Notes on Fluidized Bed Design

CHEE 331

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E.W. Grandmaison

1. Background on fluid-particle systems

2. Background on flow through packed beds

3. Fluid particle systems – fluidization
Fluid Particle Systems - Particle Settling Velocity

For theoretical consideration we can examine the behaviour of a single spherical particle of diameter \(d_p\) or radius \(r_p\) and density \(\rho_p\), freely falling at a velocity \(U_\infty\) in a fluid of density \(\rho_f\) and viscosity \(\mu_f\) (we could also consider a particle suspended in a gas stream flowing at the velocity \(U_\infty\)). We will assume that the particle concentration is dilute enough that there are no particle-particle interactions and external forces acting on the particles are negligible (such forces might include electrostatic effects, etc.).

The forces acting on the particle include:

1. Gravitational forces:
   \[
   \frac{\pi}{6} d_p^3 \rho_p g
   \]

2. Buoyancy forces - equal to the weight of displaced fluid:
   \[
   \frac{\pi}{6} d_p^3 \rho_f g
   \]

3. Viscous forces - fluid-particle forces acting at the surface of the particle, commonly called "friction" drag.

4. Pressure forces - fluid pressure forces acting over the surface and normal to the particle surface, commonly called "form" drag.

In general our force balance will have the form:

\[
(1) = (2) + (3) + (4)
\]

Forces (3) and (4) above are more complex and require special attention. We start with the solution for the velocity and pressure field of a low Reynolds number flow over a sphere. These solutions are (see for example, Bird, Stewart and Lightfoot, "Transport Phenomena", Wiley, 1960):

\[
U_r = U_\infty \left\{ 1 - \frac{3}{2} \left[ \frac{r_p}{r} \right] + \frac{1}{2} \left[ \frac{r_p}{r} \right]^3 \right\} \cos(\theta)
\]

\[
U_\theta = - U_\infty \left\{ 1 - \frac{3}{4} \left[ \frac{r_p}{r} \right] - \frac{1}{4} \left[ \frac{r_p}{r} \right]^3 \right\} \sin(\theta)
\]
In order to evaluate the "friction" drag, we need to know the shear stress distribution (spherical coordinate form),

\[
\tau_{\theta\phi} = -\mu \left\{ \frac{\partial}{\partial r} \left( \frac{U_\theta}{r} \right) + \frac{1}{r} \left( \frac{\partial U_\phi}{\partial \theta} \right) \right\}
\]
evaluated at the surface of the sphere, i.e.,

\[
\tau_{\theta\phi}\bigg|_{r=p} = \frac{3}{2} \frac{\mu U_\infty}{r_p} \sin(\theta)
\]
and integrate the z-component of this force,

\[
(-\tau_{\theta\phi}) \sin(\theta)
\]
over the surface of the sphere,

\[
F_z = \int_0^{2\pi} \int_0^\pi \left( +\tau_{\theta\phi}\bigg|_{r=p} \sin(\theta) \right) r_p^2 \sin(\theta) \, d\theta \, d\phi = 4\pi \mu U_\infty \int_0^{2\pi} r_p \, d\phi = 2\pi \mu U_\infty
\]
To evaluate the "form" drag, we integrate the z-component of the pressure distribution,

\[-P \cos(\theta)\]

where, 

\[P|_{r=r_p} = P_o - \rho_i g r_p \cos(\theta) - \frac{3}{2} \frac{\mu_i U_\infty}{r_p} \cos(\theta)\]

and 

\[F_n = \int_0^{2\pi} \int_0^\theta \left(-P|_{r=r_p} \cos(\theta)\right) r_p^2 \sin(\theta) d\theta d\phi\]

\[= \frac{4}{3} \pi r_p^3 \rho_i g + 2\pi \mu_i r_p U_\infty = \frac{1}{6} \pi d_p^3 \rho_i g + \pi \mu_i d_p U_\infty\]

Hence the total force on the particle is

\[F_{\text{Total}} = F_n + F_r\]

\[= \frac{\pi}{6} d_p^3 \rho_i g + \pi \mu_i d_p U_\infty + 2\pi \mu_i d_p U_\infty\]

\[= \frac{\pi}{6} d_p^3 \rho_i g + 3\pi \mu_i d_p U_\infty\]

The first term in this relation is a buoyancy force, present even if the particle is static, the second term arises because of the relative motion between the fluid and particle. Hence the "kinetic" force (form + friction drag) is,

\[F_k = \pi \mu_i d_p U_\infty + 2\pi \mu_i d_p U_\infty = 3\pi \mu_i d_p U_\infty\]

Our original force balance then becomes,

\[\frac{\pi}{6} d_p^3 \rho_i g = \frac{\pi}{6} d_p^3 \rho_i g + \pi \mu_i d_p U_\infty + 2\pi \mu_i d_p U_\infty\]

Or 

\[\frac{\pi}{6} d_p^3 \left(\rho_p - \rho_i\right) = 3\pi \mu_i d_p U_\infty\]

And 

\[U_\infty = \frac{d_p^2 \left(\rho_p - \rho_i\right)}{18 \mu_f}\]

This relation gives the terminal velocity for a particle under the flow conditions assumed in this analysis. The form and friction drag relations are the forces that must be imparted on the particle to provide this terminal or differential velocity between the fluid and particle.
The assumptions made in the development of the particle settling velocity above included:

1. A low Reynolds number flow, i.e.,

\[
N_{Re} = \frac{d_p \rho_f U_\infty}{\mu_f} \leq 1
\]

because the relations employed to evaluate the form and friction drag are only applicable in this regime. The equations of motion for higher Reynolds numbers are too complex for an analytical solution but we can resort to experimental observations and correlations for this regime.

2. The fluid acts as a continuum medium. This assumption is important if the particle size is small \((d_p < 0.1 \, \mu m)\) thus making the form and friction drag relations inappropriate. To handle the problem of a lack of continuum between the fluid and particle, the terminal or settling velocity is multiplied by a correction factor called the Cunningham correction factor. This problem is encountered in air pollution control systems where small particles are indeed encountered. We will not meet this problem in CHEE 331.

In order to generalize our treatment for this problem, we can introduce the concept of a drag coefficient, \(C_d\), defined by the relation,

\[
\frac{F}{A_p} = C_d \frac{\rho_f U_\infty^2}{2}
\]

where \(F\) is the total form and friction drag, \(A_p\) is the "projected" area of the particle to the flow. For a sphere,

\[
A_p = \frac{\pi d_p^2}{4}
\]

\[\therefore F = \frac{C_d \rho_f \pi d_p^2 U_\infty}{8} \]

and our force balance takes the form,

\[
\frac{\pi}{6} g d_p^3 (\rho_p - \rho_f) = \frac{C_d \rho_f \pi d_p^2 U_\infty^2}{8}
\]
The corresponding values of the drag coefficient at various values of \( N_{Re} \) are,

**Laminar flow regime** (Stokes law): \( C_d = 24 \frac{N_{Re}}{-1} \) for \( N_{Re} < 1 \)

Which leads to: \( U_\infty = \frac{d_p^2 g (\rho_p - \rho_f)}{18 \mu_f} \)

**Intermediate flow regime**: \( C_d = 18.5 \frac{N_{Re}^{-0.6}}{-} \) for \( 1 \leq N_{Re} \leq 500 \)

Which leads to: \( U_\infty = 0.153 \frac{d_p^{1.14} (\rho_p - \rho_f)^{0.71}}{\mu_f^{0.43} \rho_f^{0.29}} \)

**Turbulent flow regime** (Newton’s law): \( C_d = 0.43 \) for \( N_{Re} > 500 \)

Which leads to: \( U_\infty = 1.76 \left( \frac{g d_p (\rho_p - \rho_f)}{\rho_f} \right)^{0.5} \)

Generally, we know the particle and fluid properties and we wish to estimate \( U_\infty \). In order to do this we can assume a flow regime, solve for \( U_\infty \) and check that the Reynolds number is in the correct regime (laminar, turbulent or intermediate). Or a more direct root can be used by multiplying the Stokes law regime equation by \( \frac{d_p \rho_f}{\mu_f} \) to get:

\[
\frac{U_\infty}{\mu_f} \frac{d_p \rho_f}{\mu_f} = \frac{d_p^3 g \rho_f (\rho_p - \rho_f)}{18 \mu_f^2}
\]

The left hand side is simply the Reynolds number, which should be less than 1.0 for the assumed case of the Stokes law regime. The right hand side is also a dimensionless group involving the Galileo number* (\( N_{Ga}/18 \)),

\[
N_{Ga} = \frac{d_p^3 g \rho_f (\rho_p - \rho_f)}{\mu_f^2}
\]

\( \therefore \) \( N_{Re,T} = \frac{N_{Ga}}{18} \) and for \( N_{Re,T} < 1.0, \) we need \( N_{Ga} < 18 \)

*Note: In some textbooks this dimensionless group is called the Archimedes number*
Multiplying the Newton’s law regime equation by $d_p \rho_f/\mu_f$ we obtain,

$$
\frac{U_\infty d_p \rho_f}{\mu_f} = 1.76 \frac{d_p^{3/2} g^{1/2} \rho_f^{1/2} (\rho_p - \rho_f)^{1/2}}{\mu_f} = 1.76 N_{Ga}^{1/2}
$$

This means that for the turbulent flow regime (Newton’s law), we need,

$$
N_{Ga} > \left(\frac{500}{1.76}\right)^2 \quad \text{or} \quad N_{Ga} > 80708
$$

The intermediate flow regime must then fall between these values for $N_{Ga}$.

We can then write the flow regime criteria as:

- **Laminar flow or Stokes law** ($N_{Re} < 1.0$): $N_{Ga} < 18$
- **Intermediate regime** ($1 < N_{Re} < 500$): $18 < N_{Ga} < 80708$
- **Turbulent flow or Newton’s law** ($N_{Re} > 500$): $N_{Ga} > 80708$
**Fluid-Particle Systems - Flow Through Packed Beds**

Packed beds of particles are commonly used in the chemical process industries for gas absorption/adsorption, catalytic reactors and other contacting equipment. One of the main design objectives for packed beds is to determine the pressure drop requirements. Packed beds have a relatively low porosity and the pressure drop relationships are based on flow through tortuous channels. The pressure drop through noncircular ducts is usually based on a “hydraulic” diameter, \( d_H \), defined by

\[
d_H = \frac{4 \times \text{Flow cross sectional area}}{\text{Wetted wall perimeter}}
\]

For a circular tube this concept gives the geometrical diameter, \( d \),

\[
d_H = \frac{4 \times \left( \pi \frac{d^2}{4} \right)}{\pi d} = d
\]

and for a concentric tube (inner diameter \( d_1 \) and outer diameter \( d_2 \)),

\[
d_H = \frac{4 \times \left( \frac{\pi}{4} \left[ d_2^2 - d_1^2 \right] \right)}{\pi (d_2 + d_1)} = \frac{d_2^2 - d_1^2}{d_2 + d_1} = d_2 - d_1
\]

Packed beds also provide complex flow areas and an effective diameter in such a system can be expressed as

\[
d_H = \frac{\text{Void volume of bed}}{\text{Surface area of particles}}
\]

The void volume in a packed bed with porosity \( \varepsilon \) is

\[
V_{\text{void}} = V_{\text{Total}} - V_{\text{Solids}} = \varepsilon V_{\text{Total}} = \frac{V_{\text{Solids}}}{1 - \varepsilon} - V_{\text{Solids}}
\]

\[
V_{\text{Void}} = V_{\text{Solids}} \left\{ \frac{1}{1 - \varepsilon} - 1 \right\} = V_{\text{Solids}} \left\{ \frac{\varepsilon}{1 - \varepsilon} \right\}
\]

For \( n \) spherical particles with diameter \( d_p \),

\[
V_{\text{Void}} = \frac{n \pi d_p^3}{6} \left\{ \frac{\varepsilon}{1 - \varepsilon} \right\}
\]

and the resulting hydraulic diameter becomes,
A Reynolds number based on the hydraulic diameter is

\[ N_{Re} = \frac{\rho U d_H}{\mu} \]

The velocity through the bed material, \( U_b \), and the superficial velocity \( U_s \), (the fluid velocity upstream of the bed with no packed material present) are related by

\[ U_s = \varepsilon U_b \]

A Reynolds number through the bed is

\[ N_{Re} = \frac{\rho U_b d_H}{\mu} = \frac{\rho U_s d_H}{\varepsilon \mu} \]

and substituting for the hydraulic diameter in the bed,

\[ N_{Re,b} = \frac{1}{6} \frac{\rho U_s d_p}{\mu (1 - \varepsilon)} \]

In some derivations, the factor of 4 in the original definition of the hydraulic is not used; this leads to a coefficient of 2/3 instead of 1/6 in the above relation. Since the Reynolds number will be used in friction factor empirical relations, the leading coefficient in these expressions is normally neglected and the Reynolds number is simply expresses as

\[ N_{Re,b} = \frac{\rho U_s d_p}{\mu (1 - \varepsilon)} \]

If we define a friction factor in the general form,

\[ f = \frac{\tau_w}{\frac{1}{2} \rho U_b^2} = \frac{d_H (-\Delta P/L)}{\frac{1}{2} \rho U_b^2} \]

where \( \Delta P \) is the pressure drop through a bed of depth \( L \). Substituting the relations between \( d_H \) and \( d_p \) and \( U_b \) and \( U_s \), we obtain,

\[ f = \frac{d_p \left( \frac{\varepsilon}{1 - \varepsilon} \right) (-\Delta P/L)}{\frac{1}{2} \rho U_b^2} = \frac{(-\Delta P/L) d_p \varepsilon}{3 \rho U_b^2 (1 - \varepsilon)} \]
\[ f = \frac{(-\Delta P/L)d_p \varepsilon^3}{3 \rho U_s^2 (1 - \varepsilon)} \]

The factor of 3 is usually neglected in empirical relations for the friction factor-Reynolds number relationships, i.e.

\[ f_b = \frac{\varepsilon^3}{(1 - \varepsilon)} \frac{d_p (-\Delta P/L)}{\rho U_s^2} \]

Empirical relations for the friction factor through packed beds include:

1. **Kozeny-Carman equation** (laminar flow):
   \[ f_b = \frac{150}{N_{Re, b}}, \quad N_{Re, b} < 20 \]

2. **Burke-Plummer equation** (turbulent flow):
   \[ f_b = 1.75, \quad 10^3 < N_{Re, b} < 10^4 \]

3. **Ergun equation** (both regimes):
   \[ f_b = \frac{150}{N_{Re, b}} + 1.75, \quad 1 < N_{Re, b} < 10^4 \]
Fluid-Particle Systems - Fluidization

When a fluid passes through a bed of particles at a low flow rate, the fluid initially moves the void spaces between stationary particles. This flow regime is called a fixed bed. With an increase in flow rate, the bed passes through several regimes:

**Expanded bed:** particles move apart, a few vibrate and move about in a restricted manner.

**Incipiently fluidized bed:** the frictional force between a particle and the fluid counterbalances the weight of the particle, the vertical component of the compressive force between adjacent particles disappears, and the pressure drop through any section of the bed is approximately equal to weight of fluid and particles in that section. This is also called a bed at minimum fluidization.

**Smoothly or homogeneously fluidized bed:** occurs in liquid-solid systems above minimum fluidization and results in a smooth, progressive expansion of the bed. Gross flow instabilities are damped and remain small, and large scale bubbling or heterogeneity is not usually observed.

**Aggregative or bubbling fluidized bed:** occurs in gas-solid systems above minimum fluidization and produces large instabilities with bubbling and channelling of the gas. The bed does not expand as much as the liquid-solid system.

**Dense-phase fluidized bed:** occurs in both gas and liquid systems at high flow rates as long as there is a clearly defined upper limit or surface to the bed.

**Lean-, disperse-, or dilute-phase fluidized bed:** at a sufficiently high fluid flow rate, the terminal velocity of the solids is exceeded, the upper surface of the bed disappears, entrainment becomes appreciable, and solids are carried out of the bed with the fluid stream - this leads to pneumatic transport of the solids.

The general quality of fluidization (slugging, bubbles, etc.) can also be affected by bed geometry, gas flow rate, type of gas distributor, and vessel internals such as the presence of screens, baffles, etc.
Various kinds of contacting of a batch of solids by fluid.
Fluidized Bed Particles and the Minimum Fluidization Velocity

For nonspherical particles, the diameter may be defined as:

\[ d_p = \text{(diameter of sphere with same volume)} \]

and for irregular, nearly spherical particles, a sieve screen analysis approximates this diameter; if far from spherical, it over estimates \( d_p \).

A measure of the sphericity is

\[ \phi_s = \left( \frac{\text{surface of sphere}}{\text{surface of particle}} \right) \text{ both of same volume} \]

and \( \phi_s = 1.0 \) for a sphere and \( 0 < \phi_s < 1.0 \) for other particle shapes.

The specific surface, related to the sphericity, is defined as

\[ a' = \left( \frac{\text{surface of particle}}{\text{volume of particle}} \right) = \frac{\pi d_p^2 / \phi_s}{\pi d_p^3 / 6} = \frac{6}{\phi_s d_p} \]

or,

\[ a = \left( \frac{\text{surface of particles}}{\text{volume of bed}} \right) = \frac{6 (1 - \varepsilon)}{\phi_s d_p} \]

where \( \varepsilon \) is the void fraction in the bed of particles.

With these definitions, the Ergun equation is written in the form,

\[ \frac{\Delta P}{L} = 150 \frac{(1 - \varepsilon)^2}{\varepsilon^3} \frac{\mu U_o}{(\phi_s d_p)^2} + 1.75 \frac{1 - \varepsilon}{\varepsilon^3} \frac{\rho_f U_o^2}{\phi_s d_p} \]

The force balance on the bed material involves setting the pressure drop times the cross sectional area equal to the gravitational force exerted by the mass of the particles minus the buoyant force of the displaced fluid, i.e.

\[ \Delta P A = L_{mf} A (1 - \varepsilon_{mf}) (\rho_p - \rho_f) g \]

Or,

\[ \frac{\Delta P}{L_{mf}} = (1 - \varepsilon_{mf}) (\rho_p - \rho_f) g \]

and equating this pressure drop to the Ergun equation,

\[ (1 - \varepsilon_{mf}) (\rho_p - \rho_f) g = 150 \frac{(1 - \varepsilon)^2}{\varepsilon^3} \left( \frac{\mu U_o}{(\phi_s d_p)^2} \right) + 1.75 \frac{1 - \varepsilon}{\varepsilon^3} \left( \frac{\rho_f U_o^2}{\phi_s d_p} \right) \]
Or multiplying by

\[
\frac{d^3 \rho_f}{(1 - \epsilon_{mf}) \mu^2}
\]

we obtain,

\[
\frac{1.75}{\phi_s \epsilon_{mf}^3} \left( \frac{d_p \ U_{mf} \ \rho_f}{\mu} \right)^2 + \frac{150 (1 - \epsilon_{mf})}{\phi_s^2 \epsilon_{mf}^3} \left( \frac{d_p \ U_{mf} \ \rho_f}{\mu} \right) = \frac{d^3 \rho_f (\rho_p - \rho_f) g}{\mu^2}
\]

i.e. a quadratic relation for \( U_{mf} \) or the Reynolds number. The limit for low Reynolds numbers gives,

\[
U_{mf} = \frac{\left( \frac{\phi_s d_p}{150} \right)^2 \left( \frac{\rho_p - \rho_f}{\rho_f} \right) g \left( \frac{\epsilon_{mf}^3}{1 - \epsilon_{mf}} \right)}{\mu}, \quad \text{for } N_{Re} < 20
\]

and for high Reynolds numbers,

\[
U_{mf}^2 = \frac{\phi_s d_p}{1.75} \left( \frac{\rho_p - \rho_f}{\rho_f} \right) g \epsilon_{mf}^3, \quad \text{for } N_{Re} > 1000
\]

If \( \epsilon_{mf} \) and/or \( \phi_s \) are unknown, Wen and Yu (AIChE Journal, vol. 12, 610, 1966) have shown that for a wide variety of systems,

\[
\frac{1}{\phi_s \epsilon_{mf}^3} \approx 14 \quad \text{and} \quad \frac{1 - \epsilon_{mf}}{\phi_s^2 \epsilon_{mf}^3} \approx 11
\]

Substitution into the quadratic equation for \( U_{mf} \), we obtain,

\[
24.5 \left( \frac{d_p \ U_{mf} \ \rho_f}{\mu} \right)^2 + 1650 \left( \frac{d_p \ U_{mf} \ \rho_f}{\mu} \right) = \frac{d^3 \rho_f (\rho_p - \rho_f) g}{\mu^2}
\]

and the solution to this equation is (taking the positive roots),

\[
\frac{d_p \ U_{mf} \ \rho_f}{\mu} = -33.7 + \left[ (33.7)^2 + 0.0408 \left( \frac{d^3 \rho_f (\rho_p - \rho_f) g}{\mu^2} \right) \right]^{1/2}
\]

or,

\[
N_{Re,mf} = \sqrt{0.0408 N_{Ga} + (33.7)^2} - 33.7
\]

where \( N_{Ga} \) is the Galileo number.
The low Reynolds number limit is
\[
U_{mf} = \frac{d_p^2 (\rho_p - \rho_f) g}{1650 \mu} \quad \text{for } N_{Re, mf} < 20
\]
\[
N_{Re, mf} = \frac{N_{Ga}}{1650}
\]
and the high Reynolds number case is,
\[
U_{mf}^2 = \frac{d_p (\rho_p - \rho_f) g}{24.5 \rho_f} \quad \text{for } N_{Re, mf} > 1000
\]
\[
\left( \frac{N_{Re, mf}}{24.5} \right)^2 = \frac{N_{Ga}}{24.5}
\]
and these relations have been found to give predictions of \(U_{mf}\) within a standard deviation of \(\pm 34\%\). If data for \(\varepsilon_{mf}\) and \(\phi_s\) are known, they should be used where possible.
Pinchbeck and Popper (Chem. Eng. Sci., 6, 57, 1956) have derived an equation to estimate the ratio of the particle terminal settling velocity to the minimum fluidization velocity, $U_T/U_{mf}$. We can calculate the expected values of this ratio for low and high Reynolds number conditions,

$$U_T = \left[ \frac{4 g d_p (\rho_p - \rho_f)}{3 \rho_f C_d} \right]^{1/2}$$

For $N_{Re} < 1.0$ in the terminal velocity, $C_d = 24/N_{Re}$ and

$$U_T = \frac{g (\rho_p - \rho_f) d_p^2}{18 \mu}$$

multiplying by $d_p \rho_f/\mu$ we obtain,

$$\frac{d_p \rho_f U_T}{\mu} = \frac{g d_p^3 \rho_f (\rho_p - \rho_f)}{18 \mu^2}$$

In the laminar flow regime, $1650 \frac{d_p \rho_f U_{mf}}{\mu} = \frac{g d_p^3 \rho_f (\rho_p - \rho_f)}{\mu^2}$

and $\frac{U_T}{U_{mf}} = \frac{1650}{18} = 91.7$

For $N_{Re} > 500$, $C_d = 0.43$ and

$$U_T = \left[ \frac{4 g d_p (\rho_p - \rho_f)}{3 \rho_f (0.43)} \right]^{1/2}$$

or $U_T^2 = 3.1 \frac{g d_p (\rho_p - \rho_f)}{\rho_f}$

multiplying by $\left(\frac{\rho_f d_p}{\mu}\right)^2$, we obtain,

$$\left[ \frac{\rho_f d_p U_T}{\mu} \right]^2 = 3.1 \frac{g d_p^3 \rho_f (\rho_p - \rho_f)}{\mu^2}$$

In the turbulent flow regime, $24.5 \left[ \frac{d_p \rho_f U_{mf}}{\mu} \right]^2 = \frac{d_p^3 \rho_f (\rho_p - \rho_f) g}{\mu^2}$

\[ \therefore \left[ \frac{U_T}{U_{mf}} \right]^2 = 24.5 \times 3.1 \text{ or } \frac{U_T}{U_{mf}} = 8.72 \]
\[ C_d \text{Re}_p^2 = \frac{\pi \varepsilon d_p^3 \rho \left( \rho_s - \rho_p \right)}{3 \mu^2} \]
A pressure drop-gas velocity curve for a bed of uniformly sized sand particles is shown below:

The pressure drop increases approximately linearly with gas velocity up to $\Delta P_{\text{max}}$ at the end of the fixed bed phase and when $U_{\text{mf}}$ is reached, the pressure drop may only increase slightly with further increases in the gas velocity. This relatively constant $\Delta P$ behaviour arises because the gas-solid phase is well aerated and can deform easily without appreciable resistance. At gas velocities approaching the terminal velocity of the particles, the pressure drop decreases from the point of initiation of entrainment. When the gas velocity is decreased, there may be some hysteresis at low gas velocities (fixed bed regime) due to realignment of the particles from a random orientation.

Pressure drop behaviour for less than ideal fluidization conditions are shown below for slugging and channelling:
The choice of gas distributor has an effect on the quality of bubbling fluidization:
A range of distributor systems are shown in the diagrams below: