The Determination of Minimum Bubbling Velocity, Minimum Fluidization Velocity and Fluidization Index of Fine Powders (Hematite) using Gas-Solid Tapered Beds

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Abstract: In this report the experimental determination of maximum pressure drop, minimum fluidization velocity, and minimum bubbling velocity and Fluidization index of fine Hematite powder has been done. A tapered bed having a tapered angle of 7.40º has been used in this study. Tapered beds have a velocity gradient along the axis unlike conventional beds. They have been used as an alternative to conventional cylindrical beds to deal with problems such as slugging. In this study pressure drop was measured using a U tube Manometer). Air at ambient conditions (310 K) was used as the fluidizing medium. The air flow rates were measured using a rotameter. Initially stagnant bed height of 5.5 cm was taken. The superficial velocity (Air flow rate/area) of the air passed from below the distributor was increased in small increments. The corresponding manometer readings were noted down. Pressure drop at the corresponding superficial velocity was calculated from the height difference between two arms of the manometer. The Superficial velocity at which maximum pressure drop was obtained was noted as the minimum fluidization velocity. At the onset of fluidization bubbles appeared and the superficial velocity at which the first bubbles appeared was taken as the minimum bubbling velocity. The fluidization index is the ratio of minimum bubbling velocity to the minimum fluidization velocity. Geldart in his works had shown that fluidization index is a very useful criterion to determine the hydrodynamic behavior of fine powders. The effect of variation of the tapered bed height on maximum pressure drop, minimum fluidization velocity, minimum bubbling velocity and fluidization index were studied by repeating the tests mentioned above for stagnant bed heights of 10cm and 15 cm.

Keywords: Gas–solid fluidization; Minimum fluidization velocity; minimum bubbling velocity; Maximum pressure drop; Tapered fluidized bed.

1. Introduction

Tapered beds are useful for fluidization of materials with wide particle size distribution and also for exothermic reactions. Conventional fluidized beds face severe problems like slugging which can be tackled by introduction of baffles, operation in multistage units and imparting vibrations from time to time. The introduction of tapered bed instead of a conventional cylindrical bed is an alternative technique in gas solid fluidization which can be used to tackle this problem. Tapered beds possess certain characteristics that make them of practical interest. The variation of fluid velocity with axial position allows for different types of fluidization at different positions in the column [1].

Many of the important characteristics of gas fluidized beds depend upon the behavior of gas bubbles which are generated near the distributor and rise through the beds growing in size. The no of bubbles become fewer by coalescence. The size of bubbles at a given height over and above the distributor is governed by the initial size of bubbles as well as the mode of coalescence in the upper zones of the bed [2].

Under normal operating conditions, the gas flow is concentrated in the center of bed and a circulatory motion is setup accounting for good quality of fluidization in fluidized beds [5]. Due to angled wall, random and unrestricted particle movement occurs in a tapered bed thereby reducing back mixing [3].

As the superficial velocity of the air passed below the distributor in a fluidized bed increases, the pressure drop increases linearly, then a certain point comes where with an increase in superficial velocity the pressure drop remains almost constant. The Minimum fluidization velocity is obtained when the pressure drop across the bed is maximum, at that point beyond which with an increase in superficial velocity pressure drop remains almost constant. Minimum bubbling velocity (umb) is the superficial velocity at which the bubbles first appear.

umb/umf = Fluidization index.

The fluidization index is a very important parameter as it gives us the measure of the degree to which the bed can be expanded uniformly. The ratio is relatively high for Geldart A powders [23].
In this study the fluidization index of hematite powder has been experimentally determined.

2. Literature Review

As In the course of this project we have dealt with and will be dealing with powders therefore it is essential that we understand how powders of different particle size and density behave when fluidized. This phenomena has been best explained by Geldart[4] in his classification of particles:

2.1. Geldart classification of particles

Group C: Cohesive or very fine powders. Their size varies from 20 to 30 μm. Normal fluidization is extremely difficult because of the high interparticular force of attraction in these particles. Group A: Aerable having a small mean size and low particle density (<1.4 gm/cm^3). For this group the particle size is between 20 and 100 μm .They fluidize easily. When the solids are fluidized the bed expands considerably (by factors of 2 and 3) before bubbles appear. The gas bubbles rise rapidly and coalesce and split frequently as they rise through the bed. When the bubbles grow to vessel diameter they turn into axial slugs. Group B: Sandlike, particle size 40 u_m<DP<500u_m. The density is 1.4< ρ_s< 4 g/cm^3. These solids fluidize well with vigorous bubbling. The bubbles form as soon as the gas velocity exceeds u_{mf}. Thus u_{mb}/u_{mf} = 1. Bubbles size increase roughly linearly with distance above the distributor and excess gas velocity, (u_0-u_{mf}).. Vigorous bubbling encourages gross circulation of solids.

Group D: spoutable, or dense particles. The particles in this region are above 600 μm and typically have high particle densities they are difficult to fluidize. They behave abnormally giving large exploding bubbles or severe channeling.

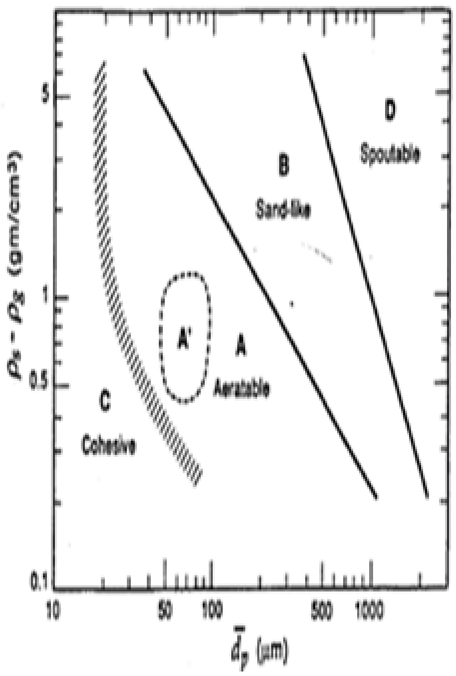

Figure 1: The Geldar Classification of particles for air at ambient conditions

2.2. Apex angle/Tapered angle

A relatively small apex angle is selected to accommodate the increase in gas volume with height in a deep bed [6]. A large apex angle is selected to suppress slugging and to reduce bed expansion and its fluctuation effectively over a much wider range of fluidization velocities [7, 8].

2.3. Flow regimes in a tapered bed

Y Peng and L.T Fan in their work on finding out the Hydrodynamic characteristics of fluidization in liquid solid tapered beds explained the various flow regimes in tapered fluidized bed in terms of pressure drop vs superficial velocity. The flow regimes were obtained during the fluidization of spherical glass beads, using water as the fluidizing medium.[16]

Prior to the experimental run, the bed of glass beads were fully fluidized, subsequently the flow-rate of water was gradually reduced until the glass beads became loosely settled to form the initial fixed bed.

When the flow rate of water was increased, the net pressure drop through the bed of particles($-\Delta p_N$) varied following the path typically described by the solid curve marked as O⇨A⇨B⇨C⇨D⇨E in the figure given in next page..

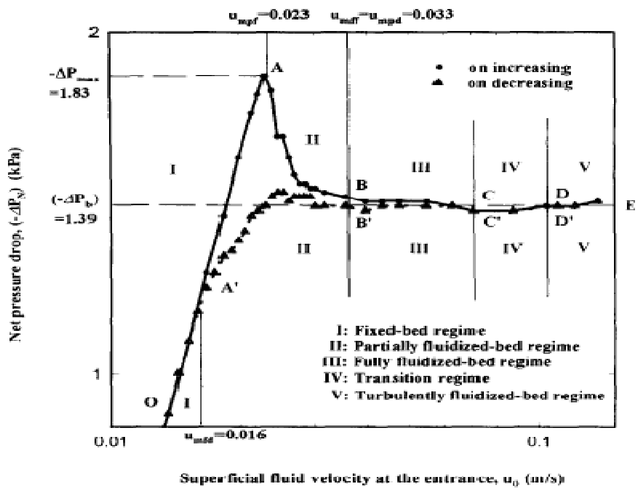

Figure 2: Pressure drop vs superficial velocity [Peng et al.1996]

1) O⇨A, Fixed bed regime. At a low flow rate the fluid simply passes upward through the bed without disturbing the particles. The bed is maintained at a constant voidage $€_0$ at a height H. The magnitude of ($-\Delta p_N$) rises steeply with increase in flow-rate as in any fixed bed of particulate matter.

2) A⇨B, Partially fluidized bed regime, At point A , the particles in immediate vicinity of the distributor are lifted since flow-rate is sufficiently high , causing any empty cavity containing a relatively small no of particles to form next to the distributor. The cavity is unstable. A fluidized zone is formed in the cavity region. The bed expands as the flow-rate increases until it reaches the top of the bed. Bed expands, net pressure ($-\Delta p_N$) drops while the flow-rate increases from max value of ($-\Delta p_{max}$) at point A to Δp_b at point B.

3) B⇨C, Fully fluidized regime, The bed reaches its critical stage of full fluidization at point B. The corresponding superficial fluid velocity through the entrance = minimum velocity of fluidization.(U_{mf}). With only a slight increase in the flowrate beyond B, the fluidized region breaks through the top surface of the particle bed.

4) C⇨D, Transition regime, $-\Delta p_N$ remains almost constant beyond C.

5) D⇨E, Turbulent fluidized bed regime, In this region particles move randomly, voidage are distributed uniformly.

2.4. Pressure drop across beds

The frictional pressure drop through fixed beds of length L containing isotropic solids of a single size d_p is given by Ergun's equation

$$\frac{\Delta p_b}{L_m} = 150\frac{(1-\epsilon_m)^2}{\epsilon_m^3}\frac{\mu u_0}{(\Phi_s d_p)^2} + 1.75\frac{1-\epsilon_m}{\epsilon_m^3}\frac{\rho_g u_0^2}{\Phi_s d_p}$$

Pressure drop across distributors is given by
$\Delta p_d = (0.2\text{-}0.4)\ \Delta p_b$ [9]

The above assumption has been verified by various analyses and experiments and represents a reasonable upper bound of the required pressure drop for smooth operations [11].

Maruyama et al in their paper on Fluidization in tapered vessels proposed analytical methods to predict the pressure drop in tapered fluidized vessel [12].

Pressure drop $(\Delta p_b) = (1-\epsilon_m)(\rho_s - \rho_g)\, g(Z_{2.0}^2 - Z_1^2)/(Z_2+Z_1)$

The following equation for pressure drop has been developed from Ergun's equation, which also includes the pressure drop , which is due to the K.E change in the tapered bed. [14]

$$-\Delta P_{max} = \frac{C_1 H_s (D_0)\, U_0}{D_1} + \frac{[C_2 H_s\, D_0(D_0^2 + D_0 D_1 + D_1^2)\ ]U_0^2}{3D_0^2} + \frac{1\,(U_0^2)\,[\,(D_0/D_1)^4 - 1]\,\rho_f}{2\ \varepsilon_0^2}$$

Jing et al [13] developed a model based on Ergun's equation for calculating pressure drop but neglecting the pressure drop due to the Kinetic energy change in the cylindrical bed.

Their equation for nearly spherical particles is given by;

$-\Delta P_t = C_1 U_0 H_s\, r_0/r_1 + C_2 U_0^2 H_s\, r_0/3r_1^2(r_0^2 + r_0 r_1 + r_1^2)$

2.5. Minimum fluidization velocity

Minimum fluidization velocity is obtained when the pressure drop across the bed is maximum. The pressure drop increases from 0 to $u_{mf,}$ after that it remains almost constant. The interception of the two lines is defined as the minimum fluidization velocity. [15]

To predict minimum fluidization velocity Kunii and Levenspiel[11] used the following analogy;

Drag force by upward moving gas = Weight of particles

Pressure drop across bed * cross sectional area of tube = Volume of bed* fraction consisting of solids * Specific weight of solids

$\Delta p_b A_t \quad = \quad W = \quad A_t L_{mf}(1-\epsilon_{mf})[\ \rho_s - \rho_g\ ]*g$

$\Delta p_b/L_{mf} \quad = \quad (1-\epsilon_{mf})[\ \rho_s - \rho_g\ ]*g$

In general, for isotropic solids, the minimum fluidization velocity, u_{mf} is given by

$$\frac{1.75(d_p u_{mf}\,\rho_g)^2}{\epsilon_{mf}^3\Phi_s\,\mu^2} + \frac{150(1-\epsilon_{mf})\,(d_p u_{mf}\,\rho_g)}{\epsilon_{mf}^3\Phi_s^{\,2}\mu} = \frac{d_p^3\,\rho_g\,(\rho_s - \rho_g)g}{\mu^2}$$

Peng and Fan [16] also developed a model for correlating minimum fluidization velocity and maximum pressure drop for solid liquid system in tapered beds.

$C_1 U_{mf} + C2\,(D0/\,D_1)^2\,U^2{}_{mf} \ - (1-\varepsilon_0)\,(\rho_s - \rho_f)\,g \times\ [(D^2{}_0 + D_0 D_1 + D^2{}_1)\,/3D^2{}_0] = 0$

Based on the experimental data obtained by Sau et al.[14] for different types of materials , in gas-solid tapered bed and by use of dimensional analysis and estimating the constant coefficients by non linear regression they gave the dimensionless correlation for U_{mf} and the correlation for pressure drop:

$Fr(\text{Froude No}) \ = \ U_{mf}/\,(gd_p)^{0.5} \ = \ 0.2714(Ar)^{0.3197}(\sin\alpha)^{0.6092}(\varepsilon_0/\,\Phi_s)^{-0.6108}$

$$\Delta P_{max} = 7.457 * \frac{(D_1)^{0.038}}{(D0)^{0.038}}\ \frac{(d_p)^{0.222}}{(_{D0})^{0.222}}\ \frac{(H_s)\ 0.642}{(D0\,)^{0.642}}\ \frac{(\rho_s)^{0.723}}{(\rho f)^{0.723}}$$

In their study of the effect of different tapered angle on minimum fluidization velocity they observed that maximum pressure drop and the minimum fluidization velocity increased with increase in tapered angles.

They also experimentally observed that U_{mf} was not a function of stagnant bed height in conical tapered beds. This phenomenon was also observed by Povrenovi et al[17]

Some studies of minimum fluidization of gas-solid 2 D tapered bed mentioned that u_{mf} decreases or increases with an increase or decrease in atmospheric pressure respectively.[18,19,20]

The method of measuring u_{mf} by Caicedo et.al. was the one normally used which consists of measuring the pressure drop across the bed as a function of increasing and decreasing gas velocity as bed passes from fixed bed to fluidized bed.[19] The results of Caicedo et al. experiments showed that there is an increase in u_{mf} with an increase in bed height. This trend is attributed to the increasing friction at the wall.[19]

2.6. Minimum bubbling velocity and fluidization index:-

Abrahmsen and geldart [21] correlated the values of minimum bubbling velocity with gas and particle properties as follows:

$U_{mb} \ = 2.07\exp(0.716\ F)\ (x_p{*}\ \rho_g^{0.06}\ /\ \mu^{0.347})$

(Where F is the fraction of powder <45 µm.)

Minimum fluidization velocity for particles less than 100 µm as given by Baeyen's equation:

$U_{mf} = (\rho_p - \rho_g)^{0.934} g^{0.934} x_p^{1.8}/(1100{*}\mu^{0.87} * \rho_g^{0.066})$

Fluidization index $\ = \ Umb/U_{mf} \ = \ 2300\ \ \rho_g^{0.126}\ \mu_g^{0.523}\exp(0.176F)/[d_p^{0.8} g^{0.934}(\rho_p - \rho_g)^{0.934}]$ [23]

The higher the ratio more is the beds capacity to hold gases between minimum fluidization and bubbling point. [22]

A high fluidization index for a catalyst implies that it has a certain plasticity and can be expanded, contracted and bent around the corners. A low fluidization index implies a brittle fluidization state in which a small change could cause a break from the uniformly fluidized catalyst to a packed bed or a bubbling regime. [22]

In some of the systems where fluidization takes place bubbles occur at velocities which is very close to u_{mf} . In some cases bubbling occurs at 3 times $u_{mf.}$. The range in which smooth trouble free fluidization occurs is extended by using fluids of high density or operating at higher pressure because the fluidization no increases slightly with pressure and viscosity of the fluidization medium. It has been observed that powders of average particle size less than 100 µm expands uniformly without bubble formation in a limited range of gas velocity greater than minimum fluidization velocity. With materials such as fine cracking catalyst the index can be around 1.2 and the ratio varies from 1 to 2 over the range of particle sizes. With greater than 2 in some special cases. [23] Davies and Richardson have obtained the values for fluidization Index up to 2.8 using cracker catalyst (dp = 55 µm, s.g. = 0.95) fluidized in air at atmospheric pressure.[10]

We may obtain differences in calculated and measured values in fluidization index . This is due to particles shape and its effect on drag and minimum fluidization velocity. Thus it is better to measure u_{mf} and u_{mb} rather than to rely on correlations.[23]

Singh et al.[23] predicted minimum bubbling velocity, fluidization index and range of particulate filtration for gas solid fluidization in cylindrical and non cylindrical beds.

Particulate fluidization exists between minimum fluidization velocity and minimum bubbling index.
Singh et al.[23] came up with the following to predict minimum bubbling velocity of different types of fluidized bed.
Cylindrical bed:

$$U_{mb}=0.5231(d_p/D_c)^{1.13}(D_c/h_s)^{-0.0384}(\rho_p/\rho_f)^{0.74}$$

Semi-cylindrical bed:-

$$U_{mb}=0.168(d_p/D_c)^{0.994}(D_c/h_s)^{-0.1849}(\rho_p/\rho f)^{0.80}$$

Hexagonal bed:-

$$U_{mb}=0.15(d_p/D_c)^{0.5733}(D_c/h_s)^{-0.0887}(\rho_p/\rho_f)^{0.5384}$$

Square bed:-

$$U_{mb}=0.168(d_p/D_c)^{0.27}(D_c/h_s)^{-0.0132}(\rho_p/\rho_f)^{0.2825}$$

Singh et al performed experiments for dolomite, Manganese ore, Chromite ore and coal. The particle size range was ($6*10^{-4}$ m to $9*10^{-4}$ m). They observed fairly comparable value with experiments using the equations they predicted. They also observed that for identical operating conditions minimum bubbling velocity and fluidization index are maximum in case of either semi-cylindrical or hexagonal bed for most of the operating conditions and least in case of square bed. As particulate fluidization is maximum in case of semi-cylindrical beds and less in case of other beds , hence when particulate fluidization is the main requirement semi cylindrical beds are most preferred.[23].

3. Experimental setup and materials used:-

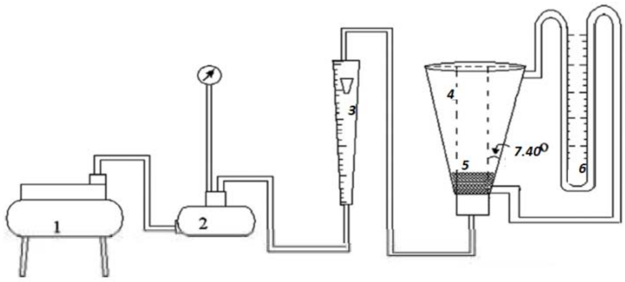

Figure 3: The experimental setup
1. Compressor
2. Receiver
3. Rotameter
4. Tapered bed with a tapered angle of 7.40º
5. Bed containing solids
6. U tube Manometer

Compressor:
An air compressor having sufficient capacity has been used. It takes air from the atmosphere and compresses the air and stores it in the receiver.

Receiver:
The receiver is a horizontal cylinder which is used to store the compressed air from the compressor. There is one G.I pipe inlet to the accumulator and one by pass from one end of the cylinder. The purpose of using the accumulator in the line is to dampen the pressure fluctuations. The accumulator is fitted with a pressure gauge.

Rotameter:
A rotameter having a range of 0-30 LPM was used to measure the air flow rates.

Tapered Bed:
A transparent tapered column was used to visually observe the fluidization and bubbling. The tapered angle of the bed used was 7.40º. Two pressure tapings are provided for observing the bed pressure drop.

U-tube Manometer:
A U tube Manometer having Carbon tetrachloride as the manometric fluid (density = 1630 Kg/m^3) was used to measure the bed pressure drop.

Material used:
Fine hematite powder was used in this study having a bulk density of 2.35 g/cm^3. The mean particle size(Mean Particle diameter) of hematite powder used was 20 µm.

4. Results and Discussion

The experiments were carried out for determination of minimum bubbling velocity, minimum fluidization velocity, the maximum pressure drop and fluidization index of hematite powder. Three experiments were conducted by varying the bed height. Bed heights of 5.5,10 and 15 cm were used and the effect of bed height on minimum bubbling and fluidization velocity and fluidization index was reported.

Before the start of each test the bed was fluidized once so that the powders were in a loosely packed state.
For a bed height of 5.5 cm the following was observed:-

Tapered Angle = 7.40	
Initial static bed height = 5.5cm.	
Superficial velocity	Pressure drop(across bed)
(in m/sec)	(in Pa)
0.003125	191.82
0.00625	249.36
0.0125	383.64
0.0375	517.92
0.0625	652.19
0.075	748.1
0.09375	652.19
0.15625	709.74

Table 1: Superficial velocity and corresponding pressure drop for bed height of 5.5 cm.

The plot of pressure drop vs superficial velocity:

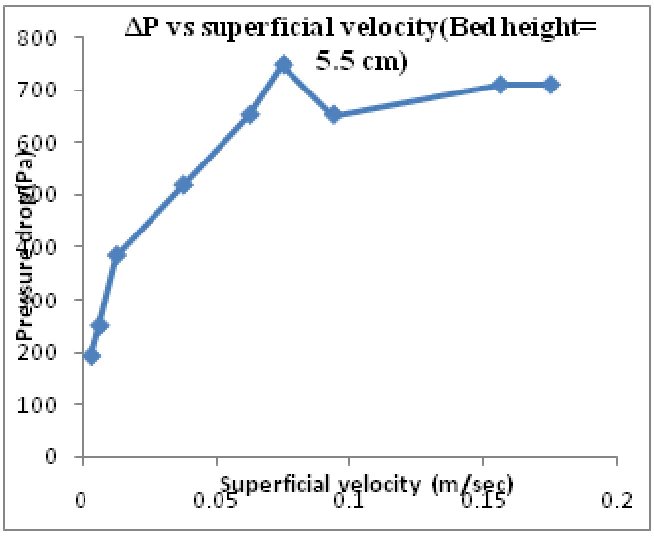

Figure 4: Plot of pressure drop vs superficial velocity for bed height = 5.5 cm

The minimum bubbling velocity obtained = minimum fluidization velocity. Both u_{mb} and u_{mf} were observed to be equal to 0.075 m/sec.

The fluidization index for initial static bed height of 5.5 cm = u_{mb}/u_{mf} = 0.075/0.075 = 1.

For a bed height of 10.0 cm the following was observed:

Tapered Angle = 7.40	
Initial static bed height = 10.0 cm.	
Superficial velocity	Pressure drop(across bed)
(in m/sec)	(in Pa)
0.025	268.55
0.05	402.82
0.9375	652.19
0.13125	863.2
0.15	767.28
0.175	786.46
0.1875	805.65
0.21875	882.38

Table 2: Superficial velocity and corresponding pressure drop for bed height of 10 cm.

The plot of pressure drop vs superficial velocity:

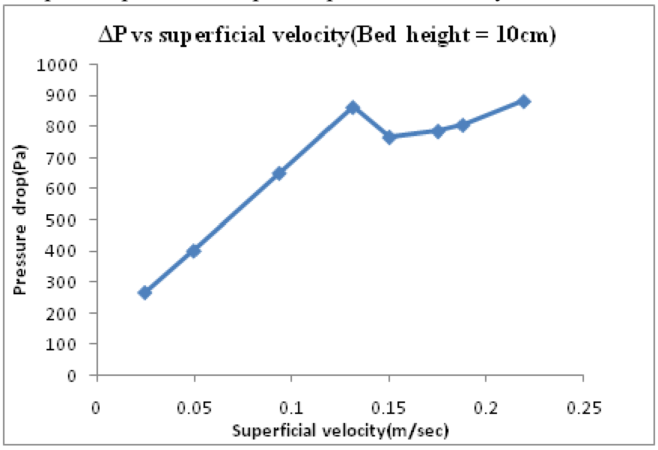

Figure 5: Plot of pressure drop vs superficial velocity for bed height = 10 cm.

The minimum bubbling velocity obtained = minimum fluidization velocity. Both u_{mb} and u_{mf} were observed to be equal to 0.131 m/sec.

The fluidization index for initial static bed height of 10 cm = u_{mb}/u_{mf} = 0.131/0.131 = 1.

For a bed height of 15 cm the following was observed:

Tapered Angle = 7.40	
Initial static bed height = 15.0 cm.	
Superficial velocity	***Pressure drop(across bed)***
(in m/sec)	(in Pa)
0.0125	306.91
0.0625	479.55
0.075	537.1
0.125	748.104
0.15625	1016.65
0.1875	1112.57
0.21875	1093.38
0.25	1150.93

Table 3: Superficial velocity and corresponding pressure drop for bed height of 15 cm.

The plot of pressure drop vs superficial velocity:

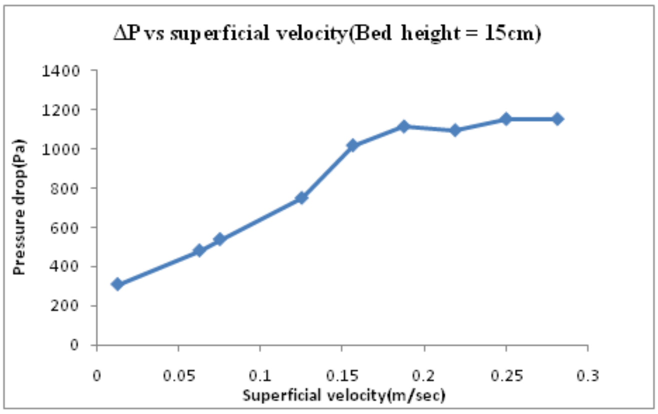

Figure 6: Plot of pressure drop vs superficial velocity for bed height = 15.0 cm.

The minimum bubbling velocity obtained = minimum fluidization velocity. Both u_{mb} and u_{mf} were observed to be equal to 0.156 m/sec.

The fluidization index for initial static bed height of 15.0 cm = u_{mb}/u_{mf} = 0.156/0.156 = 1

An increase in maximum pressure drop was observed with increase in height. The maximum pressure drop for 5.5 cm initial bed height being 748.1 Pa, that for an initial bed height of 10 cm being 863.2 Pa and that for 15 cm initial bed height being 1016.6 Pa.

The minimum bubbling velocity and minimum fluidization velocity increased with increase in bed height from 0.075 m/sec in 5.5cm bed to 0.131 m/sec in 10 cm bed to 0.156 m/sec in 15 cm bed.
The increase in minimum fluidization velocity and subsequently of minimum bubbling velocity may be due to increased frictional interaction with the walls [23].

A fluidization index value of 1 was observed in all of the cases indicating that bubbling starts from the onset of fluidization.
A fluidization index value of 1 also indicates that the bed has a very less capacity to hold gases between minimum fluidization and bubbling [22]

5. Conclusion and Future Scope

This study experimentally determined the fluidization index of fine Hematite powders. An increase in minimum bubbling velocity and minimum fluidization velocity were observed with an increase in bed height. The fluidization index value of 1 was observed in all the cases.

Till now no model has been devised for finding out the minimum bubbling velocities using tapered beds. The models must be based on the particle size, and should be different for different particle size range. This is because different bubbling behavior and fluidization index is observed for powders of different size ranges.

Future studies can be planned using different powders and by checking the accuracy of the existing models for determination of minimum bubbling and fluidization velocities. If the error between the results of existing models and the experimental results is too high than a new model can be devised. Also a CFD simulation using ANSYS software can be done to simulate the fluidization and bubbling in tapered beds.

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