Module 9: Packed beds

Lecture 33: Minimum fluidization velocity

- Minimum fluidization velocity

Minimum fluidization velocity

The minimum fluidization velocity can be calculated by equating the pressure–drop across the fixed packed–bed, calculated from Ergun's equation to that from the expression for fluidized bed under particulate (smooth) conditions.

Let us calculate the pressure-drop from the 2nd expression:
Under fluidization conditions, pressure–drop equals effective weight of solid, as intraparticle forces disappear and solids float in the bed exhibiting 'liquid–like 'behavior. For a fluidized bed of length of L and bed-porosity of $\in$,

$$\Delta p \times \left(\frac{\pi}{4} D^2\right) = \text{Weight of solid-particles–buoyancy}$$

$$= \frac{\pi}{4} D^2 L(1-\epsilon)(\rho_p - \rho_f)g$$

Or

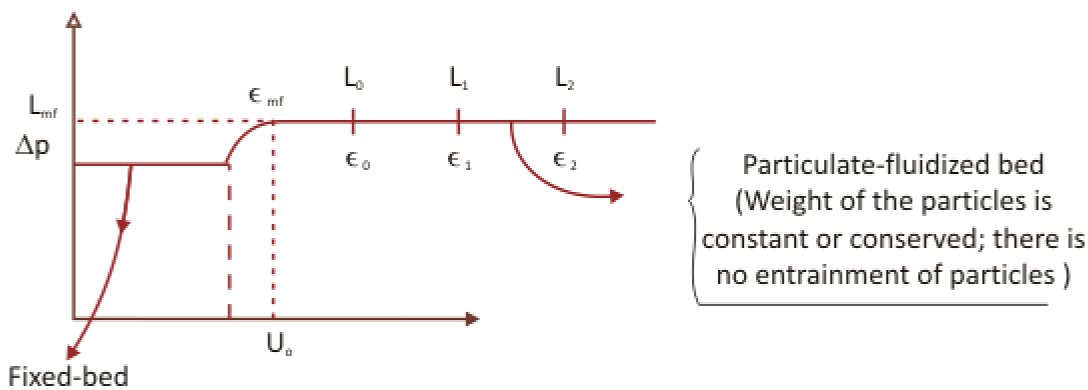
$$\boxed{\Delta p = L(1-\epsilon)(\rho_p - \rho_f)g}$$

$$= L_1(1-\epsilon_1)(\rho_p - \rho_f)g$$

$$= L_2(1-\epsilon_2)(\rho_p - \rho_f)g, \text{ etc.}$$

where $L_2 > L_1 > L$ and $\epsilon_2 > \epsilon_1 > \epsilon$

R-call:



(Fig. 33a)

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At the minimum fluidization condition:

$$\frac{\Delta p}{L_{mf}} = (1-\epsilon_{mf})(\rho_p - \rho_f)g$$

Apply Ergun's equation for 'fixed–bed' at minimum fluidization condition or at the incipience of fluidization:

$$\frac{\Delta p}{L_{mf}} = \frac{150\,\mu\,\overline{V}_{o,m}(1-\epsilon_{mf})^2}{{\phi_s}^2\,{d_P}^2\,\epsilon_{mf}^3} + \frac{1.75\rho_f \overline{V}_{o,mf}^2 (1-\epsilon_{mf})}{\phi_s\,d_P\,\epsilon_{mf}^3},$$ where $\overline{V}_{o,mf}$ ≡superficial average velocity at minimum fluidization state

Equate:

$$\frac{150\,\mu\,\overline{V}_{o,mf}\,(1-\epsilon_{mf})}{\phi_s^2\,d_p^2\,\epsilon_{mf}^3} + \frac{1.75\,V_{o,mf}^2\,\rho_f\,(1-\epsilon_{mf})\,\rho_f}{\phi_s\,d_p\epsilon_{mf}^3} = g(\rho_p-\rho_f)$$

The above-equation is quadratic on V_{mf} (minimum fluidization velocity) and may be written in the following form:

$$\boxed{\underbrace{K_1\,Re_{,Pmf}{}^2}_{\text{Inertial term}} + \underbrace{K_2\,Re_{,Pmf}}_{\text{viscous term}} = Ar}$$, where $Re_{,P} = \frac{\overline{V} d_p \rho_f}{\mu_f}$

$$Ar \equiv \text{Archimedes}\ \# = \frac{d_p^3\,\rho_f\,(\rho_p-\rho_f)g}{\mu_f^2}$$

For small particles $Re_{,p,mf} < 20$

$$\Rightarrow U_{mf} = \frac{{d_P}^2\,(\rho_P-\rho_f)g}{150\,\mu_f}\left(\frac{\epsilon_{mf}^3\,\phi_s^2}{1-\epsilon_{mf}}\right)$$

For large particles $(Re_{,Pmf} > 1000)$

$$\Rightarrow U_{mf}^2 = \frac{d_P(\rho_P-\rho_f)g}{1.75\,\rho_f}\,\epsilon_{mf}^3\,\phi_s$$

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To avoid or reduce carryover of particles form the fluidized bed, keep the gas velocity between U_{mf} *and* U_t. Recall

Terminal velocity, $u_t = \dfrac{g d_p^2 (\rho_p - \rho_f)}{18 \mu_f}$ for low Reynolds number and,

$U_t = 1.75 \dfrac{\sqrt{g d_P (\rho_P - \rho_f)}}{\rho_P}$ for high Reynolds number

With the expressions for U_{mf} and U_t known for small (viscous–flow) and large (inertial flow) particles or Reynolds number, one can take the ratio of U_t and U_{mf}:

For small $Re_{,P}$: $\dfrac{U_t}{U_{mf}} = \dfrac{150\,(1-\epsilon_{mf})}{18\,\epsilon_{mf}^2\,\phi_s^2} = 8.33 \dfrac{(1-\epsilon_{mf})}{\phi_s^2\,\epsilon_{mf}^2}$

For spherical particles, $\phi_s = 1$ and assuming $\varepsilon_{mf} = 0.45, U_t = 50\, U_{mf}$

Therefore, a bed that fluidizes at 1cm/s could preferably be operated with velocities < 50 cm/s, with few particles carried out or entrained with the exit gas.

For large Re, p: $\dfrac{U_t}{U_{mf}} = \dfrac{2.32}{\epsilon_{mf}^{1/2}}$

Or, $u_t = 7.7\, u_{mf}$ for $\varepsilon_{mf} = 0.45,$

Therefore, operating safety margin in a bed of coarse particles is smaller and there is a disadvantage for the use of coarse particles in a fluidized bed.

However, make a note that the operating particle size is also decided by the other factors such as grinding cost, pressure-drop, heat and mass-transfer aspects.

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