INTERCOMPANY MEMORANDUM

CAL CHEM CORPORATION

To: CHE Seniors

Date: Fall Quarter
File: CHE 435

From: CHE faculty
Laboratory Managers

Subject: Fluidized Bed Characteristics

Our pilot plant group wishes to insert a fluidized bed reactor in one of the tail gas streams from their new pilot plant. We have such a reactor in our laboratory which could be used to study the operating characteristics of the bed.

Please investigate the relationship between superficial vapor velocity, pressure drop across the bed, and bed height.

Obtain data in the flow regions usually associated with fluidized beds. Be sure to record the minimum fluidizing velocity and to compare this value to the value predicted by theoretical correlations.

Fluidized Bed Characteristics

The upward flow of fluid through a bed of solid particles is an important process occurring in nature and in industrial operations. The apparatus in our lab was designed to allow the study of the characteristics of flow through both fixed and fluidized beds of solid particles. At low velocities, the pressure drop increases with the fluid velocity according to the Ergun equation

$$\left(\frac{\Delta P}{h}\right)\left[\frac{D_p g_c}{\rho_f v_s}\right]\left(\frac{\varepsilon^3}{1-\varepsilon}\right) = 150\frac{1-\varepsilon}{N_{\mathrm{Re}}} + 1.75 \qquad (1)$$

where

ΔP = pressure drop through the packed bed
h = bed height
D_p = particle diameter
ρ_f = fluid density
v_s = superficial velocity at a density averaged between inlet and outlet conditions
ε = bed porosity
N_{Re} = average Reynolds number based upon superficial velocity $\frac{D_p v_s \rho_f}{\mu}$

When the packing has a shape different from spherical, an effective particle diameter is defined

$$D_p = \frac{6V_p}{A_p} = \frac{6(1-\varepsilon)}{A_s} \qquad (2)$$

where

A_s = interfacial area of packing per unit of packing volume, ft^2/ft^3 or m^2/m^3

The effective particle diameter D_p in Eq. (1) can be replaced by $\phi_s D_p$ where D_p now represents the particle size of a sphere having the same volume as the particle and ϕ_s the shape factor. The bed porosity, ε, which is the fraction of total volume that is void is defined as

$$\varepsilon \equiv \frac{volume\ \ voids}{volume\ \ of\ \ entire\ \ bed}$$

$$\varepsilon \equiv \frac{volume\ \ of\ \ entire\ \ bed - volume\ \ of\ \ particles}{volume\ \ of\ \ entire\ \ bed}$$

$$= \frac{\pi R^2 h - \dfrac{weight\ \ of\ \ all\ \ particles}{particle\ \ density}}{\pi R^2 h} \qquad (3)$$

where R = inside radius of column, A_s and ε are characteristics of the packing. Experimental values of ε can easily be determined from Eq. (3) but A_s for non-spherical particles is usually more difficult to obtain. You can find values of A_s and ε for the common commercial packing in various references. A_s for spheres can be computed from the volume and surface area of a sphere.

As the gas velocity increases, conditions finally occur where the force of the pressure drop times the cross-sectional area just equals the weigh of the particles in the bed. A slight increase in gas velocity, to increase the pressure drop, is required to unlock the intermeshed fixed-bed particles. Once the particles disengage from each other, they begin to move. The pressure drops to the point where the upward force on the bed is balanced by the downward force due to the weight of the bed particles. Further increases in gas velocity fluidize the bed, the pressure drop rises slightly until slugging and entrainment occurs.

The point of maximum pressure drop shown in Figure 1 is the point of minimum fluidization. At this point

$$(\Delta P)(S) = W = (SL_{mf})(1 - \varepsilon_{mf})[(\rho_p - \rho_f)\, g/g_c] \qquad (4)$$

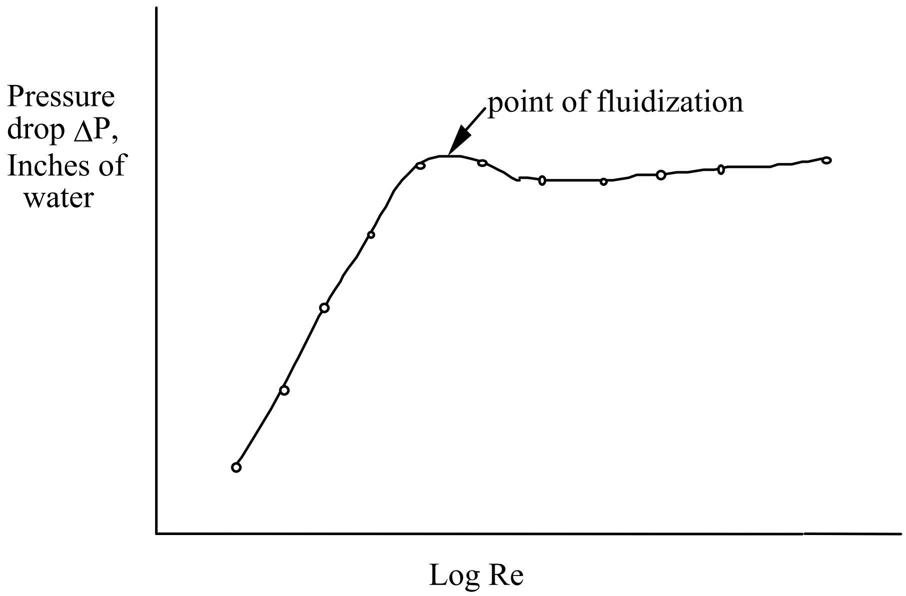
where

S = cross-sectional area of column

W = weight of bed

or

$$(\Delta P/L_{mf}) = (1 - \varepsilon_{mf})[(\rho_p - \rho_f)\, g/g_c] \qquad (4a)$$



Data can be collected evenly, however the important point of minimum fluidization should not be lost because the gas flow settings, hence Reynolds number, were not properly chosen.

Figure 1. Pressure drop in a bed of particulate solids.

Eq. (1) can be rearranged with D_p substituted by $\phi_s D_p$, v_s substituted by v_{mf}, and L substituted by L_{mf} to obtain

$$\left(\frac{\Delta P}{L_{mf}}\right) = \frac{150\mu v_{mf}}{\phi_s^2 D_P^2} \frac{(1-\varepsilon_{mf})^2}{\varepsilon_{mf}^3} + \frac{1.75\rho_f v_{mf}^2}{\phi_s D_P} \frac{(1-\varepsilon_{mf})}{\varepsilon_{mf}^3} \tag{1a}$$

The minimum fluid velocity v_{mf} at which fluidization begins can be calculated by combining Eqs. (1a) and (4a) to obtain the quadratic equation:

$$\frac{1.75}{\phi_s \varepsilon_{mf}^3}\left(\frac{D_P \rho_f v_{mf}}{\mu}\right)^2 + \frac{150(1-\varepsilon_{mf})}{\phi_s^2 D_P^3}\left(\frac{D_P \rho_f v_{mf}}{\mu}\right) = \frac{D_P^3 \rho_f (\rho_p - \rho_f) g}{\mu^2} \tag{5}$$

For many systems, Eq.(5) simplifies to:

$$v_{mf} = \frac{D_P^2(\rho_p - \rho_f)g}{1650\mu} \quad \text{for } N_{Re,mf} = \left(\frac{D_P \rho_f v_{mf}}{\mu}\right) < 20 \tag{6}$$

and

$$v_{mf} = \frac{D_P^2(\rho_p - \rho_f)g}{24.5\rho_f} \quad \text{for } N_{Re,mf} > 1000 \tag{7}$$

At high fluid velocities, when the expansion of the bed is large, the behavior of fluidization depends on whether the fluid is a liquid or a gas. With a liquid, fluidization is smooth and uniform without large bubbling. This kind of fluidization is known as “particulate” fluidization. With a gas, uniform fluidization is frequently observed only at low velocities. At high velocities, non uniform or “aggregative” fluidization will be observed with large bubbling and the bed is then often referred to as a “boiling” bed. In long, narrow fluidized beds, coalescence of the bubbles might be large enough to cover the entire cross-section of the column. These slugs of gas alternate with slugs of fluidized solids are carried upwards and subsequently collapse, causing the solids to fall back again. Slugging can cause severe entrainment problems and hence is undesirable.

Minimum Data Analysis

1. Use three packing materials. The particle size can be determined by filtering the particles through sieves. The bed porosity can be determined by: a) Weight a small sample of materials, b) Place the sample in a dry graduated cylinder and measure the dry volume of the sample, c) Empty graduated cylinder and fill it with water to a determined level, d) Place the sample in the filled graduated cylinder and note the new water level.

2. Using a log-log scale, plot the pressure drop (y-axis) versus the average superficial velocity (x-axis) for each material.

3. Using regular scale, plot the bed height versus the average superficial velocity) for each material.

4. Using data from both graphs, determine the minimum fluidization velocity. Compare this value to the value predicted by Eq. (5).

5. Using log-log scale, plot the left side of Eq.(1) versus $[(1 - \varepsilon)/N_{\mathrm{Re}}]$.

References

1. Middleman, Stanley, An Introduction to Fluid Dynamics, Wiley, (1998), pg. 411
2. Geankoplis, C. J. Transport Processes and Unit Operations, Prentice Hall, (1993), pg. 123
3. Hanesian, D. and Perna A. J., “A Laboratory Manual for Fundamentals of Engineering Design”, NJIT.