AGH University

FLUIDIZATION

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Chapter 1

Introduction to fluidization

1.1 Goal of this course

The aim of this course is to provide basic knowledge about fluidization and aerodynamics of gas-solid systems as well as mathematical tools that enable to simulate basic fluidized systems.

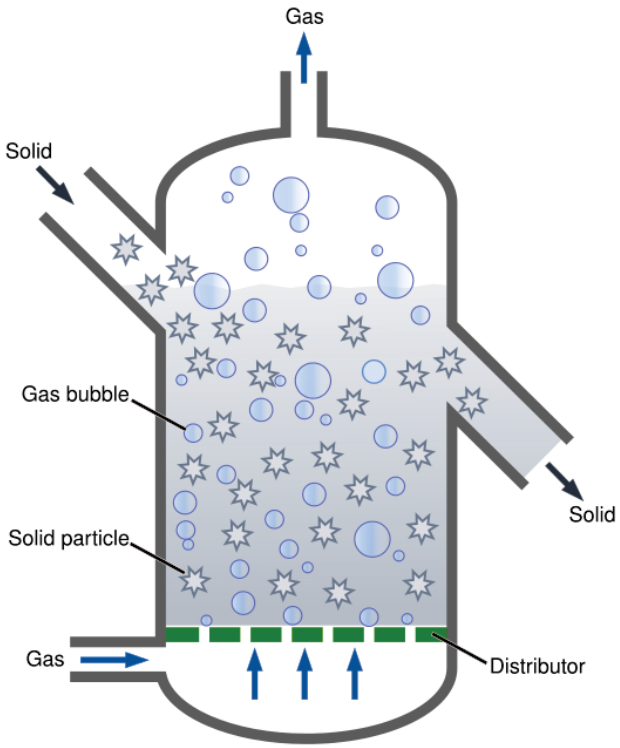
1.2 The phenomenon of fluidization

To simply decribe the idea behind fludization process one can say that it is the operation that can change a system a solid particles into fluidlike suspension in gas or liquid. This method of contacting of this two phase mixture have some unusual characteristics that are widely used in many fields of chemical industry. Simplified diagram showing the idea of fluidization is presented in Fig. 1.1. Gas is delivered from the bottom of the reactor, goes through a gas distributor to provide inform distribution through whole profile of bed and flows through packed bed of solids.

At low gas velocities the drag force is to small to lift the bed, which remains fixed. Increasing gas velocity causes solids to move upward and create fluid bed. Depending on the velocity of gas we can distinguish different modes of fluidization (Fig. 1.2) from bubbling fluidization, through turbulent and fast fluidization modes up to pneumatic transport of solids.

Another important issue concerning the fluidization proces is pressure drop through a fixed bed. Fig. 1.3 presents changes in pressure drop with changing gas velocity. At first one can observe increasing pressure drop, up to some level where it becomes constant, despite increasing gas velocity. This change in pressure drop trend can be connected

FIGURE 1.1: Schematic diagram of fluidization



with the creation of dense phase of fluidized bed and that is the moment when the fluidization occurs. The velocity at which the pressure is stabilized is called minimum fludization velocity. Pressure drop is stable in a certain range of velocities, then a slight increase can be observed which precede a drastic decrease in pressure drop. This is due to the entrainment of smaller particles which are suspended in the section over the dense fluidized bed. Further increase in gas velocity will cause more fractions to be carried over which leads to disappearance of dense phase and start of pneumatic transport. Although, as it will be shown later in some cases gas velocities exceeding the terminal velocity can be applied for the so called fast fluidization.

FIGURE 1.2: Fluidization type depending on gas velocity

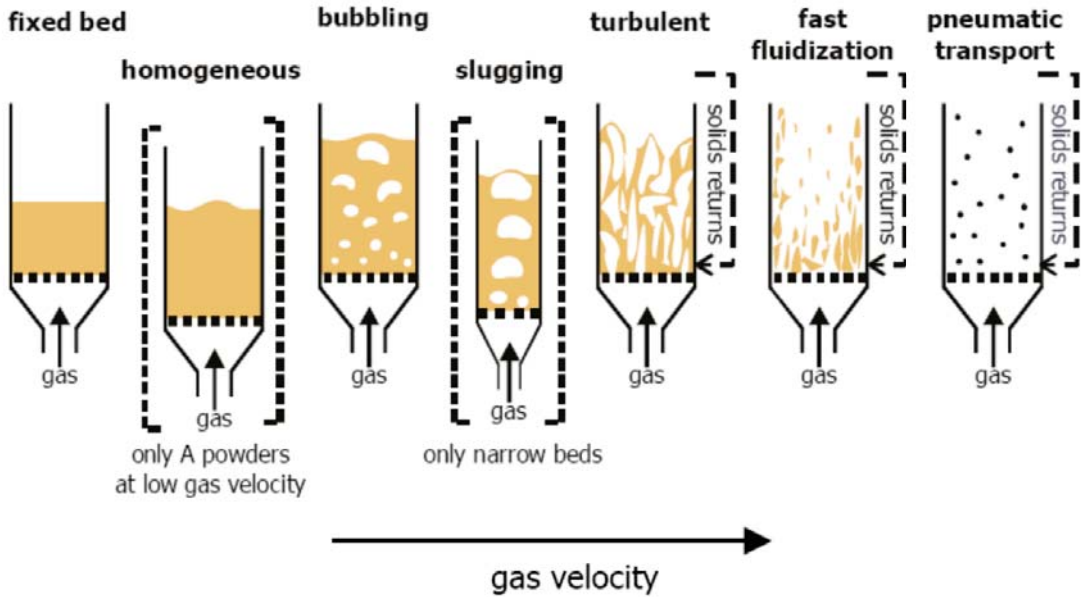
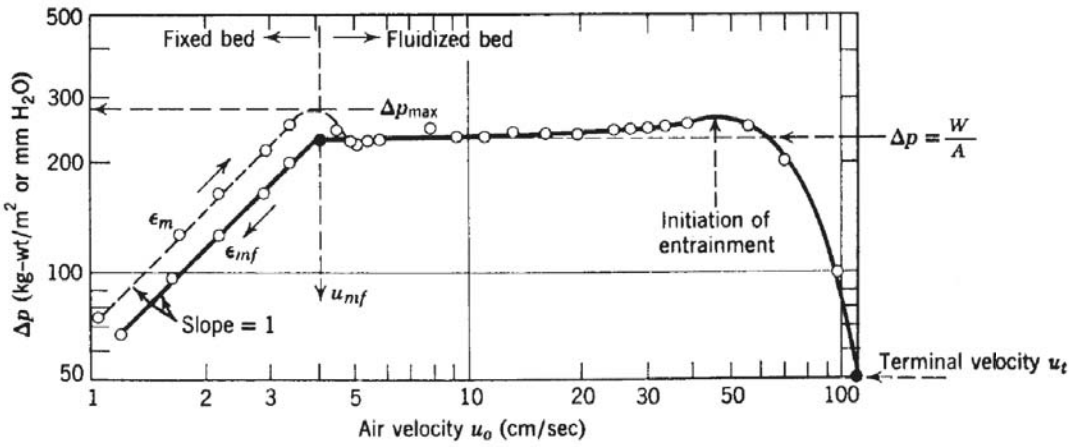


FIGURE 1.3: Pressure drop vs. gas velocity.



1.3 Discussion points

1. Definition of fluidization phenomenon.
2. Various types of beds with gas flowing through a bed of fine particles.
3. Circulating fluid bed characteristics.
4. Liquidlike features of fluid beds
5. Advantages and disadvantages of fluidized beds.
6. Heat transport phenomenon in fluidized bed.

7. Drying process in fluid bed - basic process configuration.
8. Catalytic cracking in fluid bed.
9. Fluid bed boilers for energy generation.
10. Fluid bed pyrolysis and gasification.

Chapter 2

Fluidization velocities

2.1 Characterization of particles

Usually the bed contains particles with a wide range of size and shapes, which causes the necessity to provide a proper and uniform description of size of material forming a bed. If the particles are spherical the bed can be described by means of their diameter distribution, but in real application most particles are nonspherical which yields a question about the way to decribe this kind of beds. There exists a wide range of nonsphericity measures [**?**]. However, the most wildly used is the one called sphericity (ϕ_s) defined as the ratio of the surface of sphere to the surface of particle with the same volume. For spherical particles $\phi_s = 1$ and for other shapes $0 \leq \phi_s \leq 1$. Sphericity values for some popular particles can be found in [**?**]. Other important parameter describing nonspherical particles is their specific surface, given as the ratio between surface and volume of the particle:

$$a^{'} = \frac{6}{\phi_s d_{sph}} \tag{2.1}$$

where d_{sph} is a diameter of the sphere having the same volume as the considered particle. The same concept can be applied to the whole bed of particles:

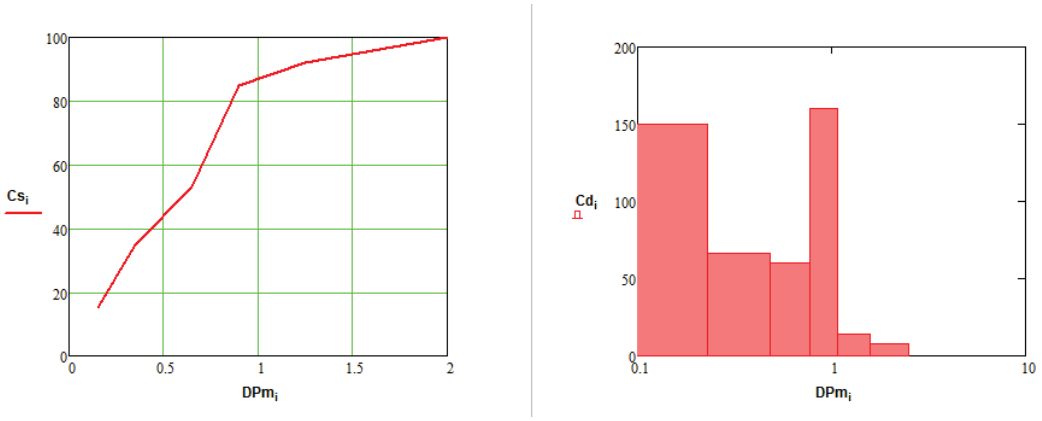
$$a = \frac{6(1-\epsilon_m)}{\phi_s d_{sph}} \tag{2.2}$$

where ϵ_m is the fractional voidage, which usually can be found experimentally for each specific system.

2.2 Particle size distribution

Variety in shapes of particles is not an only problem in describing the bed, because in most cases one also has a bed of particles with different sizes. For this description we can define two functions of size distribution **p** and **P**. Assuming that we have a bed of solids with diameters dp_i, for $i \in (1, 2, \ldots, N)$ then **p** gives the fraction (mass, volume, number) of particles that are of the diameter $d \in (dp_1, dp_{i+1})$. The function **P** gives the so called cumulative distribution, which means the fraction of solids that are smaller than the given value dp. Examples of such distributions are shown in Fig.2.1

FIGURE 2.1: Difference between **P** (left) and **p** (right)



Next issue is to provide an average size than can best describe properties of the system and can be used in further calculations. This is done by harmonic diameter:

$$\overline{dp} = \frac{1}{\sum_{i=1}^{N} \frac{x_i}{dp_i}} \tag{2.3}$$

where x_i is a fraction of solids with diameter $(dp_i, dp_{i+1}$ and $dp_i = \frac{dp_i + dp_{i+1}}{2}$. Then mean specific surface can be obtained using equation (2.1):

$$\overline{a'} = \frac{6}{\phi_s \overline{dp}} \tag{2.4}$$

2.3 Fluidization velocities

First step in the process of description of fluidized bed is to calculate the velocity of gas needed in the system. We can distinguish two basic velocities describing fluidization: minimum and terminal. In this section we present the procedure used to calculate both

of them. All calculations can be performed for a bed with single or multi size particles (in case of multi size bed it is necessary to calculate its mean diameter).

2.3.1 Pressure drop

Pressure drop through fixed bed of solids of uniform size (dp) of the length L is given by Ergun [XX] correlation:

$$\frac{\Delta P}{L} = 150\frac{(1-\epsilon_m)^2}{\epsilon_m^3}\frac{\mu u_0}{(\phi_s d_p)^2} + 1.75\frac{1-\epsilon_m}{\epsilon_m^3}\frac{\rho_g u_o^2}{\phi_s dp} \tag{2.5}$$

where μ is gas viscosity, dp is solid diameter, ρ_g is gas density, u_0 is superficial gas velocity.

2.3.2 Minimum fluidizing velocity

At the beginning of this section one has to revise definitions of two dimensionless numbers: Reynolds(2.6) and Archimedes (2.7)

$$Re = \frac{d_p u_{mf} \rho_g}{\mu} \tag{2.6}$$

$$Ar = \frac{d_p^3 \rho_g (\rho_s - \rho_g) g}{\mu^2} \tag{2.7}$$

Now, remembering that the phenomenon of fludization occurs when drag force created by the upward flow of gas is a least equal to the weight of particles in the bed. Mathematically it can be presented with the following equation:

$$\Delta P_{bed} A_t = A_t L_{mf}(1-\epsilon_{mf})[(\rho_s - \rho_g)g] \tag{2.8}$$

Rearranging and combining with equation (2.5) gives a quadratic in u_mf which can be presented in dimensionless form of the equation (2.9) (for the details see "Problem solving", ex. 3)

$$\frac{1.75}{\epsilon_{mf}^3 \phi_s} Re_{p,mf}^2 + \frac{150(1-\epsilon_{mf})}{\epsilon_{mf}^3 \phi_s^2} Re_{p,mf} = Ar \tag{2.9}$$

Solving equation (2.9) can be laborious but gives reliable estimation of u_{mf} if sphericity and voidage are known. For rough estimation without knowledge of voidage and sphericity of the system some simplifications can be used. For fine particles expression (2.10) proposed by Wen and Yu [xx] can be used to obtain reynolds number in minimum

fludization conditions:

$$Re_{p,mf} = (33.7^2 + 0.0494Ar)^{1/2} - 33.7 \tag{2.10}$$

2.3.3 Terminal fluidization velocity

Terminal fluidization velocity can be calculated from the equation given in the dimentionless form (2.11)

$$C_D Re_t^2 = \frac{4}{3} Ar \tag{2.11}$$

where C_D is drag coefficient, which can be obtained experimentally or calculated from one of many empirical relationships, C_D is a function of Reynolds number. One of the correlation that enables to calculate drag coefficient for a wide range of Reynolds numbers ($10^{-1} \div 10^6$) was proposed by Kaskas [xx] (2.12

$$C_D(Re) = \frac{24}{Re} + \frac{4}{\sqrt{Re}} + 0.4 \tag{2.12}$$

Substituting (2.12) to (2.11) we obtain equation (2.13). Solving numerically for Re_t we can obtain terminal velocity for the system.

$$Re^2(\frac{24}{Re} + \frac{4}{\sqrt{Re}} + 0.4) = \frac{4}{3} Ar \tag{2.13}$$

To avoid numerical calculations of equation (2.13) some simplification can be used, depending on the type of low in the reactor. For laminar flow we obtain the only analytical solution to the equation (2.13)

$$C_D = \frac{24}{Re} \text{ for } Re < 0.4 \tag{2.14}$$

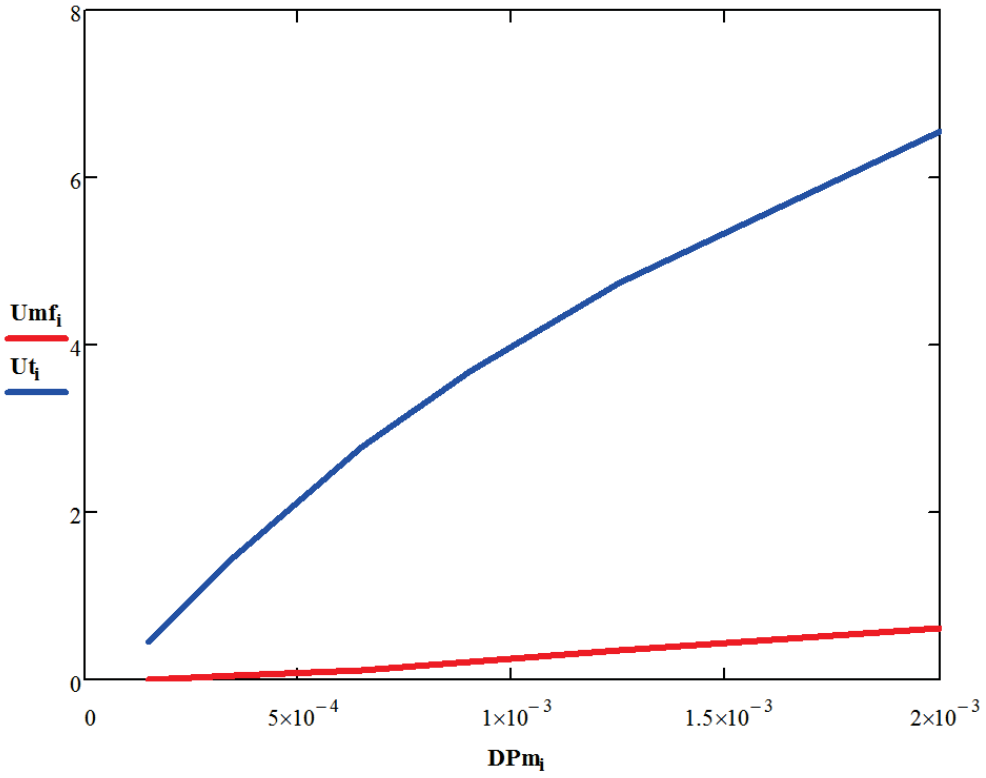
For larger Reynolds numbers we can use one of the following approximations:

$$Cd = \frac{10}{\sqrt{Re}} \text{ for } 0.4 < Re < 500 \tag{2.15}$$

$$Cd = 0.43 \text{ for } 500 < Re < 2 * 10^5 \tag{2.16}$$

An example of minimum fluidization velocity and terminal velocity as a function of particle diameter is presented in Fig. 2.2

FIGURE 2.2: Minimum (red) and terminal (blue) velocity as a function of particle diameter.



2.4 Gas distributor design

The design of fluid bed gas distributors may have a marked influence on the performance of a fluid bed reactor. The primary physical reason for this influence is that the distributor design influences the hydrodynamics and thus the gas/solid contacting pattern in the fluidized bed. Particle and gas properties play a key role in successful design together with the critical pressure drop ratio, and hole size, geometry and spacing; these strongly influence jet penetration, dead zones, particle sifting, attrition and mixing. [Geldart, 1985; Bauer, 1981].

This section deals with the simple algorithm that enables to design a perforated plate distributor using just an orifice theory.

1. Calculate pressure drop across the distributor (**??**):

$$\Delta p_d = (0.2 \div 0.4)\Delta p_b \tag{2.17}$$

where Δp_b can be calculated form (2.5) or (2.8).

2. Calculate the vessel Reynolds number($Re_t = \frac{d_t u_0 \rho_g}{\mu}$, where d_t is tube(reactor) diameter) and select the corresponding drag coefficient from the table below

Re_t	100	300	500	1000	2000	> 3000
$C_{D,or}$	0.68	0.70	0.68	0.64	0.61	0.60

3. Calculate gas velocity through the orifice

$$u_{or} = C_{D,or} * (\frac{2\Delta p_d}{\rho_g})^{1/2} \tag{2.18}$$

Check the ratio $\frac{u_0}{u_{or}}$ which gives the fraction of open area in the distributor and should be less than 10%.

4. Assume orifice diameter (d_{or} and calculate the number of orifices per unit area of distributor using equation (2.19)

$$u_o = \frac{\pi}{4} d_{or}^2 u_{or} N_{or}. \tag{2.19}$$

2.5 Discussion and problem solving

1. Dimensionless numbers - define Archimedes and Reynolds number.
2. Describe the procedure of calculating minimum and terminal velocity for a poly-dispersed system.
3. Prove, that starting from combined equations (2.5) and (2.8) and using following assumptions one can obtain equation (2.10).

$$\frac{1}{\phi_s \epsilon_{mf}^3} = 14 \quad \frac{1 - \epsilon_{mf}}{\phi_s^2 \epsilon_{mf}^3} = 11$$

4. Prove that in steady state condition equation

$$U_s \frac{dU_s}{dz} = \frac{3}{4} C_D \frac{\rho_s}{\rho_s d_p} (U_g - U_s)^2 - g \frac{\rho_s - \rho_g}{\rho_s} isidenticalto(2.11). \tag{2.20}$$

5. Calculate mean diameter for the system of particles presented below

(a)

d_i, mm	$0.1 - 0.2$	$0.2 - 0.5$	$0.5 - 0.8$	$0.8 - 1.0$	$1.0 - 1.5$	$1.5 - 2.5$
%	15	20	18	32	7	8

(b)

d_i, mm	$0.2 - 0.4$	$0.4 - 0.6$	$0.6 - 1.0$	$1.0 - 1.5$	$1.5 - 2.0$	$2.0 - 3.0$
%	32	20	18	15	7	8

6. Calculate minimum fluidization velocity for presented system. Perform the calculation on mean diameter. Check if applying calculated velocity wile cause any fraction to be carried over?

7. Design gas distributor for the given system.

8. Describe Geldart's powder classification.