

QUASI-THREE DIMENSIONAL EXPERIMENTS ON LIQUID-SOLID FLUIDIZED BED
OF THREE DIFFERENT PARTICLES IN TWO DIFFERENT DISTRIBUTORS

Chukwuyem Charles Obuseh, B.S.

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APPROVED:

Zhi-Gang Feng, Major Professor
Sandra Boetcher, Committee Member
Tae-Youl Choi, Committee Member
Miguel F. Acevedo, Interim Chair of the
Department of Mechanical and Energy
Engineering
Costas Tsatsoulis, Dean of College of
Engineering
Michael Monticino, Dean of the Robert B.
Toulouse School of Graduate Studies

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This thesis is an experimental study of the fluidization of binary mixture in particulate flows. A fluidized bed with two distributors was built with water being used as carrying fluid. Three types of solid particles of nylon, glass and aluminum of the same size and different densities are used in the experiments. The wall effect on a single particle fluidization, the fluidization of binary mixture of large density difference (nylon and aluminum of density ratio of 0.42), and the fluidization of binary mixture of close density (glass and aluminum with density ratio of 0.91) were investigated. Also, the effect of distributors on mono-disperse and bi-disperse particle fluidization was investigated. Results show that the presence of narrow walls reduces the minimum fluidization velocity for a single particle by as much as nearly 40%. Also, in the case of binary mixture of close density particles, uniform mixing was easily achieved and no segregation was observed, but in the case of large density difference particles, there exists significant segregation and separation. At high velocity, the uniform distributor behaves like a transport bed. To achieve a full bed in the single jet, it requires 1.5 times velocity of the uniform distributor. This behavior determines their application in the industries.

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CHAPTER 1

INTRODUCTION AND LITERATURE REVIEW

1.1. Fluidized-Bed Systems

In early times, the term “teetering” was applied to sorting and sizing of particles based on differences in densities and size. The name was changed from teetering to liquid-fluidized bed when particles at high velocities moved upward in a contacting medium of gas column was described with the word “fluidization”[1].

In 1926, Fritz Winkler built the first commercial large scale fluidized bed reactor. This has given birth to other fluidized bed processes for different operations. It was called Winkler coal gasifier and was used in producing carbon monoxide from coal.

In the Second World War, research in catalytic cracking of heavy oils was desperately needed as petrol demand was on the increase. Exxon engineers based on the idea which was proposed and confirmed experimentally by a group of researchers at the Massachusetts Institute of Technology (MIT) that a completely pneumatic circuit consisting of fluidized beds and transport lines could be operated stably for a satisfactory catalytic cracking process[2], built a large-scale pilot plant of an up-flow cracking unit [3]. The fluid catalytic cracking (FCC) process was then in the pipeline.

Just like liquefaction denotes the act of making liquid, fluidization confers fluid-like properties of solid particles. It is a process of transforming solid particles by suspension using gas or liquid into fluid-like state. A simple example is seen in a vacuum cleaner in Figure 1.1. The airflow from it is used to suspend dust particles into another chamber and later disengaged from the fluid.

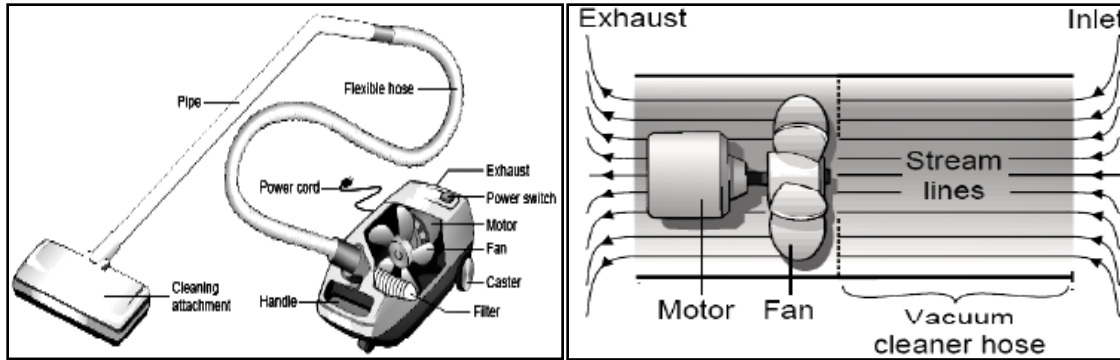


Figure 1. 1: Vacuum cleaners [4].

Upward flow of fluid through a bed of particles above a certain fluid velocity causes the particles to be suspended. These suspended particles have many of the properties of a fluid it seeks a bed height and assumes the shape of the container vessel.

At low rate of flow, there is no much drag from the fluid to overcome gravity, fixed or packed bed results in this case. These particles move apart with a few vibrations as the flow rate rises leading to an expanded bed. The expansion of these solid particles is directly related to the superficial velocity of the liquid and this relationship is useful in fundamental understanding of the behavior of fluidization and subsequent application.

Further increase in velocity gives rise to a point where the frictional force between particles and fluid equals the weight of the particles, all the particles become suspended by the upward flow of fluid. The pressure drop through any section of the bed equal to the weight of the fluid and the particles, on the other hand, fluid drag plus buoyancy overcomes gravity force. This is called minimum fluidization and also referred to as incipient fluidization. At this point, the bed particles achieve the appearance and many properties of a true fluid, the pressure drop increases and become constant

but the bed height is constant until this point. This is very important in deciding the fluid flow required to expand the bed of particles and its behavior under a number of flow conditions.

Fluidization is of great importance in industries. The processes of fluid– solid fluidization are widely utilized in the fields of hydrometallurgy, food Technology, biochemical processing and water treatment. It has also been applied in dealing with biomass because of their advantages of high heat transfer, controllable and uniform temperatures, ability to handle wide variety of particle properties and good fluid-solid contacting medium [5].

Disadvantages include difficulties in prediction of the behavior of a fluidized bed reactor due to the complexity and ambiguity of some fundamental properties (size, density and shape of the particle) that play an important role in prediction and calculation of hydrodynamics of fluidized beds and possibility of particle collision leading to container wall corroding.

1.2. Applications

Two distinct category of fluidization mechanism research are the engineering view which involves applied research in laboratory units and tend to imitate process plants and also mathematician view which involves theoretical analysis [6].

Fluidized bed finds its application in petroleum refining processes, for example in catalytic cracking reactors, particles are usually fluidized by vaporization. The purpose of this catalyst is to break down large crude petroleum molecules into smaller constituents. Also gas fluidized bed finds its use in chemical reactors and also as combustors in power generations by raising steam. Moreover liquid fluidized beds are majorly employed in water treatment, mineral processing and fermenting technology.

In fluid particle systems, a application is on gas-solid, liquid solid and gas-liquid solid systems, and in reactor scheme and operations it is grouped as stationary fluidized bed (SFB), circulating fluidized bed and spouted bed reactors. The spouted bed involves formation of high velocity upward stream of fluid and particle and downward movement of particles. This is

achieved by injection of fluid through the bottom of the bed to a central nozzle into the bottom of a cylinder vessel. Application is found in pneumatic transportation, drying of particles and coating.

SFB is also called dense-phase fluidized bed since the bed has a dense phase of particulate materials, particles stay in the bed. In a gas-solid system, flow regimes are smooth or particulate fluidized bed, bubbling fluidized bed, slugging fluidized bed and turbulent fluidized bed. In the liquid-solid systems, only one regime is present and it is called the particulate or smooth fluidized bed regime.

Excess fluidizing gas in gas-solid fluidized bed forms bubbles and some other problems, but in liquid solid fluidized bed, bed expansion resulting from excess liquid is involved. This favors uniform contact between the particles in the entire bed. Increasing superficial velocity can create better liquid-solid control in the bed [7].

Gas-liquid-solid fluidized bed (GLSFB) results from the injection of gas into liquid-solid fluidized bed (LSFB). They are usually grouped into bubble of gas phase and dense phase containing only liquid and solid. Regimes are dispersed bubble, coalesced bubble and slug regimes. The differences in gas-liquid-solid fluidized bed and gas-solid fluidized bed is that gas is not present in the dense phase in the gas-solid fluidized bed. Applications of GLSFB involve upgrading heavy stocks like hydro cracking and artificial lifting in the oil and gas industries. Other applications of fluidized bed include the following.

1.2.1. Polymerization Reactor

The gas-phase polymerization reactor is a system where smaller particles (catalyst) are introduced in the bed to react with the monomer gas to produce polymers with a higher size distribution. Segregation plays an important role when the fully grown polymers will settle in

the bottom of the bed and collected from the reactor and particles with smaller size distribution and catalyst continuously migrate to the top for further reaction with the monomers [8].

1.2.1. Filter Backwashing

Before drinking water goes into a clear well, its cleanliness decides if the filter requires a filter backwashing. In drinking water systems, filters are used to collect, catch, or gather particles from an incoming flow. Backwashing a drinking water system filter means reversing and increasing the velocity to flush out clogged particles. Backwashing is not only vital to the life of a filter, it is fundamental to the quality of water coming out of the filter hence sooner or later; all filters need to be backwashed or replaced.

1.2.2. Particle Classification

This phenomenon has been made possible by segregation characteristics of liquid fluidized bed by size, density and shape hence, liquid fluidization has been mainly used for mineral separation and density is a dominating factor compared to size in particle sorting. Particles are grouped in order of particle behavior and size ranging from the smallest to the largest [9].

- Group A materials have small mean particle size ($d_p < 30 \mu m$) and/or low particle density $< 1.4 g/cm^3$.
- Group B falls under the size range of $150 \mu m$ to $500 \mu m$ and density from 1.4 to $4 g/cm^3$
- Group C have sizes usually less than $30 \mu m$.
- Group D has sizes in the range of $500 \mu m$ to $4000 \mu m$ and usually referred to as spoutables.

1.2.3. Fluidized Bed Electrodes

This consists of a bed of electrically conducting particles in a reservoir of electrolyte connected to a current feeder which produces a D C current. Without so much resistance, a uniform electrical contact is provided between the feeder and the particle and also between the particles themselves. This uniform electrical contact has helped in boosting the current density due to the particle contact electrodes.

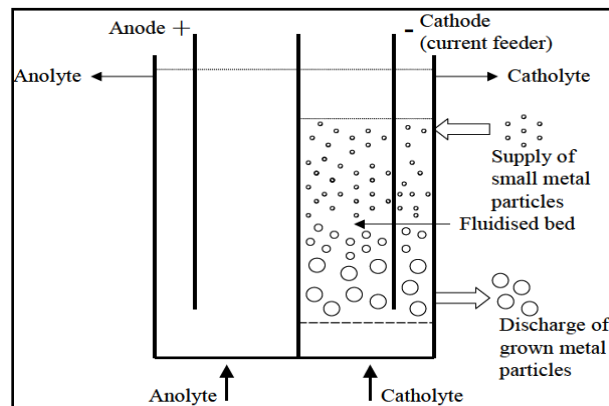


Figure 1. 2: Fluidized bed electrodes [10].

1.2.4. Biofluidization

This is a process of cultivation of microorganism by methods of fluidization. Just above the distributor is a rotating agitator which helps to prevent defluidization in the lower portion of the bed rotating separator is also included at the top to return the removed particles back to the bed and an electrode which acts as a sensor to detect the water content of the particles.

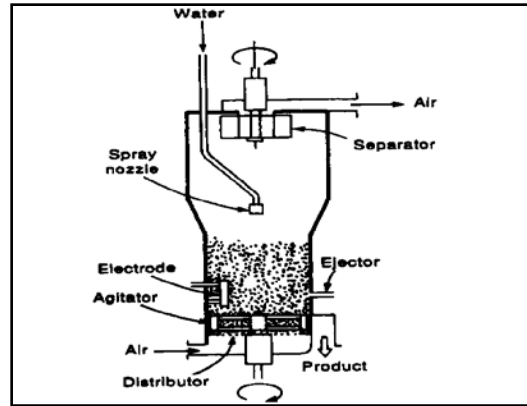


Figure 1. 3: Fluidized bed cultivators [2].

1.3. Distributors or Grid

The success or failure of a fluidized bed is determined by the distributor performance. Difficulties in fluidized beds are as a result of careless design of distributors. To improve the design success of a distributor for fluidization processes, some parameters have to be considered which include the pressure drop ratio, orifice size, geometry, spacing, and dead zones, distributor strength to resist deformation and the ability to operate for long periods without blocking. They usually come in different forms or designs and are grouped depending on direction of entry. They could be downward, upward, laterally or horizontally depending on the use and cost. Selection criteria are based on process application.

Fluidized bed distributors can take variety of shapes and sizes. But a major design criterion is jet flow. This is important because bed erosion is reduced. Also particle erosion can be increased or decreased at the distributor depending on application. To achieve some particular jetting region, size of the orifice and jet velocity of fluid is needed. Jet flow has made it possible to achieve this goal [11].

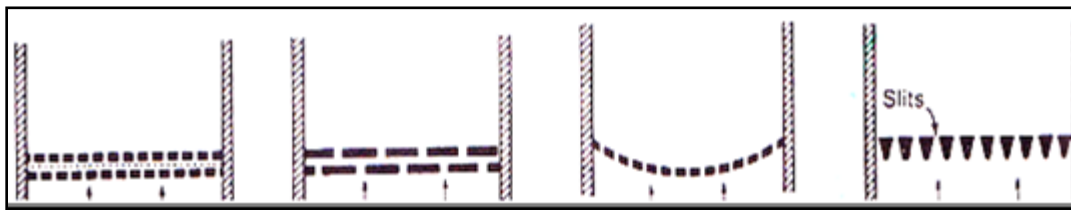
In their study of jet flow, Hong et al. concluded that the jet velocity is increased when the cross sectional area of a jet is reduced and the jet momentum per unit cross sectional area is

then increased [12]. The structure of the distributor has been conducted by many researchers over the past few years, with the objective of improving and optimizing the fluidization quality [13].

Kiarash Vakhsouri investigated the influence of distributors and plenum chamber volume on gas bubble behavior and size and concluded that this was due to pressure drop which is a function of open area ratio of the distributor and plenum [14].

1.3.1 Types of distributors

1.3.1.1 Perforated Distributors. / Multi-orifice



Sandwiching plate staggered plates dished plate grate bars

Figure 1. 4: Perforated Distributor Types [2]

Perforated plate distributors are simple and inexpensive and account for their use in industries. It is also easy to scale up or down. Choice is based on the ease of fabrication, cost, ease of modification, cleaning and shape. Set backs in this kind of distributors are high pressure drops, and sealing of surfaces to the bed.

1.3.1.2. Bubble Cap

This takes the shape of a laterally directed flow and choice of design is based on lower pressure drop. They can be stiffened by attaching a cap, it has minimum weeping and a better turndown ratio and its setbacks are cost, cleaning difficulty, sealing of surfaces is needed and modification is difficult.

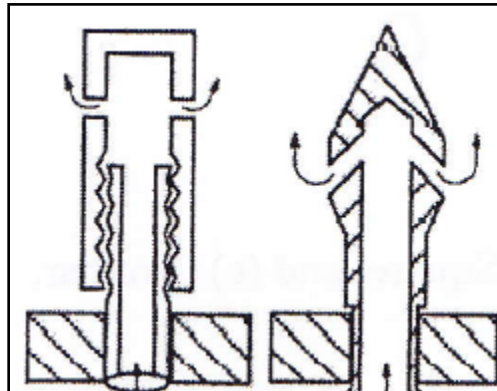


Figure 1. 5: Bubble cap distributor [15].

1.3.1.3. Sparger

This is a flow that has a lateral and downward pattern. Choice of this design includes low pressure drop, it can hold thermal expansion without failure and ease of flow of particle from top to bottom of the distributor. Major setbacks are defluidization under the distributor. Figure 1.6. Shows a downward pointing nozzle used to prevent particles from clogging the sparger during inflow of fluid and the Sohio acrylonitrile process which is a pipe grid used to feed fluid into a second carrier from the bottom of the grid.

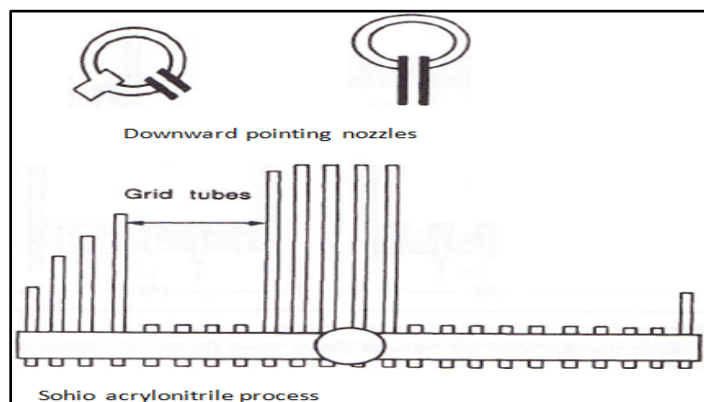


Figure 1. 6: Sparger Distributor Types [2].

1.3.1.4. Touyeres and Caps

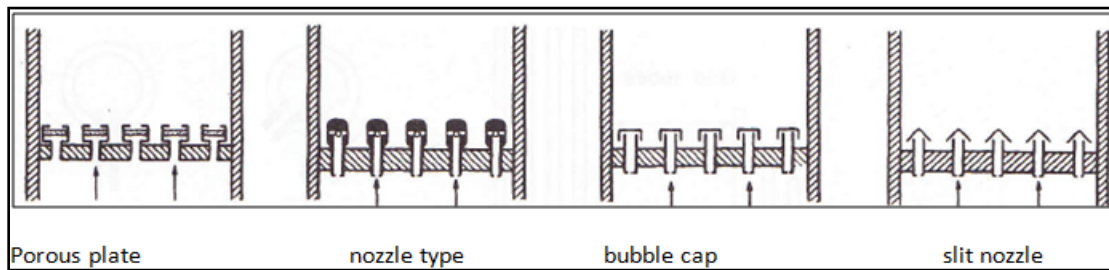


Figure 1. 7: Touyeres and Caps Distributor Types [2].

This is used when operations involve high temperature and a highly reactive environment and its drawback is that particles may settle and stick on the distributor plate. They are usually designed with a high-resistance orifice at the inlet section and around the caps, orifice is designed to create a uniform fluidization by a build-up of pressure drop. Other designs are pipe grid and spargers which are designed with a heat exchanger tubes to improve fluid-solid contacting by preventing gross circulation of solids and gulf streaming.

1.4. Mono-dispersed Fluidization

Mono-disperse fluidization is that which involves spheres that have approximately normal distribution with a considerable spread in diameters. Uniform fluidization is characterized by a homogenous random motion of particles by a constant average velocity of fluid and momentum throughout the bed [16]. This is usually achieved in the presence of gulf streaming (i.e. separate regions of upward and downward solids motion) [17]. For mono-disperse spheres, a homogenous expanded bed whose height changes with flow rate is usually achieved. Going by Courderc assumption with a uniform distribution of particles in a liquid solid fluidization, homogenous bed with stable operation is achieved [18].

Particle pair interactions are fundamental mechanisms in all practical applications of particulate flows. Classification of local mechanism of sphere rearrangements in fluidization process is based on the type of fluid. For Newtonian liquids, this pairing of neighboring spheres can be described as drafting, kissing and tumbling and called drafting, kissing and chaining in viscoelastic liquids. In a Newtonian liquid, when one falling sphere enters the wake of another, it experiences reduced drag, attracts the downward leading sphere, and kisses it. Tumbling results in more stable horizontal arrays which may give rise to closed packed spheres separated by interstices of clear water [19]. In viscoelastic liquids, on the other hand, two spheres falling side-by-side will be pushed apart if their initial separation exceeds a critical value. However, if their initial separation is small enough, they will draft, kiss, turn and chain.

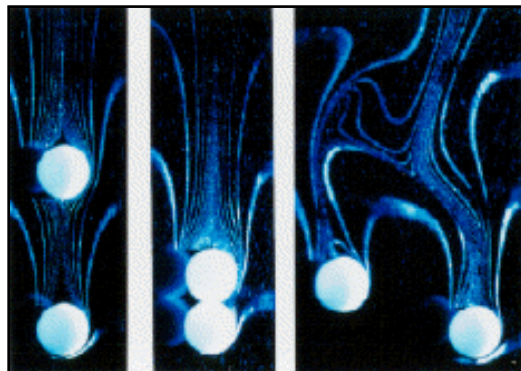


Figure 1. 8: Drafting, kissing and tumbling [19].

1.5. Poly-dispersed Fluidization

Stefan et al. defined poly-dispersed fluidization as a process in which a relatively compact bed of particles are fluidized by an upward bulk flow of fluid [20]. For a bi-dispersed fluidization, the necessary condition for existence of a fluidized bed is summarized below [21].

$$d_1 > d_2 \quad (1. 1)$$

$$d_1^2(\rho_1 - \rho_f) > d_2^2(\rho_2 - \rho_f) \quad (1.2)$$

$$\rho_1 < \rho_2 \quad (1.3)$$

In bi-disperse fluidization, density, size and superficial velocity are major mechanism for mixing and segregation.

Figure 1.9 explains what happens in a binary-solid fluidized bed. This shows that the bottom of the bed is made up of one or both kind of particle, but with the exception of figure 1.9C. Figure 1.9A is the case when the small particle has much higher density than that of large particle. Figure 1.9B is the case when some of the large size particles start to move up while some remain trapped among the small size particle. Figure 1.9C shows both type particles are able to mix uniformly. Figure 1.9D is the case when some of small size particle is easily being flow to the top of the column. Figure 1.9E is the case where clearly segregation is observed, and large particle size stay in the bottom bed and small size particles move to the top.

For little difference in size of particles in a binary fluidization, two separate regions are present which shows the separation from the bottom layer with larger particles and top layer with smaller particles fluidization [22]. Similar literatures on segregation can be found in Di Felice who demonstrated cases with larger particles separating at the top and smaller particles in the bottom under some conditions [23], Hoffmann et al., [24] and Wu and Baeyens [25] also experimented particles of equal density and different sizes and Nienow et al., also experimented on particles of the same size and different densities[26].

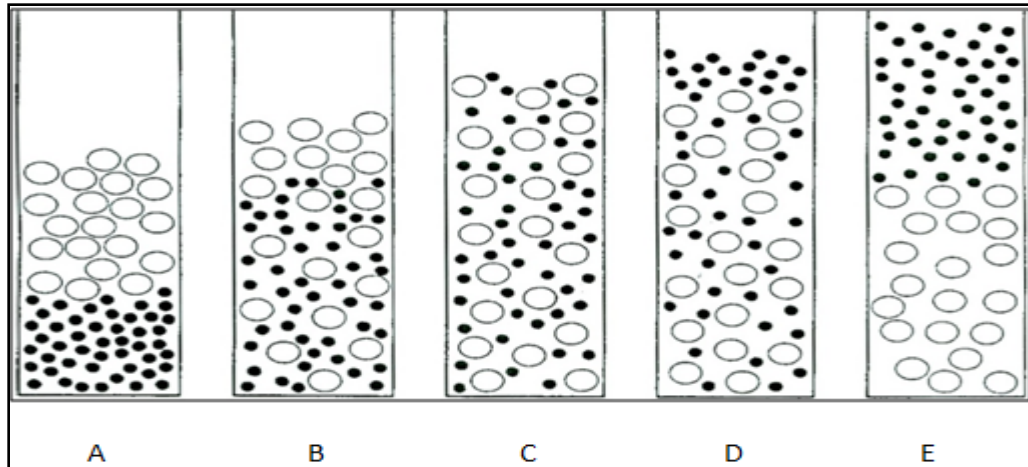


Figure 1. 9: Binary mixture fluidization pattern [25].

1.6. Problems of Fluidization

1.6.1. Bubble Coalescence

Bubbles result from high superficial gas velocities. It may be growth of bubbles in horizontal direction with neighboring bubbles, growth of bubbles in vertical direction with trailing bubbles and hydrostatic pressure decrease [27]. It is common in group B particles due to their mean particle size and particle size distribution [28].

In large industrial bed, the pattern of bubble growth is decided by bed depth and the superficial velocity [29]. Kan et al. has suggested addition of internal baffles of various designs as a means of braking down these bubbles and remedies the situation [30].

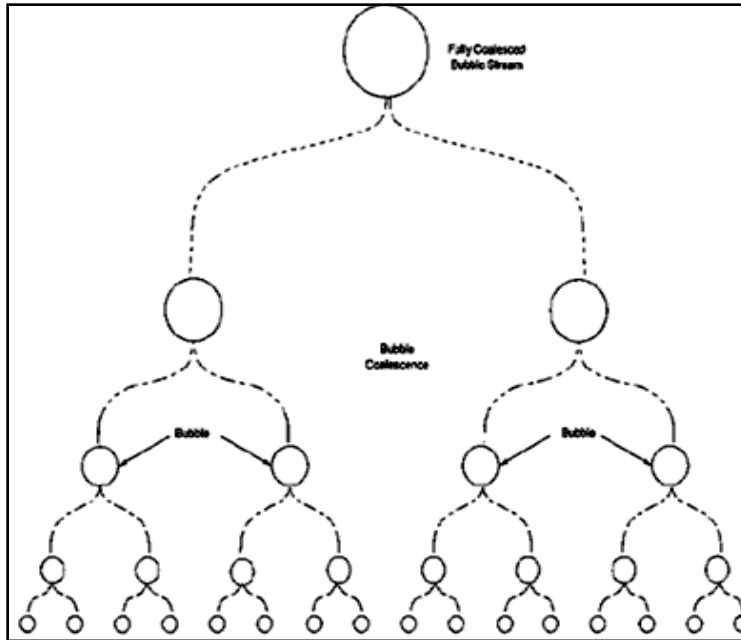


Figure 1.10: Dalton's bubble coalescence [31].

1.6.2. Agglomeration/ Sticky Particles

Bed agglomeration is the process of formation of larger particles when separate particles stick to one another. The formation of large agglomerates decreases the mixing of the bed and may result in defluidization, this is a sudden decrease of pressure drop over the bed to a lower level.

Defluidization in bubbling beds may be due to the agglomerates, it disturbs uniform mixing causing hot spots, increases particle size and inter-particle force due to the sticky coating and also increases minimum fluidization velocity.

The mechanism of agglomeration can be better understood in fuel analysis of reactive alkali and chlorine solvent and also ash melting, sintering and agglomeration temperature analysis.

1.6.3. Segregation

Segregation is defined as de-mixing or reverse mixing or separation of a component from a mixture of particles [32]. The problem of segregation has been studied by various researchers

in food, mechanical, chemical, material and agricultural engineering. In processing and manufacturing and handling industries, like food product industries, segregation has caused large variations in product packaging because of differences in bulk densities [33].

Fluidized beds are studied to determine operation conditions necessary to promote bed mixing and to minimize or eliminate particle segregation. Also for different materials, optimum conditions under which separation can be accomplished can also be studied. These two objectives are necessary for investigating particle segregation in a fluidized bed [27].

The behavior of float-sam and jet-sam of large particles is dependent on the ratio of its absolute density to the bulk density of small particles. The classification of float-sam and jetsam is prominent in multi-component bed mixing where a wide size and density difference is dominant. The jet-sam particles usually fall to the bottom of the bed while lighter and smaller float-sam particles stay at the top and interstitial spaces of the larger particles [34].

Segregation mechanism has been classified based on physical properties of the particles (size, density and shape) [35], it has also been classified based on the direction of particle movement as vertical and horizontal segregation [36]. Above all, research done identified the more generalized classification of segregation mechanism as percolation, displacement, sieving, air current, rolling, trajectory, fluidization, push-away, impact, embedding, angle of repose, concentration-driven displacement and agglomeration [37,38 and 39].

In a particle elution, the elution characteristics of well-mixed bed are different from segregated bed with good amount of particles concentrating at the surface of the bed. Segregation may occur when small and dense particles are very small and tends limits interparticle movement to areas with passing bubbles [40].

It is rare to have an ideal system of equal density mono-sized particles in fluidized bed system, because applications usually involve a binary mixture of particles, hence mixing and demixing will always occur at some operating conditions. But it is useful to separate the driving force for segregation and mixing in a given system in order to understand the process and achieve some desirable quantities of a system. One method of reducing segregation is by adjusting the physical properties of the particles.

Particle segregation can usually be prevented by operating at sufficiently high velocity [27]. The application of electric field can also modify interparticle force and remedy the problem of segregation [41]. Lastly, the improvement of physical properties, adjustment of handling steps or device parameter and the use of mass flow bin can minimize segregation [42].

1.7. Objective and Scope

Although modern technology gives a more precise prediction and application, more research is still needed in the study of fluidization. This is because of the difficulties in prediction of the behavior of a fluidized bed reactor due to the complexity and ambiguity of some fundamental properties (size, density and shape of the particle) which plays an important role in prediction and calculation of hydrodynamics of fluidized beds [43]. In related papers presented, the effect of distributor design has been given much attention in bubble size, bubble formation and fluidization quality. Pan et al have performed experiment on fluidization of 1204 spheres where values were compared with simulated values by fluidizing mono-disperse spheres of nylon in a two dimensional-like bed using a uniform distributor [44].

Particles can be fluidized by liquid or gas. So much attention has been given to gas-solid fluidized bed reactor with binary mixtures. The objective of this study is

- To investigate the behavior of three different types of particles with same size and different densities in a quasi-three dimensional liquid–solid fluidized bed by two different distributors.
- The effect of these distributors on mono-disperse and poly-disperse (bi-disperse) particle fluidization with water at room temperature.
- The hydrodynamics of a single particle behavior of these spheres in a uniform distributor and compares experimental values to numerical values.
- The effect of Wall on the single particle.

The distributors used are a multi-orifice distributor and a single orifice distributor. The multi-orifice distributor will be used for mono-disperse and poly-disperse and single particle fluidization while the single orifice will be used for mono-component fluidization.

Chapter one of these theses focuses on the introduction and literature review with emphasis on fluidized bed system, distributors, mono-disperse and poly-disperse fluidization and their application in industries, problems and remedies in fluidization. These sections also introduce some of the basic concepts and definitions that will be used throughout the thesis.

Chapter two describes the design of the fluidized bed used in this experiment with emphasis on the column and distributor design and compares with other designs in other studies. It also talks about material selection and reason for selection, procedures and equipment used.

Chapter three emphasizes on the fluidization of a single particle and analyzes the force balance, drag coefficient which was applied numerically to solve for the numerical values of the minimum velocity. It also emphasizes on the terminal velocity and compares result with the numerical values of velocity without wall effect and experimental data was compared with the

numerical and analytical values to obtain the wall effect on the particle. Particle fluidization with multi-orifice and single orifice distributors with pictures and experimental data, tables of data and graphs and experimental observations are also included in this chapter.

This thesis ended with chapter four on the comparison of the two distributors based on experimental findings, conclusion and recommendation for future study on similar research. This design is simple and parts are available and affordable at local stores.

CHAPTER 2

DESIGN

2.1. Material Selection

This is a very important factor as the success of the bed lies in the strength of the material used. At first, 0.635cm thickness of Plexiglas material was selected for this experiment, the effect manifested itself. A failed design resulted as the bed expands since it could not withstand the pressure buildup. Replacing the material with a thickness of 1.30cm, a great improvement in strength and rigidity was achieved.

2.2. Column Design

The liquid solid Fluidized bed design parameters involve the main components which are the fluidization column (bed portion, water and distributor), flow control, water inlet and return line, instrumentation and water supply. Design parameters are bed height and pressure drop across the distributor which is usually based on application. The quasi-three-dimensional rectangular column fabricated with a transparent Plexiglas plate was constructed in the mechanical engineering laboratory. The dimensions of the bed are 7 cm deep, 30 cm wide and 91 cm tall with a plate thickness of 1.30cm for structural strength and rigidity. The fluidizing water was supplied by a 3-hp submersible water pump. Aluminum, nylon glass spheres whose physical properties are summarized in table 2.2 were fluidized.

Two types of distributor were designed to determine their effects on fluidized-bed hydrodynamics. The bed was calibrated in centimeter using a flexible tape attached to one side of the bed to measure the height of the bed.

On the top of the bed is an outlet which is connected to a discharge port made of Plexiglas material. This outlet is wide enough to prevent a pressure buildup on the bed and help

prevent bed collapse. The bottom of the bed is an inlet which is mounted to the top of the distributor. This serves as an inlet for the water to flow from the outlet of the distributor outlet to the bed.

2.3. Distributor Design

The effect of the Distributor on the hydrodynamics of particles in the bed is one of the major interests of this thesis and hence it requires a careful design consideration.

Design consideration is based on distributor pressure drop determination and fluid velocity. Bad fluidization is a result of low pressure drop across distributors. This is so because there is no uniform distribution of water and hence temporary de fluidization may occur and channeling may also result. High pressure drop distributors may give a more uniform distribution of fluid, but more power is needed to run the compressor and fans [45].

In the case of upward and lateral flow, the ratio of distributor pressure drop to bed pressure drop should be 0.3 for even fluid distribution [29].

$$\Delta P_d = 0.3 \Delta P_b \quad (2.1)$$

And for downward flow, the distributor pressure drop should be 10 % of bed pressure drop [46].

$$\Delta P_d = 0.1 \Delta P_b \quad (2.2)$$

Where Δp_d represents the pressure drop across the distributor and Δp_b is the pressure drop across the bed.

The ratio of the bed height to bed width influences the critical value of the ratio of the distributor pressure drop to bed pressure drop [47]. Kiarash Vakhshouri investigation on the influence of distributor and plenum chamber volume on gas bubble behavior and sizes

showed that the behavior of the distributors was due to pressure drop which is a function of open area ratio of the distributor [14].

The design of the distributors in this experiment was in accordance with the experiment of Nidal et al. [48]. The minimum fluidization of the two different distributors differs greatly this is because of the size of the orifice or rather open area ration of the distributors. As the open area ration increases, the minimum fluidization reduces.

Two Rectangular boxes measuring 7cm × 7cm × 32cm (tall, wide and length) were built using a Plexiglas material of thickness 0.635cm it has an opening at the bottom which serves as inlet for water to flow through. On the top of the distributor is the orifice opening which has a base for the bed to sit. For the single orifice distributor, a rectangular hole measuring 0.5cm × 5cm was centrally drilled and for the multiorifice, 46 holes each with a diameter of 0.4cm was drilled on the top of the distributor. This is similar to the distributors in Kiarash Vakhshouri experiment. The difference being that on his design, the areas of both the single and the multi-orifice distributors were the same. The open-area ratio of the distributors was calculated from the formula,

$$A_o = \frac{NA}{B} \quad (2.3)$$

Where N is the total number of holes, A is the area of the orifice and B is the area of the bed. With this formula, the open area ratio of the multi-orifice turns out to be 23% and the single orifice is 10%. The Appendix gives a full description of the uniform and jet distributors and their working drawings.

2.4. Principle of Operation

The two-dimensional fluidized bed is operated by a 3hp submersible pump fully immersed in a cylindrical reservoir of water with dimensions (40 cm, 70 cm) (diameter, length) at room temperature. The pump is connected to plumbing which has a valve attached to the flowlines. The function of the valve is to control the flowrate and velocity of the water and hence was graduated at some point so one can read off the flowrate directly. This plumbing runs into the distributor which is attached to the bed and at the top of the bed is a return line which scavenges water back into the cylindrical reservoir.

When the water pump is turned on by connecting to an external AC power, regulated water runs into the flowlines, the amount of water and flowrate is regulated by the valve. The valve is gradually open until bed is fluidized and readings of bed height at every superficial velocity and flowrates were recorded. At the end of the experiment, the bed is drained of water using the drain line installed along the flow line. Video imaging was recorded using a camcorder and pictures taking at every flow velocity and bed height.

For maintenance purpose it is a good practice to always drain the reservoir at the end of the experiment as this will help reduce corrosion on the pump and increase its service life.

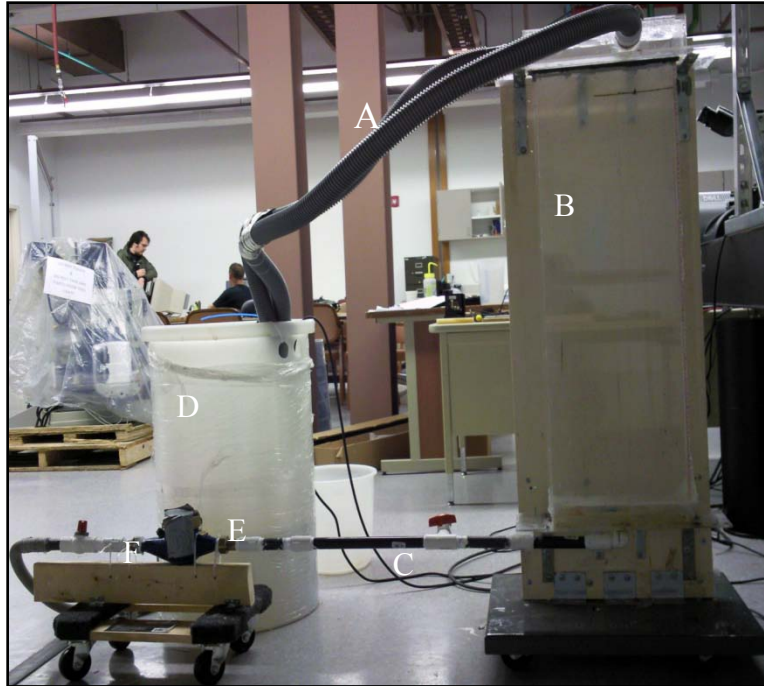


Figure 2. 1: Picture showing the fluidized bed. (A) Outlet, (B) Bed column, (C) Drain line, (D) Reservoir, (E) Inlet, (F) Flow meter

2.5. Flow Rate Measurement

Liquid flowrates were obtained by measuring the amount of water through a water meter calibrated in gallons over a fixed period of time.

The regulating valve was marked at different points to show the respective flow rates of water per minute of flow. The data collected was later converted to cm^3/secs . These values are shown in table 2.1. With these flow rates, superficial velocities were calculated using the formula $\frac{\varphi}{A}$, where φ is the flow rate and A the cross sectional area of the bed for the uniform distributor experiment on the single particle and also the cross sectional area of the distributor orifices for the single jet distributor and uniform distributors with multi-particles. Superficial velocity ranges are shown in table 2.1 for the distributors and the bed.

Table 2. 1: Superficial velocities and flowrates.

Flowrates (cm^3/sec)	Bed velocities (cm/sec)	Uniform distributor velocities(cm/sec)	Jet distributor velocities (cm/sec)
11.36	0.47	1.97	4.544
63.1	2.6	10.92	25.24
157.73	6.46	27.29	63.09
189.27	7.76	32.75	75.71
220.82	9.05	38.20	88.33
283.91	11.64	49.12	113.56
391.16	16.03	67.68	156.46
517.34	21.20	89.51	206.94
630.9	25.86	109.15	252.36

2.6. Bulk Densities

The bulk densities of particles were determined by liquid displacement using a known weight of particles (glass, nylon and aluminum spheres) was slowly added to a 40 ml of water in a graduated cylinder. The volume of water displaced is the volume of added particles. The weight was determined by adding 50 particles of aluminum, glass and nylon each into a measuring cylinder of known weight and placed on the Mettler Toledo weight scale graduated in grams in the mechanical laboratory. This is shown in figure 6 (a) and (b). After several measurements the average densities were calculated and results obtained were close to the company values of their densities. These values are shown in table 2.2.

Table 2. 2: Particles and their properties.

Particle	d (cm)	ρ_p (g/cm^3)	ρ_b ($\frac{g}{cm^3}$)
Aluminum	0.6	2.7	2.54
Nylon	0.6	1.14	1.135
Glass	0.6	2.45	2.38



Figure 2. 2:(a) Determination of the weight of alum (b) Determination of the weight of Nylon

CHAPTER 3

RESULTS AND DISCUSSIONS

3.1. Single Particle in a Uniform Distributor

3.1.1. Single Particle Fluidization

In a paper presented by Taehwan et al., they obtained correlations for lift-off of particles in Oldroyd-B fluids. The fluid elasticity reduces the critical shear Reynolds number for lift-off, the effect of the gap size between the particle and the wall, on the lift force, was also studied and the conclusion was that the particle lifted from the channel wall will attain an equilibrium height at which its buoyant weight is balanced by the hydrodynamic lift force [49].

For ease of simplicity, single particle fluidization gives a better picture of mechanism of fluidization. Assumptions made are based on the fact that particle is moving at a constant velocity relative to its immediate fluid. The force acting on this particle depends on the flow of fluid in its immediate vicinity [27]. When a single particle is fluidized in a bed of either water or gas, with increase in velocity, a minimum velocity is achieved (U_{mf}), at this point, upward hydraulic force is equal to weight of the particle or net gravity. Increasing the velocity above this point causes the particle to move upward, this implies that the hydraulic force has overcome the net gravity and the particle velocity compared with fluid velocity reduces to a point that the force of gravity is balanced. For single particle fluidization, the drag force on the particle is equal to the difference between the force of gravity and upward buoyant force.

This study analyzes the force balance on a single particle, drag coefficient which was applied numerically to solve for the numerical values of the minimum velocity. It also emphasizes on the terminal velocity and compares result with the numerical values of velocity

without wall effect. Finally, experimental data was compared with the numerical and analytical values to obtain the wall effect on the particle.

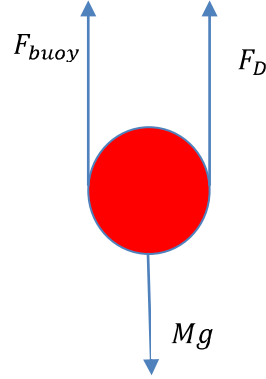


Figure 3. 1: Forces acting on a single particle.

$$F_D = Mg - F_{buoy} \quad (3.1)$$

$$Mg = V\rho_p g \quad (3.2)$$

$$F_{buoy} = V\rho_f g \quad (3.3)$$

$$F_D = V[\rho_p - \rho_f]g \quad (3.4)$$

$$V = \text{volume of the particle} = \frac{4}{3}\pi R^3 \quad (3.5)$$

This implies that at minimum fluidization,

$$F_D = \frac{4}{3}\pi R^3 [\rho_p - \rho_f]g \quad (3.6)$$

$$\text{Also, } F_D = \frac{1}{2}C_D(\pi R^2)\rho_f U_f^2 \quad (3.7)$$

Where C_D is the drag coefficient and defined as the ratio of the force on the particle and the fluid dynamic pressure caused by the area projected by the particle represented. The drag coefficient depends on the Reynolds number, which gives an indication of how the fluid is

flowing (i.e., laminar, semi-turbulent, or fully turbulent). For a spherical particle, the standard definition is expressed as

$$C_D = \frac{F_D}{(\rho_f U_f^2 / 2)(\pi d_p^2 / 4)} \quad (3.8)$$

And combining and substituting for F_D ,

$$C_D = \frac{4}{3} d_p \frac{(\rho_p - \rho_f)}{\rho_f U_f^2} g \quad (3.9)$$

$$R_e = \frac{d_p U_f \rho_f}{\mu_f} \quad (3.10)$$

Depending on particle Reynolds number, three regimes have been identified²⁷. The first is the Stokes flow regime. This is also called creeping flow where viscosity of the fluid is a dominating factor. The creeping flow in the packed bed exists due to the fact that the flow rate is very low. Hence the streamlines take the part of interstitial spaces of the space⁶.

$$C_D = \frac{24\mu}{U_f \rho_f d_p} = \frac{24}{R_{ep}} \quad R_{ep} < 0.2 \quad (3.11)$$

Secondly, the intermediate flow regime where drag coefficient is a function of the particle Reynolds number.

$$C_D = f(R_{ep}) \quad (3.12)$$

For $1 < R_{ep} < 1000$, which is referred to as the transition regime, the following expressions may be used for a sphere in a large domain [50]. This formula most is accurate since particles are constrained to Reynolds number value of this range.

$$C_D = \frac{24}{R_{ep}} [1 + 0.15 R_{ep}^{0.687}] \quad (3.13)$$

With this formula, the drag coefficient for nylon was determined.

Lastly the Inertial flow or Newton's law Regime. This is the flow at very high Reynolds number in the range of $Re_p > 1000$. The drag coefficient in this case approaches a constant value.

$$C_d \approx 0.44 . \quad (3. 14)$$

This value of the drag coefficient was used for aluminum and nylon particles which have very high Reynolds number. Fluidization regime is between the minimum fluidization velocity and the terminal velocity of the particles [51].

3.1.2 Terminal Velocity

Terminal velocity is the free falling velocity of a particle in a fluid. The weight of a particle equals drag force under terminal equilibrium conditions. The weight is usually the net effect of gravity and particle buoyancy [6]. Particles will be elurated from the bed if its velocity exceeds the terminal velocity. For spherical particles, Haider et al., summarized the terminal velocity at large domain as follows [52].

$$U_t = \left[\frac{4(\rho_s - \rho_f)gd_p}{3C_d\rho_f} \right]^{1/2} \quad (3. 15)$$

The value obtained with this formula was use to compare with the experimental results of minimum fluidization velocity to get the wall effect. Hence for nylon, the Reynolds number range applies $1 < Re < 1000$ and for glass and aluminum, $1000 < Re < 20000$ applies.

Table 3. 1: Values for the drag coefficient based on the Reynolds number

Drag coefficient, C_D	Reynolds Number, Re
$24/Re$	$Re < 1$
$\frac{24}{Re_p} [1 + 0.15Re_p^{0.687}]$	$1 < Re < 1000$
0.44	$1000 < Re < 20000$

3.1.3. Numerical Data

Assumptions are free flow without wall effects; the density of water is constant steady state flow.

Combining equ.3.6 and 3.7,

$$\frac{1}{2} C_D (\pi R^2) \rho_f U_f^2 = F_D = \frac{4}{3} \pi R^3 [\rho_p - \rho_f] g \quad (3.16)$$

$$\text{And } C_D = \frac{8R(\rho_p - \rho_f)g}{3\rho_f} = \frac{8R}{3} (s - 1)g \quad (3.17)$$

Where $s = \text{specific gravity} = \rho_p / \rho_f$, $R = 0.6 \text{ cm}$, $\rho_f = 1 \text{ g/cm}^3$,

$$g = 980 \text{ cm/s}^2 \quad \text{and } \rho_p = (2.7, 2.45, 1.14) \text{ g/cm}^3 \quad (\text{Al, glass and nylon})$$

Also from equ.3.13, for a sphere in a large domain,

$$C_{D,\infty} = \frac{24}{Re_{p,\infty}} [1 + 0.15 Re_{p,\infty}^{0.687}] \quad (3.18)$$

$$\text{Where } Re_{p,\infty} = \frac{\rho_f U_\infty 2R}{\eta_f} \quad (3.19)$$

The fluid dynamic viscosity is denoted η_f . ($\eta_f = 0.01 \text{ P}$)

$$\text{And } Re_{p,\infty} = 60 U_\infty \quad (3.20)$$

The above equations were solved numerically and the following data in table 3.2 was obtained.

Table 3. 2: Numerical data

Particles	U_{∞} (cm/s)	$Re_{p,\infty}$	$C_{D,\infty}$
Nylon	15.65	939.06	0.45
Glass	50.83	3049.76	0.44
Aluminum	55.04	3302.22	0.44

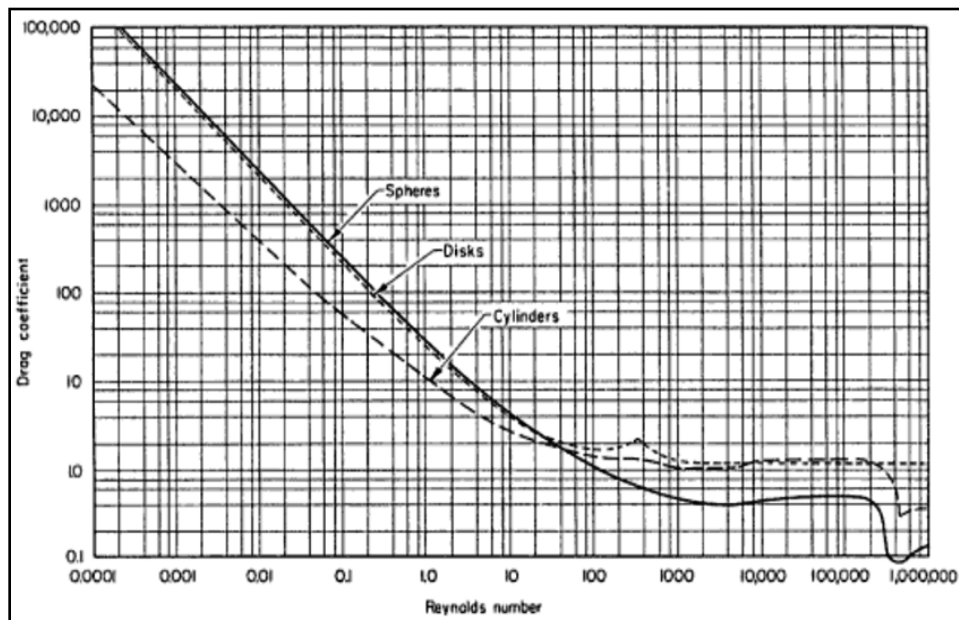


Figure 3.2: Drag coefficient of particles of various shapes moving in a fluid at various Reynolds number [53].

The numerical values of the drag coefficient and Reynolds number has also been compared with figure 3.2 and found to be true.

3.1.4. Experimental Study.

For the single particle experiment, the positioning of the sphere was of great interest. Several positions were chosen at first but the obtained results were different from the expected value from the numerical data and other similar experiments. A better result was obtained with the particle placed at a bed height just above half the height of the bed. At this point, the flow has a uniform velocity which was observed with the stream of water having no relative fluctuation. This problem is true with the conclusion that “the bed height of the sphere above the bottom increases with the fluidizing velocity” [44].

Table 3. 3: Experimental values for single particle fluidization

Particle	d (cm)	ρ_p (g/cm ³)	U_{mf} (cm/sec)	U_t (cm/sec)	$\frac{U_{exp}}{U_t} = f$
Aluminum	0.6	2.7	22.75	55.04	0.413
Nylon	0.6	1.14	7.76	15.65	0.496
Glass	0.6	2.45	21.20	49.95	0.424

Table 3.3 shows the three particles with their properties, minimum fluidization velocities, terminal velocities and wall effect. The trend is the same for the three particles but with a very distinctive difference in behavior for nylon compared with aluminum and glass both having a minimum velocity close to 22.75cm/sec and 21.20cm/sec. A lot of similarity which lies in the small density difference is common with aluminum and glass spheres. The fluidization experiment of these three particles reveals the fact that the particle is moving at a constant velocity relative to its immediate fluid and it also reveals that the effect of wall is responsible for the lesser experimental values of minimum fluidization velocity in a narrow domain compared with the terminal velocity of the same particle in a large domain.

The data for Nylon particle was compared with the findings of Pan et al., [44] as shown in figure 3.2 and found to be in agreement. In their simulation an exactly zero velocity of the sphere was not achieved. For $U_{mf} \leq 6 \text{ cm/s}$ the particle will not rise. The particle rises out of the bed when $U_{mf} > 10.5 \text{ cm/s}$

Table 3. 4: Average velocities of a sphere [44].

Inflow velocity (<i>cm/s</i>)	Average vertical speed (after it has stabilized)
6.0	-0.00522
6.5	0.00827
7.5	0.00219
9.0	-0.00178
9.5	0.00631
10.0	0.006997
10.5	0.00521
10.75	0.276
11.0	0.265
12.0	1.260

3.1.5. Wall Effect

Wall effect is defined as the retardation of the motion of an object in a column due to displacement and opposing motion of surrounding fluid [54]. It is a correction factor is a correction factor for velocities and this is needed to correct the drag coefficient and Reynolds number. Conversely, the drag experienced by the sphere between walls is much higher than that

with no walls. The simplest way to quantify wall effect by using a factor f which is a ratio of two velocities [55].

$$f = \frac{U_{exp}}{U_t} \text{ For } 0 < f < 1, \quad (3.21)$$

$$C_{D,\infty} = f^2 C_D \quad (3.22)$$

$$\text{And } R_{ep,\infty} = R_{ep}/f \quad (3.23)$$

U_{exp} Is the empirical value of the velocity of the sphere and $U_t, R_{ep,\infty}$, and $C_{D,\infty}$ are the terminal velocity, Reynolds number and the drag coefficient of the sphere without any wall effect.

3.1.6. Comparism with other Studies

Di Felice [56] in his experiment of a single particle in a tube summarized that for all cases of d/W , the minimum fluidization velocity is expressed as,

$$\frac{U_t}{U_{t\infty}} = \left(\frac{1-\lambda}{1-0.33\lambda} \right)^\alpha \quad (3.24)$$

α is a function of the bounded Reynolds number

$$\lambda = \frac{d}{W} \quad (3.25)$$

W is the width of the bed and d is the diameter of the particle.

U_t is the minimum fluidization velocity in a system with wall.

$U_{t\infty}$ is the minimum fluidization velocity in a system without wall.

In my experiment, $\lambda = 0.85$, extrapolating the data of $\lambda = 0.6$ to $\lambda = 0.85$, at $R_{ep} > 1000$,

$\frac{U_t}{U_{t\infty}} \sim 0.3$, which shows a results of ~ 0.45 . This is true since the wall effect from tube should be

greater than the narrow channel which I used in my experiment.

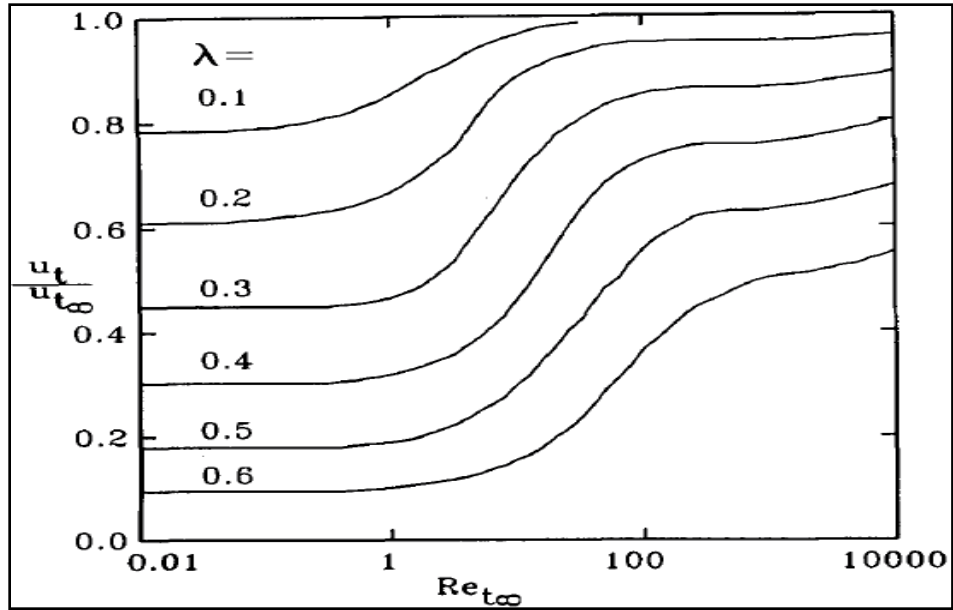


Figure 3.3: The Ratio of $U_t/U_{t\infty}$ for wide range of Reynolds Number and λ [57]

3.2. Fluidization with Different Distributors

3.2.1. Uniform Distributor

On the uniform distributor, separate particles of aluminum and nylon were fluidized for mono-disperse fluidization and binary mixture of aluminum-nylon particles and aluminum-glass particles were experimented for poly-disperse fluidization to study the hydrodynamics of mixing and segregation and applications in industries.

3.2.1.1. Mono-disperse Fluidization

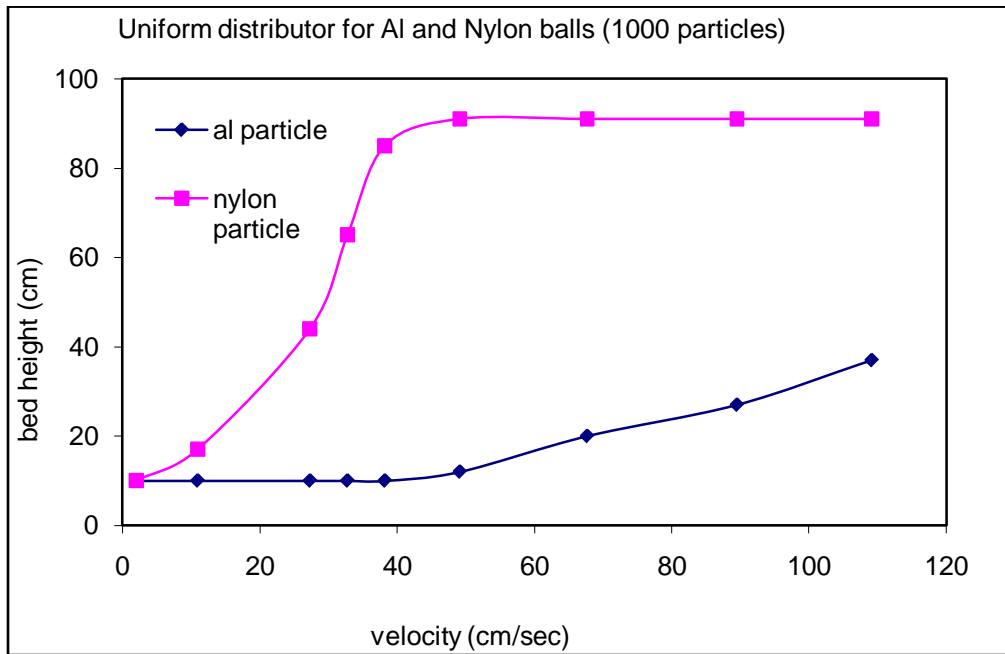


Figure 3. 4: Aluminum and nylon (1000 particles).

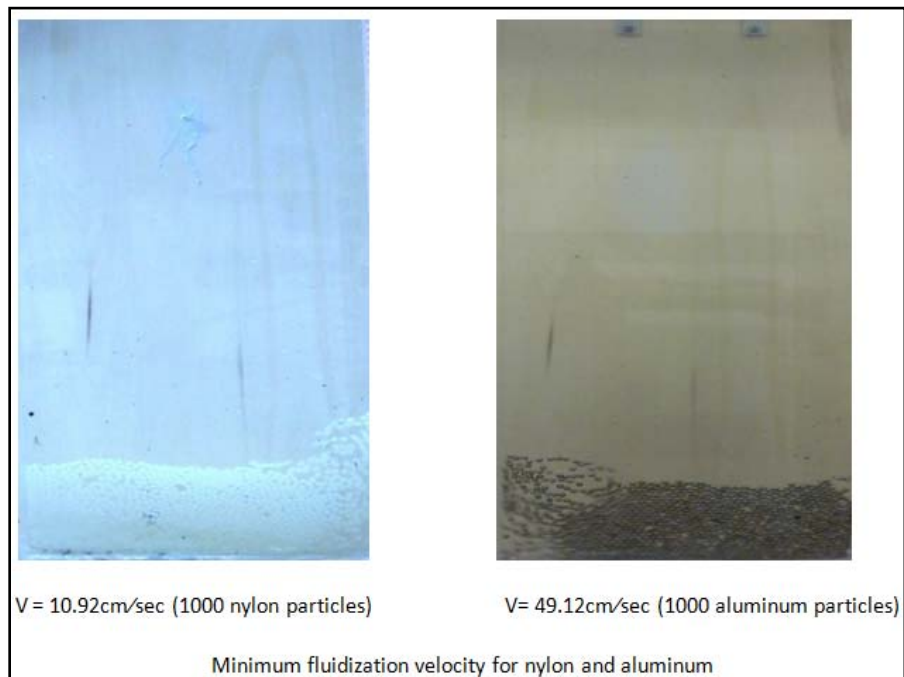


Figure 3. 5: Minimum fluidization for nylon and aluminum (1000 particles each)

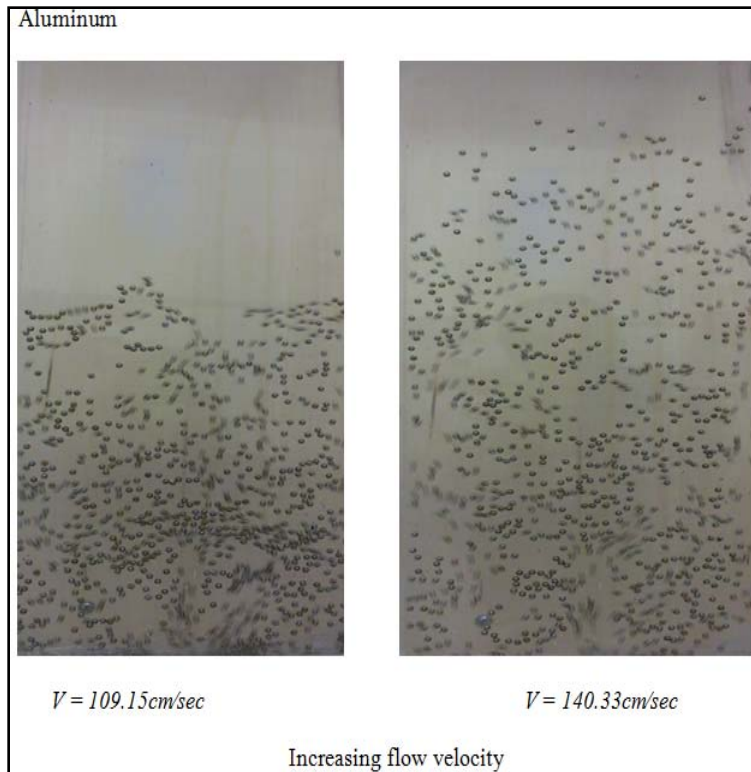


Figure 3. 6: Increase in superficial velocity for 1000 aluminum particles

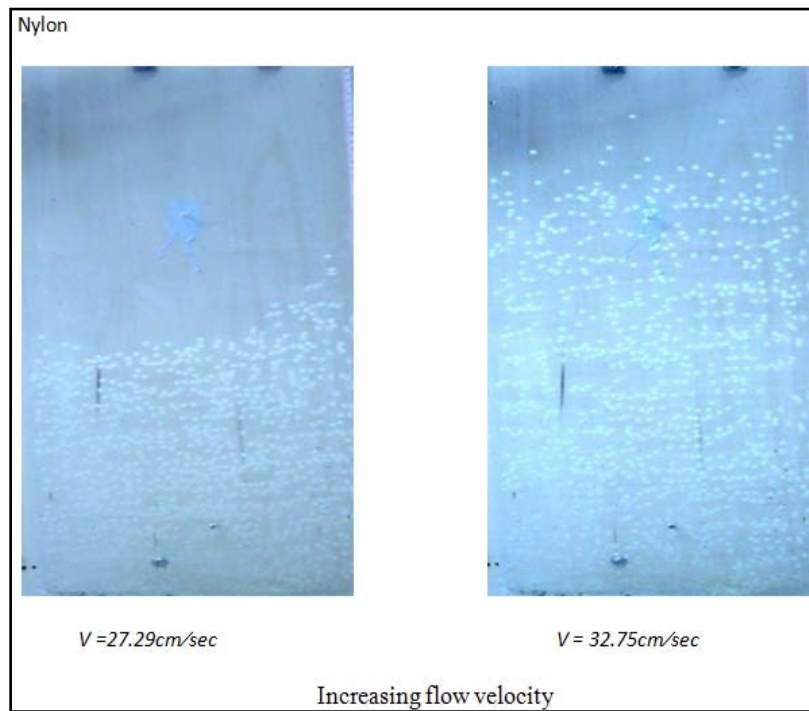


Figure 3. 7: Increase in superficial velocity for 1000 nylon particles

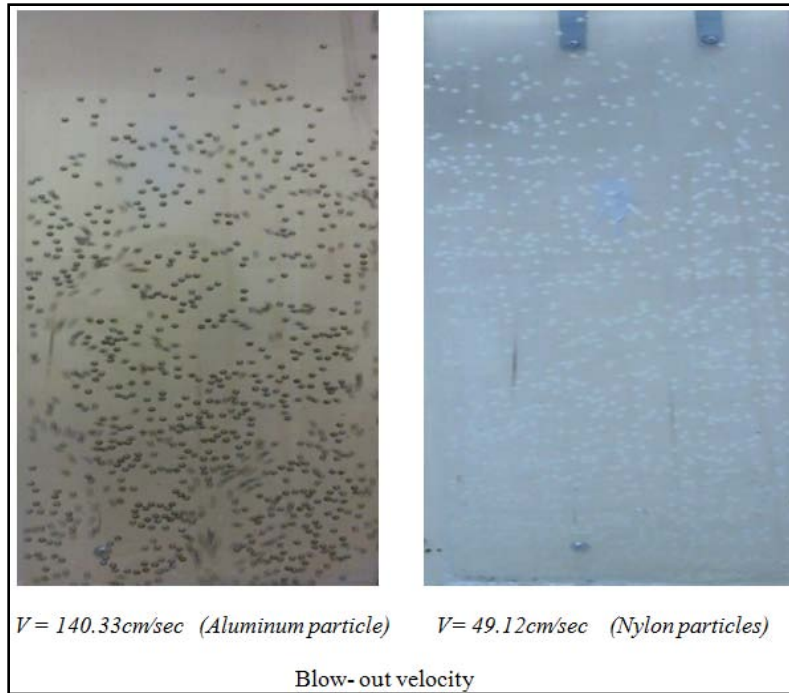


Figure 3. 8: Blow-out velocities for 1000 nylon and 1000 aluminum particles

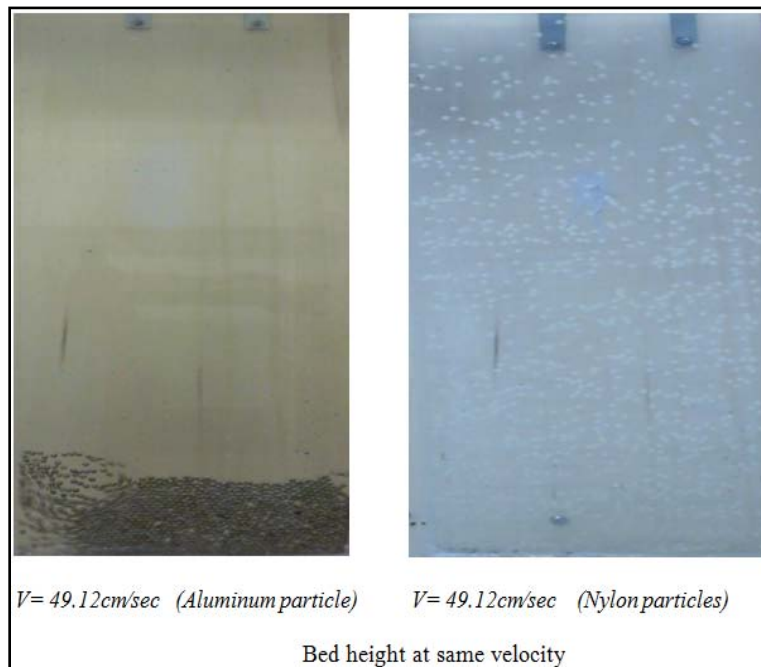


Figure 3. 9: Bed height of 1000 aluminum and 1000 nylon particles at same velocity

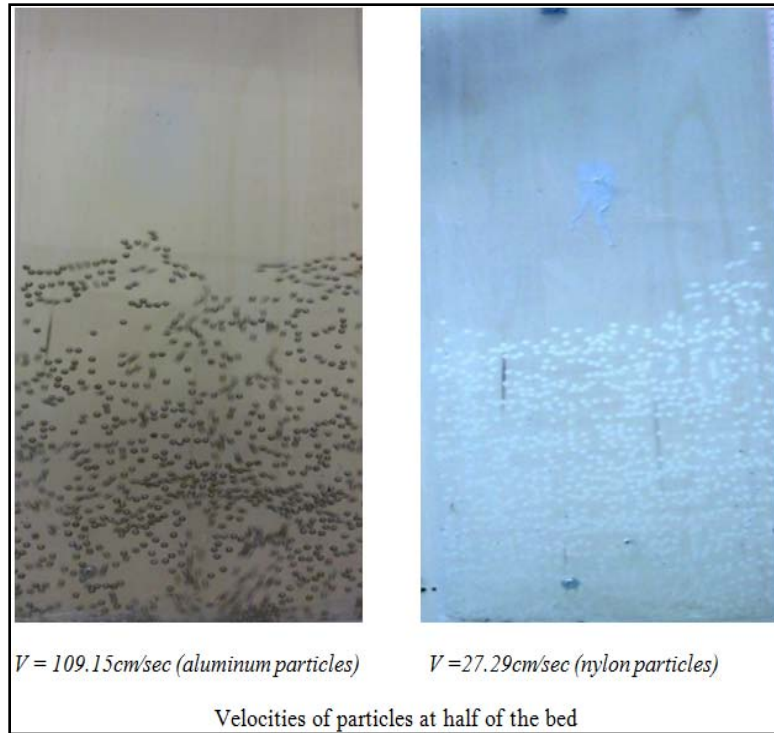


Figure 3. 10: The velocities of particles at half of the bed

Density plays an important role in the minimum fluidization velocities. Comparing the fluidization of 1000 nylon particles and aluminum particles. It is found that the minimum fluidization velocity of 1000 nylon particles is about 11 cm/s, which is about 1.5 times the minimum fluidization velocity of a single nylon particle (figure 3.4). On the other hand, to fluidize 1000 aluminum particles it needs a minimum fluidization velocity close to 50 cm/s (figure 3.4), which is 2 times of the minimum fluidization velocity of a single aluminum particle and at this velocity approximately 50 cm/s, nylon particles reach the top of the bed (figure 3.8). Increasing flow velocity causes the increase of bed height, at high velocity, the uniform distributor behaves like a transport bed.

As seen in Figures 3.7, at flow velocity 50 cm/s, nylon particles reach the top of the bed column. However, aluminum particles are only able to fill half of the bed column even at flow

velocity 140 cm/s. To reach half the bed height of aluminum, it requires 4 times velocity of nylon (figure 3.9).

A comparison between fluid velocity and bed height is shown in Figure 3.3. This relation has also been compared with the work of Wen-Ching Yang [27] in figures 3.10 on the relationship between bed height and liquid superficial velocity and found to be in agreement. This figure shows the direct relationship between bed height and superficial velocity. With increasing bed height, superficial velocity increases to a velocity U_0 when the bed height experiences an infinite value $l \rightarrow \infty$.

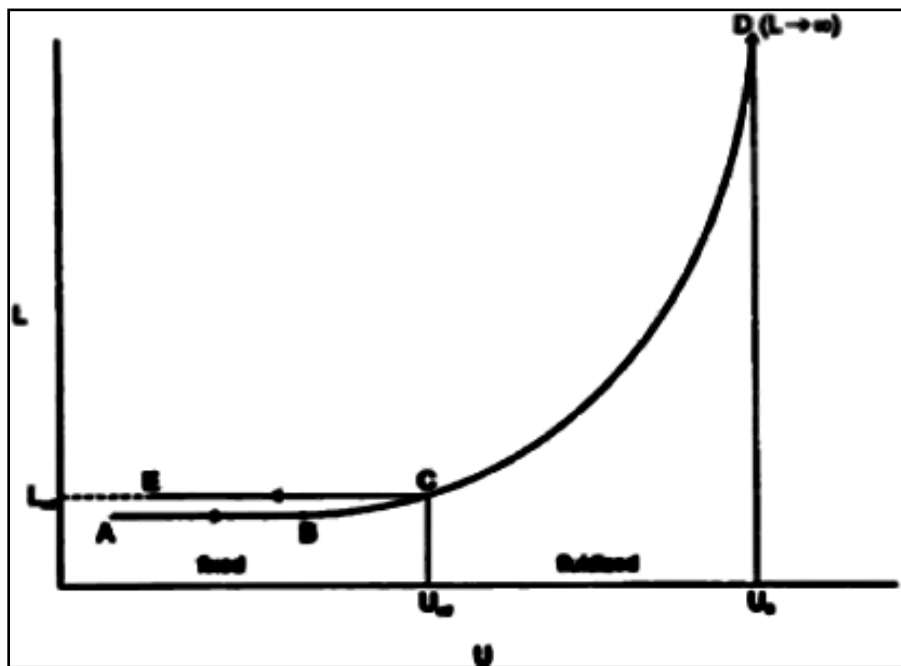


Figure 3. 11: Bed height and liquid superficial velocity [27].

At incipient fluidization, there are no mixing, mixing starts when the flow rate is increased. A characteristic homogenous random motion of particles at constant average velocity of fluid and momentum throughout the bed results with increase in superficial velocity above minimum fluidization velocity, this resulted in a homogenous expanded bed whose height

changes with flow rate. This is usually achieved in the absence of gulf streaming (i.e. separate regions of upward and downward solids motion) [58], this phenomenon is common with the single jet distributor. Going by Courderc assumption with a uniform distribution of particles in a liquid solid fluidization, homogenous bed with stable operation was achieved [59].

3.2.1.2. Uniform Distributor for Poly-disperse Fluidization

Differences in properties of the respective particles strongly influence the hydrodynamic behavior of binary fluidized beds. With relatively large differences the tendency for segregation becomes correspondingly strong and eventually causes a binary fluidized bed to separate completely [60]. For a bi-dispersed fluidized bed of aluminum and nylon, the necessary condition for existence of a fluidized bed is summarized as $(\rho_1 < \rho_2)$ [21], where ρ_1 is the density of the lighter particle and ρ_2 is the density of the heavier particle. Segregation is dependent on the differences in density and size of the components, and gas velocity in the bed [61]. In the case of particles of the same size, their density difference is the major factor of segregation. It is expected that the increase of density difference will accelerate segregation. However, it is unclear how much density difference will be able to cause particles segregation and what the role of flow velocity plays. In the case of fluidization of a binary mixture of equal number of nylon particles and aluminum particles that has a higher density ratio of 0.42, a stratified layer of nylon was observed on top of the bed, which results a complete separation as seen in figure 3.12. Low density nylon particles are suspended on the top region of the bed column while heavy aluminum particles remain in the lower portion of the bed.

With aluminum and glass spheres with ratios close to unity, the process of fluidization achieves a nearly homogeneous bed, which results to a completely mixed bed. This is shown in Figure 3.11 where binary mixture of 1000 glass particles and 1000 aluminum particles is used for

fluidization. The density ratio between glass and aluminum particles is 0.91. It is found that two types of particles are able to be mixed uniformly and there is no segregation even at high flow velocity. This shows that a density difference of below 10% will not cause significant segregation during fluidization.

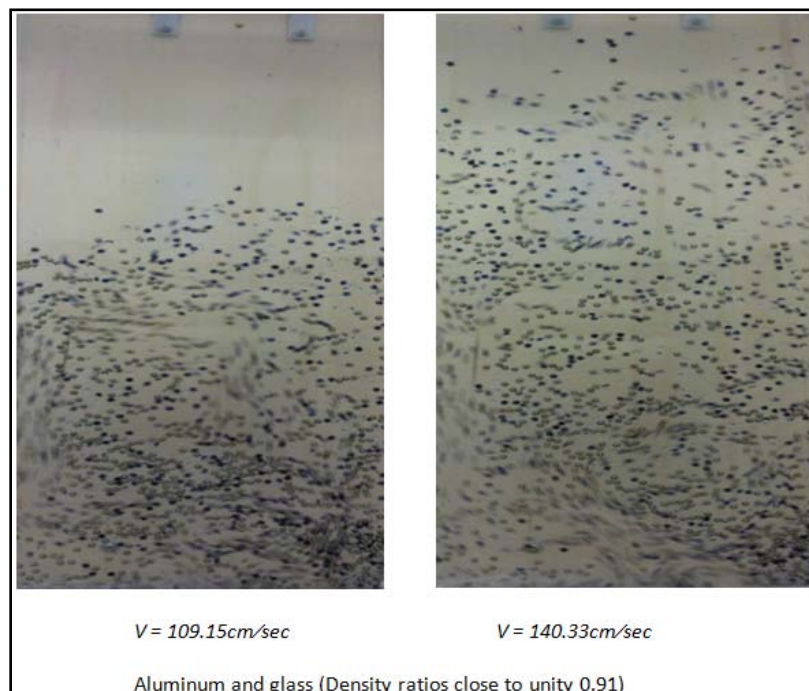


Figure 3.12: Binary mixture of 1000 glass particles and 1000 aluminum particles

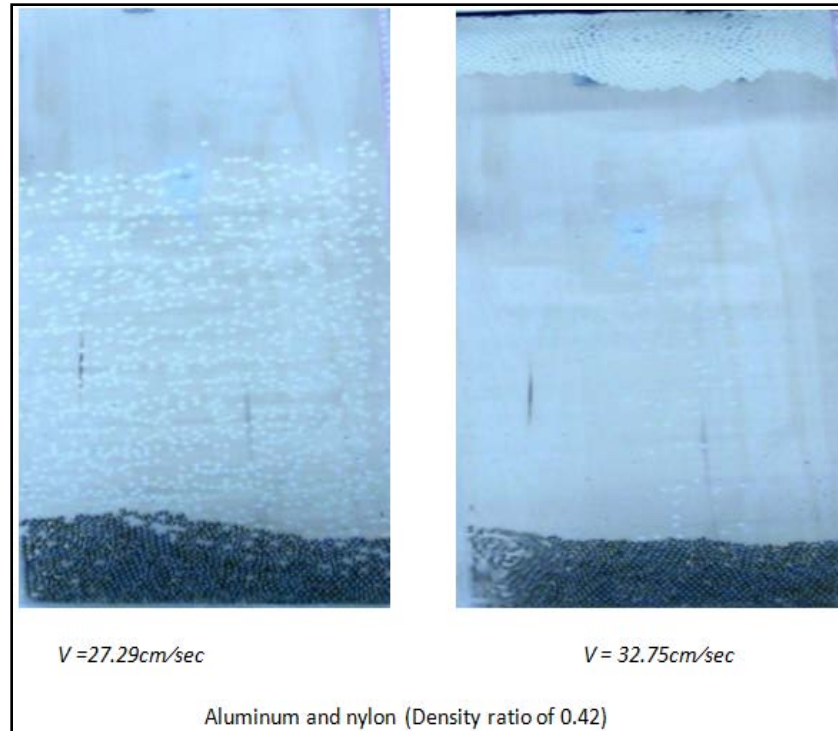


Figure 3. 13: Binary mixture of 1000 nylon particles and 1000 aluminum particles

Comparing figure 3.11 and 3.12, it can be seen that density was the major mechanism for their mixing and segregation. Comparing the density ratios ρ_2/ρ_1 , for a good or complete mixing, a ratio close to unity is needed.

3.2.2. Single Jet Distributor

Experiments were carried out in a quasi-three-dimensional rectangular liquid solid fluidized bed attached to the distributor shown in the figure below. Mono-component spheres of Aluminum and Nylon were fluidized with water at room temperature. Data obtained from this experiment include minimum fluidization velocity, blow-out velocity and bed height at different superficial velocities. The results were compared with that of uniform distributor to study the effect of distributor design on fluidization dynamics.

In some ways, this is reminiscent of the dynamics of a water fountain. The single jet can be likened to a spouted bed. In the images in figure 3.14, 3.15 and 3.16, spouting occurs as water is continuously injected with a high velocity through the distributor opening at the center of the bed. Water penetrates the whole bed and carries along particles upward. At the top of the spout, particles fall back to the top of particles located at the bottom of the bed and then recirculate just like in circulating fluidized bed.

3.2.2.1. Single Jet Distributor for Mono-disperse Fluidization

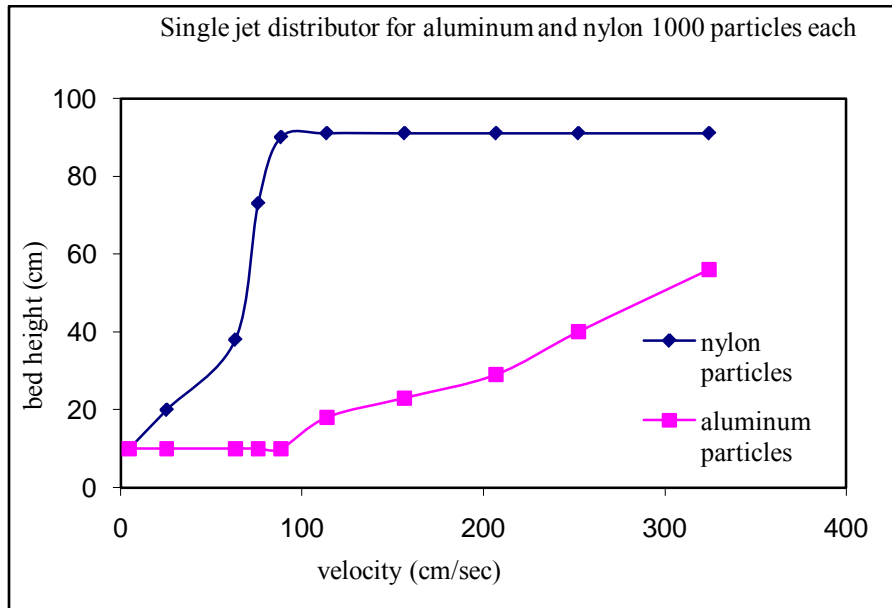


Figure 3. 14: Aluminum and nylon (1000 particles each)

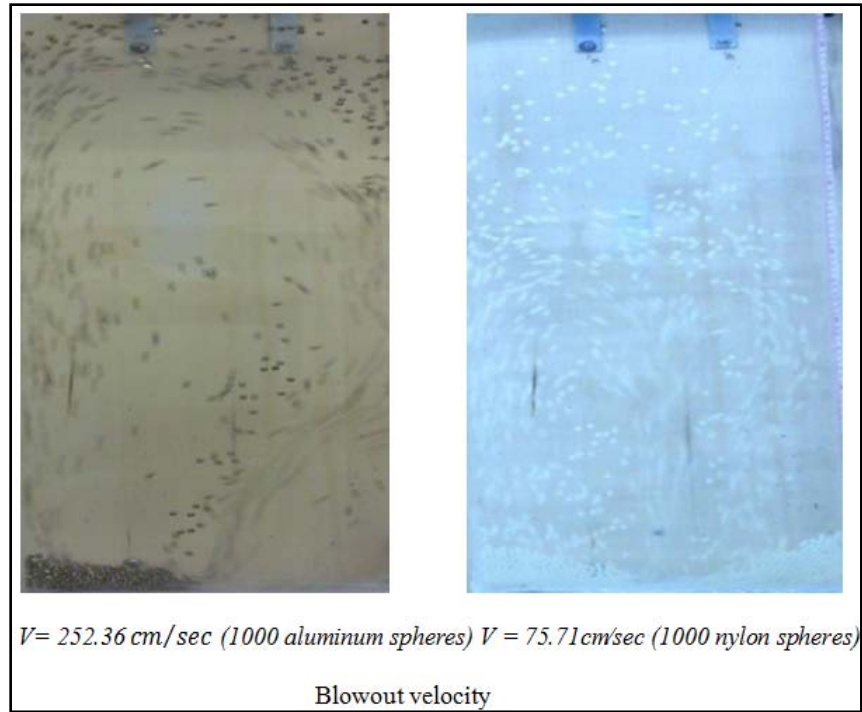


Figure 3. 15: Blow-out velocities for 1000 nylon and 1000 aluminum particles.

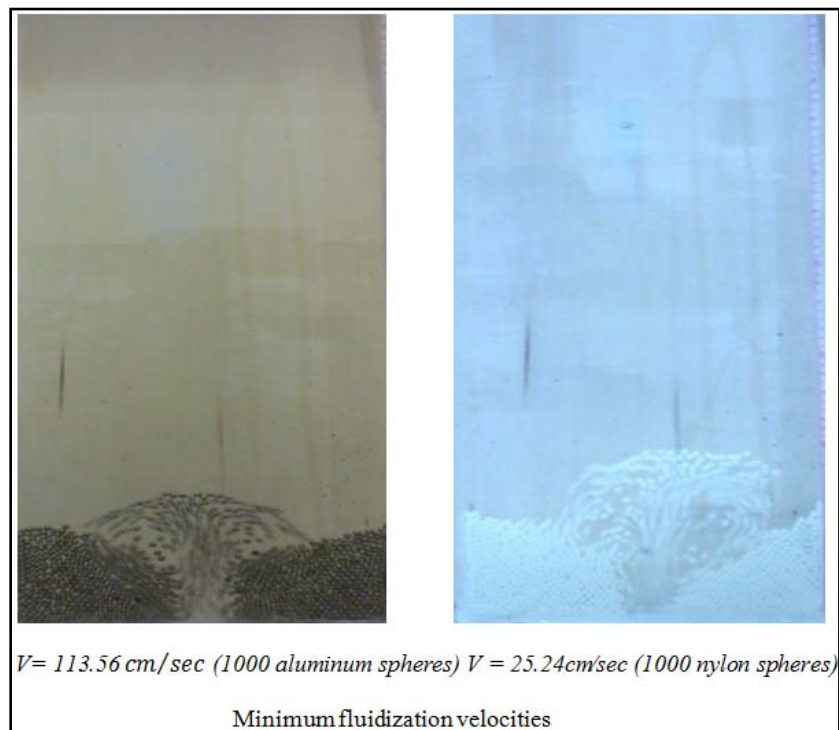


Figure 3. 16: Minimum fluidization for nylon and aluminum (1000 particles each)

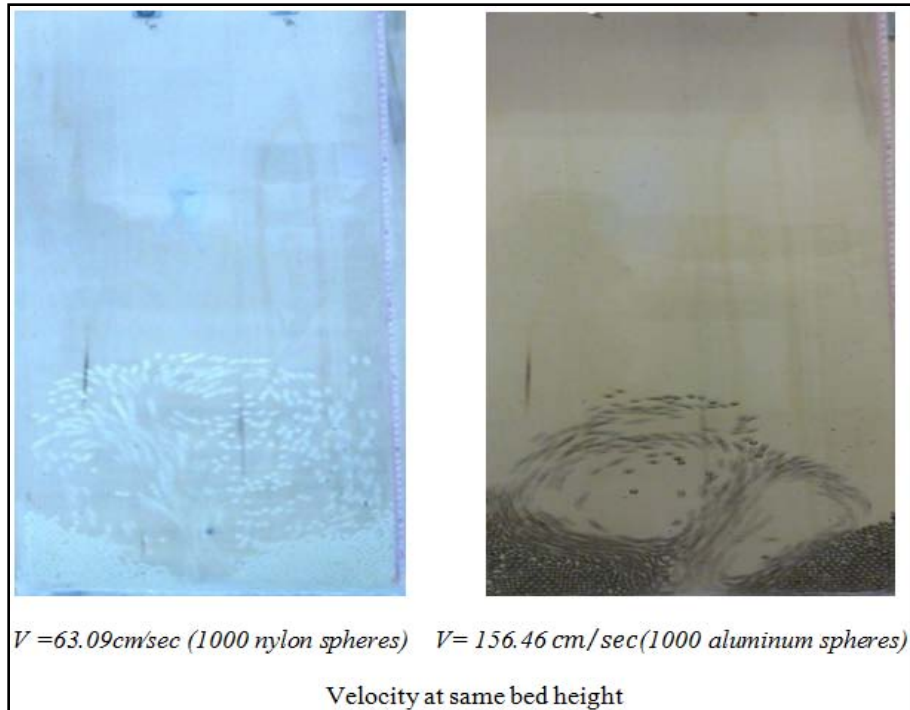


Figure 3. 17: superficial velocities at the same bed height

Summarily, it is found that the minimum fluidization velocity of 1000 nylon particles about 3.25 times the minimization velocity of a single nylon particle, and to fluidize 1000 aluminum particles it requires 5 times of the minimum fluidization velocity of a single aluminum particle as shown in figure 3.15. Increasing flow velocity causes the increase of bed height (figure 3.13). At flow velocity 75.71 cm/s , nylon particles reach the top of the bed. Aluminum particles are only able to fill the bed column at flow velocity 252.36 cm/s , this is shown in figure 3.15. At same bed height, aluminum requires 4 time velocity of nylon (figure 3.16).

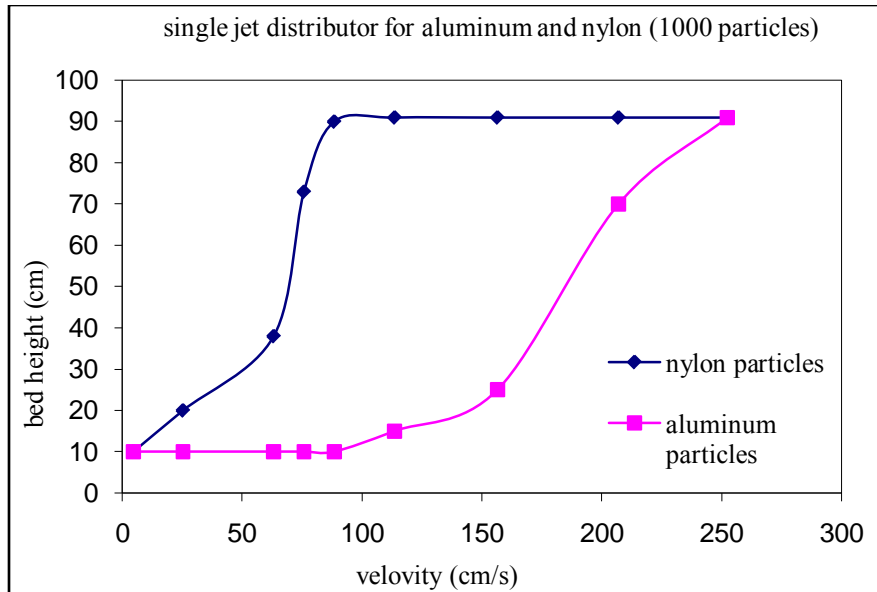


Figure 3. 18: Aluminum and nylon (1000 particles)

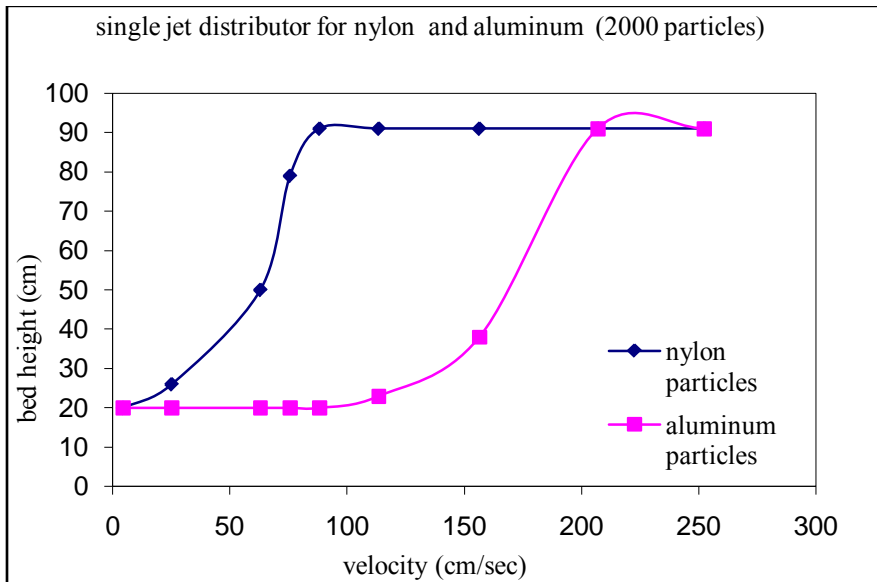


Figure 3. 19: Nylon and aluminum (2000 particles)

3.3. Discussion

3.3.1. Mixing Ratio

By adjusting the size of aluminum, better mixing could be achieved in the fluidization of both aluminum and nylon particles. This is done by balancing the respective drag forces of both particles. From figure 3.1,

$$F_{Dny} = M_{ny}g - \rho_f V_{ny} = \frac{1}{2} \rho_f U^2 C_{Dny} A_{ny} \quad (3.26)$$

$$\text{And } F_{Dal} = M_{al}g - \rho_f V_{al} = \frac{1}{2} \rho_f U^2 C_{Dal} A_{al} \quad (3.27)$$

$$\text{And } U^2 = \frac{2g(M_{ny} - \rho_f V_{ny})}{\rho_f C_{Dny} A_{ny}} = \frac{2g(M_{al} - \rho_f V_{al})}{\rho_f C_{Dal} A_{al}} \quad (3.28)$$

From my numerical data in table 3.2,

$$C_{Dny} \approx C_{Dal} \quad (3.29)$$

$$\text{And } \frac{(\rho_{ny} - \rho_f) V_{ny}}{A_{ny}} = \frac{(\rho_{al} - \rho_f) V_{al}}{A_{al}} \quad (3.30)$$

$$(\rho_{ny} - \rho_f) d_{ny} = (\rho_{al} - \rho_f) d_{al} \quad (3.31)$$

$$\text{This implies that } (1.14 - 1.0) d_{ny} = (2.7 - 1.0) d_{al} \quad (3.32)$$

$$0.14 d_{ny} = 2.7 d_{al}$$

$$d_{ny} = 12 d_{al}$$

This means that fluidizing aluminum sphere of diameter of 1mm and nylon diameter of 12mm will result in a good mixing of aluminum and nylon. This is a diameter ratio of aluminum to nylon of 0.08: 1 compared with 1:1 and also, $d_{al}/d_{ny} < \rho_{al}/\rho_{ny}$.

3.3.2. Comparison between the Two Distributors

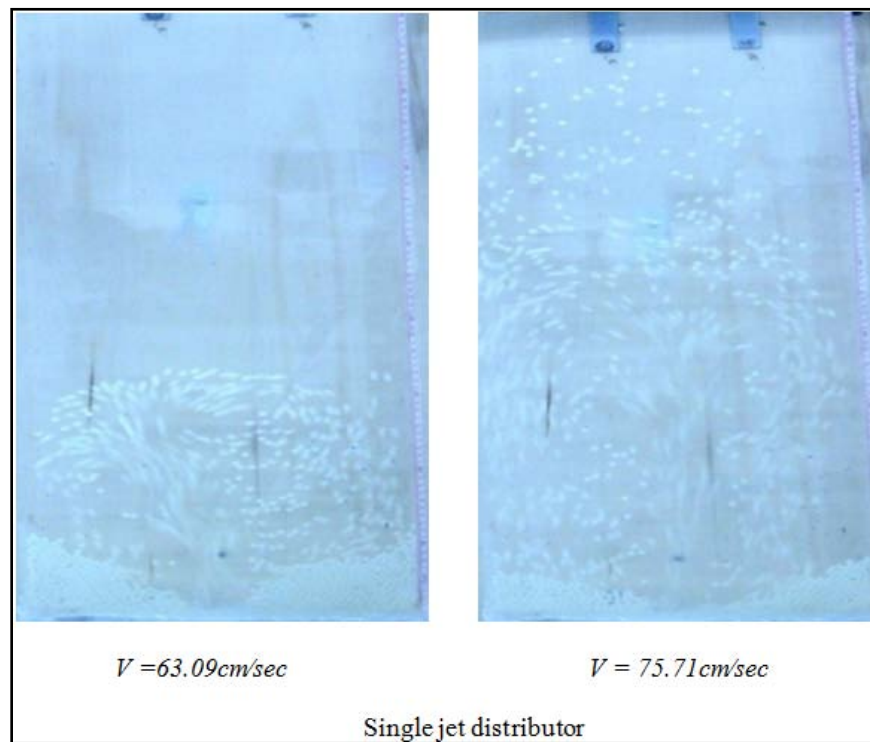


Figure 3. 20: 1000 nylon particles in a single jet distributor

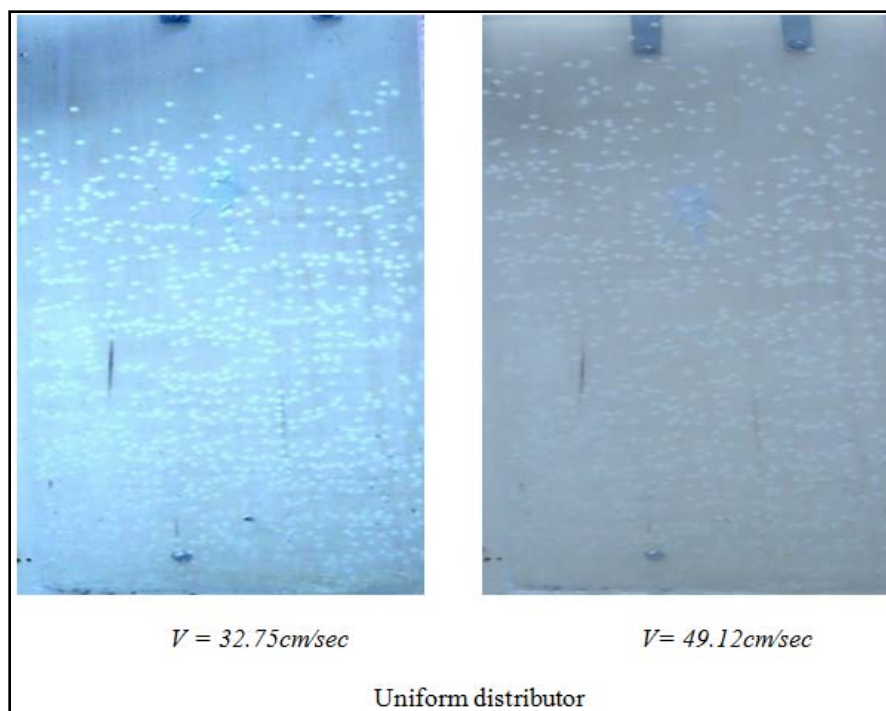


Figure 3. 21: 1000 nylon particles in a uniform distributor

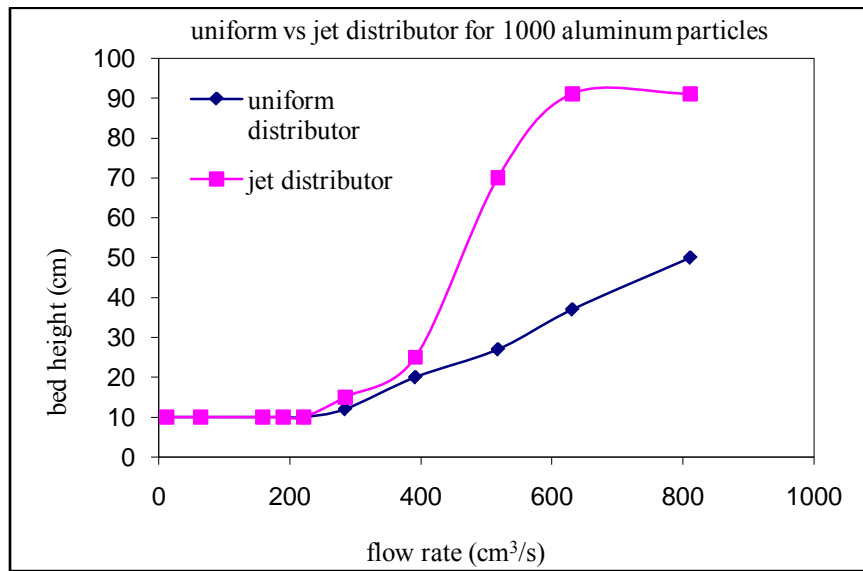


Figure 3. 22: Aluminum (1000 particles)

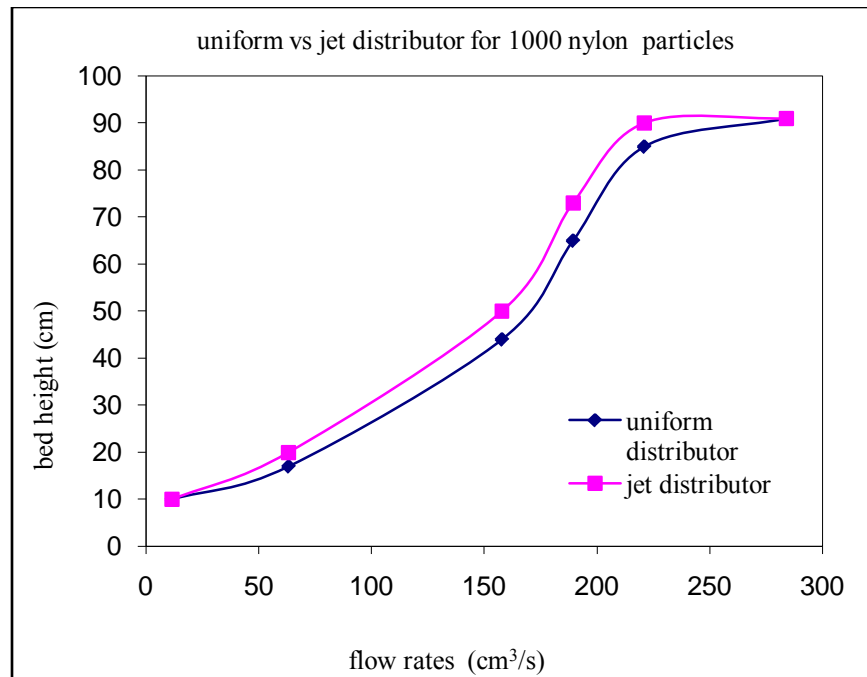


Figure 3. 23: Nylon (1000 particles)

Differences in hydrodynamics were found to exist between the two distributors. Values of minimum fluidization were found to decrease as ratio of area of distributor's orifice to the

cross sectional area of the bed increases. These differences in velocity lie in the size of the orifice.

Particle mixing in the single jet bed is a regular cyclic than the uniform distributor and for the single orifice, the jet was maintained for a certain bed height after the minimum fluidization and then collapses. This is in line with the spoutable bed of Liang-Shih Fan and Chao Zhu [62].

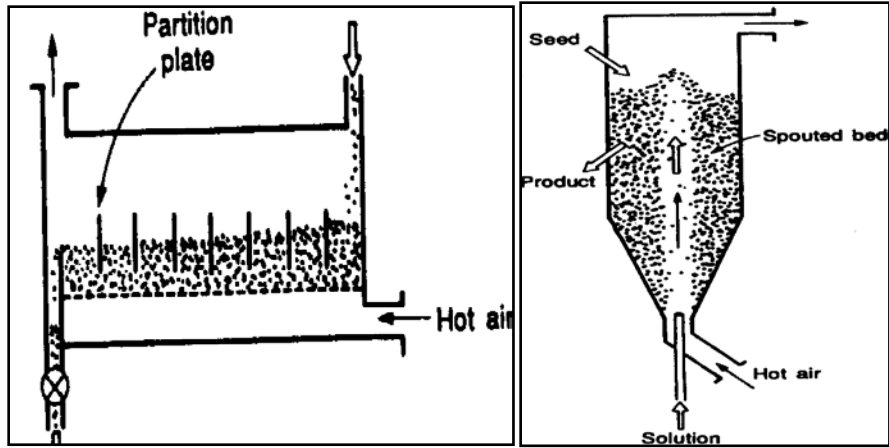
At high velocity, the uniform distributor behaves like a transport bed. To achieve a full bed in the single jet, it requires 1.5 times velocity of the uniform distributor. Application is better in particle drying where uniform mixing is required. The single jet will perform better in coating industries and the biomass industry for the production of oil where exposure time is very important. A correlation equation is derived based on the experimental data, which is expressed as;

$$Y = 1.2e^{0.006x} + 18.8 \quad (3.33)$$

Where x is the fluid velocity and Y is the bed height,

For the single jet fluidization,

$$Y = 1.2e^{0.0069x} + 18.8 \quad (3.34)$$



Drying of materials

Coating of materials

Figure 3. 24: Application of the two distributors

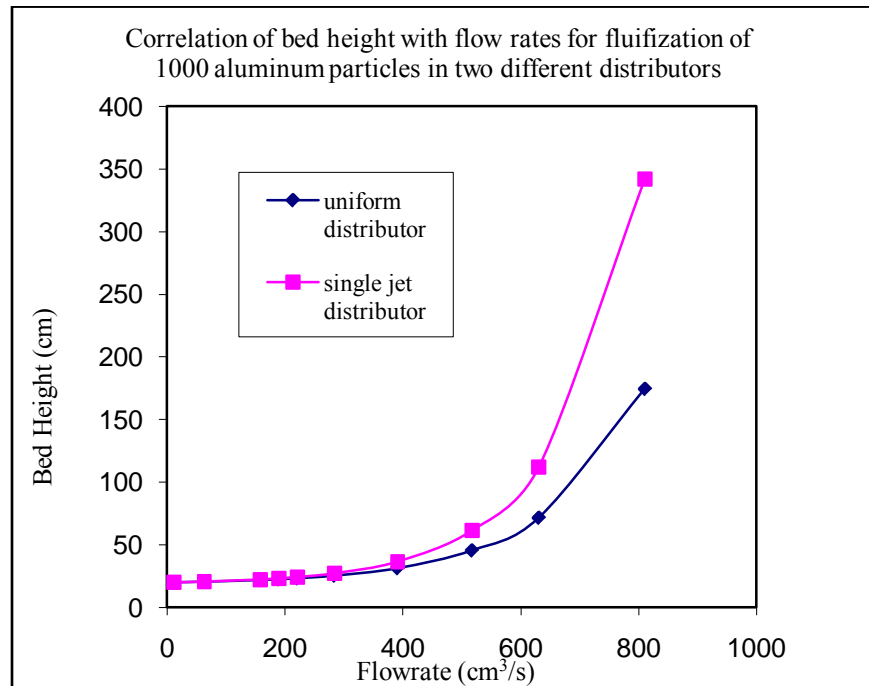


Figure 3. 25: Correlation for 1000 aluminum particles in two different distributors

3.3.3. Mono-disperse vs. Poly-disperse Mixtures at Same Velocity

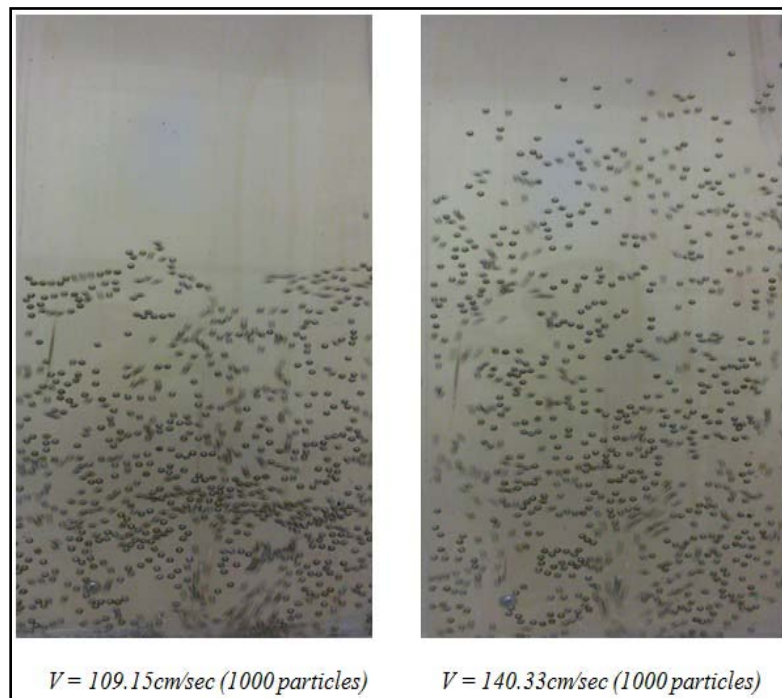


Figure 3. 26: 1000 aluminum particles at different velocities

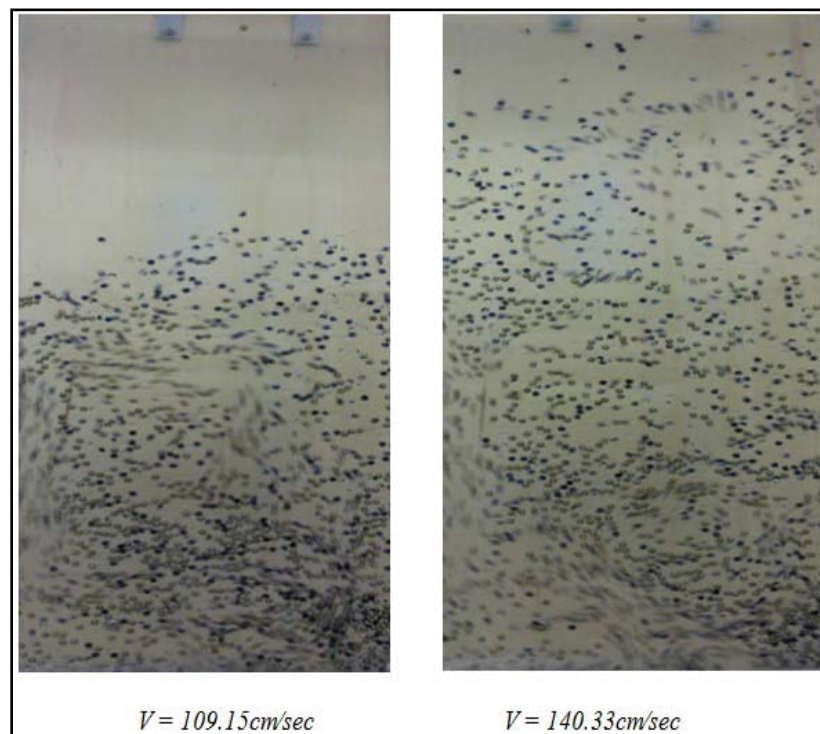


Figure 3. 27: 1000 glass and 1000 aluminum particles at different velocities

Comparing figure 3.26 and 3.27, it shows that the addition of 1000 particles of glass to 1000 particles of aluminum, it is found that the bed heights are very close at same velocity and the effect on the fluidization of aluminum particles is insignificant. This may be due to the fact that the concentration of this bi-disperse component is denser.

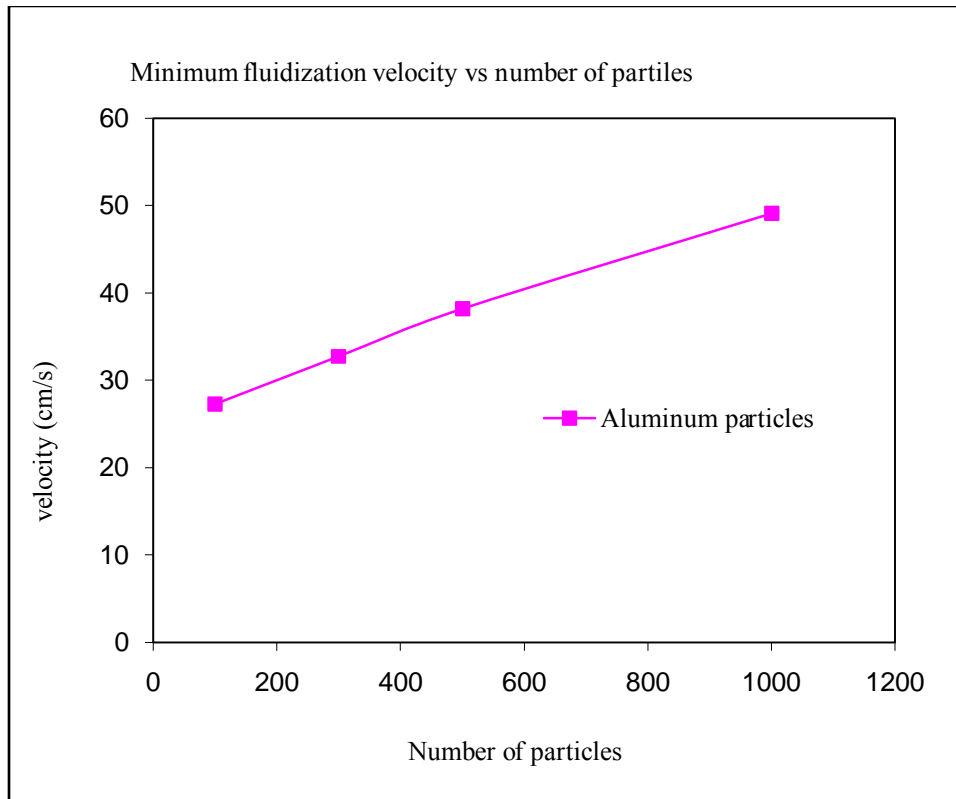


Figure 3. 28: Minimum fluidization velocity vs. number of particles for aluminum spheres

Figure 3.28 shows that a change in the number from 100 to 300 particles will result in ~20% increase in the minimum fluidization velocity of aluminum. A correlation has also been derived for the relationship between the number of particles and minimum fluidization velocity.

$$U_{mf} = 22 + 0.01x + 0.5\sqrt{x} \quad (3.35)$$

Where x represents the number of particles and U_{mf} the minimum fluidization velocity. This correlation is shown in figure 3.29.

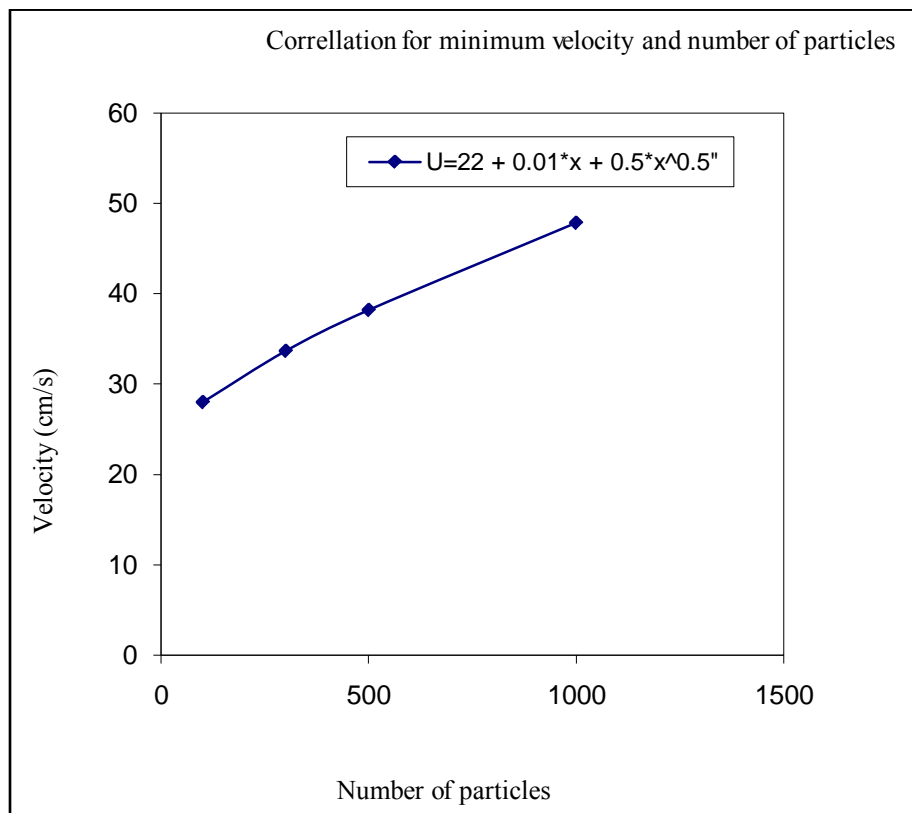


Figure 3. 29: Correlation for minimum velocity and number of particles

CHAPTER 4

CONCLUSION AND RECOMMENDATION

4.1. Achievement

This thesis has succeeded in analyzing the force balance on a single particle, drag coefficient which was applied numerically to solve for the numerical values of the minimum velocity. It also emphasizes on the terminal velocity and compares result with the numerical values of velocity without wall effect. Finally, experimental data was compared with the numerical and analytical values to obtain the wall effect on the particle. This is responsible for the increase in the hydrodynamic drag and to slow down the sphere.

It is rare to have an ideal system of equal density and size particles in a fluidized bed system, hence, mixing and demixing will always occur at some conditions. I have been able to separate the driving force for segregation and mixing in a given system in order to understand and manipulate the process to achieve some desirable qualities of a system.

The fluidization of aluminum-nylon and aluminum-glass is a density-segregation mixture of equal size and different densities. By balancing the respective theoretical drag forces of both particles and adjusting the physical properties of size, I have been able to derive a condition that may favor proper mixing for aluminum and nylon.

Another important finding from my experiments on a single is that the wall effect from front and rear walls reduces the minimum fluidization to a value between 0.4 and 0.5 of the predicted minimum fluidization of the same particle without wall effect. The main reason of reduction is due to the significantly increase of flow velocity near the particle region.

4.2. Conclusions

Experiments were performed in a quasi-three dimensional rectangular column which sits on a distributor, to study the behavior of three different particles (aluminum, glass and nylon spheres) in two different distributors (multi-orifice and a single jet distributor), hydrodynamic properties of aluminum, glass and nylon sphere in a mono-disperse and poly-dispersed. For the bi-disperse fluidization, binary mixture is experimentally simulated by using the same size of particles with different density. Also the behavior of a single particle in a uniform distributor fluidized bed was also studied with emphasis on wall effect to the minimum fluidization velocity.

It is found that the presence of narrow wall effect reduces the fluidization velocity by over 50% for the single particle case. Comparing U_{mf} and U_t in table 3.3, it is clear that the effect of the wall is to increase the hydrodynamic drag and to slow down the sphere [63]. The minimum fluidization of 1000 particles each of aluminum and nylon require roughly twice of the velocity for a single particle. It is also found that the density ratio of binary mixture is the major role for the formulation of segregation. At low density ratio of 0.42, nylon and aluminum particles show a complete segregation (separation) with light nylon particles move quickly to the top region of the bed column and heavy aluminum particles remain in the bottom region. However, at a density ratio of 0.91, glass particles and aluminum particles are able to be uniformly mixed. In other words, it can be said that a 10% density difference of a binary mixture with the same size doesn't cause segregation during fluidization process. Multi-component particles are liable to segregation or mixing when differing in density and particle size. My experiment has shown the prediction for the conditions for multi-component particles to either segregate or mix when their physical properties and composition of the bed are known. It is useful to separate the driving

force for segregation and mixing in a given system in order to understand and manipulate the process to achieve some desirable qualities of a system. A relationship for uniform mixing of binary mixture has also been derived by balancing the respective drag forces and adjusting the physical properties of size, (the diameter ratios of aluminum and nylon from 1:1 to 0.08:1).

The fluidization of aluminum-nylon and aluminum-glass is a density-segregation mixture of equal size and different densities. It is rare to have an ideal system of equal density mono-sized particles in a fluidized bed system, hence, mixing and demixing will always occur at some conditions.

Segregation and mixing can generally be either undesirable phenomenon to be prevented or a desirable process to be emphasized but in whichever case, careful understanding of particle segregation is a requirement for proper operation and design of a fluidized bed. For example, mixing and segregation is very important in industrial application as in the gas-phase polymerization, a system where smaller particles (catalyst) are introduced in the bed to react with the monomer gas to produce polymers with a higher size distribution. Segregation plays an important role when the fully grown polymers will settle in the bottom of the bed and collected from the reactor and particles with smaller size distribution and catalyst continuously migrate to the top for further reaction with the monomers

Also, in the fluidized bed biomass gasifier for the production of hydrogen which contains two binary mixtures of biomass (fuel) and a catalyst. The fuel travels to the top of the bed and the jetsam catalyst falls to the bottom. This is because of large differences in size and density of these two components. These differences in physical properties tend to promote segregation and leads to non-uniformity of the fuel particle distribution. With improper mixing of biomass and

the catalyst in the bed, volatiles are released in the rotating freeboard as the biomass floats to the top of the bed and the energy required for heating the fluidized steam may not be achieved.

At high velocity, the uniform distributor behaves like a transport bed. To achieve a full bed in the single jet, it requires 1.5 times velocity of the uniform distributor. Application is better in particle drying where uniform mixing is required. The single jet will perform better in the biomass industry for the production of oil where exposure time is very important

4.3. Recommendation

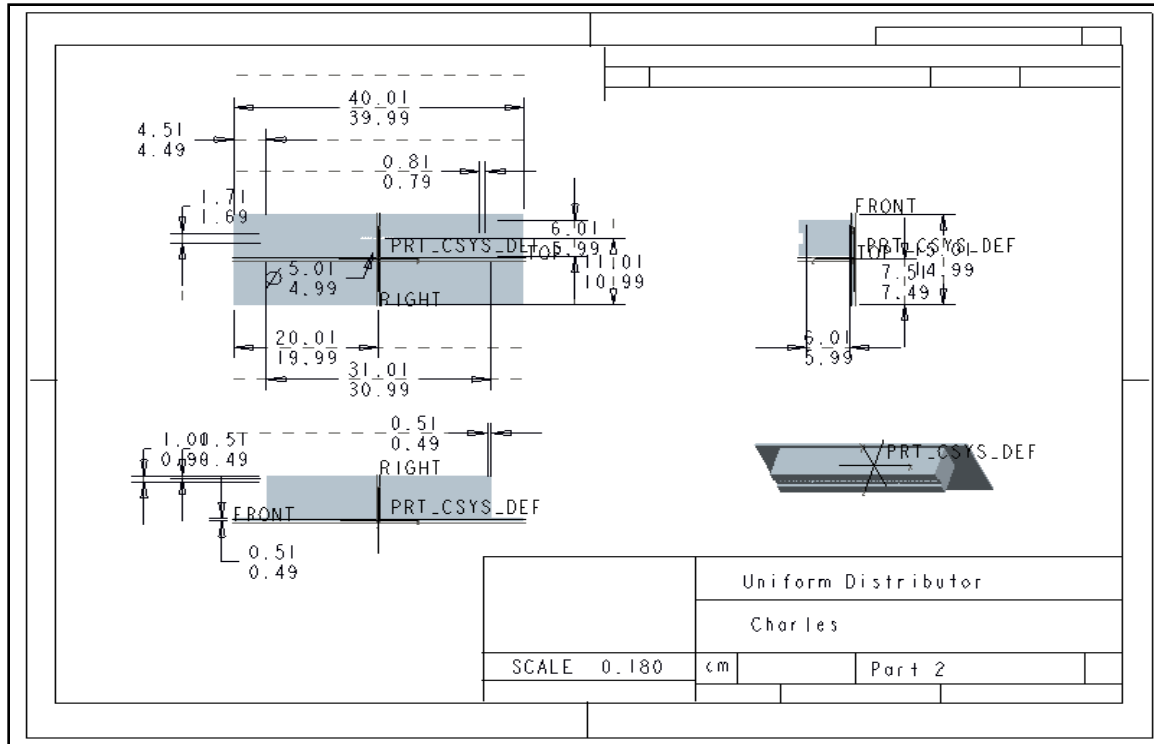
The experiment to determine the relationship between the number of particles and minimum fluidization velocity between 1000 and 2000 particles is not clear. A more sophisticated measuring instrument will be needed to obtain a more accurate result. Also more study is needed to verify the theoretical data from many diameter ratios of 0.08:1. This was obtained from a single particle analysis. More experimental studies will be required using this ratio on a multi-particles system in order to be able to ascertain the behavior of these particles under fluidization conditions.

Also, further experimental studies are required to validate the analytical comparison I made for all cases of the ratio of the particle to the width of the bed to determine the minimum fluidization velocities in larger fluidized bed systems.

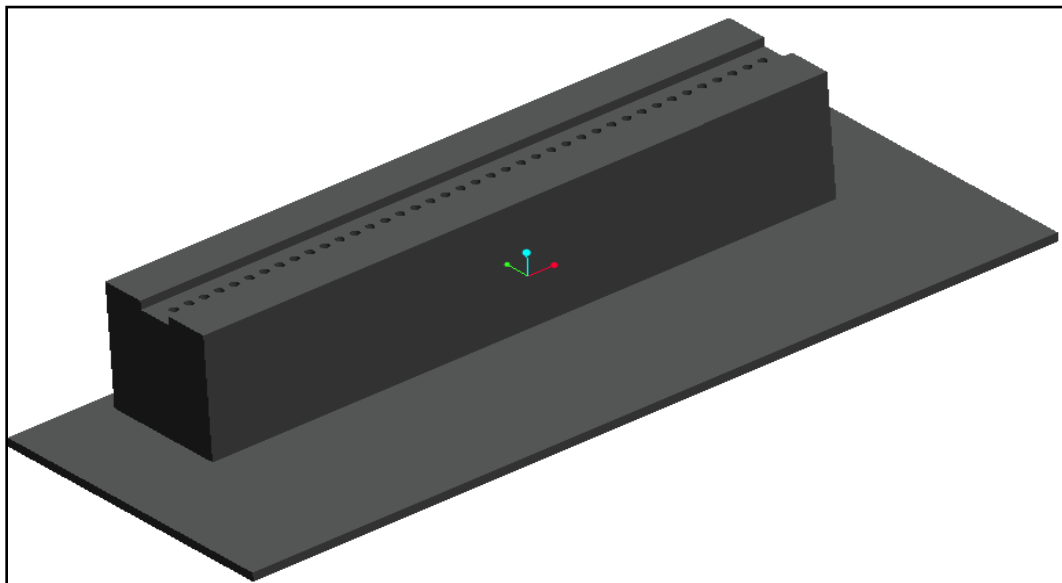
During the design of this fluidized bed, some errors which I encountered led to a better understanding and knowledge of this design. The size and height of the bed should be commensurate with the output of the pump. This will help achieve a more data from the experiment. With a bigger bed size, higher bed height would be achieved and result will be clearer to interpret and higher flow velocities can be reached with respect to bed height. Better mixing and particle segregation could be reduced to some extent.

Material selection should be a major issue for structural strength and rigidity of the entire design with consideration on plate thickness of about 2cm to prevent buckling and prevent bed leakage.

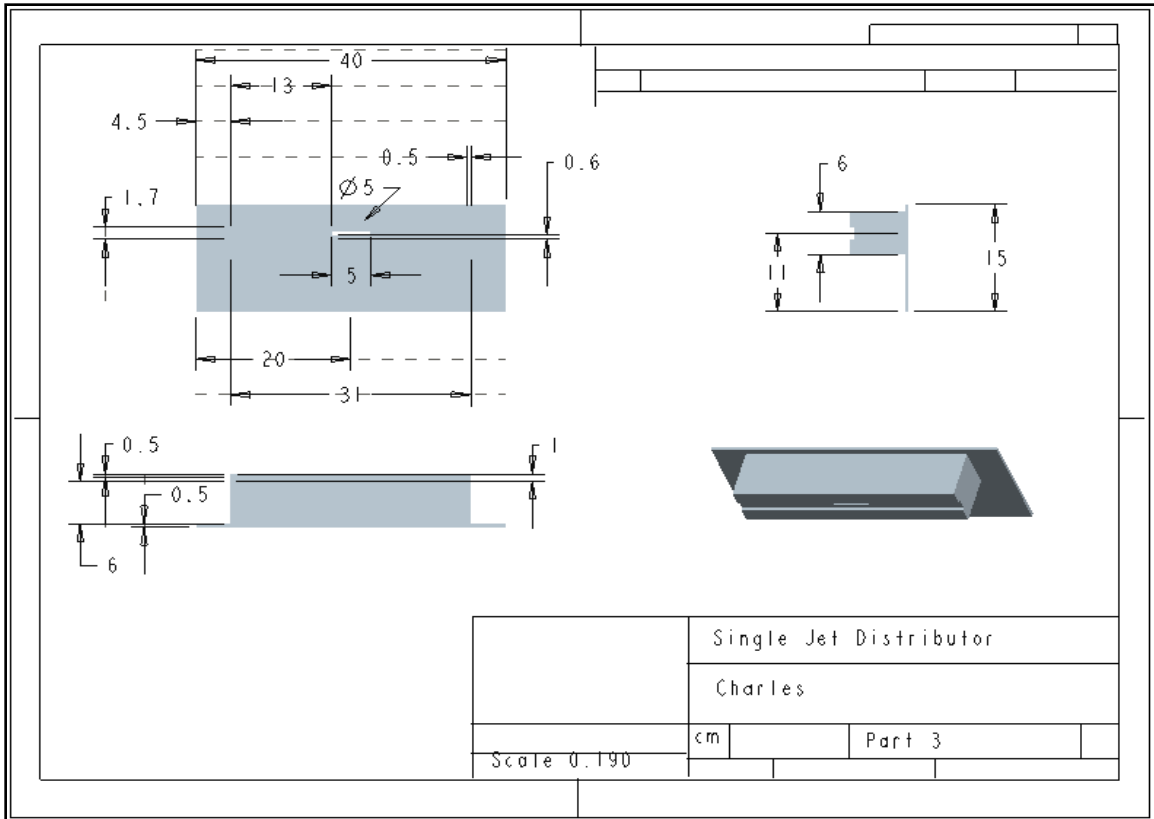
APPENDIX



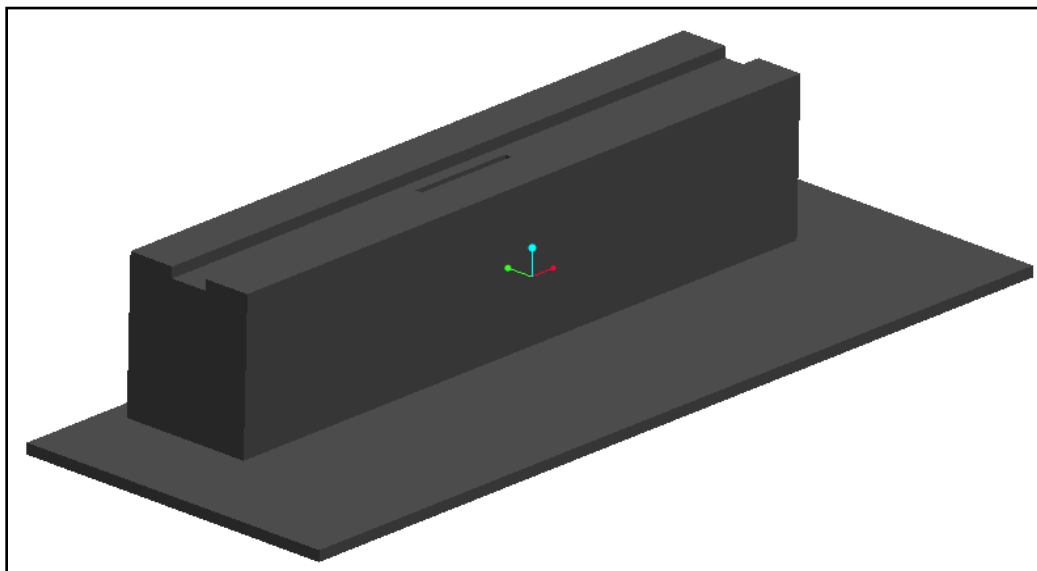
Working Drawing for the Uniform Distributor.



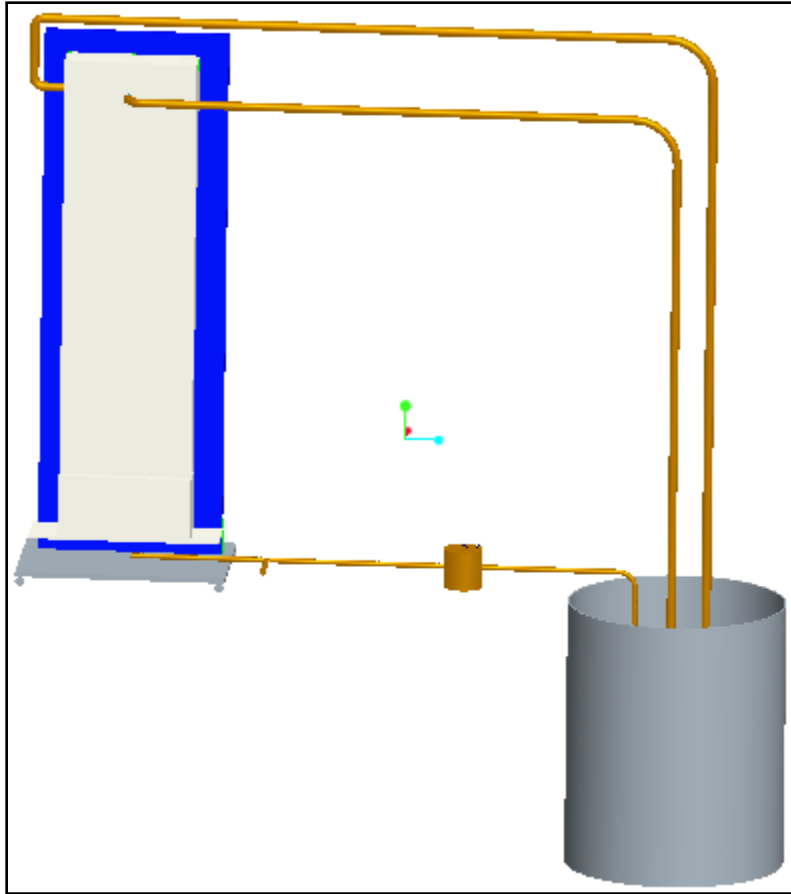
: Uniform Distributor



: Working Drawing for the Single Jet Distributor



: Single Jet Distributors



: The Full Assembly of the Fluidized Bed.

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