Centrifugal fluidization with continuous addition and removal of bed material.

All Tabatabale-Raissi
CENTRIFUGAL FLUIDIZATION WITH CONTINUOUS ADDITION AND REMOVAL OF BED MATERIAL

by

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(Date)

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NOMENCLATURE

\( a \)  Resultant acceleration
\( a_r \)  Radial acceleration
\( a_t \)  Tangential acceleration
\( C_D \)  Drag coefficient \( = \frac{\text{drag force/frontal area}}{\frac{1}{2} \rho_f U_o^2} \)
\( C_l \)  Grid coefficient
\( C_2 \)  Grid coefficient
\( \text{CFB} \)  The centrifugal fluidized bed
\( d_p \)  Particle diameter
\( d_p' \)  Mean particle diameter
\( D_N \)  Inside diameter of feeding nozzle
\( F \)  Resultant force
\( F_D \)  Drug force
\( F_r \)  Radial force
\( F_t \)  Tangential force
\( F_z \)  Axial force
\( \text{Ga} \)  Galileo number \( = \left( \frac{g_a}{g_f} - 1 \right) \omega^2 r_0 \frac{d_p^3}{v_f^2} \)
\( g_c \)  Local acceleration of gravity (9.805 m/s)
\( H \)  Bed height (distance from chamber floor to ceiling)
\( h \)  Distance above chamber floor
\( K \)  Grid coefficient
\( m_{\text{BED}} \)  Bed mass
\( m_p \)  Particle mass
\( \dot{m}_a \)  Mass flow rate of air
\( m_f \) Mass flow rate of fluid

\( m_g \) Mass flow rate of particles in the feeding line

PRT The particle residence time

P Pressure

\( P_{atm} \) Elevated pressure

\( P_i \) Pressure at inner surface of the bed

\( P_o \) Pressure at outer surface of the bed

r Radius from axis of rotation

\( r_b \) Radius of grid at chamber floor

\( r_i \) Radius from axis of rotation to bed surface

\( r_o \) Radius from axis of rotation to grid surface

\( r_T \) Radius of grid at chamber ceiling

\( R_o \) Initial radial distance of injected particle from axis of rotation

\( Re \) Particle Reynolds number - \( \frac{\rho_f \dot{V} d_p}{\mu_f} \)

\( Re_{MF} \) Minimum fluidization Reynolds number - \( \frac{\rho_f U_{MF} d_p}{\mu_f} \)

SDR The solids discharge rate

SF&R The solids feed and removal under continuous and stable condition

\( t \) Time

\( \dot{t} \) Start up time

\( r_r \) Radial velocity of fluid

\( u_t \) Tangential velocity of fluid
<table>
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<th>Symbol</th>
<th>Description</th>
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<tr>
<td>(U_i)</td>
<td>Local velocity of air</td>
</tr>
<tr>
<td>(U_{MF})</td>
<td>Minimum fluidizing velocity</td>
</tr>
<tr>
<td>(U_o)</td>
<td>Superficial air velocity at (r=r_o)</td>
</tr>
<tr>
<td>(U_r)</td>
<td>Radial velocity of air</td>
</tr>
<tr>
<td>(\dot{V})</td>
<td>Relative velocity of the fluid with respect to the particle</td>
</tr>
<tr>
<td>(V_0)</td>
<td>Initial velocity of particle</td>
</tr>
<tr>
<td>(V_r)</td>
<td>Radial velocity of particle</td>
</tr>
<tr>
<td>(V_s)</td>
<td>Average velocity of particle at feeding nozzle</td>
</tr>
<tr>
<td>(V_t)</td>
<td>Tangential velocity of particle</td>
</tr>
<tr>
<td>(z)</td>
<td>Height above chamber floor</td>
</tr>
<tr>
<td>(\alpha)</td>
<td>Grid tapper angle, size-density parameter</td>
</tr>
<tr>
<td>(\gamma)</td>
<td>Grid coefficient</td>
</tr>
<tr>
<td>(\Delta A_i)</td>
<td>Semicurcular incremental area</td>
</tr>
<tr>
<td>(\Delta H)</td>
<td>Incremental height</td>
</tr>
<tr>
<td>(\Delta P_{BED})</td>
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<td>(\Delta P_{CL})</td>
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<td>(\Delta P_{FB})</td>
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<tr>
<td>(\Delta P_{GRID})</td>
<td>Grid pressure drop</td>
</tr>
<tr>
<td>(\Delta P_i)</td>
<td>Local kinetic pressure drop</td>
</tr>
<tr>
<td>(\Delta P_{PB})</td>
<td>Packed bed pressure drop</td>
</tr>
<tr>
<td>(\Delta P_T)</td>
<td>Total pressure drop</td>
</tr>
<tr>
<td>(\varepsilon)</td>
<td>Void fraction</td>
</tr>
<tr>
<td>(\varepsilon_N)</td>
<td>Void fraction of feeding line</td>
</tr>
<tr>
<td>(n_1)</td>
<td>Grid coefficient</td>
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\[
\varepsilon_n = \varepsilon \cdot C_1 \cdot \frac{3 \rho_f}{4 \rho_s d_p}
\]
\eta_2 \quad \text{Grid coefficient} = \left[ H(r_b + r_T) \right] C_2

\Theta_o \quad \text{Initial injection direction of particle}

\mu_f \quad \text{Fluid viscosity}

\dot{\nu} \quad \text{Volumetric flow rate of air}

\dot{\nu}_F \quad \text{Volumetric flow rate of fluid}

\rho_a \quad \text{Air density}

\rho_{\text{BED}} \quad \text{Effective density of the bed} = (\rho_s - \rho_f)(1-\epsilon)

\rho_f \quad \text{Fluid density}

\rho_s \quad \text{Particle density}

\phi_s \quad \text{Sphericity of particle}

\omega_o \quad \text{Angular velocity}
ABSTRACT

This thesis involves a study of the continuous feeding and removal of solid particles with a centrifugal fluidized bed. The system consists of a conical air distributor aligned with its axis of rotation in the vertical direction. Solid particles are transferred pneumatically to the rotating bed from a pressurized tank and are injected towards the bed from the free-board region, while the fluidizing air flows radially inward through the distributor. Particles are allowed to discharge from adjustable gates in the bottom end wall of the rotating chamber. The thickness of the bed is controlled by changing the area available for the exit flow.

Experiments were performed using glass beads and air, where the experimental parameters include angular velocity of the bed, gas flow rate, bed mass, particle feed rate, and the geometry of the discharge ports. Results of batch experiments for the bed pressure drop and minimum fluidizing velocity are presented and compared with previous investigations and theoretical models. Data also are presented on system operation with solid material being fed and removed continuously. These results demonstrate stable, steady state bed operation with continuous addition and removal of solids.
I. INTRODUCTION

Fluidization is the operation by which fine solids are transformed into a fluidlike state through contact with a gas or liquid [1]. At low flow rates, the particles are not disturbed by the flow, the bed is packed, and the fluid flows through the voids between particles. With an increase in flow rate, the particles move apart and a point is reached where the particles become suspended in the flowing gas. At this point the frictional force between the particle and fluid balances the body forces acting on the particle. The bed is considered to be just fluidized and is referred to as an incipiently fluidized bed or a bed at minimum fluidization. In some respect, fluidized beds behave like liquids, and, in particular, the pressure in a fluidized bed varies hydrostatically with depth. Defining the effective density of the bed as $\rho_{\text{BED}}$, the pressure drop through a bed of height $H$ is [2]:

$$\Delta p_{\text{BED}} = \rho_{\text{BED}} H g$$

The radial pressure drop through a fluidized bed of mass $m_{\text{BED}}$ in a vertical (or horizontal) centrifuge of height (or length) $H$ rotating at angular velocity $\omega_o$ is expressed by a formula analogous to that for a liquid [2]:

$$\Delta p_{\text{BED}} = m_{\text{BED}} \omega_o^2 / 2 \pi H$$

The concept of the conventional, stationary fluidized bed is not new. The first large scale, commercially
significant use of fluidized beds was by Fritz Winkler for the gasification of powdered coal. Since that time many diverse designs have been developed and this technique has been found to be useful in a variety of industrial applications.

The centrifugal fluidized bed (CFB) is a relatively new concept which shows good promise for a variety of industrial applications [3,4,5,6]. The CFB system as shown in Figure 1 with sorbent feeding and removal has a number of advantages over conventional, stationary fluidized bed systems. The rotational nature of CFB has some interesting operational characteristics for the combustion application. The relatively high fluidizing velocities provide for compact combustors, resulting in relatively easy start-up and fewer problems with solids feed and bed mixing. By varying the speed of the rotation of the bed, the bed temperature and the fluidizing velocity, the power output of the device can be varied over an extremely wide range. Studies of centrifugal fluidized beds have been underway since 1960. Lately, the CFB has been the point of much study in the Mechanical Engineering Department at Lehigh University. A parametric analysis of the CFB from the fluid mechanics point of view was done by C. Dodge [3], and later an analytical model for prediction of bed characteristics was developed by Levy, et al. [5], for a uniform, thin CFB. W. J. Shake-
Figure 1. Sketch of centrifugal fluidized bed with solids feed and removal
Speare [4] studied some aspects of the feasibility and operability of a CFB combustion system in conjunction with other fluidized bed designs and the conventional pulverized fuel power plant. More recently, N. W. Martin [7] investigated the effects of operating a fluidized batch bed in a centrifugal flow field. His apparatus consisted of a cylindrical porous distributor, with its axis of rotation parallel to the gravitational acceleration. A theoretical model which used the liquid model assumption to calculate the shape of the bed also was presented by Martin [6,7] to compute the bed pressure drop numerically. In Section III, the results of recent studies and developments on this model are presented.

Most of the coal currently available contains a relatively high percentage of sulfur, which during combustion combines with oxygen to form $\text{SO}_2$. Sulfur removal can be accomplished to a large extent by including limestone particles in the fluidized bed. The limestone reacts with sulfur from the coal, substantially reducing the sulfur in the product gas. With bed material of dolomite or limestone to capture $\text{SO}_2$, the CFB combustor could be used to burn high sulfur coal or coal char.

The objectives of this study are to demonstrate the feasibility of continuous and stable solids feed and removal (SF&R) in a centrifugal fluidized bed apparatus,
to determine the effects of SF&R on the bed characteristics. Based on previous investigations [3,4], a CFB apparatus capable of operating with continuous SF&R was designed and fabricated, and a series of experiments was performed.
II. APPARATUS AND INSTRUMENTATION

The apparatus consists of three sections: the air supply system, the particle feeding system and the test section.

A. Air Supply System

The air supply system consists of two positive displacement air compressors, each nominally operated at 0.235 SCMS at $6.9 \times 10^5$ Pa pressure and equipped with aftercoolers and oil traps. The air flows from the compressors to an air storage tank and then to the fluidized bed laboratory. The air flow rate is controlled by a globe valve, located upstream of a 0.3 m length of pipe packed densely with commercial grade steel wool to muffle the flow.

After the muffler, a straight length of pipe 2.3 m long is used to provide sufficient length for the axial velocity profile of the air to be fully developed. Kinetic pressures used to determine velocities were measured with a United Sensor Corporation PAC-8-KL pitot-static probe having an outside diameter of 0.003175 m. The dynamic pressures are measured on a Dwyer 0.254 m (10 inches) inclined differential manometer. Flexible hoses connect the manifold with four outlets (each 0.102 m diameter) to the stationary cylindrical air plenum.

Specifications for the air supply system are given in Table 1.
Compressors

Main air supply line

Pressure regulator and air supply line to solids feed system

Globe valve

Muffler

Pitot-static tube

Test section

Figure 2. Air supply system assembly.
## TABLE 1. AIR SUPPLY SYSTEM SPECIFICATIONS

<table>
<thead>
<tr>
<th>Specification</th>
<th>Value</th>
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<tr>
<td>Minimum air flow rate</td>
<td>0.0236 SCMS</td>
</tr>
<tr>
<td>Maximum measurable air flow rate</td>
<td>0.4847 SCMS</td>
</tr>
<tr>
<td>Maximum air flow rate available</td>
<td>0.4719 SCMS</td>
</tr>
<tr>
<td>Main air supply line pressure at full load operation of compressors</td>
<td>65,500 Pa above atmospheric</td>
</tr>
<tr>
<td>Inside diameter of air supply line</td>
<td>0.102 m</td>
</tr>
<tr>
<td>Axial distance of pitot-static probe from butterfly valves</td>
<td>2.3 m</td>
</tr>
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B. Calibration of Air Flow Rate

By traversing the inside diameter of the air supply pipe (0.1016 m ID) with a pitot-static probe, kinetic pressures were measured at 0.003175 m (1/8 inch) intervals along two different diameters, perpendicular to each other. The velocity profiles were determined for both directions and were reasonably symmetric.

The local velocity $U_i$ at each semicircular incremental area $\Delta A_i$ was determined by measuring the local kinetic pressure drop $\Delta P_i$:

$$U_i = \sqrt{\frac{2g_c \Delta P_i}{\rho_a}}$$

where $\rho_a$ is density of the air at atmospheric pressure. Integrating the velocity profiles over the pipe cross section, the volumetric flow rates were determined at several different air flow rates:

$$\dot{m}_a = \rho_a \sum U_i \Delta A_i = \sqrt{2\rho_a g_c} \sum \Delta A_i \sqrt{\Delta P_i}$$

A calibration chart is shown in Fig. 3 for air flow rate versus square root of centerline kinetic pressure drop $\Delta P_{CL}$. From these data, the following linear relationship was determined by using the method of least squares fit to straight line:

$$\text{Volumetric flow rate (SCMS)} = 0.156592\sqrt{\Delta P_{CL}}$$

where $\Delta P_{CL}$ is the kinetic pressure drop as determined
Figure 3. Calibration data for air flow rate.
at the centerline of the air supply pipe in inches of water.

C. Particle Feed System

A Petrocarb ABC injector was used to inject glass beads into the rotating chamber. Solids from the injector were carried pneumatically in a stream of dry air through a flexible hose into the injector nozzle.

This system is described in detail in Fig. 4. Air which is used to carry the solid particles is supplied from the main air line and the pressure is controlled by a pressure regulator.

D. Control of the Solids Feed Rate

The solids feed rate is governed by two factors: tank pressure and conveying air flow rate. The tank pressure is controlled by adjusting the tank pressure regulator, while the air flow rate is regulated by valves. The conveying air flow rate is indicated on a Shutte and Koerting rotameter-tube No. 3HCF.

Satisfactory injector operation depends upon using the proper gas velocity, air, and solids flow rates. Control of solid feed rate was accomplished by inserting small diameter orifices upstream of the 0.00635 m (1/4 inch) ball valve. The system was always operated with the valve open, with feed control obtained by using the proper orifice opening and tank pressure.
Figure 4. ABC Injector.
The orifices were calibrated with the system shown in Fig. 5. Typical calibration data at different air flow rates and tank pressures are given in Figs. 6 and 7. The calibration was checked periodically during the experimental program.

The calibration procedure consisted of the following steps:

1. Connect the injection lance into the cyclone.
2. Open the feeding hose shut-off valve to start the feeding hose gas flow.
3. Open the feed valve.
4. To stop the injection, close the feed valve, take the lance out of the cyclone, and then close the feeding hose shut off valve.
5. When the injection operations are completed, close the tank pressurizing valve and release the tank pressure through the top of the tank.
6. Weigh the particles collected in the cyclone and measure the time of collection. The data in Figs. 6 and 7 assume steady state solids feed rates.

The calibration results show that at constant air flow rate, the solids feed rate increases with increasing tank pressure, and at constant tank pressure the solids feed rate decreases slightly with increasing flow rate of conveying air.
Solid flow from injector tank

Orifice insert

Bushing

0.00635 m (1/4 inch) ball valve

Sleeve holder

Diluted gas line

Solid-gas mixture to injector lance

Figure 5. Cross section of mixing valve assembly.
Figure 6. Calibration chart for feeding system.

\[
\bar{d}_p = 427 \times 10^{-6} \text{ m} \\
\bar{v}_a = 9 \text{ to } 16 \times 10^{-6} \text{ SCFM}
\]
Figure 7. Influence of air flow rate on solids discharge from feed tank.
In some cases the solids and gas failed to flow properly and the lance and valves were removed and cleaned out. One indication that plugging is starting is a drop in the float in the rotameter and an increase in tank pressure.

If the solids fail to flow due to plugging of the orifice in the mixing assembly, the tank pressure decreases. To remove the obstruction from the orifice in the mixing assembly, close the feed valve, disengage the disconnecting couplers, remove the object from the orifice, and reassemble.

Specifications for the solids feed calibrating system and solid feed are given in Tables 2 and 3.

E. Description of the Test Section

The apparatus consists of a rigid stationary frame constructed to support the electrical drive motor, shaft, bearings, air plenum, and all rotating members. The test section is divided into four subsystems: the injector lance, the stationary air plenum, the rotating chamber assembly and the particle collection system.

1. Injector Nozzle

The injector nozzle is made of a 0.61 m long section of bronze tubing. The tip of the nozzle is shaped to permit the particles to be fed tangentially into the bed. The design of the injector nozzle is one of the critical elements in the particle feeding system. The
TABLE 2. SOLID FEED CALIBRATING SYSTEM SPECIFICATIONS

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>Specification</th>
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<tr>
<td>Mean particle size used</td>
<td>( d = 93 \times 10^{-6} \text{m}, 427 \times 10^{-6} \text{m} )</td>
</tr>
<tr>
<td>Filter</td>
<td>used 22 gauge perforated steel, 50% open area-</td>
</tr>
<tr>
<td></td>
<td>stainless steel screen, U.S. sieve #200 mesh</td>
</tr>
<tr>
<td>Minimum measurable particle weight (scale resolution)</td>
<td>0.03125 Kg</td>
</tr>
<tr>
<td>Cyclone:</td>
<td></td>
</tr>
<tr>
<td>Air exhaust diameter</td>
<td>0.0762 m</td>
</tr>
<tr>
<td>Inlet diameter</td>
<td>0.009525 m</td>
</tr>
<tr>
<td>Cyclone exit diameter</td>
<td>0.0381 m</td>
</tr>
<tr>
<td>Height of cylindrical portion</td>
<td>0.3048 m</td>
</tr>
<tr>
<td>Height of conical portion</td>
<td>0.3048 m</td>
</tr>
<tr>
<td>Total height</td>
<td>0.61 m</td>
</tr>
<tr>
<td>Specification</td>
<td>Value</td>
</tr>
<tr>
<td>--------------------------------------</td>
<td>--------------------------------------------</td>
</tr>
<tr>
<td>Mean particle size used</td>
<td>$d_p = 427 \times 10^{-6} \text{m}$</td>
</tr>
<tr>
<td>Maximum solid particle loaded</td>
<td>100 kg</td>
</tr>
<tr>
<td>Maximum air flow rate</td>
<td>0.00125 SCMS</td>
</tr>
<tr>
<td>Minimum measurable air flow rate</td>
<td>0.000005 SCMS</td>
</tr>
<tr>
<td>Air supply line pressure</td>
<td>275790 Pa above atmos.</td>
</tr>
<tr>
<td>Feeding hose pressure</td>
<td>206843 Pa above atmos.</td>
</tr>
<tr>
<td>Tank pressure</td>
<td>between 68947.5 Pa and 124106 Pa above atmos.</td>
</tr>
<tr>
<td>Minimum measurable pressure</td>
<td>6895 Pa</td>
</tr>
<tr>
<td>Opening size of feed valve</td>
<td>0.00635 m (I.D.)</td>
</tr>
<tr>
<td>Inside diameter of orifices used</td>
<td>0.003175 m, 0.00635 m</td>
</tr>
<tr>
<td>Mixing Tee assembly size</td>
<td>0.0127 m (I.D.)</td>
</tr>
<tr>
<td>Sleeve size</td>
<td>0.0039687 m (I.D.)</td>
</tr>
<tr>
<td>Injector lance diameter</td>
<td>0.009525 m</td>
</tr>
</tbody>
</table>
final goal in the design of the injector nozzle was to feed the solid particles into the bed without particle entrainment. The important design factors were: injector nozzle diameter, injection velocity of particles, direction of injected particles and axial location of the injector nozzle. Based on theoretical studies of particle dynamics in a centrifugal field, the feed system was designed to inject solid particles tangentially in the freeboard region. To provide more successful feeding, the lance was axially located in the upper half of the rotating chamber.

2. Stationary Air Plenum

Four flexible hoses connect the air manifold to the stationary air plenum, which is made of steel. A pressure tap is attached to the plenum at the top and is connected by PVC tubing to a 1.5 m vertical manometer board. The air is distributed uniformly around the rotating chamber, by means of four inlet holes which inject the air (each 0.1016 m diameter).

The plenum must operate at elevated pressures, and pressure seals are included at the top and bottom of the rotating assembly to prevent air leakage. The seals are fabricated from circular Teflon strips. During normal operation of the system, the pressure inside the plenum is above atmospheric pressure and the free edge of the top seal is pushed against the upper rim of the rotating
chamber and the bottom seal is pressed on a well polished insert attached to the bottom of the plenum.

The design of the bottom seal is one of the critical elements in the apparatus. Air leakage through the seal would result in an air flow from the stationary air plenum down into the collecting bin. This would cause air to flow upward from the collecting bin through the discharge holes in the bottom end wall of the rotating chamber, interrupting the gravity discharge of solids from the bed.

3. Rotating Chamber Assembly

The rotating chamber assembly consists of the grid cage and grid assembly. The rotating grid cage was designed to meet the following requirements:

a. The grid assembly can be removed easily.

b. To provide clear viewing with the possibility of photographing the bed during operations, the top bed wall was made from plexiglass.

c. The particle removing system consists of four adjustable slots located on the bottom end wall of the rotating cage, each slot providing a maximum discharge area of $0.001267 \text{ m}^2$. The discharge is shown in Fig. 8. The particle discharging area is adjusted by a thin circular plate located beneath the bottom end wall of the rotating grid. The adjusting plate is welded to the hollow shaft which rotates with the cage assembly, and the
Figure 8. Geometry of discharge holes in bottom end wall.
drive shaft is located inside the hollow shaft. A screw which passes through a threaded hole on the hollow shaft connects the drive shaft and the adjusting plate during operations.

The particle removing system was specially designed so that the discharge opening can be adjusted manually by the operator. To change the discharge area, the system's screw is loosened to disengage the adjusting plate from the drive shaft. The plate is rotated manually relative to the drive shaft, and the screw is tightened.

The grid (Fig. 9) was constructed in the shape of a truncated cone with a $4^\circ$ taper angle, using perforated sheet metal with 50% open area. The inside surface of the grid was covered with stainless steel screen (U.S. Sieve #100 mesh), and the outside surface was wrapped with a single layer of canvas-like cloth stitched in place. The stainless steel screen was butt-welded to the grid using silver solder. The whole assembly was sealed into retaining grooves on the top and bottom end walls of the rotating grid cage using silicon sealant.

4. Particle Collecting System

The particle collecting system consists of a collecting chamber located below the rotating chamber and sealed from the stationary air plenum. To permit the operator to observe the particle discharging from the test section and to allow partial dismantling of the
0.3048 m Diameter

H = 0.1572 m

100 mesh stainless steel screen

22 Ga perforated metal sheet with 50% open area

Canvas-like cloth (one layer stitched outside around frame)

Figure 9. Grid assembly.
rotating cage, two plexiglass lids are bolted to the sides of the collecting chamber. Two flexible hoses (each 0.0762 m in diameter) connect the collecting chamber to the particle storage bin from which particles are removed periodically. A pressure tap is attached to the collecting chamber for measuring the pressure of the bin. This pressure is useful in determining whether the seal between the stationary plenum and the collecting chamber is operating properly.

Specifications for the test section are given in Table 4.

5. Measurement of Mean Particle Size

Potters Industries Technical Quality Glass Beads were sieved and weight percentage determined on each sieve. The following result was obtained:

<table>
<thead>
<tr>
<th>U.S. Sieve No.</th>
<th>Weight % Retained</th>
</tr>
</thead>
<tbody>
<tr>
<td>35</td>
<td>2.1</td>
</tr>
<tr>
<td>40</td>
<td>79.1</td>
</tr>
<tr>
<td>50</td>
<td>16.7</td>
</tr>
<tr>
<td>50+</td>
<td>2.1</td>
</tr>
</tbody>
</table>

The Mean Particle Size $d_p$ may be calculated from:

\[
\frac{Diam. Range \times 10^6 m}{d_p} x_1 \left(\frac{x}{d_p}\right)_1 \times 10^6
\]

\[
\begin{array}{cccc}
0-300 & 150 & 0.021 & 140 \\
300-425 & 363 & 0.167 & 460 \\
425-500 & 463 & 0.791 & 1710 \\
500+ & 550 & 0.021 & 38 \\
\end{array}
\]

Then,

\[d_p = 426 \times 10^{-6} \text{ m}\]
### TABLE 4. SPECIFICATIONS OF TEST SECTION

<table>
<thead>
<tr>
<th>Specification</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Particle size</td>
<td>427x10^{-6} m diameter</td>
</tr>
<tr>
<td>Grid diameter at top</td>
<td>0.3048 m</td>
</tr>
<tr>
<td>Grid diameter at bottom</td>
<td>0.2762 m</td>
</tr>
<tr>
<td>Grid taper angle</td>
<td>4°</td>
</tr>
<tr>
<td>Grid height</td>
<td>0.1524 m</td>
</tr>
<tr>
<td>Maximum angular velocity</td>
<td>47 rad/sec</td>
</tr>
<tr>
<td>Maximum plenum pressure</td>
<td>12500 Pa above atmos.</td>
</tr>
<tr>
<td>Maximum air flow rate</td>
<td>0.00125 SCMS</td>
</tr>
<tr>
<td>Total weight of particles</td>
<td>100 Kg</td>
</tr>
<tr>
<td>Minimum measurable plenum pressure: Manometer resolution</td>
<td>25 Pa</td>
</tr>
<tr>
<td>Minimum measurable particle weight: scale resolution</td>
<td>0.03125 Kg</td>
</tr>
<tr>
<td>Total height of the apparatus</td>
<td>2 m</td>
</tr>
<tr>
<td>Total area occupied by test section</td>
<td>1 m²</td>
</tr>
</tbody>
</table>
III. CENTRIFUGAL FLUIDIZED PARTICLES IN BATCH BEDS

A. Overview

Studies have been performed on centrifugal fluidized particles in batch beds by other investigators [7]. This chapter describes batch experiments performed with the apparatus described in Chapter II. First, the experimental results are presented, then the existing theoretical models are discussed, and finally new theoretical equations are presented.

B. Experimental procedures for determination of bed pressure drop

To check the previous results on minimum fluidization and bed pressure drop, twenty-nine experiments were run with six different bed masses. Accurate measurements of grid pressure drop is critical in determining the bed pressure drop. Grid pressure drops, measured periodically as a function of air flow rate and angular velocity, appeared to be a strong function of atmospheric conditions (temperature, moisture, etc.). The grid pressure drop was taken equal to the difference between plenum pressure and the atmosphere. The results of grid pressure drop versus air flow rate for several plenum locations are given in Figure 10. Two equations were used to correlate the grid pressure drop data:

\[ P_{\text{Grid}} = C_1 \dot{v} + C_2 \dot{v}^2 \]
and
\[ p_{\text{Grid}} = K \nu^{3/2} \]

where parameters \( C_1, C_2 \) and \( K \) are constants to be determined experimentally. The values of \( C_1, C_2 \) and \( K \) given in Figure 10 correlate the experimental data best. The values are not a strong function of the angular velocity of the grid.

The experiments on bed pressure drop were performed with a 4° grid taper angle, and 427 x 10^-6 m glass spheres. Altogether data were obtained at five angular velocities:

- 20.9 rad/s (200 rpm)
- 26.2 rad/s (250 rpm)
- 31.4 rad/s (300 rpm)
- 36.7 rad/s (350 rpm)
- 41.9 rad/s (400 rpm)

and six bed charge masses:

- 0.91 Kg (2 lbs)
- 1.36 Kg (3 lbs)
- 1.81 Kg (4 lbs)
- 2.72 Kg (6 lbs)
- 3.63 Kg (8 lbs)
- 4.54 Kg (10 lbs)

In each experiment the total bed pressure drop was measured at several air flow rates. By plotting total pressure drop and grid pressure drop versus air flow rate and using the relation:

\[ \Delta P_{\text{BED}} = \Delta P_{\text{TOTAL}} - \Delta P_{\text{GRID}} \]

29
Figure 10. Grid pressure drop.
the bed pressure drop was determined experimentally.

C. Analytical study of centrifugal particles in batch beds

Experimental results show that as the air flow rate increases, the particles in a centrifugal bed pass through three different regimes. At air flow rates below minimum fluidization, the bed is packed and $\Delta P_{BED}$ increases with air flow rate. When the bed is fully fluidized, a constant bed pressure drop is observed. Normally the transition between the packed and fluidized states is not abrupt, and a well defined transition region exists.

The net pressure gradient through a packed, stationary bed is expressed by the Ergun Equation [1]:

$$\frac{dP}{dr} = \frac{150(1-\varepsilon)^2 \mu_f}{\varepsilon^3 (\phi dp)^2} U_r + \frac{1.75(1-\varepsilon) \rho_f}{\varepsilon^3 \phi dp} U_r^2$$

Information on the radial variation of the radial velocity component is needed to integrate this equation to determine packed bed pressure drop for the centrifugal bed. If the axial component of the air velocity through the bed is assumed to be negligible, and the radial component of velocity is independent of axial location, the packed bed pressure drop may be approximated by [6]:

$$\Delta P_{PB} = \left[ \frac{150(1-\varepsilon)^2}{\varepsilon^3 (\phi dp)^2} \right] \mu_f U_o r_0 \ln \left( \frac{r_o}{r_i} \right) + \left[ \frac{1.75(1-\varepsilon)}{\varepsilon^3 \phi dp} \right] \rho_f \left( U_o r_o \right)^2 \left( \frac{1}{r_i} - \frac{1}{r_o} \right)$$
where $U_x$ is approximated from the continuity equation for a bed of height $H$:

$$U_x = \frac{\dot{m}_a}{2\pi \rho_f r H}$$

A model for predicting the shape of the fluidized bed based upon the assumption that the bed is a liquid with an effective density $\rho_{BED}$ equal to:

$$\rho_{BED} = (\rho_s - \rho_f)(1-\epsilon)$$

is used by Levy et al. [6] to determine fluidized bed pressure drop. The next section is dedicated to the new developments of this model and finally the results of this simple model are compared to predicted results using the previous model [6].

D. The liquid model

Assume the bed behaves as a liquid in rigid body rotation. The force balance on a fluid element as shown in Figure 11 is:

$$\Sigma F_r = \Sigma F_z = 0$$

Then:

$$\frac{\partial p}{\partial r} = \rho_{BED} r \omega_o^2$$

$$\frac{\partial p}{\partial z} = -\rho_{BED} g$$

Hence:

$$p = \frac{1}{2} \rho_{BED} r^2 \omega_o^2 - \rho_{BED} g z + p'$$

where $p'$ is the constant of integration and to be determined using a boundary condition for $p$ and $r$. At some
Figure 11. Differential bed element for bed pressure drop analysis.
bed height \( z_0 \):
\[
    p_0 = \frac{1}{2} \rho \text{BED} r_o^2 \omega_o^2 - \rho \text{BED} g z_0 + p'
\]
\[
    p_1 = \frac{1}{2} \rho \text{BED} r_1^2 \omega_o^2 - \rho \text{BED} g z_0 + p'
\]

Then:
\[
\Delta p_{FB}(\text{at height } z_0) = p_0 - p_1 - \frac{1}{2} \rho \text{BED} \omega_o^2 (r_o^2 - r_1^2)
\]

The average bed pressure drop may be determined by integrating \( \Delta p_{BED} \) (at height \( z_0 \)) over the entire bed height, \( H \):
\[
\Delta p_{FB} = \frac{1}{H} \int_0^H \frac{1}{2} \rho \text{BED} \omega_o^2 (r_o^2 - r_1^2) \, dz \tag{1}
\]

The mass of the fluid element is:
\[
dm_f = \rho \text{BED} \, dv = \rho \text{BED} \, r \, dr \, dz
\]

Then, the bed mass is:
\[
\text{m}_{BED} = 2\pi \int_0^H \int_0^{r_o} \rho \text{BED} \, r \, dr \, dz
\]
\[
= 2\pi \int_0^H \frac{1}{2} \rho \text{BED} (r_o^2 - r_1^2) \, dz \tag{2}
\]

Combining equations (1) and (2):
\[
\Delta p_{FB} = \frac{\text{m}_{BED} \omega_o^2}{2\pi H} \tag{3}
\]

The following boundary condition is used to determine the constant integration \( p' \):

At \( r=r_1 \), \( p=p_{\text{atm}} \)
Then

\[ p_i - p_{atm} = \frac{1}{2} \rho_{BED} r_i^2 \omega_o^2 - \rho_{BED} g z + p' \]

Hence:

\[ r_i^2 = \frac{2 g z}{\omega_o^2} + \frac{2(p_{atm} - p')}{\rho_{BED} \omega_o^2} \]  \hspace{1cm} (4)

Substituting into equation (2) gives:

\[ m_{BED} = 2 \pi \rho_{BED} \int_0^H \left( 2 r_o^2 - \frac{g z}{\omega_o^2} + \frac{p' - p_{atm}}{\rho_{BED} \omega_o^2} \right) dz \]

where:

\[ r_o = z \tan \alpha + r_b \]

Then

\[ m_{BED} = 2 \pi \rho_{BED} \left[ \frac{H^3}{3} \tan^2 \alpha + H^2 r_b \tan \alpha + H r_b^2 \right. \]

\[ - \frac{g H^2}{\omega_o} + \frac{2 H}{\rho_{BED} \omega_o} \left( p' - p_{atm} \right) \]

or

\[ \frac{2 H}{\rho_{BED} \omega_o} (p_{atm} - p') = \frac{H^3}{3} \left( r_T^2 + r_b^2 + r_T r_b \right) - \frac{g H^2}{\omega_o} \]

\[ - \frac{m_{BED}}{\pi \rho_{BED}} \]  \hspace{1cm} (5)

Combining equations (4) and (5), an expression for the shape of the fluidized bed is found to be:

\[ r_i^2 = \frac{g}{\omega_o^2} \left( 2 z - H \right) + \frac{1}{3} \left( r_T^2 + r_b^2 + r_T r_b \right) \]

\[ - \frac{m_{BED}}{\pi \left( \rho_s - \rho_f \right)(1-c)H} \]  \hspace{1cm} (6)
The minimum fluidizing velocity calculated at the outer radius of the bed is expressed by (6):

$$G_a = \frac{150(1-\varepsilon)}{\varepsilon/\phi_s} \cdot 2 Re_{MF} + \frac{1.75}{\phi_s \varepsilon/3} Re_{MF}^2$$

(7)

where:

$$G_a = \left( \frac{\rho_s}{\rho_f} - 1 \right) \omega^2 r_0 \frac{dp}{\nu_f}$$

and

$$Re_{MF} = \frac{U_{MF} dp}{\nu_f}$$

Theoretical results using equation (7) are compared to experimental data a minimum fluidization in Figure 12.

E. Numerical solutions for bed pressure drop.

The analytical procedure developed by Levy and Martin [6] assumes a value for $\Delta p_T$ and uses the equation:

$$\Delta p_T = \Delta p_{\text{GRID}} + \Delta p_{\text{BED}}$$

(8)

to compute the radial velocity of the air, $U_o(r_o)_j$ for each of the $n$ elements, at a given incremental height $\Delta H$. The total air flow rate through the distributor is:

$$\dot{m}(\Delta p_T) = \sum_{j=1}^{n} \rho_f U_o(r_o)_j (2\pi r_o)_j \Delta H$$

(9)

For a packed bed, the bed pressure drop $\Delta p_{\text{BED}}$ in equation (8) at each incremental height $\Delta H$ is calculated
Figure 12. Comparison of theoretical and experimental results for minimum fluidization.
from the equation:

\[ \Delta p_{PB} = \left[ 150(1-\varepsilon)^2/\varepsilon^3 (p_d^2) \right] u_f U_o r_o \ln(r_o/r_i) \]

\[ + \left[ 1.75(1-\varepsilon)/\varepsilon^3 (p_d) \rho_f (U_o r_o)^2 \left( \frac{1}{r_i^2} - \frac{1}{r_o^2} \right) \right] \quad (10) \]

The minimum fluidizing velocity at the corresponding element also is determined using equation (7). If the velocity \( U(r) \) computed from equation (8) using the packed bed pressure drop is greater than the minimum fluidizing velocity, then the equation:

\[ \Delta p_T = \Delta p_{GRID} + \Delta p_{FB} \quad (11) \]

is used to recalculate the radial velocity through the ring (\( \Delta H_j \)). An empirical correlation of experimental data for grid pressure drop is used for \( \Delta p_{GRID} \) in equation (11). This analytical procedure also assumes a liquid-like behavior to calculate the bed shape. A computer program was developed by N. Martin [7] which uses this analytical model to predict bed pressure drop versus air flow rate in both the packed and fluidized bed regimes.

F. Analytic solution for fluidized bed pressure drop

In section B, two profiles were used to correlate experimental data for grid pressure drop:

\[ \Delta p_{GRID} = \eta_1 U_o + \eta_2 U_o^2 \quad (12) \]
and

\[ \Delta P_{\text{GRID}} = \gamma U_o^{3/2} \]  \hfill (13)

where

\[ \eta_1 = \pi (r_b + r_T)HC_1 \]
\[ \eta_2 = [\pi (r_b + r_T)H]^2 C_2 \]
\[ \gamma = [\pi (r_b + r_T)H]^{3/2} K \]

In section D, the fluidized bed pressure drop \( \Delta p_{FB} \) at a height \( z \), was found to be:

\[ \Delta p_{FB}(\text{at height } z) = \frac{1}{2} (\rho_s - \rho_f) (1 - \epsilon) \omega_0^2 (r_o^2 - r_1^2) \]  \hfill (14)

Introducing equations (12) and (14) into equation (11) gives:

\[ \Delta p_T = \eta_1 U_o + \eta_2 U_o^2 + \frac{1}{2} (\rho_s - \rho_f) (1 - \epsilon) \omega_0^2 (r_o^2 - r_1^2) \]  \hfill (15)

Introducing equation (6) into equation (15) and changing the dependent variable \( z \) to \( z_1 \), where:

\[ z_1 = z/H \]

the expression for \( p_T \) becomes:

\[ \Delta p_T = 2(1-\epsilon) (b z_1^2 + c z_1 + d) + e + \eta_1 U_o + \eta_2 U_o^2 \]

where

\[ a = \frac{1}{2} \omega_0^2 (\rho_s - \rho_f) \]
\[ b = (r_T - r_b)^2 \]
\[ c = 2r_b(r_T - r_b) - 2gH/\omega_o^2 \]
\[ d = r_b^2 - gH/\omega_o^2 - \frac{1}{3}(r_T^2 + r_b^2 + r_T r_b) \]
\[ e = \frac{\omega_o^2}{2nH} \]

solving for \( U_o \):

\[ U_o = -\frac{n_1}{2n_2} + \frac{1}{2n_2} a n_2(1-\epsilon)(C^2-4bd) + n_1^2 + 4n_2(\Delta p_T - e) \]

\[ -4n_2ab(1-\epsilon)[z_1 + \frac{c}{2b}]^2 \]

Then:

\[ \dot{v} = \int_0^1 2\pi r_o U_o H dz_1 \]  \hspace{1cm} (16)

Finally:

\[ \dot{v} = -\frac{c_1}{2c_2} + \left[ \frac{2(r_T - r_b)}{r_T + r_b} \right] \left[ \frac{ab(1-\epsilon)/C_2}{1/2} \right] \{ S(1+G) - S(G) \} \]  \hspace{1cm} (17)

where

\[ G = c/2b \]
\[ F_1 = \left[ \frac{c^2 - 4bd}{4b^2} + \frac{c_1^2 + 4c_2(\Delta p_T - e)}{4c_2 ab(1-\epsilon)} \right]^{1/2} \]

\[ S(x) = -(F_1^2 - x^2)^{3/2} + \frac{6gCH}{(r_T - r_b)^2 \omega_o^2} \]

\[ [x(F_1^2 - x^2)^{1/2} + F_1^2 \text{Arc sin}(\frac{x}{F_1})] \]

A different result is obtained by introducing equation (13) for grid pressure drop and equation (14) into
equation (11) as:

\[ \Delta P_T = \gamma U_o^{3/2} + a(1-\epsilon)(bz_1^2 + cz_1 + d) + e \]  \hspace{1cm} (18)

Solving for \( U_o \):

\[ U_o = \left( \frac{1}{\gamma} \left[ (\Delta P_T - e) - a(1-\epsilon)(bz_1^2 + cz_1 + d) \right] \right)^{2/3} \]  \hspace{1cm} (19)

Introducing equation (19) into equation (16) yields:

\[ \dot{\nu} = \dot{\nu}_K F_2 \left( \frac{r_T-r_b}{r_T+r_b} \right) \left\{ -\frac{3}{5} F_2 \left[ 1-\left( \frac{1+G}{F_2} \right)^2 \right]^{5/3} \right. \\
\left. - \left[ 1-\left( \frac{G}{F_2} \right)^2 \right]^{5/3} \right\} + 2 \left( \frac{r_b}{r_T-r_b} - G \right) \left[ T(A_1) - T(A_o) \right] \]  \hspace{1cm} (20)

where

\[ \dot{\nu}_K = \left\{ \frac{ab(1-\epsilon)(c^2-4bd)}{4b^2 + \Delta P_T - e} / K \right\}^{2/3} \]

\[ F_2 = \left( \frac{c^2-4bd}{4b^2} + \frac{\Delta P_T - e}{ab(1-\epsilon)} \right)^{1/2} \]

\[ A_o = \text{Arc sin} \left( \frac{G}{F_2} \right) \]

\[ A_1 = \text{Arc sin} \left( \frac{1+G}{F_2} \right) \]

\[ T(x) = 0.46917 x + 0.25383 \sin(2x) + 0.002 \sin^2(2x) \]  \hspace{1cm} (2x)

Substituting equations (17) and (20) into equation (11), the batch bed pressure drop may be determined.

A digital computer solution for fluidized bed pressure drop and air flow rate was written which follows the logic path given below:

(1) Read in the grid pressure drop coefficients.
(2) The computer solves for the overall air flow rate using equations (17) and (20) (two different profiles for grid pressure drop), for several values of total pressure drop.

(3) Using equation (6) and the corresponding grid pressure drop profile, the computer calculates the fluidized bed pressure drop for each value of total pressure drop. This program is given in Appendix A.

G. Comparison of experimental and theoretical results

Detailed batch studies of minimum fluidization in a fluidized bed were made by N. Martin [7]. Since batch operation is of secondary concern in this thesis, only a few batch data were obtained. These data are compared to the theory in Figures 13 to 18. Martin's program was also used to calculate bed pressure drop and gives essentially the same results for $\Delta p_{BED}$ in the fluidized region. In these figures, solid curves resulting from the theoretical model are based on $\epsilon=0.42$. In almost all cases, the theoretical bed pressure drop over-predicts the experimental results. This type of behavior would occur if the actual bed mass or angular velocity were less than the values used in the analysis. Additional studies are needed to determine whether this over-prediction is caused by a systematic error in the experiment or an error in the model. It is interesting
Solid Lines: Analytical solutions

\[ m_{\text{BED}} = 0.91 \text{ Kg} \]
\[ \bar{d} = 427 \times 10^{-6} \text{ m} \]
\[ \alpha = 4^\circ \]
\[ \omega (\text{rad/s}) \]

\[ \Delta 41.89 \]
\[ \bullet 36.65 \]
\[ \nabla 31.42 \]
\[ \bullet 26.18 \]
\[ \square 20.94 \]

---

Figure 13. Effect of angular velocity on fluidization.
Solid lines: Analytical solutions

\[ m_{\text{BED}} = 1.361 \text{ kg} \]
\[ \bar{d}_p = 427 \times 10^{-6} \text{ m} \]
\[ \alpha = 4^\circ \]

\[ \omega (\text{rad/s}) \]
- 36.65
- 31.42
- 26.18
- 20.94

Figure 14. Effect of angular velocity on fluidization.
Solid Lines: Analytical Solutions

\[ m_{BED} = 1.81 \text{ Kg} \]
\[ d_p = 427 \times 10^{-6} \text{ m} \]
\[ \alpha = 4^\circ \]

\[ \omega (\text{rad}/s) \]
- ▲ 41.89
- ● 36.65
- ▼ 31.42
- ○ 26.18
- ■ 20.94

Figure 15. Effect of angular velocity on fluidization.
Figure 16. Effect of angular velocity on fluidization.

Solid lines: Analytical solutions

- $m_{BED} = 2.72$ Kg
- $d_p = 427 \times 10^{-6}$ m
- $\alpha = 4^\circ$

$\omega$ (rad/s)

- $\triangle$ 41.89
- $\bullet$ 36.65
- $\nabla$ 31.42
- $\bullet$ 26.18
- $\blacksquare$ 20.94

Air Flow ($m^3/s$)

$\Delta p_{BED}$ (Pa)
Solid lines: Analytical solution

- \( m_{\text{BED}} = 3.63 \text{ Kg} \)
- \( d_p = 4.27 \times 10^{-6} \text{ m} \)
- \( \alpha = 4^\circ \)

\( \omega (\text{rad/s}) \)
- \( \Delta \) 41.89
- \( \bullet \) 36.65
- \( \nabla \) 31.42
- \( \bullet \) 26.18
- \( \square \) 20.94

Figure 17. Effect of angular velocity on fluidization.
Figure 18. Effect of angular velocity on fluidization.
that the analytical results show that the fluidized bed void fraction has no significant effect on fluidized bed pressure drop (Figure 19). In addition, the computer program was used to calculate bed pressure drop with five percent less bed mass than recorded in the laboratory. This calculation revealed a predicted bed pressure drop about five percent less. This is expected because of the linear relationship between $\Delta p_{BED}$ and bed mass.

It should be noted that the liquid model for predicting bed pressure drop requires the assumption of solid body motion for the tangential component of velocity in the bed. Using an uniform tangential or any other profile gives an obvious conflict with the liquid model used to predict bed shape (see Section D).
$m_{BED} = 1.361 \text{ Kg}$

$\omega_0 = 36.65 \text{ rad/s}$

$\frac{d}{dp} = 427 \times 10^{-6}$

$a = 4^\circ$

---

**Figure 19.** Effect of voidage on bed pressure drop.
IV. MECHANICS OF FEEDING AND REMOVAL OF SOLID PARTICLES
WITH A CENTRIFUGAL FLUIDIZED BED.

A. Overview

The constraints affecting the feed and removal of solid particles to a centrifugal fluidized bed were investigated. In this section, a simple particle trajectory analysis is described. The results of the analysis are used to aid in the design of the feed injector. Although sufficient experimental data are not available to validate this theory, qualitative laboratory observations on feeding are presented. The experiments on solids removal are discussed and some recommendations are offered to modify the existing system.

B. Two dimensional model of the dynamics of a single particle in a centrifugal flow field

The general expression for the steady-state force exerted on a submerged sphere by a flowing fluid is [8]:

$$F_D = \frac{1}{2} \rho_f \frac{d_p^2}{4}(\frac{d_p}{2})c_D \bar{V}^2$$

(1)

where the drag coefficient $c_D$, is a function of the particle Reynolds number and is given by White [5]:

$$c_D = \frac{24}{Re} + \frac{6}{1 + \sqrt{Re}} + 0.4$$

(2)

where:

$$Re = \frac{\rho_f d_p \bar{V}}{\mu_f}$$

(3)
The direction of the drag force acting on each particle is assumed always to be in the direction of the relative velocity vector. The magnitude and direction of the drag force components are then obtained by multiplying the total force by \((u_r + V_r)/|\mathbf{v}|\) and \((u_t - V_t)|\mathbf{v}|\) for the radial and transverse directions, respectively.

Forces acting radially outward and in a counterclockwise direction are arbitrarily assumed positive and the direction of rotation of the bed is assumed counterclockwise throughout.

The following assumptions also are made:

1. The presence of the particles does not alter the velocity field of the fluid.
2. The particle is a non-deformable sphere.
3. The level of macroscopic turbulence in the fluid is negligible.
4. Wall effects are neglected.
5. The effects of Brownian motion are negligible.
6. Particles do not interact with one another.
7. No velocity gradient in the vertical direction is considered.
8. Magnus forces are negligible.
9. Body forces are negligible with respect to the centrifugal force.
Newton's second law of motion gives:

\[ \mathbf{F} = m_p \mathbf{a} \]

In cylindrical coordinates \((r-\theta)\), we have:

\[ \mathbf{F}_r = m_p a_r = \rho_s \left( \frac{1}{6} \pi d_p^3 \right) (\ddot{r} - r\dot{\theta}^2) \]

\[ \mathbf{F}_\theta = m_p a_\theta = \rho_s \left( \frac{1}{6} \pi d_p^3 \right) (r\ddot{\theta} + 2\dot{r}\dot{\theta}) \]

Since the ratio \( \rho_s/\rho_f > 1 \), the equation of motion reduces to:

\[ -(\pi d_p^2/4)(\rho_f/2)C_D V^2 \left[ \frac{U_r + V_r}{\rho} \right] - \rho_s \left( \frac{1}{6} \pi d_p^3 \right) \left( \frac{dV_r}{dt} - \frac{V_r^2}{r} \right) \]

\[ (\pi d_p^2/4)(\rho_f/2)C_D V^2 \left[ \frac{U_t - V_t}{\rho} \right] - \rho_s \left( \frac{1}{6} \pi d_p^3 \right) \left( r \frac{d\theta}{dt} \right) \]

\[ + 2V_r \frac{d\theta}{dt} \]

or:

\[ \frac{dV_r}{dt} = \frac{V_t^2}{r} - a C_D \bar{V}(U_r + V_r) \]

\[ \frac{dV_t}{dt} = -\frac{V_t V_r}{r} + a C_D \bar{V}(U_t - V_t) \]

where:

\[ a = 3\rho_f/4\rho_s d_p \]

These equations can be expressed as

\[ \frac{dV_r}{dt} = \frac{V_t^2}{r} - a C_D (U_r + V_r) [(U_r + V_r)^2 + (U_t - V_t)^2]^{1/2} \] (4)

\[ \frac{dV_t}{dt} = -\frac{V_t V_r}{r} + a C_D (U_t - V_t) [(U_r + V_r)^2 + (U_t - V_t)^2]^{1/2} \] (5)

The radial and tangential variations of the particle velocities depend upon the distribution of radial
and tangential velocities of the air throughout the rotating chamber.

Assuming the boundary layers on the top and bottom end walls of the rotating chamber are thin and the chamber is free of significant secondary flow patterns, the radial velocity of the gas in the freeboard region may be evaluated from the continuity equation:

$$u_r = \frac{\dot{m}_f}{2\pi \rho_f H r} \quad (6)$$

The effect of operating conditions on the radial variations of the tangential velocity of the fluid have been investigated [9]. It has been shown that a free vortex ($u_t = \omega_o r_o^2/r$) develops over most of the chamber except within the viscous core region at the axis of rotation where the fluid tends towards solid body rotation ($u_t = \omega_o r$). If the particle is fed from the region outside the viscous core, then the radial variation of the tangential velocity component of the fluid may be evaluated using the free vortex assumption:

$$u_t = \omega_o r_o^2/r \quad (7)$$

Substituting equations (6) and (7) into equations (4) and (5), we have:

$$\frac{dV_r}{dt} = \frac{V_t^2}{r} - \alpha C_D \left( \frac{\dot{m}_f}{2\pi \rho_f H r} + V_t \right) \left( \frac{\dot{m}_f}{2\pi \rho_f H r} + V_t \right)^2$$

$$+ \left( \frac{\omega_o r_o^2}{r} - V_t \right)^2 \right)^{1/2} \quad (8)$$
\[
\frac{dV_t}{dt} = - \frac{V_r V_t}{r} + aC_D \left( \frac{\omega_o r_o^2}{r} - V_t \right) \left( \frac{\dot{m}_f}{2 \pi \rho_f H r} + V_r \right)^2 \\
\quad + \left( \frac{\omega_o r_o^2}{r} - V_t \right)^2)^{1/2}, \quad (9)
\]

A digital computer analysis was developed to solve equations (8) and (9) for several values of the following system parameters:

- Air flow rate
- Angular velocity of the chamber, \( \omega_o \)
- Initial velocity of the particle, \( V_o \)
- Initial injection direction of particle, \( \theta_o \)
- Initial radial distance from axis of rotation, \( R_o \)

This program is reproduced in Appendix B. The average velocity of particles in the exit section of the feeding nozzle may be estimated from:

\[
V_s = \frac{4 \dot{m}_s}{\pi D_N^2 (1 - c_N)(\rho_s - \rho_f)} \quad (10)
\]

This relation is derived based on a uniform voidage assumption inside the conveying line. The voidage in the feeding line \( c_N \) may be evaluated from:

\[
c_N = \frac{\dot{V}_F}{\rho_f \phi_f + \rho_s \dot{m}_s} \quad (11)
\]
Introducing equation (11) into the equation (10) gives:

\[ V_s = \frac{4V_F}{\pi D_N^2}(\rho_s - \rho_f) \]  (12)

The average velocity of particles is affected only by the air flow rate through the feeding line, because the voidage in the transporting line is very close to unity. The solid feed rate was found to be independent of the air flow rate through the feed line, and the air flow rate was set at 0.0011884 SCMS for all the experiments. For feeding rates ranging between 0.01134 and 0.02268 Kg/s, the average velocity of particles at the exit section of the feeding nozzle using equation (12) was found to be:

\[ V_s = 16.76 \text{ m/s} \]

C. Results of feed analysis

In Figures 20, 21, 22, 25 the radial distance of the projected particle is plotted against time for several values of system parameters. The results may be summarized as follows:

(1) The particle residence time, PRT, is not highly affected by air flow rate and angular velocity of the rotating chamber (Figure 20).

(2) Initial velocity of the particle has a strong effect on PRT, but it seems that there is no problem in feeding the particle at initial velocities equal to 16.76 m/s (Figure 21).
Figure 20. Effect of bed angular velocity on particle residence time at low and high air flow rates.

\[ V_0 = 7.62 \frac{m}{sec} \]
\[ \theta_0 = 90^\circ \]
\[ d_p = 427 \times 10^{-6} \text{ m} \]
\[ R_o = 0.0762 \text{ m} \]
Figure 21. Effect of the initial velocity of particle on particle residence time.

Curves $V_o$ (m/s)

<table>
<thead>
<tr>
<th>Curve</th>
<th>$V_o$ (m/s)</th>
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<tbody>
<tr>
<td>1</td>
<td>3.05</td>
</tr>
<tr>
<td>2</td>
<td>4.57</td>
</tr>
<tr>
<td>3</td>
<td>6.10</td>
</tr>
<tr>
<td>4</td>
<td>7.62</td>
</tr>
<tr>
<td>5</td>
<td>9.14</td>
</tr>
</tbody>
</table>

$v_a = 0.354 \text{ m}^3/\text{sec}$

$\omega_o = 36.7 \text{ rad/sec}$

$d_p = 427 \times 10^{-6} \text{ m}$

$\theta_o = 90^\circ$ (Tangentially)

$R_o = 0.0762 \text{ m}$
(3) The PRT is a strong function of the direction of injection of the particle. The best injection angle providing the minimum PRT, is $0^\circ$ (particle injected radially toward the bed), Figure 22.

(4) The direction of injection is an important parameter in particle feeding. Significant particle entrainment from the feed jet was observed even at relatively high angular velocities. The jet of particles and air spreads as shown in Figure 23. Some of the particles flow to radii smaller than $r_e$ and may be carried from the system, depending upon the axial velocity of the gas and the vertical location of the injector inside the rotating chamber. In Figure 24, the path of the particle inside the rotating chamber is plotted in polar coordinates for several values of system parameters. The calculated particle residence time in the exhaust port section is relatively small (about 0.004 seconds) for an injection direction equal to $120^\circ$ (see Figure 25).

(5) The effects of initial injection direction on the PRT when the particle is injected $R_0 = 0.0127$ m from the axis of rotation, are shown in Figure 26 for two different tangential velocity profiles. For these assumed system parameters, it seems that
Figure 22. Effect of initial injection direction of particle on particle residence time.
Figure 23. Geometry of feeding system.
\[ \dot{v}_a = 0.354 \text{ SCMS} \]
\[ \omega_0 = 36.65 \text{ rad/s} \]
\[ d_p = 4.27 \times 10^{-6} \text{ m} \]
\[ v_0 = 7.62 \text{ m/s} \]

Figure 24. Effect of position of injection on particle trajectory.
Figure 25. Effect of initial injection direction of particle on particle residence time.

\[ v_s = 0.354 \, \text{m}^3/\text{s} \]
\[ \omega_o = 36.7 \, \text{rad/s} \]
\[ V_o = 16.76 \, \text{m/s} \]
\[ d_p = 427 \times 10^{-6} \, \text{m} \]
\[ R_o = 0.0762 \, \text{m} \]
Figure 26. Effect of initial injection direction of particles on particle residence time.

<table>
<thead>
<tr>
<th>Curves</th>
<th>$\theta_0$</th>
</tr>
</thead>
<tbody>
<tr>
<td>1,2</td>
<td>0°</td>
</tr>
<tr>
<td>3</td>
<td>90°</td>
</tr>
<tr>
<td>4</td>
<td>90°</td>
</tr>
</tbody>
</table>

Solid body assumption, $u = \omega r$, Curves 2, 4

Free vortex assumption, $u = \omega o r^2 / r$.

Curves 1, 3

$\dot{V}_a = 0.354 \text{ m}^3/\text{s}$

$\omega_0 = 26.2 \text{ rad/s}$

$V_0 = 16.76 \text{ m/s}$

$d_p = 427 \times 10^{-6} \text{ m}$

$R_o = 0.0127 \text{ m}$
the PRT is not highly affected by initial injection direction of the particle.

D. Particle removal system

In this section the effects of solids discharge rate and geometry of the discharging parts on bed characteristics are described. Experiments were performed to measure bed pressure drop, minimum fluidization velocity and bed mass as functions of bed angular velocity, air flow rate, solids discharge rate and geometry of the discharging holes. All the experiments were performed using technical quality glass particles with mean particle size:

$$d_p = 427 \times 10^{-6} \text{ m}$$

E. Experiments on bed start up

The capability of start up with continuous solids feed from an empty rotating chamber to a stable steady state fluidized bed is of prime importance. The following experimental steps were taken to show the feasibility of a steady and stable solids feed and removal operation with a CFB system:

1. Set the bed angular velocity at the desired rpm.
2. Adjust the air flow rate.
3. Adjust the feeding system for the desired solids feed rate and initiate solids flow at time t=0.
4. Record the total pressure drop every 30 seconds, until the system pressure drop stabilizes.
(5) Turn the air flow off immediately after turning off the particles feed system.

(6) Stop the rotating chamber, collect the particles from the collecting bin and weigh.

In Figure 27, the effect of particle discharge rate on the dynamics of bed start up is shown. The system reached a steady state with solids feed and removal after about 120 to 180 seconds, depending upon the particle discharge rates. The time \( \tilde{t} \) required for the system to reach a fully fluidized state with stable solids feed and removal depends upon various system parameters. In Figure 28, the effect of solid discharge rate on time \( \tilde{t} \) is plotted using the data from Figure 27.

In Figure 29, the effect of air flow rate on bed start up with continuous solids feed and removal is shown. The effect of bed angular velocity on the bed start up is shown in Figure 30, for two different angular velocities.

F. Effects of system parameters on bed characteristics with stable solids feed and removal

The effect of air flow rate on bed pressure drop is given in Figure 31 for constant angular velocity of the bed. The solid marks indicate the data taken from the batch experiments at constant bed mass. It is important to note that the data points indicated with the open symbols belong to the overflow experiments. Except
Figure 27. Effect of solids discharge rate on bed start up.
Figure 28. Effect of discharge rate on start up time.
Figure 29. Effect of air flow rate on bed start up.
Figure 30. Effect of bed angular velocity on bed start up.

Opening = 25°

SDR = 0.01606 Kg/s

\( \dot{a} = 0.2667 \text{ SCMS} \)

\( \bar{d}_p = 427 \times 10^{-6} \text{ m} \)

<table>
<thead>
<tr>
<th>( \omega_0 ) (rad/s)</th>
<th>31.42</th>
<th>36.65</th>
</tr>
</thead>
</table>

Bed pressure drop (Pa)

Time (secs)
in the packed regime, each of the overflow data points corresponds to a different bed mass. When the bed is packed, the bed pressure drop seems to increase with air flow rate. However, beyond minimum fluidization, the bed pressure drop decreases with air flow rate. This behavior is due to bubbling with the resultant increase in bed voidage. The bed pressure drop and bed mass decrease at relatively high air flow rates while the bed is fully fluidized. Note that for the first two points of this data set, the bed was packed and the maximum bed pressure drop was obtained at minimum fluidization. For the rest of the data points, the bed was either partially or fully fluidized. It is interesting to observe that these data points are located in the region bounded by 3.63 Kg and 1.81 Kg batch data, and the corresponding overflow bed masses lie between these two limits. In Figure 32, variations of the bed pressure drop are plotted versus air flow rate at 36.65 rad/s, for several values of solids discharge rates and discharge orifice areas. These data also exhibit the same behavior described above. With smaller discharge orifices and higher solids feed rates, the bed is thicker with a larger bed pressure drop. The effects of air flow rate on bed mass are plotted in Figure 33. At increasing air flow rates, the bed mass decreases and tends to level out (still more data are needed to validate
Figure 31. Effect of air flow rate on bed pressure drop.
Figure 32. Effect of solids discharge rate and discharging orifice area on bed pressure drop.
Figure 33. Effect of air flow rate on bed mass.

- \( \omega_0 = 36.65 \text{ rad/s} \)
- \( \frac{d}{p} = 427 \times 10^{-6} \text{ m} \)

<table>
<thead>
<tr>
<th>Opening (Degree)</th>
<th>SDR</th>
</tr>
</thead>
<tbody>
<tr>
<td>25</td>
<td>0.01438</td>
</tr>
<tr>
<td>25</td>
<td>0.01324</td>
</tr>
<tr>
<td>5</td>
<td>0.02194</td>
</tr>
</tbody>
</table>
this behavior) at high flow rates where the bed is completely fluidized. The transition band at which the bed progresses from a packed to a fully fluidized bed condition lies between about 0.189 to 0.307 m$^3$/s for all three data sets (Figure 32). In Figure 34, the effects of the bed mass on bed pressure drop are shown for the same angular velocity and for several values of solids discharge rate. The solid symbols again represent data taken from batch experiments. At low bed masses (high air flow rate with bed fully fluidized), the batch and overflow data points merge and are consistent with the theory (equation III.D.3). This confirms that when the bed is fully fluidized, the bed pressure drop is only a function of bed mass and angular velocity. The effect of solids discharge rate (SDR) on bed behavior is plotted at constant angular velocity and air flow rate in Figures 35 and 36. These results show very little effect of particle discharge rate on the bed mass and pressure drop. Note, however, that the range of the SDR used in the experiments is relatively small.
\( \bar{d}_p = 427 \times 10^{-6} \text{ m} \)
\( \omega_0 = 36.65 \text{ rad/sec} \)

<table>
<thead>
<tr>
<th>Opening</th>
<th>SDR (Kg/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>25</td>
<td>0.01324</td>
</tr>
<tr>
<td>5</td>
<td>0.02194</td>
</tr>
<tr>
<td>25</td>
<td>0.01438</td>
</tr>
</tbody>
</table>

Figure 34. Effect of bed mass on bed pressure drop.
Figure 35. Effect of particle discharge rate on bed pressure drop at steady state operating conditions.
Figure 36. Effect of particle discharge rate on bed mass at steady state operating conditions.

\[ \omega_0 = 36.65 \text{ rad/s} \]

\[ \dot{v}_a = 0.39 \text{ Kg/s} \]

\[ \bar{d}_p = 427 \times 10^{-6} \text{ m} \]

Opening = 25°
V. CONCLUSIONS

As a first approximation, the batch bed pressure drop may be calculated using Equation (3) Section III.D. It is shown that neither this simple equation nor the analytic solutions (presented in Section III.F) compare well with the data taken from the batch experiments. All the models over-predict the fluidized bed pressure drop significantly. This difference may be due to a systematic error in the experiments (for example, bed mass and bed angular velocity measurements) or an error in the theoretical model. Still further investigations are needed to reveal the cause of this discrepancy.

The solid feed angle is an important parameter in the design of a proper feeding system. The original design of the solids feed nozzle was such that the particles were injected in the tangential direction. Particle entrainment from the feed jet was observed in all experiments even at high bed angular velocities. The probable cause of the entrainment is improper orientation of the injector. The jet of particles and air spreads and some of the particles flow to radii smaller than \( r_e \) and are carried from the system by the large axial velocities in this region. With a 120° initial injection direction, the particle passes through the exhaust port region before flowing to the bed. During this short interval of time, if the air flow rate is
high enough, the particle may be carried out by the high axial velocity of the exhausting air. More work is needed on the particle feed phenomenon. In particular, the effect of a relatively thick bed on the feeding characteristics should be investigated.

The bed feed and particle removal systems are described and experiments on continuous particle feed and removal were performed. It is shown that the bed characteristics with PF&R are very close to the batch bed behavior. Any analytical approach to the solids removal problem is quite difficult. As a start, the simple liquid model should be adapted to analyze overflow behavior. To do this, information is needed on the discharge coefficients of the removal orifices. Several hypotheses are made in the previous section concerning the observed bed behavior, and now more experiments are needed to understand completely the effect of bed rotation on solids discharge. Following are some important observations and suggestions.

1. Because of some technical difficulties in proper operation of the system, more accurate experiments should be performed to measure bed masses in the overflow experiments.

2. Improper seal design between the air plenum and particle collecting chamber caused some difficulties with solids discharge. It was observed that the
bed did not achieve a stable thickness when the
discharge areas were below a minimum size. This
effect could be due to air flow from the stationary
air plenum to the collecting chamber and then upward
to the bed region through the discharge openings.
This would act to restrict a downward solids flow
through the discharge orifices. Less difficulties
were observed at high air flow rates, probably
because the seal works better at these conditions
(higher plenum pressure on the seal). Experiments
were performed which confirm the presence of air
leakage across the bottom seal.

(3) More experimental data should be taken with a wider
range of system parameters, especially location and
ter of discharge holes. Some investigations
should also be done with relatively thick beds.
Bubbling in the centrifugal fluidized bed should
be studied. The effect of air flow rate and other
system parameters on fluidized bed void fraction
should be investigated both analytically and ex-
perimentally. The particle mixing phenomenon in
the bed is another realm of investigation to be
explored and understood so that the sorbent waste
and unburned carbon can be minimized.
REFERENCES


Appendix A

Digital Computer Programs for Determination of Fluidized Bed Pressure Drop

```
PROGRAM ANAPD (INPUT, OUTPUT, TAPE5=INPUT, TAPE6=OUTPUT)
SUM(XX)=-(1.73)*((F2**2-XX**2)**1.5)+{(6.32.174*H)
1/((RT-RB)*OMEGA)**2)*XX*SQRT(F2**2-XX**2)+(F2**2)*
2ASIN(XX/F2))
RHO=154,
RHOF=0.075
H=6,
RT=5.875
RB=9.475
B=(RT-RB)**2
PRINT 7
7 FORMAT (////)///

NOTE THAT THIS PROGRAM ASSUMES THAT THE BED IS UNIFORMLY
FLUIDIZED EVERYWHERE THROUGH THE BED.
ALSO NOTE THAT ALL THE PARAMETERS ARE IN ENGLISH SYSTEM OF UNITS.
```

READ 8,RPM,EPS
8 FORMAT(F10.2,F10.3)
READ 81,G1,G2
81 FORMAT(2F10.3)
DO 143 J=1,10
READ 82,BMASS
IF(EOF,5)143,83
82 FORMAT(F10.3)
83 PRINT 9
9 FORMAT(/,16X,3HRPM,6X,8WBEDED,5X,4HSCFM,5X,7HTOTAL P,3X,6MGRID P,5X,2HG1,
17X,2HG2)
OMEGA=1.141599*RPM/30,
A=0.5*(OMEGA**2)*(R/OS-RHOF)/(24102.024)
C=2.*R9*(RT-RB)-2.*12.*32.174*H/(OMEGA**2)
D=R3**2+((32.174*12.174*H/(OMEGA**2))-1RT**2*40**2+RT*R8)/3.
E=(BMASS*(OMEGA**2))/(17,34844*H)
F=(C**2-4.*B**2)/(4.*B**2)
G=C/(2.*B)
UPHG=E+(A*D*(1.-EPS))-(G1**2/(4.*G2))
Appendix A (Continued)

```
DPM1G = DPMG + A*(B+G)*(1.-EPS)
OMEG = SQRT(2.32,175,12.)*(RT**2-RB**2)
IF(OMEGA.LE.0.OMEG) DPTPA = DPMG
IF(OMEGA.GT.OMEG) DPTPA = DPM1G
DPTPA = DPTPA + 0.082
DPTPA = DPTPA/500.
10 P = 0
PRINT 90,RPM,3MASS, EPS, DPTPA, G1, G2
90 FORMAT(15X,F4.0,5X,F5.1,15X,F10.5,10X,F9.1,11X,F10.7,F11.8)
DO 14 I = 1,41
91 DPT = (DPTPA+ (I-1)*500.)/249.082
X = ((G1**2) + 4.*G2*DPT-E)/(4.*G2*A*B*(1.-EPS))
F1 = F + X
IF(F1.LE.0.0) GO TO 131
F2 = SQRT(F1)
IF(F2.LE.0.0) GO TO 132
IF(I.I.EQ.21) GO TO 143
AS = ABS(G2/F2)
AS1 = ABS((1.+G)/F2)
IF(AS.LE.1.0 .AND. AS1.LE.1.0) GO TO 92
I = I + 1
DPTPA = DPTPA + 0.01*DPTPA
GO TO 91
131 CONTINUE
SQ = F2**2 - G**2
SQ1 = F2**2 - (1.+G)**2
IF(SQ.GE.0.0 .AND. SQ1.GE.0.0) GO TO 93
I = I + 1
DPTPA = DPTPA + 0.01*DPTPA
GO TO 91
132 CONTINUE
EE = (12.*(RT-H3))/(RT+R8) * ABS(SQRT((A*B*(1.-EPS))/G2))
ED = G1/(2.*G2)
SCFM = ED * ABS(E*SUM(1.+G) - SUM(G))
DPT = G1 * SCFM * G2 * (SCF**2)
DPT8D = DPT * DPT
```

On Feb 74 8 15:56 03/22/78
DPT = DPT * 249.082
DPT8D = DPT * 249.082
Appendix A (Continued)

```
DPED=DPED*2!9.082
PRINT 13,RPM,8MASS,DPT,DPG,G1,G2
13 FORMAT (15X,F4.0,5X,F5.1,5X,F6.1,4X,F10.5,F9.1,1X,F9.1,1X,
1F9.1,1X,F10.7,F11.8)
IF (SCFM,GE,1000.) GO TO 141
IF(I,EQ,1) OPTPA=500.*IPTPA
IF(IPTPA,LT,0) OPTPA=500.
GO TO 14
141 PRINT 121,RPM,8MASS,DPT,F1
121 FORMAT(15X,F4.0,5X,F5.1,13X,22HF FOR DPT EQUAL TO THIS..F9.1,5X,
17HF1=F*X=,F15.5)
GO TO 14
142 PRINT 122,RPM,8MASS,DPT,F2
122 FORMAT(15X,F4.0,5X,F5.1,13X,22HF FOR DPT EQUAL TO THIS..F9.1,5X,
113HF2=SQRT{F*X}=,F10.5)
GO TO 14
14 CONTINUE
143 PRINT 15
15 FORMAT(//////////)
STOP
END
```
Appendix A (Continued)

PROGRAM ANAPD (INPUT,OUTPUT,TAPES=INPUT,TAPE6=OUTPUT)
RHOS=154.
RHOF=0.075
H=6.
RT=5.375
RB=5.4375
3*(RT-RB)**2
PRINT 7
7 FORMAT (///////)

NOTE THAT THIS PROGRAM ASSUMES THAT THE BED IS UNIFORMALLY
FLUIDIZED EVERYWHERE THROUGH THE BED.
ALSO NOTE THAT ALL THE PARAMETERS ARE IN ENGLISH SYSTEM OF UNITS.

READ 9,RPM,EFS
8 FORMAT(F10.2,F10.3)
READ 81,CC
81 FORMAT(F10.6)
DO 140 J=1,10
READ 82,EMASS
IF (EOF,5) 140,83
82 FORMAT(F10.3)
83 PRINT 9
9 FORMAT (///,16X,3HRPM,6X,6HBED MASS,4X,5HDPBED,5X,
17HEPSILON,5X,4HSCFM,5X,7HTOTAL P,3X,6HGRIO P,1
14X,2HCC)
OMEGA=3.1415*RPM/30.
A=0.5*(OMEGA**2)*(RHOS-RHOF)/(24102.024)
C=2.*RB*(RT-RB)-(2.*12.*32.174*H)/(OMEGA**2)
D=32.174*(32.174*12.*H)/(OMEGA**2)-(RT**2+RB**2+RT*RB)/3.
E=(EMASS*(OMEGA**2))/(87.634844*H)
F=(C**2-4.*B*D)/(4.*B**2)
G=C/(2.*B)
DO 14 I=1,41
Appendix A (Continued)

```plaintext
DPT=(I/500.)/249.082
G0=(A*B*(1.-EPS)*F*DPT-E)/CC)**(0.666667)
X=(DPT-E)/(A*B*(1.-EPS))
F1=F*X
IF (F1.LE.0.0) GO TO 14
F2=SORT(F1)
IF (F2.LE.0.0) GO TO 14
G1=A3S((1.+G)/SORT(F1))
IF (G1.GT.1.) GO TO 14
G4=A3S(G/SORT(F1))
IF (G4.GT.1.) GO TO 14
A1=ASIN((1.+G)/SORT(F+X))
A0=ASIN(G/SORT(F+X))
SI1=0.46917*A1+0.25383*SIN(2.*A1)+0.002*(SIN(2.*A1))**3+0.0061*
1*SIN(4.*A1)+0.000086*SIN(8.*A1)
SI0=0.46917*A0+0.25383*SIN(2.*A0)+0.002*(SIN(2.*A0))**3+0.0061*
1*SIN(4.*A0)+0.000086*SIN(8.*A0)
CP=1./F2
SS=((2.*CO)/CP)*((RT-R8)/(RT+R8))*((R8/(RT-R8))-G)*((SI1-SI0)
Q1=(1.-((CP)**7)*((1.+G)**2))**(5./3.)
Q0=(1.-((CP)**2))**(5./3.)
SCFF=(1./5.)*(RT-RF)/(RT+R8)*((CO/(CP**2))*Q1-Q0)
SCFM=SS-SCFF
DPG=CC**((SCFM)**(1.5))
DP3ED=DPT-DE
DE=DE+1000.
DL=DE+1300.
PRINT 13,RF,MASS,TPBEO,EPS,SCFM,DPT,DPG,CC
13 FORMAT (15X,F4.0,5X,F5.1,5X,F6.1,4X,F10.5,F9.1,1X,F9.1,1X,
1F9.1,1X,F10.7)
IF (SCFM.GT.1000.) GO TO 140
CONTINUE
140 PRINT 15
15 FORMAT(/)STOP
END
```
Appendix B


SUBROUTINE INITIAL
COMMON/T/T,NFIN,NRUN/Y/Y(4)/F/F(4)/PARAM/NPAGE,NLINE
COMMON/Z/Z(5)/AREA/AF,W,DP,V0,TTA
AF=750.
W=250.
TTA=45.
DP=427.
V0=55.
VPR=V0*COS(3.14159*TTA/180.)
VPT=V0*SIN(3.14159*TTA/180.)
NPAGE=1
NLINC=0
Y(1)=VPR
Y(2)=VPT
Y(3)=1./24.
Y(4)=0.
Z(1)=(0.005233*AF)/Y(3)
Z(2)=(0.02618*W)/Y(3)
Z(3)=SQRT((Z(1)+Y(1))**2.0+(Z(2)-Y(2))**2.0)
Z(4)=0.0195288*DP*Z(3)
Z(5)=(24./Z(4))+(6./(1.+SQRT(Z(4))))*0.4
RETURN
END
SUBROUTINE DERV
COMMON/T/T,NFIN, NRUN/ Y/Y(4)/F/F(4)/PARM/NPAGE, NLINE
COMMON/Z/Z(5)/AREA/AF, W, DP, VO, TTA

A = 111.33 /DP
Z(1) = (0.005233*AF) / Y(3)
Z(2) = (0.0261*A*W) / Y(3)
Z(3) = SQRT(z(1) + y(1))**2.0 + (z(2) - y(2))**2.0
Z(4) = 0.01952 + DP*Z(3)
Z(5) = (2.0 + Z(4)) + (0.1 + SQRT(Z(4))) + 0.4
F(1) = (Y(2)**2.0 / Y(3)) - (A*Z(5)*(Z(1) + Y(1) + Z(3)))
F(2) = A*Z(5)*((Z(2) - Y(2)) + Z(3)) - (Y(1)*Y(2) / Y(3))
F(3) = Y(1)
F(4) = Y(2) / Y(3)
RETURN
END
SUBROUTINE PRINT(NI, NO)
COMMON/T,F,NPIN,NRUN/Y/Y(4)/F/F(4)/PARM/NPAGE,NLINE
COMMON/Z/Z(5)/AREA/AF,W,DP,VO,TTA
IF(NLINE GT 0) GO TO 1
WRITE(6,5)
FORMAT(1H1)
WRITE(6,2)NPAGE,AF,W,DP,VO,TTA
5 FORMAT(11X,5HPAGE ,I2, //,
1 61X,11HFLOW RATE = ,F6.0, 5H SCFM, //,
2 54X,18HANGULAR VELOCITY = ,F6.0, 4H RPM, //,
3 53X,19HPARTICLE DIAMETER = ,F6.0, 7H MICRON, //,
4 42X,30HINITIAL VELOCITY OF PARTICLE = ,F6.0,
5 7H FT/SEC, //,
6 31X,41HINITIAL INJECTION DIRECTION OF PARTICLE = ,F6.0,
7 7H DEGREE, //,/
8 26X,44TIME,4X,6HRADIUS,9X,6HDEGREE,7X,
9 3HVPR,10X,34VPT,10X,3HVGR,10X,3HVGT,9X,2HCD)
NLINE=NLINE+15
1 WRITE(6,3)T,Y(3),Y(4),Y(1),Y(2),Z(1),Z(2),Z(5)
3 FORMAT(20X,F10.4,F10.5,F15.5,F13.5,F11.4)
NLINE=NLINE+1
IF(NLINE LT 60) RETURN
NPAGE=NPAGE+1
NLINE=0
WRITE(6,4)
4 FORMAT(1H1)
RETURN
END
VITA

Ali Tabatabaie-Raissi, the son of Mahmoud and Habibeh Tabatabaie-Raissi, was born in Teheran, Iran, November 11, 1952. He achieved his bachelor of science degree in mechanical engineering from University of Teheran in 1975. During his undergraduate studies, he held scholarships for the years of 1973-1974 and 1974-1975. He was employed as a technical expert for the Industrial Credit Bank (1974) and by the Institute of Standards and Industrial Research of Iran in 1975.

He began graduate studies at Lehigh University in January 1976, in the areas of fluid mechanics, thermodynamics, and heat transfer, working as a research assistant with Drs. E. K. Levy and J. C. Chen on studies of fluidized beds. He also served as a teaching assistant with Dr. P. G. Kosky in the course "Introduction to Nuclear Engineering" (1977).

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The American Society of Heating, Refrigerating and Air Conditioning Engineers (Student Member)

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