INFLUENCE OF ANGLE OF AIR INJECTION AND PARTICLES IN BED HYDRODYNAMICS OF SWIRLING FLUIDIZED BED

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BY

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DECLARATION

I hereby declare that the synopsis of the thesis entitled **"INFLUENCE OF ANGLE OF AIR INJECTION AND PARTICLES IN BED HYDRODYNAMICS OF SWIRLING FLUIDIZED BED"** is based on the original work done by me under the supervision of Dr. Sreejith P.S., Professor, Division of Mechanical Engineering , School of Engineering, Cochin University of Science and Technology, Kochi-22. No part of this thesis has been presented for any other degree from any other institution.

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CERTIFICATE

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ABSTRACT

The thesis presented here unveils an experimental study of the hydrodynamic characteristics of swirling fluidized bed viz. pressure drop across the distributor and the bed, minimum fluidizing velocity, bed behaviour and angle of air injection. In swirling fluidized bed the air is admitted to the bed at an angle 'θ' to the horizontal. The vertical component of the velocity $v \sin \theta$ causes fluidization and the horizontal component $v \cos \theta$ contributes to swirl motion of the bed material.

The study was conducted using spherical particles having sizes 3.2 mm, 5.5 mm & 7.4 mm as the bed materials. Each of these particles was made from high density polyethylene, nylon and acetal having relative densities of 0.93, 1.05 and 1.47 respectively.

The experiments were conducted using conidour type distributors having four rows of slits. Altogether four distributors having angles of air injection (ϕ) - 0^0, 5^0, 10^0 & 15^0 were designed and fabricated for the study. The total number of slits in each distributor was 144. The area of opening was 6220 mm^2 making the percentage area of opening to 9.17. But the percentage useful area of opening of the distributor was 96.

The experiments on the variation of distributor pressure drop with superficial velocity revealed that the distributor pressure drop decreases with angle of air injection. Investigations related to bed hydrodynamics were conducted using 2.5 kg of bed material. The bed pressure drop measurements were made along the radial direction of the distributor at distances of 60 mm, 90 mm, 120 mm & 150 mm from the centre of the distributor. It was noticed that after attaining minimum fluidizing velocity, the bed

pressure drop increases along the radial direction of the distributor. But at a radial distance of 90 mm from the distributor centre, after attaining minimum fluidizing velocity the bed pressure drop remains almost constant. It was also observed that the bed pressure drop varies inversely with particle size as well as particle density.

An attempt was made to determine the effect of various parameters on minimum fluidizing velocity. It was noticed that the minimum fluidizing velocity varies directly with angle of air injection (ϕ), particle size and particle density.

The study on the bed behaviour showed that the superficial velocity required for initiating various bed phenomena (such as swirl motion and separation of particles from the cone at the centre) increase with increase in particle size as well as particle density. It was also observed that the particle size and particle density directly influence the superficial velocity required for various regimes of bed behaviour such as linear variation of bed pressure drop, constant bed pressure drop and sudden increase or decrease in bed pressure drop.

Experiments were also performed to study the effect of angle of air injection (ϕ). It was noticed that the bed pressure drop decreases with angle of air injection. It was also noticed that the angle of air injection directly influence the superficial velocity required for initiating various bed phenomena as well as the various regimes of bed behaviour.

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NOMENCLATURE

Ar	Archimedes number
a_m	Mean area of the bed
br	Width of the ridge
D	Bed diameter
c	Chord length
d	Orifice diameter
dp	Mean particle diameter
g	Acceleration due to gravity
Ha	Head of air in metres
HDPE	High density polyethylene
Hw	Head causing flow in metres of water
Hs	Static bed height
Ht	Overall bed height
K	Proportionality constant
k	Fraction of bed weight supported by fluidizing gas
L	Effective width of the distributor
l	Length of the slit
Mb	Mass of the bed
m_{fh}	Radial mass flow rate per unit of bed height
N	Total number of orifices in a distributor plate
n	Number of operating orifices

nr	Number of rows in the distributor
P.R	Pressure drop ratio
Q	Volume flow rate
R	Aspect ratio
Rc	Critical aspect ratio
Rp	Critical pressure drop ratio
Re	Reynolds number
ri	Inner radius of the bed
ro	Outer radius of the bed
rc	Radius of the distributor cone
S	Orifice spacing
T	Temperature of air
t	Distributor plate thickness
U	Superficial gas velocity
Uo	Tuyere gas velocity
Umf	Minimum fluidizing velocity
Va	Actual volume of air discharged
ΔPb	Bed pressure drop
Δpbmax	Maximum bed pressure drop
ΔPd	Distributor pressure drop
Δpdmin	Minimum distributor pressure drop
ΔPv	Venturimeter pressure drop
Є	Voidage

ϵ_s	Static voidage
ϵ	Voidage
ϵ_s	Static voidage
θ	Inclination of the slit with the horizontal
μ_f	Dynamic viscosity of fluid
ρ_a	Density of air
ρ_b	Bed density
ρ_f	Density of fluid
ρ_w	Density of water
ρ_p	Density of fluidized particle
ϕ	Inclination of the slit with the radius of the distributor
ψ	Proportionality constant
ω_o	Angular velocity

CHAPTER 1

INTRODUCTION

1.1 GENERAL

The process of imparting fluid like properties by forcing them to suspend in a fluid; either a gas or liquid is termed as fluidization. This can be accomplished by passing the fluid through a bed of particles. The easiest method to determine whether a bed is fluidized or packed is the velocity at which the fluid passes upward through an unrestrained bed of particles. Both types of bed require a containing vessel with a porous base through which the fluid can be introduced to the bed. This porous base can be of various designs ranging from a plate having a number of small holes to a block of bubble caps. The most important function of the porous base is to distribute the fluid uniformly across the base of the bed. Hence the porous base is universally called a distributor. The most desirable property of the distributor is to distribute the gas uniformly without excessive resistance to the gas flow. This is achieved by suitable choice of pressure drop across the distributor with respect to that across the bed and by proper distribution of gas inlet points. The pressure drop across the distributor is dependent on the type of orifices, the free area and the gas flow rate.

When the fluid flows through the space between the particles, it exerts a drag force on the particles. This force may be sufficiently large to disturb the arrangements of the particles within the bed. On increasing the velocity of the fluid in the upward direction, a stage will reach where the entire weight of the particles is supported by this fluid drag. The bed is then said to be incipiently fluidized and it exhibits fluid like properties. The velocity corresponding to this condition is known as minimum fluidization velocity. Point of minimum fluidization specifies the necessary minimum amount of fluid to reach the boundary between fixed bed and fluidized bed. Interparticle forces such as cohesive force, capillarity force and electrostatic interactions play a key role in determining minimum fluidization velocity. The bed can flow under a hydrostatic head; the free surface will remain horizontal if the containment is tilted and low density objects will float.

A completely fluidized state is one in which the contact between the gas and particles is perfect without any channeling. This can be achieved by an adequate supply of gas and proper distributor design. An effective design of the distributor is one which achieves uniform distribution of gas without excessive resistance to the gas flow. The quality of fluidization is specified by the pressure drop ratio which is defined as the ratio of pressure drop across the distributor to the pressure drop across the bed. The value of this ratio at minimum fluidization was found to be dependent on bed height, and its degree depends on distributor design. Investigations conducted by Medlin and Jackson (1975) revealed that for a porous distributor, bed diameter has a significant effect on the pressure drop ratio. But when the pressure drop ratio is greater than 0.15, stability was independent of diameter. It is desirable to obtain a value for the critical pressure drop ratio, for stable operation under different operating conditions. Seigel (1976) found that the value of critical pressure drop ratio is 0.072 for smaller particles for Reynolds number less than 10.

Experimental investigations were conducted by D. Sathyamoorthy et al. (2003) and concluded that aspect ratio also has a significant effect on the quality of fluidization. Aspect ratio is defined as the ratio of bed height to the diameter of the bed. The bed height is usually taken at the minimum fluidization velocity. If the aspect ratio is less than or equal to unity the bed is considered as shallow, whereas a bed having a value greater than one is regarded as a tall or deep bed. There is a critical value of aspect ratio where the quality of fluidization is maximum. This value is influenced by operating velocity and the type of the distributor. Y.K.Mohanty (2007) reported that the quality of fluidization can also be specified by the terms like fluctuation ratio and expansion ratio. Fluctuation ratio is defined as the ratio of the highest to the lowest bed heights of the fluidized bed in expansion. The expansion ratio is defined as the ratio of the average of highest and lowest bed heights to the static bed height for a particular gas flow rate. Lower static bed height and smaller particle size lead to lower fluctuation ratio which signifies better fluidization quality.

All types of beds cannot be fluidized satisfactorily. Beds of different kinds of particles can behave differently, when fluidized. Hence it is desirable to categorize particles according to the way they behave during fluidization. Geldart (1973) suggested that the mean size and density of the particles are the main parameters that influence the bed behavior. Accordingly the particles can be classified into four categories viz. type A, B, C & D.

Fluidized bed drying has the advantage of high intensity of drying and high thermal efficiency. High heat and mass transfer rates between the gas and the particles are possible due to large contact area between the particles and the gas. It also provides rapid mixing of solids and nearly uniform moisture content distribution throughout the bed. Due to rapid drying it has been considered as an economical drying method compared with other drying techniques. In addition solids may be added or removed during operation. Hence fluidization technology is extensively applied in process industries. But it is necessary to pump large volume of fluidizing medium through the bed. This can lead to various limitations in conventional fluidized beds such as slugging, channeling, elutriation of solid particles and limitation in the size of the bed materials.

Various attempts were made to surmount the limitations and improve the performance of the conventional fluidized bed. This led to the development of different types of fluidized beds, viz. vibro-fluidised bed, magneto fluidized bed, tapered fluidized bed, spouted fluidized bed, centrifugal fluidized bed, swirling fluidized bed etc.

The swirling fluidized bed is a modified version of centrifugal fluidized. Here the gas is injected into the bed at an angle 'θ' with the horizontal. The velocity of the gas can be resolved into the vertical component $v \sin \theta$ and the horizontal component $v \cos \theta$. The vertical component causes fluidization and the horizontal component in turn causes swirl motion of the bed material. If the bed is sufficiently deep this horizontal momentum of the jet deteriorates as the jet penetrates deep into the bed and finally it ceases at a certain height above the distributor. But in a shallow bed the velocity of the jet leaving the bed will have two components. In this case the bed will be swirling as a single mass. The particles will be swirling so intensively that a high shear between the particles and the gas is obtained. As a result all transport processes are enhanced. Unlike in a conventional fluidized bed, which has only two regimes of operation, three and sometimes even four regimes of operation is present in a swirling fluidized bed. For deep beds, a two layer fluidization regime occurs with a lower swirling layer and a top bubbling layer. A striking feature that distinguishes the swirling bed from a conventional fluidized bed is that the bed pressure drop in the swirling mode increases with air velocity.

Even though there is extensive commercial application of swirling fluidized bed, the literature available on this topic is regular scanty. Experiments were conducted by Vikram et al. (2003) in swirling fluidized bed and observed that the velocity of the swirling particles increased with radial distance. Paulose M.M (2005) conducted studies in swirling fluidized bed using different distributors and concluded that the distributor pressure drop was lowest for the vane type distributor. It was further noticed that the distributor pressure drop decreases with increase in vane angle.

Swirling fluidized beds have several advantages over conventional fluidized beds. In the swirling zone, no bubbles are formed and no gas bypassing occurs. The distributor pressure drop is also less compared with conventional fluidized beds. An added advantage is that the size of the reactor can be reduced. Due to improved retention of particles and temperature more thorough processing is possible. Large particles (Geldart D type) which are difficult to fluidize in a conventional bed can be effectively fluidized in swirling fluidized beds. Hence swirling fluidized bed can be effectively utilised for drying of agricultural produces as well as ayurvedic tablets. Since the swirling flow leads to the generation of turbulence which promotes mixing between particles and fluid, it is an effective method for incineration process.

However the swirling fluidized bed has certain limitations also. In a swirling fluidized bed due to vortex formation followed by swirl motion only the outer annular space could be utilised. This leads to reduction in useful area of the distributor. The superficial velocity required for starting of swirl motion increases with the weight of the bed. At higher values of air velocity a portion of the air may get by-passed through the inner region of the distributor.

Swirling fluidized bed employing conidour type distributor is a new concept which facilitates an increase in useful area of the distributor. It was observed that in this case the distributor pressure drop was comparatively lower for the distributor provided with slits opening towards the outer periphery of distributor compared with that provided with slits opening towards the inner periphery. Studies have been conducted on the effect of particle size and particle density in the bed hydrodynamics of conventional bed. But till date no such studies were conducted in swirling fluidized bed.

In the present work an experimental study was conducted to determine the effect of angle of air injection and the particle density as well as particle size in the bed hydrodynamics of swirling fluidized bed.

1.2 SCOPE OF THE THESIS

The contents of the various chapters included in the thesis are mentioned below

A general introduction of the thesis is presented in chapter 1. Chapter 2 contains a brief review of the literature on swirling fluidized bed and the objectives of the present investigation. The design details of the different distributors fabricated for the required study is revealed in chapter 3. The details of the experimental set up and the related procedure adopted for the current investigation is described in chapter 4. Chapter 5 contains the results of the various experiments conducted for the present study and the discussions based on it. The major conclusions based on the investigations as well as certain suggestions for further work constitute the contents of chapter 6.This are followed by a list of references and appendices.

CHAPTER 2

LITERATURE REVIEW

2.1 INTRODUCTION

A critical review of the available literature on the hydrodynamic characteristics of a fluidized bed, which are relevant to the scope of the present study, is presented in this chapter. The literature review is based on the various parameters like minimum fluidizing velocity, distributor pressure drop, bed pressure drop etc. that influence the behavior of the fluidized bed.

2.2 DISTRIBUTOR

The vital component in any fluidized bed is the distributor. The major function of the distributor is to distribute the fluidizing gas across the base of the bed so that it is maintained in the fluidized condition over the whole of its cross section. The stable operation of any fluidized bed is attributed to the effective design of a distributor. The various factors that influence the design of a distributor are:

- pressure drop across the distributor,
- pressure drop across the bed,
- bed weight,
- bed expansion ratio,
- fluidizing velocity,

- particle size,
- particle size distribution,
- geometry of the distributor and
- fraction of the distributor area open for gas flow.

The performance of a distributor during operation determines the success of any industrial application.

The desirable characteristics of a good distributor are

1. It should distribute the gas uniformly in the bed.
2. It should have low distributor pressure drop at the operating velocity.
3. It should not permit the drain of solids
4. It should be strong enough to withstand both thermal and mechanical stresses.
5. It should have minimum particle attrition.
6. It should have the ability to prevent particle attrition.

It is practically impossible to achieve all the above mentioned qualities in a single distributor because some of these qualities can be attained at the expense of other qualities. Depending on the type of the process for which a particular distributor is adopted the related characteristics can be achieved in it.

2.3 DISTRIBUTOR PRESSURE DROP

The basic relationship between the quality of fluidization and the distributor is governed by the various characteristics of the bubbles produced in the system during the entrance of the gas into the bed through openings in the distributor. Initial size of the bubble, bubble frequency, bubble rise velocity and bubble coalescence are some of the parameters which control the quality of fluidization. Generally small orifice openings in the distributor plate produce small bubbles which render better and uniform fluidization. But such distributors are associated with large pressure drop.

The circulation of particles in the bed is dependent on the size of the bubbles formed. This in turn is mainly influenced by the distributor. A sufficiently large gas velocity is required to overcome the pressure fluctuations above the bed. This leads to an increase in pressure drop across the distributor. Thus the pressure drop across the distributor shall be sufficiently large to ensure stable operation. But above a depth of one metre, the distributor has only little influence on the bed. The pressure drop across the distributor is determined by the type of orifices, the free area and the gas flow rate. A.E.Qureshi and D.E Creasy (1978) proposed the following equation for calculating the pressure drop in the presence of a bed.

$$\Delta P_d = 1.04\,(d/t)^{1/4}\,U_0^{\,2}/2g \qquad (2.1)$$

S.C.Saxena et al. (1978) conducted experiments using porous plate distributor, two bubble cap distributors of different geometries and four Johnson screen distributors in 30.5cm X 30.5cm square fluidized bed. They found that the distributor pressure drop increased with fluidizing velocity, decreased with percentage open area of the distributor and is independent of the bed weight or height for a given distributor design. But Chen-Song Chiyang and Cheng -Chung Huang (1991) conducted investigations using a perforated plate distributor and found that the presence of the bed increased the distributor pressure drop.

Yacono and Angelino (1978) conducted comparative studies on the influence of ball distributor and porous plate distributor in bubble behavior. They observed that a very homogenous fluidization could be achieved with ball type distributor. The distributor pressure drop was very small and the permeability to dust transported by the gas stream was high. But any unexpected fluidization of the balls will destroy the distributor.

The pressure drop ratio which is defined as the ratio of the pressure drop across the distributor plate to the pressure drop across the bed is an important parameter that affects the design of distributors. The quality of fluidization is often related to the value of pressure drop ratio. The pressure drop ratio was found to increase rapidly with increase in fluidization velocity. The value of this ratio at minimum fluidization was found to be dependent on bed height and its degree depends on distributor design.

It is found that many beds operate successfully with values of R as low as 0.1 or less. It is desirable to obtain a value for the critical pressure drop ratio, Rc for stable operation under different operating conditions. For deep beds or of high density materials, Agarwal et al. (1962) recommended a minimum value of 0.10 for 'R'. But a minimum distributor pressure drop of 350 mm of water is recommended for shallow beds.

Siegel (1976) developed a mathematical model to determine the value of Rc and found that for smaller particles, having $Re < 10$ gives a value of $Rc = 0.72$. In a theoretical paper on stability, Medlin and Jackson (1975) showed that for a porous distributor, bed diameter 'D' has a significant effect on the pressure drop ratio. They showed that small beds ($D < 0.1m$) could be stable for values of R as low as 0.01, but when $R > 0.15$, stability was independent of diameter. Hiby (1967) suggested a minimum value of $R = 0.3$ for a porous plate, to have uniform fluidization. He also reported that the value of Rc increased moderately with particle size. It is quite evident from the literature that different investigators propose different values of Rc.

Otero and Munoz (1974) conducted studies on the fluidization quality, pressure drop and solids flow back through the bubble cap type plate distributor. On the basis of these studies they reported that the solids flow back is due to the bed pulsations and depends on the particle size and the diameter and inclination of the holes in the cap.

Industries often employ multi-orifice type distributors. Experimental investigations on multi orifice type distributors revealed that the number of operating orifices is determined by the gas flow rate, bed height, type of bed material and the free area of the distributor.

Sathyamoorthy and Rao (1984) suggested the following equation to determine the superficial gas velocity at which all the orifices of a distributor in a uniformly fluidized bed become operative.

$$U_M = U_{mf} \{2.65 + 1.24 \log_{10}(Ut / U_{mf})\} \quad (2.2)$$

The pressure ratio can be calculated from U_{mf} and U_M using the equation

$$\Delta Pd / \Delta Pb = C\,(U_{mf} / U_M - U_{mf})^C \quad (2.3)$$

The value of constant C = 2.

A more precise relationship for the above equation was obtained using experimental correlation as

$$\Delta Pd / \Delta Pb = 2.7 \, (U_{mf} / U_M . U_{mf})^{2..32} \tag{2.4}$$

If 'N' is the total number of orifices present in a plate and n is the number of operating orifices at any gas flow rate for a given system, the ratio n/N can be given by the equation

$$\ln (1 - n/N) = - K \, (U . U_{mf} / U_{mf}) \tag{2.5}$$

The proportionality constant 'K' may be expressed as a function of the pressure drop ratio. Experiments were done by Sathiyamoorthy and Rao (1977) using + type and Y type distributors and revealed that for a particular bed material the value of 'K' decreases with increase in bed height and also with increase in the number of orifices in the distributor. It was also observed that at a certain gas velocity, orifices near to central region of the distributor operate first, and with a subsequent increase in the gas flow rate orifices towards the outer periphery of the distributor operate. This is due to the reason that the resistance to the flow of gas increases as the distance of an orifice from the centre of the distributor increases. When all the orifices become operative the flow in each orifice is almost the same.

Whitehead et al. (1971) conducted studies on operation of multi-orifice plate distributors. Detailed investigations were carried out to determine the minimum gas velocity at which all the orifices become operative. Fakhimi et al. (1971) proposed a mathematical model to predict the number of active orifices at any given flow rate. They found that the ratio (ΔPd min / ΔPb) is a function of the orifice spacing, overall bed height, mean particle diameter and incipient fluidizing velocity at which all the orifices are operative. Kassim (1972) noted certain discrepancies between the observations of Whitehead and the calculated values using Fakhimi's relationship, when working with materials which had a low minimum fluidizing velocity.

Experiments were conducted by Upadhyay (1981) using multijet tuyere distributors in a square fluidized bed of size 0.305 m* 0.305 m. Distributors having different number of slits and slit width were investigated and the distributor pressure drop was correlated in terms of tuyere gas velocity. Based upon the experimental investigations they developed the following relationship to predict the distributor pressure drop.

$$\Delta Pd = C\, U_0^{\,n} \tag{2.6}$$

The constants C and n will depend upon the slit width.

Investigations were done by Wen et al. (1978) on dead zone heights near the distributor plate in two dimensional and three dimensional fluidized beds. The results from two dimensional beds indicate that the dead zone height is dependent on gas velocity, distributor type, orifice pitch, and orifice diameter and particle size. The results from three dimensional beds confirm the importance of orifice pitch and indicate that the behavior of the two dimensional bed cannot be readily extrapolated quantitatively to three dimensional cases.

To eliminate severe agglomeration problems in a process for reactive modification of a polymer in an industrial fluidized bed Brik et al. (1990) designed a distributor with horizontal jets and inclined surfaces (HJIS). This distributor design produced sufficient mixing of the gas and the particles to eliminate both the mass and heat transfer problems. In this design of the distributor two factors prevented the solid polymer particles to remain stagnant on the exposed surfaces of the gas distributor. First the inclined surfaces or tents utilized gravity and the angle of repose of the particles to prevent stagnation of polymers above those areas. Second there was no stagnation in the other flat areas between those tents due to the sweeping action of the horizontal jets before they dissipated and the gas entered the bulk of the bed to fluidize the particles. They developed a design procedure and equations to determine the size and location of the tents and their orifices to produce uniformly active jets that could prevent powder stagnation and gas distributor blockages.

Hiby suggested that the pressure drop through the distributor should be at least 30% of that through the bed to prevent uniform fluidization. Agarwal et al. (1962) recommended a pressure drop across the distributor of 10% of the bed pressure drop when the bed is deep or of high density material. For shallow beds of low density material, it is recommended that the pressure drop through the distributor should not fall below 3.45 kN/m^2.

Michael Wormsbecker and Todd Pugsley (2007) conducted experimental investigations on the influence of distributor design on the hydrodynamic of a fluidized bed. Three types of distributor designs viz. Dutch weave mesh distributor, Perforated plate distributor and Punched plate distributor were investigated for bed weight varying from 1 kg to 3 kg at superficial velocities of 1.5 m/s and 3 m/s. The punched plate showed improved performance over the other designs.

2.4 BED PRESSURE DROP

The pressure drop across the bed is one of the most frequently measured variables in the studies of the properties of fluidized beds. It is a measure of the quality of fluidization. A rise in bed pressure drop with increase in gas flow over the fluidized bed region indicates slugging whereas a decrease in pressure drop suggests channeling. The pressure drop across the bed depends on the type and the quantity of bed materials and generally it is considered equal to the weight of the bed material per unit cross sectional area of the bed.

Investigations were done by J.H de Groot et al. (1967) and reported that for tall beds, the actual pressure drop across the fluidized bed is lower than that calculated on the bed weight per unit area basis while for shallow beds it is equal to the calculated value.

J.H Shing Yang et al. (1987) developed a mathematical model for predicting the pressure drop ratio in shallow fluidized beds (ie. the ratio of bed pressure drop to static bed pressure) and correlated by experiments. Static bed pressure drop represents the total weight of the fluidized particles divided by the cross sectional area of the bed.

The relationship for pressure drop ratio is given by

$$PR\ (\text{pressure- drop ratio}) = Pb/\{\rho_p(1-\epsilon s)Hs\} \tag{2.7}$$

They further observed that the pressure drop ratio generally increases with increasing particle size for large bed heights. But for shallow beds of fine particles, the pressure drop ratio was independent of fluidization conditions such as superficial gas velocity.

Luca Mazzei and Paola Lettieri (2007) developed a mathematical equation to predict the drag force developed on spherical particles in uniformly fluidized systems. The equation is based on the empirical correlation by Richardson and Zaki. In order to validate the equation its predictions were compared with experimental data obtained of spherical particles based on the work by Happel and Epstein (1954), Richardson and Zaki (1954), Rumpf and Gupte (1971) and Wilhelm and Kwauk (1948). A good agreement was found between theoretical results and experimental values.

Investigations on the variation of bed pressure drop with superficial velocity were conducted by Botterill et al. (1982), Mathur et al. (1986), Nakamura et al. (1985) and Saxena et al. (1990). The influence of different parameters such as type of distributor, bed geometry, particle size and size distribution, bed temperature and bed pressure was studied. They reported that the bed pressure drop is constant at a gas velocity greater than the minimum fluidizing velocity.

Aspect ratio (R) is defined as the ratio of bed height, 'H' to the diameter of the bed 'D'. The bed height of a fluidizing bed is usually taken at the minimum fluidizing velocity. If the aspect ratio is more than unity, the bed is usually considered to be a tall or deep bed. A shallow bed is the one that has aspect ratio equal to or less than unity. D Sathiyamoorthy et al. (2003) conducted experiments using two types of multi-orifice distributors and three bed materials having their particle size in the range 70-160 μm in an air fluidized bed. Based upon the experimental investigations they concluded that aspect ratio has a significant effect on the quality of fluidization. There is a critical value of aspect ratio where the quality of fluidization is maximum. They further observed that this critical aspect ratio is influenced by operating velocity and the type of the distributor.

Experimental study on the bed hydrodynamics of gas solid turbulent fluidized beds was conducted by N. Ellias et al. (2004). The flow dynamics were investigated by means of pressure transducers, optical probes and capacitance probes. The superficial gas velocity-Uc, at which the turbulent flow regime occurs, was found to depend on the aspect ratio. According to Bouratoua et al. (1993), the lower pressure drop after minimum fluidizing velocity is due to the existence of an additional force along the walls, contributing to support the fluidized solids. This force can be due to the wall friction exerted on particles flowing down along the walls which compensates the upflow of particles in the wake of bubbles. Upadhyay et al. (1981) noticed that there is 15-20% reduction in bed pressure drop than the static bed pressure. They found that this difference is due to partially fluidized bed and pointed out that 15-20% of the bed was not fluidized. According to Sutherland (1964) in a conventional fluidized bed, a rise in pressure drop with increasing gas flow over the fluidizing region is an indication of slugging, whereas a decrease in the pressure drop leads to channeling.

D.G.Kroger et al. (1979) developed mathematical equation for predicting the pressure drop across the bed in rotating fluidized beds. They confirmed the validity of the analytic models by conducting experiments over a wide range of operating conditions using different bed materials. According to them

$$\Delta Pb = \{150(1-\acute{\varepsilon})^2 \mu_f m_{fh} \ln(ro/ri)\}/\{2\pi\rho_f \acute{\varepsilon}^3 dp^2\} + \{1.75\,(1-\acute{\varepsilon}) m^2_{fh}(1/ri - 1/ro)\}/\{4\pi^2 \rho_f \acute{\varepsilon}^3\, dp\} \quad (2.8)$$

Due to the rotation of the bed a vortex formation occurred in the bed and ri & ro were the inner radius and outer radius of the bed respectively at the same vertical distance. They observed that fluidization commenced at the lower edge of the bed where ri has its smallest value and centrifugal forces are less than that at any other point. It was also noticed that due to the decrease in ri, for an increase in bed mass, fluidization occurred earlier in deep rotating bed than in a shallow rotating bed.

Gelperin et al. (1960) conducted investigations in a centrifugal fluidized bed and proposed the following relation to determine the maximum pressure drop across the bed which usually occurs at minimum fluidizing velocity.

$$\Delta Pbmax = W_0 \acute{\omega} o^2 / 2\pi L \tag{2.9}$$

Teruo Takahashi et al. (1983) conducted experiments in a bed rotating horizontally with speeds ranging from 400 to 800 r p m. Glass beads, magnesia, clinker, silica sand and polyethylene powder were used as the bed materials. They observed that the experimental values of maximum bed pressure drop almost agree with those of calculated values.

B. Sreenivasan et al. (2000) developed the following mathematical model to predict the pressure drop across the bed in swirling fluidized bed.

$$\Delta Pb = kMb/am + \psi \acute{\omega}^2 \tag{2.10}$$

They confirmed the model prediction by conducting pressure drop studies in a swirling fluidized bed using two different sizes of spherical particles.

Experiments were conducted by J.M.Valverde et al. (2006) to determine the effect of bed inclination on fluidization and sedimentation behaviour. The experiment was performed using 4.42 cm diameter polycarbonated cylinder. A controlled flow of nitrogen was supplied from below. It was found that the inclination promotes fluidization heterogeneity. It was also observed that bubbling occurs at superficial velocity lower than that in the case of vertical bed. Y.K. Mohanty et al. (2007) studied the effect of co-axial rod, disk and blade promoters on bed fluctuation and expansion in a gas-solid fluidized bed with varying distributor open areas. Experimentation was conducted with four different bed materials viz. dolomite, sand, refractory brick and coal. It was found that mass velocity has a larger effect on fluctuation ratio than static bed height, particle density and size. It was also observed that the fluctuation ratio increases with increasing static bed height upto about twice minimum fluidizing velocity and then reduces at somewhat higher velocities.

2.5 MINIMUM FLUIDIZING VELOCITY

Minimum fluidizing velocity (U_{mf}) can be determined from the graph of bed pressure drop versus superficial velocity. When the gas is made to flow through the bed of particles, during the initial stages the bed pressure drop increases linearly with increase in superficial velocity and reaches a maximum. This occurs in the packed bed region. At this point the bed is said to be incipiently fluidized. Any further increase in superficial velocity does not make any change in bed pressure drop. This occurs in the fluidized bed region. Then the intersection of the sloping fixed bed and the horizontal fluidized bed give the minimum fluidization velocity. Estimation of minimum fluidization velocity is a necessary step in the design and operation of fluidized beds.

Kawabata et al. (1981) conducted experimental investigations to determine the minimum fluidizing velocity in conventional fluidized beds. They observed that their results were in good agreement with the predicted values obtained from empirical relations formulated by Wen and Yu (1966) for pressures upto 0.8 MPa at ambient temperature. Masaaki Nakamura et al. (1985) conducted experimental investigations to study the variation of minimum fluidizing velocity with temperature and pressure in conventional bed. The experiments were conducted with uniformly sized glass beads having diameters between 0.2 mm and 4 mm in beds ranging in diameter from 30 to 50 mm. Nitrogen was used as the fluidizing gas and minimum fluidizing velocity measurements were made at temperatures upto 800^0 K and pressures upto 5 MPa. The experimental results revealed that the minimum fluidization velocity decreased gradually with pressure in the low pressure region but decreased more rapidly in the high pressure region. It was also observed that the minimum fluidization velocity decreased steadily with increasing temperature for 0.19 mm particles. But in the case of particles with 2 mm diameter the minimum fluidizing velocity was found to be independent of temperature change. When particles much larger than 2 mm diameter were used it was observed that the minimum fluidizing velocity increased with temperature.

J S M Botteril et al. (1982) conducted experimental investigations in hot fluidized bed contained within a 188 mm diameter stainless steel cylinder. Operating temperature varied from 250^0C to 700^0C. Sand, ash and alumina falling within the categories of Geldart's Groups B and D were used as the bed materials. The minimum fluidizing velocity was found to decrease with increase in the operating temperature for group B materials. But in the case of Group D materials, the minimum fluidizing velocity increased with increase in operating temperature. A transition between behavior characteristics of Group B and D type materials was observed around a Reynolds number of 12.5 and Archimedes number of 26000.

Wen and Yu (1966) correlated the following expression for predicting the minimum fluidizing velocity in conventional fluidized beds.

$$dp\, U_{mf}\, \rho_g / \mu = \{(33.7)^2 + 0.0408\, dp^3\, \rho_g (\rho_s - \rho_g) g / \mu^2\}^{0.5} - 33.7 \qquad (2.11)$$

J. Shu et al. (2000) conducted hydrodynamic studies in a toroidal fluidized bed reactor. The reactor was provided with a distributor consisting of blades inclined at an angle θ of 25^0 to 30^0 to the horizontal held in an annular ring at the reactor bottom. Experiments were conducted using both fine alumina having a mean diameter of 0.4 mm and coarse alumina having a mean diameter of 3 mm. It was observed that regarding the transition from fixed bed to minimum fluidization, there was no significant difference between a toroidal fluidized bed and a conventional fluidized bed. They further observed that the minimum fluidizing velocity in a tore bed can be calculated as

$$\text{Umf, Torebed} = \text{Umf}/\text{Sin}\ \theta \qquad (2.12)$$

Binod Srenivasan et al. (2000) conducted hydrodynamic studies in swirling fluidized bed. Experiments were conducted with spherical P V C particles of sizes 3.5 mm and 2.5 mm. It was observed that the correlation for minimum fluidizing velocity recommended by Chitester et al. for coarse particles in a conventional bed given below compare well with the minimum bubbling

$$\text{Rep.mf} = [28.7^2 + 0.0494\ \text{Ar}]^{0.5} - 28.7 \qquad (2.13)$$

velocity values in swirling fluidized bed. R Moreno et al. (2002) conducted experimental study on saw dust particle drying in vibro - fluidized bed. It was noted that the minimum fluidization velocity was reduced upto three times compared to those in a non vibrated bed. M. H.Shi et al. (2000) conducted an experimental study on the heat and mass transfer characteristics of wet materials in a drying process in a centrifugal fluidized bed dryer. The rotating speed ranged from 300 to 500 rpm. Wet sand, glass beads and sliced food products were used as the testing materials. It was observed that minimum fluidizing can be achieved at any gas flow rate by changing the rotating speed of the bed. It was also observed that by using a strong centrifugal field much greater than gravity, the bed was able to withstand a large gas flow rate without the formation of large bubbles.

Sobrin C et al. (2007) performed experiments to study the influence of rotational speed of the distributor plate on the hydrodynamic behaviour of the fluidized bed. A perforated plate with 2 mm diameter holes was adopted as the rotating distributor. The distributor plate was rotated at different speeds upto 100 rpm. It was observed that the minimum fluidizing velocity decreased with increase in rotational speed.

2.6 CENTRIFUGAL FLUIDIZED BED

A centrifugal fluidized bed is a relatively new concept that is introduced for special purposes like coal combustion, zinc roasting etc. where large amount of aeration is required. A centrifugal fluidized bed is a cylindrical bucket rotating about its axis of symmetry. The particles are fluidized by introducing aeration in the inward direction of the radius. The body force in a centrifugal bed is determined by the rotation speed and bucket radius. By using a strong centrifugal field considerably greater than gravity, the particle bed is able to withstand a large amount of aeration without considerable formation of large bubbles. Thus the gas - solid contact at a high aeration rate is improved.

In conventional fluidized bed, introduction of large amount of aeration leads to the formation large bubbles or slugging and particle elutriation. In such cases, gas solid contact becomes rather poor. Hence in processes where high superficial velocity is required, conventional fluidized bed is insufficient and led to the concept of centrifugal fluidized bed.

Farkas et al. (1969) proposed that the pressure drop across a centrifugal fluidized bed could be predicted by the overall balance of the centrifugal force and the drag force of the fluid. Brown et al. (1972) and Hanni et al. (1970) suggested a modified design for centrifugal fluidized bed to reduce the pressure drop by using a cross flow configuration. Various investigations were conducted by different scientists on the variation of pressure drop with superficial velocity in centrifugal fluidized beds. The investigations done by Howard et al. (1977) revealed that the pressure drop increased linearly with superficial velocity in the initial stages and then it declined to a plateau as the superficial velocity increased. But according to the investigations of Fan et al. (1985) the pressure drop across the bed increases linearly with superficial velocity, reaches a maximum and then decreases. The point of maximum pressure drop corresponds to minimum fluidizing velocity.

Different models to predict the incipient fluidization condition of centrifugal fluidized beds have been proposed. Levy et al. (1978) and Kroger et al. (1979) proposed an approach based on the local force balance at the distributor. But this theory is applicable only to shallow beds.

Ye - Mon Chen (1987) conducted investigations to explore the fluidizing phenomenon in a centrifugal fluidized bed. It was predicted that the centrifugal bed is fluidized layer by layer from the inner surface to the outwards in a range of aeration rates. The span of this range increases with the depth of the particle bed. Teruo Takahashi et al. (1984) conducted experiments in a centrifugal fluidized bed which was rotated at speeds ranging from 400 to 800 rpm. Experiments were conducted using glass beads, magnesia clinker, and silica sand and polyethylene powder. It was observed that the pressure drop in the centrifugal fluidized bed attains a maximum value at a certain fluidizing velocity viz. minimum fluidizing velocity. A further increase in gas velocity beyond minimum fluidization velocity resulted in a slight decrease in pressure drop.

Experimental Investigations were conducted by W.Y Wong et al. (1999) to optimize the fluidization performance of the rotating fluidized bed. A small scale vertical axis rotating fluidized bed was designed and fabricated with an internal diameter of 200 mm and 50 mm height. The operating parameters viz. particle size, rotation speed and bed height were optimized. A study on the flow field revealed the existence of two regimes ie. the free vortex outer region and the forced vortex flow near the axis. This leads to the generation of turbulence, which promotes mixing between particles and fluid.

M.H.Shi et al. (2000) conducted experiments in a cylindrical basket provided with 3 mm diameter holes on the side surface. The basket was 200 mm in diameter and 80 mm in width. A variable speed motor was used to rotate this basket by means of a shaft connected to the other end wall of the basket. Air was blown from a blower. Wet sand, glass beads and sliced food products were used as the bed materials. Experiments were conducted at different rotating speeds. Observations were made on the variation of bed pressure drop with superficial velocity. With sand it was found that the pressure drop increased linearly with superficial velocity in the initial stages. After reaching the critical point the pressure drop will be almost a constant. But in the case of glass and sliced food it was observed that after reaching the critical point the pressure drop decreased with superficial velocity.

2.7 SWIRLING FLUIDIZING BED

The swirling fluidized bed is similar to a Tore bed reactor. It consists of an annular bed and the gas is injected through the distributor blades in an inclined manner. This results in a swirling motion of solid particles in a confined circular path. When a jet of gas enters the bed at angle θ to the horizontal, the vertical component of the velocity, v sin θ causes fluidization and the tangential component v cos θ in turn causes swirl motion of the bed material. If the bed is sufficiently deep the horizontal momentum of the jet deteriorates with the penetration of the jet into the bed and finally it ceases at a certain height above the distributor. But in a shallow bed the velocity of the jet leaving the bed will still have two components. In this case the bed will be swirling as a single mass. There are a number of commercial applications based upon swirling fluidized bed. But a very few literature is available on this subject.

Ouyang and Levenspiel (1986) designed a spiral distributor to achieve swirl motion. This distributor was made of overlapping vanes, shaped as sectors of a circle provided with a gap between the vanes. The gap between the adjacent vanes was maximum at the outer periphery and zero at the centre of the distributor. The vanes were arranged in a manner so as to provide tangential exit of air from the gap of the vanes. It was observed that the inclined jet emerging from the opening impart a swirling motion in a shallow bed. But in a deep bed the swirling motion is restricted to the lower portion of the bed and above that bubbling occurs. They also made a comparative study on the characteristics of this distributor, such as pressure drop, quality of fluidization and heat transfer coefficient with that of a sintered plate distributor. It was observed that for low density solids, the sintered plate gives better fluidization at low superficial velocity. On the other hand at high superficial velocities the performance is better in a spiral distributor. But for high density solids better fluidization can be achieved with spiral distributor at all gas velocities. It was also noticed that the pressure drop across the spiral distributor was lower than for the sintered plate distributor.

Binod and Raghavan (2000) conducted experiments in a swirling bed having vanes inclined at an angle of 12^0 to the horizontal. The gap between the blades varied in proportion to the radius creating a trapezoidal opening for air flow. A hollow metal cone was located centrally at the base of the bed. Experiments were performed with spherical P V C particles having sizes 3.5 mm and 2.5 mm. It was observed that the pressure drop in the swirling regime is not constant but increased with gas flow rate. Unlike in a conventional bed four regimes of operation could be observed in a swirling fluidized bed viz. bubbling, wave motion with dume formation, two - layer fluidization and stable swirling. It was also observed that as the angle of injection 'θ' increased the pressure drop in the fluidized regime decreased.

G Vikram et al. (2003) conducted hydrodynamic studies in a swirling fluidized bed. It was observed that the velocity of the swirling particles increased with radial distance. It was also observed that there was a sharp decay of both gas and particle velocity with height. Investigations were conducted by Paulose and Narayanan Nampoothiri (2004) to compare the performance of different types of distributors in swirling fluidized bed. Experiments were performed using three types of distributors viz. single- row vane type distributor, three- row vane type distributor and inclined hole type distributor. It was found that the percentage useful area was highest for the inclined hole type distributor. But the distributor pressure drop was lowest for the single-row vane type distributor. Paulose and Narayanan Nampoothiri (2005) conducted further hydrodynamic studies in swirling fluidized bed. Single-row vane type distributors having vane angles 150 and 200 were adopted. Experimentation was conducted using coffee beans (Geldart D-type) having an average diameter of 8.82 mm and relative density of 0.75. It was found that the minimum fluidizing velocity is constant irrespective of the vane angle. The experimental studies further revealed that the distributor pressure drop decreases with increase in vane angle.

T. Madhiyanon et al. (2007) developed a new cold model combustor named cyclonic fluidized bed combustor which has the advantage features of both swirl flow and fluidized bed flow. Primary air or vortexing air was injected into the top of the combustor while secondary air or fluidized air was projected upward through a perforated air distributor. They noticed the formation of vortex rings in the bed which promoted fluidization by enhancing the upward axial flow velocity without increasing the fluidizing air velocity. Parametric analysis of an analytical model of a swirling fluidized bed was conducted by M. Kamil et al. (2007). The influence of blade angle, gas density and particle density on the hydrodynamic characteristics of swirling fluidized bed was investigated. The results of the study revealed that the particle density and gas density are the most influential parameters followed by blade angle and superficial gas velocity.

The review of the literature presents an overall picture of the different types of distributors developed for different processes. An optimum design of the distributor is the one which ensures uniform distribution of air combined with rather low distributor pressure drop. The pressure drop ratio is often considered as an important criterion in the design of a distributor. This ratio is dependent upon the dimensions of the distributor. For stable fluidization in deep fluidized bed, the minimum value of pressure drop ratio ranges from 0.02 to 0.05. But Agarwal et al. recommended a minimum distributor pressure drop of 350 mm of water for shallow conventional beds.

In the case of conventional fluidized bed, almost all researchers agree that the bed pressure drop becomes constant at gas velocity greater than minimum fluidizing velocity and variation in the pressure drop after fluidization is an indication of undesirable characteristics such as slugging or channeling. But there is a difference of opinion in the case of centrifugal fluidized bed. While Howard et al. (1977) reported a constant bed pressure drop, Fan et al. (1985) reported a decrease in bed pressure drop after minimum fluidization. In the case of it was reported that there is an increase in the bed pressure drop with superficial velocity in the swirl regime.

On the basis of experimental investigations conducted on multi-orifice type distributors it was observed that the number of operating orifices is determined by the gas flow rate, bed height, type of bed material and the free area of the distributor. Fakhimi et al. (1971) proposed a mathematical model to predict the number of active orifices at any given flow rate. Whitehead (1971) conducted experimental investigations on this matter and noticed certain discrepancies. Investigations conducted with porous plate distributors revealed that the quality of fluidization is controlled by various characteristics of the bubbles produced. D. Sathyamoorthy et al. (2003) conducted experimental investigations and concluded that aspect ratio has a significant effect on the quality of fluidization.

Various investigations were conducted by different scientists to study the variation of minimum fluidization velocity with temperature. It was observed that the minimum fluidizing velocity decreased steadily with increasing temperature for 0.19 mm particles. But in the case of particles with 2 mm diameter the minimum fluidizing velocity was found to be independent of temperature. With particles much larger than 2 mm in diameter it was observed that the minimum fluidizing velocity increased with temperature.

Only a very few literature is available on swirling fluidized bed. Experiments were conducted by Ouyang and Levenspiel (1986) on swirling fluidized bed and noticed that the pressure drop decreased with increase in angle of injection. G. Vikram et al. (2003) noticed that the velocity of swirling particles increased with radial distance. By conducting experiments in swirling fluidized bed Paulose and Narayanan Nampoothiri noticed that similar to conventional fluidized bed, minimum fluidizing velocity in swirling fluidized bed is constant (2005). Analytical studies conducted by M. Kamil et al. (2007) revealed that particle density and gas density are the most influential parameters in a swirling fluidized bed. Till date no experimental investigations has been conducted on the effect of angle of air injection and particles on the hydrodynamic characteristics of swirling fluidized bed.

2.8 OBJECTIVES OF THE PRESENT STUDY

Based upon the brief review of the available literature, the following objectives have been proposed for the present study.

To study the basic the basic hydrodynamic characteristics of swirling fluidized bed using large size particles and by varying the following major variables

1. Density of the particles
2. Size of the particles
3. Angle of air injection

CHAPTER 3

DESIGN AND FABRICATION OF DISTRIBUTOR

3.1 INTRODUCTION

Distributor is the vital part in any fluidized bed. As the name indicates it distributes the air uniformly to the bed. An effective distributor design is one in which distributes the gas uniformly without excessive resistance to the gas flow. In a conventional fluidized bed air is admitted vertically upwards to the bed, whereas in a swirling fluidized bed, air enters the bed at an angle. This is achieved by providing slits inclined at an angle with the radius of the distributor. In conventional fluidized beds a distributor pressure drop as high as 350 mm of water is required for good fluidization. However in a swirling fluidized bed, effective fluidization can be achieved with a comparatively lower distributor pressure drop. This chapter deals with the design and fabrication of distributors.

3.2 DESIGN

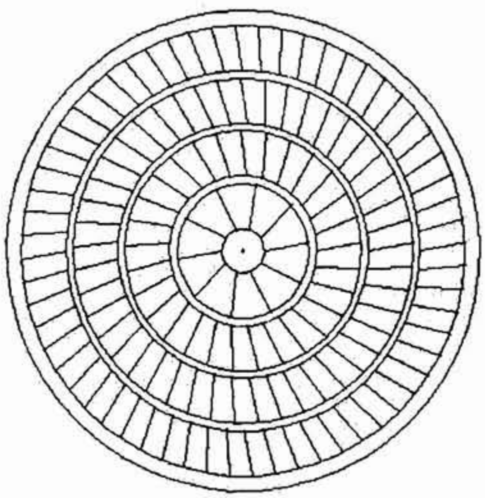


Figure 3.1 Details of a distributor

The distributor has an outer diameter of 300 mm. A 40 mm diameter wooden cone was provided at the centre of the distributor. The slits were arranged in four rows along four concentric circles. A ridge having a width of 11 mm was provided in between each row, which facilitated the pressure tappings to be conveniently located there. The length of the slit can be determined from the following relation

$$\text{length of the slit 'l'} = [rd - rc - (3 x br)]/ nr \tag{3.1}$$

where rd – radius of the distributor = 150 mm

rc - radius of the cone at the centre = 20 mm

br – width of the ridge = 11 mm

nr - number of rows = 4

Thus length of the slit 'l' = [150 – 20 – (3x11)]/ 4 = 24 mm

The inner radius of each row of slits can be determined from the relation

$$\text{The inner radius of the row 'ri'} = rc + (ns - 1) x [l + br] \tag{3.2}$$

$$= 20 + (ns - 1) x [24 + 11]$$

where ns – position of the row starting from the centre

The inner radius of each row is given in table 3.2.1

Table 3.2.1 Inner radius of each row of slits

Sl No.	Position of the row 'ns'	Inner radius of the row 'ri' = 20+(ns– 1)x [24 + 11] (mm)
1	1	20
2	2	55
3	3	90
4	4	125

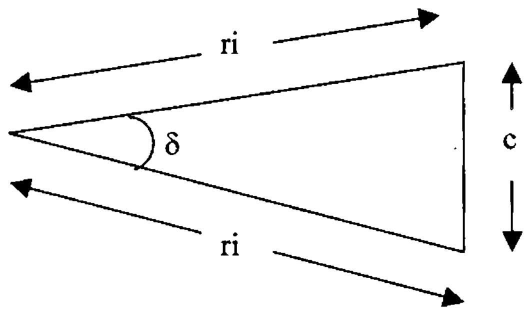


Figure 3.2 Geometry of slits in distributor

The number of slits in each row can be found out from the following relation

$$ri^2 + ri^2 - 2ri^2 \cos \delta = c^2$$

$$2\, ri^2 (1 - \cos \delta) = c^2$$

$$\cos \delta = 1 - (c^2/2\, ri^2) = (2\, ri^2 - c^2)/ 2\, ri^2$$

$$\delta = \cos^{-1} [(2\, ri^2 - c^2)/ 2\, ri^2]$$

where chord length 'c' = 12 mm

$$\text{Number of slits in each row} = 360/ \delta \qquad (3.3)$$

The number of slits in each row is given in table 3.2.2

Table 3.2.2 Number of slits in each row

Sl. No.	Position of the row	Inner radius 'ri' (mm)	Angle 'δ'= $\cos^{-1} [(2\, ri^2 - c^2)/ 2\, ri^2]$	Number of slits = 360^0 / (δ)
1	1	20	36	10
2	2	55	12.5	29
3	3	90	7.7	46
4	4	125	5.8	62

Table 3.2.3 Design details of the distributor

Sl. No.	Particulars	
1	Minimum gap between slits (mm)	12
2	Length of the slit (mm)	24
3	Width of the slit (mm)	1.8
4	Area of opening of each slit	43.2
5	Total number of slits	147
6	Total area of opening (mm^2)	6350.4
7	Percentage area of opening	9.15
8	Percentage useful area of distributor	99.5

Percentage area of opening $= (6350.4 \times 100)/\pi(rd^2 - rc^2)$

$= (6350.4 \times 100)/\pi \times 88400 \quad = 9.15$

Percentage useful area of distributor $= [(rd^2 - rc^2) \times 100]/ rd^2$

$= (150^2 - 20^2) \times 100/150 \quad = 99.5$

3.3 FABRICATION

The slits of the distributor were sheared by pressworking. A 6 mm hole was drilled at the centre of each distributor through which a bolt of the same size can be inserted. The mating nut was suitably located inside the centre of the wooden cone. Thus the cone could be conveniently attached to any distributor. Suitable holes were also drilled outside the periphery of outermost row of slits for securing the distributor to the plenum

Figure 3.3 Photograph of a typical distributor

chamber by means of nuts and bolts. The die used for the fabrication of the distributor is shown in figure 3.4

The bed materials were fabricated by injection moulding. The set of dies employed for the production of different size bed materials is given in figure 3.5.

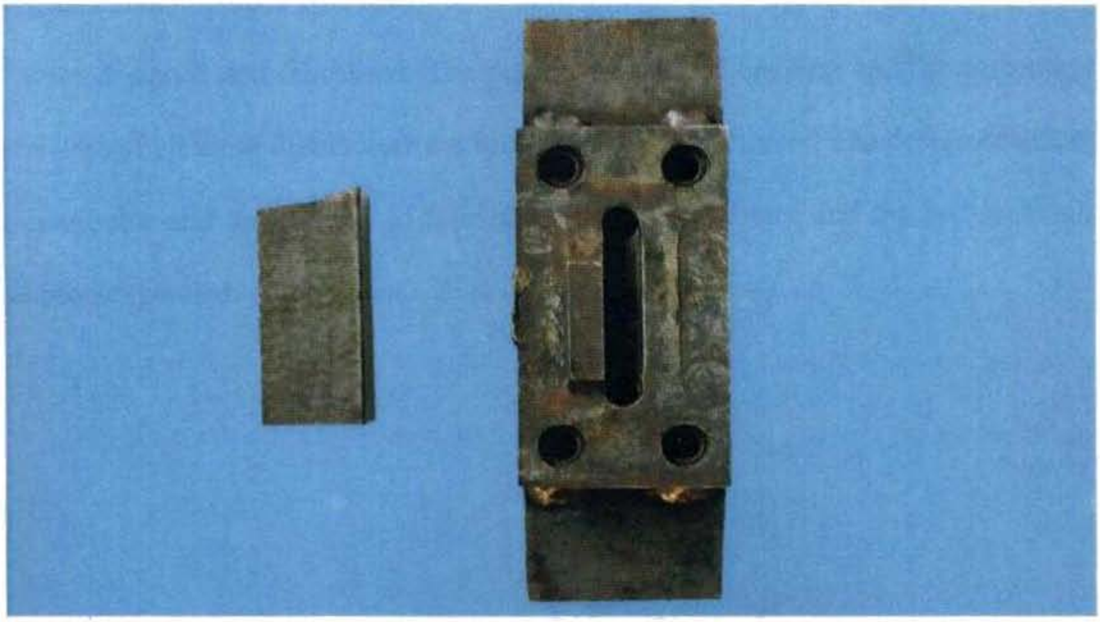


Figure 3.4 Photograph of the die used for the fabrication of the distributor

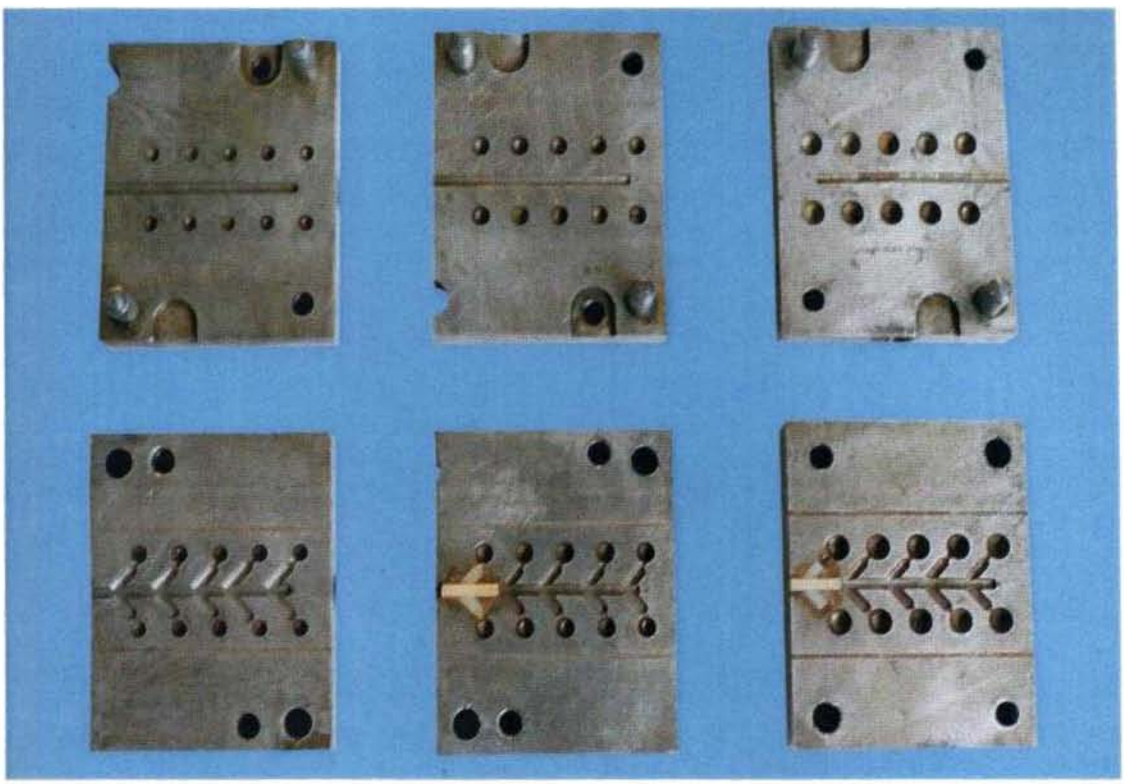


Figure 3.5 Photograph the set of dies employed for the production of different bed materials

3.4 CONCLUSION

Altogether four conidour type distributors having angle of air injection (Φ) 0^0, 5^0, 10^0 & 15^0 were designed and fabricated. The percentage area of opening and the percentage useful area of all these distributors are 9.15 and 99.5 respectively. The design details of the distributor and the method of fabrication of the distributor and the bed materials have been explained.

CHAPTER 4

EXPERIMENTATION

4.1 INTRODUCTION

The details of the experimental set up and procedure to determine the various parameters is presented in this chapter. The procedure adopted for determination of various physical properties of bed particles such as bed density, particle density and voidage is described here. A detailed description of the procedure adopted to determine the various characteristics viz. distributor pressure drop, bed pressure drop, bed height and minimum fluidizing velocity is also presented in this chapter.

4.2 EXPERIMENTAL SET UP

A schematic diagram describing the main parts of the experimental set up is given in figure 4.2.1 and the photograph of the complete view of the set up is presented in figure 4.2.2

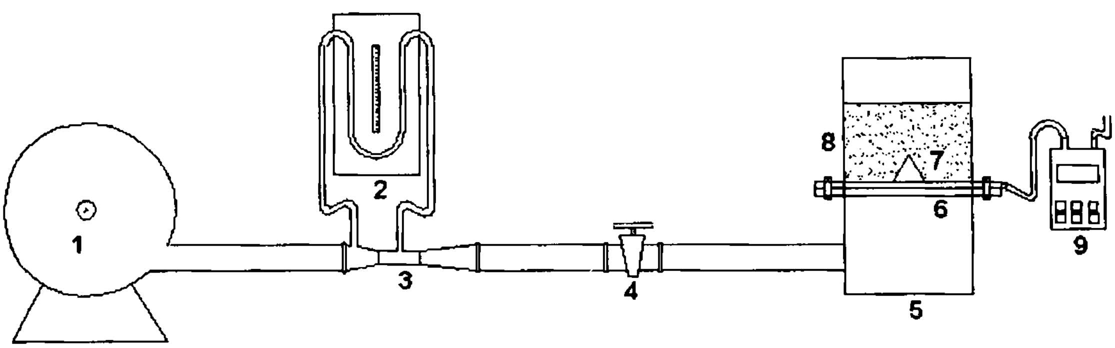
The air required for fluidization was supplied by a single stage centrifugal type 7.5 H P blower. The slightly compressed air was allowed to pass through a spiral case, before it came out by the outlet. The air was supplied from the blower to the plenum chamber through a 125 mm diameter G.I pipe. The rate of flow of air was controlled by means of the gate valve. The flow rate was measured with the help of a calibrated venturimeter whose calibration equation is given by

$$Va= 0.021\sqrt{Ha} \tag{4.1}$$

A differential U tube water column manometer connected to the venturimeter enabled to measure the head causing flow in metres of water (Hw) from which head of air was calculated from the equation.

$$Ha = Hw^*\rho w/\rho a \tag{4.2}$$

The venturimeter was fixed at a distance of 1000 mm away from the outlet of the blower and the gate valve was located at 1000 mm downstream with respect to the venturimeter. This arrangement ensured uniform flow in the venturimeter. In order to avoid parallax error the readings on the manometer were measured by a pointer attached to a rack and pinion arrangement. The air was made to enter the plenum chamber in a tangential direction so as to have a clockwise air circulation within this chamber. This was to reduce the pressure loss at entry to the distributor since the inclined vane type distributors were also designed to have a clockwise entry of air into the bed.



1. Blower 2. Differential Manometer 3. Venturimeter 4. Valve
5. Plenum Chamber 6. Distributer 7. Solid Cone 8. Bed Column
9. Micromanometer

Figure 4.2.1 Schematic diagram of the experimental set- up

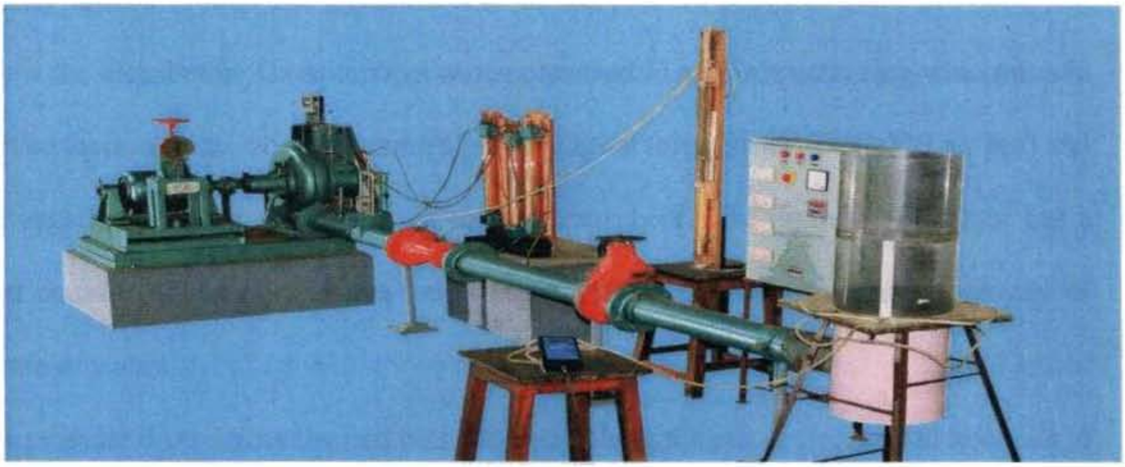


Figure 4.2.2 Photograph of the experimental set- up

The plenum chamber was fabricated with mild steel and it has an internal diameter of 300 mm and height of 600 mm. It was provided with a flange on the top for attaching the bed column. The bed column was also cylindrical and has an internal diameter of 300 mm and a height of 600 mm. This was made from perspex to facilitate visual observations while conducting experiments. It was also provided with a 10 mm thick flange that enabled it to be joined to the plenum chamber by means of nuts and bolts. The distributor was held in position by means of a distributor holder. This was made from 12mm thick perspex square plate of 450mm.The same was provided with a 300mm diameter central opening and inner annular groove of 10mm depth and 15mm width. The assembly was made airtight by providing rubber packing at top and bottom. The temperature of air flowing through the pipe was measured by calibrated temperature sensor of type K thermocouple (Alumel-Chromel) provided inside the pipe just before the venturimeter. This temperature was measured to an accuracy of 0.1^0 with a digital temperature indicator.

For measuring the distributor pressure drop three 0.8mm diameter pressure tappings, equally spaced around the circumference of the plenum chamber were provided just below the distributor. These tapings were connected to a piezometric ring which was in turn connected to the positive terminal of the digital micrometer (FCO 520 air pro) and the negative terminal was left open to the atmosphere. The digital micrometer has a least count of 0.01mm of water and is capable of measuring a maximum pressure of 60mm of water.

The pressure drop across the bed was measured along the radial direction at intervals of 30 mm. These measurements were made at distances of 60 mm, 90 mm, 120 mm and 150 mm from the centre of the distributor. Similar to distributor pressure drop measurement, at each of these radial distance three points equally spaced around the circumference of the distributor were located. These locations were then provided with 0.8mm diameter pressure tappings flush with the top surface of the distributor. The pressure tappings at the same radial distance were connected through a piezometric ring to the positive terminal of the digital micromanometer and the negative terminal was left open to the atmosphere.

4.3 EXPERIMENTAL PROCEDURE

4.3.1 Determination of physical properties of bed particles.

The bed material is specified by determining the following physical properties of the bed material.

-mean particle size

-particle density

-bed density

-bed voidage

4.3.1.1 Mean particle size

Since the bed materials were made by injection moulding with the help of suitable dies having required openings their size need not be experimentally determined.

4.3.1.2 Particle density

The particle densities of the materials were determined with the help of a standard pycnometer. Distilled water was employed as the liquid in the pycnometer for determining the densities of acetal and nylon, whereas the density of HDPE was determined using turpentine. The weight was measured using a digital weghing machine having a least count of 0.1 gm. The details of the observations made for determination of particle density in the case of acetal & nylon is presented in tables 4.3.1, while that of H D P E is presented in table 4.3.2. The relative density of HDPE with respect to water was then obtained by multiplying the relative density of HDPE with respect to the turpentine by the relative density of turpentine which is 0.84.

Table 4.3.1 Calculation of particle density of acetal & nylon

Sl. No.	Description	acetal Particle size (mm)			nylon Particle size (mm)		
		3.2	**5.5**	**7.4**	**3.2**	**5.5**	**7.4**
1	Weight of pycnometer, p (gm)	464.5	464.8	464.1	464.3	464.7	464.1
2	Weight of pycnometer + particle, q (gm)	790.5	860.1	841.0	708.7	729.8	705.0
3	Weight of pycnometer + particle + water r (gm)	1374.7	1393.0	1389.6	1280.6	1285.3	1278.0
4	Weight of pycnometer + water, s (gm)	1268.1	1268.5	1269.0	1268.3	1269.3	1268.7
5	Relative density of particle $(q-p)/[(s-p)-(r-q)]$	1.48	1.46	1.47	1.05	1.06	1.04
6	Average particle density	1.47			1.05		

Table 4.3.2 Calculation of particle density of H D P E

Sl. No.	Description	Particle size (mm)		
		3.2	5.5	7.4
1	Weight of pycnometer, p (gm)	464.9	464.2	464.5
2	Weight of pycnometer + particle, q (gm)	640.3	640.9	640.6
3	Weight of pycnometer + particle + turpentine, r (gm)	1171.2	1171.7	1169.0
4	Weight of pycnometer + turpentine, s (gm)	1154.1	1153.0	1153.5
5	Particle density relative to turpentine (r.d = 0.84), t (q- p)/[(s – p) – (r – q)]	1.107	1.119	1.095
6	Relative density of particle, t*0.84	0.93	0.94	0.92
7	Average particle density	0.93		

4.3.1.3 Bed density

The bed densities of the materials were determined using a calibrated container having a capacity of three litres. The materials filling the container was weighed to an accuracy of 0.1 gm.The details of the observations made for calibration of bed density is given in table 4.3.4

4.3.1.4 Bed voidage

After determining the particle density and bed density of the materials, the bed voidage(Є) can be calculated from the relation.

$$\epsilon = 1- \rho b/\rho_p \qquad (4.3)$$

Table 4.3.3 Calculation of bed density and voidage

Particle	Sl. No	Bed density (gm/litre) For particle size (mm)		
		3.2	5.5	7.4
Acetal	1	890	875	872
	2	885	880	877
	3	883	879	873
	Mean	886	878	874
	Bed density (ρb)	0.886	0.878	0.874
	Particle density (ρp)	1.47	1.47	1.47
	Voidage (Є)	0.397	0.403	0.410
Nylon	1	636	631	619
	2	632	627	624
	3	637	629	620
	Mean	635	629	621
	Bed density (ρb)	0.635	0.629	0.621
	Particle density (ρp)	1.05	1.05	1.05
	Voidage (Є)	0.395	0.401	0.409
H D P E	1	561	558	550
	2	568	561	547
	3	566	558	556
	Mean	565	559	551
	Bed density (ρb)	0.565	0.559	0.551
	Particle density (ρp)	0.93	0.93	0.93
	Voidage (Є)	0.393	0.399	0.408
Mean voidage (Є)		**0.395**	**0.401**	**0.409**

Table 4.3.4 Physical properties of particles

Sl. No.	Description	acetal Particle size (mm)			nylon Particle size(mm)			H D P E Particle size(mm)		
		3.2	5.5	7.4	3.2	5.5	7.4	3.2	5.5	7.4
1	Average particle density (ρp)	1.47	1.47	1.47	1.05	1.05	1.05	0.93	0.93	0.93
2	Bed density (ρb)	0.886	0.878	0.874	0.635	0.629	0.621	0.565	0.559	0.551
3	Mean voidage (Є)	0.395	0.401	0.409	0.395	0.401	0.409	0.395	0.401	0.409

4.3.1.5 Particle specification

The various physical properties of acetal, nylon and HDPE can be noticed from tables 4.3.4

4.3.2 Determination of distributor pressure drop

Determination of distributor pressure drop is an important factor that influences the quality of fluidization. The energy consumption in any fluidization process is very much dependent on distributor pressure drop. The distributor pressure drop varies directly with the superficial velocity [Saxena, 1979]. Distributor pressure drop can be determined by observing the pressure difference across the distributor in an empty bed. The distributor was screwed properly between the plenum chamber and bed column by bolting them together. The joint was made leak proof with the aid of suitable packing materials. Horizontality of the distributor was ensured with the help of a spirit level. Pressure tapping from the bottom of the distributor was connected through a piezometeric ring to the positive terminal of the digital micro manometer and the negative terminal was left open to the atmosphere. The air flow rate through the pipe was varied at regular intervals by regulating the gate valve. This was made possible by regular variations in the differential U-tube water manometer connected to the venturimeter provided in the pipe. For each value of the venturimeter reading (venturimeter pressure drop, ΔPv) the distributor pressure drop (Δ Pd) and temperature of air (T) were observed. The measurements were made only after the readings were stabilized.

The volume flow rate of air (Q) was calculated by the equation

$$Q = 0.021\ \Delta Pv\ X\ \rho w\ (273 + T)\ /\ \rho a\ X\ 273 \qquad (4.4)$$

The superficial velocity was then calculated from the volume flow rate by the equation

$$U = 4\ X\ Q / (D^2 - D^2{}_C) \qquad (4.5)$$

The experiment was repeated for all the distributors.

4.3.3 Determination of bed pressure drop

The quality of fluidization was also assessed by the variation of bed pressure drop with superficial velocity. The bed pressure drop was determined by observing the pressure difference across the bed. The pressure difference across the bed was determined along the radial direction at distances of 60 mm, 90 mm, 120 mm, and 150 mm from the centre of the distributor.

The required distributor for the particular experiment was fixed in position as explained in section 4.2. At each of these radial distances the three pressure tappings from the top of the distributor was connected through a piezometric ring to the positive terminal of the digital micro manometer and the negative terminal was left open to the atmosphere. For each value of superficial velocity ten micro manometer readings were taken at regular intervals of ten seconds. The average of these ten readings was considered as the bed pressure drop (ΔPb) for the present study. It was observed that beyond a weight of 2.5 kg of the bed material two layer formation occurs in the bed. Hence the experiments were conducted using 2.5 kg of bed material. The volume flow rate of air was varied by regulating the gate valve. These variations were made at regular intervals by making a difference of 2 mm between successive readings on one limb of the U-tube differential water manometer connected to the venturimeter. For each value of the venturimeter reading the following observations were made, after the readings became steady

-venturimeter pressure drop (ΔPv)

-bed pressure drop (ΔPb)

-temperature of air (T)

Volume flow rate of air and superficial velocity was calculated using equations (4.4) and (4.5). The experiment was repeated with all the distributors using different bed materials.

4.3.4 Determination of bed height

The bed height was measured by three symmetrical scales fixed on the outer periphery of the bed column. The least count of these scales was 1.0 mm and the average of these three readings on the scale, rounded to 1.0 mm accuracy was regarded as the bed height. At higher values of superficial velocity there was considerable oscillation for the top surface of the bed. Therefore a measurement of bed height at the higher values of superficial velocity was not accurate.

4.3.5 Determination of minimum fluidizing velocity

Minimum fluidizing velocity has a vital role in the design of fluidized bed. It depends on the physical properties of the particles.

The bed pressure drop was measured with a calibrated digital micro manometer. The details of the same are given in appendix VI. The positive terminal of the manometer was connected through a piezometric ring to three equidistant pressure tappings provided at the top of the distributor; at equal spacings around the circumference of the distributor. The negative terminal was left open to the atmosphere. There was considerable fluctuation in the bed pressure drop readings. For each value of the gate valve opening ten readings were taken at equal intervals of ten seconds and the average was considered as the bed pressure drop. The time interval was set correctly with the help of a buzzer. 2.5 kg of bed material was placed in the bed column. The gate valve was initially opened to such an extent that the bed fluidized vigorously and the following observations were made.

-venturimeter reading (venturimeter pressure drop ΔPv)

-bed pressure drop (micro manometer readings)

-temperature of air

These observations were made only after the readings were stabilized. The experiment was repeated for lower air flow rates by controlling the gate valve opening.

4.4 CONCLUSIONS

A description of the experimental set up and equipments is given above. The procedure adopted to determine the various physical properties of the bed particles and the various bed parameters is also described.

CHAPTER 5

RESULTS AND DISCUSSION

5.1 INTRODUCTION

This chapter deals with the results obtained and the discussions thereon based on the experiments conducted using conidour type distributors having slits opening towards the outer periphery of the distributor inclined with the radius (ϕ) as mentioned below

1) at an angle $\phi = 0^0$
2) at an angle $\phi = 5^0$
3) at an angle $\phi = 10^0$
4) at an angle $\phi = 15^0$

Particles of specifications as mentioned in section 4.3.1.5 were employed as the bed materials.

The various hydrodynamic studies conducted are listed below.

1) Variation of distributor pressure drop (ΔPd) with superficial velocity.
2) Minimum fluidizing velocity.
3) Variation of bed pressure drop (ΔPb) with superficial velocity at different radial positions of the above mentioned distributors using the various bed materials of different sizes.
4) Variation of bed pressure drop (ΔPb) with superficial velocity at different radial positions for different particle size with distributor having $\phi = 15^0$, using different bed materials.

5) Variation of bed pressure drop (ΔPb) with superficial velocity at different radial positions for different density materials with distributor having $\phi = 15^0$ using bed materials of different size.

6) Variation of bed pressure drop (ΔPb) with superficial velocity at different radial positions for different angles of air injection using acetal of different size as the bed material.

Based on the results of the experiments conducted and discussions, certain conclusions were arrived at and it is also presented in this chapter.

5.2 DISTRIBUTOR PRESSURE DROP (ΔPd)

Distributor pressure drop plays a predominant role on the energy consumption and quality of fluidization. Energy consumption increases with distributor pressure drop. But a minimum value of distributor pressure drop is required for uniform fluidization. It has been reported that [Agarwal, 1962] that this value is 350 mm of water in a shallow conventional bed. But in a swirling fluidized bed, uniform fluidization can be achieved even with a comparatively low distributor pressure drop [Binod & Raghavan, 2002]. It has been further reported that [Paulose M.M, 2006] distributor pressure drop decreases with increase in percentage area of opening of the distributor.

The pressure drop across the distributor was measured for all the four distributors using micromanometer. Figure 5.2.1 shows the variation of distributor pressure drop with superficial velocity for different angles of air injection. It is quite evident from the figure that the distributor pressure drop decreases with increase in angle of air injection (ϕ). The velocity of the jet (v) emerging from the vanes of the distributor can be resolved into tangential component v_t and radial component v_r. The radial component of

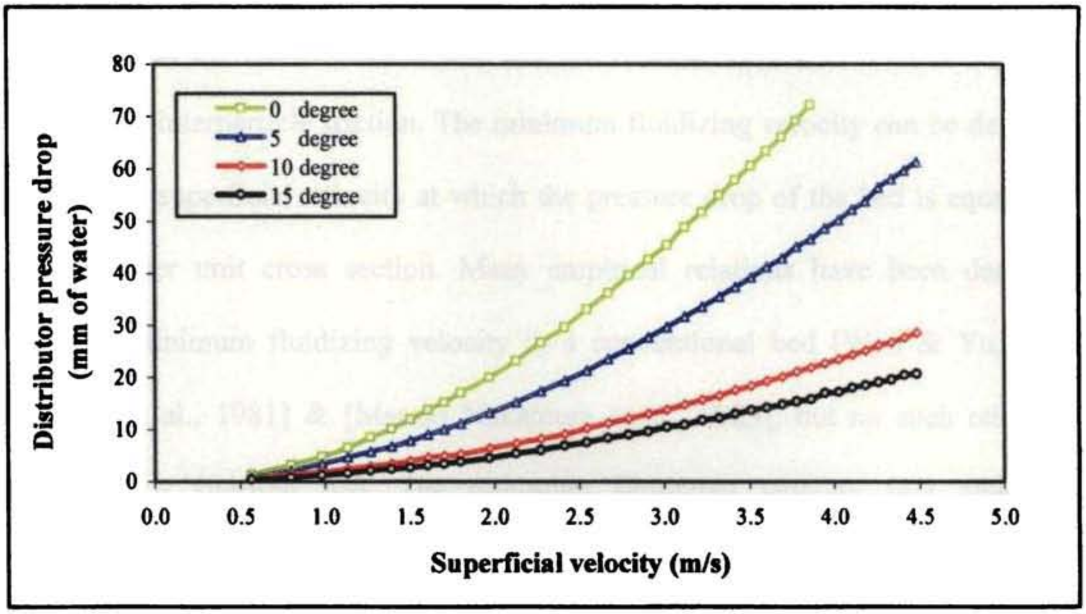


Figure 5.2.1. Variation of distributor pressure drop with superficial velocity

velocity is given by v sin ϕ. With increase in angle of air injection (ϕ) the radial component of velocity will increase. When the jet of air emerging from the vanes strikes the walls of the bed column this radial component of velocity will vanish. This in turn leads to the build up of pressure head above the distributor. This pressure head which is proportional to the radial component of velocity increases with increase in angle of air injection. Since the area of opening is the same for all these distributors the pressure head below the distributors remain the same irrespective of the angle of air injection. Hence the distributor pressure drop which is the difference between the pressures below and above the distributor decreases with increase in angle of air injection (ϕ).

5.3 MINIMUM FLUIDISING VELOCITY

The swirl motion of the particles can be achieved only by allowing the movement of the particles. So the bed has to fluidize first and the particles have to unlock themselves to overcome the interparticle friction. The minimum fluidizing velocity can be defined as the minimum superficial velocity at which the pressure drop of the bed is equal to the bed weight per unit cross section. Many empirical relations have been derived to predict the minimum fluidizing velocity in a conventional bed [Wen & Yu, 1966], [Kawabata et al., 1981] & [Masaki Nakamura et al., 1985], but no such relation is available in a swirling bed. The minimum fluidizing velocity was determined experimentally for all the four distributors. Spherical particles of acetal, nylon & HDPE each of different size all coming under "D – type" as per Geldarts classification [Howard, 1985] were adopted for this study.

The minimum fluidizing velocity can be determined from a plot showing the variation of bed pressure drop with superficial velocity and the methods of determination of these values are described in section 4.3.5. The minimum fluidizing velocity is the superficial velocity corresponding to the intersection of two straight lines drawn appropriately on the above plot. A horizontal line is drawn connecting constant bed pressure drop values after fluidization. Another straight line is drawn by connecting the last few points before fluidization (in the packed region). Since the shape of the curve in the packed region is not a straight line only the last few points are considered to draw the second line. It has been reported that for "D – type" particles, the graph in the packed region will be non – linear as the flow of the gas in the interstices is more laminar [Howard, 1985] and the same behavior was observed in the present study also.

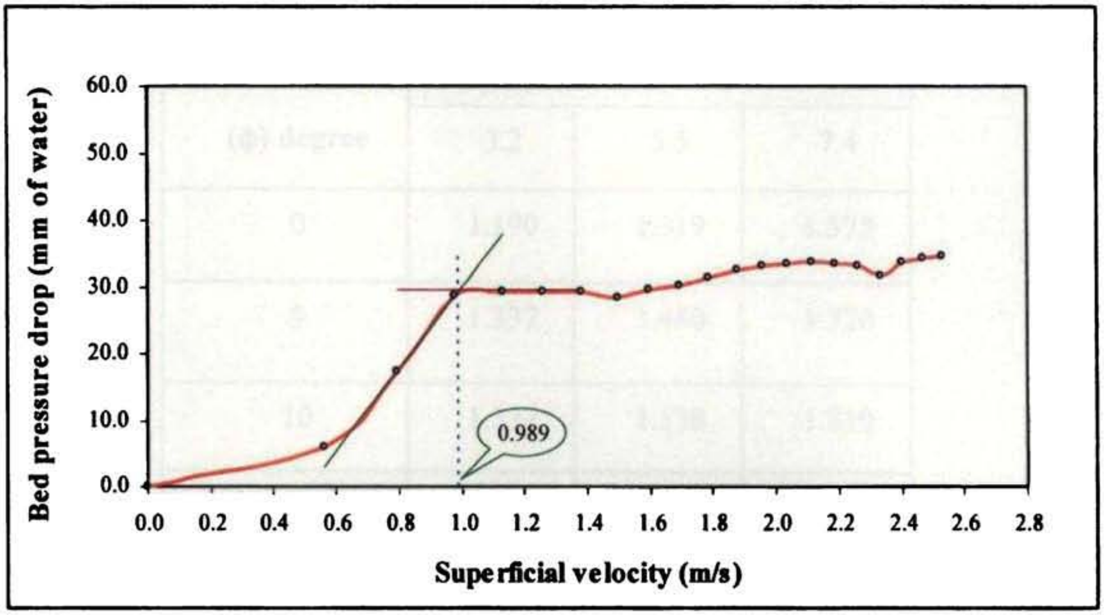


Figure 5.3.1 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 3.2 mm HDPE as bed material

A typical graph depicting minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ using 3.2 mm H D P E as bed material is presented in figure 5.3.1.The graphs showing the determination of minimum fluidizing velocity for various distributors using various bed materials of different is given figures from 5.3.2 to 5.3.24 (Appendix – I). The minimum fluidizing velocity employing four distributors using acetal of different sizes is compared in table 5.3.1, while table 5.3.2 gives a comparison of the minimum fluidizing velocity for various bed materials of different sizes employing distributors having angle $\phi = 0^0$ and angle $\phi = 15^0$.

Table5.3.1 Minimum fluidizing velocity (m/s) of acetal particles of varying particle size using distributors of different angles of air injection (ϕ)

Vane Angle (ϕ) degree	Particle size (mm)		
	3.2	5.5	7.4
0	1.190	1.319	1.575
5	1.337	1.480	1.720
10	1.337	1.538	1.810
15	1.392	1.546	1.846

It is evident from table 5.3.1 that the minimum fluidizing velocity increases with angle of air injection. In conventional bed the jet of air emerging from the distributor is in the vertical direction and strikes against the particles. On striking against the particles the velocity of the jet is converted into pressure force acting vertically upwards. The remaining part of the velocity flowing upwards through the interstices of the particles exerts a drag force in the upward direction. These two forces together causes fluidization. When the jet of air enters the bed at an angle, only its vertical component contributes to lifting of particles or fluidization of particles. With increase in angle of air injection (ϕ) this component decreases. Hence the minimum fluidizing velocity increases with increase in angle of air injection (ϕ). From table 5.3.2 it is clear that the minimum fluidizing velocity increases with increase in particle density as well as particle size.

Table 5.3.2 Minimum fluidizing velocity (m/s) for different particles of varying size using distributors having angles of air injection (ϕ), 0^0 & 15^0

Vane Angle (ϕ) degree	Material	Relative density	Particle size (mm)		
			3.2	5.5	7.4
0	H D P E	0.93	0.989	1.100	1.282
	Nylon	1.05	1.066	1.294	1.426
	Acetal	1.47	1.190	1.319	1.575
15	H D P E	0.93	1.178	1.285	1.463
	Nylon	1.05	1.180	1.301	1.516
	Acetal	1.47	1.392	1.546	1.846

5.4 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY

Bed pressure drop (ΔPb) is a crucial factor that gives a measure of the quality of fluidization. Experiments were conducted by [Botteril et al., 1982], [Mathur et al., 1986], [Nakamura et al., 1985] & [Saxena et al., 1990] in swirling fluidized bed and found that the behaviour of a swirling fluidized bed is determined mainly by the factors such as swirl velocity, relative velocity of particles, lateral mixing of particles and torroidal motion. But till date no experimental data is available on the influence of these parameters. In this experiment an attempt has been made to study the influence of some of these parameters.

The variation of bed pressure drop with superficial velocity was determined using all the four distributors. Acetal, nylon and high density polyethylene having particle size of 3.2 mm, 5.5 mm & 7.4 mm each were used as the bed materials. The bed weight in all the cases was 2.5 kg. Measurements of pressure drop across the bed were made at radial positions of 60 mm, 90 mm, 120 mm & 150 mm (the outermost periphery) from the centre of the distributor.

5.4.1 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DISTRIBUTOR ($\phi = 0^0$)

The variation of bed pressure drop with superficial velocity for distributor having angle $\phi = 0^0$ using acetal having particle size of 5.5 mm, 3.2 mm & 7.4 mm is given in figures 5.4.1, 5.4.2 & 5.4.3 respectively. 2.5 kg of bed material was used in all the cases.

It can be seen from the figure 5.4.1, that at a radial position of 60 mm from the centre of the distributor the bed pressure drop increases almost linearly upto a superficial velocity of 1.39 m/s, then it remains more or less the same till a superficial velocity of 1.97 m/s is attained, after which the bed pressure drop declines suddenly. At a radial position of 90 mm from the distributor centre the bed pressure drop varies almost linearly upto a superficial velocity of 1.39 m/s, thereafter it remains almost constant till a superficial velocity of 1.97 m/s is attained, after which it decreases slowly. At a radial position of 120 mm from the centre of the distributor the bed pressure drop increases almost linearly till a superficial velocity of 1.5 m/s is attained, thereafter it increases slowly upto a superficial velocity of 1.97 m/s, beyond which the rate of increase in bed pressure drop with superficial velocity is considerably more. At a radial position of 150 mm from the centre of the distributor more or less linear variation of bed pressure drop

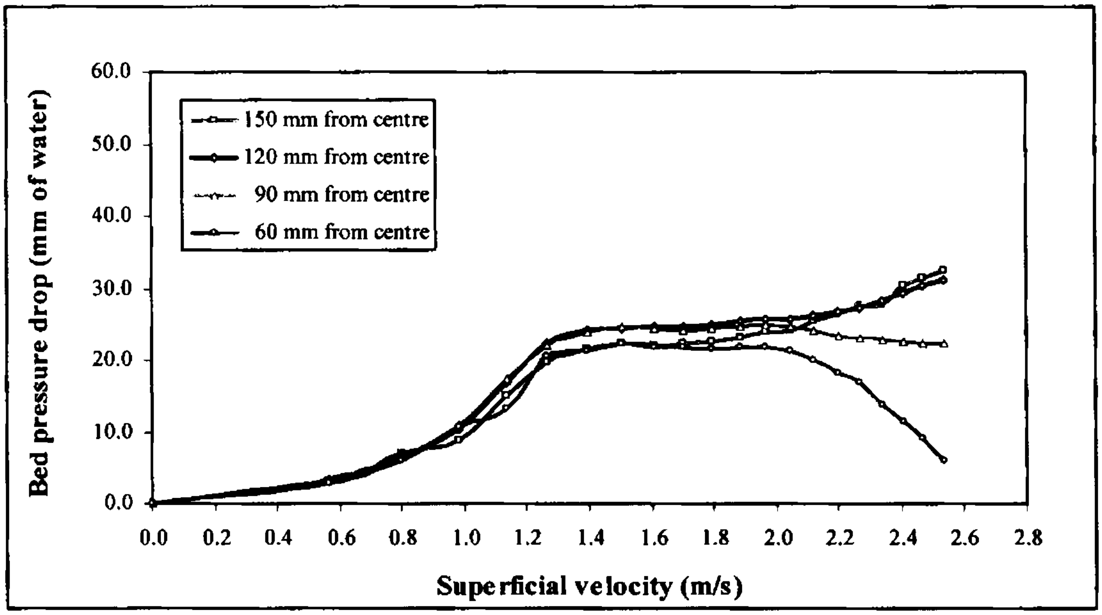


Figure 5.4.1 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection(Φ) = 0^0, using 5.5 mm acetal

with superficial velocity exists upto 1.27 m/s, thereafter it increases slowly till a superficial velocity of 1.87 m/s is attained, beyond which it increases sharply. It can also be noticed from the figure that, for the same superficial velocity the bed pressure drop increases as the distance from the centre of the distributor increases.

It is evident from the figure that during the initial stages the bed pressure drop increases almost linearly with the superficial velocity. This is due to the reason that the bed is not fluidized (packed bed) during these stages. After increasing the bed pressure drop remains almost constant for a period of 0.48 m/s of superficial velocity. It is due to the reason that after achieving the minimum fluidizing velocity the bed pressure drop remains almost constant. It can be noticed from the figure that after this stage, at a radial position of 60 mm from the centre of the distributor there is a sharp decrease in bed pressure drop, whereas there is a sharp increase corresponding to radial positions of 120 mm and 150 mm from the distributor centre .It can be further noticed that both these phenomenon occur almost at the same value of superficial velocity. This is due to the reason that a parabolic shape will be formed at the surface of the bed as a result of vortex formation in the bed. When this happens bubbling starts through the annular space. In addition to this, at the outer periphery of the distributor the wall friction also contributes to increase in bed pressure drop. It may also be noticed from the figure that, after achieving the minimum fluidizing velocity, the bed pressure drop at the radial position of 90 mm from the centre of the distributor remains almost constant. This is due to the reason that during vortex formation there is no change in the height of the bed at this position of the distributor.

While conducting the experiments it was observed that at a superficial velocity of 1.13 m/s very few particles started vibrating. When the superficial velocity reached 1.39 m/s more particles were seen to vibrate and bubbling was also observed. This velocity corresponds to the end of linear variation for bed pressure drop at radial positions of 90 mm, 120 mm &150 mm from the centre of the distributor. The particles started swirling slowly when the velocity reached 1.5 m/s. This velocity corresponds to the end of linear variation for bed pressure drop for particles at a radial position of 120 mm from the centre of the distributor. When the superficial velocity reached 2.41 m/s, the particles were completely separated from the cone. It can be noticed from figure 5.4.1 that this velocity fall within the declining regions of bed pressure drop at radial positions of 60 mm & 90 mm from the distributor centre, whereas these come within the increasing regions of bed pressure drop at radial positions of 120 mm & 150 mm from the distributor centre. When the superficial velocity reaches 2.54 m/s, the depth of the parabola formed at the surface of the bed due to vortex formation increases and the distributor plate is visible upto a radius of 60 mm from the centre. This velocity falls within the region of sudden decline in pressure drop in the figure for bed materials at a radial position 60 mm from the distributor centre. It can also be noticed from figure 5.4.1 that at this velocity the bed pressure drop is shooting downwards and beyond this it is almost zero.

The results of the experiments conducted with the same distributor having particle sizes of 3.2 mm & 7.4 mm are given in figures 5.4.2 & 5.4.3(Appendix – II). It can be observed that these results are similar to that of particles having 5.5 mm size.

5.4.2 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DISTRIBUTOR ($\phi = 5^0$)

Figures 5.4.4, 5.4.5 & 5.4.6 reveal the variation of bed pressure drop with superficial velocity for distributor having angle $\phi = 5^0$, employing 2.5 kg of acetal having particle sizes of 5.5 mm, 3.2 mm & 7.4 mm respectively as the bed material.

It can be seen from figure 5.4.4 that, at a radial position of 60 mm from the centre of the distributor the bed pressure drop varies almost linearly upto a superficial velocity of 1.39 m/s, thereafter it decreases very slowly till a superficial velocity of 2.05 m/s is reached, beyond which it decreases suddenly. At a radial position of 90 mm from the distributor centre, more or less linear variation of distributor pressure drop prevails upto a superficial velocity of 1.5 m/s, thereafter it remains almost constant. It is quite evident from the figure that at a radial position of 120 mm from the centre of the distributor, the bed pressure drop increases almost linearly upto a superficial velocity of 1.61 m/s, thereafter it remains almost the same, beyond which it increases suddenly. At a radial position of 150 mm from the distributor centre it can be noticed from the figure that

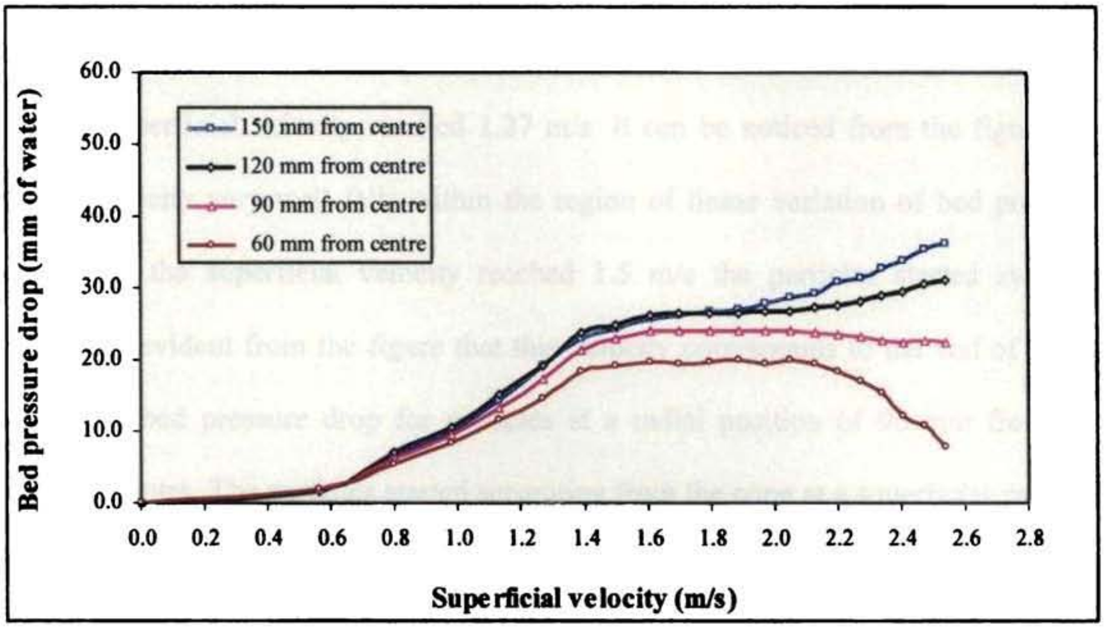


Figure 5.4.4 Variation of bed pressure drop with superficial velocity for having angle of air injection (Φ) = 5°, using 5.5 mm acetal

almost linear variation of bed pressure drop occurs upto a superficial velocity of 1.7m/s, after which it increases slowly till superficial velocity attains 1.9 m/s, thereafter it increases sharply (at a faster rate than that in the case of 120 mm from the distributor centre).

This is due to the reason that in addition to the increased bed height the fluidizing gas has to overcome the wall friction also. It can also be noticed from the figure that, similar to the previous case (section 5.4.1), for the same superficial velocity the bed pressure drop increases as the distance from the centre of the distributor increases.

Thus it is quite clear that the variation of bed pressure drop with superficial velocity is almost similar as mentioned in section 5.4.1. Hence the reasons for these variations are also the same as that for the distributor having angle $\phi = 0^0$. But it may be noticed here that the superficial velocity at the beginning and end of various regimes (linear variation of bed pressure drop, constant bed pressure drop, sudden decrease or sudden increase of bed pressure drop) get delayed compared to the distributor having angle $\phi = 0^0$.

While conducting the experiments it was observed that few particles started vibrating when the superficial velocity reached 1.27 m/s. It can be noticed from the figure 4.5 that this velocity very well falls within the region of linear variation of bed pressure drop. When the superficial velocity reached 1.5 m/s the particles started swirling slowly. It is evident from the figure that this velocity corresponds to the end of linear variation of bed pressure drop for particles at a radial position of 90 mm from the distributor centre. The particles started separating from the cone at a superficial velocity of 2.48 m/s. It can be seen from the figure that similar to the previous case this velocity fall within the region of sudden decrease or sudden increase in bed pressure drop in the figure depending upon the distance of the radial position from the centre of the distributor. At a superficial velocity 2.6 m/s the depth of the parabola at the bed surface increased to such an extend to make the distributor plate visible upto a radius of 60 mm from the distributor centre. This velocity falls within the region of sudden decline in pressure drop in the figure for bed materials at a radial position 60 mm from the distributor centre. It is well evident from the figure that, at this velocity the bed pressure drop is shooting downwards and beyond which the bed pressure drop is almost zero. It is clear from the above observations that the superficial velocity required for the starting of various bed phenomena such as swirl motion & separation of particles from the centre cone get delayed when compared with the distributor having angle $\phi = 0^0$.

The results of the experiments conducted with the same distributor using acetal having particle size of 3.2 mm and 7.4 mm are depicted in figures 5.4.5 & 5.4.6 respectively

(Appendix – II). This is almost similar to the results presented in figure 5.4.4.

5.4.3 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DISTRIBUTOR ($\phi = 10^0$)

Experiments were also conducted to determine the variation of bed pressure drop with superficial velocity for distributor having angle $\phi = 10^0$. 2.5 kg of acetal having particle sizes of 5.5 mm, 3.2 mm & 7.4 mm were employed as the bed materials and the results of these experiments are depicted in figures 5.4.7, 5.4.8 & 5.4.9 respectively. It is quite clear from figure 5.4.7, that at a radial position of 60 mm from the centre of the distributor the bed pressure drop varies almost linearly upto a superficial velocity of 1.5

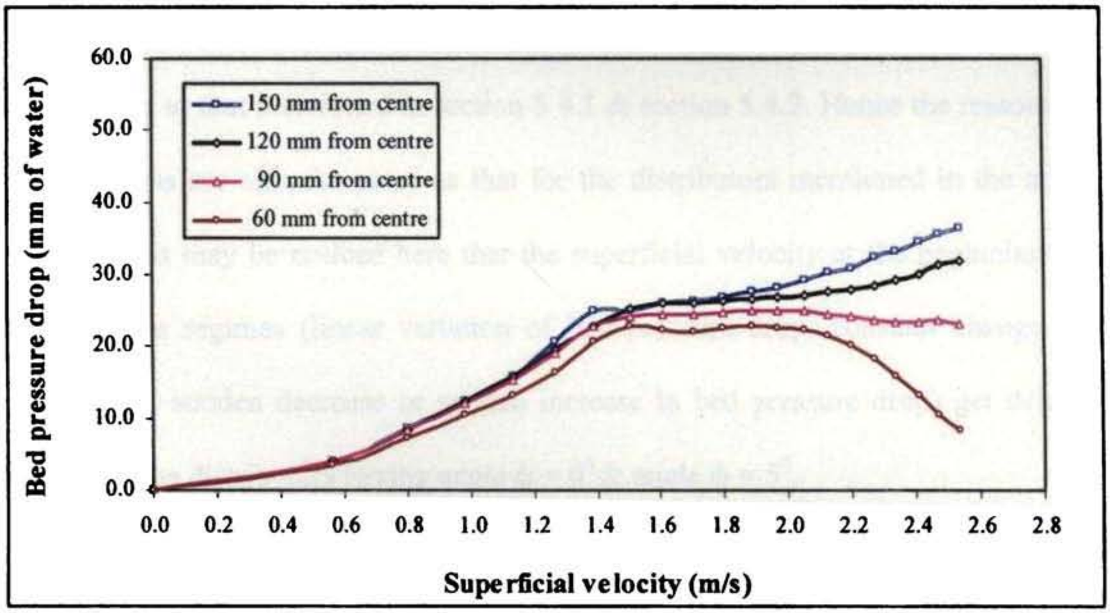


Figure 5.4.7 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 10^o, using 5.5 mm acetal

m/s, thereafter it decreases slowly till a superficial velocity of 1.97 m/s is attained, after which it decreases suddenly. At a radial position of 90 mm from the distributor centre more or less linear variation of bed pressure drop prevails till a superficial velocity of

1.5 m/s is attained, thereafter it increases slowly upto a superficial velocity of 1.89 m/s, beyond which it decreases slowly. At a radial position of 120 mm from the distributor centre, the bed pressure drop varies almost linearly upto a superficial velocity of 1.61 m/s, after which it increases slowly till a superficial velocity of 1.97 m/s is reached, beyond which it increases at a faster rate. At a radial position of 150 mm from the centre of the distributor the bed pressure drop increases almost linearly upto a superficial velocity of 1.61 m/s, thereafter it increases slowly till a superficial velocity of 1.97 m/s is attained, beyond which it increases suddenly. It is also quite evident from the figure that, similar to the previous cases (section 5.4.1 & section 5.4.2) for the same superficial velocity the bed pressure drop increases as the distance from the centre of the distributor increases.

Thus it is quite clear that the variation of bed pressure drop with superficial velocity is almost similar to that mentioned in section 5 4.1 & section 5.4.2. Hence the reasons for these variations are also the same as that for the distributors mentioned in the above sections. But it may be noticed here that the superficial velocity at the beginning and end of various regimes (linear variation of bed pressure drop, constant change bed pressure drop, sudden decrease or sudden increase in bed pressure drop) get delayed compared to the distributors having angle $\phi = 0^0$ & angle $\phi = 5^0$.

The following observations were made while conducting the experiments. When the superficial velocity reached 1.27 m/s, few particles were seen to start vibrating. It can be noticed from figure 5.4.7 that this velocity very well comes under the linear variation of bed pressure drop. When the superficial velocity reached 1.61 m/s, particles started swirling. From the figure it can be seen that this velocity corresponds to the end of linear variation of bed pressure drop for particles at a radial position of 120 mm from the distributor centre. It was observed that the particles were completely separated from the cone at the centre at a superficial velocity of 2.48 m/s. This velocity fall within the region of sudden decrease in bed pressure drop for particles at a radial position of 60 mm from the distributor centre and simultaneously the region of sudden increase in bed pressure drop for particles at radial positions of 120 mm & 150 mm from the distributor centre. When the superficial velocity reached 2.6 m/s the depth of the parabola at the bed surface increased to a value high enough to make the distributor plate visible upto a radius of 60 mm from the distributor centre. From the figure it can be seen that corresponding to this velocity the bed pressure drop is shooting downwards. From the above observations it is evident that the superficial velocity required for initiating various bed phenomena are comparatively higher than that of the distributor having angle $\phi = 5^0$.

The results of the experiments conducted with the same distributor using acetal having particle size of 3.2 mm and 7.4 mm which are given in figures 5.4.8 & 5.4.9 respectively (Appendix – II) are also similar to that conducted with the same distributor using acetal of particle size 5.5 mm.

5.4.4 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DISTRIBUTOR ($\phi = 15^0$)

The variation of bed pressure drop with superficial velocity for distributor having angle $\phi = 15^0$ using acetal having particle sizes of 5.5 mm, 3.2 mm & 7.4 mm is given in figures 5.4.10, 5.4.11 & 5.4.12 respectively. The bed weight adopted in all these cases was 2.5 kg.

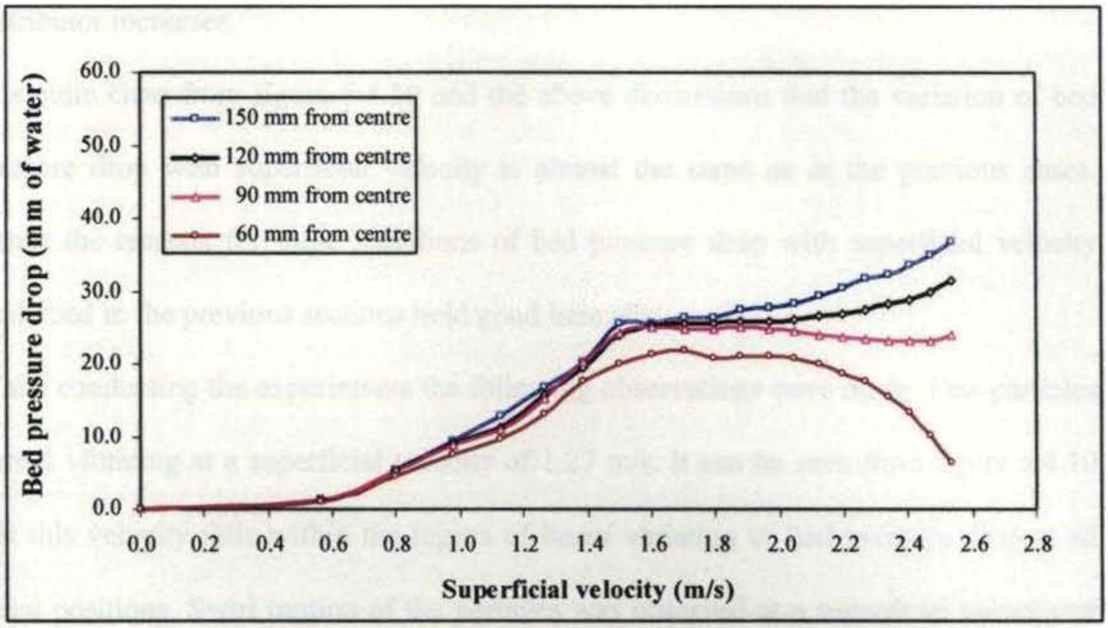


Figure 5.4.10 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 5.5 mm acetal

It can be seen from figure 5.4.10 that, at a radial position of 60 mm from the centre of the distributor the bed pressure drop increases almost linearly upto a superficial velocity of 1.5 m/s, after that it increases slowly till a superficial velocity of 1.88 m/s is reached, after which it decreases suddenly. At a radial position of 90 mm from the distributor centre the bed pressure drop increases more or less linearly till a superficial velocity of 1.6 m/s is attained, where after it decreases slowly. At a radial position of 120 mm from the centre of the distributor an approximate linear variation of bed pressure drop exists upto a superficial velocity of 1.6 m/s, thereafter it remains constant till a superficial velocity of 1.8 m/s is reached, beyond which it increases suddenly. At a radial position of 150 mm from the distributor centre bed pressure drop varies almost linearly upto a superficial velocity of 1.6 m/s, after which it remains unchanged till a superficial velocity of 1.8 m/s is attained, afterwhich it increases suddenly. It can also be noticed from the figure that, similar to the previous cases for the same superficial velocity the bed pressure drop increases as the distance from the centre of the distributor increases.

It is quite clear from figure 5.4.10 and the above discussions that the variation of bed pressure drop with superficial velocity is almost the same as in the previous cases. Hence the reasons for these variations of bed pressure drop with superficial velocity explained in the previous sections hold good here also.

While conducting the experiments the following observations were made. Few particles started vibrating at a superficial velocity of 1.27 m/s. It can be seen from figure 5.4.10 that this velocity falls within the region of linear variation of bed pressure drop at all radial positions. Swirl motion of the particles was observed at a superficial velocity of 1.7 m/s. It can be noticed from the figure that this velocity falls after the region of linear variation of particles at all radial positions from the distributor centre. The particles were completely separated from the centre at a superficial velocity of 2.48 m/s. It is quite clear from the figure that this point fall within the region of sudden decrease in bed pressure drop for particles at a radial position of 60 mm from the distributor centre and simultaneously the region of sudden increase in bed pressure drop for particles at radial positions of 120 mm & 150 mm from the distributor centre. When the superficial velocity reached 2.6 m/s, the depth of the parabola formed at the bed surface increased to a value high enough to make the distributor plate visible upto a radius of 60 mm from the centre. It can be seen from the figure that corresponding to this velocity the bed pressure drop is shooting downwards and almost approaching zero. It is evident from the above observations that the superficial velocity required for the starting of various bed phenomena such as swirl motion & separation of particles from the centre cone is slightly more when compared with distributor having angle $\phi = 10^0$.

The results of the experiments conducted with the same distributor using acetal having particle sizes of 3.2 mm & 7.4 mm which are given in figures 5.4.10 & 5.4.12 (Appendix – II) respectively are found to be similar to the results given in figure 5.4.10.Experiments on the variation of bed pressure drop with superficial velocity were also conducted with the same distributor using 3.2 mm diameter, 5.5 mm diameter & 7.4 mm diameter high density polyethylene and nylon. The results of the experiments conducted using high density polyethylene are given in figures from 5.4.13 to 5.4.15 (Appendix – II) and the results of the experiments conducted using nylon are depicted in figures from 5.4.16 to 5.4.18 (Appendix – II). These figures reveal that results of these experiments are similar to that of acetal.

Based on these experimental results the following conclusions can be made. The bed pressure drop increases as the distance from the distributor centre increases. At a radial position of 90 mm from the distributor centre, after attaining minimum fluidizing velocity the bed pressure drop remains constant irrespective of superficial velocity.

5.5 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT PARTICLE SIZE

Bed pressure drop being a crucial factor in fluidization process, an attempt is made to determine the effect of particle size on the variation bed pressure drop with superficial velocity. Till date no experimental data is available on the effect of particle size on the variation of bed pressure drop with superficial velocity in swirling fluidized bed.

On the basis of experiments conducted to determine the variation of distributor pressure drop with superficial velocity, it was found that the distributor pressure drop was minimum for the distributor having angle $\phi = 15^0$ (section 2). Hence to determine the effect of particle size on the variation bed pressure drop with superficial velocity, experiments were conducted with the same distributor. The bed pressure drop measurements were made at radial positions of 60 mm, 90 mm, 120 mm & 150 mm from the centre of the distributor using micromanometer. Experiments were conducted using acetal, high density polyethylene and nylon having particle sizes of 3.2 mm, 5.5 mm & 7.4 mm as the bed material.

5.5.1 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT PARTICLE SIZE USING HDPE AS THE BED MATERIAL

The variation of bed pressure drop with superficial velocity for different particle size at radial positions of 90 mm, 60 mm, 120 mm & 150 mm from the centre of the distributor is given in figure 5.5.1, 5.5.2, 5.5.3 & 5.5.4 respectively. The experiment was conducted with distributor having angle $\phi = 15^0$ and using 2.5 kg of high density polyethylene having relative density of 0.93 as the bed material. It may be noticed from figure 5.5.1 that at a radial position of 90 mm from the centre of the distributor, after attaining minimum fluidization velocity the bed pressure drop remains unchanged since there is no change in the height of the bed at this position of the distributor. It is quite evident from figure 5.5.1 that, for the same superficial velocity the bed pressure drop decreases with increase in particle size. It is due to the reason that the voidage decreases with decrease in particle size (table 4.3.4). So the velocity of gas flow through the interstices of the bed will be comparatively more for smaller size particles. In addition to this the surface area of the bed particles also will increase as the size of the particles decreases. Hence the frictional resistance offered by these particles to the gas flow will be comparatively more.

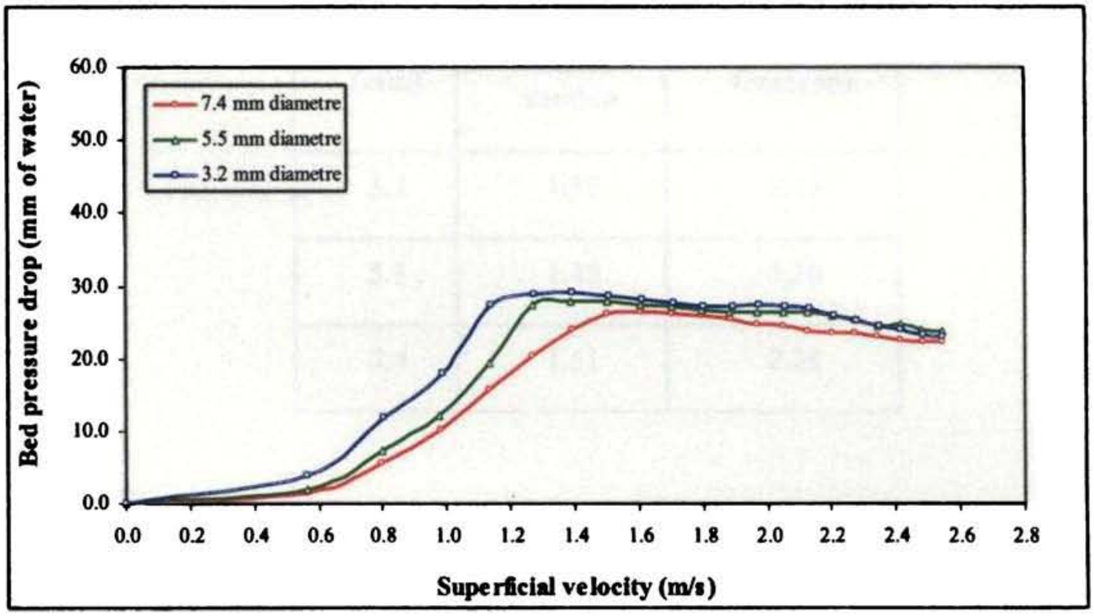


Figure 5.5.1 Variation of bed pressure drop with superficial velocity for different particle size at 90 mm from the centre of the distributor (Φ= 15°), using HDPE

This in turn leads to an increase in bed pressure drop for smaller size particles. It can also be observed that after swirl motion vortex formation occur in the bed. This leads to the formation of a parabolic shape at the surface of the bed.

It is quite evident from figure 5.5.1 that the minimum fluidizing velocity increases with increase in particle size. It may also be noticed from table 5.5.1 that the superficial velocity required for initiating various bed phenomena such as swirl motion & separation of particles from the cone at the centre increases with increase in particle size. This is due to the reason that the drag force decreases when the particle size increases.

Table 5.5.1 Bed phenomena of high density polyethylene particles

Particle size (mm)	Superficial velocity at the beginning of (m/s)	
	Swirl motion	Seperation fromcone
3.2	1.39	2.13
5.5	1.45	2.20
7.4	1.51	2.28

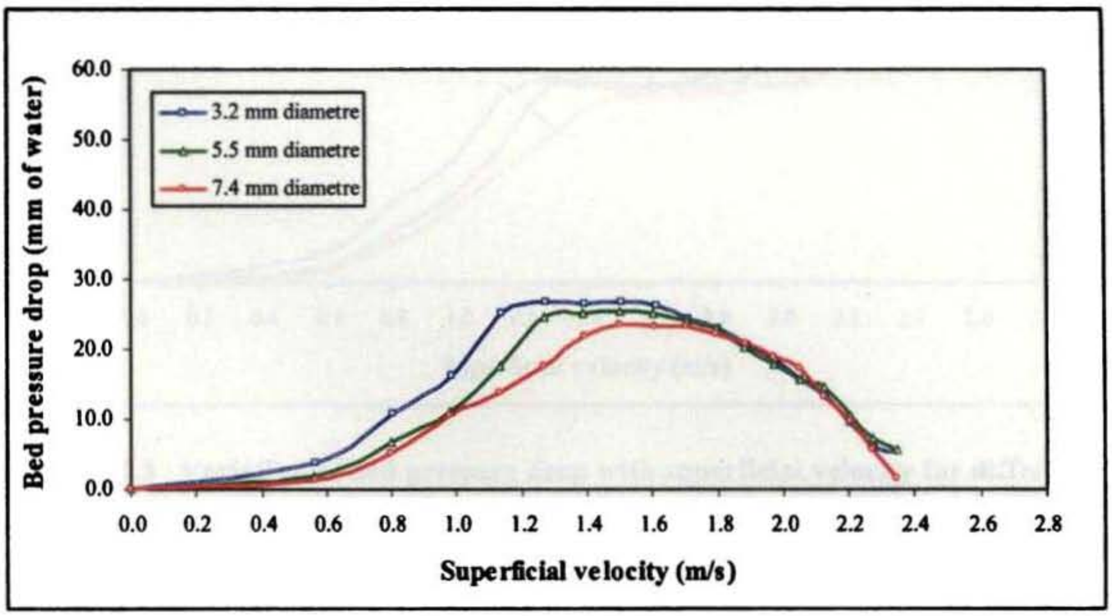


Figure 5.5.2 Variation of bed pressure drop with superficial velocity for different particle size at 60 mm from the centre of the distributor (Φ= 15°), using HDPE

The results of the experiments regarding the effect of particle size on the variation of bed pressure drop with superficial velocity conducted with the same distributor using the same bed materials at a radial position of 60 mm from the distributor centre is presented in figure 5.5.2. It can be seen from the figure that the variation of bed pressure drop with superficial velocity is almost similar to that in figure 5.5.1. It may be noticed here that the bed pressure drop decreases suddenly after a superficial velocity of 1.6 m/s.

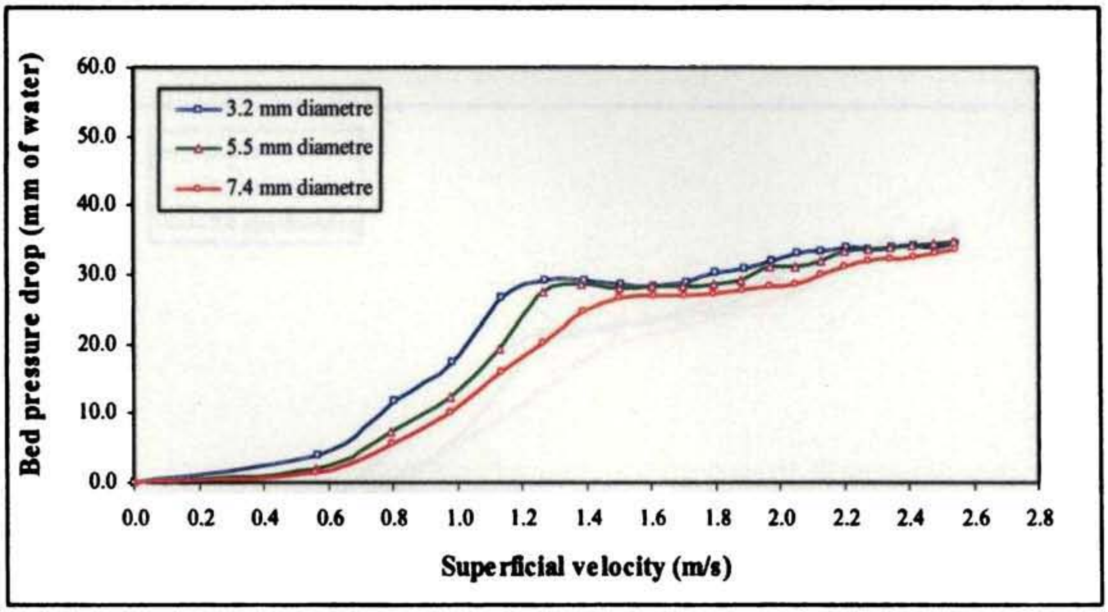


Figure 5.5.3 Variation of bed pressure drop with superficial velocity for different particle size at 120 mm from the centre of the distributor (Φ= 15°), using HDPE

The results of the same experiments conducted with the same distributor and the same bed materials at a radial position of 120 mm from the centre of the distributor is depicted in figure 5.5.3. A similar behaviour of bed pressure drop can be experienced here also. It can be noticed here that after a superficial velocity of 1.6 m/s there is an increase in bed pressure drop.

Figure 5.5.4 reveals the results of these experiments at a radial position of 150 mm from the distributor centre. It is quite evident from the figure that the variation of bed pressure drop is similar as in the previous cases. But it may be noticed here that the bed pressure drop increases suddenly after a superficial velocity of 1.6 m/s. This radial position corresponds to the outermost region of the distributor. In addition to the resistance due to increase in bed height wall friction also will be experienced here.

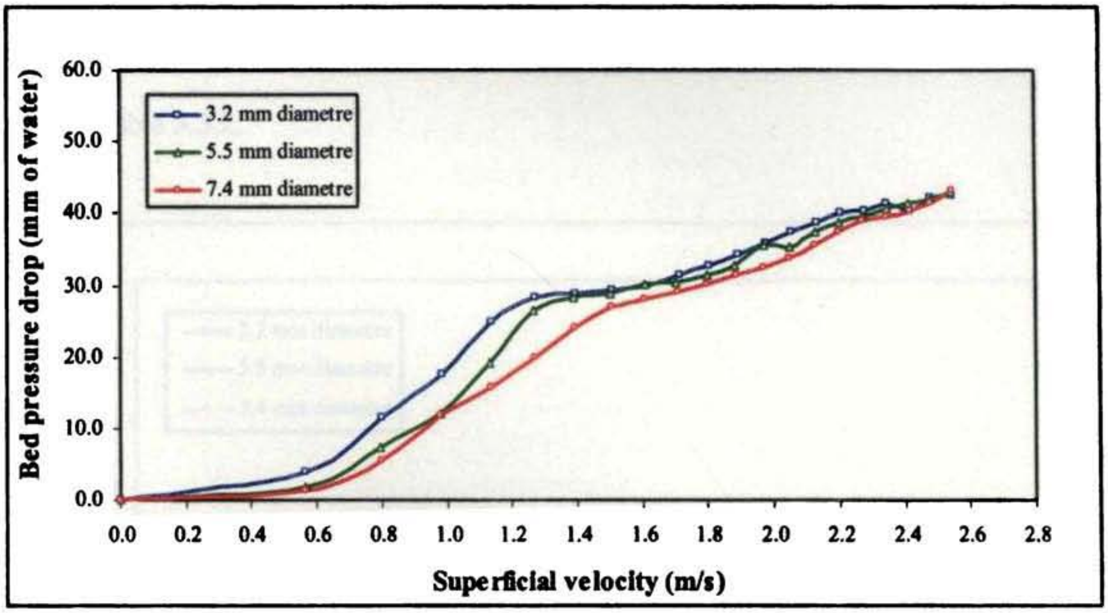


Figure 5.5.4 Variation of bed pressure drop with superficial velocity for different particle size at 150 mm from the centre of the distributor (Φ= 15°), using HDPE

These lead to sharp increase in bed pressure drop at this radial position of the distributor.

5.5.2 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT PARTICLE SIZE USING ACETAL AS THE BED MATERIAL

Experiments were conducted to determine the effect of particle size on the variation of bed pressure drop with superficial velocity employing distributor having angle $\phi = 15^0$. 2.5 kg of acetal having relative density of 1.47 and particle sizes of 3.2 mm, 5.5 mm & 7.4 mm were used as the bed material. Bed pressure drop measurements were made at radial positions of 60 mm, 90 mm, 120 mm & 150 mm from the centre of the distributor. The results of these experiments are depicted in figure 5.5.5 and figures from 5.5.6 to 5.5.8 (Appendix – III). The various bed phenomena for these particles are given in table 5.5.2.

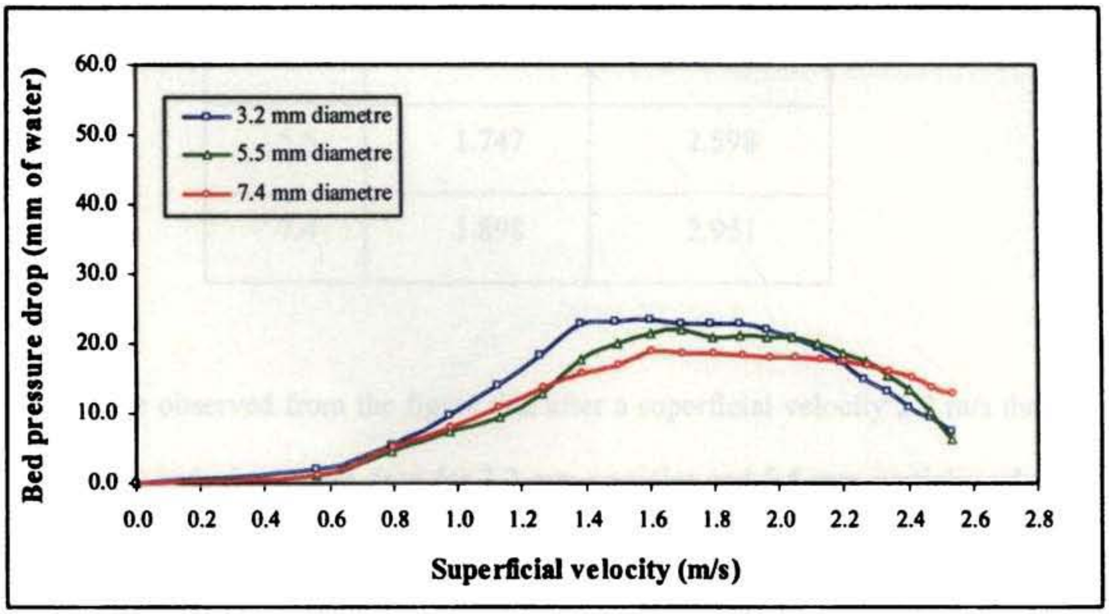


Figure 5.5.5 Variation of bed pressure drop with superficial velocity for different particle size at 60 mm from the centre of the distributor (Φ= 15°), using acetal

It is quite clear from these figures that in the case of acetal also the minimum fluidizing velocity increases with increase in particle size. It may also be noticed from table 5.5.2 that the superficial velocity required for initiating various bed phenomena such as swirl motion & separation of particles from the cone at the centre increases with increase in particle size. The reasons for this behaviour have been already explained in the previous section. It is also evident from these figures that the variation of bed pressure drop with superficial velocity is similar to that for high density polyethylene. It may be noticed from figure 5.5.5 that the sudden drop in bed pressure drop in the case of acetal occur at a later stage compared with that of HDPE. This is due to the reason that the relative density of acetal is more than that of HDPE.

Table 5.5.2 Bed phenomena of acetal particles

Particle size (mm)	Superficial velocity at the beginning of (m/s)	
	Swirl motion	Seperation from cone
3.2	1.548	2.405
5.5	1.747	2.598
7.4	1.898	2.951

It may also be observed from the figure that after a superficial velocity 2.2 m/s there is a sudden drop in bed pressure drop for 3.2 mm particles and 5.5 mm particles whereas the drop in bed pressure drop corresponding to the same point is rather slow for 7.4 m particles. It is clear from the figure that beyond a superficial velocity of 2.2 m/s the bed pressure drop increases with increase in particle size. This point corresponding to superficial velocity falls in the region of separation of particles from the distributor centre. The drag force decreases with increase in particle size. So the increase in the depth of the parabola formed at the surface of the depth as a result of vortex formation will be delayed as the particle size increases. Hence beyond a superficial velocity of 2.2 m/s the bed pressure drop increases with increase in particle size. A similar pattern of behaviour was noticed at radial positions of 90 mm, 120 mm & 150 mm from the centre of the distributor which are given in figurers 5.5.6, 5.5.7 & 5.5.8 respectively (Appendix – III).

5.5.3 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT PARTICLE SIZE USING NYLON AS THE BED MATERIAL

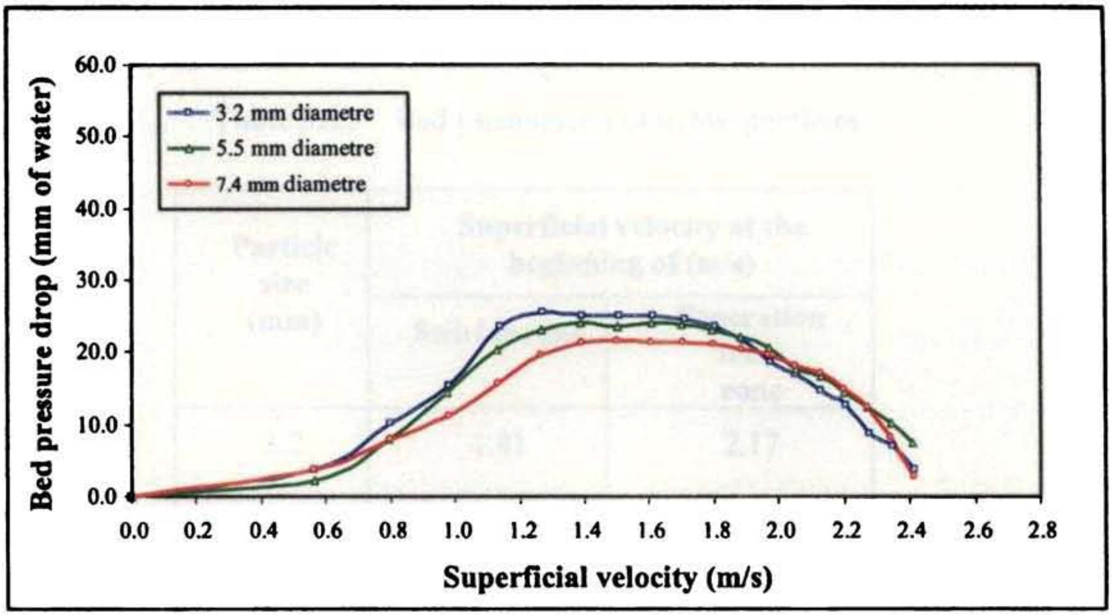


Figure 5.5.9 Variation of bed pressure drop with superficial velocity for different particle size at 60 mm from the centre of the distributor (Φ= 15°), using nylon

As an attempt to determine the effect of particle size on the variation of bed pressure drop with superficial velocity using nylon particles, experiments were conducted with distributor having angle $\phi = 15^0$. 2.5 kg of nylon having relative density of 1.05 and particle sizes of 3.2 mm, 5.5 mm & 7.4 mm were used as the bed materials. Similar to the previous cases measurements of bed pressure drop were made at radial positions of 60 mm, 90 mm, 120 mm & 150 mm from the distributor centre and the results of these experiments are given in figure 5.5.9 and figures 5.5.10, 5.5.11 & 5.5.12 respectively (Appendix – III). The various bed phenomena for these particles are given in table 5.5.3.

It may also be noticed from table 5.5.3 that in the case of nylon particles also the superficial velocity required for initiating various bed phenomena such as swirl motion & separation of particles from the cone at the centre increases with increase in particle size. The reasons for this behaviour have been already explained in section 5.5.1. It can be noticed from these figures that the variation of bed pressure drop with superficial velocity is similar to that for high density polyethylene and acetal. It is evident from

Table 5.5.3 Bed phenomena of nylon particles

Particle size (mm)	Superficial velocity at the beginning of (m/s)	
	Swirl motion	Seperation from cone
3.2	1.41	2.17
5.5	1.48	2.35
7.4	1.55	2.42

figure 5.5.9 that the sudden decrease in bed pressure drop occur at a superficial velocity higher than that of high density polyethylene, but lower than that of acetal. This is due to the reason that nylon has a relative density higher than that of high density polyethylene, but lesser than that of acetal. It is also evident from the figure that the sudden fall in bed pressure drop occurs at a superficial velocity of 1.9 m/s, which is lower than that for acetal particles at the same radial position of the distributor. It may be further observed that after a superficial velocity of 1.9 m/s the variation of bed pressure drop with superficial velocity for different particle size is just the opposite compared with that before a superficial velocity of 1.9 m/s, similar to that of acetal particles, ie. after a superficial velocity of 1.9 m/s the bed pressure drop decreases with decrease in particle size. The reason for this change in variation has been already explained in section 5.5.2.

The variations of bed phenomena for the different particles are given in table 5.5.4. It can be noticed from this table that the rate of change of superficial velocity/unit size required for various bed phenomena increases with particle density.

The experiments conducted and the results lead to the following conclusions. The bed pressure drop increases with decrease in particle size for constant bed weight. The superficial velocity required for initiating various bed phenomena such as swirl motion & separation of particles from the cone at the centre increases with increase in particle size.

The superficial velocity required for various regimes such as linear variation of bed pressure drop, constant bed pressure drop and sudden increase or decrease in bed pressure drop increases with increase in particle size. The rate of change of superficial velocity per unit size required for required for various bed phenomena increases with increase in particle density.

Table 5.5.4 Variation of bed phenomena for the different particles

Range of particle size (mm)	Rate of change of superficial velocity/unit size required for (m/s/mm)					
	H D P E (r.d = 0.93)		nylon (r.d = 1.05)		acetal (r.d = 1.47)	
	Swirl motion	Seperation from cone	Swirl motion	Seperation from cone	Swirl motion	Seperation from cone
3.2 – 5.5	0.026	0.030	0.030	0.078	0.086	0.084
5.5 – 7.4	0.032	0.042	0.037	0.037	0.079	0.186
Mean	0.029	0.036	0.034	0.058	0.083	0.135

5.6 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT DENSITY MATERIALS

This is a venture to determine the effect of particle density on the variation of pressure drop across the bed with superficial velocity. Till date no experimental data is available on the variation of bed pressure drop with superficial velocity for different density materials in swirling fluidized bed.

Similar to that mentioned in section 5.5, experiments were conducted with the distributor having angle $\phi = 15^0$. Measurements of pressure drop across the bed were made at radial positions of 60 mm, 90 mm, 120 mm & 150 mm from the centre of the distributor using micromanometer. Experiments were conducted using 2.5 kg of high density polyethylene, nylon and acetal (having relative densities of 0.93, 1.05 and 1.47 respectively) as the bed materials. Particles having diameters 3.2 mm, 5.5 mm & 7.4 mm each of these materials were utilized for the experiment.

5.6.1 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT DENSITY MATERIALS USING 3.2 MM SIZE PARTICLES AS THE BED MATERIAL

The variation of bed pressure drop with superficial velocity for different density materials using 3.2 mm particles at radial positions of 60 mm, 90 mm, 120 mm & 150 mm from the distributor centre is given in figures 5.6.1, 5.6.2, 5.6.3 & 5.6.4 respectively. It is quite evident from figure 5.6.3 that, for the same superficial velocity the pressure drop across the bed increases with decrease in particle density. It is due to the reason that, for constant bed weights the height of the bed increases with decrease in particle density. This increases the distance to be traversed by the fluidizing gas through the bed. Therefore the resistance offered to the flow of gas through the bed increases with decrease in particle density. It can be observed from the figure that during the initial stages upto a superficial velocity of 1.2 m/s the bed pressure drop increases almost linearly. The bed is in a packed condition till this value of superficial velocity. Therefore the variation of bed pressure drop with superficial velocity will be similar to that of packed bed. After that the bed pressure drop remains almost constant till a superficial velocity of 1.8 m/s is reached. This is due to the reason that eventhough swirl motion has just started there; there is no change in bed height. But a sharp rise in

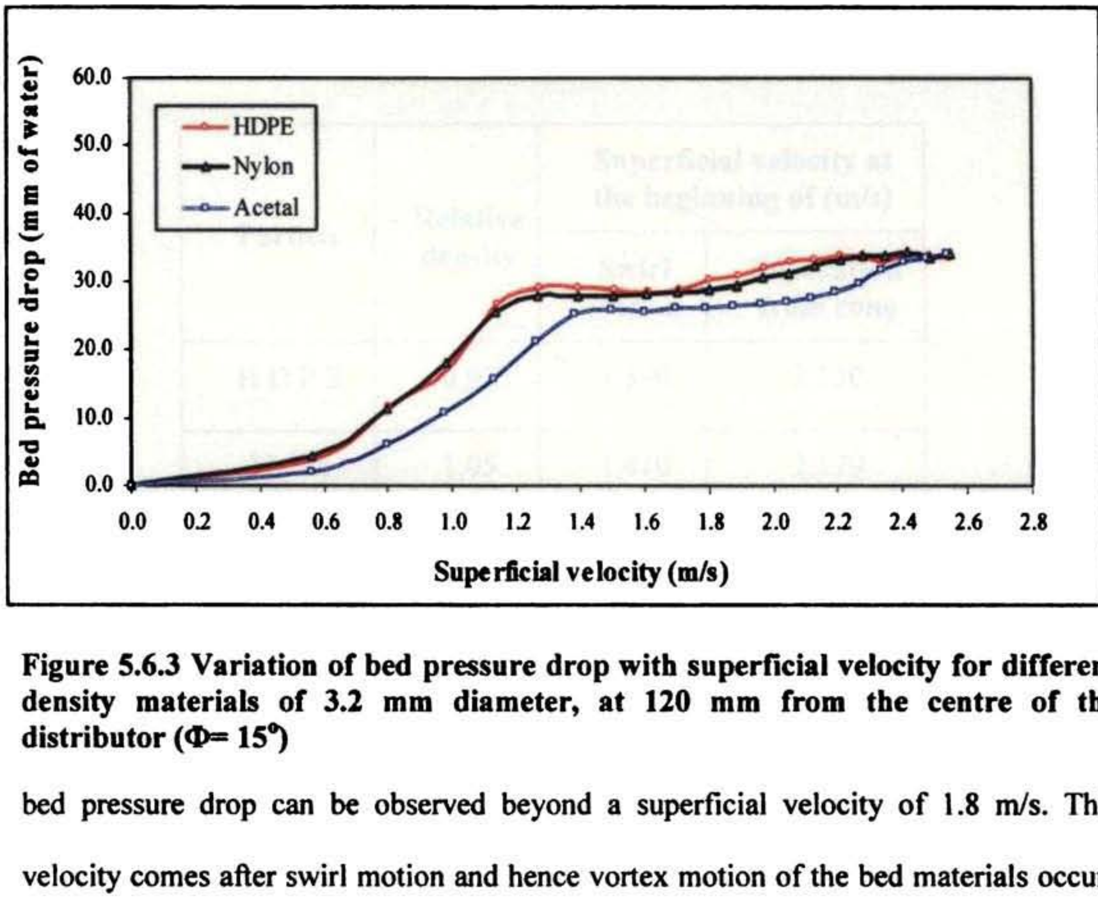


Figure 5.6.3 Variation of bed pressure drop with superficial velocity for different density materials of 3.2 mm diameter, at 120 mm from the centre of the distributor (Φ= 15°)

bed pressure drop can be observed beyond a superficial velocity of 1.8 m/s. This velocity comes after swirl motion and hence vortex motion of the bed materials occurs in this region. As a result of this the bed materials from the inner periphery of the distributor (at radial positions of 60 mm & 90 mm from the distributor centre) get shifted towards the outer periphery of the distributor (at radial positions of 120 mm & 150 mm from the distributor centre). This is quite evident from figure 5.6.1 & figure 5.6.2 (Appendix-IV) which corresponds to the variation of bed pressure drop with superficial velocity for different density materials using particles of the same size at radial positions of 60 mm & 90 mm from the distributor centre respectively.

From figure 5.6.4 (which corresponds to the variation of bed pressure drop with superficial velocity for different density materials using particles of the same size at a radial position of 150 mm from the distributor centre) - (Appendix-IV) it may be observed that beyond a superficial velocity of 1.8 m/s, the increase in bed pressure drop with superficial velocity is more sharp. This is due to the effect of wall friction.

Table 5.6.1 Bed phenomena of 3.2 mm particles

Particle	**Relative density**	**Superficial velocity at the beginning of (m/s)**	
		Swirl motion	**Seperation from cone**
H D P E	0.93	1.390	2.130
Nylon	1.05	1.410	2.170
Acetal	1.47	1.548	2.405

The superficial velocity required for the starting of various bed phenomena such as swirl motion and separation of particles from the cone was observed while conducting the experiments and it is given in table 5.6.1.It is quite evident that the superficial velocity required for initiating these phenomena increases with increase in particle density.

5.6.2 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT DENSITY MATERIALS USING 5.5 MM SIZE PARTICLES AS THE BED MATERIAL

The variation of bed pressure drop with superficial velocity for different density materials using 5.5 mm particles at radial positions of 60 mm, 90 mm, 120 mm and 150 mm from the distributor centre are given in figures 5.6.5, 5.6.6, 5.6.7 & 5.6.8 respectively. Table 5.6.2 gives the bed phenomena for these particles. It is evident from these figures that the effect of particle density on the variation of bed pressure drop with superficial velocity for 5.5 mm particles is similar to that of 3.2 mm particles.

It is clear from figure 5.6.5 that at a radial position of 60 mm from the distributor centre, the bed pressure drop varies almost linearly with superficial velocity upto 1.3 m/s, thereafter it decreases slowly till a superficial velocity of 1.9 m/s is reached, beyond which it decreases suddenly. At a radial position of 90 mm from the distributor centre it can be seen from figure 5.6.6 that the bed pressure drop increases almost linearly with superficial velocity upto 1.3 m /s, where after it decreases slowly. The variation of bed pressure drop with superficial velocity at a radial position of 120 mm from the distributor centre is given in figure 5.6.7. It may be noticed from this figure that the bed pressure drop increases almost linearly with superficial velocity till 1.3 m/s, then it remains constant upto a superficial velocity of 1.9 m/s, after which it increases slowly.

Table 5.6.2 Bed phenomena of 5.5 mm particles

Particle	Relative density	Superficial velocity at the beginning of (m/s)	
		Swirl motion	Seperation from cone
H D P E	0.93	1.450	2.200
Nylon	1.05	1.480	2.350
Acetal	1.47	1.747	2.598

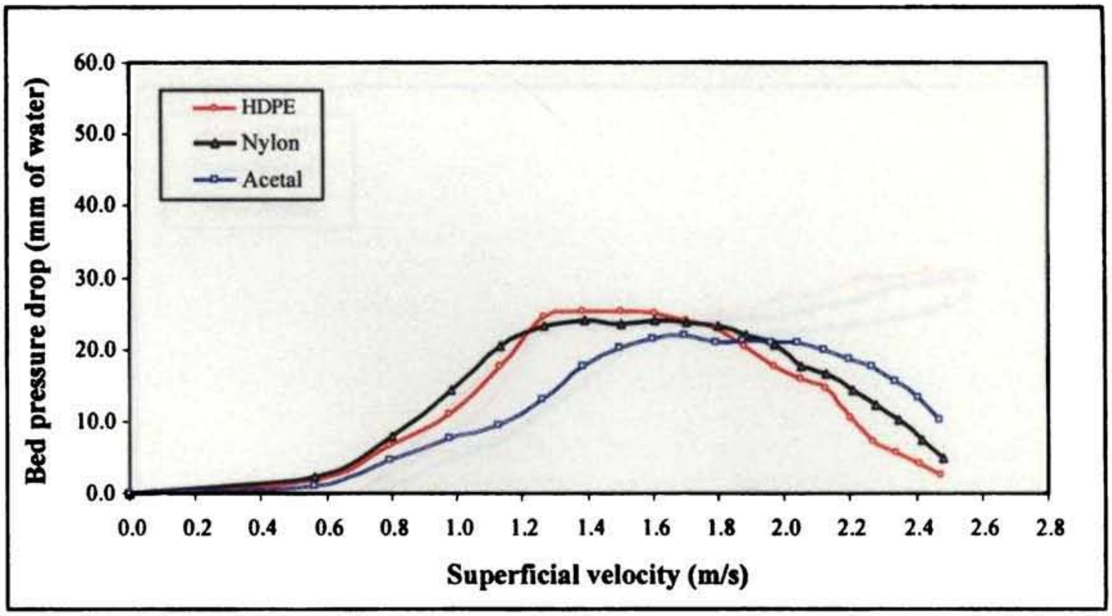


Figure 5.6.5 Variation of bed pressure drop with superficial velocity for different density materials of 5.5 mm diameter, at 60 mm from the centre of the distributor (Φ= 15°)

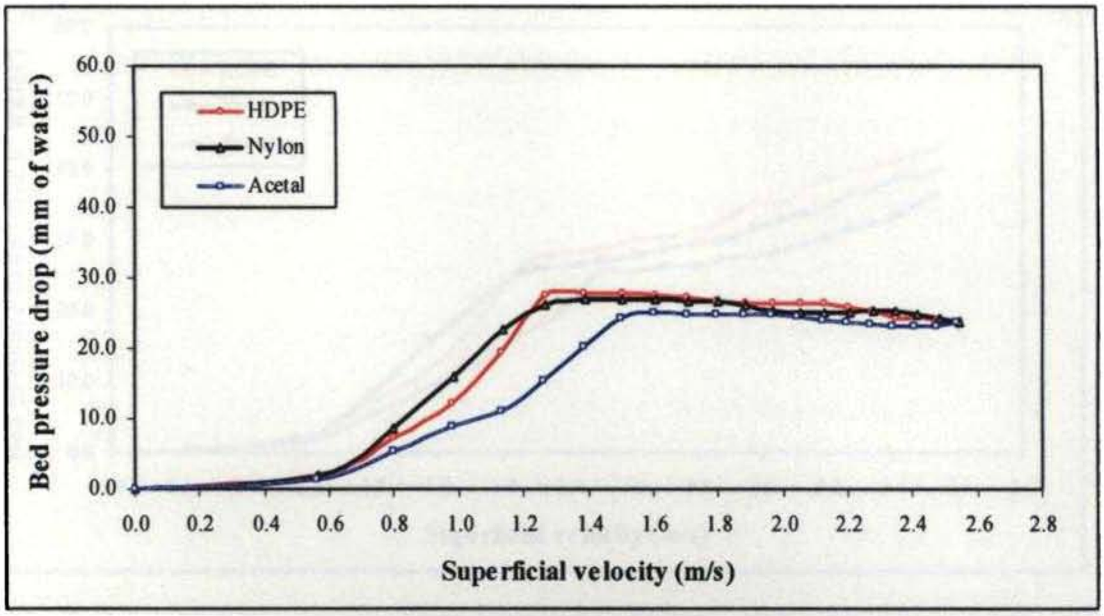


Figure 5.6.6 Variation of bed pressure drop with superficial velocity for different density materials of 5.5 mm diameter, at 90 mm from the centre of the distributor (Φ= 15°)

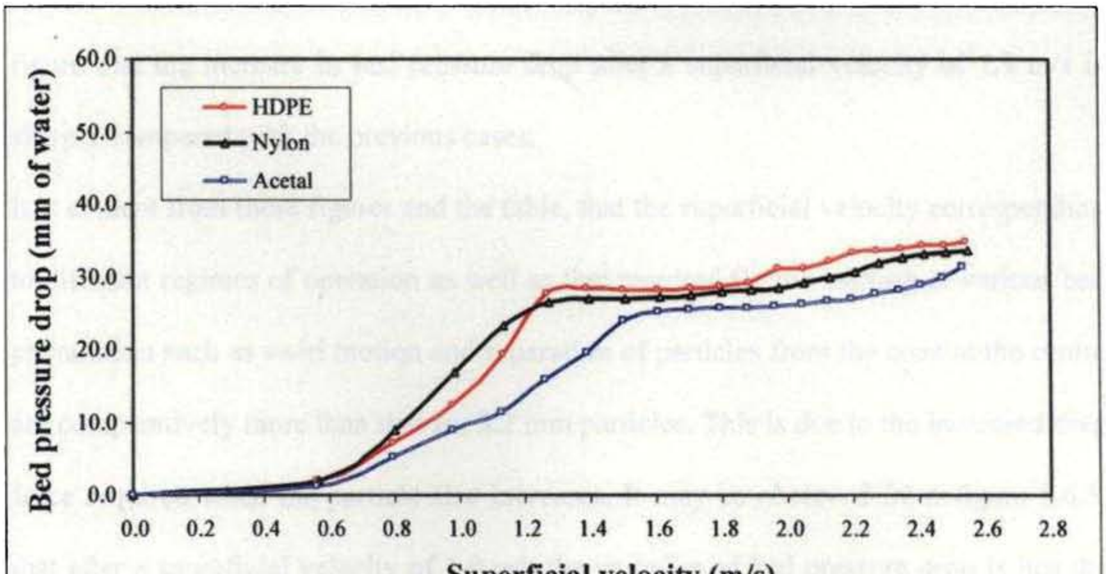


Figure 5.6.7 Variation of bed pressure drop with superficial velocity for different density materials of 5.5 mm diameter, at 120 mm from the centre of the distributor (Φ= 15°)

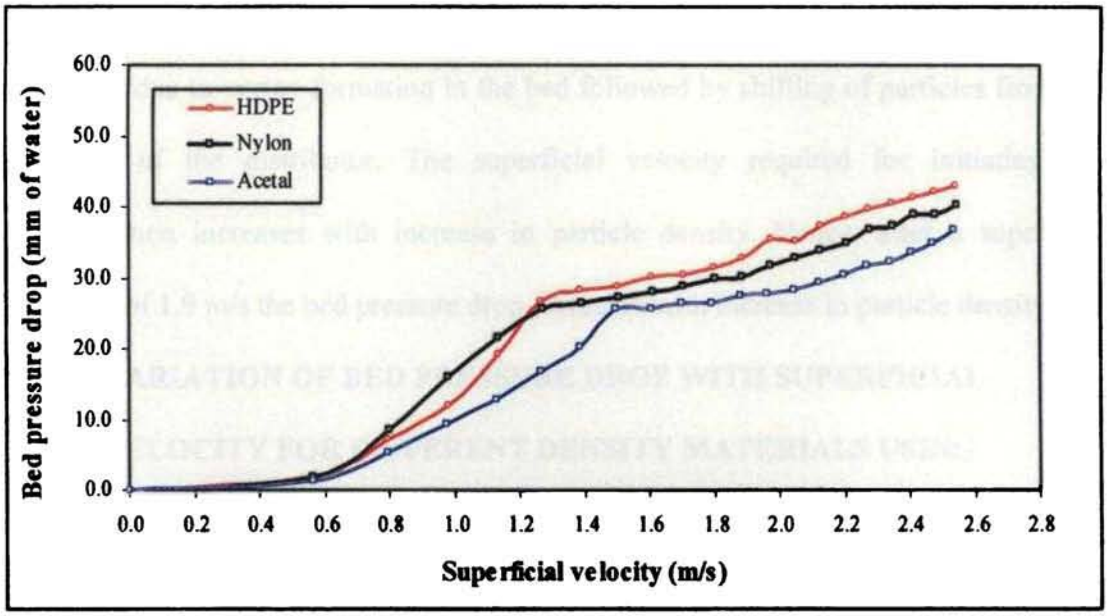


Figure 5.6.8 Variation of bed pressure drop with superficial velocity for different density materials of 5.5 mm diameter, at 150 mm from the centre of the distributor (Φ= 15°)

Figure 5.6.8 reveals the variation of bed pressure drop with superficial velocity at a radial position of 150 mm from the distributor centre. It can be observed from this figure that the increase in bed pressure drop after a superficial velocity of 1.9 m/s is sharper compared with the previous cases.

It is evident from these figures and the table, that the superficial velocity corresponding to different regimes of operation as well as that required for the starting of various bed phenomena such as swirl motion and separation of particles from the cone at the centre are comparatively more than that for 3.2 mm particles. This is due to the increased drag force required when the particle size increases. It may be observed from figure 5.6.5, that after a superficial velocity of 1.9 m/s the variation of bed pressure drop is just the opposite to that before, ie. the bed pressure drop increases with increase in particle density. This region falls after the vortex formation in the bed. It may also be noticed that this position of the bed corresponds to the inner periphery of the distributor. The decrease in bed pressure drop occur in this portion of the bed after a superficial velocity 1.97 m/s due to vortex formation in the bed followed by shifting of particles from this position of the distributor. The superficial velocity required for initiating this phenomenon increases with increase in particle density. Hence, after a superficial velocity of 1.9 m/s the bed pressure drop increases with increase in particle density.

5.6.3 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT DENSITY MATERIALS USING 7.4 MM DIAMETER SIZE PARTICLES AS THE BED MATERIAL

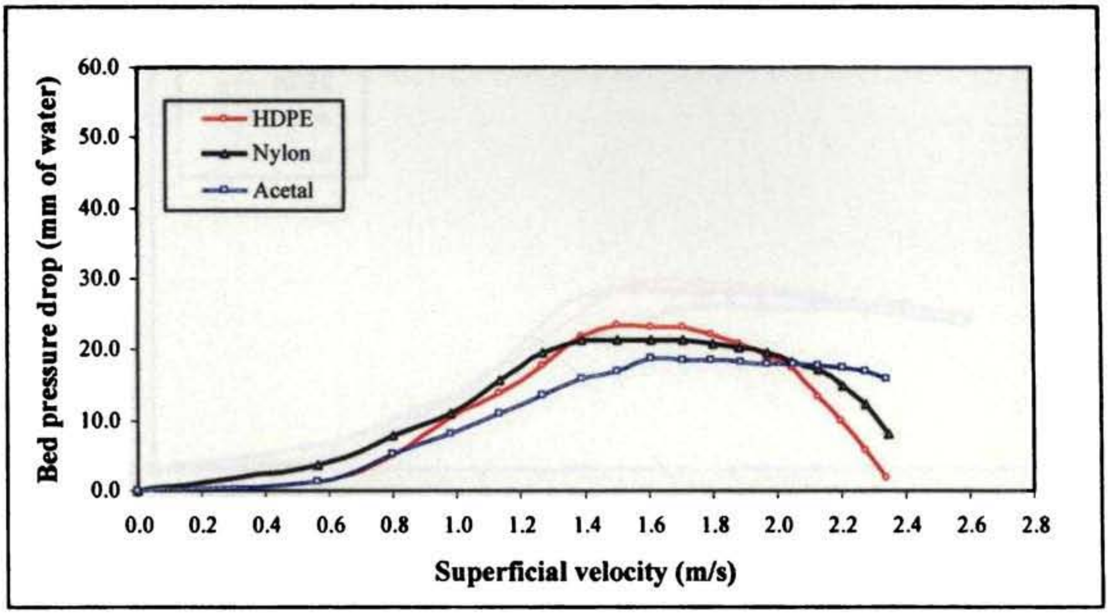


Figure 5.6.9 Variation of bed pressure drop with superficial velocity for different density materials of 7.4 mm diameter, at 60 mm from the centre of the distributor ($\Phi = 15^o$)

Figures from 5.6.9 to 5.6.12 reveal the variation of bed pressure drop with superficial velocity for different density materials using 7.4 mm particles at radial positions of 60 mm, 90 mm, 120 mm and 150 mm from the distributor centre. The various bed phenomena such as swirl motion and separation of particles from the cone at the centre for these particles are given in table 5.6.3. A phenomenon similar to that of 3.2 mm particles and 5.5 mm particles can be observed in the case of 7.4 mm particles also. Figure 5.6.9 corresponds to the variation of bed pressure drop with superficial velocity at a radial position of 60 mm from the distributor centre. It can be seen from this figure that the bed pressure drop increases linearly upto a superficial velocity of 1.4 m/s, after which it decreases slowly till a superficial velocity 2 m/s is reached, beyond which it decreases suddenly. It is evident from figure 5.6.10, which depicts the variation of bed pressure drop with superficial velocity at a radial position of 90 mm from the

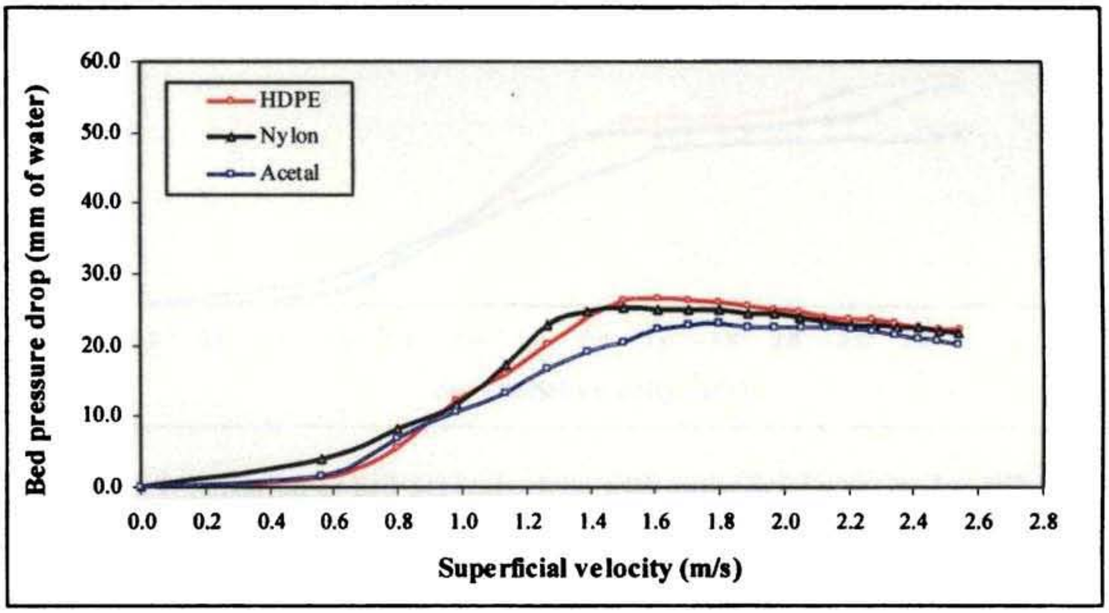


Figure 5.6.10 Variation of bed pressure drop with superficial velocity for different density materials of 7.4 mm diameter, at 90 mm from the centre of the distributor (Φ= 15°)

distributor centre, that after the linear variation of bed pressure drop upto 1.4 m/s of superficial velocity the bed pressure drop decreases gradually. The variation of bed pressure drop with superficial velocity at a radial position of 120 mm from the distributor centre is given in figure5.6.11. It may be noticed from this figure that after almost linear variation of bed pressure drop upto a superficial velocity of 1.4 m/s, it increases gradually. Figure 5.6.12 reveals the variation of bed pressure drop with superficial velocity at a radial position of 150 mm from the distributor centre. It is evident from this figure that after the linear variation of bed pressure drop upto a superficial velocity 1.4 m/s, the increase in bed pressure drop is sudden.

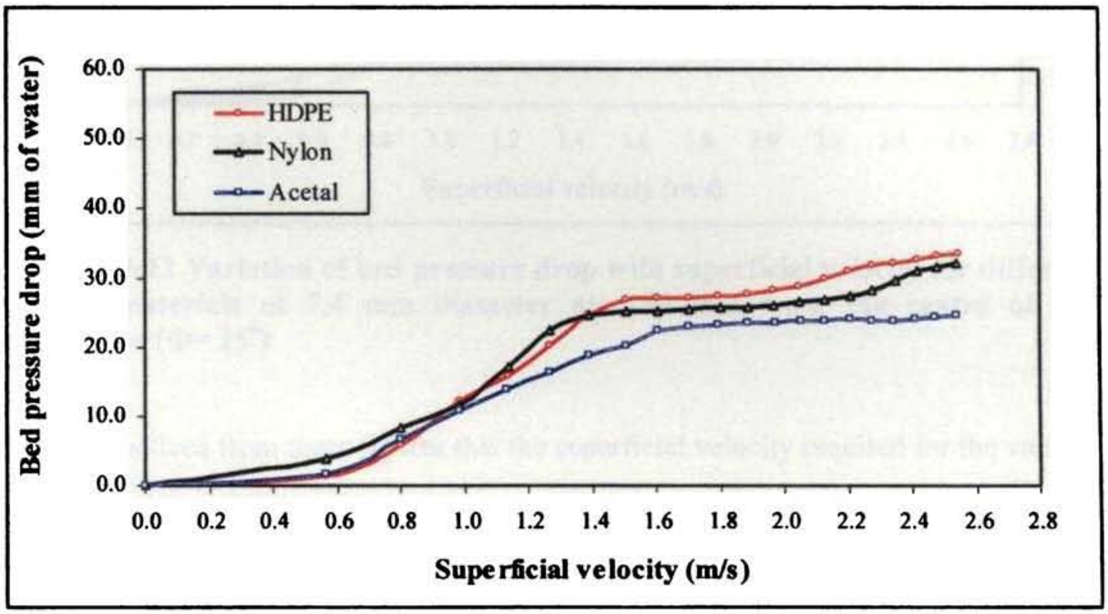


Figure 5.6.11 Variation of bed pressure drop with superficial velocity for different density materials of 7.4 mm diameter, at 120 mm from the centre of the distributor (Φ= 15°)

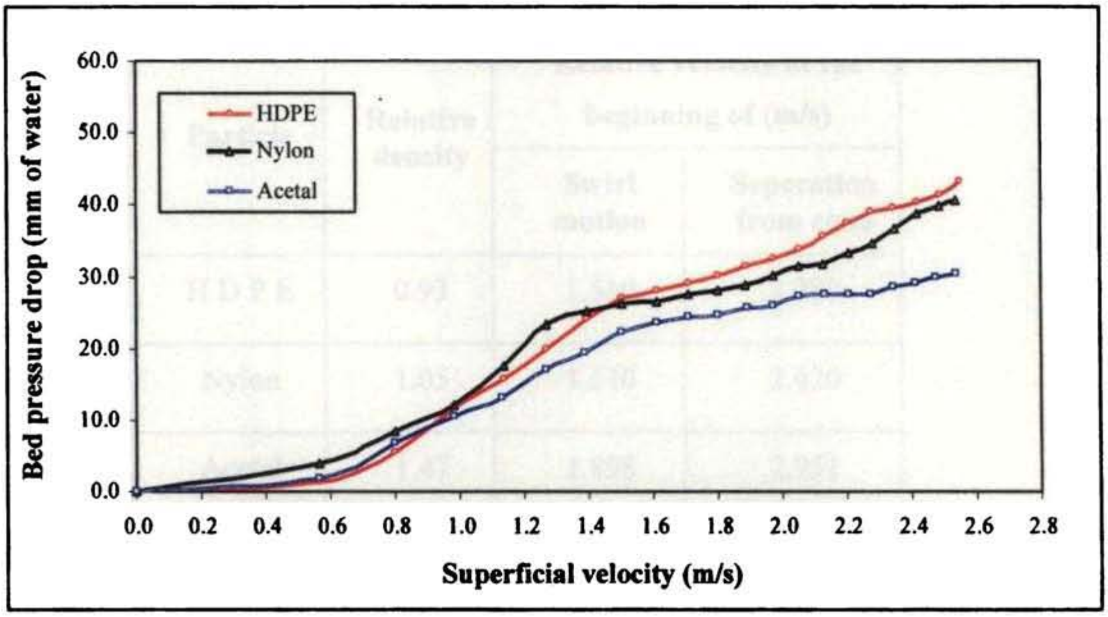


Figure 5.6.12 Variation of bed pressure drop with superficial velocity for different density materials of 7.4 mm diameter at, 150 mm from the centre of the distributor (Φ= 15°)

It can be noticed from these figures that the superficial velocity required for the various regimes is comparatively higher than that required for 3.2 mm particles and 5.5 mm particles. It may be further observed from table 5.6.3, that the superficial velocity required for initiating various bed phenomena is relatively higher than that required for 3.2 mm particles and 5.5 mm particles. The reason for this phenomenon is already explained in section 5.6.2.It may also be noticed from figure5.6.9 that, similar to that of 5.5 mm particles, after a superficial velocity of 2 m/s the variation of bed pressure drop is just the opposite to that before. The reason for this change in pattern of behaviour is already explained in section 5.6.2.

The variation of bed phenomena for particles having different sizes is given in table 5.6.4. It is quite evident from this table that the variation of bed phenomena such as the rate of change of superficial velocity / unit density required for swirl motion and

Table 5.6.3 Bed phenomena of 7.4 mm particles

Particle	Relative density	Relative velocity at the beginning of (m/s)	
		Swirl motion	Seperation from cone
H D P E	0.93	1.510	2.280
Nylon	1.05	1.510	2.420
Acetal	1.47	1.898	2.951

separation of particles from the cone at the centre increases with particle size. On the basis of experimental results and observations made while conducting the experiments the following conclusions can be made. The bed pressure drop increases with decrease in particle density. Further the superficial velocity required for initiating various bed phenomena such as swirl motion and separation of particles from the cone at the centre increases with increase in particle density. The superficial velocity required for various regimes such as linear variation of bed pressure drop, constant bed pressure drop and sudden increase or decrease in bed pressure drop increases with increase in particle density. The rate of change of superficial velocity per unit size required for required for various bed phenomena increases with increase in particle size.

Table 5.6.4 Variation of bed phenomena for the different particles

Range of particle density	Rate of change of superficial velocity/unit size required for (m/s/mm)					
	3.2 mm particles		5.5 mm particles		7.4 mm particles	
	Swirl motion	Seperation from cone	Swirl motion	Seperation from cone	Swirl motion	Seperation from cone
0.93 – 1.05	0.167	0.333	0.250	1.250	0.333	1.167
1.05 – 1.47	0.376	0.560	0.707	0.590	0.924	1.264
Mean	0.272	0.440	0.479	0.920	0.629	1.216

5.7 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT ANGLES OF AIR INJECTION (ϕ)

From the experiments conducted on the variation of distributor pressure drop with superficial velocity it was found that the distributor pressure drop decreases with increase in angle of air injection (ϕ). However no experimental data is available on the on the effect of angle of air injection on the variation of bed pressure drop with superficial velocity in swirling fluidized bed. Hence an attempt is made to determine the variation of bed pressure drop with superficial velocity for different angles of air injection. Experiments were conducted with distributors having angle of air injection (ϕ) 0^0, 5^0,10^0 & 15^0. 2.5 kg of acetal (relative density 1.47) having particle sizes 3.2 mm, 5.5 mm & 7.4 mm were used as the bed materials. Measurements of pressure drop across the bed were made by digital micromanometer at radial positions of 60 mm, 90 mm, 120 mm & 150 mm from the distributor centre.

5.7.1 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT ANGLES OF AIR INJECTION (ϕ) USING 3.2 MM PARTICLES

The variation of bed pressure drop with superficial velocity for different angles of air injection (ϕ) for 3.2 mm particles at radial positions of 60 mm, 90 mm, 120 mm & 150 mm from the distributor centre is depicted in figure 5.7.1 and figures 5.7.2, 5.7.3 & 5.7.4 respectively (Appendix –V). It is evident from these figures that for the same superficial velocity the bed pressure drop decreases with increase in angle of air injection. The reason for this is the same as that for distributor pressure drop variation which is explained in section 5.2.

While conducting the experiments observations were made regarding the superficial velocity required to initiate various bed phenomena such as swirl motion and separation of particles from the cone at the centre. These observations are given in table 5.7.1. It can be noticed that the superficial velocity required for initiating these bed phenomena increases with increase in angle of air injection. The tangential component of velocity v_t which is given by $v \cos \phi$ is responsible for initiating these various bed phenomena. With increase in angle of air injection (ϕ), the tangential component decreases. Hence the superficial velocity required for initiating these parameters increases with increase in angle of air injection.

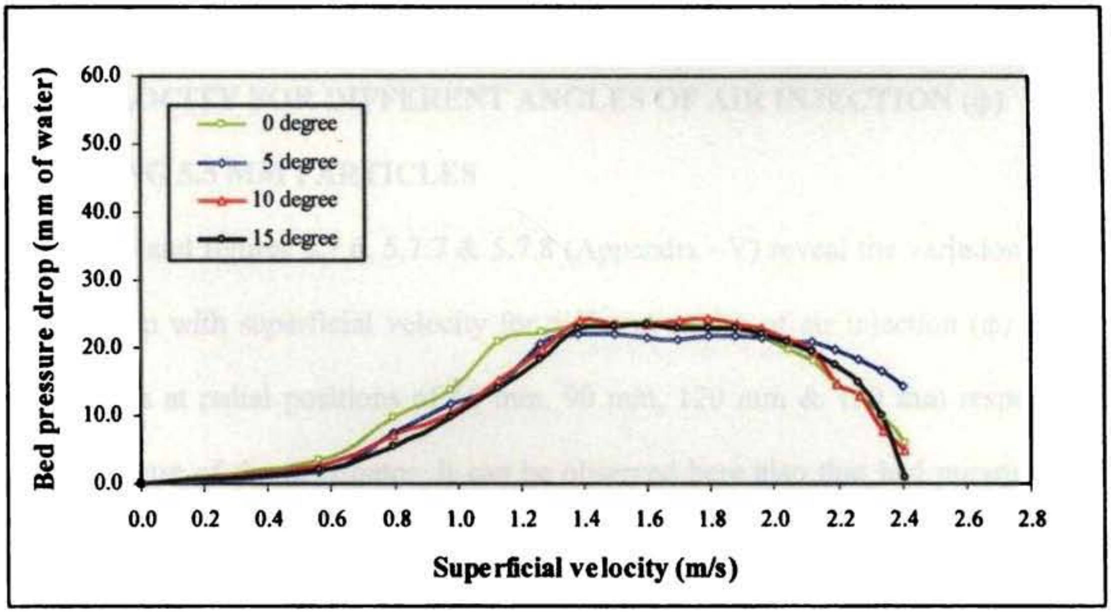


Figure 5.7.1 Variation of bed pressure drop with superficial velocity for different angles of air injection using 3.2 mm diameter acetal, at 60 mm from the distributor centre

Table 5.7.1 Bed phenomena for 3.2 mm acetal particles for different angles of air injection

Angle of air injection (ϕ) degree	Superficial velocity at the beginning of (m/s)	
	Swirl motion	Seperation from cone
0	1.386	2.71
5	1.441	2.306
10	1.495	2.341
15	1.548	2.405

5.7.2 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT ANGLES OF AIR INJECTION (ϕ) USING 5.5 MM PARTICLES

Figure 5.7.5 and figures 5.7.6, 5.7.7 & 5.7.8 (Appendix –V) reveal the variation of bed pressure drop with superficial velocity for different angles of air injection (ϕ) for 5.5 mm particles at radial positions of 60 mm, 90 mm, 120 mm & 150 mm respectively from the centre of the distributor. It can be observed here also that bed pressure drop decreases with increase in angle of air injection. The reason for this is already explained in section 5.7.1.

The various bed phenomena for these particles are given in table 5.7.2. Here also it may be noticed that superficial velocity required for initiating various bed phenomena increases with increase in angle of air injection. The reason for this is also explained in section 5.7.1.

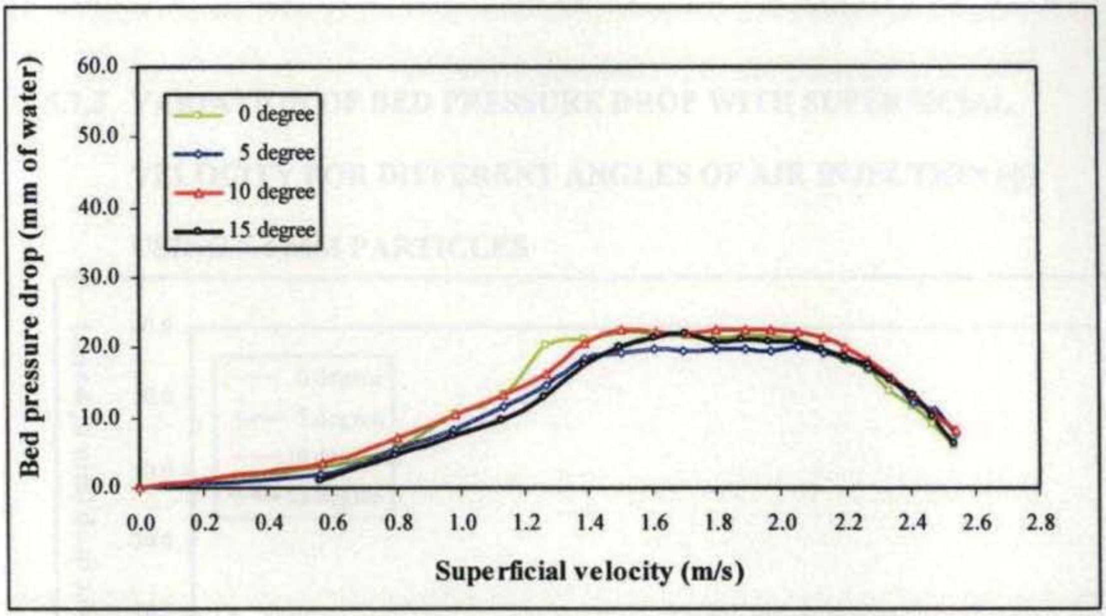


Figure 5.7.5 Variation of bed pressure drop with superficial velocity for different angles of air injection using 5.5mm diameter acetal, at 60 mm from the distributor centre

Table 5.7.2 Bed phenomena for 5.5 mm acetal particles for different angles of air injection

Angle of air injection (ϕ) degree	Superficial velocity at the beginning of (m/s)	
	Swirl motion	Seperation from cone
0	1.502	2.405
5	1.606	2.539
10	1.698	2.571
15	1.706	2.598

From table 5.7.2 it can also be observed that the superficial velocity required for initiating various bed phenomena are relatively more than that for 3.2 mm particles.

5.7.3 VARIATION OF BED PRESSURE DROP WITH SUPERFICIAL VELOCITY FOR DIFFERENT ANGLES OF AIR INJECTION (ϕ) USING 7.4 MM PARTICLES

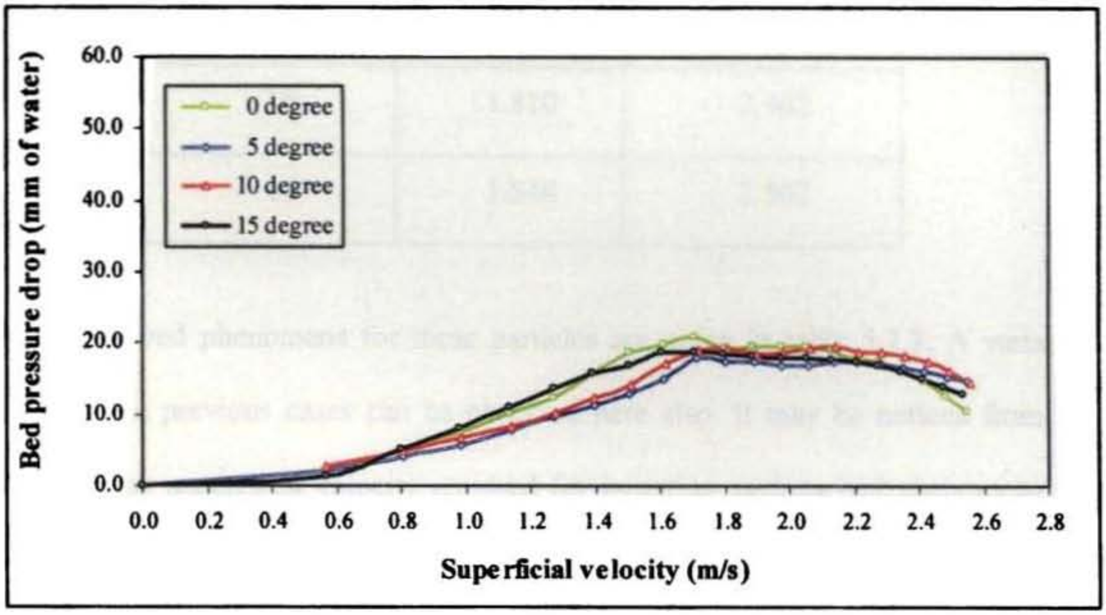


Figure 5.7.9 Variation of bed pressure drop with superficial velocity for different angles of air injection using 7.4 mm diameter acetal, at 60 mm from the distributor centre

The variation of bed pressure drop with superficial velocity for different angles of air injection (ϕ) for 7.4 mm particles at radial positions of 60 mm, 90 mm, 120 mm & 150 mm from the distributor centre is depicted in figure 5.7.9 and figures 5.7.10, 5.7.11 & 5.7.12 respectively (Appendix – V).

Table 5.7.3 Bed phenomena for 7.4 mm acetal particles for different angles of air injection

Angle of air injection (ϕ) degree	Superficial velocity at the beginning of (m/s)	
	Swirl motion	Seperation from cone
0	1.575	2.348
5	1.720	2.421
10	1.810	2.462
15	1.846	2.502

The various bed phenomena for these particles are given in table 5.7.3. A variation similar to the previous cases can be observed here also. It may be noticed from the table, that the superficial velocity required for initiating various bed phenomena are relatively higher than that for 3.2 mm & 5.5 mm particles. The variation of bed phenomena for different angles of air injection is given in table 5.7.4.

In the table the variations of acetal particles of different sizes are given. It is evident from the table that the rate of change of superficial velocity per angle of air injection required for various bed phenomena such as swirl motion and separation from the cone decreases with increase in angle of air injection.

Table 5.7.4 Variation of bed phenomena for different angles of air injection

Particle size mm	Range of angle of air injection (ϕ) degree	Rate of change of superficial velocity/angle of air injection required for (m/s/degree)	
		Swirl motion	Seperation From cone
3.2	0-5	0.0110	0.0192
	5-10	0.0108	0.0070
	10-15	0.0106	0.0128
5.5	0-5	0.0208	0.0268
	5-10	0.0184	0.0064
	10-15	0.0016	0.0054
7.4	0-5	0.0290	0.0146
	5-10	0.0180	0.0082
	10-15	0.0072	0.008

Based on the experiments conducted it can be concluded that the bed pressure drop decreases with increase in angle of air injection (ϕ). However the superficial velocity required for initiating various bed phenomena such as swirl motion and separation of particles from the cone at the centre increases with increase in angle of air injection (ϕ). The various regimes such as linear variation of bed pressure drop, constant bed pressure drop, sudden increase or decrease in bed pressure drop also increases with increase in angle of air injection (ϕ). However the rate of change of superficial velocity/angle of air injection required for various bed phenomena decreases with increase in angle of air injection (ϕ).

5.8 CONCLUSIONS

Based on the hydrodynamic studies conducted and the above discussions, the following conclusions are made.

5.8.1 Distributor pressure drop

1. Distributor pressure drop decreases with increase in angle of air injection (ϕ).

5.8.2 Minimum fluidizing velocity

1. Minimum fluidizing velocity increases with increase in angle of air injection (ϕ).
2. Minimum fluidizing velocity increases with increase in particle size.
3. Minimum fluidizing velocity increases with increase in particle density.

5.8.3 Bed pressure drop

1. After attaining minimum fluidizing velocity, the bed pressure drop increases along the radial direction of the distributor.
2. Bed pressure drop increases with decrease in particle size for constant bed weight.
3. Bed pressure drop increases with decrease in particle density.
4. After attaining minimum fluidizing velocity, the bed pressure drop is almost constant at a radial position of 90 mm from the centre of the distributor.

5.8.4 Bed Behaviour

1. The superficial velocity required for initiating various bed phenomena such as swirl motion & separation of particles from the cone at the centre increases with increase in particle size.

2. The superficial velocity required for initiating these bed phenomena increases with increase in particle density.
3. The superficial velocity required for various regimes such as linear variation of bed pressure drop, constant bed pressure drop and sudden increase or decrease in bed drop increases with increase in particle size.
4. The superficial velocity required for these regimes increases with increase in particle density.
5. The rate of change of superficial velocity/unit size required for initiating various bed phenomena such as swirl motion & separation of particles from the cone at the centre increases with increase in particle density.
6. The rate of change of superficial velocity/unit density required for initiating these bed phenomena increases with increase in particle size.

5.8.5 Angle of air injection (ϕ)

1. The bed pressure drop decreases with increase in angle of air injection for the same bed weight.
2. The superficial velocity required for initiating various bed phenomena increases with increase in angle of air injection.
3. The superficial velocity required for various regimes such as linear variation of bed pressure drop, constant bed pressure drop and sudden increase or decrease in bed drop increases with increase in angle of air injection.
4. The rate of change of superficial velocity/ angle of air injection required for initiating various bed phenomena decreases with increase in angle of air injection (ϕ).

CHAPTER 6

CONCLUSIONS

Conidour type distributors having angle of air injection (ϕ) - 0^0, 5^0, 10^0 &15^0 were designed and fabricated. The percentage area of opening and the percentage useful area of opening was the same for all these distributors.

Spherical particles having sizes 3.2 mm, 5.5 mm & 7.4 mm (Geldart D type) were selected for the hydrodynamic study. Particles having these sizes were moulded from high density polyethylene, nylon and acetal.

Basic hydrodynamic studies were conducted with these distributors and using the various bed materials. The major variables considered for the present study include particle size, particle density and angle of air injection.

This chapter reveals the major conclusions obtained from the present investigation.

6.1 CONCLUSIONS

1. The distributor pressure drop decreases with increase in angle of air injection.
2. The pressure head above the distributor increases with increase in angle of air injection.
3. Since the area of opening is independent of angle of air injection, the pressure below the distributor is independent of angle of air injection.
4. The hydrodynamic characteristics of the swirling fluidized bed are quite different from that of a conventional bed.
5. The sequence of bed phenomena in conidour type swirling fluidized bed are packed bed, fluidized bed, swirl motion and vortex motion.

6. It was observed that the minimum fluidizing velocity increases with increase in angle of air injection.
7. In a swirling fluidized bed it was observed that the particle size as well as particle density directly influence the minimum fluidizing velocity.
8. It could be observed that, after attaining minimum fluidizing velocity the bed pressure drop increases along the radial direction.
9. The sequence of variation of bed pressure drop with superficial velocity are linear increase, no change followed by sudden increase or sudden decrease.
10. At a radial position of 90 mm from the distributor centre, the bed height remains unchanged irrespective of superficial velocity. Hence after attaining minimum fluidizing velocity, the bed pressure drop at this location of the distributor remains almost constant.
11. Vortex formation in the bed leads to increase in wall friction.
12. The frictional resistance between gas- solid interface increase with decrease in particle size. As a result of this the bed pressure drop increases with decrease in particle size.
13. For any constant weight of the bed the height of the bed increases with decrease in particle density. So the bed pressure drop increases with decrease in particle density.
14. The fluid drag exerted on the materials decreases with increase in particle size. Hence the superficial velocity required for initiating various bed phenomena increase with increase in particle size.

15. Since the drag resistance offered by the materials increases with increase in particle density, the superficial velocity required for initiating various bed phenomena increase with increase in particle density.
16. The superficial velocity required for various regimes of operation increases with increase in particle size as well as particle density.
17. The radial component of velocity increases with angle of air injection, whereas the tangential component decreases. Since the tangential component is responsible for swirl motion, the superficial velocity required for initiating various bed phenomena increases with angle of air injection.
18. The bed pressure drop is found to decrease with increase in angle of air injection for any constant weight of the bed.

6.2 SUGGESTIONS FOR FUTURE WORK

The distributors adopted for the present study were designed to have the same angle of air injection from the slits in all the four rows. It is expected that by increasing the angle of air injection progressively from the inner rows towards the outer, the vortex formation in the bed can be delayed without affecting swirl motion. An investigation can be conducted in this aspect.

Every experiment was conducted using a particular size of the particle made from the same material. A detailed hydrodynamic study may be conducted by intermixing these particles of different sizes and relative densities in varying predetermined proportion.

An investigation may be conducted to determine the effect of angle of emergence of the jet with the horizontal (θ) in combination with angle of air injection (ϕ).

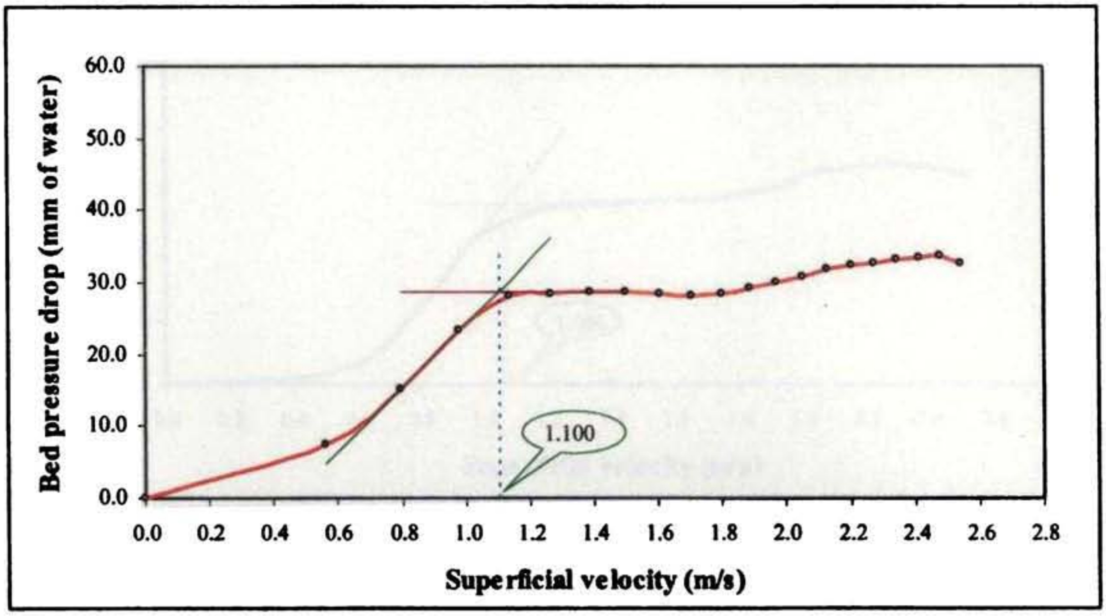


Figure 5.3.2 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 5.5 mm HDPE as bed material.

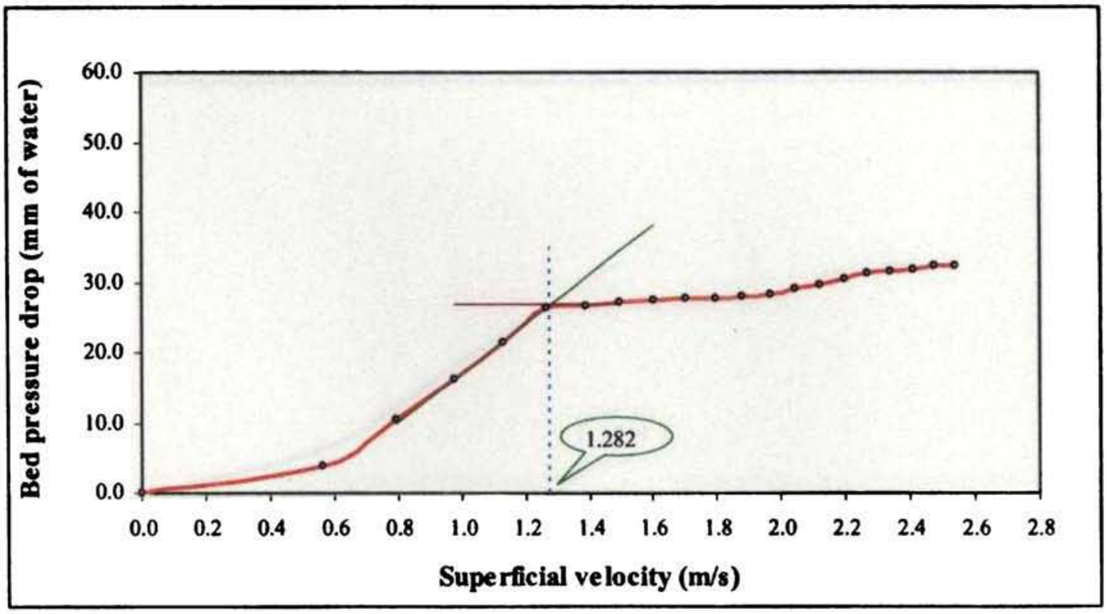


Figure 5.3.3 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 7.4 mm HDPE as bed material.

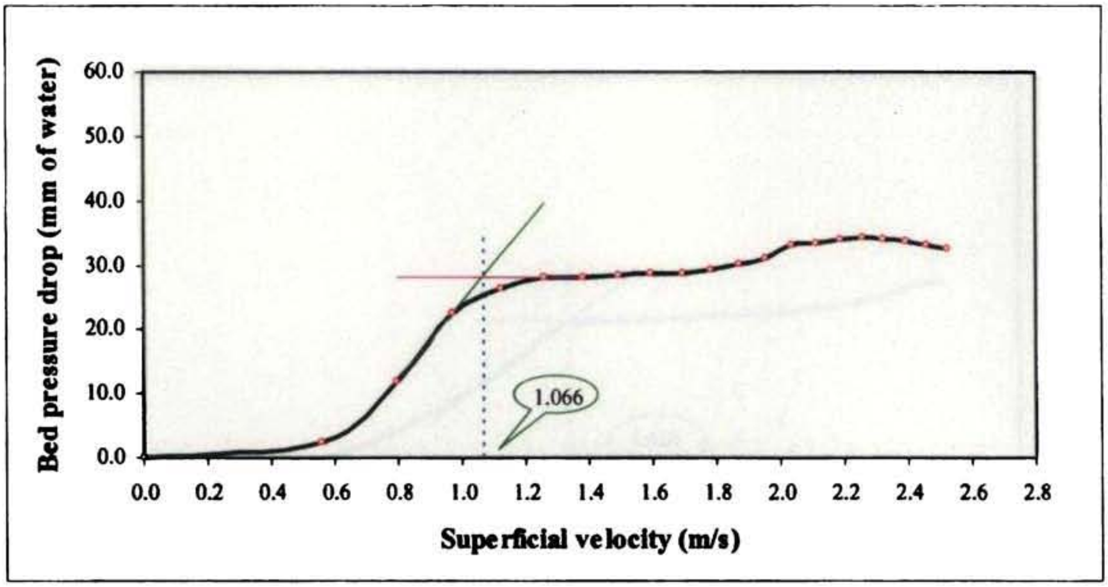


Figure 5.3.4 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 3.2 mm nylon as bed material.

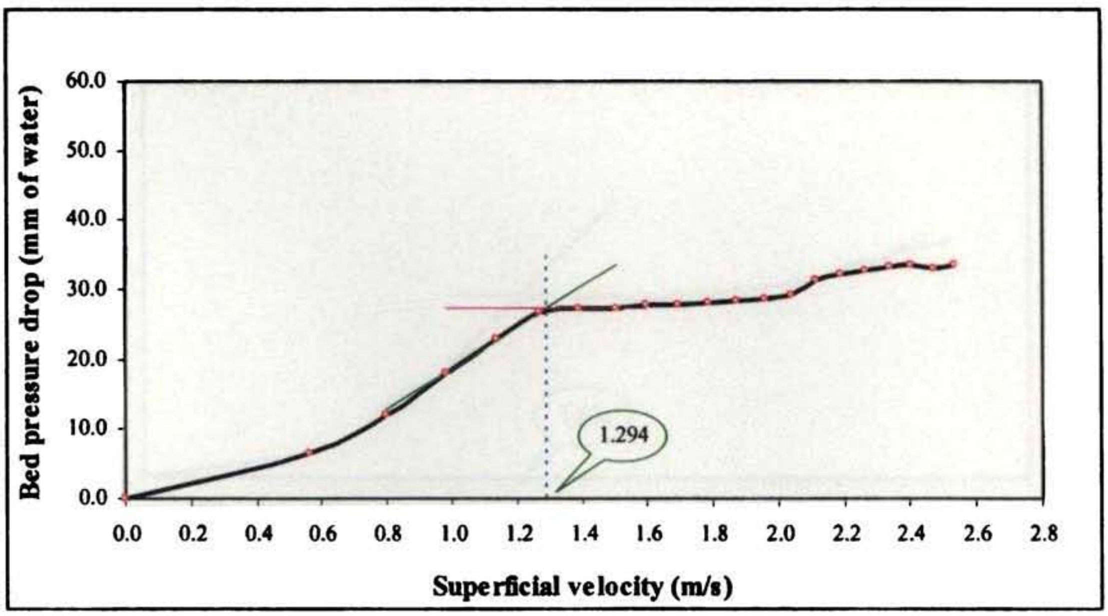


Figure 5.3.5 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 5.5 mm nylon as bed material.

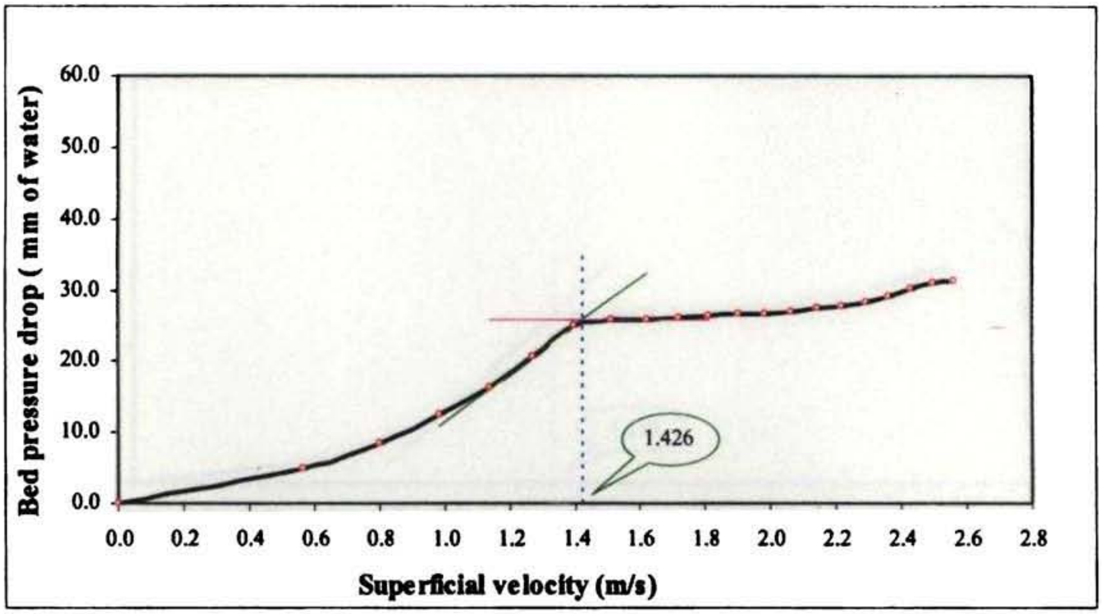


Figure 5.3.6 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 7.4 mm nylon as bed material.

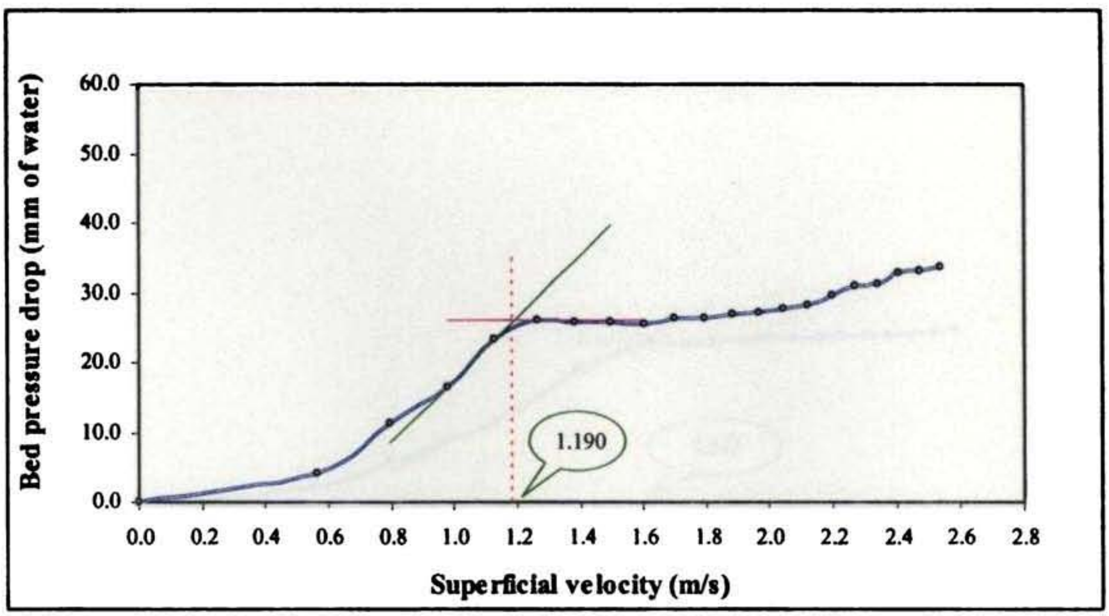


Figure 5.3.7 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 3.2 mm acetal as bed material.

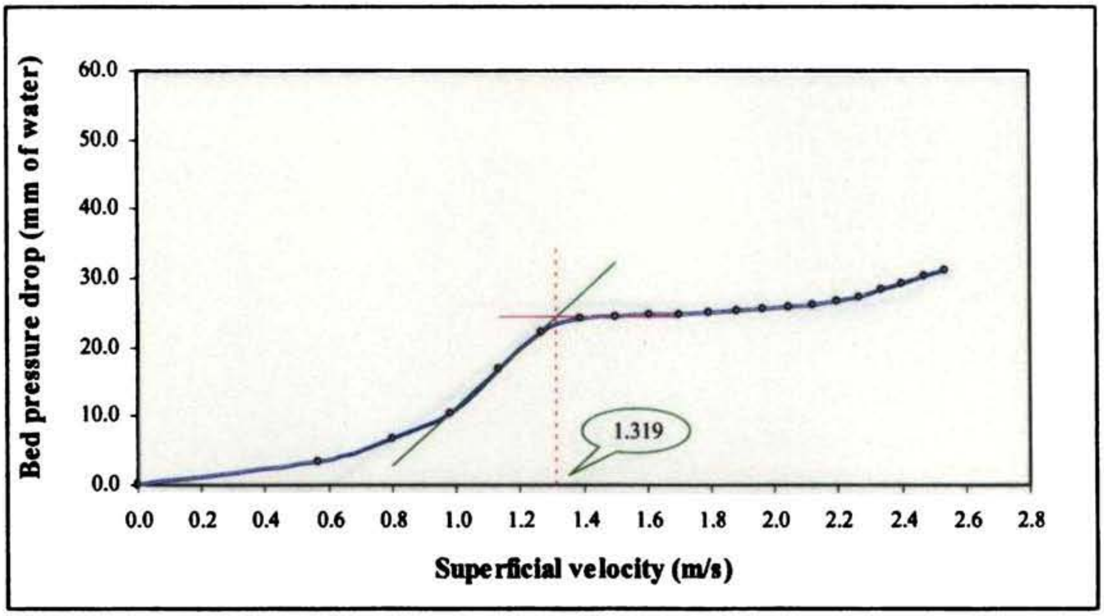


Figure 5.3.8 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 5.5 mm acetal as bed material.

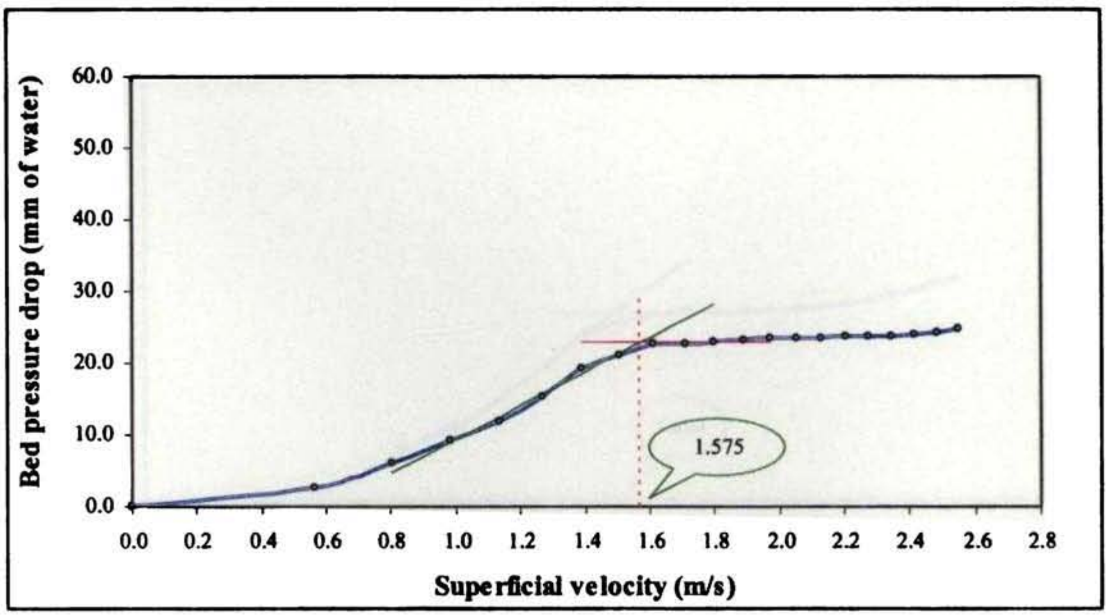


Figure 5.3.9 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 0^0$ and using 7.4 mm acetal as bed material.

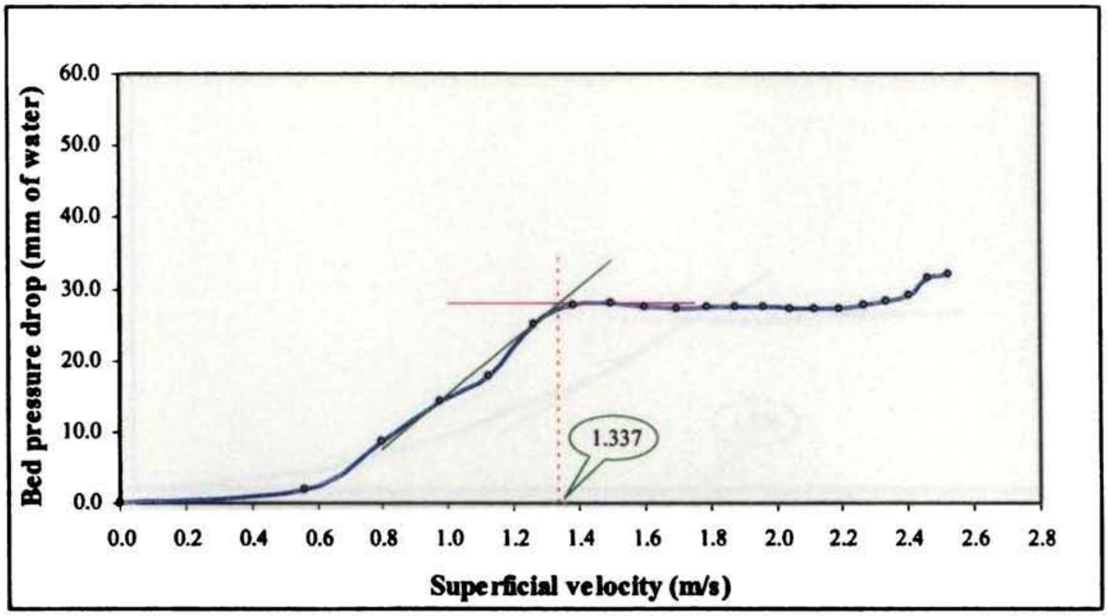


Figure 5.3.10 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 5^0$ and using 3.2 mm acetal as bed material.

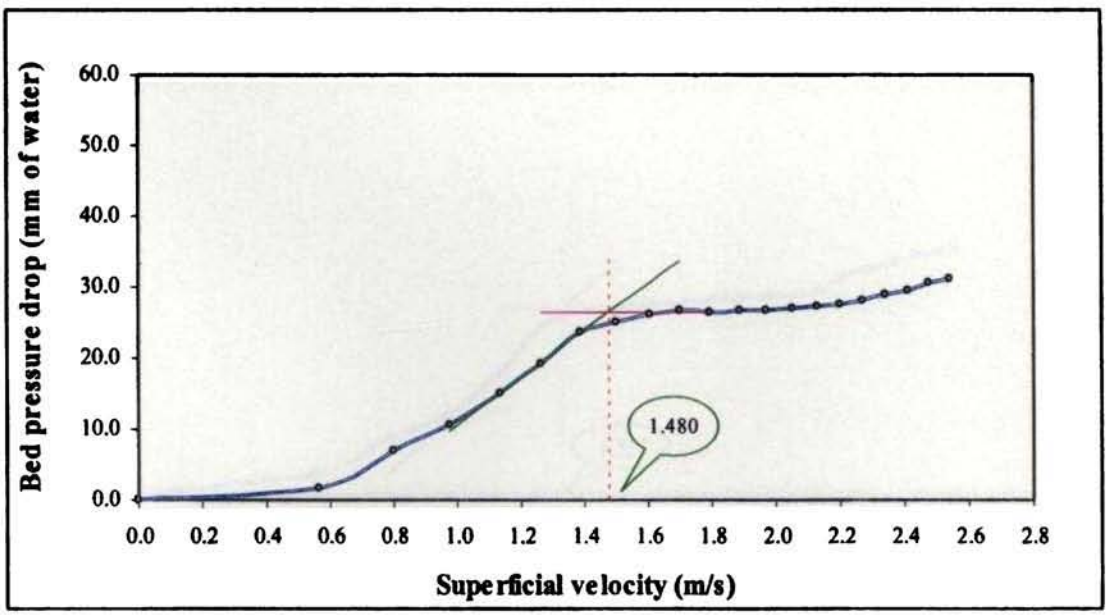


Figure 5.3.11 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 5^0$ and using 5.5 mm acetal as bed material.

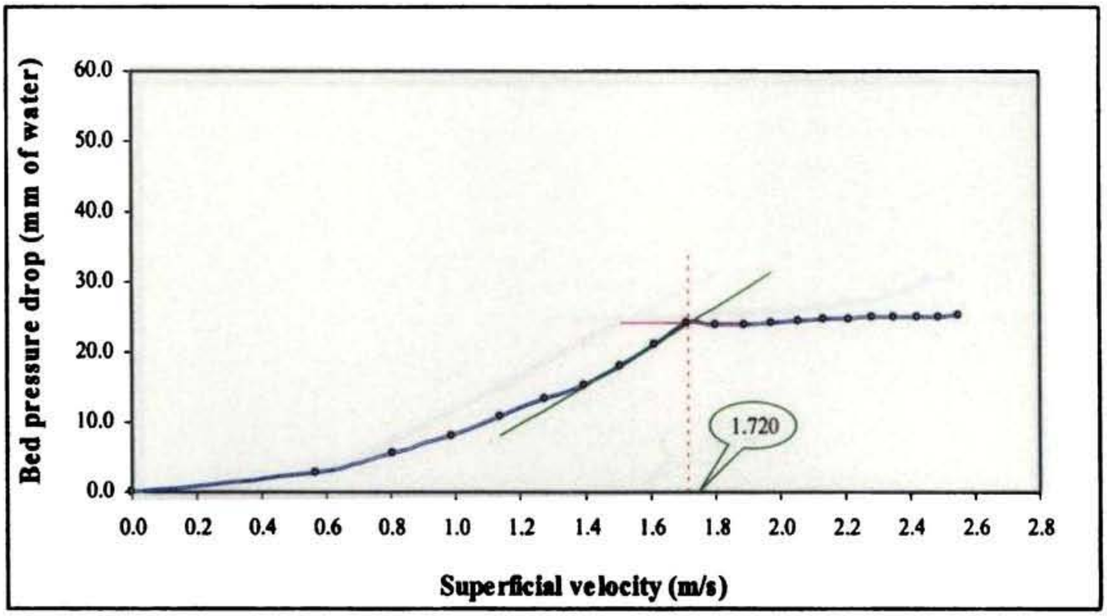


Figure 5.3.12 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 5^0$ and using 7.4 mm acetal as bed material.

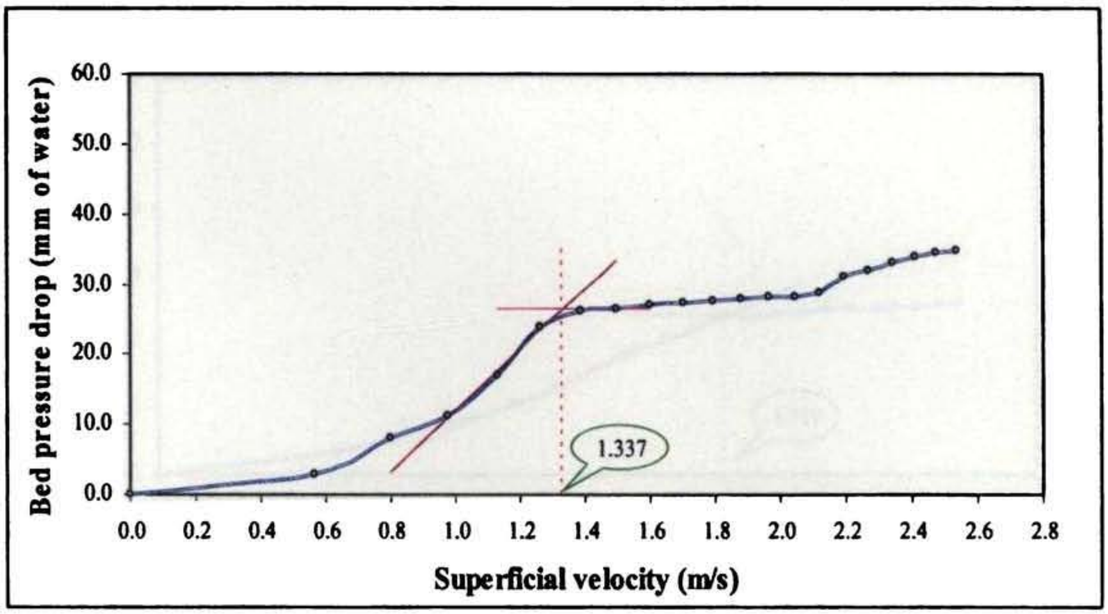


Figure 5.3.13 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 10^0$ and using 3.2 mm acetal as bed material.

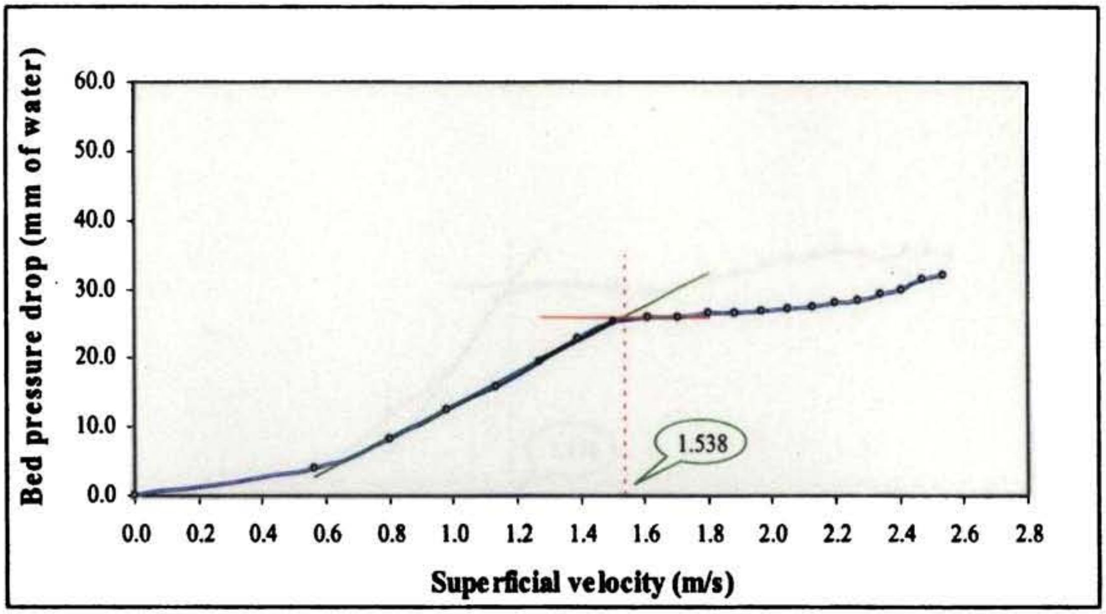


Figure 5.3.14 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 10^0$ and using 5.5 mm acetal as bed material.

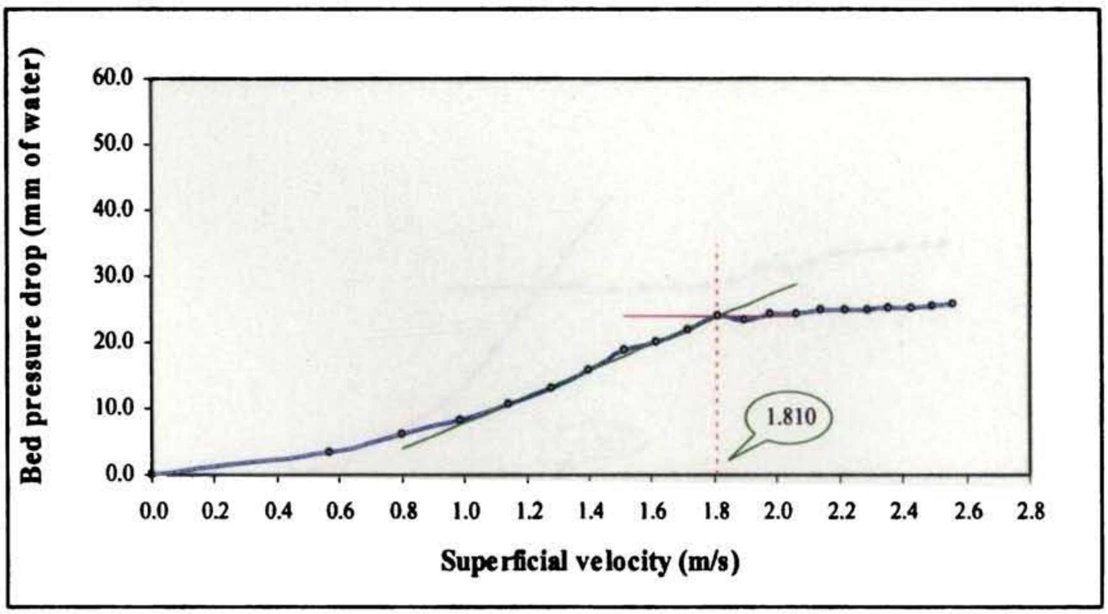


Figure 5.3.15 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 10^0$ and using 7.4 mm acetal as bed material.

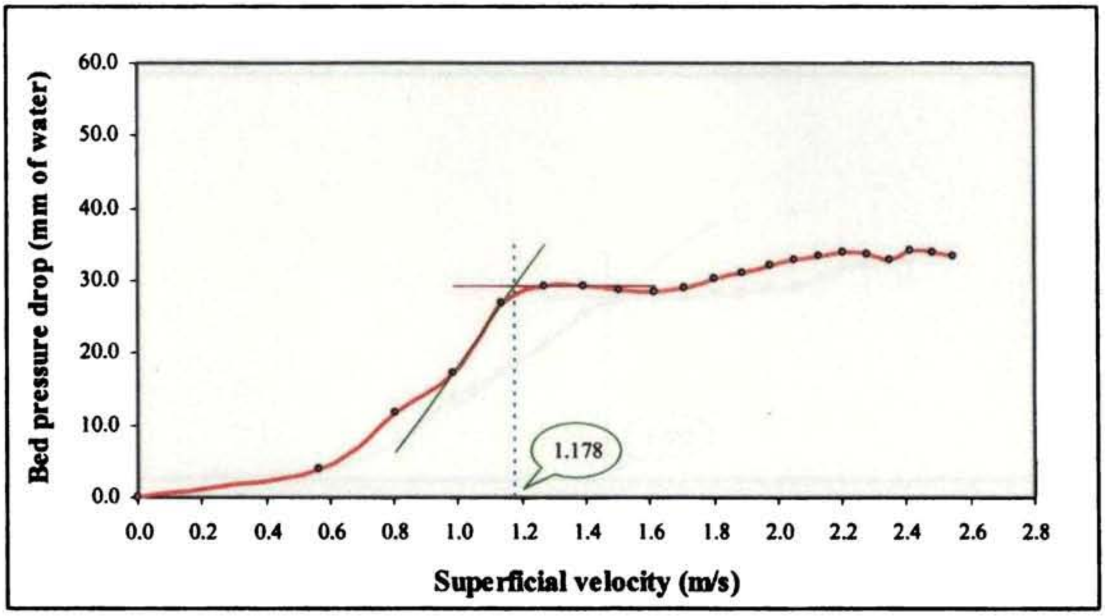


Figure 5.3.16 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 3.2 mm HDPE as bed material.

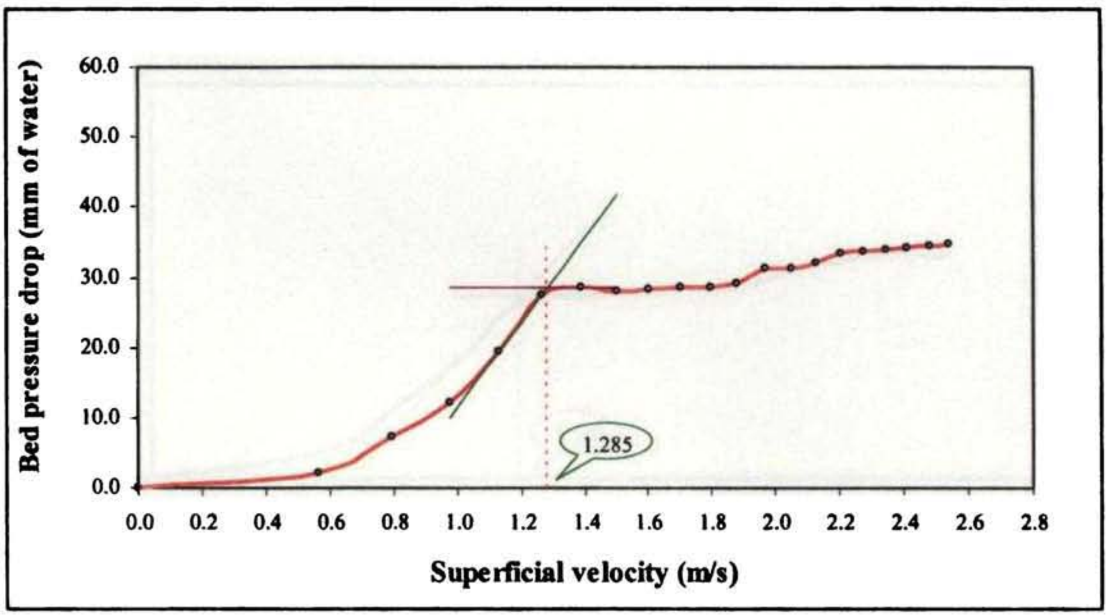


Figure 5.3.17 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 5.5 mm HDPE as bed material.

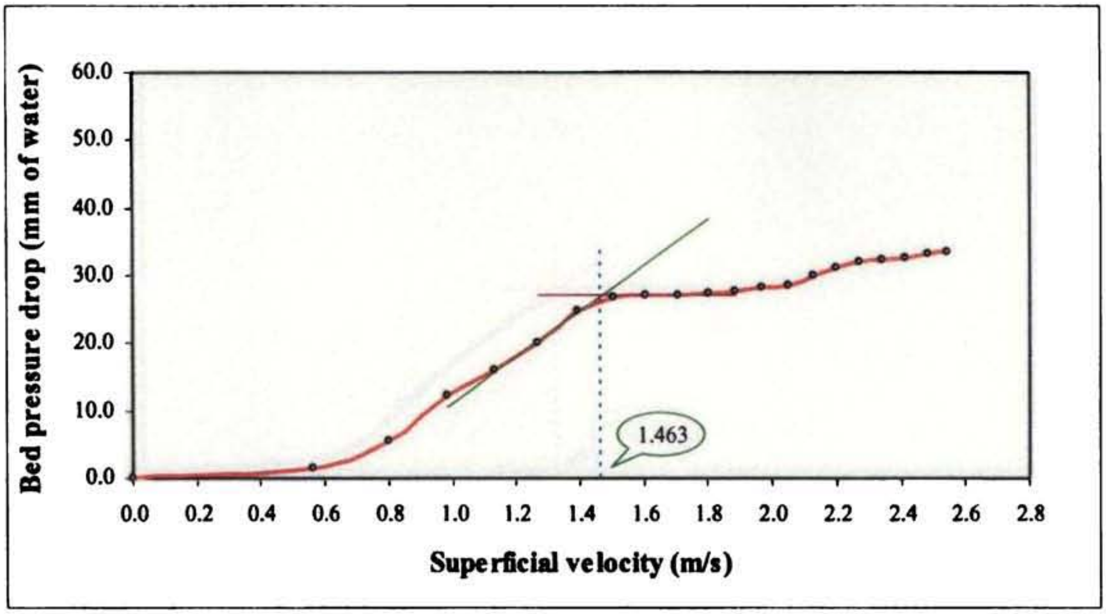


Figure 5.3.18 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 7.4 mm HDPE as bed material.

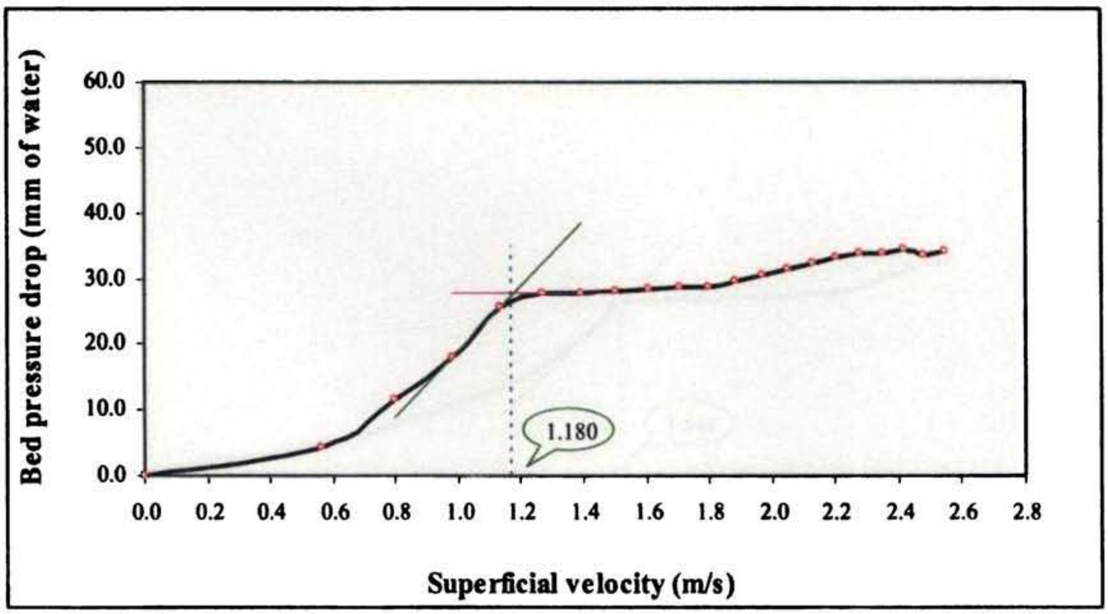


Figure 5.3.19 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 3.2 mm nylon as bed material.

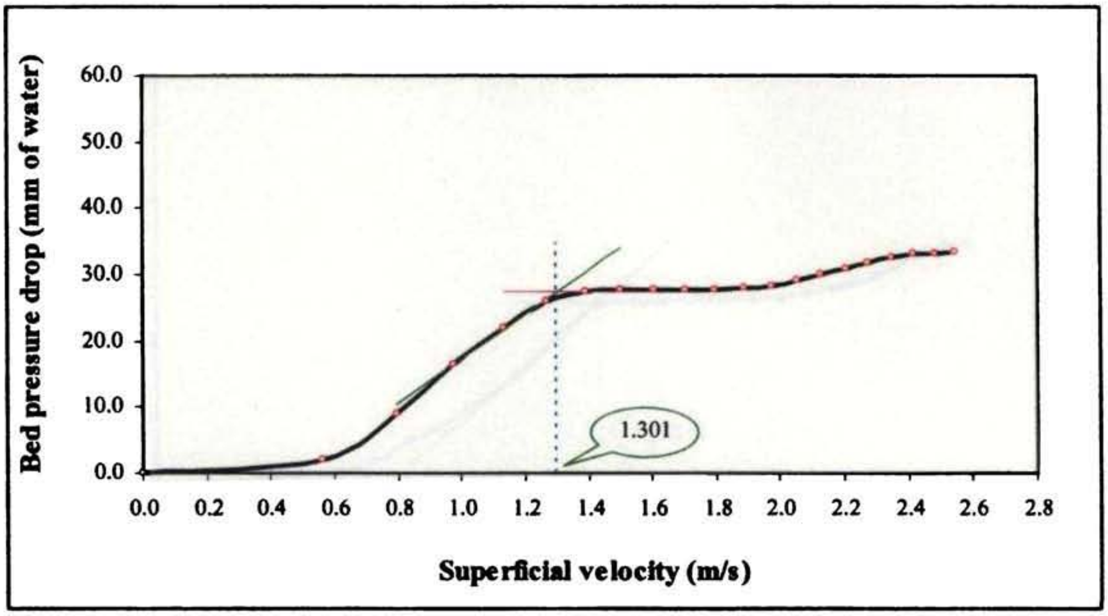


Figure 5.3.20 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 5.5 mm nylon as bed material.

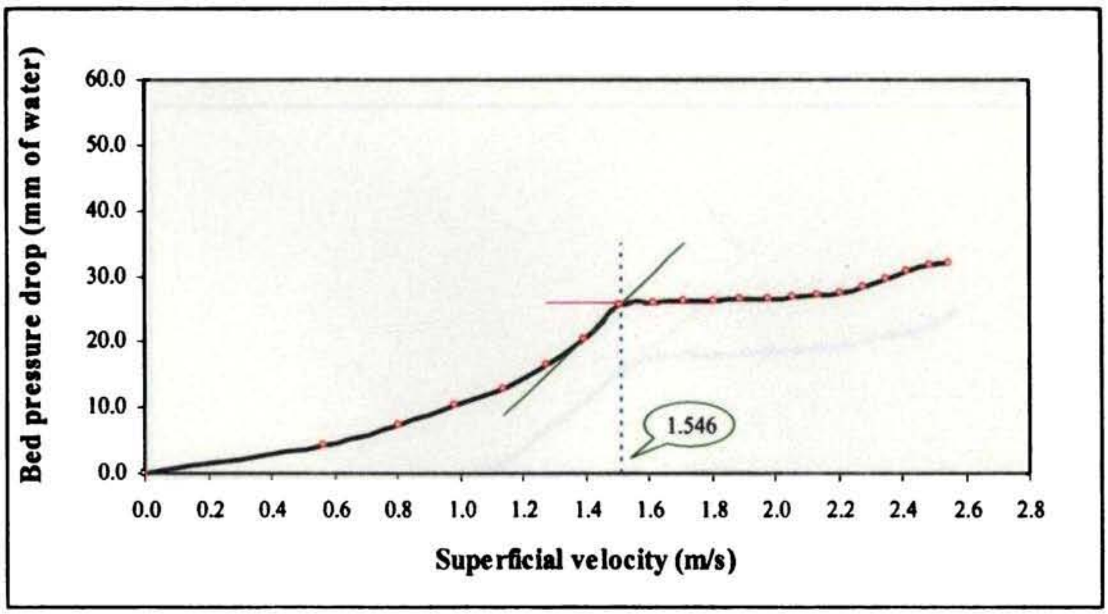


Figure 5.3.21 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 7.4 mm nylon as bed material.

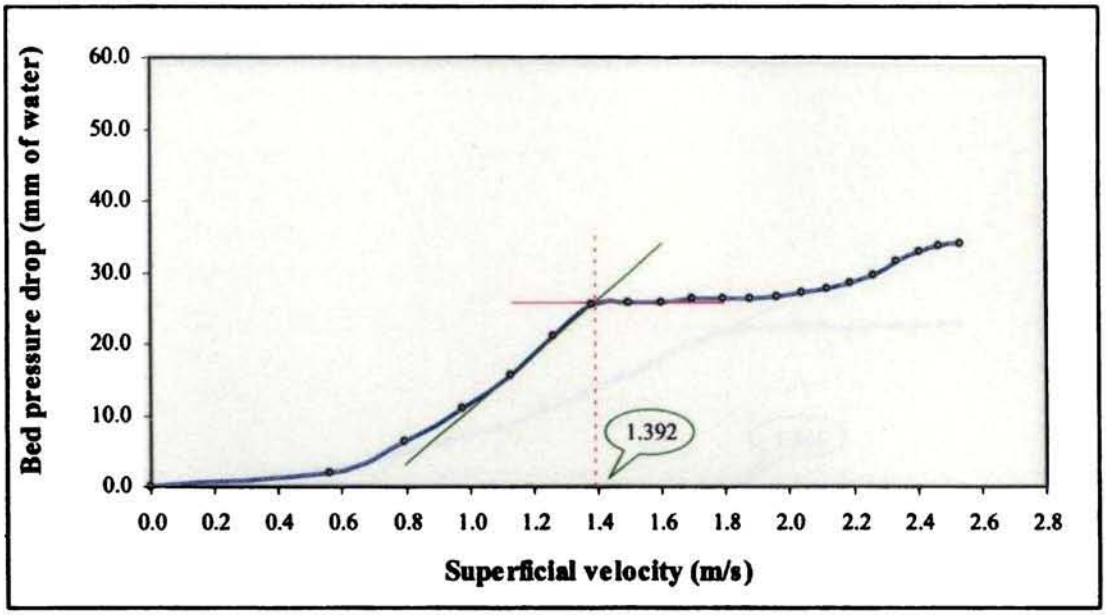


Figure 5.3.22 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 3.2 mm acetal as bed material.

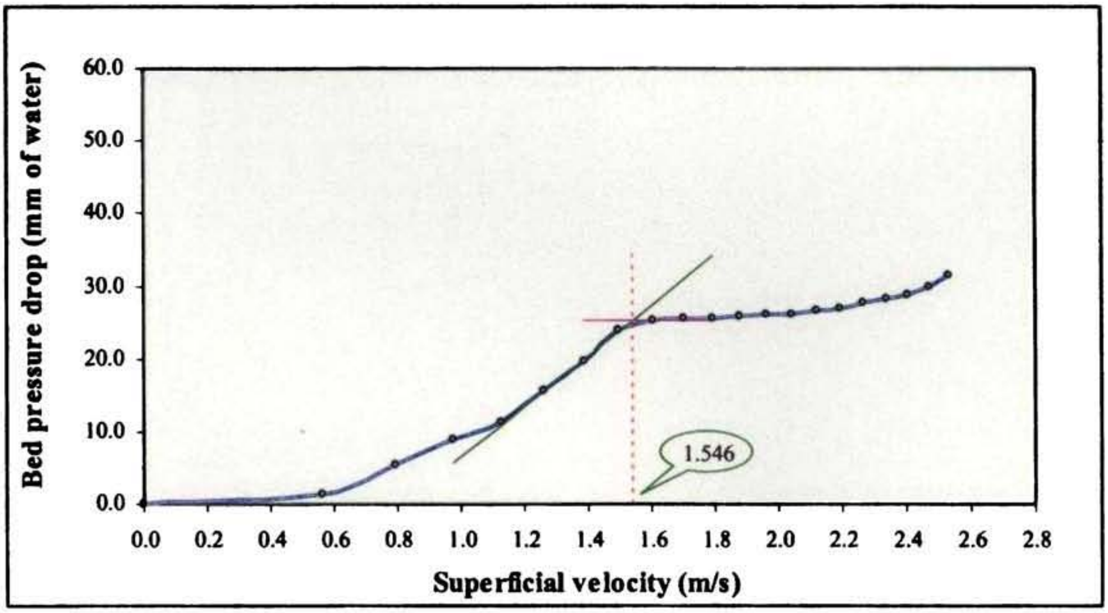


Figure 5.3.23 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 5.5 mm acetal as bed material.

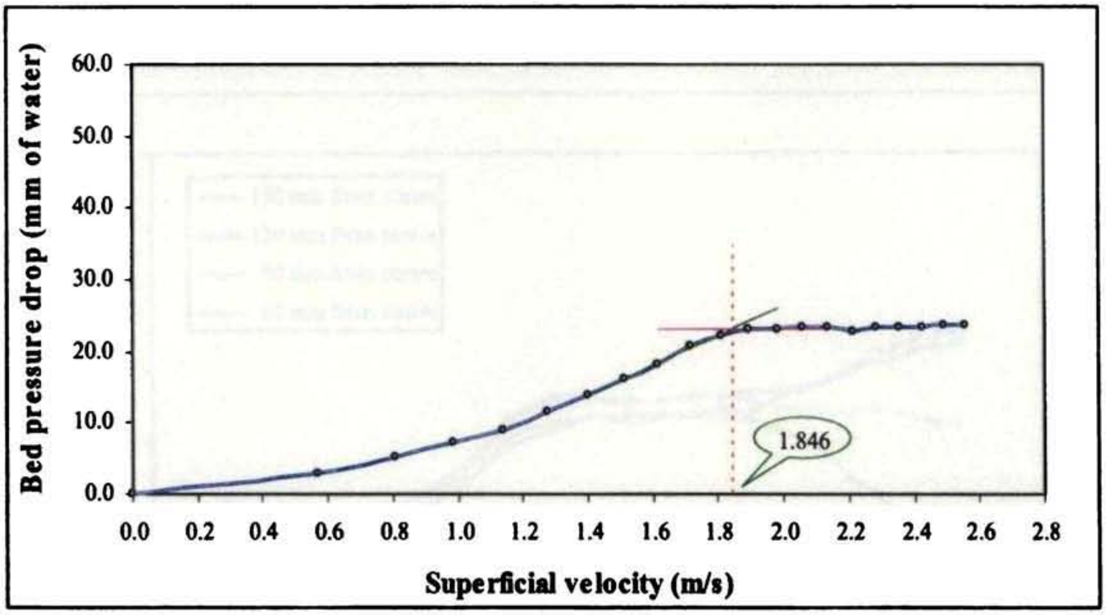


Figure 5.3.24 Minimum fluidizing velocity with distributor having angle of air injection $\phi = 15^0$ and using 7.4 mm acetal as bed material.

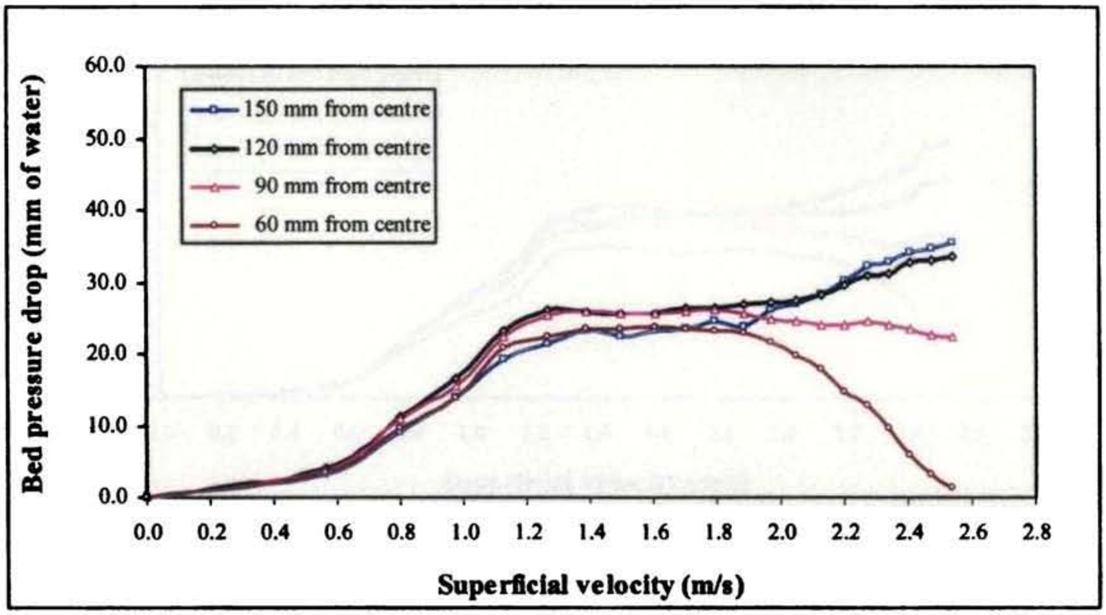


Figure 5.4.2 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection(Φ) = 0^0, using 3.2 mm acetal

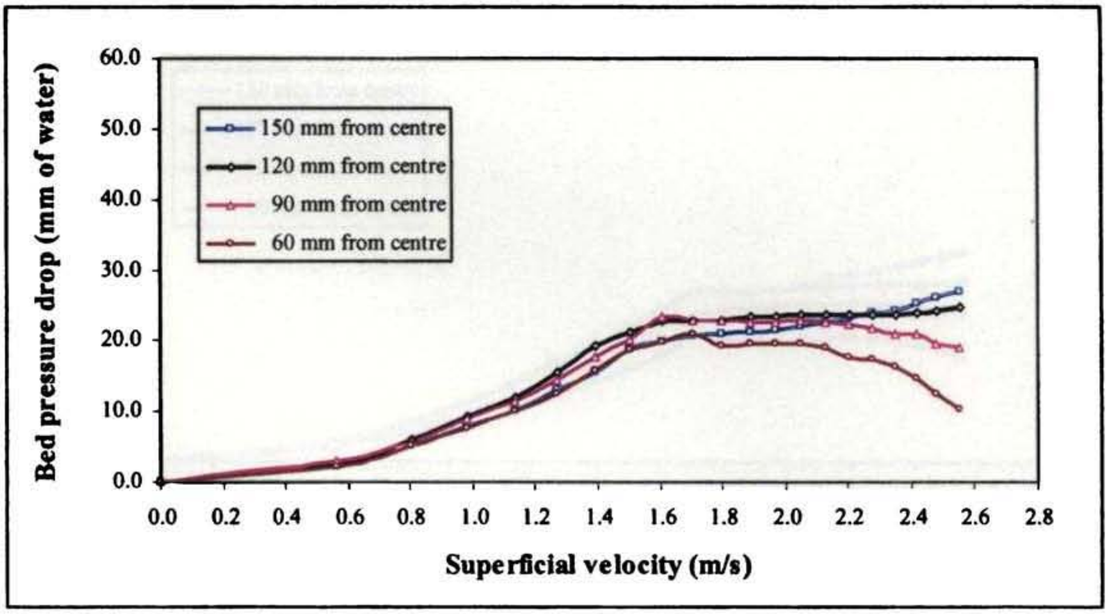


Figure 5.4.3 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 0^0, using 7.4 mm acetal

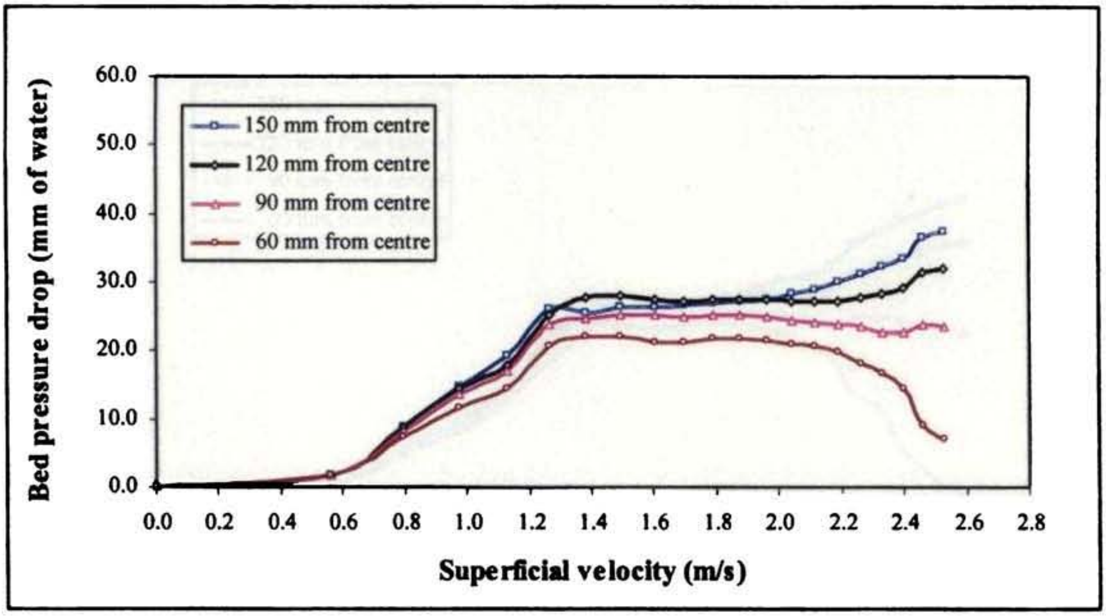


Figure 5.4.5 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 5°, using 3.2 mm acetal

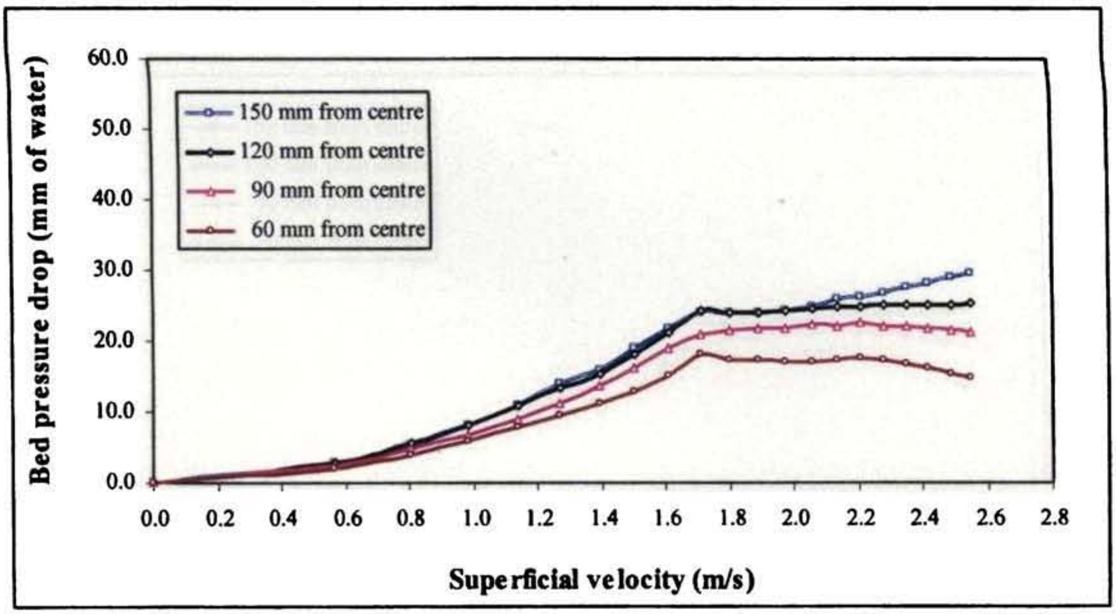


Figure 5.4.6 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 5°, using 7.4 mm acetal

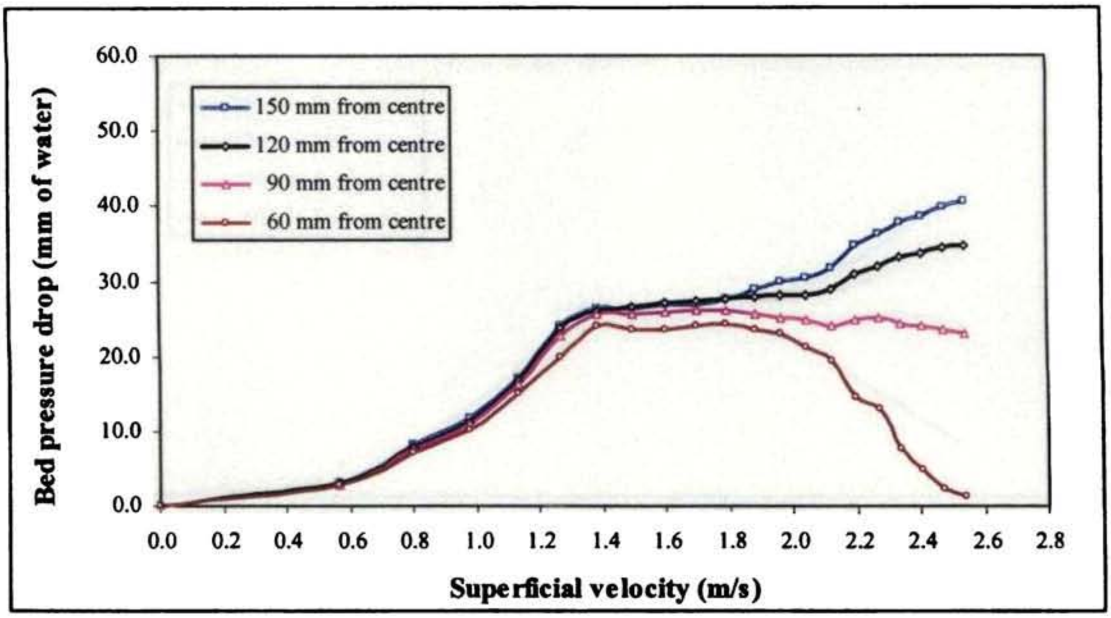


Figure 5.4.8 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 10°, using 3.2 mm acetal

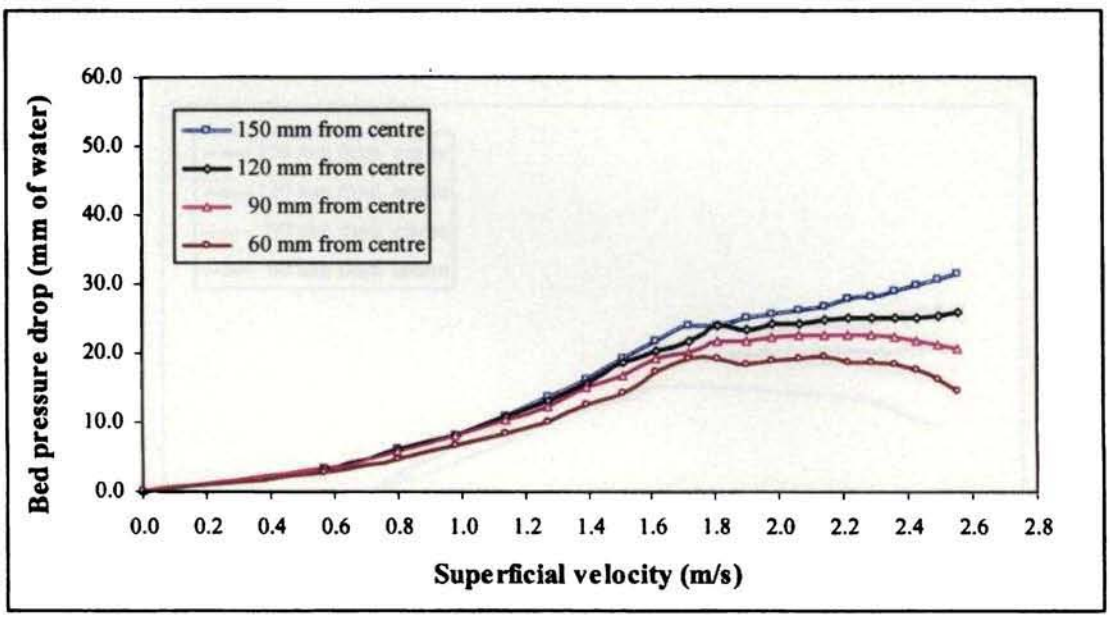


Figure 5.4.9 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 10°, using 7.4 mm acetal

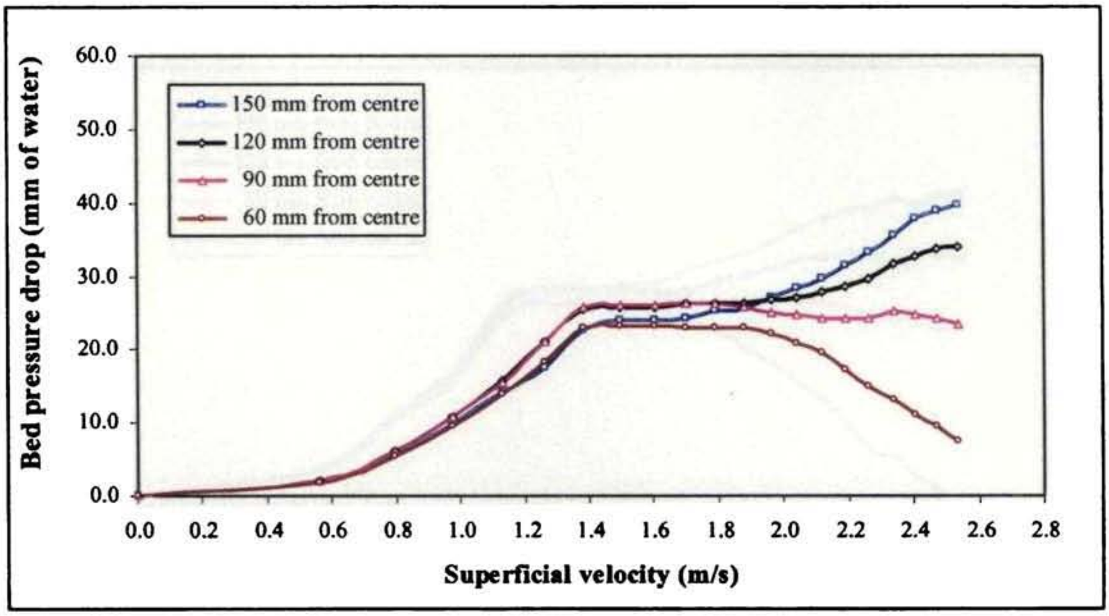


Figure 5.4.11 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 3.2 mm acetal

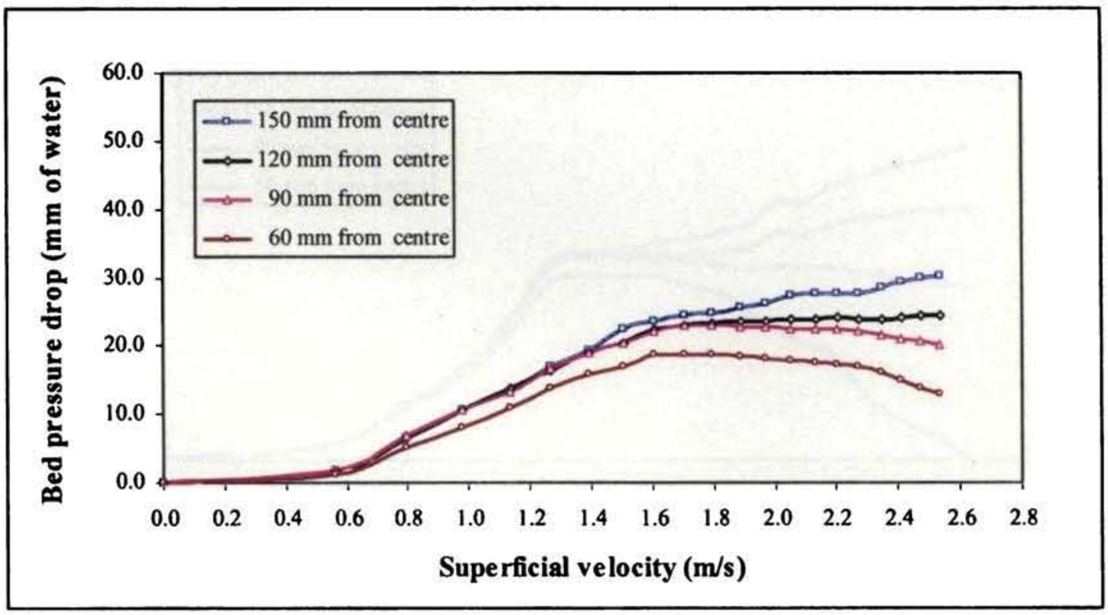


Figure 5.4.12 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 7.4 mm acetal

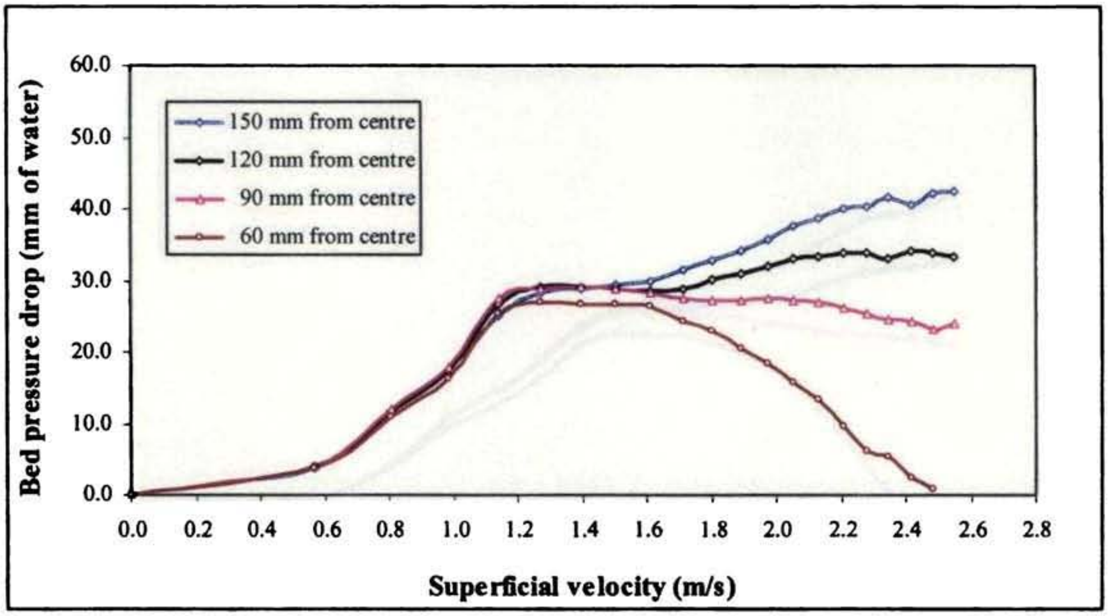


Figure 5.4.13 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 3.2 mm HDPE

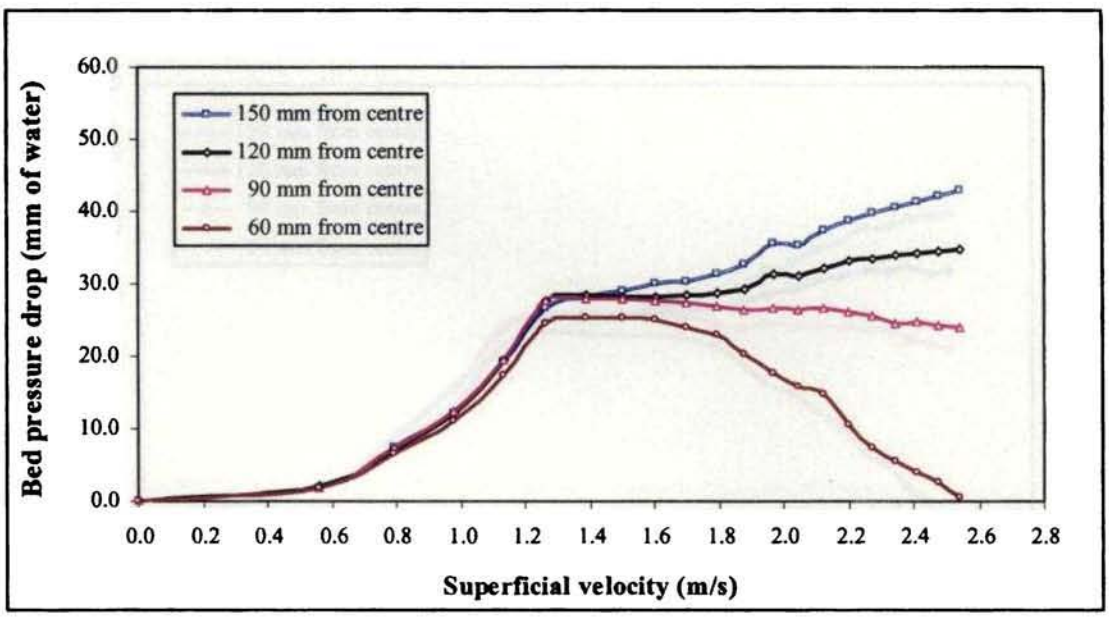


Figure 5.4.14 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 5.5 mm HDPE

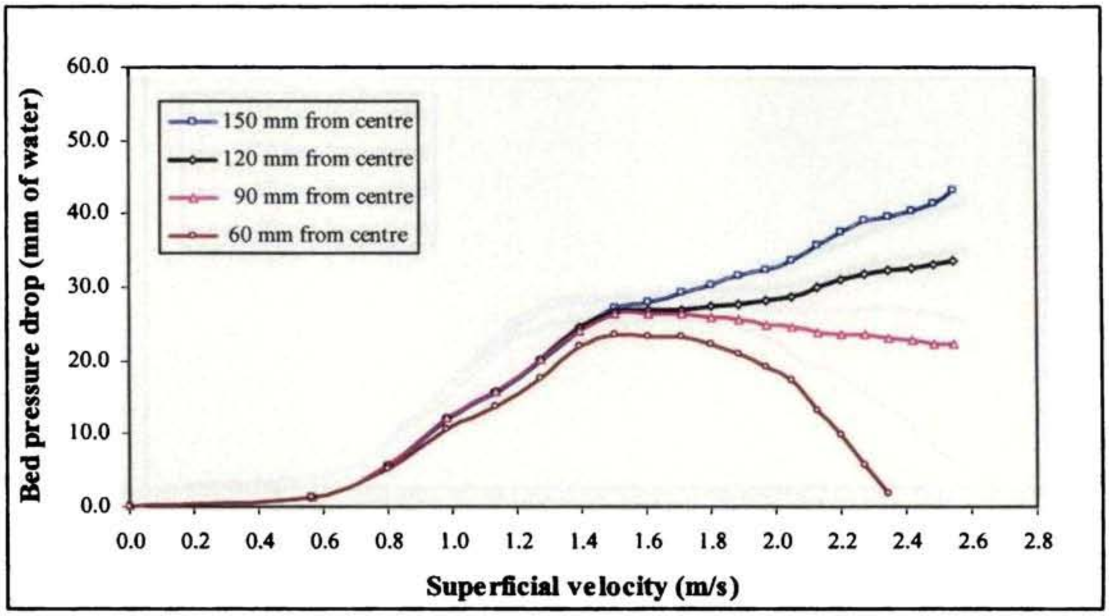


Figure 5.4.15 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 7.4 mm HDPE

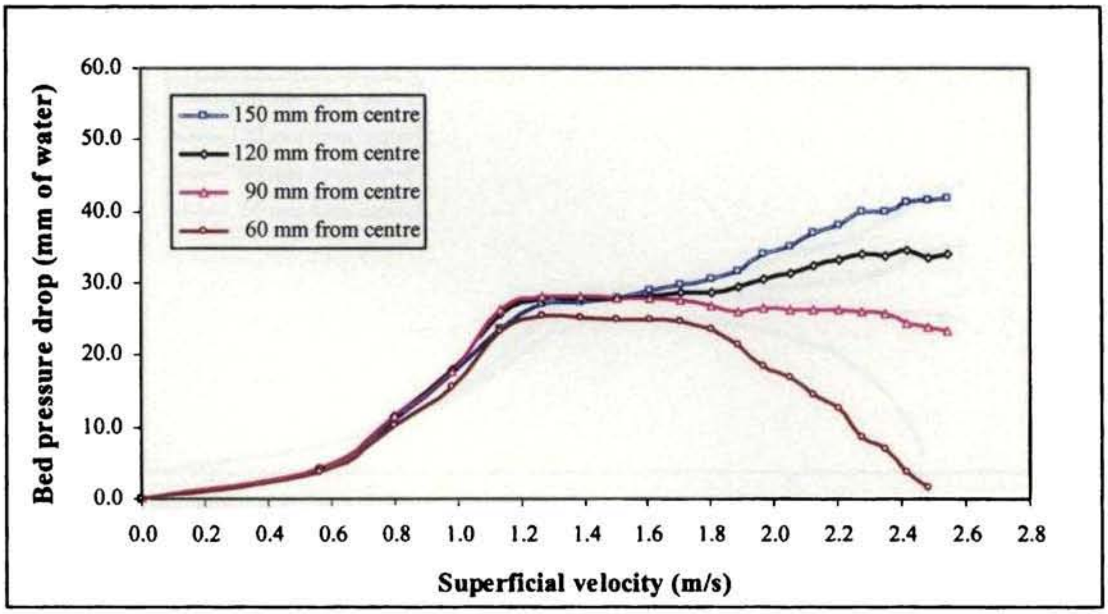


Figure 5.4.16 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 3.2 mm nylon

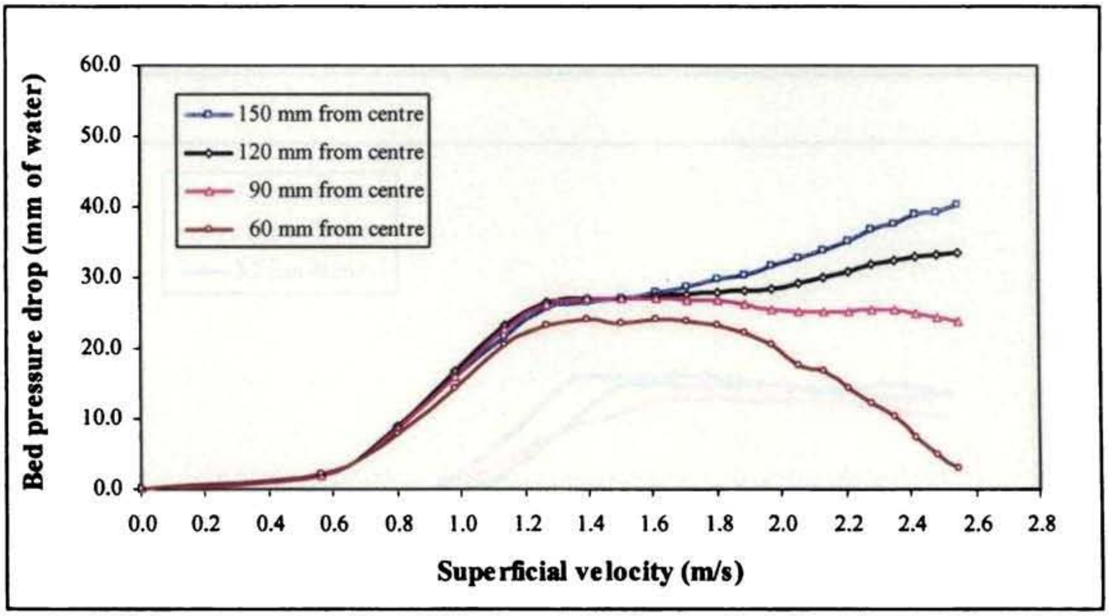


Figure 5.4.17 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 5.5 mm nylon

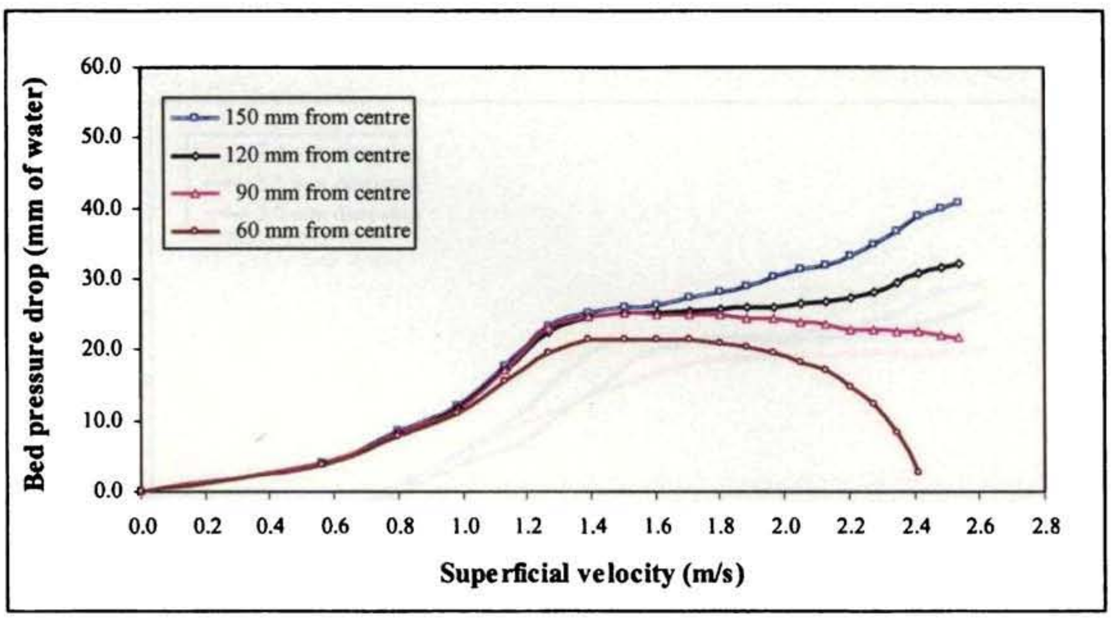


Figure 5.4.18 Variation of bed pressure drop with superficial velocity for distributor having angle of air injection (Φ) = 15°, using 7.4 mm nylon

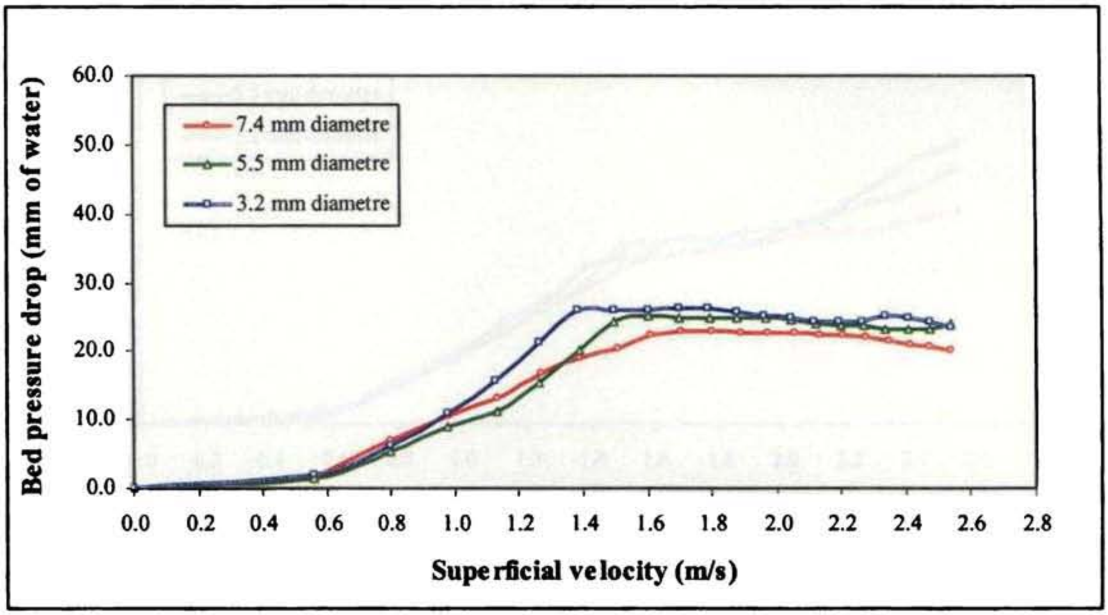


Figure 5.5.6 Variation of bed pressure drop with superficial velocity for different particle size at 90 mm from the centre of the distributor (Φ= 15°), using acetal

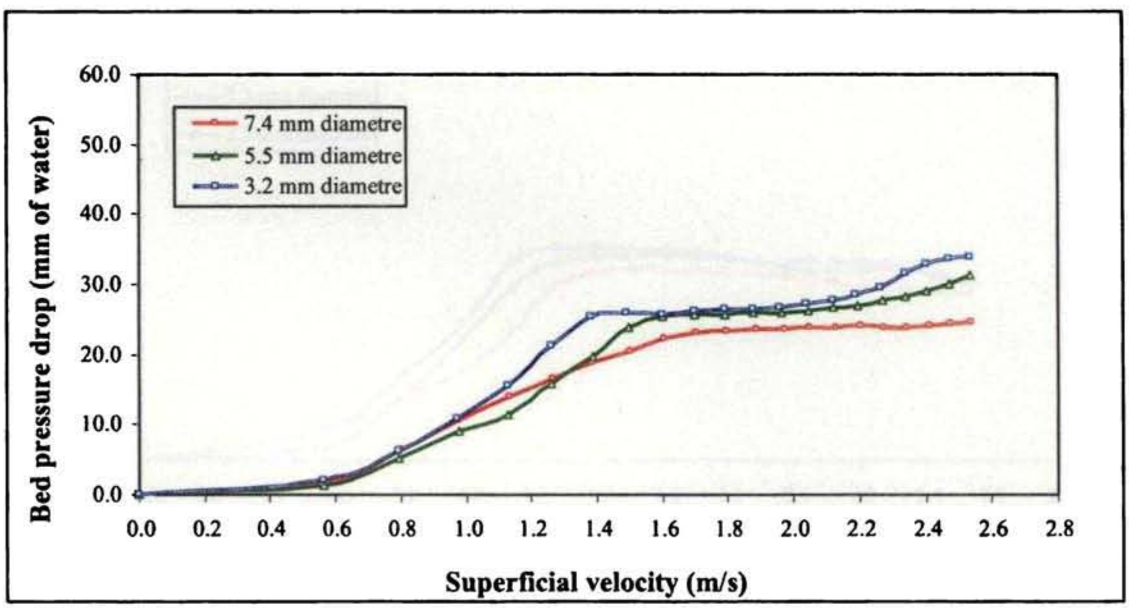


Figure 5.5.7 Variation of bed pressure drop with superficial velocity for different particle size at 120 mm from the centre of the distributor (Φ= 15°), using acetal

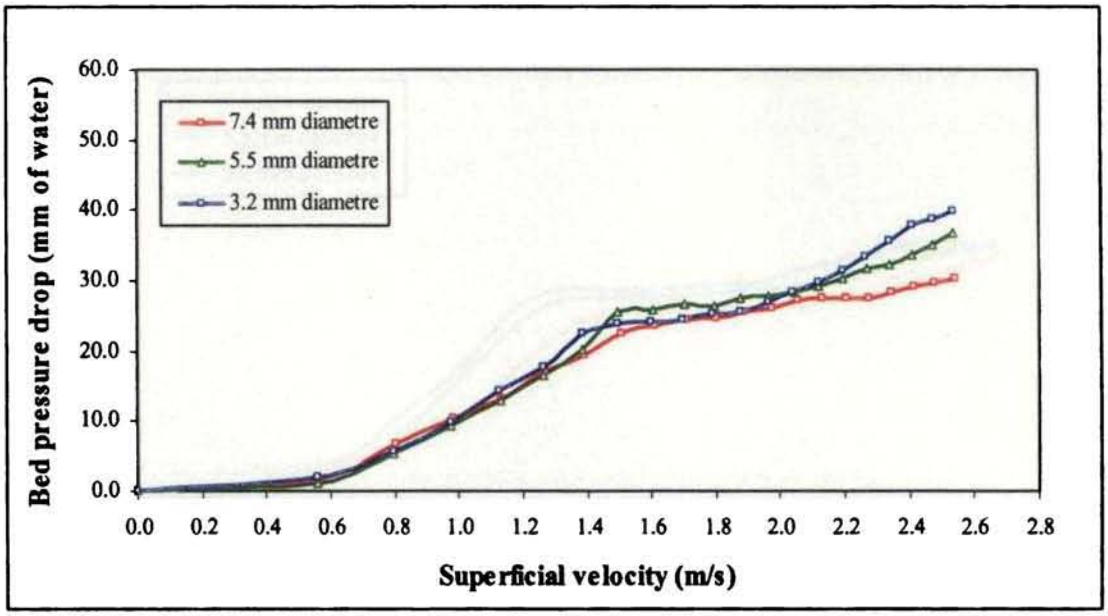


Figure 5.5.8 Variation of bed pressure drop with superficial velocity for different particle size at 150 mm from the centre of the distributor (Φ= 15°), using acetal

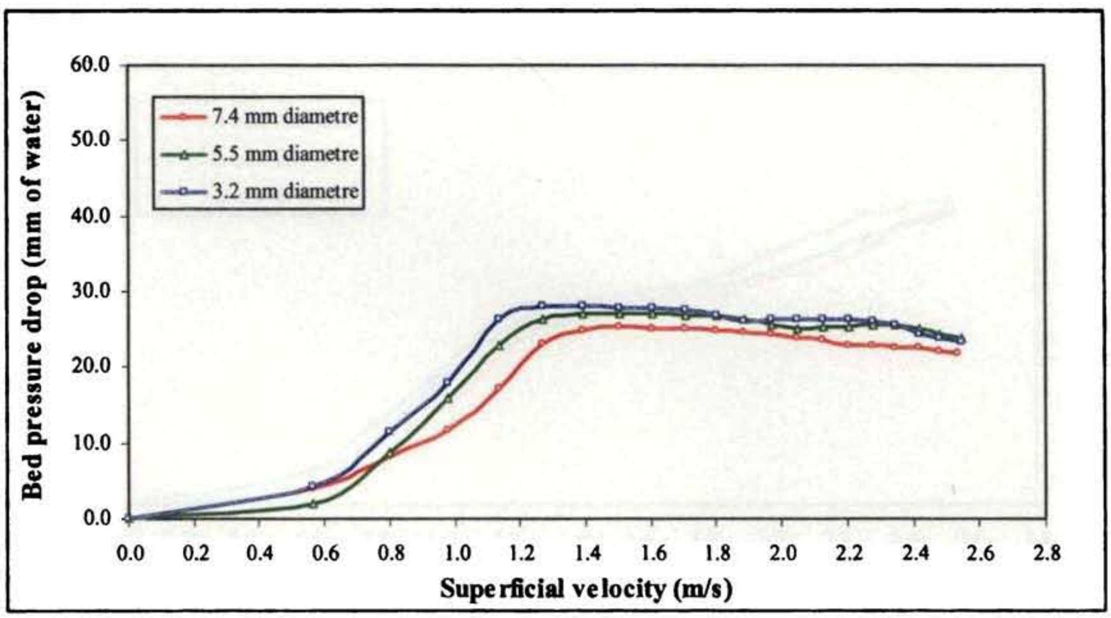


Figure 5.5.10 Variation of bed pressure drop with superficial velocity for different particle size at 90 mm from the centre of the distributor (Φ= 15°), using nylon

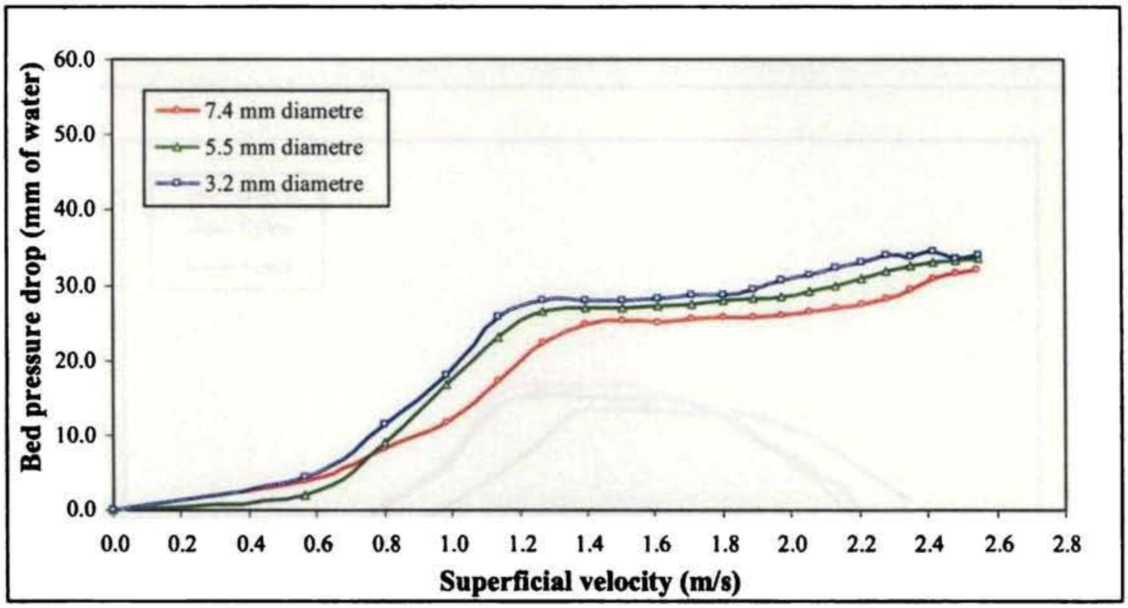


Figure 5.5.11 Variation of bed pressure drop with superficial velocity for different particle size at 120 mm from the centre of the distributor (Φ= 15°), using nylon

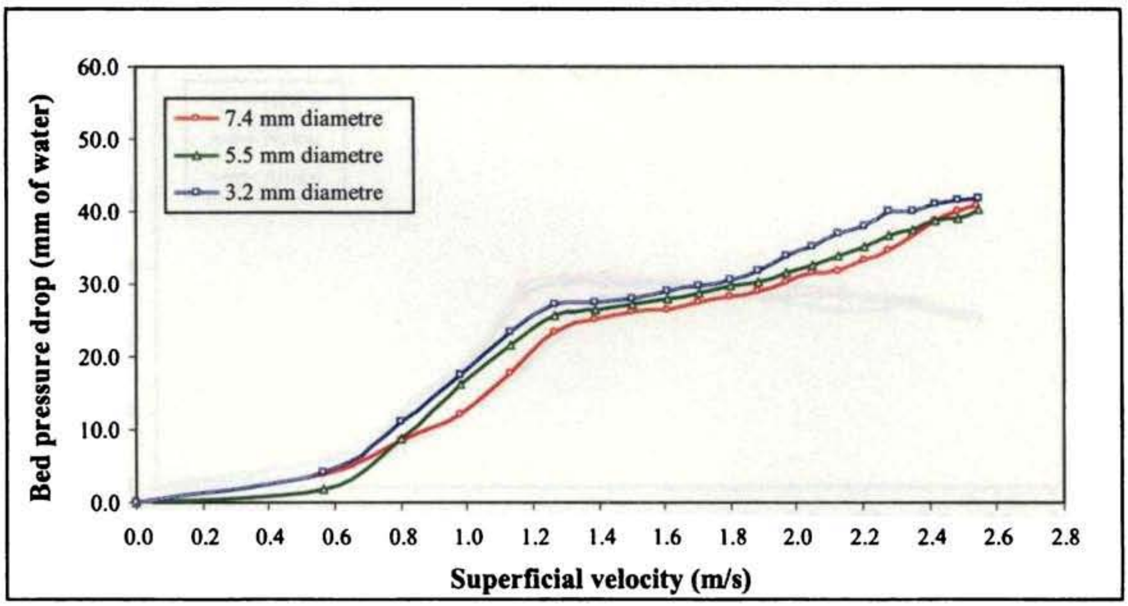


Figure 5.5.12 Variation of bed pressure drop with superficial velocity for different particle size at 150 mm from centre of the distributor (Φ= 15°), using nylon

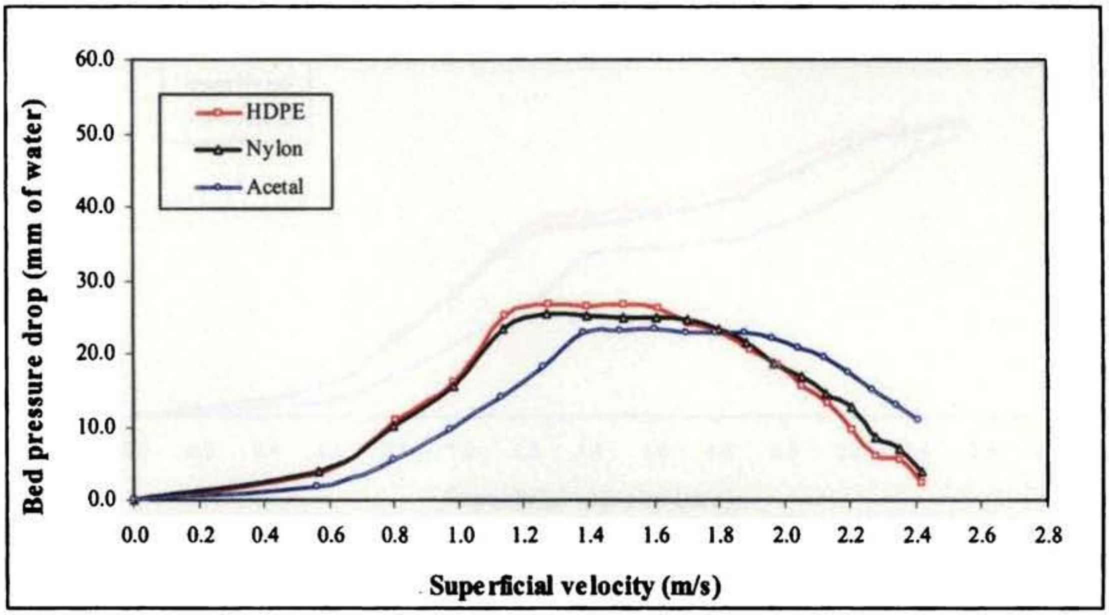


Figure 5.6.1 Variation of bed pressure drop with superficial velocity for different density materials of 3.2 mm diameter, at 60 mm from the centre of the distributor ($\Phi = 15^{\circ}$)

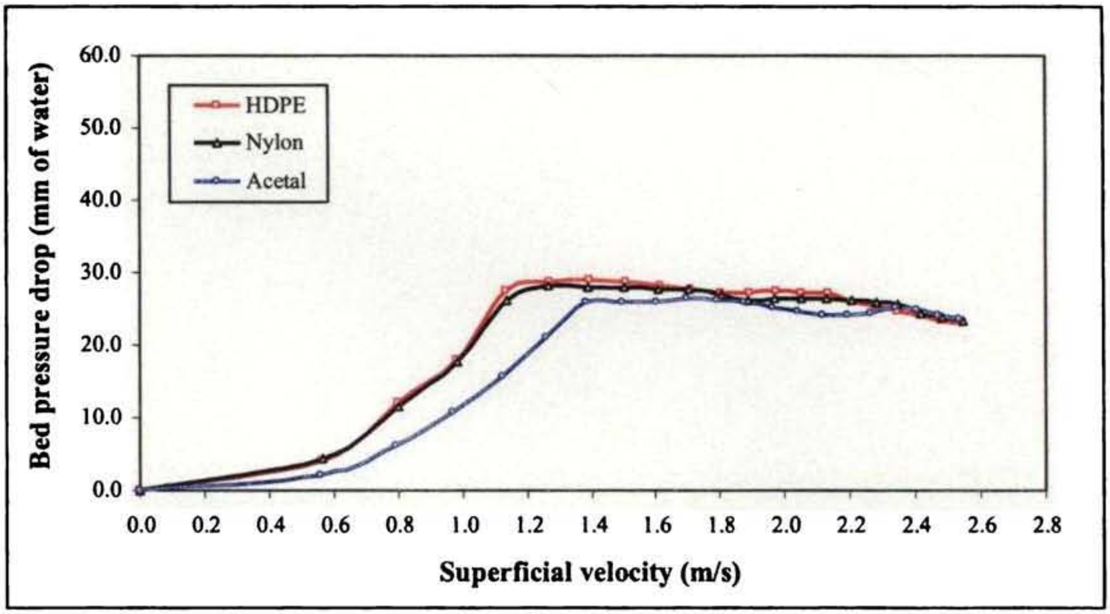


Figure 5.6.2 Variation of bed pressure drop with superficial velocity for different density materials of 3.2 mm diameter, at 90 mm from the centre of the distributor ($\Phi = 15^{\circ}$)

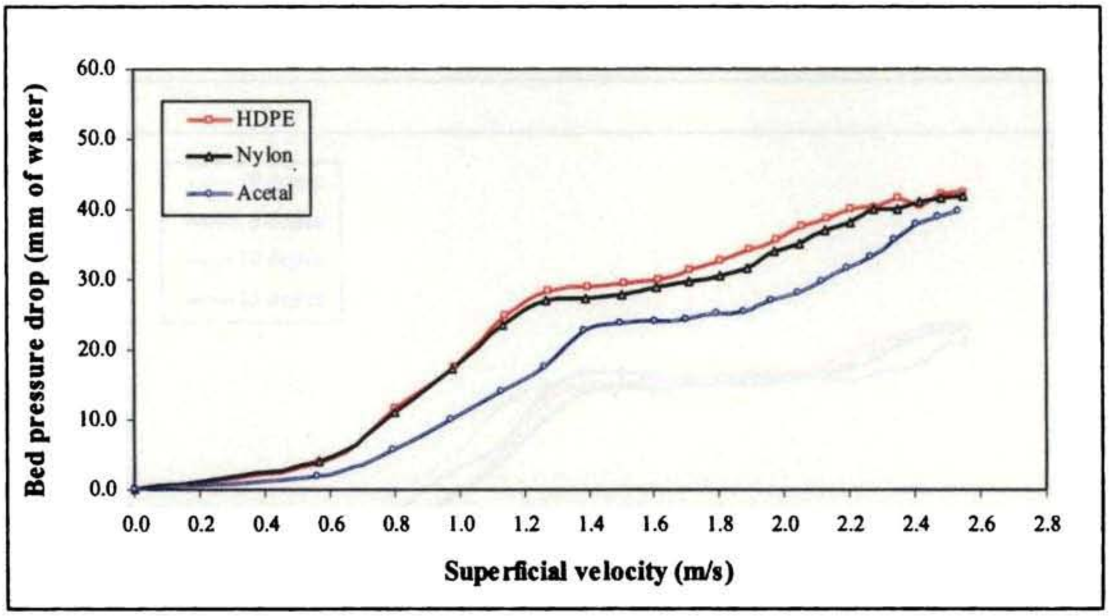


Figure 5.6.4 Variation of bed pressure drop with superficial velocity for different density materials of 3.2 mm diameter, at 150 mm from the centre of the distributor (Φ= 15°)

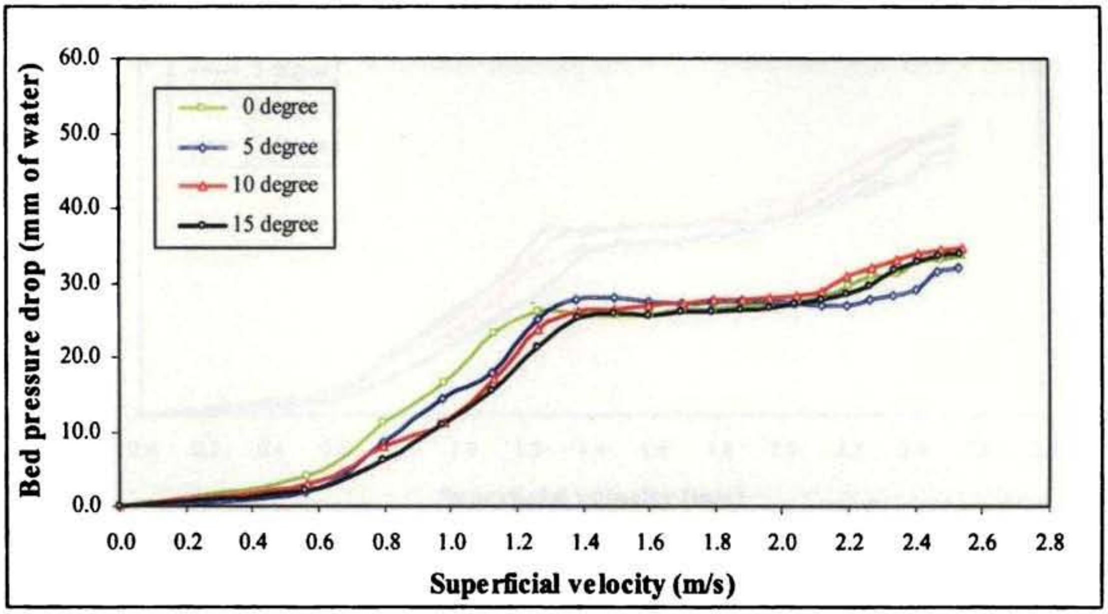


Figure 5.7.2 Variation of bed pressure drop with superficial velocity for different angles of air injection using 3.2 mm diameter acetal, at 90 mm from the distributor centre

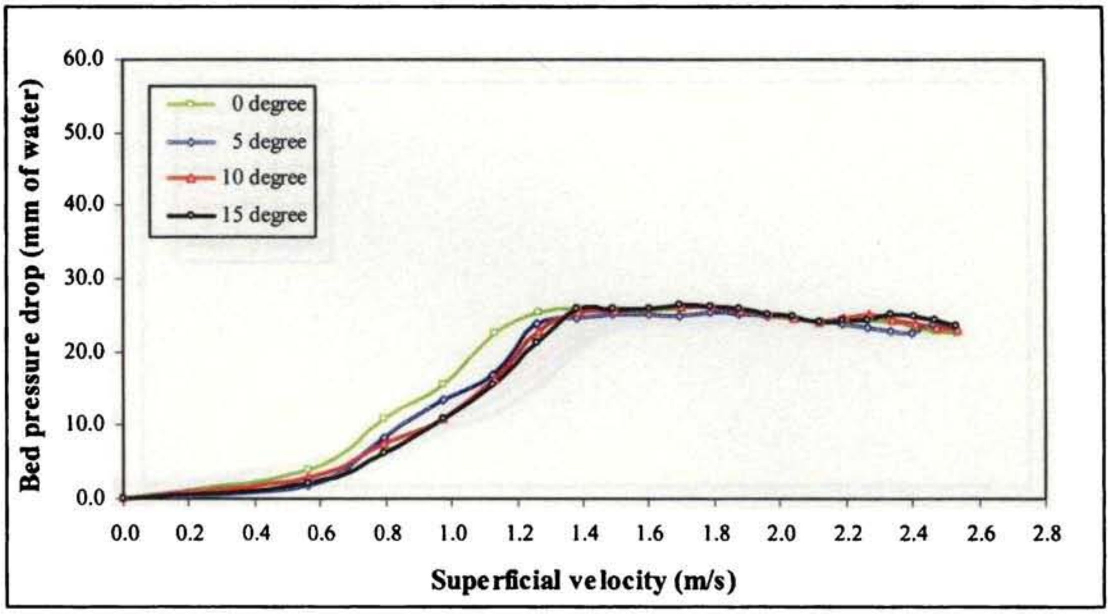


Figure 5.7.3 Variation of bed pressure drop with superficial velocity for different angles of air injection using 3.2 mm diameter acetal, at 120 mm from the distributor centre

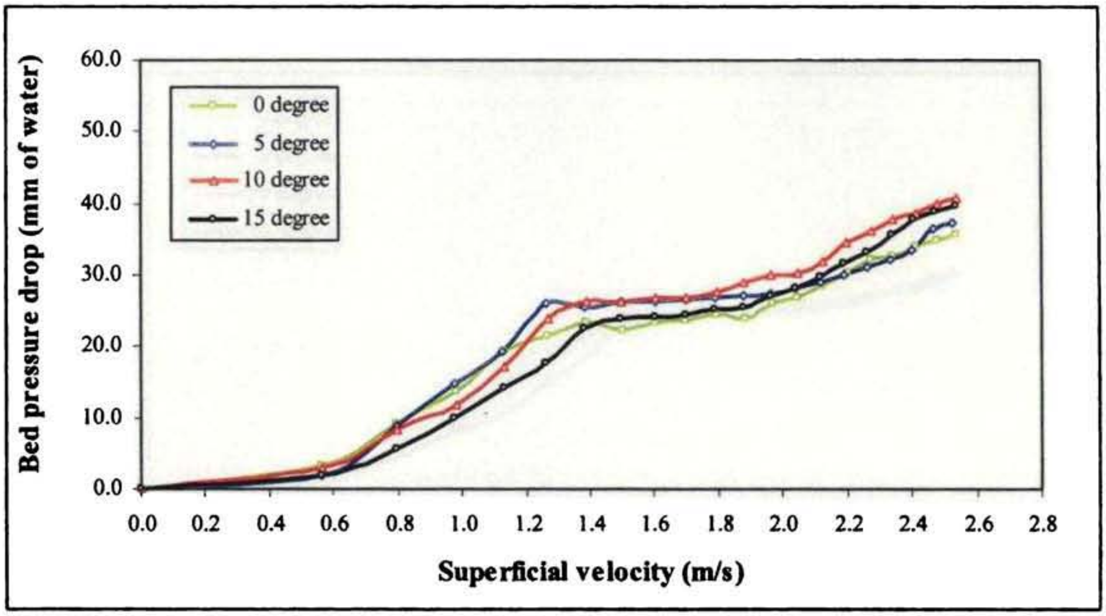


Figure 5.7.4 Variation of bed pressure drop with superficial velocity for different angles of air injection using 3.2 mm diameter acetal, at 150 mm from the distributor centre

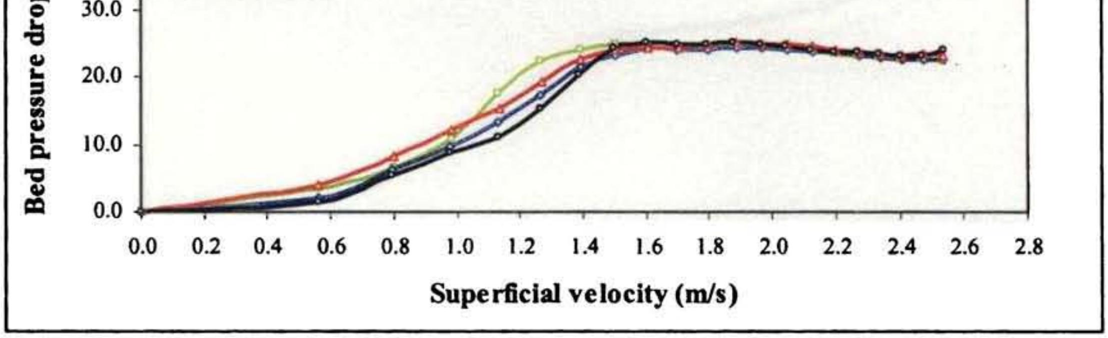


Figure 5.7.6 Variation of bed pressure drop with superficial velocity for different angles of air injection using 5.5 mm diameter acetal, at 90 mm from the distributor centre

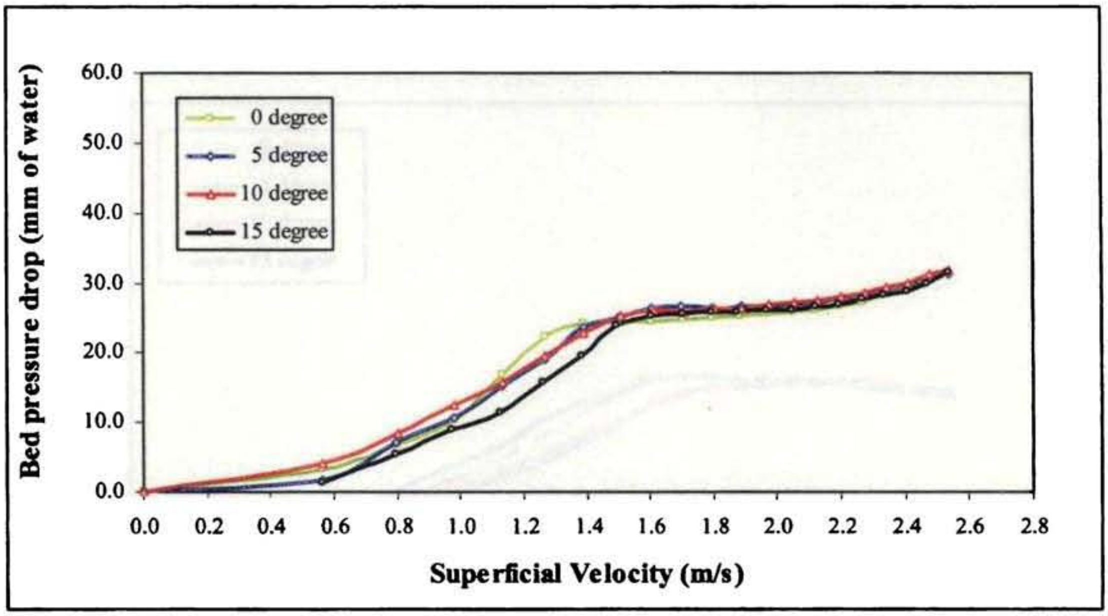


Figure 5.7.7 Variation of bed pressure drop with superficial velocity for different angles of air injection using 5.5 mm diameter acetal, at 120 mm from the distributor centre

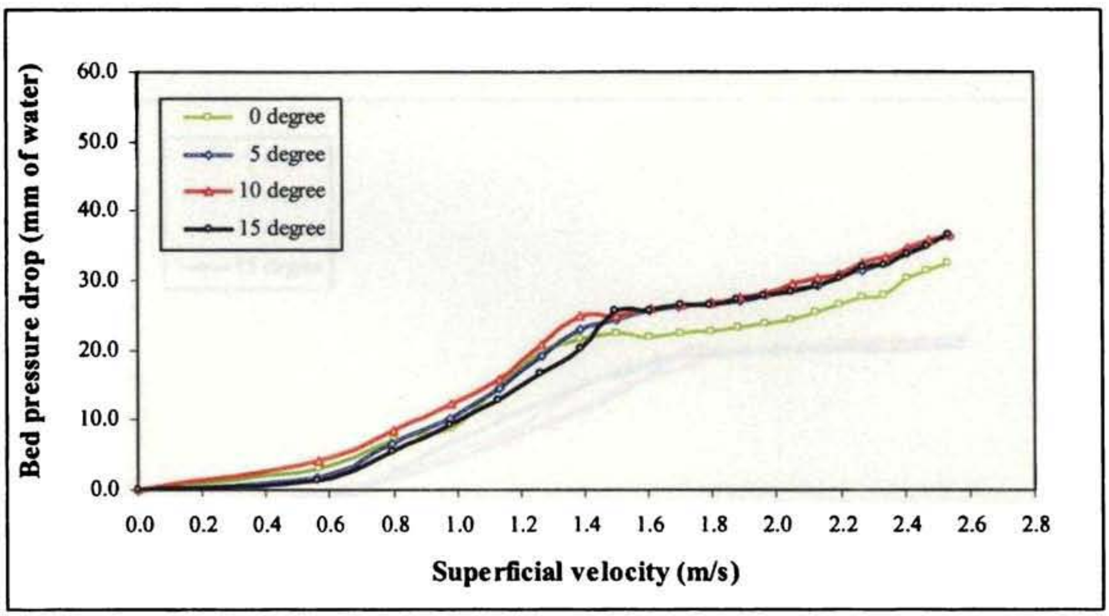


Figure 5.7.8 Variation of bed pressure drop with superficial velocity for different angles of air injection using 5.5 mm diameter acetal, at 150 mm from the distributor centre

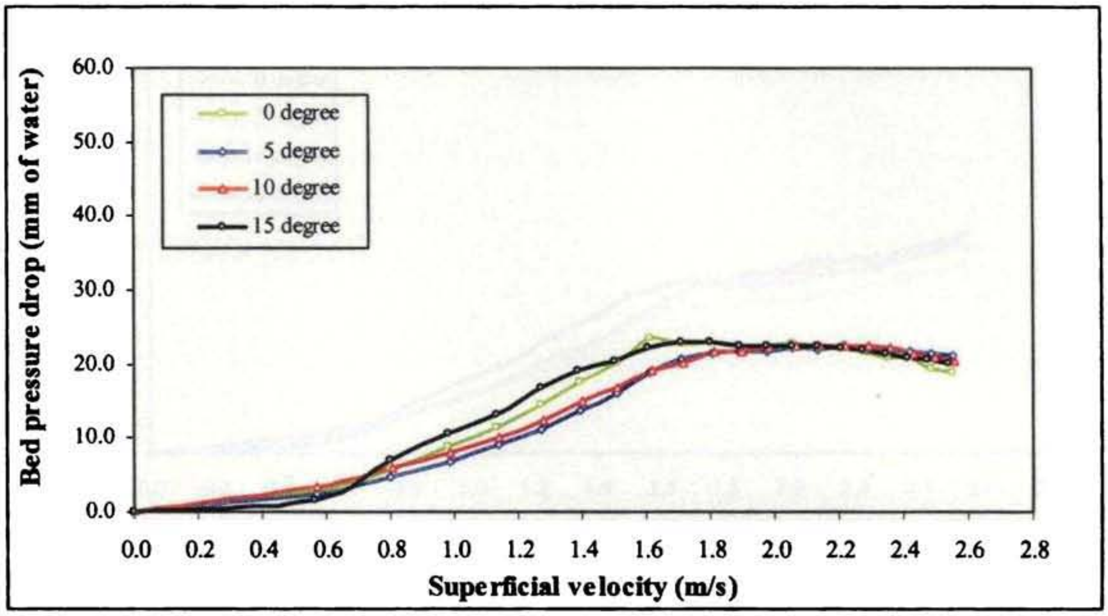


Figure 5.7.10 Variation of bed pressure drop with superficial velocity for different angles of air injection using 7.4 mm diameter acetal, at 90 mm from the distributor centre

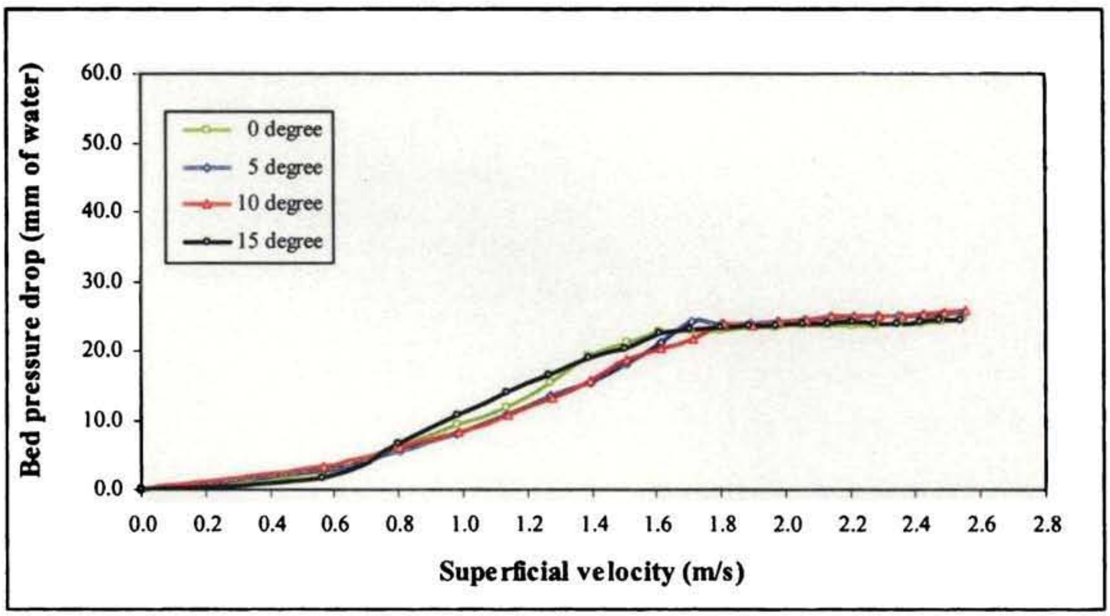


Figure 5.7.11 Variation of bed pressure drop with superficial velocity for different angles of air injection using 7.4 mm diameter acetal, at 120 mm from the distributor centre

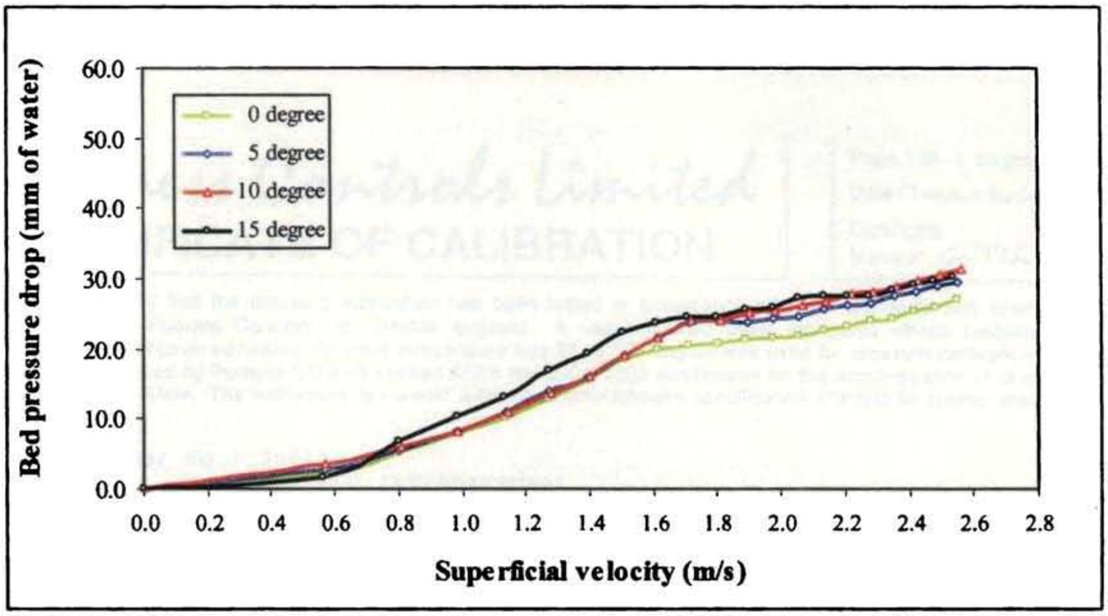


Figure 5.7.12 Variation of bed pressure drop with superficial velocity for different angles of air injection using 7.4 mm diameter acetal at 150 mm from the distributor centre

APPENDIX – VI

CALIBRATION CERTIFICATE OF MICRO-MANOMETER

The calculation details of digital micro-manometer are presented below.

Furness Controls Limited

CERTIFICATE OF CALIBRATION

Page 1 of 1 pages
Date 17 march 2004
Certificate Number CL7921

This is to certify that the following instrument has been tested in accordance with the standards laid down and adhered to by Furness Controls Ltd., Bexhill, England. It was calibrated using standards whose calibration is traceable to European standards. Ambient temperature was 20 ±2 °C. Dry air was used for pressure calibration. This certificate is issued by Furness Controls Limited which holds ISO 9002 certification for the administration of its quality management system. The instrument is / ~~is not~~* within the manufacturer's specification. (*delete as appropriate)

Sales Order No : 36018
Customer : SARTECH INTERNATIONAL
Instrument : AIRPRO
Instrument type: FCO520
Supply voltage : Battery
Serial No : 0402176

Pressure range : -600 to 600 Pa

Reference Pressure Pa	Instrument Display Pa	calibrated deviation 0	% of R
600.130	-599.7	-0.4	-0.07
480.197	-480.0	-0.2	-0.04
359.757	-360.2	0.4	0.11
239.623	-240.0	0.4	0.17
120.239	-120.4	0.2	0.17
0.000	0.0	0.0	
120.202	120.2	-0.0	0.00
240.049	240.3	0.3	0.12
360.113	360.1	-0.0	0.00
480.758	480.6	-0.2	-0.04
601.539	600.5	-1.0	-0.17

Reference instruments used to carry out calibration of the above unit.

Pressure standard : FRS4HR serial number (FCPS017), RS36, Calibration uncertainty better than 0.01% of reading + 0.002 Pa

Electrical multimeter : Thurlby 1905a - FCL ref. number WS96 Calibration accuracy 0.01%r + 1digit.

Approved sign. : G Thorogood sign : ....

History of traceability to National Standards and a list of approved signatories are available from Furness Controls Ltd.

The uncertainties are for a confidence level of not less than 95%

REFERENCES

1. Agarwal J.C., Davis W.L. and King D.T., (1962), "Fluidized -bed coal dryer", Chemical *Engineering Progress*, Vol. 58, pp 85-90.

2. Binod Srenivasan and Vijay R. Raghavan, (2003), "Hydrodynamics of a swirling fluidized bed", *Chemical Engineering and Processing*, Vol. 41, pp 99-106.

3. Birk R.H., Camp G.A and Hutchinson L.B., (1990), "Design of an agglomeration resistant gas distributor", *All India Chemical Engineering Symposium,* Vol. 86, pp 16-21.

4. Botterill J.S.M., Teoman Y. and Yuregir K.R (1982), "The effect of operating temperature on the velocity of minimum fluidization, bed voidage and general behaviour", *Powder Technology*, Vol. 31, pp 101-110.

5. Bouratoua R., Molodtsuf Y. and Koniuta A (1993), "Hydrodynamic characteristics of a pressurized fluidized bed". *Proceedings of the 12th International Conference on Fluidized bed Combustion,* San Diego, California, 9-13 May, Vol.1, pp 63-70.

6. Brown G.E. and Farkas D.F. (1972), "Centrifugal fluidized bed", *Food Technology*, Vol.26, pp 23-30.

7. Cheng-Sang Chiyang and Cheng-Chong Huang (1991), "Pressure drop across a perforated-plate distributor in a gas-fluidized bed", *Journal of Chemical Engineering of Japan,* Vol. 24, pp 358-362.

8. Chen Y.M. (1987), "Fundamentals of a centrifugal fluidized bed", *All India Chemical Engineering Journal*, Vol. 23, pp 249-252.

9. De Groot J.H and Hasset N.J (1967), *Proceedings of International Symposium on Fluidization*, Endhoven, Vol. 24, pp 32-35.

10. Ellias.N, Bi.H.T., Lim C.J and Grace J.R (2004), "Hydrodynamics of turbulent fluidized beds of different diameters", *Powder Technology*, Vol. 98, pp 124-136.

11. Fakhimi S., Sohrabi S and Harrison D (1983), "Entrance effect of a multi-orifice distributor in gas fluidized beds", *The Canadian Journal of Chemical Engineering*, Vol. 61, pp 364-369.

12. Fan L.T., Cheng C.C. and Yu Y.S., Teruo Takahashi and Zennosake Tanaka (1985), "Incipient fluidization condition for a centrifugal fluidized bed", *All India Chemical Engineering Journal*, Vol. 31, pp 999-1009.

13. Farkas D.F., Lazar M.E and Butterworth T.A (1969), "The Centrifugal fluidized bed", *Food Technology*, Vol. 23, pp125-129.

14. Geldart D and Bayens J (1985), "The Design of distributors for gas fluidized beds", *Powder Technology*, Vol. 42, pp 67-70.

15. Gelperin N.I, Einstein V.G and Zaikovski (1960), Khimicheskoe Mashinostroenic No.3,1. *Printed at the Autumn Meting of The Society of Chemical Engineers,* Japan at Nagoya, October 1982.

16. Hanni D.F., Farkas D.F and Brown G.E (1976), "Design and operating parameters for a continuous fluidized bed dryer", *Journal on Food Science* Vol. 41, pp 1172 – 1176.

17. Happel J and Einstein N (1954), "Viscous flow in multiparticle systems: cubical assemblage of uniform spheres". *Industrial Engineering and Chemistry,* Vol. 46, pp 1187.

18. Hiby J.W (1967),"Periodic Phenomena Connected with Gas-Solid Fluidization". pp 99-106.

19. Howard J.R and Metcalfe C.I (1977), "Fluidization", Cambridge University Press, Cambridge, pp 296-327.

20. Howard J.R (1989), "Fluidized Bed Technology Principles and Applications", Adam Higher, Bristol, New York.

21. Kamil M, Fakhimi M and Raghavan V.R (2007), "Parametric analysis of an analytical model of a swirling fluidized bed", *Proceedings on Asian Power and Energy Systems* – 2007, pp 560.

22. Kao J, Pfeffer R and Tardas G.I (1987), "On partial fluidization in rotating fluidized beds", *All India Chemical Engineering Journal*, Vol. 33, pp 858-866.

23. Kassim W.M.S (1972), "Flowback of solids through distributor plate of fluidized bed", *PhD. Thesis*, University of Aston, Birmingham.

24. Kawabata, Yumiyama J.M and Tazaki Y (1981), "Characteristics of gas fluidized beds under pressure", *Journal of Chemical Engineering of Japan*, Vol. 14, pp 85-89.

25. Kroger D.G, Levy E.K and Chen J.C (1979), "Flow characteristics in packed and fluidized rotating bed", *Powder Technology*, Vol. 24, pp 9-12.

26. Luca Mazzei and Paola Letteri (2007), "A drag force closure for uniformly dispersed fluidized suspensions", *Chemical Engineering Science*, Vol.62, pp 6129-6142.

27. Madhiyanon T, Piriyarungroj and Soponronnarit S (2007), "Cold flow behaviour study in novel cyclonic fluidized bed combustor", *Energy Conservation and Management*, Vol. 49. pp 1202- 1210.

28. Masaaki Nakamura, Yoichi Hamada and Shigeki Toyama (1985), "Experimental investigation of minimum fluidization velocity at elevated temperatures and pressures", *The Canadian Journal of Chemical Engineering*, Vol. 63, pp 8-13.

29. Medlin J and Jackson R (1978), "Fluid mechanical description of fluidized beds- The effect of distributor thickness on convective instabilities", *Industrial Engineering Chemical Fundamentals*, Vol.14, pp 315-320.

30. Michael Wormsbecker, Todd S Pugsley and Helen Tanfara (2007), "The influence of distributor design on fluidized bed dryer hydrodynamics", *2007 ECI Conference on The 12th International Conference on Fluidizastion*, Vancouver, Canada, Vol. RP 4, May 13-17, 2007.

31. Mohanty Y.K, Biswal K.C, Roy G.K and Mohanty B.P (2007), "Effect of promoters on dynamics of gas-solid fluidized bed-statistical and ANN approaches", *China Particuology*, Vol.5, pp 401-407

32. Moreno R and Rios R (2002), "Study on sawdust drying techniques in fluidized bed", *Biosystems Engineering*, Vol. 82, pp 321-329.

33. Ouyang F and Levenspiel O (1986), "Spiral distributors for fluidized beds", *Industrial Engineering Chemical Process Design and Development*, Vol. 25, pp 279-287.

34. Paulose M.M and Narayanan Nampoothiri V.N (2004), "Comparison of different distributors in swirling fluidized bed", *Proceedings of the National Conference on Advances in Mechanical Engineering Science*, Sidhartha Institute of Techology, Tumkur, Karnataka, September 24, 2004.

35. Paulose M.M and Narayanan Nampoothiri V.N (2005), "Experimental study of wave motion in swirling fluidized bed", *Proceedings of the National Conference on Avances in Mechanical Engineering in The Era of Globalisation*, Tagore Engineering College, Rathinamangalam, Chennai, February 11, 2005.

36. Qureshi A.E and Creasy D.E (1979), "Fluidized bed gas distributors", *Powder Technology*, Vol. 22, pp 113-119.

37. Richardson J.F and Zaki W.N (1954), "Sedimentation and Fluidization: Part-I", *Transactions of the Institution of Chemical Engineers,* Vol. 32, pp 35-38.

38. Sathyamoorthy D and Rao C.H (1977), "Gas distributors in fluidized beds", Powder *Technology*, Vol.20, pp 47-52.

39. Sathyamoorthy D and Rao C.H (1981), "The choice of distributor to bed pressure drop ratio in gas fluidized beds", *Powder Technology*, Vol. 30, pp 139- 143.

40. Sathyamoorthy D and Masayuki Hori (2003), "On the influence of aspect ratio and distributor in gas fluidized beds", *Chemical Engineering Journal,* Vol. 93 pp 151-161.

41. Saxena S.C, Chatterjee A and Patel R.C (1979), "Effect of distributors on gas-solid fluidization", *Powder Technology*, Vol. 22, pp 191- 198.

42. Siegel R (1976), "Effect of distributor plate-to- bed resistance ratio on onset of fluidized bed channeling", *All India Chemical Engineering Journal*, Vol. 22, pp 590-594.

43. Shi M.H, Wong H and Hao Y.L (2000), "Experimental investigation of the heat and mass transfer in a centrifugal fluidized bed dryer", *Chemical Engineering Journal,* Vol. 78, pp 107-113.

44. Shing Yang J.Y.H, Liu Y.A and Squires A.M (1987), "Pressure drop across shallow fluidized beds: theory and experiment", *Powder Technology*, Vol.53, pp 79-89.

45. Shu J, Lakshamanan V.I, and Dodson C.E (2000), "Hydrodynamic study of a toroidal fluidized bed Reactor", *Chemical Engineering and Processing,* Vol.39, pp 499-506.

46. Sobrino C, Almendros-Ibanez, Sanchez-Delgado S and Santana D (2009), "Hydrodynamic characteristics of a fluidized bed with rotating distributor", *2007 ECI Conference on The 12th International Conference on Fluidization,* Vancouver, Canada, Vol. RP 4, pp 767-774.

47. Teuro Takahashi, Zennosuke Tanaka and Akira Itoshima (1984), "Performance of a rotating fluidized bed", *Journal of Chemical Engineering of Japan*, Vol. 17, pp 132-136.

48. Upadhyay .N, Saxena S.C and Ravello F.T (1981), "Performance characteristics of multijet tuyere distrbutor plate", *Powder Technology*, Vol. 30, pp 155-159.

49. Valverde J.M, Castellanos A, Quintanilla M.A.S and Gilabert F.A (2006), "Effect of inclination on gas –fluidized beds of fine cohesive powders", *Powder Technology* Vol.182, pp 398-405.

50. Vikram G, Martin H and Raghavan V.R (2003). "The swirling fluidized bed-an advanced hydrodynamic analysis", *Proceedings of the 4th National Workshop on CFB Technology and Revamping of Boilers in India"*, Bangalore Engineering College, June 3-4, pp 9-19.

51. Wen C.Y and Yu Y.H (1966), "A generalised method for predicting the minimum fluidization velocity", *Chemical Engineering Journal,* Vol.12, pp 610-612.

52. Wen C.Y, Krishnan R, Khosravi R and Dutta S (1978), "Dead zone height near the grid of fFluidized bed in fluidization", *J.F. Davidson and D.L Keairns*, Cambridge University Press, pp 32.

53. Whitehead A.S, Davidson J.F and Harrison D, "Problems in large scale fluidized beds in fluidization", *Eds Academic Press*, pp 781.

54. Wilhem R.H and Kwauk H (1948), "Fluidization of solid particles", *Chemical Engineering Progress,* Vol. 44, pp 201-218.

55. Yacono C and Angelino H (1978), "An analysis of the flow between phases in a gas fluidized bed, Fluidization", *Chemical Engineering Science,* Vol.34, pp 789- 800.

56. Ye-Mon Chen (1987), "Fundamentals of a centrifugal fluidized bed", *American Institute of Chemical Engineering Journal*, Vol. 33, pp 273-278.

LIST OF PUBLICATIONS BASED ON THE RESEARCH WORK

1. Josephkunju Paul and Sreejith P.S, "Experimental analysis of conidour type distributors in swirling fluidized bed", *Proceedings of the National Conference on Manufacturing and Management,* M.A College of Engineering, Kothamangalam, 25^{th} and 26^{th} August 2005.
2. Josephkunju Paul and Sreejith P.S, "Role of conidour type distributors in fluidized bed", *Proceedings of the National Conference on Recent Advances in Mechanical Engineering,* P.A College of Engineering, Mangalore, 1^{st} – 3^{rd} December 2005.
3. Josephkunju Paul and Sreejith P.S , " Effect of particle density in swirling fluidized bed – an experimental investigation", *Proceedings of the National Conference on Emerging trends in Physics, Electronics and Engineering science*, J S S College, Mysore, 25^{th} and 26^{th} September 2006.
4. Josephkunju Paul and Sreejith P.S , " Effect of particle size in swirling fluidized bed – an experimental investigation", *Proceedings of the International Conference on advances in mechanical engineering*, Indian Institute of Science, Bangalore, 2^{nd} - 4^{th} July 2008.

CURRICULUM VITAE

1. Name Josephkunju Paul C

2. Date of Birth : 29-05-1959

3. Educational Qualifications

Bachelor degree (B Sc. Engg.) - 1981

Institute	: M. A. College of Engineering, Kothamangalam
University	: Kerala University
Specialisation	: Mechanical Engineering

Masters degree (M Sc. Engg.) - 1984

Institute	: Government Engineering College, Thrissur
University	: Calicut University
Specialisation	: Production Engineering

Doctor of Philosophy (Ph.D)

University	: Cochin University of Science and Technology.

ERRATUM

1. Units of all the terms given in the nomenclature have to be added
2. The notations should have followed a uniform pattern. For example R is used for aspect ratio, the same R is used for critical pressure drop ratio by denoting it as Rp, such overlappings have to be changed. Again, for pressure drop ratio, another notation viz. P.R is used.
3. For certain variables suitable subscripts have to be used. For example, r_i, r_o, r_c may be denoted using subscripts, now, they are not so.

The nomenclature is modified incorporating the above suggestions. The necessary modifications are made in the related equations too.

NOMENCLATURE

Ar	Archimedes number
a_m	Mean area of the bed in mm^2
b_r	Width of the ridge in mm
D	Bed diameter in mm
c	Chord length in mm
d	Orifice diameter in mm
dp	Mean particle diameter in mm
g	Acceleration due to gravity in m/s^2
Ha	Head of air in metres
HDPE	High density polyethylene
Hw	Head causing flow in metres of water
Hs	Static bed height in mm
Ht	Overall bed height in mm
K	Proportionality constant
k	Fraction of bed weight supported by fluidizing gas
L	Effective width of the distributor in mm
l	Length of the slit in mm
Mb	Mass of the bed in gms.
m_{th}	Radial mass flow rate per unit of bed height in kg/s per unit mm
N	Total number of orifices in a distributor plate
n	Number of operating orifices
n_r	Number of rows in the distributor
P	Pressure drop ratio
Pc	Critical pressure drop ratio
Q	Volume flow rate in m^3/s
R	Aspect ratio
Rc	Critical aspect ratio
Re	Reynolds number
r_i	Inner radius of the bed in mm
r_o	Outer radius of the bed in mm
r_c	Radius of the distributor cone in mm

S	Orifice spacing
T	Temperature of air in ^0C
t	Distributor plate thickness in mm
U	Superficial gas velocity in m/s
Uo	Tuyere gas velocity in m/s
U_{mf}	Minimum fluidizing velocity in m/s
Va	Actual volume of air discharged in m^3/s
ΔP_b	Bed pressure drop in mm of water
ΔP_{bmax}	Maximum bed pressure drop in mm of water
ΔP_d	Distributor pressure drop in mm of water
ΔP_{dmin}	Minimum distributor pressure drop in mm of water
ΔPv	Venturimeter pressure drop in mm of water
ϵ	Voidage
ϵs	Static voidage
θ	Inclination of the slit with the horizontal in degrees
μ_f	Dynamic viscosity of fluid in poise
ρ_a	Density of air in gram/cm^3
ρ_b	Bed density in gram/cm^3
ρ_f	Density of fluid in gram/cm^3
ρ_w	Density of water in gram/cm^3
ρ_p	Density of fluidized particle in gram/cm^3
ϕ	Inclination of the slit with the radius of the distributor in degrees
ψ	Proportionality constant
ω_0	Angular velocity in rad/s

4. In the first sentence of the introduction, the word "them" has to be changed by stating its meaning.

The process of imparting fluid like properties to solid particles by forcing them to suspend in a fluid; either gas or liquid is termed as fluidization.

5. All the equations and variables have to be written using an equation editor. Now, equations have been written using text editor, which convey wrong meanings at several locations. For example, the equation 2.7 in page no. 15 seems to be inconsistent.

All the equations and variables rewritten using equation editor are given below.

Page 9
$$\Delta P_d = 1.04\left(\frac{d}{t}\right)^{1/4}\frac{U_0^{\,2}}{2g} \quad (2.1)$$

Page 11
$$U_M = U_{mf}\left\langle 2.65 + 1.24\log_{10}\left(\frac{U_t}{U_{mf}}\right)\right\rangle \quad (2.2)$$

Page 11
$$\frac{\Delta P_d}{\Delta P_b} = C\left(\frac{U_{mf}}{U_M - U_{mf}}\right)^C \quad (2.3)$$

Page 12	$\frac{\Delta P_d}{\Delta P_b} = 2.7\left(\frac{U_{mf}}{U_M - U_{mf}}\right)^{2.32}$	(2.4)
Page 12	$\log\left(1-\frac{n}{N}\right) = -K\left(\frac{U - U_{mf}}{U_{mf}}\right)$	(2.5)
Page 13	$\Delta P_d = C\,U_0^n$	(2.6)
Page 15	$P(\text{pressure - drop raio}) = \frac{P_b}{\langle \rho_p (1\text{-}\varepsilon_s)\ H_s \rangle}$	(2.7)
Page 17	$\Delta Pb = \frac{\left[150(1-\varepsilon)^2 \mu_f m_{fh} \log\left(\frac{r_0}{r_i}\right)\right]}{\left[2\pi\rho_f \varepsilon^3\ dp^2\right] + \frac{\left[1.75(1-\varepsilon) m_{fh}^2 \left\{\frac{1}{ri} - \frac{1}{ro}\right\}\right]}{\left[4\pi^2 \rho_f \varepsilon^3 dp\right]}}$	(2.8)
Page 18	$\Delta Pb_{max} = \frac{W_0 \omega_0^2}{2\pi L}$	(2.9)
Page 18	$\Delta Pb = \frac{kMb}{am} + \psi\omega^2$	(2.10)
Page 21	$\frac{dp\, U_{mf}\ \rho_g}{\mu} = \left\{\frac{(33.7)^2 + 0.0408\ dp^3\ \rho_g(\rho_s - \rho_g) g}{\mu^2}\right\}^{0.5} - 33.7$	(2.11)
Page 21	$U_{mf}, \text{Torebed} = \frac{U_{mf}}{\sin\theta}$	(2.12)
Page 21	$\text{Rep.mf} = \left[28.7^2 + 0.494\ Ar\right]^{0.5} - 28.7$	(2.13)
Page 32	length of the slit $'\ell' = \frac{(r_d - r_c - [3 \times b_r])}{n_r}$	(3.1)
Page 32	The inner radius of the row $'r_i' = r_c + (n_s - 1) \times (\ell + b_r)$ $= 20 + (n_s - 1) \times (24 + 11)$	(3.2)

Page 33	$\cos\delta = 1-\left(\frac{c^2}{2r_i^2}\right) = \frac{\left(2r_i^2-c^2\right)}{2r_i^2}$	
	$\delta = \cos^{-1}\left(\frac{2r_i^2-c^2}{2r_i^2}\right)$	
Page 33	Number of slits in each row $= \frac{360}{\delta}$	(3.3)
Page 34	Percentage area of opening $= \frac{6350.4 \text{ x } 100}{\pi\left(r_d^2 - r_c^2\right)}$	
	$= \frac{(6350.4 \text{ x } 100)}{\pi \text{ x } 88400} = 9.15$	
	Percentage useful area of distributor $= \frac{\left(\left[r_d^2 - r_c^2\right] \text{ x } 100\right)}{r_d^2}$	
	$= \frac{\left(150^2 - 20^2\right) \text{x } 100}{150} = 99.5$	
Page 38	$V_a = 0.021\sqrt{H_a}$	(4.1)
Page 39	$H_a = \frac{H_W \text{ x } \rho_W}{\rho_a}$	(4.2)
Page 44	$\varepsilon = \frac{1-\rho_b}{\rho_p}$	(4.3)
Page 48	$Q = \frac{0.021 \, \Delta P_V \text{ x } \rho_W (273+T)}{\rho_a \text{ x } 273}$	(4.4)
Page 48	$U = \frac{4 \text{ x } Q}{\left(D^2 - D_c^2\right)}$	(4.5)

6. Lot of results have been presented in the form of graphs and the nature variations have been explained. However, it is difficult to extract some useful information by looking from several curves. The conclusions are giving only a qualitative nature of variations. No typical values are given in the conclusions. The candidate has to make a data analysis to present relevant parameters of interest. In my opinion, correlations of bed pressure drop and minimum fluidizing velocity may be obtained in terms of appropriate variables.

The conclusions are revised as per suggestion. Typical values are given wherever possible. Correlations of bed pressure drop and minimum fluidizing velocity are given below in terms of appropriate variables.

CONCLUSIONS

1. The distributor pressure drop decreases with increase in angle of air injection. (At a superficial velocity of 3.8 m/s the rate of decrease is 5.2 mm of water per unit degree in the range of 0^0 to 5^0 ; 4.6 mm of water per unit degree in the range of 5^0 to 10^0 and 1.42 mm of water per unit degree in the range of 10^0 to 15^0 .
2. The pressure head above the distributor increases with increase in angle of air injection.
3. Since the area of opening is independent of angle of air injection, the pressure below the distributor is independent of angle of air injection.
4. The hydrodynamic characteristics of the swirling fluidized bed are quite different from that of a conventional bed.
5. The sequence of bed phenomena in conidour type swirling fluidized bed are packed bed, fluidized bed, swirl motion and vortex motion. (In the case of 7.4 mm acetal particles using 15^0distributor, fluidization starts at a superficial velocity of 1.846 m/s; swirl motion at 1.898 m/s and vortex motion at 2.951 m/s.)
6. It was observed that the minimum fluidizing velocity increases with increase in angle of air injection. (In the case of 7.4 mm acetal particles the rate of increase is 0.029 m/s per unit degree in the range of 0^0 to 5^0 ; 0.018 m/s per unit degree in the range of 5^0 to 10^0 and 0.007 m/s per unit degree in the range of 10^0 to 15^0)
7. In a swirling fluidized bed it was observed that the particle size as well as particle density directly influences the minimum fluidizing velocity.
8. On the basis of experimental results the relationship of minimum fluidizing velocity can be expressed as

$$U_{mf}=\sqrt{g \times d_p}\quad \varepsilon^{8.32}\left(\frac{\rho_p}{\rho_f}\right)^{2.31}\phi^{0.39}\left(\frac{\mu_f}{\rho_p\, g^{1/2}\, d_p^{3/2}}\right)^{0.65}$$

9. It could be observed that, after attaining minimum fluidizing velocity the bed pressure drop increases along the radial direction. (In the case of 7.4 mm acetal particles using 15^0distributor, the rate of increase is 0.115 mm of water per unit mm along the radial distance from 60 mm to 90 mm; 0.005 mm of water per unit mm along the radial distance from 90 mm to 120 mm and 0.044 mm of water per unit mm along the radial distance from120 mm to 150 mm.)
10. The sequence of variation of bed pressure drop with superficial velocity is linear increase, no change followed by sudden increase or sudden decrease. (In the case of 7.4 mm acetal particles using 15^0distributor, the bed pressure drop increases linearly upto a superficial velocity of 1.798m/s; afterwhich it remains almost constant till a superficial velocity of 2.206m/s is reached and thereafter there is a sudden decrease.)

11. At a radial position of 90 mm from the distributor centre, the bed height remains unchanged irrespective of superficial velocity. Hence after attaining minimum fluidizing velocity, the bed pressure drop at this location of the distributor remains almost constant.(In the case of 7.4 mm acetal particles using 15^0distributor this value of bed pressure drop is 22.5 mm of water)
12. Vortex formation in the bed leads to increase in wall friction.
13. The frictional resistance between gas- solid interface increase with decrease in particle size. As a result of this the bed pressure drop increases with decrease in particle size.
14. For any constant weight of the bed the height of the bed increases with decrease in particle density. So the bed pressure drop increases with decrease in particle density.
15. The experimental results lead to the following relationship for bed pressure drop

$$\Delta P_b = d_p \left(\frac{r_d}{d_p}\right)^{20.71} \left(\frac{1}{\varepsilon}\right)^{6.45} \left(\frac{\rho_p}{\rho_f}\right)^{50.29} \phi^{18.74} \left(\frac{\mu_f}{\rho_p\, g^{1/2}\, d_p^{3/2}}\right)^{123.05}$$

16. The fluid drag exerted on the materials decreases with increase in particle size. Hence the superficial velocity required for initiating various bed phenomena increase with increase in particle size. (In the case of acetal particles using 15^0distributor, the superficial velocity required to initiate swirl motion for 3.2 mm particles is 1.548 m/s; for 5.5 mm particles it is 1.747 m/s and for 7.4 mm particles it is 1.898 m/s)
17. Since the drag resistance offered by the materials increases with increase in particle density, the superficial velocity required for initiating various bed phenomena increase with increase in particle density. (In the case of 7.4 mm particles using 15^0distributor, the superficial velocity required to initiate swirl motion for HDPE particles is 1.510 m/s; for nylon particles it is 1.516 m/s and for acetal particles it is 1.898 m/s)
18. The superficial velocity required for various regimes of operation increases with increase in particle size as well as particle density.
19. The radial component of velocity increases with angle of air injection, whereas the tangential component decreases. Since the tangential component is responsible for swirl motion, the superficial velocity required for initiating various bed phenomena increases with angle of air injection. (In the case of 7.4 mm acetal particles, the superficial velocity required to initiate swirl motion for distributor having angle of air injection 0^0 is 1.575 m/s; for distributor having angle of air injection 5^0 it is 1.72 m/s; for distributor having angle of air injection 10^0 it is 1.810 m/s and for distributor having angle of air injection 15^0 it is 1.846 m/s)

7. In chapter 3 on design, the reasons for selecting the dimensions such as outer diameter, ridge width etc may have to be justified or the reasons for fixing such dimensions may be explained.

Survey of the literature reveals that the size of the distributor to obtain uniform air distribution in a swirling fluidized bed is limited to 400 mm. All the previous experimental studies in this area of research were conducted using distributors having diameters of 300 mm. Hence the outer diameter of the distributor was fixed as 300 mm. The width of the ridge was selected as 11 mm for conveniently providing the pressure tappings.