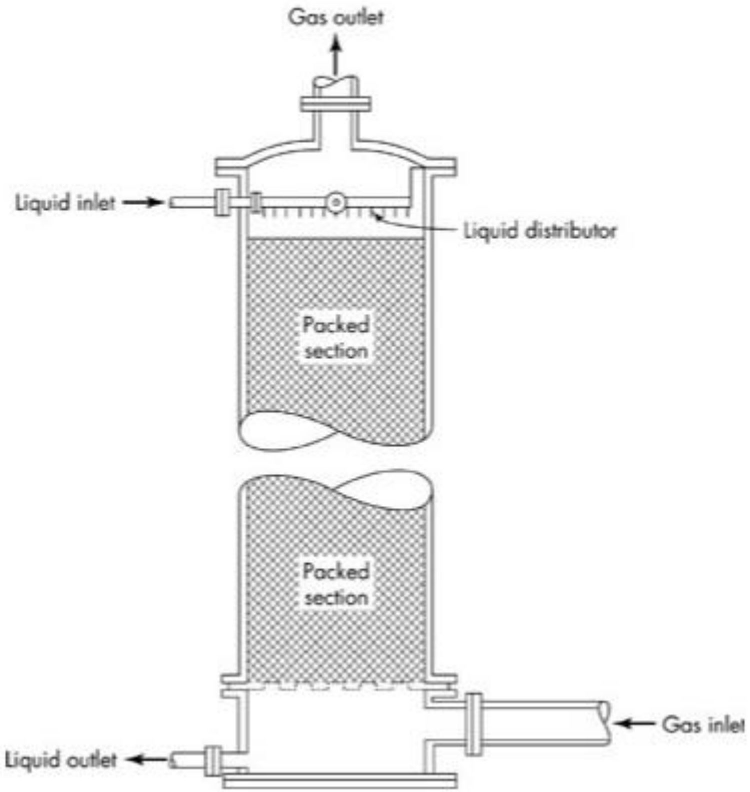
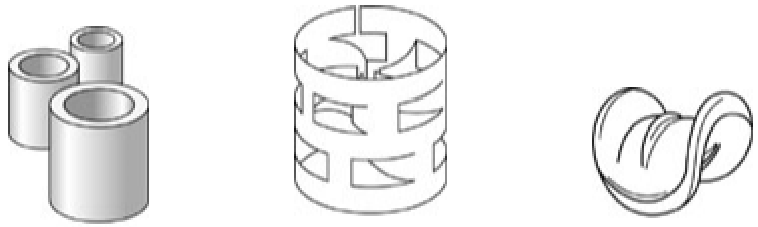
Packed Towers



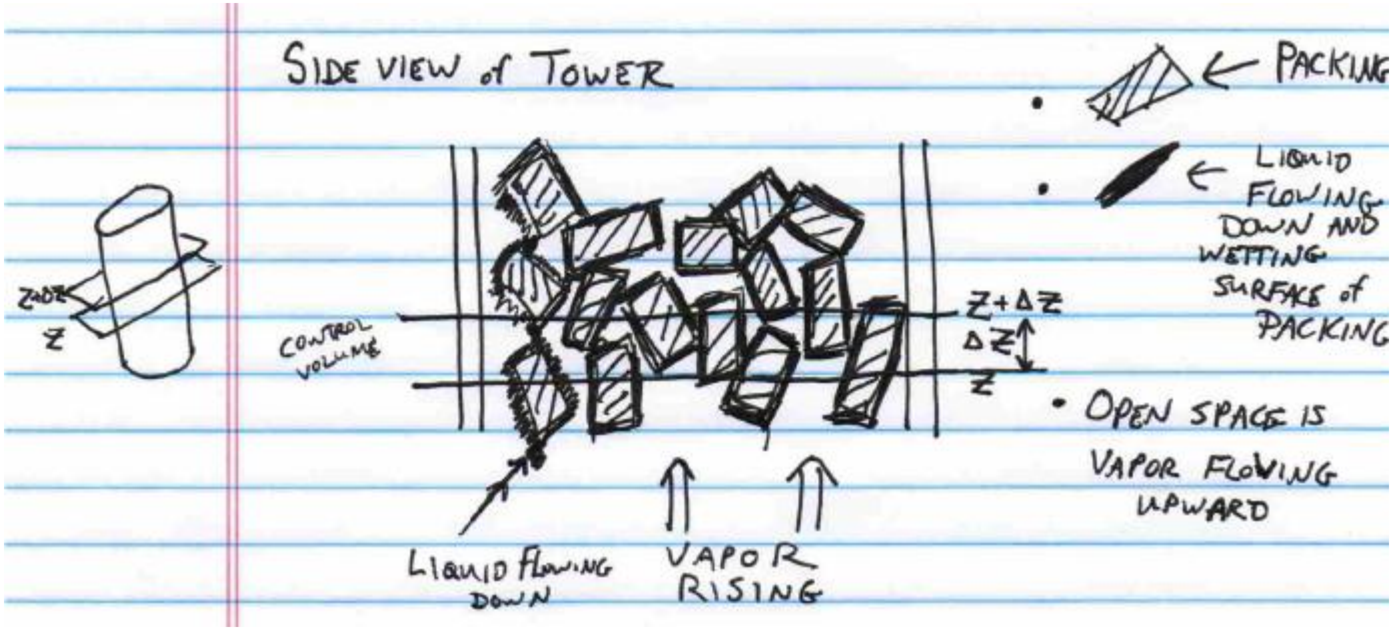
Packed Towers





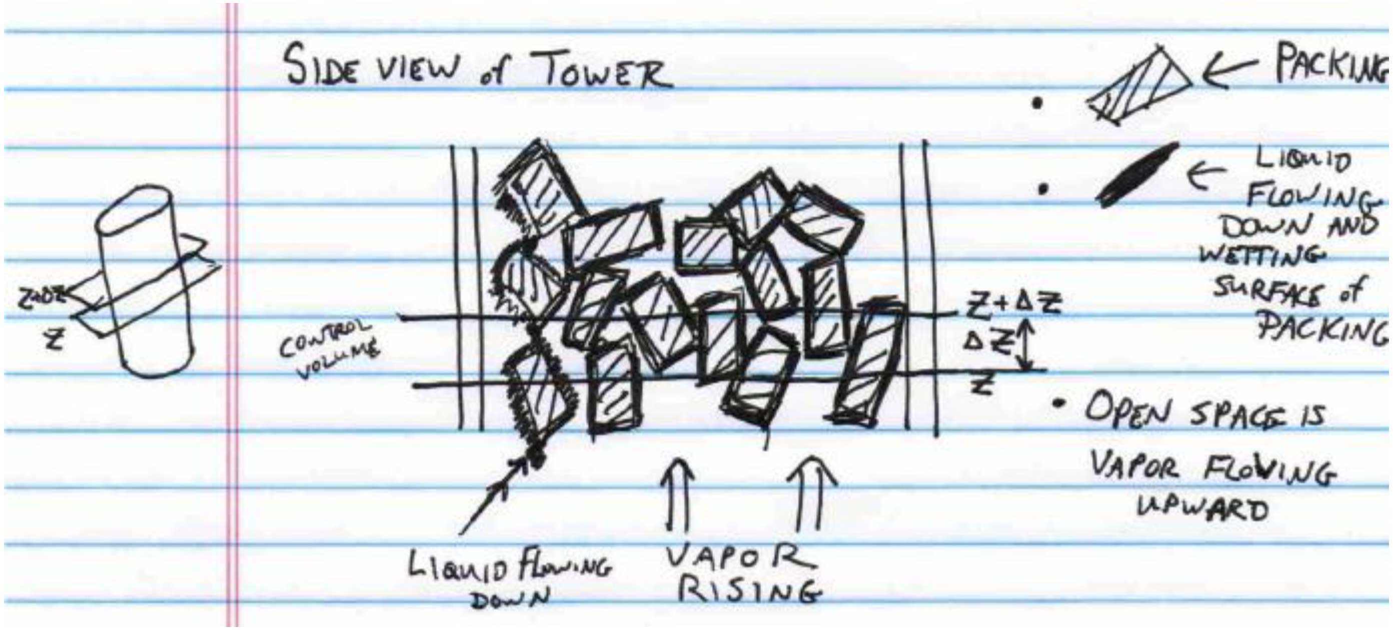
Packed Towers

- Analyze a slice of the tower from height z to $z + \Delta z$
- The control volume is the irregularly shaped volume around the wetted packing
  - i.e. the gas around the wetted packing



Control Volume Analysis

- Rate of solute entering control volume from below (via the gas) $= \boldsymbol{Vy}|_{\boldsymbol{z}}$
  - Where $\boldsymbol{V}$ is the molar flow rate of gas and $\boldsymbol{y}$ is the bulk vapor mole fraction of solute evaluated at height $\boldsymbol{z}$
- Rate of solute exiting control volume at top (via the gas) $= \boldsymbol{Vy}|_{\boldsymbol{z}+\Delta\boldsymbol{z}}$
  - Evaluated at height $\boldsymbol{z} + \Delta\boldsymbol{z}$



Control Volume Analysis, continued

- Rate of solute exiting the gas due to absorption across the gas/liquid interface

$$= \boldsymbol{k_y(y - y_i) * a\ S\,\Delta z}$$

$flux = \frac{moles}{area * time}$

cross sectional area, S, times thickness of slice = volume of slice

interfacial area per volume of packed tower

- $\boldsymbol{a\ S\,\Delta z}$ therefore is the area available for mass transfer in the control volume
- $\boldsymbol{k_y(y - y_i) * a\ S\,\Delta z}$ therefore has dimensions of moles/time – rate of mass transfer
- NOTE: $\boldsymbol{a}$ is NOT just the surface area/volume of the packing. It is the gas/liquid interfacial area per packed volume of the wetted packing and is a function of flow rate
  - The thickness of the liquid layer depends on the flow rate and the actual surface area of the liquid wetting the packing depends on the thickness of that layer.
- $\boldsymbol{y_i}$ is the mole fraction of the vapor phase at the gas/liquid interface

Control Volume Analysis, continued

- At steady state:

Moles solute in = moles solute out

$$Vy|_z = Vy\big|_{z+\Delta z} + k_y(y - y_i) * a\ S\,\Delta z$$

- Therefore

$$\frac{Vy|_{z+\Delta z} - Vy\big|_z}{\Delta z} = -k_y\, a\, S\,(y - y_i)$$

- Take the limit as $\Delta z \to 0$

$$\frac{d}{dz}(V\,y) = -k_y\, a\, S\,(y - y_i)$$

- Note: gas is losing solute as you go up the tower (increasing z), which agrees with the fact that the right hand side of the equation is negative

Control Volume Analysis, continued

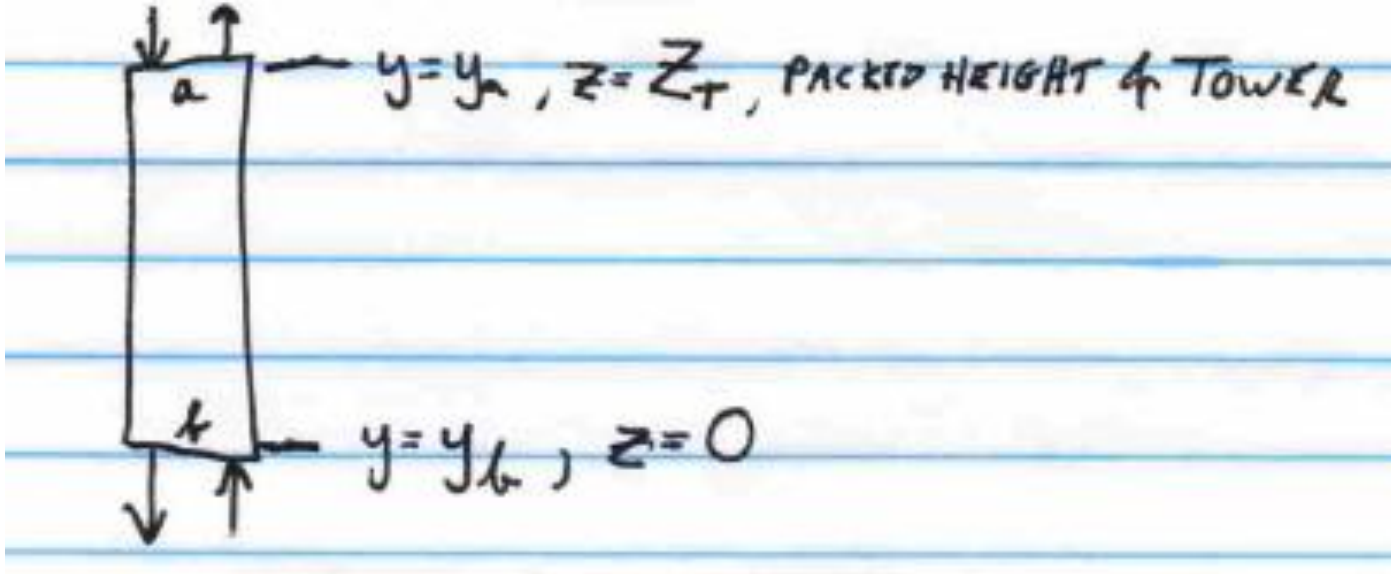
- For a dilute mixture $V \approx constant$, so we can take it out of the integral

$$V\frac{dy}{dz} = -k_y\, a\, S\,(y - y_i)$$

- Now we separate the variables

$$dz = -\frac{V/S}{k_y\, a}\frac{dy}{y - y_i}$$

- Integrate from the bottom of the tower



Control Volume Analysis, continued

- Integrate left hand side of the equation

$$\int_0^{Z_t} dz = Z_t$$

- Evaluate the RHS

$$Z_t = -\frac{V/S}{k_y\, a} \int_{y_b}^{y_a} \frac{dy}{y - y_i}$$

- Reversing the limits on the integral will change the sign

$$Z_t = \underbrace{\frac{V/S}{k_y\ a}}_{\text{Height of a Transfer Unit}} \underbrace{\int_{y_a}^{y_b} \frac{dy}{y - y_i}}_{\text{Number of Transfer Units, } \mathbf{N_y}}$$

Height of a Transfer Unit   Number of Transfer Units, $\mathbf{N_y}$

- $$H_y = \frac{V/S}{k_y\ a}$$ Because we have been working in Vapor Phase mole fractions, this carries the subscript y

Control Volume Analysis, continued

$$Z_t = H_y * N_y \qquad H_y = \frac{V/S}{k_y\ a} \qquad N_y = \int_{y_a}^{y_b} \frac{dy}{y - y_i}$$

- The height of packing required (Z_t) is the product of the height of a transfer unit (H_y) times the number of transfer units (N_y)
- Height of a transfer unit can be thought of as: Given the flows / mass transfer coefficient / available surface area per volume – how effective is a packing
- Number of transfer units can be thought of as: how much mass transfer do we need to accomplish
- This may look straightforward to solve, except:
  - How are we going to determine a ?
  - How do we determine $y - y_i$ as a function of y in order to evaluate the integral?

Number of Transfer Units

$$N_y = \int_{y_a}^{y_b} \frac{dy}{y - y_i}$$

- If $y - y_i$ is constant then we can take it out of the integral
  - $N_y = \frac{1}{y-y_i} \int_{y_a}^{y_b} dy = \frac{y_b - y_a}{y - y_i}$ which is $\frac{total\ concentration\ change\ in\ tower}{driving\ force\ for\ mass\ transfer}$

- If $y - y_i$ is not constant
  - one can numerically integrate: will need multiple data points for y_i vs y
  - one can use an average value of $y - y_i$: use Logarithmic Mean

$$\overline{(y - y_i)}_{lm} = \frac{(y - y_i)_a - (y - y_i)_b}{\ln\left[\frac{(y - y_i)_a}{(y - y_i)_b}\right]}$$

$$N_y = \frac{y_b - y_a}{\overline{(y - y_i)}_{lm}}$$

How do we sort out the interfacial mole fraction?

- At Steady State

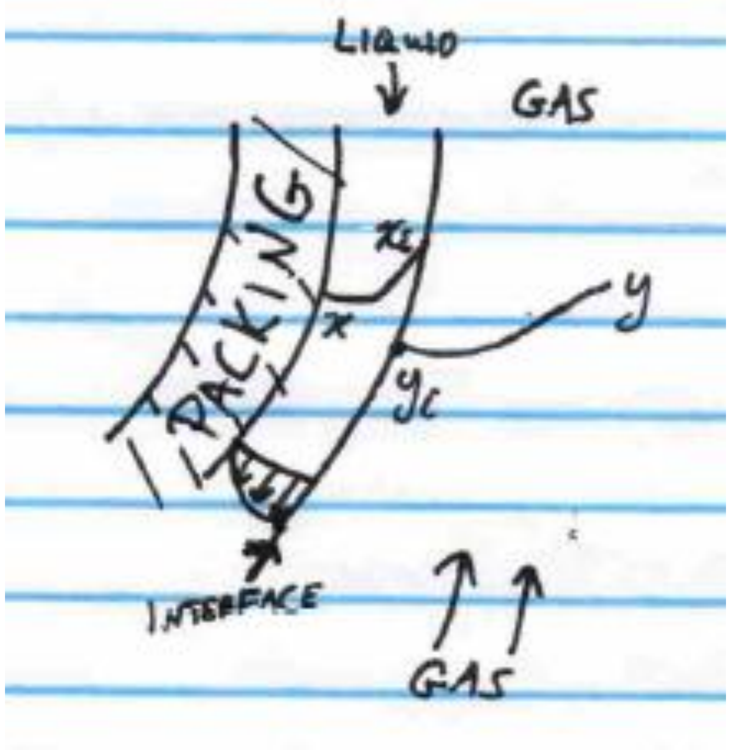
$$\begin{bmatrix} \textit{Flux of Solute from Bulk} \\ \textit{Gas to the Interface} \end{bmatrix} = \begin{bmatrix} \textit{Flux of Solute from Interface} \\ \textit{to Bulk Liquid} \end{bmatrix}$$

- and… $y_i = y^*(x_i)$ (Interfacial Mole Fractions are in Equilibrium)
- Which we can express as…

$$k_y(y - y_i)\, \underbrace{a\, \Delta V}_{\text{Area available for mass transfer}} = k_x(x_i - x)\, a\, \Delta V$$

- Which can be rearranged to…

$$y - y_i = -\frac{k_x}{k_y}(x - x_i)$$



How do we sort out the interfacial mole fraction?

- We know that: $H_y = \frac{V/S}{k_y\ a}$ Eq 22-19
- Analogously: $H_x = \frac{L/S}{k_x\ a}$ Eq 22-20
- So… $k_x = \frac{L/S}{H_x\ a}$ and $k_y = \frac{V/S}{H_y\ a}$
- Therefore: $\frac{k_x}{k_y} = \frac{L/S}{H_x\ a} * \left(\frac{V/S}{H_y\ a}\right)^{-1} = \frac{L/S}{H_x\ a} * \frac{H_y\ a}{V/S} = \left(\frac{L}{V}\right)\frac{H_y}{H_x}$
- For dilute systems $\frac{L}{V} \approx \boldsymbol{constant}$ and is the slope of the nearly straight OP Line
- $\frac{L}{V} \approx \frac{y_b - y_a}{x_b - x_a}$ from previous lectures

How do we sort out the interfacial mole fraction?

- Now $$y - y_i = -\frac{k_x}{k_y}(x - x_i)$$

$$y - y_i = -\left(\frac{L}{V}\right)\left(\frac{H_y}{H_x}\right)(x - x_i)$$

- Use equilibrium relationship (Raoult's Law, etc) to relate y_i to x_i
  - $y_i = mx_i \rightarrow x_i = y_i/m$
- Now we have one equation and one unknown so we can solve for y_i in terms of x and y at any point in the tower
- Solve for y_i at the top (a) and bottom (b) of the tower and take log mean of $(y - y_i)_a$ and $(y - y_i)_b$
- Now you can solve for number of transfer units $N_y = \frac{y_b - y_a}{(y - y_i)_{lm}}$
- Then the height of packing required is $Z_t = H_y\, N_y$

- **We've still got some issues – we don't know a and we will need to determine k_x and k_y**

Various Forms to Solve for Z_t

- Gas Film:

$$H_y = \frac{V/S}{k_y\, a} \qquad N_y = \int \frac{dy}{y - y_i}$$

- Liquid Film:

$$H_x = \frac{L/S}{k_x\, a} \qquad N_x = \int \frac{dx}{x_i - x}$$

- Overall Gas:

$$H_{Oy} = \frac{V/S}{K_y\, a} \qquad N_{Oy} = \int \frac{dy}{y - y^*}$$

- Overall Liquid:

$$H_{Ox} = \frac{L/S}{K_x\, a} \qquad N_{Ox} = \int \frac{dx}{x^* - x}$$

- All of these are equivalent and will lead to the same answer for Z_t
- **We still need to figure out mass transfer coefficients and a !**

Mass Transfer Correlations for Packed Towers

- In the previous lecture we saw some methods for estimating k_x and k_y
- When it comes to packed towers there are some issues
  - The geometry of the packing is not like the simpler cases where we have existing correlations
  - a is dependent on the flow rates, packing design, surface tension, viscosity, etc
- Fortunately there are correlations for H_x and H_y directly

$$H_x = (0.9\ ft)\left[\frac{G_x/\mu}{\left(1500\ {}^{lb}/_{ft^2\ hr}\right)/(0.891\ cP)}\right]^{0.3}\left(\frac{S_c}{381}\right)^{0.5}\frac{1}{f_p}$$

- This was arrived at by taking experimental data for O_2 in water
  - This system is dominated by liquid film resistance so the experimental measurements are essentially that of transport through the liquid film versus the combination
  - G_x is mass velocity and must be the same units as appear in the correlation, ${}^{lb}/_{ft^2\ hr}$
- Data correlated to show that $H_x \propto \left(\frac{G_x}{\mu}\right)^{0.3}(S_c)^{0.5}$
- A value of 0.9 feet corresponds to G_x= 1500 ${}^{lb}/_{ft^2\ hr}$, μ = 0.891 cP, S_c= 381, and f_p = 1

Mass Transfer Correlations for Packed Towers

- The correlation on the previous page was developed using water as the liquid – use caution when applying it to other liquids
- f_p accounts for the type of packing used
  - Be sure to use f_p and not F_p
  - F_p is used in calculations of pressure drop

TABLE 18.1
Characteristics of dumped tower packings[12,15b,27]

Type	Material	Nominal size, in.	Bulk density,† lb/ft³	Total area,† ft²/ft³	Porosity ε	Packing factors‡ F_p	Packing factors‡ f_p
Raschig rings	Ceramic	$\frac{1}{2}$	55	112	0.64	580	1.52§
		1	42	58	0.74	155	1.36§
		$1\frac{1}{2}$	43	37	0.73	95	1.0
		2	41	28	0.74	65	0.92§
Pall rings	Metal	1	30	63	0.94	56	1.54
		$1\frac{1}{2}$	24	39	0.95	40	1.36
		2	22	31	0.96	27	1.09
	Plastic	1	5.5	63	0.90	55	1.36
		$1\frac{1}{2}$	4.8	39	0.91	40	1.18
Berl saddles	Ceramic	$\frac{1}{2}$	54	142	0.62	240	1.58§
		1	45	76	0.68	110	1.36§
		$1\frac{1}{2}$	40	46	0.71	65	1.07§
Intalox saddles	Ceramic	$\frac{1}{2}$	46	190	0.71	200	2.27
		1	42	78	0.73	92	1.54
		$1\frac{1}{2}$	39	59	0.76	52	1.18
		2	38	36	0.76	40	1.0
		3	36	28	0.79	22	0.64
Super Intalox saddles	Ceramic	1	—	—	—	60	1.54
		2	—	—	—	30	1.0
IMTP	Metal	1	—	—	0.97	41	1.74
		$1\frac{1}{2}$	—	—	0.98	24	1.37
		2	—	—	0.98	18	1.19
Hy-Pak	Metal	1	19	54	0.96	45	1.54
		$1\frac{1}{2}$	—	—	—	29	1.36
		2	14	29	0.97	26	1.09
Tri-Pac	Plastic	1	6.2	85	0.90	28	—
		2	4.2	48	0.93	16	—

†Bulk density and total area are given per unit volume of column.
‡Factor F_p is a pressure drop factor and f_p a relative mass-transfer coefficient. Factor f_p is discussed on page 603 in the paragraph "Performance of Other Packings." Its use is illustrated in Example 18.7.
§Based on NH_3–H_2O data; other factors based on CO_2–$NaOH$ data.

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Mass Transfer Correlations for Packed Towers

$$H_y = (1.4\,ft)\left[\frac{G_y}{500\ {}^{lb}\!/_{ft^2\,hr}}\right]^{0.3}\left[\frac{1500\ {}^{lb}\!/_{ft^2\,hr}}{G_x}\right]^{0.4}\left(\frac{S_c}{0.66}\right)^{0.5}\frac{1}{f_p}$$

- Correlation similarly derived for an air-ammonia-water system
  - High solubility of ammonia in water leads to system being dominated by gas film resistance
- G_x and G_y are mass velocities and must be in the same units as appear in the correlation, ${}^{lb}\!/_{ft^2\,hr}$
- Notice that G_y appears in the H_y correlation but not in the H_x correlation
  - This is because gas flow rates are specified to avoid flooding in the tower and therefore are usually in a set range for a given liquid flow

Mass Transfer Correlations for Packed Towers

- Use arithmetic averages of mass velocities at the top and bottom of the tower

$$G_x = \frac{(G_x)_a + (G_x)_b}{2}$$

$$G_y = \frac{(G_y)_a + (G_y)_b}{2}$$

Overall Mass Transfer Coefficients

$$H_{Oy} = H_y + \frac{m}{L/V} H_x$$

$$H_{Ox} = H_x + \frac{L/V}{m} H_y$$

- $y_i = mx_i$
- **L and V are molar flow rates**
- **Textbook shows these as L_M and G_M which are molar velocities but the area factor cancels out**

Packed Tower HTU Example – problem statement

- 720 mol/hr stream of toluene contaminated oil (95 mole percent oil, 5 mole percent toluene) is to be cleaned by countercurrent contact with air in a stripping tower operating at 25 C and atmospheric temperature.
- Tower is packed with 1” plastic Pall rings
- Exiting liquid must have a toluene mole fraction equal to no more than 0.001
- Entering air is pure and is at 1.078 times the minimum.
- The tower diameter is 17”
- Under the proposed operating conditions $H_x = 1.0\ ft$
- Toluene will follow Raoult’s Law and has a vapor pressure of 0.0380 atm
- The oil has MW = 170, $\rho = 0.730\ \frac{gm}{cm^3}$, $\mu = 0.86\ cP$
- Due to low toluene mole fractions the physical properties may be approximated as those of pure oil

- Using $\boldsymbol{H_{Oy}}$ and $\boldsymbol{N_{Oy}}$, determine the required Packed Height
  - Use the “Usual Assumptions”

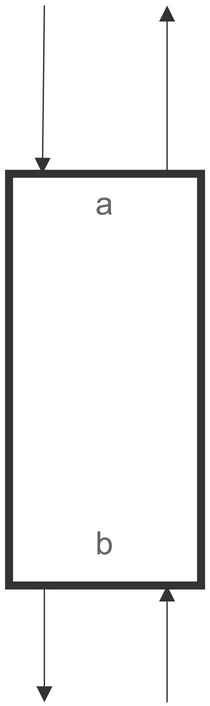
Packed Tower HTU Example – Preliminary calculations

- 1 hour basis

L_a = 720 mol
x_a = 0.05

L_c = 0.95 * 720 = 684 mol oil

$(L_{tol})_a$ = 0.05 * 720 = 36 mol tol



$V_c = same$
$(V_{tol})_a = 36 \ - 0.685 = 35.315\ mol$
y_a = ?

L_c = 684 mol oil
x_b = 0.001

$V_b = V_c = ?$
$y_b = 0$

- $x_b = 0.001 = \frac{(L_{tol})_b}{(L_{tol})_b + 684} \rightarrow (L_{tol})_b = 0.685\ \text{mol}$

Packed Tower HTU Example – minimum and actual air flow

- Due to the dilute nature and the fact that this is a stripping operation, minimum air can be calculated with the assumption that

$$(y_a)_{min} = y^*(x_a) = \frac{P_{tol}^{sat}}{P}\, x_a$$

$$(y_a)_{min} = y^*(x_a) = \frac{0.038\, atm}{1\, atm}\, 0.05 = 0.0019$$

$$(y_a)_{min} = 0.0019 = \frac{(V_{tol})_a}{(V_{tol})_a + (V_c)_{min}} = \frac{35.315}{35.315 + (V_c)_{min}} \rightarrow (V_c)_{min} = 18551.527 \text{ mol}$$

- Actual Air Flow $V_c = 1.078 * (V_c)_{min} = 19998.546\, mol \approx 20000\, mol$

$$y_a = \frac{(V_{tol})_a}{(V_{tol})_a + V_c} = \frac{35.315}{35.315 + 20{,}000} = 0.001763$$

Packed Tower HTU Example -Mass rates and Mass Fluxes

- $(SG_x)_a = \left[\left(684 \frac{mol\ oil}{hr}\right) * \left(170 \frac{g}{mol\ oil}\right) + \left(36 \frac{mol\ tol}{hr}\right) * \left(92.14 \frac{g}{mol\ tol}\right)\right] * \frac{1\ lb_m}{453.6\ g} = 263.7 \frac{lb_m}{hr}$
- $(SG_y)_a = \left[\left(20{,}000 \frac{mol\ air}{hr}\right) * \left(28.84 \frac{g}{mol\ air}\right) + \left(35.315 \frac{mol\ tol}{hr}\right) * \left(92.14 \frac{g}{mol\ tol}\right)\right] * \frac{1\ lb_m}{453.6\ g} = 1279 \frac{lb_m}{hr}$
- $(SG_x)_b = \left[\left(684 \frac{mol\ oil}{hr}\right) * \left(170 \frac{g}{mol\ oil}\right) + \left(0.685 \frac{mol\ tol}{hr}\right) * \left(92.14 \frac{g}{mol\ tol}\right)\right] * \frac{1\ lb_m}{453.6\ g} = 256.5 \frac{lb_m}{hr}$
- $(SG_y)_b = \left[\left(20{,}000 \frac{mol\ air}{hr}\right) * \left(28.84 \frac{g}{mol\ air}\right)\right] * \frac{1\ lb_m}{453.6\ g} = 1272 \frac{lb_m}{hr}$
- $\overline{(SG_x)} = arithmetic\ mean$ of liquid flow at a and b $= 260.1 \frac{lb_m}{hr}$
- $\overline{(SG_y)} = arithmetic\ mean$ of vapor flow at a and b $= 1275.5 \frac{lb_m}{hr}$
- S = Superficial Cross-sectional area = $\frac{\pi D^2}{4} = \frac{\pi\,(^{17}/_{12})^2}{4} = 1.576\ ft^2$
- $\overline{G_x} = \frac{\overline{(SG_x)}}{S} = \frac{260.1 \frac{lb_m}{hr}}{1.576\ ft^2} = 165 \frac{lb_m}{ft^2 hr}$ Liquid Mass Flux
- $\overline{G_y} = \frac{\overline{(SG_y)}}{S} = \frac{1275.5 \frac{lb_m}{hr}}{1.576\ ft^2} = 809 \frac{lb_m}{ft^2 hr}$ Vapor Mass Flux

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DIFFUSIVITIES AND SCHMIDT NUMBERS FOR GASES IN AIR AT 25 °C AND 1 ATM

Gas	Volumetric diffusivity D_v, ft²/h	$N_{Sc} = \frac{\mu}{\rho D_v}$ ‡
Acetic acid	0.413	1.24
Acetone	0.32§	1.60
Ammonia	0.836	0.61
Benzene	0.299	1.71
n-Butyl alcohol	0.273	1.88
Carbon dioxide	0.535	0.96
Carbon tetrachloride	0.26§	1.97
Chlorine	0.43§	1.19
Chlorobenzene	0.24§	2.13
Ethane	0.49§	1.04
Ethyl acetate	0.278	1.84
Ethyl alcohol	0.396	1.30
Ethyl ether	0.302	1.70
Hydrogen	2.37	0.22
Methane	0.74§	0.69
Methyl alcohol	0.515	1.00
Naphthalene	0.199	2.57
Nitrogen	0.70§	0.73
n-Octane	0.196	2.62
Oxygen	0.690	0.74
Phosgene	0.31§	1.65
Propane	0.36§	1.42
Sulfur dioxide	0.44§	1.16
Toluene	0.275	1.86
Water vapor	0.853	0.60

* By permission, from T. K. Sherwood and R. L. Pigford, *Absorption and Extraction*, 2nd ed., p. 20. Copyright 1952, McGraw-Hill Book Company, New York.
‡ The value of μ/ρ is that for pure air, 0.612 ft²/h.
§ Calculated by Eq. (21.23).

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*Bulk density and total area are given per unit volume of column.
†Factor F_p is a pressure drop factor and f_p a relative mass-transfer coefficient. Factor f_p is discussed on page 603 in the paragraph "Performance of Other Packings." Its use is illustrated in Example 18.7.
§Based on NH_3–H_2O data; other factors based on CO_2–NaOH data.

Mass Transfer Coefficients

$$H_y = (1.4\,ft)\left[\frac{G_y}{500\ {lb}/{ft^2\,hr}}\right]^{0.3}\left[\frac{1500\ {lb}/{ft^2\,hr}}{G_x}\right]^{0.4}\left(\frac{S_c}{0.66}\right)^{0.5}\frac{1}{f_p}$$

$$H_y = (1.4\,ft)\left[\frac{809\,\frac{lb_m}{ft^2 hr}}{500\ {lb}/{ft^2\,hr}}\right]^{0.3}\left[\frac{1500\ {lb}/{ft^2\,hr}}{165\,\frac{lb_m}{ft^2 hr}}\right]^{0.4}\left(\frac{1.86}{0.66}\right)^{0.5}\frac{1}{1.36} = 4.8\,ft$$

$H_x = 1.0\,ft$ given in problem statement

Overall Mass Transfer Coefficient

$$H_{Oy} = H_y + \frac{m}{L/V} H_x$$ where $y_i = mx_i$

- $L/V = \frac{720}{20{,}035} = 0.0359$ at a
- $L/V = \frac{684}{20{,}000} = 0.0324$ at b
- $L/V = 0.0342$ average

- $m = 0.0384$ vapor pressure of toluene

$$H_{Oy} = 4.8\,ft + \frac{0.0384}{0.0342} * 1.0\,ft = 5.9\,ft$$

Number of Transfer Units

$$N_{Oy} = \frac{y_b - y_a}{\overline{(y - y^*)}_{lm}}$$

$$\overline{(y - y^*)}_{lm} = \frac{(y - y^*)_a - (y - y^*)_b}{\ln\left[\frac{(y - y^*)_a}{(y - y^*)_b}\right]}$$

- $y_a = 0.001763$
- $y_b = 0$
- $y_a^* = m * x_a = 0.038 * 0.05 = 0.0019$
- $y_b^* = m * x_b = 0.038 * 0.001 = 0.000038$
- $y_a - y_a^* = 0.001763 - 0.0019 = -0.000137$
- $y_b - y_b^* = 0 - 0.000038 = -0.000038$

Number of Transfer Units and Packed Height

$$\overline{(y-y^*)}_{lm} = \frac{(y-y^*)_a - (y-y^*)_b}{\ln\left[\frac{(y-y^*)_a}{(y-y^*)_b}\right]}$$

$$\overline{(y-y^*)}_{lm} = \frac{-0.000137 - (-0.000038)}{\ln\left[\frac{-0.000137}{-0.000038}\right]} = -7.72 * 10^{-5}$$

$$N_{Oy} = \frac{y_b - y_a}{\overline{(y-y^*)}_{lm}} = \frac{0 - 0.001763}{-7.72 * 10^{-5}} = 22.84$$

- $Z_t = H_{Oy} * N_{Oy} = 5.9\, ft * 22.84 = 135\, ft\ packed\ height$