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Single and Two-Phase Flow in Packed Beds

--Manuscript Draft--

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Overview

The goal of this experiment is to determine the magnitude of maldistribution in typical packed beds, in both single phase and two-phase (gas-liquid) flow, and the effects of this maldistribution on pressure drop. The concepts of residence time distribution and dispersion are introduced through the use of tracers, and these concepts are related to physical maldistribution.

~~Gas-liquid countercurrent vertical packed beds (packed columns) are frequently used in separation processes such as distillation, absorption, and stripping.[1] Cocurrent horizontal packed beds are often used as reactors or adsorbers with a solid catalyst or adsorbent. In both cases (as separators or reactors), the packing increases the surface area of vapor-liquid contact.[1] Packing can exist in two forms: dumped packing consisting of random or simple geometric shapes of materials such as clays, metals or ceramic oxides, or structured packing from common metal and plastics, consisting of highly defined interconnected geometric networks (usually corrugated) that can reduce pressure drop compared to most dumped packings.[1] However, whether horizontal or vertical, maldistribution (channeling) can degrade the performance of the separator, reactor or adsorber; sometimes, various types of flow distributors can be employed to mitigate the effects.[2]~~ Channeling~~, which~~ in a single-phase flow can occur along walls or just by preferential~~the~~ flow through ~~occupying only~~ a~~the lower~~ larger portion of the bed cross-section~~a horizontal bed~~. Channeling in two-phase flow can result from even more complex causes, and simple two-phase flow theories seldom predict pressure drops in packed beds. A goal of design is always to minimize the extent of channeling by finding the optimal bed and particle diameters for the design flow rates, and by packing a bed in a way to minimize settling. It is always important to quantify how much maldistribution might occur and to over-design the unit to account for its occurrence.

The permeameter apparatus measures pressure drop ΔP and tracer (dye) output concentration at the exit of horizontal packed beds of armored glass, for either water flow, air flow, or two-phase flow (Figure 1). Water enters through a control valve and can be routed through manual valves to any of five beds (48” long, 3” I.D.) with different size glass bead dumped (random) packings. The pressure drop is measured using a pressure transmitter. The water flow is measured by a differential pressure ~~(DP)~~ (DP, orifice) transmitter and the air flow by a dry test meter (similar to a home gas meter). The dye sample is injected upstream by an automated sampling valve. The dye exit concentration from a bed is measured using a UV-Vis spectrometer. Residence time distributions are calculated from the tests and compared to the predictions of theories of dispersion in packed beds. Two-phase flow can be studied in bed 5, which contains the largest particles.

Commented [HK1]: How do you pronounce this?

Commented [KMD2R1]: Just won’t use the word.

Commented [HK3]: Please incorporate the second paragraph from the summery in here to emphasize the significance of the maldistribution determination and the effect of channeling.

Commented [KMD4R3]: done

Commented [HK5]: Please define briefly.

Commented [HK6]: I would shorten this paragraph a little and introduce the majority in the principle section in order to make for a nicer flow of and easier scriptwriting using animations.

Commented [KMD7R6]: done

Commented [HK8]: Could you please provide a picture of the apparatus and ideally a cross-section, so that proper depiction will be created during animation.

Commented [KMD9R8]: It’s pretty big and difficult to get it all in a single view. We have put in a 3-D drawing from AutoCad instead.

Commented [HK10]: Is it possible to obtain pictures/cross sections here as well, it will be great to create an animation comparing these packed beds.

Commented [KMD11R10]: Sorry, no animations.

Commented [HK12]: I would mention the material it is often made of, such as corrugated sheets made of metal, plastic or wire gauze, creating a honeycomb structure.

Commented [KMD13R12]:

Commented [HK14]: Why? I would briefly mention why structured packing has this ability over to dumped packing.

Commented [HK15]: This is great to explain the significance of measuring maldistribution in the packed beds and I would put this section into the overview.

Commented [KMD16R15]: It’s being moved to overview.

Commented [HK17]: Do you have a picture of this?

Commented [HK18]: Could you indicate in the figure where exactly the permeameter would be connected to measure this?

Commented [KMD19R18]: Done

Commented [HK20]: Where is this valve on the diagram? Could you provide a more descriptive diagram/figure 1 please.

Commented [KMD21R20]: It is labeled now in Fig. 1.

Commented [HK22]: Is this dumped packing or structured packing?

Commented [HK23]: DP = differential pressure, please write out abbreviations for the first time.

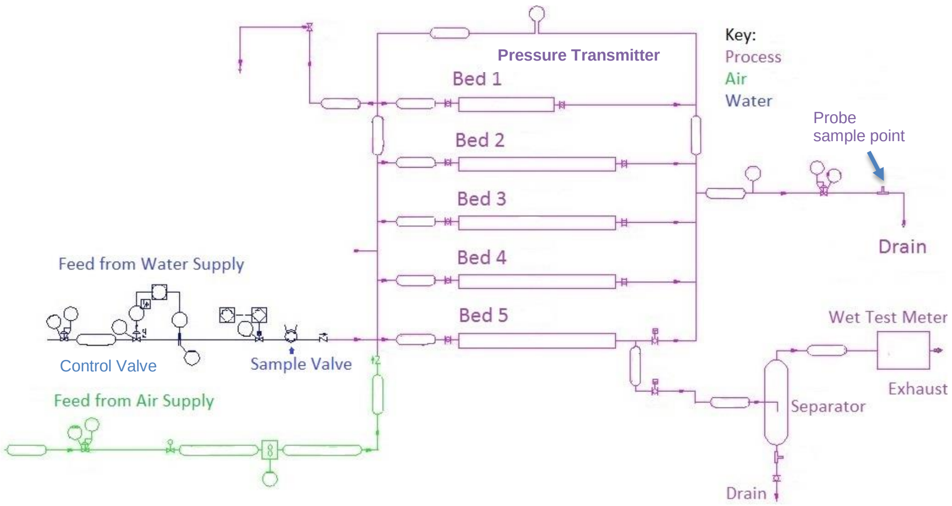

Figure 1: Process and instrumentation diagram of the ~~permeameter~~apparatus.

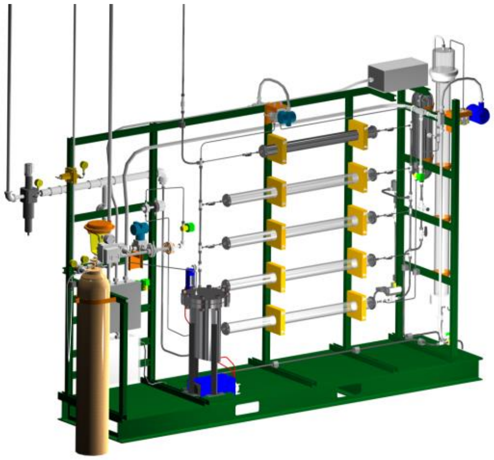

Figure 2. 3-D rendering of the apparatus. Bed #1 is at the top, bed #5 at the bottom. The water control valve is on the left (red bonnet). The DP transmitter is at the top center (blue).

Principles:

Gas-liquid countercurrent vertical packed beds (packed columns) are frequently used in separation processes such as distillation, absorption, and stripping.[1] Cocurrent horizontal packed beds are often used as reactors or adsorbers with a solid catalyst or adsorbent. In both cases (as separators or reactors), the packing increases the surface area of vapor-liquid contact.[1] Packing can exist in two forms: dumped packing consisting of random or simple geometric shapes of materials such as clays, metals or ceramic oxides, or structured packing from common metal and plastics, consisting of highly defined interconnected geometric networks (usually corrugated metals or plastics) that can reduce pressure drop compared to most dumped packings.[1] However, whether horizontal or vertical, maldistribution (channeling) can degrade the performance of the separator, reactor or

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Commented [HK24]: I would shorten this paragraph a little and introduce the majority in the principle section in order to make for a nicer flow of and easier scriptwriting using animations.

Commented [KMD25R24]: done

Commented [HK26]: Could you please provide a picture of the apparatus and ideally a cross-section, so that proper depiction will be created during animation.

Commented [KMD27R26]: It's pretty big and difficult to get it all in a single view. We have put in a 3-D drawing from AutoCad instead.

Commented [HK28]: Is it possible to obtain pictures/cross sections here as well, it will be great to create an animation comparing these packed beds.

Commented [KMD29R28]: Sorry, no animations.

Commented [HK30]: I would mention the material it is often made of, such as corrugated sheets made of metal, plastic or wire gauze, creating a honeycomb structure.

Commented [KMD31R30]: Nowadays they can be made out of almost any common metal or plastic.

Commented [HK32]: Why? I would briefly mention why structured packing has this ability over to dumped packing.

adsorber; sometimes, various types of flow distributors can be employed to mitigate the effects.[2]

The single phase packed bed ΔP's can be compared to the predictions of the Ergun equation.[3]

Tracers are dyes that are injected instantaneously into the upstream flow, and whose composition as a function of time is measured in the flow downstream of a bed.[4] The measurable tracer molecules ~~(measurable)~~ are assumed to be characteristic of all the molecules making up the liquid flow. The volume of the injected tracer~~pulse~~ must be small relative to the system volume. If perfect plug-flow (no axial mixing) occurs in a packed bed, then the~~a pulse~~ tracer injected at time zero~~over a short period δt~~ would exit the bed at a later time as a spike~~pulse of the same concentration and duration~~. For any real bed, the tracer~~pulse~~ will disperse~~spread~~, exiting the reactor at ~~a~~ lower concentrations ~~and~~ over a longer time period. If the flow is not maldistributed, the spreading will be described by the Gaussian (Normal) distribution, with the peak of the curve observed at the average residence time. The more the tracer spreads out in time, the worse the maldistribution, and typically the poorer the separation or reaction process.

The residence time distribution (RTD) ~~is the residence time distribution, which~~ describes the distribution of times that molecules can spend in the bed. If M is the total mass of tracer injected into the system, Q the volumetric flowrate, and $C(t)$ is the effluent concentration, then the mass balance on the ~~pulse~~ tracer is:

$$\frac{M}{Q}=\int_0^\infty C(t)dt \qquad (1)$$

The left-hand side of Equation (1) represents tracer mass in, the right-hand side mass out, $E(t)$ is the bed-exiting~~external~~ residence time distribution (RTD), a probability ~~distributio~~distribution. Using Equation 1 for the integral, it can be calculated as:

$$E(t)=\frac{QC(t)}{M}=\frac{C(t)}{\int_0^\infty C(t)dt} \qquad (2)$$

$E(t)dt$ is the fraction (probability) of molecules in the exit stream of residence time between t and $t+dt$. The terms E-curve and RTD are synonymous. For packed beds, the residence time is related to the void volume (product of total reactor volume V and porosity ε) divided by volumetric flow rate Q. ~~We can define~~ T~~t~~he mean residence time τ can be defined and related~~ it~~ to $E(t)dt$, the probability a given molecule entering the bed at t = 0 will exit at t:

$$\tau=\frac{\varepsilon V}{Q}=\int_0^\infty t\,E(t)dt \qquad (3)$$

As seen from Equation~~.~~ ~~(~~3~~)~~, E(t) has units of inverse time. Sometimes the dimensionless E-curve is plotted instead of the E-curve. This dimensionless E-curve, E(t/τ) is obtained by multiplying the E-curve by τ. Its average is 1.

Commented [HK33]: aka sample?

Commented [HK34]: What exactly does it mean? No mixing? What are perfect plug flow conditions?

Commented [HK35]: Briefly explain.

Commented [HK36]: Stands for?

Commented [HK37]: Please elaborate on this. Meaning it will be mixed, adsorbed?

Commented [KMD38R37]: There are many mechanisms of dispersion. Best to get to this later.

Commented [HK39]: And consequently the worse the separation process?

Commented [HK40]: Please write out the first time : RTD = Residence time distribution

Commented [HK41]: Can you briefly explain the relationship.

Commented [HK42]: How is this different to RTD? What does external mean in this case?

Commented [HK43]: Briefly explain the relationship in order for an easier scriptwriting. If we understand what is dependent on what or influenced it will be easier to animate /explain these equations.

Commented [HK44]: Please avoid using "we"

Commented [HK45]: Could you provide the units, as reviewer/scriptwriter might be unfamiliar with the equations and variables.

Commented [KMD46R45]: That is the unit.

Commented [HK47]: When and why, what is the benefit of the one vs. the other?

Commented [K48R47]: There is no major benefit. The dimensionless curve immediately shows the center of mass (τ= 1).

Another good way to quantify the deviation from plug flow (the "dispersion") is to compute the variance of the E-curve (σ^2) divided by its mean squared.

$$\sigma^2 = \int_0^\infty (t-\tau)^2 E(t)\,dt \qquad (4)$$

This quantity should be invariant with respect to flow rate for a packed bed, if maldistribution is not present. The range of values due to molecular diffusion only should be:

$$\sigma^2 / \tau^2 = (4 - 6) \; \left(d_p / L \right) \qquad (54)$$

for $Re_p < 40$, where Re_p is the particle Reynolds number, d_p the average particle diameter and L bed length. Greater values of experimental σ^2 than predicted by Equation 5 ~~of σ²/τ²~~ and deviations from the Gaussian distribution indicate flow maldistribution, as does an 'early' peak in the E(t)-curve, or~~and~~ a long tail on the main peak.

In some cases, the nature and magnitude of the maldistribution can be observed visually. This is especially true in two-phase flow.

Two simple models exist for two phase flow, the homogenous model and the stratified model.[3,5] For homogeneous flow the basic assumptions are that the actual gas velocity, U_G, actual liquid velocity, U_L and averaged velocity of the liquid-gas mixture, U_{tp} are equal:~~the velocities~~

$U_L = U_G = U_{tp}$~~,~~ (6)

Then~~that~~ the two-phase density is given by G/U_{tp} (G is mass velocity), and that the averaged two-phase viscosity, μ_{tp} is given by:

$$\mu_{tp}^{-1} = \mu_L^{-1}(1 - X) + \mu_G^{-1} X \qquad (7\cancel{5})$$

Where X is the quality (weight fraction of vapor in a vapor-liquid mixture) and μ_L, μ_G are the viscosities of the respective liquid phase and gas phases.

For stratified flow, once the pressure drop, the total porosity, and both volumetric flow rates are known, the gas volume fraction in active (i.e., not stagnant) flow α can be computed by setting equal the Ergun equations (or similar equations for ΔP) for both phases. One can then predict the $\Delta P / L$. No matter the type of flow both phases must have equal pressure drops, because they are in parallel. The mass balance relates the two phase velocity to the actual~~interstitial~~ phase velocities:

$$U_{tp} = U_L(1 - \alpha) + U_G(\alpha) = G[(1 - X)/\rho_L + X/\rho_G] \qquad (\cancel{6)}9)$$

The gas flow's effect on the liquid is both to reduce its effective cross sectional area and to provide a nearly zero-shear interface. The liquid flow's effect on the gas is also to reduce its effective cross sectional area. Therefore actual two-phase flow pressure drops

Commented [HK49]: How do I do this?

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Commented [HK50]: Is this number given somewhere? Where is it coming from?

Commented [KMD51R50]: It's not a single #, it's a calculation. Anyone doing a fluids experiment has been exposed to the concept and in particular the equation for Re_p. For one thing, it's always introduced in conjunction with the Ergun equation.

Commented [HK52]: What is the comparison? What are greater values?

Commented [HK53]: What is an early peak or a long tail?

Commented [HK54]: What is U_L, U_G and U_{tp}?

Commented [JL55]: Put this as equation.

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Commented [HK56]: What is quality?

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Commented [HK57]: This is not very clear. The relationship is unclear, the variables not all defined, how does velocity and viscosity belong together? What is the difference between homogenous and stratified flow? Which model is better and why especially for the two-phase system?

Commented [HK58]: Can this be more explained in the relationship and how one affects the other?

typically~~can~~ exceed ΔP's calculated simply on the basis of measuring or calculating α and applying a packed bed ΔP equation (using α instead of ε).

Procedure:

1. The apparatus~~Permeameter~~ is primarily operated through the ~~Honeywell (~~distributed control system~~)~~ ~~*Experion*~~ interface. A **Perm** P&ID schematic appears and opening / closing automated valves is point and click.

2. To establish water flow to either bed #4 or #5, open the inlet and exit valves to the bed being tested and the water supply solenoid. Use the flow controller to start water flowing through the bed, raising it gradually. Good starting points are 400 mL/min for bed #4 and 500 mL/min for bed #5. Monitor the differential pressure across the beds. Vary the flow to cover ~~over~~ the entire possible range of the DP transmitter.

3. Power up the spectrometer equipment and establish communication with the control console. Spectrometer procedures are detailed in the operating manual (SpectraSuite). The ~~C~~calibration~~e~~ of the spectrometer for~~using~~ the fluorescent dye standards will be provided. ~~The spectrometer fiber optic probe can be inserted directly into the test tubes. bottles.~~

4. Perform one tracer test each~~s~~ on beds #4 and #5 using 50 ppm~~the~~ dye in DI water as the tracer, at a single average flow rate for each bed. ~~This may involve some trial and error to get the dye concentration in a correct range such that the outlet dye concentrations do not saturate the spectrometer.~~ Insert the spectrometer probe into the probe sample point (Fig. 1). At the PERM interface, ~~C~~change injection valve status from “Running” to “Charging.” Inject the tracer~~sample~~ using the syringe provided into the sample~~injection~~ valve. ~~Inject sample - c~~Change status to “Running”. Clean out the injection chamber of the sample valve by changing its status back to “Charging~~.~~”, ~~D~~detaching (it screws off) and loading the syringe with water, then injecting at least 100 mL of~~the~~ water into the valve~~chamber~~. When the injected sample has completely exited the bed~~,~~ (spectrometer absorbance returns to base line), change the valve state back to “Running” and let the water flow through the valve for 10-15 min at a high flow rate before using it again.

Conducting Two-Phase Flow Pressure Drop Experiments:

1. Be sure that the water valves to the beds are closed. Be sure that the inlet and exit valves to bed #5 are open. Be sure the drain valve is open. Be sure the manual valve for the air to the beds is closed.
2. Slowly open the air regulator to establish an air flow (< 5 psig at first). Open the manual valve for the air to the beds.
3. Set water flow controller at desired setpoint (700 mL/min) and open manual valve. Route water/air flow to gas-liquid separator (see valving in Fig. 1).
4. Confirm that water is exiting to drain. ~~You may want to~~ C~~c~~lose the valve to the drain for a period of time to build up a liquid head in the gas-liquid separator. This will result in better separation of the air and water.

Commented [HK59]: Please avoid using trademarks, if possible describe purpose of the software/machine.

Commented [HK60]: Why bed #4 or #5 and not the other ones? What are typical values for a good water flow?

Commented [KMD61R60]: The normal assignment for this lab is 2, 3, 4, 5. They would only have time to do 2 beds for the video. Bed 5 is the only one set up for 2-phase flow. Bed 4 gives more interesting behavior than the other 2.

Commented [HK62]: Provide some information in order to enable scriptwriting. What would the person enact? Which buttons to push? What are the signs that communication between the apparatuses was established?

Commented [HK63]: How is this done?

Commented [HK64]: Which bottles? What are they used for? How large? Glass or plastic bottles?

Commented [HK65]: How often? How many samples are taken?

Commented [HK66]: What is the probe?

Commented [HK67]: Sample? The tracer? How much? What is the dye?

Commented [HK68]: From where?

Commented [K69R68]: It's a simple screw mechanism.

Commented [HK70]: Which chamber? The injection chamber?

Commented [HK71]: How do I know it completely exited it?

Commented [HK72]: Where are the UV/Vis readings taken?

Commented [HK73]: Why is only bed #5 used in this experiment?

Commented [KMD74R73]: It is the only one set up for 2-phase flow.

Commented [HK75]: What would be a desired set point?

Commented [HK76]: How? This needs more description.

Commented [K77R76]: The manual valves that would have to be manipulated are shown in Fig. 1. You are just switching from one drain to the other.

Commented [HK78]: Why is this beneficial?

5. Adjust air flow (typically < 2 SCFM) as desired using the pressure regulator and the dry test meter on the gas exit line. C~~You can c~~lose the drain valve for short periods of time to get a correct gas flow reading on the wet test meter.
6. Conduct two-phase flow pressure drop (use DP transmitter) experiments using bed #5, at ~~a single low liquid flow rate but~~ multiple air rates. Try to cover the range of the DP transmitter. Disconnect the dry test meter if you see water exiting from the gas exit line.

Representative Results

Obtain the RTDs (E-curves, using Equations~~.~~ 1-2) after subtracting an appropriate baseline (if necessary) from the spectrometer signals. An example of baseline correction for Bed #3 (not used here) is in Figure 3. Using Equations 1-3, ~~C~~calculate average porosity, tracer mass, mean residence time, variance and variance divided by mean squared from the RTDs. Compare calculated tracer mass with injected mass – if they aren't within expected precision, examine how you are determining the baseline in the spectrometer measurements (and maybe determine differently). Examine how the variance ~~results~~ compares to the prediction from dispersion theory (Equations 4-5); deviations denote excessive channeling.

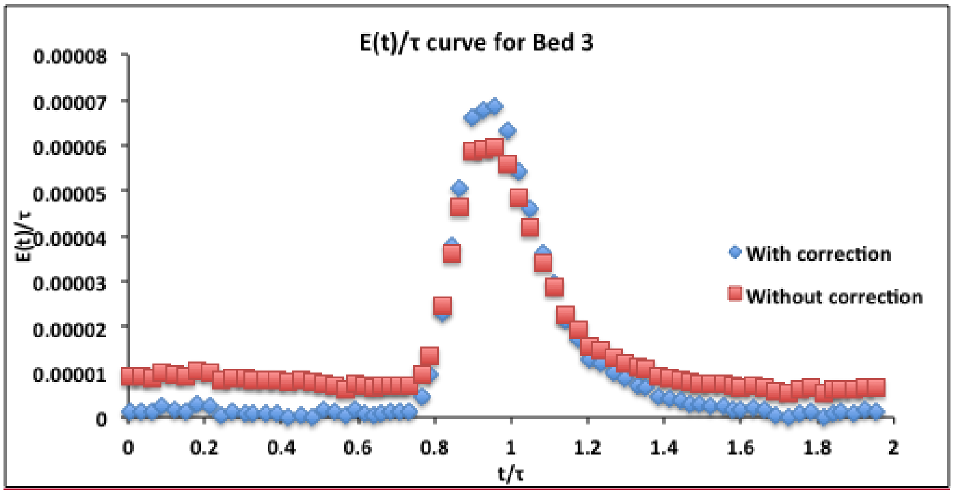

Figure 3. Bed #3 dimensionless RTD E-curve (390 mL/min, 50 ppm tracer injection) with and without baseline correction. The calculated τ from Equations 2 and 3 was 3.6 min. The baseline correction was made by subtracting two average baseline values, one before and one after the maximum. The one before was subtracted from all values prior to the maximum, the other after was subtraced from all values after the maximum.

Once the porosities of the beds (~~Eq. 3~~Equation 3) have been found, the Ergun equation can be used to predict the ΔP's ~~and friction factors~~ for the water flow experiments. The

Commented [HK79]: What is the desired airflow? What should I be looking for?

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Commented [HK80]: Elaborate/describe. How is this conducted? What is a single low liquid flow rate? What are multiple air rates? How often do you take samples, how to inject, where to get the readings from? How to operate the machine? Etc. It needs to be a step by step description.

Commented [KMD81R80]: No samples to take. Have specified a single liquid flow rate and a range for the air.

Commented [HK82]: Why is this used In the first place? What does it tell me if the water is exiting, what does it mean for the experiment?

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Commented [HK83]: Could you provide a graph for this to somehow visualize this step?

Commented [HK84]: The equations used for this would be great so that they could be presented again on the screen.

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Commented [HK85]: While these are great instruction, the purpose of the video is not only to say what needs to be done for obtaining the results, but also how to interpret them and what are the conclusions. As for here it is unclear what the comparison would look like and what is the "expected" precision?

average particle diameter must first be calculated. Because particle drag is related to area for flow, the surface-area (d^2) weighting is usually the best way to obtain the average diameter for a range of particles. The average diameter can be computed as follows, obtaining the particle diameters from the information in the Materials List (ω_i is the wt fraction of particles of diameter d_i).

$$\bar{d} = \left(\frac{\sum_i \omega_i d_i^2}{\sum_i \omega_i}\right)^{0.5} \qquad (10)$$

The calculated porosities can be used to pinpoint the cause of any discrepancies between predicted (by the Ergun equation) and measured ΔP's. For example, the minimum porosity for close-packed spheres is 0.36. It is unlikely that any real ε of an entire bed is less than 0.3. Predicted ΔP's >> actual ΔP's suggest channeling (short-circuiting) along the walls or in the upper portion of the bed when settling occurs. Such phenomena would result in a low ε calculated from the E-curve, leading to high predicted ΔP's. This is the case in Figure 4 for both beds #3 and 4. Note that the more expected ε = 0.36 reproduced the Ergun equation results except at a very high flow rate where a high percentage of the flow was through low voidage regions. This channeling can actually be observed in the experiment.

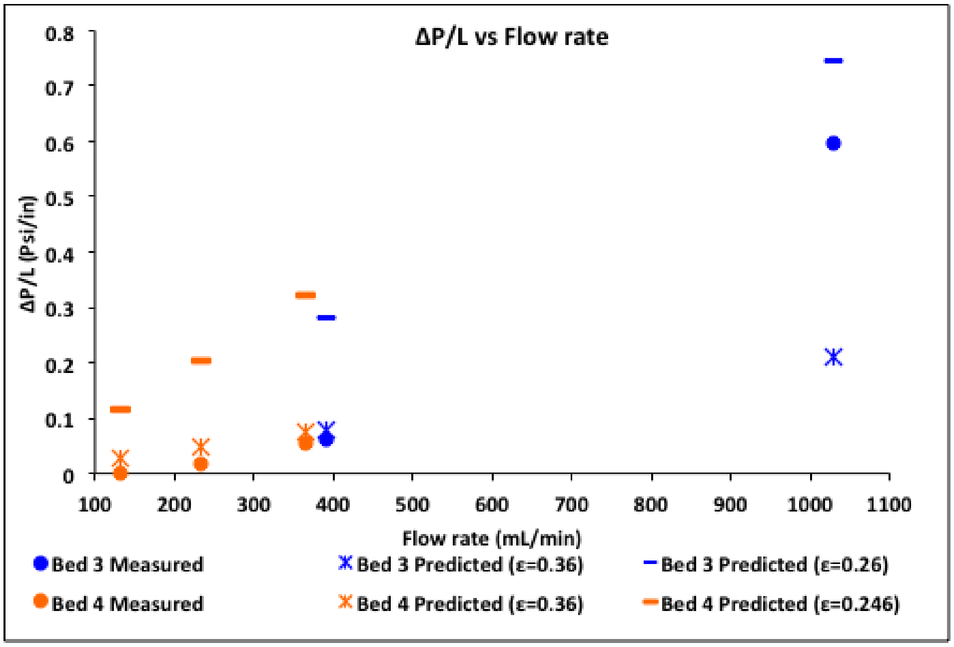

Figure 4. Experimental ΔP's compared to the predictions of the Ergun equation, both at ε = 0.36 and the ε values determined from the E-curves.

Predicted ΔP's << actual ΔP's suggest channeling only through the lower half of the bed, or partial bed blockage. For these beds, this is unlikely.

For the two-phase flows, compute predicted ΔP's by both homogeneous flow and stratified flow theories using Equations 6-9. For stratified flow, one must solve the Ergun equation and Equation 9 simultaneously to obtain α, setting Ergun ΔP/L (liquid) = Ergun ΔP/L (gas). Then ~~C~~compare computed vs. actual ΔP's and see which theory applies best, or if in fact either theory applies. Other flow regimes (e.g., slug, mist or inhomogeneous bubbly flows) are possible, as are major flow distortions due to channeling, which is often more prevalent in two-phase flows.

~~A 50 ppm concentration tracer (dye) solution injected into bed 3 at 390250 mL/min gave the dimensionless E-curve shown in Figure 2. The mean residence time τ computed from the curve was 3.62.75, which is less than what would be computed from the bed geometry using ε = 0.25.~~

~~The ΔP /L friction factors calculated using the experimental, and predicted (Ergun equation, but using the experimental ε) and typical ε for packed beds (ε=0.36) are shown in Figure 3. The predicted friction factors (and ΔP/L's's) are much higherlower for both bed #3 and #4., in complete agreement with the assumption of flow only through the lower part of the bed. Conversely, for bed #4 the experimental ΔP's are either less than or about the same as those predicted by~~

For two-phase flows through bed #5, the ΔP's ~~gas volume fractions~~ calculated using homogeneous flow theory prove to be better than those using stratified flow theory (Table 1), although as seen neither theory applies exactly~~but not by much~~. The high actual ΔP's ~~is~~ suggest~~s~~ severe channeling in a horizontal bed during two-phase flow – the liquid is confined to a small portion of the cross-sectional area~~, resulting in high ΔP's~~. Indeed, the gas volume fractions estimated by visual inspection looked to be at least 0.90. The liquid was also confined to the non-wall region of lower voidage, which increases~~ing~~ ΔP. The results reflect the limitations of the simpler rheological models for two-phase flow, and why far more sophisticated microrheological models are finding more use today.

Table 1: Gas volume fractions α and pressure drops in two-phase flow, bed #5.

Q water	Q air	Q air	α (stratified)	ΔP (stratified) psi	ΔP (homogeneous) psi	ΔP (actual) psi
mL/min	ft^3/min	mL/min				
1100	1.62	45900	0.58	2.2	12	17
1100	1.26	35700	0.47	1.7	10	14
1100	1.11	31400	0.38	1.5	9.5	11

Commented [HK86]: How would I know which applies and what if neither theory works? What are the results/values we would be looking for?

Commented [KMD87R86]: Neither theory does work. We will try to make that evicdent now.

Commented [HK88]: The procedure never mentioned bed #3, where is this coming from?

Commented [KMD89R88]: We're not using bed #3 for the video, but the E-curve is a good example of baseline correction, so I left the result in.

Commented [HK90]: What are the axes? What does skewed toward t=0 mean? What would be a perfect curve for no channeling?

Commented [KMD91R90]: It would be a perfect Gaussian and the variance would conform to Equation (5). I hope we made that clearer now.

Commented [HK92]: Describe the figure? Again bed #3 was never mentioned in the procedure. What is the difference between bed #3 and 4, how do the result support this? What is the relationship of friction factor to flow rate? Does high friction factor mean less channeling because particles are equally distributed?

Commented [HK93]: Can you specify where to look at the table and which values to compare and what they tell us?

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1100	0.930	26300	0.19	1.3	8.6	8
500	0.73	20700	0.58	0.66	3.4	12
500	0.50	14200	0.47	0.50	2.7	9
500	0.39	11000	0.38	0.40	2.7	6
500	0.16	4250	0.19	0.29	1.4	3

Commented [KMD94]: I deleted the column for α in homogeneous flow because you don't need it to do the Ergun calculations and it's a meaningless number (the assumption in homogeneous flow is that gas + liquid form a pseudo single phase). For stratified flow, you need α to solve for ΔP using the Ergun equation.

Summary

In this experiment the real flow behavior of horizontal packed beds, both in single and two-phase flow, was contrasted to the simpler theoretical models for pressure drop and dispersion (flow spreading in the axial direction, deviating from plug flow). The utility of tracer tests in probing for maldistribution ("channeling") in such beds has been demonstrated, and it has even been shown that certain metrics calculated from the tracer tests can give some idea of the cause of the channeling. These calculations using the tracer tests, such as computing the E-curve, are normally known as "residence time distribution" (RTD) theory.

Channeling~~, which~~ in single-phase flow can occur along walls or any other low voidage region, for example, if settling occurs in a horizontal bed~~just by the flow occupying only the lower portion of a horizontal contactor, degrades the performance of a reactor or adsorber~~. Channeling in two-phase flow can result from even more complex causes, and as seen simple two-phase flow theories seldom predict pressure drops in packed beds. Channeling increases downstream separation costs or can ruin the product. A goal of design is always to minimize the extent of channeling by finding the optimal bed and particle diameters for a given desired Q, and by packing a bed in a way to minimize settling.

Commented [HK95]: This is great to explain the significance of measuring maldistribution in the packed beds and I would put this section into the overview.

Commented [KMD96R95]: It's being moved to overview.

The tracer method of testing is a simple way to quantify the RTD. However, the tracers are seldom the same molecules as used in the process (although they can be close, if isotopes are used). Therefore tracer molecules may not behave in exactly the same ways as reactant or adsorbate molecules in the fluid phase. In particular, it is important that the tracer not adsorb on the solid particles, because then it cannot be fully characteristic of a fluid molecule.

Applications

The time each molecule of a reactant spends inside a chemical reactor is ~~obviously~~ an important determinant of the macroscopic conversion and selectivity to the desired product. The occurrence of "dead zones" (regions of stagnant flow) often lead to poorer than expected selectivities even when the conversions are not much affected. This is one reason why RTD theory is so important in reactor design.[4]

Tracers are also used by environmental and petroleum engineers to help characterize subsurface solid packing structure. In these applications, two wells are drilled some distance apart; a tracer is injected into one and recovered at the other. Because the earth's subsurface is highly heterogeneous, the effluent profiles (E-curves) are typically nonsymmetrical, indicating the presence of preferential flow paths. This information helps characterize the structure of the subsurface strata, which is important for modeling petroleum recovery and contaminant transport in groundwater.

In environmental engineering, the use of partitioning tracers can be used to locate and quantify organic contaminants in subsurface strata. An inert tracer is injected to characterize the flowing (aqueous) phase between two wells. A partitioning tracer is then injected, partitioning preferentially into an organic contaminant phase if one exists. The tracer is light enough that it will eventually diffuse out of the organic phase ~~once the pulse has passed~~ . This behavior manifests itself as a time delay as compared to the inert tracer, and comparison of these two can be used to deduce the volume of stagnant organic phase present.

Two-phase flows are also commonly found in power plants, in non-reactor, non-adsorber applications. An example is boiling heat transfer, with the steam created in a boiler. They are also found in all distillation columns, absorbers and strippers, although in vertical rather than horizontal configuration.

Legend

Figure 1: Process and instrumentation diagram of the apparatus~~permeameter~~.

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Figure 2: 3-D rendering of the apparatus. Bed #1 is at the top, bed #5 at the bottom. The water control valve is on the left (red bonnet). The DP transmitter is at the top center (blue).

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Figure 3: Bed #3 dimensionless RTD E-curve (390 mL/min, 50 ppm tracer injection) with and without baseline correction. The calculated τ from Equations 2 and 3 was 3.6 min. The baseline correction was made by subtracting two average baseline values, one before and one after the maximum. The one before was subtracted from all values prior to the maximum, the other after was subtraced from all values after the maximum.

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Figure 4: Experimental ΔP's compared to the predictions of the Ergun equation, both at ε = 0.36 and the ε values determined from the E-curves.

Table 1: Gas volume fractions α and pressure drops in two-phase flow, bed #5.

Materials List:

Name	Company	Catalog Number	Comments
Equipment			
Bed #3 – glass beads	Grainger	25-40 mesh (50%) 60-120 mesh (50%)	Packed in parallel
Bed #4 – glass beads and blast sand	Grainger	60-120 mesh (90%) - glass 80-120 mesh (6%) - glass 120-200 mesh (4%) - sand	Mixed together

Bed #5 – glass beads	Grainger	5-10 mesh	
Dry test meter	Singer	Model 803	
Fiber-optic UV-Vis spectrometer	Ocean Optics	Model USB2000	Includes Ocean Optics DT-1000 light source
Test tubes	VWR	10 mL	For calibration
Reagents			
Yellow/green fluorescent dye	Cole-Parmer	0298-17	Used to make up tracer solutions

References:

1. "Encyclopedia of Chemical Engineering Equipment." *Distillation Columns*. http://encyclopedia.che.engin.umich.edu/Pages/SeparationsChemical/DistillationColumns/DistillationColumns.html. Accessed 9/22/16.

2. "Encyclopedia of Chemical Engineering Equipment." *Absorbers*. http://encyclopedia.che.engin.umich.edu/Pages/SeparationsChemical/Absorbers/Absorbers.html. Accessed 9/22/16.

3. The ~~use of the~~ Ergun equation and the particle Reynolds number Re_p are discussed ~~can be found~~ in most Chemical Engineering fluids texts, for example: N. de Nevers, *Fluid Mechanics for Chemical Engineers*, 3rd Ed., McGraw-Hill, 2004, ~~c~~Ch. 11. A derivation can be found in: M.M. Denn, "Process Fluid Mechanics", Prentice-Hall, 1980, ~~c~~Ch. 4.

4. A more complete treatment of RTD theory can be found in most Chemical Engineering reactor design texts, for example: Fogler, H.S., "Elements of Chemical Reaction Engineering", Prentice-Hall, 2006, ~~c~~Ch. 13.1-13.3 and 14.3-14.4 (dispersion models); Levenspiel, O., "Chemical Reaction Engineering", 3rd Ed., John Wiley, 1999, ~~c~~Ch. 11 and 13 (dispersion models); Missen, R.W., Mims, C.A., and Saville, B.A., "Introduction to Chemical Reaction Engineering and Kinetics", John Wiley, 1999, ~~c~~Ch. 19 and 20.1.

5. Levy, S., "Two Phase Flow in Complex Systems", John Wiley, 1999, ~~c~~Ch. 3.

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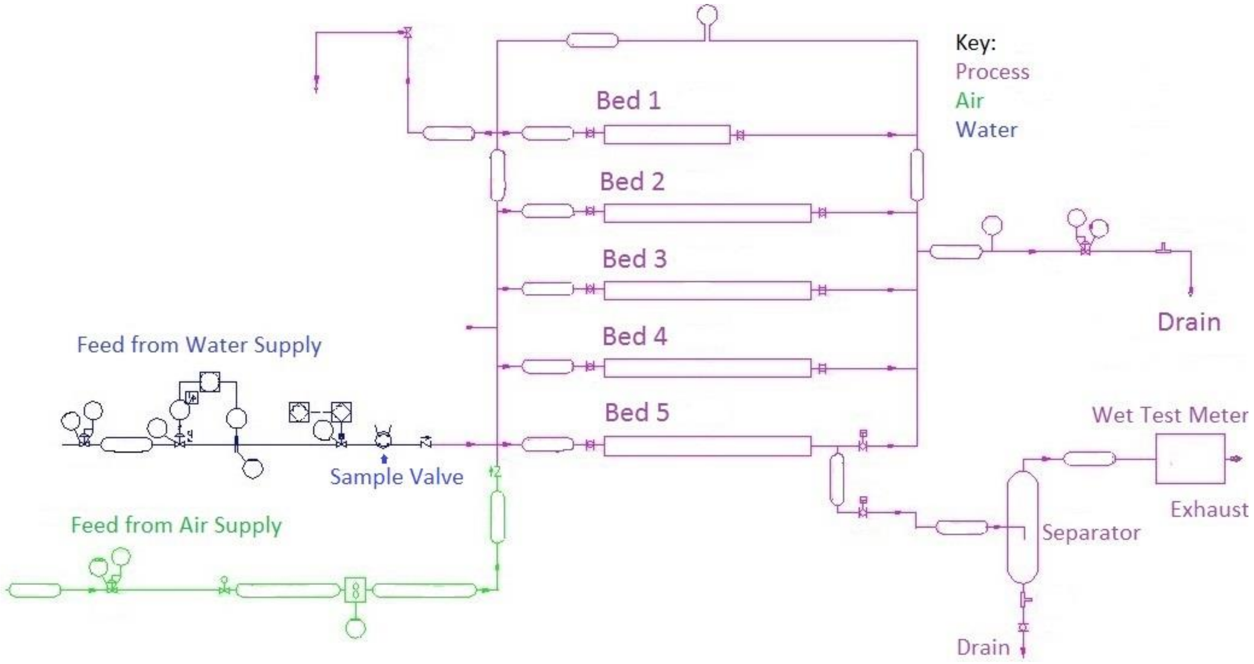

Figure 1

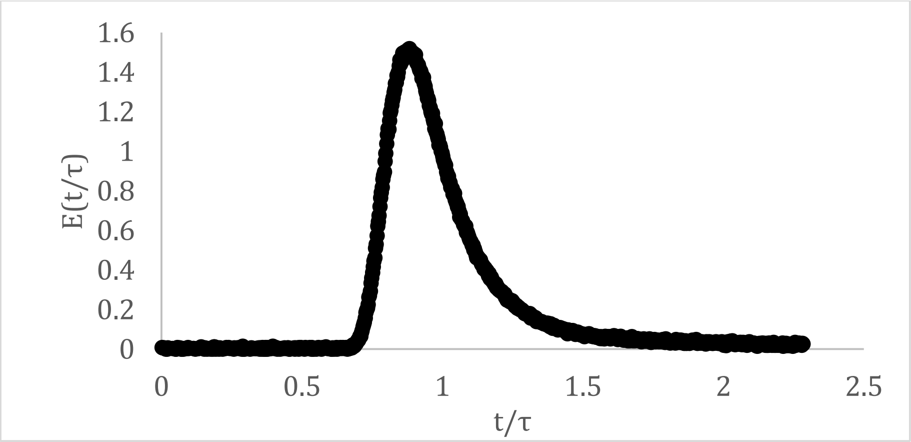

Figure 2

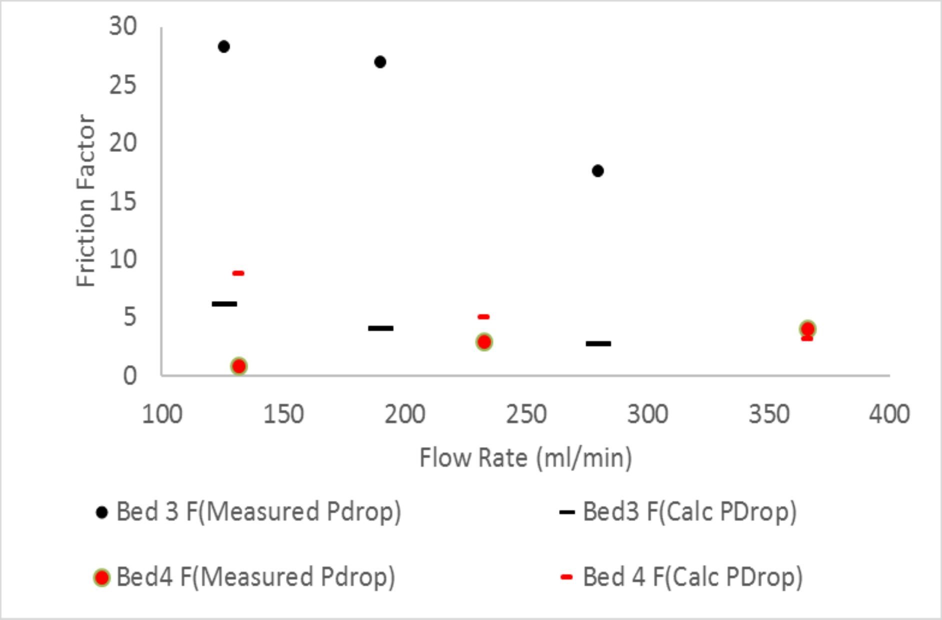
Friction Factor
30
25
20
15
10
5
0
100
150
200
250
300
350
400
Flow Rate (ml/min)
● Bed 3 F(Measured Pdrop)
— Bed3 F(Calc PDrop)
● Bed4 F(Measured Pdrop)
— Bed 4 F(Calc PDrop)

Figure 3