A HYDRODYNAMICS MODEL OF A CIRCULATING FLUIDISED BED REACTORS

by

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ABSTRACT

The principle of circulation of solids in a gas-solid system had led to the development of a number of processes. These processes in the form of circulating fluidised beds (CFB) are usually large-scale operations, with the fluid catalytic cracking process (FCC) and combustion as the major examples.

In this work pressure drop and solid circulation rate measurements have been conducted during the operation of a circulating fluidised bed (CFB) consisting of a riser, a solid collecting system (cyclones), a slow bed, a ball valve and a riser distributor. Most of the experiments have been conducted with FCC powder of average particle diameter of 64μm and particle density of 884 kg/m³. Some further experiments were conducted with either pure Poralox powder of particle diameter of 127μm and density of 1850 kg/m³ or a 50/50 v/v mixture of Poralox and FCC. The gas velocities were varied between 1.5 and 4 m/s. The three riser distributor plates had the same free percentage of open area, divided into 1 nozzle, 7 and 13 holes.

The pressure profiles around the circulating fluidised system should be such that the higher pressure drop values should be located above and below the ball valve for safe operation of the riser type reactors.

Increasing the slow bed inventory from 21 to 60 cm resulted in an increase in the riser pressure drop and hence in the axial solids fraction present in the riser.

The design of the distributor plate affects the performance of the CFB. For the same operating conditions using the 7 and 13 hole distributors much higher solid circulation rates were observed compared with the 1 nozzle distributor.

Increasing the density and the size of particles from FCC to Poralox leads to higher solid circulation rates but also higher pressure drops.

Operating the CFB at high opening areas of the ball valve will normally leads to a downflow of solids in an annulus adjacent to the riser wall.

If the ball valve is fully open then the CFB cannot be operated without instabilities.

The Dense Annulus Lean Core model (DALC) is proposed which relates the pressure drop along the axis of the riser to particle and fluid properties. The DALC model is shown to give good predictions, both qualitatively and quantitatively when the CFB is operated so as to achieve high solid circulation rates.

The DALC model is then applied to the case in which a first order catalytic reaction takes place in the riser. By increasing the reaction rate constant to 15s⁻¹ the model indicates a conversion of 55 % in the downflow region within a distance of less than 3m above the distributor. By increasing the gas velocity in the riser the model predicts that the conversion of a first order reaction decreases.
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CHAPTER 1
INTRODUCTION

1.1 DEFINITION OF STATES OF FLUIDISATION

Fluidisation is the operation by which fine solids are transformed into a fluidlike state through contact with a gas or liquid. Suppose a quantity of finely divided solid particles is placed into a bed (i.e., a cylindrical container) supported with a porous plate and air is passed through the base in the upward direction. At relatively low air velocities, fluid merely percolates through the void spaces between stationary particles, and this is the state of the fixed bed. With an increase in fluid flow rate, particles move apart and a few are seen to vibrate and move about in restricted regions. This is the expanded bed. With further increase in the upward fluid velocity, the expansion of the bed continues until a point is reached when the particles are all just suspended in the upward flowing gas. At this point the drag (frictional) force between particles and fluid counterbalances the weight of the particles, the bed takes on the appearance of a liquid having a flat surface and the particles are then said to be 'fluidised'. The superficial gas velocity at which this occurs is called the 'minimum fluidisation velocity' $U_{mf}$ and the bed is said to be in the state of incipient fluidisation. With an increase in gas velocity beyond minimum fluidisation, it is observed experimentally that instabilities and cavities are
developed in the form of gas bubbles. These bubbles form within the bed, rise to the surface where they burst in the same way as gas bubbles in a boiling liquid. At this gas velocity, the bed is divided essentially into two phases - the dense or emulsion phase where gas percolates through and the lean or bubble phase where much of the gas is out of contact with solids. Such a bed state is called an aggregative fluidised bed or a bubbling fluidised bed. A further increase in gas velocity leads to more vigorous bubbling with bigger gas bubbles appearing at the surface until at some velocity, the diameter of the surface bubble becomes equal to the diameter of the bed. This is the state of 'slugging' fluidisation and it is characterised by strong pressure fluctuations, with many particles being thrown up into the space above the bed known as the 'freeboard'. At these high velocities much of the solid is lost from the bed by entrainment and elutriation, so that cyclones are needed to return particles to the bed. At relatively high gas velocities the particles will be transported out of the bed and only a low concentration of rapidly moving particles remain in contact with the gas. This state is known as vertical dilute phase flow and is sometimes found in pneumatic conveying systems. However, between the regions of slugging and dilute phase flow, there is evidence for the existence of another regime of flow, the so-called "fast" fluidisation regime, characterised by the highly turbulent motion of individual particles and aggregates or clusters of particles, high
solid concentrations compared with the condition obtained in dilute phase flow and extensive backmixing of solids. It may be possible to maintain a fast fluidised bed at atmospheric pressure and with velocities of 2-4.5 m/s with fine powders. The various states of gas-solid contacting as they have been described above are shown in Figure 1.1.

1.2 CHARACTERISTICS OF CIRCULATING FLUIDISED BEDS

The discovery of how to maintain a stable and continuous circulation of solids in a gas-solid system has led to the development of a number of processes using the principle of circulation. These processes are usually large-scale operations, especially in the petroleum industry with the fluid catalytic cracking process (FCC) as the major example.

Different forms of gas-solid contacting and flow occur in the various sections of any circulating system in order to achieve stable circulation. These forms may include pneumatic conveying (low solids concentration), moving beds ($U_{\text{gas-solid}} < U_{\text{mf}}$), fast fluidised solids transport (high solids concentration, $U_{\text{gas-solid}} > U_{\text{mf}}$) and nearly always one or more fluidised beds.

The basic problem then is to decide what type of flow and what pressures are to be maintained in the different sections of a circulating system, the size of pipes and vessels, the gas flowrate so as to design an assembly which is stable and has a circulation rate meeting the demands of the particular process.
Solid circulation systems have been used with liquid-solids (ion exchange, adsorption-desorption systems); however the most important applications are with gas-solids systems.

The most important and extensive use of circulation systems has been in the field of solid catalysed gas phase reactions, where they are used in one of the following two situations. The first is where the catalyst deactivates rapidly and requires frequent regeneration. In this situation, a steady-state solid circulation condition is achieved with the catalyst being continuously regenerated and returned to the reactor. A second use is where much heat must be brought into or removed from the reactor. Gases have a very small specific heat capacity relative to heats of reaction, whereas solids have a relatively large heat capacity. Hence a continuous circulation of solids between two vessels (i.e., reactor and heat exchanger) can effectively transport heat from one vessel to the other and also control the temperatures in the units.

The circulation system can be treated as a whole, if the circulation rate for solids and the frictional losses for both the dense and lean mixtures are known. The principles involved in the circulation can be illustrated using as an example the water-steam system (Kunni and Levenspiel, 1973).

Consider the circulation loop ABCDEA of Figure 1.2. Section AB is 1m high and filled with water. Section BC is horizontal and contains a heater that vaporises all the
water moving to the left from B to C. Section CD is filled with steam and section DA has a heat remover such that all steam moving to the right from D to A is condensed. The static head around the loop starting with point A can be estimated as follows:

\[ P_B = P_A + 100 \text{ cmH}_2\text{O} \]
\[ P_C = P_B = P_A + 100 \text{ cmH}_2\text{O} \]

If the density of steam is negligible with respect to the density of water then,

\[ P_D = P_C = P_A + 100 \text{ cmH}_2\text{O} \]

and hence,

\[ P_E = P_D = P_A + 100 \text{ cmH}_2\text{O} \]

Thus there is a pressure head of 100 cmH\text{O} between E and A tending to cause convection (because of the density difference of the fluid in the two sides of the loop) in the clockwise direction around the loop. If the valve is then opened, the fluid will circulate at a rate such that the pressure drop around the loop just matches the static head of 100 cm.

These principles of circulation apply to gas-solid systems, and from the above example one can see that such circulation systems must have regions of high bulk density and others of low bulk density.

A typical flow diagram of a circulating fluidised bed (CFB) with gas-solid contact is shown in Figure 1.3. It consists of a vertical tube, the riser, a separation system, the cyclones, and a bubbling fluidised bed. The latter is connected through an inclined pipe with
the high gas velocity riser.

A valve is used to regulate the solid rate within the riser. To illustrate the principle of the circulating fluidised bed loop of Figure 1.3, a typical pressure balance around the system, is presented in Figure 1.3. For all restrictions to the solid flow due to the valve, the maximum pressure drop is located above the valve at the bottom part of the bubbling bed. This prevents leakage of the high velocity gas from the riser side into the bubbling bed. To ensure proper operation of the riser type of reactor, the gas leakage should be avoided at all costs.

According to Geldart's classification, Group A powders are characterised by a relatively small particle size ($d_p = 30-150 \mu m$) and a low particle density ($< 1500 \text{ kg/m}^3$). Frequently these powders show an increase in the void fraction of the emulsion phase as the fluidising gas velocity increases. Furthermore the minimum bubbling velocity $U_{mb}$ is always greater than $U_{mf}$. They easily form bubbles which rise at velocities much in excess of that of the emulsion phase gas. Fluid cracking catalyst is a typical example of a Group A powder.

Squires et al. (1985) constructed a regime flow diagram for a Group A powder which is illustrated in Figure 1.4. This particular diagram is a plot of the solid volume fraction versus the superficial gas velocity, and introduces the different flow regimes of fluidisation. In each flow regime a schematic diagram of an industrial application is presented.
Diagram (I) shows the upflow design of the original fluid catalytic cracker (FCC model). Diagram (II) indicates the downflow design with a meniscus. The first downflow FCC's (in the 1940's) operated at about 0.4 m/s, with a dust loading of about 1.7 kg per cubic meter of gas entering a single stage cyclone. Before 1955, FCC regenerators similar to (II) in mechanical arrangement, operated at about 0.75 m/s with about 70 kg/m^3 dust loading into a three stage cyclone. Diagram (III) shows a CFB for Fischer-Tropsch synthesis at Sasol; steam tubes within the bed remove the heat of reaction. Diagram (IV) shows a staged turbulent bed for the production of acrylonitrile. Diagram (V) indicates a modern fluid cracker (Mobil-UOP) with a riser cracking zone and a CFB catalyst regenerator. Diagram (VI) shows a CFB boiler.

The major advantages of gas-solid processing in CFB reactors in comparison to that of bubbling fluidised beds are the following:

1) A highly expanded bed of circulating fine-grain solids provides excellent gas-solid contact.

2) The high mean relative velocity between gas and solid results in high rates of heat and mass transfer.

3) The fluidised gas in the riser section of a CFB shows a plug flow behaviour.

4) The staged addition of one or more gaseous reagents at different levels in the riser is possible.
Although many of the above advantages describe the CFB reactor as an excellent gas-solid type of reactor, the high gas velocities and the recirculation of solids, may make the CFB system more expensive in terms of power requirement and investment in comparison with conventional fluidised bed reactors.

One of the main areas where CFBs have found commercial application is in combustion. Although initially most research and commercial development was centred on the bubbling form of fluidised bed, in more recent years much interest has focussed on circulating systems in particular for the finer and less dense forms of solid fuel such as some types of biomass. The Ahlstrom 'Pyroflow' system, which is based on the circulating fluidised bed principle, was initially developed in Finland for burning biomass fuels such as peat, wood wastes and paper mill sludges as well as coals. It is shown schematically in Figure 1.5.

This circulating bed is characterised by high fluidising velocities of the order of 5 m/s with entrainment of the majority of the bed particles which are collected by cyclone and returned to the combustor. The primary combustion air is fed through the main fluidised air distributor at the bottom while secondary air is introduced further up the reactor. Fine grains of sand are used as the bed material.

Fuels, such as peat, bark, sawdust and pulverised domestic waste are screw fed into the combustor which operates between 700°C and 1000°C depending on the fuel
being burnt. Because the circulating material is in a highly turbulent state and mixes rapidly, bed temperatures remain very uniform. Boiler tubes are situated in the upper part of the combustor and downstream after the cyclone.

In particular, it has proved to be a good system for burning biomass fuels of low density and small size which can be difficult in the more conventional 'bubbling' type of fluidised bed combustor as well as low reactivity coals.

Circulating fluidised bed systems for burning coal are also available from Lurgi (West Germany), Battelle (Columbus, USA) and Studsvik (Sweden). Lurgi have developed their system based on fast fluidised bed technology for the calcination of aluminium hydroxide to alumina. The coal is crushed to about 300 μm size. A novel feature of the Studsvik development is the use of a 'Labyrinth' particle separator instead of a cyclone for the recirculation of the bed material.

The Battelle process is called Multisolid Fluidised Bed Combustion. Essentially it superimposes the circulating bed of fine particles on a fluidised bed of large particles as is shown in Figure 1.6. This enables coal to be fed with a top size of up to 35 mm. By decoupling the combustion and heat transfer operations, this system allows greater fuel flexibility and better load following characteristics than a bubbling fluidised bed combustor.

Further technical advantages of this type of combustion system concern environmental matters. In particular, because of the low combustion temperatures employed,
emissions of sulphur dioxide to the atmosphere may be controlled by retaining much of the sulphur in the fuel using a fluidised bed containing crushed limestone or dolomite. Also, because of the low combustion temperatures, the formation of nitrogen oxides from atmospheric nitrogen is virtually eliminated and the conversion of fuel nitrogen to nitrogen oxides can be minimised by using appropriate operating conditions. There is also evidence that emissions of many of the trace elements associated with fuels are less than those obtained using alternative combustion techniques.

For over 10 years there has been considerable commercial interest worldwide in developing fluidised bed combustion, particularly for burning coal for power generation, industrial steam raising and furnaces producing hot gases. Much attention is also being directed to the use of biomass and lower grade fuels because of the increasing costs of premium fuels. At the same time, there are increasing numbers of combustible waste materials which require utilisation or disposal for environmental reasons. Frequently, these wastes contain sufficient combustibles to make incineration an attractive proposition, particularly if useful energy can be recovered.

Both bubbling and circulating types of fluidised bed combustion display certain characteristics, each having particular merit for different fuel types and applications. For example, very finely divided solid fuels of low density will be more suitable for circulating fluidised bed
combustion whereas a bubbling fluidised bed would be preferred for coarser, denser fuels.

The selection of the most suitable combustion equipment for each fuel will depend on its properties, such as chemical analysis, volatile content, calorific value, particle size and ash characteristics, as well as the specific application.

The choice between the circulating fluidised bed combustor and the "bubbling" bed combustor ultimately rests on their relative capital and operating costs. Where emission standards are not too onerous, the simplicity of the 'Shallow Bubbling Bed' technology is favoured; for example, at the smaller industrial scale, up to about 50 MW. The more complex and expensive circulating fluidised bed systems are more appropriate for the largest industrial and small power generation boilers where tight emissions standards need to be met.

1.3 OBJECTIVES OF THIS STUDY

The objective of this work is to study a circulating fluidised bed with regard to its hydrodynamics. This means performing experiments to measure axial pressure differences and solid circulation rates. Both monocomponent and binary component systems are to be investigated.

An important part of the work is taken up with theoretical aspects of predicting the axial pressure profiles and the axial distribution of solids.
Other researchers have attempted to predict pressure profiles and voidages along the axis of the riser. This has normally been by a combination of theory and empiricism. The resulting relationships can rarely be applied for general design purposes with satisfactory results. In some cases, the modelling has resulted in implications which are contradicted by visual observations.

It is hoped that this work will give rise to a more general means of accurately predicting the performance of a circulating fluidised bed (CFB).

Having developed a hydrodynamic model of a CFB, it is the intention to use this model for predicting the conversion of a chemical reaction taking place in the riser.
Figure 1.1 Principal flow regimes for upward flow of gas through solid particulate materials (Grace, 1986).
Figure 1.2 Sketch illustrating the principal of a circulating loop (Kunii & Levenspiel, 1969).
Figure 1.3 Pressure profile around a CFB.
Figure 1.4 Regime diagram for fine powder
(Squires et al., 1985).
Figure 1.5 Circulating bed boiler system (Cooke, 1989).
Figure 1.6 Battelle multisolid fluidised bed boiler (Cooke, 1989).
CHAPTER 2
LITERATURE SURVEY

The circulating fluidised bed (CFB) has been an important part of many industrial processes for the last decade. The first known review of the CFB dates back to the Second World War. Squires (1985) devoted a paper in the First International Conference on CFBs which told the history of the development of the fluid catalytic cracking (FCC) process. According to this a consortium led by the Standard Oil Company of New Jersey and M.W.Kellogg Company used data obtained at the Massachusetts Institute of Technology (MIT) to design a fluid bed substitute for a 100 barrel per day pilot cracking reactor for the catalytic cracking of oil feed to gasoline.

However the first real attempt to study systematically the subject of circulating fluidised beds was carried out by Yerushalmi et al. (1976). This work, in common with much subsequent research, dealt only with the pressure gradient measurements over a range of solid circulation rates and gas velocities. A wide range of experimental conditions was examined by Yerushalmi et al. and the conclusions drawn from their studies are applicable for design purposes.

The first successful attempt to produce general rules for predicting the voidage and hence the pressure gradient in the riser of a CFB came with the work of Li and Kwauk (1980) who conducted over a 1000 experiments using solids of different sizes and densities.

Although a great deal of empirical work has been
published, work of a theoretical nature has been limited, and in the last 10 years Yerushalmi's and Kwauk's work has remained as a basis with which subsequent research has been frequently compared.

For the purpose of this study, the literature survey is limited to work published on low concentration two-phase flows.

2.1 DEFINITIONS OF THE STATE OF TWO-PHASE FLOW IN CFBS

The riser of a circulating fluidised bed operates in gas velocity and bed density region between the extremes of bubbling fluidisation and vertical pneumatic transport. This regime has been subdivided into "turbulent fluidisation", "fast fluidisation" and "riser transport" Yerushalmi et al. (1976), but the criteria proposed to mark the boundaries of these regions are not yet fully developed. This can result in confusion and in order to avoid this these terms will not used in the discussion section. However the description of this subdivided regime as it is described in literature is considered below.

2.1.1 Definitions of the State of the Turbulent Regime

There are three main definitions of the turbulent regime which have been described in the literature. (i) Increase in the gas velocity in the riser above that causing slugging, causes the fluidised bed to pass into a state in which continuous coalescence of slugs occurs leading to a channelling state with tongues of fluid darting in a zigzag fashion through the bed. This is the
state of turbulent fluidisation, as defined by Kehoe and Davidson (1971).

(ii) According to Yerushalmi et al. (1978) the turbulent regime is that which exists between two characteristic velocities $U_c$ and $U_k$. These velocities are determined from variations in the amplitude of pressure fluctuations across a section of the fluidised bed. During observations the amplitude of pressure fluctuations first peaked and then fell to a relatively low level as the gas velocity was increased. The superficial velocity at the peak was called $U_c$ and the velocity at which the amplitude levelled out was called $U_k$.

(iii) Rhodes (1986) defined the turbulent regime as that in which solids are transferred from the fluidised bed to freeboard by two competing processes: increase in bed height with increasing gas velocity due to bed expansion and decrease in bed height with increasing gas velocity as more solids are held in the freeboard.

2.1.2 Definitions of the Fast Bed Regime

There are three main definitions in the literature.

(i) Yerushalmi et al. (1976) defined as fast fluidisation a technique for bringing gas at high velocity into intimate contact with a finely divided solid in an entrained dense suspension characterised by extreme turbulence and extensive refluxing of dense packets and strands of particles. This technique was primarily orientated towards gas-solid reactor applications where it offered several important advantages over the conventional low-velocity
bubbling fluidised bed.

(ii) Fast fluidisation was defined by Rhodes (1986) as the transition at higher velocities where conditions in the riser of a CFB are relatively insensitive to changes in gas velocity or imposed solids flux. This results in a stable condition in which the proportions of dilute-phase and dense phase transport in the vertical column varied only gradually with changes in gas velocity or imposed solids flux.

(iii) According to Schnitzlein and Weinstein (1988), fast fluidisation is a two-phase flow with two distinct main regions, a dense region of solid at the bottom of the fluidisation column with similarities to bubbling and turbulent fluidised beds and a dilute region at the top. The solids are present in the dense region of the high velocity fluidised bed in at least three different forms; a) stagnant or down-flowing very near the wall at a void fraction close to minimum fluidisation conditions, b) travelling upwards in the form of dense packed solid aggregates with constant average velocity adjacent to the above region and c) dispersed dilutely in the upward flowing gas.

2.1.3 Definitions of the Dilute Phase Regime

Two definitions have been given in the literature.

(i) A dilute phase is said to be reached when all the particles are in motion and no particles remains at the bottom of the riser for more than a short time, while the gas is in the continuous phase.
(ii) The term circulating fluidisation is used in a general way to describe the gas-solid up-flow systems that exhibited carry-over of solid with an external recirculation leg.

2.2 UNIVERSITY WORK ON HYDRODYNAMICS OF CIRCULATING FLUIDISED BEDS

Although CFBs have been exploited commercially by a number of companies since the 1940's, university research was only started in the late 1970's. Yerushalmi and co-workers at City College New York, investigated the behaviour of a fine powder, (Group A according to Geldart's Classification), in fluidised beds operated at high gas velocity. They published a number of papers in which it was claimed that high velocity beds are favourable as gas-solid reactors compared to conventional fluidised beds. They referred to such a system as a "fast fluidised bed". Since then, a large number of researchers have investigated the phenomenon of fast fluidisation.

The studies of Yerushalmi et al. (1976-1979) focused on the definitions of different regimes of fluidisation as the gas velocity and solid circulation rate is increased.

Lanneau (1960) was the first to recognise the transition from bubbling to turbulent fluidisation by studying the fluidisation characteristics of a fine solid as the fluidising air velocity was increased to 1.7m/s. He used small capacitance probes and reported that the
heterogeneous character of the fluidised bed was changed to a condition of increasing homogeneity as the gas velocity increased beyond 1m/s.

Kehoe and Davidson (1971) also described the breakdown of the slugging regime at similar velocities into a state described as turbulent. Since then, the transition from bubbling to turbulent fluidisation has been the subject of investigation by Yerushalmi et al. (1976).

The first papers published by Yerushalmi et al. (1976, 1977) focused on the three different regimes of fluoridation, bubbling, turbulent and fast.

As already mentioned the transition from the bubbling to the turbulent state, was characterized by two gas velocities, $U_c$ the velocity at which the pressure fluctuations in the bed peak and $U_k$ the gas velocity which marked the decay in the magnitude of pressure fluctuations in the bed. Visual observations have shown that the transition was also accompanied by the breakdown of large bubbles and slugs which continually coalesce and split. The solid in the bed was rearranged into clusters and streamers of particles whose motion was mostly downwards with smaller clusters and individual particles that were entrained upwards.

Yerushalmi et al. strongly supported the idea of the existence of clusters when a fine solid was used in a bed in the turbulent state. They claimed that the terminal
velocities of the clusters were greater than the gas velocity in the bed and individual particles are present due to the erosion of clusters.

It was stated, however, that no particle carry-over will be experienced provided that the freeboard is sufficiently high and hence there is no need for recirculation of the solid.

The studies of Yerushalmi et al. were carried out using high speed photography in three experimental set-ups: a rectangular two dimensional bed, a 75 mm diameter bed and a 152 mm diameter bed.

They presented their results in terms of pressure gradient across a range of sections of the riser, as a function of superficial gas velocity and solid circulation rate. They claimed that for a given gas velocity, the solid concentration in the riser increased with increasing solid circulation rate and they considered the possibility of achieving solid concentrations comparable to those in a bubbling bed. They also found that the pressure gradients were normally higher in the bottom section of the riser and lower towards the top.

They claimed high slip velocities in the fast bed and they proposed as in the above mentioned case of turbulent flow that the presence in the bed of dense packets of particles which had an effective free-fall velocity considerably in excess of that of a single particle. Because of its size the packet of particles would not be sustained by the upward flowing gas but would fall back into the bed and
disintegrate.

According to Yerushalmi et al. (1976), the fast bed condition was marked by high solid concentrations, clustering of individual particles which break apart and reform, extensive backmixing of solid and slip velocities much higher than the terminal velocity of an individual particle.

The slip velocity was defined as: \( U_{sl} = \frac{U_g}{\epsilon} - U_p \)

The fast bed condition was stabilized only if the solid emerging from the top part of the reactor was recirculated by means of cyclones and a standpipe.

At velocities in excess of the transport velocity, \( U_{tr} \), the solid was fully entrained and its concentration within the column depended not only on the gas velocity but also on the solid circulation rate (Figure 2.1). When the solid circulating rate was low the dilute phase flow regime existed in which particles moved upwards in relatively straight paths and the slip velocities were of the same order as the terminal velocities of the individual particles. If the solid rate was increased further, the suspension became denser and at sufficiently high rates the fast bed state was established.

The structure of the fast bed was also explained in terms of clusters of particles (Yerushalmi et al., 1978) with high slip velocities. The equivalent diameter of a spherical cluster was calculated by neglecting the
acceleration and friction effects and assuming that:

a) all the particles were in densely packed clusters,
b) clusters were discreetly distributed in the bed and
c) clusters had a voidage equal to that at minimum fluidisation.

Based on these assumptions, the pressure drop per unit length was described by the following equations:

Pressure drop due to gravity in the vertical pipe:

\[
\frac{dP}{dL} = (1-\varepsilon) (1-\varepsilon_m) \rho_s \ g
\]  

(2.1)

where: \( \varepsilon \) is the effective voidage
\( \rho_s \) is the solid density

Pressure drop due to the drag exerted by the gas on the cluster:

\[
\frac{dP}{dL} = \frac{1}{8} N A \rho_g \left( \frac{U_g}{\varepsilon} - U_p \right)^2 c_{dp} f(\varepsilon) D_e^2
\]  

(2.2)

where: \( N \) is the number of clusters in the length \( dL \)
\( A \) is the cross-sectional area of riser
\( \rho_g \) is the gas density
\( U_p \) is the mean velocity of clusters
\( c_{dp} \) is a drag coefficient related to that of a single particle, \( c_{ds} \) by the expression:

\[
c_{dp} = c_{ds} \left( \rho_g D_e \left( \frac{U_g}{\varepsilon} - U_p \right) / \mu \right)
\]  

(2.3)

where: \( \mu \) is the gas viscosity

The function \( f(\varepsilon) \) accounted for the hindered settling
caused by the closeness of the aggregates. This function was related to the Richardson-Zaki (1954) index \( n \) by:

\[
f(\varepsilon) = \varepsilon^{-2n}
\]  

(2.4)

The value of \( n \) ranges from 4.65 to 2.4 for terminal Reynolds number, \( \text{Re}_t, 0.2<\text{Re}_t<500. \)

Hence, \( D = \frac{3}{4} \rho_s \left( \frac{U_g}{\varepsilon} - U_p \right)^2 c_{dp} \varepsilon^{-2n} \rho_s g (1-\varepsilon_{mf}) \)  

(2.5)

The pressure drops obtained from these equations agreed well with the experimental data. The cluster voidage was an adjustable parameter which was used to allow the model to fit the data. The cluster diameter \( D_c \) was calculated and plotted as a function of the solids concentration, \( (1-\varepsilon) \), for the 0.15m diameter pipe. Cluster diameters in the range 2 to 35 mm were calculated with this equation for the solids FCC and Dicalite 4200, though no visual observation suggesting the existence of such large cluster diameters was reported. The disadvantage of this model is that it required the use of many unmeasured parameters such as \( N, \varepsilon, D_c \) and \( U_{sl} \).

Li and Kwauk (1980), emphasized the significance of the solid distribution in the axial direction of the riser and they presented experimental data obtained using a CFB, in the form of axial average voidage distribution. The values of average voidage were obtained by a detailed analysis of
the pressure gradient, entrance and acceleration effects being neglected. The experimental vertical average voidage profiles were of S-shaped form (Figure 2.2).

The characteristic voidage profile of the fast bed consisted of a dense phase at the bottom of the bed, a dilute phase at the top and an inflection point in the middle, that was the transition from a low voidage at the bottom to a high voidage at the top of the bed. Based on the experimental evidence of the voidage profile, Li propose a hydrodynamic model, based on the concept of clusters for powders belonging to Group A of Geldart's classification (Geldart 1973).

According to the model, the clusters at any height of the bed moved upwards through a diffusive mechanism from a relatively dense region in the lower section, and then, when they arrived at a higher region along the bed where the average density was lower, they tended to fall back to the lower region because of buoyancy. Under steady-state conditions dynamic equilibrium existed and the fluxes due to diffusion and buoyancy were equal. Figure 2.3 shows the physical picture of the model. The equation was of the form:

\[ \ln(\varepsilon - \varepsilon_a/\varepsilon^* - \varepsilon_a) = -1/z_0 (z - z_i) \]  

where: \( \varepsilon^* \) is the voidage for the dilute phase section
\( \varepsilon_a \) is the voidage for the dense phase section
\( z_i \) is the position of the inflection point in the
voidage curve

$z_0$ is the characteristic length defining the ease of mixing and segregation for the clusters

The solution of the above equation required the determination of the four unknown parameters.

Li et al. (1982) correlated the above four parameters from experimental data. The following correlations were found to fit the data well:

$$
\varepsilon_0 = 0.756 \ (18\text{Res} + 2.7\text{Res}^{1.687} / \text{Ar}^{0.0741}) \quad (2.7)
$$

$$
\varepsilon^* = 0.924 \ (18\text{Res} + 27 \text{Res}^{1.687} / \text{Ar}^{0.0286}) \quad (2.8)
$$

where:

$$
\text{Res} = \frac{\rho_g d_p (U - V \varepsilon / (1-\varepsilon))}{m} \quad (2.9)
$$

$$
V = \frac{G}{\rho_s} \quad (2.10)
$$

$$
\text{Ar} = d_p^3 \frac{\rho_g g (\rho_s - \rho_g)}{\mu^2} \quad (2.11)
$$

The position of the inflection point, $z_i$, was found to depend on the average particle residence time in the dense region, which was correlated from the experimental data by the expression:

$$
t_\infty = 175.4 \ ((\rho_g/\rho_s-\rho_g) (U-V_p)^2/d_p g)^{-1.922} \quad (2.12)
$$

and since $z_i = L - t_\infty V_p$
then, the inflection position

\[ z_i = L - 175.4 \left( \frac{(\rho_g/(\rho_s-\rho_g))d_p g}{V_p \left( U-V_p \right)} \right)^{1.922} (U-V_p)^{-3.844} \]  \hspace{0.5cm} (2.13)

where: \( L \) is the total length of the riser

The characteristic length, \( z_o \), was given by the empirical correlation:

\[ z_o = 500 \exp \left( -69(\varepsilon^* - \varepsilon_0) \right) \] \hspace{0.5cm} (2.14)

Kwauk et al. (1985) presented modified correlations for the four unknown parameters and a full picture of the modeling of the fast fluidisation circuit.

The pressure drop in the downcomer, \( dP_d \), was balanced by the pressure drops through the solid valve, \( dP_v \), the fast bed, \( dP_f \), and the overhead cyclone, \( dP_c \).

Hence,

\[ dP_d = dP_v + dP_f + dP_c \] \hspace{0.5cm} (2.15)

This equation is written in terms of voidage and riser height and it was claimed that the major portion of the solids existed in the fast bed and downcomer. Hence, they presented an equation for determine the average bed voidage in the fast bed, for fixed rates of solid circulation.

\[ \varepsilon = 1 - ((I/A) - (a/A)C) / (1+a/A)Z \] \hspace{0.5cm} (2.16)
where: $I$ is the total inventory of solids

$A$ is the c/s area of the fast bed

$a$ is the c/s area of the downcomer

$C = (dP_v + dP_c)/\rho_s$

$Z$ is the height of the fast bed

This mathematical model represented the best quantitative description of the fast fluidised bed, if the system was defined as consisting of two main regions, a dense region at the bottom of the bed and a dilute region on the top.

Cankurt and Yerushalmi (1978) investigated gas mixing in a CFB, using a tracer technique. They claimed that the radial concentration profile was uniform if a group A powder was used. For a riser diameter less than 0.1m however wall effects enhanced the gas velocity profile causing radial segregation. From those observations they concluded that the gas flow in the fast regime was of plug flow type.

Avidan (1980) using the same equipment, and a radioactive tracer technique, investigated solid mixing in the CFB. He found that the axial dispersion coefficients were higher near the top of a turbulent bed compared with a fast bed at the same gas velocity. He also claimed that the axial dispersion coefficient showed a minimum at a specific gas velocity of 3.5m/s and this value was constant in the CFB. Avidan also reported that the Peclet number increases
considerably when the gas velocity rises above 4m/s but below this value it remains constant. Since the Peclet number \( \text{Pec} = \frac{G_L}{\rho_s D_s} \) gives the relative importance of solids movement by the mechanisms of convection and dispersion, Avidan proposed that below this critical gas velocity, the two mechanisms were of equal magnitude and due to this fact, the claims of good gas-solid contact and high heat transfer coefficients for the fast bed were basically true. However, he did not report anything on the radial distribution of solids.

Weinstein et al. (1983) found that the position of the inflection point was also dependent on the 'imposed' pressure drop set by the recirculation leg of the circulation system. In their experiments, an unrestricted standpipe was used for solid recirculation and it was stated that the solid inventory above the standpipe leg affected the position of the inflection point. Moreover the pressure profile in the bottom and the top of the riser seemed to be unchanged.

The radial voidage distribution in CFBs was first investigated by Weinstein et al. (1984). They made measurements with an X-ray source and a plate film with time exposures of about 5 s. To obtain density variations from the plate film an absorption technique was used. They plotted the experimental data as local value of voidage versus the radial position. For a set of flow parameters,
they claimed that the radial profiles showed a dilute core of high voidage in the range 0.9-0.99 surrounded by a dense annular region of voidage 0.6 to 0.85, along the axis of the riser. Under identical flow conditions, they also measured the variation of voidage along the riser height. From these observations they claimed that a dense phase occupied half the riser height and a transition to a dilute phase occurred near the top of the bed. Although the above mentioned radial profiles have been measured in positions, of what they claimed to be the dense lower part of the bed and within the transition region from dense to dilute phase, the radial profiles exhibited the same shape and variation ie high voidage of around 0.9 in the center of the riser and low voidage of around 0.7 in the annulus (Figure 2.4).

In a different paper, Weinstein et al. (1985), claimed the importance of the inflection point in the axial voidage profile and the behavior of this inflection point under different flow conditions. Hence, their approach towards describing the CFB system, was that according to their design of equipment, the riser two-phase flow consisted of: a) a dense region in the bottom part of the bed which occupied half of the riser height, with a radial profile of very high voidage in the center of the riser and a low voidage in the annulus, b) a dilute phase in the top part of the riser, where the radial voidage was not examined and c) an inflection point which showed the transition from the
dense phase towards the dilute phase and the position of which was examined under different conditions of solid flowrate, gas velocity and inventory. The radial voidage profile within the region of the inflection point for one set of flow conditions showed a high voidage in the center of the bed and a low voidage in the annulus of the bed. Hence the conclusion from the two papers was that there was strong radial variation and strong axial variation along the so called 'fast bed'.

In another paper, Weinstein et al. (1984) investigated the effect of particle density on the axial hold-up of solids in the fast bed. They used two powders with the same value of mean particle diameter, but whose densities differed by 35%. From their experimental data, they found two main differences: a) that the maximum solid rate was higher for the denser powder and b) that the variation of the solid fraction along the bed was smaller for the lighter solid. They offered an explanation based on the interchange of kinetic and potential energies between the gas and solid and the dissipation of the rest of the energy in the form of frictional losses. They concluded that the average solid loading in the fast bed was almost independent of solid particle density under identical operation conditions, however, they emphasized the magnitude of the variation which they said was greater for the denser solid.

It should be noted that Weinstein et al. (1984) used a
riser as part of the CFB system with no gas distributor at the lowest part of the bed, although, the U-tube connected to the riser from one side and to the slow bed from the other side, was aerated by means of nozzles.

Later Hartge et al. (1985) investigated the local hydrodynamics of two fast fluidised beds. They used a γ-ray absorption technique to measure the average solids hold-up and the data compared well with the measurements derived from pressure drop data for both fluidised beds. They then measured horizontal voidage profiles with a fibre optic probe. Since they accepted that the fast bed consisted of two distinct regions, a lower dense region and an upper dilute region, measurements of radial voidage profiles in both regions showed that in the dilute region, the profile was relatively flat with a wall zone of increased solids concentration and hence lower voidage. In the dense region of the bed, the voidage in the wall zone was as low as 0.7, while in the middle of the column the voidage was significantly higher. They therefore inferred that the segregation of gas-solid flow was present. This means that a dilute core region where the solids were transported upwards was surrounded by a denser annular region with downflow of the solids. These radial voidage profiles are comparable to the ones presented by Weinstein (1984).

Monceaux et al. (1985) reported axial pressure profiles in a CFB using Group A powders. According to these
authors, the axial pressure profile was divided into three sections, the acceleration section at the bottom of the riser, a section of constant pressure gradient ie the fully-developed flow, and a disengagement section, which only appeared at high loading ratios. In the fully-developed flow region, the pressure loss was due mainly to the solid hold-up. In this region, the radial solid flux profiles showed a downflow of solids near the wall in balance with the solid flux in the center of the riser.

All the above work supports the much earlier observations of Gajdos and Bierl (1978).

Rhodes and Geldart (1985) studied experimentally the hydrodynamics of a CFB. They presented plots of pressure gradient at different solid rates and over a range of gas velocities between 2.5 to 5m/s for a Group A powder. In order to explain the shape of the plots, they used the variation of entrainment flux E with distance h above the dense phase of the riser. The empirical equation for the rate of entrainment due to Wen an Chen (1982) was chosen to describe the flow of particles:

\[ E = E^* + (E_0 - E^*) e^{-ah} \]  

(2.17)

The value of \( E^* \), the entrainment rate above TDH was calculated from Geldart's (1979) correlation:
\[ E^* / \rho_g U = 23.7 \exp (-5.4 V_t / U) \] (2.18)

Values of \( E_0 \), the entrainment rate at the bed surface, were calculated from the correlation of Wen and Chen (1983)

\[ E_0 / A_{db} = 3.04E-9 \rho_g^{0.5} g^{0.5} / \mu^{2.5} (U-U_{mf})^{2.5} \] (2.19)

The value of the coefficient in the exponent of equation (2.17) was taken to be 4 as recommended by Wen and Chen (1982). Hence they obtained the variation of entrainment flux with distance above the bed surface at high and low gas velocities. From their plots of axial solids concentration profiles they suggested that these profiles were the result of the changes in the bed level with solid circulation rate and gas velocity. The profiles of pressure gradient represented the rise in the level height with increase in solid circulation rate. Hence the variation of entrainment rate with distance was used to explain the characteristics of fast fluidisation.

Horio et al. (1985) investigated the hydrodynamics in a CFB. Optical fibre techniques were used for the measurement of particle velocity distributions, particle concentration and cluster structures. The cross correlation method was applied to determine the particle velocity in the riser. Their results showed that in the dilute phase flow, the average particle velocity was of the order of 2.8m/s when the gas velocity was 4m/s. They suggested that
clusters were surrounded by the dilute phase where fine particles were moving faster than the particles in the clusters. However, they were not capable of distinguishing dense phase particles from dilute phase particles, so they did not present any variation of particle velocity and hence any quantitative description of the fast fluidisation condition.

Later Rhodes et al. (1987) proposed an empirical model to predict the axial distribution of solids along the riser height. They applied the Wen and Chen (1982) correlation of the form:

\[ G = E^* + (E_0 - E^*) \exp(-ah^*) \]  \hspace{1cm} (2.20)

This describes the variation of the entrainment flux with the distance above the surface of a bubbling fluidised bed which occupies a considerable distance within the riser of the CFB system. The above correlation was combined with a system pressure balance to predict the variation of voidage within the riser. Although the model was based on empirical correlations and it enables the axial solid distribution to be predicted, it might not be completely applicable to the riser conditions because it compares the fast fluidised bed with the freeboard of a bubbling bed and it underestimates the voidage in the dilute phase.

More recently, Schnitzlein et al. (1988), studied flow characteristics of fast fluidisation, by a method of instantaneous pressure signals in a fast bed.
They claimed that the mean voidage within both regions was constant and independent of height. The voidage in the dilute region was found to be independent of gas velocity and solid rate and they reported the presence of a wave within the dense region of constant speed of 1.6 m/s. Because of this wave, they considered that the solid within the dense region was present in the form of a stagnant region near to the wall (down-flowing), as an upward movement of aggregates and as a dispersed solid of an upward flowing gas.

Bolton and Davidson (1988), studied the flowrates of particles down the wall of a high velocity fluidised bed using a sampling technique. By taking as a basis the model proposed by Rhodes et al. (1987) and from experimental evidence, they claimed that the flowrate of particles in the wall region decreased exponentially with increasing height above the gas distributor. They proposed a correlation similar to that of Wen and Chen (1982) of the form:

\[ W = W_0 + dW e^{-kz} \]  

(2.21)

where \( W_0 \) and \( dW \) were constants. The exponential decay coefficient \( k \) accounted for the turbulent diffusion of particles in the freeboard since they reported that the solid layer in the wall region was due to deposition of particles from the dilute core to the wall because of the concentration difference between the two regions. Hence the diffusion mechanism was controlling the process of
backmixing of particles in a region where there was concentration equilibrium. According to this approach, they suggested a value of $k$ in the range $3.5 \text{m}^{-1}$ to $6.4 \text{m}^{-1}$ which is contrary to the low value of $0.5 \text{m}^{-1}$ presented by Rhodes (1986).

They explained the flow of the solids down the wall as being due to the geometry of the gas-solid separation system due to the increased particle concentration in the core with the increase of solid rate.

They proposed the use of an impingement device at the riser outlet in order to increase the mean solid concentration in the riser. Fortunately enough, they increased the solid concentration by increasing the depth of the bubbling bed and the backmixing of solids in the wall region was also further increased. However, they suggested that the most efficient operation of the riser of a CFB could be achieved if the depth of the bubbling bed at the bottom of the riser was reduced to zero.

Finally, they defined a fast fluidised riser as one consisting of two parts a) the lower part of the riser occupied by a violently bubbling or a slugging fluidised bed and 2) the upper part containing a core moving upwards in which the particles are entrained by the gas and a region adjacent to the wall in which a thin dense film of particles falls under gravity.

Horio et al. (1988) studied pressure distribution and solid circulation rates in a CFB with no riser distributor.
The pressure was measured by using semi-conductor sensors. They also measured cluster velocities by using a fibre optic probe technique. The conditions under investigation were: Ug in the range of 1.17 to 1.29 m/s and Gs in the range of 11.7 to 11.25 kg/m²s. Under such low superficial gas velocities in the riser section, they presented plots of the variation of the riser mean axial voidage with height and radial profiles of the variation of solid velocity. At the center of the riser the solid velocity for the above experimental conditions reached a maximum value in the range of 2.2 to 2.8 m/s, while near the wall region the solid velocity was below 2 m/s. From their measurements of radial velocity distributions, they claimed the formation of annular flow in the riser, where the thickness of annulus was not changed along the riser height and also that rates of internal and external circulation of solids were equal. They reported that for the values of Ug between 1.17 to 1.29 m/s, the cluster diameter in the dilute core was small (5 mm) and constant under the operating conditions, whereas in the annulus the size of clusters was greater than 10 mm and it became larger as the solid moved downwards in the annulus. They concluded that a) the downflow circulation rate was higher than the external rate b) by using the cluster model (Yerushalmi et al., 1978) and by changing the value of cluster voidage they claimed that the cluster size was constant in the core irrespective of height, while it was large (10 mm) in the downward flow, when a cluster voidage of 0.2 to 0.3 was used.
Brereton and Stromberg (1985) studied two aspects of the fluid mechanics in CFB, macroscopic density profiles in small and large units, and micro-structure of the flow. With regard to the first aspect, the CFB was described empirically by the following equations:

$$ P = a h^b $$

(2.22)

where: $P$ is the static pressure
$h$ is the riser height
$a, b$ are constants depending on the gas velocity, solid circulation rate and solid and riser properties.

When this equation was differentiated, it yielded the bulk density distribution:

$$ g \rho_o = \frac{dP}{dh} = - (ab) h^{b-1} $$

(2.23)

The plots of equations (2.22) and (2.23) on log/log paper gave straight lines for constant gas velocity and variable solid circulation rate. They stated clearly from equation 2.2.7 that the product '$ab$' was independent of gas velocity and was a strong function of solid rate, while the slope '$b-1$' was constant for the specific geometry of CFB.

Although the above equations were applicable to CFB with no dense phase above the distributor, they suggested that if a dense saturated zone was formed in the CFB with secondary air then the Li and Kwauk (1982) equation could be used with good agreement between the experimental data and the
predicted line. As a result of this they showed that the secondary air did not influence the basic properties of the system. What is of interest to mention is the influence of geometry with respect to the separator and the recycle leg. The above researchers claimed without showing any graphs that by using six different separator designs they obtained different pressure drops for a given circulation rate. From their visual observations they reported that the different pressure drops were due to the clusters of particles reflecting off the top of the riser, colliding with other clusters and initiating a chain of downflow.

With respect to the second aspect of their studies, the microscopic structure of a CFB was examined by studying the solids distribution at $U_g$ of 2.5m/s. They reported that at low circulation rates, there was no evidence for the two phase structure of the flow and that the solid was in the pneumatic transport regime; they calculated that the slip velocity ($U_g/c - U_p$) was twice the terminal velocity of a single particle. As the circulation rate was increased the high resolution trace of the capacitance signal showed a well defined two phase structure to be formed, where a refluxing of solids existed along the riser wall and small clusters of particles were present in the center of the column. Further studies of the radial distribution of solids in the riser using the same technique, proved that the cluster or dense phase concentration had its highest downflow value in the wall region. The explanation of the experimental evidence was due to transportation of clusters
and individual particles from the dilute core of the riser to the wall by the mechanism of turbulent diffusion, and at a critical point, the gas velocity was equal to the effective terminal velocity of the average cluster. At this point either the clusters were suspended and re-entrained into the core flow or fell along the wall as wave like sheets.

By offering this interpretation of the experimental evidence, they did not argue about the stability of the average cluster diameter because of a time averaged motion, but they suggested that the dense phase rose in an inner core and descended in an outer boundary layer.

This picture from the capacitance probe evidence, was also observed by Weinstein et al. (1984) and Bierl et al. (1980).

During the second conference on circulating fluidised bed technology (1988), the approach of most researchers was primarily to use more accurate techniques for the hydrodynamic measurements of a CFB and hence to obtain a clearer picture of the local structure of the riser. Secondly they used basic equations or well established models to explain theoretically or describe empirically the structure of two-phase flow in the riser.

Arena et al. (1988) studied and compared the hydrodynamic behavior of CFBs. Experimental techniques such as measurement of pressure gradients and slide valves
along the riser were used to determine voidage. They took
as a basis, the elutriation theory as introduced by Horio
et al. (1980) and Pemberton et al. (1983) and assumed that
the solid was entrained upwards due to the high gas
velocity and that part of it was transferred to the wall
region by turbulent diffusion. Furthermore by neglecting
the descending solid with respect to that in the ascending
zone, they reported that the voidage above the inflection
point is determined by the equation:

\[
\frac{d\varepsilon}{dz} = C \rho_g (1-\varepsilon)^2 / G_s D (g/D)^{0.5} - \\
C \rho_g (1-\varepsilon)^2 (g/D)^{0.5} / G_s
\]  

(2.24)

where: C is a constant

- \( G_s \) is the solid mass flux
- D is the riser diameter
- \( \varepsilon \) is the voidage in the dilute phase

Of course, this was an empirical equation because it
contains the term \((1-\varepsilon)\) which has to be evaluated from the
model of Li and Kwauk (1980) and even so the equation was
applicable only to the dilute part of the riser. The fact
is that it takes into account the particle density and the
riser diameter.

Rhodes et al. (1988) reported measurements of radial
and axial solid fluxes in a riser using a sampling probe
technique. Their results showed that there was an upward
and a downward local solid flux, which decreased with
increasing height in the riser. This type of solid movement was explained as being due to the transfer of solids from the dilute core of the riser to the region of low gas velocity near the wall. When the particles reached the wall region they fall down the column height producing the downward flux. They also stated that this downward flux of solid would strongly influence the gas velocity profile causing a higher gas velocity in the core. Then they argued that although there was a continuous transfer of solid from the core to the wall, at the interface between these regions there would be an exchange of particles in both directions and that since the flux decreases with height, one might expect the upward flux in the core to be exponential in form. Their results did not indicate this trend. This might be an approach to the problem. However it should be noted that neither the velocities of solid and gas nor the voidage were measured. The local structure and in particular the motion of particles is unknown. The existence of the downward flux is however not accounted for. The transfer mechanism of solid from the core to the wall has been held responsible for the downward flux and the shape of the radial gas velocity profile which has been modified from the flat turbulent gas profile to that parabolic profile of laminar flow because of the great effect of solids in the wall region. If this is true it should be expected that the downward solid flux is equal to or greater than the upward flux. However, experimental measurements have shown that at three different axial riser
positions, the upward flux was at least twice the magnitude of the downward flux (Rhodes et al. 1988).

Kefa et al. (1988), proposed a two-dimensional gas-particle flow model to explain the performance of the CFB. The model takes into account the fluctuation of particle movement which is treated as a random Fourier series based on a gas-flow fluctuation spectrum. It ignores the influence of solid flow on the gas motion and is based on the turbulent diffusion of particles due to the fluid motion. The model predicts velocity particle trajectory for a range of particle sizes and particle concentration distribution in the axial direction. However, the researchers reported only one set of experimental conditions, that is at $U_g^* = 9 \text{m/s}$, to prove the agreement between theory and experimental data. The question arises whether this gas velocity is representative of the so called "fast" bed or dilute phase transport. The circulation rate is also unknown. Hence, the approach of mathematical modeling may be good but the physical picture of the CFB is not well described experimentally and in this case the CFB was operated in the vertical pneumatic type of mode.

Son et al. (1988), studied the hydrodynamics of a CFB. It is interesting to note that these researchers measured the solid circulation rate in the CFB, then used the model of Do et al. (1972) which was applicable to the bubbling fluidised bed and its freeboard, in order to determine the
variation of particle carry-over flux with gas velocity. They claimed that good agreement existed between their experimental data and the model prediction and they recommended the use of the model for determining the solid circulation rate.

Adams (1988) studied gas mixing in a CFB. A tracer gas technique was employed for determining gas phase diffusion coefficients, based on the mechanism of eddy diffusion. From the experimental evidence, it was reported that the presence of particles greatly reduced the gas diffusion coefficient, hence he confirmed the poor gas mixing in the riser reported by Cankurt and Yerushalmi (1978) and Yang et al. (1983). Then Adams investigated the effect of the riser gas velocity and solid circulation rate on gas diffusivity. Based on the above, it was reported that the diffusion coefficient was proportional to the gas velocity as in the case of single-phase flow but that it was inversely proportional to the solid circulation rate and bed density. This was explained as being due to the fact that the presence of particles suppressed the turbulent intensity of two-phase flow according to the size and density of the particles. As a result of the above findings, one would expect in a two-phase system to record either an increase or a decrease in diffusion coefficient with solids concentration depending on the particle properties.

During the Second International Conference on CFBs (1988), Burkell et al. (1988) presented a critique of five
different techniques for determining solid circulation rates in a CFB. The five methods were:
(i) closing a permeable valve in the return leg,
(ii) timing the descent of particles in the vertical standpipe of the return system stem,
(iii) recording the force imparted by returning solids as they cascade downward from the primary cyclone,
(iv) measuring the pressure drop across a constriction in the return loop and
(v) estimating the solids flow from a heat balance on a calorimetric section in the vertical standpipe.

They investigated the above five methods by using a CFB.

Method (i) was the favored method, used previously by Yerushalmi et al. (1976), Hartge et al. (1986) and Arena et al. (1986). Burkell et al. stated that this method was acceptable for the cold model unit as long as the duration of a measurement was less than 30s. During this time interval, there was no discernible influence on the pressures in the riser.
The method (ii) involved measuring the time for identifiable particles to descend through a known distance in a section of a standpipe. This method was suggested to work well in both cold and hot circulating bed units.
The method (iii) included an impact flow-meter which was designed to measure the force of recirculating solids falling into a 'pan'. The impact flow-meter was located far enough below the primary cyclone so that falling particles
approached their terminal velocities before striking the pan. Forces transmitted through the lever and pivot were measured by a sensitive load beam. The signals obtained were seen to exhibit pulses. This was a sensitive method over a limited range of circulation rates.

The method (iv) involved the measurement of pressure drop across a constriction type flow-meter and was successfully demonstrated by Carlson et al. (1948) and Farbar (1953). Reh (1984) suggested that this technique could be adapted to measuring solid circulation rates in CFB units. The modified orifice meter did not appear to be successful under circulating bed conditions employing counter-current gas-solid motion in the return leg.

Finally method (v) used a jacketed heat transfer section in the vertical return standpipe of the high temperature CFB unit. In practice, this method was subject to uncertainty because of radial temperature gradients in the descending solids.

Takeuchi et al. (1986) examined the fluidisation state in a CFB based on data of the static pressure distribution. They presented their results in terms of pressure drop distributions along the riser height for different solid circulation rates. They concluded that when the solid circulation rate was controlled independently of the gas velocity, the state of fast fluidisation was characterized by two critical gas velocities $U_{ff}$ and $U_{dt}$. $U_{ff}$ was the minimum gas velocity to keep the solid circulation rate at
a steady level and $U_{dl}$ was the velocity below which a gap appeared in the pressure drop distribution along the height when the velocity was reduced from that for dilute transport.

Yang (1988) developed a mathematical model which described the dynamics around the CFB loop. The dilute phase region of the fast bed, including the acceleration zone, was correlated with pneumatic transport equations developed previously (Yang, 1977). A dense phase pneumatic transport model with annular region was employed to characterize the dense phase region of the fast bed. The height of the dense phase was then obtained by pressure balance around the complete circulating fluidised bed loop. It was claimed that the model provided a good description of literature data reported by Kwauk et al. (1985), Weinstein et al. (1983) and Hartge et al. (1985).

Breault and Mathur (1989) investigated a CFB loop from the pressure data measurements for the development of a mathematical model to predict the pressure profile in CFBs. The CFB loop consisted of several gas-solid flow regimes such as aerated gas-solids downflow, aerated solids downflow and gas up-flow, horizontal pneumatic transport, vertical pneumatic transport and standpipe flow. Research workers have modeled the pressure drop in each of these regimes as the sum of individual contributions due to acceleration, kinetic energy, potential energy and
frictional effects. The model required solid flowrate, gas flowrate, equipment geometry and solids fraction as input to predict the pressure at any point in the system. They claimed that the predicted pressure profiles compared well with the experimental data.

In a recent paper Kato et al. (1989) studied particle holdup and axial pressure drop in fast fluidised beds. A perforated plate was used as the gas distributor. They claimed that the fast bed consisted of a dense region of particles in the lower part of the bed and a dilute region of particles in the upper part. The height of the inflection point between the dense region of particles and the dilute region was determined by the axial pressure drop profile. The location of the inflection point was affected by the superficial gas velocity, the circulation rate of particles and the particle Reynolds number. The following empirical equation was obtained:

\[ Z = 360 \left( \frac{M_s}{\rho_p U_t} \right)^{1.2} \left( \frac{U_a - U_t}{U_t} \right)^{1.45} \left( \frac{\text{Re}_p}{\rho} \right)^{-0.2} \]  

The application range of Eqn. 2.3.1 was the following:

\[ 12 \leq M_s \leq 200, \quad 1.6 \leq U_a \leq 6.0 \]
\[ 0.45 \leq \text{Re}_p \leq 10, \quad 2.5 \leq U_a / U_t \leq 50 \]

A particle sampling technique was used to determine the particle holdup. From their experimental data, they found that the particle holdup was affected by the superficial gas velocity, circulation rate of particles, particle
Reynolds number, tube diameter and axial distance above the distributor. The following empirical equation was obtained for the particle holdup:

\[
\varepsilon/(1-\varepsilon) = 0.048 \exp(Z-Z_t)/(\exp(Z-Z_t)+1) (U_a-U_t/U_t)^{1.35} \\
x (M_s/\rho_p U_t)^{-1.28} (Re_p)^{0.29} (D_t)^{-1.28} (2.26)
\]

The equation (2.26) was applicable under the conditions:

\[
\begin{align*}
1.4 & \leq U_a \leq 6.0 & 0.1 & \leq U_t \leq 0.9 & 30 & \leq M_s \leq 130 \\
0.45 & \leq Re_p \leq 10 & 0.04 & \leq D_t \leq 0.1
\end{align*}
\]

More recently Louge and Chang (1990) developed a model to explain the discrepancy observed by Arena et al. (1985) between the voidage profiles inferred from pressure gradients and those measured by the quick-closing valve technique. The proposed model does not take into account radial profiles of voidage because it was claimed they did not contribute significantly to the transfer of momentum in the vertical direction. By neglecting the shear in the gas and solid phases the flow was considered one-dimensional. They claimed excellent agreement between the predicted voidage and Arena's voidage measurements using quick-closing valves although for their predictions the pressure gradient at a given section of the riser was required.

Ishii et al. (1989) presented the clustering annular
flow model to describe the annular flow structure in the riser of a CFB. This model was analyzed on the basis of an appropriate expression for the drag force acting on clusters and the minimum pressure gradient used previously by Nakamura and Capes (1973). In order to estimate the flow parameters including the dimensionless radius of the core, the values of both the diameter and the voidage of the cluster phase should be given. This model was based on assumptions such as:

(a) the solid suspension in the CFB consisted of clusters and a leaner continuous network

(b) the drag coefficient of the cluster was estimated from the spherical cluster approximation

(c) the particles in the dilute core were neglected in the material balance

(d) in the cluster the gas moved with the cluster and particles were uniformly suspended

(e) solids were ascending in the core and descending in the annulus

(f) clusters were homogeneously suspended both in the core and in the annulus

(g) the Richardson-Zaki equation was applicable to a cluster.

The most significant difference of the present model from that of Yerushalmi et al. (1978) was that annular flow was an essential assumption. Furthermore in this model, the voidages in the cluster were treated as variables for clusters in the core and annulus. The experimental
technique used involved optical probes for measuring cluster velocity and particle concentration in the cluster phase. From the experimental results an average cluster diameter of 10mm and a voidage of 0.8 were obtained, at operating conditions of gas velocity of 1.29 m/s and solid rate of 10.7 kg/m²s. The origin and the mechanism of formation of clusters in the annulus and the core were not clearly explained. However they stated that most of the clusters in the riser were not transported but were only fluidised. The inflection point reported by Li and Kwauk (1980) in the voidage distribution was present at the surface of fluidised clusters. Above that surface clusters of small size were transported to the column exit. They concluded that the particle friction factor gave a good prediction for the core diameter and they claimed that the core-annulus shear stress cannot be neglected from calculations of the dimensionless core diameter.

As a continuation to the theoretical work Rhodes (1990) studied radial solids flux profiles in a CFB under a wide range of operating conditions. He developed a sampling probe to measure radial and axial solids flux variations within the riser. From the experimental results he presented radial solid flux profiles at three different axial positions of the riser. These profiles have a parabolic form and exhibited strong downwards solids flow near to the riser wall. The solid flux profiles were representative of what the author terms 'the refluxing dilute-phase transport
regime'. In order to explain this regime, the author proposed a model based on the following assumptions:
a) the flow structure consisted of a core region of solids rising in the dilute phase which was surrounded by the annulus of solids falling in the dense phase suspensions
b) all the gas passes through the core, giving no net flow of gas in the annulus
c) at any axial position there was a net transfer of solids from the core to the annulus. This rate of transfer was directly proportional to the product of the concentration of solids in the core and the interfacial area between the core and the annulus regions.

The last assumption was supported by the mechanism for particles transfer from a fast moving core to the slow moving annulus. Thus the rate of transfer increased with increasing solid concentration in the core and increasing interfacial area. The interface concentration between the core and annulus was neglected.
d) It was further assumed that the solids in the core behaved as individual particles such that the slip velocity between gas and particles was equal to the single particle terminal velocity
e) the superficial gas velocity was greater than the single particle terminal velocity
f) the voidage in the core did not vary with radial positions within the core
g) the downflow velocity of solids in the annular region was determined by a balance between gravitational forces
and wall friction forces and was constant for given powder properties

h) the voidage of the downward flowing suspension in the annulus was independent of axial position in the riser
i) the gas passing through the core exhibited a laminar velocity profile
j) the downward solids flow in the annulus was evenly distributed over the annulus area.

On the basis of these assumptions the core-annulus model successfully predicted radial solids flux profiles. However it is important to mention that the model could not predict the recirculation ratio and radial flux profiles from a knowledge of gas velocity, solid circulation rate, gas and particle properties and system geometry.

A summary of the experimental conditions used by the above mentioned authors is given in the following table.
<table>
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<th>REFERENCE</th>
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2.3 CONCLUSIONS ON HYDRODYNAMICS OF CFBs

With the help of the above review of fast fluidisation, some conclusions can be drawn that will provide a clear definition of fast fluidisation itself. Fast fluidisation is the flow pattern which seems to be evident under the following conditions:

1) Solid circulation is necessary to maintain a steady state flow in the fluidisation column where the solid circulation rate strongly affects the solid hold-up.

2) The pressure gradient changes along the riser height in a dense lower section and in a dilute upper section of the bed.

3) The axial variation of voidage along the riser height shows that the dense phase occupies the lower region of the riser height and the transition to the dilute phase occurs near the top of the riser.

4) The radial voidage profile in the dense phase has a very high voidage in the centre of the riser and a low voidage in the annulus.

5) In the dilute phase, the radial voidage profile is relatively flat with a wall zone of increased solids concentration and hence a lower voidage.

Hence, the description of the riser of a CFB leads to the fact that a segregation of gas-solid flow is present so that the riser consists of a dilute core region where the particles are transported upwards and is surrounded by a denser annular region with downflow of solids.
Figure 2.1 Qualitative representation of the pressure gradients measured in the modified 15.2cm system (a) across the bottom section, (b) between pressure taps no.1 and 2 (Yesushalmi et al., 1976).
Figure 2.2 Vertical voidage distribution in fast fluidisation (Kwauk et al., 1985).
Figure 2.3 Modelling for fast fluidisation

(Kwauk et al., 1985).
Figure 2.4A Radial variation of void fraction at different velocities, open valve (Weinstein et al., 1985).

Figure 2.4B Radial variation of void fraction, constant solids rate (Weinstein et al., 1985).
CHAPTER 3
MATHEMATICAL MODELLING

In this chapter a model is presented which relates pressure drop along the axis of a circulating fluidised bed (CFB) riser to two of the main operating variables. These are;

(1) superficial gas velocity, \( u_g \) and
(2) solid circulation rate; \( R_c \) for dilute core and \( R_{CD} \) for the downflow region.

The model takes into account the effect of physical properties of particles and the gas.

With the CFB used for this study (see Chapter 4 for further details), visual observations indicated two distinct regions in the riser. Within the central core of the riser, particles are entrained by the upward flowing gas. This core is characterised by a low particle concentration (high voidage). The second region found in the immediate vicinity of the riser wall (forming an annulus) is characterised by a relatively high concentration of particles (low voidage) which fall downwards in the opposite direction to that of the gas flow. These observations have also been noted by other researchers [Bierl et al., 1980, Weinstein et al., 1984].

As a consequence of these visual observations made during the operation of a CFB, the development of the proposed model proceeds by examining each region in turn. The forces on single particles and clusters in both regions are examined. This is then followed by the
examination of the pressure drops in each region
and finally the total pressure drop for the riser is
considered.

### 3.1 FORCES ON A SINGLE PARTICLE

Consider a spherical particle at some axial distance x
from the base of the riser (Figure 3.1). Neglecting the
effect of buoyancy, there are two forces that will
influence the motion of the particle: (i) the force of
gravity, mg, acting downwards, and (ii) the drag force $F_D$,
acting upwards caused by the relative speed between the air
and the particle.

![Diagram of forces on a single particle](image)

**FIGURE 3.1 FORCES ON A SINGLE PARTICLE**
At a distance \( x \),

\[
m \left( \frac{d^2 x}{dt^2} \right) = mg - F_D
\]

3.1.1

taking the convention that downwards is the positive \( x \)-direction. In eqn. 3.1.1, \( m \) is the mass of the particle.

The drag force may be written in terms of a drag coefficient, \( c_D \). Thus,

\[
F_D = \frac{1}{2} \pi \left( \frac{d_p}{2} \right)^2 u^2 c_D \rho_g
\]

3.1.2

where: \( u \) is the slip velocity

\( d_p \) is the particle diameter

\( c_D \) is the drag coefficient, and

\( \rho_g \) is the gas density

\[
u = -(u_g/\varepsilon - u_p)
\]

\[
c_D = f (Re_p)
\]

3.1.3

For small particles, \( c_D \) can be approximated by its value in the Stoke's regime, calculated from:

\[
c_D = \frac{24}{Re_p}
\]

3.1.4

where:
\[
\text{Re}_p = \frac{\rho_f u d_p}{\mu_f} \tag{3.1.5}
\]

For particles of the order of 100\(\mu\)m in diameter and slip velocity of the order of 1m/s in air, the particle Reynolds number is about 1.

Substituting for \(\text{Re}_p\) in eqn. 3.1.4 from eqn. 3.1.5, then

\[
c_d = \frac{24 \mu_f}{\rho_f u d_p} \tag{3.1.6}
\]

Replacing \(c_d\) in eqn. 3.1.2 using eqn. 3.1.6 gives:

\[
F_D = 3 \pi d_p \mu_f u \tag{3.1.7}
\]

Substituting for \(F_D\) into eqn. 3.1.1

\[
m \frac{d^2x}{dt^2} = mg - 3 \pi d_p \mu_f u \tag{3.1.8}
\]

Dividing by \(m\)

\[
\frac{d^2x}{dt^2} = g - \frac{3\pi d_p \mu_f u}{m} \tag{3.1.9}
\]

Noting that

\[
m = \pi \frac{d_p^3}{6} \rho_p \tag{3.1.10}
\]

and substituting into eqn. 3.1.9 for \(m\) using eqn. 3.1.10 and replacing the slip velocity \(u\), with eqn. 3.1.3 and
rearranging, gives

\[
\frac{d^2x}{dt^2} = g - \frac{18\mu_f (u - u_g)}{d_p \rho_p}
\]

Noting that \( u = \frac{dx}{dt} \)

Hence,

\[
\frac{d^2x}{dt^2} + \left( \frac{18\mu_f}{d_p \rho_p} \right) \frac{dx}{dt} - \left( \left[ \frac{18\mu_f u_g}{d_p \rho_p} \right] + g \right) = 0
\]

Letting \( A = \frac{18 \mu_f}{d_p \rho_p} \),

eqn. 3.1.12 reduces to

\[
\frac{d^2x}{dt^2} = -A \frac{dx}{dt} + (A u_g + g)
\]

Eqn. 3.1.13 is applicable both in the bulk of the riser where the particles are entrained by the air stream and in the region of the riser immediately adjacent to the vertical walls where the solids are flowing downwards. The differences between the solutions for the two regions are due to the initial conditions. Knowing the initial boundary conditions and \( u_g \), eqn. 3.1.13 can be solved for each region using a numerical method, outlined below.
To solve eqn. 3.1.13 for the bulk of the riser it is necessary to calculate the initial particle velocity at \( t=0 \) and \( x=0 \)(see Fig. 3.1). This was assumed to be equal to the terminal velocity of the particles i.e. \( u_p = 0.2 \text{ m/s} \).

Additionally, it is also assumed that \( \Delta x/\Delta t \rightarrow dx/dt \) as \( \Delta x \rightarrow 0 \). In this case \( \Delta t \) will be incremented by steps of 0.01s. Using smaller increments of time would increase the accuracy of the prediction at the expense of computational time.

The general algorithm for the Solution of eqn. 3.1.13 is as follows:

1. Set \( u_g, \rho_p, \mu_f, d_p \) and \( D_T \)
2. Calculate \( A \) from eqn. 3.1.12a
3. Set initial value for \( dx/dt \) at \( t=0, x=0 \) \( (u_p = 0.2 \text{ m/s}) \)
4. Calculate \( d^2x/dt^2 \) from eqn. 3.1.13
5. Calculate \( \Delta x \) \((dx/dt \times \Delta t) \) i.e. \((dx/dt)0.01\)
6. Calculate the new value of \( dx/dt=\Delta x/n\Delta t \) where \( n=2,3,4... \)
7. Hence, go to 4

The results of this numerical solution are shown in Fig. 3.2 for the FCC powder and for \( u_g \) of 3 m/s. At first the particles appear to fall by a small distance before rising upwards with the gas. After about 0.1 s, corresponding to a distance of about 0.12 m, the particles appear to reach its terminal slip velocity \( u_{ts} \); \( u_{ts}=0.2 \text{ m/s} \), calculated from standard correlations:

\[
u_{ts} = \frac{d_g^2 (\rho_p - \rho_f)}{18\mu} \quad \text{(Shamlou, 1988)}.
\]
3.2 FORCES ON A PARTICLE IN AN ASSEMBLY

The presence of other particles will modify the effective drag coefficient for a single particle in the gas flow. Work by Wen and Yu (1966) and Wen and Galli (1971) has indicated that this modification can be adequately described by:

\[ c'_D = c_D \varepsilon^{-4.7} \] 3.2.1

where \( c_D \) is the single particle drag coefficient predicted by eqn. 3.1.4. Combining eqns. 3.1.4, 3.1.12a and 3.2.1 it can be shown that:

\[ \frac{d^2 x}{dt^2} = -A \varepsilon^{-4.7} \frac{dx}{dt} + (A \varepsilon^{-4.7}u + g) \] 3.2.2

\( \varepsilon \) can be obtained by dividing the solid superficial velocity by its linear velocity. Thus:

\[ 1 - \varepsilon = \frac{V_c}{u_p} = \frac{V_c}{dx/dt} \] 3.2.3

where \( V_c \) is the superficial solid velocity (m/s)

\[ V_c = \frac{R_c}{\rho_p A} = \frac{4R_c}{\rho_p \pi D_T^2} \] 3.2.4

where: \( D_T \) is the riser diameter (m)

\( \rho_p \) is the particle density (kg/m³)

In eqn. 3.2.4, \( R_c \) is the solid circulation rate.

For the dilute core, \( R_c \) is obtained from experimental
measurements of solid flowrate. These measurements indicated that $R_c$ is a function of the gas velocity and the ball valve opening. It was found that $R_c$ varied in the range of $3 \times 10^{-3}$ kg/s to $30 \times 10^{-3}$ kg/s for gas velocities in the range 2-4 m/s and ball valve openings between 40° to 70°.

The numerical solution of eqn. 3.2.2 for the dilute core is similar to that described previously for eqn. 3.1.13, except that the values of the solid circulation rate, $R_c$, and column diameter, $D_T$, (eqn. 3.2.4) are now part of the input data. The algorithm for the solution of eqn. 3.2.2 is as follows:

1. Set $u_g, \rho_p, \mu_f, d_p, D_T$, and $R_c$
2. Calculate $A$ from eqn. 3.1.12a
3. Set initial values for $dx/dt$ at $t=0, x=0$ ($u_p = 0.2$ m/s)
4. Calculate $d^2x/dt^2$ from eqn. 3.1.13
5. Calculate $V_c$ using eqn. 3.2.4
6. Calculate voidage $\varepsilon$ using eqn. 3.2.3
7. Calculate $\Delta x$ ($dx/dt \times \Delta t$) i.e. $(dx/dt) 0.01$
8. Calculate the new value of $dx/dt = \Delta x/n\Delta t$ where $n=2,3,4...$
9. Hence go to 4

Figures 3.3 and 3.3A show calculated values of the voidage plotted against riser height with gas velocity and particle diameter as variables. Evidently, the voidage and particle velocity reach their steady state values within the first few centimeters after the point of entry. The voidage appears to be relatively insensitive to the changes in particle diameter. Figure 3.3B shows predicted values of
the voidage plotted against riser height with the gas velocity and particle density as variables.

For the downflow region, the solution of eqn. 3.2.2 is complicated by two factors. Firstly, close to the vertical wall of the riser, calculations (see Appendix C) show that particles are likely to fall not as individual primary grains, but rather as clusters. As far as the solution of eqn. 3.2.2 is concerned, this implies that the magnitude of parameter $A$ (eqn. 3.1.12a) is determined by cluster diameter, $D_c$, and not by primary particle diameter $d_p$. Unfortunately, it is not possible at this stage to estimate the size of these clusters and therefore for the purpose of solving eqn. 3.2.2 a trial and error procedure has to be adopted. Secondly, solution of eqn. 3.2.2 for the downflow region requires information about solid circulation rates $R_{CD}$ near the wall. This is obtained as shown below.

**Calculation of solid circulation rate in the downflow region**

At time, $t=0$, and for a fixed ball valve opening position, setting the gas velocity to a pre-determined value will cause a change in the height of the material in the slow bed. For a given gas velocity, let the height of material in the slow bed fall by $\Delta H_{SB}$ (see Fig. 3.1). This means that during the steady-state operation the mass of solids in motion is given by:

$$\Delta M_{SB} = \Delta H_{SB} \frac{\pi D_{SB}^2}{4} \rho_{SB}$$  \hspace{1cm} (3.2.5)
where:

\[ \rho_{SB} = (1 - \varepsilon_{SB}) \rho_p + \varepsilon_{SB} \rho_g \]  

\( \varepsilon_{SB} \) is the voidage corresponding to the point of minimum fluidisation.

If the total volume \( V_T \) of the system is known, and the volume of the riser \( V_R \) is known then as an approximation the mass of the solids in the riser, \( M_R \), will be:

\[ M_R = \frac{V_R}{V_T} \Delta M_{SB} \]  

From a simple mass balance:

\[ M_R = M_{DC} + M_D \]  

where: \( M_R \) is the total solid hold-up in the riser  
\( M_{DC} \) is the solid hold-up in the dilute core  
\( M_D \) is the solid hold-up in the downflow.

If it is also assumed that all the solid initially goes to the dilute core region, then the solid hold-up in that zone is equal to the product of solid circulation rate and its residence time. Since the acceleration zone is relatively short (see Fig. 3.2), the residence time, \( t_R \), may be estimated by using:

\[ t_R = \frac{H_R}{u_{slip}} \]  

where: \( u_{slip} = u_g - u_i \)

Thus, the solids hold-up in the dilute core is given by:
Substituting for $M_{DC}$ and $M_R$ in eqn. 3.2.7a using eqns. 3.2.9 and 3.2.8 gives:

$$M_D = M_R - M_{DC} = \frac{V_R}{V_T} \Delta M_{SB} - \frac{R_C H_R}{u_{\text{slip}}}$$  \hspace{1cm} 3.2.10

To calculate the mass flowrate of solids, $R_{CD}$, the residence time of solids in the downflow region must be known. Assuming an estimated mean value of 1m/s based on visual observations (Bolton, 1988) for the solids velocity in the downflow region, the residence time in that zone will be given by (Section 3.4):

$$t_{CD} = \frac{H_R}{1}$$  \hspace{1cm} 3.2.11

and thus $R_{CD}$ is given by:

$$R_{CD} = \frac{M_D}{t_{CD}} = \frac{M_D}{H_R} = \frac{M_D}{H_R}$$  \hspace{1cm} 3.2.12

The difference between the initial and steady-state height in the slow bed, $\Delta H_{SB}$, has to be known in order to use eqn. 3.2.5. The variation of $\Delta H_{SB}$ was measured experimentally as a function of gas velocity and ball valve opening.

Equation 3.2.12 provides the value of the solid
circulation rate in the downflow region. This together with eqn. 3.2.4 and an estimated value of cluster diameter are used to solve eqn. 3.2.2 in order to obtain the voidage and solid velocity profiles in the downflow region. As explained previously, the solution of eqn. 3.2.2 involves a trial and error procedure involving cluster diameter \( D_c \).

The numerical solution adopted involves solving eqn. 3.2.2 for an initial guess of \( D_c \). Having obtained the voidage profile distribution for the downflow as well as for the dilute core, the computer program then calculates the overall pressure drop profile for the riser (see next section). The predicted pressure is then compared (visually) with the measured pressure distribution for the same operating conditions. If the two profiles do not match then a new value of \( D_c \) is used and the whole procedure repeated, until the calculated pressure profile adequately matches the measured distribution. The voidage profile and solid velocity distribution in the downflow region are then plotted routinely as part of the solution.

Figures 3.4 and 3.5 shows plots of solid velocity and voidage in the downflow region. It is interesting to note that the voidage in the downflow region (Fig.3.5) is lower than that for the dilute core.

3.3 PREDICTION OF PRESSURE DROP IN THE RISER

The pressure losses in the riser have four origins: (1) particle acceleration (2) change in static pressure (3) gas-wall friction and (4) solid-wall friction.
The pressure drop due to particle acceleration, although is calculated by the program, is relatively small compared to the other terms in eqn. 3.3.1

The pressure drop due to the static head of bulk material may be split into the contributions of particles and fluid. Thus

$$\Delta P_{\text{STAT}} = \rho_p (1-\varepsilon) g \Delta x + \rho_g \varepsilon g \Delta x \quad 3.3.2$$

Calculation show that, the extra pressure contribution due to the gas flow alone ($\rho_g \varepsilon \Delta x$) is significantly smaller than that due to the solid particles. While in practice the contribution due to gas flow alone may be ignored, it was routinely calculated and its contribution was included in this analysis.

Using the Fanning friction factor versus Reynolds number relationship for turbulent flow, the pressure drop due to the gas-wall friction can be calculated in the
standard way (Kay, 1960):

\[ \Delta P_{aw} = 2 \, c_f \, \rho_g \, \Delta x \, \frac{u_p^2}{D_T} \]  

The pressure drop due to the flow of solids is obtained using a solid-wall friction factor (Zenz and Othmer, 1960):

\[ \Delta P_{sw} = 2 \, f_s \, \rho_p \, (1-\varepsilon) \, \Delta x \, \frac{u_p^2}{D_T} \]  

where: \( f_s \) is the solid-wall friction factor similar to gas-wall friction factor. \( f_s \) is normally obtained from experimental data. For the purpose of this discussion the following correlation is used (Breault et al., 1989):

\[ f_s = 12.2 \, \frac{(1-\varepsilon)}{\varepsilon^3} \, \frac{u_p^2}{D_T} \]  

The local values of \( \varepsilon \) in eqns. 3.3.2 and 3.3.5 and the magnitude of the local particle velocity \( u_p \) in eqn. 3.3.5 are obtained from the solution of eqn. 3.2.2. for the dilute core and for the downflow regions. The total pressure drop is calculated by adding the pressure drops for the dilute core and the downflow region.

\[ \Delta P_{\text{TOTAL}} = \Delta P_{\text{DILUTE}} + \Delta P_{\text{DOWNFLOW}} \]  

Both \( \Delta P_{\text{DILUTE}} \) and \( \Delta P_{\text{DOWNFLOW}} \) are obtained by using eqns. 3.3.2 to 3.3.5. For each region local \( \varepsilon \) and \( u_p \) are obtained by solving eqn. 3.2.2 as described previously.

Figures 3.6 to 3.11 show predicted pressure profiles with an initial solid velocity in the downflow region of 1m/s and for a range of assumed cluster diameters (eqn.3.3.1). Small changes in cluster diameter as shown in
Figs. 3.6 and 3.7 make significant differences for the solution of eqn. 3.2.2 and the subsequent pressure drop calculations.

Typical plots of the pressure distribution in the dilute core, the downflow region and the total pressure for three gas velocities (4, 3.5 and 2.5 m/s) for the FCC powder are presented in Figs. 3.9 to 3.11 for a mean solid velocity of 1 m/s. Figure 3.12 shows an example of predicted pressure profiles in the dilute core, downflow region and total pressure with a mean downflow solid velocity of 0.7 m/s. This difference (from 1 m/s to 0.7 m/s) has resulted in lowering the pressure drops in the downflow region by 20%.

Appendix (F) contains the algorithm and listing of the computer program that has been developed for predicting the pressure profile in the riser.

Chapter 4 contains details of experiments conducted in this study. The results of these experiments are discussed in Chapter 5 and also compared with the predictions made using the model developed in this Chapter.
FIGURE 3.2 PLOT OF SOLID VELOCITY VERSUS TIME ACCORDING TO EQ. 3.1.13

\[ u = 3 \text{ m/s} \]
\[ d_p = 64 \mu \text{m} \]
\[ \rho_p = 884 \text{ kg/m}^3 \]
\[ \mu_f = 10^{-5} \text{ kg/ms} \]
Figure 3.3 Theoretical voidage of dilute core versus riser height

\[ D_T = 0.05 \text{ m} \]

FCC

BVO = 70°

\[ d_p = 64 \mu \text{m} \]

\[ \text{Re} = 27 \times 10^{-3} \text{ kg/s} \quad \text{Re} = 24 \times 10^{-3} \text{ kg/s} \quad \text{Re} = 17 \times 10^{-3} \text{ kg/s} \]

\[ \text{Ug} = 4 \text{ m/s} \quad \text{Ug} = 3.5 \text{ m/s} \quad \text{Ug} = 2.5 \text{ m/s} \]
FIGURE 3.3A Theoretical Voidage of dilute core versus riser height

\[
\begin{align*}
\text{FCC} \\
\rho_p &= 884 \text{kg/m}^3 \\
\text{BVO} &= 70^\circ \\
\end{align*}
\]

\[
\begin{align*}
\text{assumed} \\
R_e &= 27 \times 10^{-3} \text{ kg/s} \\
R_e &= 24 \times 10^{-3} \text{ kg/s} \\
R_e &= 17 \times 10^{-3} \text{ kg/s} \\
\end{align*}
\]

\[
\begin{align*}
U_g &= 4 \text{ m/s} \\
U_g &= 3.5 \text{ m/s} \\
U_g &= 2.5 \text{ m/s} \\
\end{align*}
\]

point of solid entry

HEIGHT (m)
FIGURE 3.3B Theoretical voidage of dilute core versus riser height

FCC

\[ \theta = 70^\circ \]

\[ \bar{d} = 127 \mu m \]

\[ \bar{\rho} = \frac{1840 \text{ kg/m}^3}{\text{kg/m}^3} \]

- \( Re = 27 \times 10^{-3} \text{ kg/s} \)
- \( \text{Ug} = 4 \text{ m/s} \)
- \( Re = 24 \times 10^{-3} \text{ kg/s} \)
- \( \text{Ug} = 3.5 \text{ m/s} \)
Figure 3.4 Theoretical solid velocity of downflow region versus riser height

Solid Velocity (m/s)

- FCC
- BVO=70°

\[ D_c = 4.4312 \times 10^{-4} \text{ m} \quad \text{Ug} = 4 \text{ m/s} \]

\[ D_c = 2.2927 \times 10^{-4} \text{ m} \quad \text{Ug} = 3.5 \text{ m/s} \]
Figure 3.5 Theoretical voidage of downflow region versus riser height

- FCC
- BVO=70°

- $D_c=4.4312E-4 \text{ m}$, $U_g = 4\text{ m/s}$
- $D_c=2.2927E-4 \text{ m}$, $U_g = 3.5\text{ m/s}$
- $D_c=3.8168E-4 \text{ m}$, $U_g = 2.5\text{ m/s}$

Voidage versus Height (m)
Figure 3.6 Plot of pressure gradient versus riser height

\( U_g = 4 \text{ m/s} \)
\( \Delta \rho = 0.3 \text{ kg/m}^3 \)
\( \alpha = 70^\circ \)

Pressure gradient (Pa/m)

Height (m)
Figure 3.7 Plot of pressure Gradient versus riser height

FCC

Ug = 2.5 m/s

Dc = 3.8168E-4 m

BVO = 70°
Figure 3.8 Plot of pressure gradient versus riser height

FCC

$U_g = 3.5 \text{ m/s}$

$D_c = 4.2927 \times 10^{-4} \text{ m}$

$BVO = 70^\circ$
FIGURE 3.9 Pressure gradient versus riser height

FCC
Ug=4m/s

- Dilute core
- - - - Downflow region
- - - - Total pressure

BVO = 70°
D_c=4.4312E-4 m
FIGURE 3.10 Pressure gradient versus riser height

FCC
Ug=3.5 m/s

- - - - Dilute core
- - - - Downflow region
- - - - Total pressure

BVO=70°
D_m=2.2927E-4 m
FIGURE 3.11 Pressure gradient versus riser height

FCC
Ug=2.5 m/s

- - - - Dilute core
- - - - Downflow region
- - - - Total pressure

BVO = 70°
D_c=3.8168E-4 m
FIGURE 3.12 Pressure gradient versus riser height

- FCC
- Dashed line: Dilute core
- Dotted line: Downflow region
- Dashed-dotted line: Total pressure

- \( U_g = 4 \text{ m/s} \)
- \( \text{BVO} = 70^\circ \)
- \( D_c = 4.306 \times 10^{-4} \text{ m} \)
Previous research concerning circulating fluidised beds has usually involved gas velocities in the range 1 - 5m/s for Groups A and B powders. The purpose of this experimental study was to extend the range of gas velocity to 4m/s safely entering the fast fluidisation regime, and measure the corresponding pressure profile along the height of the riser.

The pressure drop measurements yielded data that can be used for both qualitative and quantitative comparisons with previous work, as well as providing a means by which to test the "Dense Annulus/Lean Core" model proposed in Chapter Three.

Apart from the gas velocity, the other important independent variable is the opening of the ball valve between the slow bed and the riser. Experiments have been conducted to measure the solid circulation rate, and the changes in the slow bed inventory, as well as the pressure drop for various combinations of gas velocity and ball valve openings.

Data has also been collected to determine the performance of the CFB when using a 50/50 v/v mixture of two different powders.
4.1 THE CIRCULATING FLUIDISED BED SYSTEM

The design of the circulating fluidised bed apparatus is schematically shown in Figure 4.1.1, (front elevation and side elevation).

It consisted of a 50mm diameter, 3m high fast fluidised column, the riser, a solid collecting system and 150mm diameter bubbling bed, the slow bed. The riser was composed of two perspex sections, each 1.5m in height connected to each other by flanges. Pressure tappings located at 0.18m intervals were provided along the riser. The slow bed, also made of perspex, had the purpose of storing solids and providing the force for solid circulation in the system. Fluidised air for the slow bed fed through a porous plate distributor at the side of the bed in order to maintain the solids in the state of minimum fluidisation. Solids fed to the riser by means of a plastic pipe 25 mm in diameter inclined at an angle of 30 degrees. In general the solid circulation rate to the riser was determined by the ball valve installed between the riser and the slow bed as well as by the gas velocity. Three different riser gas distributors were used.

The top of the riser was connected to the inlet of the primary cyclone by a standard elbow. The primary cyclone was designed according to the Stairmand (1959) high efficiency (H.E.) design (see Appendix E) for an operation of 10m/s riser velocity and by assuming that the riser cross-sectional area and the cyclone inlet area are the
FIGURE 4.1.1 DESIGN OF THE CIRCULATING FLUIDISED BED
same. Solids leaving the primary cyclone were directed either into the slow bed or the load cell system. Diverting two-way valves were placed along the discharge lines of both cyclones to allow accurate solid flowrate measurements to be made. Gas leaving the primary cyclone was fed into the secondary cyclone which was also based on the Stairmand H.E. design by considering that the cyclone efficiency may be improved if the pressure drop inside this cyclone does not exceed the upper pressure drop limit of 1200 N/m², as recommended by Svarovsky (1986).

Both cyclones were fabricated from copper. Gas leaving the secondary cyclone was discharged into a duct.

The air was taken from the laboratory compressed air supply and air flowrates were measured using rotameters.

The design of the ball valve (Figure 4.1.2) was such that between 0 and 15 degrees there was no opening and consequently no possible solids flow. There was need to determine the open area of the ball valve at a particular angle \( \phi \) of the stem. The ball valve was fully open at 90 degrees (\( \phi_{\text{max}} \)). The minimum opening was 15 degrees (\( \phi_{\text{min}} \)). Diameter of ball valve, \( D_v = 25 \text{ mm} \). Assume that

\[
\frac{D_v}{90 - \phi_{\text{min}}} = 2x
\]

Then \( 25/75 = 2x = 0.33 \text{ mm/degree} \)
Hence \( x = 0.167 \text{ mm/degree} \)

**CASE I** \( 2x < R \)

For an angle \( \phi > \phi_{\text{min}} \)

1) Find \( 2x \)

2) Find distance \( y = R - 2x \)
3) Find \( \cos \theta = \frac{y + x}{R} \)

4) Find \( \theta \) in degrees

5) Calculate sector area \( \frac{2\theta}{360} \pi R^2 \)

6) Calculate \( z = (R^2 - (x+y)^2)^{0.5} \)

7) Calculate open area

\[
A_{op} = 2 \left( \frac{2\theta}{360} \pi R^2 - \frac{z(x+y)}{2} \right) \tag{4.1}
\]

**CASE II 2x > R**

Radius of ball always constant.

Symmetry applicable.

\( x = 0.167 \text{ mm/degree} \)

\( y = 2x - R \)

Use the same algorithm to determine the open area.

For an angle \( \phi > \phi_{\text{min}} \), \( \phi > 15 \)

(1) Find 2x

(2) Calculate \( y = 2x - R \)

(3) Calculate \( z = R^2 - (x-y)^2 \)

(4) Find \( \sin \theta = z/R \)

(5) Find \( \theta \) in degrees

(6) Calculate the sector area \( \frac{2\theta}{360} \pi R^2 \)

(7) Calculate the open area

\[
A_{op} = 2 \left( \frac{2\theta}{360} \pi R^2 - z(x+y) \right) \tag{4.2}
\]

\[
2x = \frac{0.167}{90-\phi}
\]

The open area of the valve, \( A_{\text{op}} \), is related to the angle \( \phi \),
from equations 4.1 and 4.2. A non-linear regression of the calculated values of $A_v$, gives the following expression:

$$A_v = 1.29 \, (\phi)^{1.38}$$  \hspace{1cm} 4.3

Above 15 degrees the area of the opening can be calculated from equation 4.3. For the range of gas velocities used in this study, operation of the CFB with a ball valve opening greater than 70 degrees, led to an unstable situation caused by flow reversal of solids through the ball valve.

FIGURE 4.1.2 DESIGN OF THE BALL VALVE
4.2 POWDERS USED

The powders used in the 50mm diameter riser of the circulating fluidised bed were; calcined FCC silica alumina, Paralox and a 50/50 v/v mixture of the two. The mean particle sizes were determined by using a sieve analysis. The densities of particles were measured using mercury pychnometry. This method involved the following steps: a) cells were filled with mercury and weighted b) cells were emptied and cleaned and a known weight of sample was introduced to the cell c) the cell with the sample was placed under vacuum to exclude all the air d) mercury was introduced into the cell by releasing the vacuum e) the cells were then re-weighted and f) the density of the sample $\rho_s$ was estimated as following:

$$\rho_s = \frac{W_s}{W_s - (W_{sHg} - W_{Hg})}$$

where: $W_s$ is the weight of the sample

$W_{sHg}$ is the weight of the sample and mercury

$W_{Hg}$ is the weight of mercury.

Values of relevant particle properties are given in TABLE 4.2.1.
4.3 PRESSURE DROP EXPERIMENTS

4.4.1 APPARATUS

For this series of experiments, the circulating fluidised bed described in section 4.1, was used.

4.4.2 MATERIALS

The pressure drop experiments were conducted with three different powders all of them belonging to Group A according to Geldart's classification. The properties of these powders are shown in Table 4.2.1

4.4.3 PROCEDURE

The riser was first emptied of solids by conveying any remaining powder into the slow bed by supplying a large air velocity. The required air velocity, $U_g$, for the experiment was then set constant using one of the rotameter valves, and the ball valve was set to a particular angular opening. This allowed a flow of solids to run into riser. The system was then allowed to settle to a steady state which deemed to exist when the solid height in the slow bed shows no variation with time.

<table>
<thead>
<tr>
<th></th>
<th>FCC</th>
<th>PORALOX</th>
</tr>
</thead>
<tbody>
<tr>
<td>$d_p, (\mu m)$</td>
<td>64</td>
<td>127</td>
</tr>
<tr>
<td>$\rho_p, (kg/m^3)$</td>
<td>884</td>
<td>1850</td>
</tr>
<tr>
<td>$U_{mf}, (m/s)$</td>
<td>1.36E-2</td>
<td>1.6E-2</td>
</tr>
</tbody>
</table>

TABLE 4.2.1.
Properties of powder used
Normally, when making any change to the operation of the CFB (either by altering $U_g$ or ball valve opening), approximately half an hour was required before the system approached steady state.

While the system approached a new steady state, pressures within the riser cause backpressures on the piping upstream of the distributor, causing fluctuations to the rotameter reading. This required slight adjustments to be made to ensure that the desired flowrate of air was admitted to the riser.

Confirmation that the steady state been achieved came from monitoring manometer levels and the height of the slow bed. After steady state achieved, the pressure profile determined by recording each of the 15 pressure tappings along the riser.

The pressures were detected using U-tube manometers filled with water dyed with rhodamine. The difference in the water levels in each arm of the U-tube was measured with a graduated scale with measurements made to the nearest millimeter.

4.4 SOLID CIRCULATION RATE EXPERIMENTS

4.4.1 APPARATUS

For this series of experiments, the circulating fluidised bed described in section 4.1, together with the load cell system used.

4.4.2 MATERIALS
The solids used for these sets of experiments are shown in Table 4.2.1.

4.4.3 PROCEDURE

With a desired gas velocity $U_g$, and ball valve opening $\phi_{BVO}$, the CFB allowed to achieve steady state. The two-way valves (see Figure 4.1.1) manipulated to divert the solids through flexible tubing to a vessel supported by a platform which in turn formed part of the load cell apparatus, shown in Figure 4.4.1.

The deflection caused by the weight on the load cell measured by a chart-recorder. This recorder was previously calibrated using the procedure explained in section 4.5. Knowing the weight on the load cell and the time that had elapsed since the solids were diverted then the rate at which the solids circulated through the CFB was estimated. It should be noted that the weight recorded by the load cell was also a function of the distance between the cell and the pivot. This means that the weight has to be divided by 5.

The diversion of solids via the two-way valve disrupted the steady state operation of the CFB, because of the intrusive nature of the sampling technique. This means that a balance had to be achieved so that the diversion of solids took place for a long enough period to record a reproducible deflection on the chart-recorder, but whilst also ensuring that the diversion should not continue for so long that the CFB allowed to deviate markedly from its original steady state condition. In practice, this
diversion of solids was allowed for approximately 15 seconds.

4.5 CALIBRATION OF LOAD CELL

The load cell was connected to the chart-recorder. The chart-recorder was set up to 10mV and 30mm/min. In order to obtain a direct calibration of the load cell, a known weight was added periodically on top of the vessel, and the deflection caused by the additional weight was recorded on paper.

Readings of the chart-recorder paper produce the deflection of weight in terms of mV, mm of paper and time.

A graph of the chart-recorder readings in millivolts is plotted against the known weight in grams. The straight line equation is given by:

\[(\text{mV}) = A \times (\text{Weight}) + C\]

A is the slope of the line

\[A = 2.91E-3 \text{ mV/g}\]

C is the intercept, set to zero by adjusting the chart-recorder and corresponds to the value of the vessel weight.

SPECIFICATION OF THE LOAD CELL

The load cell made within the Chemical Engineering Department. It contained strain guages which respond to changes in the electrical resistance of the metal within the load cell.

The maximum load of the load cell was estimated to
correspond to 16mV equivalent to 0.9 kg weight.

4.6 RISER DISTRIBUTOR DESIGNS

Three different designs of riser distributor were used with the same fraction of free open area of 4%, but with a variable number of orifices.

Riser Distributor No.1: 1 nozzle centrally located of

\[ \text{id=}10\text{mm.} \]

Riser Distributor No.2: 7 orifices, each one of

\[ \text{id=}3.8\text{mm, arranged in a triangular pitch.} \]

Riser Distributor No.3: 13 orifices, each one of

\[ \text{id=}2.8\text{mm, arranged in a triangular pitch.} \]

Designs of the riser distributors 1, 2 & 3 are presented in Figure 4.6.1.
correspond to 16mV equivalent to 0.9 kg weight.

4.6 RISER DISTRIBUTOR DESIGNS

Three different designs of riser distributor were used with the same fraction of free open area of 4%, but with a variable number of orifices.

Riser Distributor No.1: 1 nozzle centrally located of id=10mm.

Riser Distributor No.2: 7 orifices, each one of id=3.8mm, arranged in a triangular pitch.

Riser Distributor No.3: 13 orifices, each one of id=2.8mm, arranged in an triangular pitch.

Designs of the riser distributors 1, 2 & 3 are presented in Figure 4.6.1.
CHAPTER 5
RESULTS AND DISCUSSION

5.1 INTRODUCTION

This chapter describes experiments carried out in a circulating fluidised bed to investigate (1) the effect of the inventory on the solid circulation rate, slow bed height, pressure profiles and voidage and (2) the effect on riser pressure drop and the solid circulation rate of the four main design variables of the system:

a) Distributor Design
b) Gas Velocity
c) Ball Valve Opening and
d) Particle Properties.

The results of these experiments are then compared with the "Dense Annulus/Lean Core" model developed in Chapter Three to test its accuracy and predictions. The model is then applied to the case in which a first order reaction is taking place in the riser of a CFB.

5.1.1 Effect of bed inventory on solids circulation rates and slow bed height.

The solids circulation rates and slow bed heights measured at each of the five ball valve openings are plotted in Figures 5.1.1.1 and 5.1.1.2. Four bed inventory levels in the range of 21 to 60 cm were used and for each inventory and constant gas velocity, the solids circulation rates and slow bed heights were measured.
At constant slow bed inventory level as the opening of the ball valve increases the solid circulation rate (SCR) also increases. This increase of SCR is significantly higher at the inventory levels of 45 and 60 cm.

The slow bed height at a constant inventory level decreases slowly by increasing the ball valve opening. The change in the slow bed height, from its initial position, has a mean average value of 0.1 cm/degree for the four different inventory levels. Although the decrease of the slow bed height is small for the four different inventory levels, at a nearly constant value, the inventory levels clearly affect the flow in the riser and increase the solid circulation rate.

5.1.2 Effect of inventory and ball valve opening (BVO) on pressure profiles around the system

The pressure profiles around the fluidisation system may be generalised into three diagrams as shown in Figures 5.1.2.1 a-c.

Three different slow bed inventories were chosen in the range of 33 to 60 cm and four openings of the ball valve, at a constant gas velocity of 2.2 m/s.

When the ball valve is fully opened (BVO=90) the pressure profiles around the system for the three different inventories are of the same form as is shown in Figure 5.1.2.1b.

The higher pressure drop is located just below the ball valve due to acceleration effects that are present at the
bottom of the column. On the other hand, the pressure drop in the exit section of the bed decreases with the increase of inventory. Hence the bed height in the riser increases. If however, the gas velocity in the riser is increased to higher values at constant BVO and inventory this height will reduce to zero not only because of the flow conditions but also because of the constraints of the equipment.

A completely different picture is now presented, Figure 5.1.2.1a and 5.1.2.1c, if the solid flow is restricted by means of the ball valve.

At the lower inventory levels (33 or 45 cm) and low openings of the ball valve (50 or 60) the system behaves as shown in Figure 5.1.2.1c. The pressure drop values above and below the ball valve are lower than those corresponding to the riser side. This means that the gas entering the riser at high velocity has two paths; an upward movement through the riser and a diverted upward movement through the valve. This is not desirable because it prevents the solids flow in the riser and disturbs the equilibrium of the system.

For the highest inventory level of 60 cm and all restrictions to the solid flow due to the ball valve, Figure 5.1.2.2a represents the pressure balance around the system. Because of the pressure drop across the valve, the maximum pressure is located above the valve. The bed height in the riser increases by increasing the BVO and obviously it will decrease if higher gas velocities are used.

The pressures at the lowest part are identical which
FIGURE 5.1.2.1 PRESSURE PROFILE AROUND THE CFB

CONDITIONS

(a) H=33cm BVO=70
(b) H=33cm BVO=90
(c) H=33cm BVO=50,60
H=45cm BVO=70
H=45cm BVO=90
H=45cm BVO=50,60
H=60cm BVO=50
H=60cm BVO=90
BVO=60
BVO=50
BVO=70

SLOW BED SIDE
RISER SIDE
means that the solid is falling in free fall from the ball valve to the riser.

Therefore, for safe operation of the riser reactors, the pressure profiles shown in Figure 5.1.2.1c should be avoided at all costs.

Based upon the above information, for the safe operation of the riser of a circulating fluidised bed, it was decided to operate the riser at a slow bed inventory of 60 cm.

5.1.3 Effect of the inventory and ball valve opening (BVO) on the voidage profiles

The pressure drop measurements are converted into voidage $\varepsilon$ by assuming no acceleration or frictional effects and by using the equation:

$$\frac{\Delta P}{\Delta L} = (1-\varepsilon)\rho_s g$$

$\Delta P$ is the pressure difference between successive pressure tappings and $\Delta L$ the distance apart.

The voidage profiles along the riser, for constant superficial gas velocity of 2.2 m/s, constant opening of the ball valve and various slow bed inventory levels in the range of 21 to 60 cm are shown in Figures 5.1.3.1a-b.

Each plot presents the axial profiles for 4 different inventory levels of the slow bed at constant gas velocity and BVO.

At the lower BVO of 60 only the highest inventory shows an inflection point and the presence of both dense and dilute phases. At this BVO, the voidage profiles of the lower inventory levels (ie 21, 33 and 45 cm) are constant
FIGURE 5.1.3.1a
VOIDAGE VERSUS RISER HEIGHT

FCC
Ug=2.2 m/s

BUO: 60°
- H: 21 cm
+ H: 33 cm
* H: 45 cm
□ H: 60 cm

FIGURE 5.1.3.1b
VOIDAGE VERSUS RISER HEIGHT

FCC
Ug=2.2 m/s

BUO: 70°
- H: 21 cm
+ H: 33 cm
* H: 45 cm
□ H: 60 cm
along the riser which operates in the dilute phase transport regime Figure 5.1.3.1a. The higher the BVO, and hence the higher the solids circulation rate, the higher the position of the inflection point, Figure 5.1.3.1b. The increased inventory imposed a greater pressure drop on the riser which was shown by the bed filling with dense phase. The dense phase regions in the riser have solids fractions of 0.15-0.05 characteristic of the fast fluidisation regime while the dilute phase region has solid fractions of about 0.03.

This finding is in contradiction to that of Rhodes et al. (1987) ie the inventory of solids has no effect on the axial profiles in the riser at constant values of \( U_g \) and solid circulation rate (SCR). The reason for such a difference might be due to the ball valve used in this work to control the solid rate. Weinstein et al. (1984) reported that the axial solids profiles are independent of the solids inventory but are dependent on solids rate and gas velocity.

As we already reported, at constant \( U_g \), the solid circulation rate increases with the BVO and slow bed inventory level. Since the axial solids profiles are dependent on solids rate and the solids circulation rates are increased with the increase of the slow bed inventory therefore it would seem that the axial solids profiles in the riser will be affected by the inventory level of the slow bed.
5.2 EFFECT OF DISTRIBUTOR DESIGN ON PRESSURE DROP AND SOLID CIRCULATION RATE

The effect on pressure drop and solid circulation rate of the different design of distributors mentioned in Chapter 4 will now be discussed.

5.2.1 EFFECT ON PRESSURE DROP

The pressures recorded at each of the 13 pressure tappings, above the solids inlet, along the riser are plotted in Figures 5.2.1 to 5.2.18. Six gas velocities in the range 1.5 to 4 m/s were used and at each velocity four openings of the ball valve (BVO), shown as the parameter BVO on the plots, in the range 40 to 70 degrees were studied for the FCC powder. From these observations the following conclusions may be drawn:

(a) For a given gas velocity as the ball valve opening is increased the pressure drop between adjacent pressure tapping in the riser also increases for all three designs of distributor.

(b) The rate of change of pressure drop decreases with riser height, at constant gas velocity, for each one of the three riser distributors.

(c) At low values of the ball valve openings (BVO<60), the pressure profile is relatively flat along the entire height of the riser and decreases slightly with the riser height. This is demonstrated very clearly for all distributor designs in Figures 5.2.1 to 5.2.18.

(d) Visual observations indicate that the relatively flat pressure profile corresponds to the fact that the riser is
operating in the dilute phase mode. In this mode, the particles are moving upwards in relatively straight lines with the absence of any significant downflow.

(e) As the ball valve opening increases (BVO>60), there is a marked increase in the pressure drop at the bottom part of the riser and a levelling off near the top. Under these conditions a dense slugging phase is formed in the lower part of the riser and strong fluctuations are observed in the pressure drops. Further up the riser the solid particles appear to be divided into two phases, a dense phase near the riser wall where particles are falling downwards and a dilute core in the centre of the column where particles are transported upwards.

(f) At higher gas velocities a large increase of solid circulation and hence concentration by increasing the ball valve opening is required to bring the riser into the dense phase transport regime. This dense phase region which occupies the lower part of the riser remains nearly constant for values of the ball valve openings greater than 60 degrees.

By increasing the gas velocity from 1.5 to 4 m/s each pressure profile plot consists of two regimes, a lower regime at low ball valve openings where the pressure profile is flat and decreases only slightly with the height, and an upper regime where there is an initial sharp decrease of pressure drop with the height.

The axial pressure profile measurements have been performed along the riser for several operation conditions.
For the air gas velocities of 1.5 and 2m/s and the seven hole distributor, the profiles are shown in Figures 5.2.7 and 5.2.8. From these curves the existence of three different sections, according to Monceaux et al. (1985), can be present: a) the acceleration section at the bottom of the riser b) a section of constant pressure gradient which is characteristic of a fully-developed pipe flow and c) a disengagement section where the pressure gradient is lower than that in fully-developed flow section, and appears only for high loading ratios.

The shape of the pressure profile changes slightly with the ball valve opening and hence with solid flowrates. Although Monceaux et al. (1985) presented a linear pressure profile at Ug=2.7m/s, the profiles showing in Figures 5.2.7 and 5.2.8 are not linear. They explained the non-linearity due to the increase of the lengths of acceleration and disengagement regions as the solid flowrate increases. They also claimed the disappearance of the fully-developed region at sufficiently high solid flowrates.

The axial pressure profiles are in agreement with those reported by Yerushalmi et al. (1976), Weinstein et al. (1982). Li et al. (1980) observed a transition from a high concentration at the bottom of the riser to a low concentration at the top with an inflection point in between the two regions. Monceaux et al. explained the inflection point as being due to the fact that the acceleration length becomes very large and even sometimes the fully-developed flow is not obtained.
FIGURE 5.2.1 RISER DISTRIBUTOR: 1 NOZZLE
PRESSURE DROP VERSUS RISER HEIGHT

- BUO: 40°
+ BUO: 50°
* BUO: 60°
□ BUO: 70°

$U_g = 1.5 \text{ m/s}$
FCC

RISER HEIGHT (m)

FIGURE 5.2.2 RISER DISTRIBUTOR: 1 NOZZLE
PRESSURE DROP VERSUS RISER HEIGHT

- BUO: 40°
+ BUO: 50°
* BUO: 60°
□ BUO: 70°

$U_g = 2 \text{ m/s}$
FCC

RISER HEIGHT (m)
FIGURE 5.2.5  RISER DISTRIBUTOR: 1 NOZZLE
PRESSURE DROP VERSUS RISER HEIGHT

- BUO: 40°
- BUO: 50°
- BUO: 60°
- BUO: 70°
Ug=3.5m/s
FCC

FIGURE 5.2.6  RISER DISTRIBUTOR: 1 NOZZLE
PRESSURE DROP VERSUS RISER HEIGHT

- BUO: 40°
- BUO: 50°
- BUO: 60°
- BUO: 70°
Ug=4m/s
FCC
FIGURE 5.2.7 RISER DISTRIBUTOR: 7 HOLES
PRESSURE DROP VERSUS RISER HEIGHT

FIGURE 5.2.8 RISER DISTRIBUTOR: 7 HOLES
PRESSURE DROP VERSUS RISER HEIGHT
FIGURE 5.2.9 RISER DISTRIBUTOR: 7 HOLES
PRESSURE DROP VERSUS RISER HEIGHT

FIGURE 5.2.10 RISER DISTRIBUTOR: 7 HOLES
PRESSURE DROP VERSUS RISER HEIGHT
FIGURE 5.2.11 RISER DISTRIBUTOR: 7 HOLES
PRESSURE DROP VERSUS RISER HEIGHT

- BUO: 40°
- BUO: 50°
- BUO: 60°
- BUO: 70°
Ug = 3.5 m/s
FCC

FIGURE 5.2.12 RISER DISTRIBUTOR: 7 HOLES
PRESSURE DROP VERSUS RISER HEIGHT

- BUO: 40°
- BUO: 50°
- BUO: 60°
- BUO: 70°
Ug = 4 m/s
FCC
**FIGURE 5.2.13** RISER DISTRIBUTOR: 13 HOLES  
PRESSURE DROP VERSUS RISER HEIGHT

- BVO: 40°  
- BVO: 50°  
- BVO: 60°  
- BVO: 70°  

$U_g = 1.5 \text{m/s}$  
FCC

**FIGURE 5.2.14** RISER DISTRIBUTOR: 13 HOLES  
PRESSURE DROP VERSUS RISER HEIGHT

- BVO: 40°  
- BVO: 50°  
- BVO: 60°  
- BVO: 70°  

$U_g = 2 \text{m/s}$  
FCC
FIGURE 5.2.15 RISER DISTRIBUTOR: 13 HOLES
PRESSURE DROP VERSUS RISER HEIGHT

FIGURE 5.2.16 RISER DISTRIBUTOR: 13 HOLES
PRESSURE DROP VERSUS RISER HEIGHT
FIGURE 5.2.17 RISER DISTRIBUTOR: 13 HOLES
PRESSURE DROP VERSUS RISER HEIGHT

- BUO: 40°
+ BUO: 50°
* BUO: 60°
□ BUO: 70°

Ug = 3.5 m/s
FCC

FIGURE 5.2.18 RISER DISTRIBUTOR: 13 HOLES
PRESSURE DROP VERSUS RISER HEIGHT

- BUO: 40°
+ BUO: 50°
* BUO: 60°
□ BUO: 70°

Ug = 4 m/s
FCC
However calculations have showed that the acceleration length of a particle is not more than a few centimeters of riser height before it reaches its steady state (Figure 3.2, Chapter 3).

The pressure drop values for the 7 and 13 holes riser distributors are of the same order of magnitude to each other.

The main difference between the 1 nozzle distributor and the 7 and 13 hole distributors lies in the order of magnitude of pressure drops. The pressure profiles obtained with the 1 nozzle distributor have lower pressure drop values under identical conditions of operation of the system. This means that less solid is present in the riser.

Whether or not there is a high or low pressure drop in the riser is at least partly determined by the solid circulation rate which in turn is influenced by the distributor design. If it is assumed that each hole of the distributor design behaves as a fluid jet, then an introduction to fluid jets is given in Appendix G.

By considering the design of the three distributors used during the experimental work and by assuming that the jet angle is of 14 degrees, then the potential core length can be determined for each orifice diameter of the distributor.

The effect of the distributor design on the pressure profile is shown in Figures 5.2.19 and 5.2.20. From those observations the following conclusions may be drawn:

a) For a given gas velocity and constant ball valve opening, the pressure drop at a particular level in the
FIGURE 5.2.19 PRESSURE DROP PROFILES
AT UG=3M/S & BUO=40°

FIGURE 5.2.20 PRESSURE DROP PROFILES
AT UG=3.5M/S & BUO=70°
riser increases with the increase of the number of holes of the distributor due to a change in voidage in the riser.
b) At high ball valve openings, Figure 5.2.20, there is a marked increase in pressure drop at the bottom of the riser up to 1 m riser height and then a linear decrease at higher levels. This sudden increase of pressure drop is higher for the 7 hole distributor than for the distributors consisting of 1 nozzle and 13 holes.
c) At low ball valve openings, Figure 5.2.19, there is a continuous decrease of pressure drop along the riser height. This decrease is sharper for the 7 hole distributor and rather smoother for the 1 nozzle and 13 hole distributors.

For the case of 1 nozzle, the local radial gas velocity profile is as shown in Figure 5.2.21(a). The air as it passes the 1 nozzle, expands in the upward direction. The local velocity in the centre reaches a maximum value at first and then decreases along the height until it approaches the final turbulent state. This turbulent state is reached at a riser height of 16.3 cm if the jet angle is assumed to be 14 degrees. After that, the radial profile does not change but remains constant along the height. The particles are introduced in the riser and are accelerated until they reach their final linear velocity at a vertical height. The linear velocity is smaller than the linear gas velocity by the value of terminal velocity. According to the local gas profile, as soon as particles are distributed in the region near the
FIGURE 5.2.21  GAS FLOW PATTERNS THROUGH DISTRIBUTOR

(a) 1 nozzle

(b) 7 holes

(c) 13 holes

135
wall, they start falling in the opposite direction of the upwards gas movement because the local gas velocity reaches its minimum value near the wall. Hence a stagnant region of particles starts to accumulate in the area above the 1 nozzle distributor. This stagnant region extents about 20cm up from the distributor plate.

In the case of the 7 hole distributor, the local gas velocity profile radially and axially is as shown in Figure 5.2.21(b). The local gas radial profile reaches its final vertical state in a short axial length that is of axial length of 4.6 cm if the jet angle is taken as 14 degrees. As a result the particles when they enter the riser reach their final linear velocity in a shorter vertical distance and time. Because of the minimum value of the local gas profile next to the wall region, a downward movement of particles and some accumulation of particles will occur in the 7 hole distributor area.

The case of the 13 holes distributor is similar with that of 7 hole distributor, with the only difference that the local gas profile is reached at a riser distance of about 4 cm.

The values of pressure profiles can be explained as being due to the axial and radial gas profile.

For the 7 hole distributor, the pressure values are of higher order in comparison to the 13 hole and 1 nozzle distributors. These high values are due to the higher solid concentration present in the riser. For the 1 nozzle distributor the low pressure values might be due to the
stagnant areas of particles present in the flat area next to the nozzle, which limits the gas distribution.

5.2.1.1 Solid concentration

Graphs 5.2.1.1 to 5.2.1.14 are plots of the variation of solid concentration fraction (1-ε) along the riser height at constant gas velocity, for the three different gas distributors.

Graphs 5.2.1.4, 5.2.1.5 and 5.2.1.6 of the 1 nozzle gas distributor, 5.2.1.9 and 5.2.1.10 of the 7 hole gas distributor and 5.2.1.13 and 5.2.1.14 of the 13 hole gas distributor show a common trend with respect to the shape of the graph, 'an exponential decay', that is a sharp decrease of solid concentration fraction up to a riser height of about 1m and then a constant value of solid concentration fraction along the rest of the riser height.

Comparison of the graphs 5.2.1.4, 5.2.1.9 and 5.2.1.13 at the same gas velocity of 3 m/s reveals that for the three gas distributors the solid concentration fraction along the riser varies from 18% at the bottom part of the riser to 2% at the top part of the riser, for all three distributors, although the solid circulation flux achieved varies from 6 kg/m²s to 10 kg/m²s. The measured axial profiles of solids fraction within the 5cm riser for the FCC powder are in agreement to the axial solids fraction.
FIGURE 5.2.1.1  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

1 NOZZLE
BVO: 70'
Ug=1.5 m/s

FIGURE 5.2.1.2  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

1 NOZZLE
BVO: 70'
Ug=2 m/s
FIGURE 5.2.1.3  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

1 NOZZLE
BVO: 70°
Ug = 2.5 m/s

FIGURE 5.2.1.4  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

1 NOZZLE
BVO: 70°
Ug=3 m/s
FIGURE 5.2.1.5  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

1 NOZZLE
BVO: 70°
Ug=3.5 m/s

FIGURE 5.2.1.6  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

1 NOZZLE
BVO: 70°
Ug=4 m/s
FIGURE 5.2.1.7  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

7 HOLES
BVO: 70°
Ug=1.5 m/s

FIGURE 5.2.1.8  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

7 HOLES
BVO: 70°
Ug=2 m/s
FIGURE 5.2.1.9  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

7 HOLES
BUO: 70'
Ug = 3 m/s

FIGURE 5.2.1.10 SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

7 HOLES
BUO: 70'
Ug = 4 m/s
FIGURE 5.2.1.11  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

13 HOLES
BVO: 70°
$U_g = 1.5 \text{ m/s}$

FIGURE 5.2.1.12  SOLID CONCENTRATION FRACTION VERSUS RISER HEIGHT

13 HOLES
BVO: 70°
$U_g = 2 \text{ m/s}$
profiles presented by Rhodes and Geldart (1987) with respect to the shape of the graph and magnitude although much higher solid circulation fluxes of the order 10 to 80 kg/m²s were achieved by using a denser powder (1800 kg/m³ compared to 884 kg/m³) within a riser of 15.2 cm diameter.

Work presented by Youchou and Kwauk (1980) and Weinstein et al. (1983) using pressure data reveals the presence of the inflection point between the dense region and the dilute region of the riser. The fact that such inflection point is not present in the concentration profiles presented in Figures 5.2.1.4, 5.2.1.5 and 5.2.1.6, 5.2.1.9 and 5.2.1.10 and 5.2.1.13 and 5.2.1.14 might be related to Youchou and Kwauk (1980) observation that its location is displaced towards the top of the column as the gas velocity decreases.

Graphs 5.2.1.1, 5.2.1.7 and 5.2.1.11 are the corresponding axial solid concentration profiles at a gas velocity of 1.5m/s for the gas distributors with 1 nozzle, 7 holes and 13 holes respectively. The three graphs show a continuous decrease of solid concentration fraction along the riser height. Since the solid fraction varies from 42% to 1%, well above the normal operation of the riser of a CFB system (25%-1%), this particular gas velocity might be either the choking velocity or the transition velocity from turbulent to fast regime.

During the experimental procedure, it was observed that the riser was in the dense form and slugs were travelling upwards within the riser height. Strong fluctuations of the
water manometers were recorded. The values recorded in the graph are averaged \((1-\varepsilon)\) values between a maximum and minimum amplitude of the water manometers.

**CHOKING**

Leung and Wiles (1976) have identified two types of transition from dilute phase flow to dense-phase flow in vertical transport. In the first type, the choking condition is approached as the gas velocity is reduced at a fixed solid rate, the uniform dilute phase flow is replaced by a dense-slugging flow. In the second type a gradual increase in suspension density is observed before the dense-phase flow is obtained.

There is a degree of confusion in literature over the state of choking. In vertical pneumatic transport, choking is the transition from the dilute-phase flow to the dense-phase flow with increase in solid concentration and pressure fluctuations. With coarse or heave particles slugging occurs while with small light particles a large amount of internal recirculation at the walls of the column take place.

Correlations are available in the literature to classify a system as choking or non-choking. Two well known are those of Yousfi and Gau (1974) and Yang (1976). Yousfi and Gau proposed the following criterion for choking: \(\frac{V_i^2}{gd_p} > 140\)

For the FCC powder used during the experiments \(\frac{V_i^2}{gd_p} = 64\)
the above criterion does not exhibit the characteristic of choking.

Yang proposed that choking is avoided provided the following criterion is met: \[ \frac{\sqrt{v_t^2}}{gD} < 0.35 \]

For the FCC powder \[ \frac{\sqrt{v_t^2}}{gD} = 0.082 \]

Yousfi and Gau criterion does not take into account the effect of pipe diameter. Neither of the above two criteria take into account the effect of the gas velocity for the transition from dilute to dense phase flow.

**TRANSPORT VELOCITY**

According to Yerushalmi and Cankurt (1979), the transport velocity is regarded as the boundary which divides vertical gas-solid flow regimes into two groups. Below it lie the bubbling and turbulent fluidised bed and above it lies the transport regime which includes a wide range of stages from dilute-phase flow to the fast bed condition. For the FCC the transport velocity was estimated to be in the range of 1.2 to 1.5 m/s. The approach of transport velocity is characterised by a sharp increase in the rate of particle carryover.

The transport velocity, according to the above authors, might be determined from the plots of \( \Delta P/\Delta L \) along a section of the riser as a function of the solid rate at different gas velocities.
It is not clearly defined how the transport velocity can be determined. The above plots indicate that at transport velocity a further increase in the solid rate results to a constant ΔP/ΔL value across a section of the riser of a circulating fluidised bed.

For the FCC powder used during the experiments it is observed that above 2 m/s, the solid rate is increased considerably. If it is assumed that the transport velocity lies between 1.5 to 2 m/s, close enough to the values suggested by the above authors, then this velocity might represent the transport velocity as well as the choking phenomenon.

5.2.2 EFFECT ON SOLID CIRCULATION RATE

The solid circulation rate (SCR) is measured using a load cell system, as is reported in Chapter 4. The solid circulation rate is defined as the weight of solids recirculated around the circulating fluidised bed during a time interval. Since it is assumed that the system reaches a steady state condition, after it is operated for half an hour, each SCR is constant and reproducible. Graphs of solid circulation rates versus superficial gas velocities are plotted for a range of ball valve openings (30-70) in Figures 5.2.22, 5.2.23 and 5.2.24, for the three different distributors.

Riser Distributor No.1 (1 nozzle)

The SCR does not seem to be affected by the increase of
FIGURE 5.2.22 PLOT OF GAS VELOCITY VERSUS SOLID CIRCULATION RATE

- BVO: 30°
- BVO: 40°
- BVO: 50°
- BVO: 60°
- BVO: 70°

1 NOZZLE

FCC

SOLID CIRCULATION RATE \times 10^{-3} (kg/s)

GAS VELOCITY (m/s)
FIGURE 5.2.23 PLOT OF GAS VELOCITY VERSUS SOLID CIRCULATION RATE

- **BU0: 30°**
- **BU0: 40°**
- **BU0: 50°**
- **BU0: 60°**
- **BU0: 70°**

7 HOLES

FCC

SOLID CIRCULATION RATE x 10^-3 (kg/s)

GAS VELOCITY (m/s)
FIGURE 5.2.24 PLOT OF GAS VELOCITY VERSUS SOLID CIRCULATION RATE

- BVO: 30°
- BVO: 40°
- BVO: 50°
- BVO: 60°
- BVO: 70°

13 HOLES

FCC
the superficial gas velocity, \( U_g \), for the values of \( U_g \) greater than 1.5m/s at constant opening of the valve of 30, 40, 50 and 60 degrees. However on increasing the opening of the solid-rate control valve above 60 degrees, the SCR for a constant value of \( U_g \) does slightly increases.

Riser Distributor No.2 (7 holes)

The SCR increases with \( U_g \) up to a value of 1.5m/s and then remains constant for constant valve openings of 30, 40 and 50 degrees. The general trend of increasing SCR with the increase of \( U_g \) is obtainable for the valve openings of 60 and 70 degrees up to a gas velocity of 2.5m/s. Above this value of \( U_g \), the SCR is constant for all values of \( U_g \) and is only affected by the opening of the ball valve.

Riser Distributor No.3 (13 holes)

The SCR increases with \( U_g \) up to a value of 1.5m/s and then remains constant even if \( U_g \) is further increased to a value of 4m/s, at the valve openings of 30 and 40 degrees. For the cases of constant valve openings of 50, 60 and 70 degrees an increase of SCR with \( U_g \) exists up to 2.5m/s. Above that for each valve opening the SCR remains constant but increases with the opening of the ball valve.

5.3 EFFECT OF GAS VELOCITY ON PRESSURE DROP AND SOLID CIRCULATION RATE

5.3.1 EFFECT ON PRESSURE DROP

For Figures 5.3.4, 5.3.5 and 5.3.6 the shapes of the curves at each gas velocity were generally such that the
FIGURE 5.3.4 PRESSURE DROP VERSUS RISER HEIGHT

- $U_g = 1.5 \text{ m/s}$
- $U_g = 3 \text{ m/s}$
- $U_g = 4 \text{ m/s}$
- $BVO = 70^\circ$

1 NOZZLE
FCC

PRESSURE DROP (Pa)

RISER HEIGHT (m)
FIGURE 5.3.5 PRESSURE DROP VERSUS RISER HEIGHT

- $U_g=1.5m/s$
- $U_g=3m/s$
- $U_g=4m/s$
- $BVO=70°$
- 7 HOLES
- FCC

PRESURE DROP (Pa)

RISER HEIGHT (m)
FIGURE 5.3.6  PRESSURE DROP VERSUS RISER HEIGHT

- $U_g=1.5\text{m/s}$
- $U_g=3\text{m/s}$
- $U_g=4\text{m/s}$

$B_UO=70^\circ$

13 HOLES
FCC

PRESSURE DROP (Pa)

RISER HEIGHT (m)
rate of change of pressure drop with height decreases as the top of the riser is approached.

For all three distributor designs, a common pattern emerges. At heights below about 1m, the greatest pressure drops occur at the lowest gas velocities. The variation at the lowest pressure tappings can exceed 2000 Pa when comparing the pressure drop curves for the 1.5 and 4 m/s gas velocities (a factor of about 3).

There is a height of approximately 1 to 1.5 m common to all three design of distributors, at which there is a convergence of all three pressure profiles. The relationship between pressure drop and gas velocity is reversed at this point of convergence. Visual observations made during the course of the experiments showed that below a height of approximately 1 m the riser appeared to contain a relatively dense phase. Above the point of convergence the voidage will be close to unity with the result that the hydrodynamics of the riser will be dominated by the properties of the gas alone. This explains why the pressure drops will be greater for the higher values of gas velocities.

Figures 5.3.1, 5.3.2 and 5.3.3 show the variation of pressure drop with riser height for the three distributor designs at a ball valve opening of 40 degrees. As with a ball valve opening of 70 degrees, the general shape of the pressure profile is such that the rate of change of pressure drop with height is greater in the lower half of the riser as compared with the top half of the riser.
FIGURE 5.3.1 PRESSURE DROP VERSUS RISER HEIGHT

- $U_g=1.5\text{m/s}$
- $U_g=3\text{m/s}$
- $U_g=4\text{m/s}$

NOZZLE
FCC
FIGURE 5.3.2 PRESSURE DROP VERSUS RISER HEIGHT

- $U_g=1.5\text{m/s}$
- $U_g=3\text{m/s}$
- $U_g=4\text{m/s}$
- $BVO=40^\circ$

7 HOLES
FCC
FIGURE 5.3.3 PRESSURE DROP VERSUS RISER HEIGHT

- U_g=1.5 m/s
- U_g=3 m/s
- U_g=4 m/s

BVO=40°

PRESSURE DROP (Pa)

0  100  200  300  400

RISER HEIGHT (m)

0  0.5  1  1.5  2  2.5  3

13 HOLES FCC
For the gas velocities investigated, the variation in pressure drops, that corresponds to each gas velocity, is much less than when the ball valve opening was at 70 degrees. The pressure drops themselves are almost an order of magnitude less than the pressure drops shown in Figures 5.3.4, 5.3.5 and 5.3.6.

Unlike the curves presented in Figures 5.3.4, 5.3.5 and 5.3.6, no clear trends emerge to characterise the relationship between each pressure profile and the gas velocity. Allowing for some degree of experimental uncertainty it might be appropriate to conclude that at low ball valve opening of 40 degrees the gas velocity has little influence on pressure drop.

5.3.2 EFFECT ON SOLID CIRCULATION RATE

The comparison of solid circulation rates (SCR) versus gas velocity plots of the riser distributors No.1, No.2 and No.3 shows that the SCRs using the distributor No.1 have much lower values than those of distributors No.2 and No.3, for all the cases of constant valve opening. A typical SCR-Ug plot is shown in Figure 5.3.7.

At low flowrates of air, the limiting condition is the energy contained within the air stream that can be transferred to the solid in the riser as it is being forced upwards, line AB of Figure 5.3.7.

At high air flowrates, ie point B of Figure 5.3.7, the air has now sufficient energy to impart to the solids so that the rate at which the solids are forced upwards
FIGURE 5.3.7 TYPICAL PLOT OF SOLID CIRCULATION RATE VERSUS GAS VELOCITY
matches the rate at which solids can pass through the ball valve. Increasing the air flowrate beyond point B will not increase the solid circulation rate because although the air stream has more energy out the rate at which the solids pass into the air stream does not change because this rate is determined solely by the ball valve.

The influence of the air in the riser is more confined radially causing a region of relative stagnation between the air stream and the wall of the riser immediately above the plane of the nozzle.

Increasing the number of holes in the distributor is likely to improve the radial gas distribution and minimise the stagnant regions especially near the base of the riser where the solids from the ball valve first come into contact with the air stream. Therefore there is more chance that the solid will come into contact with a fast moving air velocity from which it can gain energy in order to move upwards. Hence by improving the radial distribution of air one is likely to increase the solid circulation rate.

At very low gas flowrates, ie Region OA, the solid circulation rate is negligible because the air has insufficient energy to entrain the solids. This means that at low flowrates, Region AB, it is the rate at which the air has a capacity to entrain the solids, so as to carry them up the riser, which will determine, at steady-state, how much solid can pass through a given cross-section of the circulated bed, ie the solid circulation rate.

From the shape of the graph shown in Figure 5.3.7, it
is apparent that within this region AB the solid circulation rate is a very sensitive function of gas velocity. However for the region BC, the solid circulation rate is independent of gas velocity. If the system is operated at such gas velocity greater than B, the air flowrate will have sufficient capacity to entrain more solids, but the opening of the ball valve is such that it is already allowing the maximum possible flowrate for the dynamic equilibrium of the system. Therefore, the extra capacity of the air stream will not be utilised. It can be said that under these conditions the ball valve opening, (BVO), is determining the solid circulation rate. Hence the plateau BC shown in graph Figure 1. Increasing the BVO rather than increasing the gas flowrate will lead to a faster solid circulation rate.

5.4 EFFECT OF BALL VALVE OPENING ON SOLID CIRCULATION RATE

Figures 5.4.1, 5.4.2 and 5.4.3 show graphs of solid circulation rate (SCR) against ball valve opening (BVO) for the three design distributors. Not unexpectedly, an increase in gas velocity leads to an increase in SCR at a given BVO (see previous discussion in Section 5.3.2). Again at a given gas velocity, increasing the BVO leads to an increase in SCR. As with the graphs of SCR versus Ug (Figures 5.2.22 to 5.2.24), each curve resembles an S-shape with three distinct regimes. At low BVO (below 50 degrees) the rate at which the SCR increases with the BVO is
FIGURE 5.4.1 SOLID CIRCULATION RATE
VERSUS BALL VALVE OPENING

- $U_g = 1.5 \text{ m/s}$
- $U_g = 2 \text{ m/s}$
- $U_g = 2.5 \text{ m/s}$
- $U_g = 3 \text{ m/s}$
- $U_g = 3.5 \text{ m/s}$
- $U_g = 4 \text{ m/s}$

FCC

1 NOZZLE

SOLID CIRCULATION RATE $10^{-3}$ (kg/s)

BALL VALVE OPENING (degrees)
FIGURE 5.4.2  SOLID CIRCULATION RATE VERSUS BALL VALVE OPENING

- $U_g = 1.5 \text{ m/s}$
- $U_g = 2 \text{ m/s}$
- $U_g = 2.5 \text{ m/s}$
- $U_g = 3 \text{ m/s}$
- $U_g = 3.5 \text{ m/s}$
- $U_g = 4 \text{ m/s}$

FCC
7 HOLES

SOLID CIRCULATION RATE E-3 (kg/s)

BALL VALVE OPENING (degrees)
FIGURE 5.4.3  SOLID CIRCULATION RATE
VERSUS BALL VALVE OPENING

- $U_g=1.5\ m/s$
- $U_g=2\ m/s$
- $U_g=2.5\ m/s$
- $U_g=3\ m/s$
- $U_g=3.5\ m/s$
- $U_g=4\ m/s$

FCC
13 HOLES

SOLID CIRCULATION RATE $E^{-3}$ (kg/s)
BALL VALVE OPENING (degrees)
relatively low, whilst between 50 and 60 degrees (between 60 and 70 degrees in the case of 1 nozzle distributor), the slope of the curve dramatically increases, before reaching some plateau. The reasons why the curves follow this characteristic S-shape are very similar to the arguments discussed in Section 5.3.2.

The main differences in the performance of the three distributors manifest themselves at the higher gas velocities (above 2.5m/s). For the 7 and 13 hole distributors both the shape of the graphs and the actual solid circulation rates measured are very similar. However for the 1 nozzle distributor the curves at each gas velocity do not appear to display the emergence of a plateau. Furthermore, the solid circulation rates are significantly lower (above half) than the solid circulation rates achieved with the 7 and 13 hole distributors.

5.5 EFFECT OF PARTICLE PROPERTIES ON PRESSURE DROP AND SOLID CIRCULATION RATES

5.5.1 EFFECT ON PRESSURE DROP

Two different types of particles were used for the experimental investigation of the hydrodynamics of the riser when using a 1 nozzle distributor. The properties of these particles are shown in Table 5.5.1.
Experiments were conducted using only the Poralox powder as well as with a 50/50 v/v mixture with FCC powder.

Figure 5.5.1 shows the pressure drops with riser height for pure Poralox, pure FCC and a 50/50 v/v mixture, at a ball valve opening of 70 degrees and a gas velocity of 3 m/s. The shape of the graphs are similar to those presented in Figures 5.2.1 to 5.2.18. For all riser heights and especially near to the base the pressure drops measured with Poralox are much higher than the pressure drops using pure FCC powder. Not surprisingly, a mixture of the two powders gives rise to a pressure drop curve between the two pure powder curves.

At a riser height of 40 cm the Poralox pressure drop is well over twice the FCC pressure drop (3600 compared to 1500 Pa). When the pressures are measured near to the top of the riser the variation in pressure drop has narrowed considerably (1100 compared with 650 Pa).

The explanation for the larger pressure drops measured when using Poralox lies in the increased density and diameter of the Poralox particles. As mentioned before (see Section 5.4.1), the lower part of the riser will experience lower voidages and therefore higher particle concentrations especially at high ball valve openings. If the voidages at the base of the riser are similar for the FCC and Poralox.
FIGURE 5.5.1  RISER DISTRIBUTOR: 1 NOZZLE

PRESSURE DROP VERSUS RISER HEIGHT

* PORALOX
+ MIXTURE
* FCC

BVO=70°
Ug=3m/s
powder, then the increased density of the Poralox powder will imply a greater static pressure.

Figure 5.5.2 shows the variation of cumulative pressure difference with the ball valve opening (gas velocity of 3 m/s), for the pure Poralox, pure FCC and the mixture of the two. As the ball valve opening is increased the variation in cumulative pressure drop increases markedly for the three curves. At a ball valve opening of 40 degrees the variation between the Poralox curve and the FCC curve is about 1050 Pa but at a ball valve opening of 70 degrees this variation has increased to 1400 Pa.

Figures 5.5.3 and 5.5.4 show pressure drops along the height of the riser with Poralox at gas velocities of 2 and 4 m/s.

Comparison with the equivalent graphs for FCC presented in Figures 5.2.2 and 5.2.6 show that for the same ball valve opening and gas velocity the fluidisation of Poralox lead to much greater pressure drops.

At a gas velocity of 2 m/s and a ball valve opening of 60 degrees the Poralox pressure drop was nearly ten times greater than the FCC pressure drop (5000 compared with 700 Pa).

At a gas velocity of 4 m/s and a ball valve opening of 70 degrees the Poralox pressure drop was about twice the equivalent FCC pressure drop (2700 compared with 1200 Pa).

Using Poralox powder it was found that instabilities occurred when the ball valve opening was increased beyond 60 degrees at low gas velocities ie 2 and 2.5 m/s.
FIGURE 5.5.2 PRESSURE DIFFERENCE VERSUS BALL VALVE OPENING

- FCC
- MIXTURE
- PORALOX

Ug=3m/s
This unstable condition corresponds to a situation when a very dense phase appears at the base of the riser promoting a flow reversal of solids through the ball valve at ball valve openings greater than 60 degrees.

5.5.2 EFFECT ON SOLID CIRCULATION RATE

Figure 5.5.5 shows the variation of solid circulation rate with gas velocity at different ball valve openings using pure Poralox powder. As in Section 5.5.1 the circulating fluidised bed was operated using a 1 nozzle distributor. The curves at each ball valve opening are similar to the S shape curves presented in Figure 5.2.22 to 5.2.24, so that at high gas velocities the solid circulation rate reaches some plateau value. This has already been discussed and explained in Section 5.3.2.

Figure 5.5.6 shows a comparison of the solid circulation rate versus ball valve opening for the pure Poralox, pure FCC and 50/50 v/v mixture, at a gas velocity of 3 m/s. Generally, the solid circulation rates are higher with the pure Poralox powder as compared with the pure FCC powder. Whilst the experimental data for the pure Poralox and the 50/50 v/v mixture are well fitted to a single straight line, the FCC data might be better described by a curve, whose slope increases dramatically when the ball valve opening is greater than 60 degrees.

Figure 5.5.7 shows the variation of solid circulation rate with gas velocity using pure FCC powder and pure
FIGURE 5.5.5  SOLID CIRCULATION RATE VERSUS GAS VELOCITY FOR PORALOX
FIGURE 5.5.6 SOLID CIRCULATION RATE
VERSUS BALL VALVE OPENING

FCC
MIXTURE
PORALOX

Ug=3 m/s

SOLID CIRCULATION RATE E-3 (kg/s)

BALL VALVE OPENING (degrees)
FIGURE 5.5.7 PLOT OF SOLID CIRCULATION RATE VERSUS GAS VELOCITY

SOLID CIRCULATION RATE $E^{-3}$ (kg/s)

- BVO: 40°
- BVO: 60°
- BVO: 40°
- BVO: 60°

PORALOX

FCC

GAS VELOCITY (m/s)
Poralox powder. For the pure FCC powder at ball valve openings between 40 and 60 degrees there is very little variation in solid circulation rate as the gas velocity is increased. At a ball valve opening of 40 degrees the solid circulation rate of FCC powder was barely more than 1 g/s for the range of gas velocities investigated. At a ball valve opening of 60 degrees the solid circulation rate stayed at about 2 g/s.

Operating the circulating fluidised bed using Poralox powder meant much higher solid circulation rates, and greater variation in solid circulation rate with increasing gas velocity. At higher gas velocities the solid circulation rates at ball valve openings of 40 and 60 degrees appear to reach a plateau. At a ball valve opening of 60 degrees the plateau is about 34 g/s. This can be compared with a value of 2 g/s measured for pure FCC powder at the same ball valve opening of 60 degrees.

5.6 COMPARISONS OF PRESSURE DROP DATA WITH PREDICTIONS MADE WITH MODEL

Figures 5.6.1 to 5.6.6 show the pressure drop per unit length at different heights in the riser. The graphs represented can be used to show a comparison between the experimentally measured pressure gradients and the theoretical predictions made using the model developed in Chapter 3.

Figure 5.6.1 is an example of where prediction and measurement are closely matched. For the range of gas
FIGURE 5.6.1 PRESSURE GRADIENT VERSUS RISER HEIGHT AT $U_g=4\text{m/s}$ & $BVD=70^\circ$

- FCC
- $D_c=4.4312\times10^{-4} \text{ m}$
- ○ experimental points
- — theoretical line
FIGURE 5.6.2 PRESSURE GRADIENT VERSUS RISER HEIGHT AT \( U_g=4\text{m/s} \) & \( \theta_0=50^\circ \)

Experimental points and theoretical line
FIGURE 5.6.3 PRESSURE GRADIENT VERSUS RISER HEIGHT AT \( U_g = 4 \text{m/s} \) & \( \beta = 70^\circ \)

- FCC
- \( U_c = 1 \text{m/s} \)
- \( U_c = 2 \text{m/s} \)
- \( U_c = 3 \text{m/s} \)
- \( U_c = 4 \text{m/s} \)

\( D_c = 4.4312 \times 10^{-4} \text{m} \)

- Experimental points

Theoretical lines
FIGURE 5.6.4 PRESSURE GRADIENT VERSUS RISER HEIGHT AT \( U_g = 3.5 \text{ m/s} \) & \( BV_0 = 70^\circ \)

- FCC
- \( \text{Dc} = 4.2927 \times 10^{-4} \text{ m} \)
- ○ experimental points
- — theoretical line
FIGURE 5.6.5 PRESSURE GRADIENT VERSUS RISER HEIGHT AT $U_g=2.5\text{m/s}$ & $BVO=70^\circ$

- FCC
- $D_c=3.8158E^{-4}\text{m}$
- ○ experimental points
- — theoretical line
FIGURE 5.6.6 PRESSURE GRADIENT VERSUS RISER HEIGHT AT Ug=4m/s & BVO=70°

- FCC
  - \( R_c = 19 \text{g/s} \)
  - \( R_c = 17 \text{g/s} \)
  - \( R_c = 15 \text{g/s} \)

- Theoretical lines

- \( D_c = 4.4482E-4 \text{m} \)

- Experimental points
velocities and ball valve openings experimentally investigated, the predicted curve always followed the same "exponential decay" type graphs as those that arise from plotting the experimental data.

There is a good qualitative comparison between model and experiment in terms of graphical shape and order of magnitude. However, at low ball valve openings and gas velocities (ie those conditions that give rise to a low solid circulation rate), there is a wider quantitative divergence between model and experiment (see Figure 5.6.2).

During the development of the proposed Dense Annulus Lean Core (DALC) model many assumptions and approximations were required.

Figures 5.6.3 and 5.6.4 show experimental data points superimposed on to the curves representing DALC predictions at different initial cluster velocities. The accuracy of prediction does not appear to vary widely when the initial (downward) cluster velocity, \( U_c \), is varied from 1 to 4 m/s. In general however using lower \( U_c \) will lead to a lesser divergence between experiment and model.

Figure 5.6.5 shows experimental data superimposed on the curves representing DALC predictions at \( U_g = 2.5 \text{m/s} \) BVO = 70°. There is a good agreement between experimental points and the theoretical line.
Figure 5.6.6 shows the sensitivity of predicted pressure profiles with changing solid circulation rates. The actual measured solid circulation rate under the experimental conditions described in Figure 5.6.6 is 19.22 g/s. The other two predicted curves represent calculated pressure profiles that correspond to using solid circulation rates 10 and then 20% less than the measured solid circulation rate. The three predicted curves are very similar with a maximum deviation of 10% between the curves representing a solid circulation rate of 19.22 and 14.79 g/s.

5.7 IMPLICATIONS OF THE HYDRODYNAMIC MODELLING

The hydrodynamic model described in Chapter 3 is now applied to the case in which the riser of the CFB is used as a reactor. Suppose a first-order catalytic reaction of the form $A \rightarrow_{k} B$ takes place in the riser reactor at 400°C and that the fluidising gas velocity $u$ is many multiples of $u_{mf}$. By considering the model of Yates and Rowe (1977) and assuming the reactant gas to be in plug flow, the rate of change of concentration of $A$ within the cell of gas surrounding a particle will be determined by the mass transfer coefficient $h_{m}$, the surface area of the particle $A_p$ and the concentration driving force $C_{Ap} - C_{A}$. Thus
\[- \frac{dC_A}{dt} = \frac{h_m A_p}{V_c} (C_{A_h} - C_{A_p})\]

The rate of change of concentration of A at the particle surface will be determined by the chemical reaction rate which is assumed to be of the first-order with respect to A:

\[\frac{dC_{A_p}}{dt} = k C_{A_p} \left(\frac{1-\varepsilon b}{1-\varepsilon_m}\right)\]

where: \(\varepsilon_m\) is the fixed bed voidage on which the value of \(k\) is based

\(k\) is the fixed bed rate constant

Let \(\varepsilon_1 = \frac{1-\varepsilon b}{1-\varepsilon_m}\)

then \[- \frac{dC_{A_p}}{dt} = k C_{A_p} \varepsilon_1\]

At any height \(h\) above the bed surface

\[\frac{dC_A}{dt} = \frac{dC_{A_p}}{dt} = \frac{dC_{A_h}}{dt}\]

Hence:

\[- \frac{dC_{A_h}}{C_{A_h}} = \frac{dt}{\frac{V_c}{h_m A_p} + \frac{1}{k \varepsilon_1}}\]

The above equation may be integrated with boundary condition that at \(t=0\), \(C_{A_h}=C_{A_s}\) where \(C_{A_s}\) is the concentration of A at the surface of the fluid bed.

Thus

\[\frac{C_{A_h}}{C_{A_s}} = \exp \left[- \left(\frac{t}{\frac{V_c}{h_m A_p} + \frac{1}{k \varepsilon_1}}\right)\right] \quad (5.7.1)\]
where: \[ V_c = \frac{V_p}{1 - \varepsilon_b} \]

Equation 5.7.1 enables conversions in the dilute core and downflow region to be calculated for the case of plug flow of the reactant gas.

Figures 5.7.1, 5.7.2 and 5.7.3 are the results of the calculations for the three values of the rate constant, \( k \), 0.5 s\(^{-1}\), 5 s\(^{-1}\) and 15 s\(^{-1}\) at a gas velocity of 4 m/s and by using the 7 hole distributor. It is clear from the results that the conversion in the downflow region is higher than the conversion in the dilute core and that this mainly depended on the value of reaction rate constant. The most extreme example is shown in Figure 5.7.3 for \( k = 15 \text{s}^{-1} \) where the model indicates a conversion of 55% in the downflow region within a distance of less than 3m of the riser height, although the correspondent conversion in the dilute core is less than 1%. This is explained due to the fact the particle concentration in the dilute core is very low in comparison to that in the downflow region. In both regions the particle concentration decreases with increasing riser height. In the model described, the particles are assumed to be of uniform size but in practice there will be a distribution of small and large particles. Certainly some of the assumptions on which the model is based represent over-simplifications. But it is worth mentioning here that this model takes into consideration that the solid concentration fraction decreases along the riser height. Measurements of other workers agree well with
FIGURE 5.7.1
Ca/Cao VERSUS RISER HEIGHT

--- DILUTE CORE
- - - - DOWNFLOW REGION

Ug=4 m/s
BVO=70°
k=0.5 s⁻¹
FIGURE 5.7.2
Ca/Cao VERSUS RISER HEIGHT

DILUTE CORE
-.-.-. DOWNFLOW REGION

Ug = 4 m/s
BVO = 70°
κ = 5 s⁻¹
FIGURE 5.7.3
Ca/Cao VERSUS RISER HEIGHT

- - - - DILUTE
- - - - DOWNFLOW

Ug = 4 m/s
BVO = 70°
k = 15 s⁻¹
this fact. This is the main difference between this model which calculates the variation of voidage with the riser height and the model presented by Yates and Rowe (1977). It is clear from the results that the downflow region next to the riser wall, exerts a considerable influence on the course of the reaction compared with that of the dilute core.

Figures 5.7.4, 5.7.5 and 5.7.6 are the results of calculation when the 1 nozzle distributor is used at a constant value of the reaction rate constant \( (k=15\text{s}^{-1}) \) and at constant value of the solid circulation rate \( (10.98\text{g/s}) \), and only the gas velocity is varied for 3 to 4 m/s. The effect of increasing the gas velocity from 3 to 4 m/s, in the dilute core, is small since the conversion in this region of the riser has varied from 0.5 to 0.3 % respectively. However, the effect of increasing gas velocity is very significant for the 'Downflow Region', where the conversion at the top of the riser ie at a height of 2.7 m reaches a value of 85% at a gas velocity of 3 m/s and this value decreases to 70 % at a gas velocity of 4 m/s. The results show that the conversion of a 1st order reaction decreases by increasing gas velocity. This is explained that under the above mentioned conditions by increasing the gas velocity the solid concentration within the riser decreases and hence the voidage increases. Since the conversion is inversely proportional to voidage, an increase in gas velocity will result to a decrease of the conversion.
FIGURE 5.7.4
Ca/Cao VERSUS RISER HEIGN

- DILUTE
- - DOWNFLOW

U_g = 3 m/s
BVO = 70°
\( k = 15 \text{ s}^{-1} \)
FIGURE 5.7.5
Ca/Cao VERSUS RISER HEIGHT

- DILUTE
- - DOWNFLOW

$U_g = 3.5 \text{ m/s}$
$BVO = 70^\circ$
$k = 15 \text{ s}^{-1}$
FIGURE 5.7.6
Ca/Cₐₒ VERSUS RISER HEIGHT

- DILUTE
- - - DOWNFLOW

U_g = 4 m/s
B_{vo} = 70°
k = 15 s⁻¹
Up to date, there is no literature available to compare these results with respect to the variation of conversion with height. In a recent publication, Kunni and Levenspiel (1990), reported that a fast fluidised bed reactor can be treated as having a) a lower dense zone where $\varepsilon_{\text{bed}} = 0.75-0.85$ and this behaves as a bubbling bed where similar conversion equations are applicable and b) the upper freeboard zone, the freeboard model applies for the extra conversion. Hence this model determines one conversion in the dense bed and an extra conversion at the exit of the reactor, without taking into account the variation of conversion along the reactor height due to the voidage variation, which was the main objective of this exercise.
6.1 CONCLUSIONS

Pressure drop and solid circulation rate measurements have been made during the operation of the circulating fluidised bed. The riser of the circulating fluidised bed was made of transparent perspex with a height of 3m and internal diameter of 50mm. Most of the experiments have been conducted with FCC powder of average particle diameter of 64\(\mu\)m and particle density of 884 kg/m\(^3\). Some further experiments were conducted with either pure Poralox powder of particle diameter of 127\(\mu\)m and density of 1850 kg/m\(^3\) or a 50/50 v/v mixture of Poralox and FCC. The fluidising medium was compressed air directed through the base of the riser using three designs of distributor plates. The gas velocities were varied between 1.5 to 4 m/s. The three distributor plates had the same free percentage of open area; divided into 1 nozzle, 7 and 13 holes.

Analysis of the experimental data leads to the following conclusions:

(1) The design of the distributor plate can effect the performance of the circulating fluidised bed. For the same operating conditions using the 7 and 13 hole distributors lead to much higher solid circulation rates in comparison with the 1 nozzle distributor. These increased solid circulation rates corresponded with an increased pressure
drops for the 7 and the 13 hole distributors.

(2) Increasing the solid circulation rate can only be achieved at the expense of higher pressure drop. For reactive systems, a high solid circulation rate is often desirable to improve conversion. The benefits of improved conversion must be weighted against the economic costs of a greater pressure drop.

(3) Increasing the density and size of particle (from FCC to Poralox) will lead to higher solid circulation rate but also higher pressure drop.

(4) Operating the circulating fluidised bed at high ball valve openings (above 60 degrees) will normally lead to a downflow of solids. This downflow region is confined to an annulus adjacent to the riser wall.

(5) If the ball valve opening is greater than 70 degrees then the circulating fluidised bed cannot be operated without instabilities. These instabilities manifest themselves by fluctuations, sometimes vigorous, of the levels in the water manometers; and also by the periodic reversal of solid flow through the ball valve.

Observations made during the operation of the circulating fluidised bed confirm that the riser consists of a dense phase, close to the base of the riser with a much leaner phase above it.

In Chapter 3 a model has been proposed (DALC) which aims to relate the pressure drop along the axis of the riser to particle and fluid properties. The DALC model has showing itself able to give good predictions, both
quantitatively and qualitatively when the circulating fluidised bed is operated so as to achieve high solid circulation rates. However at the model's present stage of development it has not been possible to predict the experimentally measured values of pressure drop with a satisfactory degree of accuracy under all conditions.

The DALC model is applied to the case in which a first order catalytic reaction takes place in the riser of a CFB. By increasing the reaction rate constant to 15 s\(^{-1}\) the model indicates a conversion of 55 % in the downflow region within a distance of less than 3m above the distributor. By increasing the gas velocity in the riser the model predicts that the conversion of a first order reaction decreases.

6.2 RECOMMENDATIONS FOR FURTHER WORK

The pressure measurements for this study were made with pressure tappings flash with the wall of the riser. It is not clear that these pressure tappings accurately reflect the pressure across the whole cross-section of the riser, at that tapping height, or, as it has been suggested in this study, the region immediately adjacent to the riser wall. For this reason experimental equipment should be designed to allow the measurement of any radial pressure profile, at a given height in the riser.

A more accurate measurement of the solid circulation
rate can only be achieved with a non-intrusive technique.

More sensitive measurements of pressure can be achieved by using pressure transducers. If the output from these transducers is fed to appropriate computer software a time average value of pressure drop can be calculated to deal with any fluctuations.

An explanation for the origin of the downflow has been proposed during the development of the DALC model. However more work is required to investigate this phenomenon. This might be accomplished using X-ray photography.

A wider range of powders should be used in order to more fully investigate the effect of particle diameter and density on the operation of the circulating fluidised bed.
NOMENCLATURE

SYMBOLS

A cross-section area of riser m²
a cross-section area of downcomer m²
a constant in eqn. 2.17
a constant in eqn. 2.20
a constant in eqn. 2.22
Ar Archimedes number
b constant in eqn. 2.23
BVO angle of the ball valve, degree
C equal to \((dP_v + dP_c)/\rho_s\) eqn. 2.16
C constant in eqn. 2.24
\(c_D\) drag coefficient of a single particle
\(c_D'\) drag coefficient of particles in assembly
\(c_{dp}\) drag coefficient of particles in assembly
\(c_{ds}\) drag coefficient of a single particle
\(C_f\) Fanning friction factor
D riser diameter m
Dc cluster diameter m
Dt tube diameter m
dp particle diameter m
\(D_{ss}\) diameter of slow bed m
\(D_R\) riser diameter m
dx length m
dP pressure drop N
dh riser length m
dp particle diameter m
pressure drop due to solid valve $ dP_v \ N$
pressure drop in the downcomer $ dP_d \ N$
pressure drop in fast bed $ dP_f \ N$
pressure drop in cyclone $ dP_c \ N$
constant in eqn. fitted to wall flow rate $ dW \ N$
entrainment flux $ E \ kg/m^2s$
entrainment flux above TDH $ E^* \ kg/m^2s$
entrainment flux at bed surface $ E_0 \ kg/m^2s$
drag force $ F_D \ N$
solid friction factor $ f_s \ N$
acceleration due to gravity $ g \ m/s^2$
mass flux of solid $ G \ kg/m^2s$
mass flowrate of air $ G_s \ kg/s$
solid mass flux $ G_s \ kg/m^2s$
distance above the dense phase of riser $ h \ m$
riser height $ H_R \ m$
total solid inventory $ I \ kg$
exponential decay coefficient $ k \ m$
length of riser $ L \ m$
mass of a particle $ m \ kg$
solid hold-up in the annular region $ M_D \ kg$
total solid hold-up in the riser $ M_R \ kg$
solid flux $ M_s \ kg/m^2s$
solid hold-up in the dilute core $ M_{DC} \ kg$
mass of cluster $ m_c \ kg$
number of clusters in length $ dL $ $ N_c \ kg$
number of particles per unit length $ N_p \ kg$
$R_c$ solid circulation rate in dilute core kg/s
$R_{CD}$ solid circulation rate in downflow kg/s
$Re_p$ particle Reynolds number
$Re_t$ terminal Reynolds number
$t_o$ average particle residence time s
$t_R$ residence time s
$t_{CD}$ residence time of solids in the downflow s
$u$ slip velocity m/s
$U_g$ superficial velocity m/s
$U_g$ superficial gas velocity m/s
$U_p$ linear particle velocity m/s
$u_g$ superficial gas velocity m/s
$u_p$ linear particle velocity m/s
$U$ superficial gas velocity m/s
$U_{mf}$ gas velocity at minimum fluidisation m/s
$U_a$ superficial gas velocity m/s
$U_{sl}$ slip velocity m/s
$U_p$ linear solid velocity m/s
$U_t$ terminal solid velocity m/s
$u_{ts}$ terminal slip velocity m/s
$V_C$ superficial solid velocity m/s
$V$ solid velocity m/s
$V_t$ terminal velocity m/s
$V_T$ total volume of system m$^3$
$V_R$ riser volume m$^3$
$V_p$ particle velocity m/s
$W$ flowrate of particles kg/s
flowrate of solids down the wall \( \text{kg/s} \)

height of fast bed \( \text{m} \)

height \( \text{m} \)

position of inflection point \( \text{m} \)

characteristic length \( \text{m} \)

riser height \( \text{m} \)

GREEK SYMBOLS

\( \rho_b \) \quad \text{bed density} \quad \text{kg/m}^3

\( \rho_g \) \quad \text{density of gas} \quad \text{kg/m}^3

\( \rho_f \) \quad \text{density of fluid} \quad \text{kg/m}^3

\( \rho_p \) \quad \text{density of particle} \quad \text{kg/m}^3

\( \rho_{SB} \) \quad \text{slow bed density} \quad \text{kg/m}^3

\( \varepsilon \) \quad \text{voidage}

\( \varepsilon_{mf} \) \quad \text{voidage at point of minimum fluidisation}

\( \varepsilon^* \) \quad \text{voidage for dilute phase-section}

\( \varepsilon_a \) \quad \text{voidage for dense phase - section}

\( \varepsilon_i \) \quad \text{dilute phase voidage}

\( \varepsilon_{SB} \) \quad \text{voidage value at } u_{mf}

\( \Delta P \) \quad \text{pressure drop} \quad \text{N}

\( \Delta P_{\text{STAT}} \) \quad \text{static pressure drop} \quad \text{N}

\( \Delta P_{\text{gw}} \) \quad \text{pressure drop due to gas-wall friction} \quad \text{N}

\( \Delta P_{\text{sw}} \) \quad \text{pressure drop due to solid-wall friction} \quad \text{N}

\( \Delta H_{\text{SB}} \) \quad \text{change of slow bed height} \quad \text{m}

\( \Delta M_{\text{SB}} \) \quad \text{mass of solids in motion} \quad \text{kg}

\( \mu \) \quad \text{gas viscosity} \quad \text{kg/ms}
REFERENCES


Rhodes M.J. and Geldart D. (1987), "A model for the


Ruden P. (1933), Naturwissenschaften, 21, p.375.


Squires A.M., Kwauk M. and Avidan A.A. (1985), "Fluid beds:
at last, challenging two entrenched practices", Science, 230, p.1329.


APPENDIX A

SUMMARY OF EXPERIMENTAL RESULTS
RISER DISTRIBUTOR: 1 NOZZLE

POWDER: FCC

Ug = 1.5 m/s  P: N/m²

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RISER DISTRIBUTOR: 1 NOZZLE
POWDER: FCC

### Ug=3m/s  P: N/m³

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RISER DISTRIBUTOR: 7 HOLES
POWDER: FCC

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RISER DISTRIBUTOR: 7 HOLES
POWDER: FCC

Ug=3m/s  \( P: \text{N/m}^2 \)

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POWDER: FCC

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A7
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POWDER: PORALOX

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POWDER: PORALOX

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POWDER: 50/50 v/v MIXTURE

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A10
POWDER: FCC
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POWDER: FCC
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POWDER: FCC

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POWDER: PORALOX

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RISER DISTRIBUTOR: 1 NOZZLE

POWDER: 50/50 v/v MIXTURE

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APPENDIX B

POWDERD USED

SIEVE ANALYSIS
### POWDER: FCC

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</tr>
<tr>
<td>&lt; 75</td>
<td>48.75</td>
</tr>
</tbody>
</table>

### POWDER: PORALOX

<table>
<thead>
<tr>
<th>SIEVE SIZE (µm)</th>
<th>% WEIGHT</th>
</tr>
</thead>
<tbody>
<tr>
<td>&gt;353</td>
<td>0.16</td>
</tr>
<tr>
<td>353-300</td>
<td>1.09</td>
</tr>
<tr>
<td>300-250</td>
<td>1.09</td>
</tr>
<tr>
<td>250-212</td>
<td>6.37</td>
</tr>
<tr>
<td>212-180</td>
<td>14.83</td>
</tr>
<tr>
<td>180-150</td>
<td>22.12</td>
</tr>
<tr>
<td>150-125</td>
<td>22.0</td>
</tr>
<tr>
<td>125-106</td>
<td>10.76</td>
</tr>
<tr>
<td>63-53</td>
<td>1.85</td>
</tr>
<tr>
<td>53-45</td>
<td>0.93</td>
</tr>
<tr>
<td>&lt; 45</td>
<td>0.56</td>
</tr>
</tbody>
</table>
APPENDIX C

HYDRODYNAMICS OF DOWNFLOW REGION WITHOUT CLUSTERING
In chapter three during the development of DALC model it was assumed that the origins of the downflow were caused by aggregation of individual particles into clusters. It is the purpose of this Appendix to consider the hydrodynamics of individual particles in the immediate vicinity (within $10d_p$) of the riser wall.

The universal non-dimensional representation of turbulent boundary layer flow is given by

$$u_y = \frac{y^*}{u_\infty}$$

viscous sublayer $y^* \leq 5$

$$10 \arctan(0.1y^*) + 1.2$$

buffer layer $5 \leq y^* \leq 30$

$$5.5 + 2.5 \ln y^*$$

turbulent layer $y^* \geq 30$

where $u_\infty$ is the friction velocity defined by

$$u_\infty = (\tau_v / \rho_f)^{0.5} = u_g (c_f / 2)^{0.5}$$

and the non-dimensional wall distance given by

$$Y = \frac{y u_\infty}{\nu} = \frac{y u_\infty \rho_f}{\mu}$$

For a particle diameter of 64\,$\mu$m in air at room temperature and pressure, the terminal particle settling velocity is 0.2 m/s. The particle will start to fall in the downflow region if the gas velocity is less than 0.2 m/s. For air flowing with a bulk velocity of 3 m/s in a riser of diameter 0.05 m the pipe Reynolds number is 18000 safely in the turbulent regime. Using Fanning friction factor chart the friction factor $c_f$ was estimated as 0.006.
The friction velocity \( u^* = 3 (0.006/2)^{0.5} = 0.164 \text{ m/s} \)

When the dimensionless distance \( y^* = 5 \) then the velocity

\[
\frac{u}{y} = 5 \Rightarrow u = 5 \times 0.164 = 0.82 \text{ m/s}
\]

When \( u = 0.2 \text{ m/s} \)

\[
y^* = \frac{0.2}{0.82} \times 5 = 1.22
\]

This corresponds to a dimensional distance of

\[
y = \frac{1.22 \times 10^{-5}}{0.164 \times 1.2} = 6.2 \times 10^{-5} \text{ m} = 62 \mu \text{m}
\]

Individual particles of diameter of 64 \( \mu \text{m} \) will only fall if they are within the very small annulus 62 \( \mu \text{m} \) wide. This would imply that only very few particles could constitute a downflow regime one or at best two particle diameter thick. This does not agree with visual observations which indicate that the downflow involves a large number of particles flowing downwards in an annulus of discernible thickness.
APPENDIX D

GAS VELOCITY MEASUREMENTS
GAS VELOCITY MEASUREMENTS

The laboratory compressed air supply was used at air flowrates less than 1200 lt/min and the flowrates were monitoring using rotameters. The rotameters were piped such that the pressure of air inside them could be held constant. This is done by using a pressure regulator and by positioning of the air control valve downstream of the rotameter. The rotameters were calibrated with the air inside the float chamber held at a constant 10 psig. Mass type gas flowmeters were used for calibration depending on the flowrate.

During the calibration the volumetric flowrate was measured at ambient pressure and temperature. Bed temperature and pressure were monitored and used in calculating the actual gas velocity in the bed from the rotameter reading.
APPENDIX E

CYCLONE DESIGN

MINIMUM AND MAXIMUM CLUSTER DIAMETER
Stairmand high efficiency cyclone is used for the design of the primary cyclone, for an operation of 10 m/s riser velocity and by assuming that the riser cross-sectional area and the cyclone inlet duct area are equal.

Area of the inlet duct = \( \frac{\pi D^2}{4} = \frac{\pi \times 0.05^2}{4} = 0.002 \text{ m}^2 \)

From figure E1, duct area = 0.5 \( D_c \times 0.2 \ D_c \)

or \( 0.002 = 0.10 \ D_c^2 \)

Hence \( D_c = 140 \text{ mm} \)

The design of the secondary cyclone is also based on the Stairmand's high efficiency design, by assuming that the cyclone efficiency may be improved when the maximum pressure drop inside the cyclone do not exceed 1200 N/m².

Area of the inlet duct = \( \frac{\pi \times 0.037^2}{4} = 1.075 \times 10^{-3} \text{ m}^2 \)

From Figure E1, duct area = 0.5 \( D_c \times 0.2 \ D_c \)

or \( 1.075 \times 10^{-3} = 0.10 \ D_c^2 \)

Hence \( D_c = 105 \text{ mm} \).
MINIMUM AND MAXIMUM CLUSTER DIAMETER

The minimum and maximum cluster diameters for a range of operation conditions are listed in Table E1, with an initial cluster velocity of 1 m/s.

<table>
<thead>
<tr>
<th>BVO</th>
<th>(U_g) m/s</th>
<th>Riser Distributor</th>
<th>Minimum (d_c) m</th>
<th>Maximum (d_c) m</th>
</tr>
</thead>
<tbody>
<tr>
<td>70</td>
<td>3</td>
<td>7 HOLES</td>
<td>3.834E-4</td>
<td>3.9E-4</td>
</tr>
<tr>
<td>40</td>
<td>3</td>
<td>7 HOLES</td>
<td>3.66E-4</td>
<td>3.81E-4</td>
</tr>
<tr>
<td>70</td>
<td>4</td>
<td>7 HOLES</td>
<td>4.43085E-4</td>
<td>4.4655E-4</td>
</tr>
<tr>
<td>70</td>
<td>2</td>
<td>7 HOLES</td>
<td>3.82035E-4</td>
<td>3.8276E-4</td>
</tr>
<tr>
<td>70</td>
<td>2.5</td>
<td>7 HOLES</td>
<td>3.816865E-4</td>
<td>3.8495E-4</td>
</tr>
<tr>
<td>70</td>
<td>3.5</td>
<td>7 HOLES</td>
<td>4.29272E-4</td>
<td>4.31E-4</td>
</tr>
<tr>
<td>70</td>
<td>4</td>
<td>13 HOLES</td>
<td>4.6565E-4</td>
<td>4.6658E-4</td>
</tr>
<tr>
<td>70</td>
<td>3.5</td>
<td>13 HOLES</td>
<td>4.403E-4</td>
<td>4.4155E-4</td>
</tr>
<tr>
<td>70</td>
<td>3</td>
<td>13 HOLES</td>
<td>4.1724E-4</td>
<td>4.1868E-4</td>
</tr>
<tr>
<td>70</td>
<td>4</td>
<td>1 NOZZLE</td>
<td>4.82893E-4</td>
<td>4.8316E-4</td>
</tr>
<tr>
<td>70</td>
<td>3.5</td>
<td>1 NOZZLE</td>
<td>4.507835E-4</td>
<td>4.51423E-4</td>
</tr>
<tr>
<td>70</td>
<td>3</td>
<td>1 NOZZLE</td>
<td>4.364E-4</td>
<td>4.3677E-4</td>
</tr>
</tbody>
</table>
APPENDIX F

COMPUTER PROGRAMMES
FLOW CHART
1 Initial i se Var- i ab 1 es

1 I n i t i a l i s e Var i a b 1es

I

I-----------

-------------------1
I

2 Calculation

of SCR

2 Calculate

o-f v o i d a g e

3 Calculate

I
I

3 Calculation
constant

I

Calculate

I-------------

o-f DSBH

I

I

and v o i d a g e

4r

down-flow SCR

4 Choose

B

cluster

d i a m et e r

1= 1+ 1
1

Ca.I cul ate a c c e l er ati on
at time t

— IC a l c u l a t e v o i d a g e
5 and v o i d a g e

c on st an t

i
I

time by At

I n cr e m e n t

1

1
1

C a l c u l a t e d i s t a n c e Ax
t ravel 1ed in At

I

1
1

C a l c u l a t e c h a n q e in
v e l o c i ty at t im e At

1

9 C a l c u l a t e n umber of p a r t i c l e s
per unit length b e t we e n x and
l
10 C a l c u l a t e gain in K.E.
1 p a r t i c l e at x + Ax

" of
"

x+Ax

1

i
r~
1 1 C a l c u l a t e gain of K.E.
N p a r t i c l e s at x+Ax

of

I
12 C a l c u l a t e total d i s t a n c e
tr av e 1 1 ed at x + Ax

1

1

12 C a l c u l a t e c h a n g e in p a r t i c l e
v e l o c i t y in s t a g n a n t air at Ax

T
13 C a l c u l a t e e f f e c t i v e d e n s i t y
of p a r t i c l e s at x+Ax

13 C a l c u l a t e K.E. d i f f e r e n c e for
c l u s t e r b e t w e e n 11 and 12 at Ax

I

I
I

1
14 C a l c u l a t e l oss of
th e gas over Ax

14 C a l c u l a t e P.E. ga in ed by air
t r a v e i l i n g f rom x t o x + Ax

F2

K.E.

of

"1


15 Calculate rate of loss of K.E. by gas from \( x \) to \( x+\Delta x \)

16 Calculate change in gas velocity from \( x \) to \( x+\Delta x \)

17 Calculate static pressure S.P.

18 Calculate dynamic pressure from Bernoulli's equation

19 Calculate gas-wall frictional losses as the gas moves from \( x \) to \( x+\Delta x \)

20 Calculate solid-wall frictional losses as the solid moves from \( x \) to \( x+\Delta x \)

21 Calculate total pressure drop from \( x+\Delta x \)

22 Calculate new solids velocity at \( x+\Delta x \)

23 Calculate accumulative pressure drop from previous pres. tap. over \( x+\Delta x \)

24 If \( x > \) distance between pres. tap. go to 22

25 Calculate pressure drop per unit length

26 Go to 3

27 Put \( x=0 \)

15 Calculate 'change' in gas velocity over \( \Delta x \)

16 Calculate dynamic pressure drop from Bernoulli's eqn. over \( \Delta x \)

17 Calculate static pressure drop over \( \Delta x \)

18 Calculate total pressure drop 16+17 over \( \Delta x \)

19 Calculate accumulative pressure drop from previous pres. tap. over \( x+\Delta x \)

20 If \( x> \) the length of pres. tapping go to 22

21 Go to 5

22 Calculate the pressure profile \( 19/x \)

23 Put \( x=0 \)

24 If \( I=14 \) go to 26

25 Go to 5

26 STOP
28 If I=14, Go to 30
29 Go to 3
30 STOP
10 PRINT "WHAT IS PARTICLE DIAMETER"
20 INPUT DP
30 PRINT "DENSITY OF PARTICLES"
40 INPUT ROEP
50 ROEP=531
60 MP=3.142*ROEP*DP^3/6
70 PRINT "PARTICLE MASS, MP"
80 PRINT "DENSITY OF GAS"
90 INPUT ROEG
100 ROEG=1.2
110 PRINT "VISCOITY OF GAS"
120 INPUT VIS
130 VIS=1E-5
140 PRINT "WHAT IS INITIAL GAS VELOCITY"
150 INPUT UGAS0
160 PRINT "WHAT IS GAS MASS FLOWRATE"
170 INPUT GS
180 A=(18*VIS)/(DP^2*ROEP)
190 PRINT "VALUE OF A", A
200 PRINT "WHAT IS INITIAL SOLIDS VELOCITY, DOWNWARDS POSITIVE"
210 INPUT VELIN
220 VELIN=1
230 PRINT "DIAMETER OF RISER"
240 DR=0.05
250 PRINT "SOLID CIRCULATION RATE"
260 INPUT SCR
270 VOIDC=4*SCR/(ROEP*3.142*DP^2)
280 PRINT "VOIDAGE CONSTANT", VOIDC
290 UGAS1=UGAS0
300 EPS=1-(VOIDC/VEL)
310 B=A*EPS^-4.7
320 PRINT "B", B
330 DEL2=-(B*VEL)+(B*UGAS0+9.81)
340 PRINT "ACCELERATION", DEL2
350 DELTIM=0.01
360 DELX= DELTIM*VEL
370 PRINT "DISTANCE", X
380 DELX=X-DELX
390 DVEL=DELX+DELX
400 PRINT "DISTANCE", X
410 PRINT "ACCELERATION", DEL2
420 DVEL2=9.81*DELTIM
430 VEL=DVEL+VEL
440 VEL2=VEL+DVEL2
450 KE1=0.5*MP*H
460 PRINT "GAIN OF PARTICLE K.E.", KE1
510 ENRT = KE1
520 ENRATE = NP*ENRT*VEL + DELX*GS*9.81
530 PRINT "RATE OF LOSS OF K.E", ENRATE
540 PRINT "NP", NP
550 PRINT "P.E. GAINED BY AIR", DELX*GS*9.81
560 J = UGAS0^2 - (2*ENRATE/GR)
570 UGAS2 = -(J^0.5)
580 KELOSS = 0.5*GS*(UGAS0^2 - UGAS2^2)
590 PRINT "RATE OF LOSS OF K.E", KELOSS
600 PRINT "UGAS", UGAS
610 PRINT "VOIDAGE", EPS
620 DELGAS = UGAS0 - UGAS2
630 PRINT "DELGAS", DELGAS
640 SP = (1-EPS)*(ROEP-ROEG)*9.81*DELX
650 PRINT "STATIC PRESSURE DROP", SP
660 ROEF = (1-EPS)*ROEP+EPS*ROEG
670 Delp = DELGAS^2*ROEG/2
680 PRINT "ROEF", ROEF
690 Delp = Delp + SP
700 PRINT "PRESSURE DROP", Delp
710 PTOT = PTOT + Delp
720 PRINT "SOLIDS VELOCITY", VEL
730 PL = Delp/DELX
740 PRINT "PRESSURE PROFILE", PL
750 PRINT "TOTAL PRESSURE", PTOT
760 IF X < 2.6 GOTO 330
770 STOP
780 END
10 DIM XSTPD(10)
20 DIM XSTPD(11)
30 DIM PL(10)
40 DIM PLD(10)
50 DIM PR(20)
60 DIM EPS(10)
60 PRINT "HOW MANY PROBES" 
70 PRINT "IN THE DILUTE CORE"
80 INPUT NP
90 NP = 13
100 FOR I = 1 TO NP
110 PRINT "HOW FAR FROM PREVIOUS PROBE"
120 INPUT XSTPD(I)
130 NEXT I
131 XSTPD(1) = 0.0107
132 XSTPD(2) = 0.18
133 XSTPD(3) = 0.18
134 XSTPD(4) = 0.18
135 XSTPD(5) = 0.18
136 XSTPD(6) = 0.18
137 XSTPD(7) = 0.18
138 XSTPD(8) = 2.25
139 XSTPD(9) = 0.25
140 LSTP = 2.57
141 XSTPD(10) = 0.25
142 XSTPD(11) = 0.25
143 XSTPD(12) = 0.26
144 XSTPD(13) = 0.26
150 XMIN = 0 : XMAX = LSTP
160 YMIN = 0 : YMAX = 1000
161 PRINT "WHAT IS GAS DIFFUSIVITY"
162 INPUT DF
163 PRINT "WHAT IS FIXED BED VOIDAGE"
164 INPUT EPSF
165 PRINT "WHAT IS REACTION RATE CONSTANT" 
166 INPUT K1
167 PRINT "WHAT IS INLET CONCENTRATION"
168 INPUT CAS
170 PRINT "WHAT IS PARTICLE DIAMETER"
180 INPUT DP
190 DP = 64E-6
200 PRINT "DENSITY OF PARTICLES"
210 INPUT ROEP
220 ROEP = 884
230 MP = 3.142 * ROEP * DP * 3 / 6
240 PRINT "PARTICLE MASS", MP
250 PRINT "DENSITY OF GAS"
260 INPUT ROEG
270 ROEG = 1.2
280 PRINT "VISCOSITY OF GAS"
290 INPUT VIS
300 VIS = 1E-5
310 SC = VIS / (ROEG*DF)
320 PRINT "WHAT IS INITIAL GAS VELOCITY"
330 INPUT UGAS0
340 PRINT "WHAT IS GAS MASS FLOWRATE"
350 INPUT GS
360 A = (16*VIS) / (DP^2*POEP)
370 PRINT "VALUE OF A", A
380 PRINT "WHAT IS INITIAL SOLIDS VELOCITY, DOWNWARDS POSITIVE"
390 INPUT VELIN
400 VELIN = -1.08
410 PTOT = 0
420 VEL = VELIN
430 PRINT "DIAMETER OF RISER"
440 DR = 0.05
450 PRINT "SOLID CIRCULATION RATE"
460 INPUT SCR
470 VOIDC = 4*SCR / (ROEP*3.142*DR^2)
480 PRINT "VOIDAGE CONSTANT", VOIDC
490 UGAS1 = UGAS0
500 CAH = CAS
510 I = 0
520 X = 0
530 XL = 0
540 I = I + 1
550 XL = XL + XSTPD(I)
560 PTOT = 0
570 X = 0
580 DELP = 0
590 DELX = 0
600 EPS = 1 + (VOIDC / VEL)
610 B = A*EPS^(-4.7)
620 DEL2 = -(B*VEL) + (B*UGAS0 + 9.81)
630 DELTIM = 0.01
640 DELX = DELTIM * VEL
650 X = X + DELX
660 DELVEL = DEL2 * DELTIM
670 NP = 6*(1-EPS)*DR^2 / (4*DP^3)
680 H = (VEL + DELVEL)^2 - VEL^2
690 KE1 = 0.5*MP*H
700 PE1 = -MP*9.81*DELX
710 ENTOT = KE1 + PE1
720 ENRATE = NP*ENTOT*VEL - (DELX*GS*9.81)
730 J = UGAS0^2 - (2*ENRATE/GS)
740 UGAS2 = -(J^0.5)
750 KELOSS = 0.5*GS*(UGAS0^2 - UGAS2^2)
760 DELGAS = UGAS0 - UGAS2
770 DEL = -(1-EPS)*(ROEP-ROEG)*9.81*DELX
780 ROEP = (1-EPS)*ROEP + EPS*ROEG
820 DELP = DELGAS^2 * ROEG / 2
830 FS = (-12.1 * (1 - EPS)) / (VEL * EPS^3)
840 SWF = (1 - EPS) * ROEP * DELX * VEL^2 / 0.05
850 DELP = DELP + SWF + FR + SP
860 VEL = VEL + DELVEL
870 RET = (UGAS + VEL) * DP * ROEG / VIS
880 EPS1 = (1 - EPS) / (1 - EPSM)
890 VP = 4 / 3 * 3.14 * (DP / 2)^3
900 EPS = 4 / 3 * 3.33 * EPS
910 T = T + DELT
920 AP = 3.14 * DP / 4
930 CAH = CAS * EXP(-T / (VC / (HM * AP)) + 1 / (K1 * EPS1))
940 PTOT = PTOT + DELP
950 IF (X < XSTPD(I)) GOTO 910
960 GOTO 610
970 STOP
980 X2 = -X
990 PRINT "PRESSURE PROFILE", PLD(I)
1000 PRINT "CAH", CAH
1010 PRINT "EPS", EPS
1020 LSTP = 2.58
1030 XSTP(10) = 0.1
1040 XSTP(11) = 0.18
1050 XSTP(12) = 0.18
1060 END
1070 PRINT "HOW MANY PROBES"
1080 PRINT "IN THE DOWFLOW REGION"
1090 INPUT NP
1100 NP = 13
1110 FOR I = 1 TO NP
1120 PRINT "HOW FAR FROM PREVIOUS PROBE"
1130 INPUT XSTP(I)
1140 NEXT I
1150 XSTP(1) = 0.04
1160 XSTP(2) = 0.26
1170 XSTP(3) = 0.26
1180 XSTP(4) = 0.26
1190 XSTP(5) = 0.26
1200 XSTP(6) = 0.26
1210 XSTP(7) = 0.26
1220 XSTP(8) = 0.18
1230 XSTP(9) = 0.18
1240 LSTP = 2.58
1250 XSTP(10) = 0.1
1260 XSTP(11) = 0.18
1270 XSTP(12) = 0.18
1424 DESTP(13)=0.1775
1410 PRINT "WHAT IS PARTICLE DIAMETER"
1420 INPUT DP
1430 PRINT"INPUT GAS DIFFUSIVITY"
1431 INPUT DF
1432 PRINT"INPUT FIXED BED VOIDAGE"
1433 INPUT EPSM
1434 PRINT"INPUT REACTION RATE CONSTANT"
1435 INPUT K1
1436 PRINT"INPUT INLET CONCENTRATION"
1437 INPUT GAS
1440 PRINT"DENSITY OF PARTICLES"
1450 INPUT ROEP
1460 ROEP = 531
1470 MP=3.142*ROEP*DP^3/6
1480 PRINT "DENSITY OF GAS"
1490 INPUT ROEG
1500 ROEG=1.2
1510 PRINT"VISCOSITY OF GAS"
1520 INPUT VIS
1530 VIS=1E-5
1531 SC=VIS/(ROEG*DF)
1540 PRINT"WHAT IS GAS MASS FLOWRATE"
1550 INPUT GS
1560
1570 A=(19*VIS)/(DP^2*ROEP)
1580 PRINT"VALUE OF A", A
1590 PRINT"INITIAL SOLIDS VELOCITY"
1600 INPUT VELIN
1610 VELIN = 1
1620 VEL=VELIN
1630 DR=0.05
1640 GOTO 2270
1670 VOIDC = 4*SCRD/(ROEP*3.142*DR^2)
1680 PRINT"VOIDAGE CONSTANT", VOIDC
1690 I=0
1700 XL=0
1710 I=I+1
1720 XL=XL+XSTP(I)
1730 PTOT=0
1740 X=0.004
1750 DELP=0
1760 EPS=1-(VOIDC/VEL)
1770 B=A*EPS^(-4.7)
1780 DEL2=-(B*VEL)+(3*UGAS0+9.81)
1790 DELTIM = 0.01
1800 DELX=DELTIM*VEL
1810 X=X+DELX
1820 DVEL=DEL2*DELTIM
1830 DVEL2=9.81*DELTIM
1840 VEL=VEL+DVEL
1850 VEL2=VEL+DVEL2
1860 NF=6*(1-EPS)*DR/DP^2
J 870 P i V E L 2 A 2 - V E L ' 2
1330 KE*0.5*MP*H
1900 J=UGAS0^2-(ENRATE*2/GS)
1910 UGAS2=-(J*0.5)
1920 DELGAS=UGAS0-UGAS2
1930 SP=(1-EPS)*(ROE+ROEG)*9.81*DELX
1940 ROEF=(1-EPS)*ROE+EPS*ROEG
1950 DELP=DELGAS^2*ROEG/2
1960 DELP=DELPGAS2*ROEG/2
1970 PTOT=PTOT+DELPGAS2
1980 IF X>XSTP(I) GOTO 2010
1990 GOTO 1760
2000 STOP
2010 X2=X
2015 EPS(I)=EPS
2020 PT(I)=PTOT/X2
2030 PRINT "PRESSURE PROFILE", PL(I)
2040 IF X<LSSTP GOTO 1710
2050 PRINT "RETURN FOR TOTAL"
2060 INPUT YES
2070 YES=0
2100 FOR I = 1 TO 13
2110 J=14-I
2120 PT(I)=PL(J)+PLD(I)
2125 PRINT "TOTAL PROFILE", PT(I)
2130 NEXT I
2140 FOR I=1 TO 12
2141 J=14-I
2142 AVEPS=0.5*(EPS(J)+EPS(J-1))
2143 PRINT "AVEPS", AVEPS
2144 EPS11=(1-AVEPS)/0.55
2145 TRES=-XSTP(J)/UGAS0
2146 CAH=CAH+EXP(-TRES*K1*EPS11)
2147 PRINT "CAH", CAH
2148 NEXT I
2260 END
2270 ROEF=486.74
2280 DS=0.15
2290 VSB=0.01767
2300 PRINT "CHANGE OF S.B.HEIGHT"
2310 INPUT DH
2320 VSB=ASB*DH
2330 PRINT "HEIGHT OF SLOW BED"
2340 INPUT HBED
2350 VBED=ASB*HBED
2360 VTOT=0.038
2370 V2=0.036-VBED
2380 R=5.89E-3/V2
2390 PRINT "VOLUME RATIO", R
2400 CM=VSB*ROEF
2410 MR=R*M
2420 SCR1=SCR
2430 US=-UGAS0-0.2
2470 HR=3.0
2480 TSI=HR/V5
2490 M1=TSI*SCR1
2500 MD=MR-M1
2510 VS=1
2520 TSD=HR/V5
2530 SCRD=MD/TSD
2540 PRINT"SCR IN Downflow",SCRD
2550 GOTO 1670
2560 STOP
2570 END
APPENDIX G

FLUID JETS
It seems important at this point to introduce the fluid jets, which play an important role in many turbulent mixing operations such as mixing in piping, jet mixing in tanks.

Tollmien (1926) and Ruden (1933) both using piezometric traverse methods across an air-jet, showed that in an axially symmetrical jet, the jet appeared to be in the form of a regular cone with its origin some distance behind the nozzle-mouth. In the region close to the nozzle-mouth the axial velocity profile remains constant and the velocity profile gradually changes over from being virtually flat to the Gaussian shape. Some of the features of turbulent jets (Re > 2100) are illustrated in Figure G1. In region I termed the potential core, the velocity is equal to that in the nozzle itself. The potential core is \( z_c \). Region II is an intermediate zone in which the flow is both turbulent and nonturbulent. Region III is the mixing zone that lies between the undisturbed fluid. In region IV the mixing zone and potential core merge and form a totally turbulent pattern.

Regarding to the jet angle ie the angle subtended by the cone, various values can be derived from the published data. Tollmien (1926) recommended a complete angle of 24°.
From visual observations, the total jet angle appears to be roughly 14°. Donald and Singer (1959) observed that jet angles increased with kinematic viscosity of the fluid. By considering the design of the 3 distributors used during the experimental work and by assuming that the jet angle is of 14° or 24°, using simple geometry the potential core length can be determined for each orifice diameter of the distributor.

\[ \tan \theta = \frac{x}{r_t - r_o} \]
\[ x = \tan \theta (r_t - r_o) \]

For \( \omega = 14°, \theta = 83° \)

<table>
<thead>
<tr>
<th>No. of holes</th>
<th>( r_o ) (cm)</th>
<th>( r_t ) (cm)</th>
<th>( x ) (cm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.5</td>
<td>2.5</td>
<td>16.3</td>
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<tr>
<td>7</td>
<td>0.19</td>
<td>0.75</td>
<td>4.6</td>
</tr>
<tr>
<td>13</td>
<td>0.14</td>
<td>0.625</td>
<td>3.95</td>
</tr>
</tbody>
</table>

1 nozzle case  7 hole case  13 hole case
For \( \omega = 24^\circ \), \( \theta = 78^\circ \)

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<th>( r_t ) (cm)</th>
<th>( x ) (cm)</th>
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</thead>
<tbody>
<tr>
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