A

THESIS REPORT

ON

TWO PHASE FLUIDIZATION OF DOLOMITE PARTICLES IN A TAPERED FLUIDIZED BED

SUBMITTED

BY

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CERTIFICATE

This is to certify that the thesis entitled "**Two phase fluidization of coarse materials in a tapered fluidized bed",** which is being submitted by Bhabesh Majhi (roll no-109ch0074) in partial fulfillment for the requirements of the degree of Bachelor of Technology in Ch-mical Engineering is a bonafied record of observations carried out by him under my su-pervision and guidance.

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ROURKELA (Bhabesh Majhi)

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ABSTRACT

The object of this work is to study onthe tapered bed fluidization in tapered bed& to correlate the relation between pressure drop and volume flow rates. The quality and quantity of mixing and segregation of heterogenous materials are analysed. To study the characteristics of tapered beds , several experiments have been carried out with different tapered angles of the bed , with regular as well as irregular particles of different sizes and densities. The tapered angles of the beds have been found to affect the characteristics of the bed. Detailed visual observations of fluid and Particle behavior and measurements of the pressure drops have led to the identification of five flow regimes.Tapered fluidized beds are widely used in industrial operations to fluidize a wide range of particle sizes , and are thought to induce relatively strong particle mixing. Like their straight – sided counter parts , tapered fluidized beds are often considered as a homogenous emulsion phase through which bubble propagates. However it has been shown that gas flow through dense Particles is heterogenous , generating a central fluidized core and unfluidized peripheral regions. The time scale for particle turnover in the central fluidized region is much shorter than that of particles captured within the peripheral regions.

CHAPTER - 1

INTRODUCTION

Fluidization (or fluidisation) is the process similar to the liquefaction of granular materials from static solid state to dynamic fluid like state.This process occurs when fluid like gas or liquid passes through the granular substances (or materials).

When gas flow is introduced to the fluidized bed containing granular materials , the gas opt to flow through the granular materials considering the aerodynamic drag on each particle. Gas flow at low velocities having low aerodynamic drag to the granular materials. As the velocity increases slowly the drag forces also increases simultaneously on the gas flow and hence aerodynamic drag counteract the gravitational forces causing the bed to expand in volume as the particles move away from each other.Further increasing the velocity of the gasflow it will reach a critical value where all drag forces on fluid flow counter balances the gravitational forces causing the particles to remain suspended within the fluid. At this critical value the bed is said to be fluidized bed and exhibit the fluidic behavior. By further increasing the gas velocity the bulk density gradually decreases and the fluidization becomes more violent until the particles no longer form a bed and are conveyed upwards by the gas flow.

Over past decades fluidized bed has extensive applications compared to fixed bed and can be adaptable fluid - solid contact device in chemical and biochemical industries. Extensive use of fluidization began in the petroleum industries with the increasing development of fluid bed catalytic cracking. Presently fluidization techniques have found extensive application in various fields like roasting of ores , drying of fine solids , adsorption of gases , regeneration of catalyst , fluidization bed combustion of coal (Winkler Process) , biological treatment of wastewater , incineration of waste materials , coating nuclear fuel particles , crystallization and numerous gas – solid reactions. The chief advantages of fluidization is that solid particles can easily miscible by the fluid passing through the bed , thus resulting in little or no temperature gradiating even with highly exothermic or endothermic reactions , high rate of heat transfer to cooling tubes immersed in the bed and high rates of heat and mass transfer between the solid and the fluid.

In spite of the above – mentioned advantages , the gas – solid fluidized beds suffer from certain inherent drawbacks like channeling , bubbling and slugging which result in poor homogeneity of the bed – there by affecting the quality of fluidization. Thus maintenance of good fluidization conditions in a large diameters of commercial reactors has been a difficult task for the design engineering.

1.1 CHANNELING

The channeling bed is the fluid bed in which the air (or gas) forms the channels in the bed which they opt to free from the granular materials.

1.2 SLUGGING

The slugging bed is the fluid bed in which the entire bed is covered with bubbles emerging in the bed and there by creating two sub layers between fluid and gas.

1.3 BUBBLING BED

The fluid bed creates an air channel in which the particles are free to move. At higher air velocities thus forming the channels and hence allowing the particles to move the ag-tation becomes more violent and the movement of granular particles become more vigorous. At certain velocities the bed does not expand rather it ceases to the optimum volume at minimum fluidization. Such a bed is called aggregative bed or bubbling fluidized bed. At sufficient air flow rates the terminal velocities of the partcles exceeded and the upper surface of the bed disappears entrainment becomes appreciable, and the solids are carried out of the bed with the airstream.

1.4 VARIABLES AFFECTING THE QUALITY OF FLUIDIZATION

Some of the variables are responsible for affecting the quality of fluidization and hence the gas to solid contact may be summarized as follows:-

a) GAS ENTRY

The fluidized bed should be designed so as to facilitate good gas distribution.

b) GAS VELOCITY

The velocity should be high enough to keep the solids in fluidized state but should not be so high that can cause channeling and slugging.

c) BED HEIGHT

With other variables remaining constant the greater the bed heights can cause difficulty in fluidization. The more bed heights also results in high pressure drops across the bed.

d) GAS AND SOLID DENSITY

It is better to have relative densities of solid and gas more closer to have a smooth fluidization.

e) PARTICLE SIZE

It is easier to have particles with wide range for the fluidization than with the particles of uniform size.

CHAPTER - 2
LITERATURE REVIEW

2.1 INTRODUCTION

Gas – solid fluidized beds exhibit a wide range of behaviors depending on the type , size and density of particles used. Different size particles of low density and wide size –range distribution fluidize in a very different manner compared to closely sized , large , spherical particles.

The major disadvantages of gas – solid fluidization are its inherent characteristics like bubbling , channeling and slugging – which tend to occur at gas velocities well in excess of that necessary to fluidized the bed. Under such conditions the fluidized beds tend to separate into two zones – a dense phase containing major amount of solid particles and small gas – bubbles and a lean phase which contains mainly the gas bubbles with few solid particles.

Tapered fluidized beds have many attractive features among which are they are able to handle different size particles of irregular shapes and properties and for getting excessive particle mixing. These beds have been widely applied in various processes including boillogical treatment of wastewater , incineration of waste materials , coating nuclear fuel particles , coal gasification , crystallization and liquification and coasting sulfide ores. Nevetheless , fundamental understanding of the tapered fluidized beds appear to lag behind their applications.

A fluidized bed is formed when the particles in the bed are in dynamic equilibrium ; the drag forces of fluid on the particles and the buoyancy forces exerted on upward direction against gravitational forces causing the particles to move downwards. This drag forces is constant at any position of a columnar bed of uniform particles ; however , it decreases in the upward direction in a tapered bed accompanied by the reduction in the superficial velocity of the fluid. Thus the particles at the lower part of the bed will first be fluidized upon an increase in the flow rate , in contrast , those at the upper part of the bed remain static. This phenomenon of partial fluidization is pecular to the tapered fluidized bed.

2.2 BEHAVIOUR OF GAS – SOLID FLUIDIZED BEDS

a) CHANNELING

Channeling is the phenomenon in which the fluid tends to flow through the bed of lower solid particle concentration at the minimum fluidization velocity. This results in reduced contact between solid and gas. Channeling may result from initial non – uniformities in the bed and tends to get accentuated by stickiness of particles which prevents them from flowing in to refill the channeling region. A uniform bed of nonsticky particle tends to become well fluidized and at minimum fluidization condition ,the tendency to channeling depends upon stability considerations. With the increase in fluid velocity above that the minimum fluidization ,the pressure drop across the beds and the distributors combined together increases with increase in local velocity , the channeling formation tends to get dampened. Conversely , the channeling tends to get established if the local pressure drop across the bed – distributors system decreases with increase in velocity.

b) BUBBLING

Formation of bubbles is an inherent properties of gas – solid fluidization. This phenomennon can be attributed to the large differences in densities between the solid and the gas as opposed to that between the solid and the liquid in particular fluidization. Bubbles once formed , increase in size by draining gas from the expanded continuous phase as they rise from the bed. Since rise velocity of each bubble is much greater than superficial gas velocity when working with beds of finer particles , the total gas hold – up is decreased and the bed – height therefore drops.

Small bubbles in a small – scale bed , fluidized close to the incipient condition tends to rise close to the wall. As the gas velocity is increased or as bubbles grow in size by coalescence and stronger particle circulation develops , more and more bubbles rise through the central zone of the beds.

c) PATTERNS OF MIXING AND SEGREGATION

It is well known that the mixing degree of solids in a binary fluidized bed with a constant cross – section is always between two extremes , i.e. complete separation and complete mixing of solid phases. Completely separated binary fluidized beds consists of two distinct fluidized layers or beds , each containing one type of particle only. Conversely , particles of each type are distributed uniformly over the entire solid – phase volume in well mixed beds.

Segragation can persist when the gas flow rate is sufficiently large to fluidize the entire bed. Under such conditions it can be shown that segregation can be successfully modeled by drawing an analogy with sedimentation of particles from a turbulent flow field.

CHAPTER - 3
EXPERIMENTAL SET UP

3.1 SET UP

The experimental set up consists primarily of the following components :-

a) AIR COMPRESSOR

It is two stage air compressor of sufficient capacity.

b) AIR ACCUMULATOR

It is a horizontal cylinder used for storing the compressed air from compressor. The operating pressure in the cylinder is kept at 20 psig or at 1.5 atm.

c) PRESSURE GAUGE

A pressure gauge of the required range is fitted in the line for measuring the working pressure.

d) ROTAMETERS

Two rotameters , one for measuring low flow rates and the other for measuring high flow rates.

e) MANOMETER PANEL BOARD

One set of manometer is arranged in this panel board to measure the bed pressure drop. Carbon tetrachloride is used as the manometric liquid.

f) SUPPORTING STRUCTURE

A steel supporting structure is provided to keep the experimental set up vertical and erect.

g) SILICA GEL COLUMN

A silica gel column is provided in the line immediately after the air accumulator to arrest the moisture carried by air from the accumulator.

3.2 PROCEDURE

A weighted amount of material was charged to the fluidizer and with the fixed slant static bed height was recorded. Airflow rate was gradually increased and the correspondi-ng bed pressure drops were noted. After the point of incipient fluidization the expanded slant bed heights were also noted . As the bed fluctuated between two limits typical of gas – solid fluidization ,heights of the upper and the lower surfaces of the fluctuating bed were recorded for each fluid velocity higher than the minimum fluidizing ones.

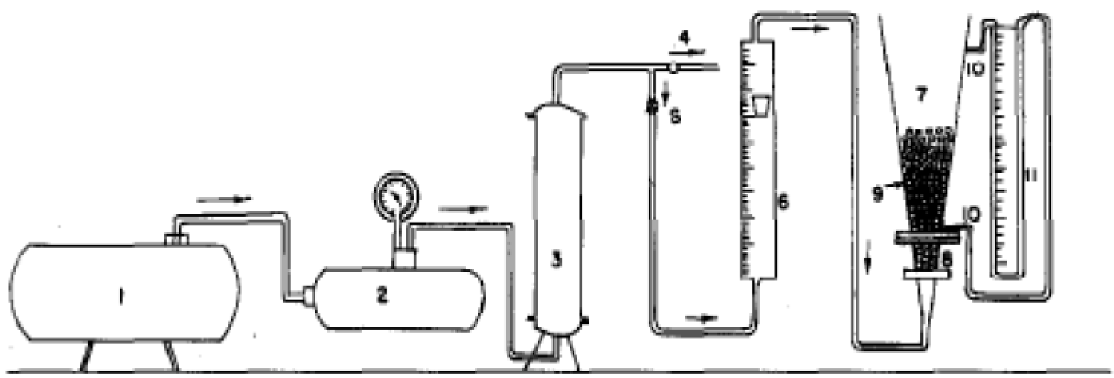


FIG - 1

1) Air compressor

2) Air accumulator

3) Silica gel tower

4) By pass Valve

5) Line valve

6) Rotameter

7) Tapered bed fluidizer

8) Coarse materials packing

9) Coarse Materials in fluidized state

10) Pressure tapings to manometer

CHAPTER – 4

OBSERVATIONS

EXPERIMENTAL DATA TABLES AND PLOTS

TABLE NUMBER - 1

DOLOMITE :- (-18 + 25) MESH SIZE

PARTICLE SIZE :- 600 microns

STATIC BED HEIGHT :- 3 cm

ROTAMETER FLOWRATE (LPM)	VELOCITY (m/s)	MANOMETER		ΔH (cm)	ΔP=δ*g*ΔH (kPa)
		MAXIMUM(cm)	MINIMUM(cm)		
2.5	0.021	22.5	21.4	1.1	0.171
11.25	0.094	22.9	21.0	1.9	0.296
13.75	0.115	23.3	20.8	2.5	0.398
18.75	0.156	23.7	20.5	3.2	0.498
23.75	0.197	24.5	19.6	3.9	0.607
28.75	0.239	24.8	19.3	5.5	0.857
36.25	0.302	25.3	18.7	6.6	1.028
46.0	0.383	25.5	18.5	7.0	1.096
52.5	0.437	25.6	18.4	7.2	1.121
55.0	0.458	25.6	18.4	7.2	1.121

TABLE NUMBER - 2

STATIC BED HEIGHT -4.5 cm

MESH SIZE :- (-18 + 25)

PARTICLE SIZE :- (600 microns)

ROTAMETER FLOWRATE (LPM)	VELOCITY (m/s)	MANOMETER		ΔH (cm)	ΔP (kPa)
		MAXIMUM(cm)	MINIMUM(cm)		
2.5	0.021	22.9	21.3	1.6	0.249
7.5	0.063	23.4	20.7	2.7	0.420
11.25	0.095	24.1	20.1	4.0	0.623
13.75	0.115	24.2	19.9	4.3	0.670
18.75	0.156	24.6	19.5	5.1	0.794
21.25	0.177	25.0	19.2	5.8	0.903
28.75	0.239	25.3	18.8	6.5	1.012
32.5	0.270	25.6	18.8	6.5	1.012
36.25	0.302	25.6	18.5	7.1	1.016
41.25	0.344	25.7	18.5	7.2	1.121

TABLE NUMBER - 3

STATIC BED HEIGHT - 2 cm

MESH SIZE :- (-18 + 25)

PARTICLE SIZE :- 600 microns

ROTAMETER FLOWRATE (LPM)	VELOCITY (m/s)	MANOMETER		ΔH (cm)	ΔP(kPa)
		MAXIMUM (cm)	MINIMUM (cm)		
2.5	0.021	24.5	20.2	4.3	0.670
7.5	0.063	24.9	19.7	5.2	0.810
11.25	0.094	25.0	19.2	5.8	0.903
13.75	0.115	25.1	19.1	5.9	0.919
18.75	0.156	25.2	18.9	6.2	0.966
21.25	0.177	25.2	18.7	6.5	1.012
28.75	0.239	25.4	18.5	6.9	1.075
32.5	0.270	25.4	18.5	6.9	1.075
36.25	0.302	25.6	18.3	7.3	1.137
41.25	0.344	25.7	18.2	7.5	1.168

TABLE NUMBER 4

PARTICLE SIZE - 400 microns

MESH SIZE :- (-25 + 32)

STATIC BED HEIGHT -3 cm

ROTAMETER FLOWRATE (LPM)	VELOCITY (m/s)	MANOMETER		∆H(cm)	∆P(kPa)
		MAXIMUM(cm)	MINIMUM (cm)		
1.25	0.011	25.6	23.6	2.0	0.311
2.50	0.021	26.0	23.2	2.8	0.436
3.75	0.031	26.1	23.1	3.0	0.467
5.00	0.042	26.2	23.0	3.2	0.498
10.5	0.088	26.3	22.9	3.4	0.529
11.25	0.094	26.3	23.0	3.4	0.529
21.25	0.177	26.4	23.0	3.4	0.529
25.00	0.208	26.5	23.0	3.5	0.545

TABLE NUMBER - 5

STATIC BED HEIGHT – 4.5 cm

PARTICLE SIZE :- 400 microns

MESH SIZE :- (-25 + 32)

ROTAMETER FLOWRATE (LPM)	VELOCITY (m/s)	MANOMETER		ΔH(cm)	ΔP(kPa)
		MAXIMUM (cm)	MINIMUM (cm)		
1.25	0.011	26.0	23.6	2.4	0.373
2.50	0.021	26.1	23.4	2.7	0.420
3.75	0.031	26.2	23.2	3.0	0.467
5.00	0.042	26.3	23.0	3.3	0.514
10.5	0.088	26.4	22.9	3.5	0.545
11.25	0.094	26.5	22.9	3.6	0.560
21.25	0.177	26.6	22.8	3.8	0.592
25.0	0.208	26.7	22.7	4.0	0.623

TABLE NUMBER - 6

STATIC BED HEIGHT – 5.5 cm

MESH SIZE :- (-25 + 32)

PARTICLE SIZE :- 400 microns

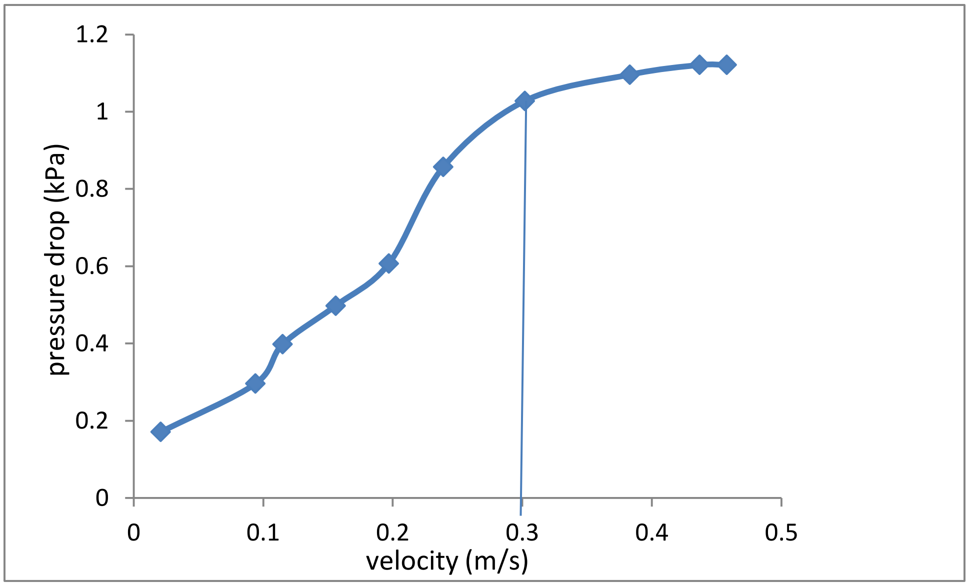
ROTAMETER FLOWRATE (LPM)	VELOCITY (m/s)	MANOMETER		ΔH(cm)	ΔP(kPa)
		MAXIMUM (cm)	MINIMUM (cm)		
1.25	0.011	25.5	23.5	2.0	0.311
2.50	0.021	25.6	23.3	2.3	0.358
3.50	0.029	25.7	23.2	2.5	0.389
3.75	0.031	25.8	23.1	2.7	0.420
5.00	0.042	25.9	23.0	2.9	0.451
8.50	0.071	26.0	22.9	3.1	0.483
10.50	0.088	26.2	22.7	3.5	0.545
11.25	0.094	26.3	22.6	3.7	0.576
20.50	0.170	26.4	22.5	3.9	0.607
25.0	0.208	26.5	22.5	4.0	0.623

PLOT NO. 01

PRESSURE DROP (kPa) VS SUPERFICIAL VELOCITY ,u (m/s)

DOLOMITE:- (-18 +25) mesh size

STATIC BED HEIGHT:- 3 cm



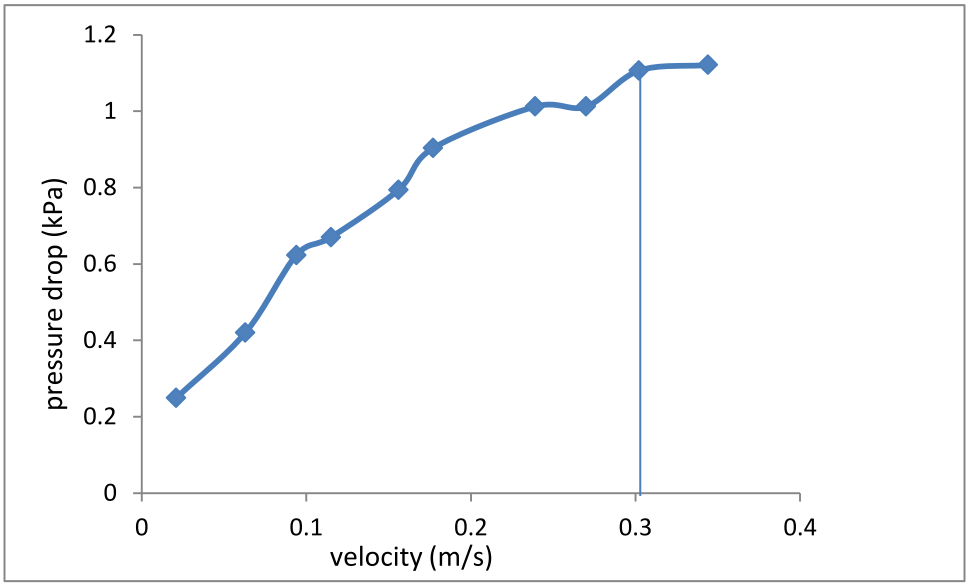
As shown in the above graph the pressure drop increases starting from **0.171 kPa** to **1.12 kPa** corresponding to the increasing superficial velocity from 0.021 m/s to the 1.121 m/s . The pressure drop tends to remain constant when the superficial velocity (u) reaches **0.3 m/s** . After then with the increasing velocity from 0.3 m/s the pressure drop independently remains constant.

PLOT NO. 02

PRESSURE DROP (kPa) VS SUPERFICIAL VELOCITY,u (m/s)

MESH SIZE (-18 +25)

STATIC BED HEIGHT :- 4.5 cm



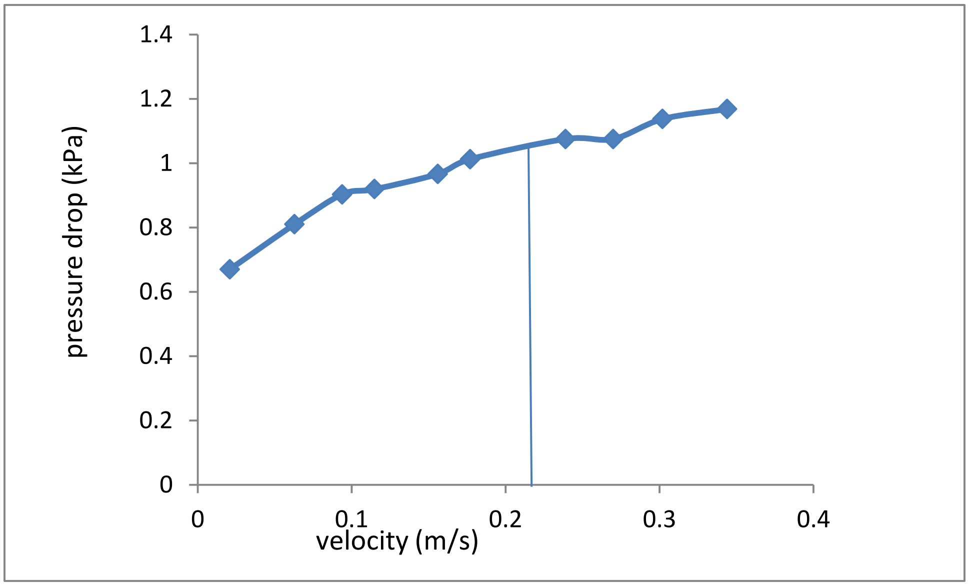
As shown in the graph the pressure drop increases from **0.249 kPa** to **1.121 kPa** with increasing superficial velocity (u) from **0.021 m/s** to **0.344 m/**s . The pressure drop remains constant when the superficial velocity (u) reaches **0.32 m/s**. After then the pressure drop does not show any effect with the increasing superficial velocity.

PLOT NO. 03

PRESSURE DROP (kPa) VS SUPERFICIAL VELOCITY,u (m/s)

STATIC BED HEIGHT :- 2 cm

MESH SIZE :- (-18 +25)



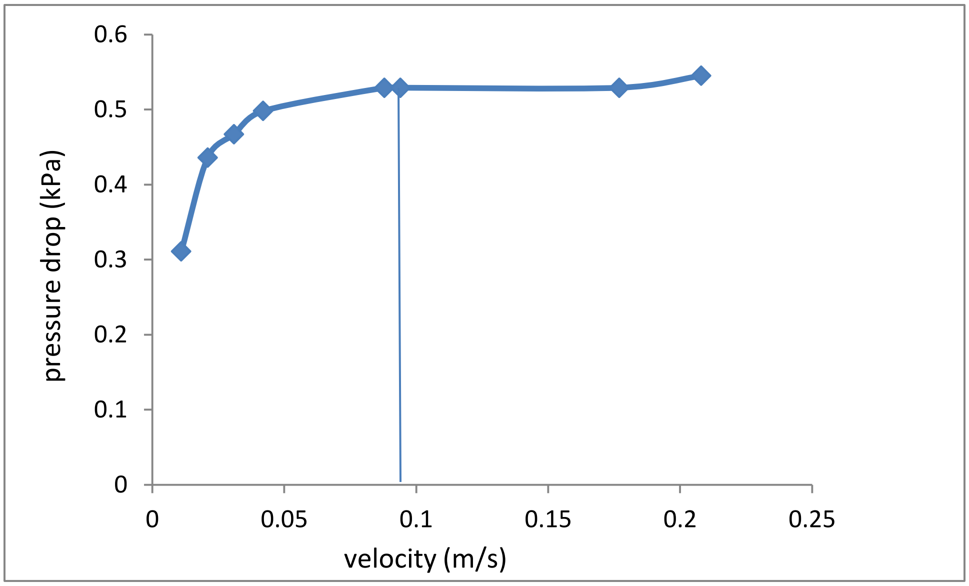
As shown in the graph the pressure drop increases from **0.670 kPa** to **1.168 kPa** corresponding to the increasing superficial velocity (u) starting from **0.021 m/s** to **0.344 m/s**. The pressure drop tends to remain constant after the velocity attains **0.22 m/s**.

PLOT NO. 04

PRESSURE DROP (kPa) VS SUPERFICIAL VELOCITY,u (m/s)

MESH SIZE (-25 +32)

STATIC BED HEIGHT :- 3 cm



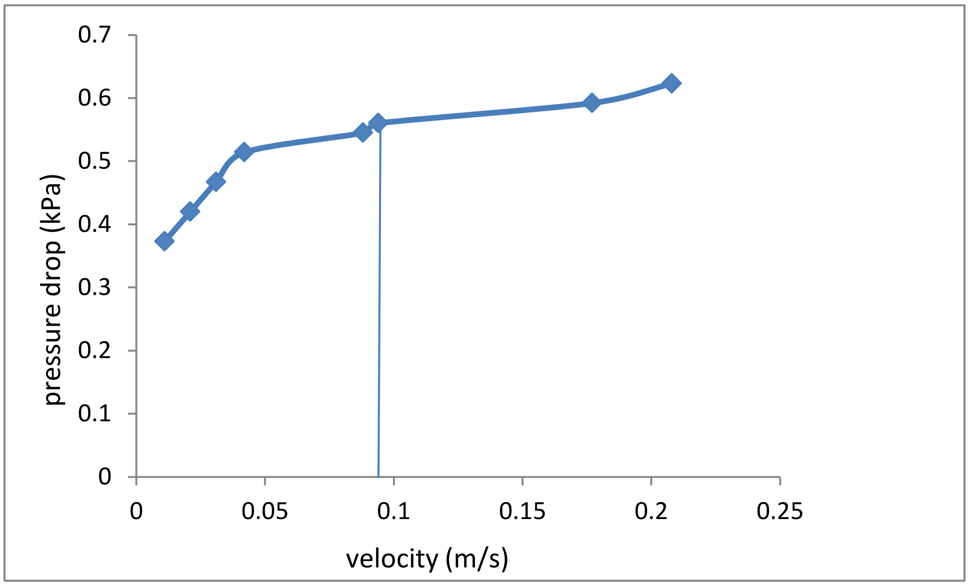
As shown in the graph the pressure drop increases with increasing superficial velocity. The pressure drop increases from **0.311 m/s** to **0.545 m/s** against the increasing velocity from **0.011 m/s** to **0.208 m/s**. The pressure drop remains constant after it attaining the superficial velocity (u) **0.08 m/s**. After then the pressure drop remains constant independent to the superficial velocity.

PLOT NO. 05

PRESSURE DROP (kPa) vs SUPERFICIAL VELOCITY,u (m/s)

MESH SIZE :- (-25 + 32)

STATIC BED HEIGHT :- 4.5 cm



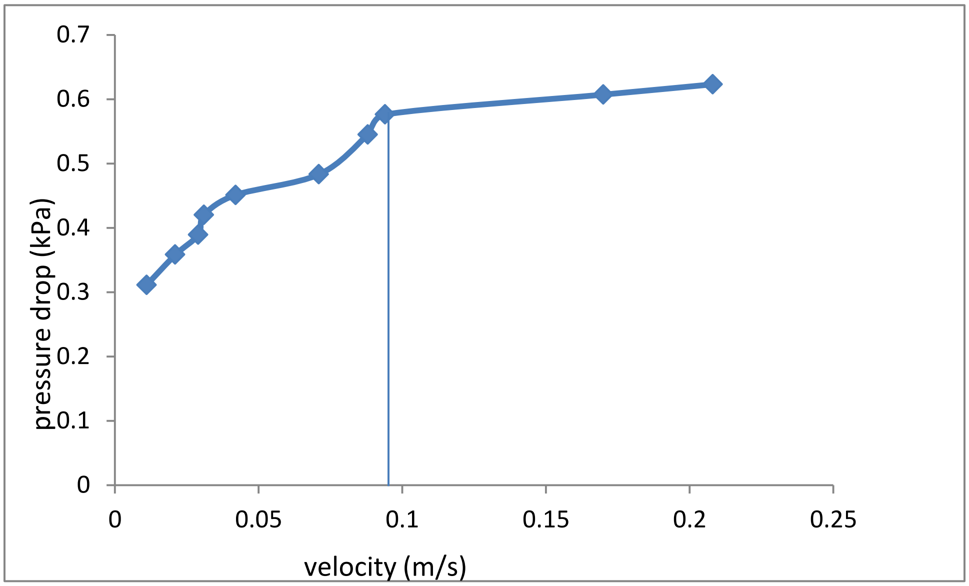
As shown in the graph the pressure drop varies proportionaly to the superficial velocity . After attaining the velocity of **0.085 m/s** the pressure drop remains constant and it independent to the increasing superficial velocity (u).

PLOT NO. 06

PRESSURE DROP (kPa) VS SUPERFICIAL VELOCITY,u (m/s) :-

MESH SIZE (-25 + 32)

BED HEIGHT 5.5 cm



As shown in the figure the pressure drop increases with the increasing superficial velocity (u). But the gas after attaining **0.08 m/s** the pressure drop remains indepe-ndent to the superficial velocity as it remains constant corresponding to the incre-asing superficial velocity.

CHAPTER - 5
RESULTS AND DISCUSSIONS

Hence it is important to relate the pressure drop with the superficial velocity considering the above graphs as shown. As the volumetric flow rate is increasing steadily , the gas velocity accordingly increases causing the pressure drop to increases simultaneously up to certain level of velocity. After then the pressure drop remains constant independent to the increasing velocity. The velocity at that point of fluidization is known as MINIMUM FLUIDIZATION VELOCITY .

CONCLUSION

For the deep beds of dense particles in tapered fluidized bed slows the axial solid mixing at gas flow rates close to the minimum primarily by reducing solids mixing in the lower part of the bed. Slower mixing was observed at gas flow rates just above the minimum .

Slugging reduces the rate at which particles entering the bed are uniformly dispersed in it. Tapering a bed that slugs increases the dispersion rate principally by increasing the mixing rate at the top of the bed.

Segragation can persist when the gas flow rate is sufficiently large to fluidize the entire bed.

Segragation is better at the large variation of the densities of the mixing material.

NOMENCLATURE

ΔH - Height difference in cm

ΔP - Pressure difference in Kilo Pascals.

kPa - Kilo Pascals (Unit of Pressure)

LPM - Litres per Minute (Unit of volumetric flowrate)

u – Superficial velocity (m/s)

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