Solid Particulates

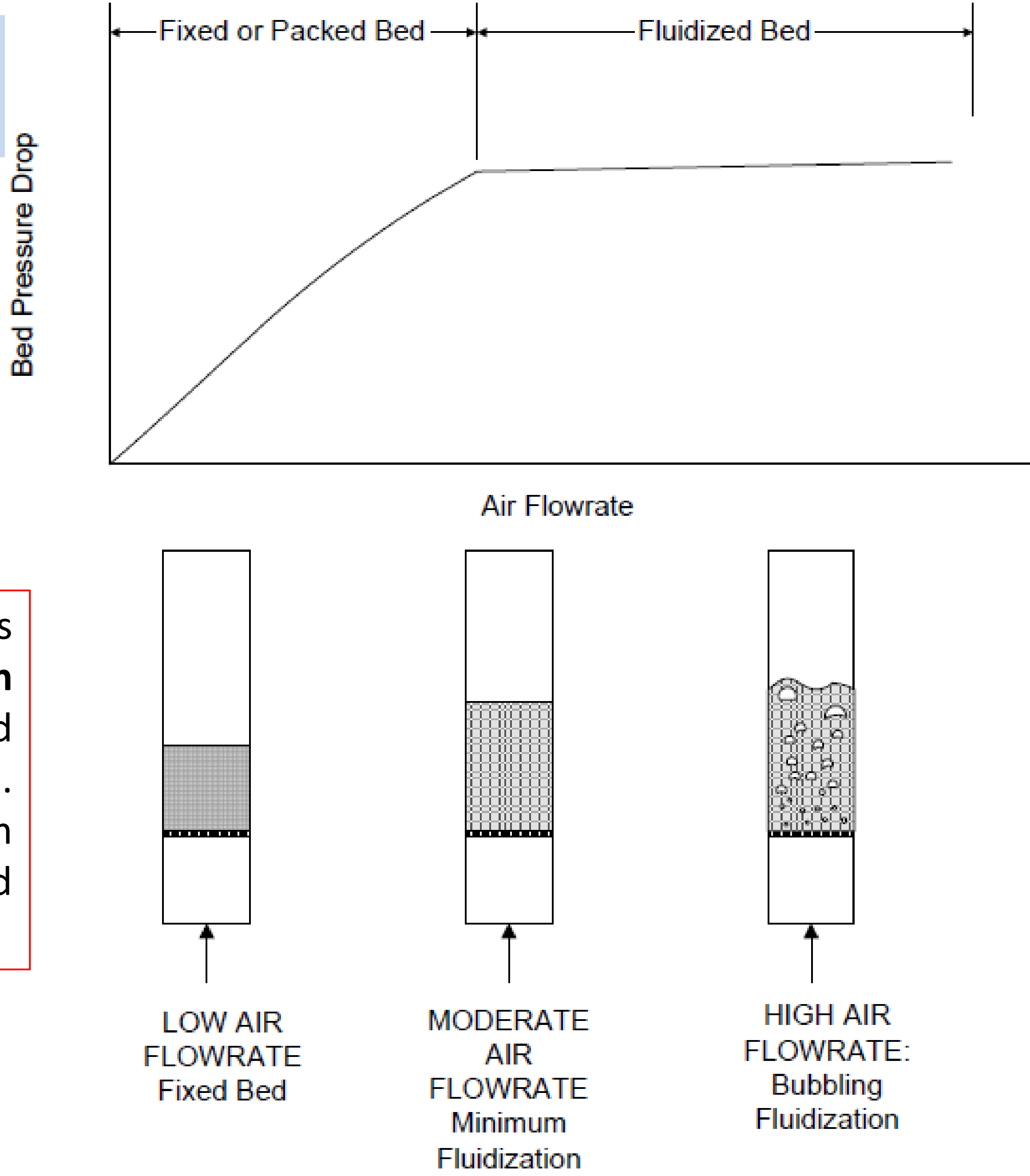
Fluidisation

Dr. Motasem Saidan

M. Saidan@gmail.com

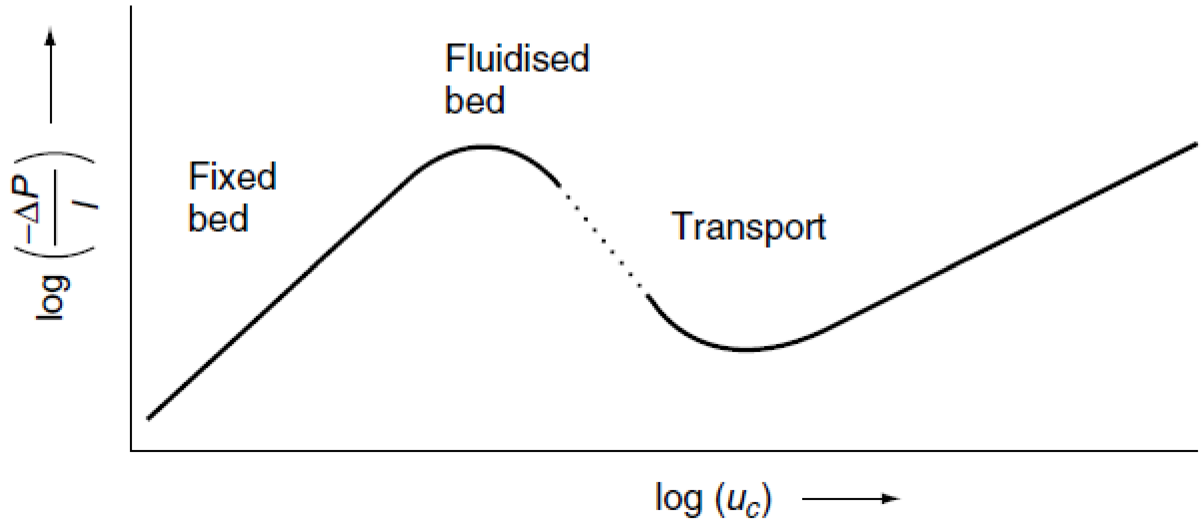
Fluidization Fundamentals

- In fluidization, a gas or liquid is passed through a bed of solid particles which is supported on a perforated or porous plate.
- In the case of fluidized bed coating, air is passed through a bed of polymer particles. When the frictional force acting on the particles, or pressure drop, of the flowing air through the bed equals or exceeds the weight of the bed, the powder particles become suspended and the bed exhibits liquid-like behavior.
- The air velocity corresponding to a pressure drop that just equals the weight of the bed is referred to as the **minimum fluidization velocity**. At this air velocity or flowrate all of the bed particles are completely suspended by the air stream. For a given system, minimum fluidization velocity can be determined from a pressure drop vs. air velocity diagram.
- As air flow is increased above the minimum fluidization velocity, the bed may exhibit behaviors ranging from smooth fluidization to bubbling fluidization to dilute fluidization in which powder can be transported by the air stream. Smooth fluidization is desirable for optimal performance in the powder coating process.
- The liquid-like nature of the fluidized powder bed allows for high heat and mass transfer rates between the gas phase and the solid phase



As shown in the figure, at gas flowrates less than the **fluidization velocity**, the bed is a fixed bed and there is no movement of particles. At flowrates above minimum fluidization the bed expands and bubbles appear.

- When a fluid flows slowly upwards through a bed of very fine particles the flow is streamline and a linear relation exists between pressure gradient and flowrate,
- As the superficial velocity approaches the minimum fluidizing velocity (u_{mf}), the bed starts to expand and when the particles are no longer in physical contact with one another the bed is fluidized,
- The pressure gradient then becomes lower because of the increased voidage and, consequently, the weight of particles per unit height of bed is smaller. This fall continues until the velocity is high enough for transport of the material to take place, and the pressure gradient then starts to increase again because the frictional drag of the fluid at the walls of the tube starts to become significant.

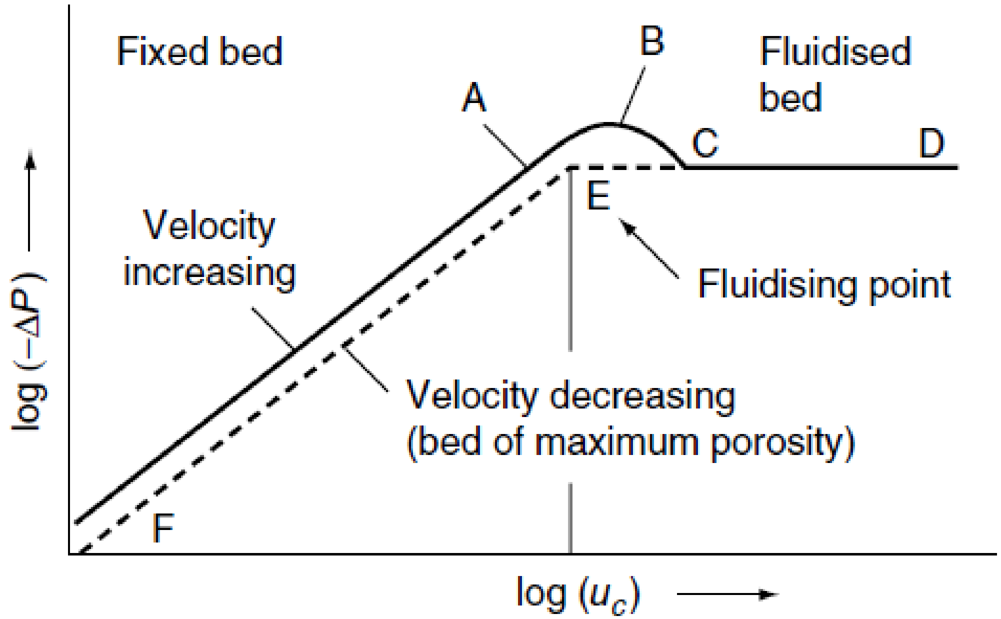


Measurement of minimum fluidizing velocity

- The minimum fluidising velocity, *umf* , may be determined experimentally by measuring the pressure drop across the bed for both increasing and decreasing velocities and plotting the results as shown in Figure.

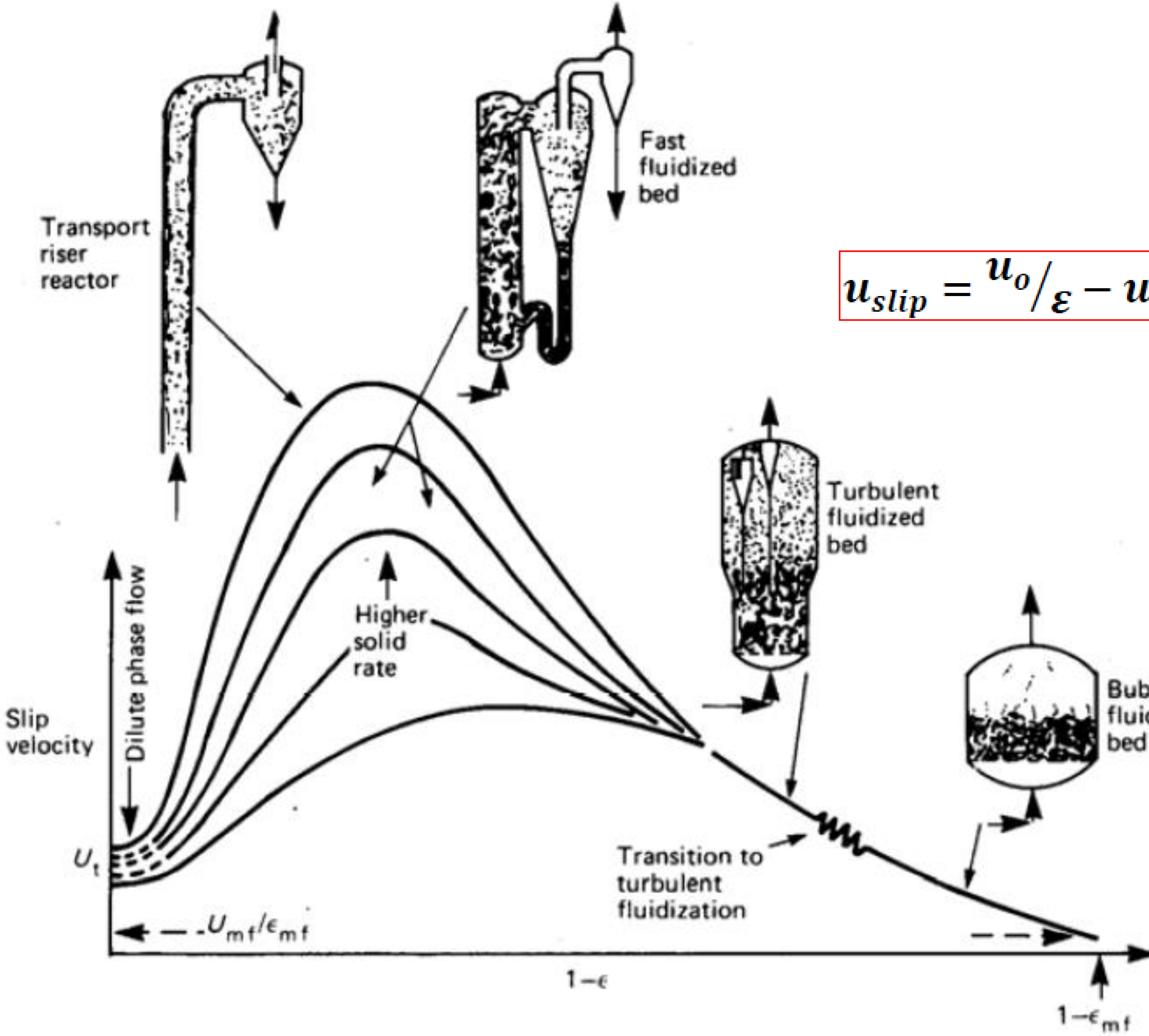
- The two 'best' straight lines are then drawn through the experimental points and the velocity at their point of intersection is taken as the minimum fluidizing velocity. Linear rather than logarithmic plots are generally used, although it is necessary to use logarithmic plots if the plot of pressure gradient against velocity in the fixed bed is not linear.

Figure 6.2., R&C page 294



Industrial Applications of Fluidized Beds

Polymeric Materials	- gas phase polymerization of polyethylene - production of silicon for the semi-conductor industry
Biochemical	- cultivation of microorganisms for the food and pharmaceutical industries
Chemical Synthesis	- phthalic anhydride - Fischer-Tropsch synthesis of hydrocarbons - acrylonitrile, maleic anhydride, activated carbon, calcination, roasting of sulfide ores, chlorination, reduction,
Petroleum Processing	- fluid catalytic cracking (FCC) for production of gasoline from oil - coal gasification - thermal cracking of naphtha petroleum fractions to produce ethylene and propylene - fluid coking
Combustion	- coal combustion - solid waste incineration - steam raising
Physical Operations	- coating metal objects - drying of solids - adsorption of solvents

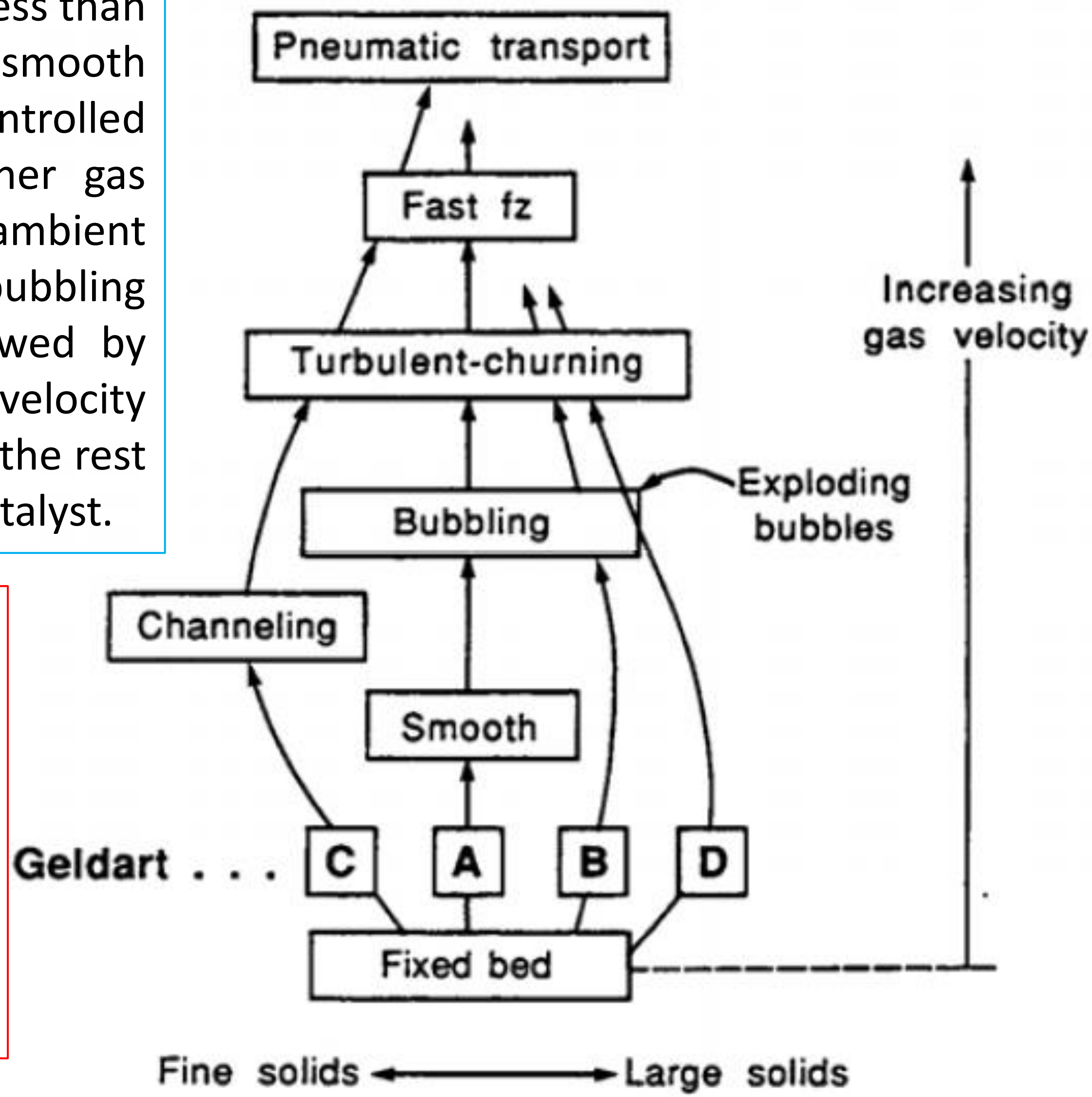


$$u_{slip} = {u_o}/{\varepsilon} - u_{particle}$$

Geldart classification of particles

Group A Small particle size or density less than 1.4 g/cm3. Easily fluidized with smooth fluidization at low gas velocities and controlled bubbling with small bubbles at higher gas velocities. When fluidized by air at ambient conditions, result in a region of non-bubbling fluidization beginning at Umf , followed by bubbling fluidization as fluidizing velocity increases. Gas bubbles rise faster than the rest of the gas. Major example is the FCC catalyst.

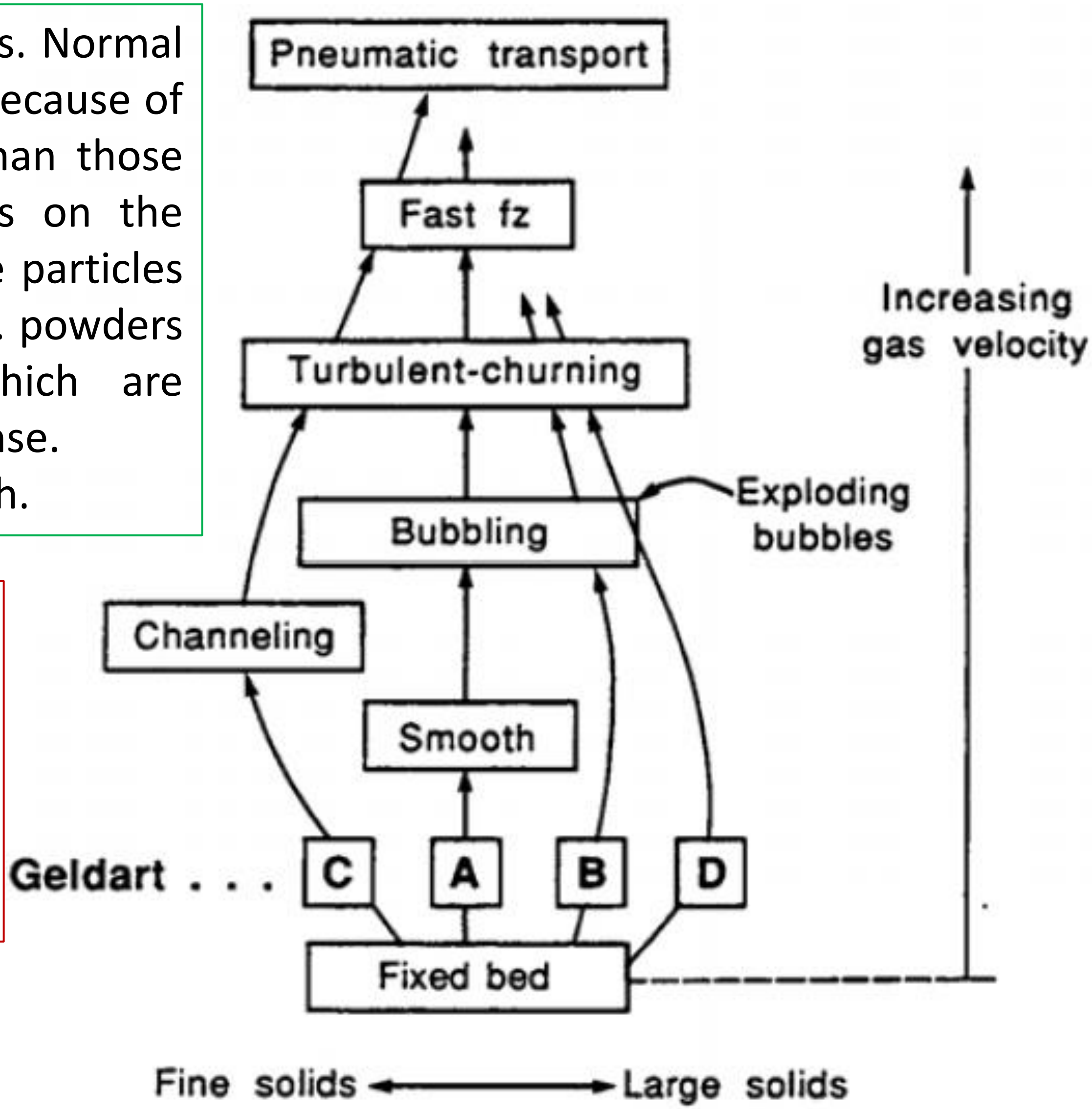
Group B are sand like powders which result in vigorous bubbling fluidization under these conditions. Bubbles form as soon as the gas velocity exceeds the minimum fluidization velocity. Majority of gas-solid reactions occur in this regime based on particle size of raw materials.



Group C cohesive, or very fine powders. Normal fluidization is difficult for these solids because of interparticle forces that are greater than those resulting from the action of the gas on the particles. In small diameter beds these particles form a plug of solids that rises upward. powders - very fine, cohesive powders which are incapable of fluidization in the strict sense.
Examples: Face powder, flour, and starch.

Group D spoutable, large and/or dense particles.

Examples include drying grains, roasting coffee beans, gasifying coals and roasting of metal ores.



Calculation of minimum fluidizing velocity

- In a fluidized bed, the total frictional force on the particles must equal the effective weight of the bed. Thus, in a bed of unit cross-sectional area, depth ***l***, and porosity ***e***, the additional pressure drop across the bed attributable to the layout weight of the particles is given by:

$$-\Delta P = (1-e)(\rho_s - \rho)lg \tag{6.1}$$

where: g is the acceleration due to gravity and

ρ_s and ρ are the densities of the particles and the fluid respectively.

If flow conditions within the bed are streamline, the relation between fluid velocity u_c, pressure drop $(-\Delta P)$ and voidage e is given, for a fixed bed of spherical particles of diameter d, by the Carman-Kozeny equation (4.12a) which takes the form:

$$u_c = 0.0055\left(\frac{e^3}{(1-e)^2}\right)\left(\frac{-\Delta P d^2}{\mu l}\right) \tag{6.2}$$

For a fluidised bed, the buoyant weight of the particles is counterbalanced by the frictional drag. Substituting for $-\Delta P$ from equation 6.1 into equation 6.2 gives:

$$u_c = 0.0055\left(\frac{e^3}{1-e}\right)\left(\frac{d^2(\rho_s - \rho)g}{\mu}\right) \tag{6.3}$$

Minimum fluidizing velocity

- As the upward velocity of flow of fluid through a packed bed of uniform spheres is increased, the point of ***<u>incipient fluidization</u>*** is reached when the particles are just supported in the fluid.
- The corresponding value of the ***minimum fluidizing velocity*** *(u_{mf})* is then obtained by substituting e_{mf} into equation 6.3 to give:

$$u_{mf} = 0.0055 \left(\frac{e_{mf}^3}{1 - e_{mf}} \right) \frac{d^2(\rho_s - \rho)g}{\mu} \tag{6.4}$$

Since equation 6.4 is based on the Carman–Kozeny equation, it applies only to conditions of laminar flow, and hence to low values of the Reynolds number for flow in the bed. In practice, this restricts its application to fine particles.

The value of e_{mf} will be a function of the shape, size distribution and surface properties of the particles. Substituting a typical value of 0.4 for e_{mf} in equation 6.4 gives:

$$(u_{mf})_{e_{mf}=0.4} = 0.00059 \left(\frac{d^2(\rho_s - \rho)g}{\mu} \right) \tag{6.5}$$

When the flow regime at the point of incipient fluidisation is outside the range over which the Carman-Kozeny equation is applicable, it is necessary to use one of the more general equations for the pressure gradient in the bed, such as the Ergun equation given in equation 4.20 as:

$$\frac{-\Delta P}{l} = 150\left(\frac{(1-e)^2}{e^3}\right)\left(\frac{\mu u_c}{d^2}\right) + 1.75\left(\frac{(1-e)}{e^3}\right)\left(\frac{\rho u_c^2}{d}\right) \tag{6.6}$$

where d is the diameter of the sphere with the same volume:surface area ratio as the particles.
Substituting $e = e_{mf}$ at the incipient fluidisation point and for $-\Delta P$ from equation 6.1, equation 6.6 is then applicable at the minimum fluidisation velocity u_{mf}, and gives:

$$(1-e_{mf})(\rho_s-\rho)g = 150\left(\frac{(1-e_{mf})^2}{e_{mf}^3}\right)\left(\frac{\mu u_{mf}}{d^2}\right) + 1.75\left(\frac{(1-e_{mf})}{e_{mf}^3}\right)\left(\frac{\rho u_{mf}^2}{d}\right) \tag{6.7}$$

Multiplying both sides by $\dfrac{\rho d^3}{\mu^2(1-e_{mf})}$ gives:

$$\frac{\rho(\rho_s-\rho)gd^3}{\mu^2} = 150\left(\frac{1-e_{mf}}{e_{mf}^3}\right)\left(\frac{u_{mf}d\rho}{\mu}\right) + \left(\frac{1.75}{e_{mf}^3}\right)\left(\frac{u_{mf}d\rho}{\mu}\right)^2 \tag{6.8}$$

In equation 6.8:

$$\frac{d^3\rho(\rho_s-\rho)g}{\mu^2}=Ga \tag{6.9}$$

where Ga is the 'Galileo number'.

and:

$$\frac{u_{mf}d\rho}{\mu}=Re'_{mf}. \tag{6.10}$$

where Re_{mf} is the Reynolds number at the minimum fluidising velocity and equation 6.8 then becomes:

$$Ga=150\left(\frac{1-e_{mf}}{e_{mf}^3}\right)Re'_{mf}+\left(\frac{1.75}{e_{mf}^3}\right)Re^{2}_{\prime mf} \tag{6.11}$$

For a typical value of $e_{mf}=0.4$:

$$Ga=1406Re'_{mf}+27.3Re'^{2}_{mf} \tag{6.12}$$

Thus:

$$Re'^{2}_{mf}+51.4Re'_{mf}-0.0366Ga=0 \tag{6.13}$$

and:

$$(Re'_{mf})_{e_{mf}=0.4}=25.7\{\surd(1+5.53\times10^{-5}Ga)-1\} \tag{6.14}$$

and, similarly for $e_{mf}=0.45$:

$$(Re'_{mf})_{e_{mf}=0.45}=23.6\{\surd(1+9.39\times10^{-5}\ Ga)-1\} \tag{6.14a}$$

$$u_{mf}=\frac{\mu}{d\rho}Re'_{mf} \tag{6.15}$$

Example 6.1

A bed consists of uniform spherical particles of diameter 3 mm and density 4200 kg/m3.

What will be the minimum fluidizing velocity in a liquid of viscosity 3 mNs/m2 and density 1100 kg/m3?

Solution

By definition:

$$\text{Galileo number, } Ga = d^3\rho(\rho_s - \rho)g/\mu^2$$
$$= ((3 \times 10^{-3})^3 \times 1100 \times (4200 - 1100) \times 9.81)/(3 \times 10^{-3})^2$$
$$= 1.003 \times 10^5$$

Assuming a value of 0.4 for e_{mf}, equation 6.14 gives:

$$Re'_{mf} = 25.7\{\sqrt{(1 + (5.53 \times 10^{-5})(1.003 \times 10^5))} - 1\} = 40$$

and: $$u_{mf} = (40 \times 3 \times 10^{-3})/(3 \times 10^{-3} \times 1100) = 0.0364 \text{ m/s or } \underline{\underline{36.4 \text{ mm/s}}}$$