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**DEMONSTRATION OF AN ADVANCED
CIRCULATING FLUIDIZED BED COAL
COMBUSTOR PHASE I: COLD MODEL STUDY**

CDO/R(D) - 86 - 55B

FINAL REPORT

PROJECT MANAGER

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OFFICE IS GRATEFULLY ACKNOWLEDGED**

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EXECUTIVE SUMMARY

This study was concerned with the research and development of a Swirling Circulating Fluidized Bed Coal Combustor system. Specifically, this report only addresses the cold model studies on swirling circulating fluidized bed. Three separate sets of experiments were conducted on a 3.5 m high and 102 mm internal diameter primary bed, a 4.2 m high and 193 mm internal diameter primary bed and a 5.1 m high and 298 mm internal diameter primary bed. Each primary bed had specially designed secondary air nozzles to provide secondary air injection at various velocities and direction.

It was found that there was a strong dependence of the density profile on the secondary air injection location and that there was a pronounced solid separation from the conveying gas, due to the swirl motion. Furthermore, the swirl motion generated strong internal circulation patterns and higher slip velocities than in the case of nonswirl motion as in an ordinary circulating fluidized bed. Radial solids flux profiles were measured at different axial locations. The general radial profile in a swirling circulating fluidized bed indicated an increased downward flow of solids near the bed walls, and strong variations in radial profiles along the axial height.

Studies of erosion in these circulating fluidized beds showed that the amount of erosion was dependant on the particles concentration and velocities moving parallel to the wall inside the wall sublayer zone. That means in an 193 mm diameter bed with the same air velocity and particle velocity, the amount of erosion is theoretically the same as in a 102 mm diameter bed.

Results obtained in this study show that particle residence times in a swirling circulating fluidized bed are substantially higher than in an ordinary circulating fluidized bed. The tangential velocity profile approach that of solid body rotation. For swirl numbers greater than 0.2, substantial secondary air penetration occurs resulting in persistence of swirl motion upto 10 diameters upstream of the secondary air injection ports. Furthermore, for swirl numbers less than 0.9, which is typical for swirling circulating fluidized beds, there is no significant increase in erosion due to swirl motion inside the bed. Pending further investigation of swirl motion with combustion, at least from our cold model studies, no disadvantages due to the introduction of swirl motion were discovered.

INTRODUCTION

The objectives of this research program were as follows:

- 1. To conduct research on the swirling concept and develop experimental data through bench scale experiments; and**
- 2. To develop an experimental plan for continuing the testing of the proposed concepts on a pilot scale at an appropriate facility to be located at the end of the research program.**

This report only addresses the first objective. To achieve this objective, the following detailed studies were conducted:

- 1.1 Measurement of local solid fluxes in the primary bed for different gas velocities and at different solid recirculation flowrates.**
- 1.2 Experiments were conducted with different particle diameters and solid densities. Initially six types of solid particles were selected for the study. These were: FCC catalyst, Coarse Alumina, Fine Alumina, Pyrites cinder, and Iron ore conc. However, after the scaling relationships were developed, it was found that none of these solids satisfied the equations. This meant that any measurements made with these solids could not be related to the hot swirling circulating fluidized bed system. Since the objective of this study was to examine the characteristics of a swirling bed and then relate them to the combustion system, two types of solid particles, namely steel grit and sand, were selected that satisfied the scaling relationships.**
- 1.3 Measurement of the axial pressure profiles at constant gas velocity and at various solid flowrates.**
- 1.4 Conduct studies on three bed sizes: 102 mm, 193 mm and 298 mm internal diameter.**
- 1.5 Quantify the extent of erosion in the cold model system by measuring the weight loss of material from a surface placed on the inside surface of the bed.**
- 1.6 Measure the penetration of the secondary air stream inside the bed. Initially it was conjectured that his penetration of the secondary air stream would be critical in maintaining the swirl flow inside the bed. However, upon experimentation it was found that the extent of penetration of the secondary air stream is not important, but rather the Swirl Number itself quantifies this effect.**

- 1.7 Investigate other methods of injecting secondary air with the goal of minimizing erosion. Since the extent of erosion in a swirling circulating fluidized bed was not found to be significant over an ordinary circulating fluidized bed, the question of minimizing erosion was found to be not important.

GENERAL BACKGROUND

1.1 The concept of circulating fluidized bed

Circulating fluidized bed (CFB) technologies have been in industrial use since the 1940's. The first commercialized circulating fluidized bed technology was the fluid catalytic cracking (FCC) process in the petroleum refinery industry. Despite the potential for widespread industrial application, very little academic research has been conducted on the circulating fluidized bed. In the 1970s, Yerushalmi and Avidan in the City University of New York started the first hydrodynamics study of circulating beds. Since that time, there have been many discussions about the hydrodynamics of the CFB system.

To describe the circulating fluidized bed, we should first understand the different regions of a fluidized bed. When a gas is flowing through a bed of packed particles, at very low gas velocities, the bed is fixed, as shown in Figure 1.1. With increase in the gas velocity, the particles start to vibrate and the bed begins to expand, with the voidage between particles increasing due to bubble formation, leading to a bubbling fluidized bed. Further increase in the fluidizing velocity leads to a turbulent bed, with a large bed expansion and very violent motion of particles. When the gas flow rate is continuously increased, at some point the dense phase solids start traveling vertically upward and are carried out of the bed. If the particles are not recirculated to the bed, the bed will be emptied in a very short time. This type of bed is called fast bed or circulating fluidized bed.

Figure 1.2 illustrates a vertical gas-solids transport line. The circulating bed is different from the conventional fluidized bed. The solid particles circulate in a closed loop in which entrained material is

Figure 1.1 The different regimes of fluidization as a function of the fluidizing velocity

Figure 1.2 Schematic diagram of a circulating fluidized bed

captured in a dust collector and then returned to the secondary bed. The valve between the secondary and primary bed is used to independently control the rate of circulation of solid particles.

Yerushalmi (1978) observed the circulating bed through a transparent wall and found that there are many strands of refluxing particles continuously coalescing and reappearing. These clusters of particles were dispersed uniformly in the reactor and were in a continuous state of formation and destruction. They behaved like large particles. The average slip velocity, which is defined as the relative velocity between gas and solids in a circulating fluidized bed, may be many times greater than the terminal velocity of an individual particle. This is illustrated in Figure 1.3. Because of the high slip velocities there are a number of potential advantages:

- (i) The ability to maintain dense beds at high gas velocities and allow a high throughput of gas in the unit.
- (ii) Intense backmixing of solids, promoting temperature uniformity.
- (iii) The ability to vary the solids concentration in the bed by independently changing the external circulation rate of solid.

These features suggested that the circulating bed is an effective unit operation for conducting many chemical reactions.

1.2 Literature review of circulating fluidized bed

It is important to know the solid density profiles in a circulating fluidized bed system. The density profile along the reactor determines the gas-solid contacting history. In catalytic reactions, the conversion and reaction rate is determined by gas-solid relative velocity and solid particle concentration. It is essential to understand the contacting

Figure 1.3 Slip velocities in high velocity fluidization according to Yerushalmi and Cankurt (1978)

history to predict the conversion. In a circulating bed combustor, the density profile will determine heat transfer, temperature distribution, furnace exit temperature, carbon and pollutant concentrations. The local density value will influence the suspension to wall heat transfer coefficients. (Kobro and Brereton, 1985)

Experimental studies conducted at City University of New York by Yerushalmi et al (1978) and similar studies by Li and Kwauk (1980) show that density profiles in circulating fluidized beds may generally be described by an approximately "S" shaped profile as shown in Figure 1.4. They stretch from a high density asymptote at the base of the column, or at some imaginary point below the base, through an inflection point to a low density asymptote at or beyond the top of the unit.

The Li and Kwauk model treats the circulating bed voidage profiles in terms of clusters of particles. The density profile is according to a balance of "diffusive" and "buoyancy" forces acting on each cluster. The profile is characterized by the equation

$$\ln \left\{ \frac{e - e_a}{e^* - e} \right\} = \frac{-1}{Z_0} (Z - Z_i) \quad (1.1)$$

where the fitted constants e_a , e^* and Z_i have a physical significance as shown in Figure 1.5. Z_0 was termed the characteristic length for the fast fluidization, Z_i is the location of the inflection point. Equation (1.1) yields an S-shaped curve, towards e^* at the top as $z \rightarrow -\infty$, and towards e_a at the bottom as $z \rightarrow +\infty$. Different factors have been found that influence the shape of the profile. The variables include

Figure 1.4 Density profiles for fast fluidization
(Li and Kwaul, 1980)

Figure 1.5 The Li and Kwauk model for density profiles in fast fluidized bed

particle size, particle density, gas viscosity, gas velocity and solids circulation rate.

A number of models which exist in the literature on circulating fluidized beds consist of an upflowing dilute core and a descending dense annulus. Many workers have reported observing this phenomena. Van Breugel et al. (1969) studied solids fluxes in an upward-moving gas-solids stream have reported a parabolic flux profile with a downward flux in the wall region. They showed that the presence of the solids significantly affected the gas velocity profile. The solids can create steep gradients of gas velocity where most of the gas is channelled through the center of the riser.

1.3 Swirling circulating fluidized bed

Industrial circulating fluidized bed combustion systems are characterized by addition of the fluidizing air at more than one location. For example, the Lurgi unit at Duisburg utilizes air introduction at three vertical locations, primary air, secondary air and a third fraction of air (Wein and Felwor, 1986). Staged air promotes staged combustion, which reduces formation of NO_x, and results in formation of a lower velocity higher density bed in the lowest or primary zone. This dense bed is advantageous for mixing of highly reactive fuels (Kobro and Brereton, 1985). However, despite its importance to practical operation, the effect of staged air introduction upon circulating bed behavior is not well understood.

The main advantage of a SCFB is to provide increased residence time of the particles in the bed due to the helical motion of the particles. The SCFB would have a higher slip velocity (i.e., the relative

velocity of air to the particles) in comparison to that of a CFB because the slip velocity vector is at an angle instead of being in the vertical direction.

1.4 Degree of swirl

The degree of swirl of air flowing through a pipe is usually characterized by the swirl number S , which is nondimensional number representing axial flux of swirl momentum divided by axial flux of axial momentum, multiplied by the equivalent radius, i.e., --

$$S = \frac{G_q}{G_x D/2} \quad (1.2)$$

where G_q = axial flux of swirl momentum

$$= \int r(u^2 - \frac{w^2}{2}) dr$$

$$G_x = \text{axial flux of axial momentum} = \int r^2 u w dr$$

D = diameter of fluidized bed column

Both the axial fluxes of angular and linear momenta can be evaluated from integration of velocity profiles. For the case of a solid body rotation flow with a uniform distribution of the axial velocity

$$w = w_{mo}(r/R)$$

$$u = u_{mo}$$

if the ratio of the maximum values of swirl and axial velocity is G ,

where $G = w_{mo} / u_{mo}$

It can be shown that

$$S = \frac{1/2 G}{1 - 1/4 G^2} \quad (1.3)$$

1.5 Erosion in the primary bed

Studies on erosion due to particle impaction were first conducted in 1931 in Germany with the erodent particles (smoke and dust) impacting different shapes. A survey of the erosion literature prior to 1946 has been presented by Wahl and Hartstein (1946). Although there is much practical information discussed in this survey, information regarding the mechanism of surface material removal during erosion is lacking. Siebel and Brockstedt (1941) investigated the erosion of various materials (hard, soft, alloyed and unalloyed steels) due to a stream of quartz sand directed perpendicular to the surfaces. The results indicated that the erosion was very much the same for all the samples. These findings were in direct contrast with actual experience. Wellinger and various co-workers (1942) found that the erosion resistance of materials changed with the angle of impingement, and hence clarified the apparent contradictions of the results obtained by Siebel and Brockstedt (1941) and actual experience. After Holtey (1939), Wellinger and various co-workers (1942) broke erosion down into two processes called "rub" and "shock" wear. It was shown that soft steels are best under "shock" wear conditions, while hard steels are superior for "rub" type wear conditions. Finnie (1958, 1960a, 1960b) was the first to take into account particle velocity rather than fluid velocity in determining the effects of particle velocity on erosion. The particle velocities were measured using photographic techniques. He concluded

that sharp edged particles cause more erosion than dull edged (rounded) particles. For cases where abrasive particles are much harder than the surface of the specimen material, the hardness of the particles is relatively unimportant. When the particle hardness is less than that of the specimen material, erosion is greatly reduced. As a result of these studies, Finnie (1958, 1960a, 1960b) developed the first mathematical or theoretical model attempting to predict and understand the erosion mechanism.

The degree of erosion damage incurred on a part during operation is highly dependant on the nature of the erosion environment (Preece, 1977; Sheldon, 1977; Shewmon, 1983; Goodwin et al., 1969; Finnie and McFadden, 1978). The type of impinging particles, the velocity of these particles, the angle of attack of the impinging particles on the target material, and the temperature of the erosion environment, are several of the variables that affect the abrasiveness of the erosive environment.

The erosive effects of the impinging particle's angle of attack on the target material has been widely studied. Figure 1.6 illustrates the dependance of the erosion rate on the attack angle for both ductile and brittle materials. The erosion rate for ductile materials approaches zero at very low angles of impact, increases to a maximum when the incident particle makes an acute angle of 20° to 30° to the target surface, and then falls to 1/2 to 1/3 of the maximum erosion rate as the normal impact to the target surface is approached (Bellman and Levy, 1981). The erosion rate for brittle materials approaches zero at very low angles of impact, and increases to a maximum as the incident particles become normal to the target surface. Note that in a swirling circulating fluidized bed, the angle of attack for the particles will be small, and hence impact erosion will be

Figure 1.6 Erosion in a typical ductile and brittle material as a function of impingement angle

minimal. There are no systematic studies on erosion in swirling circulating fluidized beds.

Recent studies of erosion surfaces using the scanning electron microscope (SEM) show that Finnie's (1958, 1960a, 1960b) analogy with cutting (removing metal by scraping it out from the target surface in the form of a chip) is not completely valid, and occurs rarely. These SEM studies have revealed that at shallow angles, where a large horizontal component of the velocity overwhelms the particle's rotational tendencies, smear craters are dominant (Bellman and Levy, 1981). As the impingement angles increase, the ploughing type of crater, which is a combination of the smearing and indentation type craters, increases in frequency. At the higher impact angles, an indentation crater prevails.

The size, shape, hardness, and type of impinging particles has a definite effect on the abrasiveness of an erosion atmosphere. As shown in Figure 1.7, there is a significant size effect which is itself dependant on impact velocity (Goodwin et al. 1969). There is a critical particle size above which erosion is not influenced by size and this value appears to increase linearly with velocity. The erosion rate varies with velocity raised to the n^{th} power, where n is usually in the range of 2.2 to 3.0 having a mean value of 2.4 (Sheldon, 1977; Goodwin et al., 1969).

The temperature of the erosion environment will obviously have a significant impact on the erosion rate of a target material. As the temperature changes, the mechanical properties of the target material can change drastically, and thus change the erosion resistance of the material. The temperature also affects the properties of the erodent material.

Figure 1.7 The effect of impingement angle and particle size on erosion

APPARATUS AND MEASUREMENT SYSTEM

2.1 The Swirling Circulating Fluidized Bed System

2.1.1 Design considerations

The experimental measurements in the present study were conducted on three circulating fluidized bed units fabricated specifically for this work. This section describes the construction of these units and the accompanying instrumentation systems. Description of other pieces of apparatus are given at the point of reference.

Each circulating fluidized bed unit was designed with the following criteria:

- (i) It should be tall enough that a substantial portion of the unit operates beyond the particle acceleration zone, since the particles accelerate at the base of a circulating fluidized bed (CFB) column before reaching developed flow in the column.
- (ii) The unit must have as large a diameter as possible to make the results amenable to scale up by minimizing the wall effects.
- (iii) The unit should be designed to operate at superficial gas velocities as high as 10 m/s, a typical pneumatic transport velocity for a fine material and the typical operating range of circulating bed combustion systems.
- (iv) The unit should be able to control the circulation rate of solids independently of gas velocity.
- (v) Accommodation should be made to permit introduction of the air at more than one level to allow swirling flow in the primary bed.

The design of the unit was constrained by the following factors:

- (i) A maximum unit height of 6 m, restricted by the available headroom in the laboratory.
- (ii) An air supply to the riser of 300 scfm available at a delivery pressure of 3 psig.

Within the framework of these constraints three units was designed which are shown in Figures 2.1, 2.2 and 2.3. Each unit comprises of three basic sections:

- (i) A high velocity transport line,
- (ii) A solids separation section,
- (iii) Solids storage and return system.

Under steady operating conditions in any of the transport regimes the circulating fluidized bed typically operates as follows: Particles are fed at a constant rate from the storage column, through an L-valve into the riser at its base. The rate of solids flow is controlled by varying the rate of aeration to the L-valve. High velocity gas then carries the solids up the riser, and discharges them through an exit at the riser top into the cyclone. Each part of the circulating bed system is now described separately.

2.1.2 The riser columns

Column with internal diameter of 102 mm:

The swirling circulating fluidized bed riser was constructed of 102 mm (4") ID 114 mm (4 1/2 ") OD cast aluminum tubing. Individual flanged sections combined to give a maximum total column height of 3.5 m (138") . The air was introduced into the column through a perforated plate and a stainless steel screen at the column base.

In addition to the primary air, introduced at the base of the unit, air could also be introduced through secondary nozzles. These nozzles are

located 85 cm above the distributor in Figure 2.1. This location is arbitrary since they may be interchanged with any other section of the riser to produce geometries of interest; however, the configuration which is shown gives a dense bed depth typical of some industrial combustion applications. The secondary air injection module was specially designed for flexibility in positioning the secondary air nozzles. A schematic of the secondary air injection module is shown in Figure 2.4. Secondary air is injected tangentially through four nozzles, each nozzle rigidly attached to a separate ball which can be positioned in any way as in a ball and socket joint. This design allows the angle and radial location of each secondary air nozzle to be varied independently. The four nozzles can be positioned to create a swirl flow pattern inside the primary bed. The four nozzles can also be positioned opposite to each other to create a nonswirl flow.

To allow measurement of static and dynamic pressure at different points along the length of the riser, and to permit insertion of various intrusive probes, access ports were located at 25 cm intervals along the length of the riser.

The column is supported structurally at the base, and by support brackets of steel dexion on each of the sections of the riser.

Column with internal diameter of 193 mm:

A schedule 40 PVC pipe with an internal diameter of 193 mm was used to construct the primary bed. Flanged sections separating the bottom portion, secondary air injection section and upper section were designed to give a total primary bed height of 4.2 m. Air was introduced into the column through a perforated plate and a PVC screen at the column base.

(a) BALL AND SOCKET ARRANGEMENT

Figure 2.4 Secondary air injection module

In addition to the primary air, introduced at the bottom of the primary section, air could also be introduced through secondary nozzles located 85 cm above the distributor as shown in Figure 2.2. The secondary air injection section was designed in the same way as for the 102 mm riser column. The secondary air section consisted of four nozzles, each nozzle attached to a separate ball placed inside a ball and socket type joint. This allowed the nozzle to be positioned at any angle and location inside the primary bed. The position of the holes for measurement of solid mass fluxes and erosion rates were the same as for the 102 mm column.

Column with internal diameter of 298 mm:

A schedule 40 PVC pipe with an internal diameter of 298 mm was used to construct the primary bed. Flanged sections separating the bottom portion, secondary air injection section and upper section were designed to give a total primary bed height of 5.1 m. Air was introduced into the column through a perforated plate and a PVC screen at the column base.

In addition to the primary air, introduced at the bottom of the primary section, air could also be introduced through secondary nozzles located 85 cm above the distributor as shown in Figure 2.3. The secondary air injection section was designed in the same way as for the 102 mm riser column. The secondary air section consisted of four nozzles, each nozzle attached to a separate ball placed inside a ball and socket type joint. This allowed the nozzle to be positioned at any angle and location inside the primary bed. The position of the holes for measurement of solid mass fluxes and erosion rates were the same as for the 102 mm column.

2.1.3 The Gas-solid separation system, storage and recirculation system

Gas and solids leaving the riser pass through a short horizontal section of flexible hosing and enter the modified cyclone. This is a 12 in. ID round vessel, connected to a 3 in. ID plexiglass column, in which the test solid is stored. Solids are fed at the bottom of the column through a venturi type injector by means of an L-valve. The solid flow rate is controlled by the aeration rate at the elbow of the L-valve.

2.1.4 Solid circulation rate

The solid powder to be fluidized is charged to the secondary bed before the bed operation is begun. The amount of solid storage in the secondary bed is decided by the bed level which can be viewed through the plexiglass column. Solid can be maintained in the primary bed at any desired fluidization condition through independent control of the primary gas velocity and of the solid rate from the secondary bed to the bottom of the primary bed.

Before every measurement is begun the bed is operated at least half an hour to ensure that the steady state has been reached. At steady state, the solid feed rate must be equal to the solid entrainment rate. When this condition is reached, the solid bed level in the secondary bed must be fixed and the bed surface level can be viewed through the plexiglass wall.

There are two ways to measure solid circulation rates : In the first method, at a given moment, the recirculation air valve is closed, and the

rate of ascent of fixed bed level is timed. In the second method, without air flowing in the primary bed, the recirculation air valve is opened, the rate of descent of bed level is measured, and the amount of solid flowing into the primary bed is discharged and weighed. From this information, the solid rate in the fast bed is calculated. The solid circulation rate can be calibrated as a function of the air flow rate, as shown in Figure 2.5.

2.2 The Measurement System

2.2.1 Design criteria for a microstructural probe

The microstructural study of the SCFB was intended to (a) determine the time averaged solids mass flux at various locations and (b) measure the time-mean velocity profiles for different swirl number and solids circulation rate.

A large number of probes have been introduced into circulating and bubbling fluidized beds with the objective of determining various parameters (Grace and Baeyens, 1986). In addition several non-intrusive techniques have been used specifically for the point-averaged suspended solids concentration.

The following characteristics are required of immersed probes in the fluidized bed:

- (a) minimal disturbance of the bed.
- (b) measurement of local properties.
- (c) Rapid response to transients.
- (d) mechanical strength.
- (e) Mobility, i.e., it should be possible to relocate the probe.

Figure 2.5 Calibration diagram of solids circulation rate

In this research , we selected a Pitot probe and solid sampling probe to make the point measurement of air velocity and solid mass flow rate.

2.2.2 Solids sampling technique

Local mass flux measurements were performed by means of a sampling probe which can be introduced into the flow at any location across the column diameter. Its nose was placed either up or down in order to measure either upward or downward directed fluxes of particles. We assign the positive sign to upward fluxes and the negative sign to downward fluxes.

Equipment is connected as shown in schematic arrangement in Figure 2.6 with a sharp edged nozzle, 1/4 in. diameter of opening. The gas is drawn through the sampling train by a vacuum pump, metered by a rotameter. The rate of sampling is controlled by a valve. The dusty gas picked up by the sampling probe is filtered in a preweighed filter bag and held in a sintered metal thimble. The particles were sucked through the probe into a filter bag during a given time, and then weighed.

It is necessary to determine the aspiration rate which gives the true value of the local mass flux. Therefore, for each radial location, a series of samples were taken for a variety of aspiration rates. The results showed that the mass of recovered particles first increases with the aspiration rate and then remains constant over a rather large range, increasing again thereafter. This is probably due to the fact that the deviation of a particle from its initial trajectory requires a large force which can be achieved only if the aspiration velocity is very different from that existing in the flow at the point under consideration.

(a). Particle sampling equipment diagram

(b). Details of measuring probes

Figure 2.6 (a) Particle sampling equipment diagram
(b) Sampler probe

Consequently, this measurement gives the value of the local mass flux of particles .

2.2.3 Five-hole Pitot probe measurement

The gas flow field has been investigated by many researchers using various approaches such as computer simulation, flow visualization and time-mean velocity measurements for single-phase fluid flow . In order to make the point measurement of air velocity in a fluidized bed, the probe should be acutely sensitive to flow, convenient to manipulate and relatively compact. The five-hole Pitot probe will be a suitable instrument to measure the magnitude and direction of the velocity in a fluidized bed.

To obtain the three components of the swirl flow velocities in the axial, radial and azimuthal directions at a given location in the fluidized bed, the yaw angle β , pitch angle δ and the magnitude of total velocity V at that point must be known . The magnitude and direction of the velocities were measured by a five hole Pitot impact tube. The probe used in this study is a model DA-250-12 manufactured by United Sensor and Control Corp as shown on Figure 2.7. The five-hole Pitot probe is one of the few instruments capable of measuring both the magnitude and the direction of fluid velocity simultaneously. The sensing head allows for probe shaft rotation without altering the probe tip location. The probe was calibrated against an outside source of known velocity. The results is very close to the calibration curve supplied by the manufacturer.

The instrumentation assembly as shown in Figure 2.8, in addition to the five-hole Pitot probe, is composed of a manual traverse mechanism, one five-way ball valves, a differential pressure transducer, a power

Figure 2.7 Three dimensional directional probe sensing head

1. Fluidized bed duct
2. Swirl air injection port
3. Manual traverse mechanism
4. Five hole pitot probe
5. Switching valve
6. Pressure trasducer
7. Voltmeter

Figure 2.8 Pressure measurement system

supply, and an integrating digital voltmeter. The differential pressure transducer has a range of from 0 to 10 inch H₂O. The voltmeter is the TSI model 1076. A traversing gear mounted on the wall of the aluminum column, was used to position the five hole probe. The traversing gear mechanism could be used to slide the probe pitchwise in the radial direction as well as allow rotation of the probe about its axis. The traverse unit can be moved to different axial location to measure the pressures at different height of the bed. Each radial traverse measurement begins from the wall, the probe was moved across the radius at each level in increments of one-tenth the wall radius, that is 0.2 inch increments up to 2 inches at the centerline.

Each probe has five measuring holes located on its tip. A centrally located pressure hole measures pressure P_1 , while two lateral pressure holes measure pressures P_2 and P_3 . Before the production measurements are taken, the five-hole Pitot probe rotary vernier must be set for zero yaw angle so that the z and q axes of the measurement coordinate frame coincide with those of the test section. This is achieved by adjusting the yaw angle to be zero at the center of the test section inlet for nonswirling flow. The first measurement for each location is the yaw angle for a zero reading of $(P_2 - P_3)$. The yaw angle of flow is then indicated by the traverse unit scale.

When the yaw angle has been determined an additional differential pressure $P_4 - P_5$ is measured by pressure holes located above and below the total pressure (P_1) hole. Pitch angle is determined by calculating $\frac{P_4 - P_5}{P_1 - P_2}$ and using the calibration curve for the individual probe. As shown in Figure 2.9 at any particular pitch angle, the velocity pressure

(a) Pitch angle calibration characteristic

(b) Velocity coefficient calibration characteristic

Figure 2.9 Calibration characteristics for five-hole Pitot Probe

(a) Pitch angle calibration characteristic

(b) Velocity coefficient calibration characteristic

coefficient $\frac{P_t - P_s}{P_1 - P_2}$ and total pressure coefficient $\frac{P_1 - P_t}{P_t - P_s}$ can be read from the calibration curve, and $P_t - P_s$ and P_s calculated.

The velocity components are easily calculated by elementary geometry using the measured yaw angle, the derived pitch angle, and the total velocity :

$$V = [2 (P_t - P_s) / r]^{1/2} \quad (2.1)$$

$$u_r = V \sin \delta \sin b \quad (2.2)$$

$$u_q = V \sin \delta \cos b \quad (2.3)$$

$$u_z = V \cos \delta \quad (2.4)$$

The raw data measured are reduced by a FORTRAN computer program and output as axial, radial, and swirl velocity components.

EXPERIMENTAL RESULTS

3.1 Density profiles - Macroscopic Aspects

3.1.1 Considerations regarding use of pressure data

As noted earlier, the macrostructure of circulating fluidized beds has generally been characterized by the variation of suspended solids density with height along the column. Density profiles are generally inferred from the gradient of the absolute pressure profile, or from direct measurement of differential pressures over regular intervals along a circulating bed. The pressure drop is then ascribed to the weight of the solids and fluid per unit area, assuming the combined effects of gas-wall friction, solids-wall friction, and solids acceleration are neglected. This assumption, leading to

$$\rho_{\text{susp}} = -1/g \frac{dp}{dz} \quad (3.1)$$

has been shown to be accurate to within 10% by Turner (1978) for a 152 mm ID column. However Arena (1985) found that the error could be as high as 70% in a smaller column (41 mm ID) at velocities of 7 m/s. In this study we have employed equation 3.1 since, in all the runs of interest, gas-wall and solid-wall frictional pressure drop was calculated by Konno and Saito (1969) correlation as equivalent to a solids hold up of 1.3 Kg/m³. This was for a solids flux of 70 Kg/m²s at a gas velocity of 7 m/s in a 102 mm dia. column. An additional calculation shows that a suspension with a density of 100 Kg/m³, accelerated or decelerated between rest and 2 m/s over a 2 m distance, introducing an error of 20 Kg/m³ to the apparent density given by equation 3.1. Hence, errors may approach this magnitude at entrances and exits.

3.1.2 Particles chosen for studies

It is well known that the behavior of gas-fluidized system is related to the properties of the bed solids. Therefore, it is necessary to establish the range within which fluidized beds behave similarly. Geldart (1973) suggested the classification of bed solids into four different groups, characterized by the density difference between particle and gas ($\rho_p - \rho_g$) and the mean size of the solids. The properties observed with one powder can be extended to all powders in the same group without particular risk. Characteristics of the four groups considered by Geldart are described below:

Group A: Powders of group A have a small mean size and a low density.

Group B: This group contains median size particles with intermediate values of density.

Group C: All cohesive powders which are very difficult to fluidize belong to group C.

Group D: The powders in group D consist of either large and/or very dense particles.

Easy identification can be performed by using the density difference versus mean particle size diagram for fluidization with air at room temperature as shown in Figure 3.1.

3.1.3 Scaling Factors

Laboratory models which are fluidized at atmospheric pressure and temperature provide a means to study the fundamental bed hydrodynamics. A set of scaling relationships have been formulated from dimensional analysis that allow a bed operating at ambient conditions to model the hydrodynamics of a bed at elevated pressure

**Figure 3.1 Geldart's classification of powders:
case of fluidization by air at room conditions**

and temperature. Both beds must be geometrically similar, and must have equal values of the Reynolds number, the Froude number, the ratio of particle to fluid density, the dimensionless particle size distribution, and the particle sphericity. The non-dimensional parameters of cold bed and hot combustor should be equal (Glicksman, 1984):

$$\frac{r_s^2 d_p^3 g}{\mu^2}, \frac{u^2}{g d_p}, \frac{r_g}{r_s}, \frac{L}{d_p}, \frac{D}{d_p}, f_s$$

Consider the case where the real combustor is operated at an elevated temperature 1100° K and one atmospheric pressure. The scaled model is to be operated with air at ambient pressure and temperature. The fluid density and viscosity will be significantly different between these two gas conditions, e.g. the gas density of the cold bed is 3.5 fold higher than the density of the hot bed. With the solid density of the model determined, the Archimedes number is used to determine the particle diameter of the model and the Froude number is used to determine the superficial velocity. Table 3.0 prescribed the proper scaling between the hot bed and the ambient model.

The specifics of this report, notably the types of solids and the range of gas velocities employed, have been directed towards the combustion application. From the above scaling relationship, for a SCFB of 12 ft height and bed diameter 4 in., the physical properties of the solids were chosen as : particle diameter $d_p = 500 \mu\text{m}$, solid density $r_s = 7200 \text{Kg/m}^3$, leading to beds solids in groups B and D.

As pointed earlier in the introduction section of this report, the type of solids initially proposed for this study did not satisfy the above scaling relationships. This meant that conducting studies with these

particles would not have allowed us to relate the cold flow studies to the actual hot combustor.

Table 3.0 Scaling factors for modeling of hot bed performance

Variable	Hot Bed Variable (800 °C)	Scaled Cold Bed Variable(15 °C) (Variable in this experiment)
Superficial velocity	$2 u_{15}$	u_{15}
Particle Diameter	$4 d_{p15}$	d_{p15}
Particle density	ρ_{s800} of coal char $=1550\text{Kg/m}^3$	$3.5 \rho_{s15}$ of iron grit $=7200\text{Kg/m}^3$
Bed Height	$4 L_{15}$	$L_{15}=12 \text{ ft}$
Bed Diameter	$4 D_{15}$	$D_{15}= 4 \text{ inch, } 8 \text{ inch, } 12 \text{ inch}$
Particle sphericity	$f_s = 0.8$	$f_s = 0.8$

3.1.4 Initial studies

Initial studies were conducted with only primary air without swirl flows. Steel particles were used as the circulating solid. These particles had a mean diameter of 525mm, density of 6900 Kg/m³. The complete properties of the steel grit is given in Table 3.1. When the density profiles were plotted for a number of steady runs, they are shown in Figure 3.2 for a series of three runs in which total gas velocity was held essentially constant and the solids circulation rate G_s varied over a wide range. For each solid recirculation rate, G_s , measurements of density were made at different heights. Three measurements were taken for each air velocity and solid recirculation rate. The measurement errors were defined as follows:

$$r_{avg} = \frac{r_1 + r_2 + r_3}{3}$$

where r_{avg} = Average density at a given air velocity and solid recirculation rate and height.

r_i = Actual i th measurement of density at the given air velocity, solid recirculation rate and height.

$$\text{Measurement error} = \text{Absolute } | r_i - r_{avg} |$$

The maximum value of Absolute ($r_i - r_{avg}$) obtained for all heights was defined as the measurement error for the given air velocity and solid circulation rate. This error was plotted as a bar with a length equal to $2 \max | r_i - r_{avg} |$ in Figure 3.2 to 3.7.

Examination of the density profiles generated with steel shows trends in the lower half of the column consistent with the results of Yerushalmi (1978). Figure 3.2 shows the characteristic acceleration

Table 3.1
Properties of solid particles used in SCFB studies

Material	steel grit	sand particle
Tyler Screen size, mesh		
100% thru	16	28
100% on	28	35
D_p, arithmetic av.		
diameter (mm)	525	235
r_p, particle density		
Kg/m³	6900	2650

Figure 3.2 Longitudinal density distributions for steel in a circulating fluidized bed, $U_g=10.2 \text{ m/s}$, $G_s = 16, 31$ and $55 \text{ Kg/m}^2\text{s}$

zone followed by a dense phase and a decaying dilute freeboard. However, higher up the column, instead of a continuous decay of density or a steady density value, the density increases near the top of the unit. At the high velocity there is a dramatic exit effect which can stretch to occupy the whole upper half of the unit. The exit effect is because of the separation of solids from gas at the top of the exit column. The separation is due to the change of direction of air velocity and solid particles hit the top and bounced back. Two further runs at lower velocity with the same circulation rates are shown in Figure 3.3. Further increase the circulation rate at the low gas velocity will cause a violent slugging at the base of the column. This is called "choking" behavior. Because of the choking behavior, the air velocity can only be operated in a limited range.

All the subsequent experiments were performed with sand. The sand particle properties are listed in Table 3.1. Figure 3.4 and 3.5 show the density profiles. The exit effect does not exist in the density profile of sand particles. This is probably because these two kind of particles belong to different group.

The density profiles all show the S shape of Li and Kwauk's model. The shape of profile of figure 3.5 is qualitatively different from Figures 3.2 to 3.4. In Figure 3.5, the high density asymptote exist at the base of the column. In Figure 3.2 to 3.4 the asymptote is at an imaginary point below the base.

Figure 3.3 Longitudinal density distributions for steel in a circulating fluidized bed, $U_g = 8.6 \text{ m/s}$, $G_s = 17$ and $32 \text{ Kg/m}^2\text{s}$

Figure 3.4 Longitudinal density distributions for sand in a circulating fluidized bed, $U_g=8.5 \text{ m/s}$, $G_s=16$ and $30 \text{ kg/m}^2\text{s}$

**Figure 3.5 Longitudinal density distributions
for sand in a circulating fluidized
bed, $U_g=5.1$ m/s, $g_s= 25$ kg/m²s**

3.1.5 The impact of secondary air addition on circulating bed density profiles

The primary zone can operate in a variety of fluidization regions in a combustion system, depending upon the load, mean particle size, and the fuel type. In this study, this region was varied by varying the primary-to-secondary air split while total gas flow rate was maintained constant. The total air flowrate and secondary to primary air ratios are listed on Table 3.2.

A series of tests were conducted with air entering the column at the primary distributor and through four secondary ports located 85 cm further up the column. Two methods of positioning the secondary air were tested, firstly with the directly opposed ports, and later with the four tangential horizontal (swirl) nozzles.

For each geometry the total gas velocity and solids circulation rates were maintained constant at 8.6 m/s and 16 kg/m²s respectively.

Results from the different runs are shown in Figures 3.6 and 3.7. They are discussed in the discussion section of this report.

Table 3.2 Total air flowrate and secondary to primary air ratios

Secondary m³/s	0.006	0.013	0.027
Primary m³/s	0.060	0.053	0.039
S/P	0.1	0.264	0.59
Total =Secondary+Primary =0.066 m³/s

Figure 3.6 Density profiles measured in circulating beds of sand at a total gas velocity of 8.6 m/s and a solids circulation rate of 16 kg/m²s with different secondary to primary air ratios. Secondary air introduced through opposed ports

Figure 3.7 Density profiles measured in circulating beds of sand at a total gas velocity of 8.6 m/s and a solids circulation rate of 16 kg/m²s with different secondary to primary air ratios. secondary air introduced through tangential direction

3.2 Experimental studies on the Microstructure

3.2.1 Solid mass flux profile

Mass flux profiles were generated by traversing a radius of the riser at different axial positions for circulating bed. These profiles are shown in Figures 3.8 . The results give information about the internal recirculation of particles: the points located under the zero value of net mass flux are in the vicinity of the wall where most of the particles flow downward. It can be seen from Figure 3.8 that the relative rate of internal recirculation decreases with increase of the height of the bed. It is because of this radial mass flux profile develops with height causing a gradual density decay in a circulating fluidized bed.

The general radial profile of a swirling bed of Figure 3.9 indicated an increased downward flow of solids near the riser wall, and strong variations in radial profiles along the axial location. This radially non-uniform structure was used to explain the macroscopic phenomena of the secondary swirl air addition of the circulating fluidized bed .

3.2.2 Results of air velocity profile

The degree of swirl of air flowing through a pipe is usually characterized by the swirl number S , which is nondimensional number representing total torque across the pipe section divided by total flux of axial linear momentum, multiplied by the equivalent radius, i.e., --

$$S = \frac{G_g}{G_x \frac{D}{2}} \quad (3.2)$$

where

Figure 3.8 Variation in solids flux profile with axial position for the circulating bed at $U_g = 8.5$ m/s and $G_s = 16$ kg/m²s

Figure 3.9 Solids mass flux profile of sand
of $U_g=8.6$ m/s and $G_s= 16\text{kg/m}^2\text{s}$
at height of 125 cm from distributor with
different secondary to primary air ratios
for swirl flow

$$G_x = 2\pi\rho \int_0^R u_z^2 r \, dr = \text{the total linear momentum across the pipe section}$$

$$G_\theta = 2\pi\rho \int_0^R u_z u_\theta r^2 \, dr = \text{integration of torque across the pipe section}$$

D = diameter of fluidized bed column

Both G_x and G_θ can be evaluated from integration of velocity profiles.

Measurements were first made without particles to understand the swirl flow pattern of air. With same total amount of air of primary and secondary flow. The calculated swirl number corresponding to each split are shown on Table 3.3.

A study of the results of tangential velocity profile (Figure 3.10.) indicates that the flow pattern can be approximated by three zones:[1] Forced vortex zone: This zone contains inner part of the tube. The tangential velocity always increase with increasing radius. [2] Transition zone: This zone usually contains the maximum tangential velocity and is located between forced and free vortex zone.[3] Free vortex zone: Near the tube wall area the tangential velocity always decrease with increasing radius. The tangential velocity profile, across the radius, showed the velocity to increase from zero at the center to a maximum at a radius of injection port location, and then fall to zero again at the wall. Because the viscous dissipation weakens the swirl the tangential velocity will steadily decrease downstream from the inlet. However the location of each zone did not change with the swirling flow moving up the pipe.

Figure 3.10 Tangential velocities for single air flow without particles in 4 inch diameter bed at different locations and different swirl numbers for total air velocity 8.5 m/s and $G_s=0$

When swirl flow is injected into the fluidized bed with particles flow, the tangential velocity will be weakened not only by viscous dissipation but also by friction between gas and solid particles as shown in Figure 3.11.

Table 3.3 Calculated swirl number for different ratio of air split

S	0.066	0.134	0.234
$Q_{tan} \text{ m}^3/\text{s}$	0.006	0.013	0.027
$Q_{ax} \text{ m}^3/\text{s}$	0.060	0.053	0.039
Q_{tan}/Q_o	0.1	0.2	0.418
$Q_o = 0.066 \text{ m}^3/\text{s}$			
S/P	0.1	0.264	0.59
($=Q_{tan}/Q_{ax}$)

Figure 3.11 Tangential velocities for air and particles flow in 4 inch diameter bed at different locations for swirl number 0.134 of different solid mass flux

3.2.3 Error estimation

The static pressure measurement error because of the cross sectional flow area blocked by the probe is experimentally found by manufacturer to be:

$$\frac{P_{s1} - P_{s2}}{P_t - P_{s1}} = 1.2 \frac{a}{A} \quad (3.3)$$

where P_{s1} is original indicated static pressure, P_{s2} is modified static pressure, P_t is total pressure. In our system the measured static pressure is calculated to be in the range of 0.01 inch H₂O.

Pitch angle errors are dependent on the calibration chart. The pressure transducer error range is 0.01 inch H₂O ,which in turn induce 2.5 degree pitch angle error.

Yaw angle errors are assumed zero when the probe is rotated $P_2 = P_3$, the error caused by rotary vernier is no more than 1 degree.

As a result, the calculated total velocity is subject to about 10 % error. The relative error of radial velocity will be larger than that of swirl velocity and axial velocity because of the yaw angle and pitch angle error.

3.3 The Effect of Bed Diameter

Separate studies were conducted with primary beds of internal diameter 193 mm and 298 mm. The purpose of these studies was to examine the effect of bed diameter on the hydrodynamics of circulating and swirling circulating fluidized beds.

Figures 3.12 and 3.13 show the relative static pressure profile for the 193 mm internal diameter riser column with air and particle flow. The pressures are reported at various axial locations for different swirl

Relative static pressure ($P-P_c$) inch H₂O, where P_c is the pressure at tube center

Figure 3.12 Relative static pressure profile for the 193 mm internal diameter riser at different axial locations for swirl number 0.134 and mass flux = 16 Kg/m²s

Figure 3.13 Relative static pressure profile for the 193 mm riser at different axial locations for swirl number 0.234 and mass flux = 16 Kg/m²s

numbers and mass flux = $16 \text{ Kg/m}^2\text{s}$. Pressure is expressed as the difference with tube center.

Figures 3.14 and 3.15 show radial variation of the net solids mass flux in the primary bed for sand at gas velocity of 8.6 m/s , $G_s = 16 \text{ Kg/m}^2\text{s}$ and $32 \text{ Kg/m}^2\text{s}$, respectively and height of 125 cm from the distributor. The results for various primary to secondary air flowrate ratios have been shown. Figure 3.16 shows the tangential velocities in the bed at different axial locations.

Studies with a riser column of internal diameter 298 mm were also conducted to investigate the impact of bed diameter. Figures 3.17 and 3.18 show the relative static pressure profile in the bed with air and particle flow. Figures 3.19 and 3.20 show the solid mass flux profile for sand at 125 cm height from the distributor. The effect of primary to secondary air flow rate ratio is shown. Figure 3.21 shows the tangential velocity in the bed at different axial locations.

It should be noted that qualitatively the hydrodynamic characteristics of the 102 mm , 193 mm and 298 mm internal diameter beds were very similar. To further investigate the similarity quantitatively, a computer simulation program was developed to calculate the pressure profile, tangential velocities and solids flux in the three beds. The development of the program was supported by Cray Research, funded through the Ohio Consortium for Supercomputing Research. The results of this simulation study showed that the hydrodynamics of circulating and swirling circulating fluidized beds of various diameters could be represented by a simplified mathematical model, presented in the following section.

Figure 3.14 Solids mass flux profile in a 193 mm riser for sand at gas velocity of 8.6 m/s and $G_s = 16 \text{ Kg/m}^2\text{s}$ at height of 125 cm from distributor with different secondary to primary air ratios for swirl flow

Figure 3.15 Solids mass flux profile in a 193 mm riser sand at air velocity of 8.6 m/s and $G_s = 32$ Kg/m²s at height of 125 cm from distributor with different secondary to primary air ratios for swirl flow

Figure 3.16 Tangential velocities for air and particles in 193 mm ID riser at different axial locations and slid mass fluxes for swirl number = 0.134

Figure 3.17 Relative static pressure profile for 298 mm ID riser at different axial locations for swirl number 0.134 and mass flux = 16 Kg/m²s

Relative static pressure ($P-P_c$)
inch H₂O, where
 P_c is the pressure
at tube center

Figure 3.18 Relative static pressure profile for 298 mm ID riser at different axial locations for swirl number 0.234 and mass flux = 16 Kg/m²s

Figure 3.19 Solids mass flux profile in 298 mm ID riser for sand at air velocity of 8.6 m/s and $G_s = 16$ Kg/m²s at height of 125 cm from distributor with different primary air ratios for swirl flow

Figure 3.20 Solids mass flux profile in 298 mm ID riser for sand at air velocity of 8.6 m/s and $G_s = 32$ Kg/m2s at height of 125 cm from distributor with different primary air ratios for swirl flow

Figure 3.21 Tangential velocities for air and particles flow in 298 mm ID riser at different axial locations and solid mass fluxes for swirl number 0.134

DISCUSSION

4.1 Possible mechanisms for solids movement in CFB

The macrostructural measurement of pressure drop showed the solid density gradually decaying with height in a circulating fluidized bed. This is supported by the results of microstructural studies illustrating how radial mass flux profile develops with height. In the lower region of the bed, the mass flux profile had a strong radial nonuniformity with upward in the center. The high density at lower region gradually gave way to a more dilute core-annular type flow up the column. One explanation for the density decay in solids concentration is that solids move from an upflowing zone in the center of the riser to the downflowing zone at the riser wall. This is due to a natural tendency for solids to move into the low gas velocity region close to the wall perhaps by convective gas flow and radial gradients in turbulence intensity. Solids will fall downward in this low velocity region. Unless they are reentrained, the cross sectional mean density will decay with height.

Figure 4.1 shows an internal circulation pattern inside the circulating fluidized bed. On the left is a schematic in which arrows indicate approximate direction of solids flow and show the development of the wall layer. On the right is a second diagram gives an idea of how up and downflow fluxes vary with height in the unit to give a net positive flux where there is a net constant upflux of the riser. In order to obtain a longitudinal profile which remains invariant with time, the rate at which solids leave the reactor must become equal to their feed rate.

Figure 4.1 Schematic diagram showing solids fluxes in a riser of a circulating fluidized bed. On the left is a schematic in which arrows indicate approximate direction of solids flow and show the development of the wall layer. On the right is a second diagram gives an idea of how up and down flow fluxes vary with height in the unit to give a net positive flux

4.2 Effects of swirling air on solids motion

The phenomena associated with secondary air introduction can be explained in terms of the radially non-uniform structure of the circulating bed. Introduction of secondary air with directly opposed ports leads to graphs of longitudinal density profiles for different primary to secondary air splits at constant total velocity as shown in Figure 3.6. The profiles are characterized by a continuous decay in density from the primary distributor to exit. There appears to be a slight discontinuity in the slope of the density profile at the secondary air ports. As one would anticipate, lower primary velocity promotes higher primary zone density, but at the top of the reactor the profiles are identical. The profiles can be explained by considering two risers on top of one another joined at the secondary air ports by an upflowing core and downflowing annular exchange of solids (refer Figure 4.2).

The profiles obtained with swirl (tangential entry) secondary air are remarkably different (refer Figure 3.7). In these cases there is a discontinuity in the gradient of the density profile and a high density zone where the secondary air is introduced. Two separate zones are formed. The lower bed operates just as in a conventional circulating bed, with a characteristic decay with height due to radial transfer. Above the secondary air port a second zone is formed with increased internal flux. It was apparent that the difference was caused by the tendency for the swirl air to pick up annulus particles by virtue of the high angular velocities. These particles force the downflowing layer upward against itself, and into the center of the reactor where it was reentrained. Thus, the swirl air prevented the downflowing solids of core and annulus to exchange between the upper and lower zones.

Figure 4.2 Variation of solids fluxes in a swirling circulating fluidized bed. The density profile on the left hand side is thought to be caused by up, down and cross fluxes as shown on the right hand side

The swirl air had a substantial cyclonic effect, tending to intensify the core-annulus phenomena of fast fluidization. One might anticipate decreased decay lengths due to increased rates of solids transfer to the wall area.

MICROSTRUCTURAL ANALYSIS IN SCFB FROM MEAN VELOCITY DISTRIBUTIONS

For a nonswirling turbulent tube flows with boundary layer assumptions, only one component t_{rz} of t is significant. However, for a swirling tube flow, again with boundary layer assumptions, more than one component of shear stress are significant. In systems with high levels of turbulence, there is difficulty in obtaining accurate turbulence measurements in comparison to the relative ease of measuring mean velocities. The analysis described in here utilizes axial and swirl velocity measurements to determine the significant shear stress and pressure terms in a gas solid swirling flow field.

The equations of motion for the flow of a gas-solid suspension have been suggested by many authors. By using a continuous approach, formal equations were suggested, for example, by van Deemter, Hinze and Soo. Because lack of particle velocity and particle concentration distribution, in the following only gas phase equation of motion is discussed.

5.1 Analysis

The basic vector-tensor stress equations of conservation of mass and momentum are

$$\frac{Dr}{Dt} + r (\text{div } u) = 0 \quad (5.1)$$

$$r \frac{Du}{Dt} = - \text{grad } p + \text{div } t + r g \quad (5.2)$$

In a turbulent flow the expression for the total stress t is

$$\mathbf{t} = \mathbf{t}^l + \mathbf{t}^t \quad (5.3)$$

where $\mathbf{t}^t$ is turbulent shear stress tensor and $\mathbf{t}^l$ is molecular shear stress tensor. The molecular stress tensor is usually omitted in fully turbulent free flows since

$$\mathbf{t}^l \ll \mathbf{t}^t.$$

Assuming incompressibility it can be shown that $\mathbf{t}^t$ is related the correlations $u'u'$ of turbulent velocity fluctuation components u' by

$$\mathbf{t}^t = -\rho u'u' \quad (5.4)$$

For the two phase gas solid flow systems the conservation of momentum equation must include interphase interaction term, the equation becomes

$$\rho \frac{D\mathbf{u}}{Dt} = -\text{grad } p + \text{div } \mathbf{t} + \rho \mathbf{g} - \mathbf{F} \quad (5.5)$$

The equations are similar to those of Jackson's (Davidson and Jackson, 1961) and can be considered as a generalization of the Navier-Stokes' equations for two phase flow. $\mathbf{F}$ is local average force exerted by the fluid on the particles, and is dependent on the void fraction and the relative velocity of the two phases. This will be assumed to act in the direction of the relative velocity.

For a single spherical particle, the drag force F_D is:

$$F_D = \frac{\rho_g}{2} (\Delta u)^2 \frac{\pi D_p^2}{4} C_D \quad (5.6)$$

The drag coefficient can be related to the Reynolds number (Rowe and Henwood, 1961) by means of the relations

$$C_D = \frac{24}{Re} (1 + 0.15 Re^{0.687}) \quad Re < 1000 \quad (5.7)$$

$$C_D = 0.44 \quad Re > 1000 \quad (5.8)$$

The density of suspension (dispersion density)

$$r_{ds} = r_p(1-e) \quad (5.9)$$

= mass of particles per unit volume of dispersed phase

where e is bed voidage

$$\text{mass per particle} = \frac{1}{6} \pi D_p^3 r_p \quad (5.10)$$

$$\text{number of particles per unit volume} = \frac{r_p(1-e)}{\frac{1}{6} \pi D_p^3 r_p} \quad (5.11)$$

$$\text{Drag force per unit volume} = \frac{3}{4} r_p (\Delta u)^2 C_D (1-e)/D_p \quad (5.12)$$

Different empirical equations of drag force have been proposed for different air velocities and cluster size. In a complicated flowfield like swirling bed, single empirical correlations can not satisfy the whole bed. For the purpose of approximation only single particle drag force is used here.

Considering now the quasi-steady turbulent equation system in a cylindrical polar coordinate system (z,r,q), assuming axisymmetry ($\partial/\partial q = 0$) and no body force ($rg \neq 0$) the basic equations become

$$r \left(u_z \frac{\partial u_z}{\partial z} + u_r \frac{\partial u_z}{\partial r} \right) = \frac{\partial}{\partial z} (t_{zz}) + \frac{1}{r} \frac{\partial}{\partial r} (r t_{rz}) - \frac{\partial p}{\partial z} - F_z \quad (5.13)$$

$$r \left(u_z \frac{\partial u_r}{\partial z} + u_r \frac{\partial u_r}{\partial r} - \frac{u_q^2}{r} \right) = \frac{\partial}{\partial z} (t_{rz}) + \frac{1}{r} \frac{\partial}{\partial r} (r t_{rr}) - \frac{t_{qq}}{r} - \frac{\partial p}{\partial r} - F_r \quad (5.14)$$

$$r \left(u_z \frac{\partial u_q}{\partial z} + u_r \frac{\partial u_q}{\partial r} - \frac{u_q u_r}{r} \right) = \frac{\partial}{\partial z} (t_{qz}) + \frac{1}{r^2} \frac{\partial}{\partial r} (r^2 t_{rq}) - F_q \quad (5.15)$$

The continuity equation is

$$\frac{\partial}{\partial z} (r u_z) + \frac{1}{r} \frac{\partial}{\partial r} (r r u_r) = 0 \quad (5.16)$$

The measured value of axial velocity u_z is about 10 m/s and u_q is about 3 m/s which means u_z is the same order of u_q . The radial velocity u_r is about 10^{-3} m/s and can be neglected. The order of magnitude of axial directional gradient values of velocities, pressures and shear stresses are much smaller than those in the radial direction, i.e. $\frac{\partial}{\partial z} \ll \frac{\partial}{\partial r}$.

Equation (5.13) to (5.15) can be simplified by these assumptions. They are shown on the following:

(1) For z equation:

Equation (5.13) can be modified to

$$r \left(u_z \frac{\partial u_z}{\partial z} + u_r \frac{\partial u_z}{\partial r} \right) = \frac{1}{r} \frac{\partial}{\partial r} (r t_{rz}) - \frac{\partial p}{\partial z} - F_z \quad (5.17)$$

From measured velocity profiles we find that $r \left(u_z \frac{\partial u_z}{\partial z} + u_r \frac{\partial u_z}{\partial r} \right)$ is about $10 \text{ Kg/m}^2\text{s}^2$ and $\frac{\partial p}{\partial z} \int 400 \text{ Kg/m}^2\text{s}^2$. Assuming the viscous term is the same order of magnitude as the inertial term, then both the viscous and inertial term can be neglected compared to the pressure term, and then $-\frac{\partial p}{\partial z} \approx F_z$. This proves the assumption that axial directional pressure drop is mostly due to drag force between gas and solid particles.

Because lack of particle velocity and particle concentration distribution, it's not possible to calculate precisely z directional drag force distribution F_z . But qualitatively, the nonuniform pressure drop distribution means that the drag force is not uniformly distributed, leading to a nonuniform particle velocity and concentration.

(2) For q equation:

Equation (5.15) can be simplified to

$$r \left(u_z \frac{\partial u_q}{\partial z} + u_r \frac{\partial u_q}{\partial r} - \frac{u_q u_r}{r} \right) = \frac{1}{r^2} \frac{\partial}{\partial r} (r^2 \tau_{rq}) - F_q \quad (5.18)$$

The left hand side of equation (5.18) $\int 20 \text{ Kg/m}^2\text{s}^2$

for $\Delta u_q = 2 \text{ m/s}$ and take $C_D = 0.44$

$$F_q = \frac{3}{4} r_p (\Delta u_q)^2 C_D (1-e)/D_p = 65 \text{ Kg/m}^2\text{s}^2 \quad (5.19)$$

All the terms in the q equation are the same order of magnitude and then the shear stress τ_{rq} is significant in the analysis.

(3) r equation:

Equation (5.14) becomes

$$\frac{r u_q^2}{r} = \frac{\partial p}{\partial r} + F_r \quad (5.20)$$

The measured r directional velocity is as small as 10^{-3} m/s and the predicted r directional drag force is small compared to pressure term. So we conclude that

$$\frac{r u_q^2}{r} = \frac{\partial p}{\partial r} \quad (5.21)$$

This means the centrifugal force of the swirling air is balanced by the radial pressure drop and is not affected by the presence of solid particles as shown on Figure 5.1. The decay of swirl velocity along the height of the bed as shown on Figure 3.11 can thus be explained as decay of radial pressure gradient along the axial direction of the pipe as shown on Figure 5.2.

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Figure 5.1 Comparison of calculated values of centrifugal force and radial pressure gradient for air and particles flow at 15 cm above the swirl air inlet port for swirl number 0.134 and mas flux of 16 $\text{Kg/m}^2\text{s}$

Figure 5.2 Relative static pressure profiles for air and particles flow at different axial locations for swirl number 0.134 and mass flux $G_s = 16 \text{ kg/m}^2\text{s}$ pressure is expressed as the difference with tube center

EROSION STUDY

6.1. Experimental apparatus and procedure:

Rates of material loss of fluidized bed surface were determined by accurately weighing a small sample specimen after exposure to fluidized particles under room temperature conditions in the 102 mm, 193 mm and 298 mm internal diameter primary bed columns. The procedures for measuring weight loss were the same for all three bed diameter studies.

Wear rates were calculated from the weight loss of 1 square cm area specimen (diameter = 1.13 cm) which can be screwed into a swagelok valve. Initially, we had planned to machine the weight loss sample to be 1 cm in diameter; however, when the system was being constructed, it was discovered that if the sample size is kept at 1 cm diameter, the sample would not be flush when placed in a standard swagelock fitting. Hence, the sample size was modified so that the sample area exposed in the bed would be 1 square cm in area. The valve acts as a holder and can be adjusted to make the edge of the specimen flush with the combustor wall. Figure 6.1 illustrate the basic test section geometry and the test specimen holder. Figure 6.2 shows the sample location in the fluidized bed systems. The weight loss were used to determine the erosion rate, which is defined as the weight loss of material per unit area of wall surface exposure to the particles per hour.

The specimen was carefully weighed before and after exposure in the bed since the weight loss was very small. It was critical to have a very accurate and reliable measurement technique. The balance used was a

Figure 6.1 Schematic of the erosion test section and specimen

Figure 6.2 Schematic showing the different locations of the erosion specimen sample (circled numbers) along the height of the primary bed

digital balance Mettler H20T with an accuracy of 0.05 mg. Two special aluminum specimens were kept separately and used as standard weights for comparison. After being removed from the holder, the specimen was wiped clean using tissue paper and cleaned in acetone. The experiments were designed to last long enough to produce a weight loss of at least two significant figures to ensure accuracy. Each sample was measured twice, and, if the difference of the two measurements did not agree to within 0.05mg, the measurement was repeated until agreement within this tolerance range was attained.

The time period for which the weight loss specimen was kept inside the primary bed varied for studies with the 102 mm internal diameter riser. Initially, when the experimental plan was developed, a 6 hour specimen exposure was designed, since preliminary indications were that sufficient weight loss would have occurred during this 6 hour time period. However, when the experiments were conducted, it was found that the duplicate data was not consistent. This was mainly due to difficulties in maintaining the bed at a fixed operating condition, i.e., solids mass flux rate and air velocity. The primary difficulty was in controlling the pressure and flowrate of the air entering the bed from the blower. Small variations in air flowrate had a major impact on the bed operating condition. Hence, as long as measurable weight loss was attained, the test specimens were removed as soon as the bed operating condition changed due to blower speed variability.

For studies with 193 mm and 298 mm internal diameter beds, variability in blower speed did not have a major impact on the bed's operating condition. This can be explained by the fact that as the bed

diameter increases, the bed capacity has a moderating influence on hydrodynamic variabilities due to small variations in either air flowrate or inlet air pressure. Hence for the larger diameter riser columns, the test specimen exposure time was kept constant at 6 hours, as originally planned.

6.2. Studies on Erosion in CFB

For wear of materials caused by impact of solid particles, the rate of material loss is closely related to the particle impact velocity, particle mass flow rate(mass of particles striking unit surface area per unit time) and particle properties (Tilly,1979). In a fluidized bed, these parameters are determined by the operating conditions of the bed and the gas and particle properties. In our system with steel particles the air superficial velocity can be varied from 20 to 48 ft/sec to maintain the fast fluidized bed condition.

In this research, consistent with usual practice in the fluidization literature, we use erosion rate to describe the total material loss, even though abrasion in addition to erosion may contribute to the overall wear.

The experimental data are shown in Tables 6.1 through 6.5 for the 102 mm internal diameter riser column, Tables 6.6 through 6.10 for the 193 mm diameter riser column, and Tables 6.11 through 6.14 for the 298 mm internal diameter riser column.

Table 6.1 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 0.388 g/cm²sec
(102 mm ID riser)

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.42070	1.42013	5.7	7.26	0.0785
2	1.41836	1.41810	2.6	4.618	0.0563
3	1.43132	1.43105	2.7	4.615	0.0585
4	1.42865	1.42845	2.0	4.608	0.0434
5	1.42075	1.42048	1.7	10.465	0.0258
6	1.41847	1.41824	2.3	4.713	0.0488

Compared Sample: Initial Weight 1.48020
Final Weight 1.48025
Weight Loss -0.00005

Table 6.2 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
25 ft/sec. and solid mass flowrate 0.388 g/cm²sec
(102 mm riser)

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.42785	1.42750	3.5	5.224	0.0670
2	1.41673	1.41646	2.7	4.720	0.0572
3	1.41762	1.41737	2.5	4.710	0.0531
4	Data deleted due to round-off errors in calculation				
5					
6					

Compared Sample: Initial Weight 1.42264
Final Weight 1.42271
Weight Loss -0.00007

**Table 6.3 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 1.355 g/cm²sec**

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm)x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.42514	1.42422	9.2	7.857	0.1171
2	1.41872	1.41789	8.3	7.860	0.1056
3	1.41337	1.41286	5.1	7.878	0.0649
4	Data deleted due to round-off errors in calculation				
5					
6					

Compared Sample: Initial Weight 1.42116
Final Weight 1.42120
Weight Loss -0.00004

**Table 6.4 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 2.1 g/cm²sec**

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm)x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.42025	1.41900	12.5	7.850	0.1592
2	1.42475	1.42388	8.7	8.080	0.1077
3	1.43712	1.43646	6.6	7.830	0.0843
4	Data deleted due to round-off errors in calculation				
5					
6					

Compared Sample: Initial Weight 1.42116
Final Weight 1.42118
Weight Loss -0.00002

Table 6.5 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 0.388 g/cm²sec
with swirl flow S1 = 20 cfm

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x 10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.42118	1.42085	3.3	4.376	0.0754
2	1.41863	1.41821	4.2	5.079	0.0827
3	1.43121	1.43093	2.8	4.714	0.0594
4	1.42852	1.42837	1.5	4.717	0.0318
5	1.42095	1.42068	2.7	4.631	0.0583
6	1.41844	1.41824	2.0	4.684	0.0427

Compared Sample: Initial Weight 1.42963
 Final Weight 1.42965
 Weight Loss -0.00002

Table 6.6 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 0.388 g/cm²sec
(193 mm ID riser)

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x 10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.41081	1.41034	4.7	6	0.0782
2	1.42640	1.42607	3.3	6	0.0554
3	1.47171	1.47136	3.5	6	0.0581
4	1.48002	1.47976	2.6	6	0.0437
5	1.43126	1.43111	1.5	6	0.0257
6	1.41614	1.41599	1.5	6	0.0254

Compared Sample: Initial Weight 1.47231
Final Weight 1.47232
Weight Loss 0.00001

Table 6.7 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
25 ft/sec. and solid mass flowrate 0.388 g/cm²sec
(193 mm riser)

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x 10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.42107	1.42066	4.1	6	0.0680
2	1.41162	1.41128	3.4	6	0.0573
3	1.41638	1.41606	3.2	6	0.0530
4	1.47114	1.47086	2.8	6	0.0470
5	1.48642	1.48626	1.6	6	0.0267
6	1.43741	1.43728	1.3	6	0.0216

Compared Sample: Initial Weight 1.46754
Final Weight 1.46752
Weight Loss -0.00002

**Table 6.8 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 1.355 g/cm²sec
(193 mm riser)**

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.47116	1.47046	7.0	6	0.1172
2	1.42182	1.42118	6.4	6	0.1060
3	1.43789	1.43750	3.9	6	0.0650
4	1.44614	1.44582	3.2	6	0.0529
5	1.43718	1.43637	3.1	6	0.0510
6	Outlier data neglected				

Compared Sample: Initial Weight 1.45342
Final Weight 1.45344
Weight Loss 0.00002

**Table 6.9 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 2.1 g/cm²sec
(193 mm riser)**

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.45128	1.45033	9.5	6	0.1591
2	1.47620	1.47555	6.5	6	0.1079
3	1.46221	1.46171	5.0	6	0.0840
4	1.48111	1.48065	4.6	6	0.0764
5	1.46702	1.46661	4.1	6	0.0680
6	1.41608	1.41565	4.3	6	0.0719

Compared Sample: Initial Weight 1.45643
Final Weight 1.45644
Weight Loss 0.00001

Table 6.10 Weight changes due to erosion at different locations 95
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 0.388 g/cm²sec
with swirl flow S1 = 20 cfm
(193 mm riser)

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.46112	1.46066	4.6	6	0.0761
2	1.48781	1.48732	4.9	6	0.0820
3	Outlier data neglected				
4	1.41121	1.41103	1.8	6	0.0308
5	1.47612	1.47577	3.5	6	0.0589
6	1.46741	1.46715	2.6	6	0.0429

Compared Sample: Initial Weight 1.41265
Final Weight 1.41266
Weight Loss 0.00001

Table 6.11 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 0.388 g/cm²sec
(298 mm ID riser)

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.42960	1.42913	4.7	6	0.0787
2	1.47511	1.47477	3.4	6	0.0569
3	1.48348	1.48313	3.5	6	0.0585
4	1.47341	1.47315	2.6	6	0.0431
5	1.41161	1.41146	1.5	6	0.0256
6	1.42622	1.42593	2.9	6	0.0489

Compared Sample: Initial Weight 1.45641
Final Weight 1.45639
Weight Loss -0.00002

**Table 6.12 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
25 ft/sec. and solid mass flowrate 0.388 g/cm²sec
(298 mm riser)**

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Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm)x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.41181	1.41143	3.8	6	0.0640
2	1.46782	1.46745	3.4	6	0.0560
3	1.45711	1.45679	3.2	6	0.0539
4	1.47821	1.47792	2.9	6	0.0481
5	1.46118	1.46101	1.7	6	0.0280
6	1.42620	1.42607	1.3	6	0.0220

Compared Sample: Initial Weight 1.41150
Final Weight 1.41149
Weight Loss -0.00001

**Table 6.13 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 1.355 g/cm²sec
(298 mm riser)**

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm)x10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.41172	1.41101	7.1	6	0.1180
2	1.46532	1.46468	6.4	6	0.1070
3	1.45731	1.45692	3.9	6	0.0651
4	1.44641	1.44608	3.3	6	0.0542
5	1.45732	1.45704	2.8	6	0.0474
6	1.41628	1.41599	2.9	6	0.0488

Compared Sample: Initial Weight 1.47892
Final Weight 1.47893
Weight Loss 0.00001

Table 6.14 Weight changes due to erosion at different locations
along the bed height for superficial air velocity
48 ft/sec. and solid mass flowrate 0.388 g/cm²sec
with swirl flow S1 = 20 cfm
(298 mm riser)

Sample Location	Initial Weight (gm)	Final Weight (gm)	Weight Loss (gm) x 10 ⁴	Time (hrs)	Erosion Rate (mg/cm ² hr)
1	1.46751	1.46706	4.5	6	0.0757
2	1.47984	1.47935	4.9	6	0.0820
3	1.46711	1.46676	3.5	6	0.0591
4	1.41184	1.41165	1.9	6	0.0320
5	1.46582	1.46547	3.5	6	0.0589
6	1.47981	1.47955	2.6	6	0.0429

Compared Sample: Initial Weight 1.46598
Final Weight 1.46598
Weight Loss 0.00000

Discussion of results for 102 mm internal diameter riser column:

Erosion data for all six locations were obtained at the following operating conditions:

Superficial air velocity 48 ft/sec

- Solid mass flowrate 0.388 g/cm²sec
- Solid mass flowrate 1.355 g/cm²sec
- Solid mass flowrate 2.1 g/cm²sec
- Solid mass flowrate 0.388 g/cm²sec and swirl air flow of 10, 12, 15, and 20 SCFM

Superficial air velocity 25 ft/sec

- Solid mass flowrate 0.388 g/cm²sec

These operating conditions were selected because they provided data on the effect of superficial air velocity, solid mass flowrate and secondary swirl air flowrate.

Data in Table 6.2, 6.3 and 6.4 for locations 4,5 and 6 were deleted since the initial and final weights matched exactly, and this was believed to be due to round-off error in the weight measurements. It can be seen that the erosion rate decreased from location 1 to location 6.

It should also be noted that the time duration of each sample varied with each sample. As stated earlier, this was due to operational instability

in the bed, caused by small variations in the blower's speed. This

variability caused the bed operating condition to vary and hence to obtain erosion data at a given operating condition, the samples were kept long enough to obtain a measureable weight gain and until the bed's solid mass flowrate could be maintained at a specific value.

Discussion of erosion data for 193 mm and 298 mm internal diameter beds

Data at the same operating condition as for the 102 mm internal diameter bed were taken for the 193 mm bed. The procedure of obtaining this data was the same as for the 102 mm diameter bed. It should be noted that the time duration for each sample was kept constant at 6 hours. This was possible mainly because the 193 mm and 298 mm diameter beds were fairly stable even when variabilities in blower speed occurred, mainly due to the larger capacitance in the beds.

It should also be noted that the time was recorded with an accuracy of ± 10 minutes. This was due to the fact that the time duration of the sample did not have a major impact on the final results, as was the case with weight loss. Since the weight loss was being measured very accurately, the erosion rate could be determined to the desired accuracy. It should be noted that we were investigating significant changes in erosion rates rather than their precise values.

The experimental results show the following:

(1) From Figures 6.3 (102 mm internal diameter), 6.4 (193 mm ID) and 6.5 (298 mm ID), wear rate was found to be weakly influenced by increasing superficial air velocity. This is consistent with the literature:

Figure 6.3 Erosion at different locations along the bed height for different air velocities for 102 mm ID riser

Figure 6.4 Erosion at different locations along the bed height for superficial air velocities of 25 and 48 ft/s and solid mass flowrate = 0.388 g/cm²s for 193 mm ID riser

Figure 6.5 Erosion at different locations along the bed height for superficial air velocities of 25 and 48 ft/s and solid mass flowrate = 0.388 g/cm²s for 298 mm ID riser

Parkinson (1985) found that at higher velocities (turbulent regimes), there was a decrease in wear. Woodford (1983) concluded that erosion is almost independent of gas superficial velocity in a high superficial velocity range where within the turbulent regime the size of voids is insensitive to gas velocity.

(2) Most metal loss occurred near the bottom of the column i.e., location 1 which was located near the solid injection port. This is mostly due to the effect of impingement angle of attack. Metals generally exhibit ductile type of erosion and the maximum metal removal occurs at angles of particle impact near 40° .

(3) Increasing solid mass flow rate caused a higher solid concentration in the primary bed. Figures 6.6 (102 mm ID), 6.7 (193 mm ID) and 6.8 (298 mm ID) show that the wear rate increased with increasing solid mass flow rate. Figure 6.9 shows that swirl injection did not appear to increase the extent of erosion appreciably. Note that the bars indicate the extent of experimentation error due to impreciseness in sample weighing. Further long term experiments are necessary before a final conclusion can be made regarding the impact of swirl flow on erosion.

(4) In the circulating fluidized bed system the solid particles move either straight upward or downward near the bed wall. The amount of erosion is dependant on the particle concentration and velocity moving parallel to the wall inside the wall sublayer zone. In an 8 inch diameter bed, with the same air velocity and particle velocity, the amount of erosion is theoretically the same as in the 4 inch diameter bed. From the above analysis, it was concluded that 8 inch bed study was not necessary.

Figure 6.6 Erosion at different axial locations for various solid mass flow rates for 102 mm ID riser

Figure 6.7 Erosion at different axial locations for various solid mass flow rates for 193 mm ID riser

Figure 6.8 Erosion at different axial locations for various solid mass flow rates for 298 mm ID riser

Figure 6.9 Comparison of erosion rates with and without swirl flow for 102 mm ID riser

6.3 Studies on Erosion in SCFB

The method of weight loss of small sample specimen on the fluidized bed wall were used again. The location of the sample was 6 inch above the ball and socket of the nozzle. When the primary air flowrate and solid recirculation rate are fixed, the erosion rates were measured by varying the secondary air flow rate and angle θ and distance d . The data are listed in Tables 6.15, 6.16 and 6.17 for the 102 mm internal diameter column.

Figure 6.10 shows a plot of the erosion rate versus the swirl air flowrate in the 102 mm internal diameter riser column . It can be seen that for the same primary air flowrate the smallest erosion rate happens when only primary air is flowing in the bed i.e., no swirl. Furthermore, as shown in Table 6.18, as the bed diameter increases from 102 mm to 193 mm and 298 mm, the effect of swirl air flowrate on erosion is the same as for the smaller diameter column.

The effect of the angle of the secondary air nozzle with the horizontal is shown in Figure 6.11. The result show that when theta is greater than 25 degrees there is a dramatic decrease in erosion rate.

The impact of secondary air nozzle angle for the larger diameter beds is summarized in Table 6.19. The erosion rate increases with increase in swirl number as the nozzle angle is decreased from 90° to 0° .

The plot of erosion rate versus the secondary nozzle radial location dimensionless parameter d/D is shown in Fig 6.12. The coresponding data

Table 6.15 The erosion rate and relative erosion increase for various swirl air flowrates for primary air velocity of 48 ft/s and solid mass flowrate 1.355 g/cm²s with $O = 0$ and $d/D = 0.8$. The relative increase is computed based on a nonswirl value of 0.123 mg/cm²hr.

Swirl air flowrate Q_2 (cfm)	Swirl number S	Erosion rate (mg/cm ² hr)	Relative erosion increase
20.0	0.653	0.183	0.49
18.0	0.529	0.179	0.46
16.0	0.418	0.178	0.44
14.0	0.320	0.166	0.35
12.0	0.235	0.164	0.33
10.0	0.163	0.153	0.26

Table 6.16 The erosion rate and relative erosion increase of different angle with horizontal for primary air velocity 48 ft/sec. and solid mass flowrate 1.355 g/cm²sec. with $Q_2 = 20$ cfm and $d/D = 0.8$. The relative increase is computed based on a nonswirl value of 0.123 mg/cm²hr.

Angle θ (degree)	Swirl number S	Erosion rate (mg/cm ² hr)	Relative erosion increase
0.0	0.653	0.172	0.47
15.00	0.631	0.171	0.49
25.0	0.592	0.170	0.42
35.0	0.535	0.138	0.31
45.0	0.462	0.132	0.26
90.0	0.0	0.123	0.0

Table 6.17 The erosion rate and relative erosion increase of different d/D for primary air velocity 48 ft/sec. and solid mass flowrate 1.355 g/cm²sec. with $Q_2 = 20$ cfm and $d/D = 0.8$

The relative increase is computed based on a non-swirl value of 0.123 mg/cm²hr.

d/D	Swirl number S	Erosion rate (mg/cm ² hr)	Relative erosion increase
0.8	0.653	0.183	0.49
0.7	0.571	0.182	0.48
0.6	0.489	0.172	0.40
0.5	0.408	0.177	0.44
0.4	0.326	0.149	0.21
0.3	0.245	0.145	0.18
0.2	0.163	0.135	0.10
0.1	0.082	0.132	0.08
0.0	0.0	0.123	0.0

Table 6.18 The effect of swirl air flowrate on the relative increase in erosion rates. The relative increase is defined with respect to the erosion that occurs when the swirl flow rate is zero.

Swirl air flowrate (SCFM)	Relative Erosion Increase	
	193 mm ID	298 mm ID
20.0	0.48	0.49
18.0	0.45	0.47
16.0	0.45	0.46
14.0	0.33	0.35
12.0	0.32	0.33
10.0	0.24	0.27

Table 6.19 The effect of swirl number on the relative increase in erosion rates. The swirl number changes either due to changes in the angle of secondary air injection or varying the position of the nozzle (d/D)

Swirl Number	Relative Erosion Increase	
	193 mm ID	298 mm ID
0.163	0.25	0.27
0.320	0.34	0.37
0.592	0.45	0.41
0.653	0.48	0.49

Figure 6.10 A plot of erosion rate versus the swirl air flow rate for the primary air velocity 48 ft/sec and solid mass flow rate 0.388 g/cm²sec. with $\theta = 0$ degree and $d/D = 0.8$ in 102 mm ID riser

Figure 6.11 A plot of erosion rate versus theta, the angle subtended with the horizontal by the secondary air in a 102 mm riser

Figure 6.12 A plot of erosion rate versus the secondary air location parameter, d/D for 102 mm ID riser

is shown in Table 6.17. Results shows that increasing the radial distance increase the erosion rates. For larger diameter beds, the relative erosion increase is shown in Table 6.19.

In order to reduce the number of variables in the system, we can combine the secondary air nozzle variable into one parameter: swirl number.

Also, we can define relative erosion increase as follows:

$$RE = \frac{E - E_0}{E_0}$$

where E : erosion rate with swirl number S

E_0 : erosion rate without swirl flow of the same primary
air rate and solid recycle rate

Figure 6.13 includes all the data for the 102 mm internal diameter riser column, in terms of these two variables, i.e. relative erosion increase versus the swirl number. From Figure 6.13 the following empirical equation can be obtained :

$$RE = 0.58 S^{0.4}$$

when $\theta < 25$ and $d/D > 0.5$

The data for 193 mm and 298 mm columns was not plotted in Figure 6.13 to prevent cluttering the diagram. Clearly, examination of the data shows that all three diameter beds can be represented by the same empirical equation, presented above.

Figure 6.13 A plot of relative erosion increase versus the swirl number for 102 mm ID riser

It should be emphasized that these findings are only valid for cold models. Even though, the experimental plan was devised using dimensionless numbers, so that the results can be translated to hot combustors, erosion can be significantly different, due to the presence of reducing conditions during actual combustion. The main motivation for conducting cold model erosion studies is that it offers a better understanding of solids movement near the column wall, where measurements of solids flux are difficult to make accurately.

CONCLUSIONS

The minor conclusions from this study are as follows:

- (1) The phenomena arising from secondary air introduction in a CFB can be explained in terms of the radially non-uniform structure of the circulating bed. The swirl air has a substantial cyclonic effect, tending to intensify the core-annulus phenomena of fast fluidization. One might anticipate decreased decay lengths due to increased rates of solids transfer to the wall area.
- (2) When swirl flow is injected into the fluidized bed with particle flow, the tangential velocity will be weakened by viscous dissipation. The forced vortex zone decreased faster than the free vortex zone along the bed height. This velocity decay can be confirmed as radial pressure gradient decay along the bed.
- (3) From the order of magnitude analysis of equation of motion of air for the gas-solid suspension flow, the only significant turbulent shear stress term is τ_{rq} . For the purpose of evaluation u_r' , u_q' a more accurate experimental technique has to be developed for the measurement of solid velocity determination.
- (4) The five hole Pitot probe technique is a useful cost-effective tool to investigate turbulent swirling flow in a circulating fluidized bed. It has, however, some inherent problems:
 1. Weak sensitivity to small velocities, which might permit large relative errors.
 2. The measurement of turbulent stresses is not possible.

The results may be compared with laser doppler anemometers measurements in corresponding flow situations.

- (5) Residence time distributions are necessary to understand the mixing pattern of gas and solid particles in the bed. In a circulating bed, the residence time is expected to be as short as several seconds. Special technique like radioactive particles tracer has to be used to measure RTD's. These measurement have not been performed in this research. It will be considered as an important part of future plan for extension of this topic.

The major conclusions from this study were as follows:

1. The scaling relationships developed in this study are essential to relate the cold flow studies to the actual hot combustor. In this regards, it was found that the selection of the solid particles was very important and this led us to the conclusion that the results obtained using solids, such as FCC catalyst, Coarse alumina, or fine alumina, which have a low density, would not yield insights useful for hot combustors.
2. The extent of erosion in a swirling circulating fluidized bed was higher than in an ordinary circulating fluidized bed. This was mainly due to the fact that the solid particle velocities in the SCFB system are higher than in an ordinary circulating bed and that the erosion was mainly caused by the downflow of particles at the wall rather than by impaction of the particles. There was an increase of erosion with swirl number and an empirical correlation was developed between the relative increase in erosion rate and swirl number. It was also found that this correlation could be used for all three diameter beds tested in this study.

3. Since the extent of erosion in the SCFB system was mainly due to the downflow of particles at the wall, it was independent of the bed diameter. Hence, the extent of erosion in a 193 mm diameter bed would be the same under the same hydrodynamic conditions as in a 102 mm internal diameter bed.
4. The bed hydrodynamics (velocity profile, local solid fluxes, axial pressure profile) could be predicted from a detailed simulation model, which was implemented as a computer program running on a CRAY computer. An extensive study was initiated to develop the 3D computer model and verify it using a hot swirl combustor operated at the Wright Patterson Air Force base in collaboration with Rolls-Royce, Inc. Velocity measurements in the hot combustor were made using a two component LDV (Laser Doppler Velocimetry) system. It was found that close agreement between the model and the experimental data could be obtained using the proper boundary conditions. Further research in this area is currently in progress.
5. The dimensionless Swirl Number quantifies the impact of swirl in the circulating flow. It was found that measurement of the secondary air penetration was not necessary, since the swirl number incorporated quantitatively the extent of momentum transfer from the secondary to the primary air flowrate.
6. The major impact of swirling flow is to increase the residence time of the particles without increasing the combustor height, which is already large for a circulating fluidized bed. This is evident from the increased radial solid mass fluxes with a finite

swirl number and increases in the net solids flux as the swirl number increases (refer Figure 3.9). The impact of this increased solid residence time on combustor efficiency and NO_x generation needs to be further investigated.

NOTATIONS

C	=	Drag coefficient of particles
D	=	diameter of fluidized bed
d_p	=	diameter of particle
F_D	=	drag force
g	=	acceleration of gravity
G_s	=	solid mass flux
G_q	=	axial flux of swirl momentum
G_x	=	axial flux of axial momentum
L	=	bed height
ΔP	=	pressure drop measured across a section of the bed
r	=	radial distance from the center of the bed
S	=	swirl number
u_z	=	axial air velocity
V	=	magnitude of total velocity
u_q	=	swirl air velocity
u_r	=	radial velocity
b	=	yaw angle
d	=	pitch angle
e	=	voidage of bed
m	=	viscosity of air
r_{ds}	=	solid dispersion density
r_p	=	density of particle
r_g	=	density of air
f_s	=	particle sphericity

REFERENCES

- Arena, U., Cammarota, A., and Pistane, L. (1985). High velocity fluidization behaviour of solids in a laboratory scale circulating bed, in "Circulating fluidized bed technology," 119-126, P. Basu ed., Pergammon Press, Toronto.
- Avidan, A.A. (1980), Bed expansion and Solid Mixing in High Velocity Fluidized Bed, Ph.D. dissertation., City University of New York, New York.
- Bellman, R. and A. Levy (1981) Erosion Mechanism in Ductile Metals, Wear, 70, 1.
- Bierl, T.W., Gajdos, L.J., McIver, A.E., and McGovern, J.J. (1980). Studies in support of recirculating bed reactors for the processing of coal, DOE report Ex-C-76-01-2449, July.
- Brereton, C., and Stromberg, L. (1985). Some aspects of the fluid dynamic behaviour of fast fluidized beds, in "Circulating fluidized bed technology," P. Basu ed., 133-144, Pergammon Press, Toronto.
- Burkell, J. (1985). Solids circulation rate measurements in circulating fluidized beds. M.A.Ac. Thesis, University of British Columbia, Vancouver.
- Capes, C.E. (1974). Particle agglomeration and the value of the exponent n in the Richardson-Zaki equation, Powder Technol., Vol. 10, 303-306.
- Crowther, M.E., and Whitehead, J.C. (1978). In "Fluidization," J.F. Davidson and D.L. Keairns eds., 65-70, Cambridge University Press, London.
- Doig, I.D., Roper, G.H., Ind. Eng. Chem. Fundamentals 6, 247, 1967.

- Engstrom, F., Osakeyhtio, A.A., and Sahagian, J. (1985) Operating experience with circulating fluidized bed boilers, in "Circulating fluidized bed technology," P. Basu ed., Pergammon Press, Toronto.
- Finnie, I. (1958) The Mechanism of Erosion of Ductile Metals, *Proceedings of Applied Science (ASME)*, 527.
- Finnie, I. (1960a) An experimental study of Erosion, *Proceedings of the Society for Experimental Stress Analysis*, Vol. XVII No.2, 65.
- Finnie, I. (1960b) Erosion of surfaces by solid particles, *Wear*, 3, 87.
- Finnie, I. and D.H. McFadden (1978) On the Velocity Dependence of the Erosion of Ductile Metals by solid Particles at Low Angles of Incidence, *Wear*, 48, 181.
- Geldart, D. (Ed.), *Gas Fluidization Technology*, John Wiley & Sons, 1986.
- Geldart, D. (1973). Types of gas-fluidization, *Powder technology*, Vol. 7, 285-292.
- Glicksman, L.R., "Scaling Relationships for Fluidized beds," *Chemical Engineering Science*, Vol.39, No.9, pp.1373-1379, 1984.
- Goodwin, J.E., W. Sage, G.P. Tilly (1969) Study of Erosion by Solid Particles, *Proc. Instn. Mech. Eng.*, Vol. 184, Pt. 1, No. 15, 281-188.
- Govind, R. et.al., Research Project Plan - Demonstration of an Advanced Circulating Fluidized Bed Coal Combustor, Submitted to the Ohio Coal Development Office, Columbus, Ohio, October 1988.
- Grace, J.R. (1986). contacting modes and behaviour classification of gas-solid and other two-phase suspensions, *Can. J. Ch.E.* Vol. 64, 353-363.
- Gupta, Lilley, Syred, "Swirl flows", Abacus Press, 1984.

- Hariu, O.H., and M.C. Molstad, Pressure drop in vertical transport of solids by gases, Ind.Eng.Chem., 1949, Vol.41(6), 1148-60.**
- Hiett, G.F. and G.E. Powell, Three -Dimensional Probe for Investigation of Flow Patterns, The Engineer, Jan. 1962, p.p. 165-170.**
- Hinze, J.O. (1959) . Turbulence, McGraw Hill, New York.**
- Horio, M., Taki, A., Hsieh, Y.S., and Muchi, I. (1980). Elutriation and particle transport through the freeboard of a gas-solid fluidized bed, in "Fluidization," J.R. Grace and J.M. Matsen eds., Plenum Press, New York.**
- Hotley, H. (1939) Uber den Abnutzungsvorgang in Blasversatzrohr und die Frage der Bekampfung des Verschleisses, Geologie und Mjnbouw, I, 209.**
- Jackson, R., (1985). Hydrodynamic stability of fluid-particle systems, in "Fluidization", 2nd Ed, J.F. Davidson eds., Academic Press.**
- Konno, H., and Saito, S.J. (1969), Chem. Eng. Japan, 2, 211-17.**
- Kunii, D., and Levenspiel, O. (1969). Fluidization Engineering, John Wiley and Sons Inc., New York.**
- Leung, L.S. (1980). The ups and downs of gas-solid flow- a review, in "Fluidization," J. R. Grace and J.M. Matsen, eds., Plenum Press, New York.**
- Li, Y., and Kwauk, M. (1980). The dynamics of fast fluidization, in "Fluidization," J.R. Grace and J.M. Matsen, eds., 537-544, Plenum Press, New York.**
- Monceaux, L., M. Azzi, Y. Molodtsov, J.F. Large, Overall and local characterization of flow regimes in a circulating fluidized bed, in Circulating Fluidized Bed Technology, Prabir Basu (Ed.), pp 185-191, 1985, Pergamon Press, New York.**

- Nakamura, K. and Capes, C.E. (1973). Vertical pneumatic conveying: a theoretical study of uniform and annular particle flow models, Can. J. Ch. E. Vol. 51, 39-46.**
- Pratt, K.C., and B.J. Byrne, Measurements of radial velocity and temperature distribution in the flow of dense suspensions, Chemistry and Industry, 1978, 343-46.**
- Preece, C.N. and N.H. Macmillan (1977) Erosion, Ann. Rev. Mater. Sci., 7, 95.**
- Reddy, K.V.S., and D.C.T. Pei, particle dynamics in solids-gas flow in vertical pipe, Ind. Eng. Chem. Fund., 1969, Vol 8, n 3, 490-97.**
- Rhodes, M.J., and Geldart, D. (1986). A model for the recirculating fluidized bed, presented at annual AIChE meeting, Miami, December.**
- Rose, H.E., and Barnacle, H.E. (1957). Flow of suspensions of non-cohesive spherical particles in pipes, Engineer, June 14, 898-901, June 21, 939-941.**
- Schlichting, H. (1979). Boundary Layer Theory, McGraw-Hill, New York.**
- Sheldon, G.L. (1977) Effect of Surface Hardness and Other Material Properties on Erosive Wear of Metals by Solid Particles, Journal Of Engineering materials and Technology, 133.**
- Shewmon, P. and G. Sundarajan (1983) The Erosion of Metals, Ann. Rev. Mater. Sci, 13, 301.**
- Sloan, D. G. Modeling of Swirl in turbulent systems, Ph.D. Dissertation, Chemical Engineering Department, Brigham Young University, Provo, UT (1985).**
- Soo, S.L. (1982). Pneumatic conveying, Chapter 7 in "Handbook of multiphase systems," G. Hetsroni ed., McGraw Hill, New York.**

- van Breugel, J.W., Stein, J.J.M., and de Vries, R.J. (1969-70). Isokinetic sampling in a dense gas-solids stream, *Proc. Instn. Mech. Engrs.*, Vol. 184, Pt. 3C, 18-23.
- van Deemter, J.J. (1967). the counter-current flow model of a gas-solids fluidized bed, in "Proceedings of the international symposium on fluidization," A.A.H. drinkenburg ed. 334-347, Netherlands University Press, Amsterdam.
- Wahl, H. and F. Hartstein (1946) *Strahlrerscleis Franckh'sche Verlagshandlung, Arurrgart.*
- Wein, W., and Felwor, P. (1986), Circulating fluidized bed combustion, *Proceedings of VDI conference, "Fluidized bed combustion (CFB)", 155-170, 3rd-5th June, essen.*
- Weinstein, H.M. Meller, M.H. Shao, R.J. Parisi (1982), The effect of particle density on hold-up in a fast fluidized bed, *AIChE Symposium Series, Vol 80, n 234, 52-59.*
- Wellinger, K. and H.C. Brockstedt (1942) *Versuche zur Ermittlung des Verschleisswiderstandes von Werkstoffen für Blasversatzrohr sowie des Einflusses der Rohrverlegung bei Blasversatzanlagen Gluckauf, 78, 130.*
- Yerushalmi, J., D.H. Turner and A.M. Squires (1976), The fast fluidized bed, *Ind. Eng. Che., Process des. Dev., Vol. 15, n 1, 47-53.*
- Yerushalmi, J. and Cankurt, N. T. (1979). Further studies of the regimes of fluidization, *Powder Technol., Vol. 24, 187-205.*
- Yerushalmi, J., Cankurt, N.T., Geldart, D., and Liss, B. (1978). Flow regimes in vertical gas-solid contact systems, *A.I.Ch.E. symp., Ser., No. 176, Vol. 74, 1-13.*
- Yoon, H.K., Five-hole pitot probe time mean velocity measurements in confined swirling flows. M.S. Thesis. Mechanical Engineering Department. Oklahoma State University. Stillwater, OK (1982).

**APPENDIX I: REPORT ON MODEL DEVELOPMENT
AND VALIDATION**

Detailed Simulation of Entrained Flow Coal Combustion

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INTRODUCTION

Due the abundance of high sulfur coals in many states including Ohio, there is renewed interest in developing fluidized bed designs to burn such fuels efficiently for power generation as well as to meet the stringent environmental regulations. In this study, a new fluidized bed combustor, termed Swirling Circulating Fluidized Bed (SCFB) system was investigated through a detailed mathematical model. In the SCFB system, the secondary air is injected tangentially which creates a vortex pattern in the combustion zone, thereby increasing the mixing between the coal particles and the gas to achieve higher heat transfer, overall combustion efficiencies, and lower emissions of sulfur and nitrogen oxides.

Specifically, the objective of this project was to develop a general purpose simulation program for entrained flow coal combustion and then to test the program with experimental data. The basic building block which we started out with was COM3D, a general purpose simulation program for combustion (gaseous or liquid fuel combustion) developed by NASA.

It was recognized very early in this research program that initially modeling the combustion of gaseous fuels in swirling flow and its verification through experimental data would be attempted before embarking on the combustion of coal. This was because, combustion of coal is very complex and its kinetics itself is not completely understood. Furthermore, Wright Patterson Air Force Base, Dayton, Ohio, had the laser facilities to measure the velocity profiles in swirling flow gas combustors. Hence we established a joint research program with Wright Patterson Air Force Base personnel and Rolls Royce, Inc., who supplied the glass combustor, so that velocity measurements using Laser Doppler Velocimetry (LDV) techniques could be made in swirling flow gas combustors. This experimental data was used to validate the computer model.

MODEL DEVELOPMENT

The 3-D combustor performance computer program that forms the basis of the present work, is briefly described here. For complete details, refer to Report No. USARTL-TR-55C. The program is general and is capable of predicting recirculating turbulent flow in most gas-turbine type combustor chambers. Reacting or non-reacting, swirling or nonswirling, diffusion and/or premixed flames, and gaseous and/or liquid fuel combustion can be handled by the program. The program computes the following variables in the region of interest:

- o Axial, radial, and swirl velocity components
- o Pressure
- o Enthalpy (temperature); in conjunction with the equation of state, the temperature determines the density variations in the flow field
- o Turbulent kinetic energy and its dissipation rate
- o Mass fractions of total fuel (mixture fraction), unburned fuel, oxygen, carbon monoxide, carbon dioxide and water
- o Three radiation flux vectors.

The program employs the following physical models to solve the variables mentioned above:

- o Turbulence: Two-equation (k-ε) turbulence models to obtain turbulent kinetic energy and its dissipation rate.

- o Chemistry: Four step chemical reaction scheme:

- o Chemical Reaction Rate: Fuel and CO consumption rate are assumed to be governed by either the time-averaged Arrhenius model or the turbulent eddy break-up model.
- o Radiation: A six-flux model of radiation.

The transport equations for all dependant variables are written in the following general form:

$$\text{div} \left(\rho u \phi - \frac{\mu_t}{\sigma_\phi} \text{grad } \phi \right) = S_\phi$$

where ρ denotes the mixture density, u the velocity vector, μ_t the effective or turbulent viscosity, σ_ϕ the effective Prandtl/Schmidt number, S_ϕ the sources of ϕ ; i.e., S_ϕ includes the creation/destruction of ϕ plus other quantities that do not fall under the convective and diffusive terms.

An iterative, general finite-difference solution procedure suitable for 3-D elliptic flows in complex geometries is used to solve the above system of coupled, nonlinear partial-differential equations. The solution procedure involves discretizing the differential equations by integration over elementary finite-difference control volumes surrounding grid nodes that are nonuniformly spaced over the flow field.

RESULTS AND DISCUSSION

The computation was performed on an OSCA Cray Y-MP with approximately 3.6 CPU seconds per iteration for 50*49*6 fine nodes in

(z, r, θ) coordinates. Since the flow is axisymmetric, the choice of 6 grids in θ direction calculation is for checking the consistency in this direction. The sum of the continuity residuals for all the control volumes was less than 0.85% of the total flow rate and the maximum continuity residual for any control volume was less than 0.0017% of the total flow rate after 450 iterations.

Since the flow is assumed to be axisymmetric, mainly the patterns of the flow properties are shown in the r - z plane. As shown in Figure 1, the velocity vector plot in this plane shows a good flow pattern agreement with the experimental data. The stagnation point occurred at about $1.2 x/D$ which is a little longer than that measured ($0.8 x/D$). After the stagnation point, the reverse axial velocities disappear and further downstream the velocity continuously increases due to volume flow rate increase by combustion. The peak of the axial velocity profile shifts toward the outwall as the result of diminishing swirl and combustion.

For a large swirl number at 0.78 (from approximately straight annular vane calculation), the static pressure in the central core just downstream of the swirler becomes low enough to create flow recirculation. The maximum levels of turbulent kinetic energy were found in the shear layer region between annular jet and main recirculation zone and at the downstream end of the corner recirculation zone.

Within the main recirculation region, the calculated axial components are slightly small and curve near the swirler and too large near the center line region of the main reverse flow zone. Near the exit plane, the axial velocities are slightly underpredicted from centerline to wall. The swirl component is again under-predicted overall. The temperature contour plot in $^{\circ}\text{K}$ is shown in Figure 2. However, the combustion seems to introduce additional problems downstream of the main recirculation zone. The prediction shows a hot spot near the exit and a comparatively cold main recirculation zone. This is consistent with the calculated species concentration and mixture fraction and indicates the calculated mixing rates were too slow. Calculated contours of carbon monoxide concentration are shown in Figure 3.

The presently used turbulence model is capable of reproducing the overall features of swirling flow fairly well. However, discrepancies arise with the prediction of the strength and size of the recirculation zone and additional problems occur downstream of the recirculation regions when combustion takes place. The calculations were made with a relatively fine finite difference mesh and converged quite well. It is therefore unlikely that large discrepancies result from numerical errors. A large part of the discrepancies is often attributed to insufficiently specified inlet conditions. The present initial data was based on estimation and calculation from inlet mass flow rate and geometry of swirler. The exit velocity profile is much more peaked and biased radially outboard than that of the corresponding estimated straight vane swirler.

Appendix I is a technical paper which details the velocity measurements in the swirl combustor using a two component LDV system.

CONCLUSIONS

A set of experimental velocity data in a turbulent, swirling combustor has been obtained. The general flow features of the combustor include: the fuel-air jet coming from the swirler, a large recirculation zone in the center between $z = 0$ and $z = 100$ mm, and a net upward flow above the recirculation zone.

The experimental data was used to test turbulence and combustion models which have been proposed to solve the time-averaged conservation equations. The turbulence model used for these calculations reproduced the overall features of swirling flow fairly well. The numerical simulation also calculated temperatures and compositions as a function of position and time. All of the primary features of the combustor were present in the simulation. However, discrepancies arose with the prediction of the strength and size of the recirculation zone and with the magnitudes of the velocity components. Additional experiments are planned to provide more extensive comparisons of velocity data, temperatures, and composition.

Measurement

20 m/s
↑

Prediction

FIGURE 1: Velocities vector plot in r-z plane.

FIGURE 2: Temperature contour plot for combustor

Level	CO
A	0.010
9	0.009
8	0.008
7	0.007
6	0.006
5	0.005
4	0.004
3	0.003
2	0.002
1	0.001

FIGURE 3:

Contour plots of carbon monoxide concentration.

Appendix

**Technical Paper Presented at the Health Monitoring
Conference Held in Cincinnati, Ohio in November 1990**

VELOCITY MEASUREMENTS IN AN AXISYMMETRIC SWIRL COMBUSTOR USING A TWO COMPONENT LDV SYSTEM

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ABSTRACT

As part of a joint experimental and computational research effort, velocity data has been measured for a turbulent, axially symmetric, gas fueled, fixed vane, swirl combustor. The data was measured with a two component LDV system and includes cross correlation terms for the radial-axial and tangential-axial components.

The combustor utilized in this study was a cylindrical pyrex tube 300 mm long and 95 mm in diameter operated in a vertical orientation. Fuel and air were introduced through a fixed vane swirler built by Rolls-Royce Inc. The fuel was propane flowing at 41 lpm, and the total air flow rate was 1514 lpm.

The general flow features of the combustor include: the fuel-air jet coming from the swirler, a large recirculation zone in the center between $z = 0$ and $z = 100$ mm, and a net upward flow above the recirculation zone. The data set includes axial (u), radial (v), and tangential (w) velocity components; velocity fluctuations (u' , v' , w'); and the axial-radial ($u'v'$) and axial-tangential ($u'w'$) cross correlation terms.

These data have been used to evaluate time averaged CFD models. The turbulence model used for these calculations reproduced the overall features of swirling flow fairly well. However, discrepancies arose with the prediction of the strength and size of the recirculation zone and with the magnitudes of the velocity components. It should be possible to improve the agreement between the experimental and calculated velocity data by modifying the inlet velocity profile.

INTRODUCTION

Turbulent combustion is an energy-production process that results in well-mixed, efficient combustion for propulsion systems. As part of a joint research effort, velocity measurements have been made on a turbulent, axially symmetric, gas-fueled, fixed-vane swirl combustor. These experiments use propane as the fuel and air as the oxidizer to test turbulence and combustion models which have been proposed to solve the time-averaged conservation equations. As experimental methods and computer simulations are tested, it is possible to use methane or hydrogen as fuels and replace air with oxygen. The velocity data reported here were taken with a two-component Laser Doppler Velocimeter and include cross-correlation terms for the radial-axial and tangential-axial velocity components.

OBJECTIVE

The purpose of this work is two-fold. The first objective is to obtain a set of experimental velocity data in a turbulent, swirling combustor. The second objective is to test turbulence and combustion models which have been proposed to solve the time-averaged conservation equations. As computer models are developed which can accurately predict the operating characteristics of combustors, it will be possible to compare alternative combustors or alternative operating conditions and reduce the expense associated with the development of propulsion systems. Operating characteristics of interest include velocities, temperatures, and composition as a function of position and time.

EXPERIMENTAL

The combustor utilized in this study was a cylindrical Pyrex tube, 300 mm in length, with an inside diameter of 95 mm, and operated in a vertical orientation. Fuel and oxidizer were introduced at room temperature through a fuel injector and fixed-vane swirler built by Rolls-Royce, Inc. Figure 1 is a schematic diagram of the combustor and swirler. The swirler consisted of 20 vanes positioned between an inner and outer hub. The diameter of the inner hub was 21 mm and that of the outer hub was 42 mm. The fuel injector was located in the center of the swirler and introduced the fuel into the combustor in a conical sheet with an included angle of 90 deg. A

more detailed description of the swirler is contained in the dissertation of Wilhelmi(1).

The fuel utilized in this study was propane flowing at 41 lpm. The oxidizer was air flowing at 1514 lpm (1364 lpm non-seeded air and 150 lpm of seeded air for velocity measurements). Alumina particles having a nominal diameter of 1 μm were employed as the seed.

Velocity data were collected with a two-component Laser Doppler Velocimeter (LDV). Figure 2 is a schematic diagram of the LDV system. For this system the laser beam was split into three legs, with the polarization of each beam being oriented such that two sets of orthogonal fringes were formed when the beams were crossed. For this study a 9-MHz shift was employed between the reference beam and the other two beams to cover a -50 to +100 m/s velocity range.

The Doppler bursts from each set of fringes were isolated by a Glan Thompson prism and fed into fiber optics which transmitted the light to photomultipliers (pmt's). The output from each pmt was amplified, filtered, and fed into a TSI 990 Burst Counter for analysis. The experimental period of the Doppler burst was determined by the counters and stored in computer memory for analysis. The analysis consisted of conversion of the measured periods to velocity, rotation of the experimental axis, and detailed statistical analysis of the data. Goss, et al(2) provide a more detailed description of the LDV system.

The combustor was mounted on an x,y,z traversing mechanism to translate it relative to the LDV system. When the LDV sample volume passes through the center of the combustor in a direction perpendicular to the axis of the combustor and in the plane of the fringes, axial and radial velocity components are measured; when it passes through the center of the combustor in a direction perpendicular to the axis of the combustor and perpendicular to the plane of the fringes, axial and tangential velocity components are measured.

RESULTS AND DISCUSSION

A graph of axial and radial velocity components is shown in Figure 3. The general flow features of the combustor include: the fuel-air jet coming from the swirler, a large recirculation zone in the center between $z = 0$ and $z = 100$ mm, and a net upward flow above the recirculation zone.

Figure 4 shows radial and tangential components at three different heights($z = 10\text{mm}$, $z = 100$ mm, and $z = 200$ mm) above

the swirler. These profiles were obtained by combining radial and tangential components having the same value of r and z . These components were measured by separate traverses of the LDV system along axes which were perpendicular to each other. Each plot contains a set of up to eight radial-tangential pairs which have been rotated around the centerline to permit visualization of the profile. At $z = 10$ mm the radial velocity component produced by the swirler is clearly evident. At the higher locations the tangential velocity component becomes nearly uniform as the hot gases spiral upward. Based on the axial and tangential velocities at $z = 100$ mm, gases at $r = 5$ mm will rotate through xx deg and gases at $r = 35$ mm will rotate through yy deg before exiting the combustor.

Figure 5 shows axial velocity as a function of z for fixed values of r . As the height above the swirler increases, the velocity profile becomes uniform and the total volumetric flow rate remains constant. Above $z = 100$ mm the axial velocity increases as r increases. There is no data available in the region between $r = 35$ mm and the combustor wall ($r = 42.5$ mm) due to interference between the curved combustor walls and the LDV beams. In the region between $z = 0$ mm and $z = 100$ mm, the high positive axial components associated with the air exiting the swirler and negative axial components associated with the recirculation zone are evident.

Figure 6 shows radial velocity as a function of z for fixed values of r . As air and fuel exit the swirler-injector, there is an outward component of radial velocity. At about $z = 50$ mm, the radial component essentially "bounces" off the combustor wall. The inward component from $z = 60$ mm to $z = 100$ mm is associated with the central recirculation zone shown in Figure 3. Above $z = 100$ mm the radial components decrease and are near zero above $z = 150$ mm. The exception is for $r = 5$ mm where the positive radial component indicates flow away from the center of the combustor.

Figure 7 shows tangential velocity as a function of z for fixed values of r . Near $z = 0$ mm the positive tangential components associated with the swirler are evident, and note that for values of r less than the swirler radius (10.5 mm) the tangential velocity is near zero. Above $z = 100$ mm, the tangential velocity decreases as z increases. At the decay rate shown it may be estimated that the velocity would have only axial components if the combustor extended to about $z = 600$ mm. Above $z = 50$ mm the maximum tangential velocity is between $r = 10$ mm and $r = 20$ mm. Figure 8 shows tangential velocity profiles as a function of r for two values of z . These profiles are characteristic of swirling combustors and show

near rigid body rotation (velocity proportional to r) in the center and free vortex rotation (velocity proportional to $1/r$) as r increases.

An obvious key to efficient combustion is good mixing of the fuel and oxidizer. The fuel injector and fixed-vane swirler used in this study were built by Rolls-Royce, Inc. to provide mixing similar to that of the turbine combustors used in commercial propulsion systems. The velocity profiles described above show that the fuel and oxidizer are well-mixed in the experimental combustor used in this study.

NUMERICAL SIMULATION

Model

The 3-D combustor performance computer program that forms the basis of the present work, is briefly described. For complete details, refer to Report No. USARTL-TR-55C.

The 3-D program is general and is capable of predicting recirculating turbulent flow in most gas-turbine type combustor chambers. The program is formulated in cylindrical coordinates with the boundary being circular. The radius of the boundary may vary with the axial coordinate. Reacting or nonreacting, swirling or nonswirling, diffusion and/or premixed flames, and gaseous and/or liquid fuel combustion can be handled by the program. The program computes the following variables in the region of interest:

- o Axial, radial, and swirl velocity components;
- o Pressure;
- o Enthalpy (temperature); in conjunction with the equation of state, the temperature determines the density variations in the flow field;
- o Turbulent kinetic energy and its dissipation rate;
- o Mass fractions of total fuel (mixture fraction), unburned fuel, oxygen, carbon monoxide, carbon dioxide and H_2O ;
- o Three radiation flux vectors;

The program employs the following physical models to solve the variables mentioned above:

- o Turbulence - Two-equation (k-e) turbulence model to obtain turbulent kinetic energy and its dissipation rate.
- o Chemistry - 4 step chemical reaction scheme:

$$C_X H_y \longrightarrow C_X H_{y-2} + H_2$$

$$C_X H_{y-2} + \frac{X}{2} O_2 \longrightarrow X CO + \frac{y-2}{2} H_2$$

$$CO + 1/2 O_2 \longrightarrow CO_2$$

$$H_2 + 1/2 O_2 \longrightarrow H_2 O$$
- o Chemical Reaction Rate - Fuel and CO consumption rate are assumed to be governed by either the time-averaged Arrhenius model or the turbulent eddy break-up model.
- o Radiation - A six-flux model of radiation.

The transport equations for all dependent variables ϕ are written in the following general form:

$$\text{div} \left(\rho u \phi - \frac{\mu_t}{\sigma_\phi} \text{grad } \phi \right) = S_\phi$$

where ρ denotes the mixture density, u the velocity vector, μ_t the effective or turbulent viscosity, σ_ϕ the effective Prandtl/Schmidt number, S_ϕ the sources of ϕ ; i.e., S_ϕ includes the creation/destruction of ϕ plus other quantities that do not fall under the convective and diffusive terms.

An iterative, general finite-difference solution procedure suitable for 3-D elliptic flows in complex geometries is used to solve the above system of coupled, nonlinear partial-differential equations. The solution procedure involves discretizing the differential equations by integration over elementary finite-difference control volumes surrounding grid nodes that are nonuniformly spaced over the flow field.

Methods

This computation was performed on the Ohio State Cray Y-MP with approximately 3.6 CPU seconds per iteration for $50 * 49 * 6$ fine nodes in (z, r, θ) coordinates. Since the combustor is axisymmetric, the choice of 6 grids in θ direction calculation is for checking the consistency in this direction. The sum of the continuity residuals for all the control volume was less than 0.85% of the total flow rate and

the maximum continuity residual for any control volume was less than 0.0017% of the total flow rate after 450 iterations. The calculations were made with a relatively fine finite difference mesh and converged quite well. It is unlikely that large discrepancies result from numerical errors, but the calculated values have not been shown to be grid independent.

Results and Discussion

Since the flow is assumed to be axisymmetric, the patterns of the flow properties are shown in r - z plane. The velocity vector plot in this plane (Figure 9) shows a qualitative flow pattern agreement with the experimental data. The stagnation point in the calculated flow field occurred at about $1.2 x/D$ while it is at about $0.8 x/D$ in the experimental flow field. After the stagnation point the reverse axial velocities disappear and further downstream the velocity continuously increases due to volume flow rate increase by combustion. The peak of the axial velocity profile shifts toward the outwall as the result of diminishing swirl and combustion. Near the exit plane the calculated axial velocities were slightly lower than the experimental values. The calculated swirl component was generally lower than the experimental value throughout the combustor.

The turbulence model used for these calculations reproduced the overall features of swirling flow fairly well. However, discrepancies arose with the prediction of the strength and size of the recirculation zone and with the magnitudes of the velocity components. Part of the discrepancies may be attributed to insufficiently specified inlet conditions even though the initial data was based on estimation from experimental velocity data and calculation from inlet mass flow rate and geometry of the swirler. It should be possible to improve the agreement between the experimental and calculated velocity data by modifying the inlet velocity profile. Additional experimental data near the inlet over a finer grid would be most helpful.

As was noted above, the model is capable of calculating considerably more information than the velocity components, velocity standard deviations, and velocity correlation coefficients. The temperature contour plot that corresponds with the velocity data in Figure 8 is shown in Figure 10. The prediction shows a hot spot near the exit and a comparatively cold main recirculation zone. This is consistent with the calculated species concentration and mixture fraction and indicates the calculated mixing rates were too slow. Calculated contours of carbon monoxide concentration are shown in

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Figure ~~10~~. Turbulent kinetic energy levels were also calculated, and the maximum levels of turbulent kinetic energy were found in the shear layer region between annular jet and main recirculation zone and at the downstream end of the corner recirculation zone.

Analytical models increase the understanding of the phenomena affecting combustor performance and provide the basis for designing better combustors. The optimization of the design process will require a judicious blend of the emerging analytical tools (correlated and updated with test data) with the established empirical design techniques.

CONCLUSIONS

A set of experimental velocity data in a turbulent, swirling combustor has been obtained. The general flow features of the combustor include: the fuel-air jet coming from the swirler, a large recirculation zone in the center between $z = 0$ and $z = 100$ mm, and a net upward flow above the recirculation zone.

These data have been used to test turbulence and combustion models which have been proposed to solve the time-averaged conservation equations. The turbulence model used for these calculations reproduced the overall features of swirling flow fairly well. The numerical simulation also calculates temperatures and composition as a function of position and time. All of the primary features of the combustor are present in the simulation. However, discrepancies arose with the prediction of the strength and size of the recirculation zone and with the magnitudes of the velocity components. Additional experiments are planned to provide more extensive comparisons of velocity data, temperatures, and composition.

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The computational work was performed on the Ohio State Cray Y-MP and support from Cray Research through the Ohio State Research Foundation is gratefully acknowledged.

REFERENCES

1. J. Wilhelmi, Ph.D. Dissertation, Imperial College, London, 1984.
2. L. P. Goss, D. D. Trump, and W. M. Roquemore, "Combined CARS/LDA Instrument for Simultaneous Temperature and Velocity Measurements," Exp. Fluids 6, 189 (1988).
3. T. W. Bruce, H. C. Mongia, and R. S. Reynolds, "Combustor Design Criteria Validation," Report No. USARTL-TR-55C, Feb 1979.

Figure 1 (below) is a schematic diagram of the cylindrical combustor and swirler.

Figure 1

Figure 2 (below) is a schematic diagram of the two-component laser Doppler velocimeter employed for the combustor studies.

Figure 2

FIGURE 3

FIGURE 4

$Z = 10 \text{ nm}$

$Z = 100 \text{ nm}$

$Z = 200 \text{ nm}$

FIGURE 5

FIGURE 6

FIGURE 7

FIGURE 8

Measurement

20 m/s
→

Prediction

FIGURE 9

Velocities Vector Plot in R-Z Plane

Level T	
A	1706
9	1562
8	1417
7	1272
6	1127
5	982
4	837
3	692
2	547
1	403

FIGURE 10

Level	CO
A	0.010
9	0.009
8	0.008
7	0.007
6	0.006
5	0.005
4	0.004
3	0.003
2	0.002
1	0.001

FIGURE 11

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