Momentum Transfer in Fluidized Bed

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***Abstract*— This paper includes the experimental work involved in fluidization- the most efficient method for contacting fluid and particles. Mustard seed and sago are used one after the other and their transformed behavior from fine solids into a fluid like state are observed through contact with air or methyl alcohol. Calculation of minimum fluidization velocity is accompanied by Graphs for both test materials between (a) bed pressure drop and velocity and (b) Reynolds number and Bed expansion ratio.**

***Index Terms*—Bed expansion ratio, Bed pressure drop, fluidization, minimum fluidization velocity, Reynolds number, sphericity, voidage**

I. Introduction

"The arrival time of a space probe traveling to Saturn can be predicted more accurately than the behavior of a fluidized bed chemical reactor." Even though the above quotation (Geldart, 1986) is almost 20 years old it remains true in the new millennium of fluidization engineering. The difficulties in prediction stem in part from the complexity and ambiguity in defining the fundamental parameters such as size, shape and density of the particles. These parameters play an important role in the calculation and prediction of dynamic behavior in fluidized beds. Most physical properties of the particles are estimated indirectly, such as estimating particle shape by the bed voidage. All factors are explicitly and implicitly significant in the estimation of the behavior of fluidization operations. Although new technology is helping us to understand and give more precise prediction in fluidization, more research is still needed.

II. Review of Fluidization Basics

Fluidization is a process in which solids are caused to behave like a fluid by blowing gas or liquid upwards through the solid-filled reactor. Fluidization is widely used in commercial operations; the applications can be roughly divided into two categories, i.e.,

• Physical operations, such as transportation, heating, absorption, mixing of fine powder, etc. and

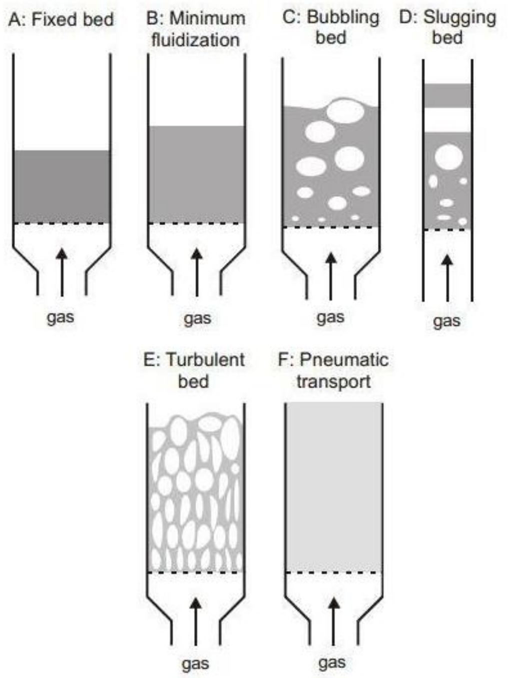
• Chemical operations, such as reactions of gases on solid catalysts and reactions of solids with gases etc.

The fluidized bed is one of the best known contacting methods used in the processing industry, for instance in oil refinery plants.

III. Fluidization Regimes

When the solid particles are fluidized, the fluidized bed behaves differently as velocity, gas and solid properties are varied. It has become evident that there are number of regimes of fluidization, as shown in Fig 1. When the flow of a gas passed through a bed of particles is increased continually, a few vibrate, but still within the same height as the bed at rest. This is called a fixed bed. With increasing gas velocity, a point is reached where the drag force imparted by the upward moving gas equals the weight of the particles, and the voidage of the bed increases slightly. This is the onset of fluidization and is called minimum fluidization with a corresponding minimum fluidization velocity, U_{mf}.

Fig 1 Schematic representation of fluidized beds in different regimes (based on Kunii and Levenspiel, 1991)



Now, with increase in the gas flow further, formation of fluidization bubbles sets in. At this point, a bubbling fluidized bed occurs. As the velocity is increased further still, the bubbles in a bubbling fluidized bed will coalesce and grow as they rise. If the ratio of the height to the diameter of the bed is high enough, the size of bubbles may become almost the same as diameter of the bed. This is called slugging. If the particles are fluidized at a high enough gas flow rate, the velocity exceeds the terminal velocity of the particles.

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The upper surface of the bed disappears and, instead of bubbles, one observes a turbulent motion of solid clusters and voids of gas of various sizes and shapes. Beds under these conditions are called turbulent beds. With further increases of gas velocity, eventually the fluidized bed becomes an entrained bed in which we have disperse, dilute or lean phase fluidized bed, which amounts to pneumatic transport of solids.

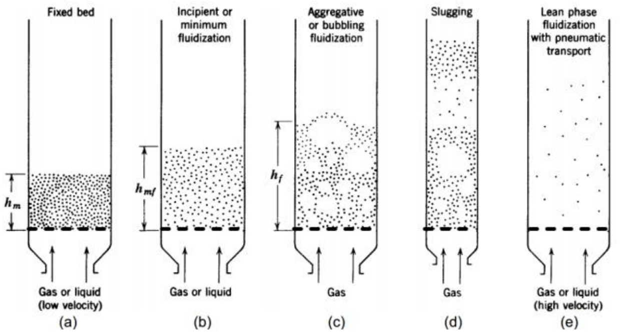
IV. DESCRIPTION OF THE PHENOMENA

We consider a vertical bed of solid particles supported by a porous or perforated distributor plate, as in figure. The direction of gas flow is upward through this bed. There is a drag exerted on the solid particles by the flowing gas, and at low gas velocities the pressure drop resulting from this drag will follow the Ergun equation. When the gas velocity is increased to a certain value however, the total drag on the particles will equal the weight of the bed, and the particles will begin to lift and barely fluidize. If ρ_c is density of the solid catalyst particles, A_c is the cross sectional area, h_s, is the height of the bed settled before the particles start to lift, h, is the height of the bed at any time, and ε_s and ε are the corresponding porosities, of the settled and expanded bed, respectively; then the mass of solids in the bed, W_s, is

$$W_s = \rho_c A_c h_s (1-\varepsilon_s) = \rho_c A_c h (1-\varepsilon)$$

This relationship is a consequence of the fact that the mass of the bed occupied solely by the solid particles is the same no matter what the porosity of the bed. When the drag force exceeds the gravitational force, the particles begin to lift, and the bed expands (i.e., the height increases) thus increasing the bed porosity. This increase in bed porosity decreases the overall drag until it is again balanced by the total gravitational force exerted on the solid particles. If the gas velocity is increased still further, expansion of the bed will continue to occur; the solid particles will become somewhat separated from each other and begin to jostle each other and move around in a restless manner. Increasing the velocity just slight amount further causes instabilities, and some of the gas starts bypassing the rest of the bed in the form of bubbles. These bubbles grow in size as they rise up the column.

Fig 2. Various kinds of contacting of a batch of solids by fluid



Coincidentally with this, the solids in the bed begin moving upward, downward, and around in a highly agitated fashion appearing as a boiling frothing mixture. With part of the gas bubbling through the bed and the solids being moved around as though they were part of the fluid, the bed of particles is said to be ―fluidized. It is in a state of aggregative, non-particulate, or bubbling fluidization. A further increase in gas velocity will result in slug flow and unstable chaotic operation of the bed. Finally at extremely high velocities, the particles are blown or transported out of the bed.

V. MINIMUM FLUIDIZATION VELOCITY

The superficial gas velocity, at which the bed of powder is just fluidized, is normally called the minimum fluidization velocity or designated by U_{mf}. This state of incipient fluidization can be described by an equation giving the pressure drop in a gas flowing through a packed bed, such as the so-called Ergun equation:

$$\frac{\Delta P}{L} = 150\,\frac{(1-\varepsilon_{mf})^2}{\varepsilon_{mf}^3}\,\frac{\mu U_{mf}}{(\phi_s d_p)^2} + 1.75\,\frac{(1-\varepsilon_{mf})}{\varepsilon_{mf}^3}\,\frac{\rho_g U_{mf}^2}{\phi_s d_p}$$

in which ΔP is equal to the bed weight per unit cross-sectional area, and the particle sphericity, ϕ_s, is defined as the surface area of a volume equivalent sphere divided by the particle‘s surface area.

When applying the Ergun equation, one has to know the minimum fluidization voidage, ε_{mf}, although it is frequently an unknown. Wen and Yu (1966) developed an expression for the minimum fluidization velocity for a range of particle types and sizes by assuming the following approximations to hold based on experimental data:

$$\frac{1-\varepsilon_{mf}}{\phi^2 \varepsilon_{mf}^3} \cong 11 \; and \; \frac{1}{\phi \varepsilon_{mf}^3} \cong 14$$

They combined these with the Ergun equation and obtained the relation:

$$Re_{mf} = \frac{d_p U_{mf} \rho_g}{\mu} = \sqrt{33.7^2 0.0408\,\frac{d_p^3 \rho_g (\rho_s - \rho_g) g}{\mu^2}} - 33.7$$

Leva (1959) obtained empirically another widely used expression:

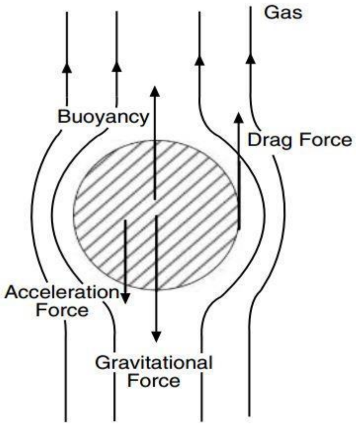
$$U_{mf} = 7.90x10^{-3} D_p^{1.82} (\rho_s - \rho_f)^{0.94} \mu_f^{-0.88}$$

This equation is valid for $Re_{mf} <= 10$, whereas for higher values of Re_{mf} a correction factor must be applied.

VI. TERMINAL VELOCITY OF A PARTICLE

In the regimes discussed so far, solids are generally retained within a certain height above the grid. Except for some entrainment, there is no large-scale migration of particles with the gas, so, these regimes (fixed, bubbling and turbulent) are in the captive stage. Subsequent discussions will concentrate on transport stage which involves large-scale migration of particles out of the vessel.

Fig 3. Force balance on a particle moving in an upward gas stream.



Consider a particle falling freely from rest and accelerating under gravity in an infinite and stationary medium. The buoyancy force and the fluid drag oppose the effect of gravity. The particle accelerates until it reaches an equilibrium velocity called the terminal velocity.

When a fluid flows over a stationary particle or travels at a velocity higher than that of an upward-moving particle, the particle experiences an upward fluid drag. It is also subjected to a buoyancy force, and downward gravitational force, as shown in figure. The drag force on the particle is related to the kinetic energy of the fluid and the projected area of the particle,

Drag Force is given by:-

$$F_D = C_D \frac{\pi d_p^2}{4} \left(\frac{\rho_g U^2}{2}\right)$$

where, U is the free stream gas velocity, and C_D is the drag coefficient, which is a function of the
Reynolds number

$$Re = \left(\frac{d_p U \rho_g}{\mu}\right)$$

The characteristic relation between C_D and Re for particles of different shape factors is expressed as:

$$C_D = \frac{a_1}{Re^{b1}}$$

where, the constants $a_1$and $b_1$can be approximated as below (Howard, 1989).

Table 1: Law corresponding to Reynolds number

Range of Re	Region	a_1	b_1
0<Re<0.4	Stoke's Law	24	1.0
0.4<Re<500	Intermediate Law	10	0.5
500<Re	Newton's Law	0.43	0.0

VII. EXPERIMENTAL PROCEDURE

Materials of uniform sizes are separated from the given mixture. The separated material is then screened through different mesh sizes particles are obtained then avg. particle size is obtained. Voidage is determined by pouring the material in burette along with kerosene oil. Reading is then noted down and voidage is then determined.

Grid pressure drop is found out from the fluidization column when there was no bed present. Material is poured into the column and bed is created then we started the centrifugal pump and by the flow of air through the bed and process of fluidization starts which is determined by the observing the reading of manometer and rotameter.

Same procedure was repeated for Sago.

VII.A MATREIAL SPECIFICATION

Table 2: Physical properties of test material

Material	Density(gm/cc)	Particle diameter (cm)	Voidage	Sphericity
1.Mustard Seed	1.25	0.192	0.42	1
2.Sago	1.35	0.253	0.49	1

Table 3: Physical properties of fluid

Fluid	Viscosity (poise)	Density (gm/cc)
1.Air	0.00018	0.00123
2.Methyl Alcohol	0.0059	0.791

VIII. EXPERIMENTAL SETUP

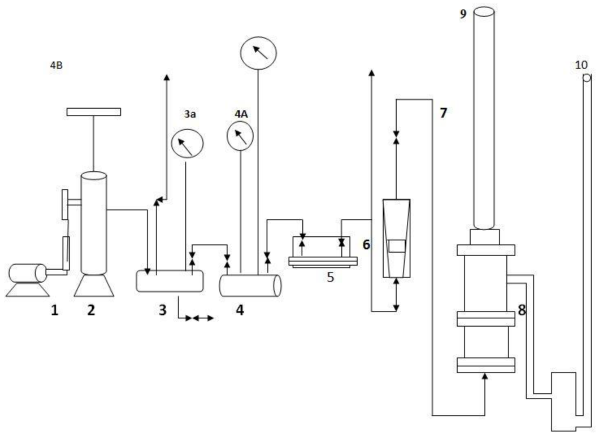
A diagram of the experimental set up is shown in figure. The apparatus consists of air compressor, air receiver, surge vessel, silicon gel drier, Rotameter, mild steel calming section, perpex fluidizer column, manometer and device for measuring temperature. A fine pore ceramic grid was used as bottom grid for first set of reading while a fine mesh brass wire screen was used as bottom grid for second for set of reading.

The fluidization on column was made of transparent perpex cylinder (6.3 cm internal diameter and 152 cm long). The cylinder was placed over the mild steel calming vessel. The ceramic or brass grit were placed between the calming section and the fluidization column to support the particles. The calming section was filled with glass balls 6 cm in diameter to ensure proper distribution of fluidized air.

The air supply was fitted with a pressure gauge and a temperature gauge to monitor the line pressure and temperature. As silica gel drier removed the moisture from the fluidizing air. The receiver is provided with a drawn valve to drain any compressor oil carried along with the fluidizing air. The air flow rate was measured with a valve. The pressure drop across the fluidization column was measured with a manometer.

Red mustard seed were used as solid particles for conducting the experiment which were charged in the fluidizing column. Mercury was used as monometer liquid while ceramic grid was used as first set of readings while methyl alcohol was used as manometer liquid with brass grid for second set of reading. Air was used as fluidizing medium.

Fig. 4: Schematic diagram of the loop for momentum transfer in a fluidized bed



Sl. no. as per Fig. 5	Name of part
1	Induction Motor
2	Air Compressor
3	Air Receiver
3a	Pressure gauge
4	Surge vessel
4A	Pressure gauge
5	Silica Gel Drier
6	Rotameter
7	Connections Pipe
8	Supporting Grid
9	Fluidization Column
10	Manometer

Table 4: Details of construction setup

IX. OBSERVATION

Table 5: Observation for Mustard Seed

Sl.no	Velocity (m/s)	Reynolds Number (R_{ep})	Bed Pressure Drop (P_{bed})	Bed expansion ratio ($R=h/h_o$)
1	0	-	1162.770	1.000
2	8.79	11.71	1550.360	1.000
3	20.52	27.35	2092.980	1.000
4	33.71	44.53	2558.094	1.000
5	45.45	60.58	3100.720	1.000
6	60.10	80.11	3720.860	1.000
7	73.29	97.69	4341.008	1.000
8	87.96	117.24	4950.012	1.000
9	101.14	134.81	5245.973	1.009
10	107.01	142.64	5170.810	1.060
11	114.34	152.41	5098.880	1.109
12	120.21	160.23	5130.423	1.154
13	127.53	169.99	5148.734	1.187
14	133.40	177.80	5249.792	1.200
15	134.86	179.78	5289.980	1.245

Table 6: Observation for Sago

Sl.no.	Velocity (m/s)	Reynolds Number (R_{ep})	Bed Pressure Drop (P_{bed})	Bed expansion ratio ($R=h/h_o$)
1	20.52	44.52	1937.950	1.000
2	45.45	98.62	2325.540	1.000
3	73.31	154.08	3100.720	1.000
4	101.17	219.53	3875.900	1.000
5	127.56	276.81	4651.080	1.000
6	155.42	373.26	5426.260	1.000
7	181.81	394.52	6201.440	1.000
8	193.54	419.98	6589.030	1.000
9	208.21	451.81	7051.800	1.030
10	219.94	477.26	7126.980	1.058
11	234.60	509.08	6914.570	1.147
12	249.26	540.89	6892.186	1.196

X. GRAPHS

Fig. 5. Pressure Drop vs Velocity (For Mustard Seeds)

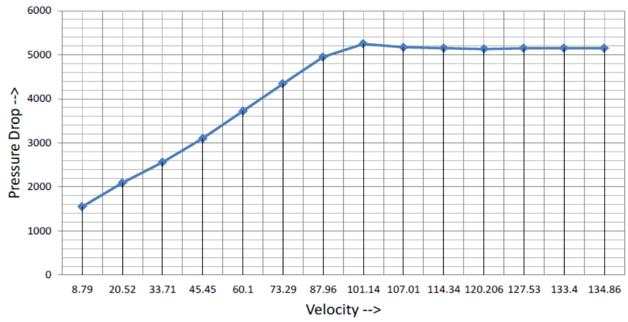


Fig. 6. Pressure Drop vs Velocity (For Sago)

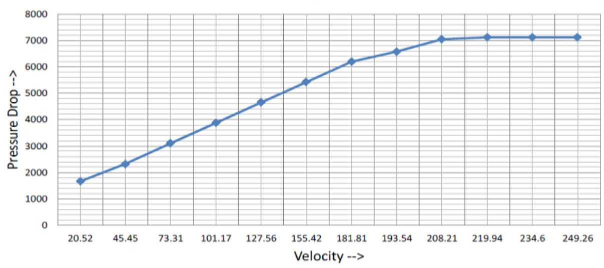
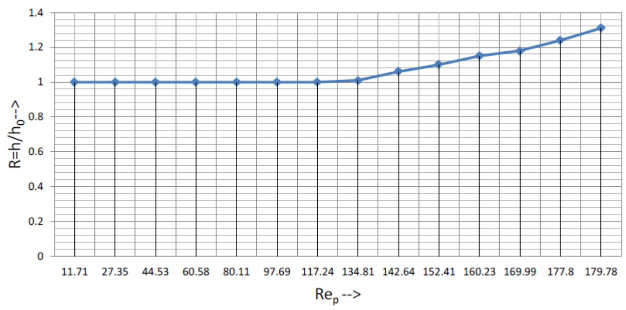


Fig 7. Bed Expansion Ratio vs Reynolds number (For Mustard Seed)



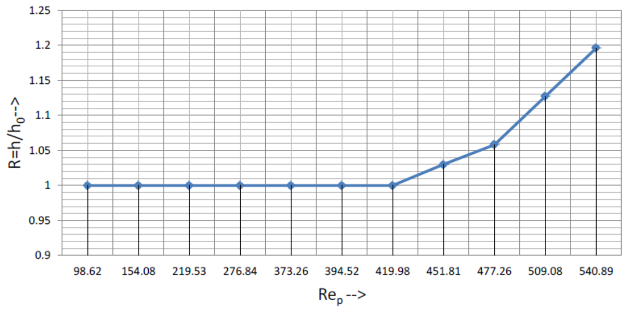


Fig 8. Bed Expansion Ratio vs Reynolds number (For Sago)

XI. CONCLUSION

From the experimental values of mustard we find that U_{mf} calculated is 73 cm/sec and U_{mf} from graph is 100 cm/sec. The experimental values of Sago suggest that U_{mf} calculated is 292.32 cm/sec and U_{mf} from graph is 219 cm/sec. In both cases the deviation are 27% and -25%. We also deduce from the experiment that bed expansion ratio for different materials depend on their densities and nature of fluidizing fluid. High density material have low bed expansion ratio while low density material have larger bed expansion ratio. Also spherical particles with larger diameter have higher minimum fluidizing velocity as compared to spherical particles with lower diameter. After minimum fluidizing condition, the pressure drop across the bed is found to remain constant for large range before bubbling condition. Research in this area must move towards modeling, simulation and optimization of fluidized bed operation-process. Therefore, the fluidization engineering scenario is enhanced intensively and the literature must report and accompany this progress.

NOTATION

U_{mf}	minimum fluidization velocity
A_c	cross-sectional area
h_s	height of the bed (before particles start to lift)
h	height of the bed anytime
W_s	mass of the solids in the bed
ΔP	bed weight per unit cross-sectional area
Re_{mf}	Reynolds number at minimum fluidization velocity
U	free stream gas velocity
C_D	drag coefficient

Greek Letters

ρ_c	density of solid catalyst particle
ε_s	porosity of settled bed
ε	porosity of expanded bed
Φ_s	particle sphericity

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