

EXPERIMENTAL STUDY ON FLUIDIZATION OF SILICA SAND

Submitted in partial fulfillment of the requirements for the award of
Bachelor of Technology Degree in Chemical Engineering

By
VISHNUPRASAD S (39690022)
LENINLAL P (39690006)



**DEPARTMENT OF CHEMICAL ENGINEERING
SCHOOL OF BIO & CHEMICAL ENGINEERING**

SATHYABAMA

**INSTITUTE OF SCIENCE AND TECHNOLOGY
(DEEMED TO BE UNIVERSITY)**

**Accredited with Grade "A" by NAAC
JEPPIAAR NAGAR, RAJIV GANDHI SALAI, CHENNAI – 600 119**

OCTOBER 2022



SATHYABAMA

INSTITUTE OF SCIENCE AND TECHNOLOGY
(DEEMED TO BE UNIVERSITY)

Accredited "A" Grade by NAAC | 12B Status by UGC | Approved by AICTE

www.sathyabama.ac.in

DEPARTMENT OF CHEMICAL ENGINEERING

BONAFIDE CERTIFICATE

This is to certify that this Project Report is the bonafide work of **VISHNUPRASAD S (Reg. No. 39690022)** and **LENINLAL P (Reg. No. 39690006)** who carried out the project entitled "**EXPERIMENTAL STUDY ON FLUIDIZATION OF SILICA SAND**" under our supervision from July 2022 to October 2022.

Internal Guide

Dr. S SATHISH, M.E., Ph.D

Associate Professor

External Guide

Dr. G. SUNEEL, M.Tech., Ph.D

Supdt(O) WIP-UOF & VWSF

NRB, BARCF kalpakkam

Head of the Department

Dr. S. SATHISH, M.E., Ph.D

Associate Professor

HEAD OF THE DEPARTMENT
DEPARTMENT OF CHEMICAL ENGINEERING
SATHYABAMA
INSTITUTE OF SCIENCE AND TECHNOLOGY
(DEEMED TO BE UNIVERSITY)
Jeppiaar Nagar, Rajiv Gandhi Salai
Chennai - 600 119

Submitted for Viva voce Examination held on _____

Internal Examiner

External Examiner

DECLARATION

I **VISHNUPRASAD S (Reg. No.39690022)** hereby declare that the Project Report entitled **EXPERIMENTAL STUDY ON FLUIDIZATION OF SILICA SAND** done by us under the guidance of **Dr.S.SATHISH,M.E.,Ph.D.**, Associate Professor, Department of Chemical Engineering, Sathyabama University and **Dr. G. SUNEEL,M.Tech., Ph.D.**, Supdt. (O) WIP, UOF& VWSF Bhabha Atomic Research Centre kalpakkam is submitted in partial fulfillment of the requirements for the award of Bachelor of Technology degree in **Chemical Engineering**.

DATE:

PLACE:

SIGNATURE OF THE CANDIDATE

DECLARATION

I **LENINLAL P (Reg. No.39690006)** hereby declare that the Project Report entitled **EXPERIMENTAL STUDY ON FLUIDIZATION OF SILICA SAND** done by us under the guidance of **Dr.S.SATHISH,M.E.,Ph.D.**, Associate Professor, Department of Chemical Engineering, Sathyabama University and **Dr. G. SUNEEL,M.Tech.,Ph.D.**, Supdt. (O) WIP, UOF& VWSF Bhabha Atomic Research Centre kalpakkam is submitted in partial fulfillment of the requirements for the award of Bachelor of Technology degree in **Chemical Engineering**.

DATE:

PLACE:

SIGNATURE OF THE CANDIDATE

ACKNOWLEDGEMENT

We are pleased to acknowledge our sincere thanks to Board of Management of SATHYABAMA for their kind encouragement in doing this project and for completing it successfully. We are grateful to them.

We convey our thanks to **Dr. S. SATHISH, M.E., Ph.D.**, Head of the Department of Chemical Engineering for providing us necessary support and details at the right time during the progressive reviews.

We would like to express our sincere and deep sense of gratitude to our Project Guides **Dr.S SATHISH.M.E.,Ph.D.**, Associate Professor, Department of Chemical Engineering Sathyabama University and **Dr. G. SUNEEL, M.Tech., Ph.D.**, Supdt. (O) WIP, UOF & VWSF, Bhabha Atomic Research Centre, Kalpakkam for his valuable guidance, suggestions and constant encouragement paved way for the successful completion of our project work.

We wish to express our thanks to all Teaching and Non-teaching staff members of the **Department of Chemical Engineering** who were helpful in many ways for the completion of the project.

VISHNUPRASAD S

LENINLAL P

ABSTRACT

Experimental estimation of the minimal fluidization velocity is presented in this study., carryover velocity and pressure drop across the experimental set up for silica sand of different sizes (0.0003m,0.0004m,0.0005m) has been done. An experimental set up consisting of two glass columns in which one filled with silica sand of bed height 30cm has been used in this study. The medium used for fluidization was air at ambient conditions. The pressure drops and air flow rates were measured using pressure gauges at inlet and outlet of the column and rotameter respectively. The velocity (air flow rates/cross sectional area) of the air passed was slowly increased. The corresponding pressure gauge readings were noted down. At a certain velocity, the bed height of 30cm height was observed to be increasing the minimum fluidization velocity was defined as the speed at which the maximum pressure decrease was reaches. The velocity was further increased to up to the silica sand is getting carried over the next column. This velocity has been noted as carry over velocity. The experiment was repeated for above mentioned particle sizes to obtain the same parameters. Theoretical validation of the above parameters was done using four correlations to examine which correlation is more or less matching with the experimental results.

TABLE OF CONTENT

CHAPTER NO	TITLE	PAGE NO
	ABSTRACT	i
	TABLE OF CONTENTS	ii
	LIST OF TABLES	v
	LIST OF FIGURES	vi
	LIST OF ABBREVIATIONS	vii
CHAPTER 1		
1.1	INTRODUCTION.....	1
1.1.1	ADVANTAGES.....	2
1.1.2	DISADVANTAGES.....	3
1.1.3	APPLICATIONS OF FLUIDISATION.....	3
CHAPTER 2		
2.1	LITERATURE SURVEY.....	5
2.1.1	TYPES OF FLUIDISATIONS.....	5
2.1.1.1	PARTICULATE FLUIDISATION.....	5
2.1.1.2	AGGREGATE FLUIDISATION.....	5
2.1.2	BED HEIGHT.....	7
2.1.3	GELDARTS ORIGINAL POWDER CLASSIFICATION.....	7
2.1.4	MEASUREMENT OF MINIMUM FLUIDISATION VELOCITY.....	9
2.1.4.1	PRESSURE DROP VS VELOCITY.....	9
2.1.4.2	EFFECT OF PRESSURE AND TEMPERATURE ON FLUIDISATION BEHAVIOUR.....	10
2.1.4.3	MAXIMUM FLUIDISATION OR CARRY OVER VELOCITY.....	11
2.1.5	DRAW FORCE.....	12
2.1.5.1	DRAW COEFFICIENT.....	12
2.1.6	TERMINAL SETTLING VELOCITY.....	12
2.1.7	BUBBLE BEHAVIOUR.....	12
2.1.7.1	BUBBLE SIZE.....	13

2.1.7.2 FLOATING WAKE.....	13
2.1.8 FLUIDISED BED REACTORS.....	14
2.1.8.1 SPLASH ZONE.....	15
2.1.8.2 HEAT TRANSFER IN BUBBLING BED.....	15
2.1.8.2.1 FOR LARGE PARTICLES LESS TIME OF CONTACT.....	15
2.1.8.2.2 FOR SMALL PARTICLES LARGE TIME OF CONTACT.....	15
2.1.9 EQUATIONS FOR CALCULATING PRESSURE DROP.....	15
2.1.9.1 KOZENY-CARMAN EQUATION.....	15
2.1.9.2 BURKE-PLUMMER EQUATION.....	15
2.1.9.3 ERGUN EQUATION.....	16
2.1.10 MINIMUM FLUIDISATION VELOCITY CORRELATIONS.....	16
2.1.10.1 WEN AND YU.....	16
2.1.10.2 SAXENA AND VOGAL.....	16
2.1.10.3 DOICHEV AND AKHAMOV.....	16
2.1.10.4 GRACE.....	16
2.1.11 ARCHEMIDES NUMBER.....	17
CHAPTER 3	
3.1 AIM.....	18
3.2 SCOPE.....	18
CHAPTER 4	
4.1 EXPERIMENTAL STUDY METHODS AND METHODOLOGY.....	19
4.1.1 EXPERIMENTAL SET UP.....	19
4.1.2 EXPERIMENTAL PROCEEDURE.....	20
4.1.3 METHODOLOGY.....	20
4.1.4 ACTUAL SET UP.....	21
4.1.5 MATERIAL USED AND PROPERTIES.....	21
4.1.5.1 PROPERTIES.....	22
CHAPTER 5	
5.1 RESULTS AND DISCUSSIONS.....	24
5.1.1 EXPERIMENTAL OUTCOMES.....	24

5.1.2 THEORETICAL CALCULATIONS.....	29
5.1.2.1 PRESSURE DROP.....	29
5.1.2.1 MINIMUM FLUIDISATION VELOCITY.....	30
5.1.2.2.1 BY USING WEN AND YU.....	30
5.1.2.2.2 BY USING DOICHEV AND AKHAMOV.....	30
5.1.2.2.3 BY USING SAXENA AND VOGEL.....	30
5.1.2.2.4 BY GRACE.....	31
5.1.3 COMPARISON.....	32
5.1.3.1 PRESSURE DROP.....	32
5.1.3.2 MINIMUM FLUIDISATION VELOCITY.....	33
CHAPTER 6	
6.1 CONCLUSIONS.....	34
REFERENCES.....	35

LIST OF TABLES

FIGURE NO	TITLE	PAGE NO
2.1	Flow Regime During Fluidization	7
4.1	Typical Characteristics Of Silica Sand	22
4.2	Sand Composition	22
4.3	Sand Characteristics	23
5.1	Experimental Result Of Minimum Fluidization Velocity	24
5.2	Experimental Result Of Carryover Velocity	25
5.3	Experimental Result Of Pressure Drop	25
5.4	Pressure Drop Theoretical Vs Experimental	32
5.5	Minimum Fluidization Velocity Theoretical Vs Experimental	33

LIST OF FIGURES

FIGURE NO	TITLE	PAGE NO
2.1	Flow Regime	6
2.2	Geldart's Powder Classification	8
2.3	Δp Versus U_0 For Uniformly Sized Sharp Sand	9
2.4	Onset Of Fluidization	10
2.5	Voidage Profile In Various Regime	11
2.6	Drag Force	12
2.7	Bubble Shape	13
2.8	Characteristics Of Rising Bubble	14
2.9	Bubbling Fluidized Bed Reactor	14
4.1	Experimental Setup	19
5.1	Superficial Velocity Vs Bed Height (0.0003m)	26
5.2	Superficial Velocity Vs Bed Height (0.0004m)	26
5.3	Superficial Velocity Vs Bed Height (0.0005)	27
5.4	Air Flow Rate Vs Pressure Drop (0.0003m)	27
5.5	Air Flow Rate Vs Pressure Drop (0.0004m)	28
5.6	Air Flow Rate Vs Pressure Drop (0.0005m)	28
5.7	Pressure Drop Theoretical Vs Experimental	32
5.8	Minimum Fluidization Velocity Theoretical Vs Experimental	33

LIST OF ABBEREVATION

ABBEREVATION	DESCRIPTION
A	AREA OF BED
N_{Ar}	ARCHEMEDES NUMBER
N_{Remf}	REYNOLDS NUMBER AT MINIMUM FLUIDIZATION
d_p	PARTICLE DIAMETERE
U_{MF} / V_{om}^{-}	MINIMUM FLUIDISATION VELOCITY
U_{mb}	MINIMUM BUBBLING VELOCITY
U_o	SUPERFICIAL VELOCITY
U_{br}	RISING VELOCITY OF BUBBLE
U_{tr}	TERMINAL VELOCITY
L	LENGTH OF BED OF SOLID
ΔP	PRESSURE DROP
G_s	ENTRAINED FLOW OF PARTICLE
ε	BED VOIDAGE
ε_{MF}	FLUIDIZATION BED VOIDAGE
P_e	ENTRAINED PARTICLE
P_b	PARTICLE SIZE DISTRIBUTION IN BED
C_D	DRAG COFICIENT
V_b	BUBBLE VOLUME
Q_b	FLOW RATE OF FLUIDIZED BED
D_q	EQUIALENT DIAMETERE OF BUBBLE
μ	VISCOSITY
ρ_g	DENSITY OF GAS
ρ_s	DENSITY OF SOLID
ϕ	SPHERICITY

CHAPTER1

1.1 INTRODUCTION

The process of fluidization converts a solid system into a fluid, such as a suspension in a gas or liquid. Due to the peculiar properties of contacting these two-phase mixes, it is widely used in chemical industries. The capacity of fluidized beds to mix and separate materials is one of its most crucial characteristics. Due to its many advantages over alternative arrangements, this approach is one of the most popular ones, provide uniform temperature throughout the solid bed due to good solid mixing leads to better mass and heat transfer rate throughout the bed, fluidized state provides low pressure drop so pumping cost can be reduced, useful for solid handling and transport yielding large axial dispersion of phases.

Several industrial operations likely gasification, calcinations and roasting of minerals combustion, desulphurization of flue gases, chemical reactors are lies on fluidization technology due to above mentioned advantages. When a fluid is put into a bed of solid particles during the fluidization process at a low velocity, the fluid hardly bubbled upward into the void area between the solid particles. Solid particles start to move apart, few of them are in vibration and move about the restricted region, called as expanded bed. After a certain range of fluid flow rate, a point is achieved where all the particles are suspended by upward flowing fluid called bed at minimum fluidization and the velocity with which the bed fluidizes is known as minimum fluidization velocity pressure drop across the bed will increase from the region of expanded bed to the fluidized bed and it will reach maximum become constant when reaching the minimum fluidization velocity. The phenomena fluidization starts when the force exerted between the solid particles and medium used for fluidizing counter balance, he particle's actual weight, so the compressive force between the adjacent particle which act upon the vertical components will disappears, results in pressure drop (ΔP) becomes equal to weight of the fluidized particles in any the bed's portion. Fluidization will initiate at the bottom of bed and extend up to the top of the bed and the pressure drop will decline sharply. The weight of the fluid and particle in any particular portion of the bed is about equal to the pressure decrease through that area. The bed is referred to as an incipiently fluidized bed or a bed at minimum fluidization since it is only beginning to fluidize. The term "fluidization velocity" refers to any additional velocity at which the bed

fluidizes, with excess velocity denoting the difference between fluidization and minimum fluidization velocity.

Under assumptions there will be a negligible friction between the bed wall and the particles and also negligible lateral velocity of the fluid in case can be neglected, but will have a uniform vertical velocity through the bed of solid particle. Particle size upward velocity of the fluid have linear relationship, that will provide a great improvement in mixing and contact, if they matched well the particle of material will be supported by the drag force. A gas fluidized bed system may have appearance of a boiling liquid, the bubbles generated during the fluidization will be burst thereafter and also provide rapid or vigorous mixing normally in horizontal free surface. The behavior of gas solid system is different from liquid solid system in various ways. Consider a liquid solid system smooth and progressive expansion of the bed results during flow rate above minimum fluidization, gross flow instabilities are damped and remains small, and heterogeneous or large-scale void of liquid are not observed under normal conditions, such conditions result in a homogeneously or particulate fluidized bed. bed will expand vigorously due to high agitation of the solid particles so these types of system are called as heterogeneous fluidized bed or bubbling fluidized bed. commonly However, in a gas-solid system, substantial instabilities with bubbling and gas channeling are seen when the flow rate is increased the bed will expand vigorously due to high agitation of the solid particles so this system referred to as heterogeneous or bubbling fluidized bed above the point of minimum fluidization

1.1.1 ADVANTAGES

The advantages of fluidization are that the solids are agitated by when the fluid passing through the bed and obtained proper mixing of the solid irrespective of temperature gradient, for both exothermic and endothermic reactions. Fluidization system establish a smooth and liquid like flow of the particle helps for controlled automatic with ease of handling of the operations. Fluidized bed will provide rapid mixing at the same time isothermal condition throughout the reactor leads to reliable separation. The circulation of solid between two fluidized beds makes it possible to remove the vast quantities of heat produced in large reactors. The above properties of the fluidized bed help in resisting of abrupt change in temperature and operating conditions due to presence of large thermal fly wheel that produced due to well mixed solids. Properly fluidized bed will provide high heat and mass transfer rate

compare to other mode of contact, even with small surfaces areas. Fluidized beds maintain a constant temperature, considerably lower pressure drops catalysts may be added to fluidized beds continuously without affecting the hydrodynamic properties of the reactor, reduce the bed chocking and channeling due to the solid movement, helps in minimize the capital investment, ease of catalyst replacement, high conversion of reactants due to mixed flow regime.

1.1.2 DISADVANTAGES

The flow pattern of the fluidizing gas is difficult to describe due to the plug flow pattern represents inefficient contacting especially the system operating with high conversion of gaseous reactants. High upward flow rate of gaseous leads to catalyst attrition due to impact and movement of particles, entrainment and carryover of particles, size of the reactor should be increased with respect to fixed beds due to expansion of bed, catalyst fluid contact per unit volume is lower due to bed expansion, less product selectivity for complex operations and effects the driving force due to back mixing. The other disadvantages of fluidized bed such as the difficulties to describe flow of gas, with its large deviation from plug flow and the bypassing of solid by bubbles, represents an inefficient contacting system. The rapid mixing of solid in the bed provide to non-uniform residence time of solids in the reactor. Some operations handling even small particles will leads to reduction of heat of reaction due to lump formation and attrition of particles, especially in non-catalytic operation require high temperature.

1.1.3 APPLICATION OF FLUIDIZATION

Fluidization technique is widely used in industry on the basis of several advantages of fluidized bed. In the chemical industry, where better reactors, combustion, and heat transfer rates are desired, it has primarily been used as a mixing process. Particle separation can be easily achieved by this method, leads to the application in mineral separation with different physical properties. The development of fluidization technique in fluidized bed catalytic cracking (industries from metallurgical, food processing, pharmaceutical process and material processes). In contrast to slurry reactors, where solid particles are dispersed by momentum transferred from gas bubbles to liquid and from the liquid to the solids, three phase fluidized beds, where particles are supported by liquid and/or gas, have been effectively used in several process industries, such as hydro generators for the hydrogen oil production, residual oil treatment by hydro desulphurization, H-coal process for coal liquefaction

and in Fischer-Tropsch process .Chemical attrition of the particle in fluidized bed due to varying gas flow rate, chemical reaction , coalescence effect the uniform size, so the development of multi component fluidized bed, Whereas the combination of particle size, density, and shape influences fluidization behavior, proper definition of the bed dynamics for the binary and the multi component mixtures in gas solid systems is a critical pre-requisite for their effective application. . Other commonly used application of fluidized bed is flue gas desulphurization by turbulent containing absorption, fermentations, and aerobic treatment of waste water, manufacture of methanol, using in bio-oxidation, naphthalene to phthalic anhydride(catalyst) by oxidation method, petroleum industries etc.

CHAPTER 2

2.1 LITERATURE SURVEY

2.1.1 TYPES OF FLUIDIZATIONS

There are two major categories of fluidization regimes: particulate (smooth) and aggregative (bubbling) fluidization.

2.1.1.1 PARTICULATE FLUIDIZATION (LIQUID -SOLID FLUIDISATION)

When a solid is fluidized with liquid, the particles spread apart and their motion becomes more aggressive as the velocity rises, but the density throughout the bed will be uniform during fluidization. Uniform expansion of bed occurs during particulate fluidization even at high velocities. Each particle's movement is affected by the presence of other particles as solid velocities are brought into touch with the liquid and numerous particles move very closely together. As a result, it is no longer possible to model a system of numerous particles using a basic analysis of the fluid-particular behavior of a single particle. Voidage (ϵ), or the volume fraction that the fluid occupies in that situation, is taken into account. By taking into account the voidage in the system, Richardson and Zaki (1954) established the fluidized bed equation given below.

$$U_o/U_{tr} = \epsilon^n$$

where the particle Reynolds number depends on the exponent n

For $Re < 0.2$ $n = 4.65$

$0.2 < Re < 1$ $n = 4.45$

$1 < Re < 500$ $n = 4.45$

$Re > 500$ $n = 2.39$

It regulates drag, which in turn regulates the particle's terminal velocity. When there are particles of different sizes or densities, they tend to separate. While, heavier particles sink to the bottom of the bed, light particles entrain from the top. Jetsam phase refers to particles that sink to the bottom, and flotsam phase refers to particles that are floating.

2.1.1.2 AGGREGATE FLUIDIZATION (GAS -SOLID FLUIDISATION)

Bed of solid fluidized with air usually known as aggregate bubbling or boiling fluidization. This kind of fluidization happens when the densities of the fluid and solid are different. Most common refers during study of fluidization is bubbling fluidized beds. During aggregate fluidization two phases will generate dense phase known as

continuous phase and bubble phase is known as the discontinuous phase. The superficial gas velocities that create bubbles at gas fluidized beds are just actually higher than those needed to simply fluidize the bed. During fluidization bed characterized by bubble phase (like gas bubble during the boiling of water) later coalesces as they rise through the bed, and emulsion phase.

The generated bubbles have their own importance that particle movement through the bed depends on upwards movement of bubbles in the bed. Property of bubbles plays a vital role in various process occurring inside the fluidized bed reactor. Gas-solid fluidized beds can be operated in a variety of hydrodynamic flow regimes, including fixed bed pneumatic convey, homogeneous, bubbling delayed bubbling, bubbling, slugging, turbulent flow, fast fluidisation, and pneumatic transport. Because different flow regimes have varying effects on flow characteristics, mixing, heat exchange behaviour, and efficiency, it is important to distinguish between them.

Flow Regimes

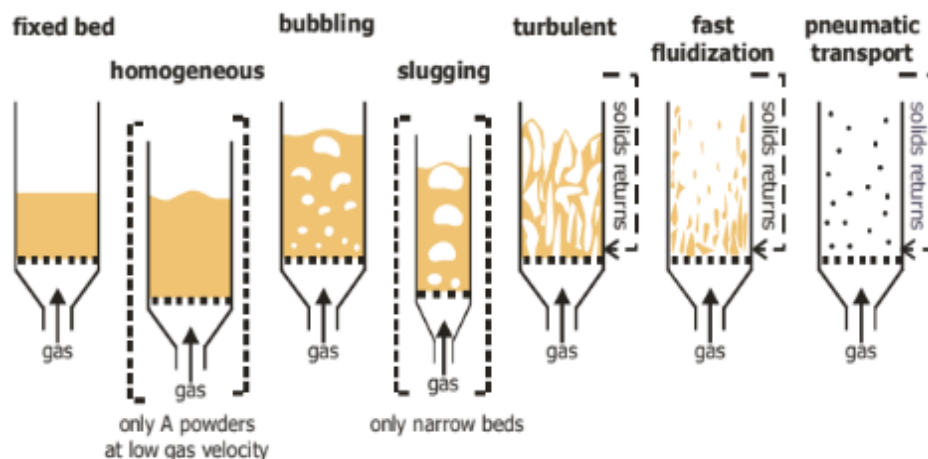


Figure 2.1 flow regime

Table 2.1 Flow regime during fluidization

Flow Regimes		
U range	Regime	Appearance and Principal Features
$0 \leq U < U_{mf}$	Fixed Bed	Particles are quiescent; gas flows through interstices
$U_{mf} \leq U < U_{mb}$	Particulate regime	Bed expands smoothly in a homogeneous manner; top surface well defined, small-scale particle motion
$U_{mb} \leq U < U_{ms}$	Bubbling regime	Gas voids form near distributor, coalesce and grow; rise to surface and break through
$U_{ms} \leq U < U_c$	Slug flow regime	Bubble size approaches column cross-section. Top surface rises and collapses with regular frequency.
$U_c \leq U < U_k$	Turbulent fluidization flow regime	Pressure fluctuations gradually decrease until turbulent fluidization flow regime is reached.
$U_k \leq U < U_{tr}$	(Turbulent Regime)	Small gas voids and particle clusters dart to and fro. Top surface is difficult to distinguish.
$U \geq U_{tr}$	Fast Fluidization	No upper surface to bed, particles transported out the top in clusters and must be replaced.
$U \gg U_{tr}$	Pneumatic conveying	No bed. All particles fed in are transported out the top as a lean phase.

2.1.2 BED HEIGHT

The bed height should be calculated in these investigations as a function of the U_0 . As bubbles burst through the surface at faster velocities, the surface will fluctuate quickly, making an exact measurement impossible. The passage of bubbles through the bed is what causes the bed to expand to a greater extent than that observed at U_{MF} .

2.1.3 GELDART'S ORIGINAL POWDER CLASSIFICATION

- Fluidization cannot be achieved for all particles. Solid particles' behavior in fluidized beds is mostly influenced by their size and density. Geldart made a detailed inspection of this phenomena, shown in Figure together with the traits of the four studies
- Following are some categories for various powder type
- Particles in Group A are classified as "aeratable." These substances have relatively Low particle density (1.4 g/cm^3) and/or the average particle size ($d_p < 30 \text{ }\mu\text{m}$). Catalysts for fluid cracking typically fall within this group. These solids fluidize effortlessly, smoothly fluidizing at slow gas velocities without the formation bubbles. U_{mb} is the greater than U_{MF} .

- Group B is known as "sand like" particles, though some people refer to it as bubbly particles. Most of particles with D_p 50 μm to 500 μm and from ρ_s 1.4 to 4 g/cm^3 . The excess gas appears as bubbles (various size depends on gas velocity) after U_{MF} , for these particles is reached. Group B components include gritty sand and glass beads
- Materials in Group C are commonly "cohesive," or very small powder. Diameters of these solids are $< 30 \mu\text{m}$ results in large interparticle force compared to other substances, fluidization is very difficult with these particles due chance of channeling in the bed, example starch, flour, and talc etc.
- Materials in Group D are called as "spoutable" and are large in size with high density. Deep beds of these substances make them harder to fluidize. In contrast to group B particles, a jet can form in the bed as gas velocity rises, and material may subsequently be blown out with the jet in a forceful motion, Spouting behavior and If there is an imbalance in the gas supply, channeling is expected. Examples of group D materials include roasting metal ores, lead shot,

Geldart's powder classification

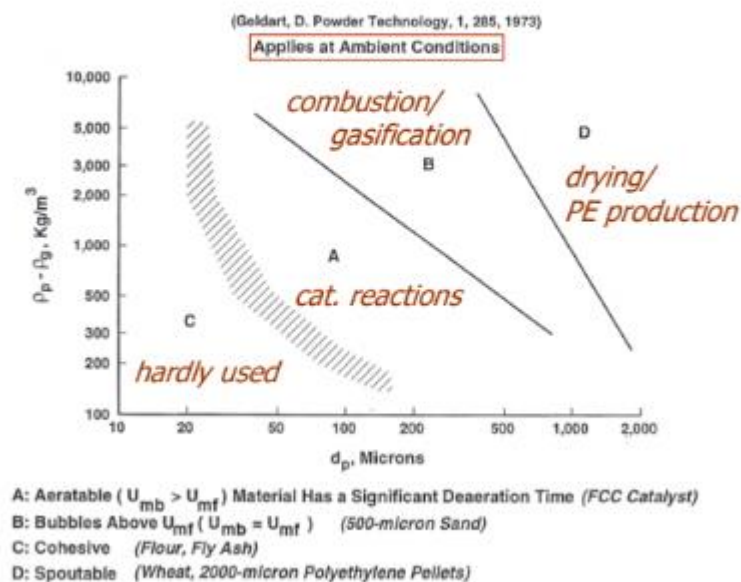


Figure 2.2 Geldart's powder classification

2.1.4 MEASUREMENT OF MINIMUM FLUIDISATION VELOCITY ($U_{MF}/\bar{V}_{om}$)

U_{MF} is the velocity (U_o) at which the solid bed is just fluidized it can be found out at ambient condition directly. In order to find out the practical value of U_{MF} , fill the container with enough particles to create a bed of solids, initially there will be an interlock or weak bonding between the particles these can be broken down by fluidization of bed for a few minutes. Pressure drop across the bed at the fluidizing velocity can be measured. Show the result in the form of bed pressure drop against superficial gas velocity. there is some deviation can be obtained from practical studies on comparison with theoretical studies, values shows that pressure drop obtained is less than bed weight per area this because some of the effect of friction between the solid particles and column wall effects the fluidization due the container wall is bearing a portion of the bed's load because of particle/wall contact

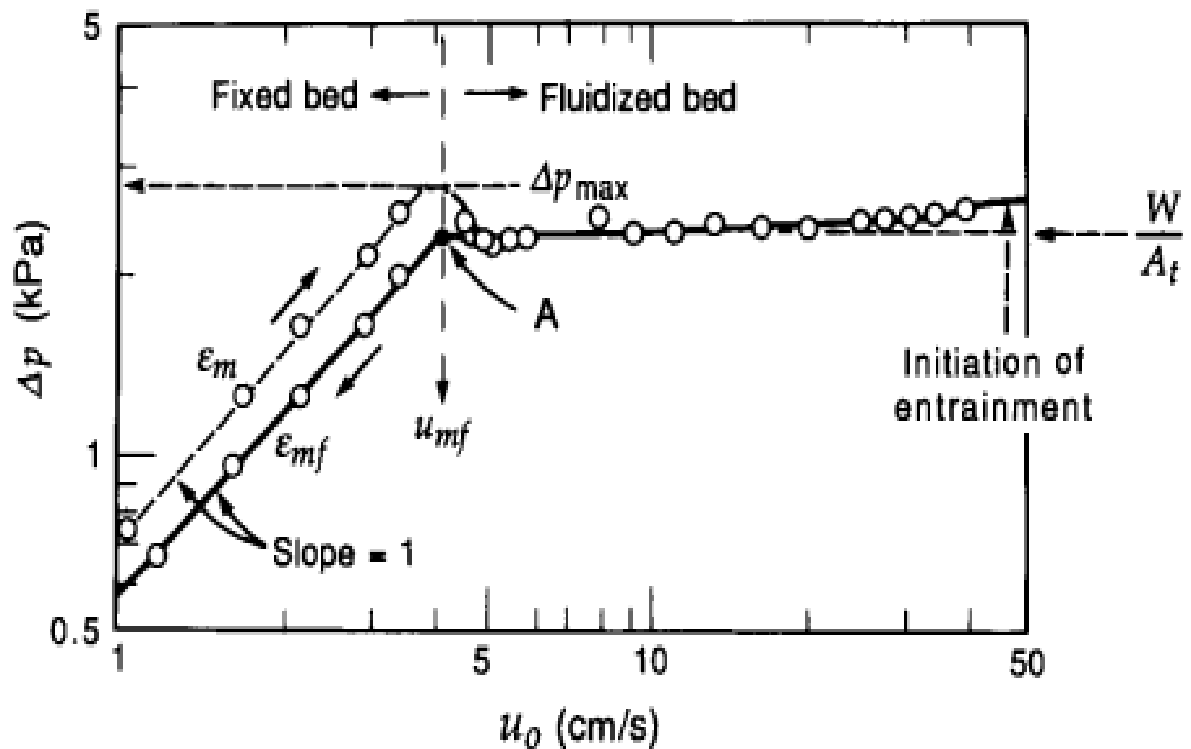


Figure 2.3 ΔP versus U_o for uniformly sized sharp sand gives

2.1.4.1 PRESSURE DROP VS VELOCITY

The ΔP throughout the bed should rise linearly with an increase in U_o from zero until the bed reaches the condition of fluidization due to laminar gas flow rate. At higher gas flow rate, results in unlocking of the particle in the bed leads to increment of void age from ϵ_m to ϵ_{mf} . Figure illustrate the against typical plot of bed pressure drop

against superficial gas velocity, point where rising and flatter portion of the curve coincide is U_{MF} for uniformly sized sand. Sharp increase ΔP can be observed in if the bed is formed by large particle during the this is because the flow pattern shifts from laminar to turbulent. We repeat the experiment in reverse direction results in tapping of the bed. The bed will become more compressed and static when you tap on it while gas is flowing through it. This situation results from the bed voidage returning to ϵ_m . Additionally, when taking progressive measurements from a particle bed, the pressure drop may initially be greater than the weight of the bed as fluidization starts, because more drag force is needed to break up particle interconnection.

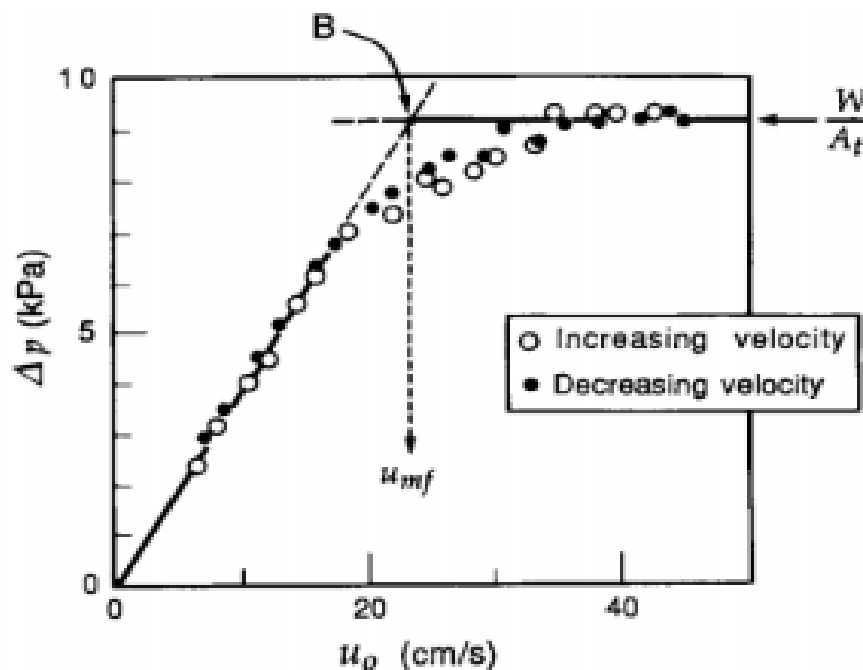


Figure 2.4 Onset of fluidization.

2.1.4.2 EFFECT OF PRESSURE AND TEMPRRATE ON FLUIDIZATION BEHAVIOUR

Numerous researchers have researched the impact of pressure, and the following sentences summaries their findings for beds of evenly sized glass beads, coal, and porous carbon powder at pressures up to 80 bar.

Operating pressure increases with a minor (1-4%) increase in ϵ_{mf} and a corresponding decrease in U_{MF} . However, this reduction becomes considerable (up to 40%) for bigger particles ($d_p = 360\mu m$) but is negligible for beds of tiny particles

($d_p < 100\mu\text{m}$). These experimental results are in line with what was predicted. At all pressures, U_{MF} may generally be accurately anticipated.

- U_{mb}/U_{MF} for coarse alumina ($d_p = 450\mu\text{m}$) increases up to 30% for a rise in operating pressure. This suggests that an increase in operating pressure widens the range of particulate fluidization in gas-solid systems.
- While ϵ_{mf} is unaffected by temperature for coarse particles, it does increase with temperature for fine particles (up to 8% for temperatures up to 500°C).

2.1.4.3 MAXIMUM FLUIDIZATION OR CARRYOVER VELOCITY

At higher gas velocity drag force become greater than that of force exerted by gravity across the individual particles and dragging it away from the bed or entrained in the fluid, called as maximum fluidization velocity. The gas leaving vessels with fluidized materials carries some fine material. The bulk density of solids in this exiting gas stream is known as the holdup, and the flow of solids is known as entrainment, G_s ($\text{kg}/\text{m}^2 \text{ s}$), or carryover. it is necessary to understand the rate of entrainment, the size distribution of the entrained particles, $P_e(d_p)$, in relation to the size distribution in the bed, $P_b(d_p)$ also the relationship between these two variables and the properties of the of the two phases, the gas flow rate, the configuration of the bed, and the position of the leaving gas stream. The bed voidage-versus-velocity diagram was provided by Avidan and Yerushalmi

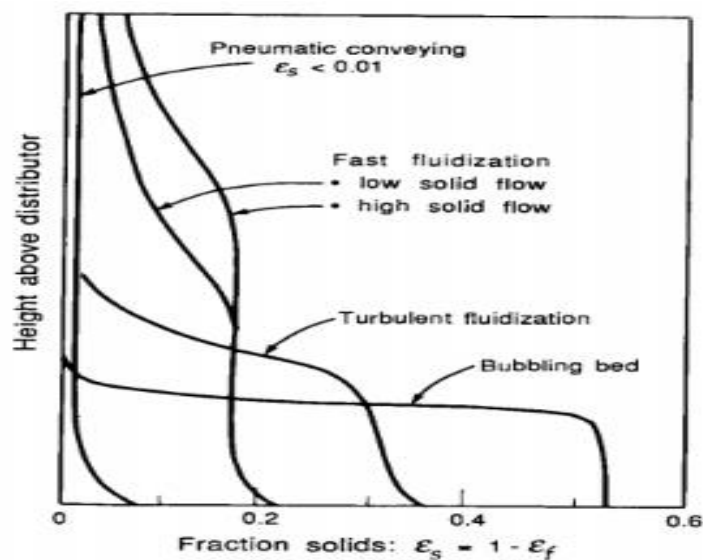


Figure 2.5 Voidage profile in various regime

2.1.5 DRAG FORCE

Drag force is the force that a fluid produces to a solid in the flow path.

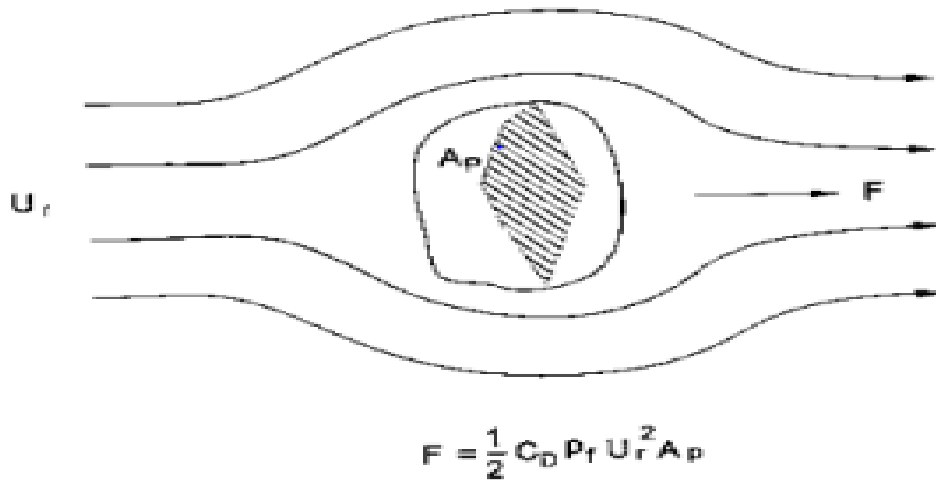


Figure 2.6 Drag force

2.1.5.1 DRAG COEFFICIENT

$$C_D = \frac{F}{(1/2) \rho_f U_f^2 A_p}$$

2.1.6 TERMINAL SETTLING VELOCITY

Terminal settling velocity is the maximum velocity of an object can reach when the sum of the drag force (F_d) and the buoyancy force equal the downward force of gravity (F_g) acting on the object. Since the net force on the object is become zero, the acceleration of the object is zero.

$$U_t = \left[\frac{4d_p (\rho_s - \rho_g) g}{3\rho_g C_D} \right]^{\frac{1}{2}}$$

The ratio U_t / U_{MF} depends on particle size: It is mentioned below

For spherical particle of one size and bed voidage = 0.4

For fine solids: ratio = 78 $dp^* < 1$

For large solids: ratio = 9.2 $dp^* > 100$

2.1.7 BUBBLE BEHAVIOUR

The bubbles are hemispherical in shape and pushed-in bottom may be seen. The area immediately beneath the bubble is referred to as the wake and contains a sizable number of solids. Additionally, each bubble has a cloud (solid content is

higher than the bubble's but lower than the emulsion's) around it that is made up of emulsion.

1. The bubble shapes are almost in similar manner; when tiny, they are nearly spherical, when bigger, they are flattened and distorted, and when very big, they are spherical with a cap
2. Large bubbles rise quickly in both systems, but little bubbles rise slowly.
3. A series of bubbles in any system may combine to produce broader bubbles. A series contact produces a distinct increase velocity, and in both situations the change's path is the same.
4. Effect of coalesce with wall is linear to the direction of bubbles' upward velocity.

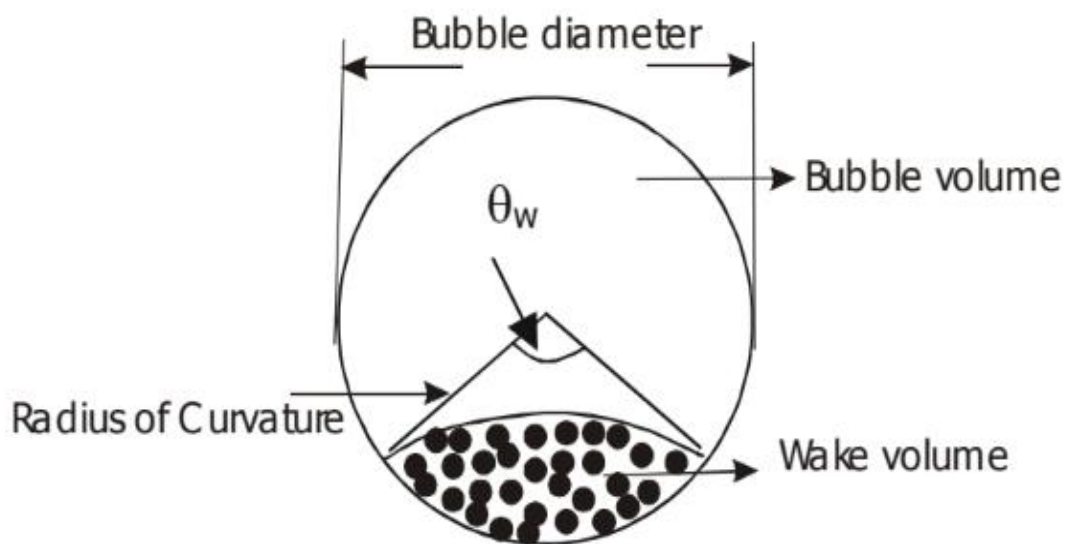


Figure 2.7 Bubble shape

2.1.7.1 BUBBLE SIZE

Due to bubble coalescence, as the distance from distributor plate mean diameter and volume, of the bubble will increase. D_q , the bubble size, can be calculated as

$$D_q = (6V_b / \pi)$$

The visible flow rate in a fluidized bed Q_b ,

$$Q_b / A = U_0 - U_{MF}$$

2.1.7.2 FLOATING WAKE

Wake is the amount of solid carried by the upward flow of bubble, The moment the bubble starts to develop, a wake immediately follows. As a result, as the bubble exits the distribution plate, it gathers up the most of the grains are at the bed's base. An idealized bubble has a bottom wake with an angle as illustrated Figure 1.8

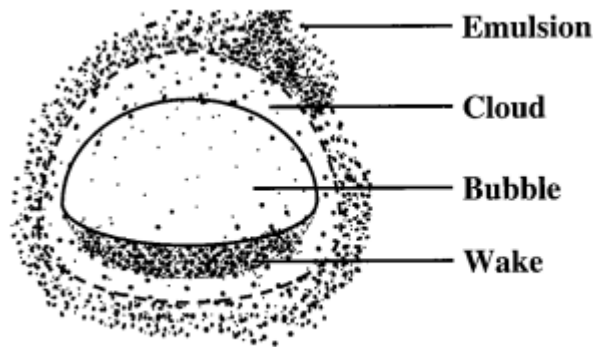


Figure 2.8 Characteristics of rising bubble

2.1.8 FLUIDIZED BED REACTORS

The most often used configurations, both in laboratories and on a commercial scale, are fluidized bed reactors, widely used most of the process industries and more successful when bio oil and feedstock are continuously produced. Additionally, a thermal carrier made of tiny inorganic particles is used in fluidized bed reactors to provide high heat-transfer rates and consistent bed temperatures. The raw material particles are injected into the fluidized bed at the bottom of the reactor by adding pressurized air, which then passes through the heated inorganic particles. Due to the pressurized air driving both the feed and inorganic particles to fluidize, the feedstock particles flow vertically through the reactor. High heat transfer rates are made possible by the nearest area.

The U_{MF} is still one of the key design considerations, despite the fact that commercial fluidized beds are frequently operated at high U_O/U_{MF} ratios. Most studies on the lowest fluidization velocity to date have concentrated on fluidization at room temperature and fluidization with common gases like nitrogen and air. On the other hand, the results on high temperature fluidization are confusing one.

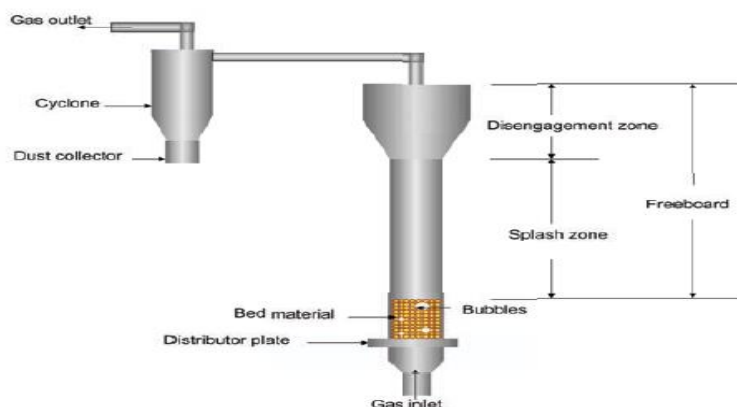


Figure 2.9 Bubbling fluidized bed gasifier

2.1.8.1 SPLASH ZONE

The overall heat transfer rate is frequently modified and managed in fluidized combustors with horizontal tube banks by raising or decreasing the bed level rather than by altering the bed temperature. This alters the percentage of tubes buried in the mattress. The difference determines how well this process performs in (Heat transfer coefficient) between tubes in the splash zone above the bed and those submerged in it. Therefore, one needs to be aware of the h values in these zones. bed surface. radiation contribution to heat transfer rises rapidly with temperature.

2.1.8.2 HEAT TRANSFER IN BUBBLING BED

The rate of heat distribution in these reactors can be explained in different ways, for the circumstances of the studies described there, these emulsion particle packets have a contact time with the surface of 0.2–0.4s. More generally, the experimental setup and the position of the heat exchange surface affect this contact time.

2.1.8.2.1 FOR LARGE PARTICLES, LESS TIME OF CONTACT

The gradient in temperature of the particles involved, occurs only in the row of particles in direct contact with the exchanger surface, and we may ignore the thermal diffusion into the rest of the emulsion packet. This is because the particles must be replaced before their mean temperature can vary noticeably.

2.1.8.2.2 FOR SMALL PARTICLE, LARGE TIME OF CONTACT

Here the particles near the surface closely approach the surface temperature, the thermal transient is felt many layers from the surface, and, hence, thermal diffusion into the emulsion packet becomes the controlling resistance

2.1.9 EQUATIONS FOR CALCULATING PRESSURE DROP

2.1.9.1 Kozeny-Carman equation

It corresponds to solid-bed flows at Reynolds numbers up to roughly 1.0.

$$\frac{\Delta p}{L} = \frac{150 \bar{V}_0 \mu (1 - \varepsilon)^2}{\Phi_s^2 D_p^2 \varepsilon^3}$$

2.1.9.2 Burke-Plummer equation

An empirical at Reynolds number >1000 for pressure drop in packed beds

$$\frac{\Delta p}{L} = \frac{1.75 \rho \bar{V}_0^2 (1 - \varepsilon)}{\Phi_s D_p \varepsilon^3}$$

2.1.9.3 Ergun equation

An equation covering entire range of flow rates

$$\frac{\Delta p}{L} = \frac{150 \bar{V}_0 \mu (1 - \varepsilon)^2}{\Phi_s^2 D_p^2 \varepsilon^3} + \frac{1.75 \rho \bar{V}_0^2 (1 - \varepsilon)}{\Phi_s D_p \varepsilon^3}$$

Ergun equation is the combination of the Blake–Kozeny equation and the Burke–Plummer equation define the limiting flow regimes for a fluid passing through a mass of spherical particles.

2.1.10 MINIMUM FLUIDISATION VELOCITY CORRELATIONS

2.1.10.1 WEN & YU

The approach of Wen and Yu involves omitting particle sphericity and bed voidage from estimations of U_{MF} at the same time. In fluidization investigations, "voidage functions" are derived by experimentally so it will minimize the study because ε_{mf} is itself a result of velocity.

$$\frac{(1 - \varepsilon_{mf})}{\phi_p^2 \varepsilon_{mf}^3} \approx 11 \quad \text{and} \quad \frac{1}{\phi_p \varepsilon_{mf}^3} \approx 14.$$

Modification of above equation is suitable for very small particle, called Wen & Yu equation

$$Re_{mf} = -33.7 + \sqrt{33.7^2 + 0.0408 Ar}.$$

$$\bar{V}_{om} = \mu / \rho_g D_p * (\sqrt{(33.7^2 + 0.0408 Ar)} - 33.7)$$

2.1.10.2 SAXENA AND VOGAL

$$\bar{V}_{om} = \mu / \rho_g D_p * (\sqrt{(25.28^2 + 0.0571 Ar)} - 25.28)$$

2.1.10.3 DOICHEV & AKHAMOV

$$\bar{V}_{om} = \mu / \rho_g D_p * (1.08 * 10^{-3} * Ar^{0.947})$$

2.1.10.4 GRACE

$$\bar{V}_{om} = \mu / \rho_g D_p * (\sqrt{(27.2^2 + 0.0408 Ar)} - 27.2)$$

2.1.11 ARCHIMEDES NUMBER

The dimensionless number known as the Archimedes number (Ar) analyses flow in respect to a system of differences in density. When dealt with the gravitational settling of fluid-filled particles, it is used.

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

CHAPTER3

3.1 AIM

The aim of the project is to collect the basic data to measure Minimum Fluidization velocity and carryover velocity of various types of resins by experimentation using an experimental setup, comparing values with those of correlations available in literature and calculating fluidization & carryover flow rates for the silica bed.

3.2 SCOPE

The project is meant to estimate certain design parameters involved in Denitrification plant. Fluidised bed thermal denitration is considered as a suitable chemical-free disposal option for the aqueous waste nitrate stream. A Denitration Plant (DNP) is to be set up for thermal decomposition of the Ammonium nitrate solution into nitrogen / NO_x and water. The DNP will be based on fluidized bed thermal denitration process technology. The impurities in the solution will get trapped onto the grains of the inert silica seed material which is used as the working powder in the fluidized bed denitration reactor. The objective of the plant is to meet the environmental discharge limits of nitrates and other impurities

CHAPTER4

4.1 EXPERIMENTAL STUDIES METHODS & METHODOLOGY

4.1.1 EXPERIMENTAL SET UP

The experimentation setup consists of a glass column which is cylindrical in shape, length 60cm, outer diameter of column is 10 cm and length and thickness 0.5 there is an inlet at the bottom for introducing water for causing bed fluidization and carryover. Actual silica sand in plant is taken and compressed air is used for the fluidization experiment. Pressure difference across the column is measured by the help of pressure gauge provided at inlet and outlet. The flow rate is measured by using rotameter at the inlet.

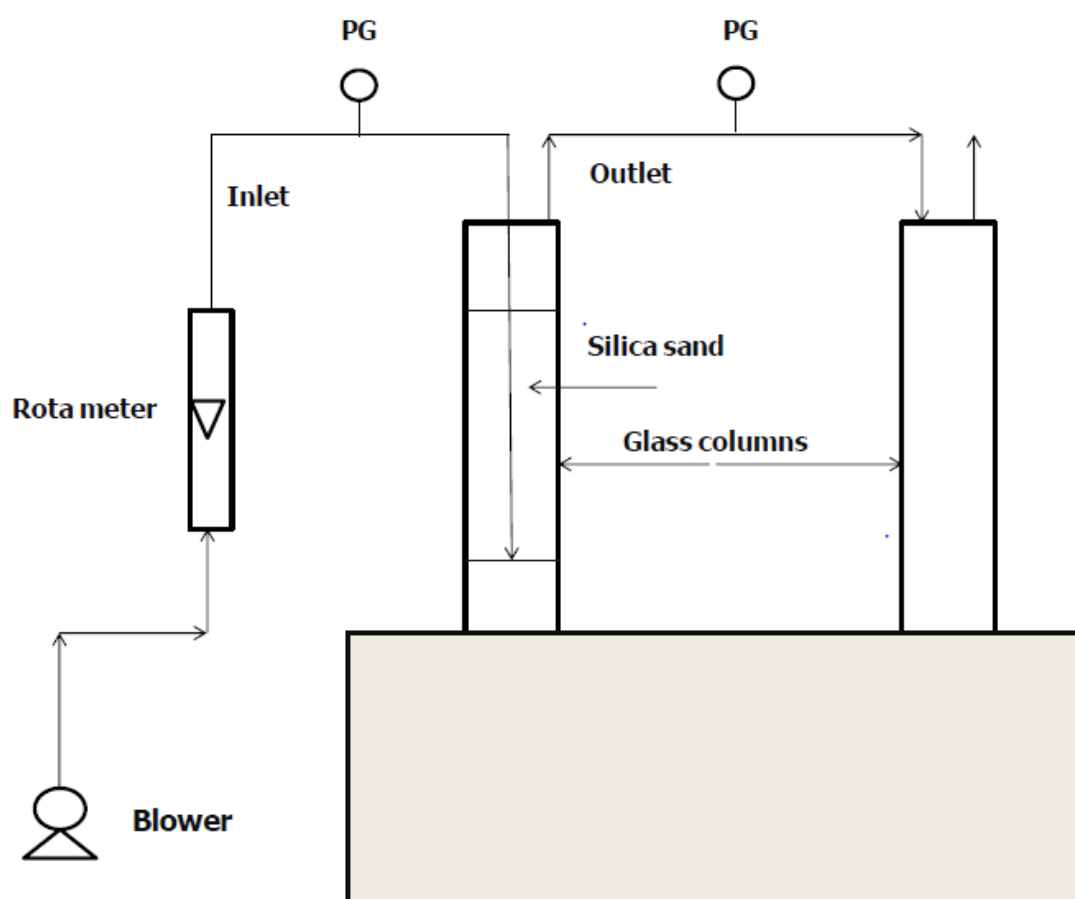


Figure 4.1 Experimental setup

4.1.2 EXPERIMENTAL PROCEDURE

- Charge sufficient quantity (bed volume) of silica sand in to the column to create a bed, then interlocking of particle to be removed by rapid fluidization for few minutes.
- interlock if exist.
- For finding out the ΔP throughout the bed, experiment to be start with Start with modest gas velocities.
- And plot the outcomes as a U_o against ΔP
- We can observe the increment in ΔP with respect to increase gas velocities.
- ΔP corresponding to the weight of the bed's weight per unit area, however, it might be smaller in practise due to the bed's weight partially being held by the column wall through particle wall interactions.
- When there is low gas flow, the ΔP across the bed should rise until fluidization is nearly attained.
- Note down the velocity at which the bed gets fluidised
- Note down the corresponding bed volume
- Increase the air flow rate gradually and note down the change in bed volume
- Increase the gas velocity until the point at which the bed is getting carried over to the next column
- Note down the corresponding velocity as carry over velocity
- Repeat the experiment with different particle sizes

4.1.3 METHODOLOGY

- Experimental estimation of minimum fluidisation velocity for silica sand bed and pressure drops across it by varying the air flow rates
- Repeating the experiment for different particle sizes
- Theoretical calculation of minimum fluidisation velocity and pressure drop using the available data
- Comparison of theoretical results with the experimental results

4.1.4 ACTUAL SETUP



Figure 4.2 Experimental setup



Figure 4.3 Experimental setup

4.1.5 MATERIAL USED AND PROPERTIES

Silica (commercial) sands are utilized for purposes other than as construction aggregates because they contain a high concentration of silica (often, but not always, more than 95% SiO_2). They come from both loosely by crushing condensed sand deposits Sandstones with poor cement. In contrast to building sands, which are

utilized for For its own properties, silica sands are valued at a physical and chemical qualities combined. These contain a significant amount of silica in the quartz-like substance and, more significantly, incredibly low amounts of harmful contaminants, especially iron oxides, refractory materials like clay, and like chromite. They usually have a limited grain size dispersion). For the majority of uses, silica sands have to be consistent with extremely well-established specifications, and consistency is of critical importance.

4.1.5.1 PROPERTIES

Table 4.1 Typical Characteristics of Silica Sand

S.NO	Characteristics	Silica sand
1	Specific gravity	2.2-2.5
2	Bulk density	2.4-2.6 g/cc
3	Moisture content in the cake	.01%
4	Sizes	.0003,.0004,.0005

Table 4.2 Sand Composition

S.NO	COMPONENT	PERCENTAGE
1	Silica as SiO ₂	97.48
2	Magnesium as MgCO ₃	0.63
3	Iron as Fe ₂ O ₃	1.14
4	Calcium as CaO	0.14
5	Alumina as Al ₂ O ₃	0.2

Table 4.3 Sand Characteristics

S.NO	Sand characteristics	Magnitude
1	Appearance	White crystalline
2	Chemical purity	Silica > 95% SiO ₂
3	Hardness	7-8 moh
4	Strength	25 kg/cm ²

CHAPTER5

5.1 RESULTS AND DISCUSSIONS

5.1.1 EXPERIMENTAL OUTCOMES

Three experimental runs for different particle sizes (300,400,500microns) were conducted as the air flow rate increased from zero, the pressure drop across the column started to increase with the same. At a certain flow rate, the pressure drop across the bed became steady and the column bed height started to increase. This was noted as the point of fluidisation. This condition shows the point at which the gas velocity becomes high enough that the drag force becomes equal to the weight of the particle

Experimentally obtained results of minimum fluidisation velocity (V_{om}), pressure drop across the bed(p), carry over velocity were tabulated as below,

Table 5.1 Experimental results of minimum fluidization velocity

MATERIAL	DIAMETER(m)	MINIMUM FLUIDIZATION VELOCITY(m/s)
SILICA SAND	0.0003	0.0891
	0.0004	0.155
	0.0005	0.246

Table 5.2 Experimental results of carry over velocity

MATERIAL	DIAMETER (m)	CARRY OVER VELOCITY(m/s)
SILICA SAND	0.0003	0.160
	0.0004	0.283
	0.0005	0.376

Table 5.3 Experimental results of pressure drop

MATERIAL	DIAMETER (M)	VELOCITY (M/SEC)	PRESSURE DROP (Kg/cm ²)
SILICA SAND	0.0003	0.445	0.021
	0.0004	0.077	0.036
	0.0005	0.123	0.055

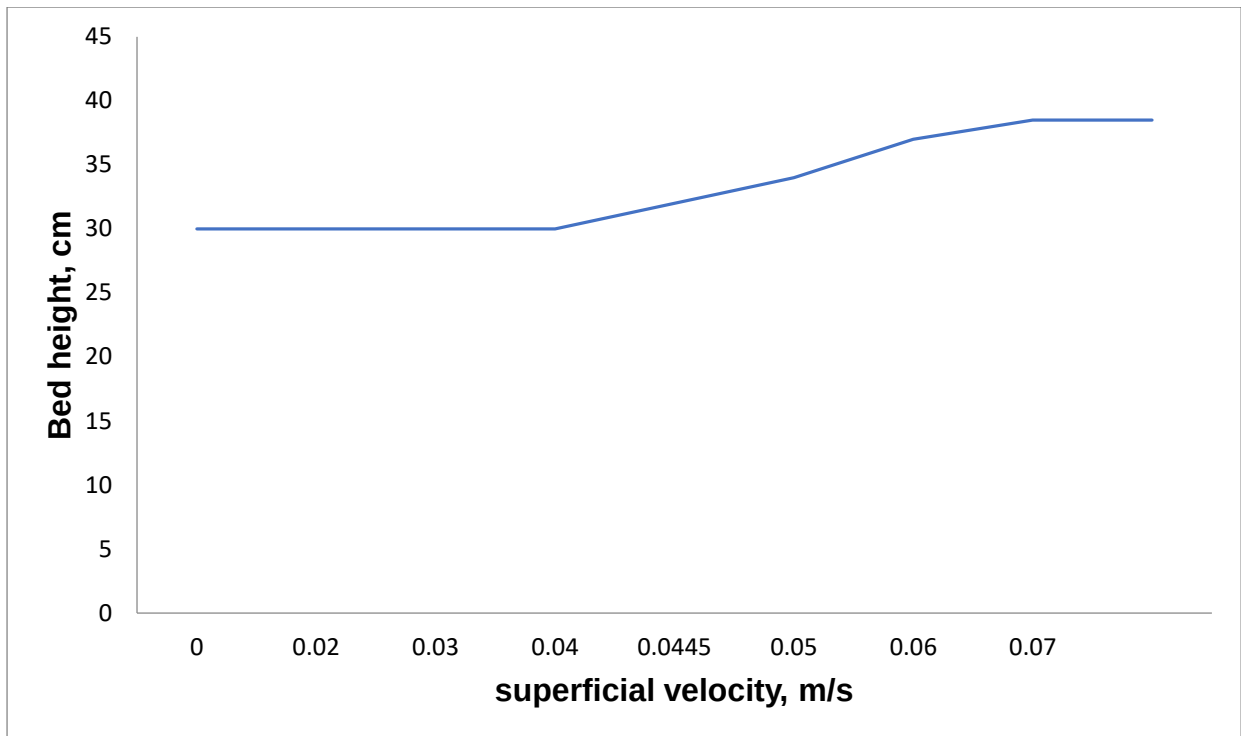


Figure 5.1 Superficial velocity vs Bed height (0.0003m)

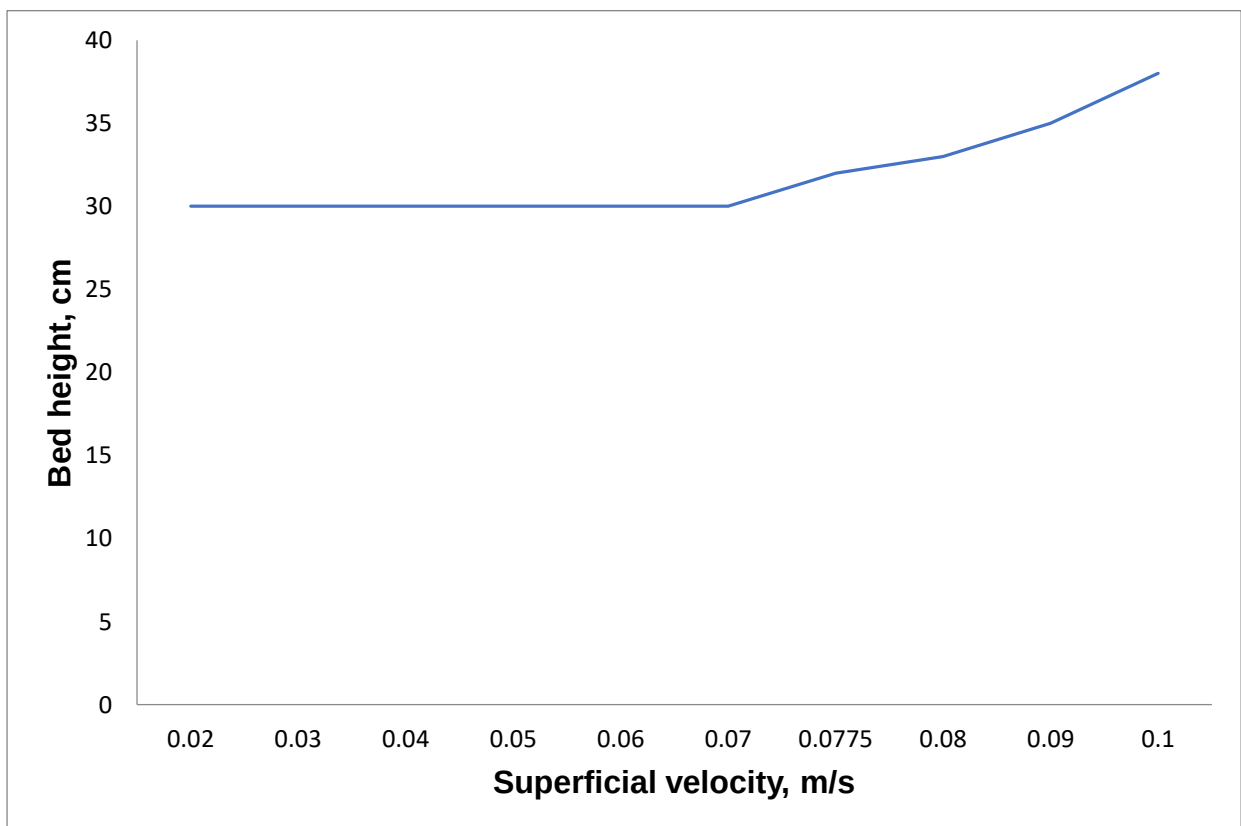


Figure 5.2 Superficial velocity vs Bed height (0.0004m)

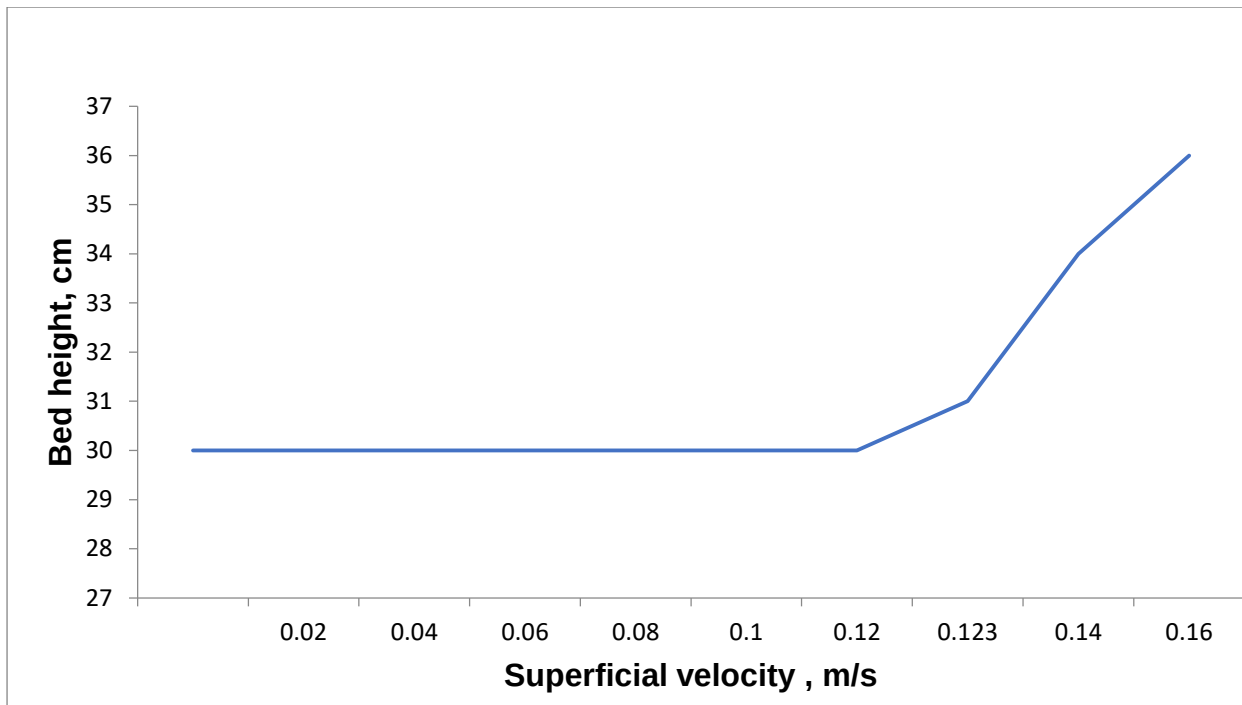


Figure 5.3 Superficial velocity vs bed height (0.0005m)

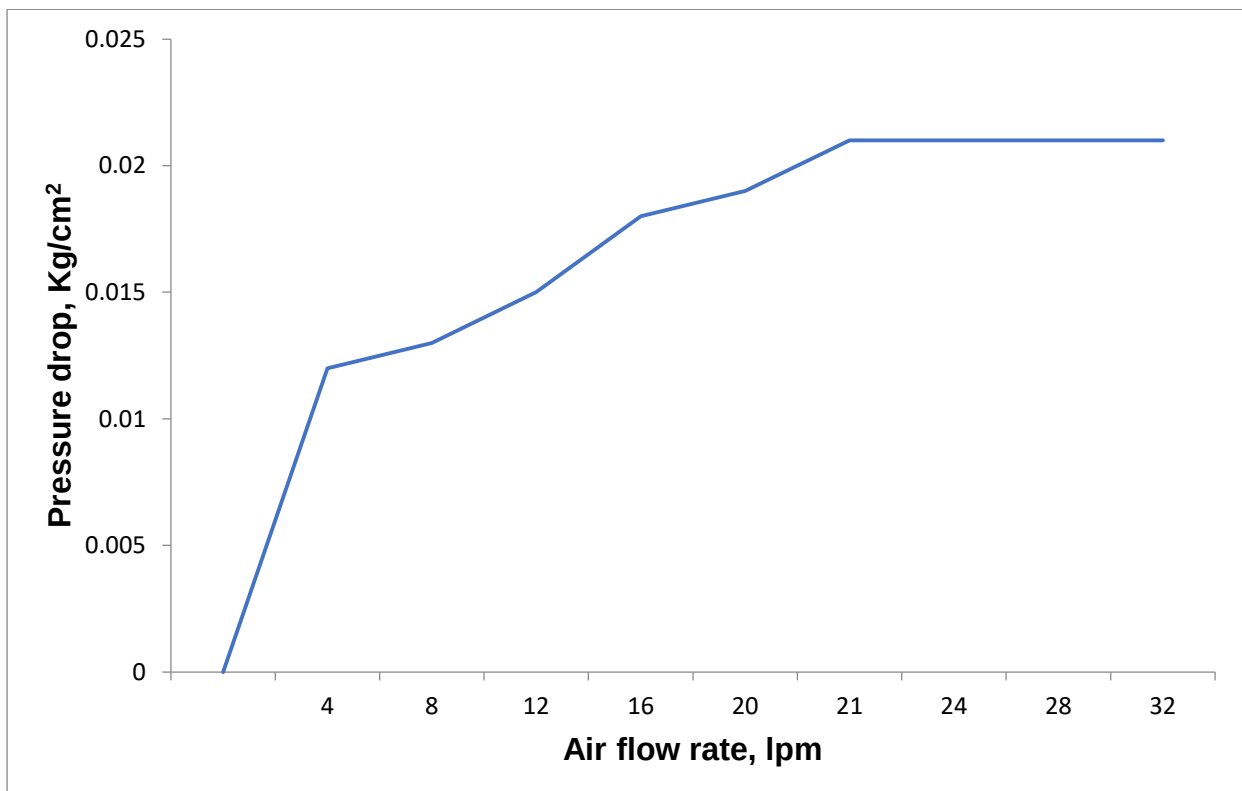


Figure 5.4 Air flow rate vs Pressure drop(0.0003m)

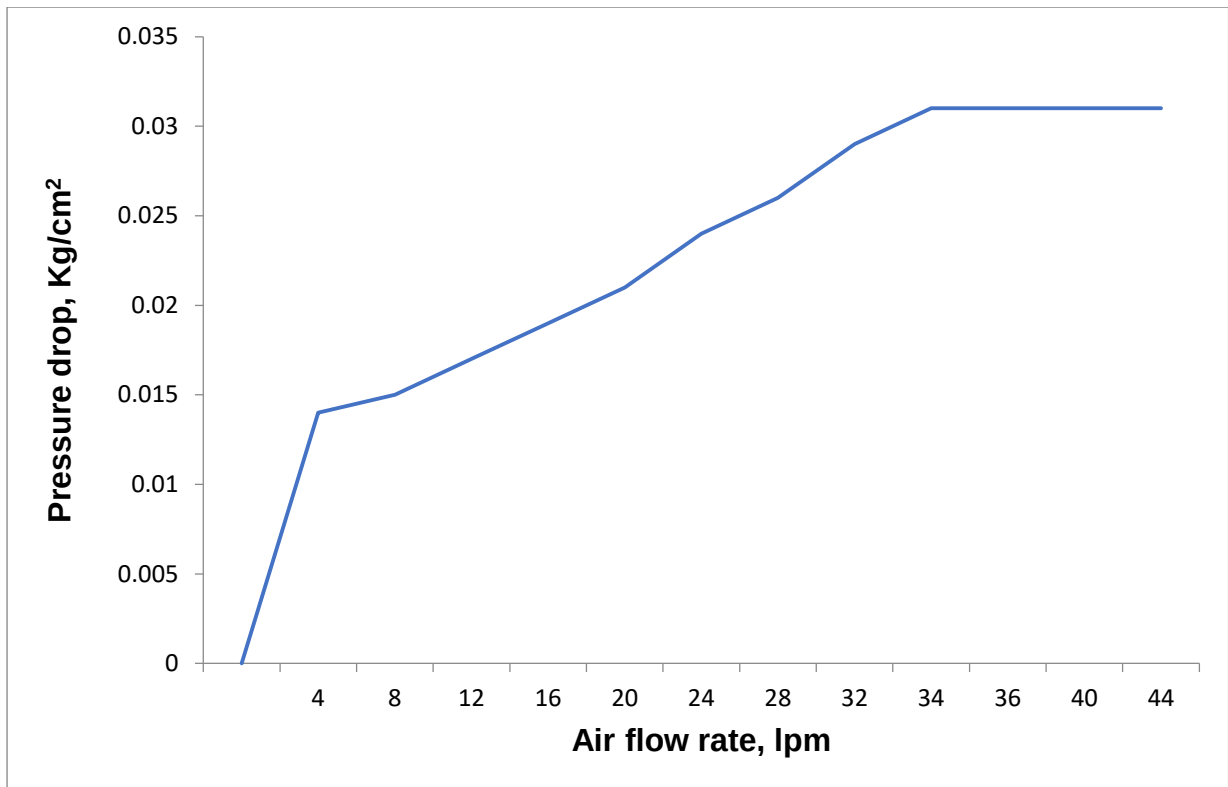


Figure 5.5 Air flow rate vs Pressure drop (0.0004m)

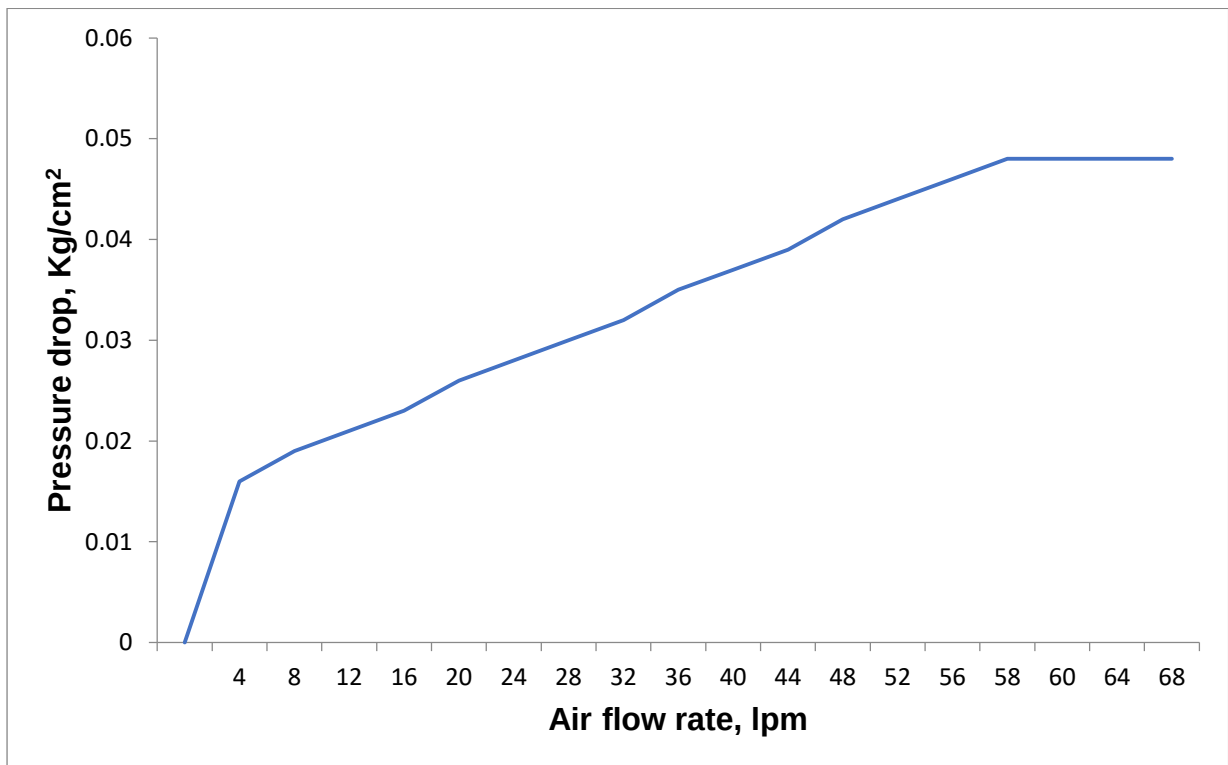


Figure 5.6 Air flow rate vs Pressure drop (0.0005m)

5.1.2 THEORETICAL CALCULATIONS

Volumetric flow rate, $Q = 21 \text{ lpm} = 0.00035 \text{ m}^3/\text{s}$

Velocity, $V_o = Q/A$

Area = $7.85 \times 10^{-3} \text{ m}^2$ (Día- 0.1m, Length- 0.6m)

$V_o = 0.00035 / (7.85 \times 10^{-3}) = 0.0445 \text{ m/s}$

$N_{re} = D_{up}/\mu = (0.1 \cdot 0.0445 \cdot 1.2) / (1.85 \cdot 10^{-5})$

$$= 288$$

Where ρ (air) = 1.2 Kg/m^3

μ (air) = $1.85 \cdot 10^{-5}$

5.1.2.1 PRESSURE DROP

- **For 0.0003m**

Pressure drop (Δp) is given by Ergun equation

$$\Delta p / L = (150 \bar{V}_o \mu (1-\epsilon)^2) / (\phi^2 D_p^2 \epsilon^3) + (1.75 \rho V_o (1-\epsilon)) / (\phi D_p \epsilon^3)$$

Where $L = 0.6 \text{ m}$

$\rho = 1.2 \text{ Kg/m}^3$

$\bar{V}_o = 0.0445 \text{ m/s}$

$\epsilon = 0.5$

$\Phi = 0.8$

$D_p = 0.0003$

$$\begin{aligned} \Delta p / 0.6 &= (150 \cdot 0.0445 \cdot 1.85 \cdot 10^{-5} (1-0.5)^2) / (0.8^2 \cdot 0.0003^2 \cdot 0.5^3) + (1.75 \cdot 1.2 \cdot 0.0445 \cdot (1-0.5)) / (0.8 \cdot 0.0003 \cdot 0.5^3) \\ &= 4356 \end{aligned}$$

$$\Delta p = 4356 \cdot 0.6 = 2613 \text{ N/m}^2$$

$$= 0.0257 \text{ Kg/ cm}^2$$

- **For 0.0004 m**

$$\Delta p / 0.6 = (150 \cdot 0.0775 \cdot 1.85 \cdot 10^{-5} (1-0.5)^2) / (0.8^2 \cdot 0.0004^2 \cdot 0.5^3) + (1.75 \cdot 1.2 \cdot 0.0775 \cdot (1-0.5)) / (0.8 \cdot 0.0004 \cdot 0.5^3)$$

$$\Delta p = 0.0259 \text{ Kg/ cm}^2$$

- **For 0.0005 m**

$$\Delta p / 0.6 = (150 \cdot 0.123 \cdot 1.85 \cdot 10^{-5} (1-0.5)^2) / (0.8^2 \cdot 0.0005^2 \cdot 0.5^3) + (1.75 \cdot 1.2 \cdot 0.123 \cdot (1-0.5)) / (0.8 \cdot 0.0005 \cdot 0.5^3)$$

$$\Delta p = 0.0271 \text{ Kg/ cm}^2$$

5.1.2.2 MINIMUM FLUIDIZATION VELOCITY

5.1.2.2.1. By using Wen and Yu correlation

- **For 0.0003 m**

$$\bar{V}_{om} = \mu / \rho_g D_p * (\sqrt{(33.72 + 0.0408 Ar)} - 33.7)$$

Where Archimedes number, $Ar = g \rho_g (\rho - \rho_g) D_p^3 / \mu^2$

$$Ar = (9.8 * 1.2 * (2500 - 1.2) * 0.00033) / (1.85 * 10^{-5})^2$$

$$= 2318$$

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0003) * (\sqrt{(33.72 + (0.0408 * 2318))} - 33.7)$$

$$= 0.0706 \text{ m/s}$$

- **For 0.0004 m**

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0004) * (\sqrt{(33.72 + (0.0408 * 5495))} - 33.7)$$

$$= 0.1224 \text{ m/s}$$

- **For 0.0005 m**

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0005) * (\sqrt{(33.72 + (0.0408 * 10732))} - 33.7)$$

$$= 0.1840 \text{ m/s}$$

5.1.2.2.2 By using Doichev & Akhmakov

- **For 0.0003 m**

$$\bar{V}_{om} = \mu / \rho_g D_p * (1.08 * 10^{-3} * Ar^{0.947})$$

$$= (1.85 * 10^{-5}) / (1.2 * 0.0003) * (1.08 * 10^{-3} * 2318^{0.947})$$

$$= 0.0853 \text{ m/s}$$

- **For 0.0004 m**

$$\bar{V}_{om} = \mu / \rho_g D_p * (1.08 * 10^{-3} * Ar^{0.947})$$

$$= (1.85 * 10^{-5}) / (1.2 * 0.0004) * (1.08 * 10^{-3} * 5495^{0.947})$$

$$= 0.1449 \text{ m/s}$$

- **For 0.0005 m**

$$\bar{V}_{om} = \mu / \rho_g D_p * (1.08 * 10^{-3} * Ar^{0.947})$$

$$= (1.85 * 10^{-5}) / (1.2 * 0.0005) * (1.08 * 10^{-3} * 10732^{0.947})$$

$$= 0.2185 \text{ m/s}$$

5.1.2.2.3 By using Saxena & Vogel

- **For 0.0003 m**

$$\bar{V}_{om} = \mu / \rho g D_p * (\Gamma(25.282+0.0571 Ar) - 25.28)$$

Where Archimedes number, $Ar = g \rho g (\rho - \rho g) D_p^3 / \mu^2$

$$Ar = (9.8 * 1.2 * (2500-1.2) * 0.00033) / (1.85*10^{-5})^2$$

$$= 2318$$

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0003) * (\Gamma(25.282 + (0.0571*2318)) - 25.28)$$

$$= 0.128 \text{ m/s}$$

- **For 0.0004 m**

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0004) * (\Gamma(25.282 + (0.0571*5495)) - 25.28)$$

$$= 0.215 \text{ m/s}$$

- **For 0.0005 m**

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0005) * (\Gamma(25.282 + (0.0571*10732)) - 25.28)$$

$$= 0.3114 \text{ m/s}$$

5.1.2.2.4 By Grace

- **For 0.0003 m**

$$\bar{V}_{om} = \mu / \rho g D_p * (\Gamma(27.22+0.0408 Ar) - 27.2)$$

Where Archimedes number, $Ar = g \rho g (\rho - \rho g) D_p^3 / \mu^2$

$$Ar = (9.8 * 1.2 * (2500-1.2) * 0.00033) / (1.85*10^{-5})^2$$

$$= 2318$$

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0003) * (\Gamma(27.22 + (0.0408*2318)) - 27.2)$$

$$= 0.0866 \text{ m/s}$$

- **For 0.0004 m**

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0004) * (\Gamma(27.22 + (0.0408*5495)) - 27.2)$$

$$= 0.148 \text{ m/s}$$

- **For 0.0005 m**

$$\bar{V}_{om} = (1.85 * 10^{-5}) / (1.2 * 0.0005) * (\Gamma(27.22 + (0.0408*10732)) - 25.28)$$

$$= 0.2194 \text{ m/s}$$

5.1.3 COMPARISON

5.1.3.1 Pressure Drop

Table 5.4 Pressure drop theoretical vs experimental

DIAMETER (M)	THEORETICAL (Kg/cm ²)	EXPERIMENTAL (Kg/cm ²)
0.0003	0.0257	0.021
0.0004	0.0259	0.0312
0.0005	0.0271	0.048

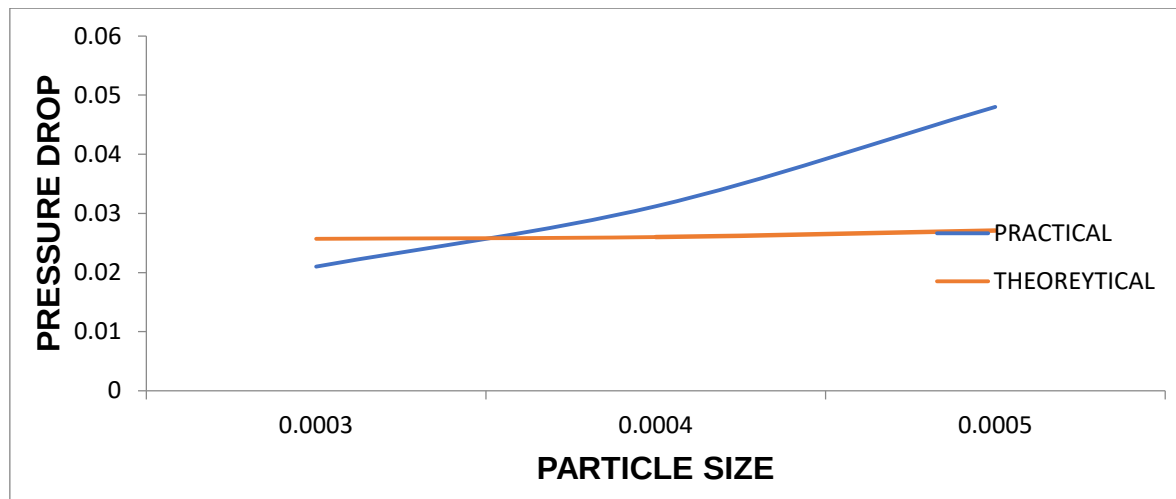


Figure 5.7 Pressure drops theoretical vs experimental

5.1.3.2 Minimum fluidization velocity

Table 5.5 Minimum fluidization velocity Theoretical vs experimental

DIAMETER (M)	GRACE (m/s)	WEN & YU (m/s)	DOICHEV& AKHMAKOV (m/s)	SAXENA& VOGEL (m/s)	EXPERI MENTAL (m/s)
0.0003	0.0866	0.0706	0.0853	0.1280	0.0891
0.0004	0.148	0.1224	0.1449	0.2150	0.1550
0.0005	0.2194	0.1840	0.2185	0.3114	0.2460

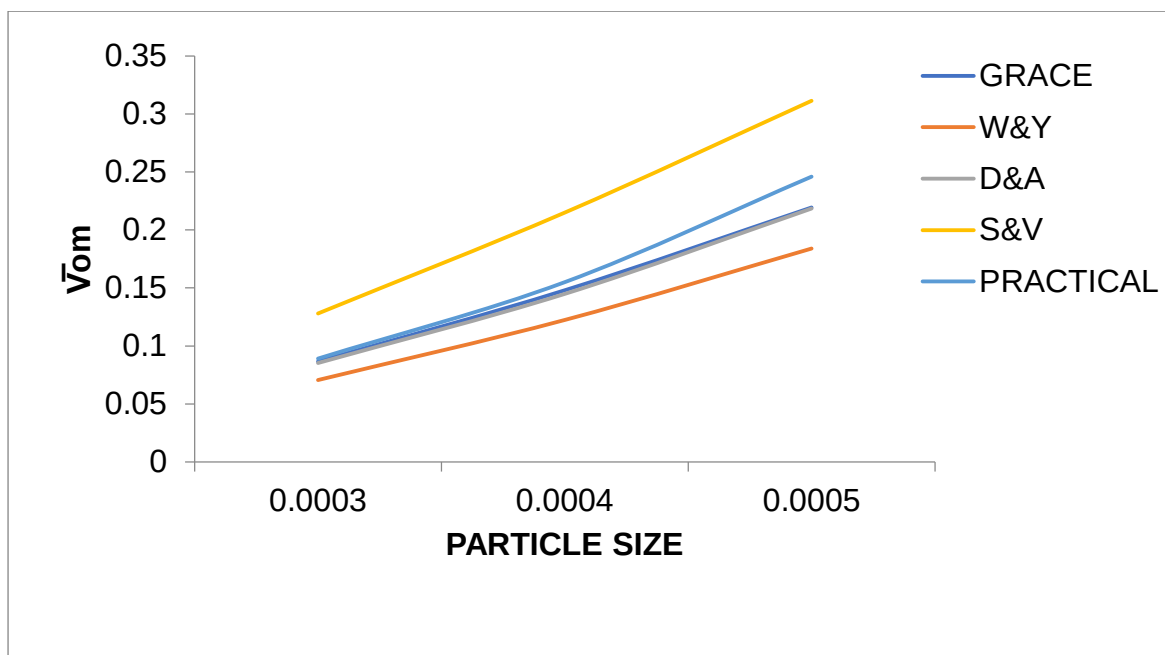


Figure 5.8 Minimum fluidization velocity Theoretical vs practical

CHAPTER6

6.1 CONCLUSIONS

The experimental results show that the minimum fluidization velocity ($\bar{V}_{om}$) as well as the pressure drop (Δp) across the column increases with the increasing particle geometric mean size. This validates the nature of the theoretical results obtained. Out of four theoretical correlations used, experimental values of minimum fluidisation velocity ($\bar{V}_{om}$) are more closely matching with Doichev& Akhmakov, Grace Correlations with a deviation of ~5%. It has been observed that the theoretical result obtained by Saxena & Vogel correlation is showing more deviation with the experimental results when compared with the other correlations. Practical values of pressure drop (Δp) obtained is observed to be having large deviation when it is compared with the theoretical results. This deviation may be due to the fact that a small change in porosity (ϵ) of the bed will have very large impact on ΔP . This makes it difficult to predict ΔP accurately and to reproduce experimental values after the bed is repacked. From the studies we can come to a conclusion that Doichev& Akhmakov, Grace Correlations are preferred for the plant scale minimum fluidisation velocity ($\bar{V}_{om}$) calculations.

REFERENCES

1. Biswal K.C., Bhowmik T., Roy G.K. – (1985) Prediction of minimum fluidization velocity for gas solid fluidization of regular particles in conical vessels, Chem. Eng. J. 30, 57–62.
2. D. Geldart. and A.C.Y Wong, Fluidization of Powders Showing. Degrees of Cohesiveness-I. Bed Expansion, Chemical Engineering Science, 39, 1481 (1984)
3. D. Kunii , A. Levenspiel , Fluidization engineering second edition, Elsevier, 2005
4. D.C. Sau, S. Mohanty, K.C. Biswal, Experimental studies and empirical models for the prediction of bed expansion in gas–solid tapered fluidized beds, Chemical Engineering and Processing: Process Intensification, 49(4), (2010), 418-424.
5. D.C. Sau, S. Mohanty, K.C. Biswal, Minimum fluidization velocities and maximum bed pressure drops for gas–solid tapered fluidized beds, Chemical Engineering Journal, 132(1–3), (2007), 151-157
6. Depypere F., Pieters J.G., Dewettinck K. - Expanded bed height determination in a tapered fluidised bed reactor, Journal of Food Engineering 67 (2005) 353–359
7. Epstein N., Pruden B.B. - Liquid fluidization of binary particle mixtures—III Stratification by size and related topics, Chemical Engineering Science 54 (1999) 401—415.
8. Escudiéa. R., Epsteina. N., Gracea. J.R., Bi. H.T. - Effect of particle shape on liquid-fluidized beds of binary (and ternary) solids mixtures: segregation vs. mixing, Journal of Chemical Engineering Science 61 (2006) 1528 – 1539.
9. F.J. Zuiderweg, Proc. Int. Symp. on Fluidization, A.A.H. Drinken berg, ed, p.739, (1967)
10. Jena, H.M., 2009. - Hydrodynamic of Gas-Liquid-Solid Fluidized and Semi-Fluidized Beds, PhD thesis. National Institute of Technology, Rourkela
11. Juma A. K. A. and Richardson J. F., - Segregation and Mixing in Liquid Fluidized Beds, Chemical Engineering Science, vol. 38, No.6, pp.955-967, 1983.
12. K Takeda, Kagaku Kogaku , Chem Eng Jpn, 21(1957), 124

13. K.C. Biswal, S. Sahu and G.K. Roy, Prediction of the fluctuation ratio for gas solid fluidization of regular particles in a conical vessel Chem Eng J,23,(1982) 97
14. K.S. Sutherland, Fluidized bed dynamics, Tran.s Inst Ch.em Eng , 39 (1961), 188
15. L. Davies and J.F. Richardson, Trans. Inst. Chem. Eng.,44 (1996), p. T293.
16. Maureen A. Howley, Benjamin J. Glasser. - Hydrodynamics of a uniform liquid-fluidized bed containing a binary mixture of particles, Chemical Engineering Science 57 (2002) 4209 – 4226.
17. Nidal Hilal, The dependence of solid expansion on bed diameter, particles material, size and distributor in open fluidized beds, Advanced Powder Technology,16(1),2005,73-86.
18. R.K. Singh, A. Suryanarayana, G.K. Roy, Prediction of minimum velocity and minimum bed pressure drop for gas–solid fluidization in conical conduits, Can. J. Chem. Eng. 70 (1992) 185–189.
19. S. Jing, Q. Hu, J.Wang, Y. Jin, Fluidization of coarse particles in gas–solid conical beds, Chem. Eng. Process.39 (2000) 379–387.
20. S.H. Schaafsma, T. Marx, A.C. Hoffmann,Investigation of the particle flowpattern and segregation in tapered fluidized bed granulators, Chemical Engineering Science,61(14),(2006), 4467-4475
21. T. Maruyama and T. Koyanagi, Fluidization in tapered vessel, The Chemical Engineering Journal, 51 (1993)121-128
22. Trambouze, P., Euzen, J., 2004. Chemical Reactor, from Design to Operation. Technip, Paeis.
23. Wen- Ching Yang - Handbook of fluidization & Fluid particle systems

