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EXPERIMENTAL AND COMPUTATIONAL STUDIES OF HYDRODYNAMICS IN THREE-PHASE AND TWO-PHASE FLUIDIZED BEDS

BY

MITRA BAHARY

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LIST OF SYMBOLS

Abbreviation	Term
A	coefficient (equal to $\kappa \rho l$)
A_{∞}	initial value of upward acceleration of a particle
C	fluctuating velocity of particle
C_D	drag coefficient
c_y	mean particles velocity
D_o	diameter of an orifice
d_b	bubble diameter
d_k	characteristic diameter of particulate phase k
e	coefficient of restitution
Fr	Froude number
$f()$	frequency
$G()$	solid compressive stress modulus
g	gravity
g_o	radial distribution function
h	bubble height
I	intensity of transmitted radiation
I_o	intensity of incident radiation
$[I]$	unit tensor
L	shear layer thickness
l	path length
n	rotation rate
p	pressure
q	flux vector of fluctuating energy
R	spindle radius
Re	Reynolds number
Re_k	Reynolds number of phase k

Abbreviation	Term
t	time
U_r	relative velocity between the colliding bubble and particle
u	velocity in x -direction
v	velocity in y -direction
$\mathbf{v}$	velocity vector
v_Y	mean particles velocity fluctuation
X_k	reduced volume fraction $\varepsilon_k / (\varepsilon_k + \varepsilon_m)$
x	x-direction in cartesian coordinate
y	y-direction in cartesian coordinate

Greek letters

α	coefficient in particle-particle drag coefficient equation
β_{km}	interphase momentum transfer between phase k and phase m
γ	collision energy dissipation; shear rate
ε	volume fraction
ζ	vorticity
$3/2\Theta_s$	fluctuating energy
κ	conductivity of fluctuating energy; attenuation coefficient
μ	viscosity
μ_s	(shear viscosity x volume fraction) of solids
ξ_s	(bulk viscosity x volume fraction) of solids
ξ	bulk viscosity
ρ	density; density of material
σ	surface tension of liquid; standard deviation of velocity fluctuations
τ	shear stress
τ_0	yield stress
Φ_s	viscous energy dissipation
ψ_s	particle sphericity

Abbreviation	Term
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ω	angular velocity;
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Subscripts

a	apparent
bed	effective bed
dil	dilute phase
g	gas phase
k	phase k
l	liquid phase
m	phase m
max	maximum
mf	minimum fluidization
p	particle
s	solid phase
X	x-ray
x	x-direction in cartesian coordinate
y	y-direction in cartesian coordinate
z	z-direction in cartesian coordinate
γ	γ -ray

Superscripts

T	transpose
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ABSTRACT

The objective of the present study was to investigate the hydrodynamics of three-phase fluidized beds, their rheology, and experimentally verify a predictive three fluid hydrodynamic model developed at the Illinois Institute of Technology, Chicago. The recent thorough reviews by L.S. Fan, B.L. Tarmy and C.A. Coulaloglou (1990) show that there exist no such models in the literature.

The IIT hydrodynamic model computes the phase velocities and the volume fractions of gas, liquid, and particulate phases. Model verification involves a comparison of these computed velocities and volume fractions to experimental values.

In this thesis, a three fluid model is presented. The input into the model can be particulate viscosities either measured with a Brookfield viscometer or derived using the mathematical techniques of kinetic theory of granular flows pioneered by Savage and others. The computer simulation of a three-phase fluidized bed in an asymmetric mode qualitatively predicts the gas, liquid and solid hold-ups (volume fractions) and flow patterns in the industrially important churn-turbulent (bubbly coalesced) regimes. The computations in a fluidized bed with a symmetric distributor incorrectly showed no bubble coalescence.

A combination of X-ray and γ -ray densitometers was used to measure the solids and the liquid volume fractions in a two dimensional bed in the bubble coalesced regime. There is a good agreement between the theory for an asymmetric distributor and the experiments.

The porosity fluctuation signals inside two phase and three phase fluidized beds were detected by using a γ -ray absorption technique and analyzed by means of the fast Fourier transform (FFT) power spectrum method. The measurements indicated that the bubble coalescence and break-up could be detected by using such techniques. The experimental and simulation porosity fluctuations and bubble frequencies show a fair comparison.

A high resolution micro-imaging / measuring system apparatus was used to measure instantaneous and time averaged particle velocities. The fluctuations of particle velocities around their average values were found to be approximately Maxwellian. The measured time averaged velocities compare well with the predicted velocities for an asymmetric distributor used in this study.

A Brookfield viscometer was used to measure the "apparent" bed viscosity in two three-phase fluidized beds. The viscosities were also calculated from measurements of particle oscillations (granular temperatures) using a high resolution micro-imaging /measuring system. To obtain these viscosities a dense phase kinetic theory formula derived in D. Gidaspow's book (1994) was used. There is an excellent agreement between the macroscopically measured viscosities using a Brookfield viscometer and those computed from particle fluctuations and collisions using kinetic theory.

CHAPTER I

INTRODUCTION

The three-phase (gas-liquid-solid) fluidized bed is a device in which the gas phase moves in the form of bubbles relative to the liquid phase, and eventually reactive solid (catalyst) is fluidized in the liquid phase. The commercial applications of three-phase fluidization systems are in heavy oil, resid hydroprocessing, synthetic crude processing, coal liquefaction in the presence of catalyst, biological waste water treatment and fermentation (Fan, 1989).

The hydrodynamic behavior of gas-liquid-solid fluidized bed is a complex subject and one of the most important for basic understanding of certain refinery and petrochemical industrial applications. Overall three-phase fluidized bed behavior, bubble motion and breakage have been experimentally studied by a number of researchers (Fan, 1989; Kim et al., 1975; Henrickson and Ostergaard, 1974). Also, considerable fundamental work has been successfully demonstrated for petroleum resid processing at Kuwait Shuaiba, Texaco Convent, and Amoco Texas City Refineries. However, recent reviews by Fan (1989), Tarmy and Coulaloglou (1992) show that there exist no hydrodynamic models for three-phase fluidized beds in the literature.

The Illinois Institute of Technology (IIT) three-phase fluidized bed is a two-dimensional rectangular vertical column. The solids (lead glass beads) are present inside the bed, the gas (air) enters at the bottom through a distributor and liquid (water) is also fed to the bed through another distributor which is located at the bottom of the

bed. In this study, the overall hydrodynamics of the three-phase fluidized bed are investigated. The details of experiments and multiphase computer simulations are described in the following chapters.

In Chapter II, the three-phase fluidized bed experimental setup, x-ray and γ -ray radiation densitometer assembly, and procedures for concentrations (volume fraction) measurements using radiation (x-ray and γ -ray) absorption techniques are described. The visual observations of bubble coalescence and break-up in three phase fluidized bed and their comparison with observations in literature are also narrated.

Chapter III describes the principle of interpretation, calibration for concentration measurements using x-ray and γ -ray absorption techniques, experiments relating to time averaged concentrations, and instantaneous porosity oscillations measurements for gas, liquid and solids phases. A discussion is also included on the bubble sizes and coalesced bubble flow regime used in the experiments of the three phase fluidized bed.

In Chapter IV, the porosity fluctuation signals detected by using a γ -ray absorption technique are reported. The frequencies of porosity oscillations at various locations inside the bed have been analyzed by means of the fast Fourier transform power spectrum method for both two phase and three phase fluidized beds. The experimental porosity fluctuations and bubble frequencies are compared with those obtained using computer simulations of the beds.

Chapter V describes the high resolution micro-imaging / measuring system apparatus for particles velocity measurements; experimental setup for fluidized beds with uniform distributor, and central jet; and experimental procedure. The instantaneous and

time averaged particle velocity measurements using CCD video camera are also recounted.

In Chapter VI, the experimental setup and the measurement of bed viscosity of the three-phase fluidized bed using a Brookfield viscometer are reported. The computed particle fluctuating velocities based on measured bed viscosities are compared with those measured directly using a CCD video camera.

Chapter VII describes the predictive hydrodynamic model including the numerical scheme, initial and boundary conditions and computer simulations of three-phase fluidized bed in symmetric and asymmetric modes. The comparison between experiments and simulation results is also given.

The future work in fluidized beds related to volume fractions, solids velocity and bed viscosity measurements, and hydrodynamic simulations are summarized in Chapter VIII.

Appendix A includes the raw experimental data for measured bed intensities using x-ray and γ -ray densitometers, and corresponding gas, liquid and solids instantaneous volume fractions at different locations inside a three phase fluidized bed.

Appendix B consists of the raw experimental data for instantaneous particles velocity measurements using a CCD video camera at many locations, under different operating conditions inside three phase fluidized beds with a uniform distributor and central jet. The statistical analysis, such as mean, median and variance values of the raw data, is also included.

Appendix C contains raw experimental data for bed viscosity using a Brookfield viscometer at different operating conditions inside two phase and three phase fluidized beds.

Appendix D describes the predictive multiphase hydrodynamic computer model used in the computer simulations, including continuity and momentum equations. The fluctuating energy equations for kinetic theory of granular solids are also included.

CHAPTER II

DESCRIPTION OF EXPERIMENTAL APPARATUS

FOR THREE-PHASE FLUIDIZED BED

2.1 Introduction

Over the past ten years, extensive experimental research on two-phase fluidization systems, which consisted of only gas and solid phases, has been carried out at the Illinois Institute of Technology (IIT) by a number of researchers (Miller and Gidaspow, 1992; Gamwo, 1992; Seo and Gidaspow, 1987; Gidaspow and Ettehadieh, 1983 and others). These experiments led to a better understanding of the hydrodynamics of both the bubbling fluidized beds (Seo and Gidaspow, 1987; Gidaspow and Ettehadieh, 1983; Gamwo, 1992), and the circulating fluidized beds (Miller and Gidaspow, 1992). The experimental research was vital to the development and extension of predictive hydrodynamic models to multiphase flows and fluidization at IIT.

However, for the first time, the experimental work on gas-liquid-solid fluidization, gas-liquid fluidization and liquid-solid fluidization is undertaken at IIT. The experiments involve the measurements of the bubble sizes, time average volume fractions, instantaneous porosity fluctuations, particle velocity measurements, and bed viscosity measurements. The results of these experiments were also used to validate existing multiphase hydrodynamic models for gas-liquid-solid, liquid-solid and gas-liquid fluidization systems (Jayaswal, 1991). These hydrodynamic models could be used as cost-effective design and troubleshooting tools by industries.

2.2 Experimental Apparatus and Procedure

The apparatus used in this experiment consisted of two major components : the fluidization equipment, and the densitometer assembly. A schematic diagram of the apparatus is shown in Figure 2.1.

2.2.1 Fluidization Equipment. A two-dimensional bed was constructed from transparent acrylic (plexiglas) sheets to facilitate visual observation and video recording of the bed operations such as gas bubbling and coalescence, and the mixing and segregation of solids. The bed height was 213.36 cm and cross-section was 30.48 cm by 5.08 cm. A centrifugal pump was connected to the bottom of the bed by a 1.0 inch (2.54 cm) I.D. stainless steel pipe. A compressor was connected to the sides of the bed. Liquid was stored in and recycled back to a fifty-five gallon storage tank.

The liquid and gas distributors were located at the bottom of the bed. The liquid was distributed by two perforated plexiglas plates with many 0.277 cm diameter holes. They were placed at 35.56 cm and 50.8 cm above the bottom of the bed, with 0.25 cm size glass bead particles filling the inside. The gas distributor consisted of six staggered 15.24 cm long porous tubes of 0.28 cm diameter. The fine pores of porous tubes had a mean diameter of 41.88 μm . The porous tubes were placed at the bottom of the bed between the liquid distributor.

2.2.2 Densitometer Assembly. Two densitometers were used alternatively for measuring both the time-averaged and the instantaneous volume fractions of three phases by means of the x-ray and γ -ray adsorption techniques. The assembly consisted of radioactive sources as well as detecting and recording devices and a positioning table.

Figure 2.1. Schematic Diagram of the Experimental Apparatus

A schematic diagram of the detector and recording devices assembly is shown in Figure 2.2.

2.2.2.1. Radioactive Source. The source is a 200-mCi Cu-244 having a 17.8-year half-life. It emitted X-rays with a photon energy between 12 and 23 keV. The source was contained in a ceramic enamel, recessed into a stainless steel support with a tungsten alloy packing, and sealed in a welded Monel Capsule. The device had a brazed Beryllium window. For the γ -ray densitometer, a 20-mCi Cs-137 source having a single γ -ray energy of 667 keV and a half-life of 30 years was used. The source was sealed in a welded, stainless steel capsule. The source holder was welded, filled with lead, and provided with a shutter to turn off the source. This is the same unit we used previously (Gidaspow et al., 1983).

2.2.2.2. Detecting and Recording Devices. The intensity of the X-ray beam was measured by using a NaI crystal scintillation detector (Teledyne, ST-82-I/B). It consisted of a 2-mm thick, 5.08 cm diameter tube with a 0.13-mm thick Beryllium window. For the γ -ray densitometer, the intensity of the γ -ray beam was detected by another NaI crystal detector (Teledyne, S-44-I/2). The dimensions of the crystal were as follows : 5.08 cm thick and 5.08 cm in diameter. The two detectors could be switched for use with different sources. The photomultiplier of the detector was connected sequentially to a preamplifier, an amplifier and a single-channel analyzer, a rate meter, and a 186 IBM compatible personal computer. The rate meter has a selector and a 0-100-mV scale range.

2.2.2.3. Positioning Table. Both the source holder and detector were

Figure 2.2. Schematic Diagram of Source-Detector-Recorder Assembly for X-ray and γ -ray Densitometer

affixed to either side of the bed on a movable frame and could be moved anywhere up-or-down or back-and-froth by means of an electric motor.

2.2.3. Experimental Procedure. The liquid from the storage tank was fed to the bed from the bottom of the bed using the centrifugal pump. The gas was fed to the bed through a compressor. Both gas and liquid from the top of the bed were directed through three openings of 1.0 inch (2.54 cm) I.D. stainless steel pipes back to the storage tank, where the gas was separated from the liquid.

In order to achieve a uniform fluidization, the liquid distributor section was designed in such a way that the pressure drop through the distributor section was 10-20 % of the total bed pressure drop. The gas was distributed in the fluidized bed through the six staggered porous tubes.

Air and water were used as the gas and liquid, respectively, in this experiment. Ballotini (leaded glass beads) with an average diameter of 800 μm and a density of 2.944 g/cm^3 were used as the solids.

The experimental volume fractions measurements were made at 240 different locations at every 1.27 cm in both -x and -y directions on the left portion of the bed.

The experimental operating conditions are shown in the Table 2.1.

Table 2.1. Experimental Operating Conditions

Temperature	23.5 °C
Pressure Drop	0.083 atm
Initial Bed Height	25.4 cm
Particle Mean Diameter	800 μm (0.8 mm)
Particle Density	2.94 g/cm^3
Minimum Fluidization Velocity	0.76 cm/s
Maximum Packing Porosity	0.39

2.3 Visual Observations of Bubble Coalescence and Break up

The hydrodynamic behavior of a multiphase fluidization system is due to complex interactions between the individual phases, and is also affected by the gas and liquid distributor designs. The phenomenon of bubble behavior at any axial location within a three-phase fluidized bed starts in the distributor region where bubbles are formed. The initial bubble sizes depend on the distributor design as well as buoyancy, viscous drag and surface tension. By using the surface tension equation, Azbal (1981) described the bubble diameter (d_b) at the instant of release as :

$$d_b = \left[\frac{6\sigma D_o}{(\rho_f - \rho_g)g} \right]^{1/3} \quad (2.1)$$

where ρ_g and ρ_l are the densities of the gas and liquid phases, respectively; D_o is diameter of the orifice (42 μm); σ is the surface tension and g is the gravitational constant.

The average bubble diameter at the moment of release from the orifice was calculated to be 0.124 mm. As soon as the bubbles left the orifice the mean bubble sizes were increased with an increase in the axial distance. These changes were also found by Massimilla et al. (1961) and Ostergaard (1969). Fan (1989) described this behavior of the bubble as "bubble coalescence".

Another behavior of the bubble that Fan (1989) described was bubble break-up. This behavior was also observed in this experiment. He defined this mechanism as an interaction between a bubble and individual particles. According to and Fan (1989):

$$1 > \frac{h\omega^2}{A_{co} - (A_{co}^2 - U_r^2 \omega^2)^{\frac{1}{2}}} \quad (2.2)$$

where,

$$A_{co} = g \frac{\rho_l - \rho_s}{\rho_s} + \frac{6\sigma}{\rho_s d_p^2} \left(1 + \frac{d_p}{2r_o} \right) \quad (2.3)$$

and

$$\omega = \left(\frac{3}{2} \frac{g}{d_p} \frac{\rho_l}{\rho_s} \right)^{1/2} \quad (2.4)$$

A particle will penetrate through the bubble if equation (2.2) satisfied. However, this criterion constitutes only a necessary condition for bubble breakup with a single particle. The calculations for our experiment showed that the right hand side of equation

Figure 2.3. Mechanism of Coalescence and Disintegration of Bubbles in Gas-Liquid-Solid Fluidized Beds (Fan, 1989)

(2.2) was 0.28, which is less than 1.0. Therefore particles could penetrate through the bubbles in our fluidized bed, and break up the bubbles as well. Figure 2.3 pictorially shows the mechanism of coalescence and disintegration of bubbles in gas-liquid-solid fluidized beds (Fan, 1989).

CHAPTER III

MEASUREMENTS OF VOLUME FRACTIONS IN TWO-PHASE AND THREE-PHASE FLUIDIZED BEDS

3.1 Introduction

X-ray and γ -ray densitometers have been used to measure porosities of fluidized beds (Miller and Gidaspow, 1993; Gidaspow, et al., 1983) and solids concentrations in nonaqueous suspensions (Jayaswal, et al., 1990). These techniques are based on the fact that the gas, liquid and solid phases under consideration have different absorptivities for x-rays and γ -rays. The same concept was adopted to measure concentration profiles inside our three-phase (air-water-lead glass bead) fluidization systems.

3.2 Principle of Interpretation

The absorption of x-ray or γ -ray by any material depends mainly on the molecular weight of the material. The amount of radiation that is absorbed by a material can be explained by the Beer-Bourget-Lambert Law :

$$I = I_0 \exp(-\kappa \rho l) \quad (3.1)$$

where I is the Intensity of transmitted radiation, I_0 the intensity of incident radiation, κ is the attenuation coefficient, ρ the density of material, and l is the path length. The attenuation coefficient is calculated from a calibrated curve. The logarithmic form of equation (3.1) is :

$$\ln\left(\frac{I_o}{I}\right) = A \quad (3.2)$$

where $A = \kappa pl$.

For a mixture of n phases, the intensity of the transmitted X-rays or γ -rays could be described as a linear function of the volume fractions of the phases as :

$$\ln\left(\frac{I}{I_o}\right) = A_1\epsilon_1 + A_2\epsilon_2 + \dots + A_n\epsilon_n \quad (3.3)$$

where ϵ_n is the volume fraction of phase n . The transmitted energy of radiation from both densitometers was placed into logarithmic form.

Another relation for ϵ_n is :

$$\epsilon_1 + \epsilon_2 + \dots + \epsilon_n = 1 \quad (3.4)$$

For three-phase fluidized beds, there are three unknowns, volume fraction of gas (ϵ_g), volume fraction of liquid (ϵ_l), and volume fraction of solid (ϵ_s). In order to estimate the unknowns, the equations (3.3) and (3.4) are simplified to form the following three equations :

$$\ln\left(\frac{I_x}{I_{o,x}}\right) = A_{g,x} \epsilon_g + A_{l,x} \epsilon_l + A_{s,x} \epsilon_s \quad (3.5)$$

$$\ln\left(\frac{I_\gamma}{I_{o,\gamma}}\right) = A_{g,\gamma} \epsilon_g + A_{l,\gamma} \epsilon_l + A_{s,\gamma} \epsilon_s \quad (3.6)$$

$$\epsilon_g + \epsilon_l + \epsilon_s = 1 \quad (3.7)$$

where I_x and I_γ are the intensity readings of X-ray and γ -ray densitometers.

3.3 Calibration of X-ray and γ -ray Densitometer

Before any measurements of the gas, liquid and solids concentration during the fluidization could be made, the x-ray and γ -ray densitometers had to be calibrated for the system. The coefficients in equations (3.5) and (3.6) were statistically calculated using the least square error technique from the calibration measurements of the intensity readings of X-ray and γ -ray densitometers at known concentrations of gas, liquid and solids in three-phase mixtures. However, these coefficients were found to have values with 20% of error for X-ray and 2% of error for γ -ray. The complete calibration curves for both X-ray and γ -ray in a three-phase fluidized bed were obtained using the following calibrated mixtures :

a. Gas :

$$\epsilon_g = 1.0, \epsilon_l = 0.0, \epsilon_s = 0.0$$

b. Gas-Solid packed :

$$\epsilon_g = (1.0 - \epsilon_{s, \max}), \epsilon_l = 0.0, \epsilon_s = \epsilon_{s, \max}$$

c. Liquid-Solid :

$$\epsilon_g = 0.0, \epsilon_l, \epsilon_s = \text{a range of average volume fractions values based on visual observations of initial and fluidized bed heights, including,}$$

c-1. Liquid :

$$\epsilon_g = 0.0, \epsilon_l = 1.0, \epsilon_s = 0.0$$

c-2. Liquid-solid packed:

$$\epsilon_g = 0.0, \epsilon_l = (1.0 - \epsilon_{s, \max}), \epsilon_s = \epsilon_{s, \max}$$

Typical linear calibration curves for the liquid-solid (water-lead glass beads) system for x-ray and γ -ray densitometers are shown in the Figures 3.1 and 3.2, respectively. Each point in the plot represents an average of 15-20 readings of the intensity of the x-ray or γ -ray transmitted through the bed.

To obtain the time averaged volume fractions at a designated position inside the three-phase fluidized bed, an integration time of 40 seconds was used. The x-ray and γ -ray densitometers intensities through the bed were obtained. From the calibration curves, the time average values of volume fractions for gas, liquid and solid phases were computed.

3.4 Results and Discussion

3.4.1 Flow Regime. Three different flow regimes normally occur in three phase fluidized beds : the coalesced bubble, the dispersed bubble, and the slugging regimes (Fan, 1989). Figure 3.3 shows a typical flow regime map (Deckwer et al., 1980) for a liquid-batch bubble column or a slurry-batch slurry bubble column containing a low viscosity liquid phase.

Figure 3.4 shows a typical photograph of the three-phase fluidization experiment. The experimental conditions are shown in the Table 2.1. The superficial velocities for gas and liquid phases were 3.37 cm/sec and 2.03 cm/sec, respectively. As seen in Figure 3.4, three distinct bubbles of about 3.0 cm diameter are visible on the bottom left, bottom right and top center of the bed. As reported in Chapter 2, the mean bubble diameter at the time of its release from the gas distributor was only 0.124 mm. Thus, the two-

dimensional three phase fluidized bed was operating in the coalesced bubble regime, which was due to a low superficial liquid velocity and a high superficial gas velocity. The gas bubbles formed on both sides of the bed, moved towards the center and coalesced together forming bigger bubbles. The bigger bubbles moved upwards through the bed with a larger velocity.

Bouillard and Gidaspow (1991) have shown that in a gas-solid fluidized bed, when the gas at the bottom moves faster than in the rest of the bed, there is a catch-up effect and large gas voids may form. It is the packing effect of the bubbles that produces this regime. When the injected bubbles are far apart, for low gas velocities, the bubbles move up at constant velocities, as in a dispersed flow regime. However, at high gas velocities, the gas volume fraction is high. Here the paths of the small injected bubbles intersect and the bubbles coalesce. The situation is similar to that described in chapter 6 for gas-solid fluidized beds by Gidaspow (1994).

The flow regimes for concurrent air-water-particle fluidized beds have been mapped out by Muroyama and Fan (1989) in terms of gas and liquid superficial velocities. Based on the experimental operating conditions, Figure 3.5 (Muroyama and Fan, 1985) also shows that the flow regime is the coalesced bubble regime.

3.4.2 Volume Fractions. The experimental volume fraction measurements were made at 240 different locations at every 1.27 cm in both x and y directions on the left portion of the bed as shown in Figure 3.6. Figures 3.7(a) to 3.7(c) show the experimental results for the time-averaged volume fractions of gas, liquid and solid, respectively.

Figures 3.7(b) and 3.7(c) show that there exists a region of a maximum value of

time averaged volume fraction of gas, slightly shifted away from the center line, that has a value of approximately 0.7. This region can be interpreted as a region through which most of the coalesced bubbles pass. Figures 3.7(a) and 3.7(c) show that there are solids close to the wall. This is due to the downward flow of the gas and the liquid near the wall, causing the solids to entrain and move down in that region.

Figure 3.1. An x-Ray Calibration Curve for Lead Glass Beads

Figure 3.2. A g-Ray Calibration Curve for Lead Glass Beads

Figure 3.3. Flow Regime Map for a Liquid-batch Bubble Column or a Slurry-batch Slurry Bubble Column Containing a Low Viscosity Liquid Phase (from Deckwer et al. 1980)

Figure 3.4. Photograph of Coalescence Regime in a Gas-Liquid-Solid Fluidized Bed, with Uniform Liquid and Gas Injection

Figure 3.5. Flow Regime Diagram for Cocurrent Gas-Liquid-Solid Fluidized Beds (from Muroyama and Fan, 1985)

Figure 3.6. The Portion of The Bed Where Measurements Were Obtained

Figure 3.7(a). Experimental Time Average Volume Fraction Contours for liquid

Figure 3.7(b). Experimental Time Average Volume Fraction Contours for Gas

Figure 3.7(c). Experimental Time Average Volume Fraction Contours for Solid

CHAPTER IV

POWER SPECTRA ANALYSIS OF BUBBLE FREQUENCY

4.1 Introduction

The porosity or concentration fluctuation and bubble frequency measurements are important from the point of view of erosion of fluidized bed combustor heat transfer tubes. Baumgarten and Pigford (1960) used the radiation densitometer for porosity fluctuation measurements. Recently, Gidaspow (1986) used the γ -ray densitometer to measure porosity fluctuations in gas-solid fluidized beds. Gidaspow et al. (1988) used a spectral analysis program in conjunction with an IBM compatible personal computer to analyze the porosity fluctuation data. Figure 4.1 (Gamwo, 1992) illustrates the setup developed for the porosity measurements for gas-solid fluidized bed experiments. This unit was used previously by other researchers (Gidaspow et al, 1983; Gamwo and Gidaspow, 1992) for experiments conducted at the Illinois Institute of Technology, Chicago.

In this chapter, the concentration fluctuations measurements in two-phase and three-phase fluidized beds are described and analyzed. The concentration fluctuation signals were detected by using a γ -ray absorption technique. The frequencies of concentration oscillations at various locations have been analyzed by means of the fast Fourier transform (FFT) power spectrum method.

4.2 Experimental Setup and Procedure

The experimental setup used for porosity acquisition and FFT power spectra

Figure 4.1. Schematic Diagram of Porosity Acquisition System (Gamwo, 1992)

analysis consisted of three major components : the fluidization equipment, a densitometer assembly and an FFT analyzer. The descriptions of the fluidization equipment and the densitometer assembly are given earlier in Chapter 2. A schematic diagram of the experimental setup is shown in Figure 2.1.

4.2.1 Experimental Procedure. The instantaneous porosity fluctuation measurements were made for both the three-phase (gas-liquid-solid) and the two-phase (gas-liquid) fluidized beds. In the experiments, a high speed camera with 60 frames per second was used to record the bubble paths, bubble diameter and bubble velocity. The experimental measurements were taken at many different locations. The dominant frequencies of porosity fluctuations were calculated using the power spectrum analyzer. The experimental procedure used in these measurements is described below.

4.2.1.1 Gas-Liquid-Solid Fluidized Bed. In order to measure concentration fluctuations in the gas, liquid and solid phases, the radiation source and detector were placed at a desired location by moving them horizontally and vertically using built-in tracks. All the measurements were initiated when the air stream was opened while the liquid was continuously flowing into the bed, and the particles were in a fluidized state as a result of liquid flow. The initial bed height was 25.4 cm. Instantaneous readings using an integrator time constant of 0.1 seconds were obtained for a sampling time of 40 seconds and a sampling frequency of 10 Hz.

4.2.2.2 Gas-Liquid Fluidized Bed. The experimental data were recorded at the instant the bed was fluidized by gas while liquid was continuously flowing into the bed. The instantaneous readings using an integrator time constant of 0.01 seconds were

obtained for a sampling time of 40 seconds, and a sampling frequency of 90 Hertz.

4.3 Experimental Concentration Fluctuation and Power Spectra Results

The major results are summarized for experimental measurements at the following three locations :

1. $X = 15.24$ cm, $Y = 3.81$ cm,
2. $X = 10.16$ cm, $Y = 12.7$ cm,
3. $X = 12.7$ cm, $Y = 12.7$ cm.

4.3.1 Gas-Liquid-Solids Fluidized Bed. Figure 4.2 shows the experimental transient concentration fluctuations in a gas-liquid-solids fluidized bed at a location 15.24 cm from the left wall (X) and 3.81 cm from the bottom of solids bed (Y). This location is on the axial centerline of the fluidized bed. Figure 4.3 shows the corresponding power spectra versus frequency diagram. Figure 4.3 shows that there is no dominant frequency indicating an absence of bubble break-up and resulting particles concentration fluctuation. This location is only 3.81 cm (1.5 inches) from the bottom of the bed. The bubbles introduced from the gas-distributor are of only 0.3 cm diameter. These small bubbles are still growing because of coalescence and no bubble break-up is taking place. Figures 4.4 and 4.5 show experimental concentration fluctuation and corresponding power spectra, respectively, at a different location $X = 10.16$ cm and $Y = 12.7$ cm. Figure 4.5 shows that there are two dominant frequencies corresponding to 0.56 Hz and 6.2 Hz. The dominant frequencies are related to concentration fluctuations in gas and solids phases, respectively. This location is 12.7 cm (5.0 inches) above the bottom of the bed where

Figure 4.2. Experimental Measurement of Fluctuation of Bubble at $X=15.24$ cm and $Y=3.81$ cm for Gas-Liquid-Solid System

Figure 4.3. Approximated Porosity Power Spectrum at X=15.24 cm and Y=3.81 cm for Gas-Liquid-Solid System

Figure 4.4. Experimental Measurement of Fluctuation of Bubble at $X=10.16$ cm and $Y=12.7$ cm for Gas-Liquid-Solid System

Figure 4.5. Approximated Porosity Power Spectrum at X=10.16 cm and Y=12.7 cm for Gas-Liquid-Solid System

some bubble break-up is already taking place.

Figures 4.6 and 4.7 also show experimental concentration fluctuation and corresponding power spectra, respectively, at a different location $X = 12.7$ cm and $Y = 12.7$ cm. This location is at the same height but 2.54 cm (1 inch) left of the previous location. At this location, there seem to be three dominating frequencies at 0.1 Hz, 0.75 Hz and 6.2 Hz. It is plausible that the third frequency at 0.1 Hz is related to the flow behavior and interaction in the center and wall regions of the fluidized bed.

4.3.2 Gas-Liquid Fluidized Bed. The experiments in gas-liquid fluidized bed were conducted to observe the differences in fluidization behavior in the absence of particles. Figures 4.8 to 4.13 show the transient porosity oscillations, and the corresponding power spectra versus frequency plots in a gas-liquid fluidized bed at the same locations where measurements were obtained for the gas-liquid-solids experiments described above.

The power spectra versus frequency at location 1 shown in Figure 4.9 shows no dominant frequency. This situation is similar to that observed in a gas-liquid-solid fluidized bed and confirms the fact that the small bubbles are still growing because of coalescence.

However, Figure 4.11 at location 2 also shows that there is no dominant frequency. In this case, the bubbles are moving up fast in the absence of particles. Thus, a low residence time is available to bubbles, and bubble coalescence is still continuing. Also, there are no particles present to cause bubble break-up. Figure 4.13 at location 3 shows a dominant frequency at 6.0 Hz. In the absence of bubble break-up, the flow interactions

between the center and wall regions of the bed are a likely reason. This interaction is similar to that presented for third dominant frequency in the gas-liquid-solid fluidization experiment. The dominant frequency in a gas-liquid fluidized bed is large signifying a slower rate of interaction.

Overall, the figures show that even though there were interactions between the gas and the liquid phases because of the absence of particles, the disintegration (break-up) of bubbles was occurring at a lower rate in the gas-liquid fluidized bed as compared to those in the three-phase fluidized bed. Also, the bubbles were moving at higher speeds, and larger bubble diameters were developed. Hence, the bubble propagation, collision, and bursting in a fluidized bed generate the void fluctuations.

Although, in a gas-liquid fluidized bed, we visually observed more frequencies of bubbles with larger diameter and faster speed than those in the three-phase fluidized bed, due to the low sensitivity of the detector and the high bubble velocity, Figures 4.8 to 4.13 show that hardly any bubble occurred in those positions.

Figure 4.14 shows the locations where the measurements for porosity oscillations were obtained. The data of major frequencies, seen in Figure 4.14, are also listed in Table 4.1. The dominant frequencies of porosity fluctuations were calculated using the power spectrum analyzer.

Figure 4.6. Experimental Measurement of Fluctuation of Bubble at $X=12.70$ cm and $Y=12.7$ cm for Gas-Liquid-Solid System

Figure 4.7. Approximated Porosity Power Spectrum at X=12.70 cm and Y=12.7 cm for Gas-Liquid-Solid System

Figure 4.8. Experimental Measurement of Fluctuation of Bubble at $X=15.28$ cm and $Y=3.81$ cm for Gas-Liquid System

Figure 4.9. Porosity Power Spectrum at X=15.28 cm and Y=3.81 cm for Gas-Liquid System

Figure 4.10. Experimental Measurement of Fluctuation of Bubble at $X=10.16$ cm and $Y=12.7$ cm for Gas-Liquid System

Figure 4.11. Porosity Power Spectrum at X=10.16 cm and Y=12.70 cm for Gas-Liquid System

Figure 4.12. Experimental Measurement of Fluctuation of Bubble at $X=12.7$ cm and $Y=12.7$ cm for Gas-Liquid System

Figure 4.13. Porosity Power Spectrum at X=12.70 cm and Y=12.70 cm for Gas-Liquid System

Figure 4.14. Bubble Frequencies along the Bubble Path

Table 4.1 Bubble Frequencies in Three-Phase Fluidized Bed at Different Locations

Point	Location		Frequency (Hz)
	X (cm)	Y (cm)	
1.	12.7	17.78	None
2.	15.24	2.54	None
3.	7.62	2.54	0.5, 5.4
4.	12.7	3.81	None
5.	15.24	5.08	0.48, 5.0
6.	12.7	5.08	None
7.	7.62	5.08	0.75
8.	10.16	7.62	None
9.	12.7	7.62	None
10.	12.7	10.16	8.0
11.	10.16	12.7	0.56, 0.62
12.	12.7	12.7	0.1, 0.75, 6.2
13.	12.7	20.32	None
14.	7.62	3.81	0.3, 1.0, 5.6
15.	15.24	22.86	None
16.	10.16	20.32	None
17.	7.62	7.62	None

4.4 Computed Concentration Fluctuation and Power Spectra Results

The experimental concentration fluctuations were compared with those obtained using hydrodynamic simulations of the fluidized bed. The details of gas-liquid-solids fluidized bed simulations are given in Chapter 7. The computational results shown in this chapter are for the simulations in asymmetric mode.

Figures 4.15 to 4.20 show the computational results of porosity distributions and power spectra diagrams.

In order to evaluate the validity of power spectra analysis, the predicted time-average gas volume fractions shown in the Figure 7.1 were compared with those in the experiment. Figure 7.1(b) shows that the computed time averaged gas volume fraction of gas, at location $X = 10.16$ cm and $Y = 12$ cm, is very low, 0.3. As a result, at that location, there is no dominant frequency as seen in Figure 4.18. However, Figures 3.7(b) and 4.5 agreed quite well when the experimental measurements showed that the gas volume fraction at location $X = 10.16$ cm and $Y = 12.7$ cm was 0.64. The dominant frequencies of the bubble were 0.58 Hz and 6.2 Hz.

4.5 Conclusion

1. The porosity fluctuations data provided experimentally verifiable results. Such information is useful in designing a fluidized bed with heat transfer tubes.
2. The comparison between experimental and simulation data was fair.

Figure 4.15. Computational data of Fluctuation of Bubble at $X=15.28$ cm and $Y=4$ cm for Gas-Liquid-Solid System

Figure 4.16. Porosity Power Spectrum at X=15.24 cm and Y=4 cm for Gas-Liquid-Solid System

Figure 4.17. Computational data of Fluctuation of Bubble at $X=10.16$ cm and $Y=12$ cm for Gas-Liquid-Solid System

Figure 4.18. Porosity Power Spectrum at X=10.16 cm and Y=12 cm for Gas-Liquid-Solid System

Figure 4.19. Computational data of Fluctuation of Bubble at $X=13.20$ cm and $Y=12$ cm for Gas-Liquid-Solid System

Figure 4.20. Porosity Power Spectrum at X=13.20 cm and Y=12 cm for Gas-Liquid-Solid System

CHAPTER V

PARTICLE VELOCITY DISTRIBUTION IN TWO-PHASE AND THREE-PHASE FLUIDIZED BED

5.1 Introduction

A number of researchers have investigated the movement of particles in the fluidized bed. Chen and Fan (1992) used PIV system (Particle Image Velocimetry) to characterize the flow structure in a three-dimensional Gas-Liquid-Solid fluidized bed. Carlos and Richardson (1968) measured particles velocities in a liquid-solid fluidized bed. The measurements were done by photographing the movement of tracer particles inside the bed, where dimethyl phthalate was used as liquid, glass beads with 0.8 cm in diameter as solids, and darkened glass beads with the same diameter as tracer particles. The liquid had the same refractive index as the particles. Hence, only the darkened particles were visible. They demonstrated that the velocity distribution follows an approximately Maxwellian distribution about its mean.

Recently, Devanathan et al. (1990) monitored the motion of a single neutrally buoyant radioactive particle by an array of scintillation detectors in an air-water system, and measured the time and ensemble averaged tracer particle velocities. They adopted this technique, called Computer Average Radio-active Particle Tracing (CARPT), to obtain time and volume averaged liquid flow fields in a bubble column.

The objective of this experimental investigation was to conduct well-defined fluidized bed experiments to measure the instantaneous particle velocities. It would also

provide useful data for distinguishing between the distribution function for velocity fluctuations and the Maxwellian distribution. The data provided by the experiments are particles velocities components in the axial and radial directions, and the bubble sizes. Experiments were conducted in a two-dimensional fluidized bed made of transparent acrylic (plexiglas) with a uniform air distributor and also in a two-dimensional fluidized bed with a central jet.

In this chapter, a direct imaging technique is used to measure the instantaneous velocity or velocity distribution in both the gas-liquid-solid fluidized bed as well as the liquid-solid fluidized bed experiments. The effects of gas and liquid flow rates and particle concentration on macroscopic flow behavior including particles movement, bubble size and liquid-solid circulation were examined. From these experimental data, the values for granular temperature and solids viscosity were computed, as described by Gidaspow (1994). The computed solid viscosity was in agreement with the experimentally measured solids viscosity, described in detail in the Chapter 6.

5.2 Apparatus and Experimental Procedure

The apparatus for the velocity measurements consists of essentially two units: a high resolution micro-imaging / measuring system and a fluidized bed. In this study, the experimental studies were carried out in two different types of fluidized beds, namely, a two-dimensional fluidized bed with a uniform distributor and a two-dimensional fluidized bed with central jet. The experimental apparatus is shown schematically in Figures 5.1 and 5.2.

5.2.1 High Resolution Micro-Imaging / Measuring System. This system comprises two units :

1) SONY high resolution color CCD video camera equipped with electronic shutter speed settings ranging from OFF to 1/10000 sec and SONY super fine pitch color monitor, and

2) A 486 / 33 MHz IBM compatible personal computer with a Micro-Imaging Board inside and a Micro-Imaging software, Image-Pro Plus.

5.2.2 Fluidized Bed with Uniform Distributor. Figure 5.1 shows a two-dimensional acrylic (plexiglas) rectangular fluidized bed with a height of 213.36 cm, width of 30.48 cm, depth (thickness) of 5.08 cm, and with a uniform gas distributor. The ballotini (leaded glass beads) particles, with 800 μm in diameter and a density of 2.94 g/cm^3 , are fluidized inside the bed by means of water and air. Water from a fifty-five gallon storage tank is introduced into the bed from the bottom through a rotameter by means of a centrifugal pump. Air from a compressor enters through a distributor section consisting of six porous tubes with a nominal pore diameter of 41.88 μm and a perforated plate. These six staggered tubes are located below the perforated plate. The distributor section is located at 50 cm (height) in the bed.

5.2.3 Fluidized Bed with Central Jet. Figure 5.2 shows a two-dimensional acrylic (plexiglas) rectangular fluidized bed with a height of 233.7 cm, width of 48.3 cm, depth of 2.54 cm, and with a central jet. The particles used in this fluidized bed were glass beads with a 1 mm diameter and a density of 2.55 g/cm^3 . Water is fed to the bed in the same way as that in the previous bed. The air from the compressor enters the bed

through three separate rotameters. Two rotameters are connected to the left bottom and the right bottom inlets of the bed. The third rotameter is connected to the central bottom inlet of the bed, which is used as a central jet inlet, and is located right below the distributor section. The distributor section consists of one perforated plate with many 0.08 cm diameter holes and is located at 50 cm from the bottom of the bed.

5.2.3 Experimental Procedure. The gas and liquid were injected inside the fluidized bed at several different velocities. These velocities are reported in the graphs. As the gas and liquid entered the bed through the distributor section, the particles were fluidized. Both the gas and liquid exited from the top of the bed, and were directed through three 1 inch (2.54 cm) diameter openings back to the fifty gallon storage tank, where the gas was separated from the liquid.

In order to get a good visualization of microscopic movement of particles, a fiber-optic light was reflected on the field of view in the front and the back of the bed. The field of view in most experiments was a 2 cm x 2 cm area. As the particles were fluidized inside the bed, the camera with a zoom lens with a range of 18-108 mm and close up focus transferred its field of view to the monitor with streak lines. These streak lines represented the distance traveled by the particles in a given time interval specified on the camera. The images were then captured and digitized by a micro-imaging board and analyzed using Image-Pro Plus software. Measurements were conducted at several different radial and axial locations (see Figure 5.3).

Figure 5.1. Schematic Diagram of Experimental Apparatus of Fluidized Bed W/uniform Distributor

Figure 5.2. Schematic Diagram of Experimental Apparatus of Fluidized Bed W/central Jet Distributor

Figure 5.3. The Image Captured by the Recording System

5.3 Analysis of Results

In this investigation, the computer recorded the distances traveled by the particles in both -x and -y directions. Using these data, instantaneous particles speed, axial and radial velocities, their variances, mean axial and radial velocities, and standard deviations were calculated. Also, the experimental velocity distribution was examined and analyzed with these set of data.

The experiments were conducted in two parts, (a) with a liquid-solid fluidized bed, and (b) with a gas-liquid-solid fluidized bed. In the gas-liquid-solid fluidized bed experiments, two different designs, (i) with uniform distributor, and (ii) with central jet were studied.

5.3.1 Liquid-Solid Fluidized Bed Experiment. Figures 5.4, 5.5 and 5.6 show typical experimental data that are plotted for the distribution function, f , as a function of instantaneous particles speed, axial velocity and radial velocity, respectively. Figure 5.7 also shows frequency plot for solids speed at different locations. The frequency, f , expressed as the fraction of the total number of velocities falling in the interval, divided by the size of the interval and the values for velocities were made of the mean value of 120 data at each position. As seen in Figures 5.4 to 5.7, the velocity fluctuations follow normal distribution. The normal distribution for solids fluctuating velocities, also called Maxwellian distribution, can be written as follows (Gidaspow, 1994):

$$f(c_y) = \frac{1}{\sqrt{2\pi\sigma^2}} \exp\left(-\frac{(c_y - v_y)^2}{2\sigma^2}\right) \quad (5.1)$$

where σ^2 is variance and is equal to the granular temperature Θ , c_y and v_y are the

Figure 5.4. Distribution of Mean Solid Speed at X=4 cm and Y=4.5 cm for Liquid-Solid System with $V_l=4$ cm/sec

Figure 5.5. Distribution of Axial Velocity of Particles at $X=4$ cm and $Y=4.5$ cm for Liquid-Solid System with $V_f=4$ cm/sec

Figure 5.6. Distribution of Radial Velocity of Particles at X=4 cm and Y=4.5 cm for Liquid-Solid System with $V_f=4$ cm/sec

Figure 5.7. Distribution of Mean Solid Speed at X=4 cm and Y=4.5 cm for Liquid-Solid System with $V_l=6$ cm/sec

particles velocity and mean particle fluctuations in the axial (-y) direction. Similarly, equation (5.1) can also be written for the velocity fluctuations in -x and -z directions.

Figure 5.8 corresponds to the data which were taken by Carlos and Richardson (1968). However, even taking into account that in their experiment, the overall mean particle velocities are zero, the agreement between Figures 5.4 to 5.7 with Figure 5.8, and the Maxwellian distribution is reasonably close. Figures 5.5 and 5.6 clearly show that the two phase (liquid-solid) fluidized bed is anisotropic, and the axial velocity fluctuations greatly exceed the radial velocity fluctuations.

The effect of liquid velocity on particles velocity is plotted in Figure 5.9. Figure 5.9 shows that at higher liquid velocities, more oscillation and collision between the particles occur and, as a result, the granular temperature is higher. Figure 5.10 shows the solid viscosity calculated from granular temperature using equations (7.2) to (7.4). As expected, the calculated solids viscosity increases as the solid volume fraction increases.

5.3.2 Gas-Liquid-Solid Fluidized Bed Experiment. In order to study the effects of the gas flow on the particles velocity, the experiments were performed in the following two beds : fluidized bed with uniform gas distributor, and fluidized bed with central jet.

5.3.2.1 Fluidized Bed with Uniform Distributor. The frequency plots for solids speed, and axial and radial velocities are shown in Figures 5.11, 5.12, and 5.13, respectively. The difference between these measurements and those shown in Figures 5.4 to 5.6 for two-phase fluidized bed experiments was the presence of the gas flow. Figures 5.11 to 5.13 clearly show that in three-phase fluidized bed the velocity fluctuation

distributions are also a Maxwellian distribution.

Figure 5.14 shows speed fluctuation distribution for the same gas and liquid velocity as in Figure 5.11, but at a different location. Figure 5.15 represents the speed fluctuation distribution for the same location as in Figure 5.15, but at a higher liquid flow rate. These figures clearly demonstrate that the velocity fluctuation in two and three-phase fluidized beds follow a Maxwellian distribution. Hence, the basic assumptions used to derive the Kinetic theory of granular solids applies to a fluidized bed.

Figure 5.16 shows the axial velocity profile as a function of radial distance at three different bed heights. It can be seen that the velocities near the wall are moving in a downward direction and far from the wall there are higher peaks. These peaks are due to the bubbles path. These velocity profiles are also observed visually.

The two-dimensional fluidized bed used in the experiment allows flow predominantly in two directions, namely axial (-y) and radial (-x). Hence, the instantaneous velocities measurements were made in the axial-radial (x-y) plane only. However, the -z direction also has a non-zero component of particles fluctuations because of a finite bed depth. Due to the method of selecting the steak lines for measurement it is reasonable to assume that the $\langle c_z^2 \rangle = \langle c_x^2 \rangle$, and that both the -z and -x directions have zero mean gas and liquid velocities. Thus, the velocities fluctuations can be written as:

$$\begin{aligned} \langle c^2 \rangle &= \langle c_x^2 \rangle + \langle c_y^2 \rangle + \langle c_z^2 \rangle \\ &= 2 \langle c_x^2 \rangle + \langle c_y^2 \rangle \end{aligned} \quad (5.2)$$

where $\langle c_x^2 \rangle$, and $\langle c_y^2 \rangle$ are the measured mean square velocity fluctuations in -x and -

Histogram of radial velocities.

Histogram of axial velocities.

Figure 5.8. Velocity Distribution of Particles (Carlos, and Richardson, 1968)

Figure 5.9. Solids Granular Temperature as Function of Liquid Velocity at $X=4$ cm, and $Y=4.5$ cm

Figure 5.10. Solids Viscosity As Function of Liquid Velocity in Liquid-Solid Fluidized Bed

Figure 5.11. Distribution of Mean Solid Speed at X=4 cm and Y=4.5 cm for Liquid-Solid-Gas System with $V_l=4$ cm/sec and $V_g=3.36$ cm/sec

Figure 5.12. Distribution of Axial Velocity of Particles at $X=4$ cm and $Y=4.5$ cm for Liquid-Solid-Gas System with $V_f=4$ cm/sec and $V_g=3.36$ cm/sec

Figure 5.13. Distribution of Radial Velocity of Particles at $X=4$ cm and $Y=4.5$ cm for Liquid-Solid-Gas System with $V_f=4$ cm/sec and $V_g=3.36$ cm/sec

Figure 5.14. Distribution of Mean Solid Speed at $X=7.5$ cm and $Y=9.0$ cm for Liquid-Solid-Gas System with $V_l=4$ cm/sec and $V_g=3.36$ cm/sec

Figure 5.15. Distribution of Mean Solid Speed at $X=7.5$ cm and $Y=9.0$ cm for Liquid-Solid-Gas System with $V_l=7.23$ cm/sec and $V_g=3.36$ cm/sec

y directions, respectively.

The expression for granular temperature in these experiments can be written as,

$$\Theta = \frac{1}{3} \langle c^2 \rangle = \frac{2}{3} \langle c_x^2 \rangle + \frac{1}{3} \langle c_y^2 \rangle \quad (5.3)$$

In Figures 5.17 to 5.19 granular temperature is plotted versus width of the bed at three different heights. It is clear that granular temperature is at a maximum value near the center at the bottom of the bed where the measurement for viscosity was also obtained by Brookfield viscometer (see chapter 6).

5.3.2.2 Fluidized Bed with Central Jet. Figures 5.20, 5.21 and 5.22 represent solids speed, axial and radial velocities distribution in the fluidized bed with the central jet. The measurements for these figures were obtained close to the center of the bed. Although in a fluidized bed with a central jet there is a lot of disturbance on the movement of particles, by comparing the figures from the previous section with the figures in this section, one can easily conclude that in both fluidized beds the velocity distribution is Maxwellian. In Figure 5.23, the solid viscosity is increasing with solid volume fraction, very similar behavior was also observed in the fluidized bed with a uniform distributor.

5.4 Conclusion

1. The computed solid viscosity is in agreement with the experimental data from the measured solids viscosity data from experiment (see Chapter 6).
2. The velocity distribution almost every where in the bed and in all types of gas

and liquid flow rate is Maxwellian distribution.

3. The velocity distribution is a Maxwellian distribution in liquid-solid fluidized bed as well as in gas-liquid-solid fluidized bed.

4. The velocity distribution is Maxwellian in both three-phase fluidized beds : fluidized bed with a uniform distributor, and fluidized bed with central jet.

Figure 5.16. Experimental Velocity Profile as Function of Radial Distance in Liquid-Solid-Gas System with $V_f = 3.7$ cm/sec and $V_g = 7.45$ cm/sec

Figure 5.17. Solid Granular Temperature Vs. Radial Distance

Figure 5.18. Solid Granular Temperature Vs. Radial Distance

Figure 5.19. Solid Granular Temperature Vs. Radial Distance

Figure 5.20. Distribution of Mean Solid Speed at X=23 cm and Y=10 cm for Liquid-Solid-Gas System with $V_i=439$ cm/sec and $V_{g1}=243$ cm/sec $V_{g2}=578$ cm/sec $V_{g3}=370$ cm/sec

Figure 5.21. Distribution of Axial Velocity of Particles at $X=23$ cm and $Y=10$ cm for Liquid-Solid-Gas System with $V_l=439$ cm/sec and $V_g=243$ cm/sec $V_{g2}=578$ cm/sec $V_{g3}=370$ cm/sec

Figure 5.22. Distribution of Radial Velocity of Particles at $X=23$ cm and $Y=10$ cm for Liquid-Solid-Gas System with $V_l=439$ cm/sec and $V_{g1}=243$ cm/sec $V_{g2}=578$ cm/sec $V_{g3}=370$ cm/sec

**** Calculated From Meas. of Granular Temp. Using Kinetic Theory

Figure 5.23. Solid Viscosity as Function of Liquid Velocity in Liquid-Solid-Gas Fluidized Bed with Central Jet Distributer

CHAPTER VI

SOLIDS VISCOSITY MEASUREMENTS IN IN THREE-PHASE FLUIDIZED BED

6.1 Introduction

The thermo-physical properties of three-phase fluidized beds have a major impact on the analysis and design of reactors in the biological, chemical, petroleum, food and pharmaceutical industries. The viscosity is one of these properties.

Viscosity is a measure of the internal friction when one "fluid" layer is made to move relative to another layer. For Newtonian fluids, the viscosity can simply be defined mathematically by the following formula,

$$\tau = \mu \gamma \quad (6.1)$$

where τ is the shear stress, its unit of measurement is dynes per square centimeter (dynes/cm^2); γ is the shear rate, its unit of measure is called the reciprocal second (sec^{-1}); and μ is the viscosity which has the unit of poise.

In three-phase fluidized beds, the particles can be treated as a continuum with a scale length that is much larger than the particle diameter. The rheological behavior of such continuum results from the cumulative effects of macroscopic (at the length scale of the particle) forces due to interaction between particles and hydrodynamic drag forces. Thus, the gas-liquid-solid mixtures are always more viscous than the liquid from which they are made. Unlike most of Newtonian fluids that are formed as pure substances such as water, multiphase fluids generally behave like non-Newtonian fluids and, as a result,

they do not have a single value for viscosity.

The viscosity measurement of the fluidized bed and its correlation with a glass-forming liquid model was first introduced by Hetzer and Williams (1969), when they analyzed average bed viscosity and expansion for homogeneous, water-fluidized beds of uniform glass spheres having diameters from 32 to 487 μm . A review of experimental studies for the measurements of viscosity in gas-solids fluidized beds is given by Cherimisinoff and Cherimisinoff (1984).

In a three-phase fluidized bed, the rheological behavior of the bed predominantly reflects the granular viscosity of the solids. Gidaspow (1994) expressed this term as:

$$\mu_s = \frac{2\mu_{dil}}{(1+e)g_o} \left[1 + \frac{4}{5} (1+e) g_o e_s \right]^2 + \frac{4}{5} e_s^2 \rho_s d_p (1+e) g_o \left(\frac{\Theta_s}{\pi} \right)^{\frac{1}{2}} \quad (6.2)$$

Dilute Phase Viscosity :

$$\mu_{dil} = \frac{5\sqrt{\pi}}{96} \rho_s d_p \sqrt{\Theta_s} \quad (6.3)$$

and Radial Distribution Function,

$$g_o = \left[1 - \left(\frac{e_s}{e_{s,max}} \right)^{\frac{1}{3}} \right]^{-1} \quad (6.4)$$

where Θ_s is granular temperature that can be obtained from the kinetic oscillation energy balance. It is apparent from the above expression that the particle size, density and packing also affect bed rheology.

Song and Fan (1986) expressed the viscosity of three-phase fluidized bed as "apparent" (or effective) overall viscosity, which is due to the viscosities of the gas and

liquid phases, particle-particle interactions, and bed expansion. This "apparent" (or effective) viscosity for a homogeneous three-phase fluidized bed is defined as,

$$\mu_{bed} = e_g \mu_g + e_l \mu_l + \mu_s \quad (6.5)$$

They found that the bed voidage and gas flow rate are important factors in the three-phase fluidized bed. It is also observed in this experiment that the bed voidage is influenced by the gas and liquid flow rates, and the fluctuations in "apparent" bed viscosity correlate with the bed voidage.

The objective of this chapter is focused on the measurement of viscosity of three-phase fluidized bed under actual operating conditions, using a simple viscometric flow generated by a vertical rotating cylinder. Thus the measured viscosities would be close to those required in the hydrodynamic analysis. In addition, some experiments are conducted in the two-phase fluidized bed as well.

6.2 Experimental Setup and Procedure

The experimental setup consisted of two major parts : the fluidization equipment, and a Brookfield viscometer. A schematic of the experimental setup is shown in Figure 6.1.

6.2.1 Fluidization Equipment. The experiment was conducted in a vertical plexiglas column of 30.48 cm width, 213.36 cm height and 5.08 cm breadth. The bottom of the column (or bed) consisted of gas and liquid distributors. The liquid was pumped to the bed through a 1.0 inch (2.54 cm) I.D. stainless steel pipe, located at the bottom of the bed. The liquid was distributed by two perforated plexiglas plates with many 0.277 cm diameter holes. They were placed at 35.56 cm and 50.8 cm above the bottom of the

Figure 6.1. Schematic Diagram of the Experimental Apparatus

bed. The gas was distributed to the bed through six staggered tubes with fine pores of $41.88\text{ }\mu\text{m}$ diameters. The gas distribution tubes were located between the perforated plexiglas plates.

6.2.2 Brookfield Viscometer. A Brookfield digital viscometer (model LVDV-II+) with spring torque of 673.7 dyne-cm was used to measure the effective bed viscosities in this experiment. This viscometer can produce twenty different rotational speeds ranging from 0 to 100 revolutions per minute (rpm) at four different modes, namely, LV, RV, HA, and HBDVII+. The viscometer readings were recorded with a Hewlett-Packard (HP) LaserJet series II printer.

6.2.3 Experimental Procedure. The liquid from the storage tank was fed to the bottom of the bed using a centrifugal pump. The gas was fed to the bed through a compressor. The viscometer was placed at the top of the bed, and secured over the centerline of the bed. A cylindrical spindle (#1 LV) of 0.9421 cm diameter, 7.493 cm effective length and overall height of 11.50 cm was used. The cylindrical spindle was attached to the bottom of the viscometer without the guard and was lowered inside the fluidized bed by an extension wire until it was completely immersed in the three-phase mixture.

Air and water were used as the gas and liquid, respectively, in this experiment. Ballotini (leaded glass beads) with a diameter of $800\text{ }\mu\text{m}$ and a particle density of 2.94 g/cm^3 were used as the solids. A constant superficial gas velocity of 3.364 cm/sec was used throughout the experiment, while the superficial liquid velocities were varied between two different values, 2.037 cm/sec and 4.037 cm/sec . The experiments were

conducted at a temperature of 20°C.

The measurements in this experiment were made under LV mode at six different speeds consisting of 2, 4, 6, 10, 12 and 20 rpm. At each rotational speed, between 10 and 30 readings were taken. The calibration of the viscometer-spindle apparatus was done using a Newtonian liquid, namely, water.

The bed expansion measurements were made by visual observation of the height of the bed's upper surface at two different liquid velocities as compared to the initial (close-packed) bed height. The bed void fraction was calculated by the following relation (Kim, et al., 1972) :

$$(\epsilon_1)_{u_g} - \epsilon_1 = 0.0025 \left(Fr_1 \frac{\rho_s}{\rho_l} \right)^{0.149} \left(Fr_g \frac{\rho_s}{\rho_g} \right)^{0.161} (Re_l Re_g)^{0.259} \quad (6.6)$$

$$\text{where} \quad (\epsilon_1)_{u_g} = 0.409 \left(Fr_1 \frac{\rho_s}{\rho_l} \right)^{0.193} Re_l^{0.074} \quad (6.7)$$

$$Fr_k = \frac{u_k^2}{g d_p} \quad k = l, g \quad (6.8)$$

$$Re_k = \frac{u_k d_p \rho_k}{\mu_k} \quad k = l, g \quad (6.9)$$

6.3 "Apparent" Bed Viscosity

Figure 6.2 shows the "apparent" bed viscosity as a function of spindle rotation rate (rpm) at two different superficial liquid velocities. As expected, the "apparent" bed viscosity at low liquid flow rate is higher than that at high liquid flow rate for a constant

Figure 6.2. Experimentally Measured Viscosity (cp) Versus Spindle Rotation Rate for Three Phase Fluidized Bed (solid Dia =800 micron, Sup Gas Velo. = 3.36 cm/sec)

Figure 6.3. Experimentally Measured Viscosity Vs. Shear Rate

spindle rotation rate. This effect is directly related to the solid concentration in the bed. As more liquid is distributed in the bed at a constant gas flow rate, the more dilute the three-phase fluidized bed becomes. The "apparent" bed viscosity as a function of shear rate is shown in Figure 6.3. This figure shows that the "apparent" bed viscosity decreases as the shear rate is increased.

A few of the reasonable causes for a decrease in "apparent" viscosity with increasing spindle rotation rate or shear rate are the following:

- (a) centrifugal forces acting on particles close to spindle surface, as well as shear induced migration resulting in less interaction between particles and the spindle; and
- (b) a Bingham plastic like behavior of the three-phase fluidized mixture.

The ratio of centrifugal force and gravity force acting on a particle is,

$$\frac{F_{\text{centrifugal}}}{F_{\text{gravity}}} = \frac{\omega^2 R}{g} \quad (6.10)$$

where, R is spindle radius and ω is spindle rotation rate. For a spindle diameter of 0.9421 cm and, at the maximum spindle rotation rate of 20 rpm used in the experiment, the above ratio is 0.004. Thus, the effect of centrifugal force on bed viscosity is small at low spindle rotation rate only.

It is difficult to measure solids concentration in the vicinity of spindle surface. Thus, the effect of shear induced migration on "apparent" viscosity is difficult to predict. A measurement technique which does not cause the solids segregation may permit better viscosity measurement. The effect of solid segregation can be reduced numerically by extrapolating the measured "apparent" bed viscosities at moderate shear rates to zero shear

rate as shown in Table 6.1 below. In this table, the liquid volume fractions are calculated from Kims's correlation (Kim, et al., 1972) shown in equations (6.6) to (6.9) and solid volume fractions are calculated from visually measured expanded bed heights under fluidizing conditions. Thus the reported volume fractions are averaged values.

TABLE 6.1 "Apparent" Bed Viscosity

Sup. Gas Velocity (cm/s)	Sup. Liq. Velocity (cm/s)	Bed Volume Fractions			"Apparent" Bed Viscosity at zero shear rate (poise)
		Gas	Liquid	Solid	
3.364	2.037	0.314	0.340	0.346	4.1
3.364	4.027	0.263	0.465	0.272	2.2

The experiments were also conducted in a fluidized bed with a central jet with 1.0 mm diameter particles, and different liquid velocities for liquid-solid fluidization. A description of the fluidized bed with central jet is given in Chapter 5. Figures 6.4 and 6.5 show the transient viscosity measurement readings at a speed of 6 rpm for a fluidized bed with uniform distributor and a fluidized bed with central jet, respectively. The fluctuations in the measurements are a result of change in the volume fractions of gas, liquid and solid phases in the fluidized state. In theory, the frequency of these fluctuations can be analyzed using power spectra techniques and correlated with the frequency of porosity fluctuations in the bed.

Figure 6.4. Experimentally Measured Viscosity Vs. Numbers of Run for Spindle Rotation Rate of 6 rpm for Three-Phase Fluidized Bed with Uniform Distributer (Solid Dia. = 0.08 mm)

Figure 6.5. Experimentally Measured Viscosity Vs. Numbers of Run for Spindle Rotation Rate of 6 rpm for Three-Phase Fluidized Bed with Central Jet (Solid Dia. = 1 mm)

6.4 Shear Stress Analysis

For analysis of the behavior of a Bingham fluid, consider a cylinder of radius R rotating at an angular velocity ω in a non-Newtonian fluid. In the momentum equation, we make an assumption that the cylinder radius is large as compared to the fluid shear layer and thus there is a uni-directional flow, $v_x(y)$, and no pressure gradient in the x -direction. Neglecting effects of two side walls of the bed, end effects and assuming the Couette flow, this yields

$$\frac{d\tau_{yx}(y)}{dy} = 0 \quad (6.11)$$

The constitutive equation for Bingham plastic is,

$$\begin{aligned} \tau_{yx} &= -\mu \frac{dv_x}{dy} + \tau_0 & \tau_{yx} &\geq \tau_0 \\ \frac{dv_x}{dy} &= 0 & \tau_{yx} &< \tau_0 \end{aligned} \quad (6.12)$$

where τ_0 is the yield stress. The yield stress defines a shear layer inside in which the shear stress is greater than the yield stress.

The boundary conditions are the no-slip conditions at both the cylinder surface and at the boundary of shear layer,

$$\begin{aligned} v_x(y=0) &= v_0 = \omega R \\ v_x(y=L) &= 0 \end{aligned} \quad (6.13)$$

The solution to equation (6.11) is a linear velocity profile in the shear layer,

$$v_x(y) = v_0 \left(1 - \frac{y}{L}\right) \quad (6.14)$$

The shear stress at the cylinder surface can be written both in terms of effective bed viscosity (μ) and "apparent" bed viscosity (μ_a) as,

$$\begin{aligned}\tau_{yx}(R) &= \mu \omega \frac{R}{L} + \tau_0 \\ &= \mu_a \omega \frac{R}{L}\end{aligned}\tag{6.15}$$

Thus, the "apparent" bed viscosity in terms of effective bed viscosity, yield stress and cylinder rotation rate can be written as,

$$\mu_a = \mu + \frac{\tau_0 L}{\omega R}\tag{6.16}$$

The angular velocity ω is related to revolutions per seconds (rps) of the cylinder (n), by a simple relation $\omega = 2 \pi n$. The equation (6.16) suggests that for a Bingham fluid, "apparent" viscosity is a linear function of the inverse of the cylinder rotation rate.

Figure 6.6 shows a graph of the experimentally measured "apparent" bed viscosity as a function of inverse of spindle rotation speeds. This figure is a variation of Figure 6.2. A good straight line fit of "apparent" bed viscosity can be clearly seen at two different liquid flow rates. Thus, it is evident that the fluidized bed is behaving like a Bingham fluid (Bird, Stewart and Lightfoot, 1960).

Figure 6.7 shows the result for "apparent" viscosity versus the inverse of spindle rotation rates in a two phase fluidized bed. Again, a good straight line fit illustrates that the liquid-solid fluidized bed acts like a Bingham fluid.

6.5 Granular Temperatures in Three-Phase Fluidized Beds

Based on the values shown in Figure 6.2, and equations (6.2) to (6.5), the granular temperatures were calculated from measured bed viscosities.

Figure 6.6. Experimentally Measured Shear Stress Vs. Velocity Gradient for Three-Phase Fluidized Bed (Solid Dia. = 0.08 mm, Sup. Gas Velo. = 3.364 cm/sec)

The granular temperatures were calculated to be in the range of 95 to 500 (cm/sec)² for spindle rotation rates of 6 to 20 rpm. However, at lower spindle rotation rates of 2 to 4 rpm, the calculated granular temperatures were in the range of 1000 to 5000 (cm/sec)².

The mean fluctuating velocity of particles $\langle C \rangle$ is related to the granular temperature as,

$$\langle C \rangle = \sqrt{3\theta} \quad (6.17)$$

Thus, at high spindle rotation rates of 6 to 20 rpm, the fluctuating velocity is in the range of 17 to 39 cm/sec (0.17 to 0.39 m/s); whereas, at spindle rotation rates of 2 to 4 rpm, the fluctuating velocities are in the range of 55 to 122 cm/sec (0.55 to 1.22 m/s).

Typical values of the fluctuating velocities for gas-solid fluidized beds were computed by Ding and Gidaspow (1990) to be in the range of 0.17 to 1.7 m/s. Also, based on solids discharge rates from mass flow hoppers, Gidaspow et al. (1986) obtained critical granular velocities in the range of 56.39 to 113.49 cm/sec (0.5639 to 1.1349 m/s) for 86 to 1550 μm size glass beads and 105.19 cm/sec (1.0519 m/s) for 603 μm sized sand particles.

The mean fluctuating velocities obtained from bed viscosity measurements using a Brookfield viscometer are very close to those reported in Chapter 5 based on the direct velocity measurement experiments. Figures 6.8 and 6.9 show the comparison of the measured viscosity with the Brookfield viscometer and calculated solids viscosity from granular temperature using kinetic theory by direct measurement of solids velocity using a high resolution micro image/measuring system technique. In Figure 6.8 the data for

gas-liquid-solid and liquid-solid are continuing with respect to the solid volume fraction. Also, these figures show there is reasonable agreement between the two technique, mostly in the region with a higher concentration of solids.

6.6 Conclusion

1. The role of the Brookfield viscometer in the direct measurement of apparent bed viscosity and the indirect measurement of granular temperatures or fluctuating velocities in fluidized beds is important, in spite of the limitations imposed because of volume fraction fluctuations, axial velocities, transient radial velocity profile, and degree of turbulence, the multiple phases are in contact with spindle.

2. Good agreement between solids viscosities obtained by using two different and independent measurement techniques suggests that the particles collision is an important mechanism in granular flows. This validation of the central mechanism of the kinetic theory of granular flows has far-reaching consequences. The kinetic theory of granular flows can provide an estimation of the rheological properties of multiphase mixtures. Such rheological properties are important for the modeling of the hydrodynamics of fluidized bed reactors and pneumatic transport.

3. In both two-phase and three-phase fluidized beds the predictions of the Bingham fluid match the experimental data reasonably well.

Figure 6.7. Experimentally Measured Shear Stress Vs. Velocity Gradient for Two-Phase Fluidized Bed (Solid Dia. = 1 mm, Sup. Gas Velo. = 332 cm/sec)

Figure 6.8. Comparison of Measured and Calculated Solid Viscosity In Fluidized Bed

▲▲▲▲▲ Measured by Brookfield Viscometer
***** Calculated From Meas. of Granuler Temp. Using Kinetic Theory

Figure 6.9. Comparison of Measured and Calculated Solid Viscosity In Fluidized Bed with Central Jet

CHAPTER VII

DESCRIPTION OF THE HYDRODYNAMIC MODEL AND SIMULATION

7.1 Introduction

In this study, a transient, two dimensional multiphase computer code for the solution of generalization of Navier-Stokes equations developed earlier for gas-liquid-solids fluidization was verified. The particulate viscosities are either an input into the code or are obtained from fluctuating energy equations derived from the Boltzmann equation for velocity distribution of particles. Previously, a number of investigators have studied and simulated gas-solids bubbling fluidized beds (Seo and Gidaspow, 1987; Gidaspow and Ettehadieh, 1983) and gas-liquid fluidized beds (Gidaspow et al., 1991). The emerging kinetic theory of granular flow (Jenkins and Savage, 1983; Lun et al., 1984) shows how to predict stresses caused by particle collisions. Ding and Gidaspow (1990) applied this theory to gas-solids fluidized beds. A review of modeling of the hydrodynamics of fluidization of bubbling beds by Gidaspow (1986) showed that inviscid two-fluid models were able to predict a great deal of the behavior of bubbling beds because the dominant mechanism of energy dissipation is the drag between the particles and the fluid. The formation, growth and bursting of bubbles were predicted. Predicted wall-to-bed heat transfer coefficients and velocity profiles of jets agreed with measurements.

Although the partial differential equations that we are solving are well-posed as

an initial value problem, a problem of how to prescribe inlet conditions for the gas phase needs to be understood. This problem arises due to the asymmetric circulation of gas, liquid and solids inside the bed. The asymmetric inlet conditions for gas phase give us experimentally observed bubble coalescence and velocity distributions in the three-phase fluidized bed which were not seen for a symmetric bed.

7.2 Hydrodynamic Model

The physical principles used are the laws of conservation of mass and momentum for the gas, the liquid and the solid phases as shown in Table D.1 in Appendix D. The constitutive equation for the stress is also shown in Table D.1 in Appendix D. This approach is similar to that of Soo (1967) for multiphase flow and of Jackson (1985) for fluidization. The equations are similar to Bowen's (1976) balance laws for multicomponent mixtures. The principle difference is the appearance of the volume fraction of phase "k" denoted by ϵ_k . In the case of phases, not all the space is occupied at the same time by all the phases, as it is by components. As in the case of the mixture equations for components, the mixture equations for phases show that the sum of the drag forces is zero. The fluid pressure p_l is in the liquid (continuous) phase.

The corresponding gas and solids (dispersed particulate phases) momentum equations with zero particulate viscosities are the trajectory equations for particulate flow. For gas-solid fluidized beds, Bouillard, et al. (1989) have shown that this set of equations produces essentially the same numerical answers for fluidization as did the earlier conditionally stable model which has the fluid pressure in both the gas and the solids

phases. In this model (hydrodynamic model B) the drag and the stress relations were altered to satisfy Archimedes' buoyancy principle and Darcy's law, as illustrated by Jayaswal (1991). The gas-particulate drag coefficients given in Table D.1 are for the model B reviewed by Jayaswal (1991). For the solid phase, p_s consists of the static normal stress and the dynamic stress, called solids pressure, which arises due to the collision of the particles. The added mass forces, Saffman's lift forces (Birkoff, 1960; Saffman, 1965; Drew, et al., 1979) were assumed to be negligible.

This model is unconditionally well-posed, i.e., the characteristics are real and distinct for one dimensional transient flow. It does not require the presence of solids pressure for stability and well posedness.

The details of hydrodynamic equations used in the computer simulation are given in Table D.1 of the Appendix D.

7.3 Numerical Scheme

The governing equations along with the constitutive equations are solved for p_k , ϵ_k , u_k and v_k , ($k = g, l, s$) using the Implicit Continuum Eulerian (ICE) method (Rivard and Torrey, 1977; Jayaswal, 1991) with appropriate initial and boundary conditions. The pressure drop was assumed to be in liquid (continuous) phase. Stewart and Wendroff (1984) have critically reviewed the ICE algorithm and related staggered mesh conservative schemes. The computations are carried out using a mesh of finite-difference cells fixed in two-dimensional space (Eulerian mesh). The scalar variables are located at the cell center and the vector variables at the cell boundaries. The momentum equation

is solved using staggered mesh whereas for the continuity equation the donor cell method was used.

The partial differential equations are well-posed. The time step is chosen to satisfy the Courant stability criterion (Courant, et al., 1952). The numerical stability of the equations can be obtained using the von Neumann stability analysis, as illustrated by Lyczkowski, et al. (1978) and Prosperetti (1982).

7.4. Initial and Boundary Conditions

As an initial condition, the liquid-solid bed was assumed to be at a minimum fluidization state. The gas and liquid flows at the bottom of the bed were started at specified rates at time $t = 0$. No-slip condition was satisfied at the wall.

The computer simulations of the three-phase fluidized bed were carried out in two different modes described later : symmetric and asymmetric. A complete inlet and initial conditions for both the modes are shown in Figure 7.1. In both the modes, simulations were carried out for 20 seconds with a time increment of 5×10^{-5} seconds. The computations were conducted on our HP-700 workstation.

7.5 Fluidized Bed Simulation

7.5.1 Symmetric Fluidized Bed Mode. In symmetric mode, the fluidized bed behavior was assumed to be symmetric across the centerline parallel to the axial direction. In this mode only half of the bed was simulated. The computations were carried out using a mesh of 1040 finite difference computational cells with 0.635 cm x 2 cm cell

Outlet Conditions:

$$\begin{aligned} V_z = V_s &= 0 \text{ cm/s} \\ V_1 &= 0.7 \text{ cm/s} \\ p &= 0.1013 \text{E}07 \text{ dynes/cm}^2 \end{aligned}$$

Initial Conditions:

$$\begin{aligned} V_g &= 0 \text{ cm/s} \\ V_1 &= 1.66 \text{ cm/s} \\ \epsilon_g &= 0 \\ \epsilon_1 &= 0.42 \\ \epsilon_s &= 0.58 \end{aligned}$$

Inlet Conditions:

$$\begin{aligned} \epsilon_g &= 0.1 \\ V_g &= 33.64 \text{ cm/s} \\ \epsilon_1 &= 0.9 \\ V_1 &= 2.3 \text{ cm/s} \\ p &= 0.1113 \text{E}07 \text{ dynes/cm}^2 \end{aligned}$$

15cm

Symmetric

Outlet Conditions:

$$\begin{aligned} V_z = V_s &= 0 \text{ cm/s} \\ V_1 &= 5.4 \text{ cm/s} \\ p &= 0.1013 \text{E}07 \text{ dynes/cm}^2 \end{aligned}$$

Initial Conditions:

$$\begin{aligned} V_g &= 0 \text{ cm/s} \\ V_1 &= 12.86 \text{ cm/s} \\ \epsilon_g &= 0 \\ \epsilon_1 &= 0.42 \\ \epsilon_s &= 0.58 \end{aligned}$$

Inlet Conditions:

$$\begin{aligned} \epsilon_g &= 0.4 & \epsilon_g &= 0.5 \\ V_g &= 7.4 \text{ cm/s} & V_g &= 7.4 \text{ cm/s} \\ \epsilon_1 &= 0.6 & \epsilon_1 &= 0.5 \\ V_1 &= 3.7 \text{ cm/s} & V_1 &= 3.7 \text{ cm/s} \\ p &= 0.1131 \text{E}07 \text{ dynes/cm}^2 \end{aligned}$$

30cm

Asymmetric

Figure 7.1. System for Simulation of Three Phase Fluidized Bed with Dimensions and Flow Conditions

sizes.

7.5.2 Asymmetric Fluidized Bed Mode. In asymmetric mode, it was assumed that the gas at the inlet entering the two sides of the bed had different volume fractions. This assumption is also supported by the visual observation of the experiment. In this mode, the complete bed was simulated. A total of 1504 computational cells with 1.016 cm x 2 cm cell sizes were used. The simulation in this mode was done in two different types of fluidized beds :

a) One-outlet bed : bed with completely open outlet at the top.

b) Multi-outlet bed: In this setup, the geometry was modified to include the multiple outlets of the bed in order to closely match the experimental set-up. The fluidized bed had three 1.0 inch x 1.0 inch (2.54 cm x 2.54 cm) rectangular equispaced openings at the top of the bed.

7.6 Results of Numerical Simulation

Numerical computations, using the hydrodynamic model presented in Table D.1, were carried out for the three-phase fluidized bed. Typical results of the simulation are presented below.

7.6.1 Volume Fraction Profiles. The computed time-averaged volume fraction contours for gas, liquid and solids in the two dimensional fluidized bed are shown in Figures 7.2(a) to 7.2(c), respectively.

The computed volume fractions of gas and liquid near the wall compare well with measurements; which were found to be approximately 0.3 for gas and 0.5 for liquid.

Figure 7.2(a). Computed Time Average Volume Fraction Contours for Liquid Phase

Figure 7.2(b). Computed Time Average Volume Fraction Contours for Gas Phase

Figure 7.2(d). Photograph of Bubble Coalescence in a Liquid-Gas-Solid Fluidized Bed ($U_g=3.37$ cm/sec, $U_l=2.03$ cm/sec)

Figure 7.2(c). Computed Time Average Volume Fraction Contours for Solid Phase

Computed time averaged solid volume fraction contours for the whole bed with three outflows are shown in Figures 7.2(c), as in the photograph of the experiment, Figure 7.2(d) shows that the solid is concentrated in the lower 25 cm of the bed. There is a murky interface between the solid and the liquid-gas in the upper portion of the bed. This is in part reality and in part caused by perhaps inadequate time averaging. We time averaged over about 20 seconds. Higher time averaging can not be used due to a small loss of particles observed both in the experiment and more so in the simulation. The simulation has larger particles loss due to the neglect of the walls in the two dimensional computations. To correct this deficiency we are in the process of developing a three dimensional code.

Figure 7.2(b) shows that there is more gas in the lower portion of the bed than in the region where there is only gas and liquid. Figures 7.3 to 7.5 demonstrate a comparison between the computed and measured time-average volume fractions of three phases at different widths of the bed. In most cases, near the wall and the center, there is good agreement between the computational and experimental results.

Figures 7.6 and 7.7 show computations for instantaneous gas volume fractions in the form of bubbles for symmetric and asymmetric cases, respectively. In both cases, symmetric and asymmetric, we see a tendency for gas bubbles to move midway between the center of the bed and the walls. In the symmetric case of the three-phase fluidized bed, no bubble coalescence was observed. The reason for no bubble coalescence is the downward flow in the center of the bed causing a deficiency of gas in that region. In asymmetric case, however, shows bubble coalescence.

Figure 7.3. Comparison of Computational and Experimental Time Average Volume Fractions at $X=2.0$ cm

Figure 7.4. Comparison of Computational and Experimental Time Average Volume Fractions at $X=7.0$ cm

Figure 7.5. Comparison of Computational and Experimental Time Average Volume Fractions at $X=11.0$ cm

Gas

Figure 7.6. Bubble Formation in Three Phase Fluidized Bed at Time = 1.5 sec, for Symmetric Bed

Figure 7.7. Bubble Formation in Three Phase Fluidized Bed at Time = 2 Sec, for Asymmetric Bed

The asymmetric mode, the computed gas bubbles move in snake-like fashion as observed in the experiment. This mode is setup by vorticities generated by gas flow circulation which produces an asymmetric gas volume fraction gradient inside the fluidized bed. The computed viscosity is very high in the region of the large bubbles.

These results clearly show the need for computer simulation to be in asymmetric mode to get experimentally observed bed behavior.

7.6.2 Velocity Profiles. Figures 7.8 and 7.9 show the velocity patterns in asymmetric and symmetric cases, respectively. From these figures it is clear that in the symmetric cases, all three phases have downward movement at the center of the bed and at the wall. However, in the asymmetric mode all three phases have downward movement near the wall and upward movement at the center, the same flow pattern reported by Chen, Reese, and Fan (1994). The experimentally observed flow circulation is the same as that obtained in the asymmetric case.

The downflow seen in asymmetric mode produces a rotation or particle vorticity, given by the inviscid equation below,

$$\frac{\partial \zeta}{\partial t} + u \frac{\partial \zeta}{\partial x} + v \frac{\partial \zeta}{\partial y} = - \frac{g}{(1 - \epsilon_{mf})} \frac{\partial \epsilon}{\partial y} \quad (7.1)$$

The equation shows that the vorticity produces a change in gas volume fraction ϵ . Then although we supply gas at an uniform rate, the gas is no longer uniformly distributed. Furthermore, Bouillard and Gidaspow (1991) have shown that when the gas at the bottom moves faster than in the rest of the bed, there is a catch-up effect and large gas voids may form, as depicted by Figure 7.2(d). It is the packing effect of the bubbles

Figure 7.8. Computed Velocity Pattern in Three Phase Fluidized Bed at Time = 2 sec for Asymmetric Bed

Figure 7.9. Computed Velocity Pattern in Three Phase Fluidized Bed at Time = 2 sec for Symmetric Bed

that produces this regime. When the injected bubbles are far apart, for low gas velocities, the bubbles move up at constant velocities. In this regime the drag is independent of the gas volume fraction. It is given by the corrected Stoke's law. At high gas velocities, the gas volume fraction is high and we use the Ergun equation to compute the drag in our model. Here the paths of the small injected bubbles intersect and bubbles coalesce. The situation is similar to that described in Chapter 6 for gas-solid fluidized beds by Gidaspow (1994).

Figures 7.10 and 7.11 show comparisons of time-averaged axial solids velocity profiles at heights of 7.0 cm and 13.0 cm from the bottom of the bed. These figures show excellent similarity. In Figure 7.10, there is slight disagreement between experimental and simulation velocity profiles. This disagreement mainly resulted from the positions of two liquid inlets in the bottom of the bed. In spite of a large pressure drop (about 10-20% of bed pressure drop through the liquid distributor) residual inlet effects could be seen on high experimental solids axial velocities. The residual velocity effect, however, was absent in velocity measurements conducted at a height of 13.0 cm, as seen in Figure 7.11.

7.7 Conclusion

1. Unlike gas-solid bubbling fluidized beds, the three-phase fluidized bed primarily behaves in an asymmetric fashion. The three-phase fluidized bed simulations in asymmetric mode showed promising results. Good agreement was obtained between measured and simulated solids volume fractions.

2. The asymmetric simulation also captured some characteristics of bubble coalescence and break-up. A coalesced bubbling regime was used in the experiments.
3. Excellent comparison was seen between measured and simulated time averaged solids axial velocity profiles at two different locations in the bed.
4. The multi-phase hydrodynamic model was validated for use in three-phase fluidized bed simulation.

Figure 7.10. Comparison of Experimental and Computational Results for Solid Velocity Profile at $Y=7$ cm

Figure 7.11. Comparison of Experimental and Computational Results for Solid Velocity Profile at $Y=13$ cm

CHAPTER VIII

SUMMARY AND RECOMMENDATIONS

8.1 Summary

From a high speed video of the experiments, we have observed that our fluidized bed was operating mostly in the coalesced bubble regime. Figure 8.1 shows an image of a bubble that was captured by the High Resolution Micro-Imaging/ Measuring system. The coalesced bubbles diameters were about 2 to 4 cm. This compares well with the bubble diameter (D_b) of 4.33 cm calculated using the correlation (Kim, et.al., 1977),

$$D_b = 0.142 u_I^{0.052} u_g^{0.248} \mu_I^{0.008} \sigma^{0.034} \quad (8.1)$$

in SI units. ($3 < D_b < 8$ cm).

Figure 8.2 shows that bubbles occurred mostly between 10 to 17 cm height above the distributors.

In this thesis, the three-phase concentration, the bed behavior, the bubbles frequency, the bubbles size, the bed viscosity, and the solids velocity were investigated. Also, the solids granular temperature was defined from two sources: measurement of the solids velocity, and measurement of the solids viscosity. There was fair agreement between these two methods. However, at lower solids concentration, the data do not match and this might be due to the fact that local solids concentration may differ from overall solids concentration in the bed.

A preliminary simulation of three-phase fluidized bed in asymmetric mode qualitatively predicts the bed behavior, such as bubble coalescence and flow circulation.

The computations in fluidized bed in symmetric mode incorrectly showed no bubble coalescence because of a downward velocity profile in the bed center which caused a gas deficiency in that region. The asymmetric behavior of three-phase fluidized beds differentiates its simulation from that of gas-solid fluidized beds where a conventional symmetric bed simulation was adequate to predict bed behavior. The asymmetric mode is set up by vorticities generated by gas flow circulation which produce a change in gas volume fraction inside the bed. Good agreement was obtained between measured and computed solids viscosities.

8.2 Future Work

1. The time-averaged volume fractions and porosity fluctuations measurements in a new three-phase fluidized bed with a central jet are needed in order to compare the bed behavior with that of three-phase fluidized beds with a uniform distributor.
2. The bubble frequency in gas-liquid two phase fluidization can be measured with a more sensitive detector than currently is available.
3. To completely analyze the velocity distribution in all three axes the measurements should be obtained by two CCD cameras with a smaller field of view.
4. In order to investigate the effect of solid concentration on the bed viscosity, the viscosity can be measured using a Brookfield viscometer and concentration can be measured using a CCD camera. Hence the data can be plotted for viscosity as a function of solid concentration.
5. The shear rate can be directly measured while the Brookfield viscometer is

conducting the viscosities value by using a High Resolution Micro-Imaging/ Measuring system technique.

6. The effect of particle sizes on the bed viscosity should be investigated. The particles of different size (for example, 0.08 mm) should be placed in the fluidized bed with a central jet so that measurements can be obtained.

7. Two different computer simulations using the predictive hydrodynamic model should be performed in three-phase fluidization in asymmetric mode: one using a measured bed viscosity and one using a predicted granular viscosity from the kinetic theory of granular solids. The computed results should be compared with the experimental results.

8. The computer simulation should be modified from two dimensions to three dimensions in order to resolve the loss of solid over a short period of time. This modification will also provide the effect of the wall in the third dimension.

Figure 8.1. Bubble Frequency Vs. Height above the Distributer at X=12.7 cm

Figure 8.2. The Image of a Bubble That Was Captured by A Recording System

APPENDIX A
VOLUME FRACTION MEASUREMENTS DATA

Table A-1. Experimental Data for Volume Fractions in Air-Water-Glass Beads
Fluidized Bed (Superficial Gas Velocity = 3.36 cm/s, Superficial Liquid Velocity =
2.0 cm/s)

X(cm)	Y(cm)	L_r	I_x	ϵ_g	ϵ_l	ϵ_s
2.54	3.81	9.77662	6.44495	0.346492	0.370805	0.178013
2.54	5.08	9.74908	6.41129	0.320369	0.365295	0.217466
2.54	6.35	9.77159	6.40455	0.341724	0.364188	0.178173
2.54	7.62	9.77094	6.41232	0.341111	0.365464	0.180862
2.54	8.89	9.807	6.40496	0.375225	0.364255	0.118793
2.54	10.16	9.82378	6.41582	0.391067	0.366039	0.092914
2.54	11.43	9.81581	6.41253	0.383546	0.365498	0.105584
2.54	12.70	9.7923	6.42247	0.361333	0.367128	0.147056
2.54	13.97	9.82606	6.39146	0.393209	0.362034	0.084087
2.54	15.24	9.86672	6.38787	0.431466	0.361444	0.015443
2.54	16.51	9.83044	6.415	0.39734	0.365903	0.081609
2.54	17.78	9.75176	6.44711	0.322908	0.371158	0.220315
2.54	19.05	9.7507	6.46285	0.3219	0.373724	0.225324
2.54	20.32	9.72932	6.55134	0.301575	0.388035	0.279354
2.54	21.59	9.72774	6.59797	0.300068	0.395499	0.291395
2.54	22.86	9.89368	6.52912	0.456733	0.38446	-0.000528
2.54	24.13	9.73778	6.52864	0.309625	0.384382	0.260484

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ε_g	ε_1	ε_s
2.54	25.40	9.83715	6.52961	0.403661	0.384539	0.093771
3.81	2.54	9.83326	6.55839	0.4	0.389167	0.106076
3.81	3.81	9.81312	6.50055	0.381009	0.379845	0.128088
3.81	5.08	9.78807	6.54584	0.357335	0.387152	0.179276
3.81	6.35	9.79503	6.53498	0.363918	0.385404	0.165398
3.81	7.62	9.80294	6.52912	0.371392	0.38446	0.150949
3.81	8.89	9.76941	6.56068	0.339657	0.389534	0.213643
3.81	10.16	9.73234	6.52573	0.30445	0.383913	0.269081
3.81	11.43	9.74289	6.53204	0.314479	0.384931	0.252557
3.81	12.70	9.74176	6.54833	0.313412	0.387552	0.257739
3.81	13.97	9.74749	6.54286	0.318853	0.386673	0.246979
3.81	15.24	9.75777	6.69599	0.328617	0.411018	0.260193
3.81	16.51	9.76195	6.70491	0.332583	0.412419	0.254911
3.81	17.78	9.73923	6.71659	0.311	0.41425	0.295517
3.81	19.05	9.74199	6.71145	0.313632	0.413446	0.289837
3.81	20.32	9.76498	6.48418	0.335454	0.377191	0.205615
3.81	21.59	9.78085	6.42184	0.350496	0.367026	0.166164
3.81	22.86	9.78899	6.43574	0.358204	0.369301	0.155336

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_l	ϵ_s
3.81	24.13	9.76395	6.39432	0.334485	0.362506	0.188911
3.81	25.40	9.78947	6.42603	0.358662	0.367711	0.152528
5.08	2.54	9.82408	6.49036	0.39135	0.378194	0.107661
5.08	3.81	9.78615	6.48326	0.355518	0.377043	0.169819
5.08	5.08	9.88878	6.56757	0.452145	0.390639	0.015367
5.08	6.35	9.84607	6.59797	0.412056	0.395499	0.092621
5.08	7.62	9.83393	6.62944	0.400633	0.400506	0.119182
5.08	8.89	9.83042	6.69566	0.397326	0.410966	0.138173
5.08	10.16	9.79052	6.70186	0.359653	0.41194	0.206262
5.08	11.43	9.75209	6.6799	0.323226	0.408486	0.266586
5.08	12.70	9.76677	6.69291	0.337154	0.410534	0.244433
5.08	13.97	9.79049	6.71387	0.359623	0.413825	0.208679
5.08	15.24	9.76115	6.71928	0.331829	0.414673	0.259077
5.08	16.51	9.76469	6.70732	0.335185	0.412798	0.250769
5.08	17.78	9.75233	6.71093	0.323454	0.413363	0.272299
5.08	19.05	9.76383	6.70679	0.334367	0.412714	0.252115
5.08	20.32	9.72829	6.70519	0.300588	0.412463	0.311755
5.08	21.59	9.73474	6.67933	0.306732	0.408396	0.295748

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ε_g	ε_l	ε_s
5.08	22.86	9.73819	6.6985	0.31001	0.411412	0.293714
5.08	24.13	9.74065	7.22407	0.312352	0.490902	0.389289
5.08	25.40	9.73947	6.72557	0.311231	0.415657	0.296872
6.35	2.54	9.75012	6.70088	0.321351	0.411787	0.274054
6.35	3.81	9.7404	6.70754	0.312115	0.412831	0.291758
6.35	5.08	9.74074	6.70175	0.312443	0.411923	0.290036
6.35	6.35	9.72911	6.70913	0.301369	0.413081	0.311145
6.35	7.62	9.73278	6.72929	0.304868	0.416238	0.308897
6.35	8.89	9.74714	6.71731	0.318522	0.414364	0.28231
6.35	10.16	9.76135	6.71928	0.332017	0.414673	0.258743
6.35	11.43	9.75394	6.71846	0.32498	0.414543	0.271071
6.35	12.70	9.76207	6.7325	0.332696	0.416741	0.260133
6.35	13.97	9.78252	6.7237	0.352079	0.415365	0.224002
6.35	15.24	9.74374	6.72227	0.315287	0.41514	0.289025
6.35	16.51	9.73947	6.69872	0.311233	0.411446	0.291586
6.35	17.78	9.74029	6.68172	0.31201	0.408773	0.286852
6.35	19.05	9.74976	6.67056	0.321014	0.407014	0.268664
6.35	20.32	9.73828	6.66433	0.3101	0.40603	0.286802

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_l	ϵ_s
6.35	21.59	9.73924	6.67485	0.311013	0.407691	0.287265
6.35	22.86	9.75024	6.66551	0.321468	0.406217	0.266859
6.35	24.13	9.77028	6.70283	0.340484	0.412092	0.240477
6.35	25.40	9.78144	6.71638	0.351055	0.414217	0.22438
7.62	2.54	9.77106	6.70945	0.341222	0.413131	0.240471
7.62	3.81	9.76297	6.72278	0.333549	0.41522	0.256712
7.62	5.08	9.75953	6.72237	0.33029	0.415156	0.262415
7.62	6.35	9.76395	6.68825	0.33448	0.4098	0.248259
7.62	7.62	9.75794	6.71513	0.32878	0.414021	0.263671
7.62	8.89	9.74024	6.71167	0.311958	0.413479	0.292848
7.62	10.16	9.74695	6.70326	0.318343	0.41216	0.279862
7.62	11.43	9.77345	6.66845	0.343489	0.406682	0.228355
7.62	12.70	9.76361	6.69991	0.334156	0.411634	0.251135
7.62	13.97	9.75962	6.69588	0.330378	0.411	0.257044
7.62	15.24	9.77371	6.68601	0.343737	0.409448	0.231386
7.62	16.51	9.77674	6.6961	0.346607	0.411035	0.228282
7.62	17.78	9.74407	6.72134	0.315603	0.414995	0.288283
7.62	19.05	9.78318	6.71248	0.352702	0.413607	0.220691

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_z	ϵ_1	ϵ_s
7.62	20.32	9.79477	6.71814	0.363671	0.414494	0.202336
7.62	21.59	9.78926	6.71387	0.35846	0.413825	0.210744
7.62	22.86	9.78633	6.70828	0.355685	0.412948	0.21457
7.62	24.13	9.77924	6.70272	0.348971	0.412075	0.225392
7.62	25.40	9.80177	6.69991	0.370293	0.411634	0.186992
8.89	2.54	9.76474	6.68217	0.335233	0.408844	0.245722
8.89	3.81	9.74171	6.6648	0.313358	0.406105	0.281113
8.89	5.08	9.74193	6.65764	0.313566	0.404973	0.279324
8.89	6.35	9.75464	6.64998	0.325644	0.403761	0.256365
8.89	7.62	9.76298	6.64665	0.333564	0.403234	0.241646
8.89	8.89	9.76084	6.66881	0.331535	0.406737	0.249642
8.89	10.16	9.75245	6.69467	0.323565	0.410811	0.2689
8.89	11.43	9.73627	6.70002	0.308186	0.411651	0.297251
8.89	12.70	9.77431	6.70143	0.344305	0.411872	0.233419
8.89	13.97	9.7704	6.69169	0.340601	0.410343	0.238074
8.89	15.24	9.75698	6.69014	0.327867	0.410098	0.26037
8.89	16.51	9.74594	6.68466	0.317379	0.409236	0.277904
8.89	17.78	9.72915	6.66845	0.301414	0.406682	0.303036

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_1	ϵ_s
8.89	19.05	9.72793	6.68319	0.300253	0.409005	0.308012
8.89	20.32	9.7376	6.66869	0.30945	0.406719	0.28882
8.89	21.59	9.72856	6.66076	0.300852	0.405466	0.302509
8.89	22.86	9.732	6.67807	0.30412	0.408198	0.300136
8.89	24.13	9.7653	7.06267	0.33576	0.467124	0.317908
8.89	25.40	9.84089	6.67079	0.407179	0.407051	0.115771
10.16	2.54	9.8358	6.68769	0.40239	0.409713	0.127611
10.16	3.81	9.82697	6.67853	0.394068	0.40827	0.140574
10.16	5.08	9.84005	6.68533	0.406388	0.409342	0.120051
10.16	6.35	9.82496	6.67462	0.392178	0.407655	0.143156
10.16	7.62	9.82696	6.68001	0.394062	0.408504	0.140877
10.16	8.89	9.82801	6.68936	0.395048	0.409976	0.140974
10.16	10.16	9.82768	6.69847	0.394739	0.411408	0.143319
10.16	11.43	9.82905	6.69258	0.396029	0.410482	0.139866
10.16	12.70	9.84746	6.7104	0.413367	0.41328	0.112604
10.16	13.97	9.81638	6.69434	0.384083	0.410759	0.161418
10.16	15.24	9.74574	6.67773	0.317191	0.408144	0.276869
10.16	16.51	9.76521	6.65884	0.33568	0.405164	0.240311

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_1	ϵ_s
10.16	17.78	9.77576	6.65456	0.34568	0.404486	0.221711
10.16	19.05	9.78256	6.64216	0.35212	0.402524	0.207819
10.16	20.32	9.77004	6.66693	0.34026	0.406441	0.233784
10.16	21.59	9.75072	6.67428	0.321922	0.4076	0.267787
10.16	22.86	9.80593	6.68013	0.374219	0.408522	0.17612
10.16	24.13	9.80593	6.70828	0.374219	0.412948	0.181672
10.16	25.40	9.80881	6.7038	0.376944	0.412244	0.175954
11.43	2.54	9.80842	6.68791	0.376575	0.409748	0.173477
11.43	3.81	9.79315	6.70881	0.362139	0.413031	0.203218
11.43	5.08	9.78909	6.69752	0.358299	0.411258	0.20781
11.43	6.35	9.79577	6.65968	0.364618	0.405296	0.189114
11.43	7.62	9.82227	6.67956	0.389639	0.408432	0.148638
11.43	8.89	9.8397	6.66112	0.406059	0.405523	0.115842
11.43	10.16	9.83804	6.65607	0.404498	0.404725	0.117613
11.43	11.43	9.83957	6.66528	0.405941	0.40618	0.116877
11.43	12.70	9.90087	6.71072	0.463462	0.41333	0.023749
11.43	13.97	9.91592	6.70033	0.477529	0.4117	-0.003262
11.43	15.24	9.92008	6.67532	0.481412	0.407764	-0.015095

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_l	ϵ_s
11.43	16.51	9.78637	6.69708	0.355728	0.411189	0.212288
11.43	17.78	9.77687	6.66386	0.346731	0.405956	0.221689
11.43	19.05	9.77947	6.6752	0.349196	0.407746	0.219561
11.43	20.32	9.77638	6.67323	0.346262	0.407436	0.224379
11.43	21.59	9.77048	6.67933	0.340675	0.408396	0.235502
11.43	22.86	9.79093	6.68758	0.360043	0.409695	0.202753
11.43	24.13	9.79273	6.68376	0.361747	0.409094	0.198975
11.43	25.40	9.77173	6.67126	0.34186	0.407124	0.231802
12.70	2.54	9.77109	6.66563	0.341253	0.406236	0.231764
12.70	3.81	9.77922	6.6681	0.348953	0.406626	0.218588
12.70	5.08	9.77584	6.65824	0.345754	0.405069	0.222311
12.70	6.35	9.7757	5.16944	0.345625	0.513914	-0.111626
12.70	7.62	9.77456	5.20093	0.344545	0.539745	-0.101692
12.70	8.89	9.73418	5.22002	0.306196	0.555329	-0.028786
12.70	10.16	9.7652	5.20841	0.335667	0.545857	-0.084036
12.70	11.43	9.80463	5.23563	0.372994	0.56803	-0.143407
12.70	12.70	9.79513	5.22051	0.364012	0.55573	-0.131283
12.70	13.97	9.76991	5.24455	0.340135	0.575271	-0.082837

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_l	ϵ_s
12.70	15.24	9.75654	5.23418	0.327454	0.566849	-0.062942
12.70	16.51	9.76833	5.22956	0.338636	0.563096	-0.083955
12.70	17.78	9.75544	5.26829	0.326409	0.594484	-0.052510
12.70	19.05	9.7673	5.24815	0.337658	0.578186	-0.077534
12.70	20.32	9.76007	5.24551	0.330802	0.57605	-0.066028
12.70	21.59	9.75738	6.70966	0.328252	0.413164	0.263534
12.70	22.86	9.75865	6.71669	0.329456	0.414266	0.262779
12.70	24.13	9.75795	5.24958	0.328785	0.579349	-0.061425
12.70	25.40	9.77369	5.22712	0.343717	0.561112	-0.093588
13.97	2.54	9.79675	5.26359	0.365547	0.590683	-0.123157
13.97	3.81	9.8002	5.28645	0.368804	0.609121	-0.123216
13.97	5.08	9.79733	5.24958	0.366088	0.579349	-0.127637
13.97	6.35	9.7987	5.30884	0.367386	0.627097	-0.115119
13.97	7.62	9.7955	5.24479	0.364366	0.575466	-0.125785
13.97	8.89	9.81248	5.22932	0.380401	0.562897	-0.158149
13.97	10.16	9.81915	5.25316	0.386695	0.582247	-0.163312
13.97	11.43	9.82414	5.2534	0.391403	0.58244	-0.171609
13.97	12.70	9.83746	5.24383	0.403949	0.574687	-0.196285

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_1	ϵ_s
13.97	13.97	9.77833	5.18047	0.348111	0.522977	-0.113226
13.97	15.24	9.77991	5.19138	0.349607	0.531927	-0.113103
13.97	16.51	9.79987	5.1494	0.368493	0.497393	-0.157345
13.97	17.78	9.82591	5.37015	0.393069	0.675938	-0.145544
13.97	19.05	9.84348	5.2993	0.409619	0.619448	-0.192454
13.97	20.32	9.84717	5.30136	0.413091	0.621094	-0.198107
13.97	21.59	9.78357	5.34223	0.353073	0.653768	-0.081435
13.97	22.86	9.81151	5.34289	0.379486	0.654291	-0.128155
13.97	24.13	9.80105	5.30997	0.369605	0.628002	-0.118776
13.97	25.40	9.78546	5.31313	0.354861	0.630531	-0.091821
15.24	2.54	9.76985	5.37143	0.340074	0.676952	-0.051166
15.24	3.81	9.75802	5.29999	0.328857	0.619997	-0.048936
15.24	5.08	9.76016	5.30635	0.330881	0.625102	-0.050944
15.24	6.35	9.7661	5.33232	0.33652	0.64587	-0.054505
15.24	7.62	9.75522	5.29839	0.326199	0.618715	-0.044615
15.24	8.89	9.74521	5.30567	0.316688	0.624557	-0.025921
15.24	10.16	9.74839	5.2802	0.319712	0.604088	-0.037641
15.24	11.43	9.76812	5.27181	0.338432	0.597322	-0.072969

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_s	ϵ_1	ϵ_s
15.24	12.70	9.75572	5.18378	0.326676	0.525695	-0.074335
15.24	13.97	9.76474	5.22615	0.335235	0.560318	-0.078781
15.24	15.24	9.75949	5.22883	0.330247	0.562501	-0.069249
15.24	16.51	9.76524	5.20592	0.335704	0.543826	-0.084732
15.24	17.78	9.74792	5.22737	0.319264	0.561311	-0.050125
15.24	19.05	9.75297	5.26099	0.324062	0.588583	-0.050176
15.24	20.32	9.77349	5.24767	0.343531	0.577798	-0.088080
15.24	21.59	9.79729	5.20717	0.366055	0.544842	-0.138289
15.24	22.86	9.77672	5.19567	0.346593	0.535435	-0.106664
15.24	24.13	9.77255	5.19037	0.342634	0.5311	-0.100984
15.24	25.40	9.77458	5.20318	0.344563	0.541584	-0.101153
16.51	0.00	9.728	5.14334	0.300313	0.492383	-0.037883
16.51	0.00	9.72819	5.17587	0.300497	0.519197	-0.029887
16.51	0.00	9.76993	5.35422	0.340156	0.663304	-0.055547
16.51	0.00	9.81815	5.29359	0.38575	0.614856	-0.151514
16.51	0.00	9.82459	5.33674	0.391825	0.649391	-0.151577
16.51	0.00	9.85455	5.35466	0.420031	0.663649	-0.197215
16.51	0.00	9.89614	5.31245	0.459041	0.62999	-0.276904

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_l	ϵ_s
16.51	0.00	9.90654	5.32322	0.468764	0.638598	-0.29149
16.51	0.00	9.9178	5.31785	0.479288	0.634307	-0.311502
16.51	0.00	9.92766	5.28368	0.488496	0.606888	-0.336388
16.51	2.54	9.9338	5.32789	0.494217	0.642332	-0.335541
16.51	3.81	9.93647	5.27041	0.49671	0.596188	-0.354291
16.51	5.08	9.94155	5.29382	0.501444	0.61504	-0.356841
16.51	6.35	9.94043	5.25625	0.500402	0.58475	-0.364395
16.51	7.62	9.92707	5.21287	0.487943	0.5495	-0.35322
16.51	8.89	9.88007	5.27087	0.443982	0.596566	-0.260574
16.51	10.16	9.87475	5.48756	0.438995	0.767932	-0.198524
16.51	11.43	9.97287	5.4036	0.530576	0.702353	-0.38145
16.51	12.70	9.78385	5.44628	0.353345	0.735813	-0.056455
17.78	13.97	9.77468	5.54246	0.344654	0.810274	-0.017912
17.78	15.24	9.7658	5.48967	0.336241	0.769567	-0.015615
17.78	16.51	9.76499	5.52141	0.33547	0.794095	-0.006632
17.78	17.78	9.77435	5.57009	0.344341	0.831429	-0.010789
17.78	19.05	9.77363	5.52216	0.343658	0.794669	-0.020989
17.78	20.32	9.78088	5.54501	0.350528	0.812235	-0.027730

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_l	ϵ_s
17.78	21.59	9.78058	5.50433	0.350245	0.780913	-0.036952
17.78	22.86	9.78997	5.50792	0.359131	0.783686	-0.051865
17.78	24.13	9.78904	5.51712	0.358256	0.790786	-0.048107
17.78	25.40	9.7882	5.43264	0.357456	0.725148	-0.067064
17.78	15.24	9.78514	5.4841	0.35456	0.765247	-0.049475
17.78	16.51	9.77552	5.71294	0.345448	0.939144	0.020683
17.78	17.78	9.77493	5.7153	0.344891	0.9409	0.022218
17.78	19.05	9.78199	5.79527	0.351581	1	0.028689
17.78	20.32	9.89325	5.79527	0.45633	1	-0.157253
17.78	21.59	9.87146	5.79527	0.435906	1	-0.120999
17.78	22.86	9.87622	5.79527	0.440374	1	-0.128929
17.78	24.13	9.88215	5.79527	0.445931	1	-0.138794
17.78	25.40	9.88704	5.79527	0.450514	1	-0.146929
19.05	2.54	9.89131	5.73957	0.454515	0.958919	-0.166785
19.05	3.81	9.89099	5.65	0.454216	0.892017	-0.187023
19.05	5.08	9.89393	5.62137	0.456974	0.870408	-0.198626
19.05	6.35	9.89643	5.5934	0.459305	0.849192	-0.209351
19.05	7.62	9.91428	5.51675	0.475998	0.790497	-0.257204

Table A-1 (Continue)

X(cm)	Y(cm)	I_y	I_x	ϵ_g	ϵ_l	ϵ_s
19.05	8.89	9.91412	5.52588	0.475854	0.797531	-0.254764
19.05	10.16	9.91766	5.74064	0.479157	0.959717	-0.210278
19.05	11.43	9.92813	5.79527	0.488933	1	-0.215127
19.05	12.70	9.94307	5.79527	0.502861	1	-0.239851
19.05	13.97	9.94139	5.79527	0.501296	1	-0.237072
19.05	15.24	9.93959	5.79527	0.499613	1	-0.234085
19.05	16.51	9.95065	5.79527	0.50992	1	-0.252382
19.05	17.78	9.94547	5.79527	0.505089	1	-0.243807
19.05	19.05	9.95927	5.79527	0.517935	1	-0.266609
19.05	20.32	9.9578	5.79527	0.516571	1	-0.264187
19.05	21.59	9.96653	5.79527	0.524682	1	-0.278585
19.05	22.86	9.96123	5.79527	0.519754	1	-0.161709
19.05	24.13	9.89787	5.79527	0.460653	1	-0.175430
19.05	25.40	9.89887	5.77354	0.461593	0.984026	-0.175308

APPENDIX B
VELOCITY MEASUREMENT DATA

Table B-1. Raw Solids Velocity Data for Liquid-Solid Fluidized Bed
with Uniform Distributor at X= 4.0 cm and Y= 4.5 cm
(Superficial Liquid Velocity = 4.0 cm/sec)

$V_{s,y}$ (cm/sec)	$V_{s,x}$ (cm/sec)
25.46	8.81
30.18	11.68
32.99	14.55
24.33	16.54
33.94	14.54
46.23	21.19
60.42	29.73
24.62	0.98
24.53	7.85
34.80	20.30
44.42	13.52
31.27	11.96
16.70	40.33
27.24	3.89
12.46	38.71
-6.89	22.64

Table B-1 (Continue)

$V_{s,y}$ (cm/sec)	$V_{s,x}$ (cm/sec)
33.00	13.59
33.00	13.59
30.25	6.83
30.25	7.80
30.25	6.83
2.81	42.14
15.70	13.73
19.60	12.75
21.38	18.46
25.09	27.99
29.99	21.28
24.42	12.70
28.11	19.38
40.60	15.46
29.83	32.71
23.26	20.35
28.21	14.59

Table B-1 (Continue)

$V_{s,y}$ (cm/sec)	$V_{s,x}$ (cm/sec)
19.44	18.47
27.19	17.47
33.91	16.47
-26.51	0.98
37.85	6.79
31.06	15.53
48.20	16.38
38.61	21.23
28.32	8.79
31.21	6.83
34.80	20.30
22.44	14.64
39.76	3.88
43.53	8.70
39.68	12.58
36.87	9.70
38.77	9.62

Table B-1 (Continue)

$V_{s,y}$ (cm/sec)	$V_{s,x}$ (cm/sec)
37.71	17.40
27.46	0.98
49.60	6.43
37.83	8.73
-22.71	0.98
19.98	10.90
45.43	7.73
25.45	9.78
35.03	3.89
29.30	6.83
21.67	7.88
22.42	15.59
34.04	7.78
23.48	11.74
37.77	13.56
34.00	10.69
33.82	21.26

Table B-1 (Continue)

$V_{s,y}$	$V_{s,x}$
<u>(cm/sec)</u>	<u>(cm/sec)</u>
41.57	13.53
28.32	8.79
47.28	12.54
31.91	21.27
30.16	12.65
26.51	1.96
34.91	13.57
34.21	4.88
24.53	7.85
41.62	9.67
17.03	2.00

Table B-2. Statistical Analysis on Solids Velocities Data in Table B-1

	For Axial component	For Radial component
Minimum Value	-26.51	0.98
Maximum Value	60.42	42.14
Sum	2283.91	1056.66
Mean Value	29.28	13.55
Median Value	30.25	12.67
Variance	180.37	72.82
Standard Deviation	13.43	8.53
Standard Error	1.52	0.97
Valid Observations	78.00	78.00
Total Observations	78.00	78.00
95% Confidence Interval	3.03	1.92

Table B-3. Frequency Distribution for Axial Solids Velocity from Table B-1

<u>$V_{s,y}$ (cm/sec)</u>	<u>Frequency</u>
-37.50	0.01
-25.00	0.01
-12.50	0.01
0.00	0.03
12.50	0.22
25.00	0.47
37.50	0.23
50.00	0.01
62.50	0.00

Table B-4. Frequency Distribution for Radial Solids Velocity from Table B-1

<u>$V_{s,x}$ (cm/sec)</u>	<u>Frequency</u>
-2.50	0.08
2.50	0.13
7.50	0.26
12.50	0.31
17.50	0.14
22.50	0.01
27.50	0.03
32.50	0.01
37.50	0.04
42.50	0.00

Table B-5. Raw Solids Velocity Data for Liquid-Solid Fluidized Bed
with Uniform Distributor at X= 4.0 cm and Y= 4.5 cm
(Superficial Liquid Velocity = 6.0 cm/sec)

$V_{s,y}$ (cm/s)	$V_{s,x}$ (cm/s)
26.45	6.85
35.96	4.86
41.75	0.97
33.04	11.66
-33.12	4.87
-46.40	1.93
35.98	1.94
29.21	12.65
36.92	2.91
40.70	4.84
22.72	34.08
60.37	34.50
52.85	24.98
34.88	15.50
28.37	5.86

Table B-5 (Continue)

$V_{s,y}$	$V_{s,x}$
<u>(cm/s)</u>	<u>(cm/s)</u>
41.64	8.71
28.32	8.79
26.42	8.81
22.39	16.55
14.90	8.94
29.30	6.83
16.56	21.43
37.87	1.94
23.41	14.63
27.93	28.90
29.30	6.83
27.35	9.77
29.26	9.75
23.81	3.96
16.04	4.01
34.09	0.97
24.69	0.08

Table B-5 (Continue)

$V_{s,y}$	$-V_{s,x}$
<u>(cm/s)</u>	<u>(cm/s)</u>
29.16	14.58
24.44	11.73
46.39	2.89
-29.35	0.97
34.94	11.64
50.98	16.99
-29.34	3.91
-40.71	0.96
53.95	7.70
28.05	22.24
37.86	2.91
51.13	1.92
-9.92	13.89
-20.78	4.94
49.16	15.42
23.60	68.83
57.76	0.96

Table B-5 (Continue)

$V_{s,y}$	$V_{s,x}$
(cm/s)	(cm/s)
-33.78	0.99
24.59	3.94
14.16	3.04
29.34	2.93
31.24	2.93
-22.68	4.93
-37.87	2.91
-21.77	0.99
25.55	2.95
-25.56	1.97
27.46	0.98
31.25	0.98
26.38	10.75
9.77	27.36
29.34	2.10
20.12	0.73
32.18	3.90

Table B-5 (Continue)

$V_{s,y}$ <u>(cm/s)</u>	$V_{s,x}$ <u>(cm/s)</u>
-23.66	1.97
33.14	0.97
-29.33	4.89
-35.04	0.97
24.60	9.85
18.73	9.86
-31.25	0.98
-34.08	2.92
-25.55	2.94
38.76	10.65

Table B-6. Statistical Analysis on Solids Velocities Data in Table B-5

	For Axial component	For Radial component
Minimum Value	-46.40	0.08
Maximum Value	60.37	68.83
Sum	1306.32	652.15
Mean Value	17.19	8.58
Median Value	26.90	4.86
Variance	791.89	111.85
Standard Deviation	28.14	10.58
Standard Error	3.23	1.21
Valid Observations	76.00	76.00
Total Observations	76.00	76.00
95% Confidence Interval	6.43	2.42

Table B-7. Frequency Distribution for Axial Solids Velocity from Table B-5

<u>$V_{s,y}$ (cm/sec)</u>	<u>Frequency</u>
-56.25	0.01
-43.75	0.08
-31.25	0.13
-18.75	0.01
-6.25	0.00
6.25	0.08
18.75	0.38
31.25	0.20
43.75	0.08
56.25	0.03

Table B-8. Frequency Distribution for Radial Solids Velocity from Table B-5

<u>$V_{s,r}$ (cm/sec)</u>	<u>Frequency</u>
-3.13	0.41
3.13	0.26
9.38	0.20
15.63	0.04
21.88	0.04
28.13	0.03
34.38	0.01
40.63	0.00
46.88	0.00
53.13	0.00
59.38	0.00

Table B-9. Raw Solids Velocity Data for Gas-Liquid-Solid Fluidized Bed with Uniform Distributor at X= 4.0 cm and Y= 4.5 cm (Superficial Liquid Velocity = 6.0 cm/sec, Superficial Gas Velocity = 3.36 cm/sec)

V	V _{s,y}	V _{s,x}
(cm/sec)	(cm/s)	(cm/s)
40.41	-40.38	1.76
28.55	-28.34	3.54
33.46	29.00	16.70
37.72	32.42	19.28
26.12	25.75	4.44
36.92	34.31	8.80
-2.32	29.20	2.65
27.50	26.58	7.09
31.27	30.00	8.82
31.51	30.90	6.18
26.52	5.96	25.84
19.07	-15.15	11.59
34.50	32.55	11.44
26.89	-26.73	2.95
47.06	43.70	17.48

Table B-9 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
13.33	-12.82	3.66
10.63	-9.51	4.75
38.42	34.19	17.53
12.68	10.96	6.39
14.83	14.58	2.73
23.76	-23.15	5.34
33.16	20.18	26.32
34.98	-31.53	14.94
37.88	31.52	21.02
18.58	15.18	10.72
28.59	23.79	15.86
21.68	17.76	12.43
18.54	17.99	4.50
21.21	17.79	11.56
29.36	27.38	10.60
25.09	19.42	15.89
30.05	25.54	15.85

Table B-9 (Continue)

V	V _{s, y}	V _{s, x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
28.05	24.73	13.25
16.41	16.31	1.81
24.57	19.44	15.02
30.01	27.35	12.35
46.59	46.38	4.38
56.86	56.70	4.36
37.92	36.90	8.78
34.50	32.55	11.44
41.23		
23.37	22.26	7.12
19.07	-18.88	2.70
29.41	-29.20	3.54
39.52	-39.52	0.88
25.83	-25.77	1.78
30.97	-30.92	1.77
43.26	-42.94	5.26
24.05		

Table B-9 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
29.22	29.21	0.89
42.65	-42.07	7.01
38.81	-38.65	3.51
39.96	35.92	17.52
37.02	36.94	2.64
19.76		
35.97	27.12	23.62
31.51	28.18	14.09
33.06	14.08	29.92
19.59	5.38	18.84
25.54	17.60	18.48
34.79	32.54	12.31
22.35	22.33	0.89
29.41	29.20	3.54

Table B-10. Raw Solids Velocity Data for Gas-Liquid-Solid Fluidized Bed with Uniform Distributor at X= 7.5 cm and Y= 9.0 cm (Superficial Liquid Velocity = 4.0 cm/sec, Superficial Gas Velocity = 3.36 cm/sec)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
18.70	-15.43	10.56
14.13	13.71	3.34
8.18	-2.59	7.76
16.48	-16.45	0.82
31.59	27.23	16.02
9.40	0.00	9.40
9.47	-4.24	8.47
10.66	-3.37	10.11
11.96	-9.95	6.64
8.65	-8.61	0.86
17.77	-12.96	12.15
15.08	14.87	2.48
16.01	-15.48	4.07
7.01	-5.53	6.06
26.65		

Table B-10 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
22.85	-22.72	2.43
20.35	19.53	5.70
19.98	17.87	8.94
18.45	17.99	4.04
36.31	35.22	8.81
15.08	14.87	2.48
12.76	-12.51	2.50
11.65	-10.87	4.18
11.76		
10.46	-10.14	2.54
12.27	10.80	5.82
12.76	9.02	9.02
14.41	-13.20	5.77
18.23	-16.30	8.15
18.63	-17.97	4.90
15.22	-14.86	3.30
10.66	-8.32	6.66

Table B-10 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
5.71	-1.80	5.41
7.84	-5.95	5.10
10.46	-2.54	10.14
8.94	-8.56	2.57
14.45	-10.62	9.80
10.54	-7.45	7.45
9.91	-9.32	3.39
16.31	-13.05	9.78
10.04	-6.67	7.50
15.87	-13.08	8.99
13.69	2.25	13.50
14.13	-11.50	8.21
11.14	7.45	8.28
10.89	-9.19	5.85
15.87	-5.75	14.79
11.30	-2.11	11.10
16.53	-8.98	13.88

Table B-10 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
18.88	-13.74	12.94
13.39	-4.97	12.43
17.61	-9.77	14.65
14.13	-11.50	8.21
14.13	-13.23	4.96
17.31	-17.23	1.64
16.96	-9.78	13.85
10.22	-8.37	5.86
17.12	-16.39	4.92
13.48	-8.22	10.69
11.94	-10.84	5.00
15.77	-13.94	7.38
14.73	-13.17	6.59
34.95	32.82	12.01
12.64	12.53	1.67
9.70	-9.35	2.55
7.23	-5.11	5.11

Table B-10 (Continue)

V	$V_{s,y}$	$-V_{s,x}$
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
18.04	-18.02	0.82
15.20	14.84	3.30
13.69	-12.40	5.79
12.17	11.70	3.34

Table B-11. Raw Solids Velocity Data for Gas-Liquid-Solid Fluidized Bed with Uniform Distributor at X= 7.5 cm and Y= 9.0 cm (Superficial Liquid Velocity = 7.23 cm/sec, Superficial Gas Velocity = 3.36 cm/sec)

V	V _{s,y}	V _{s,x}
(cm/sec)	(cm/s)	(cm/s)
19.98	-13.72	14.53
11.30	-9.14	6.65
16.96	-13.85	9.78
23.58	-22.67	6.48
19.98	-19.56	4.08
16.66	-11.37	12.18
13.71	7.41	11.53
12.24	-8.24	9.06
16.77	-10.58	13.02
21.41	-8.92	19.46
14.97	10.59	10.59
14.41	13.20	5.77
21.10	-17.80	11.33
12.66	-11.64	4.99
13.32		

Table B-11 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
15.40	14.84	4.12
10.46	-10.14	2.54
17.61	-16.35	6.54
7.39	-6.02	4.30
29.46	-25.68	14.44
13.35	-13.32	0.83
12.64	12.47	2.08
11.78	-11.75	0.84
19.36	17.91	7.33
18.63	17.97	4.90
13.88	4.14	13.24
14.97	-14.88	1.65
16.37	-15.61	4.93
20.15	-19.55	4.89
16.63	-16.44	2.47
25.67	25.04	5.65
25.08		

Table B-11 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
20.62	20.36	3.26
24.47	-24.26	3.23
25.53	-24.22	8.07
12.76	12.51	2.50
15.75	-15.67	1.65
23.63	-23.50	2.43
10.52	-9.23	5.04
13.35	-9.02	9.84
17.21	-14.68	8.97
9.44	-6.67	6.67
14.13	-11.50	8.21
18.96	-18.80	2.45
23.63	-23.50	2.43
15.22	14.86	3.30
27.72	-25.96	9.73
25.09	24.24	6.46
21.10	20.32	5.69

Table B-11 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
20.62	20.36	3.26
24.47	-24.26	3.23
25.53	-24.22	8.07
12.76	12.51	2.50
15.75	-15.67	1.65
23.63	-23.50	2.43
10.52	-9.23	5.04
13.35	-9.02	9.84
17.21	-14.68	8.97
9.44	-6.67	6.67
14.13	-11.50	8.21
18.96	-18.80	2.45
23.63	-23.50	2.43
15.22	14.86	3.30
27.72	-25.96	9.73
25.09	24.24	6.46
21.10	20.32	5.69

Table B-11 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
16.37	15.61	4.93
14.11		
16.88	15.56	6.55
21.41	19.46	8.92
18.45	18.15	3.30
11.96	9.95	6.64
20.15	19.55	4.89
16.16	15.63	4.11
10.46	-10.14	2.54
12.54		
16.16	-15.63	4.11
16.37	-4.93	15.61
16.61	-5.74	15.58
12.76	2.50	12.51
15.77	-7.38	13.94
17.87	-4.91	17.18
15.56	-11.40	10.58

Table B-11 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
18.63	-17.12	7.34
14.13	14.10	0.83
14.73	-13.17	6.59
19.78	-13.99	13.99
8.48	-5.09	6.78

Table B-12. Raw Solids Velocity Data for Gas-Liquid-Solid Fluidized Bed with Central Jet Distributor at X = 23.0 cm and Y = 10.0 cm (Inlet Liquid Velocity = 439.52 cm/sec, Inlet Gas Velocity = 578.88 cm/sec, Inlet Gas Velocity of Left Gas Inlet = 370 cm/sec, Inlet Gas Velocity of Right Gas Inlet 243 cm/sec)

V	V _{s,y}	V _{s,x}
(cm/sec)	(cm/s)	(cm/s)
33.76	11.27	31.82
26.45	6.56	25.63
30.25	1.74	30.20
5.15	2.19	4.66
-7.74	-4.92	-5.97
6.54	4.79	4.45
27.62	4.83	27.20
8.51	0.00	8.51
1.72	1.00	1.40
11.92	2.46	11.66
21.07	2.46	20.93
14.71	1.17	14.66
-10.87	-0.98	-10.82
-36.05	-4.47	-35.77
9.68	5.10	8.23

Table B-12 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
-44.21	-1.19	-44.19
-20.14	-1.12	-20.11
29.78	4.80	29.39
9.06	3.36	8.41
-19.36	-18.36	-6.12
-24.83	-2.25	-24.73
9.71	3.83	8.93
8.07	6.46	4.84
10.85	2.51	10.55
-19.15	-18.50	-4.93
5.17	4.70	2.14
6.65	5.32	3.99
-19.05	-9.60	-16.45
4.95	3.83	3.14
-26.81	-26.75	-1.86
9.25	1.52	9.13
5.14	0.00	5.14

Table B-12 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
10.15	2.16	9.92
12.88	5.29	11.75
-24.81	-0.00	-24.81
-22.08	-21.19	-6.23
8.41	0.84	8.37
16.54	3.05	16.26
10.77	3.84	10.07
9.25	4.76	7.93
-15.70	-11.10	-11.10
8.66	5.00	7.08
8.07	2.22	7.76
-16.40	-9.70	-13.23
-22.33	-12.65	-18.40
12.20	4.53	11.33
-20.47	-4.49	-19.98
3.84	3.24	2.06
5.70	5.50	1.50

Table B-12 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
5.25	0.94	5.16
4.76	4.70	0.78
-18.55	-6.18	-17.50
-14.21	-7.63	-11.99
-14.25	-10.95	-9.12
16.27	1.94	16.15
16.37	1.68	16.28
20.72	0.61	20.71
22.01	4.06	21.64
19.49	2.48	19.33
23.84	8.96	22.10
14.13	1.87	14.01
16.23	0.51	16.22
30.11	2.12	30.04
30.97	0.70	30.96
12.52	1.49	12.43
15.13	2.79	14.87

Table B-12 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
12.75	5.28	11.61
16.50	5.38	15.60
-19.68	-8.35	-17.82
-22.03	-10.55	-19.34
2.22	1.92	1.13
23.86	10.03	21.65
16.86	7.37	15.17
17.18	11.86	12.43
14.77	1.23	14.72
20.43	3.36	20.16
23.35	5.20	22.76
26.81	6.88	25.91
12.46	5.93	10.95
33.71	5.86	33.20
27.77	6.87	26.90
-35.25	-28.61	-20.60
14.16	7.72	11.87

Table B-12 (Continue)

V	V _{s,y}	V _{s,x}
<u>(cm/sec)</u>	<u>(cm/s)</u>	<u>(cm/s)</u>
10.96	9.04	6.21
3.88	3.47	1.74
4.55	3.59	2.79
12.19	4.17	11.46
15.94	6.74	14.45
20.39	9.35	18.12
30.04	17.74	24.24
16.69	10.17	13.23
16.59	7.42	14.84
8.27	3.87	7.31
15.31	5.69	14.22
-15.28	-5.42	-14.29
9.04	6.40	6.40

APPENDIX C
VISCOSITY MEASUREMENT DATA

Table C-1. Raw Viscosity Data at different Spindle Rotation Rate for Three-Phase Fluidized Bed with Uniform Distributor (Superficial Liquid Velocity = 2.037 cm/sec, Superficial Gas Velocity = 3.364 cm/sec, Particles Diameter of 0.08 mm)

Instantaneous Viscosity (cp) at Spindle Rotation Rates of					
<u>n = 2 rpm</u>	<u>n = 4 rpm</u>	<u>n = 6 rpm</u>	<u>n = 10 rpm</u>	<u>n = 12 rpm</u>	<u>n = 20 rpm</u>
1167.0	598.5	199.0	156.0	201.0	123.6
603.0	582.0	474.0	271.2	339.0	87.0
1050.0	388.5	627.0	349.2	475.0	152.7
2130.0	342.0	761.0	133.8	469.5	145.5
1284.0	576.0	475.0	198.6	276.5	159.9
2238.0	643.5	331.0	347.4	350.0	181.2
2043.0	213.0	218.0	117.0	232.5	247.5
1803.0	540.0	226.0	578.4	366.0	255.0
1386.0	444.0	369.0	355.8	412.5	163.2
2622.0	670.5	102.0	355.2	303.5	81.9
1947.0	663.0	178.0	156.0	241.5	144.9
2421.0	297.0	335.0	274.8	293.0	237.9
783.0	312.0	288.0	378.0	399.0	59.1
1359.0	478.5	365.0	165.6	148.0	192.6
1650.0	354.0	259.0	163.2	119.0	153.6
1320.0	183.0	364.0	571.2	273.5	103.2

Table C-1 (Continue)

Instantaneous Viscosity (cp) at Spindle Rotation Rates of					
<u>n = 2 rpm</u>	<u>n = 4 rpm</u>	<u>n = 6 rpm</u>	<u>n = 10 rpm</u>	<u>n = 12 rpm</u>	<u>n = 20 rpm</u>
1134.0	193.0	364.0	120.0	354.0	192.3
1818.0	763.5	474.0	378.6	267.0	112.2
2328.0	513.0	368.0	292.8	197.0	245.5
1755.0	574.5	584.0	457.2	147.5	271.5
2385.0	829.5	759.0	371.4	244.5	273.6
2547.0	670.5	706.0	465.6	266.5	292.8
2022.0	298.5	883.0	373.8	166.0	173.1
1509.0	331.5	587.0	166.2	73.0	249.6
1413.0	550.5	452.0	186.0	139.0	233.4
699.0	996.0	283.0	92.4		90.9
489.0	604.5	132.0	192.0		192.9
618.0	415.5	172.0	311.4		48.0
567.0	744.0	211.0	501.6		257.1
627.0	1199.0	149.0	578.4		235.0
1431.0	1154.0	358.0			92.4
591.0	1373.0	636.0			190.2

Table C-1 (Continue)

Instantaneous Viscosity (cp) at Spindle Rotation Rates of					
<u>n = 2 rpm</u>	<u>n = 4 rpm</u>	<u>n = 6 rpm</u>	<u>n = 10 rpm</u>	<u>n = 12 rpm</u>	<u>n = 20 rpm</u>
1173.0	1116.0	250.0			175.2
1815.0	937.5	174.0			
1848.0	1077.0	283.0			
1056.0	955.5	405.0			
801.0	706.5	316.0			
1029.0	229.5	346.0			
1443.0	646.5	595.0			
1110.0	478.5	182.0			
600.0	324.0				
1482.0	354.0				
1176.0	390.0				
1608.0	373.5				
	880.5				
	994.5				
	957.0				
	844.5				
	640.5				

Table C-1 (Continue)

Instantaneous Viscosity (cp) at Spindle Rotation Rates of					
<u>n = 2 rpm</u>	<u>n = 4 rpm</u>	<u>n = 6 rpm</u>	<u>n = 10 rpm</u>	<u>n = 12 rpm</u>	<u>n = 20 rpm</u>
	463.5				
	757.5				
	574.5				

Table C-2. Raw Viscosity Data at Spindle Rotation Rate of 6 rpm for Three-Phase Fluidized Bed with Central Jet Distributor at different Inlet Liquid Velocities (Inlet Gas Velocity at Right Inlet = 131.0 cm/sec, Inlet Gas Velocity at Left Inlet = 135.0 cm/sec, Particles Diameter = 1.0 mm)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
703.0	679.0
524.0	460.0
826.0	547.0
697.0	468.0
469.0	1000.0
712.0	985.0
611.0	432.0
881.0	909.0
928.0	851.0
647.0	922.0
991.0	920.0
646.0	589.0
888.0	667.0
525.0	608.0
495.0	367.0

Table C-2 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
503.0	461.0
657.0	738.0
748.0	845.0
113.0	575.0
622.0	492.0
885.0	399.0
997.0	678.0
697.0	887.0
231.0	903.0
546.0	795.0
570.0	865.0
701.0	963.0
325.0	947.0
750.0	958.0
902.0	610.0
720.0	974.0
638.0	641.0

Table C-2 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
90.0	525.0
201.0	709.0
496.0	586.0
969.0	851.0
665.0	712.0
135.0	885.0
284.0	748.0
263.0	419.0
83.0	741.0
317.0	905.0
206.0	589.0
550.0	873.0
890.0	980.0
900.0	841.0
717.0	466.0
970.0	735.0
931.0	842.0

Table C-2 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
783.0	964.0
782.0	737.0
542.0	789.0
543.0	711.0
171.0	963.0
397.0	860.0
509.0	766.0
552.0	467.0
627.0	437.0
604.0	402.0
693.0	407.0
303.0	416.0
549.0	547.0
377.0	768.0
330.0	988.0
346.0	963.0
527.0	790.0

Table C-2 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
937.0	634.0
974.0	798.0
576.0	767.0
226.0	493.0
848.0	831.0
884.0	776.0
761.0	419.0
684.0	119.0
207.0	479.0
214.0	678.0
674.0	551.0
972.0	244.0
721.0	581.0
466.0	595.0
434.0	751.0
529.0	658.0
787.0	822.0

Table C-2 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
750.0	889.0
571.0	760.0
798.0	608.0
733.0	404.0
606.0	322.0
301.0	516.0
323.0	557.0
274.0	876.0
284.0	515.0
809.0	451.0
812.0	832.0
609.0	879.0
525.0	
795.0	
981.0	
365.0	
385.0	

Table C-2 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
734.0	
384.0	
200.0	
263.0	
805.0	
873.0	
636.0	
609.0	
659.0	
385.0	
107.0	
318.0	
164.0	
737.0	
629.0	
814.0	
826.0	

Table C-2 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
909.0	
952.0	
881.0	
454.0	
648.0	
395.0	
942.0	
794.0	
146.0	
176.0	
441.0	
556.0	
967.0	
832.0	
697.0	
685.0	
985.0	

Table C-2 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>232.4 cm/sec</u>	<u>154.9 cm/sec</u>
430.0	
721.0	
814.0	
947.0	
725.0	
259.0	
179.0	
673.0	
545.0	
383.0	
392.0	
975.0	

Table C-3. Raw Viscosity Data at Spindle Angular Velocity of 6 rpm for Liquid-Solid Fluidized Bed with Central Jet Distributor (Particles Diameter = 1.0 mm)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>322.4 cm/sec</u>	<u>154.9 cm/sec</u>
913.0	119.0
599.0	104.0
295.0	103.0
628.0	110.0
904.0	100.0
529.0	116.0
219.0	198.0
670.0	176.0
347.0	123.0
269.0	145.0
467.0	121.0
630.0	175.0
746.0	187.0
229.0	101.0
174.0	202.0

Table C-3 (Continue)

Instantaneous Viscosity (cp) Inlet Liquid Velocity of	
<u>322.4 cm/sec</u>	<u>154.9 cm/sec</u>
180.0	
205.0	
150.0	
117.0	
278.0	
702.0	
873.0	
486.0	
411.0	
346.0	
711.0	

APPENDIX D
HYDRODYNAMIC MODEL

TABLE D-1. Governing Equations -- Hydrodynamic Model B
for Gas-Liquid-Solid Flow

In the following equations, the tensor is represented as [].

D.1 Continuity Equation for Phase k ($= g, l, s$).

$$\frac{\partial}{\partial t} (e_k \rho_k) + \nabla \cdot (e_k \rho_k \mathbf{v}_k) = 0$$

D.2 Momentum Equations.

a) Liquid Phase (Continuous Phase)

$$\frac{\partial}{\partial t} (e_l \rho_l \mathbf{v}_l) + \nabla \cdot (e_l \rho_l \mathbf{v}_l \mathbf{v}_l) = -\nabla p_l + \rho_l \mathbf{g} + \sum_{m=g,s} \beta_{lm} (\mathbf{v}_m - \mathbf{v}_l) + \nabla \cdot [\boldsymbol{\tau}_l]$$

b) Gas / Solid Phases k ($= g, s$) (Dispersed Phases)

$$\begin{aligned} \frac{\partial}{\partial t} (e_k \rho_k \mathbf{v}_k) + \nabla \cdot (e_k \rho_k \mathbf{v}_k \mathbf{v}_k) = & -\nabla p_k + \frac{e_k}{e_l} (\rho_k - \sum_{m=l,g,s \text{ } m \neq k} e_m \rho_m) \mathbf{g} \\ & + \sum_{m=l,g,s \text{ } m \neq k} \beta_{km} (\mathbf{v}_m - \mathbf{v}_k) + \nabla \cdot [\boldsymbol{\tau}_k] \end{aligned}$$

D.3 Constitutive Equations for Stress.

a) Fluid Phase Stress. k ($= l, g$).

$$[\boldsymbol{\tau}_k] = e_k \mu_k \left([\nabla \mathbf{v}_k + (\nabla \mathbf{v}_k)^T] - \frac{2}{3} \nabla \cdot (\mathbf{v}_k) [\mathbf{I}] \right)$$

b) Solid Phase Stress.

$$[\tau_s] = \mu_s [\nabla \mathbf{v}_s + (\nabla \mathbf{v}_s)^T] + (\xi_s - \frac{2}{3} \mu_s) \nabla \cdot (\mathbf{v}_s) [\mathbf{I}]$$

D.4.1 Kinetic Theory Model.

a) Fluctuating Energy $\Theta_s (= \frac{1}{2} \langle C^2 \rangle)$ Equation,

$$\frac{3}{2} \left(\frac{\nabla}{\nabla t} (e_s \rho_s \Theta_s) + \nabla \cdot (e_s \rho_s \Theta_s \mathbf{v}_s) \right) = -p_s \nabla \cdot \mathbf{v}_s + \Phi_k - \nabla \cdot \mathbf{q}_s - \gamma_s - 3\beta_{1s} \Theta_s$$

b) Collision Energy Dissipation Rate,

$$\gamma_s = 3(1-e^2) e_s^2 \rho_s d_s g_0 \Theta_s \left(\frac{4}{d_k} \left(\frac{\Theta_s}{\pi} \right)^{\frac{1}{2}} - \nabla \cdot \mathbf{v}_s \right)$$

c) Flux of Fluctuating Energy,

$$\mathbf{q}_s = -\kappa_s \nabla \Theta_s$$

d) Conductivity of Fluctuating Energy,

$$\kappa_s = \frac{2\kappa_{dil}}{(1+e) g_0} \left[1 + \frac{6}{5} (1+e) g_0 e_s \right]^2 + 2e_s^2 \rho_s d_s (1+e) g_0 \left(\frac{\Theta_s}{\pi} \right)^{\frac{1}{2}}$$

e) Dilute Phase ("Eddy Type") Granular Conductivity,

$$\kappa_{dil} = \frac{25\sqrt{\pi}}{128} \rho_s d_s \Theta_s^{\frac{1}{2}}$$

f) Solid Phase Shear Viscosity,

$$\mu_s = \frac{2\mu_{dil}}{(1+e) g_0} \left[1 + \frac{4}{5} (1+e) g_0 e_s \right]^2 + \frac{4}{5} e_s^2 \rho_s d_s (1+e) g_0 \left(\frac{\Theta_s}{\pi} \right)^{\frac{1}{2}}$$

g) Solid Phase Bulk Viscosity,

$$\xi_s = \frac{4}{3} \epsilon_s^2 \rho_s d_s (1+e) g_d \left(\frac{\Theta_s}{\pi} \right)^{\frac{1}{2}}$$

h) Dilute Phase Solids Viscosity,

$$\mu_{dil} = \frac{5\sqrt{\pi}}{96} \rho_s d_s \Theta_s^{\frac{1}{2}}$$

i) Solid Phase Pressure

$$p_s = \epsilon_s \rho_s \Theta_s [1 + 2(1+e) \epsilon_s g_0]$$

j) Radial Distribution Function,

$$g_0 = \left[1 - \left(\frac{\epsilon_s}{\epsilon_{s, \max}} \right)^{\frac{1}{3}} \right]^{-1}$$

k) Energy Dissipation Rate

$$\Phi_s = [\tau_s] : \nabla \mathbf{v}_s$$

D.4.2 Empirical Solids Viscosity and Stress Model.

$$\begin{aligned} \nabla p_k &= G(\epsilon_k) \nabla \epsilon_k \\ \xi_s &= 0 \\ \mu_s &= 5 \epsilon_s \text{ (poises)} \quad (\text{example}) \end{aligned}$$

D.5 Intra--Phase Drag Coefficients. k ($= g, s$)

a) for $\epsilon_i < 0.8$, (based on Ergun equation)

$$\beta_{ik} = \beta_{ki} = 150 \frac{(1-\epsilon_i) \epsilon_k \mu_i}{(\epsilon_i d_k \Psi_k)^2} + 1.75 \frac{\rho_i \epsilon_k |\mathbf{v}_i - \mathbf{v}_k|}{\epsilon_i d_k \Psi_k}$$

b) for $\epsilon_i \geq 0.8$, (based on empirical correlation)

$$\beta_{1k} = \beta_{k1} = \frac{3}{4} C_D \frac{\epsilon_k \rho_1 |\mathbf{v}_1 - \mathbf{v}_k|}{d_k \psi_k} \epsilon_1^{-2.65}$$

where,

$$C_D = \begin{cases} \frac{24}{Re_k} (1 + 0.15 Re_k^{0.687}) & \text{for } Re_k < 1000 \\ 0.44 & \text{for } Re_k \geq 1000 \end{cases}$$

$$Re_k = \frac{\epsilon_1 \rho_1 |\mathbf{v}_1 - \mathbf{v}_k| d_k \psi_k}{\mu_1}$$

D.6 Particle-Particle Drag Coefficients. k, m ($= l, g, s$)

$$\beta_{km \ k+m} = \frac{\alpha (1+e) \epsilon_k \rho_k \epsilon_m \rho_m (d_k + d_m)^2 \left[1 + 3 \left(\frac{\epsilon_{km}}{\epsilon_k + \epsilon_m} \right)^{\frac{1}{3}} \right]}{2 \epsilon_f (\rho_k d_k^3 + \rho_m d_m^3) \left[\left(\frac{\epsilon_{km}}{\epsilon_k + \epsilon_m} \right)^{\frac{1}{3}} - 1 \right]} |\mathbf{v}_k - \mathbf{v}_m|$$

where

$$\epsilon_{km} = \begin{cases} [(\phi_k - \phi_m) + (1-a)(1-\phi_k)\phi_m] [\phi_k + (1-\phi_m)\phi_k] \frac{X_k}{\phi_k} + \phi_m & \text{for } X_k \leq X_{k,o} \\ (1-a) [\phi_k + (1-\phi_k)\phi_m] (1-X_k) + \phi_k & \text{for } X_k \geq X_{k,o} \end{cases}$$

$$a = \sqrt{\frac{d_m}{d_k}} \quad (d_k \geq d_m)$$

$$X_k = \frac{\epsilon_k}{\epsilon_k + \epsilon_m}$$

$$X_{k,o} = \frac{\phi_k}{[\phi_k + (1-\phi_k)\phi_m]}$$

D.7 Definitions.

$$\epsilon_l + \epsilon_g + \epsilon_s = 1$$

The gas, liquid and solid phases are assumed to be incompressible and constant densities are used. The constitutive relations in kinetic theory model were derived by Gidaspow (1994). The constitutive relations may be easily changed by some program modification.

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