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## ABSTRACT

# FLUIDIZATION AND SOOT FILTRATION IN A ROTATING FLUIDIZED BED by Gui-Hua Qian

A horizontal rotating fluidized bed (RFB) has been investigated for the removal of soot from diesel engine exhaust. This requires necessitating an understanding of the nature of fluidization, particle mixing and filtration in the RFB.

In an RFB, particles are subject to a centrifugal force which can be much larger than the force of gravity. Therefore the Geldart classification for conventional fluidized beds must be modified for rotating fluidized beds. A theoretical analysis shows that group A particles (minimum bubbling velocity  $U_{mb}$  > minimum fluidization velocity  $U_{mf}$ ) can shift to group B ( $U_{mb} = U_{mf}$ ), and group C particles (agglomerating) can shift to group A under a centrifugal force. Therefore, certain group C particles can be fluidized in rotating fluidized beds. However, for very high "g", such particles shift to group D (spouting) and cannot be fluidized. This was verified experimentally by successfully fluidizing 7 µm alumina particles in the RFB which behave as group C in a conventional fluidized bed. Thus an important and unique feature of RFB technology is that it enables the use of very fine bed particles in a fluidization mode.

Since the RFB works as a shallow bed, the distributor plays an especially important role in its fluidization behavior. The pressure drop in an RFB was measured using slotted, perforated and sintered metal cylindrical gas distributors as a function of rotating speed, gas velocity and bed thickness with both polydisperse alumina particles and nearly monodisperse glass beads. The measured pressure drop for the different distributors depends strongly on the distributor design. A theoretical model available in the literature is used to calculate the minimum fluidization velocity and the pressure drop as a function of rotating speed, mass loading and gas velocity which are then compared to the experimental results.

Particle motion in an RFB was studied by observing the mixing of two layers of particles of different color and different density. Experiments show that bubbles are responsible for particle motion and mixing for layers of the same material. When a layer of denser particles is placed on the distributor, the mixing behavior is similar to that observed for layers of the same material. However, when a layer of less dense particles is placed on the distributor, mixing is dominated by differences in density and occurs before bubbles are visible.

A horizontal rotating fluidized bed filter (RFBF) charged with either polydisperse alumina granules or nearly monodisperse glass beads, was used to capture soot from diesel engine exhaust. The mass average filtration efficiency, calculated on the basis of the total mass of soot that was captured in the bed, was found to be up to 90% at the start of filtration at steady state flow conditions. Time and equipment limitations did not permit obtaining data with a build up of soot in the bed or the exploration of the effect of varying engine load conditions. A critical factor in obtaining high filtration efficiency is the use of fine sized bed particles (high surface area/volume) which is enabled by the use of an RFB. The filtration efficiency increased with increasing gas flow rate as the bed passes from the packed bed to the fluidized bed mode. The filtration efficiency also varied as a function of agglomerated soot size, showing a minimum for soot particles in the 0.3 to 0.6 µm range. This is a consequence of particles larger than 0.6 µm being removed mainly by inertial impaction and interception and smaller particles mainly by diffusion. Distributor plugging and fines generation due to attrition of the bed media were identified as critical issues and need to be addressed for realizing improvements in this area. by Gui-Hua Qian

A Dissertation Submitted to the Faculty of New Jersey Institute of Technology In Partial Fulfilment of the Requirements for the Degree of Doctor of Philosophy in Chemical Engineering

Department of Chemical Engineering, Chemistry and Environmental Science in Chemical Engineering

May 2000

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# FLUIDIZATION AND SOOT FILTRATION IN A ROTATING FLUIDIZED BED

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- Gui-Hua Qian, Istevan Bagyi, R. Pfeffer, H. Shaw and J. G. Stevens, "Particle Mixing in Rotating Fluidized Beds: Inferences about the Fluidized State", AIChE Journal, Vol. 45, No. 7, pp. 1401 – 1410, 1999.
- Gui-Hua Qian, Istevan Bagyi, R. Pfeffer, H. Shaw and J. G. Stevens, "A Parametric Study of a Horizontal Rotating Fluidized Bed Using Slotted and Sintered Metal Cylindrical Gas Distributors", Powder Technology, Vol. 100, No. 2, pp. 190 -199, 1998.
- Gui-Hua Qian, Ian Burdick, Istevan Bagyi, R. Pfeffer, H. Shaw and J. G. Stevens, "Rotating Fluidized Beds and Their Applications", Presented at 99' ChemShow Conference, New York, November 17, 1999.

- Gui-Hua Qian, Istevan Bagyi, R. Pfeffer, H. Shaw and J. G. Stevens, "The Effect of Gravity on Gas-Solid Fluidization", Presented at AIChE 1999 Annual Meeting in Dallas, October 31, 1999.
- Gui-Hua Qian, Ian Burdick, Istevan Bagyi, R. Pfeffer, H. Shaw and J. G. Stevens, "Soot Removal from Diesel Engine Exhaust Using a Rotating Fluidized Bed Filter", Presented at AIChE 1998 Annual Meeting in Miami, November 17, 1998.

To my beloved wife, Xiao-Ping; my parents, Mr. and Mrs. Qian; my sister, Gui-Qiong and my brother, Gui-Lin.

#### ACKNOWLEDGMENT

I would like to express my sincere gratitude to my supervisors, Professor Robert Pfeffer, Professor Henry Shaw and Professor J. G. Stevens, for their guidance, support and friendship throughout this research.

Special thanks to Professors Gordon A. Lewandowski, Norman Loney and Walter Weissman for serving as members on my committee and giving valuable suggestions.

I also want to thank Dr. Ronald Kane for his kind support and assistance throughout these years.

I appreciate the help and suggestions from my colleague, Dr. Istvan. Baygi. Many of my fellow graduate students are deserving of recognition for their support.

And finally, I want to thank my wife, Xiao-Ping Dai, for her constant support over all these years.

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## NOMENCLATURE

а	Acceleration, $m/s^2$
A	Hamaker constant , J (for most solids, A $\approx 10^{-19}$ J)
С	Cunningham correction factor
C <sub>D</sub>	Drag coefficient
$d_s$	Soot particle diameter, m
$d_i$	Diameter of particle i (i = 1, 2), m
$d_p$	Particle diameter, m
$D_t$	Bed diameter, m
Eu	Euler number
f	Correction factors for the weight of the particles, dimensionless
$F_c$	Cohesive force, N
$F_d$	Drag force on a single isolated particle, N
F <sub>dt</sub>	Drag force on a single particle falling at its terminal velocity, N
$F_{d\varepsilon}$	Drag force on a single particle in a fluidized bed, N
g	Gravitational acceleration, m/s <sup>2</sup>
"g"	Dimensionless acceleration, " $g'' = a/(9.8 \text{ m/s}^2)$
l	Diameter of asperities, µm
n	Exponent in Richardson-Zaki equation
r	Radius, m
r <sub>0</sub>	Distributor radius, m
r <sub>i</sub>	Inner surface radius, m
r <sub>m</sub>	Radius of interface between the two layers of particles, m

# NOMENCLATURE (Continued)

r <sub>pf</sub>	Radius of interface of fluidized and packed beds, m
Re <sub>t</sub>	Reynolds number for unhindered settling particle $\operatorname{Re}_{t} = \frac{\rho_{f} d_{p} u_{t}}{\mu}$
rpm	Rotating speed unit of rotating fluidized bed
RPM	Rotating speed unit of diesel engine
St	Stokes number $St = \frac{C\rho_s d_s U_0}{9\mu d_p}$
Т	Temperature
$U_{ heta_{ m c}}U_{g}$	Superficial gas velocity based on outside radius r <sub>0</sub> , m/s
и	Superficial gas velocity, m/s
$U_m$	Maximum gas velocity before particles blow out, m/s
$u_t$	Particle terminal velocity, m/s
U <sub>mf</sub>	Minimum fluidization velocity, m/s
$U_{mb}$	Minimum bubbling velocity, m/s
$U_{mfc}$	Critical minimum fluidization velocity, m/s
U <sub>mfi</sub>	Inner surface minimum fluidization velocity, m/s
$U_{mfr}$	Minimum fluidization velocity at a given value of radius (r) in the RFB, m/s
$W_b$	Buoyancy force, N
W <sub>e</sub>	Effective weight of particle, N
Wg	Weight of particle, N
ΔP	Pressure drop across the bed, Pa
ε	Voidage

## NOMENCLATURE (Continued)

$\mathcal{E}_b$	Voidage at the transition from particulate to aggregative fluidization				
$\mathcal{E}_{mb}$	Voidage at minimum bubbling condition				
$\mathcal{E}_{mf}$	Voidage at minimum fluidization condition				
μ	Fluid viscosity, Pa·s				
$ ho_a$	Density of "suspension", kg/m <sup>3</sup>				
$ ho_b$	Particle bulk density, kg/m <sup>3</sup>				
$ ho_{f,} ho_{g}$	Gas density, kg/m <sup>3</sup>				
$ ho_p$	Particle density, kg/m <sup>3</sup>				
$ ho_s$	Soot particle density, kg/m <sup>3</sup>				
$\phi_1$	Coefficient in Ergun equation, $\phi_1 = \frac{150(1-\varepsilon)^2 \mu}{\varepsilon^3 (\phi_s d_p)^2}$ , kg/m <sup>3</sup> 新				
$\phi_2$	Coefficient in Ergun equation, $\phi_2 = \frac{1.75(1-\varepsilon)\rho_g}{\varepsilon^3 \phi_s d_p}$ , kg/m <sup>4</sup>				
$\phi_s$	Sphericity of granules, dimensionless				
δ	Distance between two particles, nm				
ω	Rotating speed, rad/s				
Geldart	group A particles: Powders in group A exhibit dense phase expansion				

t group A particles: Powders in group A exhibit dense phase expansion after minimum fluidization and prior to the commencement of bubbling. When the gas supply is suddenly cut off, the bed collapses very slowly. All bubbles rise more rapidly that the interstitial gas velocity.

- Geldart group B particles: Powders in group B bubble at the minimum fluidization velocity. Bed expansion is small and the bed collapses very rapidly when the gas supply is shut off. Most bubbles rise more rapidly than the interstitial gas velocity.
- Geldart group C particles: Powders in group C are difficult to fluidize at all since the interparticle forces are greater than those which the fluid can exert on the particles.
- Geldart group D particles: Powders in group D can form stable spouted beds. Only the largest bubbles rise more slowly than the interstitial fluidizing gas, so that gas flows into the base of the bubble and out from the top of the bubble.

## **CHAPTER 1**

## **INTRODUCTION**

#### **1.1 Background**

Diesel engines are widely used in transportation and for small power applications. They operate by compressing the diesel fuel and air mixture to very high pressures, raising the gas temperature to the point of combustion. As a consequence of their overall lean operation and higher compression ratio, they tend to emit less CO per mile and unburned hydrocarbons with higher thermal efficiency.

The design of the combustion process, however, promotes the production of significant quantities of particulate matter (soot) and  $NO_X$ . There is evidence that particulates from diesel engines are biologically more active than those from spark ignition engines. They are suspected carcinogens (Russell-Jones, 1987). Therefore, diesel emission control is being addressed worldwide. More stringent standards have been proposed and promulgated to limit the emissions of particulates from diesel engines in many parts of the world (Heck and Farrauto, 1995, Burgler et al., 1992, Needham et al., 1991).

Controlling Diesel emissions is more complex than those from gasoline engines. Because HC and CO emissions are relatively low, the main problem is to reduce the particulate and  $NO_X$  (Heck and Farrauto, 1995). Diesel engine manufactures are trying to develop proved diesel engines which will produce less soot and  $NO_X$ . Nevertheless, the "after-treatment" technology will be needed to meet the projected stringent emissions standards.

## 1.2 "After-Treatment" Technology for Diesel Engine Exhaust

#### 1.2.1 Monolith

A monolith is a ceramic or metallic support having the structure of a honeycomb with numerous small parallel channels for gas flow. In catalytic applications, the surfaces of these channels are covered by a wash coat containing the active catalyst. It offers a number of advantages over a packed bed of catalyst pellets, especially lower pressure drop. Monoliths have been widely used as supports in environmental applications.

In the mid-1980s, many engine manufacturers considered using a device to physically filter or trap the particulates on the wall of a wall-flow honeycomb (Farrauto, et al., 1992, Murtagh, et al., 1994). The gases were forced to flow through the monolith wall and exit the adjoining channel. The particulates, having larger particle size than the pore size of the monolith wall, were trapped. Since this device had limited capacity before pressure drop became excessive, it was necessary to periodically regenerate it by combustion. The dry carbon particles of the particulates require at least 400-450°C for ignition. However, the engine exhaust does not reliably reach these temperatures. It was hoped that a heterogeneous catalyst would reduce the light-off temperature of the particulates. This approach has met with limited success. Generally, a burner system to periodically and reliably elevate the temperature of the exhaust to the ignition point of the particulates would be required. The best designs required an expensive and elaborate control system, which was not considered economical for most applications. Although ceramic monolith filter development has proceed straightforward, regeneration systems are problematic and costly. Therefore, these development efforts led to disappointing results. Using catalytic fuel additives to promote regeneration in a trap through lowering the incineration temperature of the collected soot has now been developed by PSA in France. Both Johnson Mathey and Engelhard have commercialized filter which employ a oxidation step to convert NO to NO<sub>2</sub> and then utilize the reaction: NO<sub>2</sub> + C  $\rightarrow$  CO + NO to oxidize soot at low temperatures. Desirable improvements to these technologies would be enhanced durability, reduced probability of plugging, reactor tolerance and total conversion of the NO<sub>2</sub> to N<sub>2</sub>.

## 1.2.2 Cyclone

Arcoumanis, et al., 1994, developed a cyclone technique to treat exhaust of stationary diesel engine. A water-cooled heat exchanger was used to cool the exhaust gas and induce particulate growth in the stream prior to the cyclone. Four smaller cyclones were used in a parallel arrangement to minimize the back-pressure in the engine. The cyclone system with the heat exchanger achieved around 50% filtration efficiency at all operating conditions except at high engine speeds and high load. The poor cyclone performance in this regime may be due to re-entrainment of the deposited material in the hoppers back into the cleaned gas flow. The best performance was obtained from the exhaust system with the heat exchanger, cyclones with long hoppers and exhaust gas recirculation stream via the standard route. Efficiencies of up to 77% were obtained. This approach cannot be used for mobile diesel applications.

## **1.2.3 Fibrous Filter**

Fibrous is a commonly used method of particulate control due to the simplicity of the filter structure and the low cost of materials. Zhu et al., 1998, tested fibrous filters

filtration efficiency for removal of soot from diesel exhaust. With a fresh filter, filtration efficiency was below 20%. With increasing time on stream, the particulate accumulation inside the filter gave rise to increased flow resistance (pressure drop) and typically resulted in increased collection efficiency. After 14 hours, efficiencies as high as 30% were obtained. Due to the drawbacks of low efficiency, difficulty in regeneration and ultimately unaccepted pressure drop, this technology does not appear promising for diesel applications.

## 1.2.4 Rotating Fluidized Beds (RFB) as Filters

Granular bed filters are widely used for aerosol and dust capture with high filtration efficiency (Clift, 1983, D'Ottavio and Goren, 1983). However, they cannot handle that typical engine effluent flow rates in a unit of sufficiently small size. Pfeffer and Hill, 1978, suggested the use of RFB as a filter device. The use of rotating fluidized bed filter (RFBF) has been shown to have many advantages, such as high collection efficiency, relatively low pressure drop, high flow rate of gas per unit area of distributor, ability to operate at high temperatures and a very small "footprint" compared to other granular bed filters, such as conventional fluidized beds, packed beds or moving beds. The minimum fluidization velocity in the RFB is much higher than that in conventional fluidized beds since centrifugal force. The high efficiency of an RFBF is due to the ability to increase the minimum fluidization velocity by increasing the rotating speed and thus the centrifugal force, so that formation of bubbles can be avoided even at very high gas flow rates (Pfeffer, et al., 1986). Diesel engines emit not only amount soot, but also a significant amount of NO<sub>X</sub>. The goal is to develop an RFBF act as a reactor to

simultaneously remove soot and  $NO_X$  from diesel engine exhaust using catalytic granules to filter the soot. The soot is captured by the catalyst granules in the rotating fluidized bed reactor (RFBR) and subsequently oxidized with the accompanying reduction of  $NO_X$  to produce N<sub>2</sub> and CO<sub>2</sub>. The desired overall catalytic reaction can be written as:

Soot + NO<sub>X</sub> + O<sub>2</sub> 
$$\rightarrow$$
 CO<sub>2</sub> + N<sub>2</sub>

With the soot acting as the reducing medium for  $NO_x$ , then the bed can become as selfcleaning, as has been discussed by Tsutsumi et al., 1994, Sun, 1996, Tang, 1997 and Ma, 1998.

#### **1.3. Objectives**

The aim of this study is to enhance our understanding of a novel air pollution control system for removal of soot and  $NO_X$  from the exhaust gas of diesel engines using an RFBR. A parametric study of the operation of needs to be developed. This should include the effect of using different distributors, particle behavior and mixing and the nature of fluidization in centrifugal fields. Furthermore an investigation of filtration efficiency for soot removal is necessary for practical application of the RFB. The four major objectives of this dissertation are presented below.

- 1. The first objective is a study of the effect of using different distributors (slotted, perforated and sintered metal) on the fluidization behavior. This will include measurements of pressure drop and minimum fluidization velocities in an RFB as a function of rotating speed and gas velocity. The experimental results will be compared to theoretical models available in the literature.
- 2. The second objective is to study particle motion and mixing in an RFB by observing

the mixing of two layers of particles of different color and or density. The experimental results will be compared to mixing studies in the literature of conventional fluidized beds.

- 3. The third objective is to study the nature of fluidization and the behavior of particles in centrifugal fields both theoretically and experimentally. The effect of acceleration force "g" on gas-solid fluidization will be determined and compared to the Geldart classification for gravity driven fluidized beds.
- 4. The fourth objective is to determine the parameters that affect the filtration efficiency in an RFBF used to capture soot from the exhaust of a diesel engine. The filtration efficiency as a function of soot size and gas velocity will be evaluated for different collector particles and rotating speed. A comparison with granular bed filtration theory will be provided.

## **CHAPTER 2**

## **DESCRIPTION OF APPARATUS**

#### 2.1 Rotating Fluidized Bed

A rotating fluidized bed as shown in Figure 2.1, is a cylinder with a porous wall that rotates around its axis of symmetry either vertically or horizontally, allowing gases to enter and preventing filter media from leaving. The granular material is introduced into the cylinder and is forced to the wall due to the large centrifugal forces produced by the rotation. The drag force of the gas, which flows radially inward through the porous wall, opposes the centrifugal force. At minimum fluidization, the drag force is balanced by the centrifugal force caused by the rotation of the bed. Therefore, the minimum fluidization velocity is much higher than that found in conventional fluidized beds, in which the drag force balances the force of gravity on the bed media. RFBs can be used by rotating the cylindrical bed either around a vertical or a horizontal axis. However, when treating diesel engine exhaust from vehicles, the bed needs to be rotated at relatively high RPM to maintain proper fluidization conditions, resulting in a centrifugal acceleration of more than 100 times that due to gravity. Therefore the force of gravity is negligible relative to the centrifugal force and the position of the bed (vertical or horizontal) is unimportant. Since it is much more convenient to configure the bed horizontally in a vehicle, our RFBR was rotated along the horizontal axis.



Figure 2.1 Schematic diagram of horizontal rotating fluidized bed.

The gas velocity going through an RFB is a function of radius, r, and decreases with increasing r resulting in a reduction in fluid drag. In contrast, the centrifugal force increases with increasing r. Consequently, fluidization occurs layer by layer starting with the inner surface outwards as the superficial air velocity increases (Chen, 1986). We define the superficial gas velocity, U<sub>g</sub>, to be based on the outer radius, r<sub>0</sub>, of the bed of particles. The inner surface minimum fluidization velocity,  $U_{mfl}$ , is the superficial gas velocity at which the inner surface of the bed is first fluidized. The critical minimum fluidization velocity,  $U_{mfl}$ , is the superficial gas velocity at which the inner surface of the superficial gas velocity at which the inner surface of the superficial gas velocity at which the entire bed is fluidized.



Figure 2.2 RFB configurations at different gas velocity ranges.

## **2.2 Distributors**

Three types of distributors, illustrated in Figure 2.3, were tested. The first consists of a slotted cylinder covered on the inside with a fine mesh screen. The second is a perforated cylinder covered on the inside with a fine mesh screen. The third is a sintered stainless steel cylinder. The slotted cylinder was made of stainless steel in which 30 slots were cut, 0.6 cm by 13.3 cm, resulting in 37% open area. The perforated cylinder was made of stainless steel in which 460 holes, 10 mm in diameter were drilled, resulting in 52% open area. The cylinder wall was covered on the inside with a 325 mesh (45 µm) stainless steel screen. The sintered metal distributor (manufactured by Pall Corporation) is composed of type 316 low-carbon stainless steel powder sintered together in an inert environment resulting in an average pore size of 55 µm. A variable speed motor provided the rotation of the RFB and the rotating speed was calibrated by a stroboscope. Tables 2.1a and b give the rotating speeds in rpm, rad/s and "g" for sintered metal, slotted and perforated

distributor, respectively. Here "g" is defined as the ratio of the centrifugal acceleration to that due to gravity.



Figure 2.3 Structure of distributors.

Table 2.1a Rot	tating speeds and	d "g"	for the	sintered	metal	distributor
		6.7				

rpm	Rad/s	"g"
325	34	7
525	55	19
700	73	34
900	94	56
1200	126	99
1600	167	176
2000	209	275

 Table 2.1b
 Rotating speeds and "g" for the slotted and perforated distributors

rpm	rad/s	"g"
325	34	8
525	55	21
700	73	37
900	94	60
1200	126	107
1600	167	190
2000	209	297

## 2.3 Particle Size and Density of Bed Material

Experiments were conduct with several different bed materials. Properties of these materials are given in Table 2.2. The size distributions for these particles were measured using an Aerosizer (Amherst Process Instruments, Inc.). Alumina particles were provided by the ALCOA Technical Center, Pittsburgh, PA. The mean particle diameter of fresh alumina particles is  $63 \mu m$ . After washed by water more than 10 times and sieving with a 325 mesh screen (45 m m), the fines were removed and the remaining washed alumina particle mean diameter is 89 m m. Q705 alumina which mean size is 68 m m is the high quality alumina particles and the fines have been removed by the manufacture. Some experiments were also done using glass beads manufactured by MO-SCI Corporation which have a much narrower size distribution. The mean size is  $80 \mu m$ .

The bulk density of alumina was measured by tapping the particles to a constant volume and weighing the particles. The particle density of the alumina was calculated from the bulk density based on a voidage of 0.4. The particle density of the glass beads was reported by the manufacturer.

	Bulk Density (kg/m <sup>3</sup> )	Density (kg/m <sup>3</sup> )	Mean Dia. (µm) (by volume)	Mean Dia. (µm) (by number)
Fresh alumina	930	1550	63	4
Washed alumina	930	1550	89	28
Glass beads		2450	80	75
Q705 alumina	730 - 850	1320	68	48
CP-5 alumina	770 - 1000	1470	7	2
White glass beads		2200	91	61
Blue glass beads		2200	84	56

Table 2.2 The properties of particles for experiments
## **CHAPTER 3**

### PARAMETRIC STUDY OF ROTATING FLUIDIZED BED

## **3.1 Introduction**

Rotating fluidized beds have been studied for more than 20 years. Levy et al., 1978, did experiments on the startup of a vertical RFB. When stationary, the bed material was at the bottom of the test section. As the distributor was accelerated, the bed material flowed part way up the vertical wall, reaching a limiting height as the angular velocity was increased. As the distributor was decelerated, the bed material stayed on the wall, until a sufficiently low value of angular velocity was reached at which the material slid back down under the influence of gravity.

Kroger et al., 1979, analyzed the flow in both packed and fluidized beds in a rotating system. Experiments to determine the pressure drop and radial flow distribution were conducted. Moreover, Kroger et al., 1980, studied the particle distribution and mixing in a centrifugal fluidized bed using 440 µm and 875 µm white and black glass beads. Visual observations showed that bubbles are the primary mechanism causing radial mixing. The effectiveness and rate of mixing were found to increase with air flow rate. The data showed that the rate of axial mixing increased rapidly with air flow rate above initial fluidization. The rate of tangential mixing of bed material was also found to increase with air flow rate.

Levy et al., 1981, studied particle elutriation from centrifugal fluidized beds. Experiments performed with glass beads bed material show that the air flow rate at which particle elutriation is initiated is a strong function of the angular velocity of the bed and the bed thickness. Once elutriation begins, the rate of particle loss depends on air flow rate, angular velocity, and bed thickness. Qualitative agreement is obtained between the experimental data and a terminal velocity model for particle loss. The model overpredicts the air flow rate at which particle loss occurs and differences appear to be due to the vertical variations in radial velocity in the chamber and to active bubbling in the bed, which lead to enhanced rates of elutriation.

Takahashi et al., 1984, studied the bed pressure drop and minimum fluidization velocity in a horizontal setup. They found the pressure drop does not remain constant after fluidization, but attains a maximum value at a certain fluidizing velocity. A further increase in gas velocity beyond this point results in a decrease in pressure drop.

Fan et al., 1985, proposed a model based on the balance between the forces, including the centrifugal and fluid frictional forces, exerted on the fluidized particles and the effective weight of the particles. Equations have been derived from the model for predicting the critical fluidization velocity and the maximum pressure drop through the centrifugal bed. Their data for the critical fluidization velocity and the maximum pressure drop through the drop of the bed indicated that the proposed model is valid and the derived equations are of practical use.

Chen, 1987, proposed a fundamental theory based on a local momentum balance to explore the fluidization phenomena in a centrifugal fluidized bed. Unlike the conventional vertical bed, a centrifugal bed is predicted to be fluidized layer by layer from the inner free surface outward, in a range of aeration rates. The span of this range increases with the depth of the particle bed. The pressure drop is predicted to exhibit a plateau, which agrees well with the observations of many investigators, but disagrees with the investigations of Takahashi et al. and Fan et al., in which the pressure drop exhibited a maximum. Predictions of the critical fluidization velocity and the corresponding pressure drop agree well with the data of Fan et al.

In order to fully develop the application of the RFB, it is necessary to understand the fluid mechanics of the device. In this chapter, experiments to measure the minimum fluidization velocity and pressure drop of bed as a function of gas velocity and rotating speed using different distributors are presented. These results are compared to theoretical calculations.

## **3.2 The Special Properties of Rotating Fluidized Beds**

The particles inside an RFB are subjected to a centrifugal field. The ratio of bed thickness to the equivalent diameter of distributor is very small. The gas velocity at inner surface of the bed is higher than that at the distributor since the radius is smaller.

Because RFBs are very thin compared to conventional fluidized beds, wall effects can be expected to be small. Consider, for example, the ratio of the area of the bed in contact with the wall (endplates for the RFB) to the area of the bed perpendicular to the gas flow at the distributor. In RFBs, that ratio is of the order of 0.1 compared to the value of 4 for a conventional bed of circular cross-section and height equal to one diameter. From another side of view, the distributor would strongly affect the fluid dynamic characteristics of the bed.

We have studied the fluid dynamic characteristics of an RFB using three different types of cylindrical gas distributors: slotted and perforated cylinders covered on the inside with a fine wire screen and a sintered metal cylinder. The results show that the pressure drop response obtained upon transition from the packed bed to the fluidized bed regime is quite different for these types of gas distributors. The structure of the gas distributor affects the way in which the bed fluidizes and provides a possible explanation for the different pressure drop behavior as a function of superficial velocity reported in the literature. We also determine the minimum fluidization velocity from the experimental curves of pressure drop versus superficial velocity. Because we are using alumina particles in most experiments which are not monodisperse, but have a relatively broad particle size distribution, these values may differ significantly from those which would be obtained using other definitions, such as change in bed height or acoustic shot noise (Cody et al., 1996). We also report some results using glass beads, which are much closer to being monodisperse and are more spherical than the alumina particles. Consequently, we expect the glass beads to "flow" better than the alumina. Therefore, the glass beads are ideal for investigating the effect of surface morphology on the fluid dynamics in the RFB.

#### 3.3 Theory

The pressure drop and minimum fluidization velocity in RFB can be obtained by using the classical relationships for conventional fluidized beds by replacing the acceleration of gravity, g, in these relationships by the radial acceleration,  $r_0$  produced by the rotation of the distributor.

## **3.3.1 Uniform Bed of Particles**

In the fixed bed regime, the pressure drop equation is given in:

$$\frac{dP}{dr} = \phi_1 U_g + \phi_2 U_g^2 \tag{3.1}$$

where 
$$\phi_1 = \frac{150(1-\varepsilon)^2 \mu}{\varepsilon^3 (\phi_s d_p)^2}$$
,  $\phi_2 = \frac{1.75(1-\varepsilon)\rho_g}{\varepsilon^3 \phi_s d_p}$ 

The superficial gas velocity relationship is:

$$U_g = \frac{r_0}{r} U_0 \tag{3.2}$$

In the fluidized bed regime, the pressure drop is equal to:

$$\frac{dP}{dr} = \left(\rho_p - \rho_g\right)\left(1 - \varepsilon\right)r\omega^2$$
(3.3)

The minimum fluidization velocity at a given value of r in the rotating fluidized bed can be found by equating the pressure drop across the rotating fluidized bed with that across the packed bed giving:

$$U_{mfr} = \frac{-\phi_1 + \left[\phi_1^2 + 4\phi_2 \left(\rho_p - \rho_g\right)(1 - \varepsilon)\omega^2 r\right]^{0.5}}{2\left(\frac{r_0}{r}\right)\phi_2}$$
(3.4)

Thus, the inner surface minimum fluidization velocity  $U_{mfi}$  (the velocity at which the inner surface of the bed is first fluidized) is obtained by letting  $r = r_i$  in Equation (3.4). Similarly, the critical minimum fluidization velocity  $U_{mfc}$  (the velocity at which the entire bed is fluidized) is obtained by letting  $r = r_0$  in Equation (3.4).

The pressure drop across the bed is a function of gas flow rate and can be calculated from one of the following three equations:

1) When  $U_g \leq U_{mfi}$ , the superficial gas velocity is less than the surface minimum fluidization velocity and the bed is packed:

$$\Delta P = \phi_1 U_g r_0 \ln\left(\frac{r_0}{r_i}\right) + \phi_2 U_g^2 r_0\left(\frac{1}{r_i} - \frac{1}{r_0}\right)$$
(3.5)

2) When  $U_{mfi} < U_g < U_{mfc}$ , the bed is partially fluidized and the pressure drop is the sum of the pressure drops across the fluidized and packed beds:

$$\Delta P = (1 - \varepsilon) (\rho_p - \rho_g) \omega^2 \frac{(r_{pf}^2 - r_i^2)}{2} + \phi_1 U_g r_0 \ln\left(\frac{r_0}{r_{pf}}\right) + \phi_2 U_g^2 r_0^2 \left(\frac{1}{r_{pf}} - \frac{1}{r_0}\right)$$
(3.6)

3) When  $U_{mfc} \leq U_g$ , the superficial gas velocity is equal to or exceeds the critical fluidization velocity, the bed is totally fluidized and the pressure drop is:

$$\Delta P = (1 - \varepsilon) \left( \rho_p - \rho_g \right) \omega^2 \frac{\left( r_0^2 - r_i^2 \right)}{2}$$
(3.7)

If the sphericity of particles is known, e. g.,  $\phi_s = 1$  for the spherical glass beads, the above equations can be used directly.

If the sphericity of particles is unknown, e. g., for fresh alumina particles, in the equations given in Kao et al., 1987, the correlations suggested by Wen and Yu, 1966, were used:

$$\frac{1}{\phi_s \varepsilon^3} \approx 14$$
 and  $\frac{(1-\varepsilon)}{\phi_s^2 \varepsilon^3} \approx 11$  (3.8)

The surface minimum fluidization velocity  $U_{mfi}$  (the velocity at which the inner surface of the bed is first fluidized) becomes:

$$\frac{U_{mfi}\rho_g d_p}{\mu} \frac{r_0}{r_i} = \left[ (33.7)^2 + 0.0408 \frac{\rho_g (\rho_p - \rho_g) d_p^3 \omega^2 r_i}{\mu^2} \right]^{0.5} - 33.7$$
(3.9)

Similarly, the critical minimum fluidization velocity  $U_{mfc}$  (the velocity at which the entire bed is fluidized) equals:

$$\frac{U_{mfc}\rho_g d_p}{\mu} = \left[ (33.7)^2 + 0.0408 \frac{\rho_g (\rho_p - \rho_g) d_p^3 \omega^2 r_0}{\mu^2} \right]^{0.5} - 33.7 \quad (3.10)$$

and the average minimum fluidization velocity  $U_{mf}$  is:

$$\frac{U_{mf}\rho_g d_p}{\mu} = \left[ (33.7 \frac{C_2}{C_1})^2 + 0.0408 \frac{\rho_g (\rho_p - \rho_g) d_p^3 \omega^2}{\mu^2} \frac{C_3}{C_1} \right]^{0.5} - 33.7 \frac{C_2}{C_1} \quad (3.11)$$
  
where  $C_1 = r_0^2 (1 / r_i - 1 / r_0), \ C_2 = r_0 \ln (r_0 / r_i), \ C_3 = (r_0^2 - r_i^2) / 2$ 

## 3.3.2 Two Layers of Distinguishable Particles

Equations for the minimum fluidization velocities and pressure drop can be derived for each of the two distinguishable layers of particles in the bed. The two layers of particles may have different minimum fluidization velocities since the density and particle size distributions may be different. If we assume layer 1 is at the distributor and layer 2 is on the inside of the bed, and the radius of the interface is  $r_m$ , we can apply equations (3.1) to (3.3) to each layer.

For layer 2, the inner surface minimum fluidization velocity is given by Equation (3.4) with  $r = r_i$  and the critical minimum fluidization velocity for that layer with  $r = r_m$ . Similarly, for layer 1 at the distributor, the inner surface minimum fluidization velocity is obtained from Equation (3.4) with  $r = r_m$  and the critical minimum fluidization velocity would be determined with  $r = r_c$ .

The bed pressure drop can be calculated from

$$\Delta P = \left. \int_{r}^{m} \left( \frac{dp}{dr} \right) \right|_{layer \, 2} dr + \left. \int_{m}^{r_0} \left( \frac{dp}{dr} \right) \right|_{layer \, 1} dr \tag{3.12}$$

where the pressure drop in each layer is determined from equations (3.5) - (3.7).

### **3.4 Results and Discussion**

The experimental system used to measure pressure drop is shown in Figure 3.1. Before the particles were loaded, the pressure drop of the empty bed due to the distributors as a function of gas velocity and rotating speed was measured as shown in Figure 3.2. If taken photographs for the bed, the experimental system is as shown in Chapter 5 Figure 5.6.



Figure 3.1 Experimental System for Pressure Drop Measurement.

### 3.4.1 Startup of Bed Using Sintered Metal Distributor

Figure 3.3 shows the photographs of the RFB for different startup conditions using the sintered metal distributor with 0.4 kg load of fresh alumina particles. For both cases presented, the final rotating speed obtained was 325 rpm. When the rotation is started slow with no gas flow, the particles are forced to the distributor wall and form a uniform thickness, as shown in Figure 3.3a. When the start rotation is very rapidly again with no gas flow, the particles have not enough time to distribute uniform along the wall, as shown in Figure 3.3b. However, gas flow started and the gas velocity increased to 0.056 m/s (the minimum fluidization velocity is 0.014 m/s at this rotating speed), the thickness of the bed becomes uniform and the bubbles are observed, as shown in Figure 3.3c.



Figure 3.2 Pressure drop of distributors as a function of superficial gas velocity and rotating speed.



a.  $U_g = 0$ , slow startup



b.  $U_g = 0$ , rapid startup

c.  $U_g = 0.056 \text{ m/s}$ 

Figure 3.3 Photographs of the RFB for different startup conditions using the 50  $\mu$ m sintered metal distributor (rotating speed: 325 rpm (34 rad/s); 0.4 kg fresh alumina particles).

# 3.4.2 Sintered Metal Distributor with Fresh Alumina Particles

The sintered metal distributor allows the fluidizing gas to flow uniformly into the bed. As

the gas velocity increases, the bed changes from a packed bed to a partially fluidized bed

with fluidization occurring radially outward starting at the inner surface at velocity  $U_{mfi}$ . With increasing velocity, the entire bed is fluidized at velocity  $U_{mfc}$ . Fluidization occurs in this manner because both the gas velocity (and thus the drag force) and the centrifugal force are a function of radial position in the bed. At the inner surface, the drag force is highest and the centrifugal force is lowest. The pressure drop across the bed increases linearly in the packed bed region and becomes constant after  $U_{mfc}$  is reached, when the bed is totally fluidized. Figure 3.4 is the typical curves of pressure drop through the bed versus gas velocity with fresh alumina particles.

Figure 3.4 shows the pressure drop for the sintered metal distributor as a function of increasing and decreasing gas flow rate for a 0.4 kg particle charge at a rotating speed of 55 rad/s (525 rpm). The pressure drop due to the sintered metal distributor with no particles in the bed is also shown in the figure. In this and all subsequent figures, the pressure drop across the bed refers to the total pressure drop less that of the distributor. The rotating bed is operated with the axis of its cylindrical distributor horizontal. Initially, the distributor is at rest with the particles at the bottom of the bed. Without any gas flow, the rotating speed of the distributor is increased slowly to obtain a uniform distribution of bed particles, i.e., uniform bed thickness. Then, as the gas flow is increased, the pressure drop across the bed is measured using a water-filled manometer at a selected set of superficial velocities. These are shown by the data points on the figure labelled Forward. As the flow rate is further increased, the pressure drop remains relatively constant, indicating that the bed has been fluidized. There is no sudden drop in pressure when the bed becomes fluidized, as was observed using the slotted distributor (shown in Figure 3.9). The results shown in Figure 3.4 are consistent with those found in

the literature (Chen, 1987) which exhibit a plateau in the pressure drop curve after fluidization has been achieved. It should be noted that the pressure drop of the bed follows a slightly different path when increasing or decreasing the gas flow rate. Thus, when the flow rate is decreased from its maximum value with the bed still rotating at 525 rpm, a different curve is followed, labelled Backward. This behavior has not to our knowledge been previously reported in the literature. But this result is in good agreement with those of Tsinontides and Jackson, 1993. for cracking catalyst (75 µm, mass mean diameter), who also established that the hysteresis effect is inversely related to the flowability of the particles in conventional fluidized beds.



Figure 3.4 Pressure drop using the sintered metal distributor as a function of increasing and decreasing flow rates at rotating speed of 525 rpm.

Figure 3.4 shows that the pressure drop of the bed increases slightly upon the transitioning from the fixed bed to the fluidized bed regime. This increase in pressure drop demonstrates that the particles in the RFB are fluidized layer by layer from the inner free surface outward because of the radial decrease in gas velocity and the increase in centrifugal force. From Figure 3.4 we can estimate the experimental surface minimum

fluidization velocity  $U_{mfi}$  from the fixed bed pressure drop line and the critical minimum fluidization velocity  $U_{mfc}$  from the fully fluidization bed pressure drop line (see Theory).

Figure 3.5 shows the pressure drop profile with the sintered metal distributor for different particle charges. As expected, the pressure drop is higher for larger particle charge both in the packed bed and in the fluidized bed regimes. Again, the minimum fluidization velocity increases slightly with increasing charge.



Figure 3.5 Pressure drop as a function of flow rate for two different alumina particle charges in the sintered metal distributor at a rotating speed of 525 rpm.



Figure 3.6 Experimental and calculated pressure drop for the sintered metal distributor with a 0.4 kg alumina particle charge (5.5 mm bed thickness) as a function of superficial gas velocity for different rotating speeds. ( $d_p = 84 \ \mu m$ ,  $r_o = 61.5 \ mm$ ,  $r_i = 56 \ mm$ ,  $\epsilon = 0.47$ ).

Figure 3.6 compares the pressure drop for a 0.4 kg particle charge as a function of flow rate at four different rotating speeds. The pressure drop is again the same for fixed bed conditions, but varies with rotating speed after fluidization is achieved. Higher rotating speed produces a larger pressure drop. The minimum fluidization velocity is also higher at the higher rotating speeds due to the increase in centrifugal force. For the sintered metal distributor with fresh alumina (unknown sphericity), the pressure drop can be calculated directly from the theoretical Equations (3.5) - (3.8). Figure 3.6 shows a comparison of theoretical calculations and the experimental data. Above the critical minimum fluidization velocity, the whole bed is fluidized and the pressure drop of the bed which is a function of the weight of the particles remains constant. The agreement is very good in the packed bed regime but the pressure drop is overpredicted in the fluidized bed regime.



Figure 3.7 Comparison of experimental and calculated minimum fluidization velocity for the sintered metal distributor with a 0.4 kg alumina particle charge (5.5 mm bed thickness) as a function of rotating speed. ( $d_p = 84 \ \mu m$ ,  $r_o = 61.5 \ mm$ ,  $r_i = 56 \ mm$ ).

Figure 3.7 shows that the theoretical calculation of minimum fluidization velocity curves agree with the experimental data reasonably well. This figure shows that the minimum fluidization velocity increases with increasing rotating speed.



**Figure 3.8** Pressure drop using the sintered metal distributor as a function of increasing and decreasing flow rates at rotating speed of 525 rpm.

# 3.4.3 Sintered Metal Distributor with Glass Beads

Figure 3.8 shows pressure drop measurements using the sintered metal distributor for a charge of 0.4 kg of glass beads at a rotating speed of 525 rpm as a function of superficial

gas velocity. As the gas flow is increased, the pressure drop across the bed increases. As the flow rate is further increased, the pressure drop becomes of a constant indicating that the bed has become fluidized. These are the data labelled Forward. When the velocity is decreased from its maximum value with the bed still rotating at 525 rpm, a different data are obtained labelled Backward. As can be observed in Figure 3.8, the forward and backward are almost the same. This study of the hysteresis effect with glass beads shows no substantial pressure drop difference occurs with fluidization or defluidization. These results are in good agreement with the results of Cody et al., 1996, for glass beads in the Geldart-A region. The hysteresis effect with glass beads is different from that obtained using alumina particles (as shown in Figure 3.5), although both of them belong to the Geldard same group. This difference is balanced to be due to differences in the size distribution and the resulting flow ability of the glass beads and the alumina particles.

## **3.4.4 Slotted Distributor with Alumina Particles**

The pressure drop was first measured across the distributor with no particles present. Then particles were added. Figure 3.9 shows the pressure drop measurements using the slotted distributor for a particle charge of 0.4 kg at a rotating speed of 700 rpm as a function of superficial gas velocity which varies between 0 to 0.12 m/s. The pressure drop due to the slotted distributor with no particles in the bed is also shown in the figure. In this and all subsequent figures, the pressure drop across the bed refers to the total pressure drop less that of the distributor. The rotating bed is operated with the axis of its cylindrical distributor horizontal. Initially, the distributor is at rest with the particles at the bottom of the bed. Without any gas flow, the rotating speed of the distributor is increased slowly to obtain a uniform distribution of bed particles, i.e., uniform bed thickness. Then, as the gas flow is increased, the pressure drop across the bed is measured using a waterfilled manometer at a selected set of superficial velocities. These are shown by the data points on the figure labelled First Forward (FF) in Figure 3.9. On this curve, the pressure drop increases steadily until it decreases suddenly by approximately 550 Pa, roughly 60% of its peak value, at the superficial velocity of 0.076 m/s. As the flow rate is further increased, the pressure drop remains nearly constant with a slight upward trend. If the flow rate is then decreased from its maximum value with the bed still rotating at 700 rpm, a different curve is followed, labelled Back/Subsequent (B/S), on which the pressure drop decreases nearly linearly without the discontinuity exhibited on curve FF. If the gas flow is then increased for a second time with the bed still rotating, the pressure drop follows the curve B/S, rather than the original FF curve. However, if the rotation of the bed is stopped and the process is repeated starting with a uniform bed, then the curve FF is again obtained. We define the average minimum fluidization velocity  $U_{mf}$  is to be the gas velocity which corresponds to the peak of the pressure drop.

We now consider the way in which the bed fluidizes and restructures when the slotted distributor is used. As the gas flow is first increased with the bed rotating, the particles form a packed bed of uniform thickness over both the slots and the metal webs (see Figure 2.4a). At the flow rate for which the bed fluidizes, the particles can, by definition, flow and as a result the bed can restructure. Because the webs are wide compared to the slots and the bed is shallow (thickness of the bed is 5 mm for 0.4 kg particle charge), the fluidization occurs primarily over the slots. Moreover, for the shallow beds used in this work, it occurs over a very small increment in flowrate, because

the surface and critical minimum fluidization velocities are nearly equal (see Theory). Thus, the bulk of the particles over the webs remains in the packed, unexpanded state. The particles in the fluidized, expanded state above the slots are thus free to flow over those regions above the webs which have not expanded (see Figure 3.10).



**Figure 3.9** Pressure drop for a 0.4 kg charge of alumina particles in the slotted distributor as a function of increasing and decreasing flow rate at 700 rpm.



Figure 3.10 Conceptualization of Pressure Drop Behavior for the Slotted Distributor



Figure 3.11 Photographs of the surface of RFB using the sintered metal (a, c, e) and slotted (b, d, f) cylindrical distributors. (a.  $U_g=0$  m/s; b.  $U_g=0$  m/s; c.  $U_g=0.085$  m/s; d.  $U_g=0.19$  m/s; e.  $U_g=0$  m/s after high gas flow; f.  $U_g=0$  m/s after high gas flow).

We believe that this restructuring of the bed explains both the magnitude of the reduction in pressure drop and its "one-time" hysteresis behavior. The abrupt thinning of the bed in the flow path causes the decrease in pressure drop to be much greater than that observed due to the change in porosity which occurs at incipient fluidization in conventional fluidized beds or rotating fluidized beds with other types of distributors. This phenomenon does not occur when the sintered metal distributor is used. Furthermore, the redistribution of bed particles is maintained by the rotational force, even in the absence of gas flow. Thus, when the gas flow is stopped with the bed still rotating, no such striking decrease is observed as the superficial velocity is subsequently increased. Only after the rotation is stopped and the bed is brought to a uniform thickness is it again observed. Photographs of the bed presented below provide striking confirmation of this description.

Figure 3.12 summarizes the pressure drop measurements using the slotted distributor as a function of flow rate for particle charges of 0.4, 0.5 and 0.7 kg. The rotating speed is 55 rad/s (525 rpm) and the superficial gas velocity varies between 0 and 0.10 m/s. As expected, the pressure drop for both the fixed bed and fluidized bed increases with increasing particle charge. The minimum fluidization velocities also show a modest increase with increasing particle charge. These curves are comparable to the FF curve in Figure 3.9 and exhibit the same behavior.



Figure 3.12 Pressure drop as a function of increasing gas flow rate for three different charges of alumina particles using the slotted distributor.

Figure 3.13 shows the pressure drop due to a 0.4 kg particle charge as a function of flow rate at four different rotating speeds in the forward direction. For the packed bed, the pressure drop profile is independent of rotating speed. After the bed is fluidized, the pressure drop increases at a higher rotating speed. The minimum fluidization velocity also increases as the rotating speed is increased. These results are all consistent with the relations presented in the Theory section below.



**Figure 3.13** Pressure drop due to a 0.4 kg alumina particle charge as a function of increasing flow rate at four different rotating speeds using the slotted distributor.

Figure 3.14 shows the calculated results compared with the experimental results using the slotted distributor RFB. The experimental data are much lower than the calculated values because only a fraction of particles are fluidized using the slotted distributor. The area of the gas going through the distributor is less than the total area of

the distributor so that the experimental superficial minimum fluidization velocity is much lower (roughly 70%) than that calculated from the theoretical equations which assume that all of the particle are fluidized. On the other hand, as shown in Figure 3.7, the experimental results for the sintered metal distributor agreed with the theory fairly well because all of the particles are fluidized. These results clearly show the different fluidization characteristics obtained with the two different distributors.



**Figure 3.14** Comparison of experimental and calculated minimum fluidization velocity for the slotted distributor with a 0.4 kg alumina particle charge (5 mm bed thickness) as a function of rotating speed. ( $d_p = 84 \ \mu m$ ,  $r_o = 67 \ mm$ ,  $r_i = 62 \ mm$ ).



**Figure 3.15** Pressure drop due to a 0.5 kg glass bead charge as a function of increasing flow rate at different rotating speeds using the slotted distributor.

# 3.4.5 Slotted Distributor with Glass Beads

The pressure drop in the slotted distributor was measured using a charge of 0.5 kg glass beads, which have a much narrower size distribution than the alumina particles. Figure 3.15 shows the bed pressure drop curve for these particles as a function of flow rate for different rotating speeds. It should be noted that at low rotating speeds, e.g., 325 rpm and

525 rpm, there is no sudden drop in pressure after fluidization and the results are similar to those obtained using the sintered metal distributor (see below). This behavior is due to the better flowability of the narrow size distribution glass beads as compared to the alumina particles. At low rotating speed, the gas velocity and the centrifugal force are both relatively smaller so that the particles over the webs can flow back to the slots. However, when rotating speed is increased to 700 rpm, the pressure drop curve is similar to that obtained with the alumina particles and shows a sudden drop-off after fluidization. This change in pressure drop behavior occurs because the centrifugal force acting on the particles at high rotating speed is large so that the glass beads in the web areas are forced to the wall and cannot redistribute over the slots. Thus for the slotted distributor the bed pressure curve is dependent on the flowability of the particles.

## **3.4.6 Perforated Distributor with Alumina Particles**

The behavior of the perforated distributor is intermediate between sintered metal distributor and slotted distributor. Although its fraction of open area is higher than that of slotted distributor, it does not distribute gas as uniformly as the sintered metal distributor. Figure 3.16 shows the pressure drop of the bed using the perforated distributor as a function of superficial gas velocity. After minimum fluidization velocity, the pressure drop decreases suddenly, but not as much as with the slotted distributor (see Figure 3.9). This difference suggest that for the perforated distributor there is less particle movement from the open area to the web area.



**Figure 3.16** Pressure drop due to a 0.4 kg alumina particle charge as a function of increasing flow rate at different rotating speeds using the perforated distributor.

Figure 3.17 compares theoretical calculations and experimental measurements of the minimum fluidization velocity as a function of rotating speed using slotted and perforated distributors. The minimum fluidization velocity for the slotted distributor is lower than that for the perforated distributor. This observation is consistent with the greater open area of the perforated distributor compared to that of the slotted distributor (52% vs. 37%)This means that there is more fluidization area using perforated distributor than

using slotted distributor.



**Figure 3.17** Comparison of experimental and calculated minimum fluidization velocity for the slotted and perforated distributors with a 0.4 kg alumina particle charge (5 mm bed thickness) as a function of rotating speed. ( $d_p = 84 \ \mu m$ ,  $r_o = 67 \ mm$ ,  $r_i = 62 \ mm$ ).

### **CHAPTER 4**

# PARTICLE MOTION AND MIXING IN ROTATING FLUIDIZED BEDS

### **4.1 Introduction**

Although much basic and applied data on fluidization has been generated during the past fifty years (Davidson et al., 1985), some fundamental questions still need to be addressed. For example Jackson, 1998, concluded that a combination of careful bed height, fluid pressure drop and fluctuation velocity measurements, in beds of several diameters should be capable of providing a substantial amount of information about the particle phase stresses. Recently, Menon and Durian, 1997, measured particle motions or local particle fluctuation velocities by diffusing-wave spectroscopy (DWS) in a conventional gasfluidized bed of glass beads and demonstrated that the motion of macroscopic bubbles is the source of particle motions in the bed. They concluded that the homogeneous state of the bed known as the uniformly fluidized state (after incipient fluidization has occurred but before bubbles are observed) is actually a weak solid in which particles are at rest. In the bubbling state, particles are carried in convection patterns with a downflow at the walls and an upflow in the middle of the bed. Superimposed on this overall convection is motion in the vicinity of rising bubbles, where particles are swept up with the bubbles. Using DWS, Menon and Durian, 1997, measured the mean squared particle fluctuation velocity or granular temperature near the wall of their fluidized bed as a function of particle diameter and gas flow rate. Similar experiments were done by Cody et al., 1996, 1997, by measuring the acoustic shot noise (ASN) excitation of the surface of the

are compared in a recent paper by Cody et al., 1998, who show that for Geldart-B

fluidized bed vessel by random particle impacts. The results of all of these experiments

(Geldart, 1986) glass spheres ( $d_p \ge 150 \mu$ m) the agreement between the two independent sets of data is excellent, but for Geldart-A glass spheres ( $d_p \le 120 \mu$ m) there is a systematic difference between the two measurements. The DWS measurements of Menon and Durian, 1997, do not show the bifurcation in the fluctuation velocity observed in the ASN measurements of Cody et al., 1996, 1997, below the Geldart B/A transition at  $d_p \sim$ 120 µm. The authors (Cody et al., 1998) attribute this difference to the fact that the experiments by Cody et al. were performed in a cylindrical vessel whereas those by Menon and Durian were performed in a square vessel together with the hypothesis that Geldart-A fluidization is dependent on particle circulation within the fluidized bed. To prove this point, they present theoretical arguments that the Geldart-A glass spheres of Menon and Durian are behaving like Geldart-B glass spheres due to the low shear circulation at the face of the square vessel. This hypothesis, that Geldart-A fluidization is dependent on circulation within the fluidized bed, if correct, has significant implications for all experimental and theoretical research in the area of fluidization.

In a conventional fluidized bed, it is difficult to avoid circulation of the particles because of wall effects. Also the minimum fluidization velocity is relatively low so that the range between the minimum fluidization and the minimum bubbling velocity is narrow even for Geldart-A particles. A rotating fluidized bed (RFB), on the other hand, has several advantages for these kinds of studies. There are negligible wall effects since the distributor comprises the cylindrical vessel and the only "walls" are the end plates. Minimum fluidization velocities are also typically much higher than for a conventional fluidized beds with initial fluidization occurring at the inner surface and proceeding rapidly outward with increasing gas velocity (Kao et al., 1987). However, it must be emphasized that there are essential differences between conventional fluidized beds and rotating fluidized beds. An important difference is the magnitude of the respective acceleration fields experienced by the bed particles. Thus our results may not shed direct light on the discrepancy between the findings of Menon and Durian, 1987, and Cody et al, 1996.

In this chapter, we present experimental results of mixing caused by Geldart-A particle motions in an RFB. Our simple experimental procedure is to load two different colored layers of particles consisting of the same or different materials. If the two colored layers of particles mix together, it is obvious that the particles have moved. At the same time, the relationship between pressure drop of the bed and superficial gas velocity is measured in the packed, partially fluidized, and totally fluidized bed operating regions. We did not measure bed expansion since in our rotating fluidized bed experiments the bed is very shallow (approximately 1 cm) and is also non-uniform both radially and axially. By comparing our observations of particle mixing and bubble formation with pressure drop, the point at which particle motion occurs can be determined. Kroger et al., 1980, have previously studied particle mixing in a centrifugal fluidized bed. Their visual observations show that bubbles are the primary mechanism causing radial mixing for Geldart-B and Geldart-D particles. Fluidization with bubbling at the freeboard surface was observed at the value of the predicted minimum fluidization velocity. However, Kroger et al., 1980, did not study mixing of Geldart-A particles.

The primary purpose of this chapter is to study particle motion and mixing as gas flow rate increases. Therefore, no data is presented on the fluidization/defluidization hysteresis effect. We previously studied this phenomenon and reported on it in Qian et al., 1998. Pressure drop across a bed of alumina powder from the same batch used in the current experiments was measured with increasing and then decreasing gas flow rate using the same experimental arrangement and conditions as reported in this current paper. Our results in Qian et al., 1998, show there is a slight difference between the pressure drop curves obtained with increasing and decreasing gas flow rate. These results are in good agreement with those of Tsinontides and Jackson, 1993, for cracking catalyst (75  $\mu$ m, mass mean diameter), who also established that the hysteresis effect is inversely related to the flowability of the particles. Studies of the hysteresis effect with glass beads (again, similar to those used in the current paper) show no substantial pressure drop difference occurs with fluidization or defluidization (Qian, 1998). These results are in good agreement with the results of Cody et al., 1996, for glass beads in the Geldart-A region.

## 4.2 Experimental System

A schematic diagram of the experimental system is shown in Figure 5.6 in Chapter 5. A Pulnix progressive scan CCD Camera (Model Number: TM-1001-477; resolution:  $768(H) \times 494(V)$ ; shutter speed from 1/60 to 1/10000 second) was used to study particle mixing in the RFB as a function of increasing gas velocity. The camera was connected to a computer in order to view and store the photographs in real time. All optical measurements using the Pulnix digital camera were taken perpendicularly through the Plexiglas end-plate. However, the inner surface of the rotating fluidized bed was also observed at an angle by eye and exhibited the same behavior that was seen in the photographs. We did not see any accumulation of particles sticking to the Plexiglas endplates in any of our experiments, indicating that electrostatic forces for our relatively large particles were negligible.

The sintered metal distributor with an average pore size of 55  $\mu$ m was used. The inner radius of the distributor, which is the same as the outer radius of the bed, was 61.5 mm. To determine the inner radius of the bed precisely is difficult. The inner radius varies with the mass of particles loaded into the bed. Upon rotating the RFB, the inner surface appeared non-uniform both radially and axially in the packed bed mode. When a second layer of particles was added to the bed, this non-uniformity remained. At the inner surface minimum fluidization velocity ( $U_{mfl}$ ), pronounced irregularities in the bed surface disappeared, causing the surface to become more uniform. Since we used shallow beds, very little bed expansion occurs. Consequently, we have measured the thickness of both layers in the bed at  $U_{mfl}$ . These values are given in Table 4.1.

The rotating speed of the RFB was set at 525 rpm (55 rad/s) which is equivalent to a centrifugal acceleration of 19 times that due to gravity. The bed is initially loaded with particles of one color, and then rotated at low speed. The other colored particles are blown into the bed using a flexible tube, so that two different layers of particles are formed in the bed. The interface between the two layers is distinct and clearly visible. The bed rotational speed was then increased to 525 rpm as determined with a stroboscope. The strobe frequency was set to match the rotational frequency of the bed so that any macroscopic bubbles present could be observed. The pressure drop as a function of gas velocity and rotating speed was measured using a U-tube water manometer. Before loading particles into the bed, the pressure drop of the distributor of the empty bed was also measured.

	Inner Layer				Particle Density	Mean Dia. (by Volume)
Outer Layer	Alumina, black	Alumina, white	Glass, blue	Glass, white	kg/m <sup>3</sup>	μm
Alumina, black		5.4, 5.8			1550	88.9
Alumina, white			5.0, 6.3		1550	88.9
Glass, Blue		4.8, 4.6		4.0, 4.2	2200	84.0
Glass, white					2200	91.3

Table 4.1. Four Combinations of Geldart A particles and their properties

Ordered pairs in table give outer and inner layer thickness (mm) in that order as measured at the end-plate.

Our experiments were conducted with four combinations of Geldart-A particles as described in Table 4.1. The bed material consisted of blue or white glass beads provided by MO-SCI Corporation and alumina provided by the ALCOA Technical Center. The alumina particles are washed and get rid of fines. The color of alumina is white. Black alumina was made by coating the alumina with submicron soot from a diesel engine exhaust. The effect of the soot coating on the particle size distribution and density of the alumina was negligible. Separate experiments were performed to ascertain whether the black alumina particles have significantly different fluidization properties from the fresh white alumina particles. We measured the pressure drop as a function of gas flow rate using a smaller 5.7 cm ID, 7.6 cm long rotating fluidized bed at two different rotating speeds. The results show that the values of the pressure drop for both materials are identical at a rotating speed of 600 rpm and are acceptably close (within experimental error) at a rotating speed of 1000 rpm where the gas flow rate is much higher. Thus, the very fine soot layer on the black alumina does not influence its fluidization properties.
The particle density of the glass beads is 2200 kg/m<sup>3</sup> and the particle density of the washed alumina is 1550 kg/m<sup>3</sup> (see Table 4.1). The particle size distribution for all of these particles was measured using an Aerosizer (Amherst Process Instruments, Inc.) and is shown in Figure 4.1. The alumina particles have a relatively broad size distribution, whereas the glass beads have a much narrower size distribution. The volume average size of the white glass beads, blue glass beads and alumina are 91.3  $\mu$ m, 84.0  $\mu$ m and 88.9  $\mu$ m, respectively. All of them belong to group A of Geldart's powder classification based on conventional gravity-driven fluidized beds (Geldart, 1986).



Figure 4.1 Particle size distribution by number for bed materials used.

## 4.3 Results and Discussion

The sintered metal distributor allows the fluidizing gas to flow uniformly into the bed. As the gas velocity increases, the bed changes from a packed bed to a partially fluidized bed with fluidization occurring radially outward starting at the inner surface at velocity  $U_{mfi}$ . With increasing velocity, the entire bed is fluidized at velocity  $U_{mfc}$ . In this chapter, we take the sphericity,  $\phi_s$ , to be unity for both the spherical glass beads and the washed alumina particles. The minimum fluidization velocity and the pressure drop curves can be calculated from the equations (4.4 – 4.7) and equation (4.12) in Chapter 4.3 theory. Presented below are the experimental results for the four different particle systems investigated.

## 4.3.1 Blue and White Glass Beads

Blue glass beads (0.3 kg) were placed on the distributor and the RFB was rotated. As the bed was rotating, white glass beads (0.3 kg) were added to the bed to produce two distinguishable layers of particles. The interface between the two colored layers is distinct and clearly visible. However, the radial location of the inner surface of the white beads is generally not uniform due to the method used to introduce the second layer. As the gas velocity was increased, the pressure drop of the bed was measured. Figure 4.2 shows the measured points of net pressure drop (total pressure drop less the pressure drop of the distributor) as a function of gas velocity. Figure 4.2 also shows the theoretical pressure drop curve with the theoretically calculated inner surface and critical minimum fluidization velocities. In order to fit the experimental data to the theoretical curves we multiplied the weight of the particles by a correction factor, f, equal to 0.85.

The application of this factor is reasonable for the following reasons. First, we conducted a large number of experiments in the RFB where particles were weighed before and after the experimental run. Data from 12 different runs using either alumina or glass beads as the bed material showed an average weight loss of the order of 5% of the bed weight. Some particles were lost during the initial fill-in procedure, some were lost due to elutriation at high gas velocities and still others were lost whenever the bed rotation was started or stopped. However, the major contribution to the correction factor, f, is due to the non-uniformity of the bed thickness both radially and axially. The effect of gravity alone produces a somewhat thicker bed at the bottom compared to the top of the horizontal rotating bed. For example, at 525 rpm the centrifugal acceleration is about 19 g's. The resulting difference of approximately 5% in the radial thickness of the bed allows more gas to percolate through the thinner part of the bed giving a lower pressure drop than predicted by theory. This phenomenon was pointed out by Chen, 1987, who compared the experimental data of the pressure drop in a rotating fluidized bed obtained by many different investigators (Fan et al., 1985, Demircan et al., 1978, Levy et al., 1978) to the theory. Chen concluded that the theory always overpredicts the data with an average error of more than 10% and sometimes as much as 20%. Thus, setting f = 0.85for the glass beads is consistent with our experimental observations as well as with previous experimental data.

By assuming a constant voidage of 0.41, the theoretical and experimental values of  $U_{mfi}$  and  $U_{mfc}$  in Figure 4.2 are seen to be in excellent agreement. However, it should be noted that in a rotating fluidized bed, as opposed to a conventional fluidized bed, there can be small radial variations in bed voidage due to differences in local velocities and centrifugal acceleration (Mutsers and Rietema, 1977, Chen, 1987).

When  $U_g < 0.105$  m/s, the predicted value of  $U_{mfi}$ , the RFB is expected to be in the packed bed regime. Several photographs were taken in this range. Figure 4.3-A ( $U_g = 0.054$  m/s) shows a typical photograph of the bed. The bed had a non-uniform surface due to the method of filling the second layer of particles with different color, as discussed above. In this condition, the particles are static relative to each other and behave as a solid.

As the gas velocity was increased to  $U_g = 0.113$  m/s, the bed surface started to become more uniform as shown in Figure 4.3-B. Particles begin to fluidize at  $U_{nff}$ . Upon increasing the gas velocity further the bed begins to fluidize outwardly until the entire bed is fluidized at  $U_{nfc}$ . Between the inner surface and the critical minimum fluidization velocities the bed is partially fluidized. In this range, particles are still relatively stable in relation to one another. The two different colored layers of particles have not mixed yet, indicating that particles have not moved in the radial direction. However, those particles that have already been fluidized close to the inner surface of the bed appear to have moved along the surface and the bed has become somewhat fluid-like. As the gas velocity is increased further the bed surface became more uniform, i.e., the inner surface of the white glass beads is sharply delineated, as can be seen in Figure 4.3-C at  $U_g = 0.134$  m/s.

In theory, the minimum gas velocity at which the first inner layer of particles can begin to move is the inner surface minimum fluidization velocity,  $U_{mfi}$ . The surface should start to become more uniform at this point. However, detecting this onset is difficult using either a camera or through visual observation. We note that the inner surface velocity is slightly lower than 0.113 m/s (Figure 4.3-B) but higher than 0.102 m/s at which a photograph (not shown) indicated that the bed was still packed. Our theoretical calculation indicates that the inner surface gas velocity ( $U_{mfi}$ ) is 0.105 m/s and the critical fluidization velocity ( $U_{mfc}$ ) is 0.134 m/s.

When  $U_g = 0.142$  m/s, the particles began to mix and the interface between the two colored layers became diffuse. Figure 4.3-D shows that the particles are partially mixed. The interface between the two colored layers disappears in the bed section between the two arrows in the figure. This condition was frozen by quickly lowering the gas flow rate. The photograph was taken and then  $U_g = 0.142$  m/s was reset. The entire bed became mixed very rapidly. There is no visible interface, as shown in Figure 4.3-E. The gas velocity at which the bed became totally mixed was slightly higher than the theoretical  $U_{mfc}$ , as would be expected for Geldart-A particles. However, this slight difference is within experimental error and does not clearly indicate a non-bubbling totally fluidized bed regime before mixing occurs. Bubbles were also observed when the mixing velocity was re-established. This observation indicates that bubbles cause the bulk motion of the particles in the radial direction, in agreement with the results of Menon and Durian, 1997. In contrast to their results, there is no clearly discernible difference between  $U_{mfc}$  and the minimum bubbling velocity,  $U_{mb}$ , i.e., they either coincide or are nearly equal.

For gas velocities between  $U_{mfr}$  for  $r = r_m$  (the radius of the interface) and  $U_{mfc}$ , the interface between the two colored layers is in the fluidized region of the bed. At velocities in this interval, particle motion and mixing of the two layers could occur by particle diffusion. However, during the time when  $U_g$  was greater than  $U_{mfi}$  and less than  $U_{mb}$  and thus could have been in this interval, we did not observe any "blurring" of the interface either visually or upon examination of the digital photographs. Mixing due to diffusion occurs on a much longer time scale than the rapid mixing observed at the onset of bubbling. Thus to the extent that particle motion occurred at the interface due to diffusion, it did so on a length scale too small to be observed in our experiments.

Particle motion and mixing can also occur due to circulation flow in a fluidized bed. However, our beds are very thin compared to conventional fluidized beds. Consider, for example, the ratio of the area of the bed in contact with the wall (end-plates for the RFB) to the area of the bed perpendicular to the gas flow at the distributor. In our beds, that ratio is on the order of 0.1 compared to the value of 4 for a conventional bed of circular cross-section and height equal to one diameter. Thus, we would not expect any significant effect due to shear at the wall. Indeed, we did not observe any circulation flow in our experiments.



**Figure 4.2** Pressure drop as a function of superficial gas velocity for two different colored layers of glass beads. Letters A to E refer to the photographs in Figure 4.3.





B.  $U_g = 0.113$  m/s

Partially mixed bed section



D.  $U_g = 0.142 \text{ m/s}$ 



E.  $U_g = 0.142 \text{ m/s}$ 

0.3 kg Blue Glass Beads



0.3 kg White Glass Beads

Figure 4.3 Photographs of RFB at different superficial gas velocities for blue and white glass beads.

# 4.3.2 Black Alumina and White Alumina

In order to verify the above results, alumina was used instead of glass beads in the same experiment. The average particle size of the alumina is almost the same as that of the glass beads, but its density is lower and its size distribution is much wider. The alumina particles, like the glass beads, fall into the Geldart-A classification for conventional fluidized beds. Black alumina (0.4 kg) was filled into the bed and while the RFB was rotating, white alumina (0.3 kg) was loaded into the bed. Figure 4.4 presents the experimental pressure drop values and the theoretically calculated curve (less pressure drop of the distributor) as a function of gas velocity. The bed voidage used was 0.425. The measured weight of the alumina was multiplied by a correction factor, f = 0.80, in order to fit the experimental points to the theoretical curve for the same reasons as given above for the glass beads. The small difference in f factors for the glass beads and alumina particles is probably due to the fact that the alumina particles have a much wider size distribution with more smaller particles (see Figure 4.1).

The experimental method was the same as was used for the glass beads. When  $U_g$  < 0.074 m/s, the predicted  $U_{mfi}$ , the RFB is in the packed bed regime. In this condition, the bed surface is not uniform due to the fill-in procedure and particles in the two layers did not mix. Several photographs were taken. These photographs all show the same structure of the bed, as illustrated in Figure 4.5-A ( $U_g = 0.043$  m/s).

The first change was observed at 0.080 m/s (Figure 4.5-B). The bed surface began to become uniform. Between  $U_{mfi}$  and  $U_{mfc}$  the bed was partially fluidized. However, the black and the white particles did not mix. Figure 4.5-C shows a photograph of the bed at  $U_g = 0.083$  m/s. Mixing had not occurred, but the inner surface of the bed had become even more uniform than in Figure 4.5-B.

Figure 4.5-D shows a photograph that was taken at  $U_g = 0.089$  m/s, when the bed was partially mixed. Again, the interface between the two colored layers disappears in the bed section between the two arrows in the figure. This condition was frozen by lowering the gas flow rate. After the photograph was taken, the same gas velocity was reset. A couple of seconds later total mixing of the bed occurred, and bubbles appeared as shown in Figure 4.5-E. Based on the pressure drop curve and this observation, the experimental  $U_{mfc}$  is estimated to be 0.089 m/s. It is the first point at which the gas velocity has no effect on pressure drop, i.e., the pressure drop becomes constant. The theoretical calculation gave a higher  $U_{mfc} = 0.111$  m/s. Even though these particles are in the Geldart-A classification, the mixing point appears to occur at the experimentally measured  $U_{mfc}$ , which, within observational uncertainty, is also the minimum bubbling velocity. Again, before bubbling neither small-scale mixing at the interface nor establishment of circulation flow were observed.



**Figure 4.4** Pressure drop as a function of superficial gas velocity for two different colored layers of alumina. Letters A to E refer to the photographs in Figure 4.5.



Figure 4.5 Photographs of RFB at different superficial gas velocities for black and white alumina.

# 4.3.3 Blue Glass Beads Outside and White Alumina Inside

Since the densities and size distribution of the alumina particles and glass beads are different, their minimum fluidization velocities are also different. In order to investigate the mixing properties for particles of different materials, having different densities, blue glass beads (0.4 kg) were loaded into the bed first. While the RFB was rotating, white alumina (0.3 kg) was filled into the bed giving two layers of particles with different colors and densities. Again, the pressure drop of the bed was measured as the gas velocity was increased. Figure 4.6 shows the experimentally measured pressure drop points and the theoretically calculated curve vs. gas velocity. The previously used voidages and correction factors for the glass beads and the alumina were used in calculating the pressure drop curve. Figure 4.7 shows the photographs that were taken during this experiment.

Until point A ( $U_g = 0.080$  m/s) in Figure 4.6, the inner bed surface was nonuniform, as can be observed in Figure 4.7-A. The next measurement point was at  $U_g =$ 0.086 m/s. As the alumina inner surface began to fluidize, the surface became more uniform very quickly, as shown in Figure 4.7-B. The calculated minimum fluidization velocities are as follows: for glass beads,  $U_{mfi} = 0.105$  m/s and  $U_{mfc} = 0.122$  m/s; for alumina  $U_{mfi} = 0.079$  m/s and  $U_{mfc} = 0.095$  m/s. We believe that the photograph at point A was taken just before the inner surface minimum fluidization velocity was reached, although the velocity at this point is approximately equal to the calculated  $U_{mfi}$  for the alumina. It appears that the entire layer of alumina began to fluidize before the glass beads started to fluidize. We were not able to take photographs showing the surface becoming progressively more uniform. However, we observed that a thin light blue layer intermediate between the darker blue glass beads inner layer and the white alumina outer layer formed when the glass beads began to fluidize. This thin layer did not grow and was seen until total mixing of the bed was obtained with bubbles appearing, as shown in Figure 4.7-C at  $U_g = 0.118$  m/s. At this point the particles mixed very quickly.

The observed behavior can be explained by the fact that the alumina particles were already totally fluidized by the time the entire glass bead layer became fluidized. Therefore, the mixing was very fast. The particles mixed when the glass beads reached the critical fluidization velocity. It appears, however, that the theoretical calculations give higher values of the minimum fluidization velocities than those observed experimentally.



**Figure 4.6** Pressure drop as a function of superficial gas velocity for blue glass beads and white alumina layers. Letters A to C refer to the photographs in Figure 4.7.



**Figure 4.7** Photographs of RFB at different superficial gas velocities for blue glass beads and white alumina powder.

# 4.3.4 White Alumina Outside and Blue Glass Beads Inside

Alumina particles (0.3 kg), having a lower density than the glass beads were placed on the outside of the bed adjacent to the distributor. Glass beads (0.5 kg) were filled on the inside of the bed. In this case the mixing behavior was entirely different from that observed in the previous experiments. As the gas velocity was increased, the bed pressure drop was measured. Figure 4.8 presents the experimental pressure drop points and the theoretically calculated curve as a function of gas velocity. The calculation method and the correction factors were the same as had been used previously. As can be seen from the figure, in this case the theoretical pressure drop curve overpredicts the experimental values in both the packed and fluidized bed regimes. Figure 4.9 shows the photographs that were taken during the experiment.

At point A ( $U_g = 0.064 \text{ m/s}$ ) in Figure 4.8, neither the glass beads nor the alumina are fluidized, i.e., both layers are in the packed bed regime. The particles were stationary relative to each other and therefore the inner surface of the bed was not uniform, as shown in Figure 4.9-A. In the packed bed regime several photographs were taken without any changes in the appearance of the bed surface.

At  $U_g = 0.097$  m/s the bed surface became more uniform. At the same time, a radial segment of the bed began to mix without observing any bubbles as shown in Figure 4.9-B. The appearance of the more uniform surface occurred very quickly. At the previous measurement point ( $U_g = 0.091$  m/s, not shown) the bed surface was still not uniform even though the glass beads may have become fluidized based on the theoretical pressure drop curves. The next three photographs (Figure 4.9-C, Figure 4.9-D, and Figure 4.9-E) show that the mixed region (between the two arrows where the interface is no longer visible) grew larger as the gas velocity was increased. Eventually, at 0.110 m/s the entire bed became mixed (no interface) as can be seen in Figure 4.9-F. Bubbles were not observed at this gas velocity. Bubbles appeared later at 0.118 m/s.

The calculated fluidization velocities of the glass beads and alumina are:  $U_{mfi}$  =

0.083 m/s and  $U_{mfc} = 0.105$  m/s, and  $U_{mfi} = 0.095$  m/s and  $U_{mfc} = 0.111$  m/s, respectively. Thus, the glass beads had already been partially fluidized when the alumina particles started to become fluidized at 0.095 m/s. However, the outer portion the inner layer of glass beads remained in the packed bed regime. This packed bed layer covered the alumina particles as a "lid" preventing the expansion of the particles in the alumina layer. Once all of the glass beads became fluidized, the "lid" was removed, and the entire bed proceeded to fluidize.

Unlike the previous experiments, the mixing of the bed began at the gas velocity at which the alumina particles started to become fluidized. At this point a small section of the bed started to mix (see Figure 4.9-B) and grow (see Figures 4.9-C, D, E) until the entire bed was fluidized (see Figure 4.9-F). In contrast to the virtually instantaneous mixing observed in the prior experiments, density driven mixing occurs gradually because the more dense glass beads move into the fluidized alumina layer due to the centrifugal force. Bubbles were not observed until  $U_g = 0.118$  m/s, which was somewhat higher then the velocity of total mixing (see Figure 4.9-G).



**Figure 4.8** Pressure drop as a function of superficial gas velocity for white alumina and blue glass beads layers. Letters A to G refer to the photographs in Figure 4.9.



Figure 4.9 Photographs of RFB at different superficial gas velocities for white alumina and blue glass beads.

#### 4.4 Discussion

For particles of the same material, the two layers of particles do not mix until bubbles appear. This result is similar to that of Menon and Durian, 1997, who concluded that bubbles are responsible for the bulk motion of particles in the conventional fluidized bed used in their study. Below  $U_{mfi}$  the bed behaves as a solid. Between the calculated  $U_{mfi}$  and  $U_{mfc}$  the inner surface is observed to become more uniform, i.e., in this respect, the bed exhibits fluid-like behavior before mixing. After  $U_{mfc}$  particles inside the bed start to move radially and mixing occurs rapidly. Bubbles are a strong source of particle motion and the bed becomes totally fluid-like. To within observational error, the mixing, minimum critical fluidization velocity and minimum bubbling velocity coincide. These results are different from these of Menon and Durian, 1997, who found bubbling to occur in their gravity-driven bed at a velocity appreciably greater than the minimum fluidization velocity for Geldart-A particles. Figure 4.10 is a schematic representation of these ideas.

For conventional fluidized beds, the minimum bubbling velocity,  $U_{mb}$ , is dependent on the size and density of the particles. For Geldart-B particles,  $U_{mb}$  is equal to the minimum fluidization velocity,  $U_{mf}$ , whereas for Geldart-A particles, which were used in our experiments,  $U_{mb}$  is larger than  $U_{mf}$  so that there is an appreciable operating region in which the bed is fluidized but not yet bubbling. Cody et al., 1998, suggest that particle circulation in a conventional circular cross-section fluidized bed which leads to shear forces may be responsible for Geldart-A behavior. Our RFB has negligible wall effects and hence presumably negligible particle circulation. The fact that we observe no mixing of particles before bubbles are observed could be a consequence of the hypothesis of Cody et al. that Geldart-A fluidization is dependent on circulation within the fluidized bed and can mimic Geldart-B like behavior when particle circulation is absent. On the other hand, the RFB differs from a conventional fluidized bed in that the acceleration field is considerably greater than gravity. Therefore, previous studies in conventional fluidized beds to determine  $U_{mb}$  may not be relevant. Indeed, the location of the Geldart A/B boundary in the diameter-density plane is likely to be a function of the magnitude of the effective acceleration field. Our results are also consistent with the hypothesis that a shift in the location of this boundary occurs that results in particles in the Geldart-A classification for conventional beds exhibiting Geldart-B behavior in an RFB. We intend to present a theoretical analysis of this issue in the future.

For particles of different density, the mixing characteristics of the two layers depend on whether or not the denser particles are near the distributor. When the denser particles are near the distributor the mixing behavior is very similar to that of particles of the same density. When the less dense particles are near the distributor the mixing behavior is quite different. Mixing occurs gradually due to the differences in density and fluidization properties of the two layers and total mixing is observed before bubbles appear.



Figure 4.10 Schematic diagram of operating region.

## **CHAPTER 5**

## GAS-SOLID FLUIDIZATION IN A CENTRIFUGAL FIELD

#### **5.1 Introduction**

The behavior of fluidized solid particles has been reported in the literature for over 50 years. Wilhelm and Kwauk, 1948, early on, described the stability of a fluidized bed of solids fluidized with either a liquid or a gas. Two modes of fluidization, termed particulate and aggregative fluidization, were observed depending on the value of the Froude number,  $u_{mf}^2/(gd_p)$ . Particulate fluidization, which was obtained in solids-water experiments, is characterized by the separation of individual particles much the same as the molecules of a gas. It is a non-bubbling, homogeneous fluidization. By contrast, in aggregative fluidization, observed in solids-air experiments, the particles tend to remain aggregated, more closely resembling a liquid than a gas. Gas rises through the bed mainly in the form of bubbles and the system is heterogeneous. Jackson, 1963, developed constitutive equations for both the gas and particle phases that showed that an ideal fluidized bed is always unstable if there are no interaction forces between the particles. Rietema, 1973, showed that the cohesive force between particles in contact could stabilize the fluidized bed.

Many studies have shown that the interaction between particles and between the fluid and the particles are closely linked to the fluidization quality. Mutsers & Rietema, 1977a, b, performed experiments using two different gases (nitrogen and hydrogen) having different densities and viscosities. They were also able to achieve different effective gravity forces by carrying out their experiments in a large centrifuge (up to 20 "g") but reported experimental results only up to 3.07 "g". They found that the dense-

phase porosity and the bubble point porosity in freely bubbling beds are influenced not only by hydrodynamic effects, but also by interparticle forces. They concluded that the effective gravity has an effect on the fluidization stability and is related to fluidization quality.

Generally, fluidization stability and quality is a function of particle size and density, fluid viscosity and density, and drag force, cohesive force, and gravity force. The gravity force is clearly a very important factor, but because almost all fluidization experiments have been conducted in vertical gravity-driven beds (1 "g"), we have not seen any references in the literature analyzing the effect of gravity on particle fluidization stability and quality. This research will show the effect of a variable acceleration field ("gravity") on the Geldart powder classification and analyzes fluidization quality at different "g" forces.

#### 5.2 Powder Classification in a Conventional Gravity-Driven Fluidized Bed

This section reviews the previous work on powder classification in gravity-driven gassolids fluidized beds. In addition, a new semi-theoretical equation is presented for the transition from Geldart group A to group C particles.

#### 5.2.1 Geldart's Powder Classification

Geldart, 1973, in his well known classification of solids fluidized by gases, empirically found that the behavior of solid particles with respect to their fluidizing properties falls into four groups: A, B, C and D. These groups are characterized by the mean particle size ( $d_p$ ) and by the density difference between the particles and the gas ( $\rho_p$ - $\rho_f$ ).

Geldart observed that:

- Powders in group A exhibit dense phase expansion after minimum fluidization and prior to the commencement of bubbling. When the gas supply is suddenly cut off, the bed collapses very slowly. All bubbles rise more rapidly that the interstitial gas velocity.
- Powders in group B bubble at the minimum fluidization velocity. Bed expansion is small and the bed collapses very rapidly when the gas supply is shut off. Most bubbles rise more rapidly than the interstitial gas velocity.
- Powders in group C are difficult to fluidize at all since the interparticle forces are greater than those which the fluid can exert on the particles.
- Powders in group D can form stable spouted beds. Only the largest bubbles rise more slowly than the interstitial fluidizing gas, so that gas flows into the base of the bubble and out from the top of the bubble.

In most applications of fluidized beds, maximal heat and/or mass transfer between the fluidizing gas and the bed solids is desired. Such transfer is optimal in beds that are uniformly fluidized and not bubbling. For this reason, the fluidization obtained using group A particles is preferable to that using group B particles and both are superior to that realized with group C and group D particles. This ranking forms the basis for our discussion of fluidization quality.

Rietema, 1984, pointed out that Geldart's classification is correct only when operating at normal temperatures and pressures and in the earth's gravitational field. Other factors that must be taken into account include cohesion, gas viscosity, the adsorption of gas on the particles and gravitational acceleration. Rietema concludes by saying: "A certain powder which shows A-powder behavior might show B-powder behavior when the effective gravitation is increased or a gas of lower viscosity is used. Similarly, a powder, which shows C-powder behavior, might evince A-powder behavior when a gas of higher viscosity is used."

# 5.2.2 Criterion for the Transition between Group A and Group C Particles

Geldart, 1973, defined powders that are in any way cohesive to belong to group C. Normal fluidization of these powders is extremely difficult; the powder lifts as a plug in small diameter tubes, or channels (rat-holes) badly with the gas passing up through voids extending from the distributor to the bed surface. This difficulty arises because the interparticle forces are greater than the drag force that the fluid exerts on the particle. The interparticle forces are generally the result of very small particle size (van der Waals forces), strong electrostatic charges (Coulombic forces), or the presence in the bed of wet or sticky material (capillary forces). Geldart presented only an empirical curve (with no defining equation) to distinguish between group A and C particles in his classification.

Molerus, 1982, derived a semi-empirical equation to describe the transition between group A and C particles assuming that the product of the Euler number for fluidization and the Reynolds number is a constant ( $\approx 5.0$ ) for small Reynolds numbers and the voidage,  $\varepsilon \approx 0.5$ . However, it is unclear whether this equation can be extended to different accelerations. Therefore, in order to analyze the behavior of the transition between group A and C particles under a centrifugal field, a new semi-theoretical equation will be developed. The critical difference between group A and C particles is that group A particles can be fluidized individually, whereas group C particles will tend to agglomerate because of interparticle cohesive forces and can only be fluidized as agglomerates. Therefore, we will focus on a single particle inside the bed, which is acted upon by a drag force, a gravity and buoyancy force, and a cohesive force. We will analyze each of these forces separately below.

**Drag force:** If the relative velocity between a single particle and the surrounding fluid is u and the drag force on the particle is  $F_d$ , then the drag coefficient,  $C_D$ , is defined by

$$F_{d} = C_{D} \frac{\rho_{f} u^{2}}{2} \frac{\pi d_{p}^{2}}{4}$$
(5.1)

At the terminal velocity,  $u_t$ , for a single falling particle, the drag force is balanced by the net force due to gravity and buoyancy, giving

$$F_{dt} = C_D \frac{\rho_f u_t^2}{2} \frac{\pi d_p^2}{4} = \frac{\pi d_p^3}{6} (\rho_p - \rho_f) g$$
(5.2)

Khan and Richardson, 1990, have shown that if a particle is in a suspension of likeparticles supported by a fluid having a superficial flow velocity, u, the drag force on a single particle in the suspension of voidage,  $\varepsilon$ , is  $\varepsilon^{-4.8}$  times that on a single isolated particle. This applies to both the Stokes' and Newton's law regions. Thus:

$$F_{d\varepsilon} = C_{D} \frac{\rho_{f} u^{2}}{2} \frac{\pi d_{p}^{2}}{4} \varepsilon^{-4.8}$$
(5.3)

Gravitational force  $W_g$  and buoyancy force  $W_b$ : When the bed is in the fluidized state, it behaves as a fluid and the density of the "fluidized bed" (solids and gas) is:

$$\rho_a = \varepsilon \rho_f + (1 - \varepsilon) \rho_p \tag{5.4}$$

The buoyancy force  $W_b$  on a spherical particle is given by:

$$W_b = \frac{\pi d_p^3}{6} \rho_a g = \frac{\pi d_p^3}{6} (\varepsilon \rho_f + (1 - \varepsilon) \rho_p) g$$
(5.5)

and the "effective weight" of the particle is:

$$W_{e} = W_{g} - W_{b} = \frac{\pi d_{p}^{3}}{6} \rho_{p} g - \frac{\pi d_{p}^{3}}{6} (\varepsilon \rho_{f} + (1 - \varepsilon) \rho_{p}) g = \frac{\pi d_{p}^{3}}{6} (\rho_{p} - \rho_{f}) g \varepsilon$$
(5.6)

in agreement with Foscolo and Gibiloro, 1984.

**Cohesive force:** The dominant interparticle force between two particles is the van der Waals force. The van der Waals force between two perfectly spherical and rigid particles of diameter  $d_1$  and  $d_2$  can be expressed as:

$$F_{c} = \frac{A}{12\delta^{2}} \frac{d_{1}d_{2}}{d_{1} + d_{2}}$$
(5.7)

where A is Hamaker's constant, which, for most solids, has been calculated to have an average value of  $A = 10^{-19}$  J (Rietema et al., 1993). Here  $\delta$  is the distance between the two particles and can be estimated to be between 0.15 and 0.4 nm (Krupp, 1967). Normally  $\delta$  is taken as 0.4 nm (Iwadate et al., 1998, Zhou and Li, 1999). If we assume that the particles have the same size,  $d_1 = d_2 = d_p$ , then Equation (5.7) becomes:

$$F_c = \frac{A}{12\delta^2} \frac{d_p}{2}$$
(5.8)

But the calculated cohesive force using Equation (5.8) is extremely high (Massimilla and Donsi, 1976, Rietema et al., 1993) when compared to experimental values. Most powders, however, have a rather rough surface with many protuberances,

generally called asperities with radii of curvature often not larger than 0.1  $\mu$ m. Therefore, the value of  $d_p$  in Equation (5.8) should be replaced by the diameter of the asperities, l(Rietema et al., 1993). This will reduce the value of the cohesive force and make it much more realistic. Thus, the effective average cohesive force depends strongly on the surface structure of the particles, i.e., the size distribution of the asperities and their surface density. Consequently, Equation (5.8) becomes:

$$F_{c} = \frac{A}{12\delta^{2}} \frac{l}{2} = \frac{Al}{24\delta^{2}}$$
(5.9)

Massimilla and Donsi, 1976, suggest using a mean value of l equal to 0.2  $\mu$ m.

**Transition equation between group A and group C:** Group A particles can be fluidized individually. At the maximum flow velocity,  $u_m$  (before the particles flow out of the bed) for group A particles,  $F_{d\varepsilon} > W_e + F_c$ , and for group C particles, since the particles form agglomerates,  $F_{d\varepsilon} < W_e + F_c$ . Figure 5.1 is a conceptual representation showing all the pertinent forces acting on a particle on the fluidized bed.

The transition equation can be expressed as:

$$F_{d\varepsilon}|_{u=u_m} = W_e|_{u=u_m} + F_c|_{u=u_m}$$
(5.10)

If we substitute Equations (5.3), (5.6) and (5.9) into (5.10), we obtain:

$$C_{D} \frac{\rho_{f} u_{m}^{2}}{2} \frac{\pi d_{p}^{2}}{4} \varepsilon^{-4.8} = \frac{\pi d_{p}^{3}}{6} (\rho_{p} - \rho_{f}) g \varepsilon + \frac{Al}{24 \delta^{2}}$$
(5.11)

The maximum velocity before the particles flow out of the bed, is  $u_m = u_t$ . At the terminal velocity  $u_t$ ,

$$C_{D} \frac{\rho_{f} u_{t}^{2}}{2} \frac{\pi d_{p}^{2}}{4} \varepsilon^{-4.8} = \frac{\pi d_{p}^{3}}{6} (\rho_{p} - \rho_{f}) g \varepsilon + \frac{Al}{24 \delta^{2}}$$
(5.12)

Substituting Equation (5.2) into (5.12) gives:

$$\frac{\pi d_{p}^{3}}{6} (\rho_{p} - \rho_{f}) g(\varepsilon^{-4.8} - \varepsilon) = \frac{Al}{24 \,\delta^{2}}$$
(5.13)

which is the transition equation for group A/C particles.

Here,  $A = 10^{-19}$  J,  $l = 0.2 \mu m$ ,  $\delta = 0.4 nm$ . The voidage for group C particles is different for different materials, and  $\varepsilon = 0.53$  is used here as an upper limit based on the measurement by Chaouki et al., 1985, for aerogels. The voidage,  $\varepsilon$ , for group A particles is approximately 0.4.



Figure 5.1. Conceptual representation for transition of group A/C.

For a gas-solid system in a conventional fluidized bed,  $g = 9.8 \text{ m/s}^2$ . A plot of Equation (5.13) for the voidage range,  $\varepsilon = 0.53$  and  $\varepsilon = 0.4$ , yields two straight lines on

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log paper with slope = -3 as shown in Figure 5.2. This figure and all subsequent figures are based on fluidization by air at ambient conditions. The two straight lines bracket the empirical result of Geldart, 1973, and are in close agreement with the semi-empirical boundary lines for soft and hard particles of Molerus, 1982, which represent the transition of group A/C. Thus Equation (5.13) is an excellent representation of the transition from group A to group C without requiring the specification of any additional empirical parameters. It should be noted, however, that Equation (5.13) is sensitive to the voidage and the voidage will change with the particle size. For example, the voidage was found to be as high as 0.67 for 1  $\mu$ m SiC particles (Zhou et al., 1999). Furthermore, in a centrifugal field, the voidage could be smaller than in a conventional gravity-driven fluidized bed since the centrifugal force will tend to pack the particles closer together (Mutsers et al., 1977).

We have replaced the particle diameter by the size of the asperities in Equation (5.13) to calculate the van der Waals cohesive force, but for very fine particles ( $d_p < 0.2$  µm), l should be replaced by  $d_p$ .

#### 5.2.3 Criterion for the Transition between Group A and Group B Particles

Foscolo and Gibilaro, 1984, emphatically rejected the idea that interparticle forces are responsible for the stabilization of a fluidized bed at gas flows below that required to initiate bubbling. Instead, they invoked the stability criterion presented by Wallis, 1969, that bubbles occur when conditions are such that the propagation velocity of a voidage disturbance reaches the velocity of elastic waves in the bed. In some respects this idea is analogous to the condition of a projectile reaching the velocity of sound in a fluid. Thus,

for voidage propagation velocities smaller than the elastic wave velocity,  $u_{\varepsilon} < u_{e}$ , disturbances will be accommodated in an essentially homogeneous manner, whereas for  $u_{\varepsilon} > u_{e}$ , this is no longer possible and bubbles are formed. This reasoning led to a relatively simple criterion for the instability or transition of Geldart A/B:

$$\frac{(gd_p)^{\frac{1}{2}}}{u_t} (\frac{\rho_p - \rho_f}{\rho_p})^{\frac{1}{2}} < 0.56 n(1 - \varepsilon_b)^{\frac{1}{2}} \varepsilon_b^{n-1}$$
(5.14)

where  $\varepsilon_b$  is the voidage at which the transition from particulate to aggregative fluidization occurs. For cases of practical interest,  $\varepsilon_b = \varepsilon_{mb}$ , where  $\varepsilon_{mb}$  is the voidage at the minimum bubbling velocity. Foscolo and Gibilaro, 1984, applied the stability criterion given by Equation (5.14) to define the boundary between powders that undergo a degree of particulate expansion ( $u_{mb} > u_{mf}$ ) when fluidized by air and those that do not ( $u_{mb} = u_{mf}$ ).

**Table 5.1:** Equations to calculate n and  $u_t$  for different  $\operatorname{Re}_t$ 

Re,	п	u <sub>t</sub>
$\operatorname{Re}_t < 0.2$	$n = 4.65 + 20(\frac{d_p}{D_t}) \approx 4.65$	$u_t = \frac{d_p^2 (\rho_p - \rho_f)g}{18\mu}$
$0.2 < \text{Re}_t < 1.0$	$n = [4.35 + 1.75(\frac{d_p}{D_t})] \operatorname{Re}_t^{-0.03} \approx 4.35 \operatorname{Re}_t^{-0.03}$	$u_t = \frac{d_p^2(\rho_p - \rho_f)g}{18\mu}$
$1.0 < \text{Re}_t < 2.0$	$n = [4.45 + 18(\frac{d_p}{D_t})] \operatorname{Re}_t^{-0.1} \approx 4.45 \operatorname{Re}_t^{-0.1}$	$u_t = \frac{d_p^2(\rho_p - \rho_f)g}{18\mu}$
$2.0 < \text{Re}_t < 200$	$n = [4.45 + 18(\frac{d_p}{D_t})] \operatorname{Re}_t^{-0.1} \approx 4.45 \operatorname{Re}_t^{-0.1}$	$u_{t} = 0.153 \frac{d_{p}^{1.14} [(\rho_{p} - \rho_{f})g]^{0.71}}{\mu^{0.43} \rho_{f}^{0.29}}$
$200 < \text{Re}_t < 500$	$n = 4.4  \mathrm{Re_t}^{-0.1}$	$u_t = 0.153 \frac{d_p^{1.14} [(\rho_p - \rho_f)g]^{0.71}}{\mu^{0.43} \rho_f^{0.29}}$
$\operatorname{Re}_t > 500$	<i>n</i> = 2.39	$u_{t} = 1.74 \left[\frac{d_{p}(\rho_{p} - \rho_{f})g}{\rho_{f}}\right]^{0.5}$

Equation (5.14) is dependent on the terminal velocity of a single particle, the Reynolds number,  $\operatorname{Re}_{t} = \frac{\rho_{f} d_{p} u_{t}}{\mu}$  and *n* which comes from the Richardson-Zaki equation (Richardson and Zaki, 1954). The terminal velocity  $u_{t}$  can be obtained from Equation (5.2) if the drag coefficient,  $C_{D}$ , is known. The drag coefficient is given by:  $C_{D} = \frac{24}{\operatorname{Re}_{t}}$ 

for  $\text{Re}_t < 0.1$  (for engineering purposes it is often used up to  $\text{Re}_t < 2.0$ ),  $C_D = \frac{18.5}{\text{Re}_t^{0.6}}$ 

for  $2 < \text{Re}_t < 5 \times 10^2$  and  $C_D \approx 0.44$  for  $5 \times 10^2 < \text{Re}_t < 2 \times 10^5$  (Schlichting, 1979). Thus for different Re<sub>t</sub> regimes,  $u_t$  and n can be expressed by the equations in Table 1 (assuming the particle size is much smaller than the diameter of the fluidized bed,  $\frac{d_p}{D_t} \approx 0$ ).

Inserting the physical properties of air and letting  $\varepsilon_b = 0.35$  corresponding to the practical minimum value of  $\varepsilon_{mf}$  for spherical particles into Equation (5.14), the boundary dividing group A/B powders for conventional gas-solid fluidized beds is obtained and plotted in Figure 5.2. The curve is in good agreement with the result of Oltrogge, 1972, given in Geldart's 1973 paper based on an equation suggested by Verloop and Heerjes, 1970, but modified empirically to fit his experimental data. Also shown in Figure 5.2 are the transition curves between group A/B given by Molerus, 1982, for soft and hard particles.

# 5.2.4 Criterion for the Transition Between Group B and Group D Particles

Molerus, 1982, gave a simple criterion for the transition from group B to group D

powders assuming that the transition occurs with coarse-grained particles at Re » 1. From the non-dimensional representation of the drag exerted on the particles in the high Reynolds number regime, experimental data has shown that the Euler number can be taken to be constant for a given porosity. In particular, at minimum fluidization:

$$Eu_{mf} = \frac{4}{3} \frac{d_p (\rho_p - \rho_f) g \varepsilon_{mf}^2}{\rho_f u_{mf}^2} = \text{constant}$$
(5.15)

With  $\varepsilon_{mf} \approx 0.5$ ,  $\varepsilon_{mf}^2 = 0.25$ ,  $Eu_{mf} \approx 5$ ,

$$\frac{d_p(\rho_p - \rho_f)g}{\rho_f u_{mf}^2} \approx 15$$
(5.16)

According to Molerus, this result has a simple physical meaning. With high Reynolds number and therefore dominance of the inertia forces in comparison with viscous forces, the dynamic pressure of the fluid,  $\rho_f u_{mf}^2$ , must at the onset of fluidization exceed a certain value in comparison to the effective weight of the particles per unit area. From that consideration it is reasonable to assume that such systems tend to spout. Thus the dynamic pressure of the fluid must exceed a critical value  $\rho_f u_{mf}^2$  = constant with onset of fluidization so that Equation (5.16) can be written as:

$$d_p(\rho_p - \rho_f)g = \text{constant}$$
(5.17)

Using the experimental data by Mathur, 1971, the final result for the transition of group B to group D particles from Equation (5.17) is obtained as:

$$d_p(\rho_p - \rho_f)g = 15.3 \text{ N/m}^2$$
 (5.18)

This result is plotted in Figure 5.2 and agrees well with the experimental data reported by Geldart, 1973.



Figure 5.2. Powder classification for fluidization by air.
# **5.3 Powder Classification in a Centrifugal Field**

Here we will extend the results presented in the previous section for the classification of powders in gravity-driven fluidized beds to apply to fluidized beds in a centrifugal field, i.e., in rotating fluidized beds (RFB).

# 5.3.1 Criterion for the Transition between Group A and Group C Particles

Equation (5.13) is our semi-theoretical result for the transition from group A to group C particles. The g in the equation refers to the gravitational acceleration acting on the particles. In a centrifugal field, g will be replaced by the centrifugal acceleration, a, which will be given in non-dimensional form as " $g'' = a/(9.8 \text{ m/s}^2)$ .

Figure 5.3 shows the calculated boundary for the transition of group A and C particles at 1 "g", 10 "g", 100 "g" and 1000 "g". It shows that the boundary curve shifts to the left with increasing "g". It can be seen, for example, that particles exist (of specific size and density) that belong to group C at 1 "g", but belong to group A at 10 "g". The figure also shows that it should be possible to fluidize some group C particles without agglomeration in a centrifugal field, depending on the value of "g".



Figure 5.3. Group A/C transition for fluidization by air at different "g".

# 5.3.2 Criterion for the Transition Between Group A and Group B Particles

Equation (5.15), suggested by Foscolo and Gibilaro, 1984, is also seen to be dependent on the gravitational acceleration. In a centrifugal field, the gravitational acceleration is replaced by the centrifugal acceleration and Equation (5.14) can be modified accordingly. Figure 5.4 shows the calculated boundary transition curves for group A and B particles at 1 "g", 10 "g", 100 "g" and 1000 "g". It shows that the boundary curve shifts to the left with increasing "g". Thus, if group A particles can shift to group B in a centrifugal field and the additional centrifugal force (above 1 "g") will reduce the fluidization quality for group A particles.



Figure 5.4. Group A/B transition for fluidization by air at different "g".

# 5.3.3 Criterion for the transition between Group B and Group D particles

Equation (5.18) suggested by Molerus is used to calculate the transition from group B/D in a centrifugal field. Figure 5.5 shows the calculated boundary transition curves for group B/D at different "g". Clearly, in a centrifugal field, group B particles can shift to group D, which are not easy to fluidize and tend to spout. Figure 5.5 also shows that group D particles can shift to group B or even to group A if "g" is less than unity. Thus, by decreasing "g", group B particles can shift to group A, and group D particles can shift

to group B or to group A. Group D particles are large or dense and are difficult to fluidize. However, if they shift to group B or A, the fluidization quality of group D particles will be improved. This effect can be achieved by fluidization under microgravity conditions such as fluidizing particles on the moon, or by applying an external force, such as a magnetic or electric field to balance the gravity force.



Figure 5.5. Group B/D transition for fluidization by air at different "g".

## **5.4 Experiments**

In order to test the ideas and equations developed for the classification of powders in a centrifugal field presented above, a series of experiments were performed in a rotating fluidized bed where the centrifugal acceleration could be varied from 7 to over 200 "g".

# **5.4.1 Experimental System**

A schematic diagram of the experimental system used in our research is shown in Figure 5.6. Dry compressed air was fed to the RFB through a bank of calibrated rotameters. A variable-speed motor provided rotating speeds between 0 to 5000 rpm. The metal distributor (manufactured by Pall Corporation) was composed of sintered 316 low-carbon stainless steel powder with an average pore size of 55  $\mu$ m. Its length was 15.2 cm with inner radius (the same as the outer radius of the bed) of 6.15 cm. Plexiglas end plates were used so that the bed and bed surface could be observed by eye or by a high-resolution digital camera.

The rotating speed of the RFB was set at 325, 525, 700, 900, 1200, 1600 and 2000 rpm (34, 55, 73, 94, 126, 167 and 209 rad/s) which is equivalent to a centrifugal acceleration of 7, 19, 34, 56, 100, 178, and 278 "g", respectively. The strobe frequency was set to match the rotational frequency of the bed so that any macroscopic bubbles present could be observed. The pressure drop as a function of gas velocity and rotating speed was measured using a U-tube water manometer. Before loading particles into the bed, the pressure drop of the distributor of the empty bed was also measured and then subtracted from the measured pressure drop of the bed loaded with particles.



Figure 5.6 Schematic diagram of the experimental system.



Figure 5.7 Particle size distribution.

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Our experiments were conducted with Geldart A and C particles with properties listed in Table 5.2. The bed material consisted of glass beads provided by MO-SCI Corporation and alumina provided by the ALCOA Technical Center. The particle size distribution for all of these particles was measured using an Aerosizer (Amherst Process Instruments, Inc.); typical results are shown in Figure 5.7. The fresh alumina has a broad size distribution with many fines; washed alumina has far fewer fines. The glass beads have a much narrower size distribution than the alumina particles and a higher density. The 7-micron alumina particles, due to their small size, belong to Geldart group C; all the others belong to Geldart group A.

Table 5.2 Particle size, density and their Geldard's group classification

Particle	Particle Density	Mean Particle	Mean Particle	Group
	$(kg/m^3)$	Diameter	Diameter	
		(by vol.) (µm)	(by no.) (µm)	
Glass Beads	2450	82	75	Geldart A
Washed Alumina	1550	89	28	Geldart A
Fresh Alumina	1550	63	4	Geldart A
Q-705 Alumina	1320	68	48	Geldart A
7-µm Alumina	1470	7	2	Geldart C

## **5.4.2 Experimental Results**

Experiments were conducted by measuring the pressure drop of the bed as a function of gas velocity and rotating speed. Visual observations of the minimum fluidization and minimum bubbling velocity and the quality of fluidization were made by eye and by high-resolution photography.

**Group A particles:** The most easily observed difference between powders in group A and B is whether or not the bed bubbles at or very close to minimum fluidization. The

minimum bubbling velocity  $(U_{mb})$  is defined as the gas velocity when bubbles are first observed. At low gas flow rate the bed is in the packed bed regime. As the gas velocity is increased, the inner surface begins to fluidize (See Chapter 3). The inner surface fluidizes first because at that point the gas velocity in the bed is highest and the centrifugal force is lowest. The velocity at which fluidization occurs at the inner surface is called the inner surface minimum fluidization velocity,  $(U_{mfl})$ . With increasing velocity, the entire bed is fluidized at critical minimum fluidization velocity  $(U_{mfl})$ . The minimum fluidization velocities,  $U_{mfl}$  and  $U_{mfc}$  were determined from the pressure drop data as described in Chapter 3. Figure 5.8 shows the pressure drop for a 0.5 kg charge of washed alumina particles as a function of the gas flow rate at different rotating speeds. At rotating speeds greater than 325 rpm, the minimum bubbling velocity,  $U_{mb}$ , was observed before the critical minimum fluidization velocity  $(U_{mfc})$  was reached, i.e., when the entire bed has fluidized. At the rotating speed of 325rpm, the values of  $U_{mb}$ ,  $U_{mfl}$  and  $U_{mfc}$  are practically the same.



**Figure 5.8** Pressure drop due to a 0.5 kg charge of washed alumina particles as a function of flow rate at 325 rpm, 525 rpm, 700 rpm and 900 rpm.

Similar experiments have been conducted using Q-705 alumina particles and glass beads. Figure 5.9 and 5.10 show the pressure drop for a 0.5 kg charge of Q-705 alumina particles and 0.5 kg charge of glass beads, respectively, as a function of flow rate at different rotating speeds. As seen in the figures, the results are very close to those obtained using the washed alumina.



**Figure 5.9** Pressure drop due to a 0.5 kg charge of Q-705 alumina particles as a function of flow rate at 325 rpm, 525 rpm, 700 rpm and 900 rpm.



**Figure 5.10** Pressure drop due to a 0.5 kg charge of glass beads as a function of flow rate at 325 rpm, 525 rpm and 700 rpm.

The above experiments indicate that in all cases  $U_{mf}$  and  $U_{mb}$  are very close or practically equal. Here  $U_{mfc}$  is the average of  $U_{mfi}$  and  $U_{mfc}$ . In Figure 5.4, we have marked the mean size and density of the particles used in each of the above experiments. According to the figure, for all of our centrifugal accelerations ("g"), the particles behave as Geldart group B. However, in Geldart's classification, for a gravity driven bed, these particles should behave as group A. Thus our experiments imply that group A particles can shift to group B in a centrifugal field.



**Figure 5.11** Pressure drop due to a 0.5 kg fresh alumina particles charge as a function of flow rate at 325 rpm, 525 rpm, 700 rpm and 900 rpm.

The effect of fines: Geldart and Abrahamsen, 1980, studied the effect of fines on the behavior of conventional gas-fuidized beds. They found that the presence of a small mass fraction of fines below 45  $\mu$ m strongly affects and increases the ratio  $U_{mb}/U_{mf}$ . Using fresh alumina particles, which contains significant fines below 45  $\mu$ m, we measured the pressure drop of the bed and observed the onset of bubbling,  $U_{mb}$ . Figure 5.11 shows the pressure drop for a 0.5 kg charge of fresh alumina particles as a function of flow rate at

**Fluidization of group C particles:** The ability to fluidize of group C particles is potentially significant because of the rapidly growing need for using ultra-fine particles in industrial applications. There are two basic methods to improve the fluidization quality of fine cohesive group C particles. One is by application of an external force, such as vibration or from a magnetic, electric or acoustic field. The other is by altering the intrinsic properties of the particles, e.g., modifying their surface characteristics or mixing them with other particles having different size or shape (Wang et al., 1998, Mori et al., 1992, Chirone et al., 1992, Zhu et al., 1994, Tung et al., 1989). From the theoretical analysis presented above, we have shown that group C particles can shift to group A in a centrifugal field and thus can be more easily fluidized. This is a new approach to improving the fluidization quality of fine particles and requires experimental verification.

We therefore carried out experiments in our rotating fluidized bed using 7  $\mu$ m Geldart group C alumina particles. These particles normally tend to agglomerate. In our experiments, a 0.3 kg charge of the 7  $\mu$ m alumina particles was loaded into the RFB. The rotating speeds were set at 325, 525, 700, 900, 1200, 1600 and 2000 rpm, equivalent to 7, 19, 34, 56, 100, 178, and 278 "g" respectively. At each rotating speed, the pressure drop of the bed was measured as a function of gas flow rate. The strobe frequency was set to match the rotational frequency of the bed and the bed condition was observed by eye.

Figure 5.12 shows the measured pressure drop in the bed as a function of flow

rate at different rotating speeds. Figure 5.13 summarizes our observations of the bed condition corresponding to the last data point at each rotating speed shown in Figure 5.12.



Figure 5.12 Pressure drop due to a 0.3 kg charge of 7  $\mu$ m alumina particles as a function of flow rate at 325 rpm, 525 rpm, 700 rpm, 900 rpm, 1200 rpm and 1600 rpm.



a. 325 rpm (7 "g"),  $U_g = 0.080$  m/s



c. 700 rpm (34 "g"),  $U_g = 0.134$  m/s



e. 1200 rpm (100 "g"),  $U_g = 0.134$  m/s



b. 525 rpm (19 "g"),  $U_g = 0.121$  m/s



d. 900 rpm (56 "g"),  $U_g = 0.161$  m/s



f. 1600 rpm (178 "g"),  $U_g = 0.161$  m/s

**Figure 5.13** Visual observation of the powder in the RFB corresponding to the data point at the highest velocity shown in Figure 5.13.

At 325 rpm (7 "g") and 525 rpm (19 "g"), the particles are observed to be agglomerated and the bed surface is non-uniform. At higher gas velocity, the gas passes through large regular channels, which are somewhat smaller at higher rpm. This is a normal observation for group C particles in a conventional fluidized bed. The pressure drop increases with gas velocity in the packed bed regime until the pressure drop curve reaches a plateau due to channelling.

At 700 rpm (34 "g"), the particles appear to have de-agglomerated in the bed, although the bed surface is still not completely uniform. At higher gas velocity, very small channels could still be observed. It is hard to say whether the particle behavior belongs to group C or group A.

At 900 rpm (56 "g") the particles have totally de-agglomerated and the bed surface is uniform. At higher gas velocity, the particles behave like group A or B particles and the bed appears uniformly fluidized without channelling as illustrated in Figure 5.13d. In Figure 5.12 the pressure drop increases with gas velocity in the packed bed regime and goes to a plateau after minimum fluidization is reached. Thus, between 700 and 900 rpm, the particles shift from group C to group A or B behavior.

At 1200 rpm (100 "g") and 1600 rpm (178 "g"), the bed surface is very uniform in the packed bed regime because of the high rotating speed. The pressure drop increases steadily, then, as the gas flow is further increased, it decreases suddenly. At this point, a "rat hole" is observed inside the bed. The gas passes through the rat-hole instead of flowing between the particles. The difference between operating at 1200 rpm and 1600 rpm is that the rat hole size at 1600 rpm is larger than that at 1200 rpm. The gas velocity at which the pressure drop suddenly falls due to the formation of the rat hole at 1600 rpm is higher than that at 1200 rpm. Similarly, the rat-hole size at 2000 rpm (278 "g") is even larger than at 1600 rpm and the corresponding gas velocity, when the pressure drop suddenly falls, is higher. These results indicate that the particles are not easy to fluidize at high centrifugal force. It should be noted that at very high rpm the pressure drop curve looks like that obtained for group D particles rather than for group A or B particles.



Figure 5.14 Powder classification for fluidization by air at 56 "g".

Since the voidage in the bed is probably lower than 0.4 due to the centrifugal field (Mutsers et al., 1977b), we used a voidage of 0.35 for the transition curve between group A and C. Figure 5.14 shows that the transition boundary between group B/D has shifted

further to the left than the transition boundary between group A/B. Thus there is no transition between group A and B at high "g". Rather, the 7  $\mu$ m alumina group C particles shift directly into group B, bypassing group A altogether. In Figure 5.15, which shows the results at a rotating speed of 2000 rpm (278 "g"), the 7  $\mu$ m alumina particles have shifted from group C to group D. This observation explains why we did not observe good fluidization of the 7  $\mu$ m alumina particles at rotating speeds of 1200 (100 "g"), 1600 (178 "g") and 2000 rpm (278 "g").



Figure 5.15 Powder classification for fluidization by air at 278 "g".

### **5.5 Discussion**

It is clear from both our theoretical analyses and our experimental results in the RFB that Geldart's powder classification cannot be used in centrifugal fields without modification to take into account gravity forces greater that 1 "g". As discussed in Iwadate and Horio, 1998, cohesive group C particles can only be fluidized in the form of agglomerates. They define this type of fluidization as "agglomerating fluidization". Agglomerating fluidization is characterized by plug formation or channeling in the bed. The transition that group C particles cannot be fluidized individually and group A particles can is the basis for Equation (5.13), which we have used to determine the transition between group A and C particles. Thus the essential characteristic of group C particles is that they are in the agglomerating fluidization regime.

The major difference between group A and B particles is whether bubbles appear at the minimum fluidization velocity or after the minimum fluidization velocity is reached. Thus both group A and B particles give rise to stable fluidization.

Geldart, 1973, obtained an equation for the transition of group B to group D by calculating the density/particle size combination of powders for which bubbles less than a given size would rise more slowly than the interstitial gas velocity. He chose a very large bubble size (25 cm) for his calculation. Molerus, 1982, developed his semi-empirical equation for the transition between group B and D from spouted bed data. Beds exhibiting very large bubbles or spouted bed are both unstable fluidized beds. Thus group D particles give rise to unstable fluidization.

Based on the definitions given above, we can rearrange Equations (5.13) and (5.18) to give:

$$(\rho_p - \rho_f)\frac{g}{9.8} = \frac{Al}{24\delta^2} \frac{6}{\pi d_p^3} \frac{1}{(\varepsilon^{-4.8} - \varepsilon)} \frac{1}{9.8}$$
(5.19)

$$(\rho_p - \rho_f)\frac{g}{9.8} = \frac{15.3}{d_p}\frac{1}{9.8}$$
(5.20)

where g/9.8 represents the dimensionless strength of the centrifugal field ("g").

If we plot the left-hand side of Equations (5.19) and (5.20) as the ordinate and  $d_p$  as the abscissa we obtain Figure 5.16. This figure shows the effect of different "g" conditions on the powder classification chart, e. g., on the earth, on the moon, in a centrifugal field and in deep space. Here we have indicated that particles in group A and B (cross-hatched area) will result in stable fluidization. For example, if we choose particles of density 1 g/cm<sup>3</sup>, the figure shows that particles between 60 to 10,000 µm would achieve stable fluidization on the moon, whereas particles between 30 to 1500 µm would achieve stable fluidization on the earth. For rotating fluidized beds, the figure shows that it will be difficult to achieve stable fluidization at above 100 "g" except in a very narrow particle size range.



Figure 5.16 Powder classification for fluidization by air at different "g".

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#### **CHAPTER 6**

# SOOT REMOVAL FROM DIESEL ENGINE EXHAUST USING ROTATING FLUIDIZED BED FILTER

### **6.1 Introduction**

In spite of a concerted effort to develop diesel engine emission control systems, none has been found that can continuously remove both soot and NO<sub>X</sub>. The emissions from diesel engines include solid, liquid, and gaseous components. The combined solid and liquid components are called particulates. Heck and Farrauto, 1995, reported that the composition of the particles is typical 43% dry soot (primarily carbonaceous material), 2% SO<sub>3</sub>+H<sub>2</sub>O, 20% fuel-derived components and 35% lube oil derived components. The soot components that are soluble in organic solvents are called the soluble organic fraction (SOF) and have their origin in fuel and lube oil. The gaseous phase consists of gases that affect human health such as unburned hydrocarbons (UHC), carbon monoxide (CO), nitrogen oxides (NO +NO<sub>2</sub>) and sulfur oxides (SO<sub>2</sub> + SO<sub>3</sub>) and are considered pollutants. The bulk of the gases emitted is considered non-hazardous and includes nitrogen (N<sub>2</sub>) oxygen (O<sub>2</sub>), water vapor (H<sub>2</sub>O) and carbon dioxide (CO<sub>2</sub>). However, CO<sub>2</sub> is a greenhouse gas that is believed to induce global climate change.

Needham, et al., 1991, measured a typical composition of diesel particulate emitted using the US Federal Transient Procedure (FTP) test. For current turbocharged and aftercooled heavy duty direct injection (DI) engines, the mass composition is: 41% carbon, 14% sulphate and water, 7% derived from fuel that was not completely combusted, 25% from lubricating oil that was not completely combusted and 13% other. The average total particulate emissions are 0.37 g/hp·h. This result is supported by results from other researchers (Cartellieri and Herzog, 1988, Springer, 1988). 
 Table 6.1 Soot size measurements in the literature

Reference	Diesel Engine	Measurement Methods	Soot Size
Baumgard et al., 1996	1988 and 1999 models	Electrical Aerosol Analyzer	Nuclei mode: 0.0075 – 0.046 µm
		(EAA)	Accumulation mode: $0.046 - 1.0 \ \mu m$
Vuk et al., 1976	Caterpillar 3150		Individual mean size: 0.025 µm
			Agglomerate size: 0.1 – 1.0 µm
Johnson et al., 1994			Solid carbon spheres: $0.01 - 0.08 \ \mu m$
			Agglomerates: 0.05 – 1.0 µm
Suarez et al., 1998	Heavy-duty	Iridium (III) tracer	0.22 – 1.8 μm
Panne et al., 1995	Diesel passenger cars		$0.14 - 0.17 \ \mu m$
Kerminer et al., 1997	Diesel passenger car		0.09, 0.2 and 0.8 µm
Richkeard et al., 1996	Europe Light duty vehicles	Quartz Crystal Microbalance	Number mean: 0.05 µm
		Scanning Mobility Particle Sizer	Volume peak: 0.1 µm, 80% less than
			0.8 μm
Arcoumanies et al., 1994	2.5 liter direct injection	TSI 3932C differential mobility	Volume distribution: $0.05 - 0.8 \ \mu m$
		particle sizer (DMPS)	
Wenninger et al., 1999		Standard dilution tunnel	Originally formed soot: $10 - 30$ nm
			Agglomerates: $0.1 - 0.2 \ \mu m$
Zhu et al., 1998	2.3 liter 4-cylinder Nissan	TEM Microscope	Mass mean: 1.23 µm
	diesel engine		

Table 6.1 provides information on the soot size measurements reported in the literature. Normally, individual diesel soot particles are very small, below 0.1  $\mu$ m. However, the small individual soot particles will tend to agglomerate in the exhaust stream. The agglomerated soot size is often between 0.1 to 1.0  $\mu$ m.

Several technologies have been studied for diesel engine emission control, including the monolith or honeycomb, fibrous filter, cyclone and packed bed collectors. These have been presented and discussed in the Chapter 1.

Konstandopoulos et al., 1989, modeled a ceramic trap as a packed bed collector and showed that the collection efficiency for nuclei mode particulate is almost 100% since the diffusion collection mechanism is extremely efficient for particles below 0.1 Very little research has been performed on the application of granular beds um. including packed and fluidized beds for diesel engine soot removal. D'Ottavio and Goren, 1983, reported single grain capture efficiency for 0.6 to 4.5 µm solid and liquid aerosol particles at high speed gas flow through clean granular beds. Their data show that when impaction is the dominant capture mechanism, the single grain capture efficiency depends on the Stokes number, Reynolds number and the fraction of collecting solids in the bed. The collection efficiency, at some conditions, can reach 90%. Clift, 1983, reviewed some fundamental problems arising in the filtration of gases by packed (fixed beds). For some applications, cake on the upstream face of the filter will build-up and increase pressure drop. Fluidized bed filters can avoid the formation of cake since the particles are fluidized. Clift et al., 1981, analyzed filtration in a fluidized bed. Once the gas velocity is above the minimum fluidization velocity, bubbles appear and aerosol bypass decreasing collection efficiency.

Pfeffer and Hill, 1978, suggested the use of a rotating fluidized bed as a filter (RFBF). The drag force of the gas, which flows radially inward through the porous wall acting on the bed particles, is balanced by the centrifugal force caused by the rotation of the bed. Therefore, the minimum fluidization velocity increases as the rotating speed is increased so that the formation of bubbles, which can cause bypass of soot particles, can be avoided even at high gas flow rate (Pfeffer et al., 1986). Another innovative advantage of using a RFBF is that catalyst granules can be used to filter soot as well as catalyze its oxidation by  $NO_X$  and  $O_2$  at high temperatures, with a concomitant reduction of  $NO_X$  to N<sub>2</sub>. This mode of operation can make the bed self-cleaning and avoid the need for regenerating the granules as they become saturated with soot. In this mode, we refer to the system as a rotating fluidized bed reactor (RFBR). We believe that the use of an RFBR has promise for treating the emissions from diesel engines which contain both particulate (soot) and gaseous pollutants by using a catalyst to promote the reduction of NO<sub>X</sub> with soot to form N<sub>2</sub>. This chapter summarises the results of a study of the filtration efficiency of soot removal from diesel engine exhaust using a horizontal RFBF as a function of exhaust flow rate and soot concentration and size distribution.

#### **6.2 Experimental System**

The experimental set-up used in this study is shown in Figure 6.1a and b. Figure 6.2 is the temperature and exhaust flow rate from our diesel engine (1.8 litre 1981 Volkswagen Rabbit Diesel Engine) as a function of engine load and RPM. In Figure 6.1a, the exhaust gas from the diesel engine (which is connected to a dynamometer) goes through calibrated orifice meters, a heater and the RFBF. In Figure 6.1b, the exhaust gas from the



a. for engine at idle conditions

b. for engine at high load & high RPM conditions

Figure 6.1 Flow sheets of experimental set-up used in this research.



Figure 6.2 1981 Volkswagen Rabbit diesel engine exhaust temperature and flow rate as function of engine load and RPM.

diesel engine goes directly to the RFBF, and then to the calibrated orifice meters (without the heater). The difference between these two systems is that in the second configuration, the RFBF is connected directly to the engine. Because the exhaust gas temperature at idle may not be high enough for a catalytic reaction, e. g., when engine load is 3 ft-lb, and rotating speed is 1000 RPM, the exhaust temperature is around 100°C as shown in Figure 6.2a, the heater in Figure 6.1a was used to raise the temperature to catalyst operating conditions. The filtration efficiency experiments for soot removal at engine idle conditions are mainly conducted in the configuration illustrated in Figure 6.1a. At high engine load and RPM condition, for example, when the engine load is 54 ft-lb, and the rotating speed is 1700 RPM, the exhaust temperature can reach more than 600°C and is sufficiently high for a catalytic reaction to occur. At these conditions, experiments were conducted in the configuration shown in Figure 6.1b. In order to reduce the heat loss, the heater was bypassed and the length between the outlet from the engine and the inlet to the RFBF was reduced to only 6 feet.

A variable-speed motor (AC Motor, Lincoln Electric, M/N: RN4S0.5TC61Q15) provides rotating speeds for the RFB between 0 to 5000 rpm. Two sintered metal distributors with average pore sizes of 55 and 100  $\mu$ m were used. In addition, a perforated distributor (open rate of 52%) covered on the inside with a 325 mesh (45  $\mu$ m) stainless steel screen was also tested. The soot concentration and particle size distribution is measured before and after the RFBF with a TSI Aerosizer system. This instrument provides particle size measurements from 0.1  $\mu$ m to 200  $\mu$ m vs. particle number distribution as well as particle mass distribution. The mass average filtration efficiency  $\eta$ , is calculated based on the particle mass concentrations before and after the RFBF:

$$\eta = \frac{C_{in} - C_{out}}{C_{in}} \times 100\% \tag{6.1}$$

where  $C_{in}$  is the particle mass concentration entering the RFBF and  $C_{out}$  is the particle mass concentration leaving the RFBF over the same period of time.

The filtration efficiency of a single size particle,  $\eta_i$ , is calculated based on the particle numbers for this size of particle before and after the RFBF:

$$\eta_{i} = \frac{N_{i,in} - N_{i,out}}{N_{i,in}} \times 100\%$$
(6.2)

where  $\eta_i$  is the filtration efficiency for the ith particle size,  $N_{i,in}$  is the number particles of size i before the RFBF over a measured period of time,  $N_{i,out}$  is the number particles of size i after the RFBF over the same period of time.



- a. direct sampling method
- b. isokinetic sampling method

Figure 6.3 Sampling methods in the experiments.

Figure 6.3 is the sampling methods used in this Chapter. Direct sampling method is "T" connection between the sampling tube and stream pipe. The Aerosizer sucks the gas at a constant sampling gas velocity. For isokinetic sampling method, the sampling gas velocity to Aerosizer equals the gas velocity in the stream pipe. For engine idle condition, the experimental results measured by direct sampling method. The comparison of these two methods is shown in Figure 6.17 and 6.18.

The filter granules consist of either glass beads obtained from MO-SCI Corporation or alumina provided by the ALCOA Technical Center. The particle size distribution for all of these particles was measured using the TSI Aerosizer; typical results are shown in Table 2.2 in Chapter 2. Bulk alumina has a broad size distribution with many fines. Q-705 alumina is a high quality alumina powder that contains fewer fines. The glass beads have a much narrower size distribution without any fines. The 7-micron alumina particles due to their small size belong to Geldart group C and also have a narrow size distribution. We also conducted experiments using Cu-ZSM-5 catalyst manufactured by ExxonMobil with a bulk density is 751 kg/m<sup>3</sup>.

#### **6.3 Results and Discussion**

As shown in Figure 6.2a, b, the engine exhaust temperature and flow rate is a function of engine load and RPM. The soot size and concentration changes with engine operating conditions. Figure 6.4 shows the soot size number distribution at the inlet of the RFBF, at engine idle and also at high load, measured with the Aerosizer over a 90 seconds period. The number of soot particles at high load and at 1600 RPM is much higher than at idle conditions. Normally, the soot particle size is below 2  $\mu$ m at idle condition, but both soot size and concentration increase at high load conditions. It should be noted that soot size from a modern diesel engine is reported to be on the order of 0.02  $\mu$ m. However, the measured soot size entering the RFBF is much larger because of agglomeration during the relatively long time of transport in the pipe between the engine and the RFBF

(Baumgard et al., 1996, Vuk et al., 1976, Kerminer et al., 1997). Another reason for the relatively large soot size is that our diesel engine is about 20 years old. Filtration efficiency of soot in the RFBF has been studied using different distributors and different filtration granules at engine idle and at high load conditions.



**Figure 6.4** Soot size number distribution at the inlet to the RFBF at different engine condition measured by the Aerosizer in 90 seconds. Sampling flow rate is 2.0 ml/min.

### 6.3.1 Sintered Metal Distributor at Engine Idle Condition

The sintered metal distributor will distribute the diesel exhaust (gas plus soot) into the RFBF much more uniformly than the slotted or perforated distributors (Qian et al., 1998). In experiments using the 50  $\mu$ m sintered metal distributor, pressure drop vs. diesel exhaust gas velocity data were measured as a function of time, and plotted in Figure 6.5. After 40 minutes, the bed pressure drop continues to increase with time even when the

gas velocity remains relatively constant, but the pressure drop continues to increase, indicating the soot plugs the pores of the distributor. Pore plugging is a consequence of the tortuous flow path in the sintered metal distributor. Soot, which is very wet and contains incompletely combusted fuel and leaking lubricating oil, tends to cake in the pores.



Figure 6.5 Variation of gas flow rate and total pressure drop (including bed and distributor) as a function of time on line.

Similar experiments were performed using the sintered metal distributor having an average pore size 100  $\mu$ m. After 60 minutes of operation (higher gas flow rate than that in Figure 6.5), the pressure drop starts to increase and the pores of sintered metal distributor eventually also become plugged. Consequently, filtration efficiency must be measured before distributor plugging occurs.

Figure 6.6 presents the mass average filtration efficiency as a function of superficial gas velocity for a filter media of 0.4 kg alumina using the 100  $\mu$ m sintered metal distributor at engine idle conditions. For the packed bed mode (U<sub>g</sub> < U<sub>mf</sub> = 0.18 m/s), the mass average filtration efficiency is seen to increase with gas velocity up to U<sub>mf</sub>. After the bed fluidize (U<sub>g</sub> > U<sub>mf</sub> = 0.18 m/s), the filtration efficiency quickly drops off to zero. This is because the alumina granules contained fines which appear in the outlet stream region after fluidization, obscuring the actual quantities of soot leaving the RFBF.



Figure 6.6 Mass average filtration efficiency as a function of superficial gas velocity.

# 6.3.2 Perforated Distributor at Engine Idle Condition

In order to understand the plugging problem, a perforated distributor (see Chapter 2) covered with a 45  $\mu$ m stainless steel screen is used. The pores of the screen present no tortuous path for the exhaust gas, but the screen is still plugged by soot after 160 minutes as shown in Figure 6.7. This plugging occurs because the soot is very wet and contains

fuel and lube SOF. At engine idle condition, the temperature of the exhaust is low and the wet soot stays on the surface of screen. The total (bed plus distributor) remains approximately constant until 160 min. At that time, the pressure drop increases dramatically. The pressure drop continues to increase from that time forward, indicating the further accumulation of sot on the distributor. This behavior is in contrast to that observed for the sintered metal distributor in which the plugging of its tortuous paths occurs more gradually. Therefore filtration efficiency experiments at engine idle condition must be done before the screen has plugged.



Figure 6.7 Variation of gas flow rate and total pressure drop as a function of time on line using the perforated distributor covered with a 45  $\mu$ m screen.

Alumina Particles: In a set of experiments using 0.5 kg alumina particles as the bed material, the soot concentration is measured at the inlet and the outlet as a function of gas
velocity. At gas velocities below the minimum fluidization velocity, we operate in the packed bed mode. When we increase the gas velocity, the inner surface begins to fluidize and as the velocity is increased, the entire bed becomes fluidized at the critical minimum fluidization velocity. The inner surface minimum fluidization velocity  $U_{mfi}$  (at the surface of the bed) and the critical minimum fluidization velocity  $U_{mfc}$  (at the distributor) are calculated based on the theoretical analysis given by Section 4.3. The filtration efficiency versus soot size distribution is given in Figure 6.8. Here, the measurement diameter of soot is based on the density of soot of 1.25 g/cc and assume the soot sphericity equal 1. It shows that the filtration efficiency for soot particles larger than 2  $\mu$ m or smaller than 0.2 µm is high in the packed bed mode. The filtration efficiency is higher at the inner surface minimum fluidization velocity (0.176 m/s) than that at a lower gas velocity (0.100 m/s). For the total fluidization mode ( $U_g > U_{mfc}$ ), the filtration efficiency for particles below 0.5  $\mu$ m is higher than that for the packed bed mode, but quickly drops off to zero after 0.5  $\mu$ m. This is because the alumina granules contained fines between 0.7 and 1.5  $\mu$ m which appear in the outlet stream region after total fluidization.

In Figure 6.9, the mass average filtration efficiency is seen to increase with gas velocity up to  $U_{mfi}$ , but drops to zero because the alumina fines are blown out with the soot. If we increase the rotating speed of the RFBF to 2000 rpm at the superficial gas velocity  $U_g = 0.24$  m/s, and then subsequently decrease the gas velocity, the bed reverts back to the packed bed mode (calculated  $U_{mfi} = 0.8$  m/s,  $U_{mfc} = 1.0$  m/s when rotating speed is 2000 rpm), the alumina fines stay inside the bed and the filtration efficiency improves. Figure 6.8 also shows that filtration efficiency goes through a minimum for particles in the 0.3 to 0.6 µm range in the packed and fluidized bed modes, and the

minimum shifts to smaller particle size as the bed becomes fluidized.



Figure 6.8 Mass average filtration efficiency as a function of soot using the perforated distributor.



Figure 6.9 Mass average filtration efficiency as a function of superficial gas velocity using the perforated distributor.

**Glass Beads:** In order to verify that the reason the filtration efficiency in the fluidization mode drop to zero is due to the effect of alumina fines, experiments were conducted using glass beads without fines present in the filtration media. In a set of experiments using 0.5 kg glass beads as the bed material, the soot concentration was measured at the inlet and the outlet as a function of gas velocity. The inner surface minimum fluidization



**Figure 6.10** Particle size number distributions at inlet and outlet to the RFBF as a function of superficial gas velocity using the perforated distributor.

The particle size number distributions at the inlet and outlet of the RFBF at different superficial gas velocity are shown in Figure 6.10. The number of soot particles at the outlet is lower than that at the inlet since much of the soot has been captured by the

glass beads. The filtration efficiency versus soot size number distribution is calculated and shown in Figure 6.11.



Figure 6.11 Filtration efficiency as a function of soot using the perforated distributor.

Figure 6.11 shows that the filtration efficiency in the fluidized bed mode is higher than that in the packed bed mode without the complication of fines emission when using alumina. The mass average filtration efficiency increases with gas velocity and it reaches 90% in the fluidization mode as shown in Figure 6.12. Figure 6.11 also shows that the filtration efficiency for soot particles larger than 1  $\mu$ m or smaller than 0.2  $\mu$ m is very high. However, the filtration efficiency again goes through a minimum for particles in the 0.3 to 0.6  $\mu$ m range. This result is consistent with Figure 6.8. This is a consequence of particles larger than 0.6  $\mu$ m being removed mainly by inertial impaction and interception and particles smaller than 0.2  $\mu$ m mainly by diffusion. This is in agreement with a number of theoretical analyses of filtration efficiency in granular beds [Clift et al., 1981, Paretsky et al., 1971].



Figure 6.12 Mass average filtration efficiency as a function of superficial gas velocity using the perforated distributor.



Figure 6.13 Schematic diagram of filtration mechanism for different size particles (Strauss, 1975).

Figure 6.13 shows a schematic diagram of the three classical filtration mechanisms for granules to capture small particles. There are inertial impaction, interception and Brownian diffusion. Some inertial impaction is a function of the Stokes number, the filtration efficiency is higher for larger particles and higher gas velocities. Interception is a function of bed porosity and the ratio of particle diameter to granule diameter. If the bed porosity is low or the ratio of particle diameter and granule diameter is large, the filtration efficiency by this mechanism will be high. For very small particles, especially under 0.1  $\mu$ m and low velocities, the Brownian diffusion mechanism is controlling. As the particles become smaller, filtration efficiency becomes higher. The total filtration efficiency is the combination of all of the mechanisms so that the filtration efficiency is high for large (inertia) or small (diffusion) particles, but goes through a minimum at around 0.3  $\mu$ m when neither capture mechanism is dominant.



Figure 6.14 Color difference due to soot deposit on filter media (80  $\mu$ m glass beads) vs. time using the perforated distributor.

During filtration experiments, glass beads from the RFBF are sampled every half hour. Figure 6.14 shows the color of glass beads samples versus time measured by a spectrophotometer. At the beginning of the experiment, the glass beads are clear and clean. As soot deposits on the surface of glass beads, they become progressively darker. The figure shows that the glass beads keep on getting darker and therefore are continually capturing soot for at least 160 minutes. The soot mass per glass beads surface area can be measured by TGA (Thermogravimetric Analyzer) according to the mass of soot and the surface area of glass beads. The mass of soot per glass beads surface area is also shown in Figure 6.14. It shows that the soot mass per surface area of glass beads increases with time.



**Figure 6.15** Mass average filtration efficiency as a function of superficial gas velocity at different rotating speeds using the perforated distributor.

Figure 6.15 is the mass average filtration efficiency as a function of superficial gas velocity at different rotating speeds using the perforated distributor when 0.5 kg glass beads are loaded. At 525 rpm, after fluidization which occurs ( $U_{mfc} = 0.10 \text{ m/s}$ ), at a superficial gas velocity of 0.14 m/s, bubbles containing soot bypass the granules and cause the filtration efficiency to drop. At 700 and 900 rpm, the filtration efficiency increases with superficial gas velocity. In our experiments, we did not reach the point where filtration efficiency dropped significantly due to bypassing the filtration media with gas bubbles. At these values of rotating speed, the minimum fluidization velocities were 0.18 m/s and 0.29 m/s.



Figure 6.16 Mass average filtration efficiency as a function of superficial gas velocity using the perforated distributor.

Figure 6.16 presents the mass average filtration efficiency when operating the RFBF with 0.75 kg 0.8 mm glass beads at 900 rpm. The filtration efficiency is very low because these large glass beads have a very small surface area and a much small Stokes number for a given velocity. For example, the surface area per collector volume of the 80  $\mu$ m glass beads is 10 times higher than that of the 0.8 mm glass beads. The Stokes number is 2 × 10<sup>-3</sup> when U<sub>g</sub> = 0.3 m/s for typical soot size 0.7  $\mu$ m. The filtration is in the diffusion regime at this low Stokes number. This is the reason why the filtration efficiency is very low when using 0.8 mm glass beads.

In order to determine the effect of particulate sampling on filtration efficiency, we

compared direct sampling with isokinetic sampling using the experimental setups shown in Figure 6.1b. Direct sampling is keep the sample gas flow rate as a constant no matter what is the gas flow rate in the main stream. Isokinetic sampling is the sample gas velocity is the same as that in the stream. Figure 6.17 compares the filtration efficiency as a function of soot size using isokinetic and direct sampling methods. The results of the two curves are very close. This is because the soot size is small. The soot concentration in the sampling tube is the same as in the stream.



Figure 6.17 Filtration efficiency as a function of soot size using isokinetic and direct sampling methods.

Figure 6.18 shows the mass average filtration efficiency as a function of superficial gas velocity using isokinetic and direct sampling methods. These two test data

are sufficiently similar that previous measurements taken before we adopted isokinetic sampling are not grossly in error. Since the sampling gas flow rate generated by Aerosizer is limited, this test is in the packed bed mode.



Figure 6.18 Mass average filtration efficiency as a function of superficial gas velocity using isokinetic and direct sampling method.

# 6.3.3 Perforated Distributor at High Engine Load and RPM Condition

At engine high load condition for our engine, the exhaust temperature is much higher than at idle, and soot concentration is much higher. Figure 6.19 presents the particle number distribution at the inlet and outlet of the RFBF loaded with Cu-ZSM-5 catalyst at engine high load conditions. The soot number at the outlet is much lower than that at inlet, especially for small soot particle diameter. At this operating condition, Cu-ZSM5 fines (probably produced by attrition at the high rpm of RFBF) blow out since there is an increase of larger particles at the outlet.



**Figure 6.19** Particle number distribution at inlet and outlet of RFBF loaded with Cu-ZSM-5 catalyst using the perforated distributor.

Figure 6.20 presents the particle number distribution at the inlet and outlet of the RFBF loaded with Q-705 alumina particles at engine high load conditions using the perforated distributor. Again, the soot number at the outlet is lower than that at the inlet except for large particles. This is caused by alumina fines being emitted at high gas flow rate.

When using the perforated distributor at engine high load condition, the pressure drop did not continue to increase after two hours of running the experiment. But when we checked the screen after the experiments, we found soot on the surface of distributor. Further more, the screen was found to be broken after approximately 3 hours due to the exposure to high exhaust temperature. The apparent reason for the breakage is that the screen became covered with soot which ignited producing a high local temperature, which caused metal deformation and fracture of the thin screen. On the other hand, when the perforated distributor was used at engine idle condition (relatively low temperature), the screen did not break.



**Figure 6.20** Particle number distribution at inlet and outlet of RFBF loaded with Q-705 alumina using the perforated distributor.

## 6.3.4 Sintered Metal Distributor at High Engine Load Condition

Figure 6.21 is the particle number distribution at the inlet and outlet of the RFBF loaded with 7  $\mu$ m alumina at engine high load condition using the sintered metal distributor (100

 $\mu$ m pore size). These small granules have a very high surface area per collector volume, 10 time higher than 80  $\mu$ m glass beads or more than 100 times higher than 0.8 mm glass beads. The filtration efficiency should be higher than when using larger particles since the Stokes number is inversely proportional to the granule size, i. g., the Stokes number is around 0.2 for 7  $\mu$ m alumina particles, 0.002 for 0.8 mm glass beads. Figure 6.21 shows that the filtration efficiency is very high for both particles smaller than 0.3  $\mu$ m soot or larger than 10  $\mu$ m soot. But we could not determine the filtration efficiency for on the order of 1  $\mu$ m soot particles because alumina fines that are approximately that size blow out. It should be noted that based on pressure drop measurements, the sintered metal distributor did not become blocked in three hours of running the engine at high load. However, inspecting the pores of the sintered metal distributor showed large quantities of soot present.



**Figure 6.21** Particle number distribution at inlet and outlet of RFBF loaded with 7  $\mu$ m alumina using the 100  $\mu$ m sintered metal distributor.

#### 6.4 Discussion

The experimental results indicate that RFBF can be used for removal of soot from diesel engine exhaust under certain conditions. The mass filtration efficiency can exceed 90%. The filtration efficiency is high for soot particles below 0.3  $\mu$ m or above 0.6  $\mu$ m because diffusion and inertial impaction become dominant mechanisms of capture for small and large particles, respectively. For a modern diesel engine, the soot size is normally below 0.1  $\mu$ m. However our engine due to agglomeration produces a peak of about 0.6 to 0.7  $\mu$ m for the entire range of engine operating conditions.

Although the RFBR has many advantages for treating diesel engine exhaust,

including high filtration efficiency and the ability to reduce  $NO_X$  with catalytic granules, there are still many problems that need to be solved. One problem is plugging of the distributor. We have found that both the sintered metal and perforated distributor will clog as a function of fine on stream. However, they do so in different ways. The sintered metal distributor due to its tortuous flow path plugs steadily. Whereas the perforated distributor with its component fine mesh screen plugs much more abruptly. These effects are more pronounced under engine idle conditions (low temperature). When moisture from combustion condenses producing wet soot, which tends to cake on the distributor. The moisture problem will not be severe for modern engines with high exhaust temperatures.

The second problem involves elutrition of catalysts fines, either initially present or produced by attrition, during fluidization. These fines are not only undesirable as exhaust emissions but cause difficulties in measuring the filtration efficiencies of the RFBR for soot. For example, when using alumina as the RFBR granules, we have measured negative filtration efficiencies even though the filtration of soot particles may be as high 90% on a mass basis. These negative efficiencies are due to the emission of around 1 µm alumina fines which are very close to the size of the soot. Using glass beads which do not contain or generate fines, we have mass average filtration efficiencies of between 50 and 85% when operating as a packed bed and as high as 90% when operating in the fluidized bed mode. However one should operate the RFBR close to minimum fluidization to avoid significant bypassing of soot in the bubble phase.

### **CHAPTER 7**

### CONCLUSIONS AND RECOMMENDATIONS

### 7.1 Conclusions

1. The minimum fluidization velocity and the pressure drop were measured as a function of rotating speed and bed charge for two different bed materials in a horizontal rotating fluidized bed using slotted, perforated and sintered metal cylindrical distributors. The fluidization characteristics were very different for the slotted and sintered metal distributors because of bed restructuring. This restructuring, verified by high resolution photography, explains both the magnitude of the reduction in pressure drop upon fluidization and its unusual hysteresis behavior. Using particles with a narrow size distribution (glass beads), the effects of restructuring were prominent only at high rotating speeds. This behavior can be attributed to the greater flowability of the glass beads. The fluidization characteristics using the sintered metal distributor, on the other hand, showed reasonable agreement with established theory. The perforated distributor exhibited intermediate behavior. Consequently, distributor design is very important in obtaining desired performance when using rotating fluidized beds.

2. For particles of the same material, the two layers of particles do not mix until bubbles appear. Below  $U_{mfi}$  the bed behaves as a solid. Between the calculated  $U_{mfi}$  and  $U_{mfc}$  the inner surface is observed to become more uniform, i.e., in this respect, the bed exhibits fluid-like behavior before mixing. After  $U_{mfc}$  particles inside the bed start to move radially and mixing occurs rapidly. Bubbles are a strong source of particle motion and the bed becomes totally fluid-like. To within observational error, the mixing, minimum critical fluidization velocity and minimum bubbling velocity coincide. For conventional fluidized beds, the minimum bubbling velocity,  $U_{mb}$ , is dependent on the size and density of the particles. For Geldart-B particles,  $U_{mb}$  is equal to the minimum fluidization velocity,  $U_{mf}$ , whereas for Geldart-A particles, which were used in our experiments,  $U_{mb}$  is larger than  $U_{mf}$  so that there is an appreciable operating region in which the bed is fluidized but not yet bubbling. Our RFB has negligible wall effects and hence presumably negligible particle circulation. However, no mixing was obtained before bubbles were observed, suggesting Group B behavior. For particles of different density, the mixing characteristics of the two layers depend on whether or not the denser particles are near the distributor. When the denser particles are near the distributor the mixing behavior is very similar to that of particles of the same density. When the less dense particles are near the distributor the mixing behavior is quite different. Mixing occurs gradually due to the differences in density and fluidization properties of the two layers and total mixing is observed before bubbles

3. The nature of fluidization and the behavior of particles in a centrifugal field was investigated. For conventional fluidized beds, Geldart classified powders into four groups: A, B, C and D based on their fluidization behavior. A simple model that can be used for variable "g" was developed to determine the transition boundary between group A and group C particles. The theoretical analysis shows that group A particles can shift to group B, and group C particles can shift to group A under a centrifugal force. Therefore, certain group C particles can be fluidized in rotating fluidized beds. However, for very high "g", such particles shift to group D. Experiments using 7  $\mu$ m alumina particles conform that Geldart group C particles will fluidize in a rotating fluidized bed operating at a sufficiently high rotating speed to shift them into group A or B. Similarly, group A

particles behave as group B confirming our mixing results which showed group B behaviors (bubbles were observed at minimum fluidization velocity) for group A particles.

4. A horizontal rotating fluidized bed filter (RFBF) charged either with polydispersed alumina granules, or nearly monodispersed glass beads, was used to capture soot from diesel engine exhaust at both idle and high load engine conditions. The filtration efficiency, calculated on the basis of the total mass of soot that was captured in the bed, was found to be in excess of 90% when using glass beads in the fluidized bed mode when operating with fresh particles (no soot build up in the bed). The filtration efficiency increased with increasing gas flow rate as the bed passes from the packed bed mode to the fluidized bed mode. The filtration efficiency also varied as a function of agglomerated soot size, showing a minimum for soot particles in the 0.3 to 0.6  $\mu$ m range. This is a consequence of particles larger than 0.6  $\mu$ m being removed mainly by inertial impaction and interception and smaller particles mainly by diffusion. However, the feasibility of using an RFB as a filter for diesel engine exhaust depends on the availability of catalytic granules that do not contain fines and resist attrition.

#### 7.2 Recommendations

1. The study in this dissertation shows that potential applications need to be explored using fine catalysts which have larger specific surface area and are more active in the catalytic reactions by using rotating fluidized beds.

2. Fines were found to blow out when using alumina particles. This might be caused by elutriation or particles attrition. This can be mitigated by enhancing attrition resistance of particles and installing a proper filter before the outlet of RFBF.

3. Both sintered metal distributor and screen were found clog by soot at engine idle condition. This problem has not been found using sintered metal distributor at engine high load condition. However, soot was stay inside the pores of the distributor and screen was broken because of the high temperature. Distributor design is very important in this application. Catalytic metal distributor is strongly recommended.

4. More theoretical analysis on filtration efficiency needs to be studied since the mechanism of soot filtration has not been well understood.

### REFERENCES

- Arcoumanis, C., L. N. Barbaris, R. I. Crane and P. Wisby, 1994, "Evaluation of a Cyclone-Based Particulate Filtration System for High-Speed Diesel Engines", *Proc. Instn. Mech. Engrs.*, *Part D: Journal of Automobile Engineering*, 208: 24-34.
- Baumgard, Kirby J. and John H. Johnson, 1996, "The Effect of Fuel and Engine Design on Diesel Exhaust Particle Size Distributions", *SAE Paper*, No. 960131, 37-50.
- Burdick, Ian W., J. G. Stevens, R. Pfeffer and H. Shaw, November 2, 1999, "Modeling Diesel Soot Filtration in a Rotating Bed Filter", Presented at AIChE 1999 Annual Meeting in Dallas.
- Burgler, L., P. L. Herzog and P. Zelenka, 1992, "Strategies to Meet US 1994/95 Diesel Engine Federal Emission Lefislation for HSDI Diesel Engine Powered Vehicles", *Proc. Instn. Mech. Engrs, Part D: Journal of Automobile Engineering*, 206: 47-54.
- Buyevich, Y. A., 1994, "Fluid Dynamics of Coarse Dispersion", Chem. Eng. Sci., 49: 1217
- Buyevich, Y. A. and S. K. Kapbasov, 1994, "Polydisperse Fluidization 1. Basic Equation", Chem. Eng. Sci., 49: 1245.
- Cartellieri, W. P. and Herzog, P. L., 1988, "Swirl supported or quiescent combustion for 1990s heavy duty DI diesel engines -- an analysis"" SAE paper 880342.
- Chaouki, J., C. Chavarie and D. Klvana, 1985, "Effect of Interparticle Forces on the Hydrodynamic Behavior of Fluidized Aerogels", *Powder Technology*, **43**: 117.
- Charles, J., 1990, "Filtration System Destroys Diesel Particulate Emissions", Design News, 46: 170.
- Chen, Y., 1987, "Fundamentals of a Centrifugal Fluidized Bed", AIChE J., 33 (5): 722.
- Chirone, R., L. Massimilla and S. Russo, 1992, "Bubbling Fluidization of a Cohesive Powder in an Acoustic Field", *Fluidization VII*, O. E. Potter and D. J. Nicklin, eds., Engineering Foundation, New York, 545.
- Clift, R., M. Ghadiri and K. V. Thambimuthu, 1981, "Filtration of Gases in Fluidized Beds", *Progress in Filtration and Separation 2*, edited by R. J. Wakeman, Elsevier Scientific Publishing Company.

- Cody, G. D., and D. J. Goldfarb, 1997, "Discontinuity in particle granular temperature of across the Geldart B/A boundary", *AIChE Symp. Ser.: Progress in Fluidization and Fluid Particle Systems*, D. King, ed., **93**: 18.
- Cody, G. D., D. J. Goldfarb, G. V. Storch Jr. and A. N. Norris, 1996, "Particle Granular Temperature in Gas Fluidized Beds", *Powder Technology*, **87**: 211.
- Cody, G. D., S. K. Kapbasov and Y. A. Buyevich, 1998, "Particle Fluctuation Velocity in Gas Fluidized Beds – Fundamental Models Compared to Recent Experimental Data", Symposium B, Session 20: Advanced Technologies in Fluid-Particle System, AIChE 1998 Annual Meeting, Miami Beach.
- Davidson J. F., R. Clift and D. Harrison, 1985, Fluidization, Academic Press, London.
- Demircan, N., B. H. Gibbs, J. Swithenbank and D. S. Taylor, 1978, "Rotating Fluidized Bed Combustor", *Fluidization*, J. F. Davidson and D. L. Kearins, eds., Cambridge University Press, Cambridge, 270.
- D'Ottavio, T. and L. S. Goren, 1983, "Aerosol Capture in Granular Beds in the Impaction Dominated Regime", *Aerosol Science Technology*, **2**: 91.
- Dullien, F. A. L., 1989, Introduction to Industrial Gas Cleaning, Academic Press.
- Fan, L.-S. and C. Zhu, 1998, Principles of Gas-Solid Flows, Cambridge University Press.
- Fan, L. T., C. C. Chang, Y. S. Yu, T. Takahashi and Z. Tanaka, 1985, "Incipient Fluidization Condition for a Centrifugal Fluidized Bed", *AIChE J.*, **31** (6): 999.
- Farrauto, Robert J., Ronald M. Heck and Barry K. Speronello, 1992, "Environmental Catalysts", *Chemical and Engineering News*, **70** (36): 34 44.
- Flagan, Richard C. and John H. Seinfeld, 1988, Fundamentals of Air Pollution Engineering, Prentice Hall.
- Foscolo, P. U. and L. G. Gibilaro, 1984, "A Fully Predictive Criterion for the Transition Between Particulate and Aggregate Fluidization", *Chemical Engineering Science*, 39 (12): 1667.
- Geldart, D., 1973, "Types of Gas Fluidization", Powder Technology, 7: 285.
- Geldart, D. And A. R. Abrahamsen, 1980, "The Effect of Fines on the Behavior of Gas Fluidized Beds of Small Particles", *Fluidization*, J. R. Grace and J. M. Matsen,

eds., Plenum Press, New York, 453.

Geldart, D., 1986, Gas Fluidization Technology, John Wiley and Sons, Inc., New York.

- Godard, K. And J. F. Richardson, 1969, "Bubble Velocities and Bed Expansions in Freely Bubbling Fluidized Beds", *Chem. Eng. Sci.*, 24: 663.
- Heck, R. M. and R. J. Farrauto, 1995, *Catalytic Air Pollution Control: Commercial Technology*, Van Nostrand Reinhold, International Thomson Publishing Inc.
- Iwadate, Yuki and Masayuki Horio, 1998, "Prediction of Agglomerate Sizes in Bubbling Fluidized Beds of Group C Powders", *Powder Technology*, **100**: 223.
- Jackson, R., 1963, "The Mechanics of Fluidized Beds", Trans. Inst. Chem. Engrs, 41: 13.
- Jackson, R., 1998, "The Nature and Role of Effective Stress in Fluidized Systems", *Fluidization IX*, ed. L. S. Fan, and T. M. Knowlton, Engineering Foundation, New York.
- Johnson, J. H., Susan T. B., Linda D. G. and David G. L., 1994, "A Review of Diesel Particulate Control Technology and Emissions Effects – 1992 Horning Memorial Award Lecture", SAE 940233.
- Kao, J., R. Pfeffer and G. I. Tardos, 1987, "On Partial Fluidization in Rotating Fluidized Beds", *AIChE J.*, **33:** 858.
- Kao, J., G. I. Tardos and R. Pfeffer, 1988, "Development of a Rotating Fluidized Bed Filter", *Fluid/Particle Separation Journal*, **1** (1): 44-52.
- Kerminen, V. M., T. E. Makela, C. H. Ojanen, R. E. Hillamo, J. K. Vilhunen, L. Rantanen, N. Havers, A. Von Bohlen and D. Klockow, 1997, "Characterization of the Particulate Phase in the Exhaust from a Diesel Car", *Environmental Science & Technology*, **31**: 1883 1889.
- Khan, A. R. and J. F. Richardson, 1990, "Pressure Gradient and Friction Factor for Sedimentation and Fluidization of Uniform Spheres in Liquids", *Chemical Engineering Science*, **45** (1): 255.
- Konstandopoulos, A. G. and J. H. Johnson, 1989, "Wall-Flow Diesel Particulate Filters-Their Pressure Drop and Collection Efficiency", SAE Paper No. 890405.
- Kroger, D. G., G. Abdelnour, E. K. Levy, and J. C. Chen, 1980, "Particle Distribution and Mixing in a Centrifugal Fluidized Bed", *Fluidization*, J. R. Grace and J. M. Matsen, eds., Plenum Press, New York, 349.
- Krupp, H., 1967, "Particle Adhesion, Theory and Experiment", Adv. Colloid Interface

*Sci.*, **1**: 111.

- Lemaire, J. and M. Khair, 1994, "Effect of Cerium Fuel Additive on the Emissions Characteristics of a Heavy-Duty Diesel Engine", SAE Paper No. 942067, 189-205.
- Levy, E. K., N. Martin and J. Chen, 1978, "Minimum Fluidization and Startup of a Centrifugal Fluidized Bed", *Fluidization*, D. L. Kearins, ed., Cambridge University Press, Cambridge, 71.
- Levy E. K., W. J. Shakespeare, A. Tabatabaie-Raissi and J. C. Chen, 1981, "Particle Elutriation from Centrifugal Fluidized Beds", *AIChE Symp, Ser., Recent Advances in Fluidization*, 77 (205): 86.
- Ma, K., X. Tang, I. Bagyi, H. Shaw, R. Pfeffer and J. G. Stevens, 1998, "Catalytic Reduction of Nitric Oxide with Solid Reducing Materials", Presented at the 2<sup>nd</sup> World Congress on Environmental Catalysis, Miami, FL.
- Massimilla, Leopoldo and Giorgio Donsi, 1976, "Cohesive Forces Between Particles of Fluid-Bed Catalysts", *Powder Technology*, **15**: 253.
- Mathur, K. B., 1971, "Spouted Beds", *Fluidization*, Chapter 17, J. F. Davidson and D. H. Harrison, eds., Academic Press, London and New York.
- Menon, N., and D. J. Durian, 1997, "Particle Motions in a Gas-Fluidized Bed of Sand", *Physical Review Letters*, **79** (18): 3407.
- Molerus, O., 1982, "Interpretation of Geldart's Type A, B, C and D Powders by Taking Into Account Interparticle Cohesion Forces", *Powder Technology*, **33**: 81.
- Mori, S., N. Iwasaki, E. Mizutani and T. Okada, 1992, "Vibro-fluidization of Twocomponent Mixtures of Group C Particles", *Fluidization VII*, O. E. Potter and D. J. Nicklin, eds., Engineering Foundation, New York, 571.
- Mutsers, S. M. P., and K. Rietema, 1977a, "The Effect of Interparticle Forces on the Expansion of a Homogeneous Gas-Fluidized Bed", *Powder Technology*, **18**: 239.
- Mutsers, S. M. P., and K. Rietema, 1977b, "Gas-Solids Fluidization in a Centrifugal Field. The Effect of Gravity upon Bed Expansion", *Powder Technology*, **18**: 249.
- Needham, J. R., S. A. Faulkner and D. M. Doyle, 1991, "Developing The Truck Engine for Ultra Low Emissions", *Proceedings of the Institution of Mechanical Engineers, Part D: Journal of Automobile Engineering*, **205**: 149-159.

Oltrogge, R. D., 1972, Ph. D. Thesis, University of Michigan.

- Panne, U., A. Petzold and R. Niessnerr, 1995, "Diesel Exhaust Particle Size Distribution Measurements Under Dynamic Conditions", *J. Aerosol Science*, **26**: S659.
- Paretsky, L, L. Theodore, R. Pfeffer, and A. M. Squires, 1971, "Panel Bed Filters for Simultaneous Removal of Fly Ash and Sulfur Dioxide: II. Filtration of Dilute Aerosols by Sand Beds", *Journal of the Air Pollution Control Association*, 21 (4): 204-209.
- Pfeffer, R. and F. B. Hill, 1978, "A Feasibility Study on the Use of a Rotating Fluidized Bed as a High Efficiency Dust Filter", BNL 50990.
- Pfeffer, R., G. Tardos and E. Gal, 1986, "The Use of a Rotating Fluidized Bed as a High Efficiency Dust Filter," *Proceedings of the Fifth Engineering Foundation Conference on Fluidization*, Denmark, 667.
- Qian, G.H., I. B□yi, R. Pfeffer, H. Shaw and J. G. Stevens, 1998, □A Parametric Study of a Horizontal Rotating Fluidized Bed Using Slotted and Sintered Metal Cylindrical Gas Distributors", *Powder Technology*, **100**: 190.
- Qian, G.H., I. B□yi, R. Pfeffer, H. Shaw and J. G. Stevens, 1999, □Particle Mixing in Rotating Fluidized Beds: Inferences about the Fluidized State", AIChE Journal, 45 (7): 1401.
- Richardson, J. F. and W. N. Zaki, 1954, "Sedimentation and Fluidization: Part I", *Trans. Instn Chem, Engrs*, **32**: 35.
- Rickeard, David J., John R. Bateman, Yeong K. Kwon, John J. McAughey and Colin J. Dickens, 1996, "Exhaust Particulate Size Distribution: Vehicle and Fuel Influences in Light Duty Vehicles", SAE Paper No. 961980.
- Rietema, K., 1973, "The Effect of Interparticle Forces on the Expansion of a Homogeneous Gas-Fluidized Bed", Chem. Eng. Sci., 28: 1493.
- Rietema, K., 1984, "Powders, What Are They?", Powder Technology, 37: 5.
- Rietema, K., E. J. E. Cottaar and H. W. Piepers, 1993, "The Effect of Interparticle Forces on the Stability of Gas-Fluidized Beds -- II. Theoretical Derivation of Bed Elasticity on the Basis of van der Waals forces Between Powder Particles", *Chemical Engineering Science*, 48(9): 1687.
- Russell-Jones, R., 1987, "The health effects of vehicle emissions", ImechE Conference on Vehicle emissions and their impact on European air quality, paper C355/87, 55-59.

Schlichting, H., 1979, Boundary Layer Theory, 7th ed., McGraw-Hill, New York.

- Springer, K. L., 1988, "Diesel lube oils -- fourth dimension of diesel particulate control", *Diesel Progress: Engines and Drives*, 17 – 19.
- Strauss, W., 1975, Industrial Gas Cleaning, 2<sup>nd</sup> ed., Pergamon Oxford.
- Suarez, A. E., P. F. Caffrey, P. V. Borgoul, J. M. Ondov and F. Divita, JR., 1998, "Use of an Iridium Tracer To Determine the Size Distribution of Aerosol Emitted from a Fleet of Diesel Sanitation Trucks", *Environmental Science & Technology*, 32 (10): 1522-1529.
- Sun, S., 1996, "The Reactions of NO<sub>X</sub> with Diesel Soot over Cu-ZSM5 Catalyst," Master's Thesis, New Jersey Institute of Technology.
- Tanaka, S., M. Morinaga, H. Yoshida, H. Takizawa, K. Sanse and H. Ikebe, 1996, "Effects of Fuel Properties on Exhaust Emissions from DI Diesel Engines", SAE Paper No. 962114, 181-192.
- Tang, X. Y., 1997, "Catalytic Control of Nitric Oxide with Gaseous or Solid Reducting Materials," Master's Thesis, New Jersey Institute of Technology.
- Tsinontides, S. C. and R. Jackson, 1993, "The Mechanics of Gas Fluidized Beds with an Interval of Stable Fluidization", *Journal of Fluid Mechanics*, **255**: 237.
- Tsutsumi, A., Mitsuru Demura and Kunio Yoshida, 1994, "Reduction of Nitrogen Oxides with Soot Emitted from Diesel Engines by a Centrifugal Fluidized Bed," *AIChE Symp. Ser.*, **90** (301): 152-156.
- Tung, Y. and M. Kwauk, 1989, "Assessing Fluidizing Characteristics of Powders", Proceedings of the 6<sup>th</sup> International Conference on Fluidization, J. R. Grace, ed., Banff, Canada, 169.
- Urtagh, M., D. Sherwood and L. S. Socha, 1994, "Development of a New Diesel Particulate Filter Composition and its Effect on Durability and Performance", SAE 940235.
- Verloop, J. and P. M. Heertjes, 1970, "Shock Waves as a Criterion for the Transition from Homogeneous to Heterogeneous Fluidization", *Chemical Engineering Science*, 25: 825.
- Vuk C. T., Jones M. A. and J. H. Johnson, 1976, "The Measurement and Analysis of the Physical Character of Diesel Particulate Emissions", SAE Trans. Section 1, SAE Paper No. 760131, 556-597.
- Wallis, G. B., 1969, One-Dimensional Two-phase Flow, McGraw-Hill, New York.
- Wang, Zhaolin, Mooson Kwauk and Hongzhong Li, 1998, "Fluidization of Fine

Particles", Chemical Engineering Science, 53 (3): 377.

- Wen, C. Y. and Y. H. Yu, 1966, "A Generalized Method for Predicting the Minimum Fluidization Velocity", Chem. Eng. Symp. Ser., 67: 100.
- Wenninger, G., Harald Polz and NJIT, 1998, Communication Fax.
- Wilhelm, R. H. and Mooson Kwauk, 1948, "Fluidization of Solid Particles", *Chemical Engineering Progress*, 44 (3): 201.
- Zhou, Tao and Hongzhong Li, 1999, "Estimation of Agglomerate Size for Cohesive Particles During Fluidization", *Powder Technology*, **101**: 57.
- Zhu, C., C.-H. Lin and C.-S. Cheung, 2000, "Inertial Impaction dominated Fibrous Filtration with Rectangular or Cylindrical Fibers", *Powder Technology*, in press.
- Zhu, C., 1998, Personal Communication, New Jersey Institute of Technology.