

CFD modelling of an ammonium nitrate fluidised bed: Effect of distribution plate geometry

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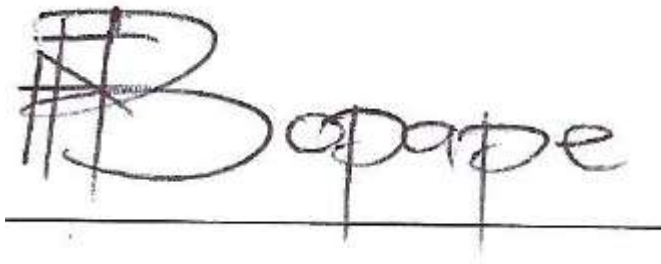
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Declaration

I, Ntsitlola Felicity Bopape. Hereby declare that this dissertation titled: “CFD modelling of an ammonium nitrate fluidised bed: Effect of distribution plate geometry”, submitted in fulfilment of the requirements for the degree in Master of Chemical Engineering, is my own work and that the published work of others has been consulted, and appropriate references have been provided. Furthermore, this work has not been submitted to any other higher education institution and copies submitted for examination are the property of the University.

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A handwritten signature in dark ink, appearing to read 'N.F. Bopape', is written over a horizontal line. The signature is stylized, with the first part being a large, looped 'B' and the last part being 'opape'.

N.F Bopape.

Abstract

Fluidisation, which is a process governed by the suspension of particles in a closed area by blowing air or another medium through the bed of particles, is gaining popularity in industries like pharmaceuticals and mining, to name a few. Contributing researchers developed theory and conducted experiments in an attempt to understand this concept of fluidisation. A wide range of investigations have been conducted, but according to current knowledge, no information was reported on the fluidisation of porous granular ammonium nitrate (PGAN). According to Geldart classification, PGAN falls under Group B particles therefore the behaviour of the particles are expected to be in agreement with that was observed for sand-like particles. The aim of this study is to investigate the effect of different distributor plates on bed hydrodynamics related to an ammonium nitrate fluidised bed in laboratory-scale and pilot-scale setup by: (i) obtaining the characteristics of different suggested distributor plates for use in the fluidisation of PGAN for modelling purposes in a CFD environment, (ii) generating a CFD model for PGAN granules fluidisation that can effectively reproduce the bed pressure drop versus superficial gas velocity curve for different distributor plates using PGAN granules with different particle sizes, and (iii) devising, using the CFD model, means to determine bed density and bed expansion for the (PGAN) fluidisation operation.

The plate pressure drop against superficial gas velocity curve was plotted to calculate Darcy-Forchheimer coefficients which were used in the CFD model environment to characterise the distributor plates. The plate pressure drop was found to increase with increasing superficial gas velocity. In the laboratory-scale investigation, the behaviour of the distributor plates was investigated for both a straight duct configuration and bend duct configuration below the distributor plate, while cone-shaped and column-shaped ducts were used in the pilot-scale setup. The model provided a good representation of the experimental observations for both laboratory and pilot-scale fluidisation setups. The velocity profiles obtained from employing the *porousSimpleFoam* solver with porous media specified by the Darcy-Forchheimer coefficients were used to generate a polynomial function which was employed to specify the boundary conditions of air in the two-phase system.

The bed pressure drop versus superficial gas velocity curves resulting from modelled data follow the trend proposed in the literature, therefore the model can be used to predict the bed pressure drop of a fluidised bed system. Fluidisation initialises quicker in the predictive model than in the experimental setups. The model overestimates the bed pressure drop for particles with small average particle diameters and underestimates the bed pressure drop for particles with higher average particle diameters. The model predicts the bed pressure drop accurately at low superficial gas velocities and the error increases with an increase in superficial gas velocity. The

range of minimum fluidisation velocity for all particle sizes in the laboratory scale fluidised bed was found to be 0.3-0.6 m/s.

Flow patterns were discussed for particles with an average particle diameter of 1.8 mm when fluidising with a superficial gas velocity of 1.1 m/s in a model for 2 seconds. The particles were specified as alpha.particles with a void fraction of 55%. Jetting was observed in a system with a straight duct configuration below the distributor plate. It was concluded that a preferential flow resulted when employing straight ducting configurations as well as bend ducting configurations in the air inlet duct below the distributor plates, both resulting in a high final particle content towards the walls of the duct after a running time of 2s.

The fluidised bed expands in general quicker in the model compared to the experiments, but little to no correspondence in behaviour was observed for a bend ducting configuration system due to preferential flow patterns observed in the system. The bed expansion ratio was found to be directly proportional to superficial gas velocity up to a certain (maximum) extent.

The bed pressure drop measured along the horizontal axis of the fluidised bed in the direction of the position of the air blower at three different probe positions above the perforated plate was used to calculate bed density across the bed. There was no significant correspondence observed between the experimental bed density and the modelled bed density using plates associated with either a straight duct configuration or a bend duct configuration. A higher bed density was observed when measuring directly above the plate due to high particle content at that position. For all the plates in the pilot-scale setup, the superficial gas velocity and probe position have influenced the bed density. An increase in e.g. the height of measurement above the plate resulted in a decrease in the measured fluidised bed density.

Keywords: *CFD-Modelling, fluidised bed, distribution plate geometry, Ammonium nitrate*

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Nomenclature

Abbreviations

AN - Ammonium nitrate

BID - Bend inlet duct

CFB - circulating fluidised bed

CFD - Computational fluids dynamics

CT - Computed technology

CUMT- China University of Mining and Technology

d - viscous loss

DEM - Discrete Elements Method

DF- Darcy-Forchheimer

d_p – particles diameter

FCCUs - Fluidised catalytic cracking units

f - inertial loss

ID - inner diameter

LHB - Left-hand blower

NGD - Nuclear Gauge Densitometry

OF – *OpenFOAM*TM

PID - Primary ideal duct

PGAN - porous granular ammonium nitrate

R - pressure drop ratio

RHB - Right-hand blower

SID - Straight inlet duct

PT - Pressure transmitter

TLW —Towards left wall

TPEF- *twoPhaseEulerFoam*

Symbols

A = total area of the distributor, m²

A_0 = area of the orifice, m²

Ar =Archimedes numbers

C_D = orifice discharge coefficient

D = the viscous loss or impermeability ,1/m²

E = total energy, J

F = fractional free-area

F_{EF} = external force vector, N

F_{IL} = the inertial loss, 1/m

h = vertical depth between two positions, m

H = enthalpy

H_s = static bed height, m

H_{max} =maximum bed expansion, m

H_{mf} = bed expansion due to minimum fluidisation velocity, m

Δh =distance between the two metal piezometric pipes, m

$I_{k,i}$ = momentum forces between phases, N

J_s = dissipation/creation of granular energy.

N_0 = total number of orifice

P = external pressure, Pa

ΔP = pressure differential measured directly by the pressure sensor (Pa)

ΔP_b = pressure drop across the bed, Pa

ΔP_d = pressure drop across the plate, Pa

q = heat flux vector

Re_{mf} =Reynolds number

s = model dependent momentum source term

U_{air} = Superficial gas velocity in OpenFOAM™, (m/s)

U_B = the bubble velocity, m/s

U_{mf} = Minimum fluidisation velocity (m/s)

$\overline{U_r}$ = relative velocity, m/s

v = volume, m³

Δv = change in volume, m³

$|v|$ = velocity magnitude, m/s

V_g = superficial gas velocity with its flow direction normal to the distributor, (m/s)

Δz = thickness of the distributor, m

Greek symbols

α = permeability

α_k = volume fraction of the fluid.

β = Forchheimer coefficient

ε = bed porosity

κ = permeability coefficient, (m²)

Θ = granular temperature, K

$\tau_{k,ij}$ = stress tensor,

μ = dynamic viscosity of the fluid, Pa. s

γ_s = dissipation due to inelastic particle-particle collisions.

ρ = fluid density, kg/m³

ρ_k = density of fluid, kg/m³

Subscripts

k = the solid and the fluid phase

P = particles

G = gas/fluid

1. INTRODUCTION

This chapter serves as an introduction to the study with background information pertaining to fluidisation. The background information of fluidisation and its applicability on broad scale is discussed in order to motivate the study. The capability of a CFD model to predict fluidisation behaviour is discussed. Following the background, the aim and objectives of the study are presented. The frame of the dissertation is outlined at the end of the chapter

1.1 Background and motivation

The fluidisation process was initiated by Fritz Winkler in 1921 (Kafle et al, 2016) promoting the commercial operation of fluidised bed processes in Germany with the introduction of the Winkler coal gasifier in the 1920s (Kafle et al., 2016). This was followed by the use of fluidised catalytic cracking units and fluidised bed reactors in the 1940s for the production of high octane gasoline and phthalic anhydride respectively (Cocco et al., 2014). The United States of America adopted its first application of large-scale fluidised beds in the 1940s. The success in the operation of the first catalytic fluidised bed plant in 1942 was the major breakthrough, paving the path for an additional thirty-one plants during world war II (Vaish, 1988). Douglas Elliot promoted the bubbling fluidised bed in the 1960s (Kafle et al., 2016). Since the 1980s, fundamental theory and practical achievements regarding fluidised bed technology have been acquired by contributing researchers at the China University of Mining and Technology (He et al., 2015).

Fluidisation processes are governed by the suspension of particles through blowing air or another medium through the particle bed in a closed area (Teunou and Poncelet, 2002; Hede, 2013; Suleiman et al., 2013), with the purpose to increase the interaction between the solids and the gas (Bailie et al., 1961). The setup of the fluidised bed is represented in Figure 1-1. Fluidised beds are used mostly for processes such as drying, coating, granulation, combustion and mixing (Lui et al., 2016; Rhodes, 2008). The increase in the employment of fluidised beds in the industry is due to the lower capital cost associated with fluidisation and the ability to perform several processes in one unit, thus saving one or more steps (Hede, 2013). Other advantages of using fluidised beds include superior heat transfer, the ability of a fluidised bed to move solid particles like a fluid and to process materials with a wide range of particle sizes (Cocco et al., 2014).

The state of a fluidised bed depends highly on both the air velocity and properties of the particles (Teunou and Poncelet, 2002; Hede, 2013). Distribution of air in a fluidised bed is the primary factor that influences the quality of fluidisation and therefore, to understand the hydrodynamics and operation of a specific fluidised bed, assessment of the airflow and its distribution is essential.

It can, therefore, be concluded that the performance of a fluidised bed depends highly on the design and operation of air distributors (Depypere *et al.*, 2004; Wormsbecker *et al.*, 2007). The hole patterns on air distributors control the rate of formation of bubbles and bubble sizes (Fasching and Utt, 1982). Guevara (2010) investigated the effect of bed height and static bed density on the hydrodynamics of a fluidised bed, and it was discovered that bed height does not have a significant effect on the minimum fluidisation velocity whereas the density of the material plays a role.

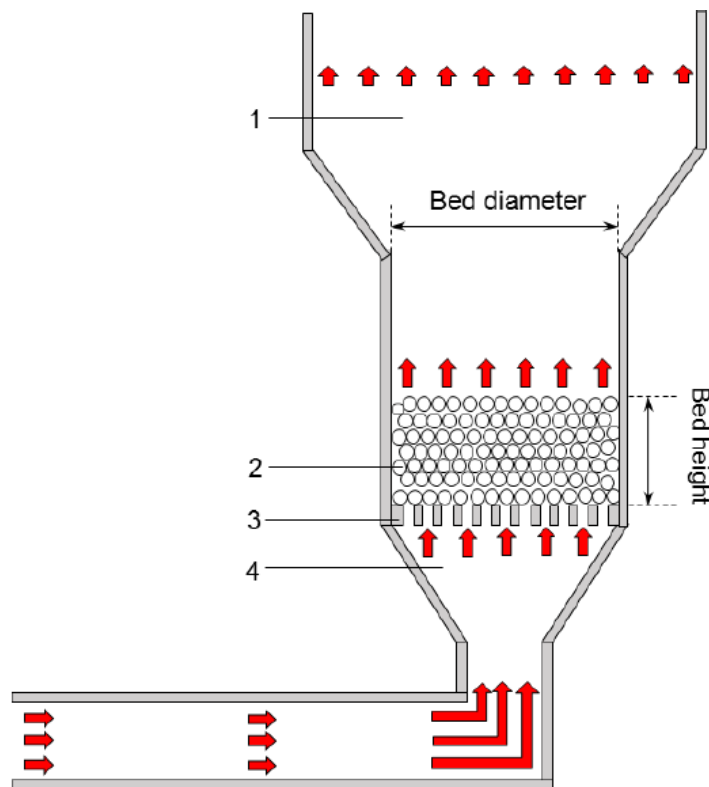


Figure 1-1: Typical fluidised bed setup

(1- Freeboard area; 2-Particles/bed material; 3-Distributor plate; 4- Plenum/wind box)

Compared to other reactor types such as packed bed reactors and stirred tank reactors, fluidised beds are complex to design and operate (Cocco *et al.*, 2014) and are characterised by complex hydrodynamics. Understanding the behaviour of the fundamental parameters, such as size, shape and density of the particles, is crucial in gaining insight into the fluidisation and reactor performance for predicting and calculating the dynamic behaviour of fluidised beds (Saayman *et al.*, 2013; Cocco *et al.*, 2014; Dechsiri, 2003).

Recently, numerical simulations are applied to study fluidisation together with experimental methods (Ma *et al.*, 2016; Zhao and Wei, 2000; Smuts, 2015). Modelling of the bubbling bed reactors originated from the development of catalytic bed reactors (Gogolek, 1998). The use of numerical models is gaining popularity as these produce results that cannot readily be obtained through experimental methods. Models are not constrained by the specific geometry of the apparatus and physical factors can easily be controlled in order to determine the effect thereof on the fluidised bed operation (Smuts, 2015).

Different techniques used for modelling particle-fluid systems include two fluid multiphase, direct numerical simulation, discrete element method and computational fluid dynamics (CFD). Techniques mentioned can also be combined, e.g. CFD and discrete elements method (DEM) coupling can be combined with continuum and discrete methods for improved prediction of fluidisation behaviour (Smuts, 2015). Yang *et al.* (2014) explored the behaviour of gas-solid flow in a spouted bed using a CFD-DEM coupled numerical model. CFD-DEM coupling was also used by Ma *et al.* (2016) to study the fluidisation properties of rod-like particles. Prediction of scaling effects in a fluidised bed used by pharmaceutical industry was investigated by Parker *et al.* (2013) using CFD simulation of three scale processor, Son *et al.* (2005) studied the effect of air distributor on characteristics of glass beads conical fluidised bed and Depypere *et al.* (2004) performed a CFD analysis on the air distribution in a fluidised bed.

Ammonium nitrate is the most widely produced chemical worldwide for fertilisation and explosion purposes (Addiscott, 2005; Speight, 2002). The granulation process is one of the steps in producing ammonium nitrate with fluidisation the most advantageous technique available (European Fertilizer Manufacturer's Association, 2000). In order to reap the full benefits of a fluidised bed, a good understanding of the operation of a fluidised bed is required. As mentioned, distributor plates have a major role in fluidisation as these determine the quality of fluidisation by influencing the air flow rate and direction of the flow. CFD modelling has proven to be a reliable method of predicting the hydrodynamics and behaviour of fluidised bed thus it is employed to investigate the effect of distributor media in the system.

1.2 Aim and objectives

Even with countless studies conducted on fluidised beds (including among others the effects of air distribution on bed density as well as predictive models on the behaviour of fluidised beds), to the researcher's knowledge, there is currently no report of such studies on the fluidisation of porous granular ammonium nitrate (PGAN) particles. The purpose of this study is to investigate the effects that different air distributor plate designs will exhibit on the behaviour of a bubbling PGAN fluidised bed system. Understanding this behaviour has two benefits; firstly, laying the

foundation for predicting fluidisation characteristics which can be modelled mathematically using a CFD platform, and secondly, it provides ground to improve and optimise industrial fluidisation processes. This study also aims to suggest a method of measuring bed density and bed height during the PGAN fluidisation process, which are considered critical operational inputs to optimising the fluidised bed. The primary objectives of this investigation are to:

- I. Obtain the characteristics of different suggested distributor plates for use in the fluidisation of PGAN granules for modelling purposes in a CFD environment.
- II. Generate a CFD model for PGAN granules fluidisation that can effectively produce the bed pressure drop versus superficial gas velocity curve for different distributor plates using PGAN granules with different particle sizes.
- III. Devise, using the CFD model, means to determine bed density and bed expansion for the PGAN fluidisation operation.

1.3 Outline of the dissertation

This report consists of five chapters of which the details are outlined below:

Chapter 1: Introduction - This is the introductory chapter that outlines the background and motivation of the study together with the issues to be addressed (aim and objectives).

Chapter 2: Literature review - This chapter summarises the detailed literature research of the necessary fundamentals and concepts significant to the study. This chapter covers published information about fluidisation processes in general, including hydrodynamics, i.e. bed pressure drop, minimum fluidisation velocity, bed height and bed density. Different kinds of distributor plates are discussed in detail accompanied by diagrammatic representations thereof. Numerical simulation methods receive attention in this chapter with a focus on the Eulerian method that birthed CFD modelling. Different solvers employed in CFD modelling are also elaborated on.

Chapter 3: Experiment and Modelling -This chapter consist of a description of the material particles, experimental setup and experimental procedure used for the experimental phase of the study. Assumptions made during the course of the study are also listed and discussed in this section. The modelling environment and information used to describe the (i) various distributor mediums employed in this study, and (ii) the actual fluidisation of the PGAN granules are also covered. Various considerations in the modelling approach receive attention in this section.

Chapter 4: Results and discussion -This chapter covers the presentation, description and discussion of the results obtained during this investigation. The results obtained from the

predictive model are validated with experimental results to compare how accurately the model will predict the behaviour of a fluidised bed system. The characterisation of four distributor plates in the laboratory scale setup of which three were also used for investigation in the pilot scale setup is discussed followed by comparison of the behaviour of distributor plates in both scale setups. The ability of the model to predict the experimental observations by making use of the bed pressure drop against superficial gas velocity curve which receives attention for different distributor plates using particles with different sizes is elaborated. This is followed by the discussion on flow patterns in a primary ideal duct setup, straight inlet duct setup and bent inlet duct setup. Bed expansion during fluidisation for a straight inlet duct setup is also discussed followed by a discussion of the bed density in both laboratory and pilot-scale fluidised bed setup.

Chapter 5: Conclusion and recommendations - At this stage of the report, conclusions drawn are listed from the results obtained. This chapter also covers the recommendations for future studies.

2. LITERATURE REVIEW

The aim of the study is to investigate the effect of the air distribution plate on the behaviour of the fluidised bed system. In order to successfully complete the study, an understanding of the fundamentals of fluidisation and the use of numerical models to predict fluidisation and fluidised bed hydrodynamics is required. This chapter summarises the necessary information for understanding the process, including the fundamental theory, measurement techniques and numerical simulations for fluidised bed systems by addressing two parts: (i) the fundamentals of fluidisation and (ii) numerical modelling. The first part of the chapter (section 2.1) provides the general overview while section 2.2 covers the fundamentals of fluidisation, fluidisation regimes, hydrodynamics and mediums of air distribution. The second part (section 2.3) focuses on the use of computational fluid dynamics in the OPENFOAM™ platform to predict fluidisation behaviour. Applicable solvers receive attention in this section.

2.1 General overview of fluidisation

Fluidisation is gaining popularity in the industry with various emerging processes, such as biomass and coal gasification, chemical looping, dehydrogenation of propane, synthesis of polycrystalline silicone and gas-to-solid conversion because of its ability to move solid particles in a fluid-like fashion (Cocco *et al.*, 2014; Pecora & Parise, 2006; Lundberg & Halvorsen, 2008; Benzarti *et al.*, 2012; Halvorsen *et al.*, 2006). Fluidised beds are used for processes such as catalytic reaction, incineration of waste, water treatment, granulation, drying and coating of particles as well as crystallisation (Rasteh *et al.*, 2015). Some of the advantages offered by fluidised bed systems include the excellent heat transfer, which can be five to ten times greater compared to the packed bed and the ability to process particles with a wide size distribution (Cocco, 2014). Fluidised beds are used to ensure adequate mixing and phase interactions. (Kelkar *et al.*, 2016; Rasteh *et al.*, 2015).

2.2 Fluidised bed operation fundamentals

2.2.1 Fluidisation regimes

During fluidisation of particles, the bed behaviour changes with changes in fluidised medium velocity as well as with different gas- and particle properties, resulting in a number of fluidisation regimes (Dechsiri, 2004). The flow regime is one of the factors that affect the performance of multiphase reactors (Nedeltchev, 2015). According to Nedeltchev (2015), the investigations dedicated to identifying the flow regimes in reactors have been active for the past 50 years.

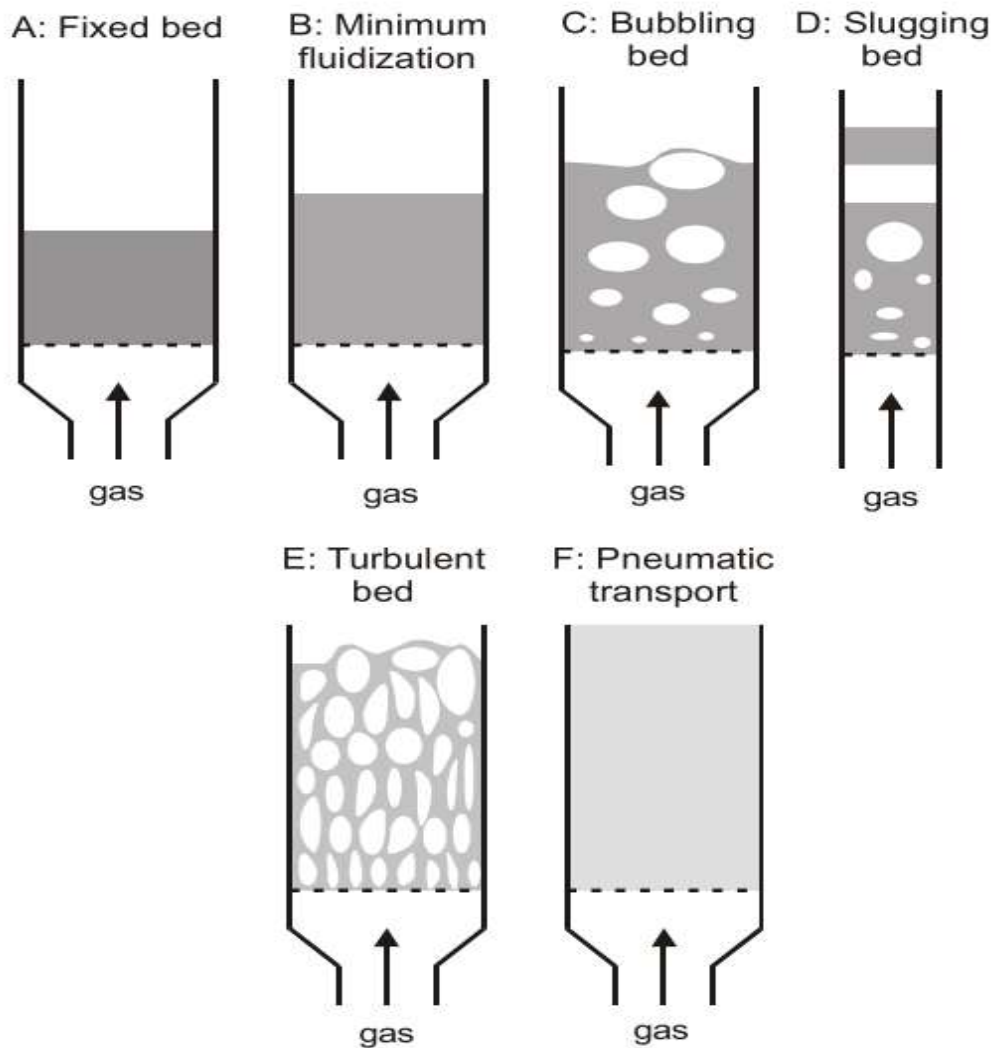


Figure 2-1: Representation of different fluidised bed regimes (Adapted from Dechsiri,2004)

Figure 2-1 depicts the behaviour of particles in a fluidised bed as the flow of the fluidising medium is increased. When the gas flow rate through a bed of particles is low enough, a fixed bed (A) results as the fluid merely penetrates through the void spaces between stationary bed particles causing few vibrations within the same height (Kunii & Levenspiel, 1991;2; Dechsiri, 2004). As the flow rate increases, the particles start to move apart, vibrates and move around in the regions resulting in an expanded bed (Kunii & Levenspiel, 1991;2). At higher fluid velocity, the upward flowing gas imposes a higher drag force to overcome the downward gravitational forces and the particles become suspended. The bed particles are then considered to be fluidised and referred to as an incipient or minimum fluidised bed (Kunii & Levenspiel, 1991:71; Dechsiri, 2004; Rhode, 2008; Cocco *et al.*, 2014).

In a gas-solid system, an unstable behaviour with bubbling and channelling of gas is observed when the velocity of the fluid is increased higher than the minimum fluidisation velocity. Part C in the figure represents a state of fluidisation wherein gas velocity has increased beyond minimum fluidisation velocity resulting in the formation of bubbles. Increasing the velocity even further will result in coalescing and growth of the bubbles formed in the fluidised bed (Dechsiri, 2004). When the flow rate is increased further, the violence of agitation, deformation of bubbles and vigorous movement of solid particles increase (Kunii & Levenspiel, 1991:3). When the terminal velocity of the particles is exceeded due to an increase of gas velocity, a turbulent bed, depicted in E in the figure, results. In this case, gas voids of various sizes and shapes and turbulent motion of solid clusters are observed instead of bubbles. Increasing the gas velocity further results in entrainment of the particles into the gas stream (Dechsiri, 2004).

Nedeltchev (2015) developed a method of identifying flow regimes in bubble columns and fluidised beds using entropy (and information entropy) extracted from computed tomography (CT) and nuclear gauge densitometry (NGD) respectively. Chen *et al.* (2017) studied the flow regimes and transitions using a mixture of cylindrical particles and silica sand in a fluidised bed. Six regimes (fixed bed, bubbling, transition, partial, complete fluidisation(s) and unable to fluidise) were identified and described by means of photographic images and schematic diagrams.

2.2.2 Fluidisation hydrodynamics

a) Pressure differentials

Fluctuations of pressure in a fluidised bed result from the action of bubbles related to bed movement (Leckner *et al.*, 2002). This chapter focuses on the various measurable pressure differentials involved in the fluidised bed system, including the bed and distribution medium.

Bed differential pressure

Upon the attempt of Hiraki (1961) to produce a degree of the quality of fluidisation, it was observed that the quality is related to the closeness between the measured values of bed differential pressure and the weight of the particle per bed section measured. Paiva *et al.* (2009) compared the resulting bed differential pressure of Group C particles with the ideal values obtained from computation based on the knowledge of the initial mass of particles charged in the bed.

When minimum fluidisation velocity is reached in a bed, the pressure drop per unit length is related to the weight of particles. The relationship can be represented by the equation 2-1 (Shaul *et al.*, 2014). Figure 2-2 is a schematic representation of bed differential measurement points in a fluidised bed system.

$$\frac{\Delta P}{H} = (1 - \varepsilon)(\rho_p - \rho_g)g \quad (2-1)$$

In Equation 2-1, the ΔP is the pressure drop, H the initial height of the bed, ε is the void fraction, ρ_g and ρ_p is the density of the gas/ fluid and particles respectively. g is the gravitational acceleration.

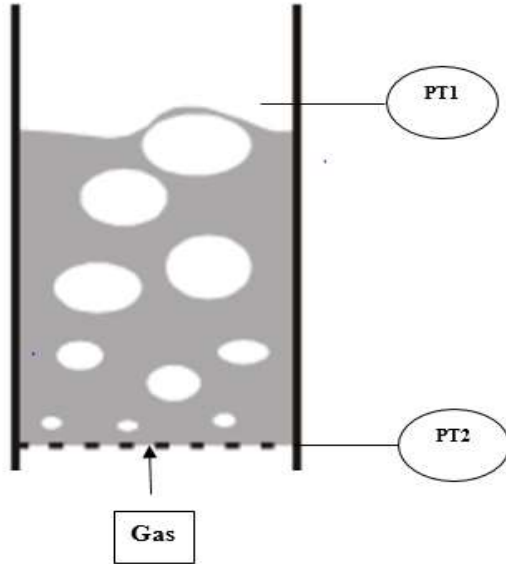


Figure 2-2: Bed pressure drop measurement points setup

Padhi *et al.* (2016) carried out experiments to predict the bed pressure drop in a conical fluidised bed by investigating the effects of superficial liquid and gas velocity, static bed height and the average size of the particles on bed pressure drop. Due to the increase in gas velocity that tends to increase the gas hold-up and decrease bed material density, it was found that an increase in superficial gas velocity along the axial direction of the bed results in a decrease in bed differential pressure. Increasing the particle size, resulted in a consequent increase in the void space in the bed material and an expected decrease in bed differential pressure. For counterbalancing the weight of the bed, an increase in differential pressure is required, and the results showed that bed pressure drop increases as expected with an increase in the initial static bed height. Lastly, it was observed that the bed pressure drop is directly proportional to the cone angle.

The hydrodynamic characteristics in a tapered fluidised bed packed with a homogeneous binary mixture of two different particles were investigated by Sau *et al.* (2008), focusing on maximum pressure drop across the bed and critical fluidisation velocities of the system. It was found that the pressure drop increased with increasing superficial velocity at a stagnant bed height of 14cm.

Distributor medium differential pressure

Bed pressure drop is one of the variables that influence the design of the distributor in terms of the required differential pressure. Other variables include weight, height, powder type, size distribution and density. Additional considerations for distributor design include distributor geometry, distributor thickness and open area fraction to guarantee undisturbed flow rate of the gas through the distributor into the bed of particles. A high distributor differential pressure is required to allow for homogeneous gas flow into the entire fluidised bed (Depypere *et al.*, 2004).

Higher pressure drops across the distributor will result in excess power consumption and construction expenses, hence the need for an optimisation procedure to choose a suitable distributor. Combination of Darcy's law with orifice theory yield typical pressure drops across perforated plates and tuyere distributors according to equation 2-2 for porous plates and equation 2-3 for perforated plates and tuyere distributors (Depypere, *et al.*, 2004):

$$\Delta P_d = \frac{\mu \Delta z}{\alpha} V_g \quad (2-2)$$

$$\Delta P_d = \frac{\rho}{2C_d^2 F^2} V_g^2 \quad (2-3)$$

With:

$$F = \frac{n_o A_o}{\alpha} \quad (2-4)$$

For the above equations, ΔP_d is the pressure drop across the plate, V_g is the superficial gas velocity with its flow direction normal to the distributor, μ the dynamic viscosity of the fluid, Δz represents the thickness of the distributor, α is the permeability of the porous media, ρ symbol of fluid density, C_d is the orifice discharge coefficient, F is the fractional free-area, N_o is the total number of orifice, A_o represents area of the orifice.

In literature, most studies regarding the influence of a gas distributor were conducted at low fluidisation velocities. The minimum distributor differential pressure required was examined in one investigation to ensure that gas is distributed uniformly (Kunii & Levenspiel, 1991:103). While investigating the influence of the distributor type at low velocities, Geldart & Kelsey (1968) observed that changing the distributor pressure drop by adding or removing porous material sheets has a direct influence on the bubble size observed. When working at high fluidisation velocities, five different perforated plates with different pitches and equal orifice diameters were used and three bubbling regimes, which were later related to the plenum pressure fluctuations, were found. Paiva *et al.* (2004) studied the influence that the distributor plate has on the bottom zone of a fluidised bed when the transition from bubbling to turbulent fluidisation was approached.

Higher pressure drop values were produced for distributor plates with higher open area in the bottom zone of a narrow fluidised bed.

Dong *et al.* (2009) found that the pressure drop across the distributor decreased with decreasing superficial gas velocity and increasing perforated ratios when investigating the effect of the perforation ratio has on the characteristics of gas-solid fluidisation. Chyang & Huang (1991) studied the discrepancies between the distributor pressure drop measured with and without the presences of bed material. The results indicate that the pressure drop measured across the distributor is dependent on the position of measurement taps. Contradictory to the distributor differential pressure being measured in the presence of bed material, the measured pressure drop across the dry plate was the same at any point where the measurement was taken, proving that there is no maldistribution (which is equivalent to a non-uniform distribution of the fluidising medium into the fluidised bed of particles) of air in the plenum.

The characteristics of the combined particle bed pressure drop and air distributor pressure drop influence the uniformity of fluidisation or homogeneous flow of air through the distributor into the system. This requirement is defined as the ratio of distributor differential pressure to bed differential pressure and is denoted with the letter R. Values of R proposed in the literature varies between 0.1 and 1 for deep and shallow beds respectively in bench-scale experiments (Depypere *et al.*, 2004). According to Kunii & Levenspiel (1991:103), the distributor to bed pressure drop ratio rule of thumb is represented by;

$$R = \frac{\Delta P_d}{\Delta P_b} = 0.2 - 0.4 \quad (2-5)$$

where ΔP_d and ΔP_b are the pressure drops across the distributor and the bed respectively. According to Depypere *et al.* (2004), R decreases with increasing superficial gas velocity and the ratio itself cannot give conclusive results with regards to the uniformity of air distribution into the fluidised bed.

According to Geldart & Kelsey (1968), an increased pressure drop ratio is necessary to prevent air maldistribution that results in dead zones in the fluidised bed system. With small or negligible pressure drop ratios as represented in equation 2-6 below, these authors observed severe channelling. Luo & Zhao (2002) proposed a criterion represented by equation 2-7 for stable and uniform fluidisation (meaning that the air is properly distributed into the fluidised bed system).

$$\frac{\Delta P_d}{\Delta P_b} = 0.001 - 0.005 \quad (2-6)$$

$$\Delta P_d > \Delta P_b = \frac{\Delta P_b^2 + \Delta P_d^2}{\Delta P_b + \Delta P_d} \quad (2-7)$$

Paiva *et al.* (2009) concluded that a higher pressure drop ratio isn't necessary for all fluidised bed systems as it might also result in maldistribution in the lower parts of the bed. Figure 2-3 is a schematic representation of maldistribution. Maldistribution occurs when dead zones are observed close to the distributor plate when the velocity of air exceeds the minimum fluidisation velocity. The pressure drop ratio has been used as a simple criterion of distributor plate designs. Different values of R have been proposed by authors in the past 50 years although a critical value is not recorded (Javier, 2015). Authors proposed a range of 0.02-1 while Kunii & Levenspiel (1991:102) proposed the 0.2-0.4 range.

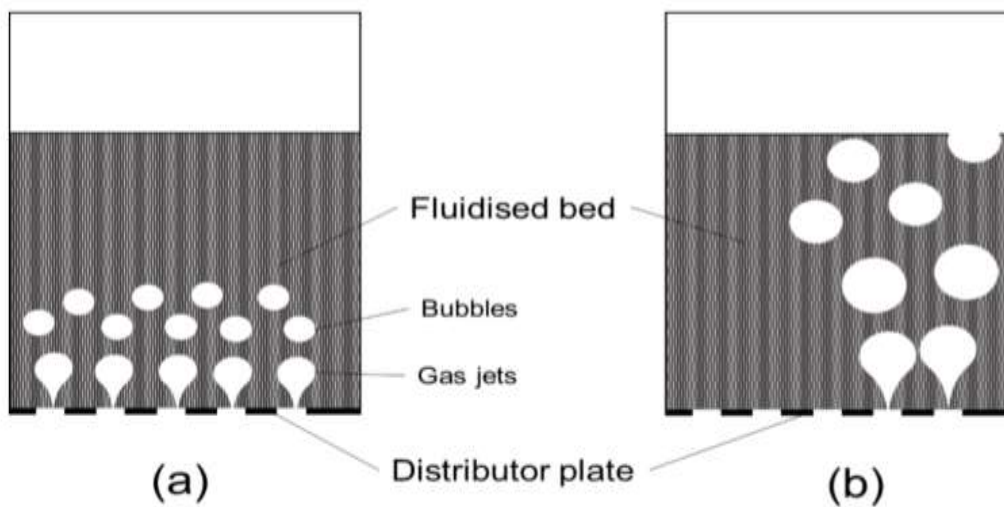


Figure 2-3: Definition of gas maldistribution (a) Uniform gas distribution (b) Maldistribution in a fluidised bed (Adapted from Thorpe, 2002)

b) Minimum fluidisation velocity

The superficial gas velocity at the onset of fluidisation of the bed is called minimum fluidisation velocity (U_{mf}) (Dechsiri, 2004; Hede, 2013). Minimum fluidisation velocity is one of the variables that plays a role in the design and operation, and determines the smooth operation of a fluidised bed (Ma *et al.*, 2013). According to Kunii & Levenspiel (1991:1), fluidisation occurs when the drag force of the gas moving upward is equal to the weight of particles in the system. When the flow rate of the gas is increased in a fixed bed, the pressure continues to rise due to the drag force until a critical value at the minimum fluidisation velocity is reached (Kafle *et al.*, 2016). The effect of the bed particles temperature and the particle size distribution on U_{mf} was explored by Ma *et al.* (2013). It was concluded from the results that for particles with a wide particle size distribution, the U_{mf} decreases with an increase in bed temperature. Sufficient care is critical when measuring

U_{mf} at high temperatures. U_{mf} can be determined by the Ergun equation (given as equation 2-8);

$$Re_{mf} = \sqrt{c_1^2 + c_2 Ar} - c_1 \quad (2-8)$$

Here, Re_{mf} and Ar are the Reynolds and Archimedes numbers respectively, which are represented by:

$$Re_{mf} = \frac{\rho_G U_{mf} d_p}{\mu_G} \quad (2-8.1)$$

$$Ar = \frac{\rho_G \Delta \rho g d_p^3}{\mu_G^2} \quad (2-8.2)$$

Here ρ_p , ρ_G , d_p , μ_p and g are particle and gas density, diameter of the particles, the viscosity of the gas and gravity, respectively. Value pairs of the empirical constants c_1 and c_2 are available in the literature, and new values pairs have been proposed which are particle-shape and species dependent (Lim *et al.*, 1995). Empirical correlations to precisely predict U_{mf} based on these value pairs have been proposed by many studies and are summarised in a table compiled by Ma *et al.* 2013 (presented as Table 2-1).

According to Dechsiri (2004), another widely used empirical expression was obtained by, provided in equation 2-9;

$$U_{mf} = 7.90 \times 10^{-3} \overline{d_p^{1.82}} (\rho_s - \rho_G)^{0.94} \mu_G^{-0.88} \quad (2-9)$$

Graphically, the minimum fluidisation velocity can be determined from the pressure drop versus superficial gas velocity experimental curves. The minimum fluidisation velocity is obtained when the pressure drop decreases and voidage increases when the fixed bed unlocks. In Figure 2-4, the point where U_{mf} is reached it is highlighted with a circle.

One of the main hydrodynamics in a fluidised bed system is the differential pressure across the bed of solid particles. The pressure difference across the bed was determined by calculating the difference between the observed or modelled pressures at the bottom and directly above the bed for different superficial gas velocities. It should be noted that this pressure differential was calculated over the bed of particles alone, thus it excludes the pressure differential of the distribution plate. Figure 2-4 shows a qualitative study of a typical bed pressure drop that can be observed from the model with a visual representation of particle behaviour at different velocities.

Table 2-1: List of typical correlation to predict minimum fluidisation velocity.

(Adapted from Ma *et al.*, 2013)

Researchers	correlation	Materials/particles	Particle diameter (mm)	Temp(K)	Other conditions
Wen & Yu	$Re_{mf} = (33.7^2 + 0.04084Ar)^{0.5} - 33.7$	Various	0.04-20	Ambient	$10^{-3} < Re_{mf} < 4 \times 10^{-3}$
Bourgeois & Grenier	$Re_{mf} = (25.46^2 + 0.0382Ar)^{0.5} - 25.46$	spherical	0.086-25.1	Ambient	$10^8 < Ar < 5 \times 10^8$
Saxena & Vogel	$Re_{mf} = (25.282 + 0.0571Ar)^{0.5} - 24.28$	-	0.088-1.14	Ambient	$6 < Re_{mf} < 102$
Nakamura et al	$Re_{mf} = (33.9532 + 0.0465Ar)^{0.5} - 33.953$	Glass beads	0.2-4	280-800	$0.08 < Re_{mf} < 1360$
Zheng et al.	$Re_{mf} = (18.752 + 0.03125Ar)^{0.5} - 18.75$	Glass beads/river sand	-	293-973	None
Doichev & Akhmakov	$Re_{mf} = 0.00108Ar^{0.947}$	Glass beads	0.09-2.2	Ambient	$\rho_p = 2650 \text{ kg/m}^3$
Ryu et al.	$U_{mf} = 2.997 \times 10^{-3} d_p^{1.636} (\rho_p - \rho_G)^{1.128} \frac{g}{\rho_G} \mu^{0.446}$	NiO/bentonite	0.181	298-1273	None
ZJC	$U_{mf} = 0.294 \times \frac{d_p^{0.584}}{v_G^{0.056}} \times \left(\frac{\rho_p}{\rho_G} - 1\right)^{0.528}$	Stone like coal	-	298-1073	$Ar = (2 - 700) \times 10^4$
NWC	$Re_{mf} = 0.129Ar^{0.54}$	Coal gangue	-	Ambient	None

HZC	$Re_{mf} = 0.0968Ar^{0.535}$	Stine-like coal ash	0.5-8	Ambient	$Ar = (2 - 1800) \times 10^4$
Chen & Zhendong	$Re_{mf} = 0.01036Ar^{0.7107}$	River sand	0.18-0.45	303-1073	$Re_{mf} < 5; Ar < 5000$
Wu & Bayens	$Re_{mf} = 7.33 \times 10^3 \times 10^{(8.241gAr - 8.81)0.5}$	River sand	0.134-0.939	293-673	None
Subramani et al.	$Re_{mf} = \frac{Ar}{1502}$	Limestone/River sand/ironstone	0.128-0.2	298-973	None

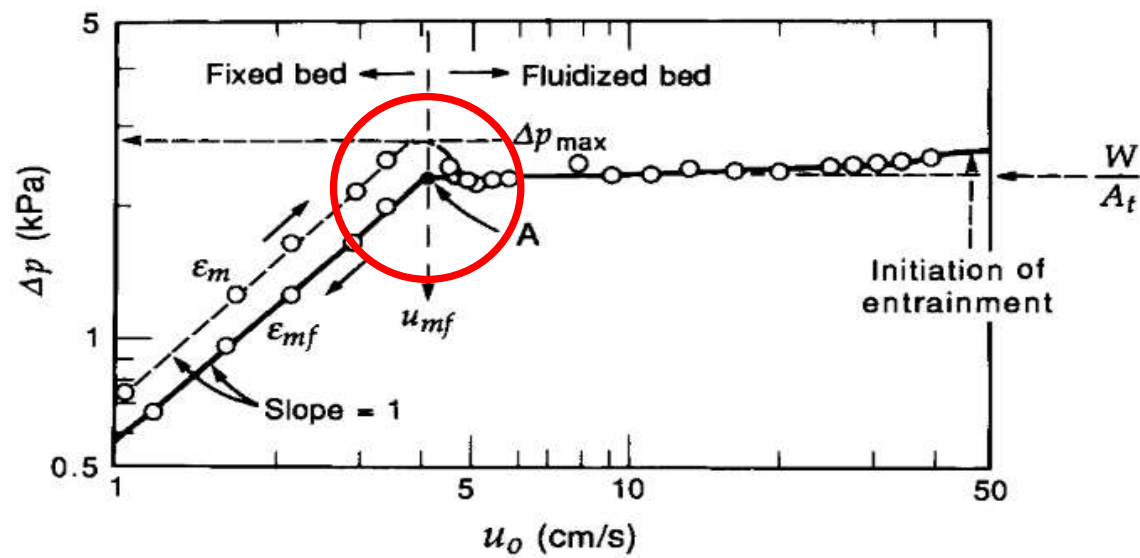


Figure 2-4: Pressure drop against superficial gas velocity curve.

(Adapted from Kunii & Levenspiel, 1991:72)

Figure 2-5 depicts the typical pressure drop against superficial gas velocities accompanied by representations of the particle movements in the system. The red and blue portions in the visual representations (a), (b), (c), (d) and (e) represent areas of high and no content of bed particles in the system. At a low superficial gas velocity, air flow through the fluidised bed system is not sufficient to suspend any of the particles, keeping the bed fixed. For these lower superficial velocities, the bed pressure drop across the system is approximately proportional to the superficial gas velocity.

As the velocity increases, the particles begin to collide with each other, resulting in a simultaneous particle-particle interaction and particle-air interaction in the system. The proportionality between pressure drop and superficial gas velocity continues until a maximum bed pressure drop (ΔP_{Bmax}) is reached. The voidage also increased as the fixed bed unlocked and the bed pressure drop started to decrease. When increasing the velocity even further, particles start to circulate, following a particular flow pattern in the system. At this point, a minimum fluidisation velocity is reached.

With superficial gas velocity beyond the minimum fluidisation velocity, the bed expands as some gas bubbles start to form. Non-homogeneity results in the system. Visibility of the bubbles becomes clear and continue to rise as the superficial gas velocity increases, but during this phase of fluidisation, the bed pressure drop remains practically unchanged. An eruptive fluidisation will result as the velocity is increased further where bubbles break and a vigorous circulation of particles follows.

A number of researchers have investigated the point of minimum fluidisation for various materials consisting of different particle sizes, bed operating temperatures and some conditions specific to an investigated application. It is observed that the prediction of the point of minimum fluidisation culminated in every case in an empirical relationship (correlation), which is indicative that the point of minimum fluidisation is strongly dependent on the material type, material properties, operating temperature and other possible factors like a fluidising medium. A summary of some correlations published by various authors is provided in Table 2-1, showing the empirical nature thereof.

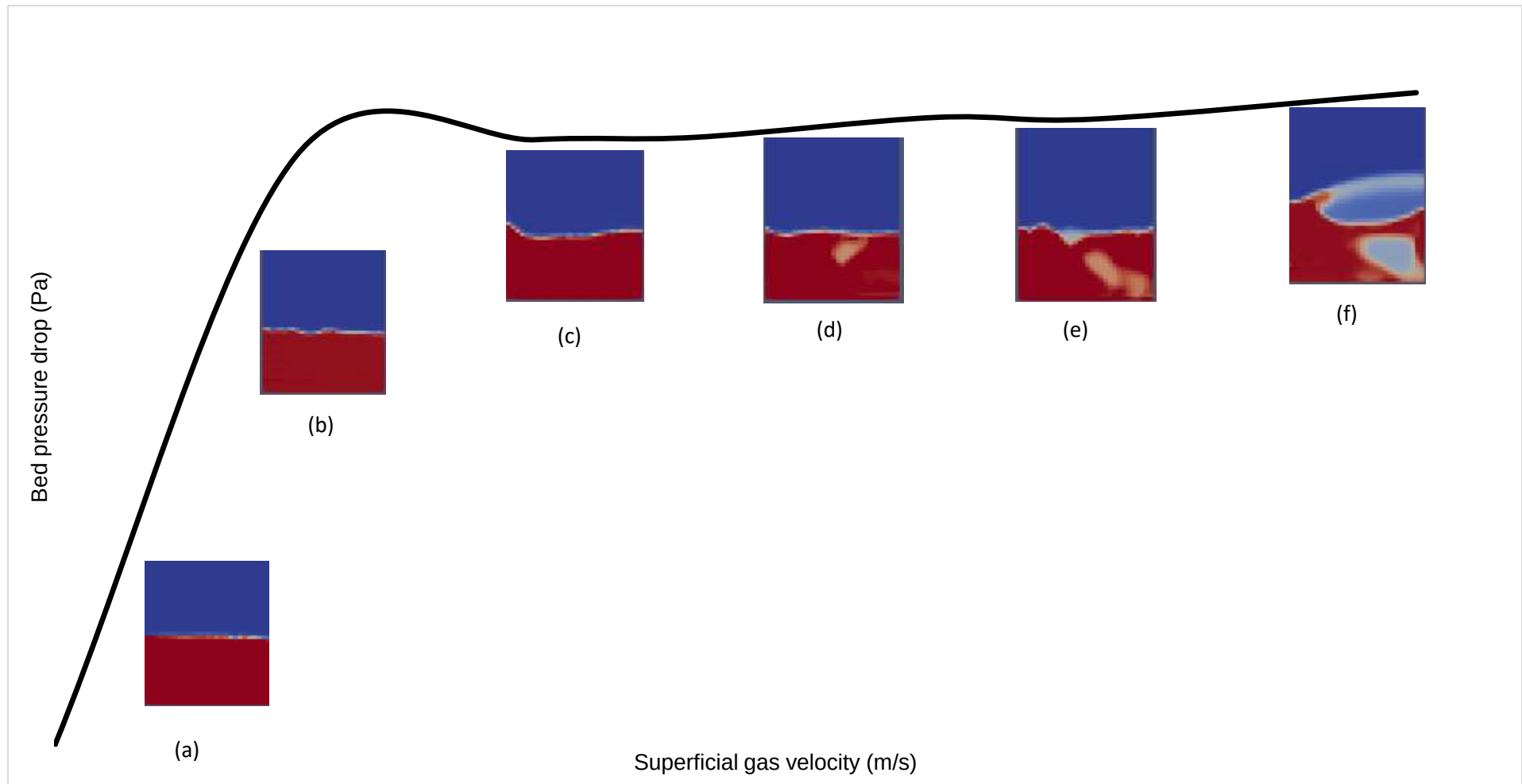


Figure 2-5: A typical bed pressure drop obtained from the Computational fluid dynamics (CFD) fluidised bed model

c) Bed density

Measurements of bed density obtained from experimental methods can be used to deduce bed hydrodynamic conditions which are essential for models (Geldart, 2004). The characteristic of bed density distribution is studied to provide theoretical structural design support to the fluidised bed, supporting *inter alia* selection of devices and control mechanisms (He *et al.*, 2013). Density fluctuations in a fluidised bed can be determined by a radiation attenuation method, capacitance probe method and detecting, measuring and recording the non-attenuated portion of the γ radiation (Bailie, 1961). Hauschild & Knochel (1995) introduced a suitable technique that employs a microwave locating reflectometer for density profile measurements.

Bed density data obtained through the mentioned measurement methods implemented during the experimental phase have the importance of being considered as reference values (He *et al.*, 2013). Pigford & Baron (1965) examined the unsteady motions equations of the particles and the uniform solids concentration in a fluidised bed and found that the small density disturbances exhibit an exponential growth as it moves upward through the bed.

Luo *et al.* (2006) proposed a regression model and bed density calculation method based on the concept of the equilibrium principle of mass. Equation 2-10 summarises this approach:

$$\rho_G = \frac{\rho_b}{1+m} \quad (2-10)$$

where $m = \frac{\Delta V}{V}$ is the expansion ratio of the fluidised bed where v is the initial volume before fluidisation and Δv is the change in volume in the bed during fluidisation

He *et al.* (2013), He *et al.* (2016), Jiang *et al.* (2016), He *et al.* (2017) used the hydrostatic head difference between two positions to determine the mean bed density using equations 2-11, 2-12 and 2-13 respectively based on the concept represented by Figure 2-6;

$$\rho = \frac{\Delta P}{gh} = \frac{P_2 - P_1}{gh} \quad (2-11)$$

$$\rho = \frac{\Delta P}{g \times \Delta h} \quad (2-12)$$

$$P = \varepsilon \rho g + (1 - \varepsilon) \rho_b \quad (2-13)$$

ΔP is the pressure differential measured directly by the manometer, h is the vertical depth between two positions, ε is the bed porosity and Δh is the distance between the two metal piezometric pipes. The distribution of density directly represents the stability of the fluidised bed. High-grade fluidisation is indicated by a uniform density throughout the bed (Jiang *et al.*, 2017).

According to He *et al.* (2013), the stability of axial positions density is better than the stability of density along varied bed height, therefore emphasis should be on maintaining as little as possible density fluctuations with bed height.

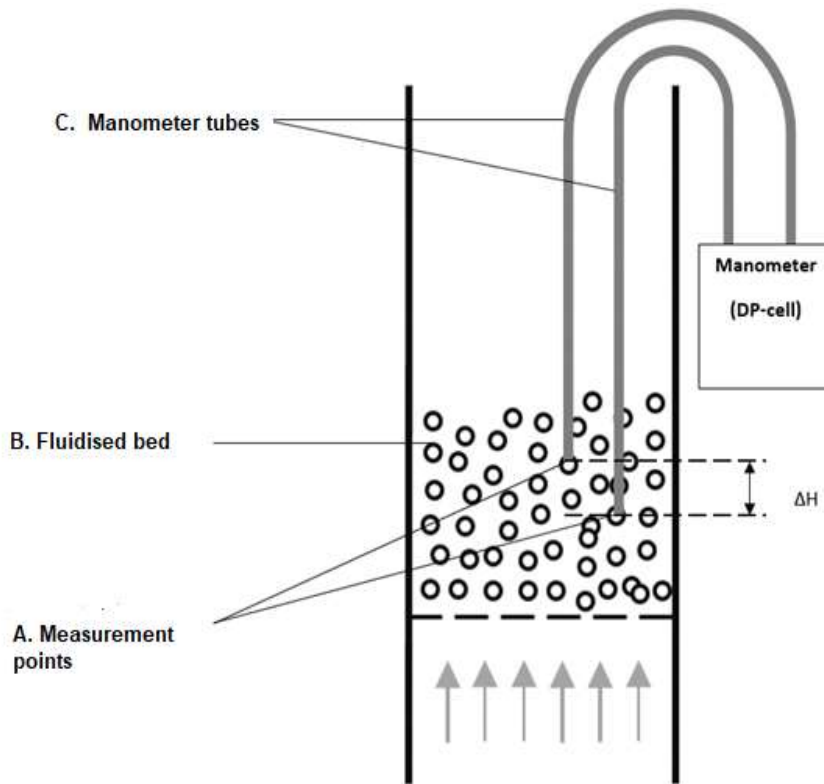


Figure 2-6: Bed density measurement setup

d) Bed expansion

Quality and performance of fluidisation depend on the gas distribution and bubble size, where large bubbles decrease the gas-solid contact. Due to the random generation and eruption of the bubble, it is difficult in bubbling fluidisation to describe the flow of gas (Horio *et al.*, 1980) (Vakhshouri, 2008). In a fluidised bed, bubbles form at the point where gas velocity has increased beyond the minimum fluidisation velocity (Cocco *et al.*, 2014).

Saxena *et al.* (1979) and Fan *et al.* (1981) observed that when a variation of the open area of the distributor was allowed by using several slot screens and perforated plates with varied hole diameters respectively, bubble size increased with a decrease in pressure drop across the distributor. Hatate *et al.* (1991) found that different distributors with the same open area have no influence on the bubble size, the influence is due to different numbers and size of holes. Besides the number and size of holes on the distributor, it is known that the characteristics of the bubble are determined by the distributor differential pressure (Depypere *et al.*, 2014). If the height to

diameter ratio of the fluidised bed is high enough, the bubble size may become almost the same size as the bed diameter. Due to the bubble coalescence, the mean bubble population size above the distributor increases with height (Dechsiri, 2004).

Measuring pressure changes or optical attenuation used to detect bubbles in a fluidised bed was found to be insufficient to explain bubble behaviour, thus, Lim *et al.* (1995) considered several approaches to determine bubble shape in greater depth. According to Dechsiri (2004), of the many researchers who attempted to predict the bubble size, Geldart (1972) used the expression proposed by Kato & Wen (1969) in conjunction with his empirical correlation for the bubble growth with bed height:

$$D_g = \frac{1.43}{g^{0.2}} \left(\frac{(U - U_{mf}) \pi d_{bed}^2}{4N_0} \right)^{0.4} + 2.05(U - U_{mf})^{0.94} h \quad (2-14)$$

The degree of dense phase expansion and bubble holdup are the two factors that influence the overall bed expansion. The knowledge of the extent of fluidised bed expansion is necessary for calculating changes in the system. The expansion of the dense phase is reported, with little data to be small in Geldarts' group B and D solid particles. The gas flow rate is the factor which bubble holdup mainly depends on (Geldart, 2004). The bed expansion in a fluidised bed system is impacted by the influence of the drag model on the flow of granular phases (Lundberg & Halvorsen, 2008). The method used to determine the bed density of fluidised bed of porous powders depends on the bed expansion measured. A modified Ergun equation (Eq. 2-15) accurately predicts the bed expansion of laminar regimes in both fixed and suspended beds (Yang, 2003).

$$\Delta P = \left(\frac{17.3}{(Re)_p} + 0.336 \right) \frac{\rho_f U^2 L}{d_p} (1 - \varepsilon) \varepsilon^{-4.8} \quad (2-15)$$

When the bubble velocity of a bubble with constant size is known, maximum bed expansion can be calculated using equation 2-16.

$$\frac{(H_{max} - H_{mf})}{H_{mf}} = \frac{(U - U_{mf})}{U_B} \quad (2-16)$$

Where H_{max} is the maximum bed expansion, H_{mf} represents the bed expansion due to minimum fluidisation velocity, U_B symbolises the bubble velocity and U_{mf} is the minimum fluidisation velocity.

2.2.3 Geldart particle type classification of Ammonium nitrate

Ammonium nitrate (AN) is manufactured as a white, crystalline solid soluble in water and hygroscopic (Speight, 2002; Pesce & Jenks, 2013). According to Nichols (2005) under certain conditions, ammonium nitrate is explosive (rated as blasting agent) and when it burns, it cannot be extinguished because it applies its own oxygen. Porous granular ammonium nitrate is produced by sparging molten ammonium nitrate nozzles into a fluidised bed granulator filled with ammonium nitrate bed particles fluidised with air. The explosive grade ammonium nitrate granules are spherical in shape with average size of 1.5mm to 3.0mm and bulk density ranging between 0.75 g/cm³ and 0.9 g/cm³ (Visagie & Pille, 2009).

Geldart devised a criterion to predict the mode of fluidisation by focusing on the particle characteristics that make them fluidise in different ways (Kunii & Levenspiel, 1991:77). The criteria predict fluidisation behaviour based on mean particle diameter and the particle density (Cocco *et al.*, 2014). According to Botha (2016) ammonium nitrate is characterised as Group B particles. Per Geldart classification, ammonium nitrate is expected to fluidise well with bubbles that grow large (Kunii & Levenspiel, 1991:77; Cocco *et al.*, 2014).

Anantharaman *et al.* (2017) the flow direction near the wall of the pilot scale circulating fluidised bed riser for Geldart Group B monodispersed particles. It was observed that the diameter of the particles was a dominant factor in indicating the annulus upward flow. It was also found that the applicability of the available flow regime maps largely developed for Geldart Group A particles are limited for the Geldart Group B particles. Flow regimes of Geldart Group B particles were also studied by Saayman *et al.* (2013) who used x-ray chromatography to quantify bubbling, turbulent and fast fluidisation flow regimes. Sluggish behaviour was observed for higher bubbling regime velocities. Rasteh *et al.* (2015) studied the behaviour of tapered fluidised beds with Geldart B (TiO₂ sand, dolomite and NaCl) particles. Numerical comparison indicated that the prediction of the model was in good agreement with the experimental data. He *et al.* (2013) studied the distribution characteristics of bed density in an air dense medium Geldart Group B magnetite powder fluidised bed based on the Euler-Euler model. It was observed that the density stability was obtained in the axial position rather than along different bed heights.

2.2.4 Fluidising medium distribution

The quality of fluidisation depends on the distribution of the fluid (Shukrie *et al.*, 2016; Depypere *et al.*, 2004). The purpose of distribution plates is;

- to provide an even gas distribution into the fluidised bed (Kale & Bisaka, 2011; Kafle *et al.*, 2016; Depypere *et al.*, 2004; Wormsbecker, 2007),
- to prevent weepage (leaking of particles through the distributor plate) of solids into the plenum,

- to support the weight of the particles, and
- to assist with proper mixing of the particles without segregation thus producing an effective contact between gas and solid and maximising movement of the particles (Kelkar *et al.*, 2016; Depypere *et al.*, 2004).

The gas distributor has a controlling influence on minimising challenging, dead zones on the plate and blockage of the plate and supporting the pressure drop forces (Depypere *et al.*, 2004). The air distributor acts as a filter (Peirano *et al.*, 2002), hence the success or failure of the fluidised bed depends on its performance. Gas distributors that are carelessly or inadequately designed malfunction in operation and are responsible for the difficulties experienced in a fluidised bed (Geldart & Baeyens, 1985) (Kelkar, 2016).

Understanding of the design of the gas distributor is crucial for the design and improvement of fluidised bed systems (Horio *et al.*, 1980; Shukrie *et al.*, 2016). According to the review compiled by Shukrie *et al.* (2016), the performance characteristics, the dynamics of air flow, solid mixing and patterns are affected by the design of distributor. The major concern in solid processing is to prevent denser particles from settling on the distributor, prevent segregation and achieve rapid dispersion of the solids being fed to the system.

Pressure drop ratio, hole size, dead zones, geometry and spacing are factors that determine the success of the design of the distributor, leading to improved fluidisation processes (Vakhshouri, 2008). Distributor design has to account for particle attrition, erosion of the vessel and other components inside the vessel and mechanical constraints such thermal expansion, in addition to pressure drop and spacing limit considerations (Cocco *et al.*, 2014; Kelkar *et al.*, 2016). There must be a good balance between the number and diameter of the orifices for good distribution and pressure drop (Cocco *et al.*, 2014). According to Kelkar *et al.* (2016), the pressure drop of the distributor in a fluidised bed affects the flow regimes, therefore, the bed will be fluidised satisfactorily with a certain magnitude of pressure drop across the distributor.

The surface area for reaction or separation in a process vessel that is available to the fluid is influenced by the distribution. To utilise the maximum surface area for the process, a uniform fluid distribution is needed, therefore enhancement of the distributor is required to ensure that the yield of the desired product is increased efficiently (Nowobilski, 1994). According to Shukrie *et al.* (2016), there are three groups of distributors (i.e. normal angle, a lateral direction and inclined angle) based on the entering air direction. The primary designs of distributors are discussed below.

Perforated (multi-orifice) plates

Perforated distributor plates consist of a solid plate with holes distributed over the entire area. The holes are arranged in either square or triangular pitches. The open area percentage, also known as perforation ratio, is determined by calculating the ratio of the open area of the total number of holes to the total distributor area (Shukrie *et al.*, 2016). The diameter of the perforated plates ranges between 1 and 2 mm in a small scale experimental setup and up to 50 mm for large units. Kunii & Levenspiel (1991) discussed several variations of perforated distributor plates.

Sandwiching perforated plates consist of two or more plates sandwiching a metal screen that prevents weepage through the perforated holes. The design that lacks rigidity is classified as staggered and consists of two staggered plates without a screen. Curved, dished perforated plates are utilised when dealing with heavy loads to provide a reinforcement structure to supports the flat plate from deflecting under heavy loads or during the leakage of thermal expansion gas. These plates also counteract the bubbling and channelling that occurs at the centre of the plate. To avoid centre channelling in upward curved plates, more holes are used at locations near the perimeter than the centre.

These type of plates are mostly used in the industry because they are affordable, easy to fabricate, scaleable, modifiable and easy to clean, can take various shapes (flat, concave, convex or double dished) and their ports are easily shrouded. Even with a number of advantages, perforated plates have drawbacks including (i) weepage of particles into the plenum, (ii) requirement for high pressure drops, (iii) subjection to the thermal distortion or buckling (preventing the use under high operating temperatures and highly reactive environments) and (iv) requirement of support over long periods of operation (Kunni & Levenspiel, 1991) (Yang, 2003).

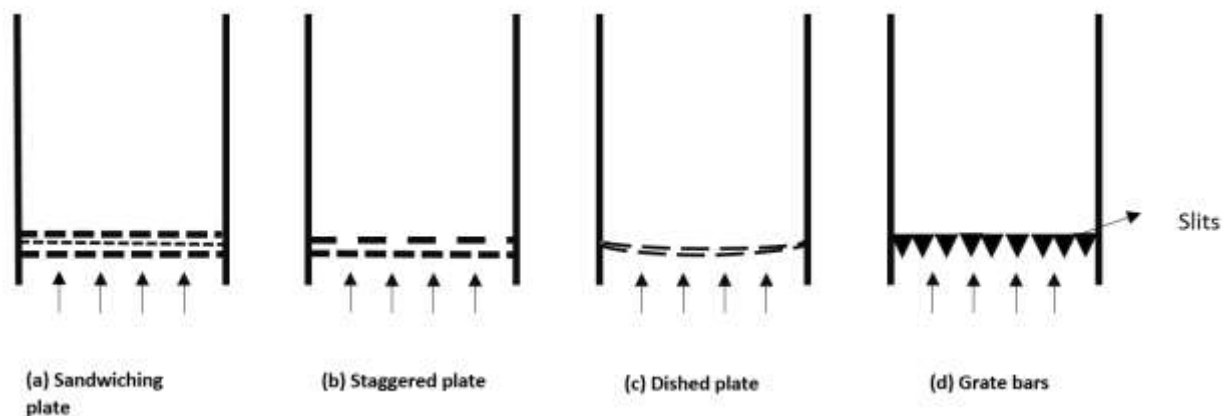


Figure 2-7: Types of the perforated/multi-orifice distributor plate.

(Adapted from Kunii & Levenspiel (1991:96))

Summary of various studies on different fluidised bed distributors

Horio *et al.* (1980) investigated the movement of particles above the perforated distributor plate in a three-dimensional fluidised bed using a tracer method for the visual observation of dead zone shape and particle motion. The results showed that the bubble motion has no significant influence on the particle motion but the flow pattern is stable. In a gas-fluidised bed, Rees *et al.* (2006) focused on the nature of fluid flow above the perforated distributor plate packed with bed material of two different particle sizes using magnetic resonance imaging. The images obtained revealed that above each orifice of the distributor in the study, there was a substantial jet.

According to Chyang & Huang (1991), there is a contradiction between the perforated-plate distributor pressure drop when measured in the absence and presence of bed particles using a manometer. It was observed from the results that the pressure drop across the distributor plate is dependent on the location of pressure taps in the presence of bed material, in contrary to measurements obtained in the absences of bed particles. The effect of superficial gas velocity was also investigated and it was seen that a smaller superficial gas velocity results in non-equal gas flow through the plate holes which subsequently results in gas maldistribution. Javier (2015) drawn the conclusion that in order to avoid gas maldistribution in fluidised bed, pressure drop across the distributor has to be kept high and can be created by increasing superficial gas velocities. The effect of air distributor on the characteristics of fluidisation was investigated by Son *et al.* (2005) using five different perforated plate distributors. It was determined that differential pressure drop across the bed increases as the velocity of the gas increases and increasing the opening fraction of the distributor.

The air supply and pressure drop of the distributor indirectly influence the pressure fluctuations in the systems through their effect on the bubbles (Leckner *et al.* 2002). The influence of air supply on the prediction of the fluid dynamics in a bubbling fluidised bed was investigated by Peirano *et al.* (2002). The results showed that in a high plate pressure drop situation, there is a good agreement between the predicted and measured data. When the air distributor has low pressure, measured dynamics cannot be reproduced in the prediction simulation. The author concluded that for the configuration of a low-pressure drop system, the plenum should be included in the simulation. According to the results obtained by Ghaly *et al.* (2015), the pressure drop is also affected by both the shape and conical angle of the distributor plate. Pressure drop decreased with the decrease in the convex angle of the distributor and increased concave angle of the plate. It was also observed that increasing the bed height resulted in an increase in the pressure drop but the ratio of the pressure drop across the plate to the bed pressure drop decreased.

Paiva *et al.* (2004) conducted the investigation on the influence that the distributor plate has in the bottom of the fluidised bed. The dynamic conditions at the bottom of the zone of the investigation resulted in a higher pressure drop for the distributors with a higher opening ratio. Sobrino *et al.* (2009) focused on the effect of the distributor on the hydrodynamics of a turbulent fluidised bed near the bottom region. Pressure and radial voidage profile measurements were obtained for a 29cm diameter column packed with Group A particles. Two different fluid distributors, bubble caps and perforated plate were used. The influence of the distributor plate on the hydrodynamics of fluidised bed was highlighted by the results when comparing experimental and simulation results of the two plates. For the perforated plate, a higher decrease of the bed height with an increase in the superficial gas velocity was presented, indicating that the rate of transfer of solids from dense bed to the freeboard of the fluidisation vessel is higher.

According to the results obtained by Dong *et al.* (2009) when investigating the effect of perforated ratios of the distributor on the characteristics of air-catalytic cracking particles fluidised bed, the pressure drop of the distributor decreases with increasing perforated ratios of the distributor and decreasing the superficial velocity of the gas. Homogeneous fluidization was observed using a distributor with a 0.46% opening compared to that of 0.86% and 1.10%.

Wormsbecker *et al.* (2007) studied the influence of distributor design on the hydrodynamics of fluidised bed dryers using perforated, punched and Dutch weave mesh distributor plates. Significant changes were studied when varying bed loadings and superficial gas velocities. It was found that the punched plate performed better than other designs when operated with lower bed loading and low gas velocity. The effect of the orifice of the distributor on the drying kinetics of a fluidised bed drier was investigated by Sutar & Sahoo (2011), using four different perforated plates of different orifice diameter and open areas. The experimental results highlighted that the rate of drying increase with increasing orifice size.

By measuring the pressure fluctuations in the bed, windbox and across the distributor, Vakhshouri (2008) studied the influence of the distributor and the volume of the plenum on the hydrodynamics of fluidised bed. It was obtained that the bubbling frequency increases with the decreasing volume of the plenum. The author concluded that the results observed may be due to the formation of larger bubbles in plenums chambers with large volume.

Paiva *et al.* (2009) examined the influence of the type of distributor plate and the pressure drop of the plate on the quality of fluidisation using perforated Perspex, metallic mesh and porous ceramic distributor plates with pressure drop ranging between 0.05 and 350 KPa. Focusing on the influence of superficial gas velocity, the results highlighted that the fluidisation quality

increases from bubbling until the transition point to turbulent regimes when distributors with low-pressure drop are used.

A perforated plate, a perforated plate covered with porous cloth on top (referred to as the porous distributor) and multi-vortex tuyere distributors were tested for their effect on the mass transfer and axial gas dispersion in fluidised beds of bubbling and turbulent flow regimes (Saayman *et al.*, 2011). It was shown that large bubbles and poor performance of the fluidised bed reactor resulted for porous distributor compared to perforated plate distributor. When investigating the effect of orifice pattern configuration of a distributor plate, Afrooz *et al.* (2016) found that a more homogeneous distribution of solid volume fraction and regular pattern of particle velocity was obtained using the triangular distributor plate. Rectangular distributor plate resulted in chaotic solid motion.

2.3 Numerical modelling

The focus of most research on the modelling of fluidised bed processing has been on hydrodynamic modelling, which is a discipline gaining attention. There are two basic types of hydrodynamic models, named the Eulerian model, commonly known as computational fluid dynamics (CFD) and the Lagrangian model, commonly referred to as discrete element models (DEM) (Hede, 2013). For the purpose of this study, the two incompressible fluid phases with one phase dispersed are solved describing both phases using Eulerian conservation equation referred to as Euler-Euler model therefore the focus will be on the Eulerian model.

2.3.1 Computational fluid dynamics (CFD)

CFD provides a powerful, cost-effective tool of real flow simulations by numerical equations solution (Sayma, 2009). By solving mathematical equations governing processes, a CFD model can be used to predict the flow of the fluid, heat and mass transfer and other related phenomena using numerical approaches, which enlarge understanding of the process (Kafle, 2009; Depypere *et al.*, 2004). Peirano *et al.* (2002) conducted experiments in a bubbling fluidised bed and used the Eulerian approach to show that the numerical simulation can predict the fluidisation pattern, focusing on the influence of the air supply on the system. It was evident from the results that numerical computation can qualitatively predict the experimentally observed values.

CFD has become one of the most applied numerical simulations for dynamic fluid systems, because of its ability to predict the hydrodynamics. Estejab *et al.* (2017) simulated the interaction between particles of coal-biomass mixtures in Geldart group A particles in a fluidised bed. The effect of the distributor orifices on the hydrodynamics of a bubbling fluidised bed was studied by Afrooz *et al.* (2006). CFD modelling was used for different applications such as;

- conducting investigations to obtain rapid and sequential performance of a fluidised bed (He *et al.*, 2016),
- modelling of the two-phase flow in a circulating fluidised bed furnace (Shah *et al.*, 2015),
- study of axial and radial solids concentration distribution by simulating the bubbling fluidised bed hydrodynamics with the particles of fluid catalytic cracking (Lv *et al.*, 2014),
- predicting of scaling effects and flow behaviour in pharmaceutical processes (Parker *et al.*, 2013),
- validation of experimental research conducted by Nagarajan *et al.* (2009) to decompose sulphur trioxide in a heat exchanger, and
- investigation of changes in a Glatt GPCG-1 fluidised bed reactor configuration to obtain homogenous air flow distribution (Depypere *et al.*, 2004), to name a few.

In a CFD simulation, different models are developed to describe the particle interaction and momentum transfer between phases in the system. A kinetic granular flow theory developed from kinetic gas theory is used to describe the particle-particle interaction. A drag model describes the interaction or momentum exchange between particles and the gas phase. The momentum equation of the granular phase is dominated by gravity and drag terms (Lundberg & Halvorsen, 2008). A model based on the Euler-Euler approach was employed by He *et al.* (2013) to obtain instantaneous information regarding bed density.

CFD modelling has two different classifications, namely the Eulerian-Lagrangian approach, which solves the Newtonian motion equation for particles individually, taking into account the collision model and the Eulerian-Eulerian approach which treats the phases as continuous and interpenetrating (Benzarti *et al.*, 2012). Equations governing CFD numerical modelling are mass conservation, momentum conservation and energy conservation equations (Depypere *et al.*, 2004; Lundberg & Halvorsen, 2008). Assuming an inertial reference frame, for an incompressible Reynolds fluid, the equations are represented as follows:

$$\nabla \cdot \vec{u} = 0 \quad (2-17)$$

$$\rho \frac{\partial \vec{u}}{\partial t} + \rho \nabla \cdot (\vec{u} \vec{u}) = -\nabla P + \rho \vec{F}_{Ex} + \nabla \cdot \vec{\tau} + \vec{S} \quad (2-18)$$

$$\rho \frac{\partial E}{\partial t} + \rho \nabla \cdot (H \cdot \vec{u}) = \nabla \cdot (\vec{\tau} \vec{u}) - \nabla \cdot \vec{q} \quad (2-19)$$

Where P is the external pressure, F_{Ex} is the external force vector, τ is the stress tensor, S is the model dependent momentum source term, E is the total energy, H is the enthalpy and q is the heat flux vector.

Table 2-2 gives a summary of the studies that employed the Eulerian approach as summarised by Estejab *et al.* (2017).

2.3.2 Applicable solvers in the CFD environment.

The leading open source code for CFD is *OpenFOAM*TM (Gu *et al.*, 2018) (Ciegis *et al.*, 2014). This section of chapter 2 provides details of the solvers used in this study and the governing equations thereof. The *twoPhaseEulerFoam* and *PorousSimpleFoam* solvers are used for two-phase fluidised bed systems simulations and characterisation of the distributor plate respectively.

twoPhaseEulerFoam

The *twoPhaseEulerFoam* solver is used when solving two incompressible fluid phases with one dispersed. Both phases are described as a continuum and inter-penetrating by Eulerian conservation averaged equations. The model is therefore referred to as a Euler-Euler model. Equations 2-20 to 2-25 are implemented in the *OPENFOAM*TM solver from Computational Fluid Dynamics of Dispersed Two-Phase Flows at high phase fractions. The momentum transfer between the two phases is accounted for by the averaged inter-phase momentum transfer (AIPMT) term. Equation 2-20 and 2-21 represent the averaged momentum and continuity equations respectively (Busch, 2015) (Baila, 2009) (Benzarti *et al.*, 2012).

$$\alpha_k \rho_k \left(\frac{\partial}{\partial t} U_{k,i} + U_{k,j} \frac{\partial}{\partial x_j} U_{k,i} \right) = \alpha_k \rho_k g_i - \alpha_k \frac{P_l}{\partial x_i} - \frac{\partial \tau_{k,ij}}{\partial x_j} + I_{k,i} \quad (2-20)$$

$$\frac{\partial}{\partial t} \alpha_k \rho_k + \frac{\partial}{\partial x_i} \alpha_k \rho_k U_{k,i} = 0 \quad (2-21)$$

Where α_k is volume fraction of the fluid. (The index k is a representative symbol for the solid and the fluid phase). ρ_k the density of the fluid. $\tau_{k,ij}$ the stress tensor and $I_{k,i}$ = momentum forces between phases.

The void fraction of different phases in the system sums up to unity. Drag forces, lift forces and turbulent dispersion are some of the forces covered by momentum forces between phases (Busch, 2015). When equation 2-21 is combined for the two phases, an implicit pressure equation can be formulated from the volumetric continuity equation obtained. To calculate the inter-phase momentum transfer, forces (drag, lift and virtual mass forces) acting on the dispersed phase particles are added, neglecting history or Basset forces. According to Crowe *et al.* (2011) the force due to the lagging boundary layer development with changing relative velocity (acceleration) of bodies moving through a fluid). Equation 2-22, 2-24 and 2-25 are the volumetric continuity, pressure equation re-casted from the volumetric continuity equation and phase continuity equations respectively (Baila, 2009).

Table 2-2: Summary of studies that employed the Eulerian approach (Summarised by Estejab *et al*,2017)

Study	Methodology	comments
Ferschneir & Mege (1996), Krishan & Van Baten (2001), Van Wachen <i>et al.</i> (2001)	Showed that there are some deficiencies in predicting behaviour of Geldart type A particles	Many attempts have been made to overcome the deficiencies and capture the fluidisation behaviour of fine particles.
Harris <i>et al.</i> (2002), Lackermeier <i>et al.</i> (2001), Mostoufi & Chaouki (2004)	Reported the existence of particle clusters in gas-solids flows due to inter-particle forces	For Geldart A particles, inter-particle forces result in particle clusters.
Hulin <i>et al.</i> (2005), Shuyan <i>et al.</i> (2008)	Proposed a model that represents the movement of cluster particles.	It is challenging to effectively predict the size of the cluster for various model applications
McKeen and Pugsley (2003), Ye <i>et al.</i> (2008)	Modified the drag model to capture the Geldart A cluster particles	The model uses non-general scaling factors, therefore factors must be determined for each application.
Parmentier <i>et al.</i> (2008), Wang (2009), Wang <i>et al.</i>(2010).	Investigated the effect of mesh size on fluidisation particles and emphasised the importance of employment of fine grids and small time steps	With regards to the current and future computer resources, the method is either expensive or non-feasible.
Wang (2009), Lu <i>et al.</i> (2011), Benzarti <i>et al.</i> (2012)	Employed a new drag model to account for the presence of particle clusters over a wide length range and time scales.	Kinetic theory incorporated in method assuming homogeneous granular flow.

$$\nabla \cdot \bar{U} = 0 \quad 2-22$$

Where
$$\bar{U} = \alpha_a \bar{U}_a + \alpha_b \bar{U}_b \quad 2-23$$

$$\left\| \nabla \cdot \left(\left(\alpha_{af} \left(\frac{1}{\rho_a(\mathcal{A}_a)_D} \right) + \alpha_{bf} \left(\frac{1}{\rho_b(\mathcal{A}_b)_D} \right) \right) \nabla [\bar{P}] \right) \right\| = \nabla \cdot (\alpha_{af} \Phi_a^* + \alpha_{bf} \Phi_b^*) \quad 2-24$$

$$\frac{\partial \alpha_a}{\partial t} + \nabla \cdot (\bar{U} \alpha_a) + \nabla \cdot (\bar{U}_r \alpha_a (1 - \alpha_a)) = 0 \quad 2-25$$

$\bar{U}_r$ = relative velocity.

The kinetic theory of granular flow is used to derive the pressure and granular viscosities, where the granular flow temperature is replaced with the thermodynamic temperature. The granular flow temperature is the measure of the fluctuating particle's velocity.

Solid phase stress is required to complete the momentum equation for the solid phase. The granular equation is represented by equation 2-26. The first right-hand term and the second terms of the equation is the creation of fluctuating energy resulting from particle phase shear and diffusion fluctuating energy, respectively (Baila, 2009). The algebraic equation, equation 2-27, is used when equilibrium in the kineticTheoryProperties in OpenFOAM™ KineticTheoryModel.C source code is on instead of the balanced equation.

$$\frac{3}{2} \left[\frac{\partial}{\partial t} (\alpha_s \rho_s \theta) + \nabla \cdot (\alpha_s \rho_s \theta v_s) \right] = (-\nabla P_s \bar{I} + \bar{\tau} s) : \nabla v_s - \nabla \cdot (k_s \nabla \theta) - \gamma_s - J_s \quad 2-26$$

$$\theta = \left(\frac{-(K_1 \alpha_s + \rho_s) \text{tr}(\bar{D}_s) + \sqrt{(K_1 \alpha_s + \rho_s)^2 \text{tr}^2(\bar{D}_s) + 4K_4 \alpha_s [2K_3 \text{tr}(\bar{D}_s^2) + K_2 \text{tr}^2(\bar{D}_s)]}}{2\alpha_s K_4} \right)^2 \quad 2-27$$

θ represents the granular temperature, γ_s is the dissipation due to inelastic particle-particle collisions and J_s the dissipation/creation of granular energy.

Drag Model

When a bed of particles is penetrated through by the flow of a fluid, it experiences drag and buoyancy force (Friedle *et al.*, 2017). Different models are specified to calculate the drag transfer coefficients which is used to calculate the drag force in a two-phase system. Values mostly applied for the drag models include Reynolds number, phase fraction, the relative velocity between phases and drag coefficient (Baila, 2008). The general representation of drag force is represented by equation 2-28 (Busch, 2011). Gidaspow combined the Ergun equation, Rowe and Wen & Yu drag model to generate a drag model that covers a wide range of porosities (Halvorsen *et al.*, 2006; Lundberg & Halvorsen, 2008). The drag model that is frequently used to determine the gradients in a fluidised bed is the Ergun equation represented by equation 2-28. The derivation of the Ergun equation is

based on basic assumptions including two model parameters, only accounting for overall effects of fluid flow (Friedle *et al*, 2017). Equation 2-29 is the Wen and Yu equation applicable for voidage above 0.8. The relationship between the friction and Reynolds number proposed by Rowe (1961) is given by equation 2-30.

$$FD^2 = \frac{A(1-\varepsilon)^2}{\varepsilon^3} + \frac{B}{\varepsilon^3} Re = \frac{150(1-\varepsilon)^2}{\varepsilon^3} + \frac{1.75(1-\varepsilon)}{\varepsilon^3} Re \quad 2-28$$

$\varepsilon = \text{porosity}$

Re is the Reynolds number, represented by $Re = \frac{\rho_g v_g D}{\mu_g}$

v_g is the Superficial gas velocity, obtained from $v_g = \varepsilon |U_g - U_s|$

$$FD^2 = \frac{3}{4} C_D \frac{\varepsilon_s \varepsilon D}{\mu} \rho_g |U_g - U_s| \varepsilon^{-2.65} \quad 2-29$$

$$C_D = \frac{24}{Re_s} (1 + 0.15 Re^{0.687}), Re_s \leq 1000 \quad 2-29.1$$

$$C_D = 0.44, Re_s > 1000 \quad 2.29.2$$

The drag model used for this study is the GidaspowErgunWen-Yu model, represented by particle Reynolds equation 2-30 (Benzarti *et al.*, 2012);

$$Re_p = \frac{\varepsilon_g \rho_g |U_g - U_s| d_p}{\mu_g} \quad 2-30$$

Geometric Agglomerated Algebraic MultiGrid (GAMG)

The solver in *OpenFOAM*TM is the GAMG algorithm that solves the Poisson equation used for pressure correction. The profiling results have shown that the smoother takes half of solving time for GAMG (Gu *et al*, 2018). Figure 2-8 illustrates an example of the concept implemented for two-level GAMG algorithm in *OpenFOAM*TM. GAMG can also operate as a preconditioner to ensure that the preconditioned system converges faster than the original. GAMG soothes the high-frequency errors the coarse grid with fast convergence to generate a fine grid solution by geometric or algebraic multi-grid (Behrens, 2009; Rynell, 2010). Ciegis *et al* (2014) studied the efficiency of parallel solvers based in *OpenFOAM*TM for heat conduction simulations. The study compared the efficiency of diagonal incomplete Cholesky (DIC) and GAMG and observed best running times when using GAMG solver.

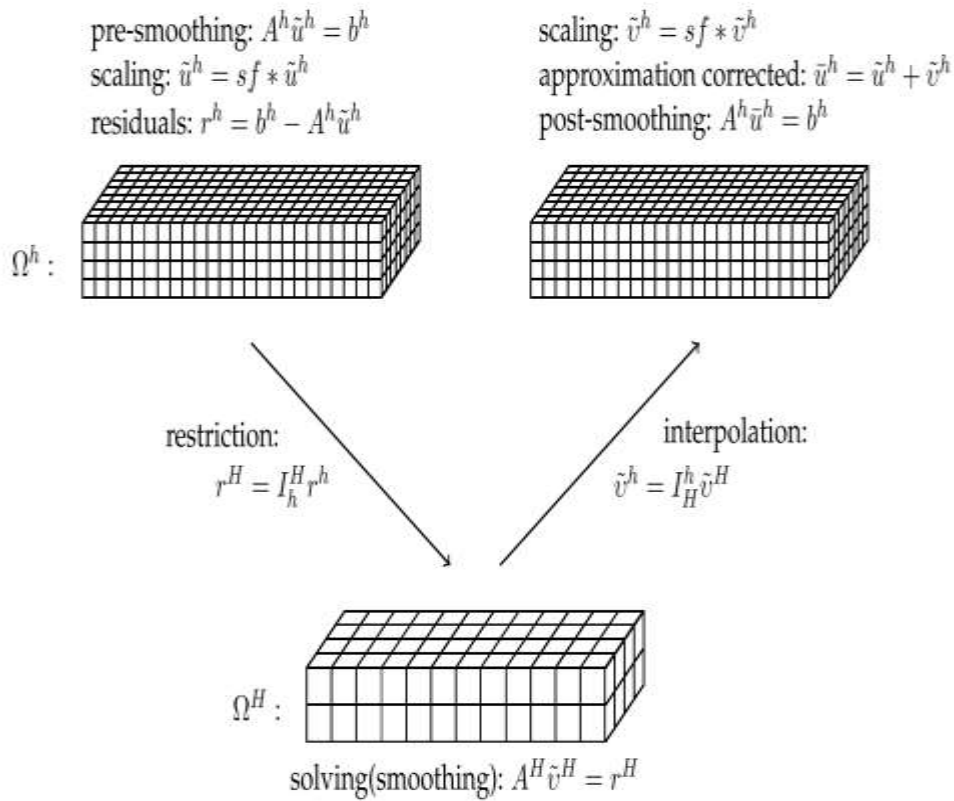


Figure 2-8: A two-level example of GAMG solver in OpenFOAM™.

SmoothSolver

SmoothSolver is a linear iterative solver that uses a smoother for matrices that are either symmetric or asymmetric (Behrens,2009). When using a *smoothSolver*, a smoother need to be chosen and a number of sweeps specified (*OpenFoam*™ V1806). The smoothers include DIC, DICGaussSeidel, DILU, DILUGaussSeidel, GaussSeidel, etc (Behrens, 2009). The efficiency is improved by evaluating the residual after every nSweeps of smoothing iterations. The call graph of the function of *SmoothSolver* as described in *OpenFOAM*™ V1806, extended code guide is represented by Figure 2-9.

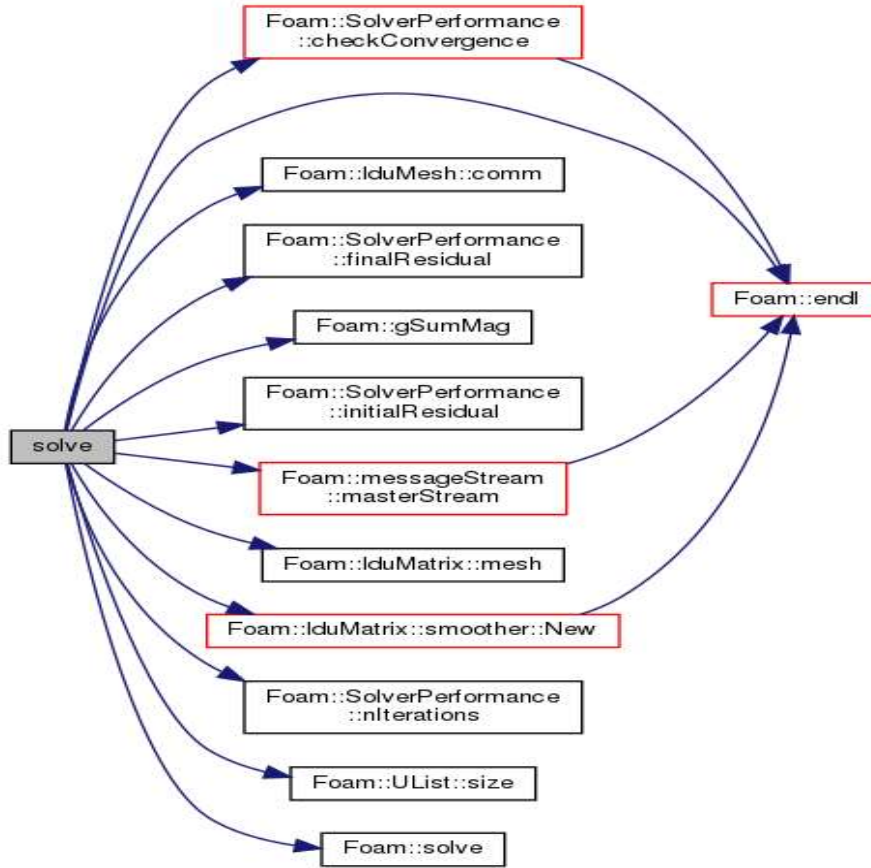


Figure 2-9: *smoothSolver* function call graph.

SIMPLE porous medium model (*porousSimpleFoam*)

Semi-Implicit Method for Pressure-Linked Equations (SIMPLE) allows coupling of Navier-Stokes equations using an eight-step procedure; Firstly, the boundary conditions are set. The momentum equation is then discretised to compute the velocity field. Mass fluxes of the cell faces are computed in order to solve the pressure equation. Correct values of mass fluxes and velocity are corrected after applying under-relaxation. Boundary conditions are then updated and the procedure repeats until convergence is reached (Ferziger & Peric, 2001). The porous media in the model determines the additional flow resistance in volume empirically, therefore it is a momentum sink of general momentum equations. The momentum sink term is composed of the viscous loss and inertial loss obtained referred to as Darcy Forchheimer equation (Eq 2-31) (Depypere *et al.*, 2004; Fithry & MohdZamri, 2013; Mlynarczyk & Cyklis, 2017).

$$S_i = -(\sum_{j=1}^3 -(D_{ij})D_{ij}\mu v_j + \sum_{j=1}^3 C_{ij} \frac{1}{2} \rho |v| v_j) \quad 2-31$$

Where $|v|$ is the velocity magnitude. D is the viscous loss or impermeability term predefined matrix and C is the inertial loss term predefined matrix in the three coordinate directions. This approach

results in a proportional relationship between the pressure drop and fluid dynamics in the porous cell. For a simple homogeneous media, equation 2-32 above takes the following form

$$S_i = -(\frac{\mu}{\alpha} v_i + C_2 \frac{1}{2} \rho |v| v_i) \quad 2-32$$

Where α is the permeability (Mlynarczyk & Cyklis, 2017). with the proportionality between the pressure drop and velocity, constant C_2 can therefore be assumed to be zero for a laminar flow and the equation (2-33) above simplifies to Darcy's law (Depypere *et al.*, 2004) (Fithy & MohdZamri, 2013) (Mlynarczyk & Cyklis, 2017). Darcy law links the pressure drop and fluid flow velocity in a porous media (Sobieski & Trykozko, 2012)

$$\nabla P = -\frac{\mu}{\alpha} \vec{v} \quad 2-33$$

which can also be expressed as

$$-\frac{dP}{dx} = \frac{1}{\kappa} \cdot (\mu \cdot \overrightarrow{v_f}) \quad 2-34$$

Where P, x, κ, μ and v_f are the pressure (pa), coordinate (m), the permeability coefficient (m^2), fluid dynamics viscosity coefficient (kg/m. s) and velocity, respectively. Darcy's law represents the porous media flow correctly for low flow velocities. In the case of high fluid velocities, a discrepancy between experimental values and Darcy's law results appears. The discrepancy was linked to inertial effects by Forchheimer by suggesting an additional term representing the kinetic energy, resulting in the following expression:

$$-\frac{dP}{dx} = \frac{1}{\kappa} \cdot (\mu \cdot \overrightarrow{v_f}) + \beta \cdot (\rho \cdot \overrightarrow{v_f}^2) \quad 2-35$$

Where β is the Forchheimer coefficient in 1/m and ρ is the fluid density.

The constant C_2 represents the correction factor of inertial loss at high flow velocities. Simplification and rearrangement of equation 2-35 above gives the following expression;

$$-\frac{\Delta P_d}{\Delta z} = \frac{\mu}{\alpha} v + \frac{1}{2} \rho C_2 v^2 \quad 2-36$$

Where C_2 is the pressure fluctuation coefficient.

Ammonium nitrate plays a role in agriculture and mining industries (Visagie & Pille, 2009) therefore the production process requires optimisation. Numerical modelling is investigated as an alternative to process optimisation. From literature, CFD modelling has been investigated its ability to predict fluidisation has been proven. Fluidisation of particles classified as Geldart group B particles have been investigated under various operating conditions by Saaymaan *et al.*, 2013, Rasteh *et al.*, 2015,

Anantharaman *et al.*, 2017, to name a few. The outcomes of the studies provided foundation to the use of fluidised bed system for investigations related to ammonium nitrate particles.

2.4 Chapter summary

This chapter covered a thorough literature study of a gas-solid fluidisation system. This chapter started with an overview of fluidisation as well as theory and fundamentals that plays a major role in the fluidisation process. Highlights of studies that have been conducted on the topic of interest, including successes, shortcomings, conclusions and recommendations are also elaborated on.

The final part of the chapter covers the concept of modelling and simulation focusing on the solvers used in *OpenFOAM*™ which are applicable to this investigation. This included the necessary mathematical models and equations.

The following chapter covers the procedure and techniques to be followed in order to carry out the investigation.

3. EXPERIMENTS AND MODELLING

The effect of air distribution plates on bed hydrodynamics was investigated using particles with different average particle diameters, and with varied superficial gas velocities. The bed hydrodynamics investigated in the laboratory-scale setup include pressure differentials, minimum fluidisation velocity, bed density and bed expansion while only bed density received attention in the pilot-scale fluidised bed setup. This chapter starts with a detailed description of fluidised bed setups and equipment in both scale (section 3.1) followed by a discussion of the properties and classification of bed material. Lastly, both the experimental phase (section 3.3) and CFD modelling of both scales in OpenFOAM™ (section 3.4) are reviewed.

3.1 Laboratory and pilot scale fluidised bed setups

A fluidised bed reactor is composed of four main components, namely a plenum, a gas distributor, a bed of particles and a freeboard region. Heating and cooling coils and cyclones are optional to the design of a fluidised bed (Cocco, 2014). For this study, a laboratory scale fluidised bed represented by Figure 3-1 with a 109 mm ID (Inner diameter) was fitted with a perspex tube with a 100 mm ID. This was done to avoid the effects of the conical angle present at the bottom of the fluidised bed. The distributor plate is represented by dashed line.

Two blowers, a left-hand blower (LHB) and right-hand blower (RHB) are connected to the fluidised bed to supply fluidisation air to the system. Ball valves were used to manually manipulate the air flow rate. The entry point of the air supplied by RHB and LHB are located below and next to the riser tube respectively as seen in Figure 3-1. Each blower has a differential pressure cell connected to measure the pressure drop across an orifice. In order to achieve a homogeneous fluidisation, the RHB to LHB ratio as measured by the DP cells based on the preliminary fluidisation trials is kept constant at 3:1. According to Perry and Green (1999), the superficial gas velocity in an empty vessel is between 0.5 m/s and 6 m/s. The pressure drop measured across the orifice of the blower was related back to velocity through a predefined velocity calibration.

A pilot scale fluidised bed setup is represented by Figure 3-2 with the tower area a 1 m x 1 m square bed and equipped with a H/5 Donkin Fan industrial blower with damper. The bed particles were loaded through the front window. To minimise particle loss, the tower walls expand to the freeboard area, which is situated directly above the fluidised bed section and which is visible as the expanded section in the top of Figure 3-2. Particles that are entrained in the air leaving the freeboard are collected into the cyclone while air leaving the cyclone passes through to a scrubber to clean it from any remnant dust particles. A 13 cm hole is located beneath the top window of the fluidised bed inside the freeboard region for the addition of extra bed material if necessary.

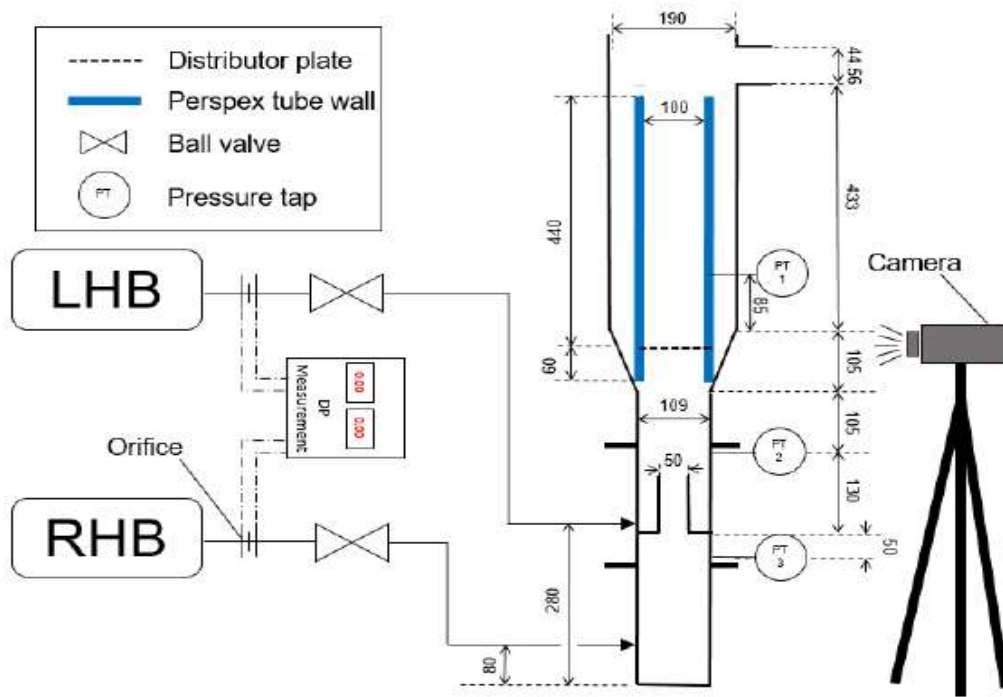


Figure 3-1: Laboratory scale fluidised bed setup.

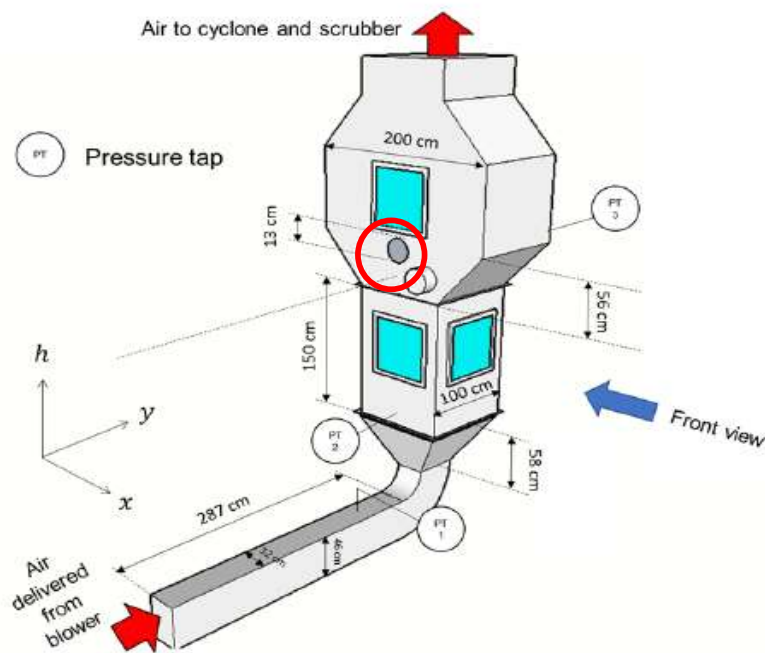


Figure 3-2: Pilot-scale fluidised bed setup.

To ensure that the fluidised bed operates reliably, constraints such as grid plates and its subjection to the pressure drop and space limit, as well as the measurement and/or estimation of entrainment rate needs to be considered (Cocco, 2014). Four multi-orifice or perforated plates supplied to and

used by the research partner were used for both laboratory scale. Due to the entrainment and massive pressure fluctuations that occurred when fluidising with plate D in the pilot scale setup, only three plates were employed for pilot scale experimental work. For the laboratory-scale setup, each plate was fixed to a Perspex tube that was fitted into the laboratory scale fluidised bed as indicated by the blue lines in Figure 3-1. Plates A, B, C and D with 8.67%, 6.12%, 2.4% and 4.2% open areas respectively employed for the experimental work in both laboratory scale and pilot scale are represented diagrammatically in Figure 3-3.

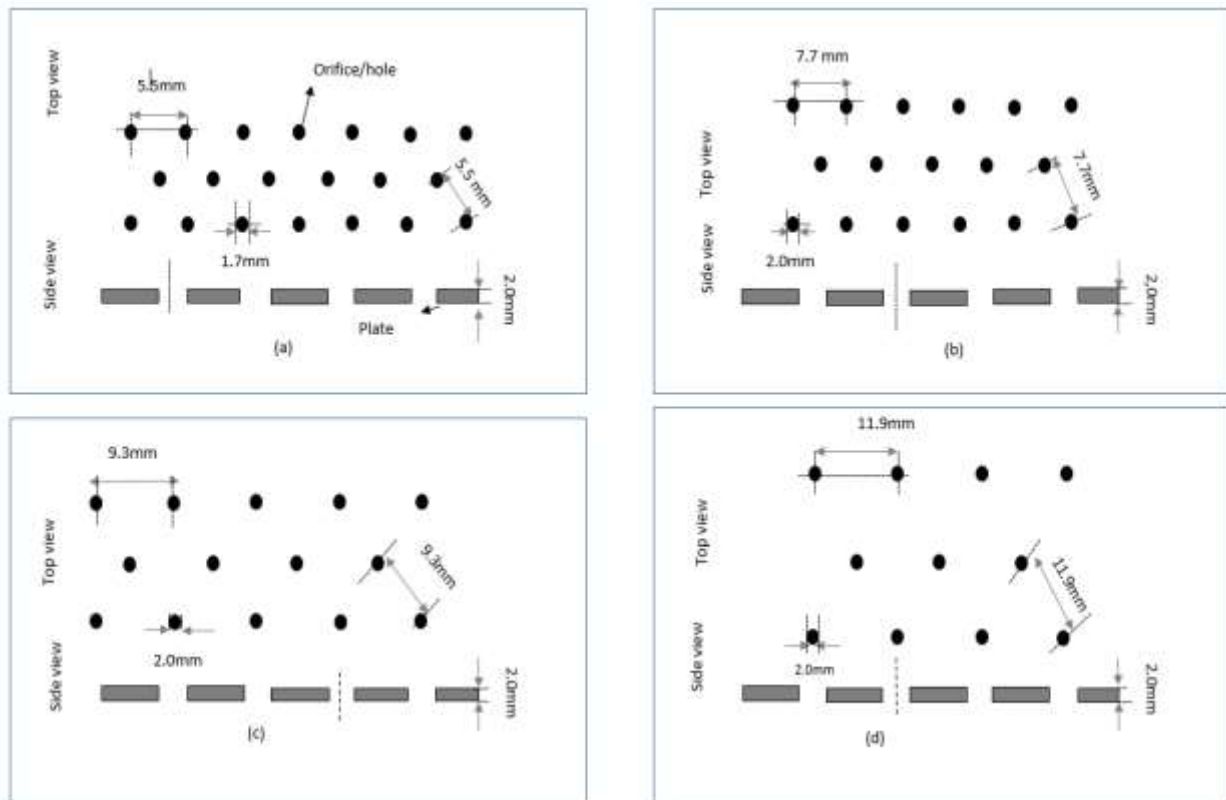


Figure 3-3: Perforated plates used in laboratory scale fluidised bed experiments.

[(a) plate A: 5.5 mm pitch, 8.67 % open area. (b) Plate B: 7.7 mm pitch, 6.12 % open area.
(c) Plate C: 9.3 mm pitch, 4,2 % open area. (d) Plate D: 11.9 mm pitch, 2.4 % open area]

3.2 Material used and particle classification

The bed material used for this study is porous granular ammonium nitrate (PGAN) that is commercially produced by the research partner. A total of 20 kg of PGAN was collected for the laboratory scale experiments from the FB granulator and 1200 kg for the pilot scale experiments. Of the 20 kg collected, two samples of 100g each were analysed using a Haver & Boecker OHG Malvern Analyser to determine the particle characteristics. The remaining particles were classified into different size ranges as detailed in Table 3-1 by using standard 8-inch sieves to obtained five particle size ranges. A final batch of unclassified particles (termed raw sample) with an average particle size

of 1.85 mm consisting of particles in the whole production particle size range are also kept as is for fluidisation experiments.

According to Visagie & Piller (2009), the explosive grade ammonium nitrate granules are smooth and spherical with an average size of 1.5 – 3.0 mm. The bulk density ranges between 0.75 and 0.9 g/cm³. The bulk density for particles used in this study were determined by dividing the mass of the particles in a cylinder of 103 mm ID by the total volume of 998mL. The particle density for each sample was obtained by calculating the ratio of the total sample mass to the volume of particles as determined by Haver & Boecker OHG Malvern Analyser. Table 3-1 summarises the weighted average particle diameter and average particle density of the particles in the samples.

Table 3-1: characteristics of the particles

Particle size range (mm)	Average diameter (mm)	Average Density (kg/m ³)
1.00-1.40	1.28	1561.5
1.40-1.80	1.60	1428
1.80-2.00	1.70	1388
2.00-2.36	2.06	1356
2.36-2.80	2.62	1428

3.3 Experimental phase

3.3.1 Experimental procedure

Pressure differential measurements

In the laboratory scale fluidised bed setup, pressure differentials were measured at two points indicated as PT1 and PT2 in Figure 3-4. Connection tubes were connected to the pressure probe points on the vessel and the pre-calibrated Fluke 922 airflow meter with a built-in pressure measurement analyser and pitot tube which in this case acted as a manometer. Five pressure readings were recorded for each flow rate increment. The pressure drop was calculated by subtracting pressure measured at PT2 with pressure at PT1. The pressure obtained is the overall pressure drop represented by equation 3-1 below. Distributor plate pressure differentials were measured in the absence of any particles at various air velocities in order to be able to eventually determine the bed pressure differential at various air velocities.

$$\Delta P_T (\text{Overall pressure drop}) = \Delta P_p (\text{distributor pressure drop}) + \Delta P_B (\text{Bed pressure drop}) \quad 3-1$$

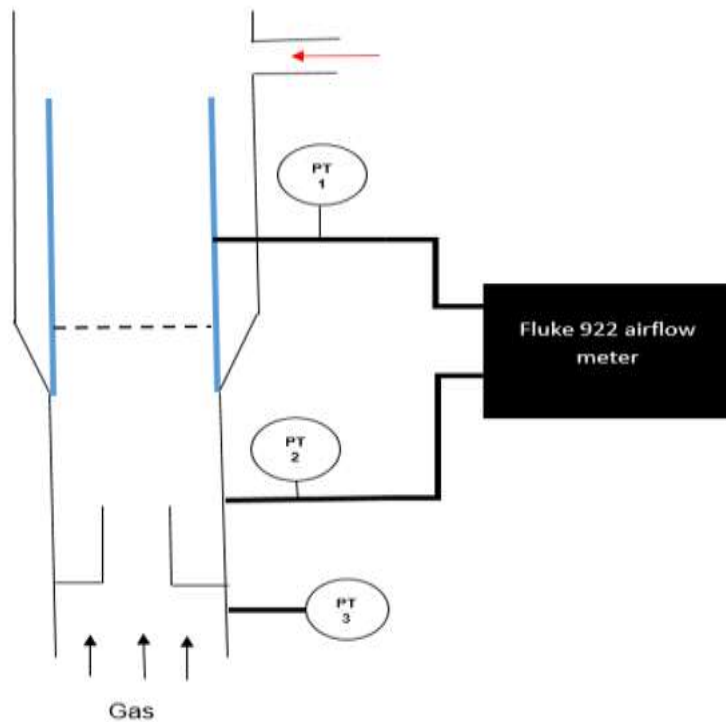


Figure 3-4: Pressure differential measurement setup in a laboratory scale fluidised bed

Bed density measurements

Figure 3-5 illustrates how the metallic tubes, connection tubes and the fluke airflow meter were mounted together in order to measure the bed density in the system. The connection tubes are inserted into respective slots on the manometer with the expected higher pressure (tube with the lowest open end into the fluidised bed) connected to the “+” port, and the other tube connected to the “-” port. The difference in height (Δh) and distance (ΔL) between the metallic tubes were kept at a constant 2 cm. In the laboratory scale setup, bed density was determined for the four distributor plates using raw (unclassified) samples. The metallic tubes were inserted at the opening indicated in figure 3-4. To avoid entering and blockage of the tube by particles, pieces of mutton cloth were used to cover the submerged ends of the tubes. Five different velocities beyond the minimum fluidisation velocity were subsequently arbitrarily chosen for the measurement of the bed density. Density was measured at ten different positions in the fluidised bed, including heights of 2, 4, 6, 8 and 10 cm in the bed and at two horizontal measuring points; one at the centre and one at the wall as shown in Figure 3-6. This procedure was followed for each particle size range on each of the distributor plates.

Dimensions	Pilot	Laboratory
ΔL - Length between tube (cm)	2	2
D- Diameter of the tubes (mm)	4	3
Δh - Height difference (cm)	1.5	2

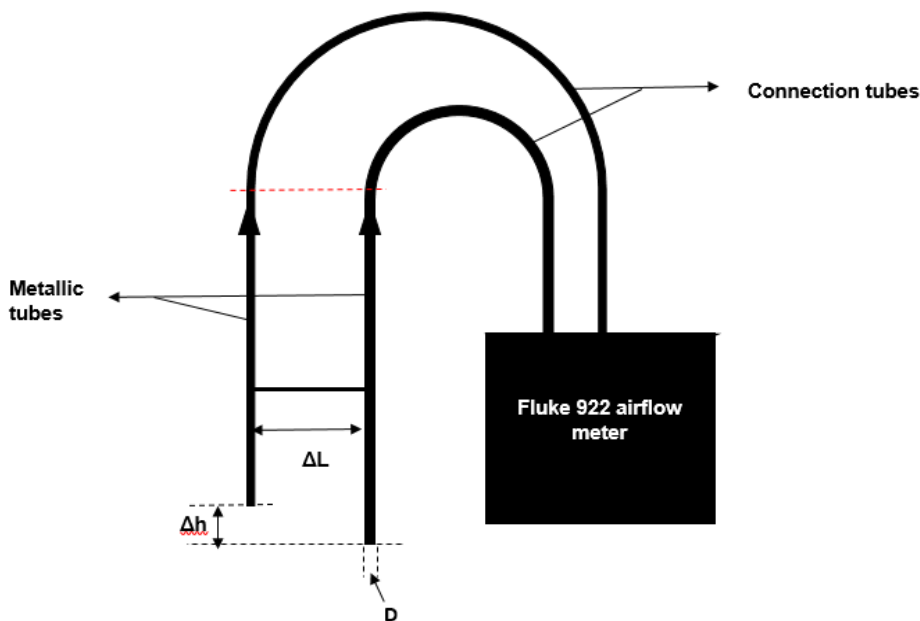


Figure 3-5: Metallic tubes, connection tubes and Fluke 922 connection for density measurement

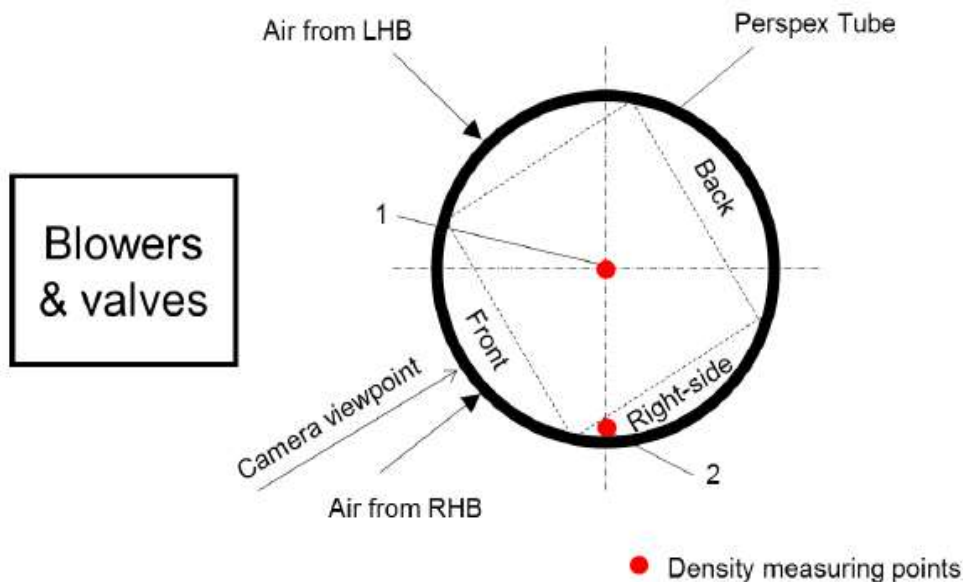


Figure 3-6: Laboratory scale experiments density measurement setup

To assist in measuring bed density in the pilot plant, a metal grid (Figure 3-7) was located at the centre of, and about 120 cm above the plate in the x-axis direction. Cross-bars were fixed at different measuring positions in the horizontal y-axis spacing to guide and ensure orientation and placement of the tubes when measuring pressure differential in the system. The correct height of measurement

was marked using wooden hooks. The density determination pressure readings were recorded in the centre of the fluidised bed along the x-axis at five different heights above the distributor plate of 5, 15, 25, 35 & 45 cm respectively and at six different distances (0, 17, 33, 49, 66 & 83 cm) from the inner wall (i.e. the wall closest to the location of the air blower) in the y-axis direction towards the outer wall (i.e. the wall furthest away from the location of the air blower). The wooden hooks are placed around the metal bar running parallel to the inner wall and are held against the top bar. To measure against the inner wall, the tubes are turned 90° so that the tube-pair is flat against that wall.

Figure 3-7: Pilot-scale experiments bed density measurement setup

3.3.2 Experimental plan

The experimental phase of this investigation was conducted to collect sufficient data that could be used for setting-up and validation of the CFD model. For every experimental run, a batch of PGAN from a specific size range was used. Fluidisation was conducted with each of the four plates and for each particle size. Particles with the selected size range were added into the perspex tube fixed with the selected distributor plate up to a static height (H_s) of 10 cm. The air flow rate was changed by

opening the valve at the RHB from 0 % opening until a point where the pressure differential indicated minimum fluidisation velocity, U_{mf} .

Pilot-scale experiments

A 1 m x 1 m pilot-scale fluidised bed is available at the research partner and was utilised for this part of the investigation. The general layout and dimensions of the fluidised bed can be seen in Figure 3-2. The performances of three different perforated distributor plates, plate A, plate B and plate C were assessed. The bed pressure drops and distributor pressure drops were measured at (PT 1 & PT 2) and (PT 1 & PT 3) respectively. The fluidised bed was loaded to a static height of 45 cm. The inlet air flow rate was manipulated manually using a damper on the blower inlet. The metallic tubes used to measure bed density in the system were inserted at the top opening (shown with a red circle in Figure 3-2) and positioned according to the metal grid (Figure 3-7).

3.4 CFD Modelling in OpenFOAM™

A computational fluid dynamics model in *OpenFOAM*™ v1712 was used to simulate a fluidised bed system with ammonium nitrate particles as bed material. A complete simulation of a fluidised bed system consists of three stages, namely (i) pre-processing, (ii) the solving stage and (iii) post-processing (Figure 3-8), which are outlined in this chapter with regards to this research project.

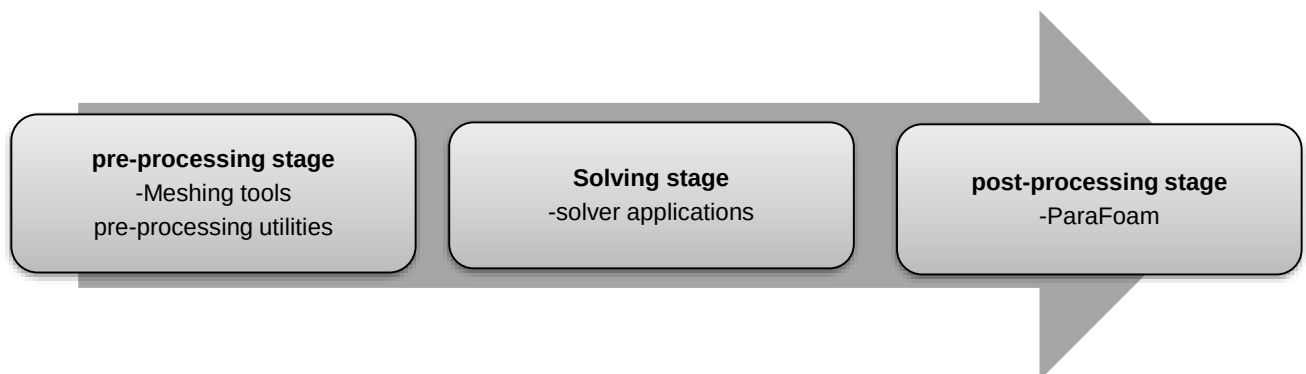


Figure 3-8: OpenFOAM™ CFD toolbox.

OpenFOAM™ operates in a three-dimensional Cartesian coordinate system, therefore all geometries are generated and cases are solved in 3D by default (*OpenFOAM*™ manual U-21). A two-dimensional solution can be instructed by specifying an empty boundary condition. This section of chapter 3 describes how the Euler-Euler multiphase model was used to predict the behaviour of a fluidised bed using the *twoPhaseEulerFoam* solver. This modelling section is structured as follows for both the laboratory and pilot-scale models:

Phase 1: Single phase model to characterise each distributor plate.

Phase 2: Two-phase fluidised bed system replicating experimentally observed behaviour.

Phase 3: Investigation of air maldistribution by measuring bed density at a different position across the duct and various heights above the plate.

3.4.1 Single-phase fluidised bed

A square fluidised bed geometry set up to scale, and without bed material, was used to determine pressure drops across the distributor plates. The geometry is represented by Figure 3-9 below.

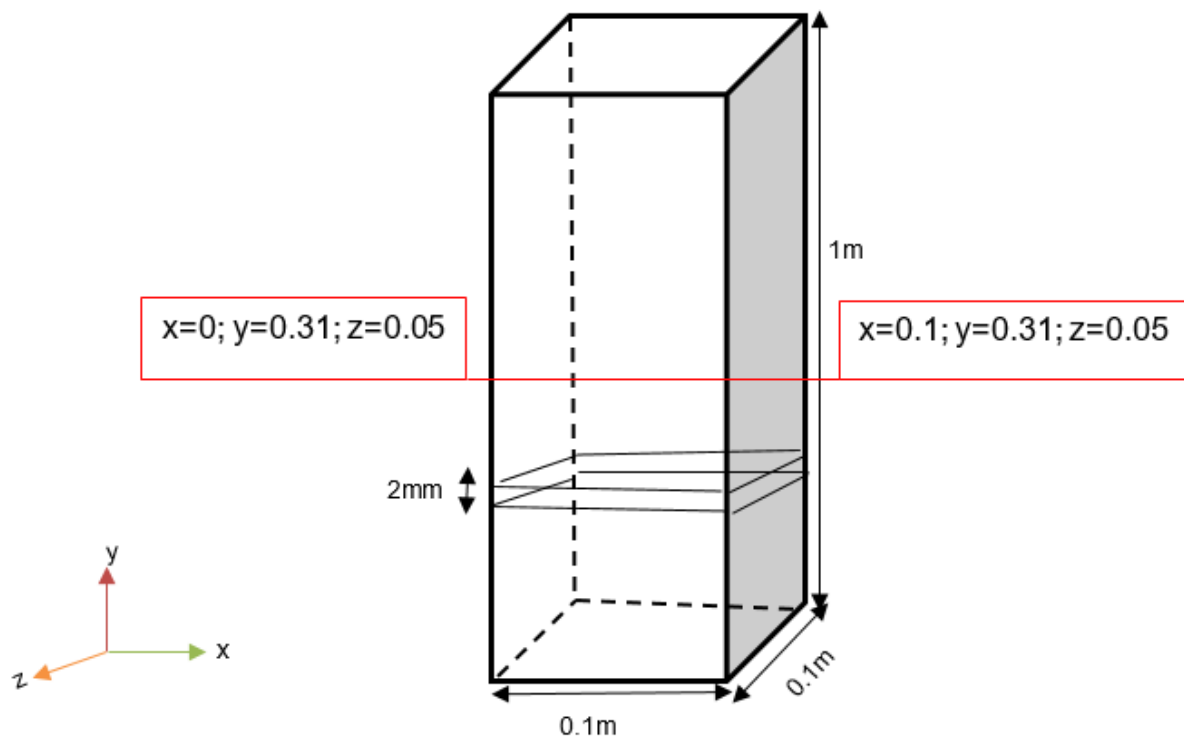


Figure 3-9: Geometry of the single-phase system with perforated plate characterisation

The distributor plate was specified as a porous medium by employing the Darcy-Forchheimer (D-F) coefficients as indicated in *OpenFOAM*[™] entry A-1.7 in Appendix A1. The case of *porousSimpleFoam* is structured diagrammatically as shown in Figure 3-10. The system folder allows for the specification of vertices and coefficients to generate the geometry of the fluidised bed modelled and also determines the run time and solver used to solve the equation and number of iterations. The properties of the porosity media are specified in the constant folder. The zero folder consists of the initial conditions of temperature (T), pressure (P), velocity (U) and wall effects (K and Epsilon) in the system.

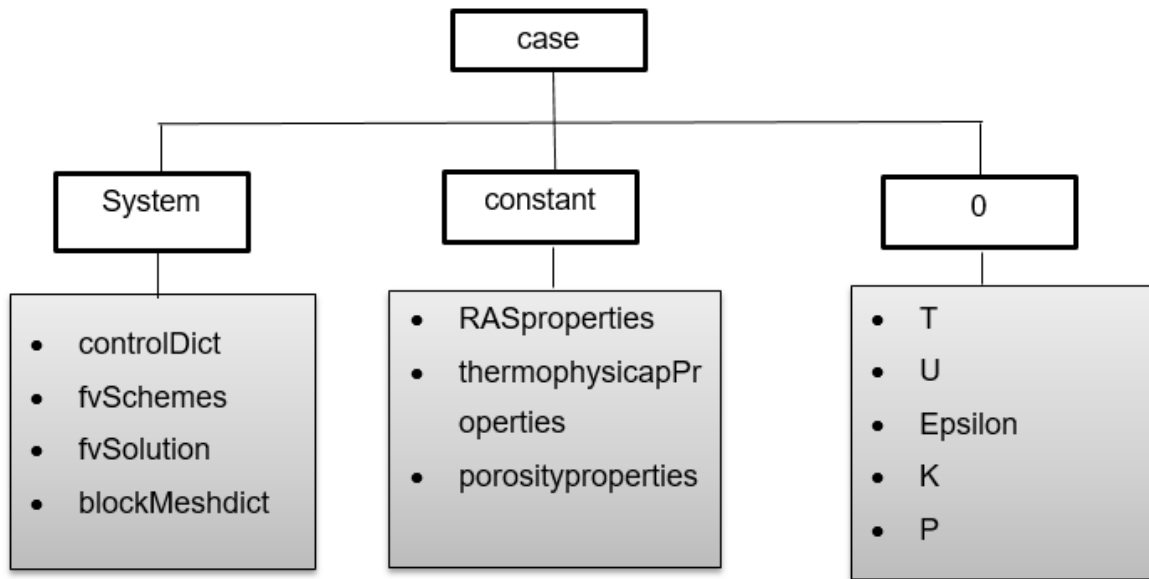


Figure 3-10: Case structure of *porousSimpleFoam*.

Pre-processing

(i) Mesh generation

A column with a length of 0.1 m, a width of 0.1 m and a height of 1 m was generated as fluidised bed geometry. The length, width and height of the block had 5, 20 and 5 cells respectively. The thickness of the perforated plate in this model is 2 mm. The *blockMesh* mesh generator supplied by *OpenFOAM*[™] generates meshes using a specified description in the input dictionary file (*blockMeshDic*) located in the system folder. The mesh is first converted to a specific unit, which is meters in this case. The coordinates of the block are specified as the vertices, followed by definitions of the three blocks. There are no edges, arcs or cyclic characteristics in a rectangle, therefore, edges are not specified. The boundary patches (inlet and outlet points of the system) and walls are stated. There are no faces, patches or walls to be merged therefore *mergePatchPairs* is not specified. To account for distributor plates in the system, porosity walls are specified both in the vertices, blocks and boundary patches.

(ii) Boundary and initial conditions

The boundary and initial conditions of the case are stored in the *constant* and *0* (zero time) folders respectively. The field dimension in m/s is specified by *dimensions* for velocity. A detailed explanation of the seven entries in dimensions found in the user-guide (section 4.2.6) are depicted by Table A-2.1 in Appendix A-2. The field in the file called "*internalField*" is used to specify the value of the air velocity into the system. For this research, the inlet air velocity in the y-direction is specified

and varies from 0.3 m/s to 1.7 m/s. Specified initial conditions include temperature, pressure, K and Epsilon wall effects. These entries can be found in OF entry A-1.1 to A-1.6 in Appendix A-1.

(iii) Physical properties

The physical properties, in this case, are stored in the *constant* folder. The porosity wall entries specify the resistance of fluid through the plate, thus accounting for different air distributor plates. Table 3-2 and Table 3-3 consists of the values of variables *a* and *b* as well as the corresponding Darcy-Forchheimer (DF) coefficients for the different plates in laboratory and pilot scale, respectively. The DF coefficients were determined from the experimental work where a graph of differential pressure against superficial gas velocity was plotted for each of the four plates. A second order polynomial curve was fitted to the graph as shown in Figure 3-10. From equation 3-1 where P_{plate} is the plate pressure drop in Pascal (Pa) and v_g is the superficial gas velocity in meters per second (m/s) , the values of *a* and *b* where obtained from the polynomial line equation and used to determine the DF inertial loss (f) and viscous loss (d) parameters using equations 3-2 and 3-3, respectively;

$$-\frac{dP_{plate}}{dX} = av_g^2 + bv_g \quad 3-1$$

$$f = -\frac{2a}{X} \quad 3-2$$

$$d = -\left(\frac{b}{\mu}\right)/X \quad 3-3$$

Here, *X* is the thickness of the distributor plate in meters and μ is the viscosity of air in meter squared per second.

Table 3-2: Darcy-Forchheimer constants and coefficient for laboratory scale distributor plates.

	Plate A	Plate B	Plate C	Plate D
a	28.9	40.0	106.5	419.7
b	1.71	7.47	7.47	80.20
Darcy-Forchheimer coefficients				
d	5.7e6	2.5e7	2.5e7	2.7e8
f	2.9e4	4.0e4	1.1e5	4.2e5

Table 3-3: Darcy-Forchheimer coefficients of distributor plates used in the pilot-scale setup.

	Plate A	Plate B	Plate C
a	30.5	367.7	190.4
b	2.2	36.5	21.0
Darcy-Forchheimer coefficient			
d	5.8e7	9.9e7	5.8e8
f	3.0e4	3.7e5	1.9e5

(iv) Control

The control dictionary (*ControlDict*) folder located in the *system* directory consists of the data relating to time control, including reading and writing of the solution data. The start, stop and step time of the case are set here. In this porous Fluidised bed case, the start time for the run is $t=0$ s for OpenFOAM™ to read data from the *0* directory folder. A time step is specified as deltaT of 0.0001 to achieve accuracy and stability when running the simulation. The simulation was run for 100s. The solver used for this application is *simplePorousFoam*.

(v) Viewing the mesh and running the application

Mesh is then viewed in a post-processing OpenFOAM™ tool called *ParaFOAM*. After the mesh is generated without errors, the application can be run in order to solve the problem using the solvers and iterations specified in *fvSchemes* and *fvSolution*.

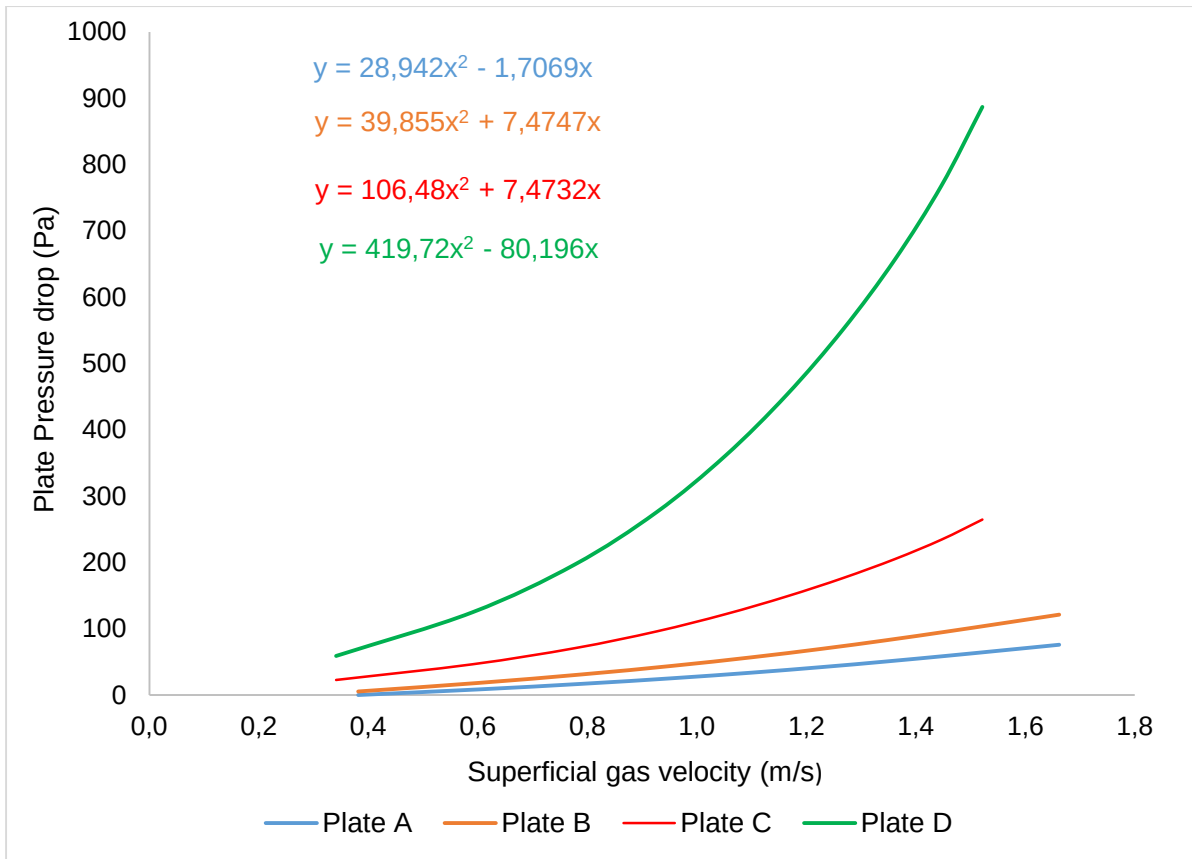


Figure 3-11: Plate pressure drop against the superficial gas velocity of model and experiment data

3.4.2 Single phase fluidised bed with the bend inlet duct

To investigate maldistribution of air through the distributor plate, a fluidised bed air-feeding duct with an elbow below the distributor plate was modelled. The behaviour of the distributor was accounted for by the velocity profile generated using this configuration. The structure of the *simplePorousFoam* is similar to that described in chapter 3.4.1. The geometry of the fluidised bed was simulated with the superficial gas flow in the x-direction. The porosity of the plate is specified with a magnitude similar as in section 3.4.1(ii), however, the only difference is the vector determining the direction of flow. The geometry of the bent inlet duct is depicted in Figure 3-12 below. The velocity profile obtained by using the four distributor plates was measured at the points indicated in Figure 3-13(a).

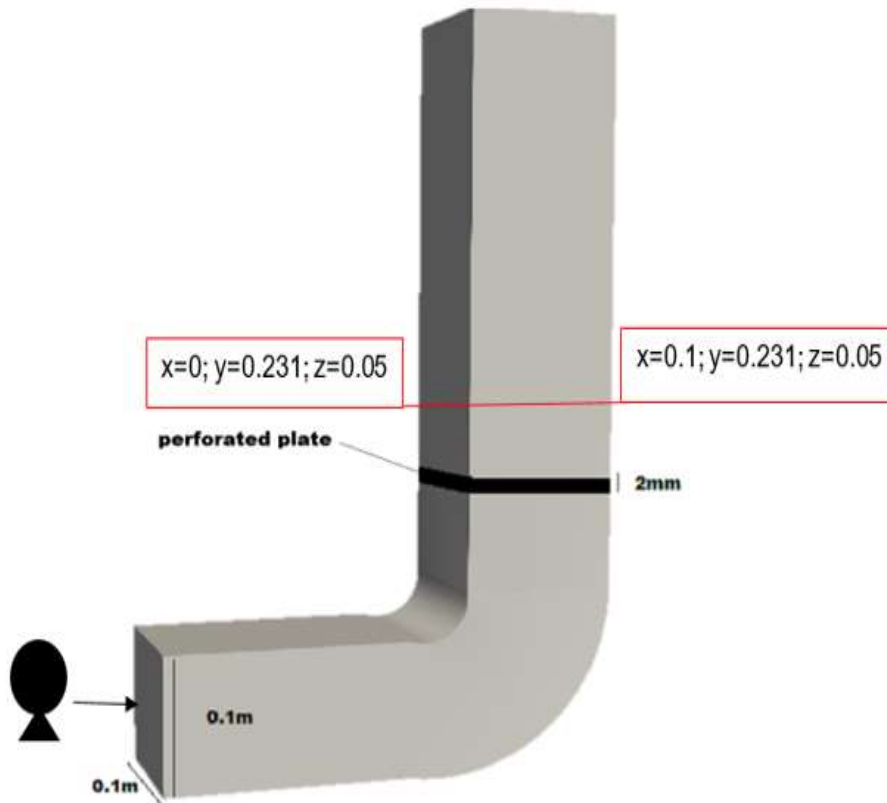


Figure 3-12: Geometry of a bend inlet duct using a perforated plate in a single-phase system.

The distributor plates are characterised in sections 3.4.1 and 3.4.2 to obtain the velocity profiles schematically represented in Figure 3-13 and Figure 3-14. The effect of the distributor plates is investigated using three solver setups; ideal plate, perforated plate in a straight inlet duct and perforated plate in a bend inlet duct for laboratory scale and two fluidised bed setup; column shape and conical shape fluidised bed for pilot-scale in *OpenFOAM*™. In the laboratory-scale setup, a primary ideal duct (PID) configuration each time refers to a basic fluidised bed characterised with an ideal plate (which is characterised by a “*distributor plate*” with a perfect block pressure and velocity profile directly above it) as well as the straight uniform section below the distributor plate. Velocity profiles associated with the straight inlet duct (SID) and bend inlet duct (BID) are also used as described in detail in section 4.1 and also outlined in the source code in *OpenFOAM*™ entry A-2.25 of Appendix A.2. A conical-shaped duct (Con-SD), which is setup of the sponsor and column-shaped duct (Col-SD) used merely to observe if there is any impact that the structure of the fluidised bed has on the performance of depicted in Figure 3-14, were used in the pilot scale investigation.

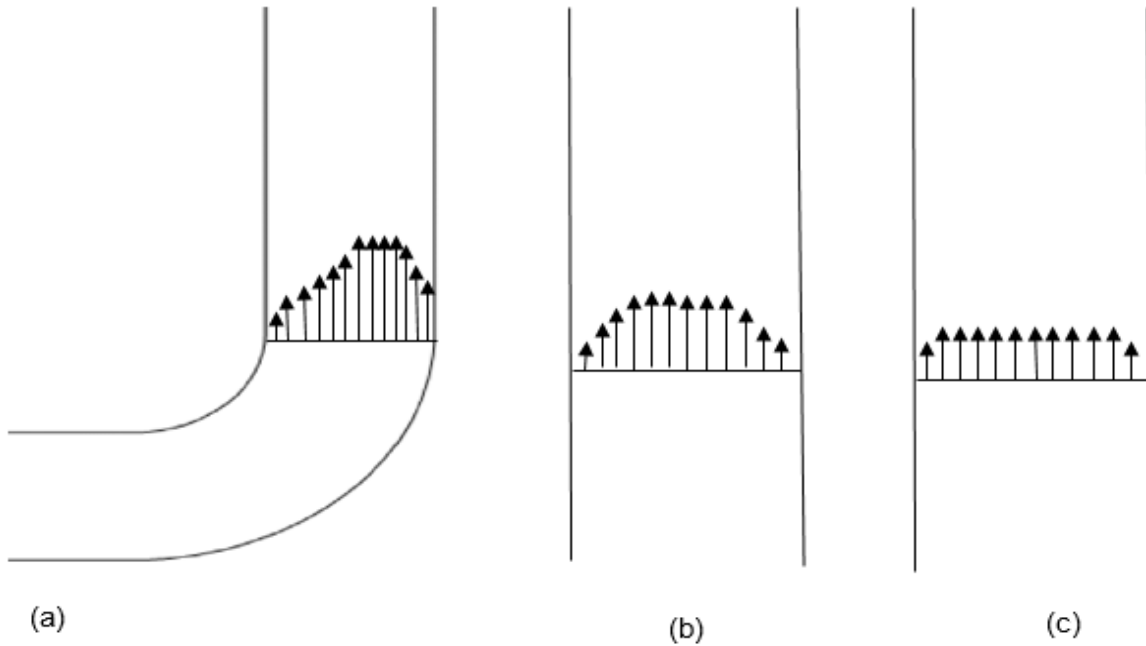


Figure 3-13: Typical velocity profiles associated with various laboratory fluidised bed duct setup (a) bend inlet duct (b) straight inlet duct and (c) primary ideal duct

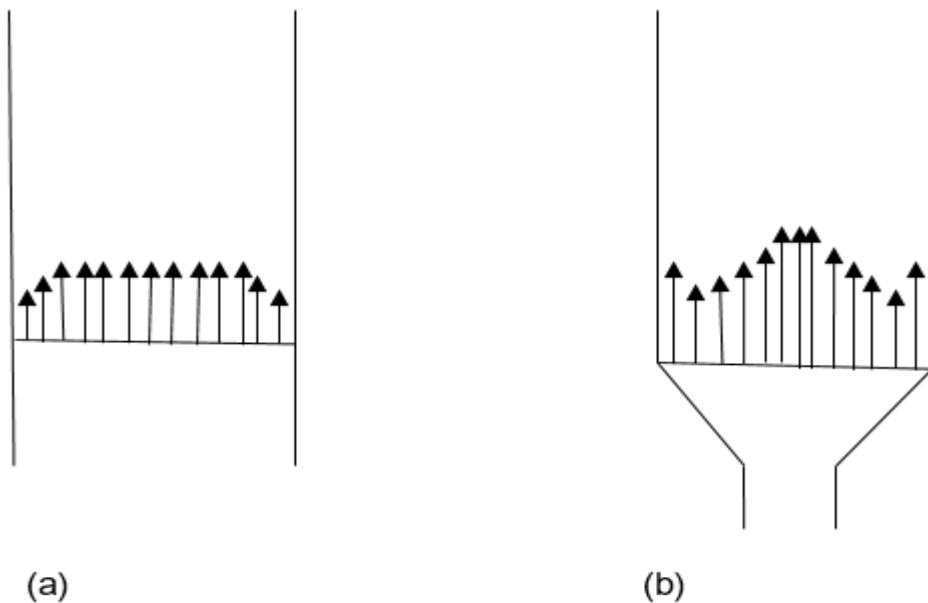


Figure 3-14: Typical velocity profiles associated with various modelled pilot-scale fluidised bed duct setup (a) column-shaped duct and (b) cone-shaped inlet duct

Figure 3-13 and Figure 3-14 are a schematic representation of typical velocity profiles obtained from the plate characterisation outlined in section 3.4.1 and 3.4.2 above. The velocity profiles are displayed, discussed and described in detail in section 4.1 for both laboratory and pilot scales. The velocity profiles were used to generate sixth order polynomials used to specify the initial boundary

condition for use in modelling of the fluidising of particles in a fluidised bed environment. The velocity of the inlet gas is provided in the file *0/U.air*. For this study, a list of vectors was generated in a compiled solver with the code found in appendix A-2. A polynomial generated in chapter 3.4.2 is specified as the velocity function in the source code. A text file consisting of 30 velocity vectors describing the flow pattern in the y-direction resulting from the polynomial within a range of time in the fluidised bed is generated. The vectors are then specified in the initial gas velocity file as demonstrated in *OpenFOAM™* entry A-3.1 of Appendix A.3.

3.4.3 The two-phase fluidised bed system

Fluidisation of granules in the fluidised bed is modelled as a two-dimensional column with a height, width (Specified as empty patch in mesh generation) and length corresponding with the actual fluidised bed. Two blowers are used to blow fluidising gas from the bottom into the system. The distributor plate is accounted for by setting a velocity profile at the bottom of the fluidised material bed as a boundary condition, as obtained from the behaviour of the plate described in section 3.4.2. Figure 3-15 outlines the structure of a *twoPhaseEulerFoam* case. The static bed height is 10 cm and 45 cm for laboratory and pilot-scale respectively, with a solid fraction of 55%. Figure 3-16 shows the basic geometry of the fluidised bed column and Table 3-3 gives the parameters that will be used.

The mesh is generated similarly as described before but refined to fit the geometry. The cell sizes are chosen to be 20, 300 and 1 in the x, y and z-direction respectively. The properties of the particles elaborated in Table 3-4 are defined in the *constant/phaseProperties* and *constant/thermophysicalProperties.particle* files. The maximum solid package is set at 62% since the general guideline for maximum solid packing is 63% to avoid dense flow and infinite collisions between particles (Bush, 2015). Initial boundary conditions are defined in the *0* folder. The volume fraction of the solid bed material in the *0/alpha.particles* file is generated through the *setFields* function above. The initial pressure is set to 1e5 Pa which is an approximated atmospheric pressure. *0/Theta* consists of the calculated granular temperature. *0/nut.air* and *0/nut.particles* define the calculated turbulent eddy viscosity of the gaseous and solid phases respectively. *Epsilon* and *K* are functions of turbulent dissipation and turbulent kinetic energy for wall treatment. The solids velocity is not specified whilst the gravitational acceleration is set to the standard value of 9.81 m.s². The turbulence modelling of the fluidised bed is turned on in the *constant/turbulentProperties.air* file. The simulation runs for an equivalent operating time of 2 seconds.

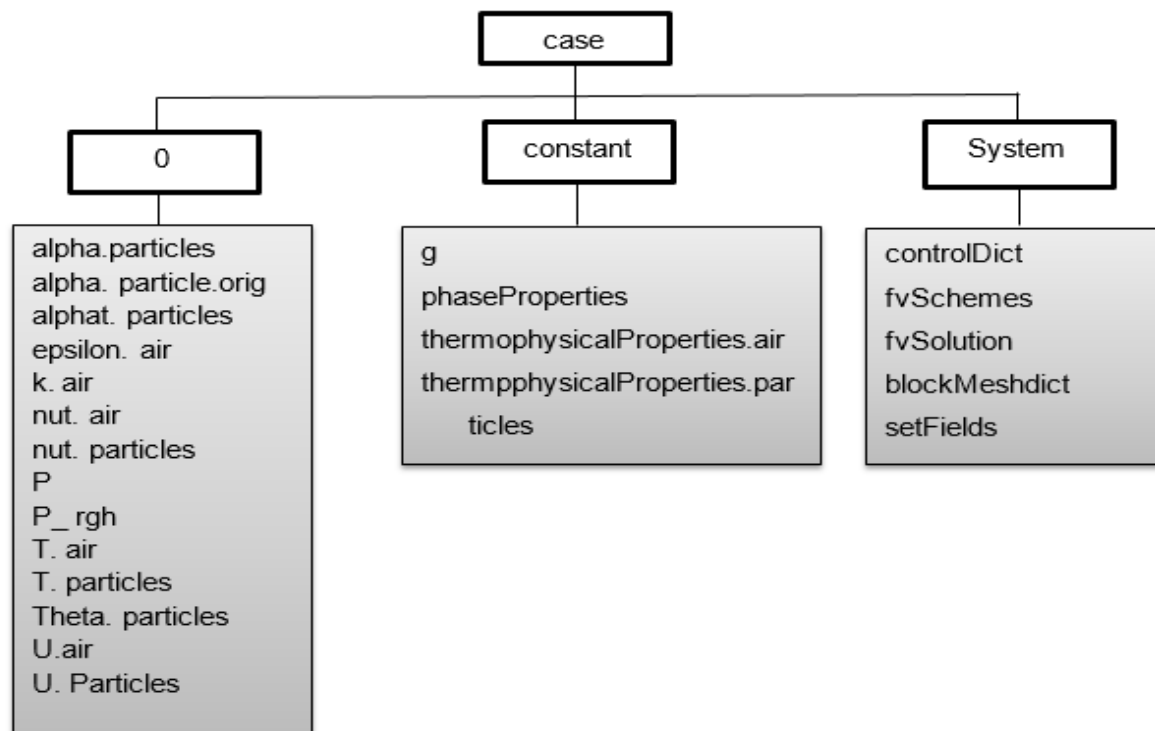


Figure 3-15: Structure of the *twoPhaseEulerFoam* case in OpenFOAM™

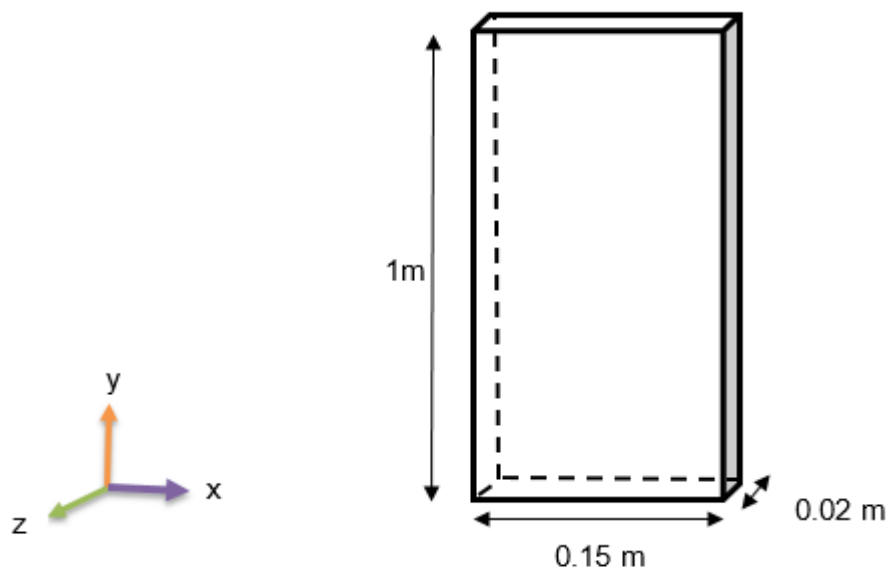


Figure 3-16: Two-phase fluidised bed geometry.

Table 3-4: Input parameters used in the CFD model

parameter	value
Solid average density (kg/m ³)	1561.5, 1428, 1388, 1356 and 1428
Solid average diameter (mm)	1.28, 1.60, 1.70, 1.85, 2.06 and 2.62
Initial solid volume fraction (%)	55
Superficial gas velocity (m/s)	0.3, 0.6, 0.8, 1.1, 1.3 and 1.7
Static bed height (m)	0.1 (laboratory scale) & 0.45 (pilot scale)
Particles molecular weight (g/mol)	80.043

The drag model used for air-particle interaction is *GidaspowErgunWenYu* with a residual of $1e-3$. A smooth solver was used to solve for alpha particles and alpha air, the velocity of air and velocity of particles, epsilon, nut, and K while GAMG was used for pressure. The static bed height in the system is defined by the “*setFields*” option in the systems folder. The void fraction of air and particles are both set in the *system* folder. The default setting of void fraction of air and particles is 1 and 0, respectively. The static bed height is set to 10 cm with a void fraction of 55 % in the specific region of the fluidised bed. The “*setFields*” of particles are backed up in *0/alpha.particles* where these are executed.

3.5 Chapter summary

This chapter covered the methods followed to carry out the investigation, starting with representation and discussion of equipment and material used for experiments in both the laboratory and pilot-scale setups. This was followed by the procedure in which the experiments were conducted and methods of measuring and collecting data.

The simulation procedure was outlined for different setups investigated for both characterising of the distributor plate and also investigating bed hydrodynamics of the AN fluidised bed. At the end of the section, a qualitative representation of the behaviour of distributor plates was outlined.

The following chapter covers the results and discussions from modelling results validated with experimental results in an attempt to realise the objectives of the study.

4. RESULTS AND DISCUSSION

The performance of the distributor plates used in fluidising ammonium nitrate granules is an important aspect in this study, as fluidisation behaviour will be a consequence of the characteristics of the distributor plate as well as the equipment geometry. In section 4.1, the generation of velocity patterns directly above the distributor plate resulting from the plate design and system geometry is discussed as an important boundary condition for the fluidisation model discussed in section 4.2. In section 4.2, details of the experimental versus modelling behaviour in the fluidisation of ammonium nitrate granules, including bed differential pressure (4.2.1), flow patterns (4.2.2), bed expansion (4.2.3) and bed density (4.2.4) are discussed for two different scales of investigation.

4.1 Characterisation of the air distributors

4.1.1 Laboratory-scale plate characterisation

Data were collected from the model by viewing the model results in the post-processor, known as *ParaView™*. Using a “plot-over-line” function, two pressure values taken at two different points, one directly below the distributor plate and the other directly above it, were recorded and subtracted from each other to calculate the plate pressure drop at a number of superficial gas velocities investigated. The pressure differential measurements were taken in a fluidised bed without bed particles. Figure 4-1 is a graph of plate pressure drop (Pa) against superficial gas velocity (m/s). The solid lines and the markers represent the model and experimental results respectively. The blue lines and markers on the graph are results for plate A (pitch = 5.5 mm, orifice = 1.7 mm), while plate B (pitch = 1.7 mm, orifice = 2 mm), plate C (pitch = 9.3 mm, orifice = 2mm) and plate D (pitch = 11.9 mm, orifice = 2 mm) are represented by the orange, red and green colours respectively.

The plate pressure drop is observed to increase with increasing superficial gas velocity for each of the plates. The pressure drop of plate A has the lowest maximum plate pressure drop of 80 Pa at the highest induced superficial gas velocity (V_g) of 1.66 m/s. The error between the experimental data and modelled data is high at low superficial gas velocities and reduces significantly at V_g values of approximately 1 m/s - 1.7 m/s. The highest plate pressure drop of plate B is 110 Pa at the highest superficial gas velocity. Contrary to plate A, the error on plate B, which represents the difference in modelled and experimental values, increases with superficial gas velocity. For plate C, the percentage error is relatively large at the superficial gas velocities below 1 m/s and at velocities higher than 1.5 m/s. The maximum plate pressure drop of plate C is three times higher than for plate A. Plate D exhibits the highest maximum plate pressure drop of 960 Pa, compared to the other three plates. The behaviour of plate D can be closely related to that of plate B with regards to the perceived error.

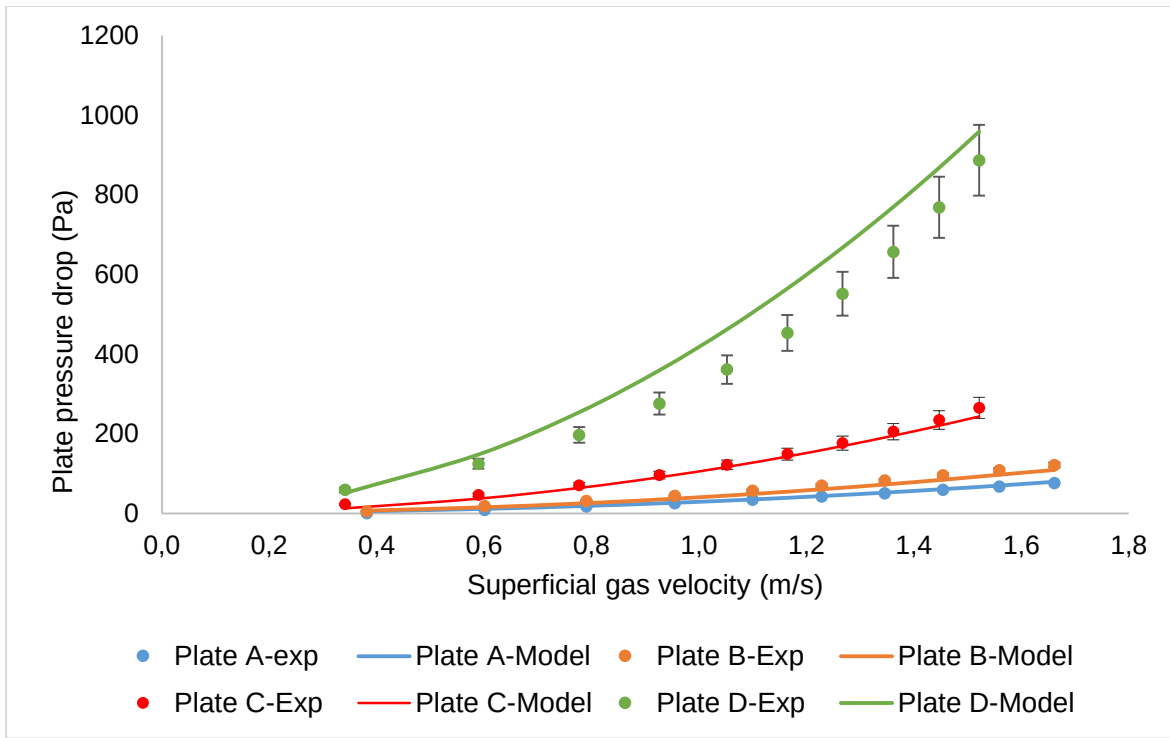


Figure 4-1: Comparison of plate pressure drops of experiments and model results against superficial gas velocity

With the results above, with the exception of plate D where the modelled data differ to some extent from the measured values, confidence can be placed in the porous medium model in representing the distributor plates' behaviour. In order to expand on the characterisation of a porous media in a straight inlet duct system (i.e. the exclusion of the usual elbow found beneath the distributor), the data obtained along a straight path across the walls of the duct at coordinated points indicated on Figure 3-10 was collected to determine the velocity profile that represents the velocity behaviour of the plate. The measurement point over the line for the bed inlet duct is marked with a red line and coordinated in Figure 3-12. The velocity profiles of plate A at different superficial gas velocities in both a straight inlet duct (SID) and a bend inlet duct (BID) configuration are represented by Figure 4-2 and Figure 4-3 respectively. Velocity profiles of plates B, C and D can be found in Appendix B-1.

In Figure 4-2 and Figure 4-3, the velocity profiles that illustrates the flow behaviour through plate A when fluidising with superficial gas velocity boundary conditions of 0.3 m/s, 0.6 m/s, 0.8 m/s, 1.1 m/s, 1.3 m/s and 1.7 m/s is represented by the blue, orange, red, yellow, purple and green profiles respectively. It can be seen that the velocity profile in a SID (Figure 4-2) is a relatively consistent block profile with velocity reduction towards the edges of the plates at the stationary walls. In the case of the SID, the velocity profile is developing very symmetrical, while in the BID, the developing profile is very much similar to the SID profile, with the exception that it is slightly skewed to the opposite side of where the blower is situated., the velocity profiles in a BID are developing parabolic

shapes, slightly skewed to the right, which is towards the opposite side from where the air blower would be situated. As the superficial gas velocity increases, the peak of the profile increases. Similar patterns were observed for plates B, C and D in a BID.

Sixth-order polynomials were fitted to the velocity profile curves to generate a velocity profile function that could be used as a boundary condition during fluidisation modelling. For a SID, the polynomial functions of different plates at different superficial gas velocity exhibited a negligible difference. This is visually observed as the maximum velocity reached was 2.5 m/s with identical shapes. As a result, instead of running simulations for each of the four plates, a generic plate profile was developed with an average polynomial function for the SID.

For the BID, even though the behaviour of the plates is similar, the polynomial functions differ significantly from each other, necessitating individual functions for all four plates. For the distributor plates investigated in a BID, the maximum velocity reached was 2 m/s. The polynomial functions of all the plates in both SID and BID configurations are included in Appendix B-1.

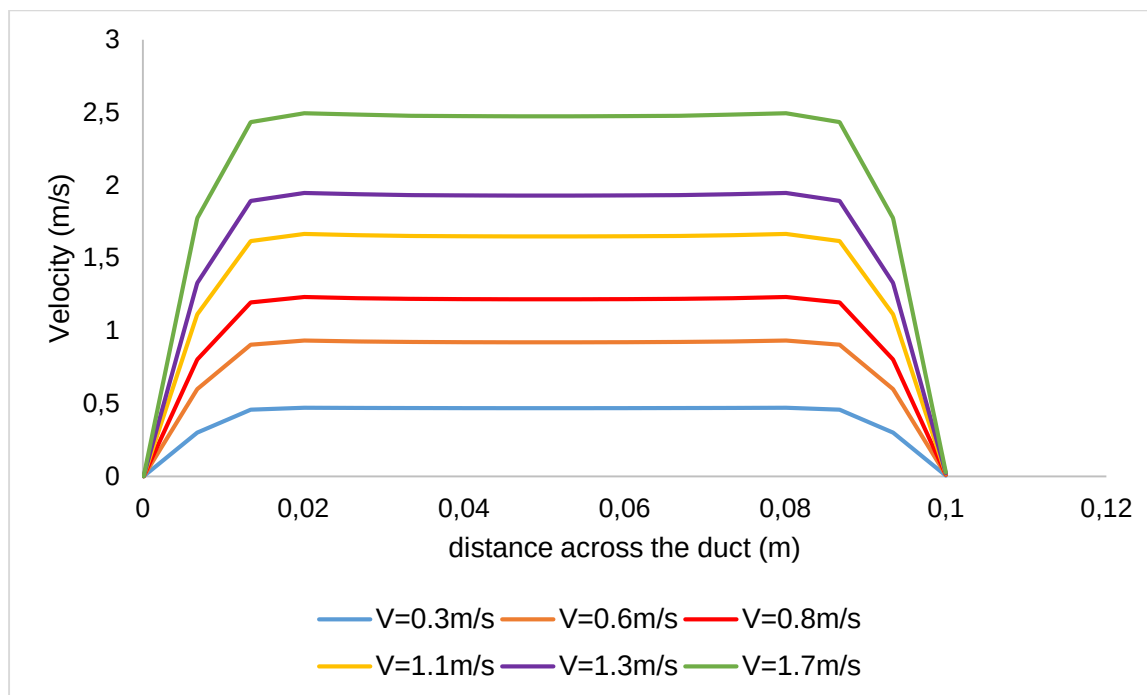


Figure 4-2: Model velocity magnitude of plate A at different superficial gas velocities in the straight inlet duct

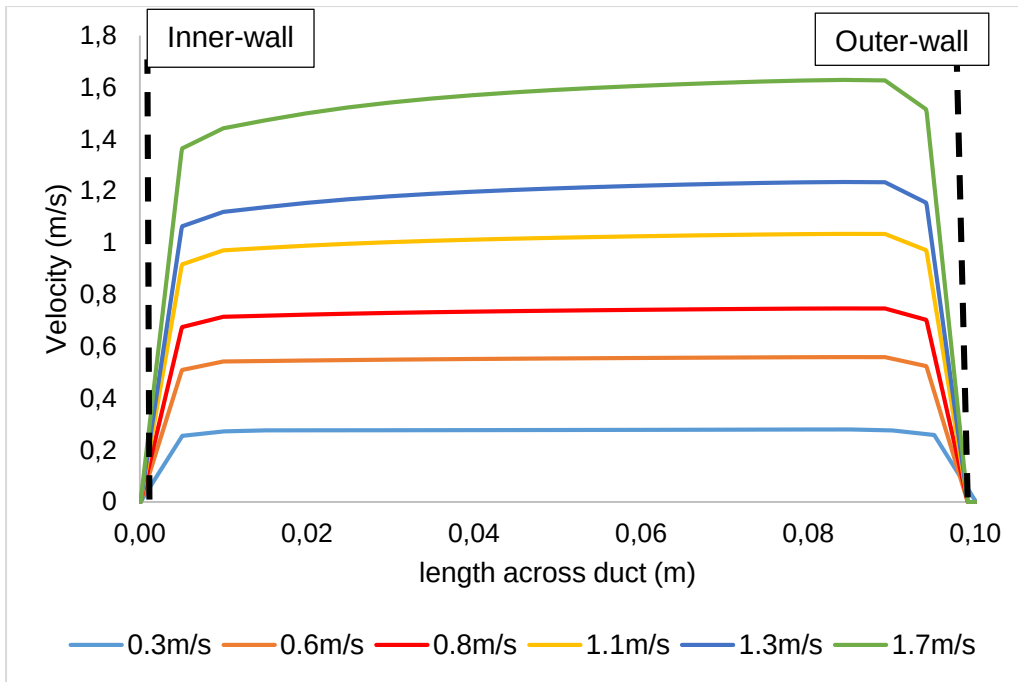


Figure 4-3: Model velocity magnitude of Plate A at different superficial gas velocities in a bend inlet duct

4.1.2 Pilot-scale plate characterisation

The three distributor plates (plate A, plate B and plate C) were characterised in a pilot-scale experimental facility. The procedure is similar to the above method followed for the laboratory scale. The Darcy-Forchheimer coefficients were calculated from the plate pressure drop against superficial gas velocity obtained from experiments, similar to the procedure elaborated on in section 4.1. The Darcy-Forchheimer coefficients were used in the *porousSimpleFoam* model in the porosity property to specify the resistance of each of the plates.

Figure 4-4 depicts the plate pressure drop against superficial gas velocity where the blue, orange and red lines represent the curves of plate A, plate B and plate C respectively. The solid lines and markers represent the modelled data and experimental curves respectively. The initial and final V_g when determining the plate pressure drop of plate A was 0.47 m/s and 2.62 m/s respectively. From Figure 4-4 it can be deduced that the plate pressure drop increases with superficial gas velocity. The model predicts experiments better at lower superficial gas velocity and the accuracy decreases with increasing V_g . When investigating plate B and plate C, the superficial gas velocity ranges between 0.21 m/s and 2.25 m/s. The trend of plate pressure drop against superficial gas velocity was also observed in the studies of Son et al (2005) and Kelkar et al (2016). The pressure drop increases with increase in velocity.

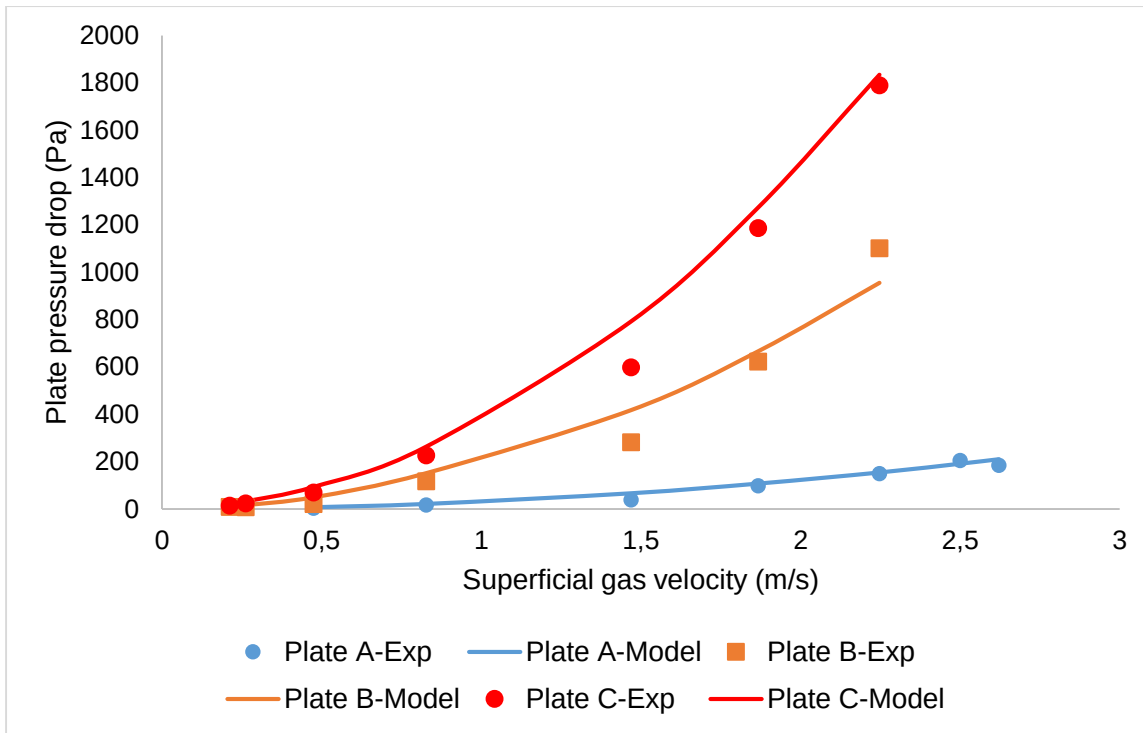


Figure 4-4: Plate pressure drop against superficial gas velocity in a pilot scale setup

The *porousSimpleFoam* solver in *OpenFOAM*TM was used to simulate cases using two different duct geometries below the distributor plate in the fluidised bed, i.e. a column duct (i.e. no widening of the duct) and cone-shaped duct (i.e. widening of the duct towards the plate). The velocity profiles were determined in the model to specify the initial behaviour assumed by the employment of each distributor plate in the system. The velocity profiles for plates A and C were determined in both a cone-shaped duct and a column duct for three superficial gas velocities. For plate B, the velocity profile was determined only in a column duct. Figure 4-5 and Figure 4-6 depict the velocity profiles of plate A produced in a column-shaped duct and cone-shaped duct fluidised bed respectively. The blue, green and red plots are velocity profiles obtained at 0.83 m/s, 1.47 m/s and 1.87 m/s, respectively.

In Figure 4-5, a rapid increase in velocity was observed between the wall of the duct, highlighted as 0 on the figure and a point 0.1m from the wall. At lower velocity boundary conditions, the expected parabolic flow profile seems to develop slower pace with multiple points of constant velocity profile development than at higher velocity boundary conditions, as the maximum velocity above the plate is achieved further from the walls at lower velocity boundary conditions.

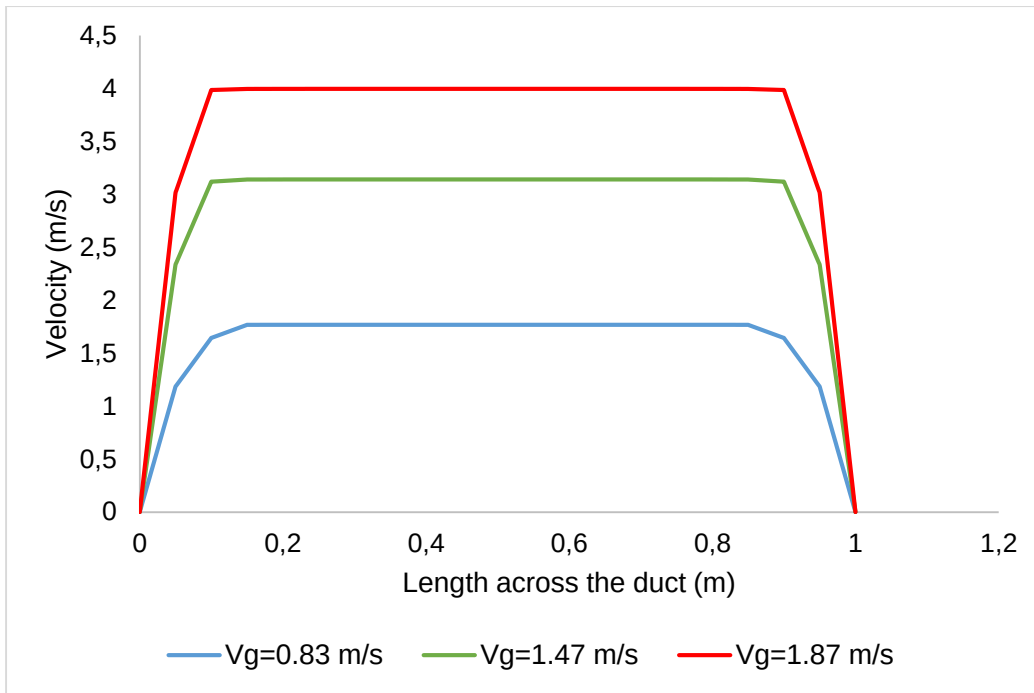


Figure 4-5: Model velocity magnitude of plate A in a column-shaped duct with different superficial gas velocities

A constant velocity area between 0.1 m and 0.9 m across the length of the duct was observed, which is indicative that the point of measurement is so close to the distribution plate that a fully developed parabolic profile is not yet perceived.

The profile described in Figure 4-5 will be termed a block velocity profile for all the superficial gas velocities investigated. The maximum velocity ($V_{\max}$) reached when fluidising with superficial gas velocities of 0.83 m/s, 1.47 m/s and 1.87 m/s were 1.4 m/s, 3.2 m/s and 4 m/s respectively. The velocity profiles produced by plate B and plate C in a column are represented by Figure B-2.5 and Figure B-2.6 included in Appendix B-2. The maximum velocity reached using plate A was 4 m/s and 5.5 m/s was obtained for plate B and plate C.

The velocity profile of plate A in a cone-shaped duct is depicted in Figure 4-6. For $V_g = 0.83$ m/s, a rapid increase in velocity was observed immediately adjacent to the walls. Immediately after the high velocity is reached adjacent to the wall, a velocity decrease towards the centre of the bed is observed with a local velocity maximum at the centre of the fluidised bed system. This phenomenon is observed to be pronounced at a velocity boundary condition of 1.47 m/s and 1.87 m/s.

Figure 4-7 depicts the velocity profile behaviour formed with place C in a conical-shaped inlet duct (i.e. a conical-shaped duct configuration immediately below the distributor plate). It can be seen that the trends follow a similar pattern, which is similar to that fashioned by plate A. The exception here is that $V_{\max}$ in plate C has quite a high magnitude and the wave-shaped profiles seem to be a bit smoother than for plate A.

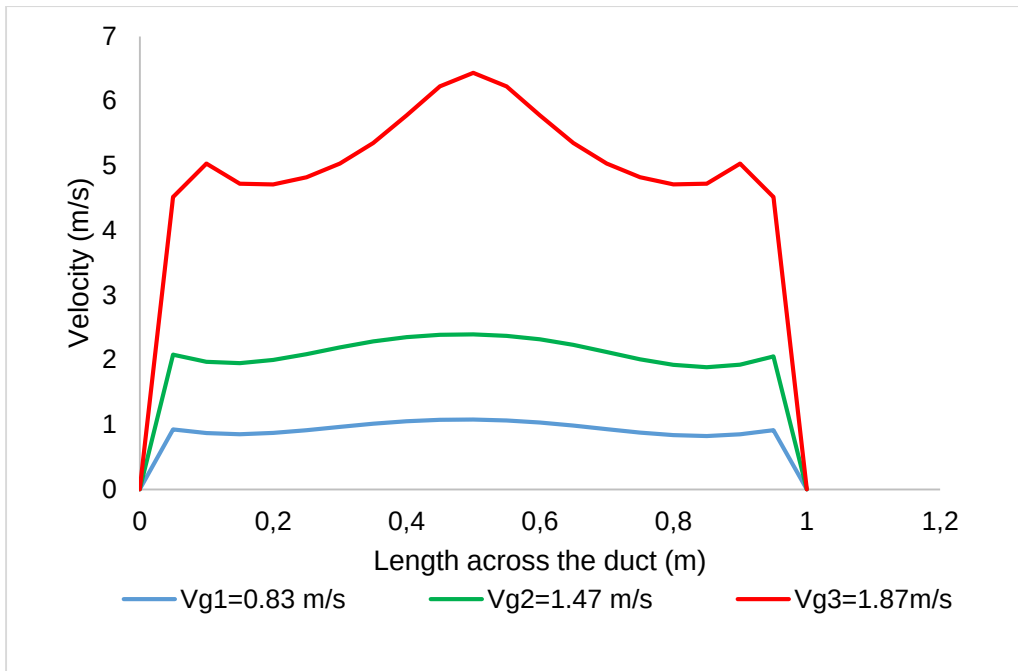


Figure 4-6: Model velocity magnitude of plate A in a conical-shaped fluidised bed duct with different superficial gas velocities

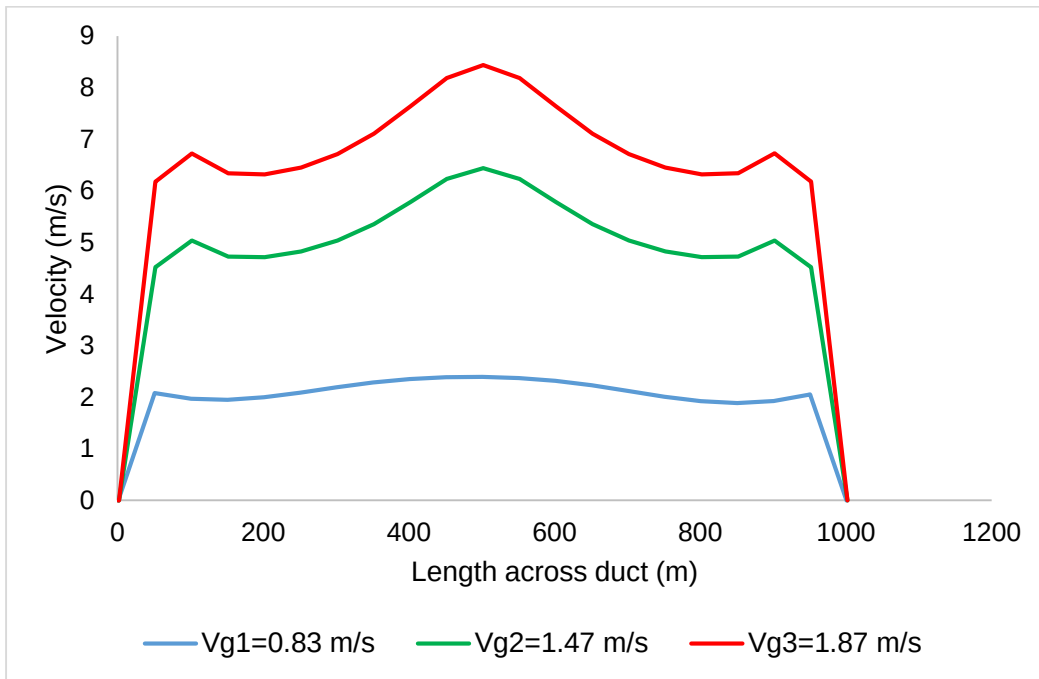


Figure 4-7: Model velocity magnitude of plate C in a conical-shaped fluidised bed duct with different superficial gas velocities

4.1.3. Comparison between plate attributes at different scales

Similar distributor plates have been used in both the pilot-scale and laboratory-scale facilities. Figure 4-1 and Figure 4-4 show that the plate pressure drop increases with superficial gas velocity. In the laboratory scale, the pitch of the plate also affects the pressure differential; an increase in distributor

plate pitch results in an increase in the plate pressure drop. This is e.g. observed at 1m/s, where the pressure drop of plate A, plate B, plate C and plate D is found to be approximately 30 Pa, 40 Pa, 100 Pa and 400 Pa, respectively.

In the pilot-scale setup, the plate pressure drop at 1m/s for plate A, plate B and plate C was found to be approximately 40 Pa, 210 Pa and 400 Pa. It can clearly be seen that the scale of the fluidisation setup affects the plate pressure differential. The plate pressure differential is observed to be higher and with a more rapid increase as a function of superficial gas velocity in the pilot-scale facility than in the laboratory-scale setup. the distributor plate pitch is perceived as a major factor influencing plate pressure drop in the pilot-scale setup. The velocity profiles observed in both the laboratory-scale and pilot-scale setups using a straight inlet duct (SID) and column-shape duct configurations below the distributor plate, display the expected block velocity profile which resulted in an identical initial boundary behaviour for modelling the fluidisation of the bed of particles in both scales.

4.2 Fluidisation of ammonium nitrate granules: comparison between model and experimental data

4.2.1 Bed pressure drop & minimum fluidisation velocity

Bed pressure drop and fluidisation velocity are closely related in such a way that the bed pressure drop against superficial gas velocity curve can be used to determine the minimum fluidisation velocity.

For a variety of different average particle sizes, the behaviour of three modelled plates was compared with actual measurements. These modelled perforated plates include (i) plates with a velocity profile associated with an ideal duct, which has a constant magnitude for the velocity vector irrespective of the position on the plate, (ii) plates with a velocity profile associated with a straight inlet duct as observed in Figure 4-2 and (iii) plates with a velocity profile associated with a bent inlet duct as observed in Figure 4-3. This was done to report on the behaviour of perforated plates in different setups, with the exclusion, on the laboratory-scale, of a conical-shaped inlet duct.

(i) Primary ideal duct

A two-phase fluidised bed employing a primary ideal duct (PID) i.e. a plate where the velocity vector has a similar magnitude irrespective of the position on the plate, was simulated with particles with characteristics listed in Table 3-1 at different superficial gas velocities. This ideal plate scenario was only subjected to modelling for the purpose of comparison with real cases, which is the reason for the absence of validation results.

Figure 4-8 depicts the bed pressure drop of particles with different sizes at different superficial gas velocities. The bed pressure drop curves of particles with average particle diameters of 1,3 mm, 1.6 mm, 1.7 mm, 2.1 mm and 2.6 mm are represented by the blue, orange, purple, yellow and red curves, respectively. The raw sample, i.e. unclassified material with a total average density of 1 432 kg/m³ and an average particle diameter of 1.87 mm, which is representative of a typical production sample, is represented by a green curve.

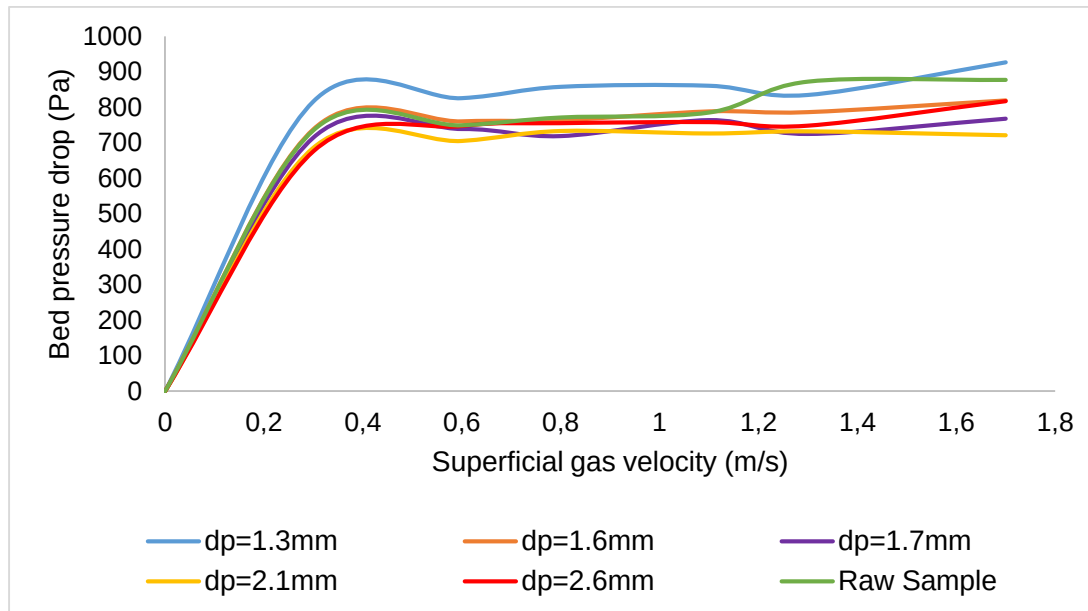


Figure 4-8: Model bed pressure drop of various particle sizes using a plate with a velocity profile associated with a primary ideal duct

Particles with a d_p of 1.3 mm exhibit a pressure drop turning point at approximately 830 Pa at a velocity of 0.4 m/s. The bed pressure drop curve stabilises at velocities higher than 0.6 m/s. With the maximum V_g of 1.7 m/s, the bed pressure drop has increased with approximately 100 Pa. The $d_p=1.6$ mm and raw sample particles have an equal bed pressure drop at 0.4 m/s, after which the curve becomes steady until 1.1 m/s where the bed pressure drop of the raw sample increases gradually while that of the $d_p=1.6$ mm particles' curve retain a somewhat constant increase. The bed pressure drop curve of the $d_p=1.7$ mm particles takes longer to stabilise. As illustrated in Figure 4-8, the steadiness initiates at 1.3 m/s. Particles with a $d_p=2.1$ mm and $d_p=2.6$ mm have the lowest bed pressure of 750 Pa at 0.6 m/s. A general pattern was observed; the minimum fluidisation velocity for all particles size ranges is observed at between 0.3 m/s and 0.5 m/s whereafter stabilisation initiates.

(ii) Straight inlet duct

The initial behaviour of the ideal PID was changed through specifying the behaviour of an actual distributor plate by means of a velocity profile function in a SID as boundary condition using the method described in section 3.4.2.

Figure 4-9 illustrates the bed pressure drop of particles with different particle sizes at various superficial gas velocities using the generic distributor plate with velocity profiles associated with a SID. The blue, orange, red, yellow, purple and green lines and markers represent the curves of particles with average particle diameters of 1.3 mm, 1.6 mm, 1.7 mm, 2.1 mm, 2.6 mm and the raw unclassified sample respectively. The lined curves and markers represent the bed pressure drop achieved from modelled and experiment data respectively.

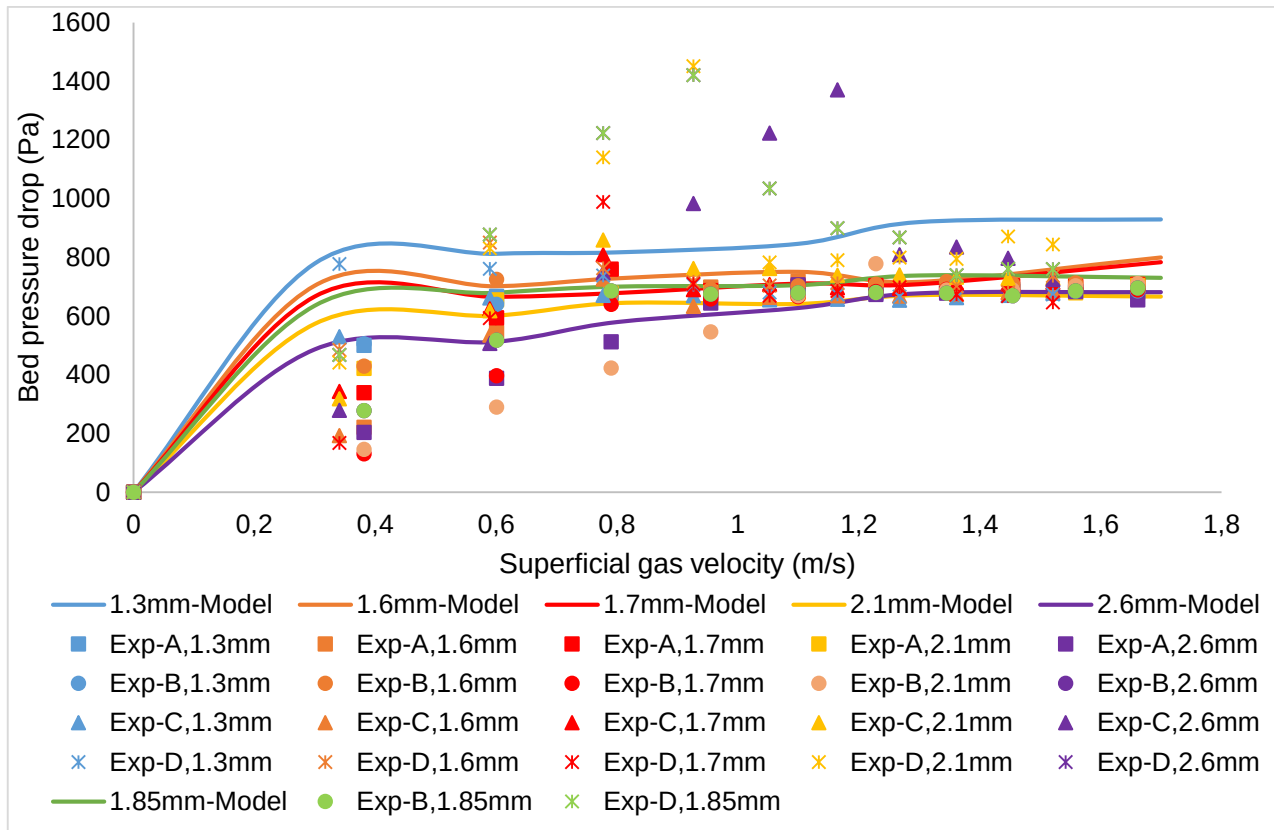


Figure 4-9: Fluidised bed pressure drop using plates with a velocity profile associated with a straight inlet duct

In the model, the highest bed pressure drop is observed at approximately 800 Pa when fluidising particles with an average particle diameter of 1.3 mm using a generic plate. The second highest bed pressure drop is reached with $d_p = 1.6$ mm and $d_p = 1.85$ mm (raw samples) simultaneously. Particles with high $d_p = 2.6$ mm showed the lowest bed pressure drop reading using both perforated plates. The trend of bed pressure drop curve mimics the typical behaviour discussed in section 2.2.2., with a rapid increase between 0 m/s and 0.3 m/s followed by turning point in the pressure drop where minimum fluidisation velocity can be graphically predicted. At 1.1 m/s, a gradual drop in bed pressure drop is observed for the generic with particle sizes $d_p = 1.6$ mm. The bed pressure drop starts to increase again at 1.3 m/s.

The model was validated with experimental data. The results show that bubbling fluidisation was achieved at lower fluidisation velocities in the model than during experiments as the minimum

velocity range was found to be less than 0.3 m/s for modelled data and above 3 m/s for experiments results. The U_{mf} range was even greater for plate B, plate C and plate D in the experiments. The highest range of 0.8 m/s – 1.3 m/s was observed for plate C when fluidising particles with $d_p = 2.6$ mm. The highest bed pressure drop of approximately 1 450 Pa was observed for plate D when fluidising particles with $d_p = 2.1$ mm and sharply drops to 750 Pa between 1 m/s and 2 m/s. The model gives a good prediction of plate D with 1.3 mm particles with an almost negligible error.

At 0.3 m/s, the bed pressure drop predicted by the model is greater than that obtained from the experiment results. The magnitude of the bed pressure drop obtained from experimental data keeps increasing until a maximum value is reached at different V_g per plate and particle size and drops between 1.1 m/s and 1.3 m/s when stability is being approached. For $d_p=2.6$ mm particles, the model underestimates the bed pressure drop of plate D. A similar outcome was observed for plate D with both 1.85 mm and 2.1 mm particles.

In general, it is observed that even though the model under-predicted the minimum fluidisation velocities of most of the particles sizes, bed pressure differentials at higher fluidisation velocities for particles with d_p less than 2.6 mm (i.e. during actual fluidisation) are described by the model.

(iii) Bend inlet duct

The behaviour of the four distributor plates in a bend inlet duct was subsequently investigated. Figure 4-10 and Figure 4-11 depicts the bed pressure drop of particles with various particle sizes using different distributor plates. The blue, orange, red, and purple curves represent the bed pressure drop of plate A, plate B, plate C and plate D respectively when fluidising particles with $d_p=1.3$ mm, $d_p=1.6$ mm, $d_p=1.7$ mm, $d_p=2.1$ mm, $d_p=2.6$ mm as well as the raw samples with $d_p=1.85$ mm. Figure 4-10 depicts the results when fluidising particles with a small average diameter whereas Figure 4-11 represents the bed pressure drop of particles with a larger average diameter. The lined curves and markers represent the modelled and experimental data respectively. It is observed that similar to the PID and SID cases discussed above, the bed pressure drop of the $d_p=1.3$ mm particles is higher compared to the other particle sizes in the superficial gas velocity range of 0.3 m/s to 0.6 m/s.

For modelled plate A, the optimum bed pressure drop of particles with $d_p=1.3$ mm is approximately 795 Pa in the velocity range 0.3-0.6 m/s, which is the range of the minimum fluidisation velocity. The bed pressure drop stabilises at 0.6-0.8 m/s and constantly increases to a value of 830 Pa at the point of 1.7 m/s. The second highest bed pressure drop at the superficial air velocity of 0.3-0.6 m/s is 660 Pa reached by using plate A in the fluidisation system, fluidising particles with an average particle diameter of 1.6 mm. The bed pressure drop increases as the superficial gas velocity increases. At the highest superficial gas velocity of 1.7 mm, the bed pressure drop is found to be 710 Pa. The bed pressure drop of the raw sample and the $d_p=1.7$ mm particles is equal from the point of $V_g = 0$ m/s until 0.3 m/s, thus the maximum bed pressure drop for both particles sizes is 630 Pa. The bed

pressure drop stays close to each other throughout. Theoretically, it is expected that the curve of the raw sample will fall between the curves of $d_p=1.7$ mm and $d_p=2.1$ mm particles, but the experiment results prove otherwise.

The particles $d_p=2.1$ mm and $d_p=2.6$ mm reach a maximum value of bed pressure drop of 550 Pa within the range of 0.3-0.6 m/s. The minimum fluidisation velocity range of the $d_p=2.6$ mm particles is 0.3 -0.8 m/s in the bent inlet duct, which is in contrast to the observations made in an ideal plate and straight inlet duct fluidised bed. The bed pressure drop decreases to the lowest point of 420 Pa at 1.1 m/s and increases gradually to a final bed pressure drop of 565 Pa at 1.7 m/s. The behaviour of the $d_p=2.6$ mm particles differs significantly to the other particle sizes investigated. The minimum fluidisation range is longer between 0.2 m/s and 0.8 m/s followed by a steep decline in bed pressure drop between superficial gas velocities 0.8 m/s and 1.2 m/s.

For Plate A in the experiments, particles with $d_p=2.6$ mm require a higher superficial gas velocity of 1.1 m/s to fluidise, followed by $d_p=1.6$ mm and $d_p=1.7$ mm with a superficial gas velocity of 0.8 m/s. The lowest superficial gas velocity of 0.6 m/s is required for $d_p=2.1$ mm and $d_p=1.3$ mm. Smaller particles seem to fluidise at lower fluidisation velocities as is observed by the minimum fluidisation velocity being lower for $d_p=1.3$ mm and highest for $d_p=2.6$ mm. Particles with $d_p=1.6$ mm and $d_p=1.7$ mm have the highest bed pressure drop of 760 Pa at their minimum fluidisation velocity point, followed by 706 Pa for particles with $d_p=2.6$ mm and $d_p=1.3$ mm, while $d_p=2.1$ mm particles exhibited a bed pressure drop of 655 Pa.

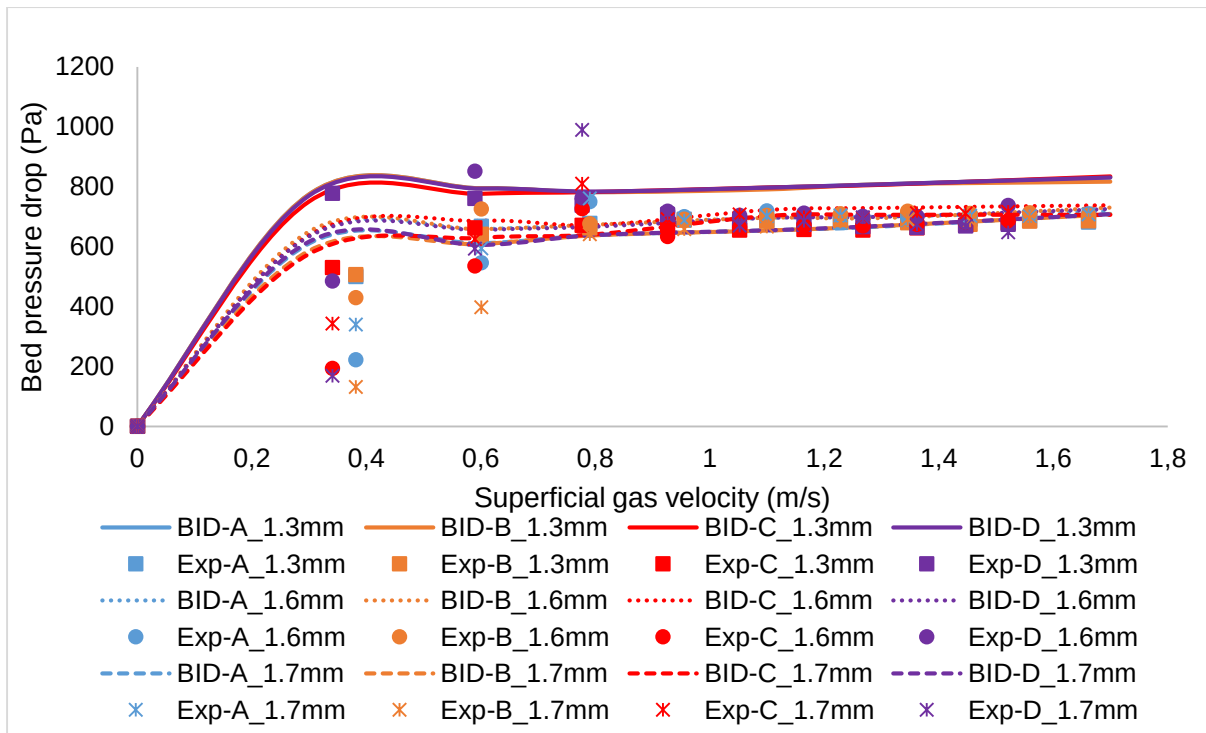


Figure 4-10: Bed pressure drop using plates with velocity profiles associated with a bend inlet duct for particles with smaller average particle diameters

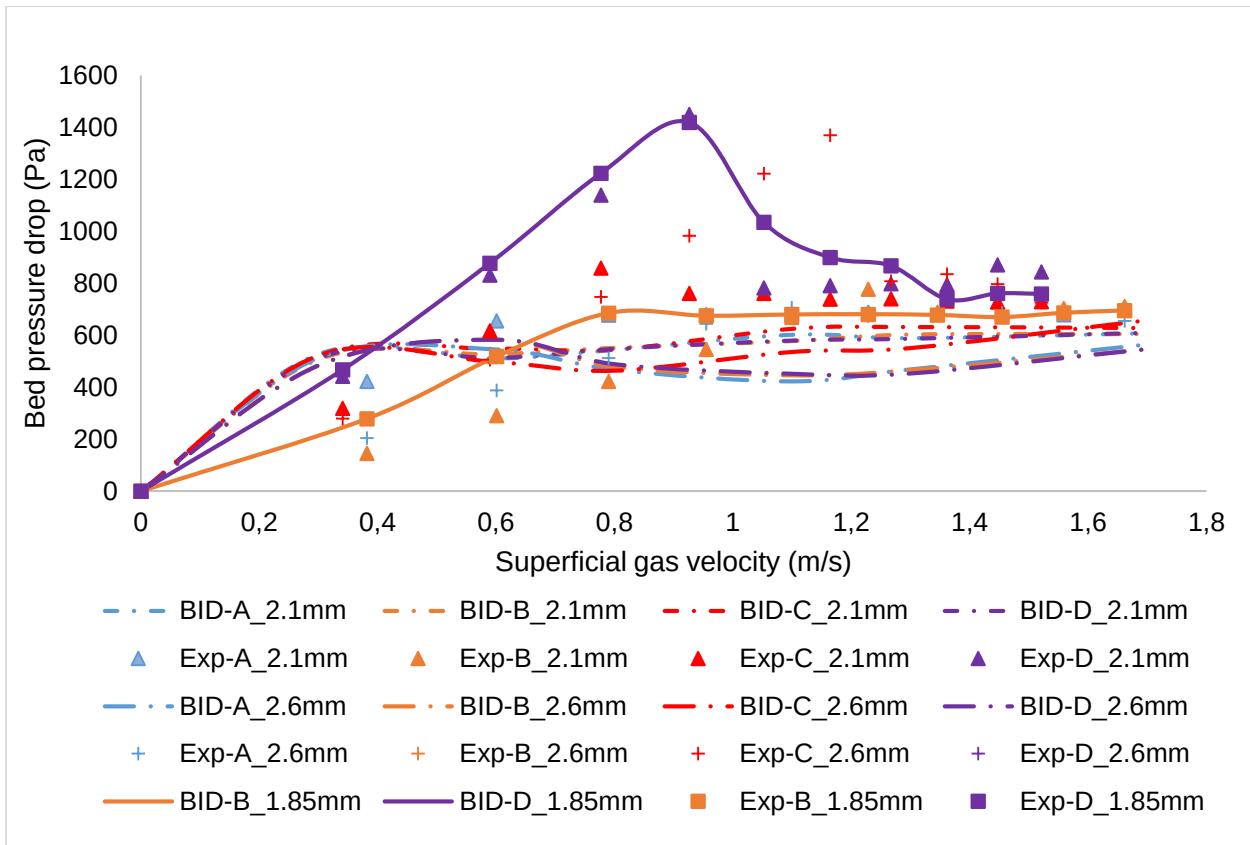


Figure 4-11: Bed pressure drop using plates with velocity profile associated with a bend inlet duct for particles with larger average particle diameters

For the modelled plate B, the behaviour of $d_p=1.3$ mm, $d_p=1.6$ mm and $d_p=2.1$ mm is similar to the behaviour of these particles on plate A. The behaviour of the raw sample meets the theoretical expectations except at the point of 1.1 m/s where the curve is between the $d_p=1.6$ mm and $d_p=1.7$ mm particles. The curve of $d_p=2.6$ mm mimics the trend in plate A except that in plate B, the maximum bed pressure is 30 Pa higher. At the final velocity of 1.7 m/s, it can be observed that, as the particle size increase, the final bed pressure drop decreases. In experiments, the behaviour of bigger particles with $d_p=2.6$ mm on Plate B is similar to the that on plate A, reaching the highest bed pressure drop of 770 Pa with a superficial gas velocity of 1.2 m/s required to initialise fluidisation. The pressure differential stabilises at $V_g=1.34$ m/s and increases steadily to a BPD of 729 Pa at a superficial gas velocity of 2.1 m/s. For plate B, particles $d_p=1.3$ mm and $d_p=1.6$ mm fluidises at the lowest superficial gas velocity of 0.6 m/s with a bed pressure drop of 640 Pa and 725 Pa, respectively. The $d_p=2.1$ mm and $d_p=1.6$ mm particles behaved differently on plate B with a comparison to plate A. The raw sample as well as the $d_p=1.85$ mm and $d_p=1.7$ mm particles fluidises with a superficial gas velocity of 0.79m/s.

For plate C, the minimum fluidisation velocity range of $d_p=2.6$ mm is 0.3-0.6 m/s, which contradicts the observation made for plate A and plate B. The optimum bed pressure drop of 500 Pa is observed within this range and the final bed pressure increases to 660 Pa which is higher than the final bed pressure drop of $d_p=2.1$ mm. It can also be seen that the curve of $d_p=2.6$ mm on plate C does not

mimic the trend of plate A and plate B above. The behaviour of the modelled raw sample meets the theoretical behaviour expectation for the superficial gas velocities.

The behaviour of particles with different particle sizes on plate D produces close similarities to the behaviour of these particles on plate A. Particles $d_p=1.7$ mm reaches the highest bed pressure drop at the superficial gas velocity range of 0.3-0.6 m/s, followed by $d_p=1.6$ mm, the raw sample, the $d_p=1.7$ mm and lastly, $d_p=2.1$ mm and $d_p=2.6$ mm. The minimum fluidisation velocity of $d_p=2.6$ mm granules is 0.3 m/s-0.8 m/s as observed in both plate A and plate B. The bed pressure drop at the final superficial gas velocity, however, replicates the pattern observed on plates A and B, where the bed pressure drop increases with a decrease in the average particles diameter.

The bed pressure drop of particles with $d_p=2.6$ mm increases rapidly until it reaches an maximum value of 1 370 Pa at a V_g of 1.16 m/s for Plate C in experiments. Particles with $d_p=1.6$ mm, $d_p=1.7$ mm and $d_p=2.1$ mm fluidises at the V_g of 0.78 m/s with bed pressure drop reading of 749 Pa and 859 Pa respectively. $d_p=1.3$ mm fluidises with the lowest V_g and stabilises to a bed pressure drop of 725 Pa at a V_g of 1.5m/s. $d_p=2.6$ mm and $d_p=2.1$ mm have an equal final bed pressure drop of 724 Pa at V_g of 1.7m/s. Contrary to the behaviour in plate B, the behaviour of raw samples is similar to that of $d_p=2.1$ mm in plate D. The V_g required to initialise fluidisation for raw samples is 0.9 m/s with a peak bed pressure drop reading of 1 420 Pa. The bed pressure drop stabilises between $d_p=2.1$ mm and $d_p=1.7$ mm. The optimum bed pressure drop of $d_p=1.3$ mm is highest in plate D with bed pressure drop of 760 Pa compared to plate A, B and C, although the U_{mf} is still obtained at the lowest V_g . It can be observed that on plate D the U_{mf} and bed pressure drop increases with the size the of the particles, with $d_p=1.3$ mm at the lowest and $d_p=2.6$ mm at the highest.

It can further be seen that the model, in general, reaches minimum fluidisation velocity at low velocity than the experiments. For $d_p=1.3$ mm particles, the local maximum bed pressure drop reached by the model is greater than that obtained during experimentation. The bed pressure drop curve behaviour in the model differs from the experimentally obtained curves. From the modelled data, bed pressure drop curves of plate A, plate B and plate D reaches the minimum fluidisation point of 794 Pa between 0.3-0.6 m/s, the curves start to stabilise after 0.6m/s. From experiments, plate D has the highest bed pressure drop at the value of 777 Pa at 0.3 m/s and approaches stability at 1.1 m/s. The minimum fluidisation velocity range obtained from the experiments is in agreement with that of the model. The BID model over-estimates fluidisation in this case as the peak and final bed pressure drop are higher than values obtained from experiments.

Experimental results for plate D and plate B for $d_p=1.6$ mm reached the highest bed pressure drop of 850 Pa and 725 Pa respectively. The minimum fluidisation velocity obtained from experiments for plates D and B is within the range of 0.6-0.8 m/s whereas the modelled minimum fluidisation velocity is around 0.3-0.6 m/s. For plate A and plate C, the minimum fluidisation velocity experimentally obtained is within the range 0.8-1.0 m/s, which contradicts the range observed in modelled data.

This is due to the fact that in experiments, detecting the initial full fluidisation tedious. It can be seen that general stability is reached on modelled and experiment data between 1m/s -1.7 m/s. The deviation between the modelled and experimental data is large at superficial gas velocities below 1 m/s and reduces as the velocity increase. The local maximum bed pressure drop for all the modelled plates is reached at approximately 660 Pa at a superficial gas velocity range of 0.3-0.6 m/s. The bed pressure drop curve shows that out of the four plates, the model predicts the behaviour of plate B better.

As the average particle size increases from $d_p=1.6$ mm to $d_p=1.7$ mm, the peak experimental bed pressure drop for plate D increased from 850 Pa to 990 Pa, 700 Pa to 810 Pa for plate B and from 590 to 760 Pa for plate A when U_{mf} ranges between 0.6 -1 m/s. The U_{mf} range for the modelled plates is between 0.3-0.6 m/s. There is an agreement between the model and experimental result at a velocity greater than 1 m/s. From the superficial gas velocity of 0.8 m/s, the model data represent experimental data closely for plate B and plate C.

The model, however, starts to under-estimate the experimental data for $d_p=2.1$ mm and $d_p=2.6$ mm particles. As illustrated, the experimental data lies well above the predicted bed pressure drop obtained from the model. In this case, the deviation is quite significant for all plates with plate D's deviation the highest of all. Similar to $d_p=1.3$ mm particles, the bed pressure drop of the modelled plate A, plate B and plate D shows a negligible deviation for $d_p=2.1$ mm particles. The peak pressure drop for experiment plate B is approximately 3 times higher than the modelled plate. For $d_p=2.1$ mm, the experiments performed on plate A and plate B has equal bed pressure drops at the superficial gas velocity of 1.7 m/s. Although the bed pressure drop is not similar for modelled and experimental runs, the modelled plate A and plate B also have equal bed pressure drops at 1.7 m/s. The model, however, gives a good representation of the experimental observations when fluidising the raw sample particles with $d_p=1.85$ mm.

The minimum fluidisation velocity (U_{mf}) determined from the bed pressure drops against superficial gas velocity curves above for different particle size fraction are given in Table 4-1. The highest U_{mf} range was determined for $d_p=2.6$ mm using plate C. Plate A and plate B maintained a constant range when fluidising $d_p=1.3$ mm and $d_p=1.6$ mm particles. A constant U_{mf} range of 0.78-0.93 m/s was observed when using plate C for $d_p=1.6$ mm, $d_p=1.7$ mm and $d_p=2.1$ mm particles. The U_{mf} range increases with the average particle diameter, as observed in the table below.

Table 4-1: Minimum experimental fluidisation velocity observed for different plates in a laboratory-scale setup

Size fraction (mm)	d_p(mm)	Range of u_{mf} (m/s)			
		Plate A	Plate B	Plate C	Plate D
-1.40+1.00	1.3	0.6-0.79	0.6-0.79	0.34-0.59	0.34-0.59
-1.80+1.40	1.6	0.6-0.79	0.6-0.79	0.78-0.93	0.59-0.78
-2.00+1.80	1.7	0.79-0.95	0.79-0.95	0.78-0.93	0.78-0.93
-2.36+2.00	2.1	0.79-0.95	0.95-1.1	0.78-0.93	0.93-1.05
-2.80+2.36	2.6	1.1-1.2	1.2-1.3	1.16-1.27	1.05-1.27
Raw Samples	1.8		0.79-0.95		0.93-1.05

The range for minimum fluidisation velocity for $d_p=2.1$ mm and $d_p=2.6$ mm particles differs between experimental and modelled plates. The range is lower for the modelled plates. The high deviation values for U_m are primarily due to experimental limitations to distinguish the starting point of full fluidisation in experiment and also the choice of drag model in the predicted output. The behaviour found in both the model and experimental results follows the trend of pressure drop against superficial gas velocity reported in the literature by Kunii & Levenspiel (1991). Jiang *et al.* (2016) investigated the effect of porous media on fluidisation, and the bed pressure drop against superficial gas velocity was found to correspond to the results in this investigation. Another correspondence was found with results obtained by He *et al.* (2015).

4.2.2 Flow patterns inside the fluidised bed

This section discusses what the particle flow patterns look like in the system and how different distributor plates affect these flow patterns. This qualitative discussion gives insight into the behaviour of the bed in terms of bed expansion and bed density which are discussed in subsequent sections. The flow patterns discussed are as observed in the model when using modelled perforated plates associated with PID, SID and BID with an average particle diameter of 1.85 mm, and with 1.1 m/s superficial gas velocity running for 2s.

The solid particles are specified as alpha particles with a void fraction of a bed of particles determined at 55%. The first part of this section describes the development of a pseudo-steady state at the start-up of a fluidised bed (which is useful to understand the particle behaviour discussed for the flow

profiles obtained in chapter 4.1) while the second and third parts of this section attempt to clarify particle behaviour in SID and BID configurations.

(i) Primary ideal duct velocity flow pattern

At the start-up of a fluidised bed setup, a pseudo-steady state in the flow pattern is developed over a short period of time. Figure 4-12 illustrates the flow pattern of superficial gas velocity in a fluidised bed system using plates associated with an ideal velocity profile plate. At the start of the simulation, there is no magnitude of air velocity observed resulting in a fixed bed. The range from the highest to the lowest air velocity magnitude in the fluidised bed is represented by a colour scale in analogy with the light spectrum, ranging from red (highest magnitude velocity) to blue (lowest magnitude velocity).

At 0.01 s, the magnitude of air is approximately 4.7 m/s until 10 cm above the plate which is the static bed height(H_s), where the particles start to rise gradually with superficial gas velocity. At 0.05 sec, the superficial air velocity is rising throughout the fluidised bed, and a localised high-velocity zone is observed rising upwards through the fluidised bed, resulting in a homogeneous steady rise of the particles on fluidisation start-up. As the bed expands, a continuous air stream enters the bottom of the bed to maintain the floating of particles. The area of high superficial gas velocity and magnitude decreases in time. It takes about 0.15 sec for another wave of air to form. At this point, the gravitational force was lower than drag forces resulting in the particles rising up in the fluidised bed duct. As the volume of high magnitude velocity of the second wave increases, the first wave starts to dissipate at 0.20 sec. As the first wave is fully dissipated, the second wave is rising with particles being suspended (0.29 sec).

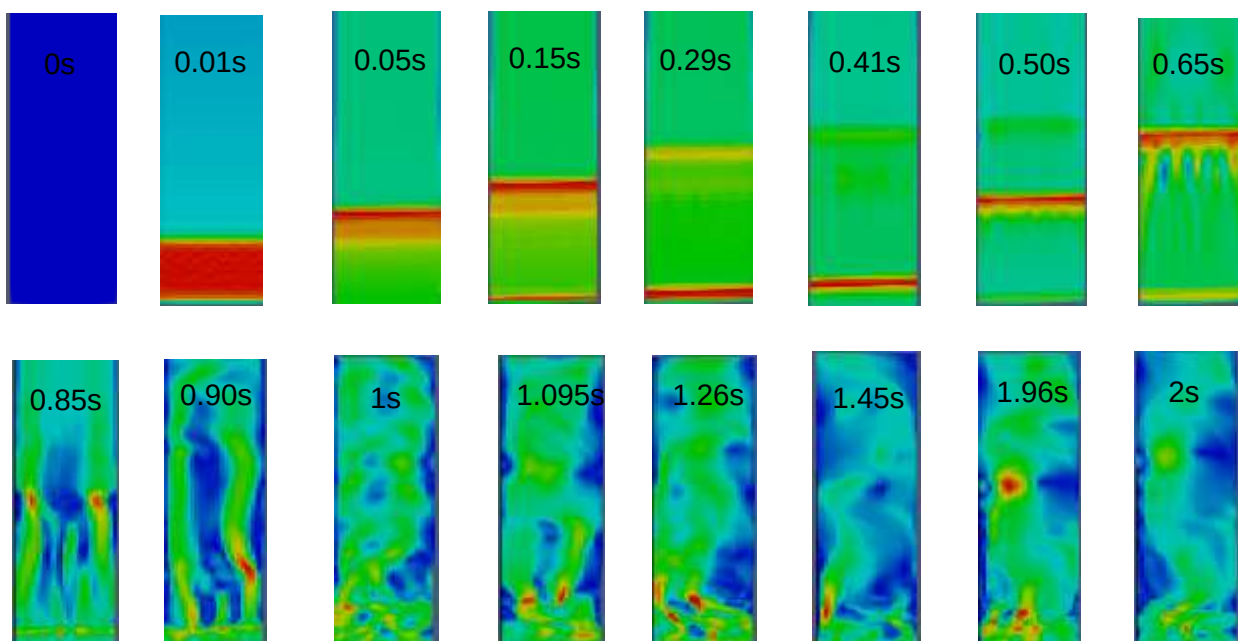


Figure 4-12: Flow patterns in a primary ideal plate fluidised bed system

Non-homogeneity is being observed at 0.41 sec after starting of the fluidisation process on the second high-velocity airwave. At 0.5 sec, localised low-velocity zones start to form in the fluidised bed as the new wave starts to form. The bubbles start to form over time when the third wave starts to rise up the duct with particles being fluidised. The shape of this third wave is not as uniform as in the first two cases, which is related to more areas of localised low-velocity air develop. Only small areas in the duct reach the velocity of 7.2 m/s at 0.6 sec, and areas of low air velocities start to extend downwards towards the distributor plate. As time progresses, the area with small velocity increase at the centre of the duct and the average air velocity in the duct at 0.85 sec is around 4m/s. At this point, bubbling fluidisation is observed with bubbles erupting violently as they reach a maximum height above the plate. Spots of high air velocities and particle concentrations were observed at the bottom of the duct.

At a runtime of 1.26 sec, the patches of high air velocities can be spotted rising with the average air velocity above the static bed height. Fluidisation is gaining its violent momentum as more bubbles form in the bed and erupt faster. The temporary dead zone can be observed below H_s towards the left wall and at the right wall at 1.26 s and 1.45 s, respectively. The initial chaotic behaviour in the system seems to be gaining an order of a steadier state at 1.45 s. As the time gets closer to 2 s, the patches of low velocity and high velocity are very similar to the patches observed at 1.96 s.

(ii) Straight inlet duct particle behaviour

At 0.01 sec, a volume of low particle concentration starts to form at the bottom right corner before any bed expansion starts taking place. This low concentration volume increases with time. The bed begins to expand at 0.07 s. The area of low- to medium particle contents represents a big bubble as a result of jetting of gas. The bed expands more towards the right wall at 0.1 sec as the area increases in size.

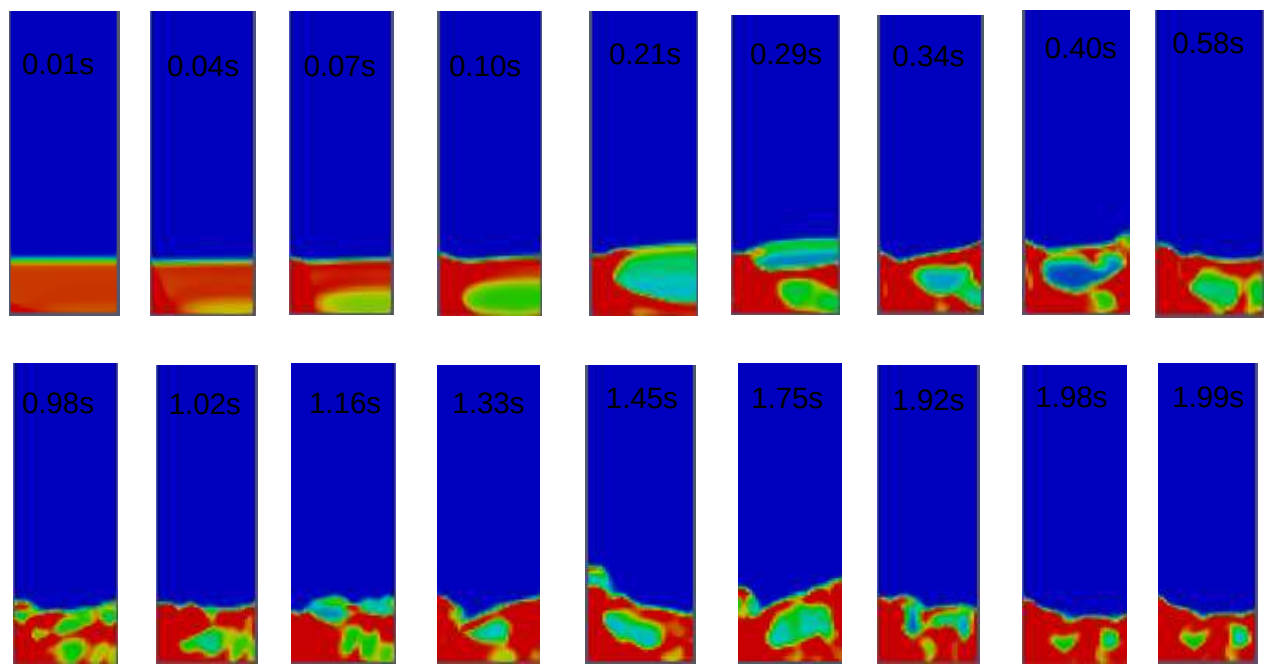


Figure 4-13: Solid particles flow pattern using distributor plates associated with SID

The first bubble eruption is observed at 0.21 sec, and simultaneously a new bubble forms at the plate. At this point, an uneven bed height is observed throughout the duct, which is deemed normal for a fluidised bed system. At 0.29 sec, the bubbles are visibly growing and at 0.34 sec, the initial bubbles have erupted. From here on, the bubbles start to form more to the centre of the duct, which is observed at 0.40 sec as new bubbles begin to form. This results in a change in direction of the flow pattern where particles now fall from left to right as the bubbles collide and coalesce before the eruption. At 0.58 sec, multiple smaller bubbles are observed which grow together until eruption. Bed expansion is observed as a pseudo-steady state is achieved.

(iii) Bend inlet duct particle behaviour

This section focuses on the comparison of the particle flow patterns produced by fluidisation of the raw sample with a superficial gas velocity of 1.1 m/s using different plates in a BID. Figure 4-14 depicts a visual description of the flow pattern employing the same scale to represent solid content in the system as in the previous part of this section. Comparison of the flow patterns using four different perforated plates in a BID configuration is outlined. The flow patterns observed when using plate A, plate B, plate C and Plate D are discussed in detail with a comparison of the observed flow patterns in these plates.

Upon starting of the simulation at 0.01 sec, an area of low particle content is observed above the plates. For all the plates at 0.05 sec, the particle concentration appears to be higher towards the

inner wall, which is the wall closest to the blower. The area of high particle content area covers a great portion of approximately one-third of the bed with a visible higher packing at the outer wall of the duct.

At the initiation of bed expansion, there is an attempt to equalise height in the system, which is evident with a rise of high particle content area towards the outer wall (i.e. the wall of the duct situated further away from the air blower) of the duct. At 0.35s, expansion height is more elevated towards the inner wall of the duct. Higher bed expansion was observed in plates D and C, followed by B & A. At 0.48 s, small bubbles start to form towards the inner wall of the duct as bed expansion continues. This behaviour is identical in all plates with the only difference being the extent of bed height. At this point, the expansion is still high towards the inner wall but levelling to a certain extent across the duct.

The results agree with the conclusion made by Javier (2015) that maldistribution is related to the distributor plate design. The flow pattern in the system varies with the geometry of both the duct and plate. The formation of gas bubbles in the system are consequences of gas velocity exceeding the incipient fluidisation. Zones near the distributor plate where bubbles are inhibited to appear are called dead zones. Sathiyamoorthy and Rao (1981) state that the stable operation of the distributor is possible when all the orifices operate similarly, and as a result, there is no preferential flow anywhere in the system.

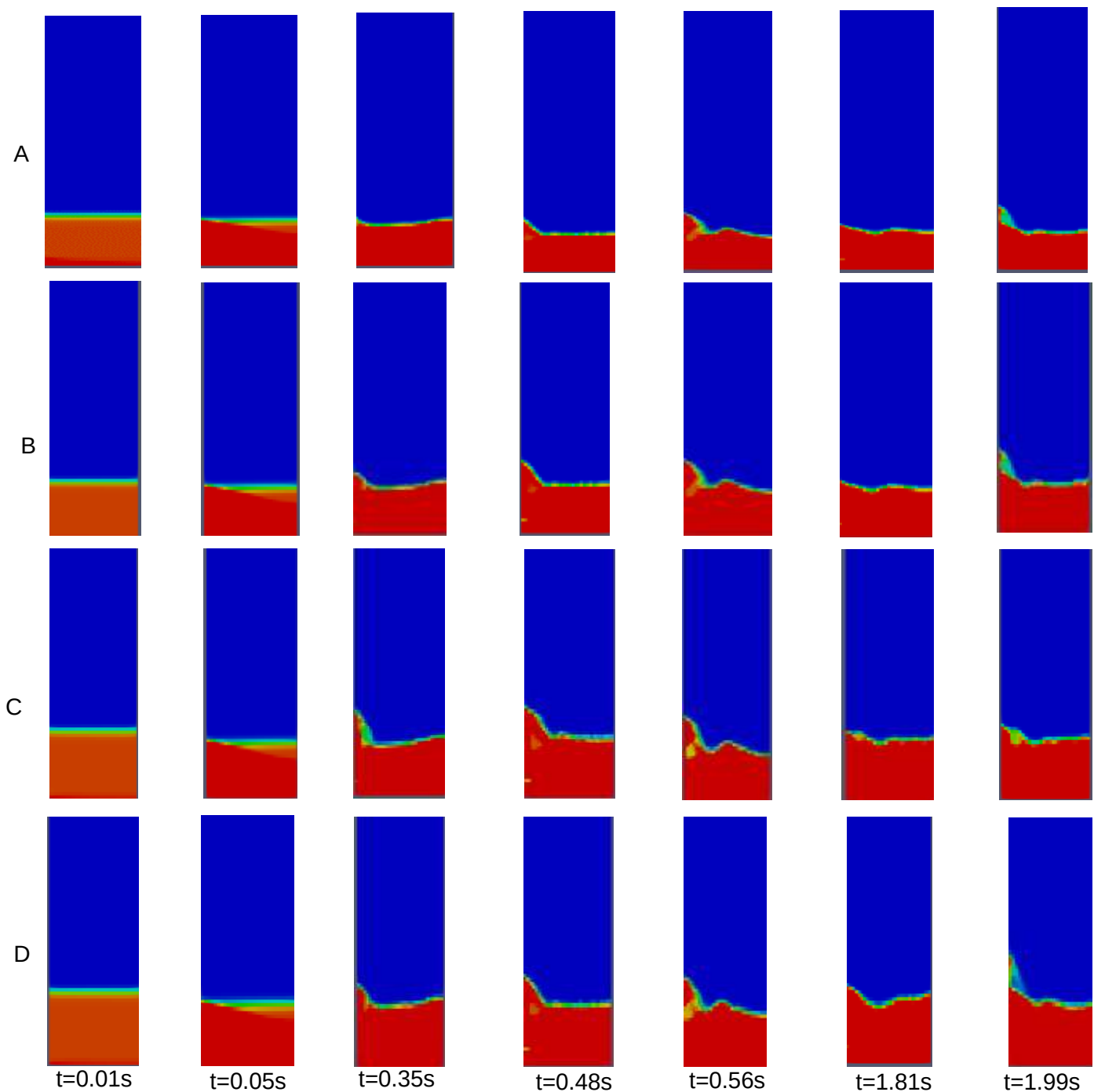


Figure 4-14: Solid particles flow patterns using distributor plates associated with a BID

4.2.3 Bed expansion during fluidisation

The raw sample (i.e. unclassified particles with an average particle diameter of 1.85 mm) was used to investigate the bed expansion in the fluidised bed system. Plates with velocity profiles associated with SID and BID configurations were modelled for this investigation and average measurements taken at 1.99 sec of runtime. The model results were validated with experimental data measured using plate B and plate D.

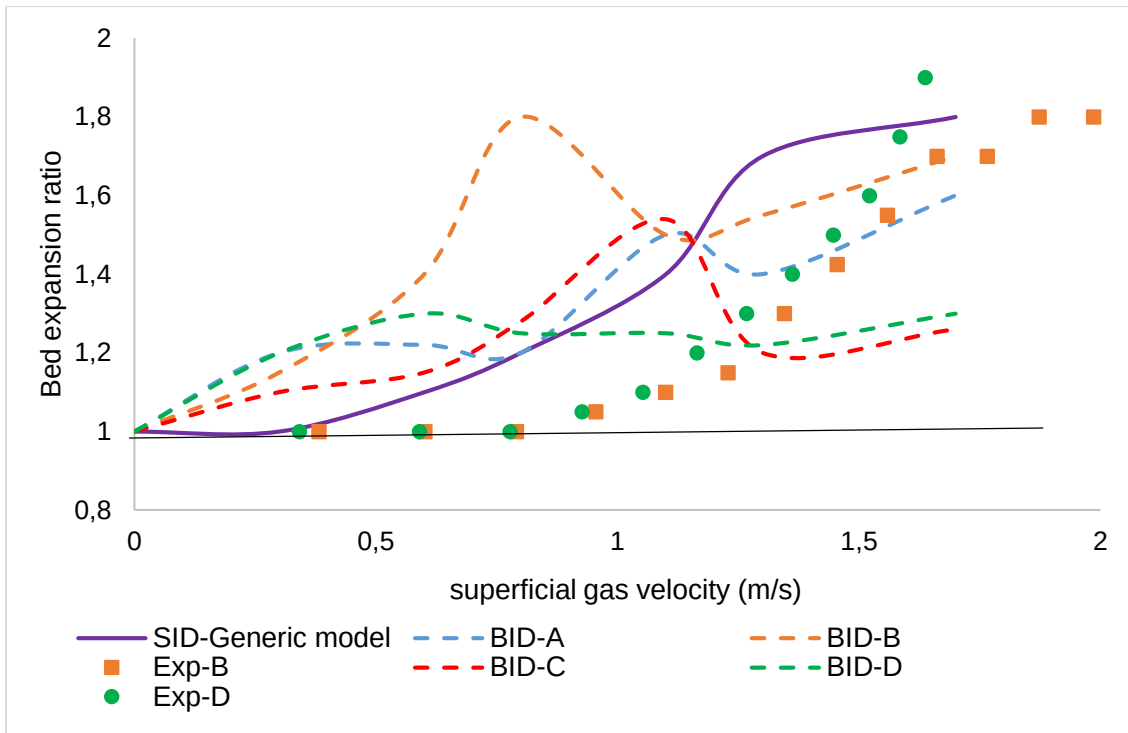


Figure 4-15: Bed expansion ratio of the raw sample particles using perforated plates associated with SID and BID

Figure 4-15 depicts the plot of bed expansion ratio (R) against superficial gas velocity. The solid green line represents the modelled bed expansion ratio curve of all four plates obtained for the SID configuration. The dotted lines represent the modelled bed expansion ratios for the four different plates in a BID configuration, with the blue representing plate A, the orange representing plate B, the red for plate C and the green for plate D. The orange and green markers represent experimental data for plates B & D respectively.

For perforated plates associated with a SID configuration, the bed starts to expand with V_g of 0.3 m/s. The graph shows that for a plate in SID, the bed remained at a static bed height of 10 cm at low superficial gas velocities of 0.3 m/s and 0.6 m/s. Using SID, the bed started to expand to 11 cm at 0.8 m/s. Gradual increase continued as the superficial gas velocity increased. The maximum bed height reached with the $d_p=1.85$ mm particles was 17 cm during fluidisation.

For experiments, the bed starts to expand at 1.0 m/s. The results show that the model gives a close representation of the experimental results at lower superficial gas velocities before expansion begins. For a BID configuration, the results indicate that the bed begins to expand at a low superficial gas velocity of 0.3 m/s with the highest expansion observed for plate B. The bed expansion reaches its highest value at 0.8 m/s and decreases to 1.1 m/s. The maximum R for plate A and plate C was reached at 1.3 m/s followed by a sudden drop.

The bed expansion observed in the model of a SID differs significantly with that of a BID. A clear contradiction between the model is observed between 0.8 m/s and 1.3 m/s where the R for a plate

with velocity profiles associated with BID decreases while for a SID, the value of R increases. The behaviour of R in both cases is due to preferential flow pattern discussed in section 4.2.3. Between 0 m/s and 0.8 m/s, there is an agreement in behaviour between modelled BID and SID expansion which in turn agrees with the behaviour observed in experiments to some extent. The R increases in the model while a negligible increase was observed in the experiments. The ability of the model to predict bed expansion is considered rather poorly.

4.2.4 Bed density during fluidisation

Bed density in a laboratory-scale setup

Bed density is one of the bed hydrodynamics that can be used to describe the quality of fluidisation although it is not well documented in the literature. Bed density in the model was measured at three different heights from the plate, i.e. at 2cm, 4cm and 6cm at both the centre of the duct and towards the left wall at the end of runtime (i.e. at 1.99 s). Distributor plates with velocity profiles associated with a SID and a BID were investigated with fluidisation of the raw (unclassified) sample with $d_p=1.85$ mm.

(i) SID

The bed density in a SID system is represented by Figure 4-16 and 4-17 where the blue, orange, red and green curves represent the bed density curves of the raw sample for both plates when fluidising with V_g of 0.8m/s, 1.1m/s, 1.3m/s and 1.7m/s, respectively. Figure 4-16 depicts the bed density measured at the centre of the duct while Figure 4-17 is the representation of bed density measured towards the wall of the duct. Since an average plate was assumed for fluidisation in a SID setup, the equivalent bed density of plate A, plate B, plate C and plate D is accepted in the system. The theoretically obtained SID modelled results were compared to the experimental results obtained for plate D on a BID (since a true SID experimental setup is not available) in order to assess the validity of the model results and bed density measurements were used to validate the model. In experiments, bed density measurements were recorded at V_g of 0.93 m/s represented by a filled black round marker, 1.16 m/s represented by filled navy blue triangular markers, 1.36 m/s represented by a yellow star marker, 1.45 m/s represented by a purple cross marker and 1.52 m/s represented by a filled purple squares.

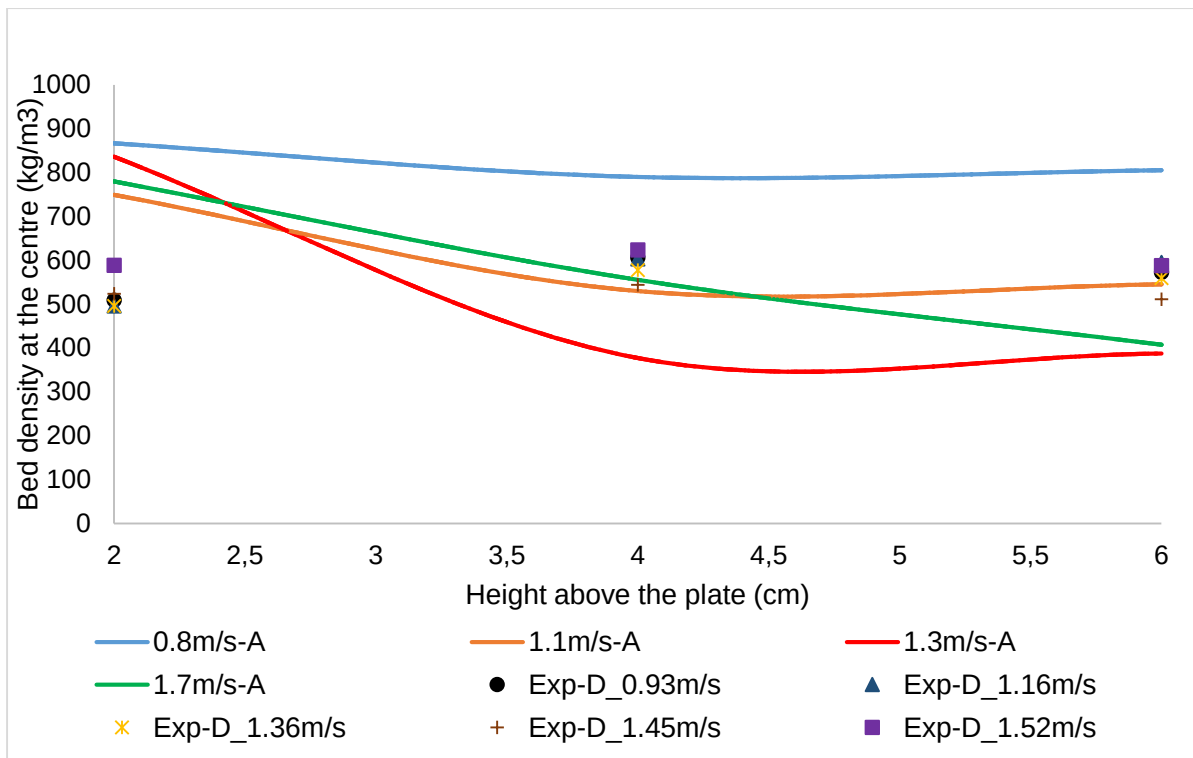


Figure 4-16: Bed density measured at the centre of the duct fluidising $d_p=1.85$ mm particles with a laboratory-scale SID

In Figure 4-16, it can be seen that the highest bed density of approximately 860 kg/m^3 was observed when fluidising with a superficial gas velocity of 0.8 m/s and measuring at 2cm above the perforated plate. The bed density decreases when the probe position for density measurement is lifted to 4cm above the plate and gradually increases as the probe position changes to 6cm above the plate. This trend is visible with velocities 1.1m/s and 1.3m/s , however, the density decreased rapidly from 2cm to 4cm above the plate with 1.3 m/s reaching the lowest bed density of approximately 380 kg/m^3 measured at the centre of the duct. With 1.7 m/s , the bed density decreases with an increase in probe position following an almost linear response.

Bed density curves constructed from experimental data shows no agreement to the bed density predicted by the model. In experiments, bed density measured at the centre of the duct reached its highest point at 4cm above the plate while the model predicts the highest density at 2cm above the plate. The trend followed begins with low bed density at 2cm above the plate followed by a gradual increase in bed density magnitude then lastly a decrease at 6cm above the plate. This behaviour at the centre of the duct varies significantly with that observed towards the wall. At the wall, the highest bed density was reached at 2cm above the plate and as the height above the plate increases, the bed density decreases. This behaviour is due to the fact that the particle concentration decreases higher up in the duct.

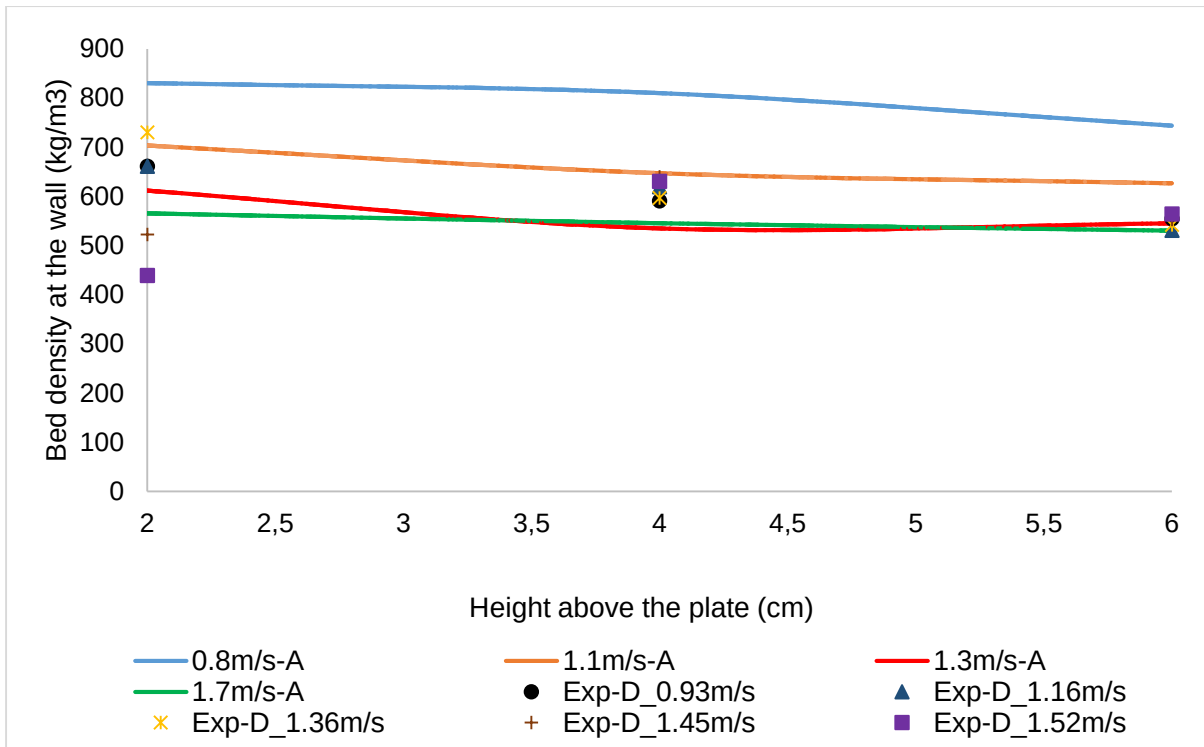


Figure 4-17: Bed density measured towards the wall of the duct fluidising $d_p=1.85$ mm particles with a laboratory-scale SID

Towards the wall of the duct (Figure 4-17), the highest bed density was again observed when fluidising with a superficial gas velocity of 0.8 m/s, followed by 1.1 m/s, 1.3 m/s and lastly 1.7 m/s. The lowest bed density is reached at a fluidisation velocity of 1.3 m/s. Contrary to the observations at the centre of the duct, the lowest bed density was measured at 6cm above the plate with 0.8m/s. This is because of the particle content at that height. Fluidisation with 1.1 m/s and 1.3m/s follows the trend observed at the centre of the duct. The response obtained with 1.7 m/s is nearly constant at all heights above the plate. An agreement between the model and experiment is obtained at 2cm above the plate where both setups show maximum bed density measurements.

(ii) BID

Figure 4-18 and Figure 4-19 show the modelled bed density of the raw unclassified sample measured at the centre and towards the wall of the fluidised bed duct respectively. The lines represent the modelled bed density curves while the markers represent the experimentally obtained results. For both these figures, the red, blue, yellow and green colours represent the bed density curves at superficial gas velocities of 0.8 m/s, 1.1 m/s, 1.3 m/s and 1.7 m/s respectively. While performing the experiments, the bed density was measured at 0.9 m/s, 1.16 m/s, 1.36 m/s, 1.45 m/s and 1.52 m/s with the data being represented by the maroon, blue, orange, green and purple markers.

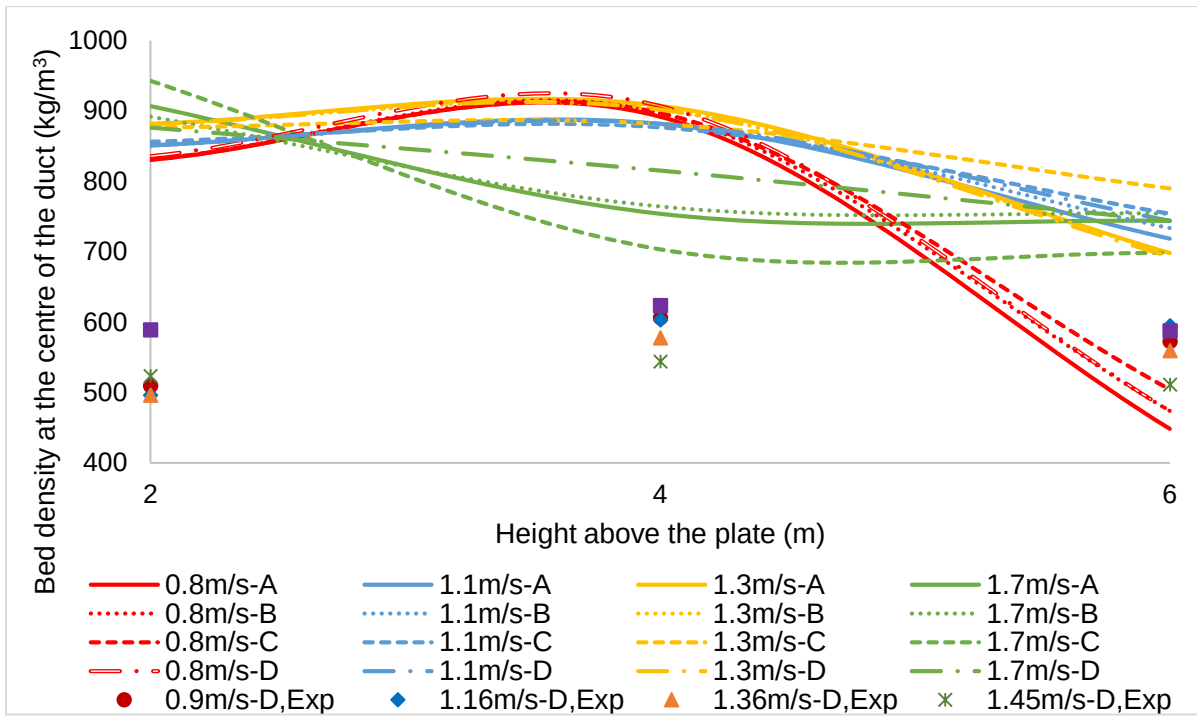


Figure 4-18: Bed density measured at the centre of the duct fluidising $d_p=1.85$ mm particles with a laboratory-scale BID

When measuring at 2cm above the plate at the centre of the duct, the highest bed density was observed when fluidising with 1.7 m/s, followed by 1.3 m/s. $V_g=0.3$ m/s produced the lowest bed density of 505 kg/m³. When increasing the height above the plate, the highest and lowest bed densities were obtained with $V_g=1.3$ m/s and 0.3 m/s, respectively. With the 6cm height above the plate, the lowest and highest bed densities were recorded at 0.8 m/s and 1.7 m/s respectively.

The general behaviour depicted by the bed density curves for plate A is that the bed density is higher at 2cm above the plate at the centre of the duct, reaches the peak at 4cm above the plate and decrease towards 6cm above the plate. At superficial gas velocities between and including 0.3 m/s and 1.3 m/s, a similar trend followed by bed density was observed. The $V_g=1.7$ m/s curve is a complete invert of the behaviour observed at lower gas velocities.

The experimental data is flat against the significant variations predicted by the model. The discrepancies between the results may be due to the fact that in experiments, the pressure differentials measurements used to calculate bed densities were measured using a straight column whereas in the model, a bend inlet setup was adopted to investigate if there will be any correspondence. Secondly, the initial superficial gas velocity in the model was considered as a polynomial determined from velocity magnitude graph rather than actual velocity magnitude in experiment. It is also suspected that overheating of the air blower during fluidisation might have impacted the experimental results, this effect will still have to be investigated.

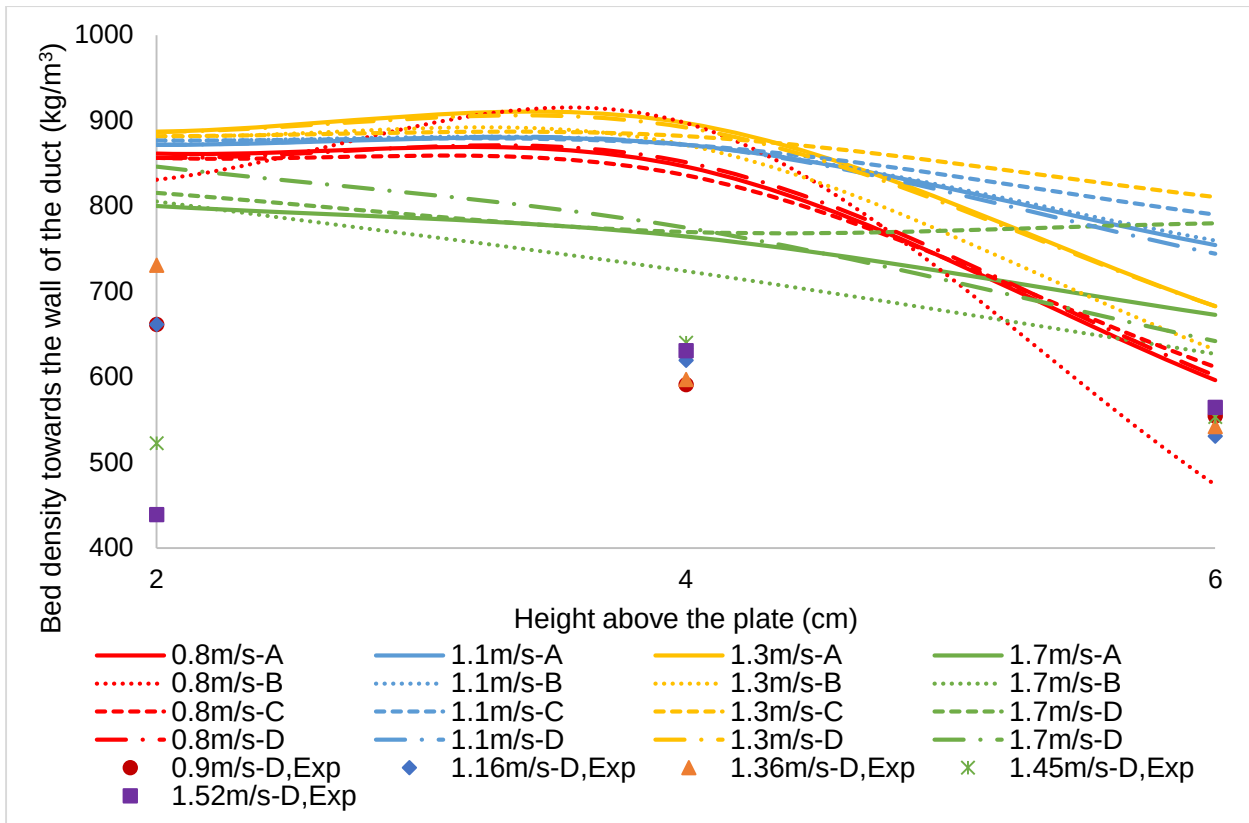


Figure 4-19: Bed density measured towards the wall of the duct fluidising $d_p=1.85$ mm particles associated with a laboratory-scale BID

At the wall of the duct, the highest bed density was reached with $V_g=1.3$ m/s at 2cm and 4cm above the plate with the lowest bed density reached at the lowest superficial gas velocity of 0.3 m/s at both measurement points. When the measurement point is further away from the plate, the highest and lowest bed density was reached with 1.1 m/s and 0.8 m/s, respectively. The trend observed at the wall for all the superficial gas velocities is similar to the trend observed for most superficial gas velocities at the fluidised bed duct centre. At $V_g=0.3$ m/s, the deviation between bed density measured at the centre and towards the wall is smaller at 2cm and 4cm above the plate and increases towards a height of 6 cm above the plate. The difference is significantly small at 2cm, 4cm and 6cm when comparing densities measured at the centre and the wall when fluidising with 1.3 m/s. Larger variances occur at $V_g=1.7$ m/s, 0.8 m/s and 0.6 m/s.

Contrary to plate A, the deviation of bed density curves between the measurements at the centre and towards the wall is high at 2cm and reaches a state of equilibrium at 4cm and 6cm for $V_g=0.3$ m/s, which is the superficial gas velocity that reached the lowest bed density. The deviation of wall bed density to centre bed density is observed to be high 1.7 m/s, 0.6 m/s and 0.8 m/s. At the centre of the bed with $V_g=1.3$ m/s and 0.8 m/s, an equal bed density of approximately 900 kg/m^3 is measured at 4cm above the plate. The centre bed density drops drastically between 4cm and 6cm when fluidising with a superficial gas velocity of 6 m/s. At the wall, the lowest bed density of 490 kg/m^3 is reached with a velocity of 0.3 m/s and 0.6 m/s at 6cm above the plate.

The unique behaviour of the bed density curve at 1.7 m/s is also visible when using plate C. The lowest bed density is reached when fluidising with $V_g=0.3$ m/s, although the difference between the wall bed density and centre bed density is high at the 2cm and 6cm measurement points above the plate, and equal at 4cm above the plate. Similar to plate A, the results show that at 2cm above the plate measuring at the centre, the bed density increases with superficial gas velocity. At 4cm above the plate at the centre, an equal bed density of 880 kg/m³ is reached with $V_g=0.6$ m/s, 1.1 m/s and 1.3 m/s.

Contrary to the behaviour of the centre and wall bed density curves with $V_g=1.7$ m/s in plate A, plate B and plate C, it can be seen that the bed density using plate D is perceived to be a straight line. The centre bed density with $V_g=0.8$ m/s decreases rapidly to the lowest density of 470 kg/m³ at 6cm. The deviation between the centre and wall bed density obtained with 0.3 m/s follows a similar pattern observed for plate C. The highest deviations are found with $V_g=0.6$ m/s and 0.8 m/s. There is a close correspondence between the centre and wall densities observed when employing superficial gas velocities $V_g=1.1$ m/s and 1.3 m/s.

To determine the ability of the model to predict the behaviour of bed density in a fluidised bed, experiments were conducted. At 2cm above the plate, the highest centre bed density is reached at $V_g=1.52$ m/s followed by 1.45 m/s then 0.9 m/s, with joined lowest centre bed densities between 1.16 m/s and 1.36 m/s. When the height above the plate was increased to 4cm, 1.52 m/s remained as the superficial gas velocity to reach the highest centre bed density. The second highest bed density was reached simultaneously with $V_g=0.9$ m/s and 1.16 m/s followed by 1.36m/s and 1.45 m/s with the lowest centre bed density. The order changed with an increase in height above the plate where the highest centre density was observed using 1.16 m/s and 1.52 m/s while 1.45 m/s produced the lowest bed density still.

The modelled bed density results show, in general, an increase in bed density from 2cm above the plate towards 4cm above the plate, and a decline in bed density towards 6cm above the plate at lower gas velocities. This behaviour is, however, not similar for higher gas velocities (around 1.7 m/s) as a steady decline in bed density is observed at this velocity.

The experimental values show a similar behaviour than the low velocity modelled behaviour for all velocities assessed, which is indicative that the model can predict the general trends to a certain extent. Experimental values were found to be consistently lower than modelled values, which can be ascribed to the approximation of a distributor plate with a porous medium in the model (and not a large number of individual orifices). The assumption of the fluidised bed system to be 2D by introducing empty patches in the model and mesh resolutions also contribute to the discrepancies between the results. The larger variation in experimental bed densities observed just above the distributor plates (at 2cm above the plates) can be ascribed to jet streams occurring through the

plate orifices, which culminates into larger bubbles at heights of 4cm – 6cm above the plates leading to less longer-term variations in measured data points.

With the current laboratory-scale bed densities being very similar at different heights above the plates and at different velocities, it is not possible to predict which distributor plate would yield the most even distribution of air into the fluidised bed. In order to have a better understanding of this behaviour, the pilot-scale study of air distribution is subsequently discussed

Bed density in a pilot-scale setup

With the plates being characterised and the behaviour of each plate being simulated in the model environment, a *twoPhaseEulerFoam* solver was used to model a fluidisation setup for the unclassified sample to determine the bed density distribution during fluidisation in a pilot scale setup. The behaviour of three of the perforated plates (i.e. plates A, B & C) was specified as the initial boundary condition following the same procedure as discussed for the laboratory-scale setup in section 3.4.3. Pressure drop was measured at five points across the duct by employing the grid and method outlined in section 3.3.1. and Figure 3-7 respectively. Measurements were also taken at four different heights (5cm, 15cm, 25cm and 35cm) above the plates during the simulation of the fluidisation of the unclassified particles with an average particle diameter of 1.85 mm.

Figure 4-20 to Figure 4-22 depict the bed density against height above plate A, plate B and plate C sequentially with superficial gas velocities of 5.7 m/s, 7.8 m/s and 11.3 m/s. In the above-mentioned figures, the lines represent modelled data whereas markers represent the experimental data. The blue, orange, red, green and purple colours represents the modelled and experimental density values along the horizontal axis of the fluidised bed at 17cm, 33cm, 49cm, 68cm and 83cm in the direction away from the location of the blower feeding the fluidised bed with air (i.e. towards the outer wall shown in Figure 3-12).

(1) Plate A

For $V_g = 5.7$ m/s and 7.8 m/s, the model results demonstrate that at 5cm above the plate, the bed density measured horizontally across the fluidisation duct remained constant although the highest density was reached with a velocity of 11.3 m/s. Bed density decreased when the probe position was moved up further from the plate hence the lowest bed density was observed at 35cm above the plate. A constant behaviour was observed at 25cm above the plate for both 7.3 m/s and 11.3m/s. The response observed at 17cm across the duct was found to be identical to the response at 83cm across the duct for $V_g = 5.7$ m/s.

In the experiments, the bed density measured at 17cm and 33cm across the duct was observed to oscillate as the measurement height above the plate increases, with the highest bed density of 1002 kg/m³ reached when measuring at 5cm above the plate and 17cm across the duct. Bed density tends to decrease when height above the plate increased to 15cm, increased as height increased to 25cm

then decreased again at a height above the plate of 35cm. At 33cm across the duct and 25cm height above the plate, an increase that was higher than that obtained at 5cm above the plate was reached. At 49cm, 66cm and 83cm across the duct, the bed density increased as the height above the plate increased with the lowest bed density reached at 66cm across the duct and 5cm above the plate. The experimental results showed that the bed density is generally high at 17cm across the duct which contradicts the observations made with model data. The difference between the model and experimental data is high at 66cm across the duct and lowest at 17cm across the duct.

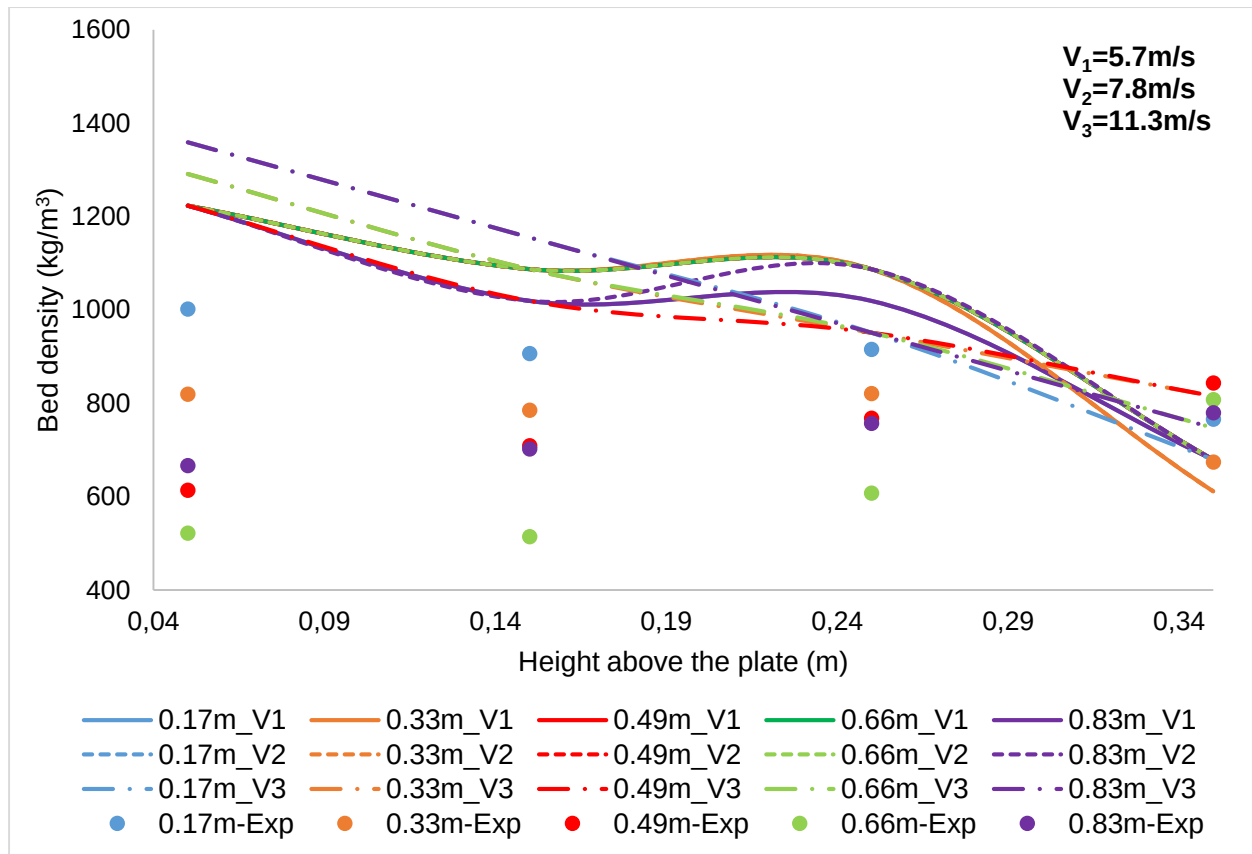


Figure 4-20: Bed density against height above plate A for various superficial gas velocities in a pilot-scale facility

It is clear that, for plate A, bed densities measured towards the inner wall (i.e. the wall closest to the position of the blower feeding air to the system) as denoted by the blue and orange markers, appear to be higher than bed densities measured towards the outer wall (green and purple markers), irrespective of the height it is measured above the plate. This behaviour is not reflected by the model, which leads to the suspicion that the porous medium approximation of a plate in the CFD environment is not a true reflection of the behaviour of this plate. The disagreement between the data may due to different probe locations for pressure differential method. In experiments, pressure differentials were measured while fluidisation is in progress whereas from the model, the pressure differentials were recorded at the end of fluidisation runtime. Measurement technique during

experiments may also have also influenced the accuracy of the results as multiple pressure differentials were measured and only average values used for the determination of bed density. These higher densities observed at the inner wall is also indicative that more air is distributed towards the outer wall (i.e. the wall further away from the location of the blower) than towards the inner wall.

(2) Plate B

From the model outputs, with $V_g = 5.7$ m/s, towards the side walls of the duct at 17cm and 83cm across the duct, a maximum bed density of 1291 kg/m^3 was reached at 5cm above the plate measuring at both 17cm and 83cm across the duct while 1223 kg/m^3 was determined at 33cm, 49cm and 66cm. At 15cm above the plate, a constant bed density of 1019 kg/m^3 was found across the duct. A constant response was also observed at 5cm and 15cm above the plate with $V_g = 7.8$ m/s and 15cm and 25cm above the plate with $V_g = 11.3$ m/s. Although the increase in height above the plate from 15cm to 25cm had no impact on bed densities at 17cm and 66cm, bed densities at 33cm, 49cm and 83cm increased to 1087 kg/m^3 followed by a drastic decrease to 679 kg/m^3 observed constantly across the duct.

Identical responses were observed when measuring at 33cm & 49cm across the duct and also at 33cm, 49cm & 83cm across the duct with $V_g = 5.7$ m/s. With $V_g = 7.7$ m/s two identical responses were observed for 17cm & 83cm across the duct as well as at 33cm, 49cm & 66cm across the duct. When fluidising at 11.3 m/s, an identical response materialised at 17cm, 33cm and 66cm across the duct. The highest and lowest bed densities of 1223 kg/m^3 and 1155 kg/m^3 for plate B were observed when fluidising with 11.3 m/s. The lowest density was observed at the centre of the duct, i.e. 49cm across the horizontal axis of the duct. A similar behaviour was observed with 5.7 m/s where the high bed density at 5cm above the plate was measured towards the inner and outer walls of the duct while lower density was measured at the centre, at 33cm, 49cm and 66cm across the duct.

The experimental results showed that a maximum bed density of 696 kg/m^3 was reached at 83cm across the duct at 5cm above the plate. Bed density at 5cm above the plate decreased to 647 kg/m^3 , 566 kg/m^3 , 551 kg/m^3 and 523 kg/m^3 at the points 66cm, 17cm, 33cm and 49cm across the duct respectively. As the height above the plate increased, the bed density also increased. With a further increase in height above the plate to 25cm, only the bed density measured at 17cm and 66cm across the duct decreased while an increase is evident at other sections. At 35cm above the plate, all bed densities measured across the duct decrease with the lowest bed density reached at 49cm across the duct.

Comparing the modelled data to experimental data, a significant difference is observed at 5cm above the plate and 17cm, 33cm and 49cm across the duct however, there is an overall significant difference in model and experiment results at all heights and distances. An error percentage of 4% was calculated for bed density at 35cm above the plate and 83cm across the duct where the final

density of the model and experiment at 33cm, 66cm and 83cm across the duct falls within the same range.

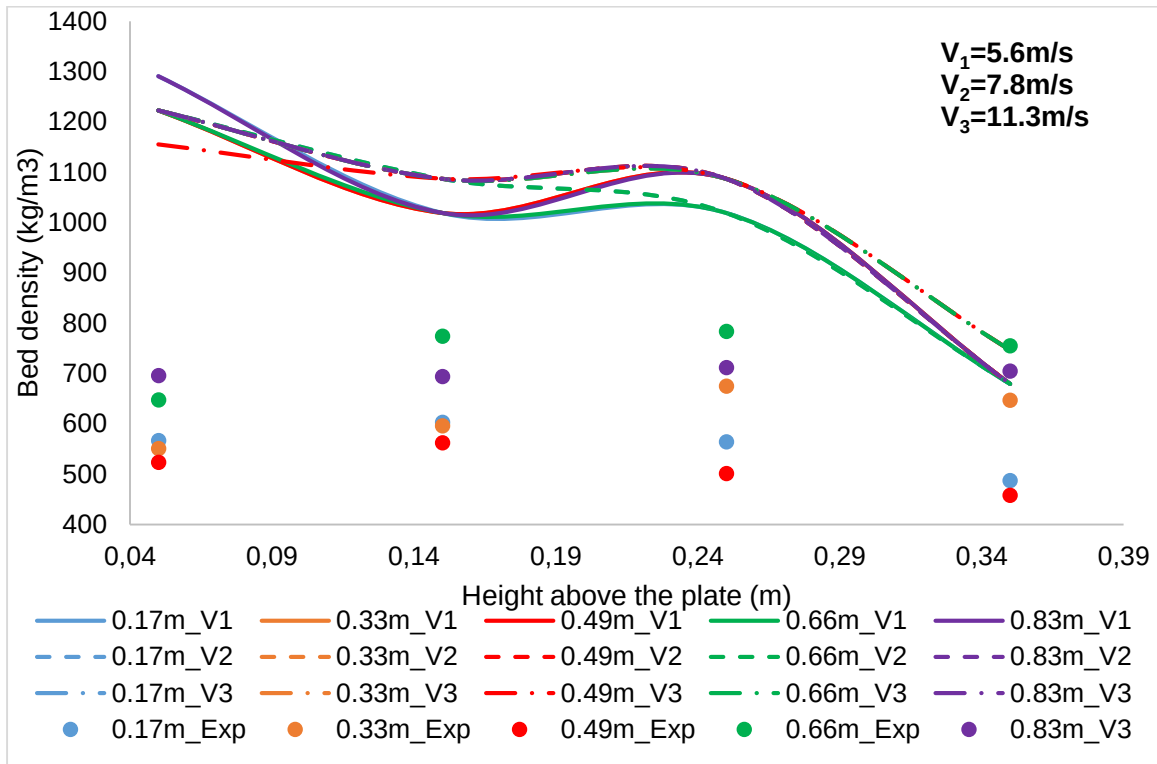


Figure 4-21: Bed density against height above plate B for various superficial gas velocities in a pilot-scale facility

For plate B, higher bed densities are observed in general towards the outer wall (denoted by the purple and green markers) than at the inner wall (denoted by the blue and orange markers). This observation is an exact opposite of what happened with plate A. The average bed density observed for plate B is slightly less than for plate A, and a narrower distribution of data horizontally across the bed is observed. This leads to the conclusion that plate B, with a higher pressure differential than plate A, exhibits a more uniform distribution of air into the fluidised bed, although more air is, in this case, channelled to the inner wall than towards the outer wall. The model, again, did not reflect the experimentally observed behaviour for reasons elaborated above.

(3) Plate C

The high bed density was measured at 5cm above the plate and 33cm, 49cm and 66cm across the duct which is essentially at the centre of the duct with $V_g=5.67$ m/s with an identical response observed at 49cm and 66cm across the duct. Shifting the probe height to 15cm resulted in a decrease in bed density to a low and constant measurement across the duct at 35cm above the plate. For both 7.8 m/s and 11.3 m/s, identical responses were observed at 17cm, 49cm and 66cm across the duct. Shifting the probe position from 15cm to 25cm above the plate had no impact on

the bed density measurement as it remained constant. However, for the three superficial gas velocities, the lowest bed density was measured at 35cm above the plate.

From experimental data, the highest bed density was reached at 5 cm above the plate and 17cm across the duct. The bed density decreased with an increase in probe position to 15cm and 25cm above the plate at 17cm, 49cm and 66cm while the density increased at the other sections at the height of 15cm above the plate and decreased at 25cm. As the height increased further, the bed density decreased at 33cm, 49cm, 66cm and 83cm across the duct and increased at 17cm. The highest bed density of 884 kg/m^3 and lowest bed density of 471 kg/m^3 was reached at 5cm above the plate and 17cm across the duct and 35cm above the plate and 49cm across the duct respectively. Comparing the predicted data to experiment data, it can be seen that the model predicts the bed density at 35cm above the plate and 33 cm and 83cm across the duct accurately with an error of 2% calculated. The most significant misrepresentation is found at 49cm and 83cm across the duct. The error decreases as the height above the plate decreases.

With visibly the narrowest distribution of experimentally obtained bed densities at different points along the horizontal axis, this plate shows the most even distribution of air into the fluidised bed system, irrespective of the position of the blower relative to the fluidised bed. Bed densities were observed to be very similar at both the inner wall and outer wall, with lower bed densities towards the centre of the bed, as would be expected. The values predicted by the model tend to be very similar regardless of the axial position in the bed, but based on the conclusions with regards to the model for the two previous plates, this prediction cannot be termed valid.

With an even higher pressure differential observed experimentally for plate D, the observation made for the three plates above can be extrapolated to predict that plate D will exhibit an even more uniform distribution of air into the system.

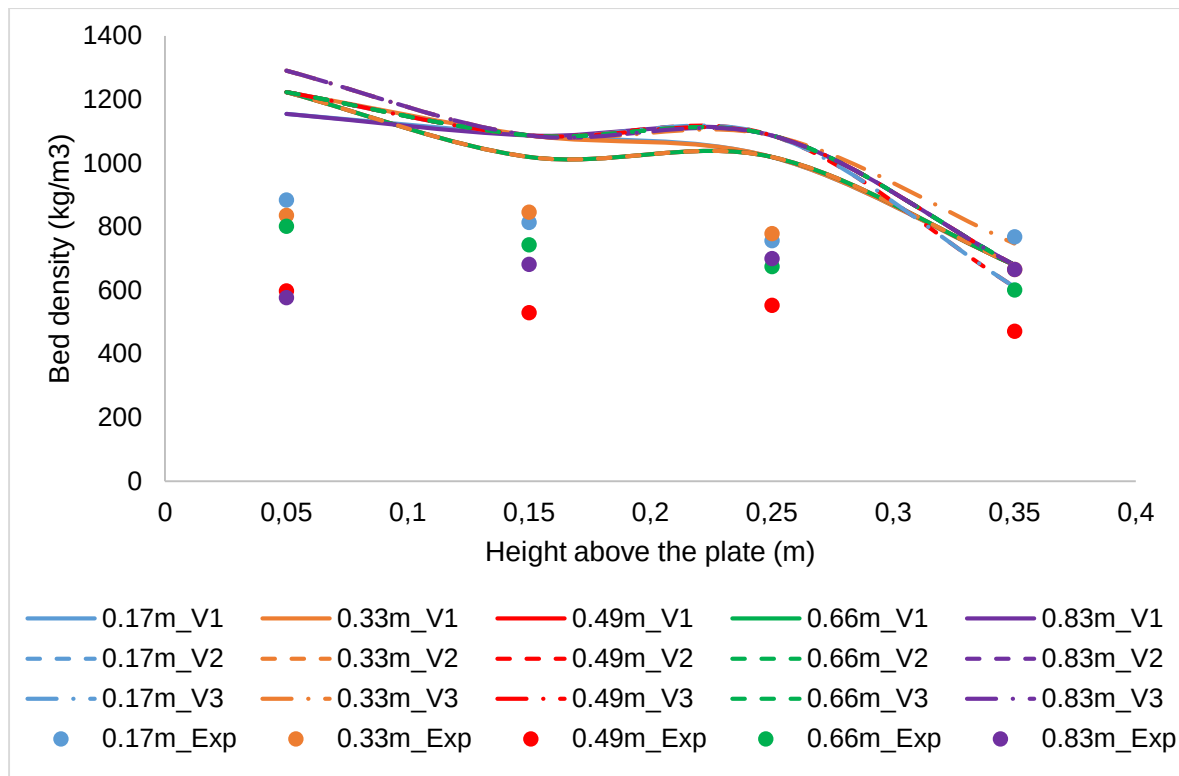


Figure 4-22: Bed density against height above plate C for various superficial gas velocities in a pilot-scale facility

4.3 Chapter summary

This chapter covered the results obtained during this investigation. At first, different distribution plates employed in fluidising ammonium nitrate were characterised in order to approximate flow profiles through these media that could be used as boundary conditions in the CFD modelling environment.

This section was followed by a description of the hydrodynamic behaviour of solid ammonium nitrate particles in a fluidised bed. At first, the total bed pressure differential and minimum fluidisation velocities were investigated using both a modelling and experimental approach. Some flow patterns of air and particles were subsequently qualitatively described for fluidisation start-up scenarios, bed expansion during fluidisation subsequently received attention and the chapter concluded with a discussion on bed density and flow distribution into the fluidised bed.

The following chapter covers the conclusions drawn from this investigation as well as some suggestions for future work related to fluidisation studies employing CFD modelling.

5. CONCLUSION AND RECOMMENDATIONS

This chapter summarises the main conclusions of the study on the effect of distributor plate on bed hydrodynamics, with emphasis on the impact of different particle sizes and superficial gas velocities. Recommendations to improve the results and issues to be addressed in future work are also discussed.

5.1 Conclusions

This study systematically evaluated the effect of the distributor plate geometry on fluidisation of ammonium nitrate particles in light of a number of objectives outlined in chapter 1.3. These objectives are subsequently listed with the conclusions drawn for each.

- (i) *Obtain the characteristics of different suggested distributor plates for use in the fluidisation of ammonium nitrate (AN) granules for modelling purposes in a CFD environment.*
 - A porous medium approach in OpenFOAM™ is capable of approximating the flow versus pressure differential behavioural characteristics of all the distributor plates with the exception of plate D, wherein some rather minor deviations are observed, in the laboratory-scale setup. This leads to the conclusion that this approach can efficiently be used to generate pressure/flow boundary conditions for subsequent modelling of fluidisation behaviour.
 - The impact of the pitch of the distributor plate holes was evident with observations of high plate pressure drop values obtained for plate C and D and lower values for plate A and B in the system. Therefore, a larger pitch resulted in higher plate pressure drops observed.
- (ii) *Generate a CFD model for AN granules fluidisation that can effectively produce the bed pressure drop versus superficial gas velocity curve for different distributor plates using PGAN granules with different particle sizes.*
 - The bed pressure drop against superficial gas velocity curve generated using the model predicted the experimentally obtained response. The trends obtained both experimentally and using the model mimic the observations by Kunii & Levenspiel (1997:72). The model can thus be concluded effective in reflecting the actual fluidisation curves.
 - The model predicted the observed higher pressure drop for particles with a low average particles diameter as well as the observed pressure differential of the unclassified sample in a bend inlet duct accurately to some extent. Based on this observation, the model can be concluded valid in predicting bed pressure differentials for various particle sizes.
- (iii) *Devise, using the CFD model, means to determine bed density and bed expansion for the porous granular ammonium nitrate (PGAN) fluidisation operation.*
 - Although the bed expansion ratios' values are in a fair agreement between the experimental and modelled observations, the experimental values seem to follow a different trend than

the modelled values. At high fluidisation velocities, the model seems to provide a better insight into bed expansion than at intermediate fluidisation velocities. The means devised using the model to determine bed expansion thus seems to be valid only at fluidisation velocities at the high end of the investigated range of velocities.

- For the bed density determination, it can be concluded that:
 - o the model is not a valid tool to determine the density of the fluidised bed, as the approximation of the plates with a porous medium in the model has led to a uniform distribution predicted by the model across the horizontal axis in the direction away from the blower position in a bent inlet duct, irrespective of the simulated plate,
 - o plates with lower differential pressures seem to allow a much more uneven distribution of air into the fluidised bed, compared to plates with a high differential pressure. This leads to the conclusion that plate D would be a much better distributor plate to use than plate A, as maldistribution of air would be eliminated.

From the above-summarised findings, it can be concluded that the target objectives have been realised to a large extent. Even though the model does not perfectly reproduce experimentally observed behaviour, there are basis with regards to the outcomes of the investigation which can be used as a foundation for future work.

5.2 Recommendations for further work

Although the model can predict the plates' differential pressure fairly well, there are significant error percentages that occurred between the model and experiment plate pressure drop magnitude. It is believed that the major cause of large errors was due to different probe positions between the model and experimental data. In the model, the pressure drop measurements were collected across the duct covering the entire cross-section whereas, in experiments, the probe positions are located at one side of the duct. It is therefore recommended that an exact replicate of the model should be employed in experiments in order to narrow down the underlying cause of errors.

The model was found to predicts a general fluidisation system. Little correlation between the model and the investigated PGAN fluidised bed was observed for the *twoPhaseEulerFoam* case. Possible adjustment to eliminate large errors in the system would be to investigate different drag models to find the model that can be used to accurately predict the actual fluidisation behaviours and to adjust the velocity of particles ($U_{particles}$) from either a fixed value to slipping.

The method of approximating the behaviour of the different distributor plates with velocity profiles in the source code is tedious and errors may occur in between transitions from one solver to the other, therefore it is recommended that the two solvers, i.e. *twoPhaseEulerFoam* and *porousSimpleFoam*

are combined into one solver in order to account for distribution effects and two-phase fluidisation behaviours in one system.

The simplification of the actual experimental setup to 2D analysis contributed to the discrepancies between model outputs and experimental results therefore it is recommended that a sensitivity analysis is conducted to compare the simulation outputs of both 2D and 3D fluidised bed setup.

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7. ANNEXURES

Appendix A: Modelling

Appendix A-1: Characterisation of air distribution medium

This chapter covers the model codes for plate characterisation in both a straight inlet duct and bend inlet duct. Firstly, the straight inlet duct added followed by the bend inlet duct.

0-epsilon

```
dimensions [0 2 -3 0 0 0 0];
internalField uniform 200;
boundaryField
{
    front
    {
        type epsilonWallFunction;
        value $internalField;
    }
    back
    {
        type epsilonWallFunction;
        value $internalField;
    }
    walls
    {
        type epsilonWallFunction;
        value $internalField;
    }
    porosityWall
    {
        type epsilonWallFunction;
        value $internalField;
    }
    inlet
    {
        type turbulentMixingLengthDissipationRateInlet;
        mixingLength 0.005;
        value $internalField;
    }
    outlet
    {
        type inletOutlet;
        inletValue $internalField;
        value $internalField;
    }
}
```

OF Entry A-1.1: Intial Epsilon wall effects for plate characterisation in SID setup

0-K

```
dimensions [0 2 -2 0 0 0 0];
internalField uniform 1;
boundaryField
{
    front
    {
```

```

    type      kqRWallFunction;
    value     $internalField;
}
back
{
    type      kqRWallFunction;
    value     $internalField;
}
walls
{
    type      kqRWallFunction;
    value     $internalField;
}
porosityWall
{
    type      kqRWallFunction;
    value     $internalField;
}
inlet
{
    type      turbulentIntensityKineticEnergyInlet;
    intensity 0.05;
    value     $internalField;
}
outlet
{
    type      inletOutlet;
    inletValue uniform 1;
    value     $internalField;
}
}

```

OF Entry A-1.2: Initial K-wall effects for plate characterisation in SID setup

0-nut

```

dimensions [0 2 -1 0 0 0 0];
internalField uniform 0;
boundaryField
{
    front
    {
        type      nutkWallFunction;
        value     $internalField;
    }

    back
    {
        type      nutkWallFunction;
        value     $internalField;
    }
    walls
    {
        type      nutkWallFunction;
        value     $internalField;
    }
    porosityWall
    {
        type      nutkWallFunction;
        value     $internalField;
    }
}

```

```

inlet
{
  type      calculated;
  value     $internalField;
}
outlet
{
  type      calculated;
  value     $internalField;
}
}

```

OF Entry A-1.3: Initial Nut wall effects for plate characterisation SID plate setup

0-P

```

dimensions [0 2 -2 0 0 0];
internalField uniform 0;
boundaryField
{
  front
  {
    type      zeroGradient;
  }
  back
  {
    type      zeroGradient;
  }
  walls
  {
    type      zeroGradient;
  }
  porosityWall
  {
    type      zeroGradient;
  }

  inlet
  {
    type      zeroGradient;
  }
  outlet
  {
    type      fixedValue;
    value     $internalField;
  }
}

```

OF Entry A-1.4: Initial pressure conditions for plate characterisation in SID plate setup

0-T

```

dimensions [0 0 0 1 0 0];
internalField uniform 293;
boundaryField
{
  front
  {

```

```

    type      zeroGradient;
}
back
{
    type      zeroGradient;
}
walls
{
    type      zeroGradient;
}
porosityWall
{
    type      zeroGradient;
}
inlet
{
    type      fixedValue;
    value     $internalField;
}
outlet
{
    type      inletOutlet;
    inletValue $internalField;
    value     $internalField;
}
}

```

OF Entry A-1.5: Initial temperature conditions for plate characterisation SID plate setup

0-U

```

dimensions [0 1 -1 0 0 0 0];
internalField uniform (0 0 0);
boundaryField
{
    front
    {
        type      noSlip;
    }
    back
    {
        type      noSlip;
    }
    walls
    {
        type      noSlip;
    }
    porosityWall
    {
        type      slip;
        value     $internalField;
    }
    inlet
    {
        // type      flowRateInletVelocity;
        //volumetricFlowRate constant 0.1;
        type      fixedValue;
        value     uniform (0.1 0 0);
    }
    outlet
    {
        type      inletOutlet;
    }
}

```

```

    inletValue    $internalField;
    value         $internalField;
  }
}

```

OF Entry A-1.6: Initial velocity conditions for plate characterisation in SID plate setup

Constant-porosityProperties

```

porosity1
{
    type          DarcyForchheimer;
    cellZone      porosity;
    d (1000 24910666.7 1000);
    f (0 106480 0);
    coordinateSystem
    {
        type      cartesian;
        origin (0 0 0);
        coordinateRotation
        {
            type      axesRotation;
            e1 (1 1 0);
            e2 (0 0 1);
        }
    }
}

```

OF Entry A-1.7: Constant properties of the porous medium in SID setup for plate characterisation

Constant-transportProperties

```

transportModel Newtonian;
nu          1.5e-05;

```

OF Entry A-1.8: Constant properties of the transport fluid in SID medium for plate characterisation

Constant-turbulenceProperties

```

simulationType RAS;
RAS
{
    RASModel      kEpsilon;
    turbulence     on;
    printCoeffs   on;
}

```

OF Entry A-1.9: Constant turbulence properties for plate characterisation in SID setup

System-blockMeshDict

```

convertToMeters 1;
vertices
(
    (0    0    0) //0
    (0.5  0    0) //1

```

```

(0.5 0.298 0) //2
(0 0.298 0) //3
(0 0 0.5) //4
(0.5 0 0.5) //5
(0.5 0.298 0.5) //6
(0 0.298 0.5) //7
(0 0.3 0) //8
(0.5 0.3 0) //9
(0.5 1 0) //10
(0 1 0) //11
(0 0.3 0.5) //12
(0.5 0.3 0.5) //13
(0.5 1 0.5) //14
(0 1 0.5) //15
);
blocks
(
  hex (0 1 2 3 4 5 6 7) inlet (5 20 5) simpleGrading (1 0.01 1) // inlet block
  hex (3 2 9 8 7 6 13 12) porosity (5 20 5) simpleGrading (1 1 1) // porosity block
  hex (8 9 10 11 12 13 14 15) outlet (5 20 5) simpleGrading (1 100 1) // outlet block
);
edges
(
);
patches
(
  wall walls
  (
    (4 5 6 7) // inlet block
    (5 1 2 6)
    (1 0 3 2)
    (0 4 7 3)
    (12 13 14 15) // outlet block
    (13 9 10 14)
    (9 8 11 10)
    (8 12 15 11)
    (7 6 13 12) // porous
    (6 2 9 13)
    (2 3 8 9)
    (3 7 12 8)
  )
  patch inlet
  (
    (0 1 5 4)
  )
  patch outlet
  (
    (15 14 10 11)
  )
);
mergePatchPairs

```

OF Entry A-1.10: Block mesh vertices for SID geometry generation

System-controlDict

```

application    porousSimpleFoam;
startFrom      startTime;
startTime      0;
stopAt         endTime;
endTime        100;
deltaT         1;

```

```

writeControl    timeStep;
writeInterval   10;
purgeWrite      0;
writeFormat     binary;
writePrecision  6;
writeCompression off;
timeFormat      general;
timePrecision   6;
graphFormat     raw;
runTimeModifiable true;

```

OF Entry A-1.11: Control conditions and specifications for plate characterisation in SID setup

System-fvSchemes

```

ddtSchemes
{
    default      steadyState;
}
gradSchemes
{
    default      Gauss linear;
}
divSchemes
{
    default      none;
    div(phi,U)   bounded Gauss upwind;
    div((nuEff*dev2(T(grad(U)))) Gauss linear;
    div(phi,epsilon) bounded Gauss upwind;
    div(phi,k)   bounded Gauss upwind;
}
laplacianSchemes
{
    default      Gauss linear corrected;
}
interpolationSchemes
{
    default      linear;
}
snGradSchemes
{
    default      corrected;
}

```

OF Entry A-1.12: Iteration functions for plate characterisation in SID setup

Systems-fvSolution

```

solvers
{
    p
    {
        solver      GAMG;
        tolerance    1e-08;
        relTol       0.05;
        smoother     GaussSeidel;
        nCellsInCoarsestLevel 20;
    }
    "(U|k|epsilon)"
    {
        solver      smoothSolver;
        smoother     GaussSeidel;
    }
}

```

```

    nSweeps      2;
    tolerance    1e-07;
    relTol       0.1;
  }
}
SIMPLE
{
  nUCorrectors  2;
  nNonOrthogonalCorrectors 0;
}
relaxationFactors
{
  fields
  {
    p           0.3;
  }
  equations
  {
    U           0.7;
    k           0.9;
    epsilon     0.9;
  }
}
}

```

OF Entry A-1.13: Solver conditions and specifications for plate characterisation in SID setup

Identical case was used for bend inlet duct the only difference was with U.air where the direction of air flow changed to x- direction and blockMesh which generates a different geometry of an elbow inlet at the duct resulting in bent inlet duct (BID). For the two phase system, a general twoPhase controls will be added and only different codes (blockMesh and U.air) will be added under SID and BID.

Appendix A-2: two-phase fluidised bed

0-alpha.particlesorg

```

dimensions      [0 0 0 0 0 0];
internalField    uniform 0;
boundaryField
{
  inlet
  {
    type         zeroGradient;
  }
  outlet
  {
    type         zeroGradient;
  }
  walls
  {
    type         zeroGradient;
  }
  frontAndBackPlanes
  {
    type         empty;
  }
}

```

OF Entry A-2.1: Initial conditions for alphas particles in the twoPhaseEulerFoam fluidisation system

0-alpha.particles

```
dimensions      [1 -1 -1 0 0 0];
internalField    uniform 0;
boundaryField
{
    inlet
    {
        type      calculated;
        value      $internalField;
    }
    outlet
    {
        type      calculated;
        value      $internalField;
    }
    walls
    {
        type      calculated;
        value      $internalField;
    }
    defaultFaces
    {
        type      empty;
    }
}
```

OF Entry A-2.2: Initial condition for the presences of alpha particles in TPEF fluidised bed

0-epsilon.air

```
dimensions      [0 2 -3 0 0 0 0];
internalField    uniform 10;
boundaryField
{
    inlet
    {
        type      fixedValue;
        value      $internalField;
    }
    outlet
    {
        type      inletOutlet;
        phi        phi.air;
        inletValue  $internalField;
        value       $internalField;
    }
    walls
    {
        type      epsilonWallFunction;
        value      $internalField;
    }
    frontAndBackPlanes
    {
        type      empty;
    }
}
```

OF Entry A-2.3: Initial epsilon to air wall effects in TPEF fluidised bed system

0-k.air

```
dimensions      [0 2 -2 0 0 0];
internalField    uniform 1;
boundaryField
{
    inlet
    {
        type      fixedValue;
        value      $internalField;
    }
    outlet
    {
        type      inletOutlet;
        phi        phi.air;
        inletValue  $internalField;
        value      $internalField;
    }
    walls
    {
        type      kqRWallFunction;
        value      $internalField;
    }
    frontAndBackPlanes
    {
        type      empty;
    }
}
```

OF Entry A-2.4: Initial K wall effect relative to air flow in TPEF fluidised bed system

0-nut.air

```
dimensions      [0 2 -1 0 0 0];
internalField    uniform 0;
boundaryField
{
    inlet
    {
        type      calculated;
        value      $internalField;
    }
    outlet
    {
        type      calculated;
        value      $internalField;
    }
    walls
    {
        type      nutkWallFunction;
        value      $internalField;
    }
    frontAndBackPlanes
    {
        type      empty;
    }
}
```

OF Entry A-2.5: Initial nut wall effects relative to air flow in TPEF fluidised bed system

0-nut.particles

```
dimensions      [0 2 -1 0 0 0 0];
internalField    uniform 0;
boundaryField
{
  inlet
  {
    type          calculated;
    value          $internalField;
  }
  outlet
  {
    type          calculated;
    value          $internalField;
  }
  walls
  {
    type          calculated;
    value          $internalField;
  }
  frontAndBackPlanes
  {
    type          empty;
  }
}
```

OF Entry A-2.6: Initial nut wall effects relative to particles in TPEF fluidised bed system

0-P

```
dimensions      [1 -1 -2 0 0 0 0];
internalField    uniform 1e5;
boundaryField
{
  inlet
  {
    type          calculated;
    value          $internalField;
  }
  outlet
  {
    type          calculated;
    value          $internalField;
  }
  walls
  {
    type          calculated;
    value          $internalField;
  }
  frontAndBackPlanes
  {
    type          empty;
  }
}
```

OF Entry A-2.7: Initial pressure conditions in a TPEF fluidised bed system

0-p_rgh

```
dimensions      [1 -1 -2 0 0 0 0];
```

```

internalField    uniform 1e5;
boundaryField
{
    inlet
    {
        type      fixedFluxPressure;
        value      $internalField;
    }
    outlet
    {
        type      prghPressure;
        p          $internalField;
        value      $internalField;
    }
    walls
    {
        type      fixedFluxPressure;
        value      $internalField;
    }
    frontAndBackPlanes
    {
        type      empty;
    }
}

```

OF Entry A-2.8: The initial density-pressure relationship in TPEF fluidised bed system

0-T.air

```

dimensions      [0 0 0 1 0 0 0];
internalField    uniform 600;
boundaryField
{
    inlet
    {
        type      fixedValue;
        value      uniform 300;
    }
    outlet
    {
        type      inletOutlet;
        phi        phi.air;
        inletValue  uniform 300;
        value      uniform 300;
    }
    walls
    {
        type      zeroGradient;
    }
    frontAndBackPlanes
    {
        type      empty;
    }
}

```

OF Entry A-2.9: Initial temperature relative to air flow in a TPEF fluidised bed system

0-T.particles

```

dimensions      [0 0 0 1 0 0 0];
internalField    uniform 600;
boundaryField
{

```

```

inlet
{
    type          zeroGradient;
}
outlet
{
    type          inletOutlet;
    phi           phi.particles;
    inletValue     uniform 300;
    value          uniform 300;
}
walls
{
    type          zeroGradient;
}
frontAndBackPlanes
{
    type          empty;
}
}

```

OF Entry A-2.10: Initial temperature relative to particles in a TPEF in fluidised bed system

0-Theta.particles

```

dimensions      [0 2 -2 0 0 0 0];
internalField    uniform 0;
referenceLevel   1e-4;
boundaryField
{
    inlet
    {
        type      fixedValue;
        value      uniform 1e-4;
    }
    outlet
    {
        type      zeroGradient;
    }
    walls
    {
        type      zeroGradient;
    }
    frontAndBackPlanes
    {
        type      empty;
    }
}

```

OF Entry A-2.11: Initial theta properties relative to particles in TPEF in fluidised bed system

0-U.air

```

dimensions      [0 1 -1 0 0 0 0];
internalField    uniform (0 0.25 0);
boundaryField
{
    inlet
    {
        type      interstitialInletVelocity;
        inletVelocity    uniform (0 0.3 0);
        alpha       alpha.air;
        value        $internalField;
    }
}

```

```

}

outlet
{
    type      pressureInletOutletVelocity;
    phi       phi.air;
    value     $internalField;
}
walls
{
    type      noSlip;
}
frontAndBackPlanes
{
    type      empty;
}

```

OF Entry A-2.12: Initial velocity condition relative to air in a TPEF fluidised bed system

0-U.particles

```

dimensions    [0 1 -1 0 0 0 0];
internalField uniform (0 0 0);
boundaryField
{
    inlet
    {
        type      fixedValue;
        value     uniform (0 0 0);
    }
    outlet
    {
        type      fixedValue;
        value     uniform (0 0 0);
    }
    walls
    {
        type      fixedValue;
        value     uniform (0 0 0);
    }
    frontAndBackPlanes
    {
        type      empty;
    }
}

```

OF Entry A-2.13: Initial velocity condition relative to particles in a TPEF fluidised bed system

Constant-g

```

dimensions    [0 1 -2 0 0 0 0];
value         (0 -9.81 0);

```

OF Entry A-2.14: Constant properties of fluidisation air

Constant-phaseProperties

```

phases (particles air);
particles
{
    diameterModel constant;
}

```

```

constantCoeffs
{
    d          1.8e-3;
}

alphaMax      0.62;
residualAlpha 1e-6;
}
air
{
    diameterModel constant;
    constantCoeffs
    {
        d          1;
    }
    residualAlpha 0;
}
blending
{
    default
    {
        type          none;
        continuousPhase air;
    }
}
sigma
(
    (particles and air) 0
);

aspectRatio
(
);
drag
(
    (particles in air)
    {
        type          GidaspowErgunWenYu;
        residualRe     1e-3;
        swarmCorrection
        {
            type        none;
        }
    }
);
virtualMass
(
    (particles in air)
    {
        type          constantCoefficient;
        Cvm           0.5;
    }
);
heatTransfer
(
    (particles in air)
    {
        type          RanzMarshall;
        residualAlpha 1e-3;
    }
);
lift

```

```
(
);
wallLubrication
(
);
turbulentDispersion
(
);
// Minimum allowable pressure
pMin      10000;
```

OF Entry A-2.15: Constant phase properties of both air and particles

Constant-thermophysicalProperties.air

```
thermoType
{
    type          heRhoThermo;
    mixture        pureMixture;
    transport      const;
    thermo         hConst;
    equationOfState perfectGas;
    specie         specie;
    energy         sensibleInternalEnergy;
}
mixture
{
    specie
    {
        molWeight 28.9;
    }
    thermodynamics
    {
        Cp        1007;
        Hf        0;
    }
    transport
    {
        mu        1.84e-05;
        Pr        0.7;
    }
}
```

OF Entry A-2.16: Constant thermophysical properties of air in the system

Constant-thermophysicalProperties.particles

```
thermoType
{
    type          heRhoThermo;
    mixture        pureMixture;
    transport      const;
    thermo         hConst;
    equationOfState rhoConst;
    specie         specie;
    energy         sensibleInternalEnergy;
}
mixture
{
    specie
    {
        molWeight 80;
```

```

}
equationOfState
{
    rho      1432;
}
thermodynamics
{
    Cp      6000;
    Hf      0;
}
transport
{
    mu      0;
    Pr      1;
}
}

```

OF Entry A-2.17: Constant thermophysical properties of particles in the TPEF system

Constant-turbulenceProperties.air

simulationType laminar;

OF Entry A-2.18: Constant turbulence properties of air in TPEF system

Constant-turbulenceproperties.particles

```

simulationType RAS;
RAS
{
    RASModel phasePressure;
    turbulence on;
    printCoeffs on;
    kineticTheoryCoeffs
    {
        equilibrium on;
        e 0.8;
        alphaMax 0.62;
        alphaMinFriction 0.5;
        viscosityModel Gidaspow;
        conductivityModel Gidaspow;
        granularPressureModel Lun;
        frictionalStressModel JohnsonJackson;
        radialModel SinclairJackson;
        JohnsonJacksonCoeffs
        {
            Fr 0.05;
            eta 2;
            p 5;
            phi 28.5;
            alphaDeltaMin 0.05;
        }
    }
    phasePressureCoeffs
    {
        preAlphaExp 500;
        expMax 1000;
        alphaMax 0.62;
        g0 1000;
    }
}

```

```
}
```

OF Entry A-2.19: Constant turbulence properties of particles in TPEF system

System-blockMeshDict

```
scale 1;
vertices
(
  (0 0 -0.01)
  (0.15 0 -0.01)
  (0.15 1 -0.01)
  (0 1 -0.01)
  (0 0 0.01)
  (0.15 0 0.01)
  (0.15 1 0.01)
  (0 1 0.01)
);
blocks
(
  hex (0 1 2 3 4 5 6 7) (30 200 1) simpleGrading (1 1 1)
);
edges
(
);
patches
(
  patch inlet
  (
    (1 5 4 0)
  )
  patch outlet
  (
    (3 7 6 2)
  )
  wall walls
  (
    (0 4 7 3)
    (2 6 5 1)
  )
  empty frontAndBackPlanes
  (
    (0 3 2 1)
    (4 5 6 7)
  )
);
mergePatchPairs
(
);
```

OF Entry A-2.20:Block mesh code for TPEF system geometry generation

System-controlDict

```
application    twoPhaseEulerFoam;
startFrom      startTime;
startTime      0;
```

```

stopAt      endTime;
endTime     2;
deltaT      2e-4;
writeControl  runTime;
writeInterval 0.01;
purgeWrite   0;
writeFormat  ascii;
writePrecision 6;
writeCompression off;
timeFormat   general;
timePrecision 6;
runTimeModifiable on;
adjustTimeStep no;
maxCo        0.9;
maxDeltaT     1e-05;
functions
{
    fieldAverage1
    {
        type      fieldAverage;
        libs      ("libfieldFunctionObjects.so");
        writeControl writeTime;
        fields
        (
            U.particles
            {
                mean      on;
                prime2Mean off;
                base       time;
            }
            U.air
            {
                mean      on;
                prime2Mean off;
                base       time;
            }
            alpha.particles
            {
                mean      on;
                prime2Mean off;
                base       time;
            }
            p
            {
                mean      on;
                prime2Mean off;
                base       time;
            }
        )
    }
}

```

OF Entry A-2.21: Control conditions and specifications of a TPEF fluidised bed system

System-setFieldsDict

```

defaultFieldValues
(
    volScalarFieldValue alpha.air 1
    volScalarFieldValue alpha.particles 0
);
regions
(
    boxToCell
    {

```

```

    box (0 0 -0.1) (0.15 0.1 0.1);
    fieldValue
    (
        volScalarFieldValue alpha.air 0.45
        volScalarFieldValue alpha.particles 0.55
    );
}
);

```

OF Entry A-2.22: Specification of position of particles in the fluidised bed system

System-fvSchemes

```

ddtSchemes
{
    default Euler;
}
gradSchemes
{
    default Gauss linear;
}
divSchemes
{
    default none;

    "div\(\phi,\alpha.*\)" Gauss vanLeer;
    "div\(\phi_{ir},\alpha.*\)" Gauss vanLeer;

    "div\(\alpha Rho \phi.*,U.*\)" Gauss limitedLinearV 1;
    "div\(\phi.*,U.*\)" Gauss limitedLinearV 1;

    "div\(\alpha Rho \phi.*,(h|e).*\)" Gauss limitedLinear 1;
    "div\(\alpha Rho \phi.*,K.*\)" Gauss limitedLinear 1;
    "div\(\alpha \phi.*,p\)" Gauss limitedLinear 1;

    div((((alpha.air*thermo:rho.air)*nuEff.air)*dev2(T(grad(U.air))))) Gauss linear;
}
laplacianSchemes
{
    default Gauss linear uncorrected;
    bounded Gauss linear uncorrected;
}
interpolationSchemes
{
    default linear;
}
snGradSchemes
{
    default uncorrected;
    bounded uncorrected;
}

```

OF Entry A-2.23: Iteration concept and method for the TPEF fluidised bed system

System-fvSolution

```

solvers
{
    "alpha.*"
    {
        nAlphaCorr 1;
        nAlphaSubCycles 2;
    }
}

```

```

    smoothLimiter 0.1;
    implicitPhasePressure yes;
    solver        smoothSolver;
    smoother      symGaussSeidel;
    tolerance     1e-9;
    relTol        0;
    minlter       1;
}
p_rgh
{
    solver        GAMG;
    smoother      DIC;
    tolerance     1e-8;
    relTol        0;
}
p_rghFinal
{
    $p_rgh;
    relTol        0;
}
"U.*"
{
    solver        smoothSolver;
    smoother      symGaussSeidel;
    tolerance     1e-5;
    relTol        0;
    minlter       1;
}
"(h|e).*"
{
    solver        smoothSolver;
    smoother      symGaussSeidel;
    tolerance     1e-6;
    relTol        0;
    minlter       1;
}
"Theta.*"
{
    solver        smoothSolver;
    smoother      symGaussSeidel;
    tolerance     1e-6;
    relTol        0;
    minlter       1;
}
"(k|epsilon).*"
{
    solver        smoothSolver;
    smoother      symGaussSeidel;
    tolerance     1e-5;
    relTol        0;
    minlter       1;
}
}

PIMPLE
{
    nOuterCorrectors 3;
    nCorrectors      1;
    nNonOrthogonalCorrectors 0;
}
relaxationFactors
{
    equations

```

```
{
    ".*"      1;
}
```

OF Entry A-2.24: The specification of solvers applied in TPEF in fluidised bed system

The source code for specifying the velocity profile as initial boundary condition.

```
FILE *f;
f = fopen("velValuesProfileA1.dat","w");
int iSum = 0;
//fprintf(f,"value    nonuniform    list<vector>\n");

forAll(mesh.Sf().boundaryField(), iregion)
{
    forAll(mesh.Sf().boundaryField()[iregion], b)
    {
        if( mesh.boundary()[iregion].name() == "inlet")
        {
            iSum += 1;
        }
    }
}

fprintf(f,"%d\n\n",iSum);

forAll(mesh.Sf().boundaryField(), iregion)
{
    double x,y,z;
    forAll(mesh.Sf().boundaryField()[iregion], b)
    {
        if( mesh.boundary()[iregion].name() == "inlet")
        {
            x = mesh.Cf().boundaryField()[iregion][b][0];
            y = mesh.Cf().boundaryField()[iregion][b][1];
            z = mesh.Cf().boundaryField()[iregion][b][2];
            U1.boundaryFieldRef()[iregion][b][0] = 0.0;
            U1.boundaryFieldRef()[iregion][b][1] = -356.04*pow(x,6)+1041.4*pow(x,5)-
            1191.3*pow(x,4)+672.38*pow(x,3)-194.93*pow(x,2)+28.476*x-0.0612;
            U1.boundaryFieldRef()[iregion][b][2] = 0.0;
            //vel = 0.25;
            //vel = -57.521*pow(x,6)+175.94*pow(x,5)-212.35*pow(x,4)+128.62*pow(x,3)-
            41.342*pow(x,2)+6.692*x-0.0335;
            printf("x,y,z = %d %le %le %le\n",b,x,y,z);
            iSum += 1;
            fprintf(f,"(0 %le 0)\n",U1.boundaryFieldRef()[iregion][b][1]);
        }
    }
}
fprintf(f,");\n");
fclose(f);
exit(0);
}
```

OF Entry A-2.25: Specification of initial velocity boundary condition in source code

Table A-2-1: Dimensions and base units for SI used in modelling

NO.	PROPERTY	SI-UNIT
1	MASS	KILOGRAMS (KG)
2	LENGTH	METRE (M)
3	TIME	SECONDS (S)
4	TEMPERATURE	KELVIN (K)
5	QUANTITY	MOLE (MOL)
6	CURRENT	AMPERE (A)
7	LUMINOUS INTENSITY	CANDELA (IS)

Appendix A-3: bend inlet duct fluidised bed

0-U.air

```

dimensions [0 1 -1 0 0 0 0];
internalField uniform (0 0.25 0);
boundaryField
{
    inlet
    {
        //type      interstitialInletVelocity;
        //inletVelocity    uniform (0 0.25 0);
        //alpha      alpha.air;
        //value      $internalField;
        type      fixedValue;
        value      nonuniform List<vector>
        30
        (
        (0 8.688013e-03 0)
        (0 2.441895e-01 0)
        (0 4.526534e-01 0)
        (0 6.364897e-01 0)
        (0 7.979506e-01 0)
        (0 9.391368e-01 0)
        (0 1.062004e+00 0)
        (0 1.168372e+00 0)
        (0 1.259925e+00 0)
        (0 1.338224e+00 0)
        (0 1.404712e+00 0)
        (0 1.460716e+00 0)
        (0 1.507456e+00 0)
        )
    }
}

```

```

(0 1.546051e+00 0)
(0 1.577523e+00 0)
(0 1.602803e+00 0)
(0 1.622734e+00 0)
(0 1.638082e+00 0)
(0 1.649535e+00 0)
(0 1.657709e+00 0)
(0 1.663155e+00 0)
(0 1.666362e+00 0)
(0 1.667760e+00 0)
(0 1.667728e+00 0)
(0 1.666593e+00 0)
(0 1.664638e+00 0)
(0 1.662104e+00 0)
(0 1.659194e+00 0)
(0 1.656077e+00 0)
(0 1.652890e+00 0)
);
}
outlet
{
    type          pressureInletOutletVelocity;
    phi           phi.air;
    value         $internalField;
}
walls
{
    type          noSlip;
}
frontAndBackPlanes
{
    type          empty;
}
}

```

OF Entry A-3.1: Initial velocity boundary conditions of air in a TPEF system with BID setup

The U.air in a straight duct was done the same way as the BID and the geometry of the fluidised bed was used to be the same as the base case.

Appendix B: Additional Results

Appendix B-1: Plate characterisation-Laboratory scale setup.

This chapter consists of dry pressure drop versus superficial gas velocity curves of modelled plate A, Plate B, Plate C and plate D compared to experiments.

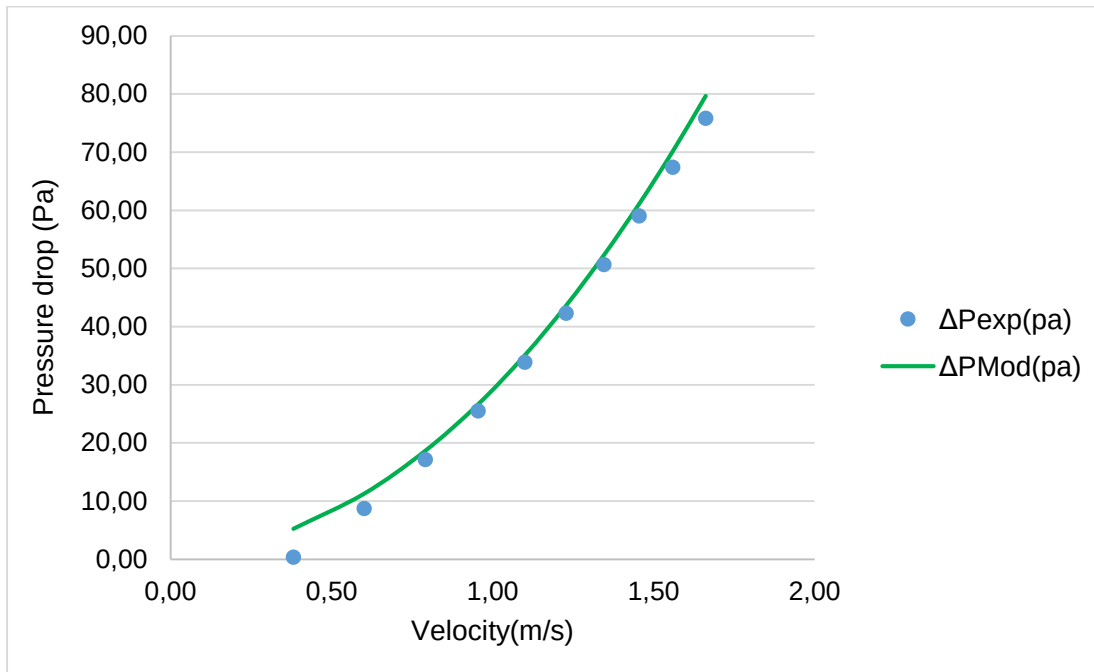


Figure B-1.1: Plate pressure drop against superficial gas velocity of plate A in laboratory scale setup.

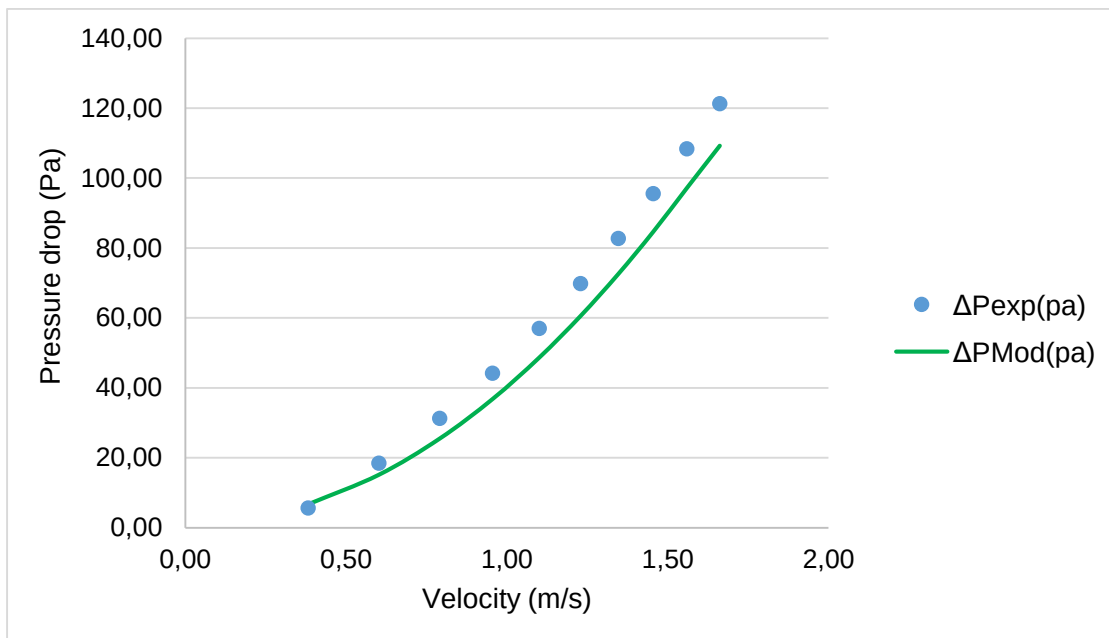


Figure B-1.2: Plate pressure drop against superficial gas velocity of plate B in laboratory scale setup

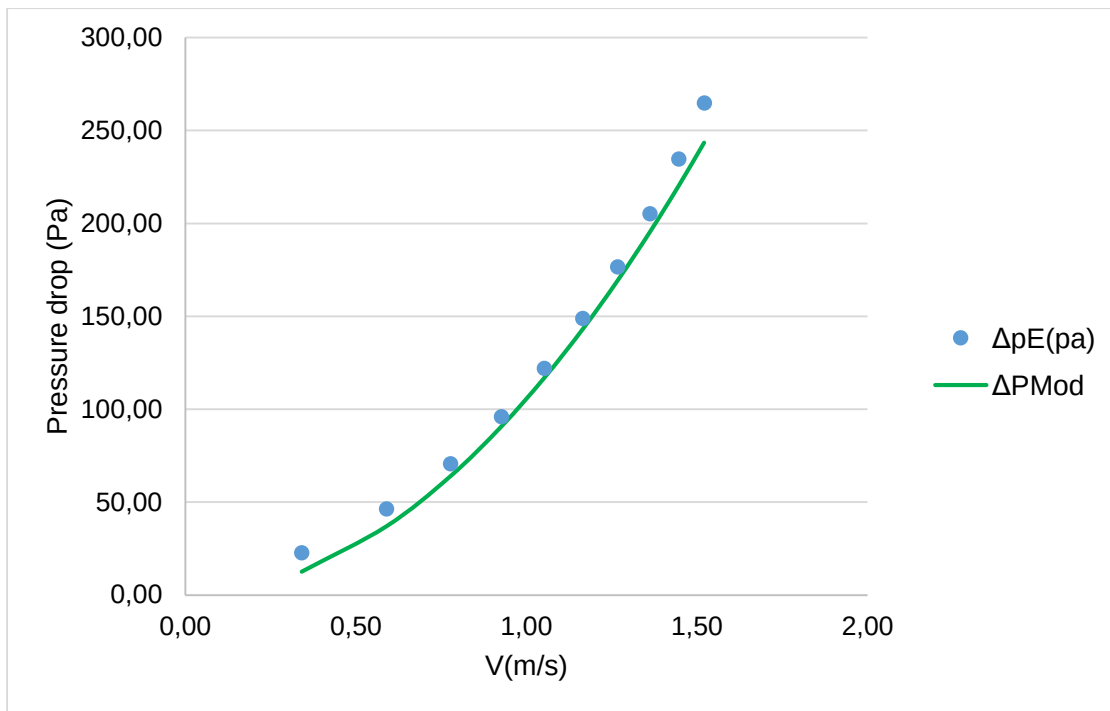


Figure B-1.3: Plate pressure drop against superficial gas velocity of plate C in laboratory scale setup.

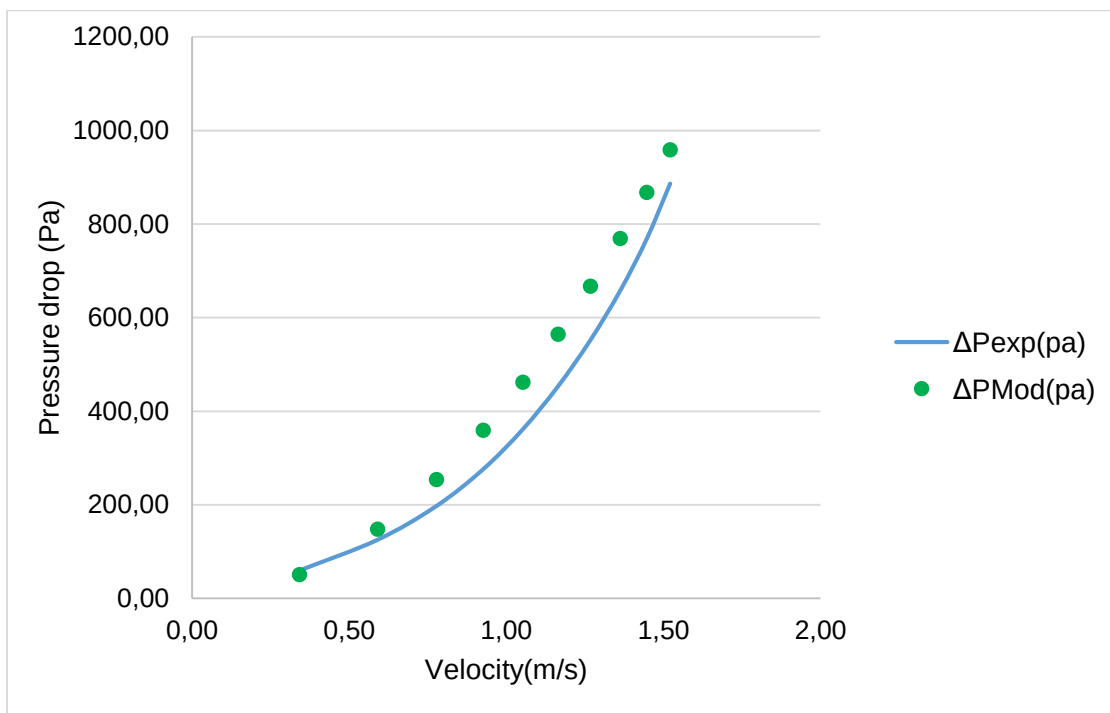


Figure B-1.4: Plate pressure drop against superficial gas velocity of plate D in laboratory scale setup

Velocity profiles of plate A, Plate B, Plate C and Plate D in a straight duct fluidised bed in a laboratory scale model

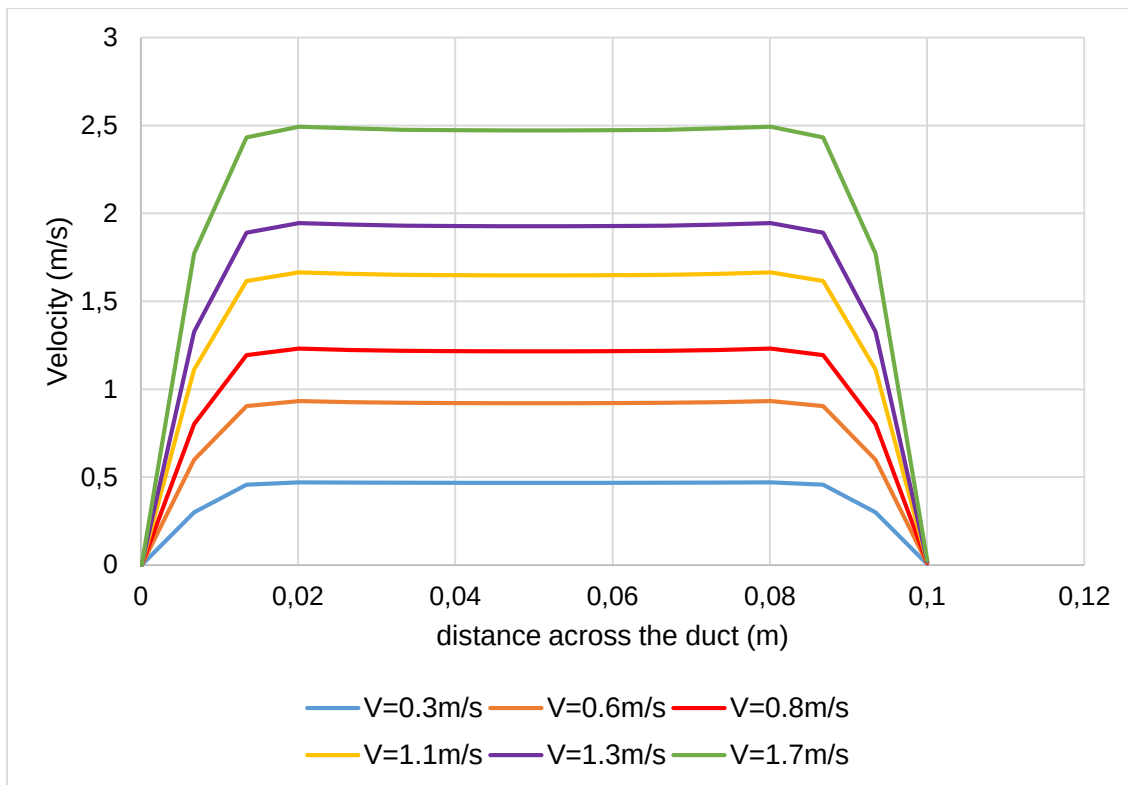


Figure B-1.5: Velocity profile of plate A in laboratory scale straight duct fluidised bed.

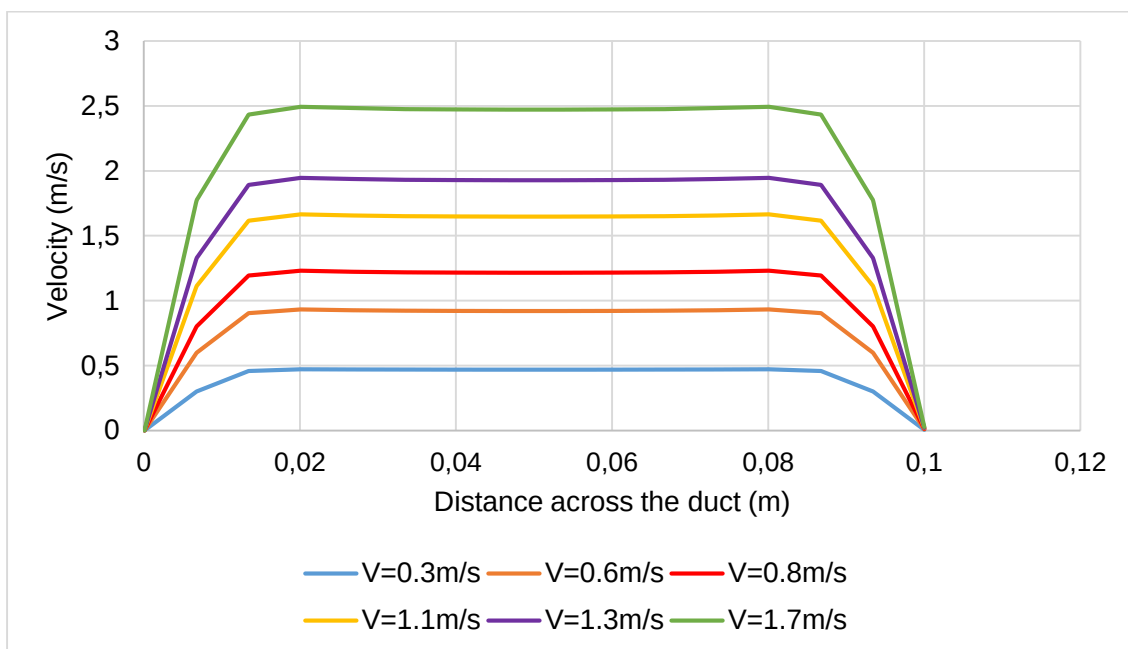


Figure B-1.6: Velocity profile of plate B in laboratory scale straight duct fluidised bed.

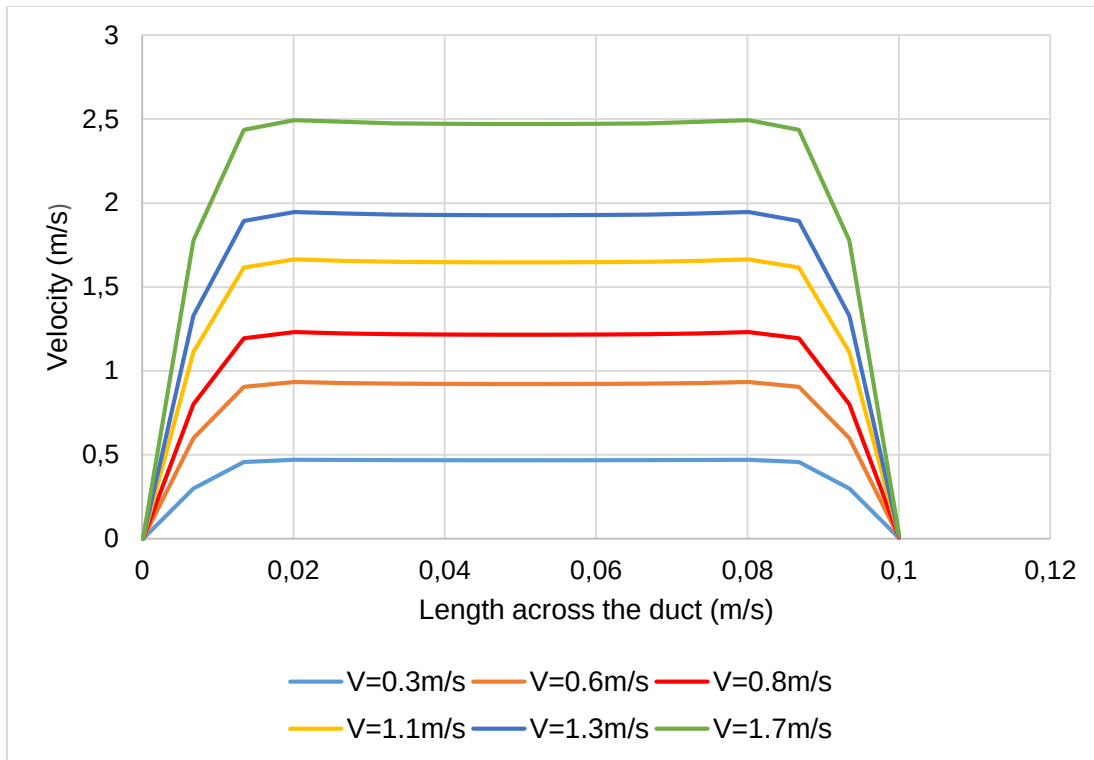


Figure B-1.7: Velocity profile of plate C in a laboratory scale straight duct fluidised bed.

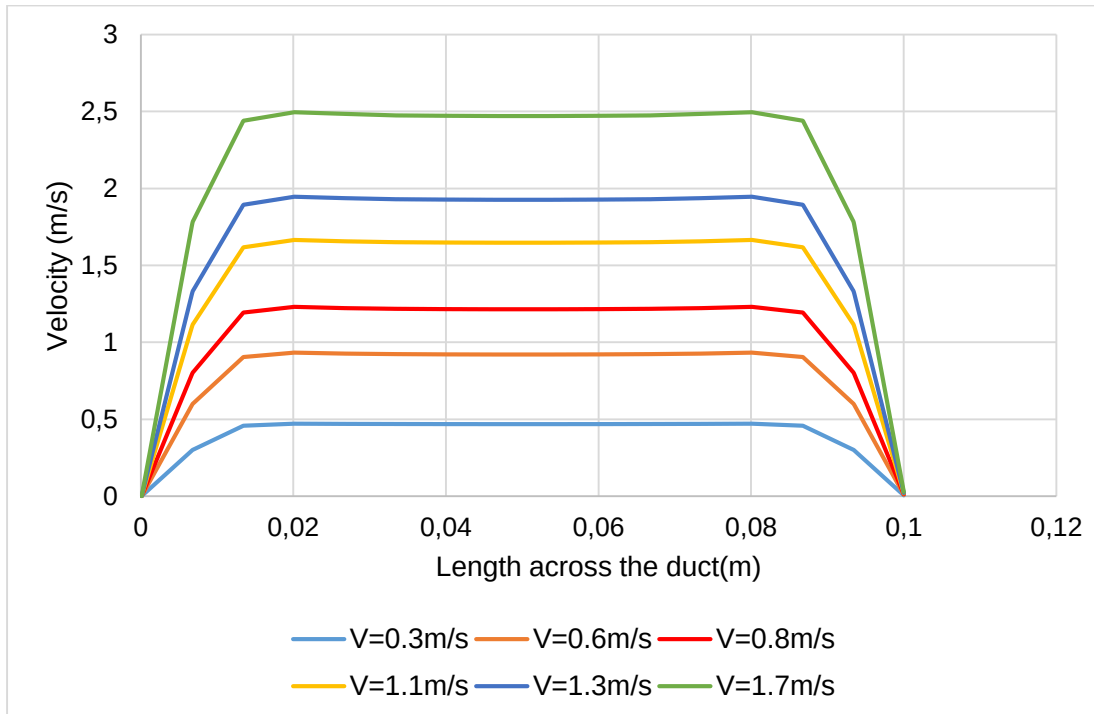


Figure B-1.8: Velocity profile of plate D in laboratory scale straight duct fluidised bed.

Velocity profile function of Plate A, Plate B, Plate C and Plate D using a straight duct fluidised bed;

Table B-1.1: Velocity profile functions of Plate A, B, C and D in a laboratory straight duct fluidised bed

v(m/s)	VP polynomial
0.3	$y = -3273.6x^6 + 4910.4x^5 - 2997.9x^4 + 951.89x^3 - 164.84x^2 + 14.582x - 0.0361$
0.6	$y = -6556.6x^6 + 9834.9x^5 - 6002.1x^4 + 1904.3x^3 - 329.15x^2 + 28.988x - 0.0704$
0.8	$y = -8967.2x^6 + 13451x^5 - 8187.6x^4 + 2583.1x^3 - 442.75x^2 + 38.596x - 0.091$
1.1	$y = -12889x^6 + 19333x^5 - 11717x^4 + 3662.1x^3 - 618.81x^2 + 53.019x - 0.1174$
1.3	$y = -15821x^6 + 23731x^5 - 14338x^4 + 4449.7x^3 - 743.67x^2 + 62.827x - 0.1326$
1.7	$y = -22304x^6 + 33456x^5 - 20100x^4 + 6159.9x^3 - 1008.8x^2 + 82.904x - 0.1591$
Plate B	
0.3	$y = -3273.4x^6 + 4910.1x^5 - 2997.7x^4 + 951.85x^3 - 164.83x^2 + 14.581x - 0.0361$
0.6	$y = -6556.6x^6 + 9834.9x^5 - 6002.1x^4 + 1904.3x^3 - 329.14x^2 + 28.987x - 0.0704$
0.8	$y = -8967.5x^6 + 13451x^5 - 8187.9x^4 + 2583.2x^3 - 442.76x^2 + 38.596x - 0.091$
1.1	$y = -12897x^6 + 19346x^5 - 11725x^4 + 3664.2x^3 - 619.09x^2 + 53.034x - 0.1175$
1.3	$y = -15839x^6 + 23759x^5 - 14354x^4 + 4454.2x^3 - 744.29x^2 + 62.86x - 0.1327$
1.7	$y = -22355x^6 + 33532x^5 - 20144x^4 + 6172.2x^3 - 1010.4x^2 + 82.989x - 0.1593$
Plate C	
0.3	$y = -3272.9x^6 + 4909.3x^5 - 2997.3x^4 + 951.73x^3 - 164.81x^2 + 14.581x - 0.0361$
0.6	$y = -6557x^6 + 9835.5x^5 - 6002.4x^4 + 1904.3x^3 - 329.15x^2 + 28.987x - 0.0704$
0.8	$y = -8972x^6 + 13458x^5 - 8191.7x^4 + 2584.3x^3 - 442.9x^2 + 38.602x - 0.091$
1.1	$y = -12920x^6 + 19380x^5 - 11745x^4 + 3669.8x^3 - 619.85x^2 + 53.074x - 0.1176$
1.3	$y = -15892x^6 + 23838x^5 - 14399x^4 + 4467.1x^3 - 746.06x^2 + 62.954x - 0.1329$
1.7	$y = -22496x^6 + 33743x^5 - 20266x^4 + 6206.3x^3 - 1015.1x^2 + 83.23x - 0.1598$
Plate D	
0.3	$y = -3272.5x^6 + 4908.8x^5 - 2997x^4 + 951.65x^3 - 164.81x^2 + 14.58x - 0.0361$
0.6	$y = -6557.6x^6 + 9836.4x^5 - 6002.9x^4 + 1904.5x^3 - 329.16x^2 + 28.987x - 0.0704$
0.8	$y = -8976.1x^6 + 13464x^5 - 8195.3x^4 + 2585.2x^3 - 443.02x^2 + 38.608x - 0.091$
1.1	$y = -12941x^6 + 19412x^5 - 11763x^4 + 3675.1x^3 - 620.59x^2 + 53.114x - 0.1177$
1.3	$y = -15939x^6 + 23908x^5 - 14441x^4 + 4479x^3 - 747.76x^2 + 63.051x - 0.1332$
1.7	$y = -22618x^6 + 33927x^5 - 20372x^4 + 6236x^3 - 1019.1x^2 + 83.443x - 0.1604$

Overall $y = -3300, +5000, -30000, +952, -165, 14.6, -0.04$

$y = -6557, +9835, -6002, +1904, -329, +30, -0.07$

$y = -8967, +13451, -8188, +2583, -443, +39, -0.09$

$y = -12889, +19333, -11717, +3662, -619, +53, -0.12$

$y = -15821, +23731, -14338, +4450, -744, +63, -0.13$

$y = -22000, +33000, -20200, +6200, -1019, +83, -0.16$

The polynomial functions were used to specify the initial behaviour of different plate in the source code. The behaviour was specified as an initial boundary condition with generates a set of velocities with the fluidised bed geometry. Since the polynomial functions in a straight duct fluidised bed are approximately equal, an average function was used instead of individual polynomials per plate and per superficial gas velocity. A block profile was observed for plates in a straight inlet duct, with maximum reachable velocity of approximately 2.5 m/s. From Figure B-5 to Figure B-8, it can be seen that the profiles of all plates are identical with regards to shape and value of the velocity profile.

Similar method was used to generate the velocity profile and functions thereof of Plate A, Plate B, Plate C and Plate D when modelling a bend inlet fluidised bed duct. The functions were again used to specify the initial behaviour of the distributor plate in the system. The velocity profile plots and polynomial functions are represented.

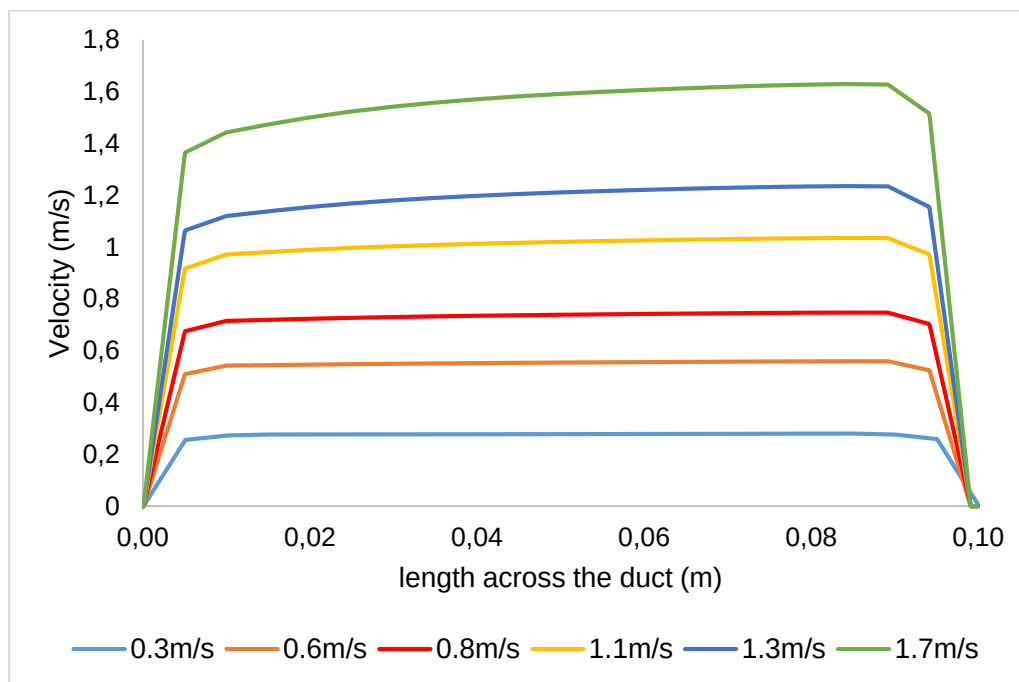


Figure B-1.9: Velocity profile of plate A in a bend inlet fluidised bed

Table B-1.2: Velocity function polynomial of generated by plate A using bend inlet fluidised bed in laboratory scale setup.

V(M/S)	VELOCITY POLYNOMIAL OF PLATE A
0,3	$y = -58.562x^6 + 173.4x^5 - 201.92x^4 + 117.17x^3 - 35.396x^2 + 5.2984x - 0.0109$
0,6	$y = -103.94x^6 + 299.19x^5 - 336.79x^4 + 187.64x^3 - 54.413x^2 + 8.2773x - 0.002$
0,8	$y = -144.5x^6 + 417.98x^5 - 472.89x^4 + 264.45x^3 - 76.555x^2 + 11.467x - 0.0144$
1,1	$y = -202.73x^6 + 589.01x^5 - 669.47x^4 + 375.78x^3 - 108.76x^2 + 16.146x - 0.0363$
1,3	$y = -246.14x^6 + 717.74x^5 - 818.66x^4 + 460.82x^3 - 133.41x^2 + 19.612x - 0.042$
1,7	$y = -320.94x^6 + 937.11x^5 - 1071x^4 + 605.04x^3 - 176.29x^2 + 26.046x - 0.059$

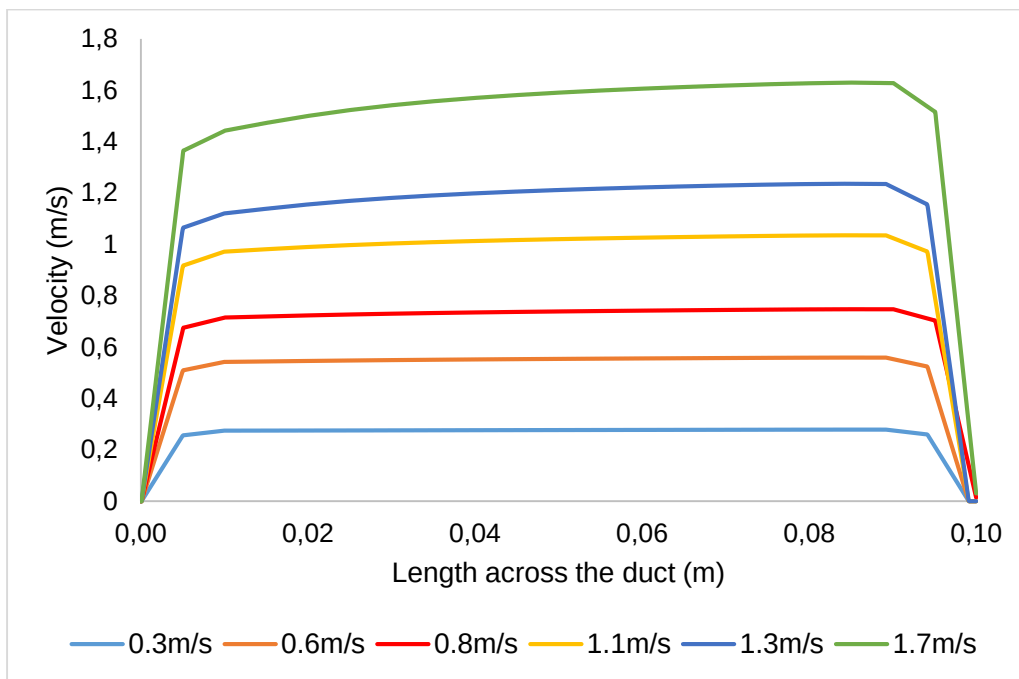


Figure B-1.10: Velocity profile of plate B in a bend inlet fluidised bed

Table B-1.3: Velocity function polynomial generated by plate B using bend inlet fluidised bed in a laboratory scale setup.

V(M/S)	VELOCITY POLYNOMIALS OF PLATE B
0,3	$y = -58.56x^6 + 173.4x^5 - 201.91x^4 + 117.17x^3 - 35.395x^2 + 5.2984x - 0.0109$
0,6	$y = -104.02x^6 + 299.33x^5 - 336.84x^4 + 187.62x^3 - 54.4x^2 + 8.2782x - 0.0019$
0,8	$y = -144.55x^6 + 418.15x^5 - 473.09x^4 + 264.57x^3 - 76.582x^2 + 11.469x - 0.0144$

1,1	$y = -202.85x^6 + 589.35x^5 - 669.85x^4 + 375.97x^3 - 108.8x^2 + 16.15x - 0.0363$
1,3	$y = -246.15x^6 + 717.79x^5 - 818.7x^4 + 460.83x^3 - 133.41x^2 + 19.611x - 0.0421$
1,7	$y = -319.94x^6 + 934.07x^5 - 1067.5x^4 + 603.15x^3 - 175.83x^2 + 26x - 0.0586$

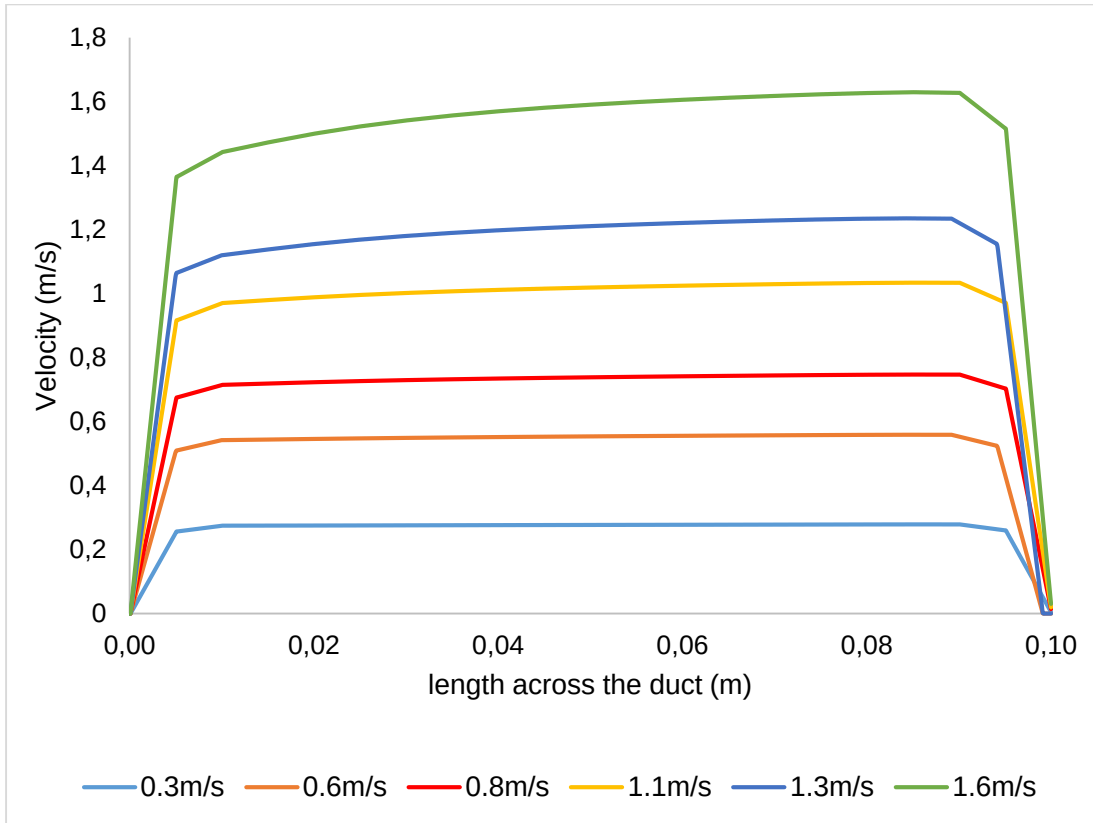


Figure B-1.11: Velocity profile of Plate C in a bend inlet fluidised bed

Table B-1.4: Velocity function polynomial generated by Plate C using bend inlet fluidised bed in a laboratory scale setup.

V(M/S) VELOCITY POLYNOMIALS OF PLATE C

0,3	$y = -52.897x^6 + 154.26x^5 - 176.52x^4 + 100.48x^3 - 29.902x^2 + 4.5664x - 0.002$
0,6	$y = -105.27x^6 + 307.8x^5 - 353.9x^4 + 203.15x^3 - 61.226x^2 + 9.4542x - 0.0204$
0,8	$y = -141.98x^6 + 410.55x^5 - 464.3x^4 + 259.78x^3 - 75.468x^2 + 11.417x - 0.0204$
1,1	$y = -201.11x^6 + 584.99x^5 - 665.8x^4 + 374.49x^3 - 108.81x^2 + 16.244x - 0.0433$
1,3	$y = -242.91x^6 + 709.06x^5 - 809.85x^4 + 456.94x^3 - 132.9x^2 + 19.672x - 0.0497$
1,7	$y = -314.11x^6 + 918.1x^5 - 1050.6x^4 + 594.87x^3 - 174.12x^2 + 25.951x - 0.0679$

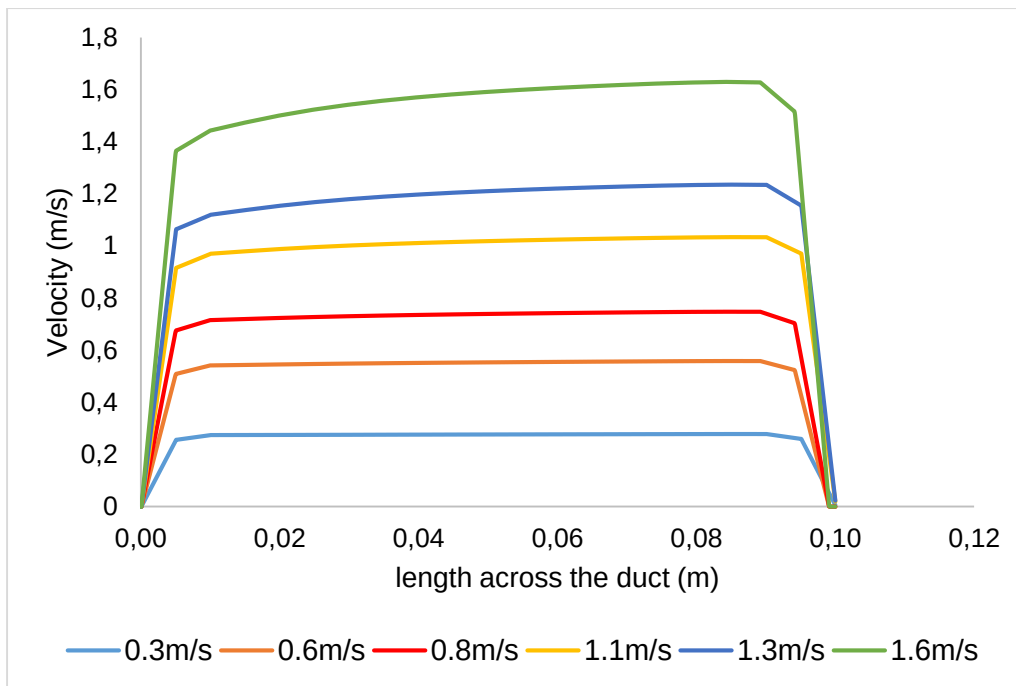


Figure B-1.2: Velocity of Plate D in a bend inlet fluidised bed

Table B-1.4: Velocity function polynomial generated by Plate D using bend inlet fluidised bed in a laboratory scale setup.

V(M/S)	VELOCITY POLYNOMIALS OF PLATE D
0,3	$y = -60.233x^6 + 178.98x^5 - 209.18x^4 + 121.78x^3 - 36.811x^2 + 5.468x - 0.0119$
0,6	$y = -103.68x^6 + 298.37x^5 - 335.79x^4 + 187.03x^3 - 54.234x^2 + 8.2575x - 0.0019$
0,8	$y = -147.71x^6 + 429.2x^5 - 488.5x^4 + 275.2x^3 - 80.17x^2 + 11.926x - 0.0147$
1,1	$y = -202.78x^6 + 589.15x^5 - 669.64x^4 + 375.87x^3 - 108.78x^2 + 16.149x - 0.0363$
1,3	$y = -246.05x^6 + 717.57x^5 - 818.56x^4 + 460.81x^3 - 133.42x^2 + 19.611x - 0.0421$

Appendix B-2: Plate characterisation-pilot scale setup

The distributor plates used in the pilot scale were characterised in order to investigate their impact on fluidisation process. The Plates have been investigated; Plate A, Plate B and Plate C.

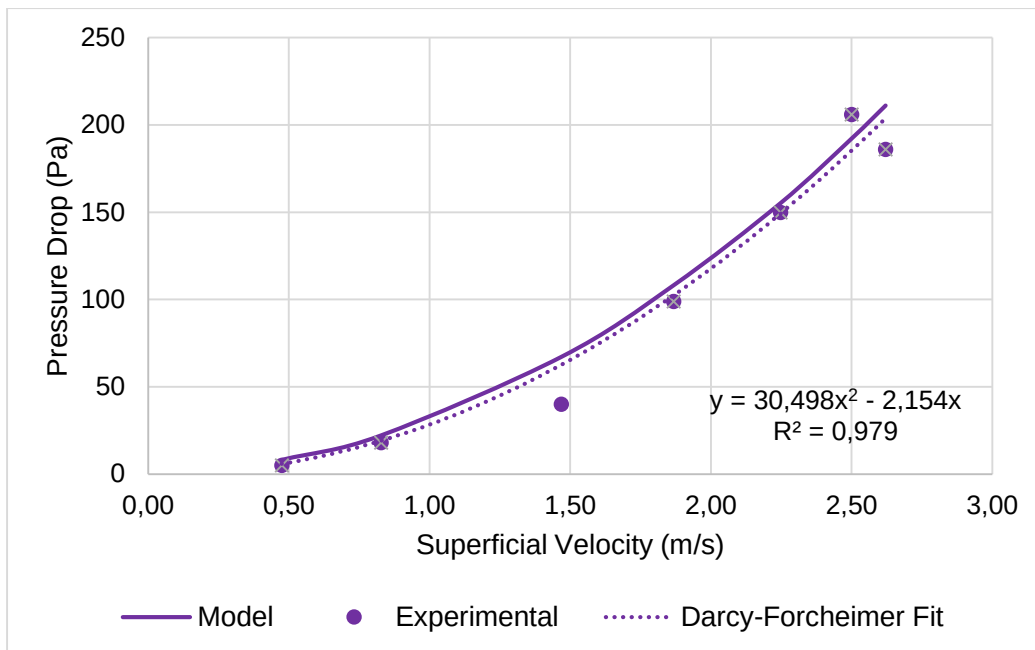


Figure B-2.1: Plate pressure drop against superficial gas velocities of Plate A comparing pilot scale against model data

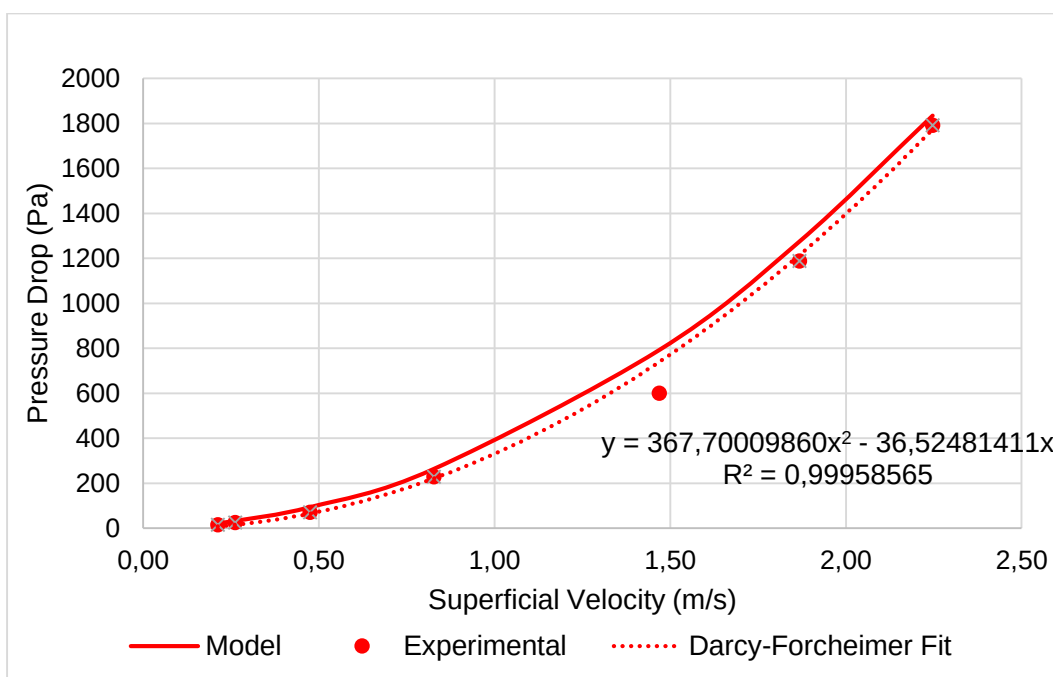


Figure B-2.2: Plate pressure drop against superficial gas velocity of Plate B comparing pilot scale against model data

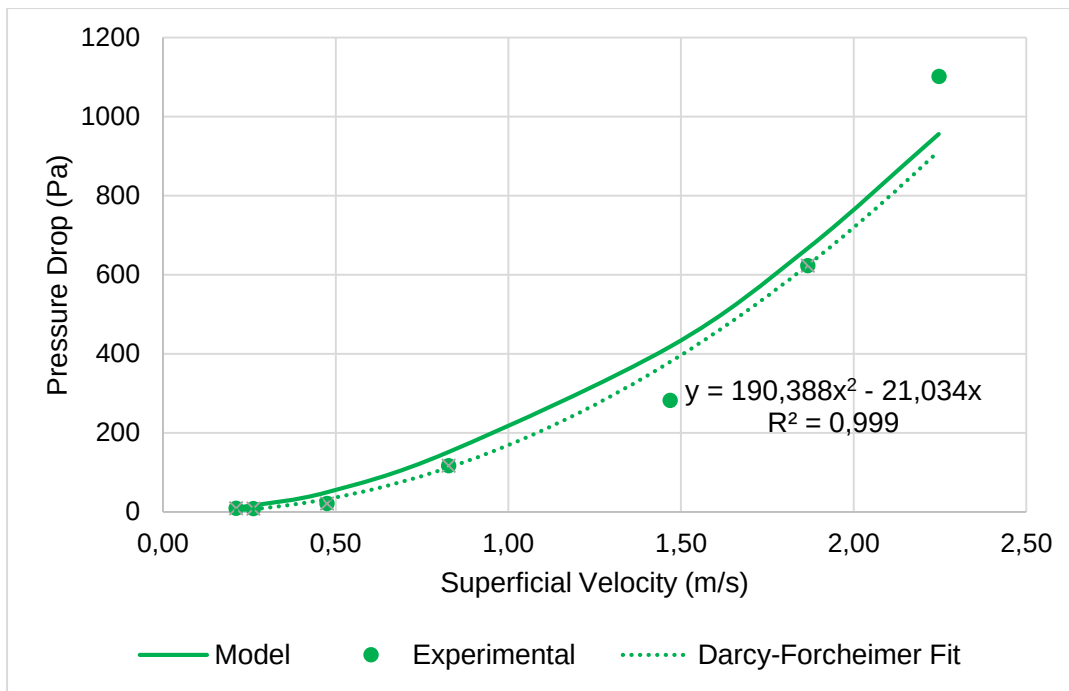


Figure B-2.3: Plate pressure drop against superficial gas velocity of Plate C comparing pilot scale against model data.

The velocity profiles of Plate A, Plate b and Plate C have been generated using a column fluidised bed and conical fluidised bed. Using the model, the distributor plates investigated using a column duct was Plate A, Plate B and Plate C.

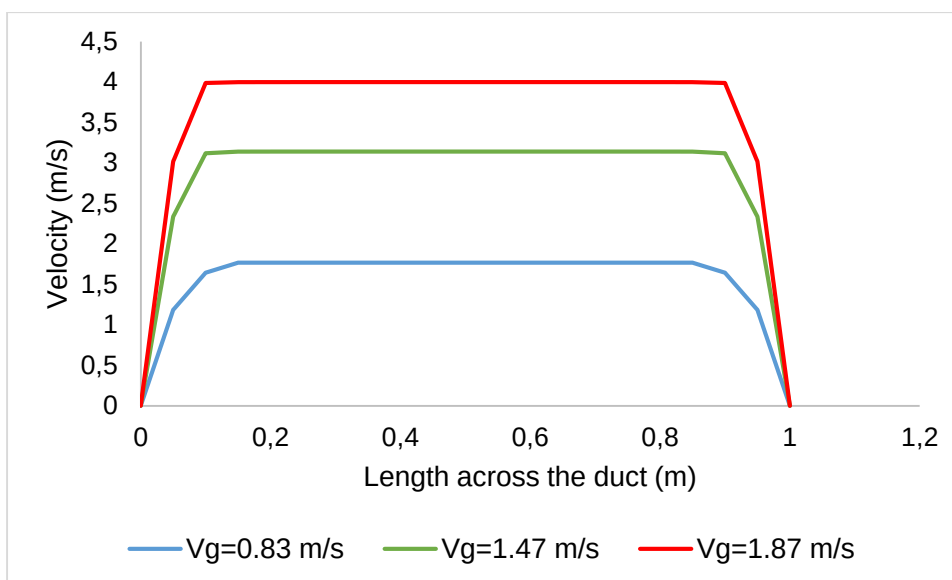


Figure B-2.4: Velocity profile of Plate A in a column pilot scale fluidised bed.

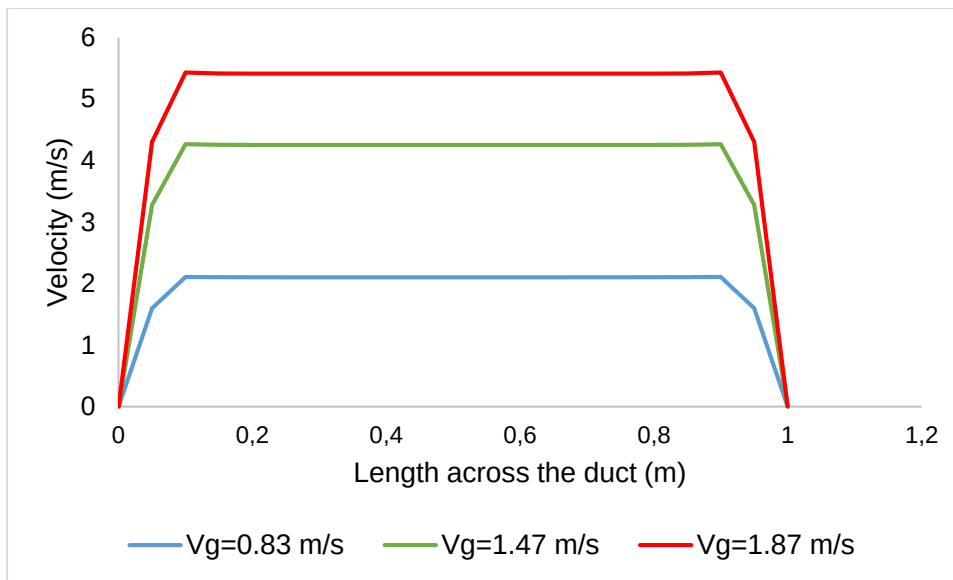


Figure B-2.5: Velocity profile of Plate B in a column pilot scale fluidised bed.

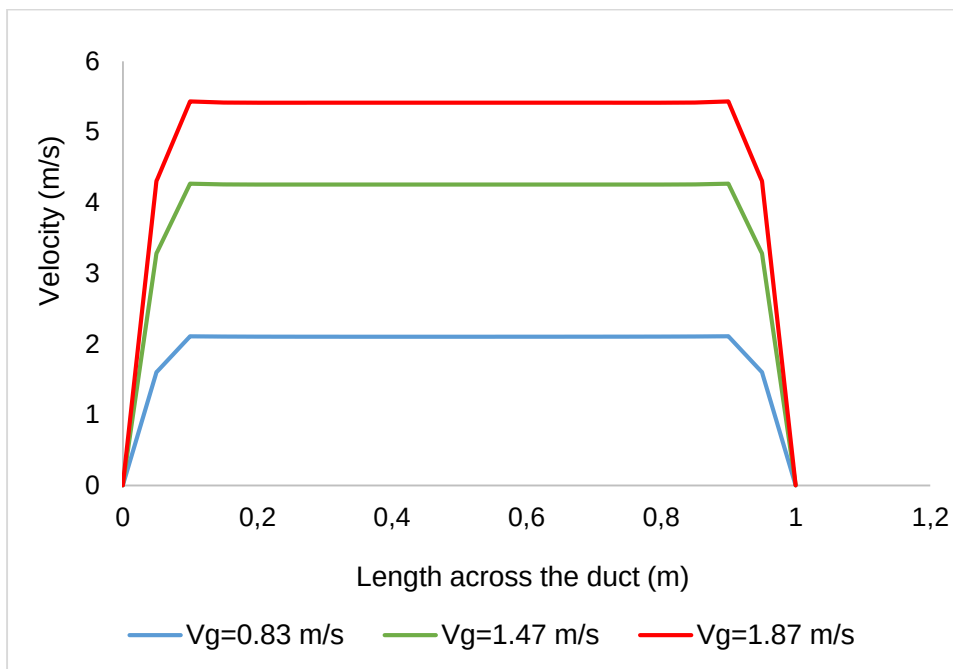


Figure B-2.6: Velocity profile of Plate B in a column pilot scale fluidised bed.

The velocity profiles of Plate A and Plate C were investigated in a conical fluidised bed system, which is an exact replicate of the pilot scale setup used at Omnia Sasolburg plant.

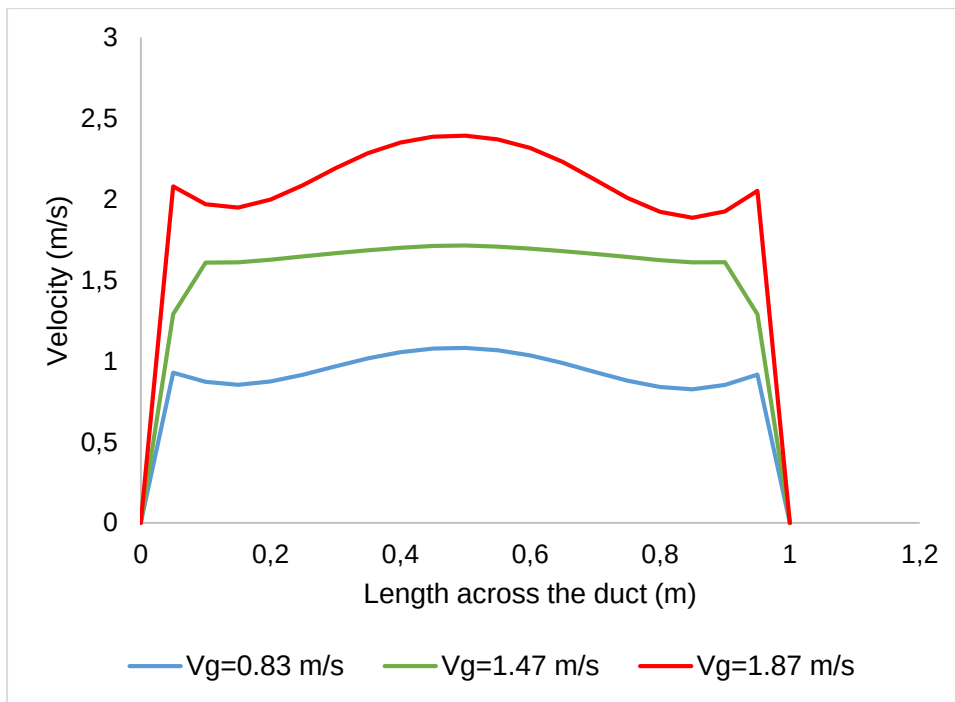


Figure B-2.7: Velocity profile of Plate A in a conical pilot scale fluidised bed.

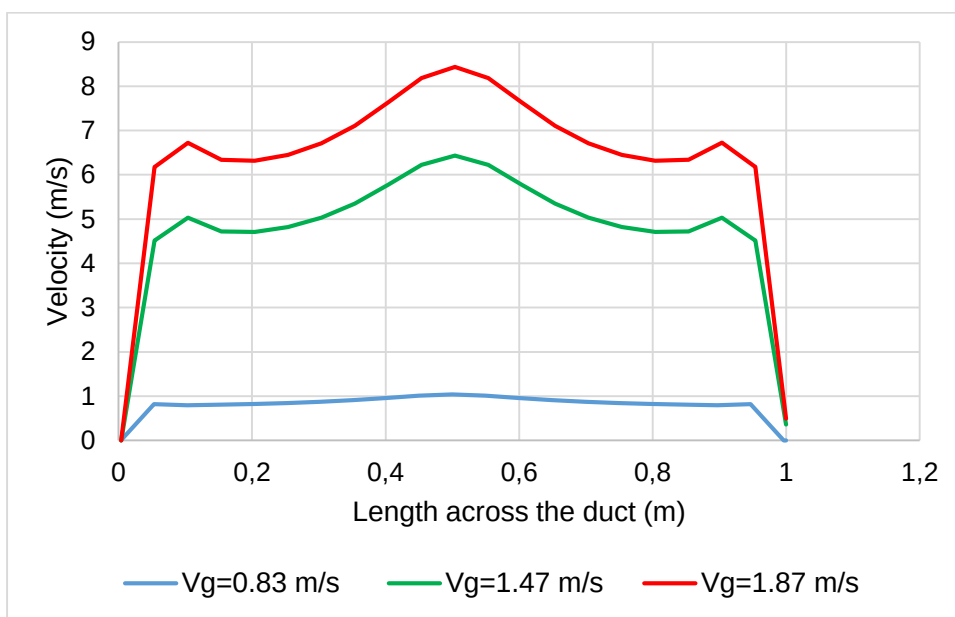


Figure B-2.8: Velocity profile of Plate C in a conical pilot scale fluidised bed.

Appendix B-3: Bed pressure drop - Laboratory scale setup.

Comparison of bed pressure drop against superficial gas velocity curves for three model plates; Ideal plate, straight duct plate and bend inlet plate A for particles with different sizes.

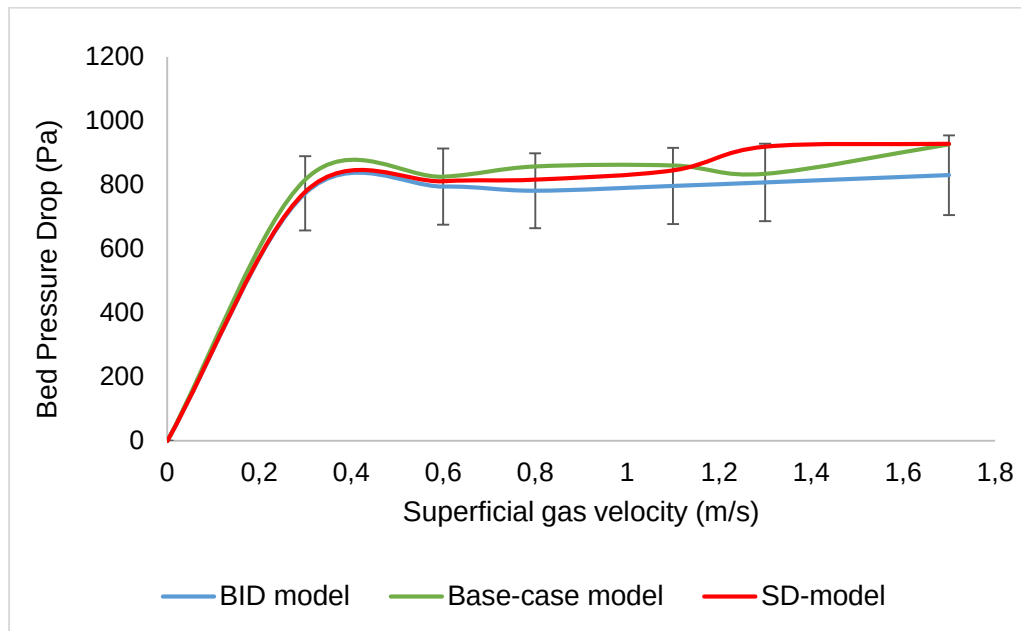


Figure B-3.1: Bed pressure drop against superficial gas velocity of $d_p=1.3\text{mm}$ particles modelled with bend inlet Plate A, Base case and straight duct fluidised bed.

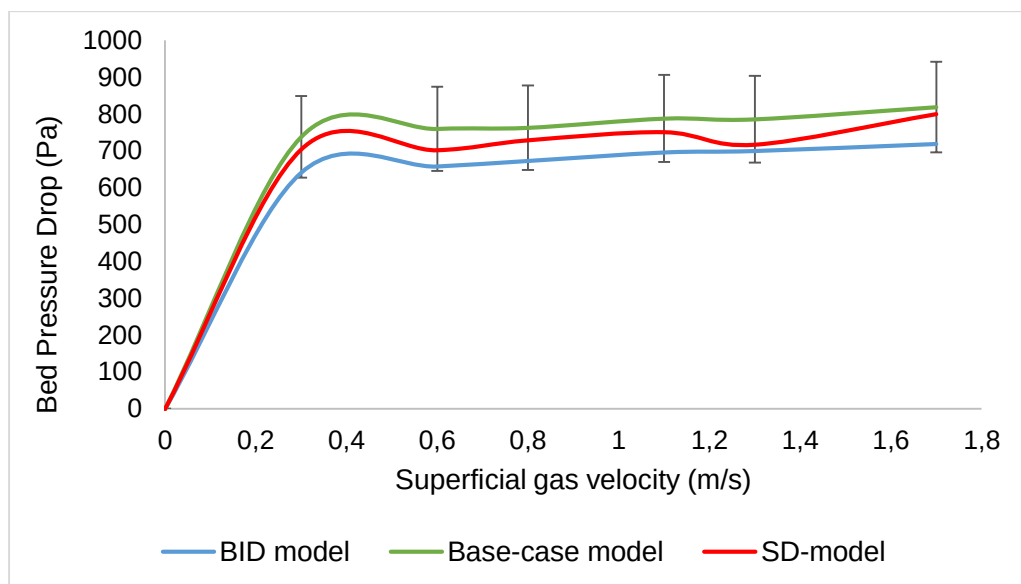


Figure B-3.2: Bed pressure drop against superficial gas velocity of $d_p=1.6\text{mm}$ particles modelled with bend inlet Plate A, Base case and straight duct fluidised bed.

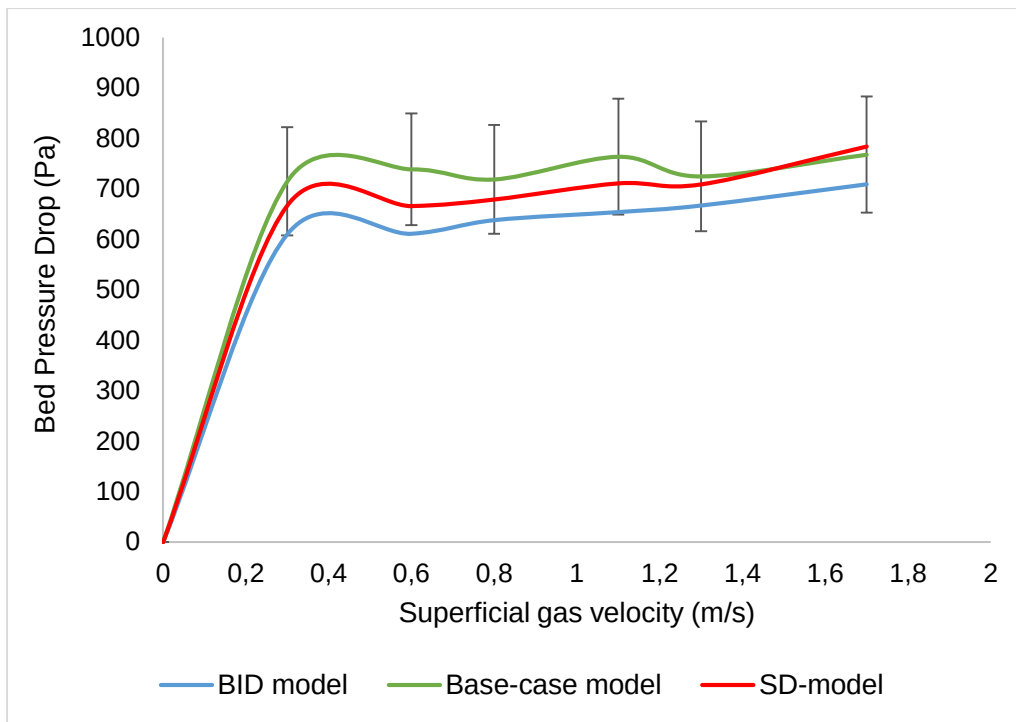


Figure B-3.3: Bed pressure drop against superficial gas velocity of $d_p=1.7\text{mm}$ particles modelled with bend inlet Plate A, Base case and straight duct fluidised bed.

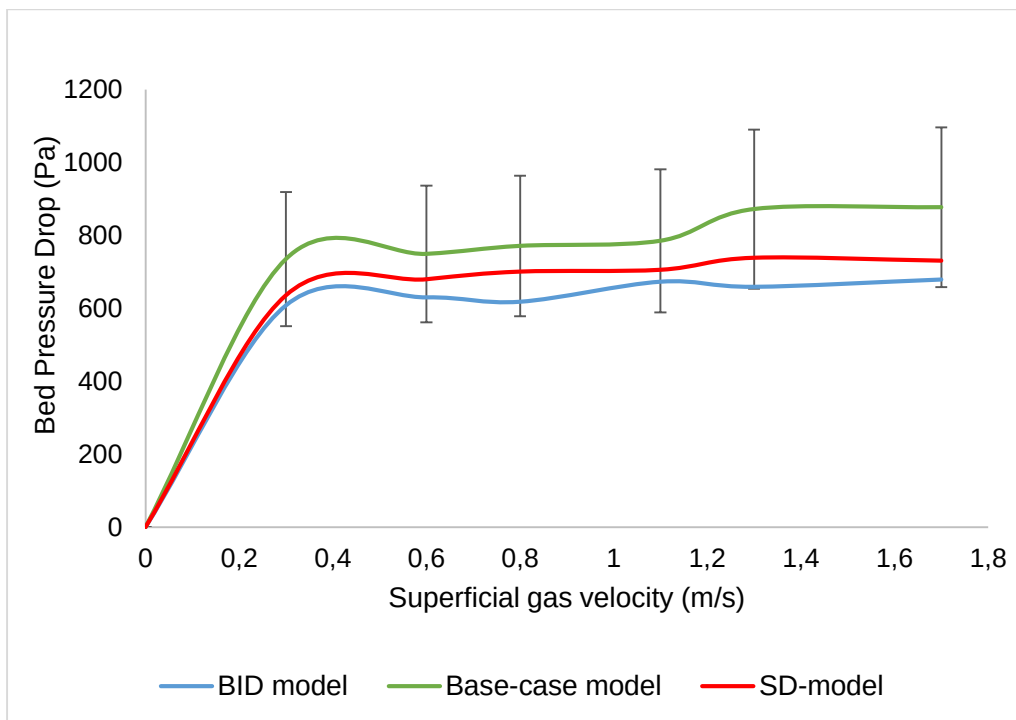


Figure B-3.4: Bed pressure drop against superficial gas velocity of $d_p=1.85\text{mm}$ particles modelled with bend inlet Plate A, Base case and straight duct fluidised bed.

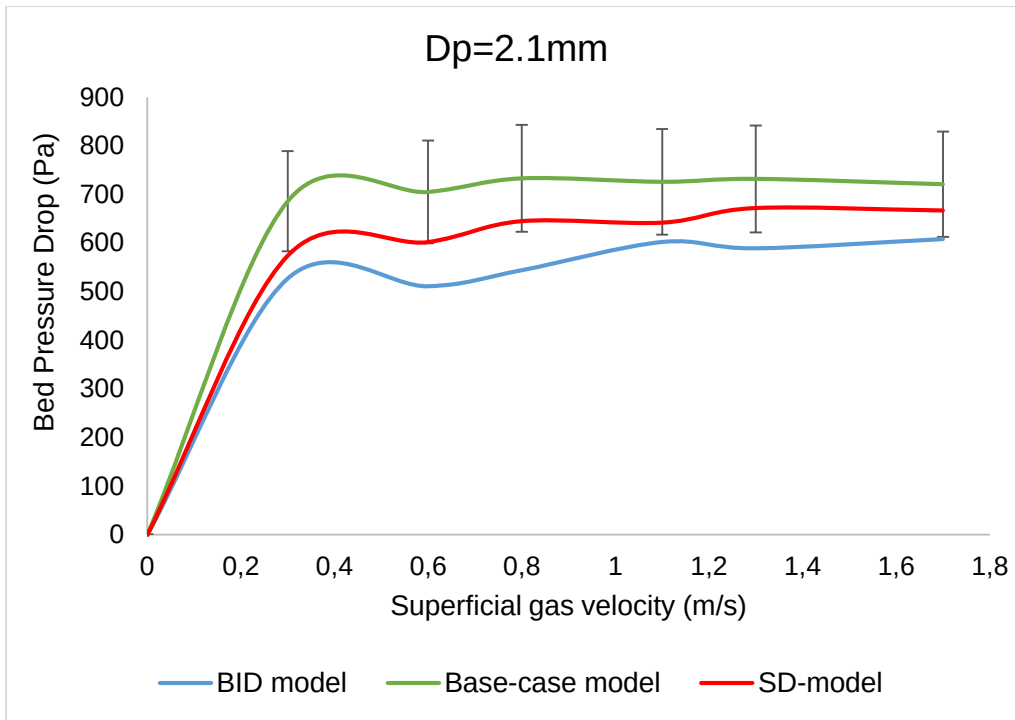


Figure B-3.5: Bed pressure drop against superficial gas velocity of 2.1mm particles modelled with Bend Inlet Plate A, Base case and straight duct fluidised bed.

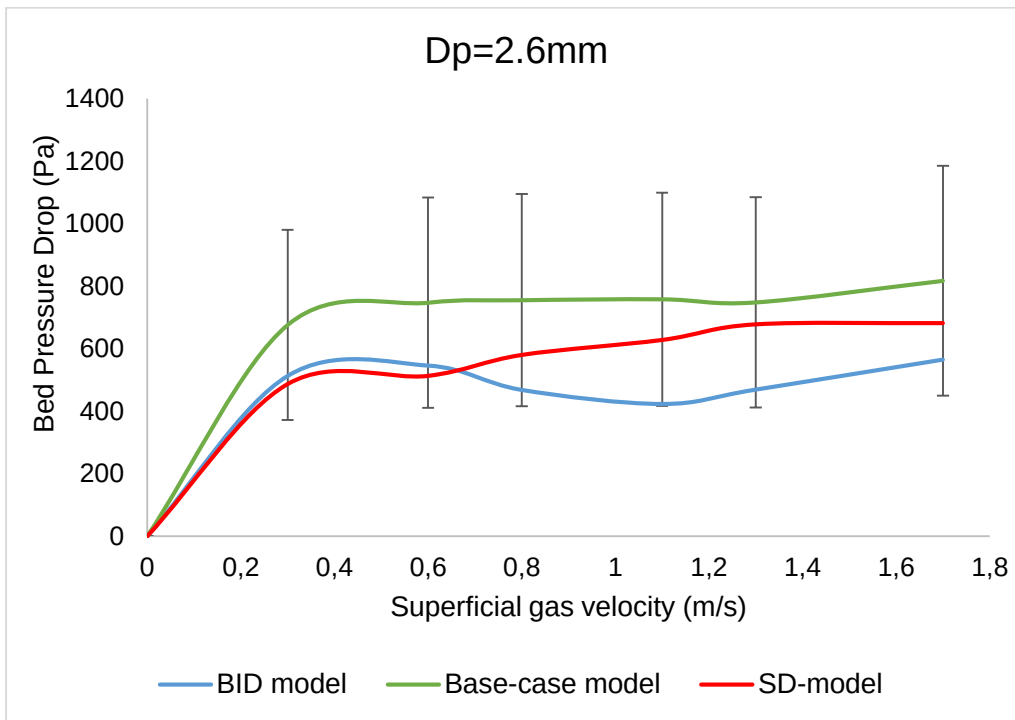


Figure B-3.6: bed pressure drop against superficial gas velocity of dp=2.6mm particles modelled with Bend Inlet Plate A, Base case and straight duct fluidised bed

Comparing bed pressure drop against superficial gas velocity curves for plate A, plate B, Plate C and Plate D in a bend inlet duct fluidised bed using particles with different particles sizes

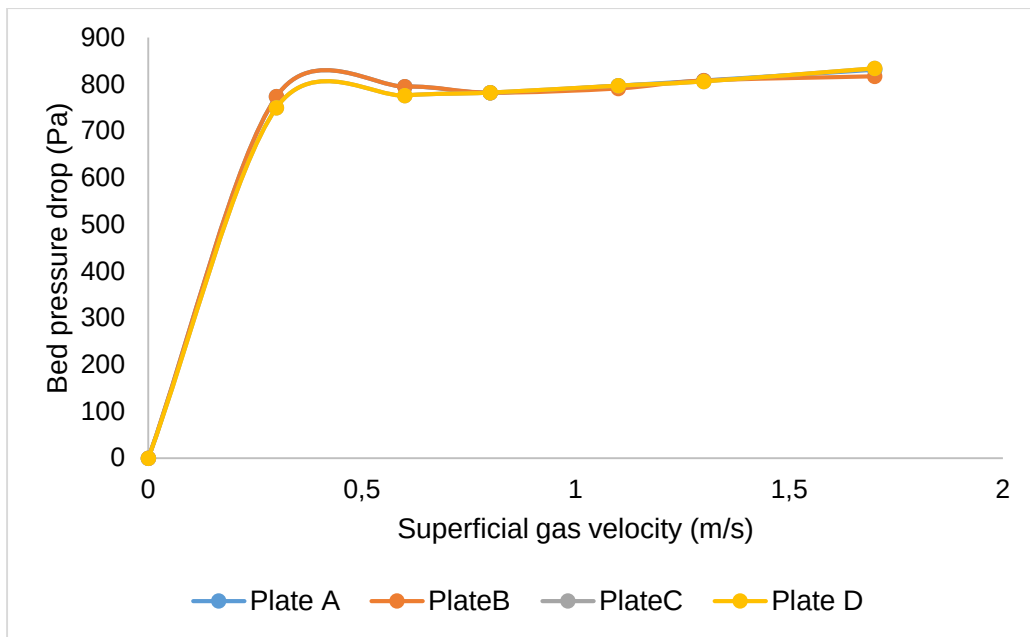


Figure B-3.7: Bed pressure drop against superficial gas velocity of bend inlet distributor plate using $d_p=1.3\text{mm}$ particles

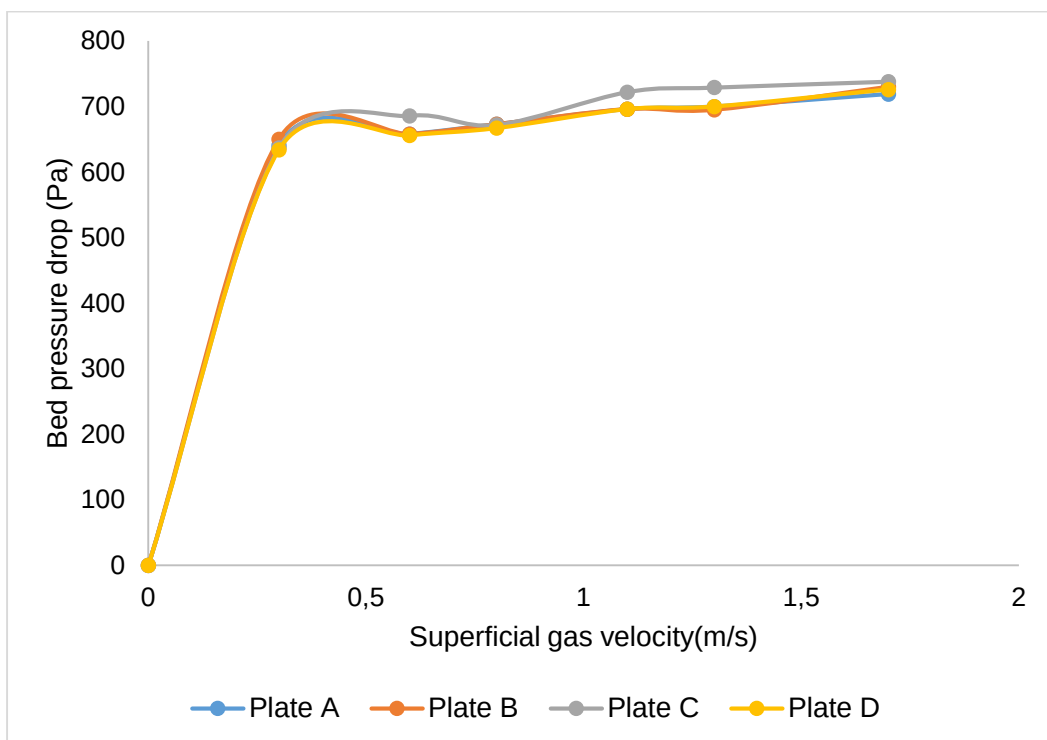


Figure B-3.8: Bed pressure drop against superficial gas velocity of bend inlet distributor plate using $d_p=1.6\text{mm}$ particles

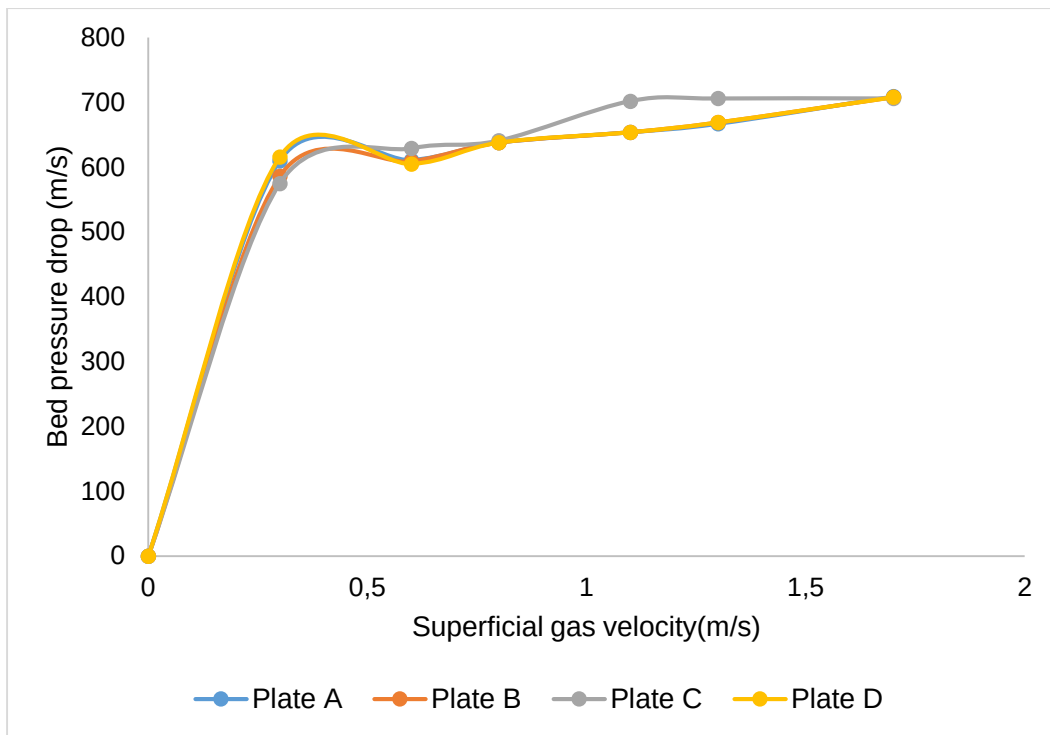


Figure B-3.9: Bed pressure drop against superficial gas velocity of bend inlet distributor plate using $d_p=1.7\text{mm}$ particles

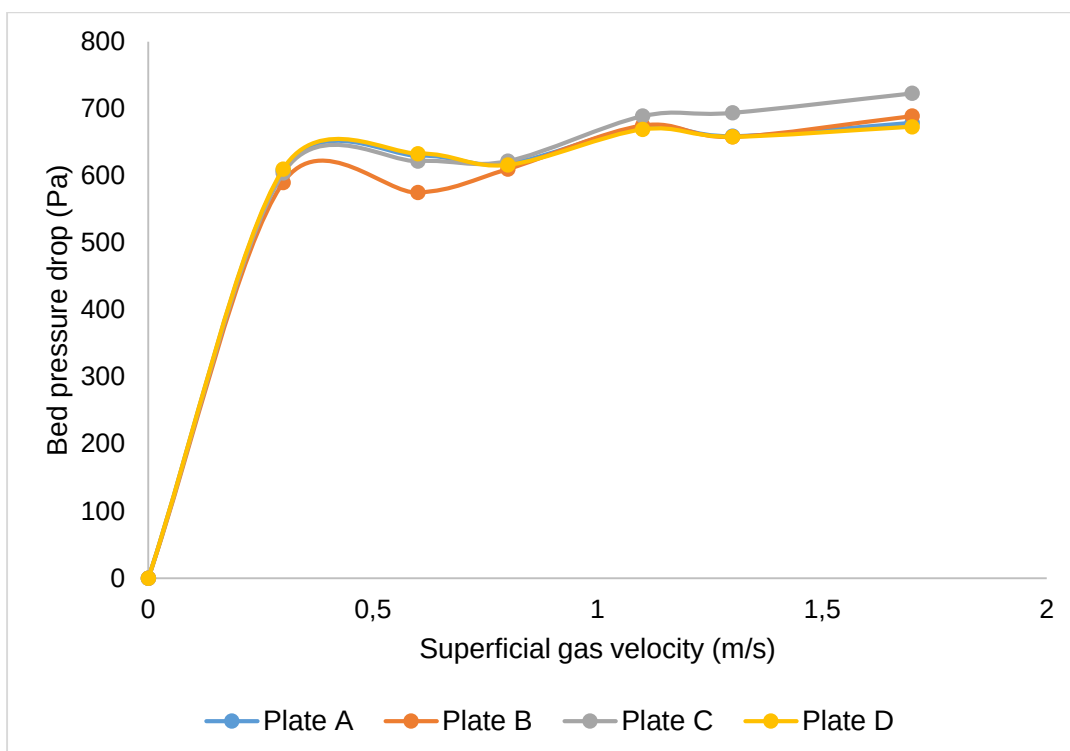


Figure B-3.10: Bed pressure drop against superficial gas velocity of bend inlet distributor plate using $d_p=1.85\text{mm}$ particles

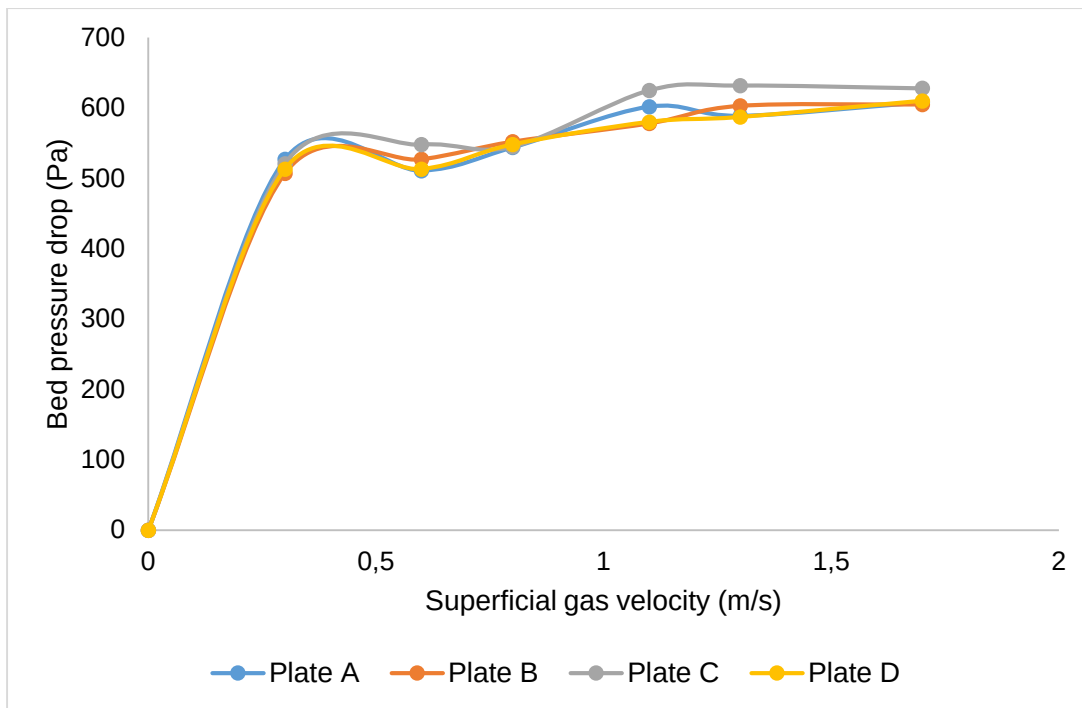


Figure B-3.11: Bed pressure drop against superficial gas velocity of bend inlet distributor plate using $d_p=2.1\text{mm}$ particles

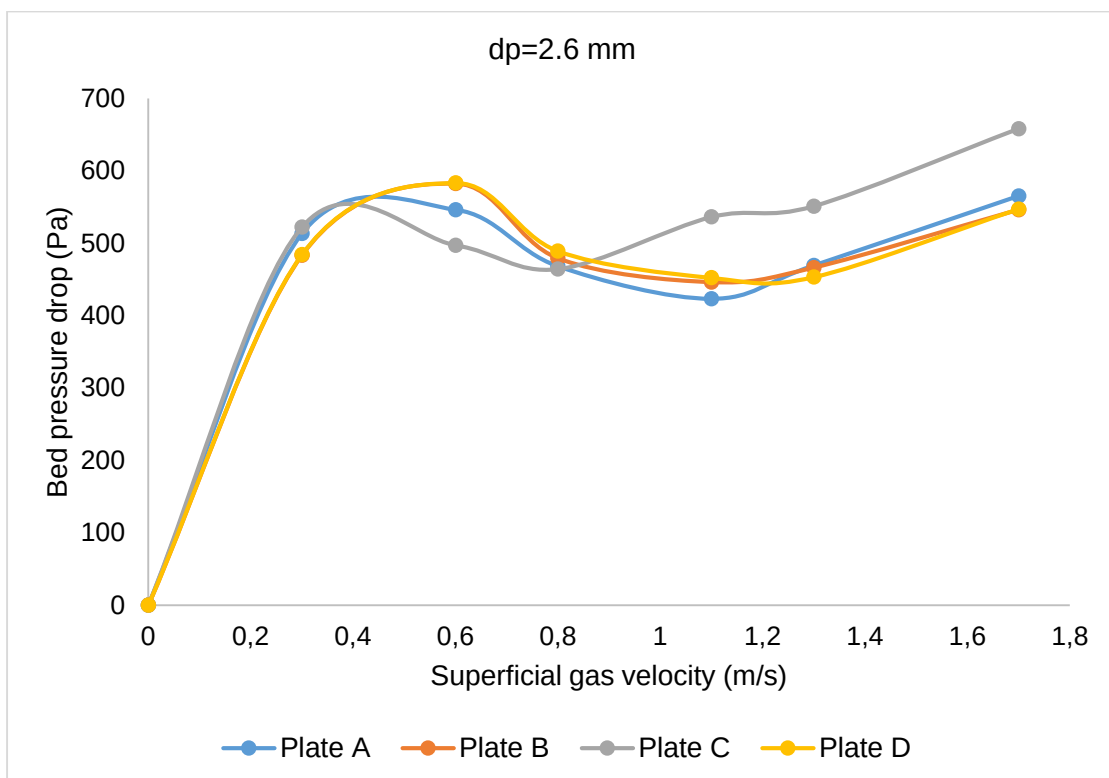


Figure B-3.12: Bed pressure drop against superficial gas velocity of bend inlet distributor plate using $d_p=2.6\text{mm}$ particles

Appendix B-4: Bed density during fluidisation in a laboratory scale setup

The model bed density of Plate A measured at the centre of the duct.

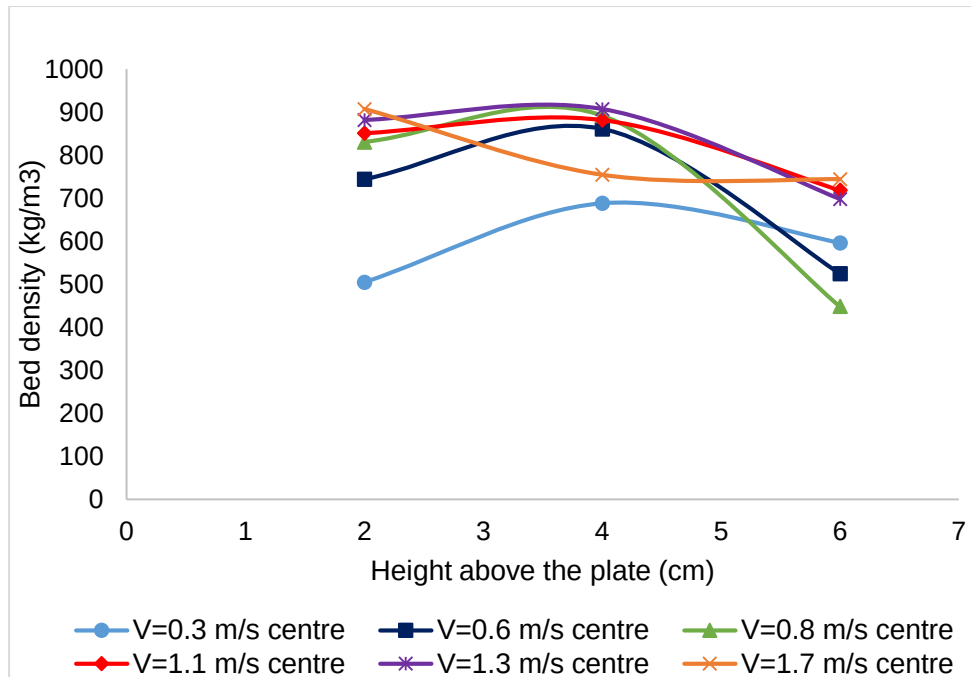


Figure B-4.1: Bed density against height above Plate A measured at the centre of the duct

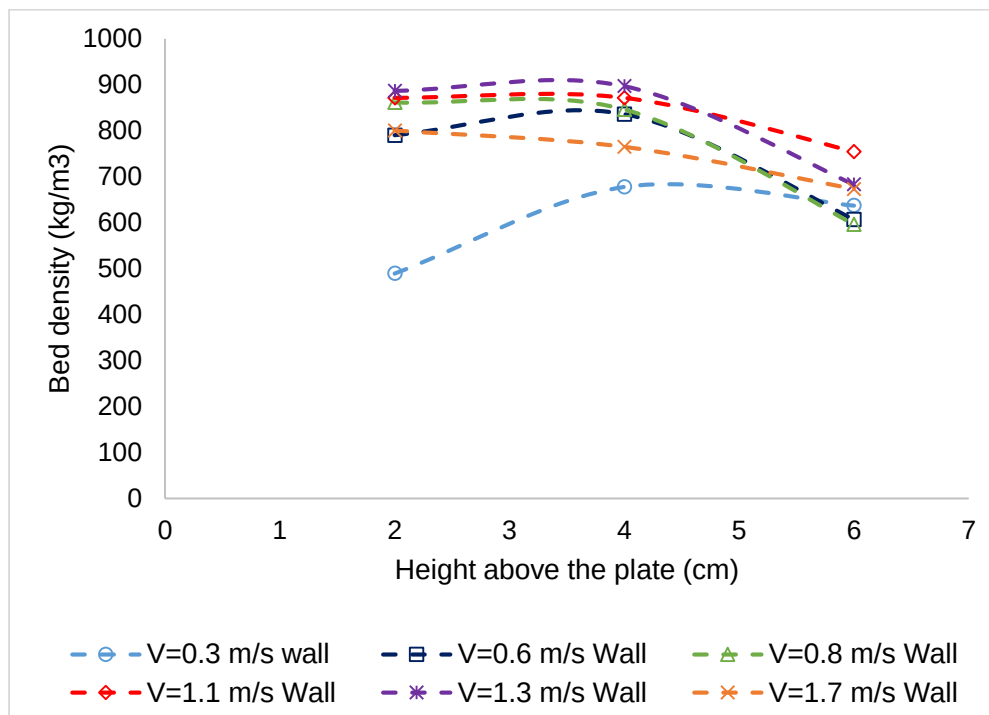


Figure B-4.2: Bed density against height above Plate A measured at the left wall of the duct

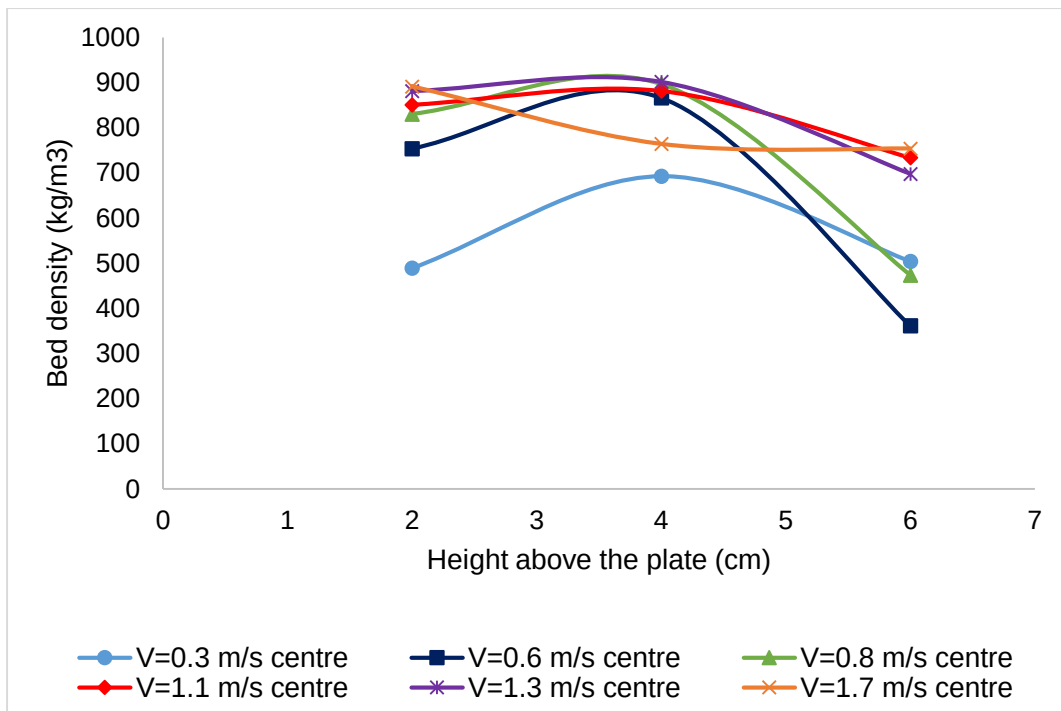


Figure B-4.3: Bed density against height above Plate B measured at the centre of the duct.

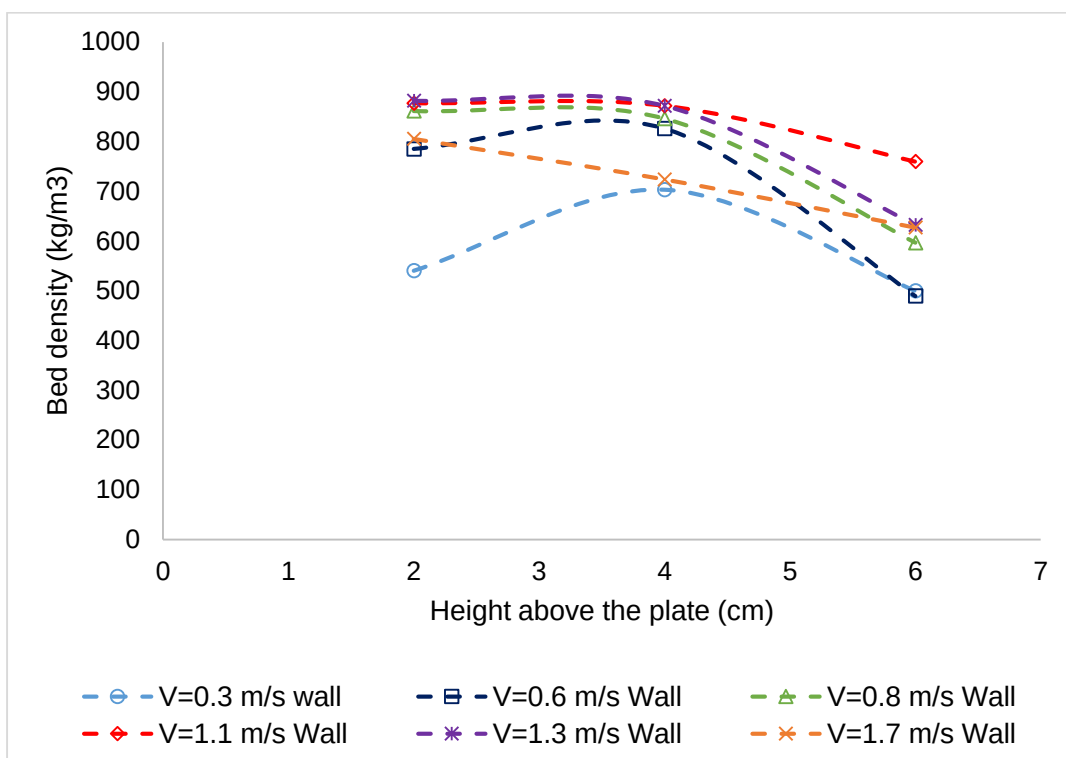


Figure B-4.4: Bed density against height above Plate B measured at the left wall of the duct

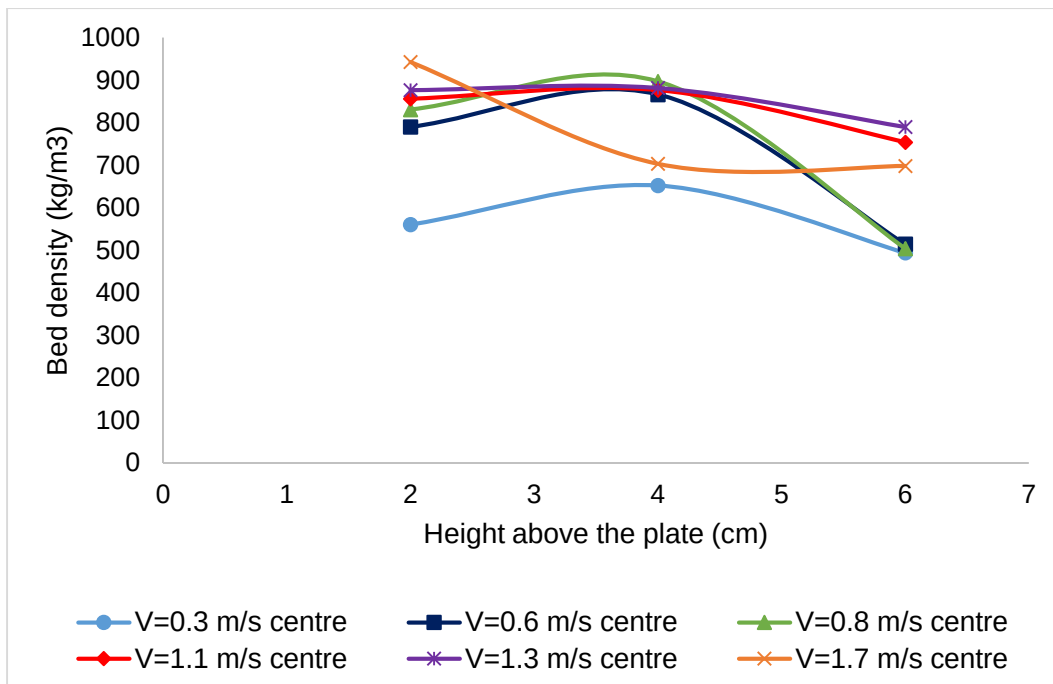


Figure B-4.5: Bed density against height above Plate C measured at the centre of the duct.

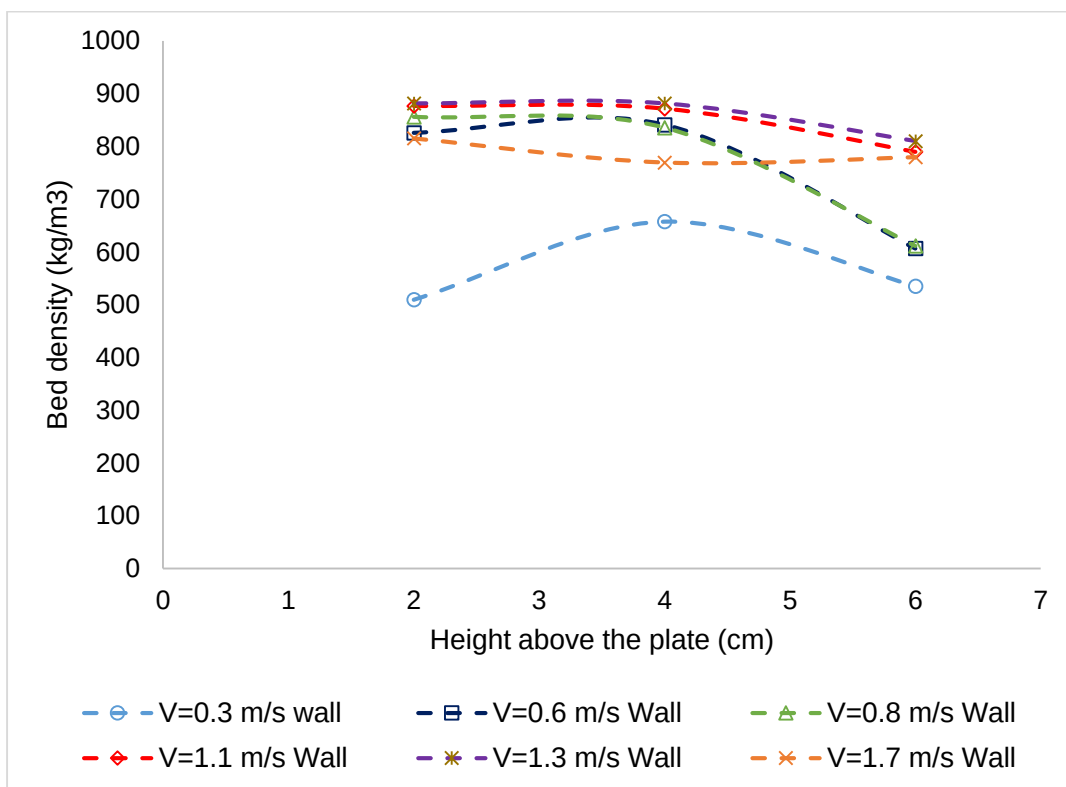


Figure B-4.6: Bed density against height above Plate C measured at the left wall of the duct.

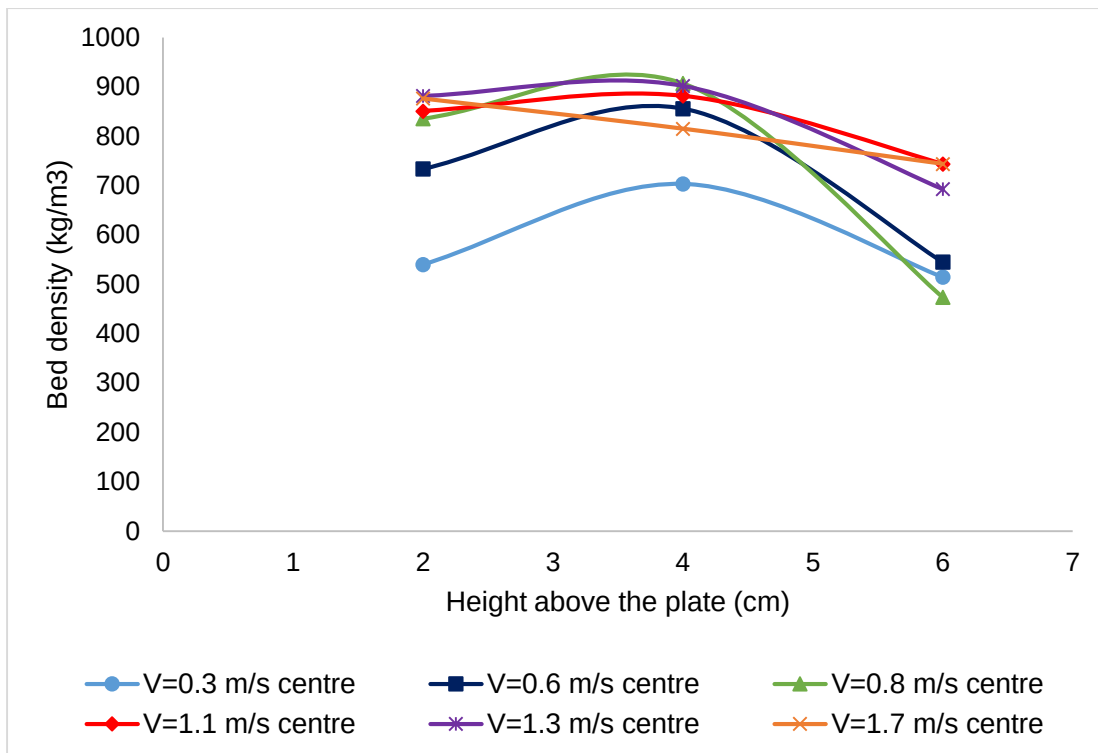


Figure B-4.7: Bed density against height above Plate D measured at the centre of the duct.

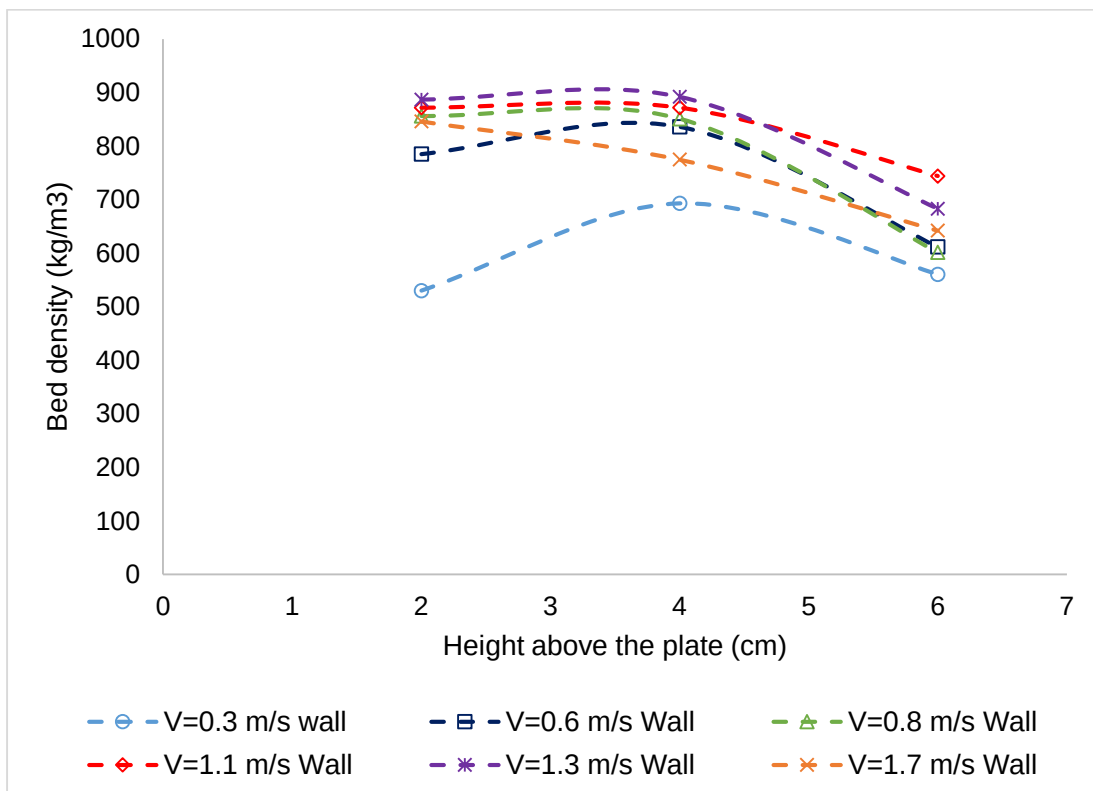


Figure B-4.8: Bed density against height above Plate D measured at the left wall of the duct.

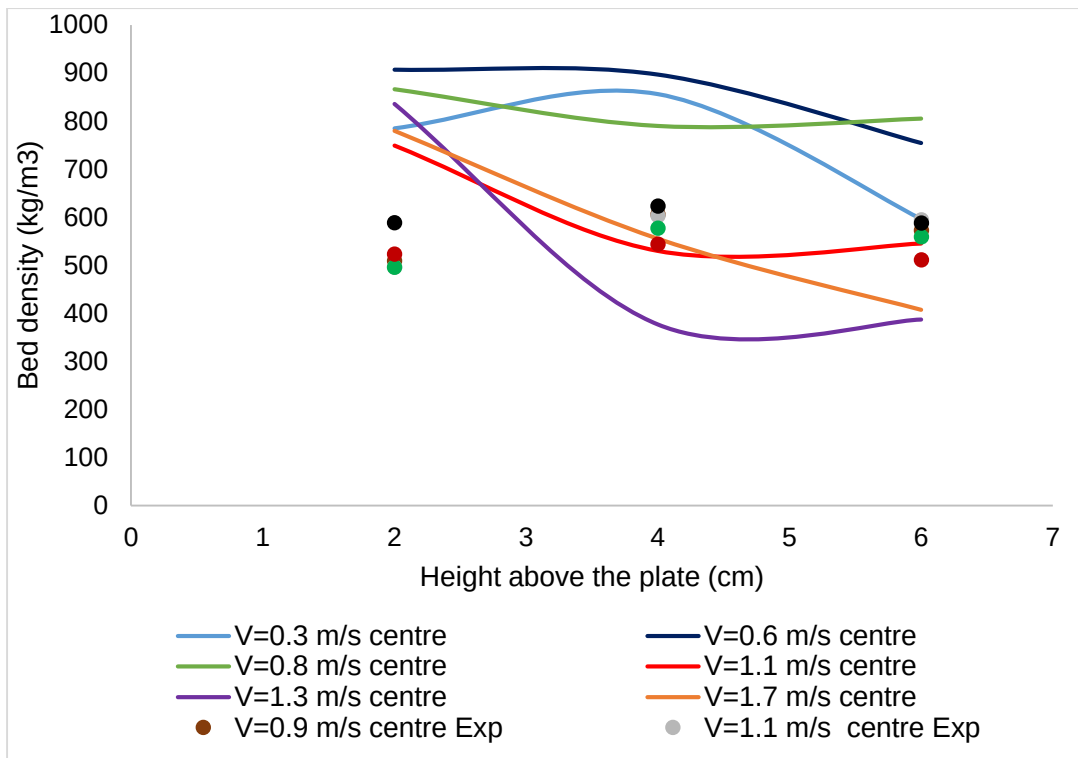


Figure B-4.9: Comparison of experiment to model Bed density against height above the plate using a block velocity profile plate.

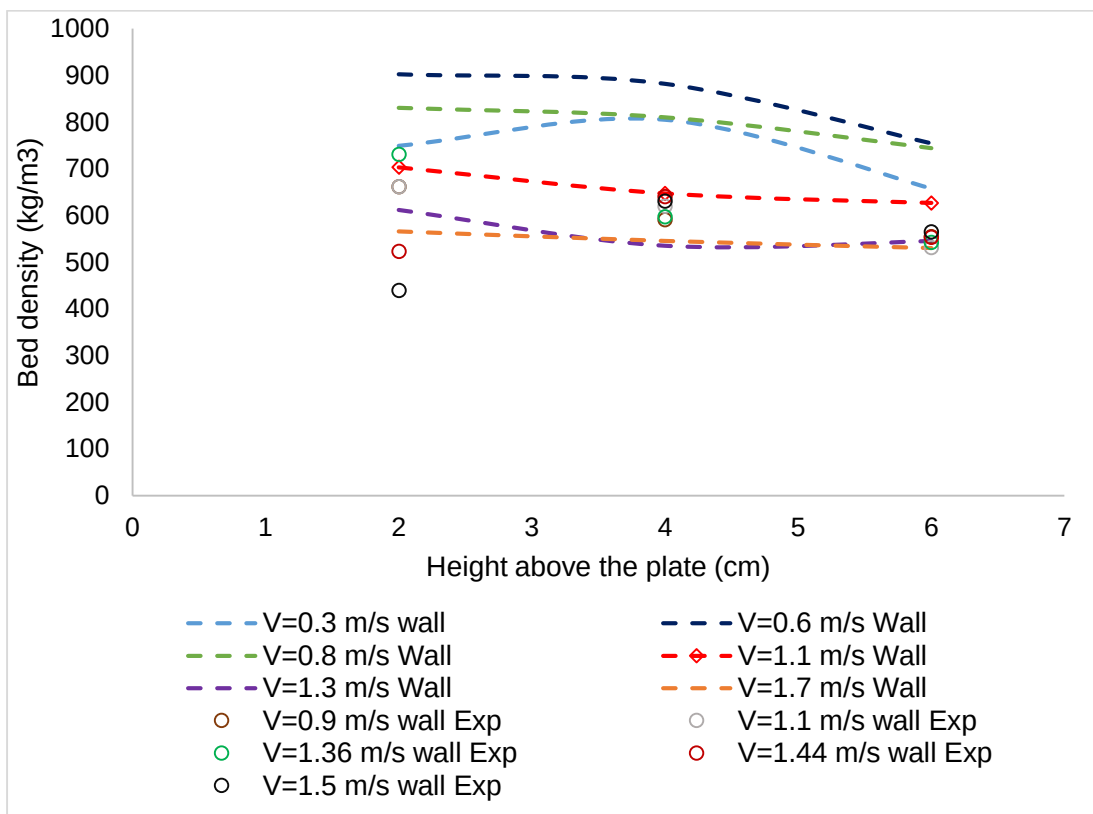


Figure B-4.10: Comparison of experiment to model bed density against height above the plate using a block velocity profile plate.

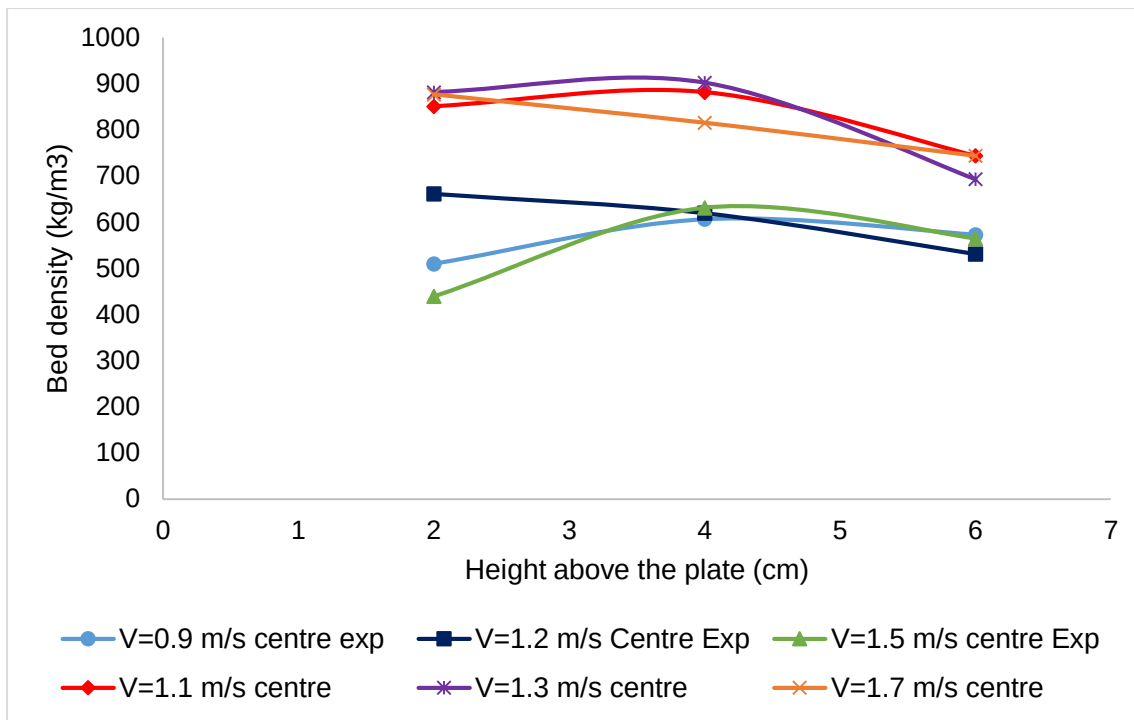


Figure B-4.11: Experiment compared to model bed density against height above the Plate D measured at the centre of the duct.

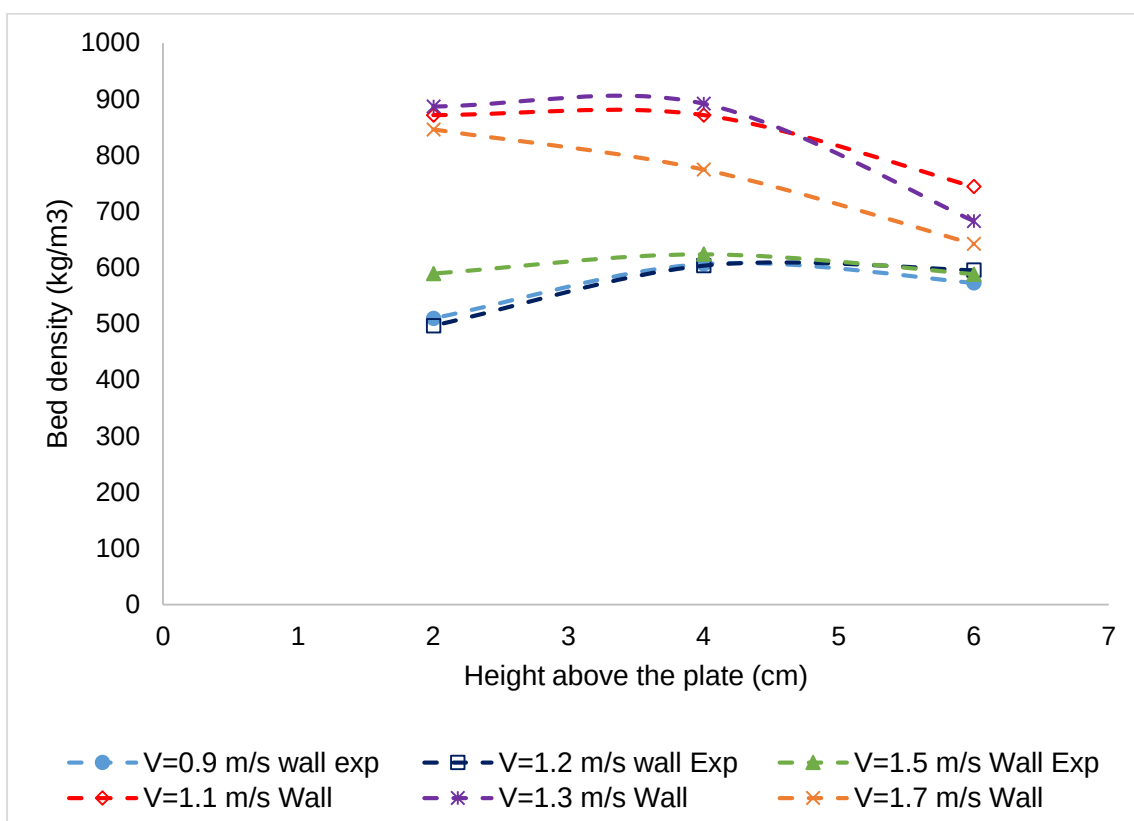


Figure B-4.12: Experiment compared to model bed density against height above the Plate D measured at the left wall of the duct.

Appendix C: Error calculations

Error percentage calculated between the plate pressure drop measured in experiments and model

Table C-1: Error percentage calculated between model and experiment plate pressure drop in laboratory scale setup.

V(m/s)	Plate A	Plate B	V (m/s)	Place C	Plate D
0.38	1283	19	0.34	45	14
0.60	29	18	0.59	20	18
0.79	9	19	0.78	10	29
0.95	4	17	0.93	6	30
1.10	3	15	1.05	4	28
1.23	3	13	1.16	4	25
1.35	3	12	1.27	4	21
1.45	3	11	1.36	5	17
1.56	4	10	1.45	6	13
1.66	5	10	1.52	8	8

Table C-2: Error percentage calculated between experiment and model bed density of Plate A at different superficial gas velocities

PRESSURE DIFFERENCE		V=5.62 M/S				
		0.17	0.33	0.49	0.66	0.83
	0.05	22%	49%	99%	134%	83%
	0.15	12%	38%	53%	111%	45%
	0.25	11%	32%	41%	79%	35%
	0.35	11%	9%	19%	16%	13%
PRESSURE DIFFERENCE		V=7.80 m/s				
		0.17	0.33	0.49	0.66	0.83
	0.05	22%	49%	99%	134%	83%
	0.15	20%	38%	53%	111%	45%
	0.25	19%	32%	41%	79%	44%
	0.35	11%	9%	19%	16%	13%
PRESSURE DIFFERENCE		V=11.38 m/s				
		0.17	0.33	0.49	0.66	0.83
	0.05	36%	57%	99%	147%	104%
	0.15	27%	38%	44%	111%	64%
	0.25	4%	16%	24%	57%	26%
	0.35	11%	21%	3%	7%	4%
PRESSURE DIFFERENCE		V=12.68 m/s				
		0.17	0.33	0.49	0.66	0.83
	0.05	42%	57%	110%	147%	124%
	0.15	27%	38%	53%	98%	64%
	0.25	4%	1%	6%	45%	17%
	0.35	11%	21%	3%	1%	4%

Table C-3: Error percentage calculated between experiment and model bed density in Plate B at different superficial gas velocities.

PRESSURE DIFFERENCE	V=5.62 M/S				
	0.17	0.33	0.49	0.66	0.83
0.05	128%	122%	134%	89%	85%
0.15	69%	71%	81%	32%	47%
0.25	81%	61%	117%	30%	53%
0.35	39%	5%	48%	10%	4%

PRESSURE DIFFERENCE	V=7.80 m/s				
	0.17	0.33	0.49	0.66	0.83
0.05	116%	122%	134%	89%	76%
0.15	80%	82%	93%	40%	57%
0.25	81%	61%	117%	30%	53%
0.35	39%	5%	48%	10%	4%

PRESSURE DIFFERENCE	V=9.97 m/s				
	0.17	0.33	0.49	0.66	0.83
0.05	116%	122%	134%	99%	76%
0.15	80%	82%	93%	49%	57%
0.25	93%	61%	131%	39%	53%
0.35	53%	5%	48%	10%	6%

PRESSURE DIFFERENCE	V=11.38 m/s				
	0.17	0.33	0.49	0.66	0.83
0.05	116%	122%	121%	89%	76%
0.15	80%	82%	93%	40%	57%
0.25	93%	61%	117%	39%	53%
0.35	53%	16%	63%	1%	4%

Table 2: Error percentage calculated between model and experiment bed density in Plate C at different superficial gas velocities.

PRESSURE DIFFERENCE	V=5.62 M/S				
	0.17	0.33	0.49	0.66	0.83
0.05	31%	46%	104%	52%	100%
0.15	34%	28%	92%	37%	59%
0.25	35%	31%	84%	51%	55%
0.35	12%	2%	44%	13%	2%

PRESSURE DIFFERENCE	V=7.80 m/s				
	0.17	0.33	0.49	0.66	0.83
0.05	38%	46%	104%	52%	123%
0.15	34%	20%	105%	46%	59%
0.25	44%	31%	97%	61%	55%
0.35	12%	2%	30%	13%	2%

PRESSURE DIFFERENCE	V=9.97 m/s				
	0.17	0.33	0.49	0.66	0.83
0.05	38%	54%	116%	61%	123%
0.15	25%	28%	105%	46%	59%
0.25	44%	31%	97%	61%	55%
0.35	20%	2%	44%	13%	2%

PRESSURE DIFFERENCE	V=11.38 m/s				
	0.17	0.33	0.49	0.66	0.83
0.05	38%	54%	104%	52%	123%
0.15	34%	28%	105%	46%	59%
0.25	44%	40%	97%	61%	55%
0.35	20%	12%	44%	13%	2%

PRESSURE DIFFERENCE	V=12.68 m/s				
	0.17	0.33	0.49	0.66	0.83
0.05	38%	46%	104%	52%	123%
0.15	34%	28%	105%	55%	59%
0.25	35%	40%	84%	51%	55%
0.35	3%	2%	58%	24%	2%