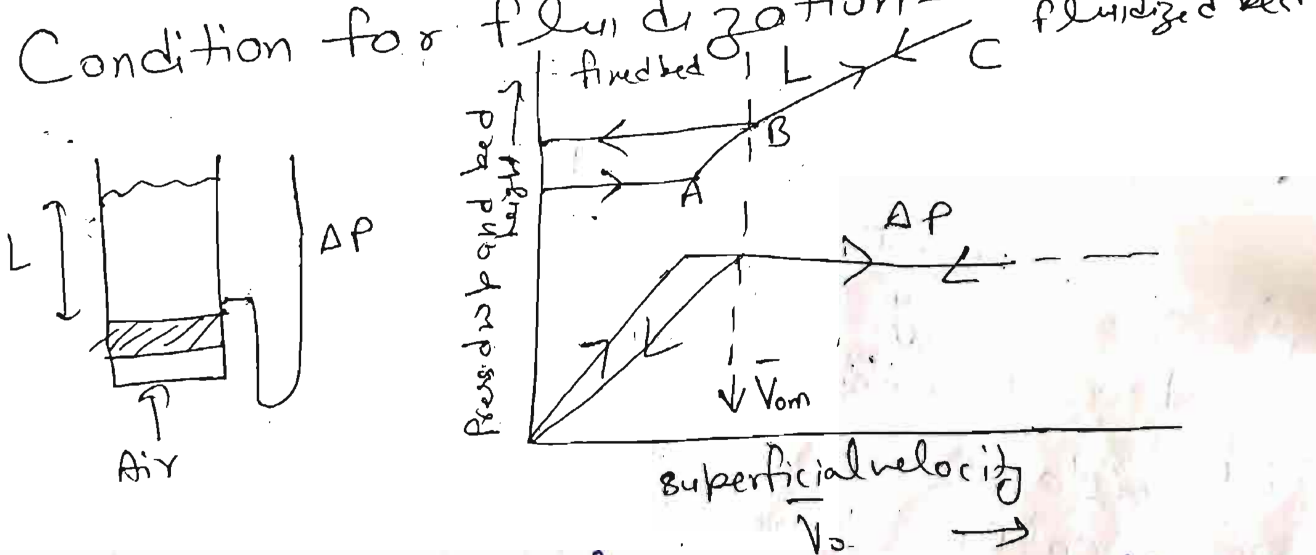
Fluidization – When a liquid or gas is passed at very low velocity up through a bed of solid particles, the particles do not move. If the fluid velocity is steadily increased, the pressure drop and the drag on individual particles increases, and eventually the particles start to move and become suspended in the fluid. The term fluidization and fluidized bed are used to describe the condition of fully ~~developed~~ suspended particles. The fluidized solids can be drained from the bed through pipes and valves just as a liquid can, and this fluidity is one of the main advantages of the use of fluidization for handling solids.

Condition for fluidization –



Consider a vertical tube partly filled with fine granular material. The tube is open at the top and has a porous plate at the bottom to support the bed of catalyst and to distribute flow uniformly over the entire cross. Air is admitted below the distributor plate at a low flow rate and passes upward through the bed without causing any particle motion. As the velocity increases the pressure drop increases, but the particles do not move and the bed height remains same. At a certain velocity, the pressure drop across the bed counterbalance the force of gravity on the particles or the weight of bed and any further increase in velocity causes the particles to move. This is point A on the graph. With a further increase in velocity ~~of~~ the particles becomes separ enough to move about in the bed, and true fluidization begins (B).

Once the bed is fluidized the ΔP across the bed stays const but bed height continues to ↑ with ↑ flow.

If the flow rate to the fluidized bed is gradually reduced, the ~~[illegible]~~ ΔP remains const and bed height ↓ following the line BC. However the final bed height may be greater than the initial value for the fixed bed due to loose packing of particles in the bed after settling from a fluidized state. The ΔP is less than from the original fixed bed. On starting again, the ΔP offsets the weight at point B and this point rather than point A, ~~~~ should be considered give minimum fluidization velocity V_{om}.

Types of fluidization-

The equations derived for min. fluidization velocity apply to liq. as well as to gases, but beyond V_{om} the apperance of beds fluidized with liquids or gases is often quite different.

Particulate fluidization- When fluidizing sand with water, the particles move farther apart and their motion becomes more vigorous as the velocity is increased, but the avg bed density at a given velocity is the same in all sections of the bed. This is called particulate fluidization and is characterized by a large but uniform expansion of the bed at high velocities.

Bubbling fluidization / Aggregative fluidizati

Beds of solids fluidized with air usually exhibit what is called aggregative or bubbling fluidization. At superficial velocities somewhat greater than V_{om} most of the gas passes through the bed as bubbles or voids which are almost free of solids, and only a small fraction of the gas

flows in the channels b/w the ba[illegible]
The particles move erratically and [illegible]
supported by the fluid, but in the [illegible]
b/w bubbles, the void fraction is ab[illegible]
the same as that at incipient fluidiz[illegible]
The nonuniform nature of the bed was
at first attributed to aggregation of the
particles and the term aggregative fluidization
was applied, but there is no evidence
that the particles stick together and the
term bubbling fluidization is a better description
of the phenomenon.

Turbulent fluidization – When the superficial
gas velocity increases to values much
above V_{om}, there is a transition from
bubbling fluidization to what is called
turbulent fluidization or fast fluidization.
The transition occurs when the bed has
expanded so much that there can no
longer be a dispersed bubble phase. The
gas phase is continuous, and there are
small regions of high or low bed density
with rapid density fluctuations at all
points in the bed. The velocity for
transition to turbulent fluidization is
in the range of 0.3 to 0.6 m/s, but it is
difficult to predict the transition velocity
because it depends on the particle
properties and the avg bubble size

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UNIT- 3 Fluidisation ①
And Motion of Solid Particle In Fluid.

Flow Through Packed Bed :-

$$\frac{(-\Delta P_b)}{L} = \frac{4 f \rho V^2}{2D} \Rightarrow f = \frac{(-\Delta P_b)\, 2D}{4 L \rho V^2} = \phi Re = \phi\left(\frac{\rho V D}{\mu}\right) \quad \text{—} ①$$

ΔP_b = bed pressure drop due to fluid friction (N/m^2)

L = bed height or bed depth (m)

f = fanning friction factor = $\phi(Re)$.

V = velocity of fluid (m/s)

D = diameter of the vessel, m

$$D_{eq} = \frac{4 S}{P_m}$$

For packed bed
$$D_{eq} = \frac{(4S)(L)}{(P_m)(L)} = \frac{\text{Total vol. of voids}}{\text{Total S.A. of particles}}.$$

$$D_{eq} = \frac{4\epsilon}{1-\epsilon} \cdot \frac{N \cdot V_P}{N \cdot A_P}.$$

ϵ = porosity i.e. fraction of the total bed vol. that is void.

N = no. of bed particles

V_P = avg. vol. of a single solid particle, m^3

A_P = " S.A. " " " " , m^2

f = Put D_{eq} in ①

$$= \frac{2(-\Delta P_b)}{4 L \rho V^2} \cdot \frac{4\epsilon}{1-\epsilon} \cdot \frac{V_P}{A_P} = \phi\left[\frac{4\epsilon}{1-\epsilon}, \frac{V_P}{A_P}, \frac{V \cdot \rho}{\mu}\right]$$

Now for spherical particles

$$\frac{V_p}{A_p} = \frac{(\pi/6)\, d_p^3}{\pi \cdot d_p^2} = \frac{d_p}{6}$$

and $V_s = \epsilon \cdot V \Rightarrow V^2 = \frac{V_s^2}{\epsilon^2}$

$$\Rightarrow f = \frac{2(-\Delta P_b)}{L \cdot \rho \cdot V_s^2} \cdot \frac{\epsilon^3}{1-\epsilon} \cdot \frac{d}{6} = \phi\left[\frac{4\epsilon}{1-\epsilon} \cdot \frac{d}{6} \cdot \right. \qquad \text{—— (2)}$$

If flow is laminar

$$f = \frac{k_1}{N_{Re}}$$

from (2)

$$\Rightarrow \frac{k_1}{N_{Re}} = \frac{2\epsilon^3}{1-\epsilon} \frac{(-\Delta P_b)}{L \cdot \rho V_s^2} \cdot \frac{d}{6}$$

$$\frac{k_1}{\frac{4\epsilon}{1-\epsilon} \cdot \frac{d}{6} \cdot \frac{V_s \rho}{\epsilon \mu}} = \frac{2\epsilon^3}{1-\epsilon} \cdot \frac{(-\Delta P_b)}{L \rho \cdot V_s^2} \frac{d}{6}$$

$$\therefore \frac{(-\Delta P_b)}{L} = \frac{k_1}{2}\left[\frac{1-\epsilon}{4} \cdot \frac{6}{d} \cdot \frac{\mu}{V_s \cdot \rho}\right]\left(\frac{1-\epsilon}{\epsilon^3}\right)\left(\rho \cdot V_s^2\right)\frac{6}{d}$$

$$= 4.5\, k_1 \frac{(1-\epsilon)^2}{\epsilon^3} \cdot \frac{\mu V_s}{d^2}$$

$$\boxed{\left(\frac{-\Delta P_b}{L}\right) = K_2 \frac{(1-\epsilon)^2}{\epsilon^3} \cdot \frac{\mu \cdot V_s}{d^2}} \qquad \text{—— (A)}$$

Carman - Kozeny Equation.

) gives the bed pressure drop per unit bed, i.e. caused to drag due to laminar flow ($N_{Re} < 1$)

the flow velocity is sufficiently large i.e. Reynolds no. high, K.E. losses become significant that solely units for bed pressure drop $(-\Delta P_b)_K$.

$$\frac{(-\Delta P_b)_K}{L} = K_3 \cdot \frac{3}{2} \frac{\rho}{d} \left(\frac{1-\epsilon}{\epsilon^3}\right) \cdot \left[\frac{V_s}{\epsilon}\right]^2$$

$$\frac{(-\Delta P_b)_K}{L} = K_4 \frac{\rho V_s^2}{d} \frac{1-\epsilon}{\epsilon^3} \quad \text{—— } B$$

B is known as Blake-Plummer eq^n

is eq^n for bed press. drop due to turbulent flow thru packed bed caused by K.E. losses.

A + B

$$\frac{(-\Delta P_b)}{L} = K_2 \frac{(1-\epsilon)^2}{\epsilon^3} \cdot \frac{\mu V_s}{d^2} + K_4 \frac{(1-\epsilon)}{\epsilon^3} \cdot \frac{\rho V_s^2}{d}$$

$$= K_2 \frac{(1-\epsilon)^2}{\epsilon^3} \cdot \frac{\mu V_{sm}}{d^2} + K_4 \frac{(1-\epsilon)}{\epsilon^3} \cdot \frac{G \cdot V_{sm}}{d}$$

V_{sm} = superficial vel. computed at avg. of inlet & outlet

$G = \rho \cdot V_s$

= mass flow rate per unit bed cross-section, $\frac{kg}{m^2 \, s}$

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~~k~~ & $k_2 = 150$, $k_4 = 1.75$

$$\boxed{\frac{(-\Delta P_b)}{L} = 150\frac{(1-\epsilon)^2}{\epsilon^3}\cdot\frac{\mu V_{sm}}{d^2} + 1.75\frac{(1-\epsilon)}{\epsilon^3}\cdot\frac{G \cdot V_{sm}}{d}}$$

This is Ergun eq^n used for Re No. b/w 1 & ...

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Minimum fluidization Velocity

An equation for the minimum fluidization velocity can be obtained by setting the pr. drop across the bed equal to the weight of the bed per unit area of cross section, & at this condition ΔP is

$$\Delta P = g(1-\epsilon)(\rho_p - \rho)L$$

At minimum fluidization velocity ϵ is given by ϵ_M. Thus,

$$\frac{\Delta P}{L} = g(1-\epsilon_M)(\rho_p - \rho)$$

Using above eqn & Ergun equation for $\Delta P/L$ at this point & $\bar{V}_0$ as $\bar{V}_{OM}$, we get

$$\frac{150\,\mu\,\bar{V}_{OM}(1-\epsilon_M)}{\phi_s^2 D_p^2\,\epsilon_M^3} + \frac{1.75\,\rho\,\bar{V}_{OM}^2}{\phi_s D_p}\,\frac{1}{\epsilon_M^3} = g(\rho_p - \rho)$$

For very small particles, only the laminar flow term of Ergun equation is significant. With $Re_p < 1$, the eqn for minimum fluidization velocity becomes

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$$\bar{V}_{OM} \simeq \frac{g(\rho_p - \rho)}{150\mu} \frac{\epsilon_M^3}{1-\epsilon_M} \phi_s^2 D_p^2$$

for roughly spherical particles, ϵ_M is generally 0.4 and 0.45, increasing slightly with decreasing particle diameter.

for higher Re_p

$$\therefore \bar{V}_{OM} \simeq \left[\frac{\phi_s D_p g(\rho_p - \rho) \epsilon_M^3}{1.75 \rho} \right]^{1/2}$$

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~~Contd.~~

Application of fluidization:- Mostly used in petroleum industry in the development of fluid-bed catalytic cracking.

- fluidization is used in synthesis of acrylonitrile & for carrying out solid-gas reactions.

Advantages of fluidization :- Because of vigorous mixing of the solids, practically no temperature gradient exists either exothermic or endothermic. It increases eat transfer rates.

Disadvantage:- uneven contacting of gas & solid. Most of the gas passes through the bed as bubbles & directly contacts only a small amount of solid. ~~In a thin shell~~

Other uses are:- Ore roasting, manufacturing cement, extraction of oil from oil shale.

Continous Fluidization:- Slurry and Pneumatic transport

When the fluid velocity through a bed of solids becomes large enough, all the particles are entrained in the fluid and are carried along with it, to give continous fluidization.

Its principal application is in transporting solids from point to point in a processing plant.

Hydraulic or slurry transport:-

Velocity also determines particle size.

Particles smaller than 50 um in diameter settles very slowly. larger particles are harder to suspend & for particles greater than 0.25 mm, a fairly large velocity is needed to keep particles moving.

The critical velocity $\bar{V}_c$ below which particles will settle down. Critical velocities are larger in big pipe than in small pipe. when the velocity is $3\bar{V}_c$

(9)

Q) A packed bed composed of uniform spherical particles of diameter 2mm & density 4150 kg/m³ is fluidized by means of a liquid of density 1000 kg/m³ & dynamic viscosity 10^{-3} Pa.s. Using Ergun's equation calculate the minimum fluidizing velocity. Take $\epsilon = 0.45$

Ergun eq^n for pr. drop through a packed bed of height L & ϵ is

$$\frac{(-\Delta P)}{L} = \frac{150\,(1-\epsilon)^2}{\epsilon^3}\cdot\frac{\mu V}{d^2} + \frac{1.75\,(1-\epsilon)}{\epsilon^3}\cdot\frac{\rho V^2}{d}$$

Also $$\frac{(-\Delta P)}{L} = (1-\epsilon)(\rho_s-\rho)g$$
$$= 16995.85 \text{ Pa/m}$$

$\therefore$ Equating above two equations

$$(1-\epsilon)(\rho_s-\rho)g = \frac{150\,(1-\epsilon)^2}{\epsilon^3}\frac{\mu V_m}{d^2} + \frac{1.75\,(1-\epsilon)\,\rho V_m^2}{\epsilon^3\, d}$$

$$\Rightarrow (\rho_s-\rho)g = \frac{150\,(1-\epsilon)}{\epsilon^3}\frac{\mu V_m}{d^2} + \frac{1.75\,\rho V_m^2}{\epsilon^3 d}$$

$$(4150-1000)\,9.81 = \frac{150(1-0.45)}{0.45^3}\times\frac{10^{-3}\,V_m}{(2\times10^{-3})^2} + \frac{1.75\;1000\times V_m^2}{0.45^3\times 2\times10^{-3}}$$

$$30901.5 = 226337.44\,V_m + 9602194.78\,V_m^2$$

$$V_m = 0.0461 \text{ m/s}$$

Q.No.2) A packed bed of solid particles of density 2500 kg/m^3, occupies a depth of 1 m in a vessel of cross-sectional area 0.04 m^2. The mass of solids in the bed is 50 kg and the surface vol. mean diameter of the particles is 1 mm. A liquid of density 800 kg/m^3 & viscosity 0.002 Pa.s flows upwards through the bed.

(a) Calculate the voidage of the bed.

(b) Calculate the pr. drop across the bed when the volume flow rate of liquid is 1.44 m^3/h.

(c) Calculate the pr. drop across the bed when it becomes fluidized.

(a) Mass of the solids in the Bed, $M = (1-\epsilon)\rho_p AL$.

$$M = \rho_p AL - \epsilon \rho_p AL$$

$$M - \rho_p AL = -\epsilon \rho_p AL$$

$$\rho_p AL - M = \epsilon \rho_p AL$$

$$\frac{\rho_p AL - M}{\rho_p AL} = \epsilon \Rightarrow \frac{2500 \times 0.04 \times 1 - 50}{2500 \times 0.04 \times 1}$$

$$\epsilon = 0.5$$

(b) Vol. flow rate = 1.44 m^3/h

superficial liq. velocity, $u = \dfrac{1.44}{0.04 \times 3600} = 0.01$ m/s.

$\dfrac{m^3}{m^2 s}$ $\quad \dfrac{m}{h} \times 3600 s$

$$u = \frac{V}{A}$$

(10)

$$\frac{\Delta P}{L} = \frac{150 (1-\epsilon)^2}{\epsilon^3} \frac{\mu u}{d^2} + \frac{1.75 (1-\epsilon)}{\epsilon^3} \frac{\rho_f u^2}{d}$$

$$\Delta P = \frac{150 (1-0.5)^2}{(0.5)^3} \times \frac{\cancel{0.0200}\ 0.002 \ (10^3) \times 0.01}{(10^{-3})^2} + \frac{1.75 (1-0.5) \times 800 \times (0.01)^2}{10^{-3} (0.5)^3}$$

$300 \times 0.02 \quad + 70$

$20 \quad 140 \quad 6000 +$

$= \cancel{6 + 70} \quad \cancel{\times 0.02}$

$\cancel{300 + 70}$

$= \cancel{70}\ \cancel{6140}$ Pa $\quad 6560$ Pa

$\cancel{6000}$

$\Delta P = L(1-\epsilon)(\rho_p - \rho_f) g$

$= 1 \times 0.5 (2500 - 800)\, 9.81$

$= 8338.5$ Pa

(Q5) A packed bed of solids of density $2000\ kg/m^3$ occupies a depth of 0.6m in a cylindrical vessel of inside diameter 0.1m. The mass of solids in the bed is 5 kg & the surface-volume mean diameter of the particles is $300\ \mu m$. $\rho_{water} = 1000\ kg/m^3$ & viscosity $= 0.001$ Pa s flows upwards through the bed.

(a) What is the voidage of packed bed?

(b) Use force balance over the bed to determine the bed pressure drop when fluidized.

(c) Assuming laminar flow and the voidage at incipient fluidization is the same as the packed bed,

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determine the minimum fluidization velocity.

for bed voidage

Now, $A = \frac{\pi}{4} d^2 = \frac{\pi}{4} \times 0.1^2$

$= 7.85 \times 10^{-3}\ m^2$

$$M = (1-\epsilon) \rho_p A L$$

$$\epsilon = 1 - \frac{5}{2000 \times 7.85 \times 10^{-3} \times 0.6}$$

$$= 0.4692$$

(b) force balance on bed

$$\Delta P = L(1-\epsilon)(\rho_p - \rho_f) g$$

$$= 0.6(1 - 0.4692)(2000 - 1000) \times 9.81$$

$$= 3124 \text{ Pa.}$$

(c) Assuming laminar flow through the bed, we apply only the laminar component of Ergun equation.

Hence $$\frac{\Delta P}{L} = 150 \frac{(1-\epsilon)^2}{\epsilon^3} \frac{\mu \cdot u}{d^2}$$

$$\frac{3124}{0.6} = \frac{150(1-0.4692)^2}{(0.4692)^3} \times \frac{0.001 \times u}{(300 \times 10^{-6})^2}$$

$$5206.66 = 409.146 \times 11111.11\, u.$$

$$u = 1.145 \times 10^{-3}\ m/s$$

for checking $Re = \frac{U_{mf} \rho_f d}{\mu(1-\epsilon)} = 0.647$

which follows laminar flow.