Design of a Multi-Staged Swirling Fluidized Bed

By

Sameer Cassim

Dissertation submitted in partial fulfillment of
the requirements for the
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(Mechanical Engineering)

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Universiti Teknologi PETRONAS
Bandar Seri Iskandar
31750 Tronoh
Perak Darul Ridzuan

CERTIFICATION

CERTIFICATION OF APPROVAL

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A project dissertation submitted to the
Mechanical Engineering Programme
Universiti Teknologi PETRONAS
in partial fulfillment of the requirements for the
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Approved by,

PROF. DR. VIJAY R. RAGHAVAN

UNIVERSITI TEKNOLOGI PETRONAS

TRONOH, PERAK

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CERTIFICATION OF ORIGINALITY

This is to certify that I am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

SAMEER CASSIM

ABSTRACT

The topic dealt with in this Final Year Project is the "Design of a Multi-Staged Swirling Fluidized Bed". Fluidization is defined as an operation through which fine solids are transformed into a fluid like state through contact with either a gas or a liquid. Circulating fluidized beds are used mostly in the Chemical process industry, Mineral processing, pharmaceutical production, energy related processes and catalysts also for drying. The aim of the project is to combine multi-staging with counter-flow operation in the novel Swirling Fluidized Bed. The research done entails deep theoretical encounters with acknowledged and published papers regarding the stated topic. Ergun's equation is the co-relationship between all the above parameters over the distance of a packed column. One main finding states that by modifying the fluidizing pattern, it is possible to improve fluidization quality and reduce elutriation simultaneously without the need of auxiliary equipment. The main design aspects were analysed by the author from various recent articles and theoretical understandings from hand books. Hydrodynamics of the novel swirling fluidized bed was studied as well as aspects from the annular spiral distributor. Most part of this project deals with the conceptual design of a multi-stage swirling fluidized bed. The prototype will have three stages. The heights of each stage decrease with each stage so as to cater for pressure drop and air flow resistance. The downcomer-outlet has a zero angle of declination; relying on centrifugal forces for particle flow thus it is mounted tangentially on the bed wall. The downcomer-inlet enters each stage extended as close to the cone as possible for longer processing time and to prevent damage to the distributor on impact. A multi-stage swirling fluidized bed model was designed in CATIA and will be fabricated and experimented in future.

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With complete humility I express appreciation to all that encouraged the completion of this project. First and foremost, for the bountiful knowledge, time and strength blessed upon me, I thank The Al-Mighty God and The beloved Prophet [S.A.W] for tranquility, order and belief in good. Herewith I would like to express my gratitude to those parties who were directly or indirectly involved in helping me achieve.

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I would like to dedicate this project as a token of love and appreciation to Sherla Amreen Dawood, her sacrifices and endless support has allowed me to gauge for success throughout life. For the opportunity to acquire this knowledge I give sincere thanks to my parents and sisters. Last but not least, I would like to thank all my friends, my lecturers, and family for their continued support and motivation.

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ABBREVIATIONS AND NOMENCLATURE

A	Area (cross-sectional)
D	Bed diameter
H	Bed height
ϵ	Bed voidage
$\mathbf{F}_{\mathbf{Drag}}$	Drag force on particle
$\boldsymbol{F}$	Force
u_f	Fluid velocity
$\mathbf{W}_{\mathbf{Bed}}$	Gravitational force on particle
u_t	Minimum fluidization velocity
u_p	Particle Velocity
$\boldsymbol{P}$	Pressure
R	Ratio of bed height to bed diameter
A_p	Surface area of particle
A_{sp}	Surface area of the equivalent-volume sphere
ϕ_p	(surface area of volume – equivalent sphere)/ surface area of particle
SFB	Swirling fluidized bed
U	Velocity
V_p	Volume of a single non-spherical particle

CHAPTER 1:

INTRODUCTION

Fluidization is a renowned technique used for contacting solids and fluids with many applications. Since the late 1970s there has been great outburst for circulating fluidized beds and the main reasons being the environmental advantages mostly in the petro-chemical industry.

New technology is needed in order to meet the requirements of today's modern age. This chapter is dedicated to introduction and explanation of the project topic,"Design of a Multi-Staged Swirling Fluidized Bed". A background about this Final Year Project is given followed by the problem statement to be addressed by the author and lastly the objectives and scope of the work are pointed out.

1.1 Background of Study

Fluidization is defined as an operation through which fine solids are transformed into a fluid like state through contact with either a gas or a liquid. It can best be described by means of a simple experiment where, particles such as sand are poured into a tube provided with a porous plate distributor. Gas or liquid is then forced upward through the particle bed. This flow causes a pressure drop across the bed, and when it becomes sufficient enough to support the weight of the particles it is said to be at minimum fluidization. [1, 11]

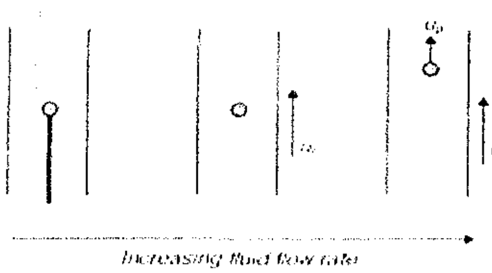

Figure 1 : Single Particle Suspension

Consider a solid sphere sitting on a small support in a vertical tube (Figure 1). A fluid is pumped up the tube so that the sphere experiences an upward force. The point at which the fluid flow balances the spheres weight is the critical point or minimum fluidization point (fluid velocity u_t) at this point the support structure can be removed and the sphere will remain stationary. If the fluid flow rate is increased beyond the critical value, the magnitude of the force becomes greater than the gravity and thus causing the sphere to accelerate upwards. As it does so the velocity relative to the fluid decreases until it reaches the critical value at which the gravitational force is again balanced, the sphere will now move upward in equilibrium at constant velocity. [1, 11]

Whereby : $u_f - u_p = u_t$

Now the constant velocity u_p: $u_p = u_f - u_t$.

Particles under the "fluidized" state experience a gravitational pull which is counter-acted by the fluid drag on them. Thus the particles remain in a semi-suspended condition. A fluidized bed displays characteristics similar to those of a liquid, below is an example of the fluidized state (Figure 2). [1, 11]

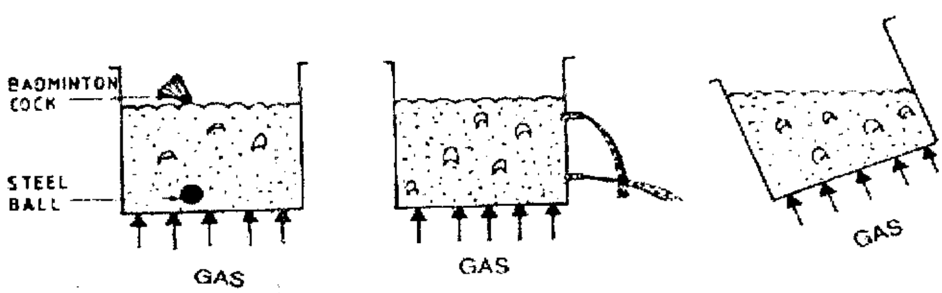

Figure 2 : A fluidized bed demonstrating all the characteristics of a fluid

The characteristics are described as follows:

- The static pressure at any height is approximately equal to the weight of bed solids per unit cross section above that level. $\boldsymbol{P = F.A}$
- The bed surface maintains a horizontal level, irrespective of how the bed is titled; also the bed assumes the shape of the vessel.

- The solids from the bed may be drained like a liquid through an orifice at the bottom or on the side.
- An object denser than the majority of the bed will sink, while the one lighter than the bed will float (e.g. a steel ball sinks in the bed, while a light shuttlecock floats on the surface.)
- Particles are well mixed, and the bed maintains a nearly uniform temperature throughout its body when heated.

An increase in the gas velocity through a bed of granular solids brings about changes in the mode of gas-solid contact in many ways. With changes in gas velocity the bed moves from one state to another. [1, 11]

These states arranged in order of increasing velocities are:

- Packed bed (fixed).
- Bubbling bed.
- Turbulent bed.
- Fast bed.
- Transport bed (pneumatic or entrained bed).

Figure 3 presents a diagram illustrating the presence of those states in different types of boilers.

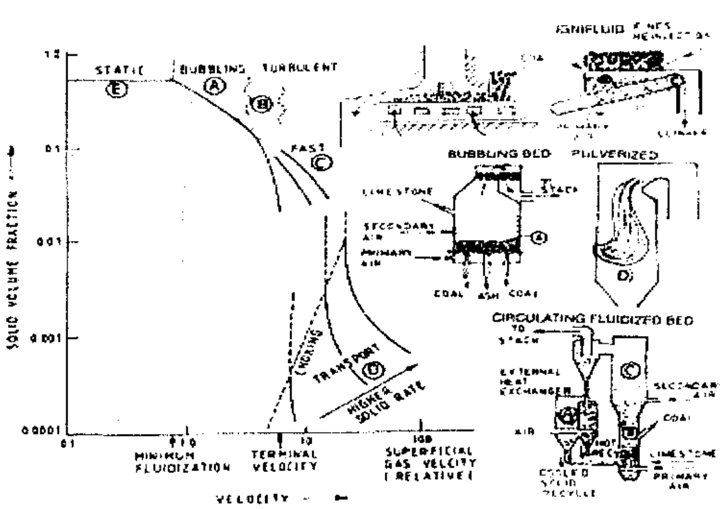

Figure 3 : Different commercial combustion systems under different flow regimes

There are many different types of fluidization with more being innovated with time in the changing modern age. There is the standard homogeneous fluidization which is used commonly throughout the petro-chemical industries. Gas fluidized beds are widely used in combustors for steam. However, the author has chosen the swirling fluidized bed as it is one of the most efficient and energy saving types. (Studies on the swirling state credited to researchers – view Chapter 2 Literature review) [1, 11]

The uses of fluidized beds are generally for:

- Mostly, the interaction of solids and fluids.
- Drying.
- Manufacturing.
- Chemical processing.
- Torrefaction.
- Energy efficient method using air as medium and power from a Blower.
- Used to produce gasoline and other fuels, along with many other chemicals.

Circulating fluidized beds are used mostly in the Chemical process industry, Mineral processing, pharmaceutical production, energy related processes and catalysts also for drying. Understanding and deficiencies of fluidized contacting and efforts to overcome them can lead to successful operation of difficult systems.

1.2 Problem Statement

In industrial processes involving gas-solid contact, it is important to ensure correct residence times for optimal performance and maximum yield in the process. To design such a system, the principles of multi-staging are useful. In addition, counter-flow operation is required for effective use of the driving potential. The aim of the project is to combine multi-staging with counter-flow operation in the novel Swirling Fluidized Bed.

1.3 Objective and Scope of Study

Objectives of this final year project are:

1. To apply the concept of serial processing to the swirling fluidized bed (multi-stage).
2. Improving fluidization processing by enhancing quality of processing with a gain in residence time.

The scope of the study is:

1. Design and construct a counter-flow, multi-staging Swirl Fluidized Bed.
2. Base the design on the study of Swirling Fluidized Beds and model a system that is energy, time and cost efficient.
3. Apply the design to software for analysis and fabricate a Multi-Staged Swirling Fluidized Bed.
4. To study and characterize the behavior of the promising Swirling Fluidized Bed in multistage mode of operation.

CHAPTER 2:

LITERATURE REVIEW

Basically during the course of this study, this project starts with researching about the concept of fluidization as a whole and then slowly approaching the desired Swirling Fluidized bed. Then research done is directed toward multi-staging and the need thereof. This entails deep theoretical encounters with acknowledged and published papers regarding the stated topic. Since there are many new advances within Fluidization the author had chosen the most specified (v.i.z. Multi-staging, Circulating Fluidized beds).

2.1 Fluidization Concept

2.1.1 Ergun Equation

The simplest form of representing the fundamental energy balance of the particles at the point of fluidisation is: $\boldsymbol{F_{Drag} = W_{Bed}}$ where $\mathbf{F_{Drag}}$ is the drag force exerted by the fluid on the particles and $\mathbf{W_{Bed}}$ is the gravitational force exerted on the particle bed.

However, the mathematical relationship of how the particles act in the fluidised state is far more vast than the above equation. Pressure drop is the determining value in order to achieve reasonable fluidisation. As mentioned before, the pressure drop is the product of the force exerted with the area occupied by the bed. This initially would mean that the pressure is dependent on the velocity of the fluid, as velocity increases (above minimum fluidisation velocity) on the contrary the effect on the pressure would be significantly low as it is now affected by the weight of the particle bed.[2]

For the correct derivation, the above items need to be combined to give us a basis on one of the governing equations relating the pressure drop to the weight of the bed or particles. According to Dr K. Sasi *et al*, the onset of fluidisation occurs when:
(pressure drop across bed) x (cross-sectional area of bed) = (Volume of bed) x (fraction of solids) x (Specific weight of solids)

He, as well as other well known researchers in this field agrees that the above equation lacks many parameters for accurate results and interpretation. The following factors need to be considered as well:

- Rate of fluid flow
- Viscosity and density of fluid
- Closeness and orientation of packing
- Size, shape and surface of particles

Ergun's equation is the co-relationship between all the above parameters over the distance of a packed column. [2]
In earlier times it was observed that pressure drop was proportional to velocity (U) at low flow rates, and proportional to the square of the velocity (U^{2}) at high flow rates. Osborne Reynolds was the first to formulate the resistance by friction on the motion of fluid as the sum of these two conditions.

Later on, Carman and Kozeny found that for viscous flow, the change in pressure was proportional to $(1-\varepsilon)^2 / \varepsilon^3$. It was also experimentally determined, by way of 640 experiments, that a constant of 150 was also a factor in the equation. These findings resulted in the Carman-Kozeny equation for change in pressure under viscous flow.

At the same time, Burke and Plummer discovered that change in pressure at turbulent flow, resulting from kinematic energy loss, was proportional to $(1-\varepsilon)/\varepsilon^3$. There was also a constant of 1.75 found to be relevant at the turbulent flow, resulting in the Burke-Plummer equation for change in pressure at turbulent flow.

It was Ergun and Orning that put these two equations together, and through much experimentation found that it was accurate for a wide range of Reynolds numbers. [2]

A tabulated form of the formulas derived follows:

Table 1 : Breakdown of Ergun Equation

PERSONS	FORMULA	CRITERIA RESOLVED
Osbourne Reynolds	$\frac{\Delta P}{L} := aU + b\rho U^2$ where a and b are representative of packing and fluid properties	Resistance by fluid flow friction
Carman and Kozeny	$\frac{\Delta P}{L} := \frac{150 \cdot (1-\varepsilon)^2 \cdot U\mu}{\varepsilon^3 \phi^2 (D_p)^2}$	Change in pressure for viscous flow
Burke and Plummer	$\frac{\Delta P}{L} := \frac{1.75 \cdot (1-\varepsilon) \cdot \rho \cdot U^2}{\varepsilon^3 \cdot D_p \cdot \phi}$	Change in pressure in turbulent flow
Orning and Ergun	$\Delta P := \frac{150 \cdot U \cdot \mu \cdot (1-\varepsilon)^2 \cdot L}{\phi^2 \cdot Dp^2 \cdot \varepsilon^3} + \frac{1.75 \cdot \rho \cdot U^2 \cdot (1-\varepsilon) \cdot L}{\phi \cdot Dp \cdot \varepsilon^3}$	All criteria combined. Φ is the particle shape factor

with $\Delta P =$ the pressure drop,

$L =$ the height of the bed,

$\mu =$ the fluid viscosity,

$\varepsilon =$ the void space of the bed,

$u_0 =$ the fluid superficial velocity,

$d_p =$ the particle diameter

and $\rho =$ the density of the fluid.

2.1.2 Particle Characterization and Dynamics

The study of the particles is very important. Every aspect of the particle needs to be taken into consideration as proven above. This means the size, shape, density and surface type all need to be properly defined.

The size is the linear dimensions of a particle like how a sphere is characterized by diameter and is estimated to the closest possible outcome of the particle as a sphere. Natural and man-made solid particles occur in almost any imaginable shape. Many factors were imposed to describe the non-spherical shapes of particles. In order to define the shape of a particle one needs to know the following parameters:

- Volume of the particle
- Surface area of the particle
- Projected area of the particle
- Projected perimeter of the particle.

In order to best describe shape, Leva (1959) defined the particle diameter of an arbitrary shape in terms of sphericity by :

$$d_p = {}^{6V_p}\!/_{A_p \phi_p} = {}^{6V_p}\!/_{A_{sp}}$$

Where; ϕ_p = (surface area of volume – equivalent sphere)/ s. area of particle

V_p = volume of a single non-spherical particle

A_p = Surface area of particle

A_{sp} = Surface area of the equivalent-volume sphere

2.2 Single-Staged Swirling Fluidized Bed

A swirling fluidized bed (SFB) is a bed of rigid particles contained by a cylindrical distributor plate that is rotating rapidly about its axis. The particles form an annular layer at the circumference due to the large centrifugal forces produced by rotation of the bed. The fluid is injected inward through the porous cylindrical wall. The wall distributor allows the fluidizing fluid to flow uniformly into the bed. Unlike gravitationally fluidized beds, the body force in a centrifugal bed becomes a controllable parameter determined by the rotation speed and the bed radius.

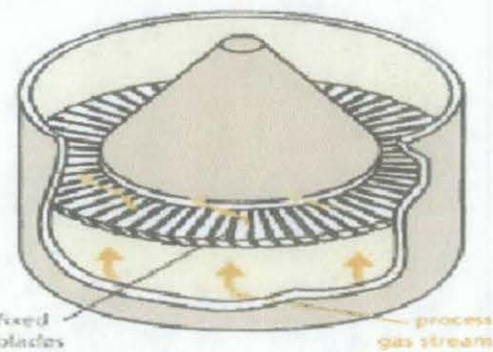

Figure 4 : An example of a swirling bed distributor

By reviewing a study done on the fluidizing pattern by C. Chyang and Y. Lin (2002), they concluded that an innovative swirling fluidizing pattern generated by a multi-horizontal nozzle distributor has been proven to produce a remarkable improvement on the fluidization quality and reduce elutriation.

Their analysis shows that with an increasing distance above the distributor the bubbles are distributed more evenly cross-sectional due to the decay of centrifugal forces. Thus comparing the conventional axial fluidization, the swirling fluidization pattern exhibits high-frequency and low-amplitude pressure change characteristics. Their main finding states that **by modifying the fluidizing pattern, it is possible to improve fluidization quality and reduce elutriation simultaneously without the need of auxiliary equipment.**

2.2.1 Hydrondynamics and Operating Parameters

As previously mentioned, in general the pressure drop in conventional fluidized beds are related to the weight and the force per cross-sectional area of the bed (apart from the other particle characteristics). However, a major feature that differentiates the swirling bed from a conventional fluidized bed is that, the bed pressure drop in the swirling mode increases with air velocity. A simple explanation for this behavior is that, the pressure drop is proportional to the centrifugal weight of the bed.

With reference to prominent researchers, V.R. Raghavan and B. Sreenivasan, the advantages of a swirling fluidized bed overtake that of the conventional system. The swirling allows for the diminishing of visible bubbles and prevents gas bypassing. Since the path traveled by the gas or fluid it is slightly longer than the height of the bed that means more time spent in the process. Once a particle enters the feed it disperses quickly and thoroughly without the need of bubbles. Large open area fractions and low pressure drops at the distributor can be employed in a swirling fluidized bed without the ill-effects. The arrangement in general caters for a cross-flow process which means stable jets are smothered by particles in the flow.

The distributor of the swirling fluidized bed can vary according to required results. The percentage useful area of the distributor was about 95 in the inclined blade type distributors, while it was 64 in the perforated plate type distributor. When the distributor pressure drops for each of the distributors were compared, the perforated plate type distributor had much higher pressure drop in compared with the inclined blade type. [15] The angle of inclination as well as amount of slits plays a huge role in the success of the process. A simple design of a distributor is basically two semi-circular plates which are welded along the diameter. In each half, gas enters through conical slits at an angle close to the horizontal, but in opposite directions in order to create large scale mixing, which is very useful in fluid bed equipment for drying, cooling or ventilation of powder or granulated products.

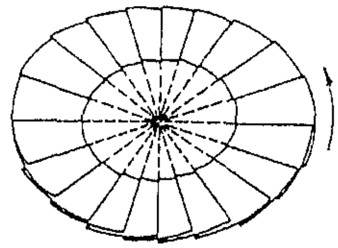

Figure 5 : An annular spiral distributor

After reviewing a report on the modeling of the hydrodynamics of swirling fluidized beds, V.R. Raghavan and B. Sreenivasan (2002) concluded that the superficial velocity and blade angle have greater influence on the swirl characteristics more than the other parameters. Deep consideration needs to be made to overcome all energy and momentum losses when designing a swirling fluidized bed.

The conventional fluidized bed has various limitations such as slugging, channeling and segregation and a high distributor pressure drop. The above mentioned limitations can however be overcome by injecting the gas at an angle to the horizontal and operating the bed in the swirling mode. Merry (1971) studied the horizontal injection of gas jet into a conventional fluidized bed to improve the lateral mixing in a shallow bed. The particles entrained into the jet cause a swirl in the bed and discourage elutriation. [15]

When the inclined blade type distributors were compared, it was noticed that the distributor pressure drop was slightly more in the case of three row type and that was due to the blade holders fixed in between the rows. [15]

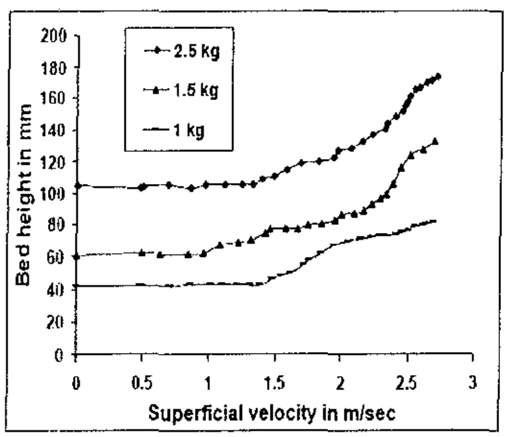

Figure 6: Variation of bed heights of single inclined blade. [15]

The variation of bed pressure drop with superficial velocity in the inclined-blade single-row distributor for different bed weights using plastic beads as bed material is depicted in Figure 6. It reveals that for each bed weight studied, the bed pressure drop was almost constant after minimum fluidizing velocity of about 1.2m/sec. [15]

2.2.2 Studies on Annular Distributor and Bed Behavior

Initially thorough research is done based on fluidization in general. Besides becoming familiar with the general concept of fluidization, it is intended to examine the various flow regimes and their stability in a SFB (swirling fluidized bed).

Recent studies show that the principle of operation is based on the simple fact that a horizontal component of gas velocity in the bed creates horizontal motion of the bed particles. The cyclone-like features resulting from the swirling motion of bed particles also contribute to this low elutriation. Hence it is possible to fluidize very fine particles and a wide variety of shapes of particles in this kind of fluidized bed. [16]

Further on the variation of bed loading and different regimes are indicated as follows:

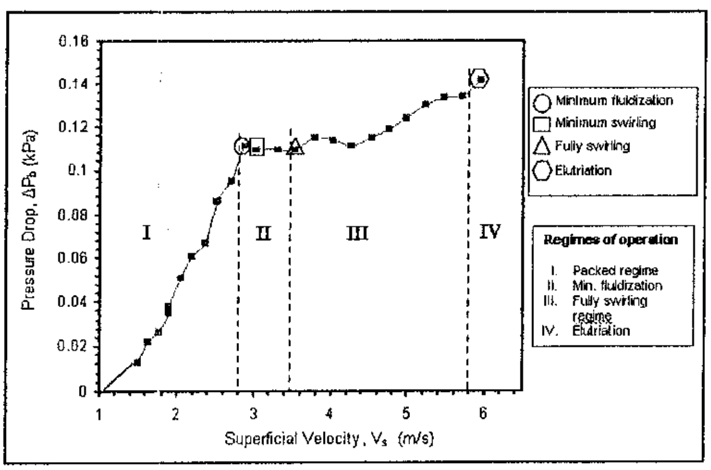

Figure 7: Effect of Variable of Bed Loading [16]

The flow regimes in swirling fluidized bed are packed bed, minimum fluidization, swirling regime, two-layer regime and finally elutriation or transport regime. Thus the bed behaviour can be summarised as:

- The pressure drop of the swirling fluidized bed increases with the mass flow rate of fluidizing gas.
- Larger particles have lower pressure drop and capable of withstanding higher superficial velocity and hence, larger swirling regime.
- Increasing the overlapping angle of the distributor causes air to flow through higher resistances which initiates higher pressure drop. This also reduces elutriation by increasing the swirling region.
- Particle size, bed weight and the number of blades (30 and 60) are the most important variables that have more influence on the bed behaviour. [16]

2.3 Multi-Staging

The idea behind the design of a multi-stage swirling fluidized bed is based mostly on the principle of residence time in the process. Since the residence time plays a huge role in the efficiency of a fluidized bed, by increasing that time and decreasing the amount of work done, by adding multiple stages of fluidization we are able to improve fluidization.

2.3.1 Some of the Reasons for Multi-Staging:

The absence of back-mixing and the large solids residence time per segment offers the opportunity to combine separate processes (e.g. gas–solids reactions), spatially divided, in a single reactor. [17]

- To prevent back-mixing of gas-solids.

  Due to its specific shape, a fluidized bed arises in the bottom cone of each segment and back-mixing of gas and solids between the segments is prevented effectively.

- To enlarge the solids residence time.

  The concept of several fluidized beds operated co-currently in series results in a ratio between the solids residence and the gas residence time being much higher than for a normal Centrifugal Fluidized Bed.

- To provide serial processing for efficient outcomes.

  The concept in which different processes are carried out in separate segments of the same reactor has tremendous benefits.

Continuous fluidized systems have different residence time for solid particles yielding non-uniform product. Despite the serious drawbacks, the compelling advantages of overall economy of the fluidized contacting system have been responsible for its successful use in many industrial operations. Continuous fluidized beds can be operated as a single-stage or multistage system. The limitations of a single-stage continuous fluidizer can be summed up in terms of wide distribution of residence time of solids and low efficiency of operation both with respect to gas and solid phase besides the chances of slugging in deep beds. [19]

2.3.2 Pressure Drop Theory Over Heights

A brief review on past journals was used to estimate what the pressure drop over heights would render. It was found that the pressure drops across different heights were measured to be approximately the same (with variation less than 2%). It was also observed that all the stages were identical in their operation as well as performance. Another noted finding was that the pressure drop due to solids across each in stage has been obtained from the difference between the pressure drop with and without solids and rendered that the pressure drop due to solids decreases with an increase in the gas flow rate and increases with increase in the solids flow rate. [20]

The reasons can be assumed that an increase in gas-flow rate increases the porosity of the bed in the system, resulting in decrease in the solids concentration and hence the pressure drop across the stage, as the height of the fluidized bed in the system corresponds to downcomer weir height. Another reason is that the pressure drop due to solids decreases with an increase in the gas-flow rate and increases with increase in the solids flow rate may be due to less frictional force, inertia and impact forces. [20]

Thus summed up, the maximum pressure drop occurred in the stage at low gas flow rate corresponding to maximum solid flow rate. [20]

2.4 Design Specific Research and Recent Findings

The initial step toward design apart from important varying parameters, is determining its configuration. Mainly, there are two types of configurations called Cross-flow and counter-flow, where:

- Cross-flow which results in a horizontal arrangement.
- Counter-flow which results in a vertical arrangement. The main focus of my study.

According to an article "Identification of stable operating ranges of a counter-current multistage fluidized bed reactor with downcomer" C.R. Mohanty, B. Rajmohana, B.C. Meikap, (2009), multistage contacting of gas and solids can be obtained by simple multiple contacting system; cross-flow contacting system; counter-current contacting system. The counter-current multistage system is an improvement over the cross-flow system.

The main design aspects were analysed by the author from various recent articles as well as theoretical understandings from hand books. Important aspects considered can be summarised into six parts, namely:

1- Operating velocity.
2- Aspect ratio.
3- Distributors.
4- Modeling aspects.
5- Downcomers.
6- Valves.

2.4.1 Operating Velocity

In gas-fluidization the operating velocity is based on the hydrodynamics, heat transfer and reaction conditions. The ideal velocity in reaction or heat transfer processes can differ vastly from that of a cold bed. In order for fluidization one needs to first attain minimum fluidization velocity and that is the fundamental velocity to base calculations on.

The hydrodynamics of the operating velocity plays a role in the stability and success of the operation. To reach stability the range of the velocity should make good gas-solid contact with optimum energy consumption. The safe operating velocity is usually theoretically applied as three to five times the minimum fluidization velocity.

The fluid undergoes two types of resistances during fluidization which cause a pressure drop. If the drop is too large the outcomes of the process would be unsatisfactory. The distributor as well as the bed both form part of the resistance experienced by the fluid. One needs to overcome the resistances by applying the right size chamber, conforming to the area for desired pressure drop.
An important gas velocity is the critical back-mixing velocity responsible for inducing back-mixing. A ratio needs to be determined between the operating velocity and minimum fluidization velocity (U/U_{mf}) . The critical ratio is dependent on the

bed voidage (ϵ) and the ratio (R) which is the volume of the displaced emulsion phase to the bubbling volume given by Stephens et al.

$$({}^{U}/_{U_{mf}})_{cr} = \frac{1+\epsilon R}{\epsilon R}$$

For small values of U_{mf}, $({}^{U}/_{U_{mf}})$ is above the critical ratio, which results in a well-mixed fluidized bed.

According to a recent article, "Identification of stable operating ranges of a counter-current multistage fluidized bed reactor with downcomer" C.R. Mohanty, B. Rajmohana, B.C. Meikap, (2009), the research shows the use of a downcoer to link the various stages in a multi-stage fluidized bed.

At a particular gas and solids flow rate, the solids height in the downcomer was almost constant. With increase in the gas rate, the height of the solids in the downcomers decreased and reached at lowest level and, at a particular gas rate, the height of the solids in the downcomers increased with time indicating instability in solids flow and no transfer of solids from downcomer to bed. The gas velocity at which instability or no transfer of solids occurred is termed as the 'upper operating gas velocity'.

2.4.2 Aspect Ratio

The aspect ratio refers to the ratio of the bed height (H) to bed diameter (D). A fluidized bed is termed as 'deep' when the ratio is above one, and 'shallow' when is it below. Generally a deep bed is better as the bubbling takes longer to travel and chances of reacting are more than a shallow bed, where short-circuiting would prevail.

According to Siegel J.H *et al*, the distributor has major impact on the bed height in terms of complete fluidization. The distributor pressure drop must be changed id the bed height or aspect ratio is changed. If the bed height is increased, the distributor pressure drop should also be increased by increasing the operating velocity or reducing cross-sectional area.

Studies show that as the aspect ratio increases, the distributor to bed pressure drops ratio (*R*) decreases indicating improvement in quality of fluidization and approaching a steady state and as *R* increases with an increase in operating velocity indicating gas channelling to some extent. Thus *R* value steeply increases with a decrease in aspect ratio revealing its influence on the uniformity of fluidization. [19]

2.4.3 Distributors

A distributor is basically a flow restrictor and a bed-supporting device. The roles it plays however are:

- Initiation of fluidization (swirling; inclined-blade spiral distributor)
- Maintaining stable operation.
- Preventing dead or de-fluidized zones of particles.
- Preserving the distributor surface and preventing solids from flowing into the plenum chamber during downtime.
- Minimizing the attrition of solid particles (wearing down by friction) and erosion of bed internals.

It is important to consider the design of the distributor as it affects the hydrodynamics of a fluidized bed. Of the three zones in a gas-solid fluidized bed, the grid zone is affected by the design of the distributor plate. Since this is the base zone, it has large influence over the characteristics other two zones above.
The type of distributor also affects the interfacial area of the bed. Bubbles and airstreams are controlled by the distributor. Thus, it is of high emphasis that the distributor defines the fluidized bed. By considering the different zones in the fluidized bed one can deduce that each zone is caused by pressure drop characteristics. The distributor determines the flow and pressure drop occurrences.

2.4.4 Modeling Aspects

The objective of a fluidized bed model is to combine the parameters and reactions mathematically to arrive at equations to determine the size of the reactor. The diameter of a fluidized bed is determined by the maximum allowable gas flow rate that would keep the bed constant. The height is determined by residence time and operating velocities as well as magnitude of the operation. The fluidized bed should consider all the possible zones caused by different patterns of gas-solid contacting. The three zones discussed are:

2.4.4.1 Grid Zone

This is the base zone as discussed earlier. The distributor directly affects this zone as it is the mechanism of gas entry into the fluidized bed. Depending on the distributor, the way the grid zone behaves affects the entire fluidization process. This is the first objective when designing a fluidized bed. After achieving stability at the grid zone the rest of the bed would be easier to handle.

2.4.4.2 Main Bubbling Bed

The zone just above the grid zone is known as the bubbling bed. It occurs when the gas jet reaches its equilibrium state. This is where the bed tends to get dense and there is no affect by bubbles and jets in the grid zone. The types of flow that occur in this zone are usually assumed to be plug, complete mixing and axially dispersed. Selecting a distinct model for a certain application will have varying zone characteristics.

2.4.4.3 Freeboard Zone

This is the last zone in the fluidized chamber. It is usually at this point where unreacted gas collects as well as entrained solids from elutriation. The split or next stage of fluidization links at this stage due to an assumed plug flow.

2.4.5 Downcomers

The downcomers form the most important component in order to gain multi-staged operation. A review on previous designs was use to estimate a model for the new prototype. The designs considered are:

Adapted from "Pressure drop characteristics of a multi-stage counter-current fluidized bed reactor for control of gaseous pollutants";C.R. Mohanty, B.C. Meikap; 2010. A three-stage counter-current gas–solid fluidized bed reactor has been designed, constructed and investigated stable operating range for different particles and the influence of downcomer for stable operation. Further the role of aspect ratio for uniform fluidization has been critically examined. An attempt has therefore been made to acquire precise knowledge on the dynamics of the distributor and downcomer used in the system and stable operating range of the two-phase flow.

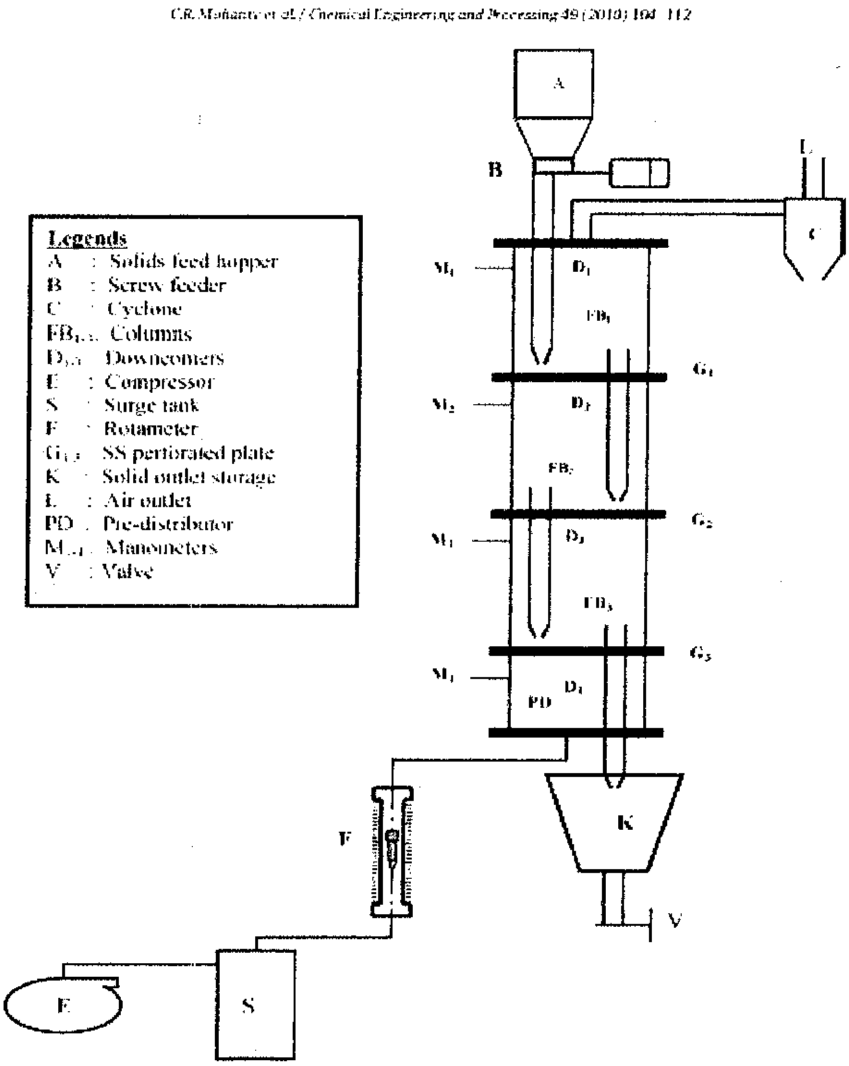

Figure 8 : Basic multi-staged fluidized bed with downcomers

More specifically the idea of the downcomers was selected from this design. A closer look into the design is as follows:

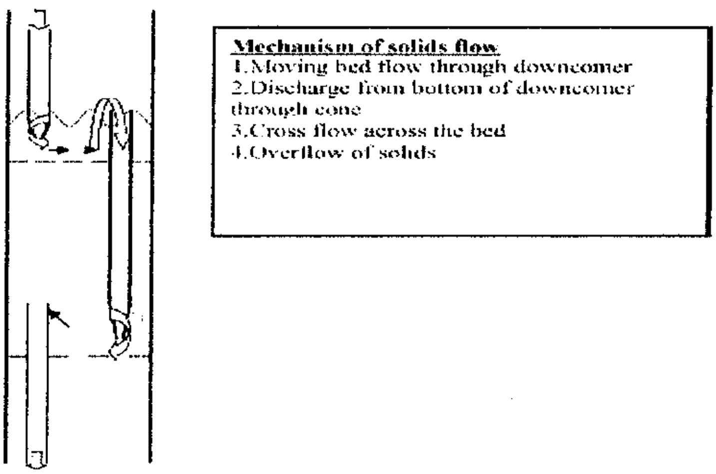

Figure 9 : Downcomers of a multi-staged fluidized bed

Another compatible design is adapted from "Stable operating velocity range for multistage fluidized bed reactors with downcomers." Martin-Gullin, A. Marcilla, R. Font, M. Asensio; 1995:

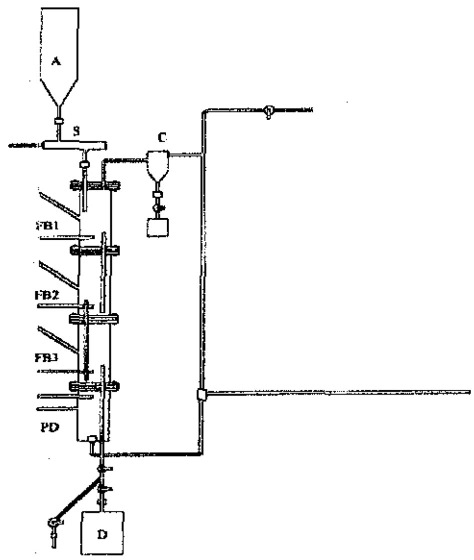

Figure 10 : A multi-stage fluidized bed with downcomers and manometers

From the designs reviewed the findings are that with an increase in the gas rate, the height of the solids in the downcomers decreased and reached at lowest level and, at a particular gas rate, the height of the solids in the downcomers increased with time indicating instability in solids flow and no transfer of solids from downcomer to bed. The decrease in solid concentration in the bed exposed the downcomer bottom to gas flow, thus not allowing solids flow to the next bed indicating flooding of downcomer followed by upper bed. Further increase in gas flow, the downcomer does not withstand the pressure drop making the lower bed emptied. [19]

It is noted that at a particular solids flow rate, the height of the solids in the downcomers deceases when gas flow rate is increased. This may be due to the fact that increasing in gas flow rate decreases the solids concentration in the bed, thereby decreasing the bed pressure drop which is counterbalanced by a decreased solid height to prevent gas flow through downcomer. The height of the solids in the downcomers increases with time indicating flooding of the bed. [19]

2.4.6 Valves

The previous designs reviewed on downcomers showed that flooding or lack of flow of solid particles are prone to downcomers since the entire regime is based on counter-current flow. In order to overcome these complications there is a dire need for valves. There is a possibility that the gas will escape upwards through the downcomer and in fact they will prevent the easy flow of particles downwards. Studies on the L-valve and J-valve concepts were done to determine a combating method for the complication. It involves a short horizontal travel of particles which discourages the bypassing of the main fluidizing gas.

In the valve mode of operation, the solids flow rate through the non-mechanical device is controlled by the amount of aeration gas added to it.The aeration gas is to aid the downward flow of particles by the injection of a small stream of gas in the direction of desired flow of solids. These devices are shown schematically in Figure 11. The primary differences between these devices are their shapes and the directions in which they discharge solids. Both devices operate on the same principle. It is harder to fabricate a smooth 180-degree bend for a typical J-valve. The most common non-mechanical valve is the L-valve, because it is easiest to construct, and also because it is slightly more efficient than the J-valve. Solids flow through a non-mechanical valve because of drag forces on the particles produced by the aeration gas. [21]

The actual gas flow that causes the solids to flow around the L-valve is not just the amount of aeration gas added to the valve. When aeration is added to a non-mechanical valve, solids do not begin to flow immediately. The initial aeration gas added is not enough to produce the frictional force required to start solids flow. Apart from the gas-flow an angle of declination would enhance flow due to gravitational forces. An experimental study was done to determine the best angel for this. [21]

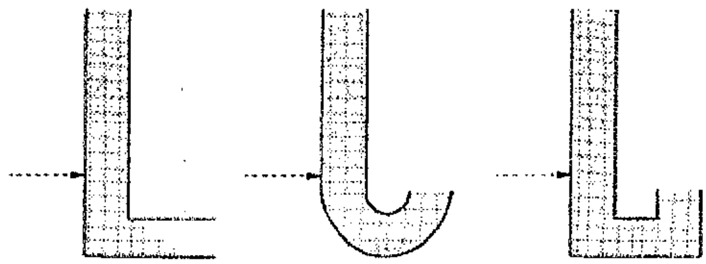

Figure 11: Schematic of non-mechanical valves

2.5 The Concept of Kinetic-Controlled and Diffusion-Controlled Reactions

The quality of fluidization depends on the time a particle spends in the process. For example if the application were to be of a drying nature, the longer a particle spends in the air the more likely it is to be dry. Residence time is the time taken up by a solid in a fluidized state within the fluidized bed.

In addition to the obvious advantages resulting from the particles behaving fluid-like, which permits them to flow freely from one location to another, with an improved residence time high level of mixing occurs, this means that heat and mass can be rapidly transferred throughout the bed. For these criteria a review on kinetic-controlled and diffusion controlled reactions were done.

2.5.1 Kinetic-Controlled Reactions

In kinetic controlled reactions also sometimes thermodynamic reactions, the reaction rate depends on the velocity of fluid in contact with the particles. The conditions of the reaction, such as temperature, pressure, or solvent, affect the reaction. The novel swirling fluidized bed with its advantages play a role in favoring a Kinetic-controlled reaction. However, large amount of energy should be spent if the process has to be done in a single stage.

2.5.2 Diffusion-Controlled Reactions

In diffusion controlled reactions, the reaction rate depends on contact time of a particle and the fluid as well as the fluid velocity. The reactions occur quickly that the reaction rate is the rate of transport of the particles through the fluid. One way is to observe whether the rate of reaction is affected by stirring or agitation, if so then the reaction is almost certainly diffusion controlled under those conditions. Hence particle residence time becomes important because as quickly as the reactants encounter each other, they react. The review indicates that with more stages in a reactor the particles reaction would be considered as if a catalyst were introduced, thus allowing faster reactions after each intermediate stage.

The required design prototype of a multi-staged swirling fluidized bed is considered with the above findings taken into account. Both Factors are covered thus enhancing reactions and quality of the fluidization.

CHAPTER 3:

METHODOLOGY

3.1 Problem Definition and Flow-Chart

The single-stage Swirling fluidized bed has many advantages in the applicable processes which triumphs over other conventional systems. However there is a need for higher residence time and higher fluidization quality, the criteria based on Kinetically controlled and Diffusion controlled systems.

Previously designed multi-stage fluidized beds cater only for centrifugal motion and not the intended swirling state. These designs make use of downcomers made to fit internally within the bed itself, which conflicts with the annular spiral distributor.
The new conventional design requires using existing parameters developed by a study on the swirling fluidized bed. The idea of the design stems from manipulating previous multi-stage designs to coincide with the current single stage swirling fluidized bed.

The following flow chart shows the progression of events that will eventually lead to the completion of the project.

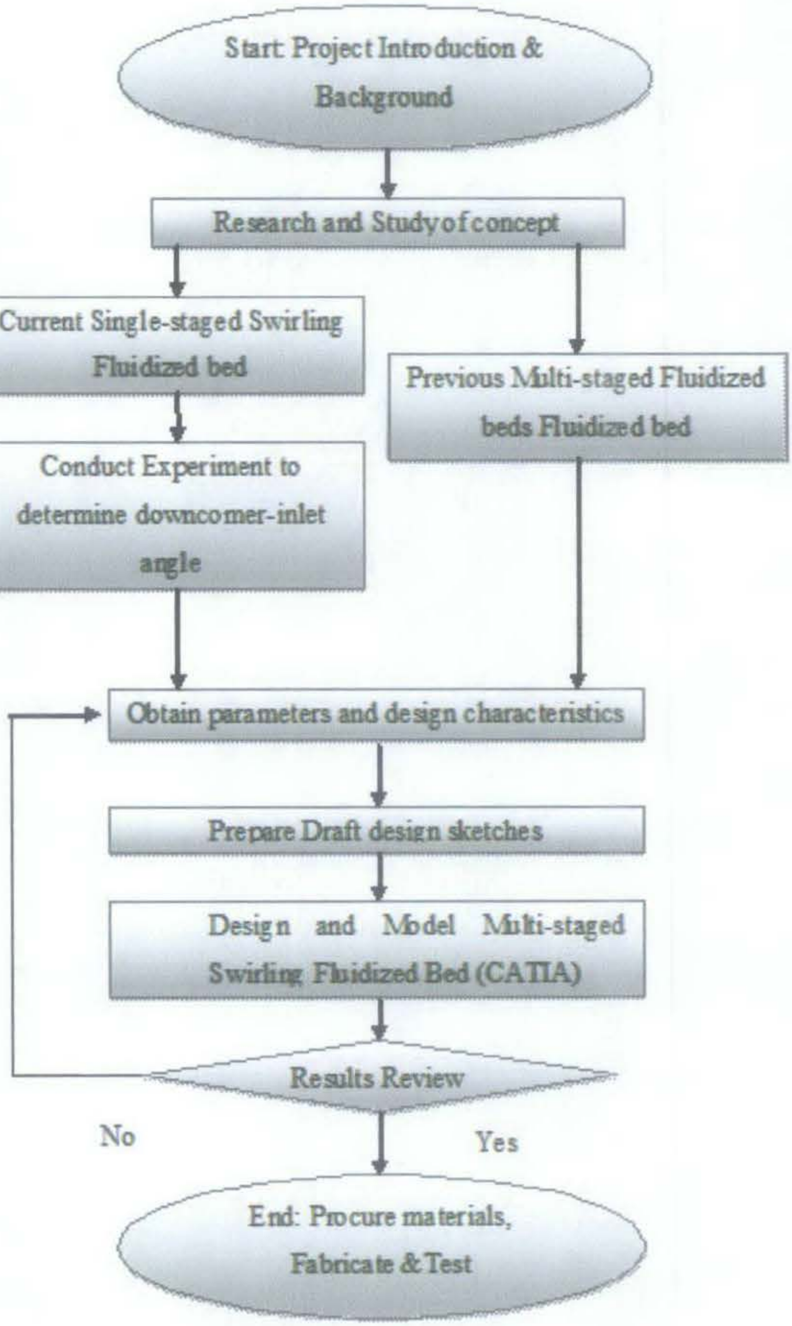

Figure 12: Project Flow Chart

3.2 Project Work

The steps in the methodology of design are as follows:

1. Research and studying the concept:

Research was done on different journals on the study of the concept of fluidization, single-stage swirling fluidized beds and multi-stage conventional fluidized beds. This relates to the study on the hydro-dynamics, residence time and downcomer placements.

2. Obtain parameters and design characteristics from previous design of single stage swirling fluidized bed:

To understand the basis of the design the single stage swirling fluidized bed was broken-down into its main components. These components were duplicated or modified to suit the design. The direction of the swirling fluid-flow maintains a clockwise path.

3. Prepare Draft design sketches:

A basic idea was developed for analysis and critique by the supervisor, graduate-assistant and fabricator. Here the Idea progressed into a workable model for improvements.

4. Conduct Experiment to determine downcomer-inlet angle:

A hopper was designed to simulate the downcomer-inlet at various stages which was tested on the single-staged swirling fluidized bed. Here the hopper was designed in two parts with the bottom part made at four different angles of declination. (0, 15, 30 and 45 degrees)

5. Familiarize with design software (CATIA):

Get familiar with CATIA to sketch full design with components.

6. Design and Model Multi-stage Swirling Fluidized Bed:

Using the parameters and constraints of the previous design and analysis of draft design sketches a full 3D design is modeled in CATIA. Here angles and locations of slots of components were established.

7. Procure design materials:

According to the designs, components are made to specifications that related to material available in the market. Procuring material from manufacturers was done with minimal deviation from design specifications so that least mechanical work was carried out to prevent any faults.

8. Fabricate prototype and test:

The last part is the most critical as it involves the actual design process as a whole. Designs would have to be tested or have firm theoretical standing. A particular set of variables (type of particles, weight, residence time, etc.) should be set in order to compare results for further analysis.

3.3 Component Break-Down of Single Stage Design and Experimental Set-Up

In order to design a Multi-staged Swirling Fluidized Bed, the concept and basis was done by analyzing the pre-existing model and its components. For this a break-down of the prototype was done. Analysis revealed:

a) The prototype bed wall is a Perspex Cylinder
b) The distributor is an annular spiral distributor made of aluminum.
c) The plenum chamber is metal and allows for smooth flow regime entry into the bed
d) A flange is welded to the plenum chamber at the hole in order to connect the chamber to the pipes.
e) The chamber is connected to the blower with PVC pipes

f) A hollow metal cone was introduced in the centre of the distributor. This prevents Dead zones and allows for particles to be directed away from bed-surfaces.

Below is a schematic drawing of the set-up with a picture of the prototype (see Appendix A):

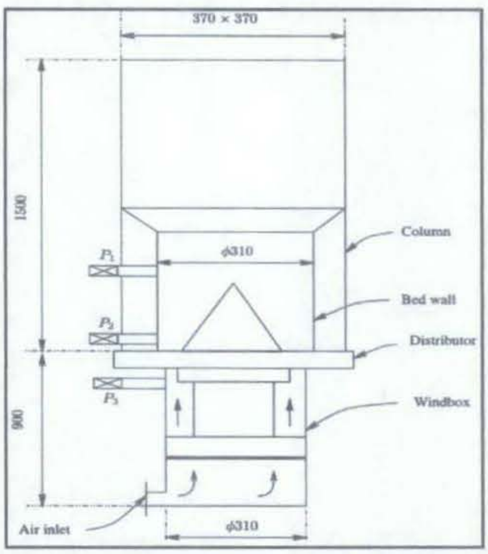

Figure 13: Schematic of Novel Swirling Bed

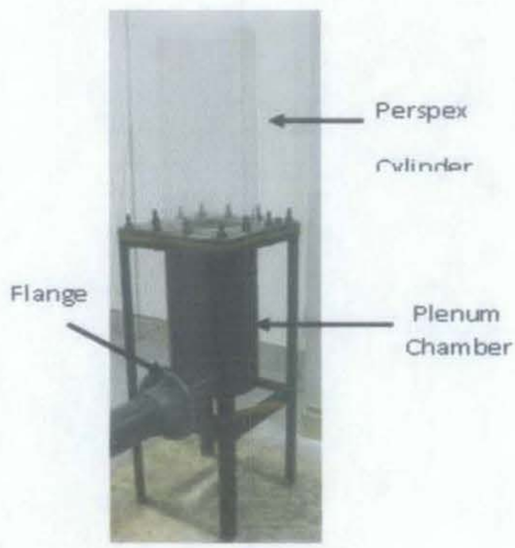

Figure 14: Single-staged Prototype

To determine the downcomer inlet angle an experiment is carried out on the above prototype. A hopper was introduced to the concept to mock a downcomer by feeding particles at various angles for observation. Tests where predicted by timing the particle flow at the various angles, visual observation was sufficient to critique the optimal angle.

The hopper was designed using CATIA and assembled at the fabrication itself. Designs of the hopper are presented below:

Figure 15 : Hopper Top (300mm Diameter cone with 20mm pipe outlet)

Figure 16: Hopper Bottom (0, 15, 30 and 45 Degrees)

CHAPTER 4:

RESULT AND DISCUSSION

The results for this project were done based on theoretical designs and the novel Single-Staged Swirling Fluidized bed. For this certain outcomes were achieved in order to build the design up, they were Data Collection, Draft Designs, CATIA models and component identification.

4.1 Data Collection

After deep theoretical research in the design aspects, modelling and recent findings the author had achieved the basic knowledge to produce a draft design for the further commencement into this project. The draft design was obtained by incorporating basic theoretical understanding to pre-existing designs of different models.

Currently the initial model is of a single-stage nature, and was designed by post-graduate students directed in the same field of study. The specifications, modelling, parameters and fabrication are solely credited to these students. Tests on the pressure drop and operating velocities are to be performed on this model.

The draft design of the multi-stage swirling fluidized bed will be based on a conceptual level with the pre-existing model of the single-stage swirling fluidized bed. In order to develop the multi-staging, much reference is given to recent findings on multi-stage fluidized beds with downcomers.

4.2 Draft Designs

There were several designs Drafted before a workable model was established. After each design critique was done to investigate the operation possibilities. Data acquired from the experiment was used to produce the last draft design. These designs were hand sketched using a scale of 1:3. It is from this progression that the final CATIA model was produced.

The first draft designs were based on assumptions and mental ideas (see Appendix B). The idea later developed into a scaled model for analysis. Below is an extract of the schematic that predicts a workable model.

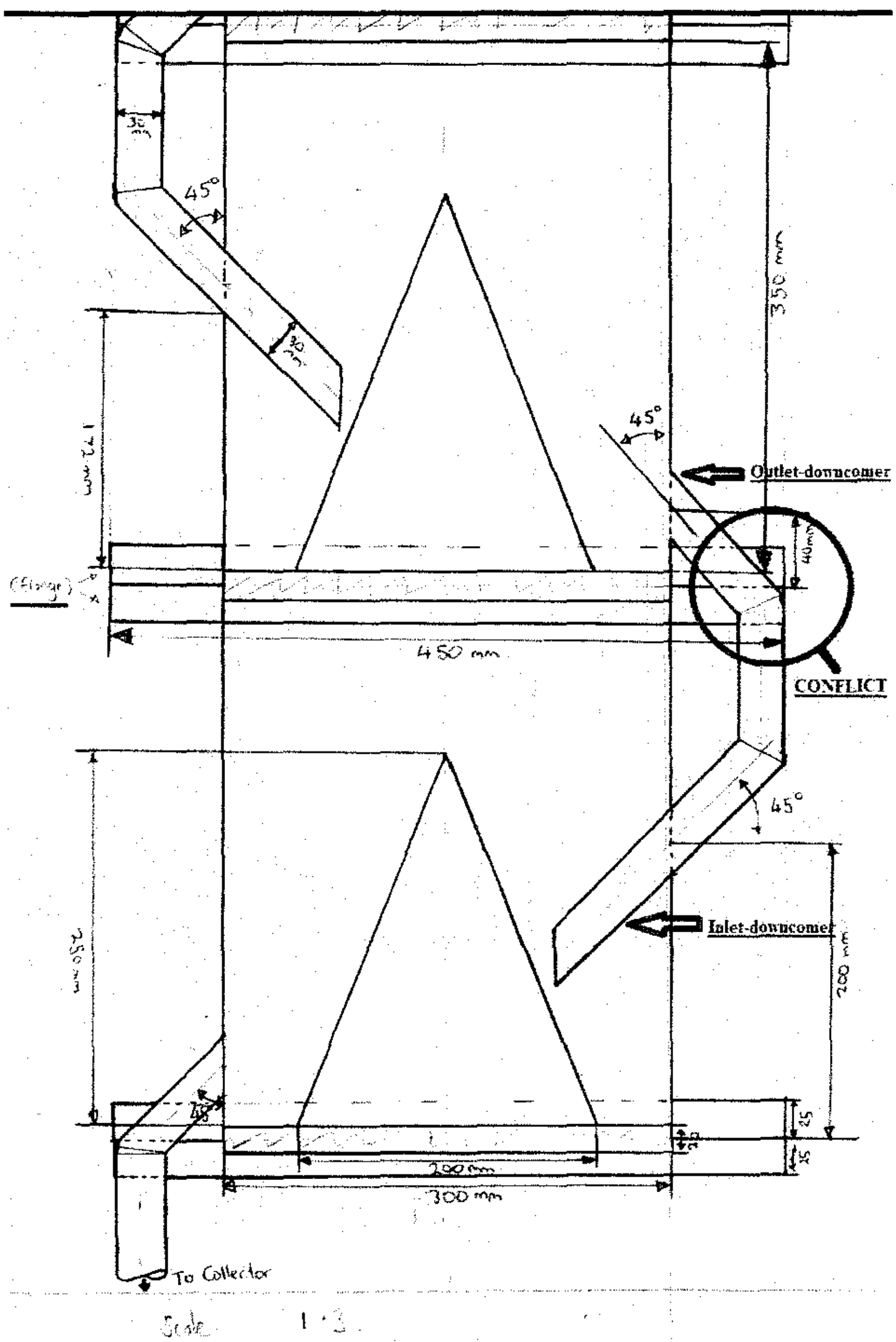

Figure 17: Scaled Draft Design with complications

There were complications which prompted a redesign. The 'fouling' being large will interfere with the downcomer-outlet. If the particles are spherical, they will run even down a 5 degree angle of declination. If less easy-flowing particles like cylinders or difficult-to-flow particles like sand (with a friction angle of close to 45 degrees) are used, then a redesign and re-fabrication of the fouling will be needed. However even a 0 degree angle of declination would suffice for the operation. The most important point for smooth flow is the outlet from the bed.

In conventional fluidized beds, the solids travel down from an upper stage to a lower stage through a vertical pipe called downcomer or standpipe which is full of solids (Figure 18). The gas does not bypass by escaping through the standpipe as the resistance for flow is higher than the normal flow thru the bed.

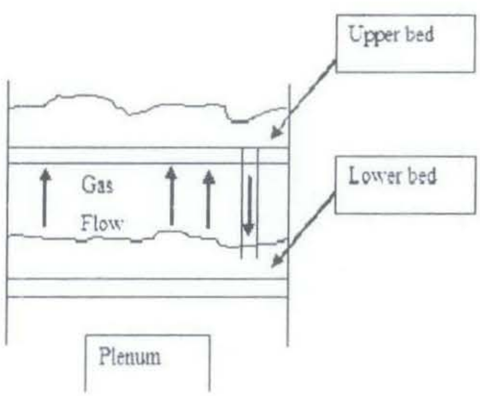

Figure 18: Schematic of conventional downcomers

Ideally for ease of flow in the downcomer outlet, the placement should be tangential to the bed-wall (cylinder). This allows for the radial and centrifugal forces of the swirling regime to assist particles through the downcomer without the need of an angle of declination. A scaled draft design of this concept can be seen in Appendices C and F.

The special features of the downcomer-inlet entail that it pierces the bed wall at an angle of 45 degrees. This was determined from the experimental observation carried out. The end of the downcomer would be placed as close to the cone as possible, this allows for particles to accumulate more residence time and also prevents damage of the distributor. At the edge a slit will be introduced as if to mock an L-valve. Further on an aeration inlet can be applied to assist flow. The slit is cut at an angel to prevent gas from back-mixing and clogging of the downcomer. A deeper description of the downcomer-inlet can be seen on the scaled drawing in Appendices D and G.

After all parameters were considered and complications worked out a complete idea was generated in another draft:

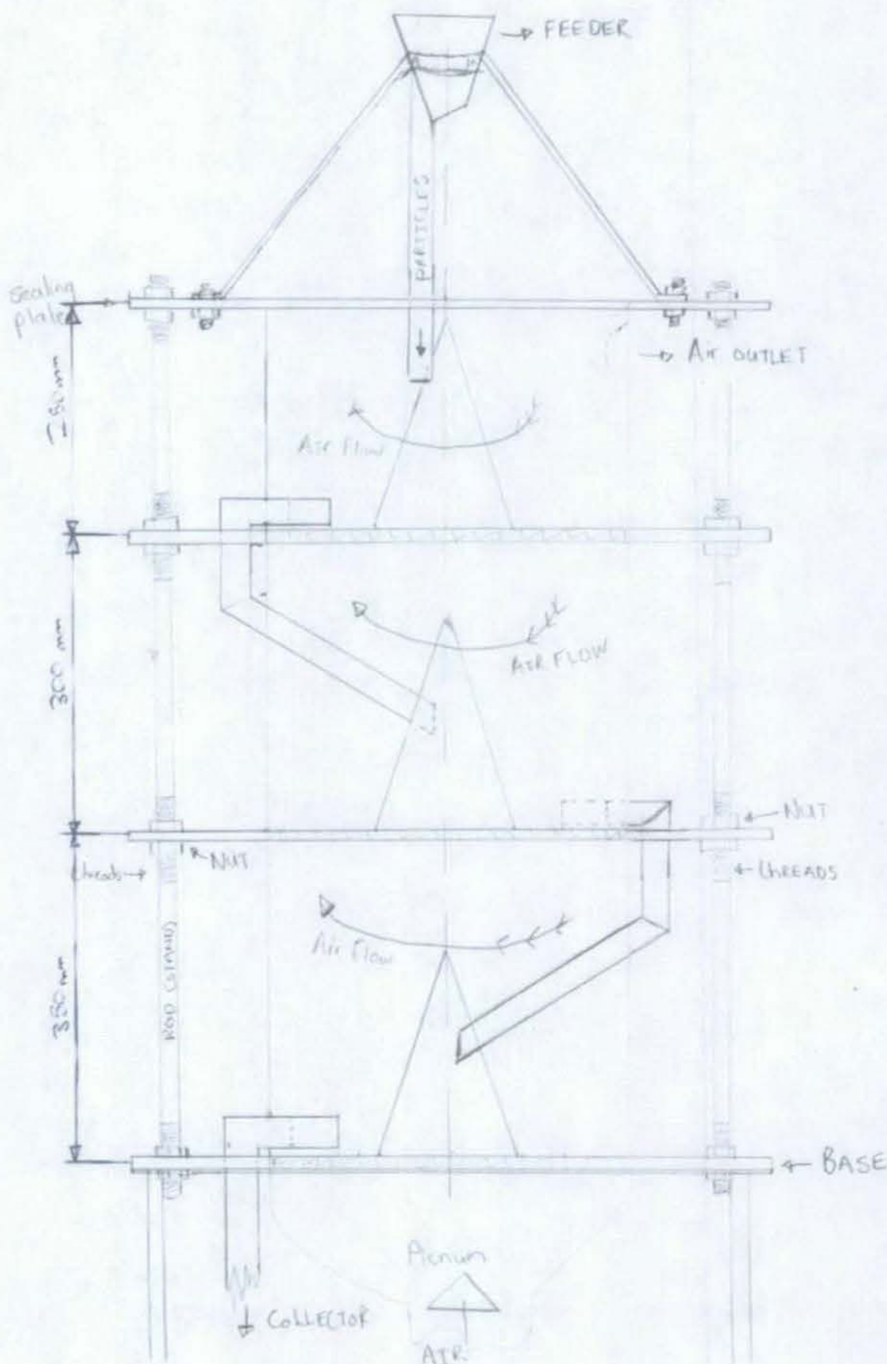

Figure 19: Final Draft Design

4.3 CATIA Model

The author familiarized himself with the workings of CATIA and had analyzed the Final Draft design (Figure 19) coupled with the scale drawings depicted in Appendices E, F and G. A graphic model was established as below:

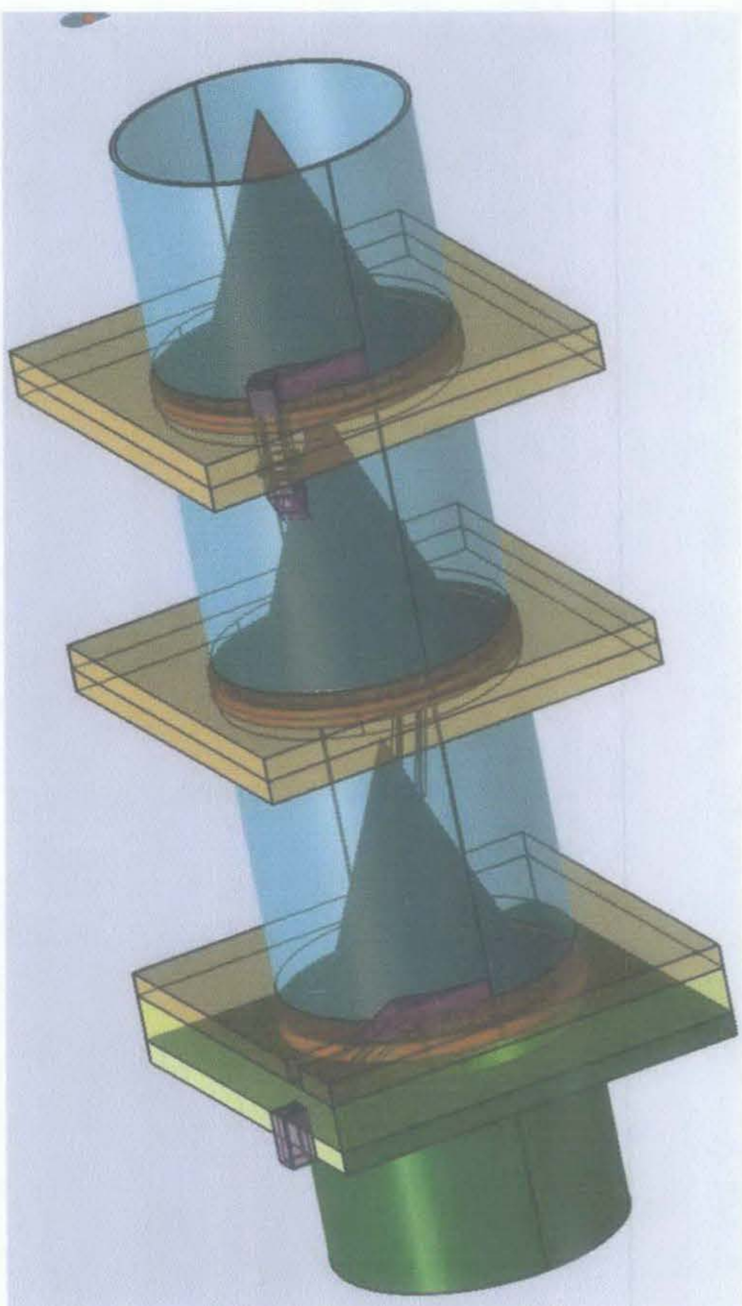

Figure 20: CATIA Model

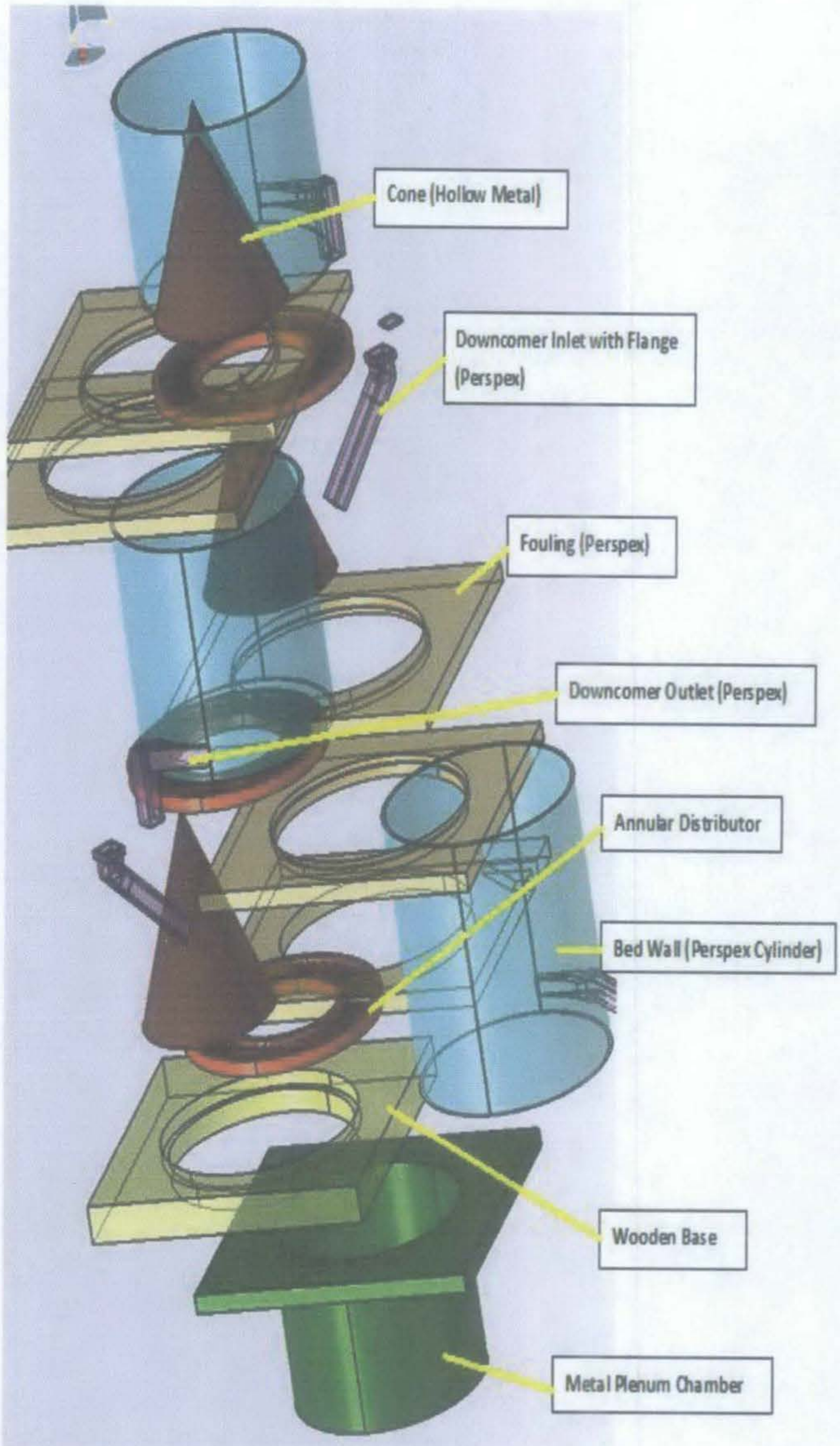

Figure 21: Exploded CATIA model with components

4.4 Multi-Staged Swirling Fluidized Bed

A brief summary on the modelled prototype is as follows:

- The prototype will have three stages. The heights of each stage decrease with each stage so as to cater for pressure drop and air flow resistance.

- The downcomer-outlet has a Zero angle of declination, relying on centrifugal forces for particle flow. It is mounted tangentially on the bed wall.

- The downcomer-inlet enters each stage extended as close to the cone as possible for longer processing time and to prevent damage to the distributor on impact. The end of each downcomer-outlet grove is made to mock an approximate non-mechanical valve (J-valve) with a planar face normal to the direction of flow to avoid gas bypassing or back-mixing.

- Components: Annular Spiral Distributor, Cone and Plenum Chamber (Metal); Cylinders, Fouling, Flanges and Down comers (Perspex)

- To support the various stages of the Perspex column, tierods will need to be used in the four corners.

4.4.1 Bill Of Materials (BOM)

Table 2: Bill of Materials (BOM)

Part No.	Quantity	Part description	Part Color	Function	Physical Parameters
1*	3	Cone (Hollow)	Brown	Eliminates Dead Zone, minimizes gravitational forces on particle entry.	Steel Sheet, 200 mm diameter x 250 mm height
2	3	Downcomer-inlet with Flange	Purple	Allows particles to enter the next stage.	Perspex, Rectangular 28x28 mm, 4 mm thickness
3	5	Fouling	Pale pink	Fouls the distributor with each stage.	Perspex, 450x450 mm, 25 mm thickness
4	3	Downcorner-outlet (Perspex)	Purple	Provides an exit for particles from a stage.	Perspex, Rectangular 28x28 mm, 4 mm thickness
5*	3	Annular Spiral Distributor (Aluminum)	Red	Causes swirling flow of fluid.	Aluminium, 320mm diameter outside, 200mm diameter inside
6	3	Bed Wall (Perspex Cylinder)	Light Blue	Contains particles and fluidized bed.	Perspex, 310 mm Diameter, 5 mm thickness, Heights: 350 mm; 300 mm; 250 mm.
7	1	Wooden Base	Beige	Absorbs unnecessary vibrations and provides support	Wood, 450x450 mm, 50 mm thickness
8*	1	Plenum Chamber (Hollow Metal)	Green	Provides even propulsion of air into the distributor.	Metal Frame, Specs. as per designer
9*	1	Blower Set-up	-	Fluid supply (air) with flow-meter and control-valves.	PVC pipes, butter-fly valves, Blower, Specs. as per designer

*Design credited to prominent researches [16]

4.4.2 Component Identification

4.4.2.1 CONE

Figure 22: Cone

The cone will be fabricated using a metal sheet. The base diameter is 200 mm and the height is 250 mm. It is fixed on the inner ring of the distributor. According to recent text the cone helps eliminate the possibility of a 'dead zone' at the centre of the bed during operation with bed materials. It also assists in transporting particles on to the bed without damaging the distributor blades. [16]

4.4.2.2 ANNULAR SPIRAL DISTRIBUTOR

Figure 23: Annular Spiral Distributor

Studies on the annular spiral distributor had proven its use to be best suited with the novel swirling fluidized bed. The design was carefully constructed by prominent researchers. The distributor consists of inner and outer rings which serve as flanges for 60 inclined blades (10 degree inclination). This forms an annular spiral configuration. The blade overlapping helps create the desired swirl. [16]

4.4.2.3 BED WALL WITH DOWNCOMER-OUTLET

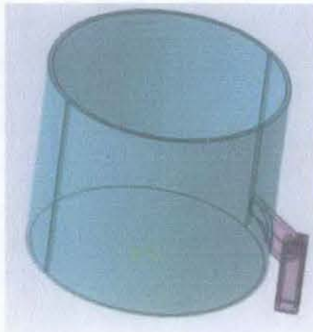

Figure 24: Bed Wall with Downcomer-Oulet

The Bed wall is formed by a Perspex cylinder with an outside diameter of 310 mm and a thickness of 5 mm. the heights for each stage varies. The bottom stage would be 350 mm the next is 300 mm and the top stage is 250 mm. this allows for less travel path of the fluid from stage-to-stage.

The downcomer outlet is placed tangentially to the bed wall so as to capture the centrifugal forces of the particles which would allow them to flow along the path of the downcomer.

4.4.2.4 DOWNCOMER-INLET

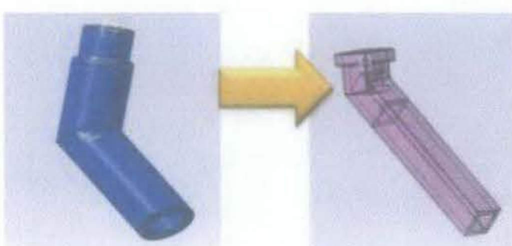

Figure 25: Downcomer-inlet

Just as the downcomer-outlet the downcomer-inlet will be rectangular Perspex. They will have an angle of 45 degrees which was observed from experiment and will pierce (Figure 25) the cylinder at a height and extend close to the cone as possible.

The particles will descend on the cone, so that they have a longer residence in the bed and do not get bypassed to the exit quickly.

At the outlet, a mock approximate J-valve will be fabricated which will prevent back-mixing of the gas or fluid. The grove is made with the planar face normal to the direction of flow. If fast drainage is required, four such downcomers can be used, on in each corner.

4.4.2.5 FOULING PLATES

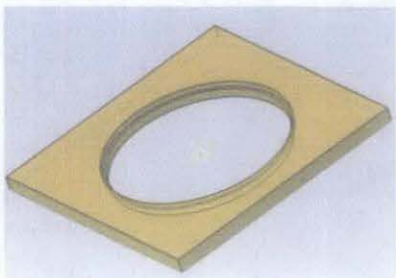

Figure 26: Fouling Plate

These plates are made of Perspex. They assist in fouling the stages together. The distributor is placed between two plates and fixed with bolts. A gasket can be used to provide a more adequate seal.

4.4.2.6 AUXILARY COMPONENTS

These components are of minor relevance as they simply provide linkage and a means to the supply of the prototype. These items include: Bolts, pipe stands, plenum chamber, gaskets, blower, PVC pipes and control-valves.

4.5 Discussion

Since the basis of the design is mostly on the characteristics and behaviour of the single-stage swirling fluidized bed, most of the data would be achieved from experiments carried out. With reference to the literature review contained within this report, there is sufficient information to allow for proper theoretical and mathematical models of a multi-stage swirling fluidized bed.

There are many advantages to a swirling fluidized bed more specifically in a multi-staged mode of operation. An idealized kinetically-controlled and diffusion-controlled system would consist of both the swirling regime and serial processing (multi-staging) combined. The swirling action of the bed overcomes certain shortcomings where no bubbles are seen and no gas bypassing occurs. Due to its helical path the gas has a slightly longer travel than the bed height this in addition with multi-staging improves fluidization on a good scale. Apart from reviews that pressure drop across the various stages does not vary that much, swirling fluidisation boasts that large open area fractions and low pressure drops at the distributor can be employed without ill-effects. [7]

Each component of the Multi-staged Swirling Fluidized bed Prototype was carefully selected to suit best requirements. Based on previous literature and design parameters of multi-staged fluidized beds one can see that the new swirling regime requires a new approach. The incorporation of downcomers and annular spiral distributor has rendered a promising multi-staged mode of operation. Downcomers are made to fit outside the bed wall to prevent air resistances and particle flooding.

Fabrication could not be carried out due to time restrictions. Materials were on hold during procurement delays. The parameters and application needs to be defined to ensure a secure design. Assumptions would begin at the easiest application which is drying. Software in order to draw scale models of the design was used

CHAPTER 5:

CONCLUSION & RECOMMENDATIONS

The rising need for alternative energy, advancements in technology and new requirements of Chemical processing encourages this project to be studied further. This prototype is meant to form curiosity and further development in fluidization which could lead to advancements in gasification. The application of such a process is vast and will be used in a wide variety of studies at a later stage.

The project proves to be very interesting and is significant for the completion of a bachelor's degree. The author attained deep theoretical knowledge on the subject matter. A draft design was developed using pre-existing designs. The prototype has yet to under-go fabrication. This system introduces a new approach to the novel Swirling Fluidized bed.

The next step to continue this promising project is to fabricate and experiment a multi-stage swirling fluidized bed. With that important factors will be studied further:

- Packing characteristics
- Flow through densely packed beds
- Bubbling beds
- Elutriation and entrainment
- Effect of pressure and temperature
- Attrition and agglomeration.

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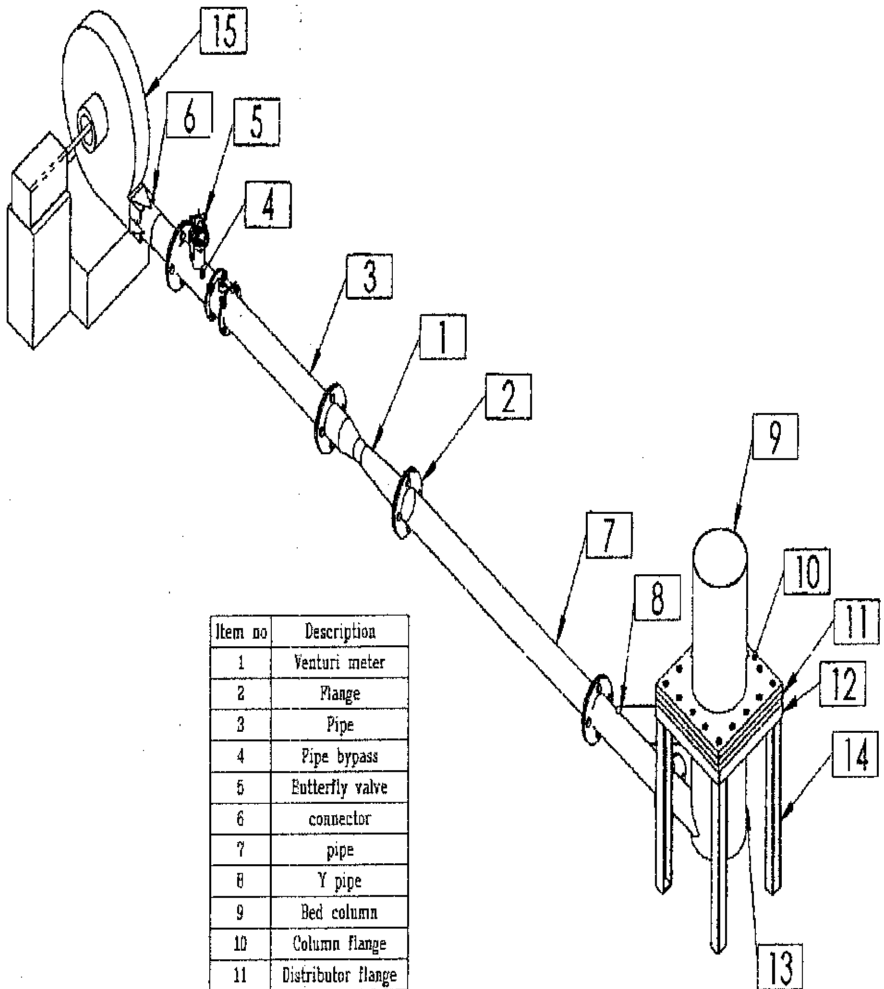
APPENDICES

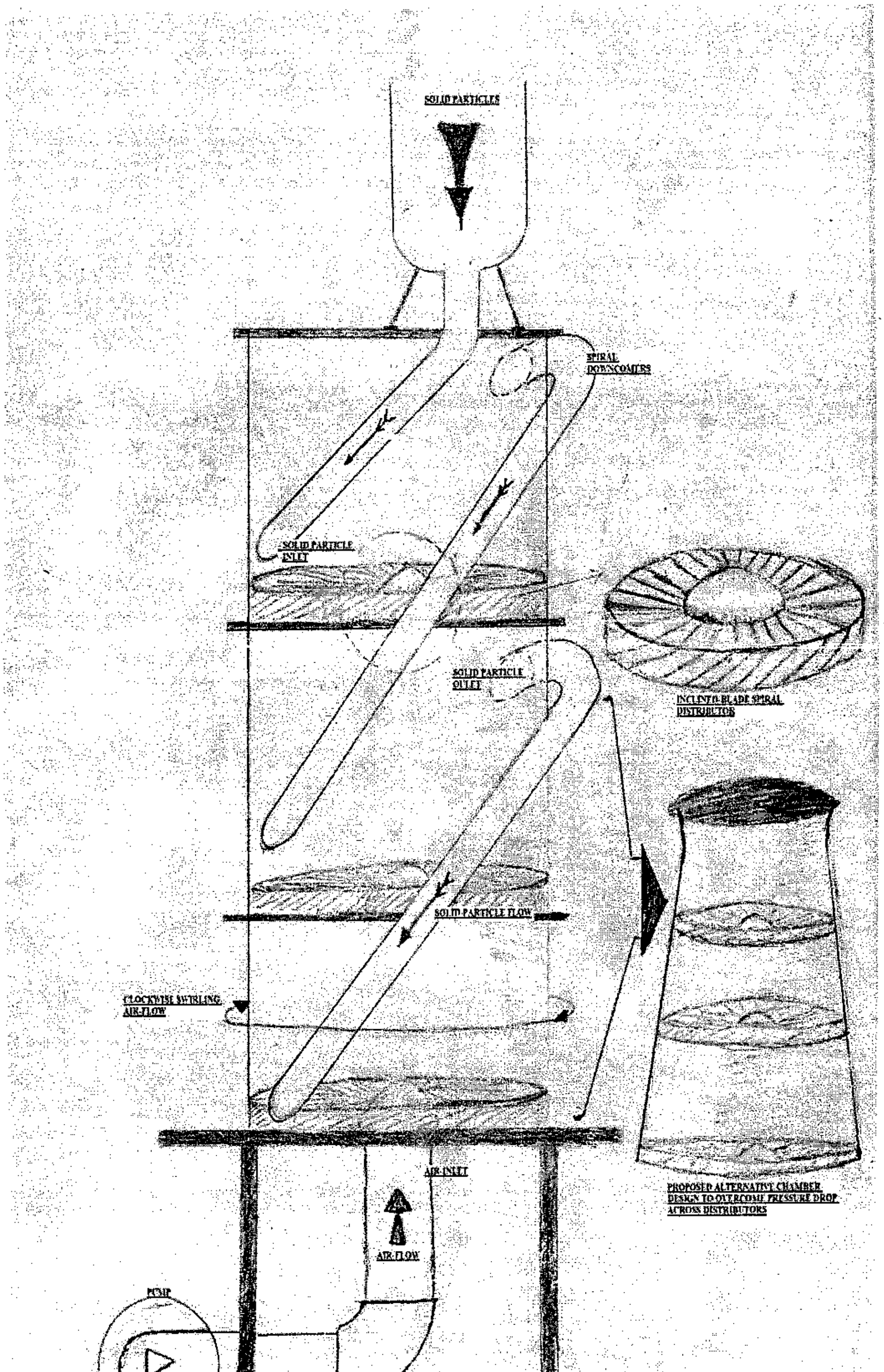
APPENDIX A



Item no	Description
1	Venturi meter
2	Flange
3	Pipe
4	Pipe bypass
5	Butterfly valve
6	connector
7	pipe
8	Y pipe
9	Bed column
10	Column flange
11	Distributor flange
12	Support frame
13	Chamber column
14	Column legs
15	Blower

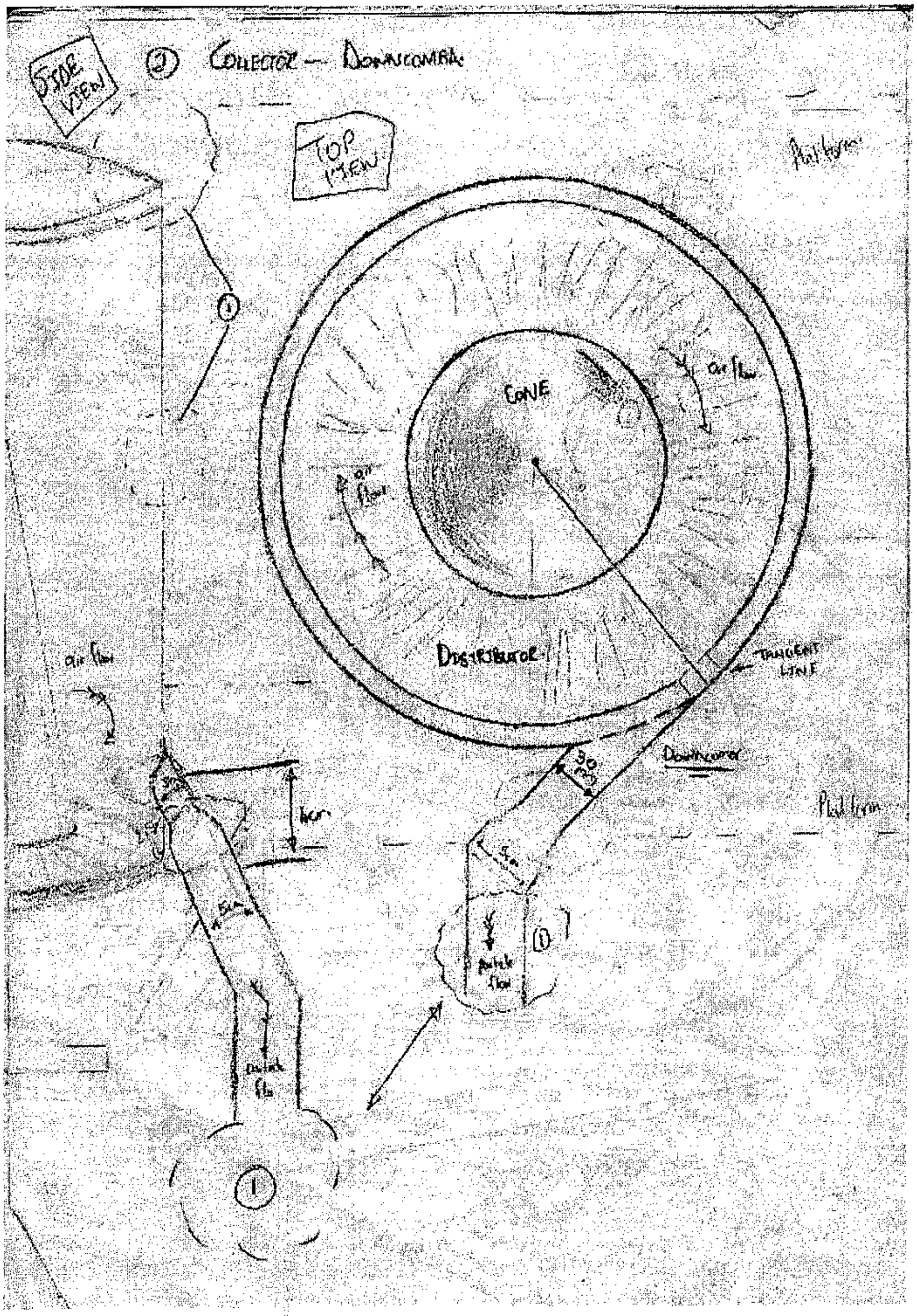
Universiti Teknology Petronas	
Title	Swirling Fluidized bed
Drawing 1	

APPENDIX B



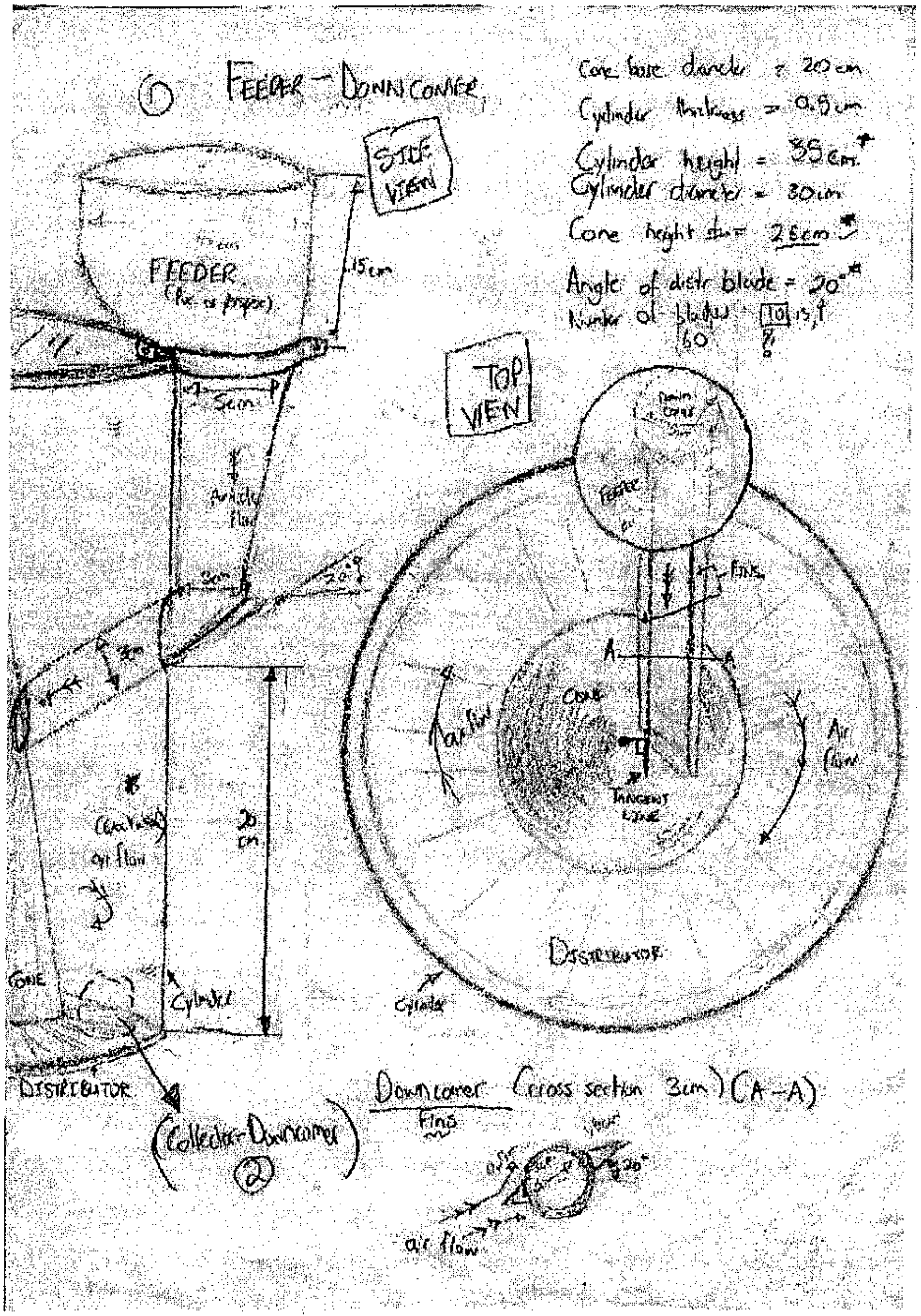
SOLID PARTICLES
SPIRAL DOWNCOMERS
SOLID PARTICLE INLET
SOLID PARTICLE OUTLET
INCLINED BLADE SPIRAL DISTRIBUTOR
SOLID PARTICLE FLOW
CLOCKWISE SWIRLING AIR-FLOW
AIR-INLET
AIR-FLOW
PUMP
PROPOSED ALTERNATIVE CHAMBER DESIGN TO OVERCOME PRESSURE DROP ACROSS DISTRIBUTORS

APPENDIX C



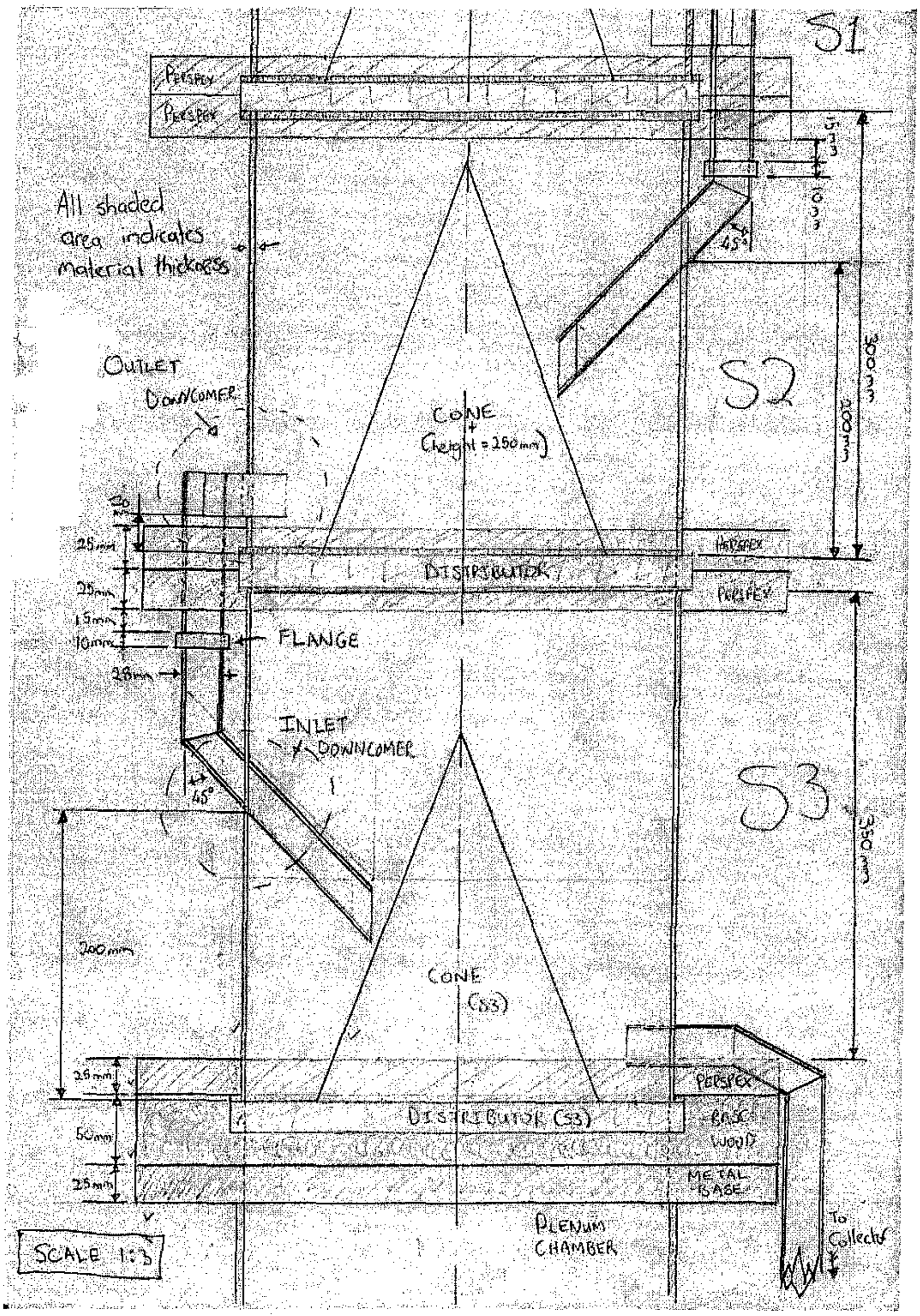
SIDE VIEW
2 COLLECTOR — DOWNCOMER
TOP VIEW
CONE
air flow
air flow
DISTRIBUTOR
TANGENT LINE
Downcomer
30 mm
air flow
particle flow
particle flow

APPENDIX D



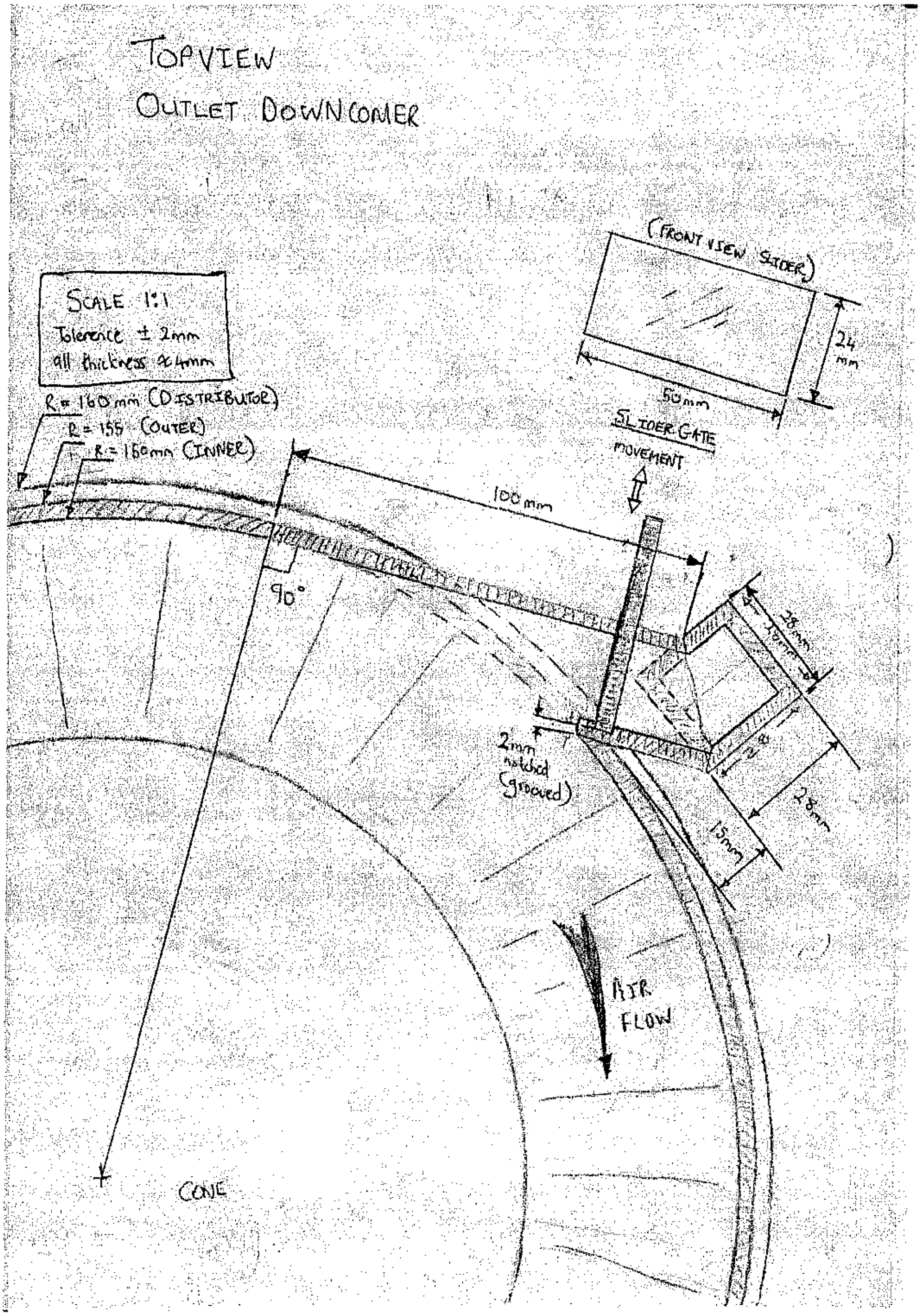
FEEDER – DOWNCOMER
SIDE VIEW
TOP VIEW
FEEDER
15cm
5cm
Particle flow
20°
20 cm
Cylinder
CONE
DISTRIBUTOR
(Collector-Downcomer ②)
Cone base diameter = 20 cm
Cylinder thickness = 0.5 cm
Cylinder height = 35 cm
Cylinder diameter = 30 cm
Cone height = 25 cm
Angle of distr blade = 20°
Number of blades 10
Fins
A
CONE
air flow
Air flow
TANGENT LINE
DISTRIBUTOR
Cylinder
Downcomer (cross section 3cm) (A–A)
Fins
20°
air flow

APPENDIX E

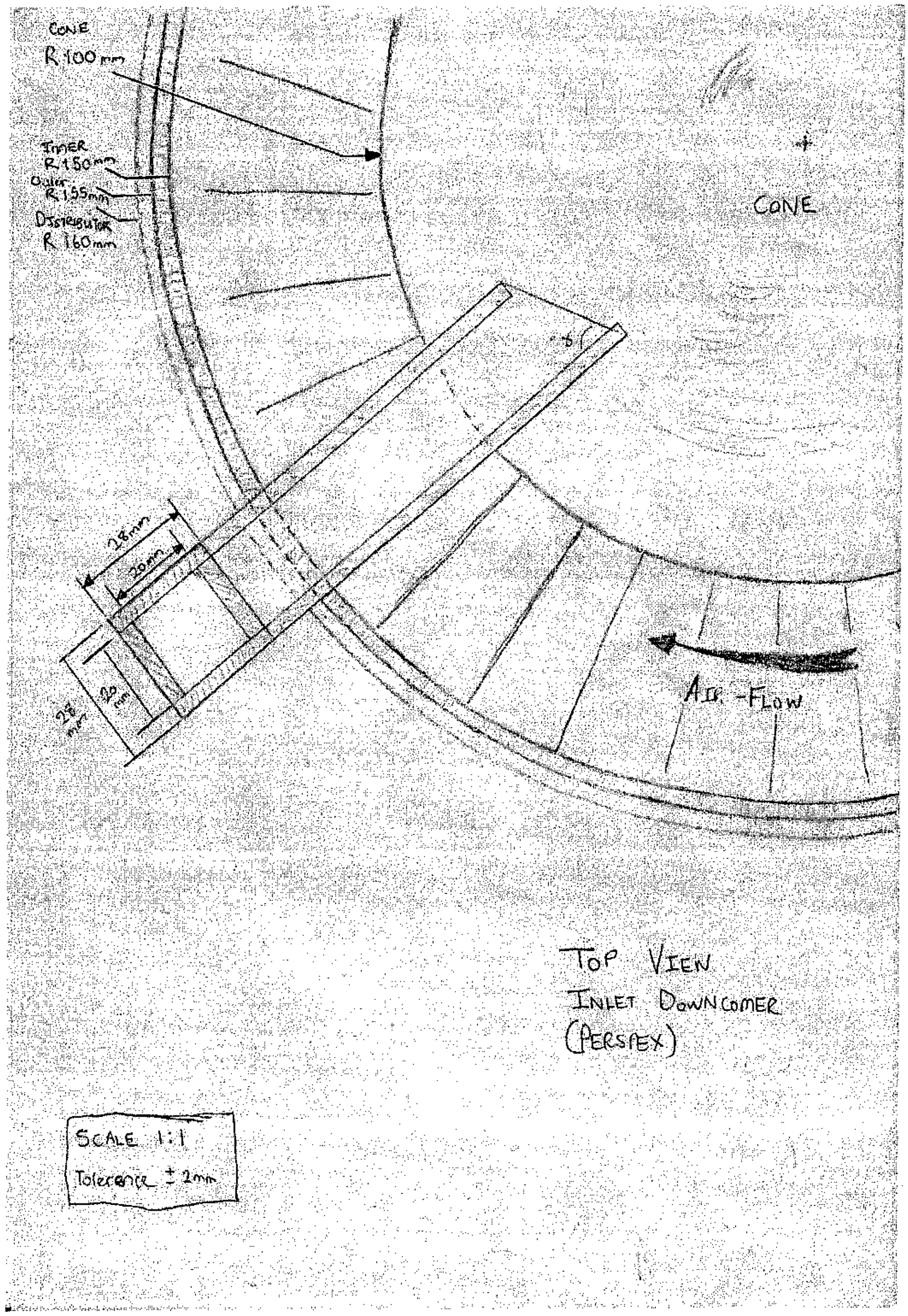


S1
Perspex
Perspex
All shaded area indicates material thickness
Outlet Downcomer
Cone (Height = 250mm)
S2
300mm
200mm
45°
Perspex
Distributor
Perspex
25mm
25mm
15mm
10mm
28mm
Flange
Inlet Downcomer
45°
S3
350mm
200mm
Cone (S3)
Perspex
Distributor (S3)
Base Wood
Metal Base
25mm
50mm
25mm
Plenum Chamber
To Collector
SCALE 1:3

APPENDIX F


TOPVIEW
OUTLET DOWNCOMER
SCALE 1:1
Tolerence ± 2mm
all thickness ≈ 4mm
R = 160 mm (DISTRIBUTOR)
R = 155 (OUTER)
R = 150mm (INNER)
(FRONT VIEW SLIDER)
24 mm
50mm
SLIDER GATE MOVEMENT
100mm
90°
2mm notched (grooved)
28mm
28mm
15mm
AIR FLOW
CONE

APPENDIX G


CONE
R 100 mm
INNER
R 150 mm
Outer
R 155 mm
DISTRIBUTOR
R 160 mm
CONE
28 mm
20 mm
28 mm
20 mm
AIR-FLOW
TOP VIEW
INLET DOWNCOMER
(PERSPEX)
SCALE 1:1
Tolerence ± 2mm