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CHEMICAL ENGINEERING LABORATORY 3, CE 427
PACKED AND FLUIDIZED BED EXPERIMENT

INTRODUCTION

Packed and fluidized beds play a major role in many chemical engineering processes. Packed-bed situations include such diverse processes as filtration, wastewater treatment, and the flow of crude oil in a petroleum reservoir. In these cases, the interest centers on the pressure drop through the bed as a function the volumetric flow rate or superficial velocity.

If the particles in the bed are loose and there is sufficient volume in the device containing the particles, the particles may fluidize at high flow rates. Fluidized beds are used extensively in the chemical process industries, particularly for the cracking of high-molecular-weight petroleum fractions. Such beds inherently possess excellent heat transfer and mixing characteristics. In the study of the fluid-mechanical behavior of these beds, the focus here is on the incipient fluidization velocity and the dependence of bed expansion on the superficial velocity.

THEORY

The theory for this experiment is covered in Chapter 7 of McCabe, Smith, and Harriott (M,S&H). The following material is a condensation of that chapter as it relates to the experiment at hand. As an aid to you, some specific equations in M,S,&H are referred to. There are three areas of interest to us: (1) Relationship between the pressure drop and the flow rate; (2) Minimum fluidization velocity, and; (3) Behavior of the expanded bed.

(1) Relationship between pressure drop and flow rate

The flow of a fluid, either liquid or gas, through a static packed bed can be described in a quantitative manner by defining a bed friction factor, f_p, and a particle Reynolds number, $N_{Re,p}$, as follows:

$$f_p \equiv \frac{\Delta p \, g_c f_s D_p e^3}{r \overline{V}_o^2 L(1-e)} \qquad \text{Similar to 7.19 in M \& H.} \qquad (1)$$

Note that this equation cannot be derived directly by extrapolating the case of flow through a circular conduit since friction factor defined in both cases is different (see McCabe and Smith 4th edition, pg. 137)

$$N_{Re,p} = \frac{\rho \overline{V}_o D_p}{\mu} \qquad (2)$$

where

ΔP	=	pressure drop across the bed
L	=	bed depth or length
g_c	=	conversion constant (= unity if SI units are used)
D_p	=	particle diameter
ρ	=	fluid density
ε	=	bed porosity or void fraction
$\bar{V}_o$	=	superficial fluid velocity
μ	=	fluid viscosity
ϕ_s	=	sphericity

The friction factor and the Reynolds number are dimensionless. Some typical sphericity factors are given in McCabe, Smith and Harriott (p. 928, Table 28.1).

For laminar flow, where only viscous drag forces come into play, $\left(N_{Re,p} < 20\right)$, experimental data may be correlated by means of the Kozeny-Carman equation:

$$f_p = \frac{150 \quad (1-e)}{N_{Re,p} f_s} \qquad \text{Similar to 7.17 MS \& H} \qquad (3)$$

Note: According to Yates ("Fundamentals of Fluidized-bed Chemical Processes," by J. G. Yates, Published by Butterworths, 1983, p. 7-8) the factor of 150 was originally given by Carman as 180 for the case of laminar flow. Ergun later suggested a better value was 150 when the particles are greater than about 150 μm in diameter.

For highly turbulent flow where inertial forces predominate, $\left(N_{Re,p} > 1000\right)$, experimental results may instead be correlated in terms of the Blake-Plummer equation:

$$f_p = 1.75 \qquad \text{Similar to 7.20 MS \& H} \qquad (4)$$

While both equations (3) and (4) have a sound theoretical basis, Ergun empirically found that the friction factor could be described for all values of the Reynolds number by simply adding the right-hand sides of equations (3) and (4). Thus:

$$f_p = \left(\frac{150(1-e)}{N_{Re,p} f_s} + 1.75\right) \qquad \text{Similar to (7 - 22) MS \& H} \qquad (5)$$

(2) <u>Minimum fluidization velocity</u>

At a sufficiently high flow rate, the total drag force on the solid particles constituting the bed becomes equal to the net gravitational force and the bed becomes fluidized. For this situation a force balance yields:

$$(-\Delta p)A = LA(1-\varepsilon_M)(\rho_p - \rho)g/g_c = M(\rho_p - \rho)g/(g_c \rho_p) \qquad (6)$$

where

ε_M	=	void fraction at the minimum fluidization velocity
A	=	cross-sectional area of the bed
ρ_p	=	particle density
g	=	gravitational constant
M	=	total mass of packing.

This is Eq. 7.48, 7.49 MS&H . The superficial fluid velocity at which the fluidization of the bed commences is called the incipient or minimum fluidization velocity, $\bar{V}_{0M}$. The incipient fluidization velocity may be determined by combining equations (1), (3), and (6) with the following result [Eq. (7.52) MS&H] for the case of small particles and consequent, $N_{Re}<1$:

$$\bar{V}_{0M} = \frac{g(\rho_p - \rho)\varepsilon_M^3 \phi_s^2 D_p^2}{150\,\mu(1-\varepsilon_M)} \qquad (7)$$

This equation is the basis for some empirical equations found in the literature. The terms can be grouped as follows:

$$\bar{V}_{0M} = \frac{\varepsilon_M^3 \phi_s^2}{150(1-\varepsilon_M)} \cdot \frac{g(\rho_p - \rho)D_p^2}{\mu} \qquad (8)$$

The first factor contains the sphericity of the particles and the bed porosity at the point of incipient fluidization. Neither of these factors is usually known with a high degree of accuracy. If spheres are assumed $(\phi_s = 1)$ and a reasonable value of voidage, say $\varepsilon_M = 0.4$, then the first factor is 0.00071. The factor is quite sensitive to ε_M. For example, if $\varepsilon_M = 0.413$, then the factor is 0.0008.

One investigator, [D. Geldhart, "Types of Fluidization," Powder Technology, **7** (1973), 285-292; Geldhart and Abrahamsen, Powder Technology, **19** (1978), 133-136] simply determined the first factor from his data and actually found 0.0008 to be the best value; that is, he reported the following correlation:

$$\bar{V}_{0M} = 0.0008\frac{g(\rho_p - \rho)D_p^2}{\mu} \qquad (9)$$

Behavior of the expanded bed

The expansion of fluidized beds is discussed in the text on Pages 170-173. The treatment to be used here is slightly different. For fluid velocities exceeding the incipient fluidization velocity, the bed expands. The porosity, ε, of an expanded bed may be related to the superficial fluid velocity, $\overline{V}_o$, by means of an empirical relation suggested by Richardson and Zaki (1,2):

$$\frac{\overline{V}_o}{u_t} = \varepsilon^n \tag{10}$$

where u_t is the terminal velocity of a spherical particle in a fluidizing medium (3). The exponent, n, depends on the flow conditions -- that is, on the Reynolds number. Thus:

$$N_{Re,p} < 0.2 \qquad n = 4.65 \tag{11}$$

$$0.2 < N_{Re,p} < 1.0 \qquad n = 4.35 N_{Re,p}^{-0.03} \tag{12}$$

$$1 < N_{Re,p} < 500 \qquad n = 4.45 N_{Re,p}^{-0.1} \tag{13}$$

$$N_{Re,p} > 500 \qquad n = 2.39 \tag{14}$$

Because the terminal velocity, u_t, is a constant for a given particle, it can be seen that Equation (10) above is essentially the same as the empirical equation in the text; namely Eq. (7.59) MS&H.

The void fraction of the expanded bed, ε, is related to that at incipient fluidization by the following equation:

$$L = L_M \frac{1-\varepsilon_M}{1-\varepsilon} \tag{7-58 MS\&H}$$

where L_M and ε_M are the bed height and void fraction at incipient fluidization, and L is the measured height of the expanded bed. Therefore, since L_M and ε_M are known, ε can be calculated from the measured height, L, of the expanded bed.

In Equations (11)-(14) the Reynolds number is based on the particle diameter, D_p, and the terminal velocity, u_t. Therefore it is necessary to know the terminal velocity. By means of a force balance it be shown that the terminal velocity for spherical particles is:

$$u_t = \sqrt{\frac{4D_p(\rho_p - \rho)g}{3C_D\rho}} \tag{15, \&.37 MS\&H}$$

where C_D denotes the drag coefficient. A graph of C_D versus $N_{Re,p}$ is shown in the text (Figure 7.6, p. 158). To find C_D, you need to know u_t so that $N_{Re,p}$ can be calculated. There are two ways of doing this: i) One could do this by trial-and-error. Thus, you could guess u_t, calculate $N_{Re,p}$, look up C_D on the graph, and put the resulting value in Eq. (15). If the calculated value of u_t did not match the guess (it surely wouldn't on the first try!), you would guess again. ii) We can also do this without trial-and-error. For this square both sides of Eq. (15) and utilize the definition of $N_{Re,p}$ (Eq. (2)) to obtain:

$$C_D N_{Re,p}^2 = \frac{4 D_p^3 \rho(\rho_p - \rho)g}{3\mu^2} \tag{16}$$

All parameters on the right are known. This suggests that a plot of $C_D N_{Re,p}^2$ versus $N_{Re,p}$ can be constructed and used to avoid the trial-and-error procedure.

The plot is prepared in the following way. Pick a series of point coordinates off the plot shown above. Some examples for spheres are:

$N_{Re,p}$	C_D	$C_D N_{Re,p}^2$
.001	22,000	22
.01	2,200	22
.1	220	22
1,000	0.48	480

Pick off a dozen similar pairs. Then plot $C_D N_{Re,p}$ as the ordinate against corresponding $N_{Re,p}$ as the abscissa. For each bed, calculate $C_D N_{Re,p}^2$ from Eq. (16). From your plot read the corresponding $N_{Re,p}$. Then use Eq. (2) to calculate u_t.

EQUIPMENT AND PROCEDURE

In this experiment the friction factor will be measured as a function of Reynolds number for the flow of air through a bed of solid particles. Experimental results will be compared with theoretical predictions for the appropriate flow regimes. A flowsheet of the experimental set-up is depicted schematically in Figure 1 below. The equipment includes two transparent beds, rotameters, manometers, a source of low pressure air, and appropriate valves and fittings.

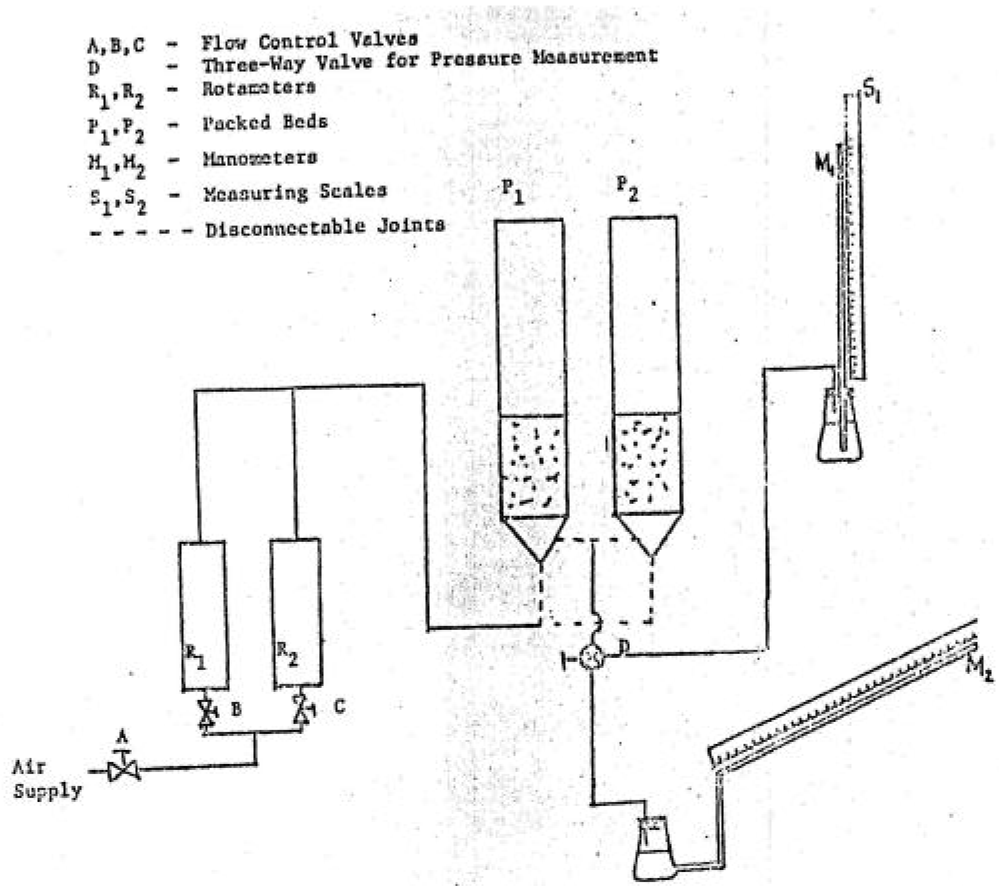
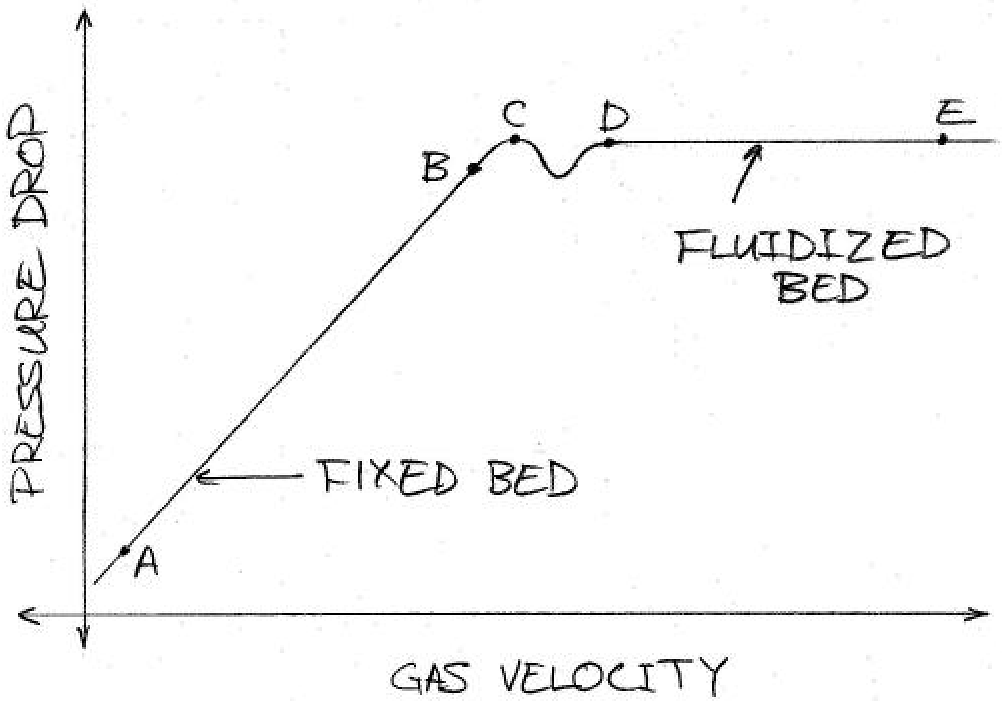


Fig. 1. Flow Sheet for Packed Bed Experiments

Measure the bed height after tapping the bed gently until no further change is observed. Close Valves B and C. Open Valve A. Control the flow of air through the system by manipulating Valve B or C depending on the rotameter used. Open Valve D for pressure drop measurements.

Increase the flow rate of air in small steps noting the rotameter and manometer readings until the bed is fluidized and the pressure drop does not change appreciably. Also, record the corresponding bed height at each flow rate. Continue the measurements until the bed is appreciably fluidized and obtain at least fifteen readings in the packed-bed region and ten readings in the fluidized-bed region. Decrease the flow rate noting the flow rate and pressure drop values.

The region where fluidization just begins -- namely, at the "minimum fluidization velocity" -- is of special interest. In Figure 2, it occurs at Point B. [This plot is based upon one in "Design for Fluidization, Part 1," by J. F. Frantz, Chemical Engineering, September 17, 1962, pp. 161-178.] At this point, where the pressure drop through the fixed bed becomes equal to the weight of the bed per unit area, a slight rearrangement of particles occurs and the particles shift position so as to present maximum flow area to the gas. Quoting from the article above, "This causes a slight decrease in pressure drop, and channeling occurs. Only at a higher gas velocity does the entire bed become fully supported by the gas stream. Leva, Shirai and Wen realized this phenomenon took place, and thus the defined minimum fluidization velocity as a gas velocity 10% greater than the point at which, with increasing velocity, the pressure drop through the fixed bed first equals the weight of the bed per unit



area."

In view of this, repeat the run to check for reproducibility, making sure that you get sufficient data in the somewhat ill-defined region between Points B and D.

The theory of this experiment is built around the assumption that at "steady state," the particles are uniformly distributed in the bed. As you increase the flow rate of air, take some notes concerning what the bed looks like at various stages. This may help to explain some discrepancies between measured and theoretical values.

Repeat the measurements on the other bed by reconnecting the appropriate lines.

Repeat the experiment at least twice (preferably thrice) with each bed to ensure that you are sufficiently sure about the quantity of the data.

In a separate experiment, the porosity of a container of solid particles was measured using the method of water displacement. This involves first weighting a measured volume of dry particles. Water is then slowly added to the particles until the upper surface is wet. The weight of the water added can then be used to calculate the porosity of the bed. This porosity corresponds to the value, ε_M, in Equation (6). These values of bed porosity and particle spherocity are provided with the experimental system.

Calculations

1. For each bed, plot the measured pressure drop (in cm of H_2O) versus the volumetric flow rate in liters/min. Note that the manometer is at an angle of 19° and therefore, to get pressure drop, the difference in manometer levels must be multiplied by sin(19°). There will likely be a "hysteresis effect," in that the pressure drop curve for increasing flow rate will differ from that for decreasing flow rate. Therefore, use different symbols for increasing and decreasing flow rates. Explain the likely cause of this effect.

2. For each bed, calculate the friction factor and corresponding Reynolds number for each data point using Equations (1) and (2). Then prepare a single plot of f_p versus $N_{Re,p}$ which combines the results for the two beds. Use a different symbol for each bed and show symbols only (no lines). On this plot, also show the predicted values from Equations (3), (4), and (5). Show these as solid or dashed lines, and do not show the points used for determining these plots.

3. From your plots in Part 1 above, determine the pressure drop at the point where fluidization begins in cm H_2O. Using Equation (6), calculate the predicted value of the pressure drop at this point in cm H_2O. Note that you have to know ε_M. Leva [Max Leva, "Chemical Engineering, November, 1957, pp. 266-270] gives a correlation that can be used. Alternatively, you could assume $\varepsilon_M \cong \varepsilon$ as provided to you by the TA for the packed bed before fluidization commences. Comment on your findings.

4. As noted earlier, the minimum fluidization velocity, $\overline{V}_{0M}$, is of considerable interest. From your results in Part 1, calculate experimental V_{0M} in m/s and ft/s for each of your runs and beds. Using the theoretical Equation (7), to also predict theoretical V_{0M} in m/s and ft/s. Compare and contrast the results.

   Several empirical equations are available in literature based on experimental data. Three of these will be used here for comparison with your experimental results.

   (a) **Leva et al.**
   Leva gives the following equation for the mass velocity at the point at which fluidization begins:

$$G_{mf} = 688 D_P^{1.82} \frac{\left[\rho(\rho_p - \rho)\right]^{0.94}}{\mu^{0.88}}$$

where: D_p = particle diameter, inches
ρ = fluid density, lbm/ft^3
ρ_s = particle density, lbm/fm^3
μ = fluid viscosity, centipoises
$G_{mf} = \rho \bar{V}_{0M}$ = mass velocity, lbm/ft^2, hr

The value of 688 was chosen by Leva as best based on 223 experimental points. Frantz noted that the standard deviation was 33% and the average deviation was 22%. The equation is valid for Reynolds numbers, $G_{mf} D_p/\mu$, less than 5. Above 5, the value of G_{mf} must be multiplied by the correction factor, F_g, given in the following plot (??).

(b) Perry's Chemical Engineering Handbook, 6th Edition, p. 20-59.

Baeyens and Geldhart ["Fluidization and Its Applications," Proc. Int. Symp. Toulouse, 253 (1973)] gives "one of the better correlations:"

$$\bar{V}_{0M} = \frac{0.0009(r_s - r)^{0.934} g^{0.934} D_p^{1.8}}{m^{0.87} r^{0.066}}$$

where $\bar{V}_{0M}$ = minimum fluidization velocity, m/s
ρ_s = particle density, kg/m^3
ρ = fluid density, kg/m^3
g = 9.81 m/s^2
D_P = particle diameter, m
μ = viscosity, kg/m,s

(c) Equation of Geldhart in "Powder Technology."

Using Equation (9).

For each bed, calculate $\bar{V}_{0M}$ from the equations of Leva, Perry's and Geldhart in m/s and ft/s. Compare the results with your experimental values.

5. Finally investigate the behavior of your beds when they are expanded. At each data point in the expanded regime, you can calculate ε from Equation (7-58). For ε_M, assume that it has the value given to you by the TA for the packed bed. Then tabulate corresponding values of $\bar{V}_0$ on the abscissa. Note that taking the logarithm of both sides of Equation (7) gives:

$$\log \bar{V}_0 = \log u_t + n \log \varepsilon$$

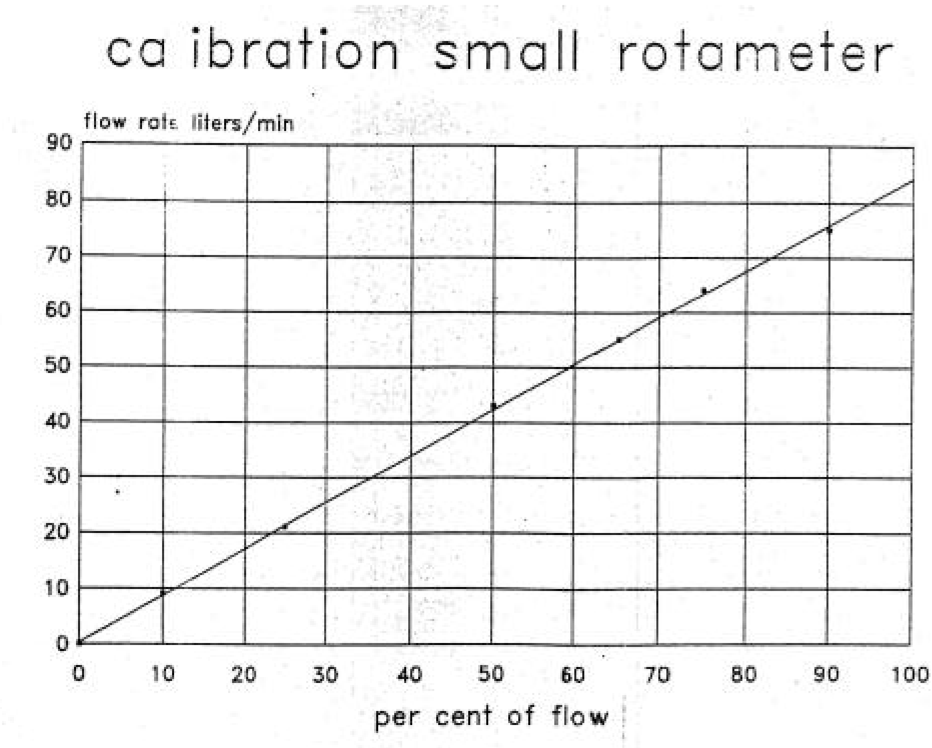
Hence the slope of each log-log plot will give n and the "intercept" should be u_t. What values of u_t do you find by this method?

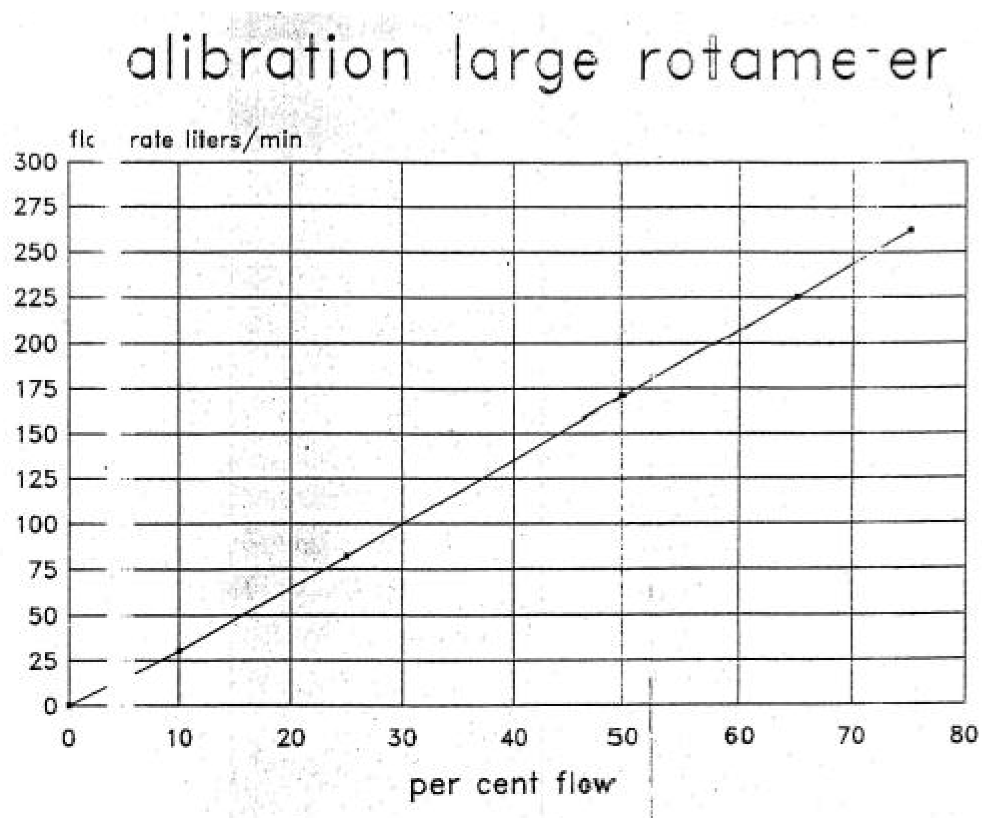
Now prepare a log-log plot of $C_D N^2_{Re,p}$ versus $N_{Re,p}$. From this plot, Equation (16), and Equation (2), determine u_t. Compare its value with that from the log-log plot of $\bar{V}_0$ versus ε.

6. Discuss your results critically.

REFERENCES

1. J.F. Richardson and W.N. Zaki, Trans. Inst. Chem. Engrs., **32**, 35 (1954).
2. J.M. Coulson and J.F. Richardson, "Chemical Engineering," Vol. II, p. 510-527, Pergamon Press, Oxford (1960).
3. R.B. Bird, W.E. Stewart, and E.N. Lightfoot, "Transport Phenomena," p. 190-194, John Wiley and Sons, New York (1960).
4. S. Ergun, Chem. Eng. Prog., **48**, 89 (1952).
5. W.L. McCabe and J.C. Smith, "Unit Operations of Chemical Engineering," 3rd Edition, p. 146-150, 159-160, McGraw-Hill, New York (1976).
6. A.S. Foust, L.A. Wenzel, C.W. Clump, L. Maus, and L.B. Anderson, "Principles of Unit Operations," 2nd edition, p. 637-547, 699-714, John Wiley and Sons, New York (1980).





Calibration curves for Rotameter: