HYDRODYNAMICS OF THREE PHASE FLUIDIZED BED USING LOW DENSITY PARTICLES

A THESIS SUBMITTED IN PARTIAL FULFILLMENT OF THE REQUIREMENTS FOR THE DEGREE OF

Master of Technology
in
Chemical Engineering

By

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Department of Chemical Engineering
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Under the guidance of
Prof. H. M. Jena

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CERTIFICATE

This is to certify that the thesis entitled, **"HYDRODYNAMICS OF THREE PHASE FLUIDIZED BED USING LOW DENSITY PARTICLES"** submitted by **Miss Aakanksha Pare** in partial fulfillment of the requirements for the award of Master of Technology Degree in **Chemical Engineering** at the National Institute of Technology, Rourkela (Deemed University) is an authentic work carried out by her under my supervision and guidance.

To the best of my knowledge, the matter embodied in the thesis has not been submitted to any other University/ Institute for the award of any degree or diploma.

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ACKNOWLEDGEMENT

I would like to express my deep sense of gratitude and respect to my supervisor Prof. Hara Mohan Jena for his excellent guidance, suggestions and constructive criticism. Working under his supervision greatly contributed in improving quality of my research work and in developing my engineering and management skills. I would also like to thank NIT Rourkela and MHRD of India for providing me research facilities and opportunity.

I am extremely fortunate to be involved in such an exciting and challenging research project. It gave me an opportunity to work in a new environment. This project has increased my thinking and understanding capability.

I would also like to thanks Mr. Rahul Omar and Mr. Sambhurisha Mishra for their extreme support during my experiments.

I would like to express my thanks to all my friends, all staffs and faculty members of Chemical Engineering department for making my stay in N.I.T. Rourkela a pleasant and memorable experience.

I would like to thank all whose direct and indirect support helped me in completing my thesis in time.

Lastly I would like to convey my heartiest gratitude to my parents for their unconditional love and support.

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CONTENTS

Abstract

Gas–liquid–solid fluidized beds are widely used in the petrochemical, pharmaceutical, biotechnology, food and environmental industries. Fluidization is dependent on effective contact between the phases. A three-phase (gas-liquid-solid) fluidized bed is used to study the hydrodynamic characteristics of low density particle (wood) using water as liquid phase and air as the gas phase. The fluidized bed consists of three sections, the test section, the gas-liquid distributor section, and the gas-liquid disengagement section. The hydrodynamic characteristics determined included the bed pressure drop, minimum fluidization velocity and bed expansion ratios. The dependence of these quantities on particle diameter, initial static bed height, liquid velocity and gas velocity has been discussed. Results indicate that the bed pressure drop and minimum fluidization velocity decreases with gas velocity but increases with particle size. Bed expansion ratio increases with increase in the values of the liquid but decreases with dp and initial static bed height. Model equation have been developed from factorial design analysis for predicting the values for bed expansion ratio and predicated values have been found to agree well with the experimental results.

Keywords: Fluidization, hydrodynamics, pressure drop, minimum fluidization velocity, bed expansion,

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NOMENCLATURE

Ar	Archimedes number
bi	coefficient in factorial design analysis
Fr_g	gas Froude number
H	height of expanded fluidized bed, m
Hs	initial static bed height, m
N	total number of treatments in factorial design analysis
ΔP	pressure drop in liquid-solid fluidized bed, Pa
R	bed expansion ratio (He/Hs) or (H/Hs) in fluidized bed
U_{Lmf}	minimum liquid fluidization velocity, ms^{-1}

Greek symbols

ε_g	gas holdup in fluidized bed,
ε_L	liquid holdup in fluidized bed, -
ε_s	solid holdup in fluidized bed, -
μ_L	liquid viscosity, Pas
ρ_g	density of gas, kgm^{-3}
ρ_L	density of liquid, kgm^{-3}

CHAPTER-1

INTRODUCTION AND LITERATURE REVIEW

1.1 INTRODUCTION

Fluidization is the operation by which solid particles are transformed in to fluid like state through suspension in a gas or liquid. It is a fluid-solid contacting technique, which has found extensive industrial applications. In fluidization process liquid or a gas is passed up through the solid particles. At very low velocity the particles do not move but as the fluid velocity is steadily increased, the pressure drop and the drag on individual particle increase, and particle start to move and become suspended in the fluid. The term fluidization describes the condition of fully suspended particles. Important design parameters for fluidization systems are: the minimum fluidization velocity, bed expansion of fluidization, and pressure drop in the bed.

Three phase fluidization also known as gas-liquid-solid fluidization. Gas-liquid-solid fluidization is defined as an operation in which a bed of solid particles is suspended in gas and liquid media due to the net drag force of the gas and/or liquid flowing opposite to the net gravitational force (or buoyancy force) on the particles. Such an operation generates intimate contact among the gas, liquid and the solid in the system and provides substantial advantages for application in physical, chemical or biochemical processing (Jena, 2010). In a three-phase fluidization system, the gas phase, in the form of bubbles, interacts intimately with the liquid and solid phases and solid particles are suspended or fluidized by the upward flow of liquid and gas bubbles. The strong interactions between the phases provide an intensive mixing which is desirable for effective heat and mass transfer and for chemical reactions.

Hydrodynamic properties of three-phase fluidized beds are important for analyzing its performance. Hydrodynamic behavior of the bed such as bed pressure drop, minimum fluidization velocity, phase holdup, bubble properties, mixing characteristics and bed expansion have to be investigated to provide the basic information required for the design of such fluidized bed (Fan,1989; Lee et al.,2001).

Fluidization is broadly of two types, viz. aggregative or bubbling and particulate. Aggregative fluidization is a characteristic of gas-solid type or gas-liquid-solid system with gas as the continuous phase. The non-uniform nature of the bed describes the aggregation of particles. While the liquid-solid system or gas-liquid-solid fluidization with liquid as continuous phase is of particulate fluidization type. This is characterized by a large but uniform expansion of the bed at high velocities (Jena, 2010).

1.1.1. Advantages of three-phase fluidized bed

1. The smooth liquid like flow of particles allows continues automatically controlled operations with easy handling.
2. The solid is vigorously agitated by the fluid passing through the bed; the violent motion of the solid also gives high heat transfer rates to the wall.
3. The rapid mixing of solids leads to close to isothermal conditions throughout the reactor; hence the operation can be controlled simply and reliably.
4. In addition the whole vessel of well mixed solids represented a large thermal fly wheel that resist rapid temperature changes, responds slowly to abrupt changes in operating conditions.
5. It is suitable for large scale operations.
6. Bubbling and circulating fluidized bed systems are becoming an increasingly important in industries for the power generation and chemical processing industries.
7. Beneficial in economic, operational and environmental terms can be achieved with fluidized bed technology over more traditional technologies.

1.1.2. Disadvantages of three-phase fluidized bed

1. For bubbling beds of fine particles, it’s difficult to describe flow of gas, with this large deviations from plug flow, represents inefficient contacting. This becomes especially serious when high conversion of gaseous reactant is required.
2. The rapid mixing of solids in the bed leads to non-uniform residence times of solids in the reactor.
3. For continuous treatment of solids, this gives a non-uniform product and poorer performance, especially at high conversion levels.

4. For catalytic reactions, the movement of porous catalyst particles, which continually captures and release reactant gas molecules, contributes to the back mixing of gaseous reactant, thereby reducing yield and performance.
5. Erosion of pipes and vessel from abrasion by particles can be serious problem.
6. For non-catalytic operations at high temperature, the agglomeration and sintering of fine particles can require a lowering in temperature of operations, reduce the reaction rate.

1.2 Modes of operation and flow regimes in three-phase fluidized bed:

Depending on the direction of the flow, fluidized beds are classified as: co-current up-flow, co-current down-flow, counter-current, liquid batch with gas up-flow (Jena, 2010).

The gas-liquid-solid fluidization is categorized into four mode of operation. These mode are co-current three-phase fluidization with liquid as continuous phase, co-current three-phase fluidization with the gas as the continuous phase, inverse three-phase fluidization and fluidization represented by a turbulent contact absorber (TCA) (Jena, 2010).

In three-phase fluidized bed flow regime has a great role and it is important for its stable operation in a particular set of operating variables. Fan (1989) identified three-phase fluidized beds can operate: bubbling, slugging and transport regime. There are two sub-categories within the bubbling regimes: the dispersed bubbles and the coalesced bubble regimes. Gas-liquid-solid fluidized beds can be operated with different hydrodynamic regimes, which depend on the gas and liquid velocities, as well as the gas, liquid and solid properties (Lauren et al., 2005). Zhang (1996) and Zhang et al. (1997) identified seven distinct flow regimes for three phase co-current fluidized bed and identified a number of quantitative methods for determining the transitions as under:

- Dispersed bubble flow: This flow corresponds to high liquid velocities and low liquid velocities. Results in small bubbles of relatively uniform size. Little bubble coalescence despite high bubble frequency.
- Discrete bubble flow: Usually occurs at low liquid and gas velocities. It is similar to the dispersed bubble flow regime with respect to small bubble and uniform size. However, the bubble frequency is lower.

- Coalesced bubble flow: Usually found at low liquid velocities and intermediate gas velocities. In this flow bubbles are larger and show a much wider size distribution due to increased bubble coalescence.
- Slug flow: This regime is characterized by large bullet shaped bubble with a diameter approaching that of the column and length that exceed the column diameter.
- Churn flow: Churn flow similar to the slug flow regime, but much more chaotic and frothy.
- Bridging flow: This is a transitional regime between the churn flow and the annular flow, where liquid and solids effectively form “bridges” across the reactor which is continuously broken and reformed.
- Annular flow: At very high gas velocities, a continuous gas phase appears in the core of the column.

Dispersed flow, discrete flow and coalesced flow are grouped under the heading “bubbling regimes”, while churn flow, bridging flow and annular flow classified in the transport regime (Jena, 2010).

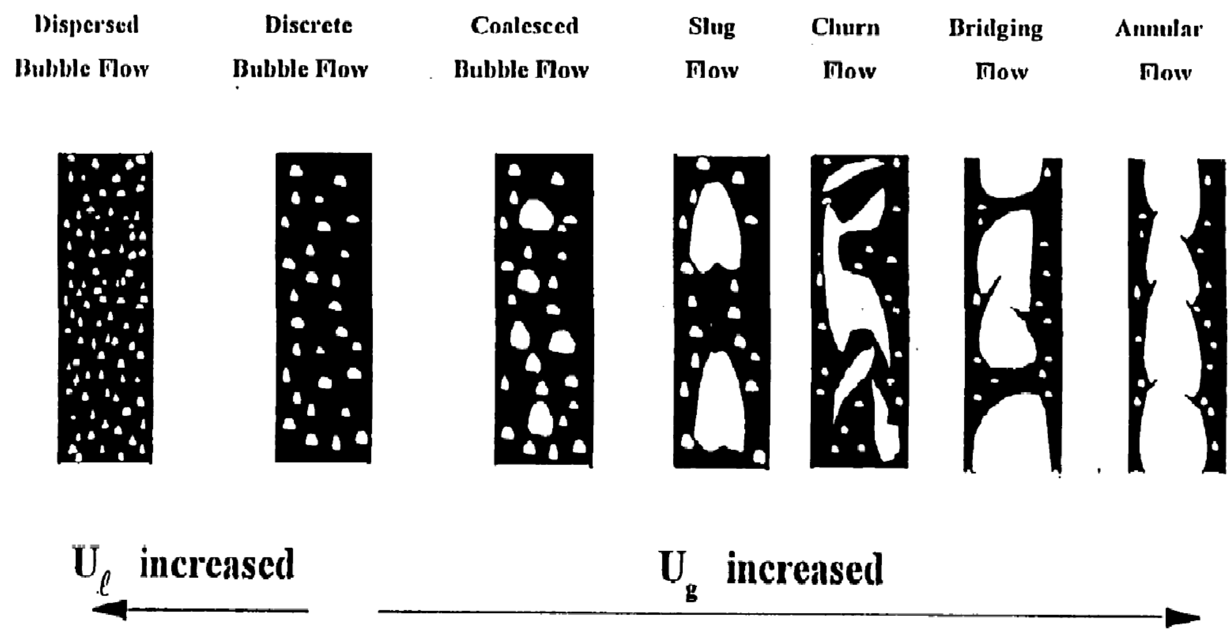

Fig.1.1.Fow regime in gas-solid-liquid co-current fluidized bed

1.3. Application of three-phase fluidized bed:

The three phase fluidized bed has found immense applications in various fields like pharmaceutical, chemical, petrochemical industry with the development of fluid bed catalytic cracking, biochemical processing, metallurgical, mineral processing and metallurgical processes such as drying, calcining, and sulfide roasting. Three-phase fluidized beds are widely used in a range of applications including hydro-treating and conversation of heavy petroleum and synthetic crude, coal liquefaction, methanol production, sand filter cleaning, electrolytic timing, conversion of glucose to ethanol, aerobic waste water treatment, and various other hydrogenation and oxidation reactions (Fan, 1989; Wild and Poncin, 1996; Jena, 2010). Fluidization has many applications with the use of ion exchange particles for the purification and processing of many industrial liquid streams. Industries such as food & beverage, hydrometallurgical, water softening, catalysis, bio-based chemical etc. use ion exchange in processing. In the waste water treatment various types of bioreactors are in use. Fluid Bed Bio-Reactor (FBBR) is mainly used in food, pharmaceutical and biological waste treatment sectors.

1.4. Some definitions of fluidization phenomena:

The successful design and operation of a gas-liquid-solid fluidized bed system depends on the ability to accurately predict the fundamental properties of the system. To design a three-phase fluidized bed chemical reactor different aspects must be predicted and quantified.

Some of the important parameters used to describe the fluidization phenomena are:

- Bed Pressure drop: Measures the drag in combination with the buoyancy and phase holdups.

$$\frac{\Delta P}{g\Delta z} = \rho_g \epsilon_g + \rho_l \epsilon_l + \rho_s \epsilon_s \tag{1}$$

- Minimum fluidization velocity: The minimum superficial velocity at which the bed becomes fluidized.

- Bed expansion ratio: Measure the extent of fluidization of the bed.

$$R = \frac{H}{H_s} \tag{2}$$

- Porosity: Measures the volume occupied by both the liquid and the gas.

$$\varepsilon = \epsilon_g + \epsilon_l = 1 - \epsilon_s \tag{3}$$

- Gas holdup: Measure the fractional volume occupied by the gas.

$$\varepsilon_g = \frac{\text{volume of gas}}{\text{total bed volume}} \tag{4}$$

- Liquid holdup: Represents the fraction of the bed occupied by the liquid phase.

$$\varepsilon_l = \frac{\text{volume of gas}}{\text{total bed volume}} \tag{5}$$

- Solid holdup: Measure the fractional volume occupied by solids.

$$\varepsilon_s = \frac{\text{volume of gas}}{\text{total bed volume}} \tag{6}$$

$$\epsilon_g + \epsilon_l + \epsilon_s = 1 \tag{7}$$

1.5. Variables affect the quality of fluidization

In gas liquid solid fluidization phenomena some of the variables affecting the quality of fluidization are as follows.

- Fluid flow rate: For the suspension of solids in fluid the flow rate should be high, but further increase in fluid flow rate channeling occurs.
- Bed height: The greater the bed height it's difficult to obtain good fluidization.
- Particle density: the closer the density of the particles to gas and liquid, then it is easier to maintain smooth fluidization.
- Fluid inlet: The distribution of the fluid in the bed is important in design the fluidized bed.

1.6. Hydrodynamic studies of three-phase fluidized beds:

Hydrodynamic studies of three-phase fluidized beds had been done by many researchers through experiments.

Soung (1976) have determined the bed expansion data from commercial cobalt-molybdenum catalysts in n-heptane and nitrogen in lucite tubes of 12.7 cm and 15.24 cm diameter for wide range of gas and liquid velocities. Three different sizes cylindrical catalysts were used. They have presented a correlation for the effect of gas velocity on bed expansion.

Fan et al., (1985) did an experimental study on air/water/binary mixtures of activated carbon, nylon, glass and alumina beads to visualize minimum fluidization velocity, bed expansion, gas holdup, mixing and segregation.

Saberian Broudjenni et al., (1987) did experiment by taking different types of system like nitrogen, helium, carbon dioxide/water, carbon tetrachloride, cyclohexane, gas oil, kerosene/glass and alumina beads, alumina extrudates and they studied gas holdup as well as liquid holdups. They also conducted experiments to study minimum fluidization, bed porosity and gas slip velocity.

Nacef et al., (1992) had taken nitrogen/water-alcohol solution/glass and propylene beads as their system and they carried out experiments to study minimum fluidization velocity, phase holdups and slip velocity.

Mieura et al., (1997) have investigated hydrodynamics and mass transfer in three-phase fluidized bed. Minimum fluidization velocity, bed velocity, gas holdup were measured in a 0.068m internal diameter fluidized bed. Glass bead of three different sizes (3mm, 5mm and 7mm) were used. They have obtained that minimum fluidization velocity decreases with increasing non-newtonian flow behaviors and gas holdup decreases with increasing non-newtonian flow behaviors and decreasing particle diameter.

Yu Rittmann et al., (1997) did experiments to study the hydraulic behavior such as bed height and overall gas, liquid and solid holdups (volume fraction) of three phase fluidized bed reactors. They have developed a predictive model. This type of model uses process such as reactor dimensions, particle properties and gas and liquid flow rates as input variables. They have used glass beads under wide range of gas and liquid velocities. This model is further applied to analyze fluidized bed with biofilm coated media.

Luo et al., (1997) investigated bed contraction and expansion in a high pressure and high temperature gas–liquid–solid fluidized bed. They have operated the bed at pressures from 0.1 MPa to 17.4 MPa and temperatures from 20^{o}C to 948^{o}C. They have visually observed the bubble dynamic behavior. They studied that the pressure and temperature affect the bed expansion and contraction, mainly through the variations in bubble behavior and the changes of liquid properties. The extent of bed contraction decreases with an increase in pressure and/or temperature. They have used 1mm and 2.1mm glass beads as solid phase. For glass beads, bed contraction is observed to be more pronounced under higher pressures for bed expansion or contraction phenomena and the generalized wake model is used for the study.

Larachi et al., (2000) thoroughly studied minimum fluidization liquid velocity in three-phase fluidized beds .The database related to over 540 measurements was used for wide-ranging fluids and bed properties. They have covered 30 various particles and 18 liquids and includes data such as aspect ratio, wall effect (or column-to-particle diameter) ratio, and Re_{Lmf} ranging from 0.8 to 27, 9 to 127, and 10-2 to 800 respectively. Multilayer perceptron artificial neural networks have been extensively used by them to generate two highly accurate, a purely dimensional and a dimensionless empirical correlation describing the U_{Lmf}. The neural network approach has been shown to predict with moderate success the minimum fluidization gas velocity, U_{Gmf}, in liquid-buoyed gas-activated three-phase fluidized beds containing coarse particles (dp > 1 mm) at high-input gas fractions.

Mieura et al., (2001) have investigated that Gas hold-up and bed expansion in three phase fluidized bed. Experimental measurements were carried out with bed of glass beads as solid and newtonian liquids and non-newtonian liquids with gas. The value of gas hold-up increased with increasing particle size and decreased with liquid velocity respectively. They have observed that the bed voidage increased with increasing superficial liquid velocities and superficial gas velocities. The increase of the viscous non-Newtonian flow behaviors resulted in an increase of the bed voidage. They have obtained from the result that the bed voidage increased with an increase in the liquid velocity and the magnitude of the increase became smaller as particle size increased. Higher bed voidage observed as increase in gas velocity. They have developed the correlation for the bed voidage in three-phase fluidized beds for gas-newtonian or non-newtonian liquid–solid by combining the generalized wake model.

Ruiz et al., (2004) have conducted the experiment to study the hydrodynamic characteristics of a gas-liquid-solid fluidized bed containing commercial hydrotreating catalyst extrudates with both water and oil as the liquid phase. Hydrodynamic characteristics such as the minimum fluidization velocity and bed expansion were calculated. Relatively smaller errors were found from empirical correlations for reproducing the systems with water than with the organic liquids.

Sivalingam et al.,(2009) have studied the hydrodynamics of the three phase fluidized bed with liquid as a continuous phase in a 54 mm internal diameter Perspex (Acrylic column) with particle size of 4.33 mm and 1.854 mm glass beads. From the results they have obtained the gas holdup and bed porosity increases with increasing gas flow rate and the system mainly depends on good contact between solid and liquid.

Sivaguru et al., (2009) have studied the hydrodynamics of three phase fluidized bed using ceramic particle of 1 mm diameter, density of 2650 kg/m^3 is used as a solid phase. Operating condition is achieved even with non-uniform air, water flow rates and with different bed heights (100 mm, 200 mm, 300 mm, 400 mm and 500 mm). Pressure drop values obtained from the CFD simulation was compared with the experimental data. They found that as the gas flow rate increases, the pressure drop in the column decreases, provided the initial bed height diameter of the column and liquid flow rate are constant. This is due to decrease in density of the fluid medium in the bed by means of gas holdup.

Sulayman et al., (2010) have studied the hydrodynamic characteristics of three-phase fluidized beds such as gas- holdup, local gas- holdup, bubble rise velocity, minimum fluidization velocity, superficial gas velocity, physical and liquid rheological properties. They have used a perforated Teflon plate as gas distributor, with 53 holes, 7 mm diameter and free surface area of 23.11%. Air, O_2 and CO_2 and Carboxy Methyl Cellulose concentrations 0.1 %, 0.5 %, 1 %, and 2 wt%, were used as non-newtonian (pseudoplastic) liquids. Activated carbon with diameter 0.25 - 0.75mm and density 770 kg/m3 and Ni-Mo/Al2O3 with diameter 1.8 mm and density 2500 kg/m3 were used as solid phase. The gas holdup was correlated with dimensionless groups and independent parameters.

Sheikhi et al., (2013) experimentally investigated hydrodynamics of three-phase gas–liquid–solid fluidized bed over a wide range of operating conditions using air, water and glass beads as gas, liquid and solid phase respectively using non-intrusive methods, vibration signature and pressure fluctuation. They have done a comprehensive study on the standard deviations of pressure fluctuation with two new statistical analyses on the pressure fluctuation, namely signal energy and average cycle frequency which presented a new method of determining minimum fluidization velocity.

Accurate prediction of minimum liquid fluidization condition is essential to the successful operation of gas-liquid- solid fluidized beds. Table 1.1.Summarizes the previous correlations on the minimum liquid fluidization velocity in gas-liquid-solid systems.

Table.1.1. Summary of correlations on minimum liquid fluidization velocity:

Reasearchers	**correlation**
Ermakova et al.(1970)	$U_{lmf} = U_{lmf}^{0}(1 - 0.5\, U_g^{0.075} - \varepsilon_{mf}\, \beta_{Gmf})$
Begovich et al.(1978)	$Re_{lmf} = 0.00512\, Ar_L^{0.662} Fr_G^{-0.118}$
Bloxon et al. (1975)	$U_{lmf} = 5.359 \times 10^{-17} U_G^{-0.14} \mu_L^{-0.497} d_c^{-0.423} \rho_s^{3.35}$
Costa et al. (1986)	$U_{lmf} = 6.969 \times 10^{-4}\, U_G^{-0.328} \mu_L^{-0.355} d_v^{1.086} d_c^{0.042} (\rho_S^{0.865} - \rho_L^{0.865})$
Fortin et al. (1984)	$U_{lmf} = 0.427 U_G^{-0.198} d_v^{1.539}\, (\rho_s - \rho_L)^{0.775}$
Song et al. (1989)	$U_{lmf} = U_{lmf}^{0}(1 - 376 U_G^{0.327} \mu_L^{0.227}\, d_v^{0.213}\, (\rho_s - \rho_L)^{-0.423}$
Nacef et al. (1992)	$ln\,(U_{lmf}) = \ln\,(\, U_{lmf}^{0}) - 13.8 Fr_G^{0.35}\, (\rho_s - \rho_L)^{-0.38}$
Lee et al.,(2001)	$Re_{Lmf} = \sqrt{\left(51.4 \frac{(1 - \epsilon_{mf})}{\emptyset_s}\right)^{2^2} + .0571 \emptyset_s \,\epsilon_{mf}{}^3\, Ar(1 - \epsilon_{gmf})^3 - 51.4 \frac{(1 - \epsilon_{mf})}{\emptyset_s}}$ $\epsilon_{gmf} = 0.16 \vert\, \epsilon_{mf}\, [U_g \vert (U_g + U_{Lmf})]$

Ramesh et al.,(2002)	$\mathrm{Re}_{Lmf} = 0.6\left[1+Fr_g\right]^{-1.85} Ar^{0.30} Mo^{-0.09} \phi_s^{0.04}$
Ruiz et al.,(2004)	$U_{Lmf} = U_{Lmf}^0 (1-0.5U_g{}^{0.075} - \in_{mf} \beta_{gmf})\phi_s^{-0.93}$ $\beta_{gmf} = \dfrac{0.16}{\in_{mf}\left[\dfrac{U_g}{U_g+U_L}\right]}$
Jena (2009)	$U_{Lmf} = 0.0065 U_g^{-0.268} d_p^{0.551} \mu_L^{0.434} (\rho_s - \rho_L)^{0.066}$

1.6.1. Hydrodynamic studies of three-phase fluidized beds with low density particles:

Low density solid particles found wide application in bio reactor for aerobic waste water treatment. Hydrodynamics study of three-phase fluidized bed with low density particles are rarely seen in literature although a tremendous work is seen for moderate or high density solid particles.

Nore et al., (1992) have studied hydrodynamics, gas-liquid mass transfer and particle-liquid heat and mass transfer in three-phase fluidized bed of light particles for biochemical application. They have used polypropylene beads with inclusion of mica and achieved a density ranging from 1130 kg/m^3 to 1700 kg/m^3 as the solid phase. They have studied the effect of liquid and gas velocities on bed porosity and liquid holdup. They have reported increase in the gas velocity and the liquid velocity bed porosity increases for both.

Briens et al., (1997) have studied phase properties evolve in biological and biochemical three phase fluidization processes and thus it is essential to monitor the fluidization regime. These type of processes use immobilized whole cells, subcellular organelles or enzymes used as the solid phase and density of these particle is slightly higher than the density of water. They have determined the minimum fluidization velocity from the coefficient of variation of V-statics of the integrated- bed pressure gradient and the derivative of the local conductivity.

Sokol and Halfani (1999) investigated low density solid support for the hydrodynamics of a gas-liquid-solid fluidized bed. They have found that value of minimum fluidization air velocity depend on the ratio of bed to reactor volume and mass of cell growth on the particles. Thy have also investigated that the air holdup depends on air velocity, ratio of bed to reactor volume and mass of biomass particles. The effect of operational parameters on biodegradation of organics in fluidized bed bioreactor with low density solid particles have been studied by Sokol (2001) and Sokol and Korpal (2004).

Briens and Ellis (2005) have characterized the hydrodynamics of three-phase fluidized bed systems by statistical, fractal, chaos and wavelet analysis. They have determined the optimum fluid velocity and ratio of volume of bed to volume of reactor for largest degradation of phenol. Gas-liquid-solid fluidized beds can be operated with different hydrodynamic regimes, which depend on the gas and liquid velocities, as well as the gas, liquid and solid properties. It is essential to know under which regime the reactor will be operating to model a proper reactor. Some studies were made to identify the flow regimes, especially on low-density particle system but that is very few.

Allia et al. (2006) studied that the solid particles covered with a biofilm are fluidized by air and contaminated water by to confirm the operating stability, to identify the nature of mode flow and to determine some hydrodynamic parameters such as the minimum fluidization velocity, the pressure drop, the expansion, the bed porosity, the gas retention and the stirring velocity.

1.7. Objectives of the present work:

Critical appraisal of the research on the hydrodynamics of three phase fluidized bed reactors shows that the majority of work has been done with high density particles. Therefore experimental work has been carried out to understand the complex hydrodynamics of three phase system using low density particles such as wood particle of different sizes. Major aim of the experiment is to better understanding of the bed behavior for industrial applications like biological treatment of waste water. Thus in this type of application, the design and operation of the fluidized bed should ensure a good quality of fluidization as well as sufficiently large residence time for the waste water in the bed.

The main objectives of the present research work are:

- Hydrodynamic study on three-phase fluidized bed with low density regular sized particles (wood particles) using water and air as liquid and gas phase. Hydrodynamic behavior like pressure drop, minimum fluidization velocity and bed expansion ratio has been studied.
- To analyze the experimental data of bed expansion ratio with that of calculated values using developed model equation by factorial design methodology.

1.8. Thesis summary:

This thesis comprises of four chapters. Introduction and literature survey, Experimental setup and technique, Result and Discussion and Conclusion and Future scope of the work.

- Chapter 1, the background information, literature review and objective of the present work is discussed.
- Chapter 2 deals with the experimental set up and detail of the system under experimental investigation. It also includes experimental procedure.
- Chapter 3, the results of various hydrodynamics properties obtained from experiment have represented graphically and discussed.
- Chapter 4 deals with overall conclusion. Future recommendations based on the research outcome are suggested. The major findings of the work are also summarized.

CHAPTER-2

EXPERIMENTAL SET-UP AND TECHNIQUES

2.1 EXPERIMENTAL SET-UP

A three-phase (gas-liquid-solid) fluidized bed is used to study the hydrodynamic characteristics (like pressure drop, minimum fluidization velocity, and bed expansion) of low density particle (wood) using water as liquid phase and air as the gas phase.

The fluidized bed column consists of three sections, the test section, the gas-liquid distributor section, and the gas-liquid disengagement section. Fig. 2.1 shows the schematic representation of the experimental setup used in the three-phase fluidization study. Fig. 2.2 gives the photographic representation of the experimental setup. The test section is the main component of the fluidized bed where fluidization takes place. It is a vertical cylindrical Plexiglas column of 0.1 m internal diameter and 1.24 m height.To prevent particle entrainment a mesh screen has been attached to the top of the column.An uniformly distributed liquid and gas mixture enters in the test section as the gas–liquid distributor is located at the bottom of the test section. The distributor section made of Perspex is fructo-conical of 0.31 m in height, and has a divergence angle of 4.5^0. The liquid inlet of 0.0254 m in internal diameter is located centrally at the lower cross-sectional end. The higher cross-sectional end is fitted to the test section, with a perforated grid made of G.I. sheet of 0.001 m thick.

Figures 2.3(a) and 2.3(b) represent the photographic view of the gas-liquid distributor section and the distributor plate. There is an antenna-type air sparger of 0.09 m diameter just below the distributor plate containing 50 number of 0.001 m holes, for generating uniform bubbles to flow throughout the cross section of the column.

In the gas-liquid distributor section, the gas and the liquid streams are merged and passed through the perforated grid. The mixing section and the grid ensured that the gas and the liquid are well mixed and evenly distributed into the bed.

1. Gas–liquid disengagement section,
2. Test section,
3. Gas–liquid distributor section,
4. Water rotameters,
5. Air rotameters,
6. Ball valves,
7. Liquid pump,
8. Air sparger,
9. Two way quick closing valves,
10. U tube manometer

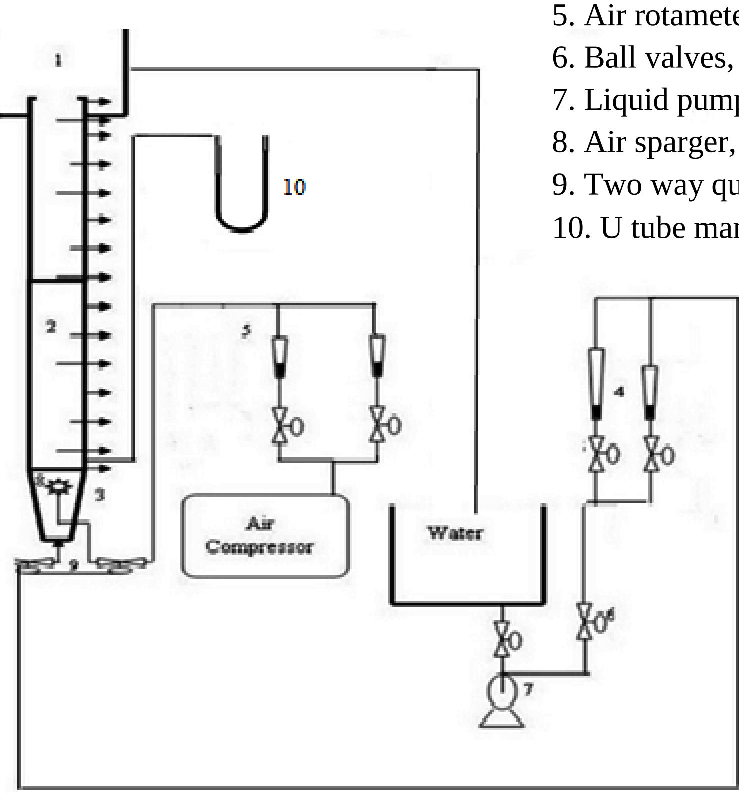

Fig.2.1. Experimental set-up of three phase fluidized bed

The gas-liquid disengagement section at the top of the fluidizer is a cylindrical section of 0.26 m internal diameter and 0.34 m height, assembled to the test section with 0.08 m of the test section inside it, which allows gas to escape and liquid to be circulated through the outlet of 0.0254 m internal diameter at the bottom of this section.

For the measurement of pressure drop in the bed, the pressure ports have been provided and fitted to the manometer filled with carbon tetrachloride as the manometric fluid.

Fig.2.2 Photographic view of the experimental set-up

(a) (b) (c)

Fig.2.3. Photographic view of: (a) the gas-liquid distributor, (b) distributor plate,(c) air sparger.

Table: 2.1 Equipment characteristics

Test section (Cylindrical Plexiglas column)	
Diameter, m	0.1
Height, m	1.24
Gas-liquid distributor section (fructo-conical)	
Height, m	0.31
Diameter of the ends, m	0.0508, 0.1
Tapered angle	4.5^0
Gas-liquid disengagement section (Cylindrical)	
Diameter, m	0.26
Height, m	0.34
Air sparger (antenna type)	
Orifice size, m	0.001 (50 nos.)
Distributor plate (GI)	
Diameter, m; thickness, m	0.12; 0.001
Liquid reservoirs	
Reservoir-1: dimension, m; capacity, lit.	ID = height = 1; 1000

2.2. Measurement of properties of the solids:

The solid particle was made from wood strips. The wood particles were furnished in carpentry shop of cubical shape with regular size of 3 mm, 5 mm and 7 mm.

2.2.1. Particle size

The equivalent diameter of cubical shape wood particles has been calculated by using the following formula

$$V_s=V_c$$

where V_s and V_c is the volume of sphere and volume of cube respectively. The diameter of cubical shape wood particles is the equivalent diameter, which is the diameter of a sphere whose volume is the same as that of the volume of cube.

$$\frac{\pi}{6}d^3 = S^3$$

where S is the length of the solid particle.

2.2.2. Particle density

The density of the wood particle has been measured using the water displacement method in which the packing voidage was obtained by displaced water volume when the particles were placed into a measuring cylinder filled with water.

Since the density of wood particle is much less than that of water it will float in water, so to increase the density of wood particle up to near to density of water i.e. 997 kg/m^3, we kept the wood particles in water for approximately 4 days. After keeping in water for this duration, the density of wood particles of sizes 3 mm, 5 mm and 7 mm increases to 1142 kg/m^3, 1111 kg/m^3 and 1111 kg/m^3 respectively. As density of wood particles is now more than that of water our solid particles can be used in fluidized bed. So, in the experiment we have used these wood particles as a solid phase.

2.3. Experimental procedure:

The three-phase solid, liquid and gas are regular size wood particles, tap water and oil free compressed air respectively. The scope of the experiment is presented in Table 2.2. The air-water flow was co-current and upwards. Accurately weighed amount of material was fed into the column and adjusted for a specified initial static bed height. Water was pumped to the column at a desired flow rate using water rotameter. The air was then introduced into the column through the air sparger at a desired flow rate using air rotameter. Three calibrated rotameters with different ranges two for water as well as one for air have been used for the accurately record of the flow rates.

Table 2.2: Properties of Bed Material

Material	**$D_{p.}$, mm**	**ρ_s, (kg/m^3)**
Wood	3.7,6.2,8.7	1142,1111,1111

Table 2.3: Mass of Bed Material

Size of particle (mm)	**Height of particles in bed(cm)**	**Mass (kg)**
3	18,23,28	0.79,1.02,1.23
5	18,23	0.785,1.04
7	18,23,28	0.79,1.04,1.28

Table 2.4: Properties of Fluidizing Medium

Fluidizing Medium	**ρ (kg/m^3)**	**μ (Ns/m^2)**
Air at 25^oC	1.168	0.00187
Water at 25^oC	997	0.095

Table 2.5: Properties of Manometric Fluid

Manometric Fluid	**ρ (kg/m^3)**	**μ (Ns/m^2)**
Carbon tetrachloride	1600	0.09

Table 2.6: Operating Conditions

Superficial gas velocity (m/s)	0-0.008
Superficial liquid velocity (m/s)	0-0.06
Static bed heights (cm)	18,23,28

In actual practice, oil free compressed air from a (1 phase, 1 hp, 1420 rpm) used to supply the air at nearly constant pressure as fluidizing gas. This was done by continuously monitoring the pressure in the compressor air tank. The air was injected into the column through the air sparger at a desired flow rate using calibrated rotameter. Water was pumped to the fluidizer at a desired flow rate using water rotameter. Centrifugal pump (1 phase, 1 hp, 2900 rpm, capacity 130 lpm) was used to deliver water to the fluidizer along with a bypass line. Three calibrated rotameters with different ranges two for water as well as one for air were used for the accurate record of the flow rates. Water rotameters used were of the range 0-10 lpm and 10-100 lpm. Air rotameter was of the range 0-10 lpm.

All experiments have been conducted with the column completely filled with water and wood particles and the initial level of manometer adjusted to have zero level. For gas-liquid-solid experiment, with a little flow of liquid close to zero, the air was slowly introduced and gradually increased to the desired flow rate after which the liquid flow rate was increased and the readings were noted down. The procedure was repeated for different values of initial static bed height, particle size and gas velocity. Measurements were conducted at 3 heights: 18 cm, 23 cm, and 28 cm. In some experiment the liquid velocity is kept constant at experimental minimum fluidization velocity and then the gas velocity is varied to see the movement of the particles in the bed.

CHAPTER-3

RESULTS AND DISCUSSION

RESULTS AND DISCUSSION

Hydrodynamic behavior of a fluidized bed is to be known for proper design and operation of a fluidized bed. Results obtained from experiment are analyzed and discussed. In the present study, experiments were conducted to study the hydrodynamics of the gas-liquid-solid fluidized bed containing low density particles i.e. wood particle having density nearer to that of water. Experiments have been conducted using different sizes of wood particles of diameter 0.0037, 0.0062, 0.0087 m having density 1142, 1111, 1111 kg/m^3. Regular size wood particle, tap water and compressed air were used as the solid, liquid and the gas phases. In this experimental study an attempt has been made to examine the hydrodynamic characteristics of fluidized bed by varying a large number of operating variables like liquid velocity, gas velocity, particle size and bed height. The hydrodynamics parameters include bed expansion, pressure drop and minimum liquid fluidization velocity.

3.1. Bed pressure drop and minimum fluidization velocity:

In the present study, pressure drop in the fluidized bed has been measured by using manometers filled with carbon tetrachloride as manometric fluid connected to pressure tapping in the column as described in chapter-2. All experiments have been started with the column completely filled with water and wood particles up to a desired height with the initial level of manometer adjusted to have zero. For gas-liquid-solid experiment with different liquid and gas flow rate has been carried out in two different manners: one with little flow of liquid close to zero, the air was slowly introduced and then gradually increased to the desired flow rate after which the liquid flow rate was increased and the readings were noted down and the other is keeping the liquid flow rate constant at some value may be minimum liquid fluidization and gas flow rate was gradually varied. As the gas velocity is steadily increased, it will reach a critical value, the drag on individual particles increase, and the pressure drop will also increase. Once the bed is fluidized the pressure drop across the bed remains constant, but the bed height continues to increase with increasing flow.

In the three phase fluidization with liquid as the continuous phase, U_{Lmf} is the superficial liquid velocity at which the bed becomes fluidized for a given superficial gas velocity and is called as the minimum liquid fluidization velocity (U_{Lmf}). The minimum liquid velocity required to achieve fluidization is determined from the bed pressure drop vs. liquid velocity plot at a constant gas velocity.

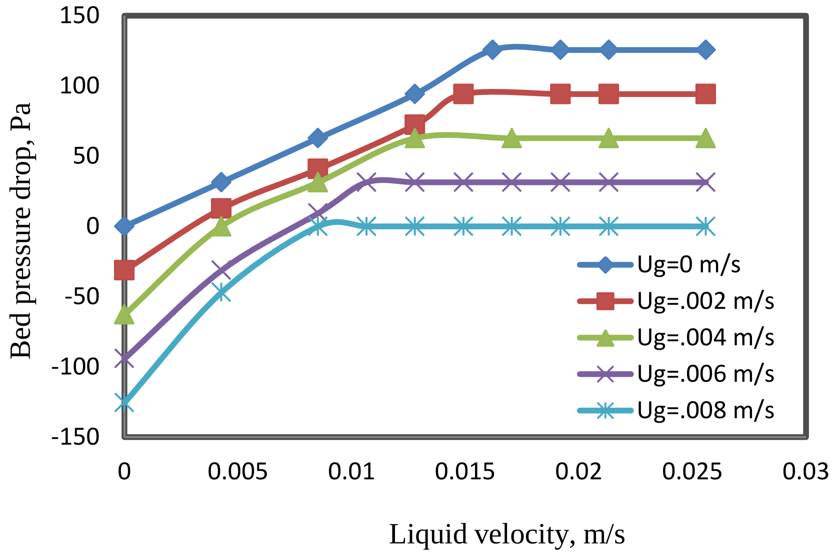

Fig.3.1. Variation of pressure drop with liquid velocity for different values of gas velocity at [Hs = 0.18 m, dp = 3.7 mm]

Visual observation determines U_{Lmf} as either the velocity at which the bed first begins to expand or as the velocity at which any particle within the bed continuously shifts position with neighboring particles (Jena, 2010). The point of intersection of the line of different slope is taken as U_{Lmf}. U_{Lmf} has been obtained from the plot of pressure drop and superficial liquid velocity (Fig.3.1) for liquid-solid system i.e. at gas velocity zero. In Fig. 3.1.the pressure drop after incipient of fluidization has been assumed to be constant. But in actual practice the pressure drop slightly increases with the increase in liquid velocity at a constant gas velocity as the liquid holdup is likely to increase.

Similarly Fig.3.2 shows the variation of pressure drop with liquid velocity at constant gas velocity for Hs = 23 cm and dp = 6.2 mm.

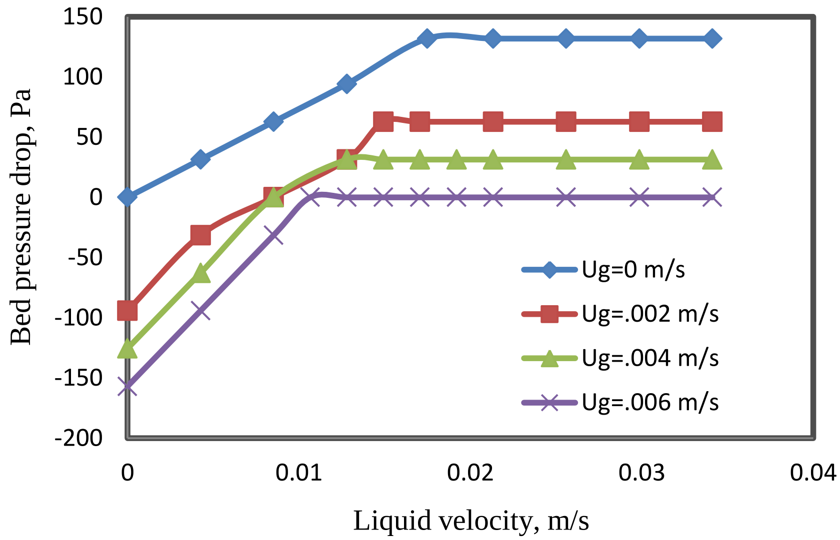

Fig.3.2. Variation of pressure drop with liquid velocity for different values of gas velocity at [Hs = 0.23 m, dp = 6.2 mm]

Fig.3.3. shows the variation of pressure drop with liquid velocity same as Fig.3.1 and Fig.3.2 for Hs=28 cm and dp= 8.7 mm. For this case pressure drop is high as compare to Fig.3.1, because it is clear that minimum fluidization velocity increases with increase in particle size for same liquid flow rate.

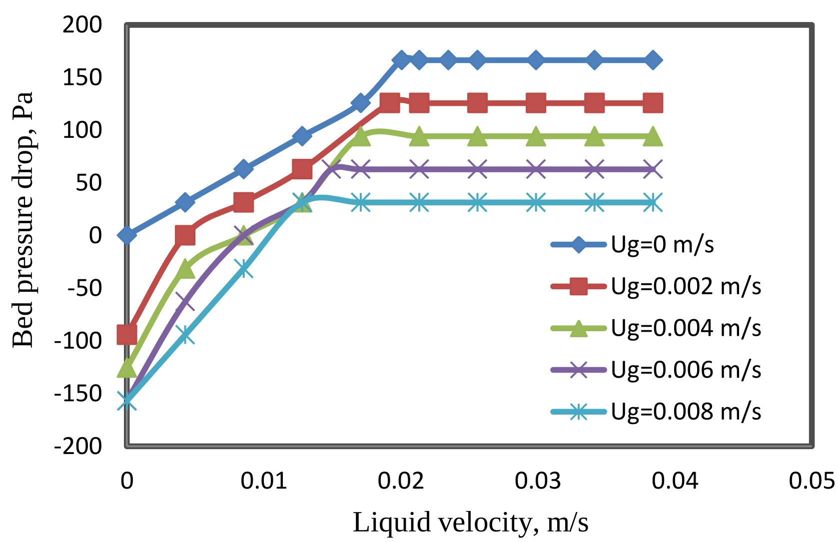

Fig.3.3. Variation of pressure drop with liquid velocity for different values of gas velocity at [Hs = 0.28 m, dp = 8.7 mm]

Experimental values of the pressure drop have been taken at different static bed height for different particle diameter.

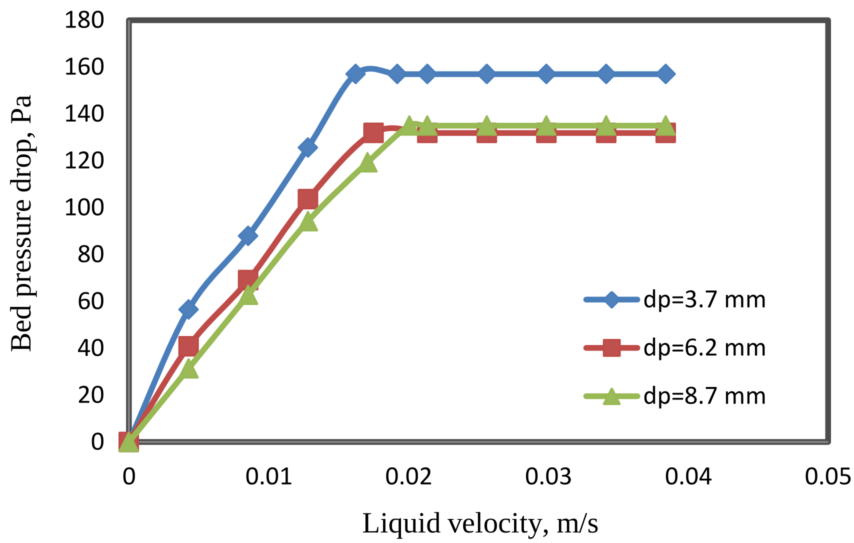

Fig.3.4.Variation of bed pressure drop with liquid velocity for different particle size. [H_s = 0.23 m, U_g=0 m/s]

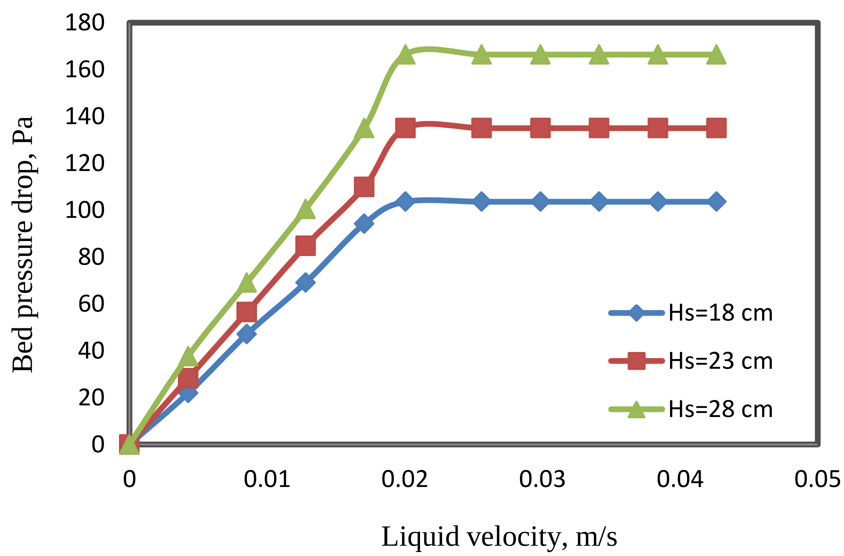

Fig.3.5.Variation of pressure drop with liquid velocity for different static bed height.

[dp=8.7 mm, U_g= 0 m/s]

Fig.3.4. shows the variation of pressure drop with liquid velocity for different particle size. There is small difference in pressure drop for different particle sizes. Smaller size particles are fluidized at higher pressure drop due to higher density and less void fraction as compared to the 6.2 mm and 8.7 mm particles. Pressure drop for 6.2 mm and 8.7 mm particle was observed to be same. . Fig. 3.5 shows variation in bed pressure drop with liquid velocity for different static bed height. Bed pressure drop is the strong function of the static bed height. It is changing for different static bed height at same liquid velocity due to increase in bed mass. From Fig.3.5 we observe that U_{Lmf} is same for different static bed height for particular particle size.

Fig. 3.6 shows a variation in U_{Lmf} with gas velocity for different particle size. Minimum fluidization velocity increases with increase in particle size. With increase in gas velocity it has been observed that the U_{Lmf} decreases.

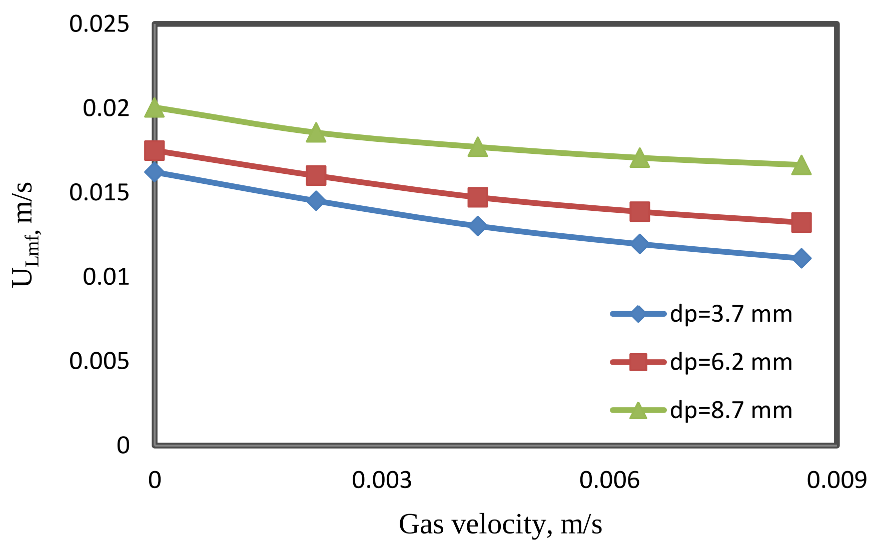

Fig.3.6. Variation of minimum fluidization velocity with gas velocity for different particle size

3.2. Bed expansion:

The knowledge of expanded bed height is essential for sizing of the system. When a liquid or gas is passed through a bed of solid particles at low velocity then the particles do not move, and thus the bed remains at fixed state. At low gas velocities drag force on each particle is also low. For the fluid velocity more than the minimum fluidization condition with increase fluid velocity the bed gradually expands to higher height. The expanded bed height in the present study has been measured by visual observation. The bed expansion study as carried out by varying liquid velocity (at constant gas velocity) at different static bed heights have been presented in graphically in this section.

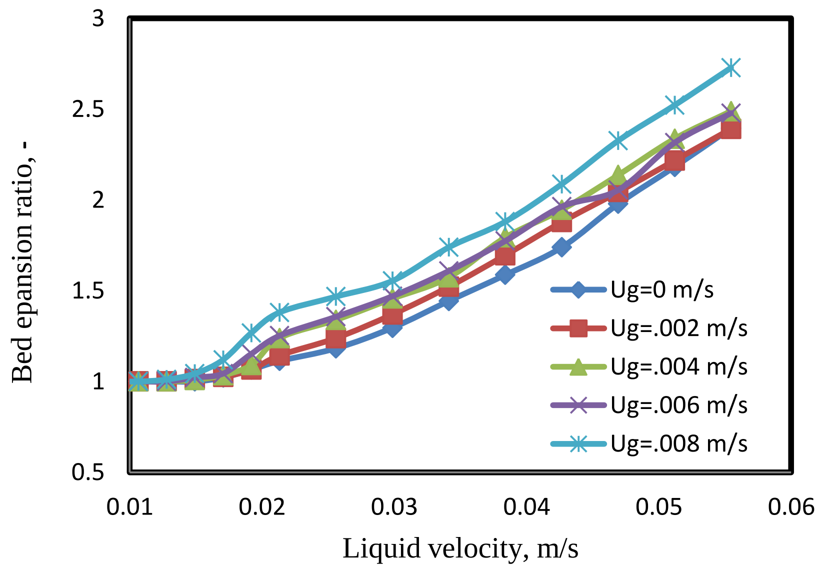

Fig.3.7. Variation of bed expansion ratio with liquid velocity for different values of gas velocity at [Hs = 0.23 m, dp = 3.7 mm]

Fig. 3.7 shows the variation of bed expansion ratio of gas-liquid-solid fluidized bed with liquid velocity at different constant gas velocities. With both increase in liquid and gas velocities the bed expansion ratio increases monotonically with varying slope. At zero liquid velocity when gas is introduced into the bed, the bed expanded but entire particles are not in fluidization, few particles driven by the gas and as the liquid velocity increased there off the bed expanded and then the rate of expansion decreased and again increased.

Figs.3.8 and 3.9 show the variation of bed expansion ratio with liquid velocity at constant gas velocity for particle size 6.2 mm and 8.7 mm at 23 cm, 28 cm height respectively. for both the cases the same phenomena as it was observed for particle size 3.7 mm (Fig.3.7) has been repeated.

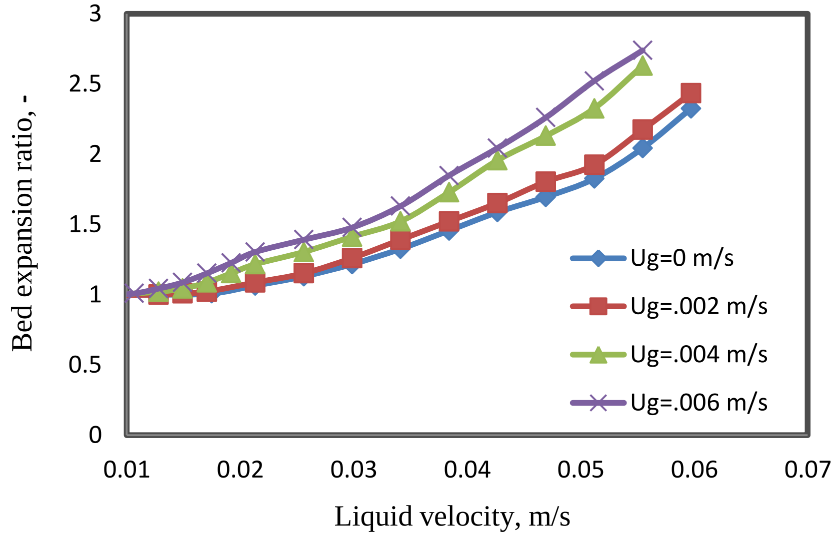

Fig.3.8. Variation of bed expansion ratio with liquid velocity for different values of gas velocity at [Hs = 0.23 m, dp = 6.2 mm]

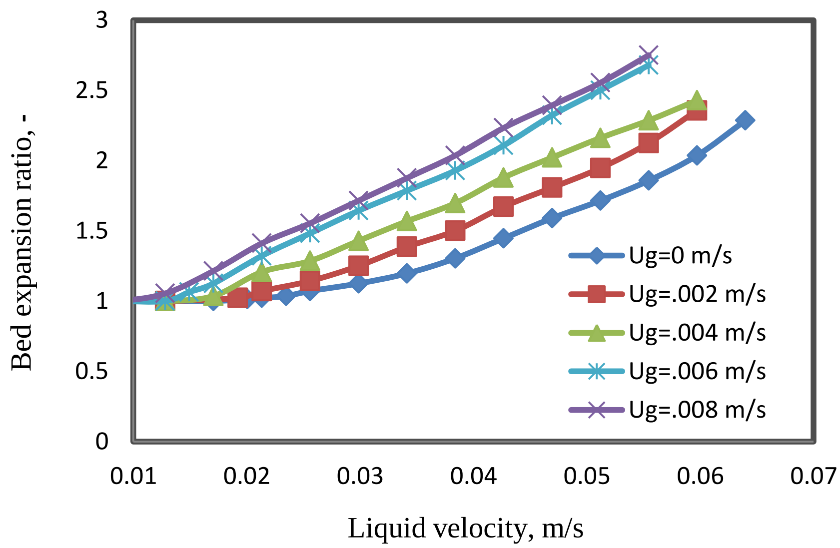

Fig.3.9. Variation of bed expansion ratio with liquid velocity for different values of gas velocity at [Hs = 0.28 m, dp = 8.7 mm]

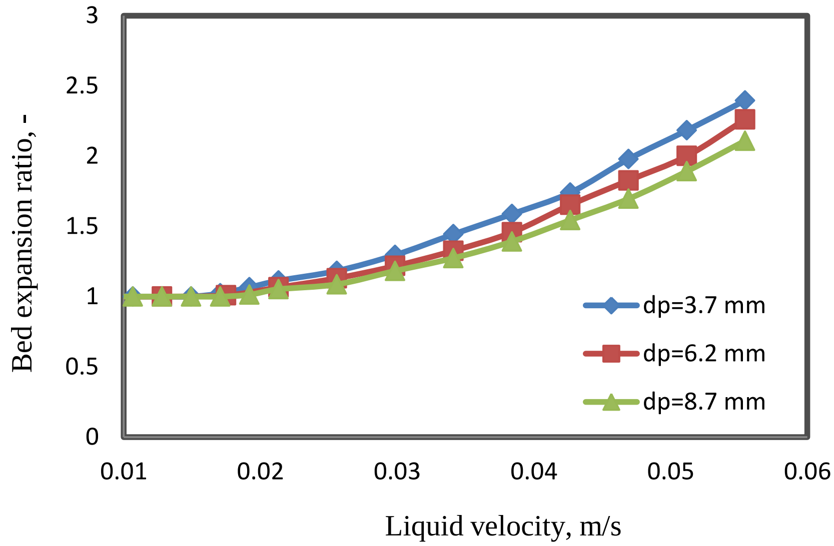

Fig. 3.10.Variation of bed expansion ratio with liquid velocity for different values of particle diameter at [Ug=0 m/s, Hs=23 cm]

Fig. 3.10 shows the variation of bed expansion ratio with liquid velocity for different particle size for the liquid-solid system. The plot shows an increase in bed expansion ratio with decrease in particle size i.e. smaller size particles lifted to higher height in the bed.. Bed expansion ratio is the function of particle diameter.

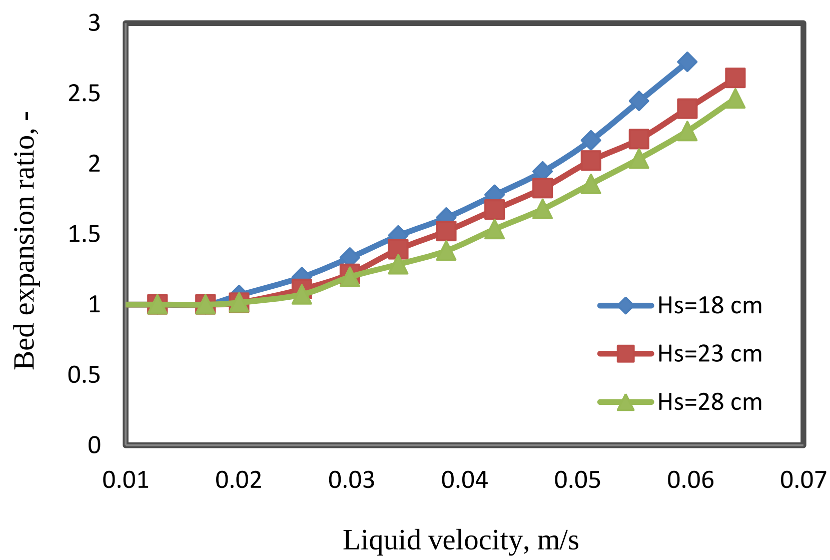

Fig. 3.11.Variation of bed expansion ratio with liquid velocity for different values of static bed height at [Ug=0 m/s, dp = 8.7 mm]

Fig. 3.11 shows an increase in bed expansion ratio with increase in liquid velocity for different static bed height. It is observed that the bed expansion ratio is a function of initial static bed height, thus for a higher initial static bed height the expanded bed height is less for a particular value of liquid velocity.

The correlation developed by Jena (2010) is as follows:

$$\left(\frac{H}{H_s}\right)^{LS} = 16.487 {U_L}^{0.811} {d_p}^{-0.418} {\mu_L}^{-0.041} (\rho_s - \rho_l)^{-0.444} {\emptyset_s}^{-0.700} \quad (8)$$

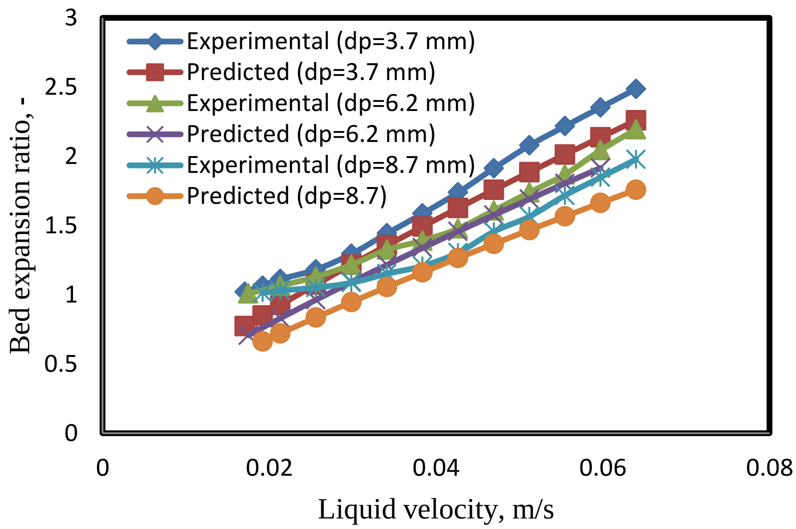

Fig.3.12.Comparison of bed expansion ratio from correlation (eq.8) at different particle size

In Fig.3.12. experimental values of bed expansion ratio has been compared with those values predicted from the correlation (eq.8) of Jena (2010) for liquid solid fluidized bed for different particle size. The experimental values of bed expansion ratio at liquid velocity are agree well with Jena (2010) as shown in Fig.3.12. It clearly shows that with increase in particle size bed expansion ratio decreases.

The experimental data for bed expansion ratio has been correlated by non-linear regression analysis and the following equation has been developed which can be used for prediction of bed expansion ratio.

$$H = 0.6443 \frac{H_s}{D_c}^{-0.0377} \frac{\rho_s}{\rho_L}^{0.187} \frac{d_p}{D_c}^{-0.0366} \frac{U_L}{U_{Lmf}}^{.424} \frac{U_g}{U_{Lmf}}^{-0.0274} \quad (9)$$

The values of experimental bed expansion ratio have been compared with those calculated from equation (9). Most of the predicted values of bed expansion ratio from eq. (9) agree within 20% with the experimental values.

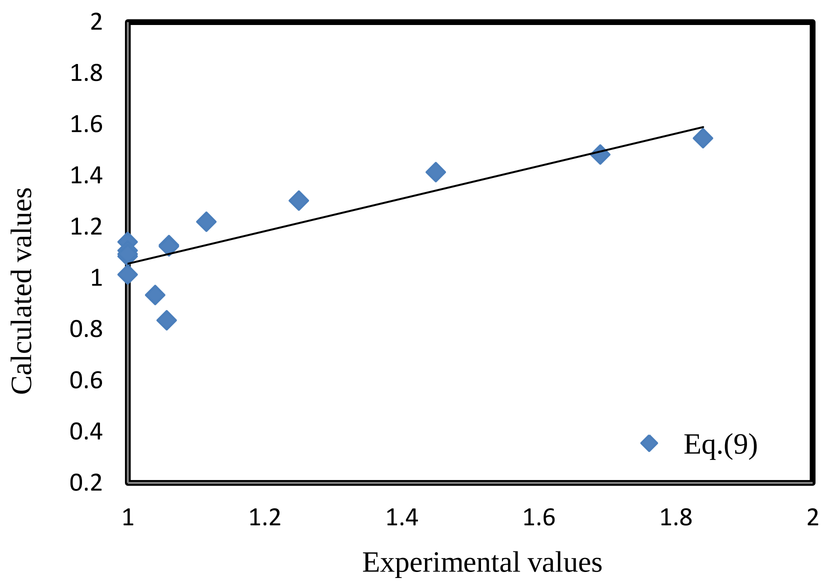

Fig.3.13. Comparison of experimental values of bed expansion ratio with calculated values (eq.9).

3.3. Development of correlation based on factorial design analysis

Factorial design analysis is used to study the interaction effects of variables on each other, which would not be found by conventional experimentation and find out the effect of each of the variables quantitatively on the response (Jena et.al. 2009). In this method the experiments have been conducted twice or thrice at two levels of each of the operating variables i.e. one at lower level (-1 level) and the other at higher level (+1 level). Factors considered for factorial design have been presented in Table. The variables, which affect the bed expansion ratio in fluidization are initial static bed height, particle size, and liquid and gas velocities. Thus total numbers of experiments required for the four variables at two levels is 16.

Table 3.1: Experimental lay out for the factorial design

S.no.	Name of variable	Variable (symbol)	Factorial design symbol	Min. level (-)	Max. level (+)	Magnitudes of variable
1	Static bed height (m)	H_s	A	0.18	0.28	0.18,0.23, 0.28
2	Particle diameter((m)	D_p	B	0.0037	0.0087	0.0037, 0.0062, 0.0087
3	Gas velocity(m/s)	U_g	C	0.0021	0.0064	0.0021-0.0064
4	Liquid velocity (m/s)	U_l	D	0.0021	0.0519	0.0021-0.0519

In this method general form of the model equation,

$$Y= (b_0+b_1A+b_2B+b_3C+\ldots+b_{12}AB+b_{13}AC+\ldots+b_{123}ABC+\ldots+b_{1234}ABCD) \qquad (10)$$

In the equation (10) first term is constant, and other terms are the main effect and interaction effect of two variables, three variables, four variables, respectively.
Coefficients are calculated by the Yates standard technique (Davies, 1978) as;

$$bi = \sum \frac{\alpha_i Y_i}{N}$$

where b_i is the coefficient, Y_i is the response, α_i is the level of the variable and N is the total number of treatments.

Calculations of the level of variables:

A: Level for static bed height= (Static bed height – 0.23)/0.05

B: Level for particle diameter = (Particle diameter - 0.005)/0.002

C: Level for gas velocity = (Gas velocity – 0.004266)/0.002133

D: Level for liquid velocity = (Liquid velocity – 0.02663)/0.0245

The experimental data based on factorial design and the nature of the effects has been presented for bed expansion ratio in Table.3.2.

Table 3.2: The effect of parameters on bed expansion ratio as per factorial design

s.no.	TC*	Exp. bed expansion	1	2	3	4	Effect (4)/8	Sum of squares $(4^2)/16$	Percent contribution
1	1	1	2	4	8	25.70			
2	A	1	2	4	17.70	-1.48	-0.19	0.14	$2.14^{\#}$
3	B	1	2	8.36	0	-0.72	-0.09	0.03	0.51
4	AB	1	2	9.34	-1.48	0.10	0.01	0.001	0.01
5	C	1	4.51	0	0	0.98	0.12	0.06	$0.94^{\#}$
6	AC	1	3.85	0	-0.72	-0.24	-0.03	0.004	0.06
7	BC	1	4.70	-0.62	0	0.60	0.08	0.02	0.35
8	ABC	1	4.64	-0.86	0.10	-0.14	-0.02	0.001	0.02
9	D	2.44	0	0	0	9.70	1.21	5.88	$91.95^{\#}$
10	AD	2.07	0	0	0.98	-1.48	-0.19	0.14	$2.14^{\#}$
11	BD	2.05	0	-0.66	0	-0.72	-0.09	0.03	0.51
12	ABD	1.80	0	-0.06	-0.24	0.10	0.01	0.001	0.01
13	CD	2.56	-0.37	0	0	0.98	0.12	0.06	$0.94^{\#}$
14	ACD	2.14	-0.25	0	0.60	-0.24	-0.03	0.004	0.06
15	BCD	2.54	-0.42	0.12	0	0.60	0.08	0.02	0.35
16	ABCD	2.10	-0.44	-0.02	-0.14	-0.14	-0.02	0.001	0.02

(TC*= Treatment combination, $^{\#}$ significant variable)

The A, C, D variables are most significant. The AD, CD interaction is significant for development of equation.

The following equation has been developed from statistical design:

$$Y=1.60625-0.0925A+0.06125C+0.60625D-0.0925AD+0.06125CD \quad (11)$$

The values of the coefficient indicate that the magnitude of the effect on the variables and the sign of the coefficients give the direction of the variable whether it is increasing or decreasing. There is a positive coefficient indicating an increase in the values of the responses with increase in the values of the variables and a negative coefficient indicates that the response decreases with increase in the value of the variable.

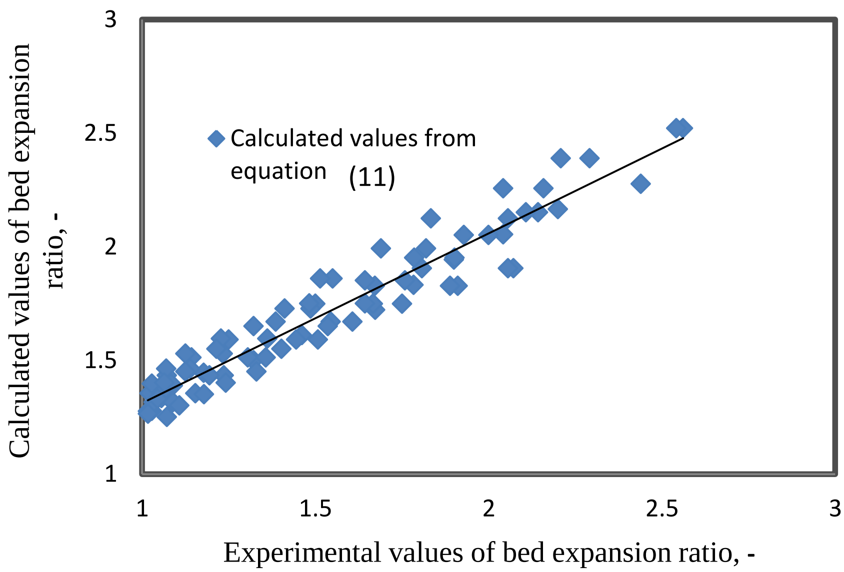

Fig.3.14. Comparison of the experimental data with calculated data

The calculated values of Bed expansion from equation (11) have been compared with experimental data. The comparison of data for bed expansion is presented in the Fig.3.14 where R-square value of 0.928 is obtained from the graph, which represents good agreement.

CHAPTER-4

CONCLUSION AND FUTURE SCOPE

4.1 CONCLUSIONS

Hydrodynamic characteristics of three-phase fluidized bed with low density wood particles have been determined by experiment.

- Pressure drop increases where liquid velocity increased at constant gas velocity and after attaining the minimum fluidization velocity the pressure drop across the bed remains constant. Pressure drop increases with initial static bed height (bed mass).
- The minimum fluidization velocity (U_{Lmf}) is not a function of initial static bed height but is function of particle size and it decrease with gas velocity but increases with particle size.
- Increase in gas and liquid velocities, the expansion ratio increases. Bed height is the strong function of liquid and gas velocity and increases monotonically.
- Bed expansion ratio decreases with increase in particle size and bed height.
- The experimental values have been compared with those predicted by the correlations and have been found to be in good agreement.

4.2 FUTURE SCOPE

The followings are the recommendations for the future work.

- Experimental observation of the different sizes of the low density particles for understanding the effect of particle size, particle density on hydrodynamic properties like bed expansion, pressure drop and minimum fluidization velocity etc.
- Development of more generalized correlations for proper study of the hydrodynamic parameters of regular size particles of different sizes.

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