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AN INVESTIGATION OF THE ENTRANCE REGION OF A RISER FLUIDIZED BED

by

Apostolos E Kostazos

**A dissertation submitted to the Graduate Faculty in Engineering in partial
fulfillment of the requirements for the degree of Doctor of Philosophy,
The City University of New York**

1997

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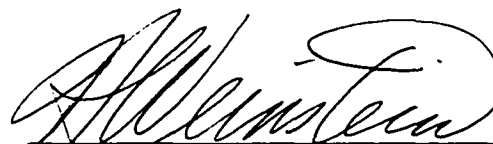
APOSTOLOS E KOSTAZOS

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This manuscript has been read and accepted for the Graduate Faculty in Engineering in satisfaction of the dissertation requirement for the degree of Doctor of Philosophy.

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Abstract

**AN INVESTIGATION OF THE ENTRANCE REGION OF A RISER
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by

Apostolos E Kostazos

Adviser: Professor Herbert Weinstein

The performance of a riser fluidized bed reactor is dependent on rapidly mixing jets of reactant fluid (gas or volatile liquid) through a flowing gas-catalyst powder mixture. Since riser reactors are typically employed with fast reaction systems, riser length is not long, typically being the 20 to 30 diameters which make up an entrance length. The rapidity of the initial mixing process and the suppression of backmixing are, therefore, critical in maximizing selectivity and yield.

An experimental work was carried out to determine the effect of the nozzle configuration on the mixing of the gas and catalyst powder. For this work the 15.2cm ID City College riser was used, equipped with a new entrance section. This added to the system two more operating conditions, the radial position of the gas injection and the ratio between the primary and secondary or "fluffing" air flowrate. The facility is equipped with the necessary instrumentation for axial and radial pressure difference measurements, radial solid density and solid flux measurements.

The results showed that the radial position of the gas injection plays the dominant role in determining the rapidity of the mixing of gas and catalyst. Gas injection close the

riser axis creates a flow with strong radial concentration and velocity gradients which increase the backmixing and reduce the degree of contact between gas and catalyst. On the other hand, gas injection close the riser wall creates a flow with much flatter density and velocity profiles and therefore good gas solid contact and minimization of backmixing which results in higher reactor performance. The database provided with this work can be used as guidance for improved phenomenological design of this kind of reactor and also for the development of a two-dimensional CFD model.

ACKNOWLEDGMENTS

This acknowledgment, unfortunately, must be written to Apostolos E. Kostazos rather than by him. He passed away after a very short illness on December 5, 1996, exactly one week before he was to defend this dissertation.

Apostolos was a model graduate student. He was very bright and eager to learn. He worked very hard but was always relaxed, easy going and a pleasure to have nearby. He accomplished a great deal: his course work, dissertation research, and two additional experimental projects in the three and one-half years he was with us at City College. He carried out all of his work with honesty, intelligence and great care.

I know that had Apostolos the opportunity to write this himself, he would have acknowledged his parents with deep love and respect. He would have also written of his closeness to his brother, a special uncle and cousin, and the other members of his family who he held dear. I must apologize for not remembering all the names he spoke of so fondly.

We will miss Apostolos and we will remember him.

H.W.

January 13, 1997

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I. INTRODUCTION

Fast or circulating fluidized bed chemical reactors have been used in a wide variety of applications in industry, especially in the manufacture of petrochemicals, the combustion of coal and in calcining processes. The main features that make the fast fluidized bed reactor better than a fixed bed or conventional fluidized bed reactor, are summarized as follows:

- ♦ High throughput and closely plug flow of gas, resulting in higher selectivity of intermediate products.
- ♦ High relative mass and heat transfer coefficients between the solid particles and the gas phase which result from the fineness of particles and the high relative gas-solid velocity.
- ♦ More uniform and controllable temperature due to the circulation of solid. The high solid flux provides either a heat source or heat sink to control the temperature of the riser.
- ♦ Ability to independently adjust the solid and gas retention time. This can help to increase the turn down ratio of the reactor or to manage those processes with varying feed quality or product requirements.
- ♦ Ability to continuously circulate catalyst through separate zones for regeneration, periodic operation or other applications. This feature makes the fast fluidized bed reactor an ideal solution for catalytic reactions where the catalyst deactivates rapidly.

The above features make fast fluidized bed reactors appropriate for:

- Reactions with high or very high reaction rate.
- Reactions with reversible catalyst deactivation calling for continuous catalyst regeneration.
- Reactions where the desired products are intermediate.
- Highly exothermic or endothermic reactions.
- Processes with varying feed and product requirements.
- Reactions with staged reactant inlets.
- Reactions that require independent solid and gas retention-time.
- Reactions that benefit from periodic operation such as oxidative reactions in a stoichiometric redox mode.
- Reactions that require temperature uniformity.
- Processes where heat removal or supply can be achieved by solid circulation.

The first steps in describing the hydrodynamics of a fast fluidized bed were reported in the last decades. However, there is very limited understanding of the flow physics. A typical fast fluidized bed is operated in two distinct regimes, a dense phase in the bottom and a dilute phase in the top which gives the solid concentration an axial S-shaped profile. Weinstein et al. (1986) and Werther (1986) showed the existence not only of an axial solid distribution but also the existence of a radial solid distribution. Weinstein et al. (1989) also found very different behavior between the dense and the dilute phase of a fast fluidized bed in terms of gas backmixing and gas-solid contact, for

various gas velocities and solid circulation rates. Although both dilute and dense phase coexist in a fast fluidized bed there are many differences between these two flow regions.

Until today a method to describe in terms of design, the flow pattern in a fast fluidized bed is not available. There is still a real need for methods to estimate the required parameters that characterize this type of flow.

The performance of a fast fluidized bed reactor depends on the distribution of the solid in the entrance region. A smooth acceleration of the solid indicates good gas-solid contact and minimization of backmixing. The good contact and mixing between gas and solid results in high reaction performance. The minimization of backmixing increases the reaction selectivity. The solid and the gas products do not return to the entrance region and therefore the reactions terminate at the desired level.

The focus of this work is fast fluidized bed reactors employed for very fast reactions ($\tau \approx 0.5-5s$). In this case the initial mixing rate between the catalyst and the reactants is very critical in determining conversion and selectivity. A fluid bed catalytic cracking unit (FCCU) riser is an important example of this type of reactor. In this riser the regenerated catalyst is carried by "fluffing" steam up to the oil nozzles elevation.

The inlet configuration of the fast fluidized bed determines the acceleration and distribution of the catalyst in the critical inlet zone. Computational Fluid Dynamic (CFD) models of the two phase flow have only been able to predict the fully developed flow characteristics and not the flow in the critical acceleration zone of the fast fluidized bed. Fully developed flow in a fast fluidized bed may exist at a height above 20 L/D. Commercial fast fluidized bed reactors typically have a height of only about 20-30 L/D

which means practically that the entire reactor is an entrance flow region. Therefore there is a real need to determine the gas-solid flow in the section from 0 to 20-30 L/D above the inlet configuration.

The scope of this work is to provide accurate and internally consistent sets of measurements that will describe completely the gas-solid flow in the critical inlet zone. These measurements can be used for the development of CFD models and as guidance for phenomenological design of a fast fluidized bed reactor.

II. PREVIOUS WORK

2.1 High Velocity Fluidization Regime

Fast fluidization, according to Li (1994), denotes a new type of bubbleless gas-solid contact, mainly characterized by the aggregation of particles into clusters. Typical axial and radial solid distributions for three types of high velocity fluidization have been drawn in Figures 2.1 and 2.2 respectively.

In the first type, Figure 2.1(a), the bubbling fluidized bed, there is a dense phase which is interrupted by bubbles which rise to the top of the solid surface. Therefore the axial profile shows the existence of a dense phase over all of the height of the bed. The radial profile for the same type of fluidized bed, Figure 2.2(a), shows the existence of the

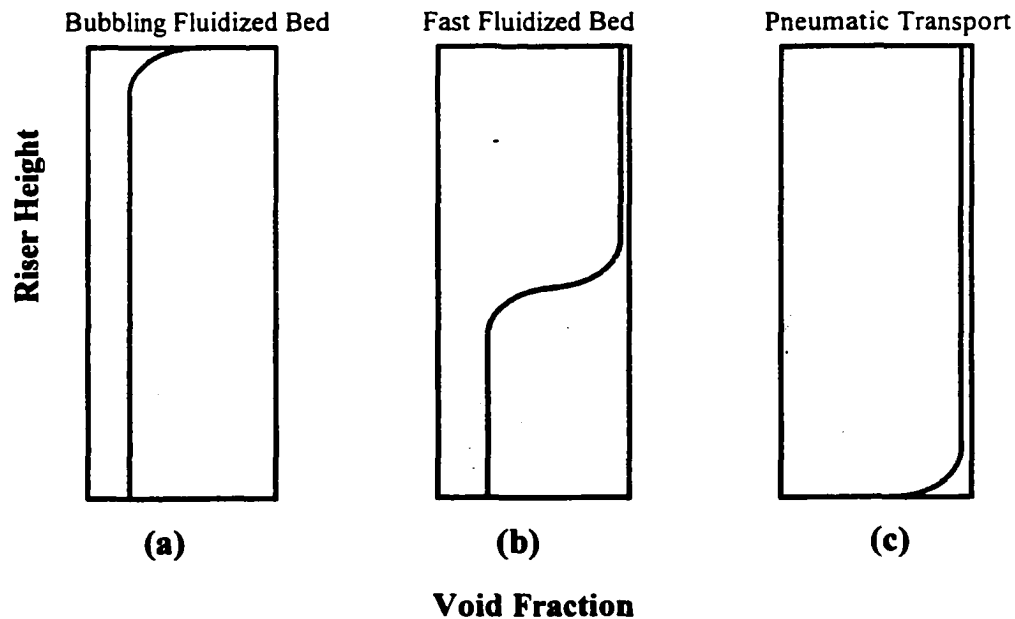


Figure 2.1: Axial Voidage Profile for High Velocity Fluidization Regime.

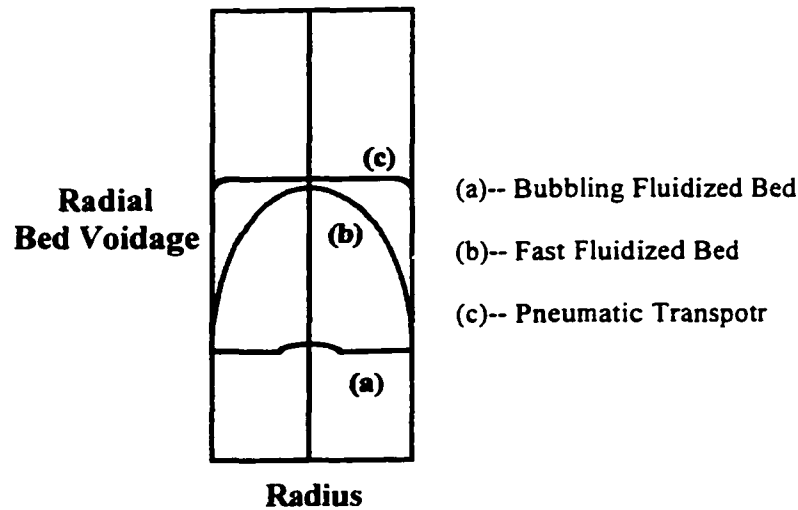


Figure 2.2: Radial Voidage Profile for High Velocity Fluidization Regime.

rising bubble. The bubbles are distributed throughout the bed in time. Spatially, the time averaged bed properties are uniform.

With increasing gas velocity the bed will enter the fast fluidized bed regime, which is characterized by the coexistence of two flow regions as the Figure 2.1(b), shows. There is a dense phase in the bottom and a dilute phase in the top. There is also a radial solid profile in a fast fluidized bed, Figure 2.2(b). Thus there is a dense annulus and a dilute core region. The average solid voidage in a fast fluidized bed depends on the superficial gas velocity U_g , the solid circulation rate G_s , and on the imposed pressure drop, ΔP . Figure 2.3 shows the hypothetical case of a very narrow fast fluidized bed or riser open to the atmosphere fed solid from a very wide slow bed, which is also open to the atmosphere. The slow bed is maintained in a bubbling fluidization condition and is wide enough such that the height of the solid in it, is independent of time. The narrow bed is a fast fluidized bed with a superficial gas velocity U_g . The rate of the solid that exits

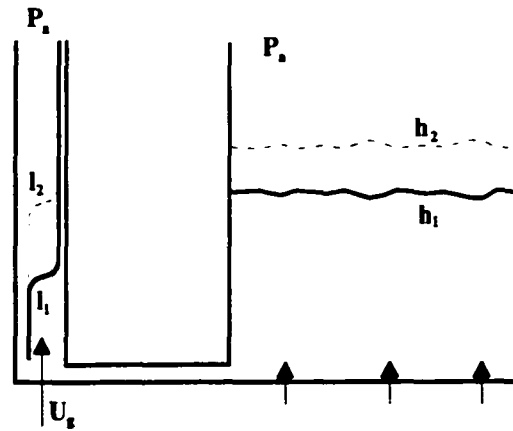


Figure 2.3: Effect of the Imposed Pressure Drop on the Axial Voidage Profile.

from the riser is low and constant since the height of the solid in the slow bed is constant. For a certain height of solid in the slow bed, h_1 (which means a certain imposed pressure drop ΔP_1) the voidage distribution in the riser will have an S-shaped profile and the inflection point will be at a certain level in the fast bed, l_1 . With increasing the height of solid in the slow bed h_2 , ($\Delta P_2 > \Delta P_1$) and keeping the superficial gas velocity constant (solid circulation rate remains constant), the height of the inflection point of the axial voidage profile in the riser will increase. Therefore the gas-solid flow in a fast fluidized bed is a function of the operating conditions of the whole system, riser and downcomer.

Matsen (1988) describes the fast fluidization regime at higher voidages and slip velocities. The continuous phase of turbulent fluidization disappears and the solid is only present as discrete particles or as ephemeral assemblages or clusters of a statistical-hydrodynamic nature. Dilute phase fluidization is characterized by having a slip velocity which decreases as voidage increases. However the boundary between the turbulent and

fast fluidized bed is veiled. A most general criterion for fast fluidized beds is that the fast bed contains both a dense phase and a dilute entrained phase while a turbulent fluidized bed holds a dense phase and a very dilute freeboard.

With increasing gas velocity we finally reach the pneumatic transport of solid. In this case the flow is a dilute flow all over the entire height of the bed and the radial solid profile is flat as it can be seen from the Figures 2.1(c) and 2.2(c) respectively.

Fast fluidization is a type of high velocity fluidization which is characterized by two flow regions in both directions, radial and axial. The average solid fraction in slow velocity fluidized beds depends only on the gas velocity. On the other hand, the average solid fraction in fast fluidized beds depends not only on the gas velocity but also on the solid circulation rate and the imposed pressure drop.

2.2. Solid Distribution in a Fast Fluidized Bed

The axial and radial solid profile in a fast fluidized bed are of great interest in terms of design because they give information about the flow structure in the bed which plays an important role in gas-solid mixing, heat and mass transfer and in conversion and selectivity when the bed is used as a chemical reactor.

Many investigations have been carried out in order to measure the axial variation in solid concentration. Figure 2.4 shows typical results of axial profile measurements (Li and Kwauk, 1980; Li et al., 1981). Experiments were conducted using different solid materials in a bed with 90mm I.D. and 8m high. The apparent voidage profile was calculated from the pressure gradient measurements using the following equation:

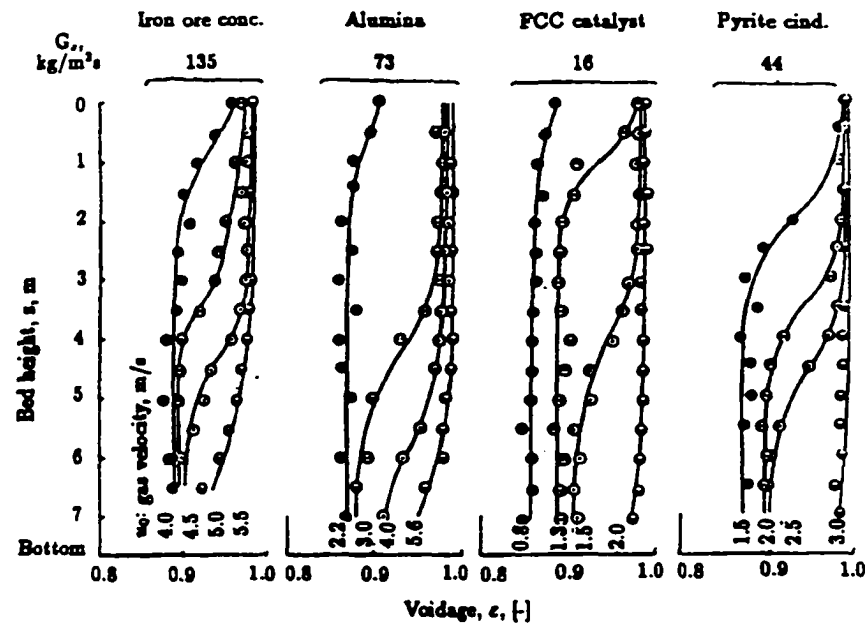


Figure 2.4: Axial Voidage Profile for Four Powders in a Fast Fluidized Bed (after Li and Kwauk, 1980a).

$$\frac{dP}{dz} = \rho_s(1 - \epsilon)_{\text{apparent}}g \quad (2.1)$$

This relation does not take into account the acceleration of solid which is important at the bottom of the riser (Weinstein and Li, 1989), the gas and solid wall friction and the gas density. It can be seen from Figure 2.4 that for any given solid circulation rate, most of the voidage profiles display a sigmoidal distribution, and the position of the inflection point becomes lower as gas velocity increases, indicating that solid inventory in the fast column diminishes with increasing gas velocity.

The position of the inflection point is dependent not only on the superficial gas velocity U_g , but also on the solid circulation rate G_s , as well as on the imposed pressure drop ΔP . For any given gas velocity an increase of the solid circulation rate has as result

the increase of the height of the inflection point. Weinstein et al. (1983) showed that for the same superficial gas velocity U_g , and solid circulation rate G_s , an increase of the imposed pressure drop causes an increase of the height of the inflection point.

Abed (1984) showed that in order to have a true steady state reading in a fast fluidized bed the pressure reading must be averaged over 3 min. The same low frequency pressure fluctuations were observed by Schnitzlein and Weinstein, (1988), using the Fast Fourier Transform of a record of instantaneous differential pressure readings in the riser. In order to determine the appropriate statistical average value of the instantaneous pressure gradient readings to be used to determine the voidage, they formulated a number of different possible methods such as: median value, mean value or minmax average value. The void fraction in a fast fluidized bed is evaluated on the basis of these three different values for the differential pressures. The results differ greatly, and the tendency to increasing void fraction with increasing height in the bottom section is more pronounced when a minmax average is used and diminishingly small when the mean and median values are used. It was concluded that only when the mean or median of the PDF's is used, it can be said that the void fraction obtained is essentially constant with height within the bottom region of the fast fluidized bed.

Feindt (1990) showed that the spacing of the pressure taps in a riser affects the pressure fluctuation readings. He showed that both the amplitude and the time average of differential pressure readings in a fast fluidized bed, increase with decreasing spacing of the pressure taps ($< 2.5 L/D$). The probability density functions of the pressure fluctuations are symmetric for a spacing greater than $5 L/D$ and become more and more

asymmetric in shape with decreasing spacing. Pressure fluctuations with a spacing less than $2.5 L/D$ are strongly influenced by the local turbulence in the flow. On the other hand an averaged effect of the turbulence was observed for spacing greater than $5 L/D$.

Local voidages for FCC catalyst at various radial positions were measured by Li et al. (1980b) using an optical fiber probe. Their results are shown in Figure 2.5. When gas velocity is less than the minimum velocity for fast fluidization (1.25m/s for that case), the radial voidage profile is relative flat; when gas velocity increases further, this profile becomes steeper: higher in center and lower near the wall. As gas velocity increases, pneumatic transport of solid takes place. The radial particle concentration in this case is

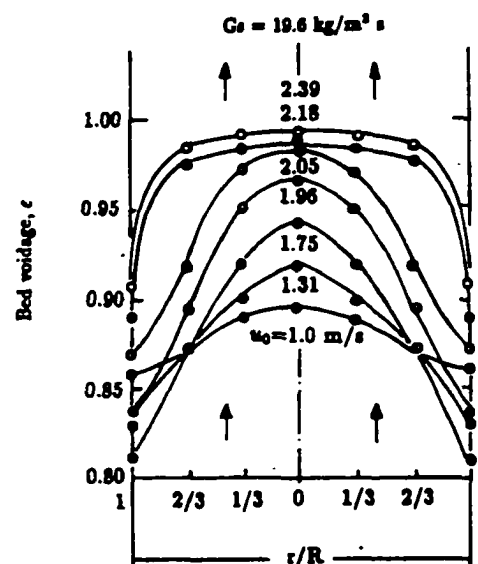


Figure 2.5: Radial Voidage for Fast Fluidized Beds (after Li et al. 1980, 1985).

flat and very dilute in the core while most of the solid aggregate in the vicinity of the wall. This core-annulus distribution is different from that of the fast fluidization, for which there is considerable variation of the radial solid density even in the core region.

Feindt (1990), measured the radial profile of a 15.2 cm I.D., 8.50m high, fast fluidized bed in both dense and dilute phases, using an x-ray absorption technique. Figure 2.6 shows the radial density profile in the dense phase for three different heights with the same cross sectional voidage. It can be observed that the radial voidage remains the same for the different heights. On the other hand in Figure 2.7 is plotted the radial density profile for three different heights in the dilute phase of the fluidized bed. It may be observed that the radial density is changing with height. The particular decrease of the solid fraction in the wall region explains this behavior. The mean cross sectional solid fraction is also decreasing with increasing height. The radial density profile in the bottom of the bed also depends on the gas inlet configuration. Weinstein et al. (1995) showed

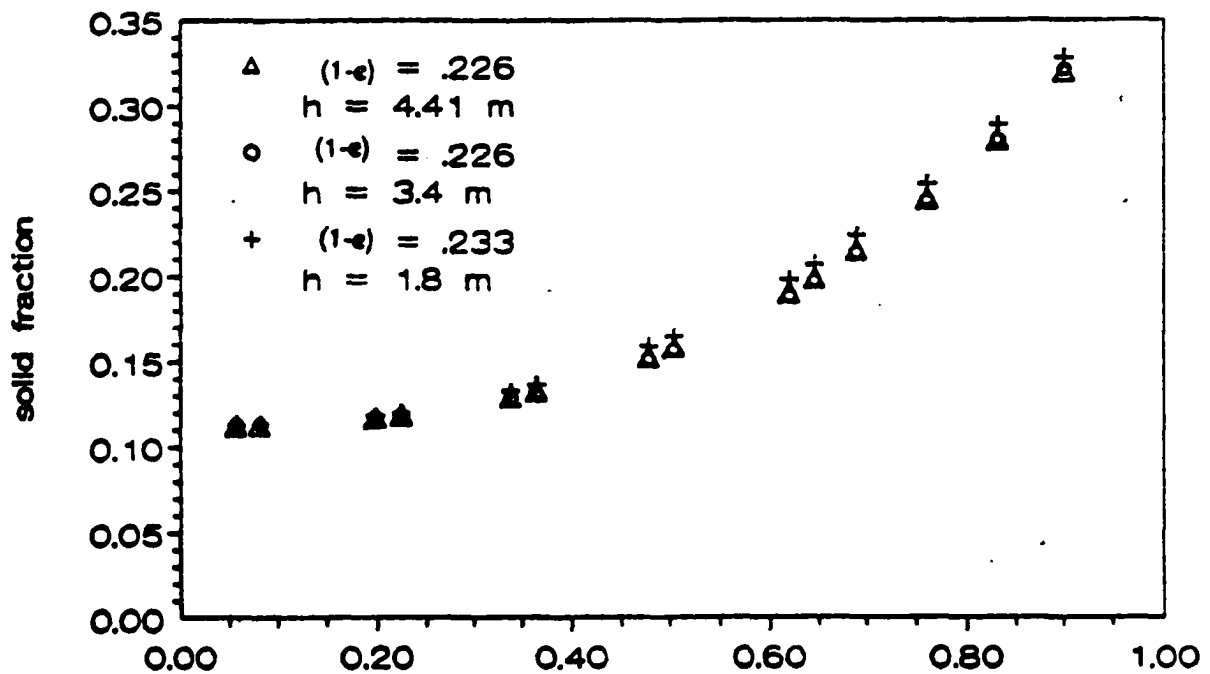


Figure 2.6: Radial Density Profile at Three Different Heights, Dense Phase, (Feindt, 1990).

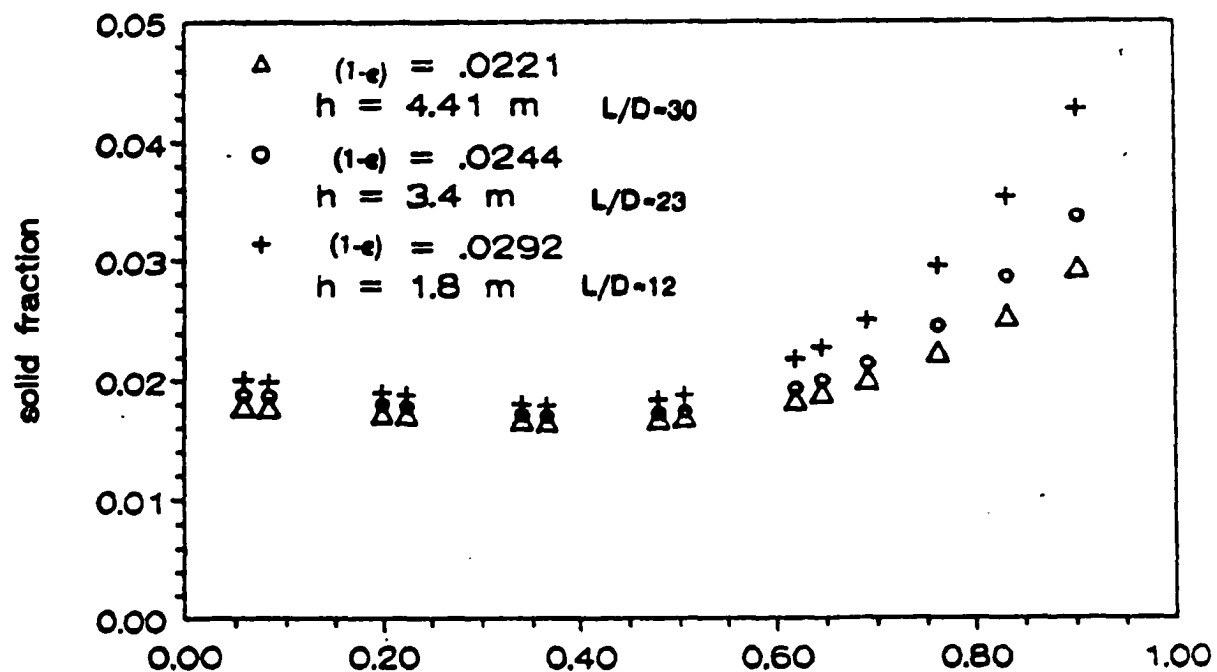


Figure 2.7: Radial Density Profile at Three Different Heights, Dilute Phase, (Feindt, 1990).

that the radial solid distribution in the entrance region of a riser is strongly influenced by the gas nozzle configuration. They used three different types of nozzles and they found a different radial density profile for each nozzle. They also showed that the influence of the nozzle configuration can be felt as far as 20 L/D downstream.

Monceaux et al. (1985) and Rhodes et al. (1988) used an non-isokinetic probe for the measurement of the local solid flux at different positions in the riser. They showed that the local solid flux profiles have a parabolic form as is shown in Figure 2.8. Close to the riser wall, the net flux is negative which means that in this region the solid is moving downward. On the other hand, in the center of the riser the solid is moving upward. Both these fluxes, the core net upward and the annulus net downward fluxes, are decreasing

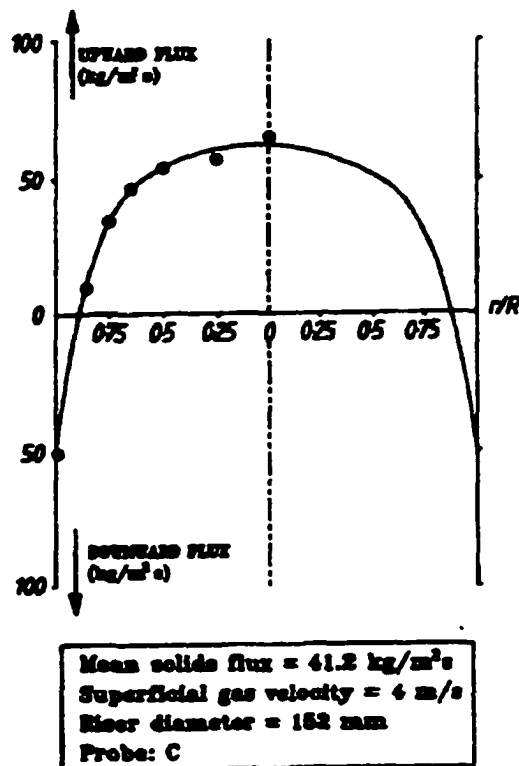


Figure 2.8: Radial Solid Flux Profile, (Rhodes et al. 1988).

with the height of the bed when the cross-sectional average void fraction is decreasing. An explanation of this behavior is the transfer of solid between these two regions, which with the radial solid distribution, imply the existence of a radial pressure difference profile.

Weinstein et al. (1995) examined the existence of this radial pressure difference using two different probe designs. Some of the results are shown in Figure 2.9. It is shown that there is a radial pressure profile with maximum difference at the center of the riser. Figure 2.10 shows that this maximum pressure difference increases with the mean cross-sectional solid fraction. It can be seen that the conclusions from these results

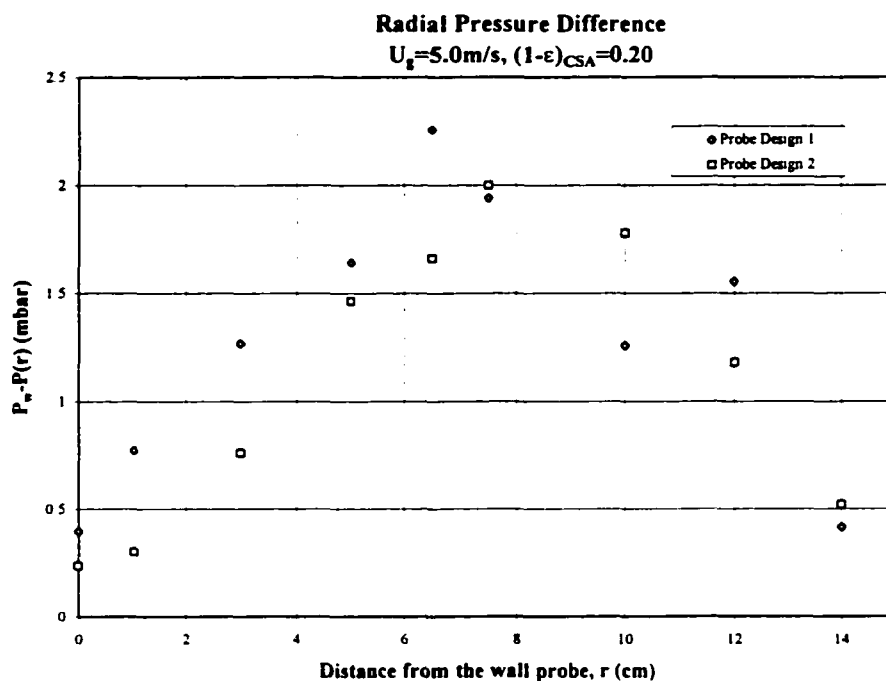


Figure 2.9: Radial Pressure Difference in a Fast Fluidized Bed.

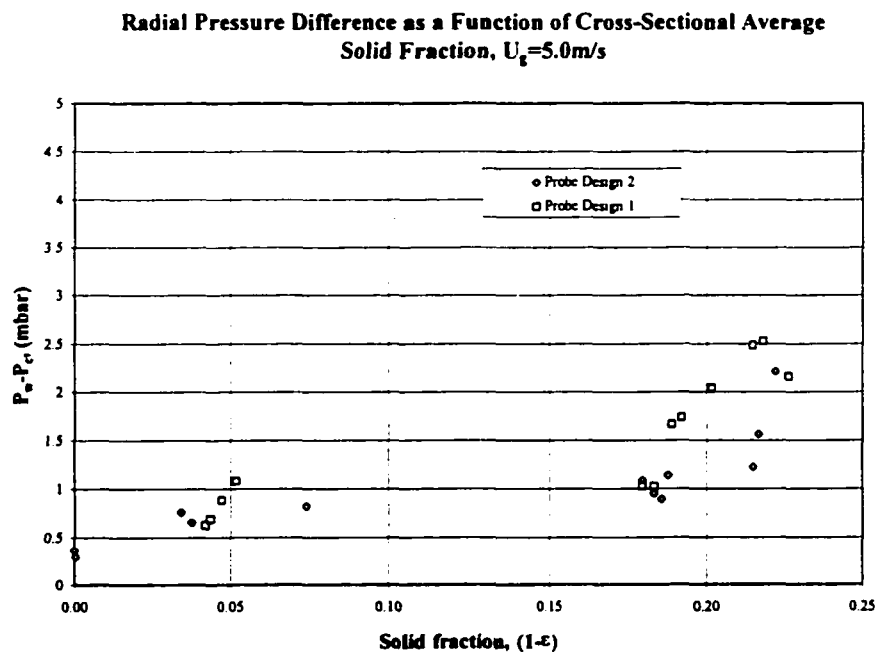


Figure 2.10: Radial Pressure Difference in a Fast Fluidized Bed as a Function of the Cross Sectional Average.

agree with the conclusions from the local solid flux measurements. When the local solid fluxes decrease the radial pressure difference decreases.

2.3 Mixing in Fast Fluidized Beds

The behavior of gas backmixing in a fluidized bed has received considerable attention since it exerts great influence on chemical reactor performance. Understanding of gas and solid mixing for fast fluidized beds is thus significant for practical design.

The presence of gas backmixing was of interest to many investigators in earlier studies. Cankurt and Yerushalmi (1974), who first studied gas backmixing in high velocity fluidized beds, concluded that gas backmixing is not significant. However recent studies proved the opposite.

Li and Weinstein (1989), used helium tracer to determine gas backmixing in the turbulent, fast and pneumatic transport fluidization regimes. The results for turbulent fluidization showed that gas backmixing is taking place through out the cross section of the bed. In turbulent fluidization is seem that even in the center of the bed gas backmixing is considerable. Figure 2.11 shows gas backmixing in the fast fluidization regime. It can be observed that backmixing is high in the dense annulus (higher than in the turbulent regime) and negligible in the dilute core. However in the transport regime, there still exists a significant amount of gas backmixing near the wall even through it is much lower than in the other regimes. The dilute core in this case is extended and the region near the wall is smaller.

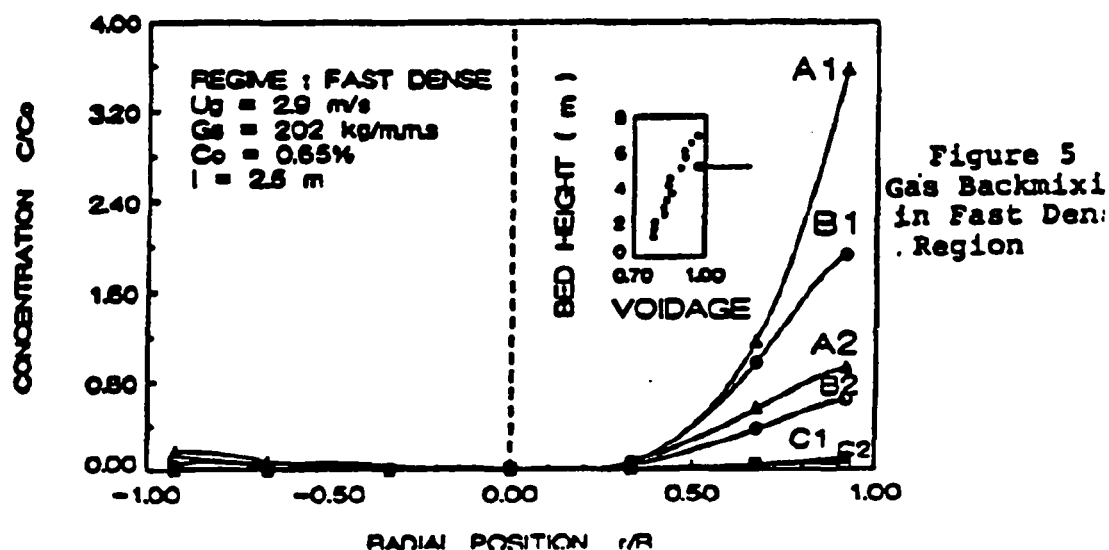


Figure 2.11: Gas Backmixing in a Fast Fluidized Bed (Li and Weinstein 1989).

The above results prove that gas backmixing exists in a fast fluidized bed although it is lower with that in a turbulent bed. In other experiments Wu (1988); Li and Wu (1991), found the existence of gas backmixing in a high velocity fluidized bed. They also found that gas backmixing increases with the solid circulation rate.

The difference in the conclusions of these studies lies in the fact that earlier studies tried to determine gas backmixing in a high velocity fluidized bed as a time dependent phenomenon. This is valid for bubbling and slugging fluidized beds where the bed structure is time dependent, but in fast fluidized beds the recent results show that gas backmixing results from the spatial dependence and not the time dependence.

The mechanism of gas mixing in a fast fluidized bed involves:

1. Gas dispersion caused by radial gas velocity distribution;
2. Solid dispersion caused by local solid circulation;

3. Gas mixing caused by interchange of solid between clusters and their surroundings with rapid succession of cluster formation and dissolution.

However contrary to bubbling fluidization, the major gas flow in fast fluidization rather than the discontinuous bubble phase, is in the continuous core phase. Therefore, it is reasonable to treat the gas mixing in terms of a pseudo-homogeneous dispersion model. On the other hand experimental results (Adams, 1988; Brereton et al., 1988) have indicated that the radial dispersion is small compared to the axial dispersion. Consequently, gas mixing for fast fluidization can be described by the following one-dimensional pseudo-homogenous dispersion dimensionless equation:

$$\frac{\partial C}{\partial \theta} = \frac{1}{Pe} \left[\frac{\partial^2 C}{\partial \xi^2} - \frac{\partial C}{\partial \xi} \right] \quad (2.2)$$

with boundary conditions:

$$C(0, \theta) = C_1(0, \theta); \quad (2.3a)$$

$$C(\xi, 0) = 0; \quad (2.3b)$$

$$C(\infty, \theta) \neq \infty \quad (2.3c)$$

Figure 2.12 shows the influence of bed voidage on D_r/UL for FCC catalyst. It can be seen that D_r/UL decreases with increasing bed voidage and superficial gas velocity, and with decreasing in solid circulation rate.

Typical results of radial dispersion measurements, in a 115 mm I.D., 8m high riser are presented in Figure 2.13. Experimental results show that the radial dispersion coefficient increases with solid circulation rate and with bed density, and decreases, though slightly, with gas velocity. Experiments are yet inconclusive as to the dependence

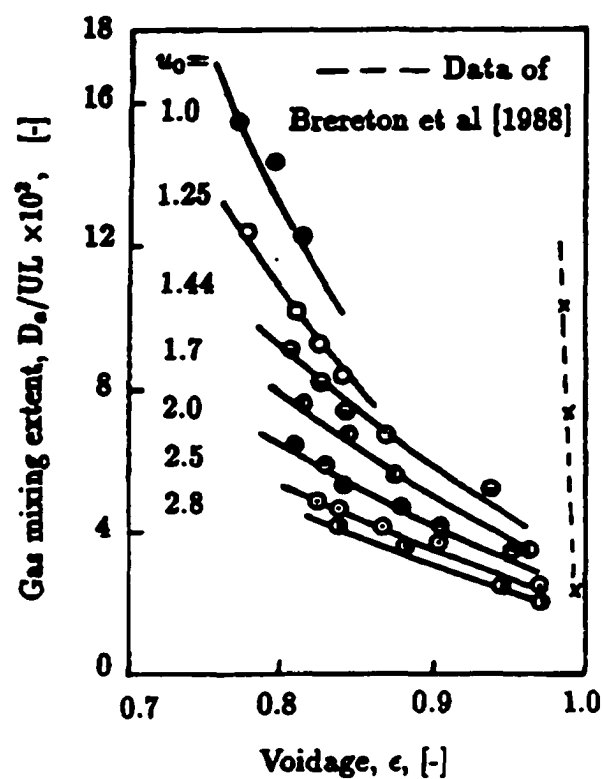


Figure 2.12: Gas Backmixing Extent as a Function of Voidage (after Li and Wu 1991).

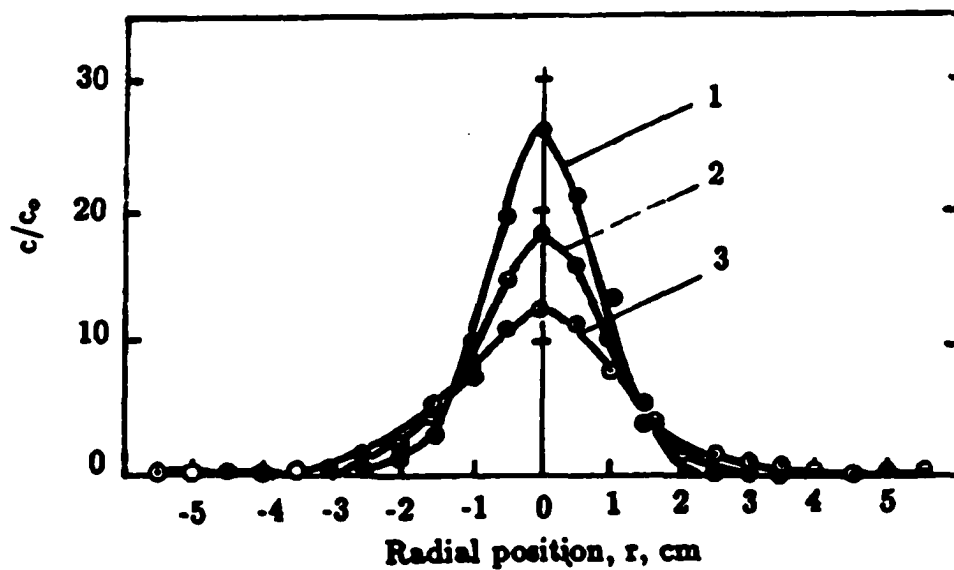


Figure 2.13: Radial Profile of Tracer Concentration, $U_f=4.1$ m/s, $G_r=(1) 43.5$; (2) 74; (3) 406 kg/m²s, (after Yang et al., 1984).

of D_r on riser diameter.

2.4 Modeling

Many efforts have been made in order to develop a mathematical model which will predict the behavior of a fluidized bed.

J.D. Murray (1963), assuming a continuum flow for both particles and fluid in a bubbling fluidized bed, developed an isothermal model consisting of the mass conservation and momentum equations that govern the system. He assumed that the bed has n identical particles of mass m and volume τ . For this system the conservation equations are:

$$\nabla n \cdot \mathbf{v}_s = -\frac{\partial n}{\partial t} \quad \text{for the particles} \quad (2.4a)$$

$$\nabla(1 - n\tau) \cdot \mathbf{v}_f = \tau \frac{\partial n}{\partial t} \quad \text{for the fluid} \quad (2.4b)$$

where: $\mathbf{v}_s$, $\mathbf{v}_f$ are the average velocities of the particles and fluid respectively over some volume large compared with the particle size.

In order to develop the momentum equations, he assumed that the drag coefficient is given by the experimental correlation by Rowe (1961), collision forces are negligible, and that the remaining forces are given from the inviscid case of an induced mass in the flow past a sphere. With these assumptions the momentum equations become:

$$n \frac{d\mathbf{v}_s}{dt_s} = -ing(m - \tau\rho) + D(n)(\mathbf{v}_f - \mathbf{v}_s) + An\tau\rho \frac{d}{dt}(\mathbf{v}_f - \mathbf{v}_s) \quad \text{for the particles} \quad (2.5a)$$

$$\rho(1 - n\tau)\frac{dv_f}{dt_f} = -ig\rho - D(n)(v_f - v_s) - \nabla P - An\tau\rho\frac{d}{dt}(v_f - v_s) \quad \text{for the fluid} \quad (2.5b)$$

The solution of the above equations gives results that agree quantitatively with the experimental observations when fully developed bubbles rise on a fluidized bed, but the model cannot predict the shape of the bubbles.

Batchelor K.G. (1988) formulated the equations governing the unsteady one-dimensional mean motion of the particles in a gas- or liquid- fluidized bed without including any major untestable hypotheses. The final form of these equations include five parameters which depend primarily on the void fraction and secondarily on the particle Reynolds number and the ratio of the particle density to the fluid density. The results agree with the experimental data but there is lack of information regarding some of these parameters.

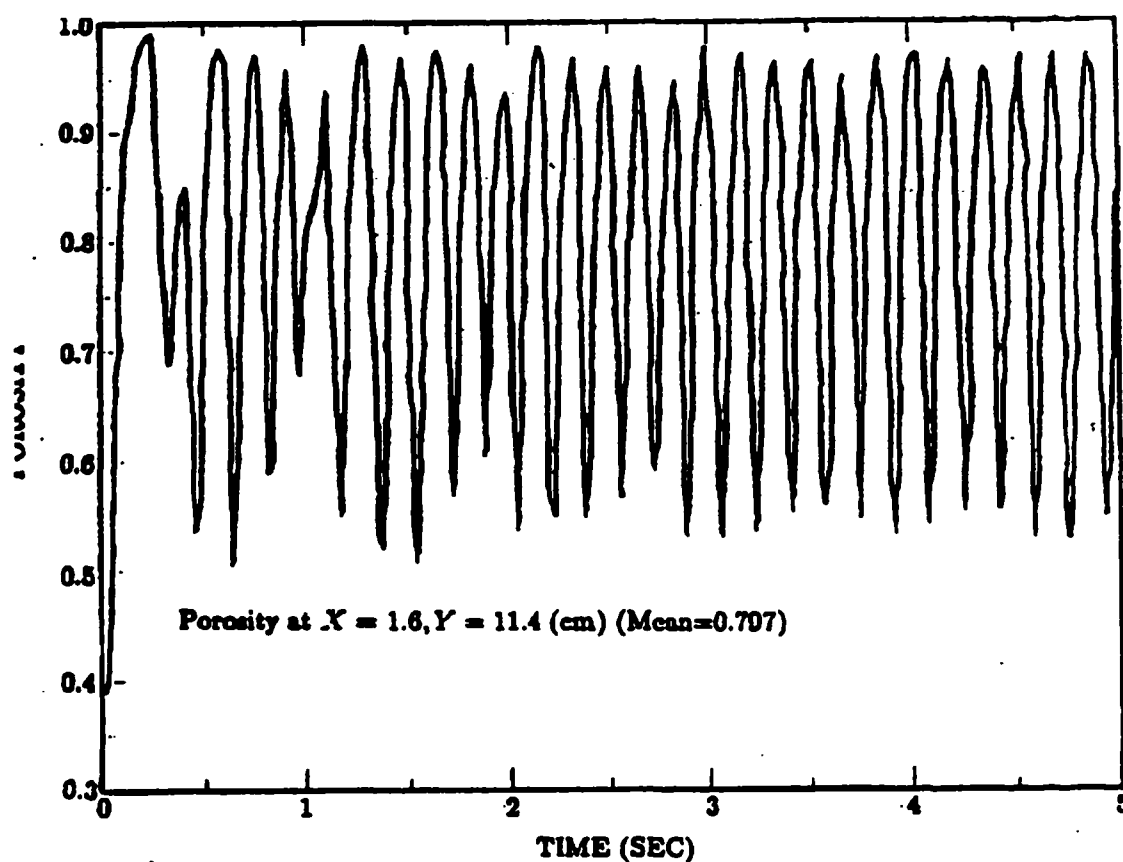
J. Ding and D. Gidaspow (1989), using the kinetic theory of granular flow and the Boltzmann equation for the distribution of the particle velocity, developed a two-dimensional model for bubbling fluidized beds. This model predicts correctly the void fraction oscillations and the formation of bubbles and the solid flow patterns for fluidization. Figure 2.14 shows the computed and calculated time dependent void fraction in a bubbling fluidized bed and Figure 2.15 shows the corresponding power spectral analysis. It can be seen that the two signals are similar. The characteristic frequency of the computed data is slightly smaller than that of the experimental data because the model does not take into account the boundary effects.

J.Sanyal and E. Cesmebasi (1994) developed a model for bubbling fluidized beds and used three different correlations for the drag coefficient, to examine the influence of

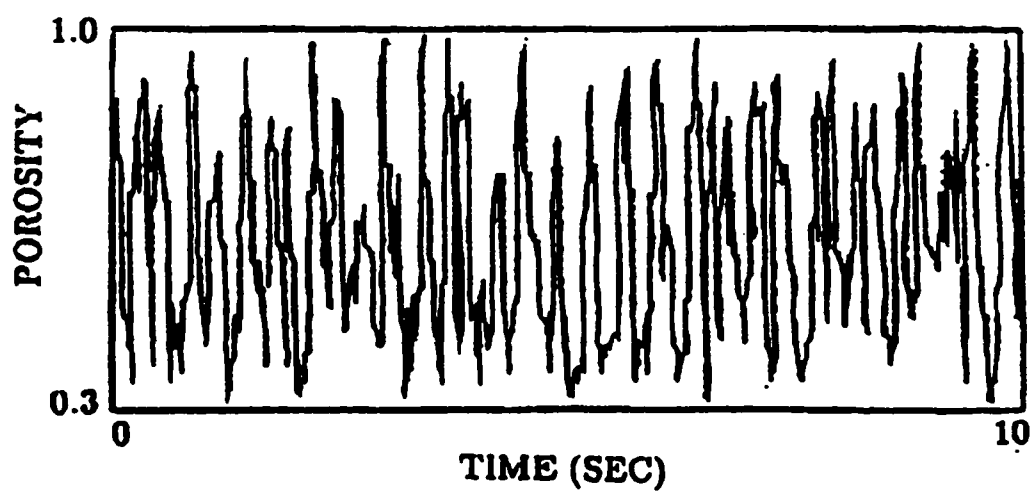
these correlations on the final results. The drag laws that they used were developed by Ma and Ahmadi (1989), Ergun (1952) Richardson and Zaki (1954), and Syamlal (1990). The results showed quite different behavior for long times.

Other investigators developed mathematical models for fully developed gas-solid flows in vertical tubes. Sinclair and Jackson (1989), pointed out that in a gas-solid flow both the fluid and the particle velocities have local average and random components. The interaction between these two components of the particle velocity generates stresses in the particle assembly. They developed a mathematical model for such flows taking into account only this interaction and neglecting the turbulence that exists in these flows. The results showed a quantitatively agreement with the experimental data. Louge (1990) tried to improve this model by inserting the effect of the turbulence. The results were closer to the experimental data.

J.A.Pita and S.Sundaresan (1991) using the above models examined the influence in the results of the inlet parameters (geometric characteristics, particle size, countercurrent or cocurrent flow). They compared the computed results with data by Bader et al. (1988). They found that the computed data were close enough to the experimental data but there was a large degree of sensitivity of the model to the inelasticity of the particle-particle collisions and the damping of particle-phase fluctuating motion by the gas.

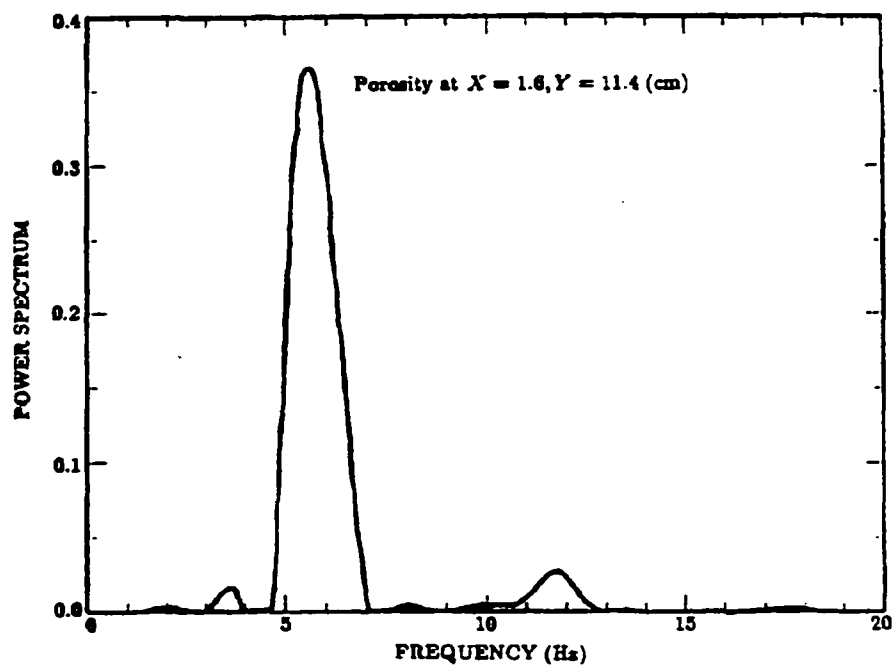


(i) Computation

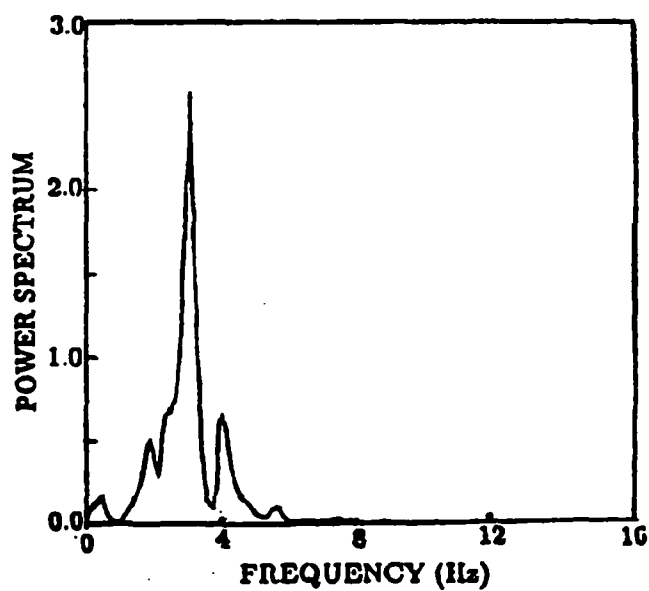


(ii) Experiment

Figure 2.14: Computed and Experimental Void Fraction in a Bubbling Bed, (J.Ding and D.Gidaspow, 1989).



(i) Computation



(ii) Experiment

Figure 2.15: Power Spectral Analysis of Computed and Experimental Void Fraction in a Bubbling Bed, (J.Ding and D.Gidaspow, 1989).

III. OBJECTIVES

From the literature review it is evident that there are many disagreements and difficulties in describing the hydrodynamics of a fast fluidized bed. There are also different experimental apparatuses and techniques which affect the results which have been reported. Nevertheless some general conclusions can be obtained.

A fast fluidized bed is characterized by a dense bottom region and a dilute top region and also dilute core and dense annulus regions. The boundaries of these regions depend on gas velocity and solid circulation rate as well as on the imposed pressure drop. Increase of the gas velocity or decrease of the solid rate or decrease of the imposed pressure drop results in the lowering of the transition point between the dense bottom and dilute top and also makes the radial density profile flatter. Geometric characteristics of an experimental apparatus such as gas inlet configuration, solid return, outlet configuration, etc., may play important roles in the flow patterns developed. Different configurations have to be examined in order to determine if they affect the measurements. It appears that there is a radial pressure difference along the diameter of the bed. Pressure is lower in the center of the riser. This pressure difference supports the dilute core-dense annulus structure. The radial pressure difference increases with the average cross sectional solid fraction of the bed.

Since each fast fluidized bed application requires a different degree of gas backmixing for optimum performance, gas backmixing is an important parameter. Gas backmixing takes place mainly in the dense annulus of a fast fluidized bed. It increases

with a decrease of gas superficial velocity and/or an increase of the solid circulation rate. Gas backmixing may also be dependent on the geometric characteristics of the bed.

Many models have been developed to predict the behavior of any kind of fluidized bed. From the literature it can be seen that all models are sensitive to the selection of the drag law even for bubbling fluidized beds. Selection of different drag correlations give quite different results. On the other hand modeling of high velocity fluidized beds has been focused in the fully developed flow region. Commercial fast fluidized bed reactors usually have a height between 20-30 L/D. Therefore there is actually no fully developed flow region. Modeling must take into account the inlet effects in order to predict what happens in the critical acceleration zone.

The scope of this work is to examine experimentally how different entrance configurations affect this initial riser section (0-20L/D). The main characteristics of a fast fluidized bed reactor are the geometry of the entrance and the gas nozzle positions. For this reason a new inlet section has been constructed and installed in the City College Fast Fluidized Bed Facility which provides the ability to change nozzle positions and gas inlet flowrates. With this new entrance region measurements were made with various entrance configurations and operating conditions (superficial gas velocity, solid circulating rate and relative nozzle flowrates for the same gross superficial gas velocity). These experiments can give a complete set of accurate and internally consistent data. The measurements include axial and radial solid distributions, and radial pressure and local solid flux profiles. The axial solid and gas velocity distributions are estimated from the measured quantities. The effect of the pressure tap spacing on the pressure measurements

is also examined. This kind of measurement can give an indication of the axial scale of any large flow structures. The measurements were carried out at selected elevations in the riser in order to cover the entire entrance section. Thus, a complete set of data is available. The data can serve as guidance for improved chemical reactor design and also for the development and testing of two dimensional CFD models which can describe the entrance flow nature of a riser reactor.

IV. EXPERIMENTAL PROCEDURE

4.1 Fluidized Powder

The solid used in this work was regenerated FCC catalyst supplied by Phillips Petroleum. The catalyst is classified as Geldart Group A powder. Table 4.1 shows the properties of this solid. Figure 4.1 shows the particle size distribution of the catalyst.

Properties	
Mean Diameter, $d_p(\mu\text{m})$	71.49
Median Diameter, (μm)	72.36
Maximum Diameter, (μm)	180.7
Minimum Diameter, (μm)	5.27
Density, $\rho_p(\text{kg/m}^3)$	1,450
Sphericity	1

Table 4.1: Catalyst Properties.

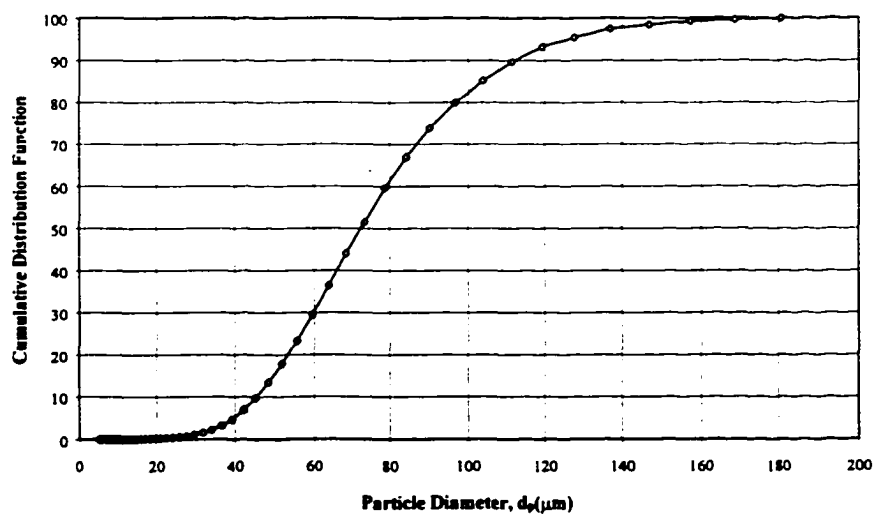


Figure 4.1: Particle Size Distribution of the Equilibrium FCC Catalyst.

4.2 Experimental Apparatus

The modified City College facility for fast fluidization was used for all the experiments of the present investigation. A schematic of this system is shown in Figure 4.2. It consists of two sections, a riser or "fast bed" on the left side of Figure 4.2 and a downcomer or "slow bed" on the right side. The riser, made of Plexiglas™, has a total length of 8.50m and an I.D. of 15.20 cm. A new entrance section in the lower part of the riser, includes a contraction of I.D. of 8.25 cm and a length of 6 cm as is shown in Figure 4.3. The expansion section has a length of 11 cm. Four nozzles of 2.54 cm I.D. each are symmetrically located after the expansion section. The nozzles can be moved to any radial position. Their axes are parallel to the axis of the riser. The reduction section also has a length of 11 cm and is located at the end of the U-bend section which connects the riser with the slow bed. The U-bend section is equipped with five nozzles (three independent lines) and a butterfly valve, V_1 which is used for the control of the solids circulation rate (Figure 4.3). The slow bed is also made of Plexiglas™ and it has an I.D. of 34.30 cm. The wall thickness of all the sections of the circulating system is 7 mm.

The solid particles are carried from the bottom to the top of the riser which terminates in a 90° Plexiglas™ elbow and enter the primary cyclone. The primary cyclone is in the top of the slow bed. The solids separated in the primary cyclone fall directly into the slow bed which is maintained at bubbling fluidization conditions. The air leaving the primary cyclone enters a secondary cyclone. The separated solids from this cyclone also return to the slow bed through a dipleg and the fines are collected in a bag filter through which the air leaves the system. The slow bed is equipped with a butterfly

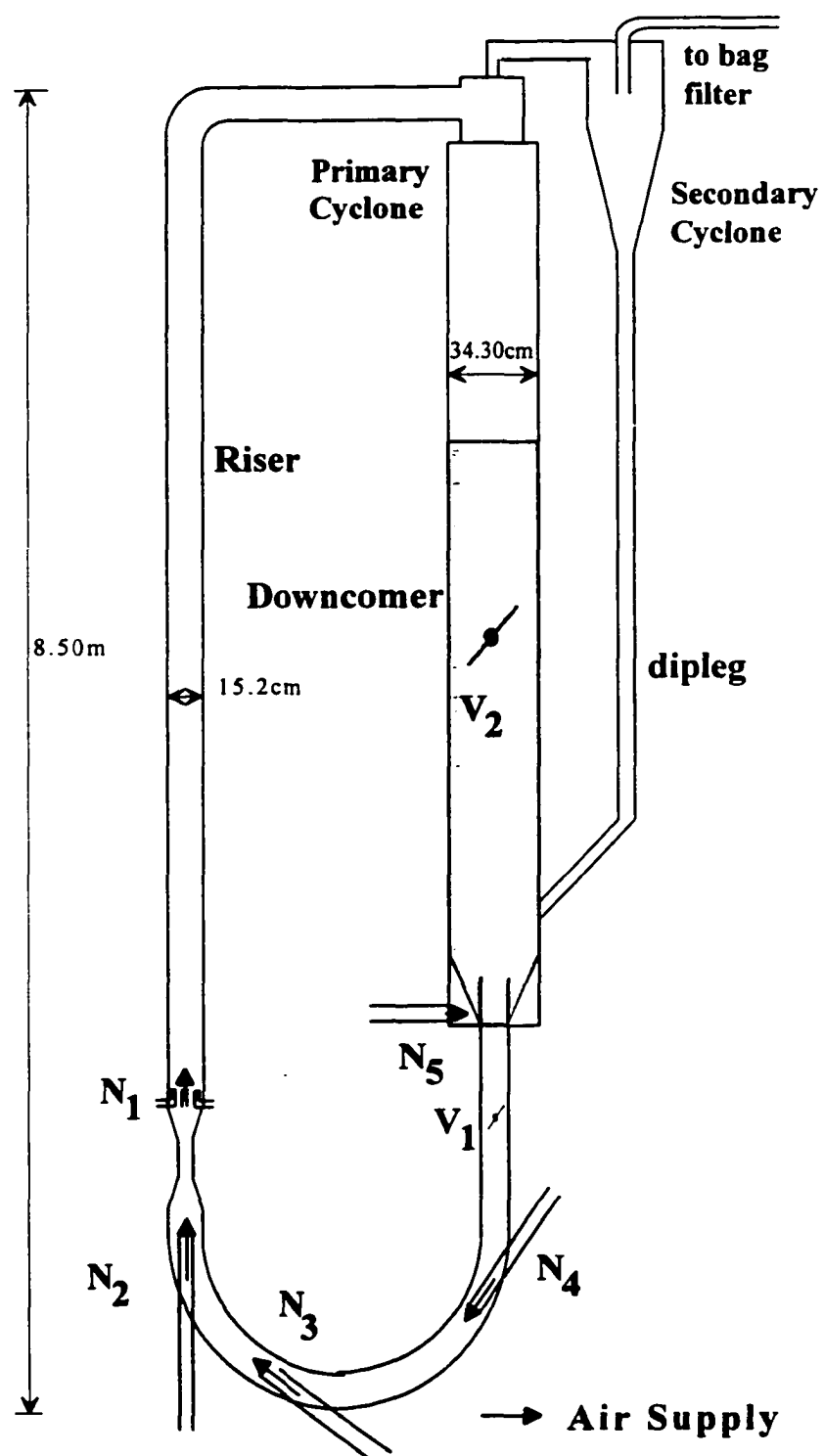


Figure 4.2: City College Fast Fluidization Facility.

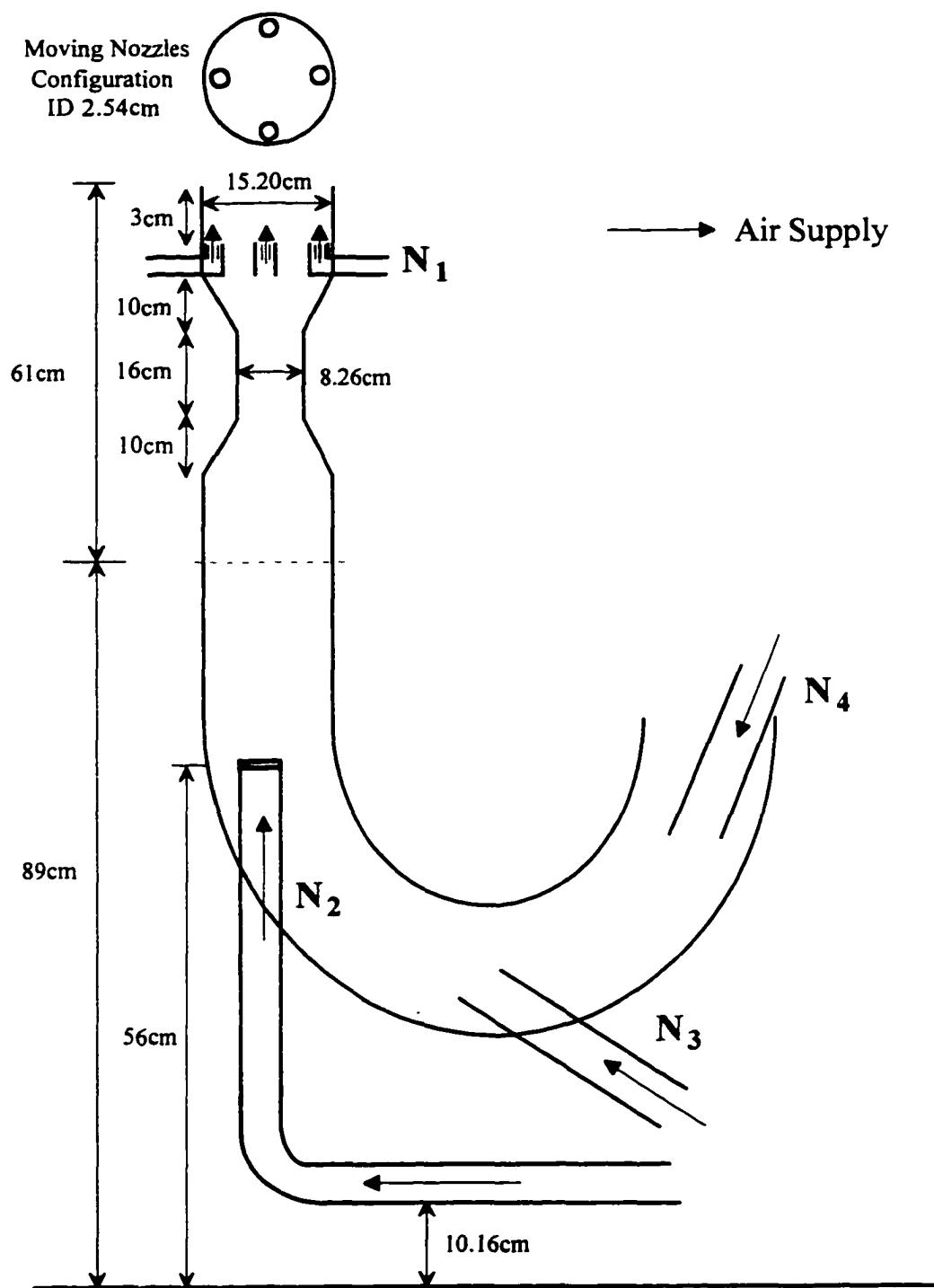


Figure 4.3: Air Nozzle Configuration.

valve V_2 , which is used for the measurement of the solid circulation rate.

Air is provided to the system by a INGERSOLL-RAND centrifugal compressor with output of $85\text{m}^3/\text{min}$ at 1.4 atm gauge. Air enters the bed through five independent lines N_1 , N_2 , N_3 , N_4 and N_5 as is shown in Figure 4.2. The last line, N_5 provides the air required to keep the slow bed at bubbling fluidization conditions, ($U_g \approx 5\text{cm/s}$). Figure 4.3 shows the configuration of the other four lines that provide air to the riser system. The main air feed is provided by the N_1 line through four tubes symmetrically located about the axis of the column. These four tubes can be moved along radii of the riser. The air provided to each of these four tubes is controlled with a separate flowmeter, so each tube receives the same volumetric flowrate of air. The air that enters the system in the U-bend section (N_2 , N_3 and N_4) is secondary air providing the necessary aeration for the flow of the solids up to the main air supply. This configuration is a mockup of the entrance section of a commercial FCC unit and provides the capability to change not only the nozzle position but also the flow split between primary and secondary air at same gross superficial gas velocity.

The butterfly valve V_1 in the beginning of the U-bend section is used to control the solid circulation rate, G_s . The second butterfly valve, V_2 , in the middle of the slow bed is used for the measurement of the solid circulation rate. This valve is made of a sintered plate, so when it is closed it has a negligible influence on the pressure balance of the system. When the valve is closed the solid fluidize above the valve and the bed below is carried away. Measuring the pressure drop as a function of time in this section below the

valve provides the slope which is proportional to the mass of the fluidized solid leaving in the storage bed per unit time.

$$G_s = \frac{A_d}{A_r} \frac{1}{g} \frac{d(\Delta P)}{dt} \quad (4.1)$$

where: G_s is the solids circulation rate, $\text{kg/m}^2\text{s}$,

A_d is the cross-sectional area of the slow bed (downcomer),

A_r is the cross sectional area of the riser.

A typical signal of this pressure measurement is shown in Figure 4.4. The error in the solid circulation rate measurement by this method has been estimated as less than 5%, (Schnitzlein and Weinstein, 1988).

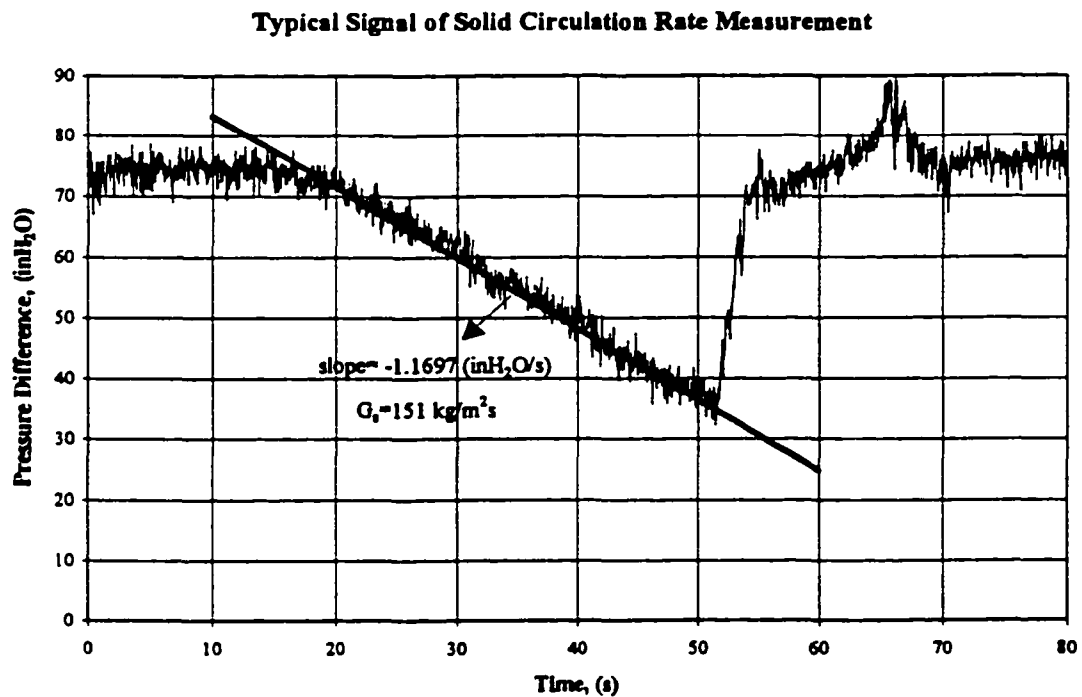


Figure 4.4: Typical Signal of Solid Circulation Rate Measurement.

4.3 X-Ray System

The circulating fluidized bed is equipped with an industrial x-ray system which is used for the measurements of the radial solid fraction. The x-ray head source is a Norelco MG 150/300 generator. The head that is used in this investigation is 150 kV with a maximum level of 30 mA at 100 kV and 20 mA at 150 kV. The x-ray tube is a fixed anode type with two focal points of 15 and 40 mm. The latter focal spot is used for the measurements because it provides a wider angle of x-rays. A schematic representation of the x-ray setup is shown in Figure 4.5. The x-ray image of the bed is projected onto a Phillips image intensifier (22.9 cm in diameter), which converts the x-ray image to visible light. The visible image from the intensifier is distributed by a mirror mounted with an

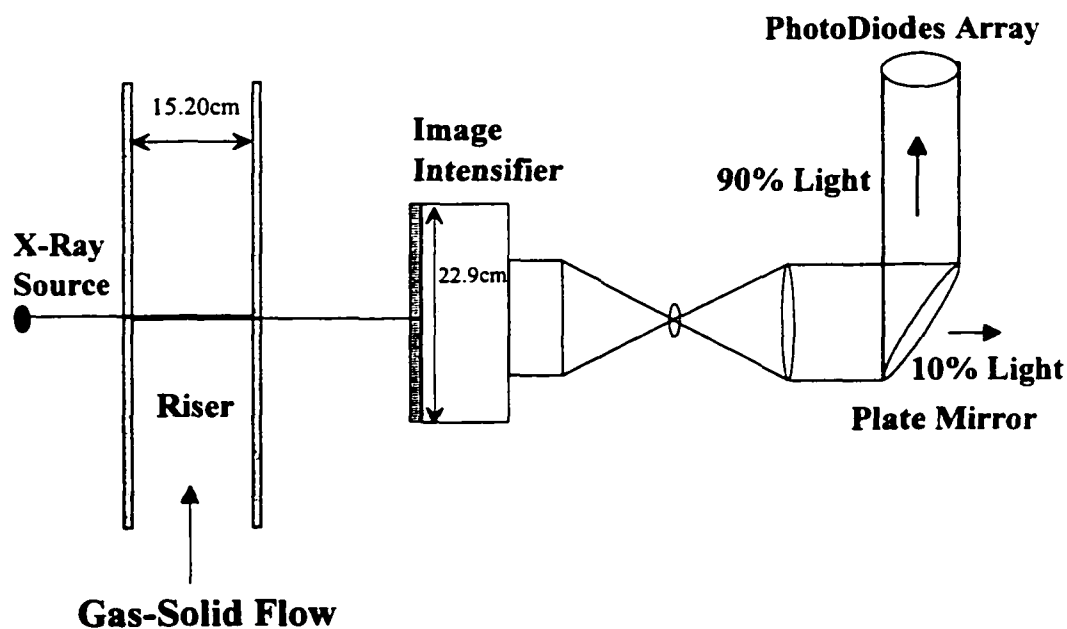


Figure 4.5: X-ray Imaging System.

angle of 45° . 10% of the light is monitored by a closed circuit T.V. camera connected to a monitor and video tape recorder in the control room. The remaining 90% of the image light is projected through lenses onto a 15.2 cm Plexiglas screen which is equipped with fifteen photodiodes. The photodiodes convert the optical signal (light), to an electrical signal (voltage). The response time for each photodiode is $1\ \mu\text{s}$. The output of each photodiode goes to an amplifier circuit which provides a low noise to signal ratio. The amplification of each diode output is 1000 as it is shown in Figure 4.6. The output of the amplifier circuit is in the range 0-10V and is hardwired to a real time computer system (Gateway 2000, 4DX2-50V) equipped with a Keithley Metrabyte DAS-20 (12 bit), analog to digital converter board.

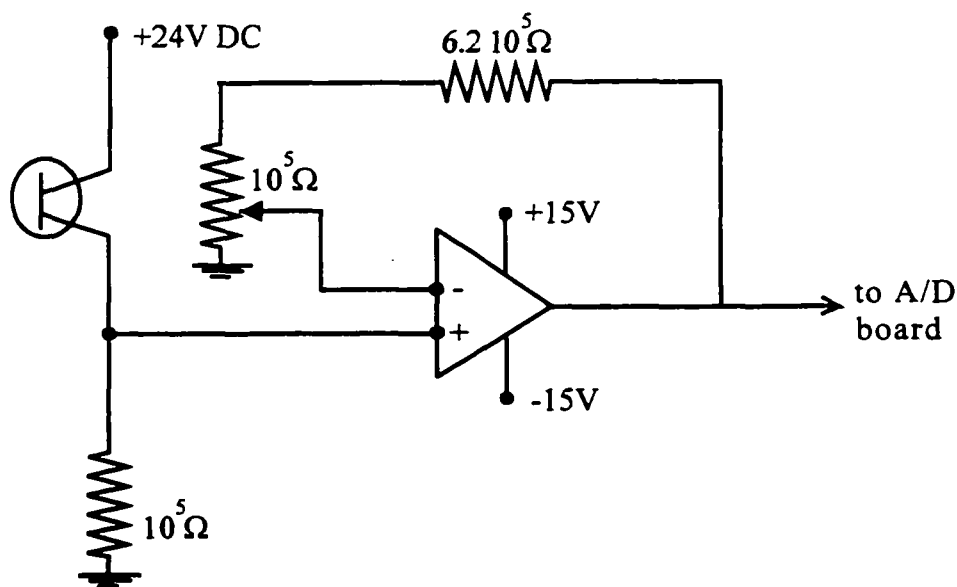


Figure 4.6: Amplifier Circuit.

4.4. Experimental Techniques

The quantities that are measured and statistically analyzed in this investigation are the axial and radial pressure differences, the radial profile of solid fraction, and the radial profile of the axial solid flux.

4.4.1. PRESSURE MEASUREMENTS

The circulating system of Figure 4.2 has 25 pressure taps for the axial differential pressure measurements. Instantaneous axial pressure differences are important measurements in a fluidized bed because the statistical analysis of a long record length gives additional information on the fluidization as well as the mean value which provides an apparent cross sectional average solid fraction.

$$\frac{(dP)_{avg}}{L} = \rho_s(1 - \epsilon)_{apparent}g \quad (4.2)$$

Figure 4.7 is a schematic of the pressure measurement setup. Pressure probes are straight tubes of 4.8 mm I.D. which terminate flush with the inside wall of the riser. The two probes that use for any pressure measurement are supplied with purge gas (air). The air for the probes is provided by a separate compressor with flowrate up to 5m³/min at 5 atm. Purge gas must be dry and clean in order to maintain the pressure transducer in good condition. The difference in the backpressures of the purge gas is measured by the pressure transducer which transforms this difference into a voltage. The transducer has to be as close as possible to the probes so that damping of the pressure signal in the tube that connects them is negligible. The transducers that are used are MicroSwitch pressure

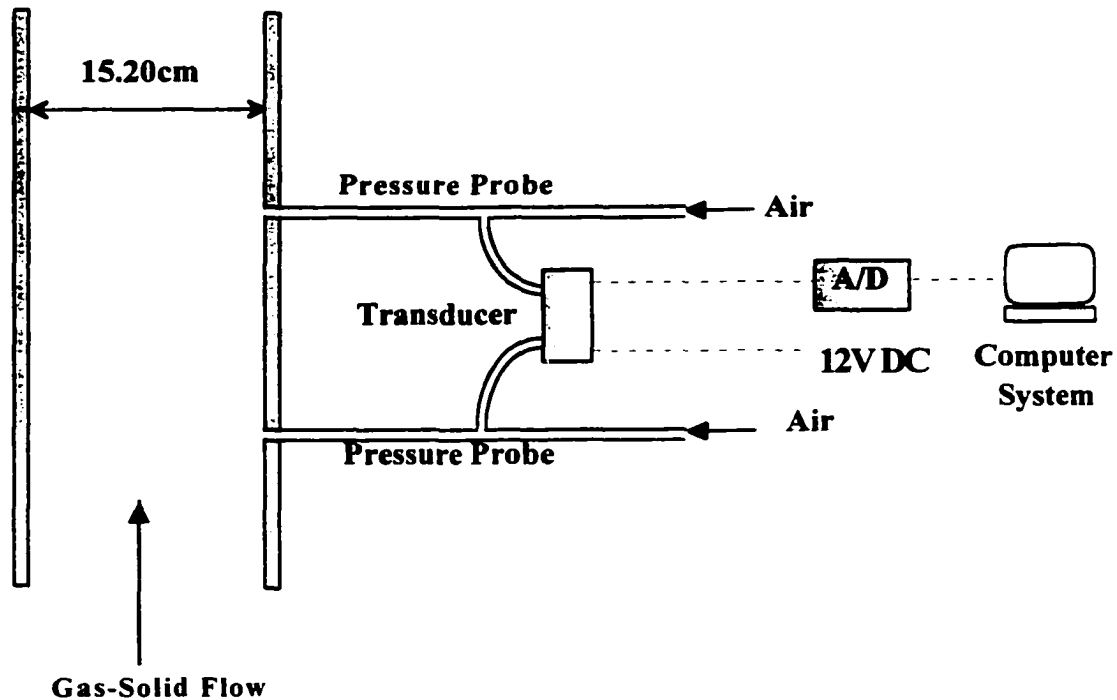


Figure 4.7: Axial Pressure Difference Measurement Setup.

transducers with full scale range 0-1, 2 or 5 psi. Each transducer is supplied with 12V DC. The outlet from the transducer has a range 0-10V DC and is hardwired to a real time computer system (Gateway 2000, 4DX2-50V) with a Keithley Metrabyte DAS-20, 12 bit (full scale $2^{12}=4096$ bytes), data acquisition board, which converts the analog signal (Volts) to a digital signal (bytes). Finally, using the calibration data of each transducer this digital signal is transformed to pressure difference data.

The radial pressure differences are measured with two different pairs of probes. Figure 4.8 shows the two different probe designs. The first pair (Probe Design 1) consists of two straight tubes of 3 mm I.D. The two probes are mounted at the same elevation in the riser. One of them is mounted at the wall of the riser and the other traverses along the

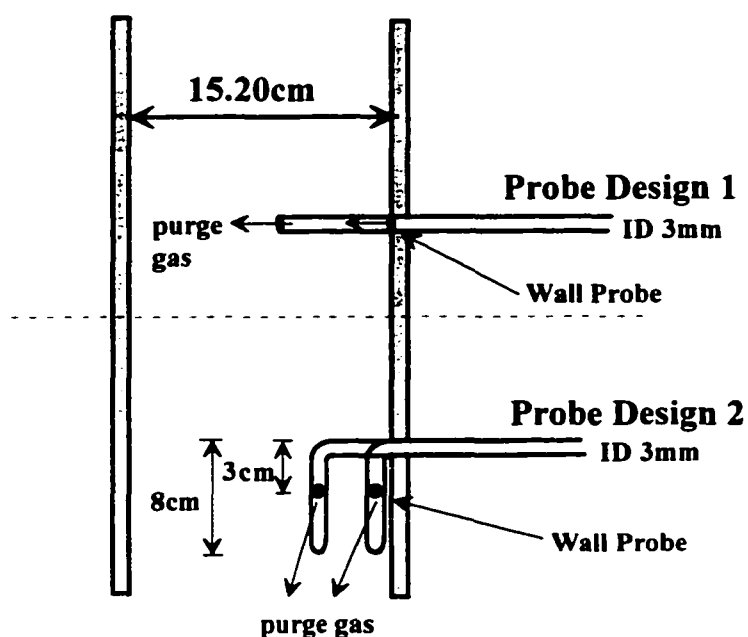


Figure 4.8: Radial Pressure Difference Measurement Setup.

diameter of the bed. The second pair of probes (Probe Design 2) have a vertical section, 8 cm long, facing downward from the horizontal. The purge gas ports are located in the vertical section as in a typical static pressure probe. The cross sectional area for flow of purge gas of the two designs is the same. The reason that two different probes are used for these pressure measurements is to examine the influence of the probe design on the measurements. Probe Design 1 was used in earlier studies by Weinstein et al. (1993). Probe Design 2 is designed to have less influence of the flow field in the riser.

The calibration of each pressure transducer was carried out using a very sensitive DWYER micromanometer. Typical calibration curves of three transducer types are shown in Figure 4.9.

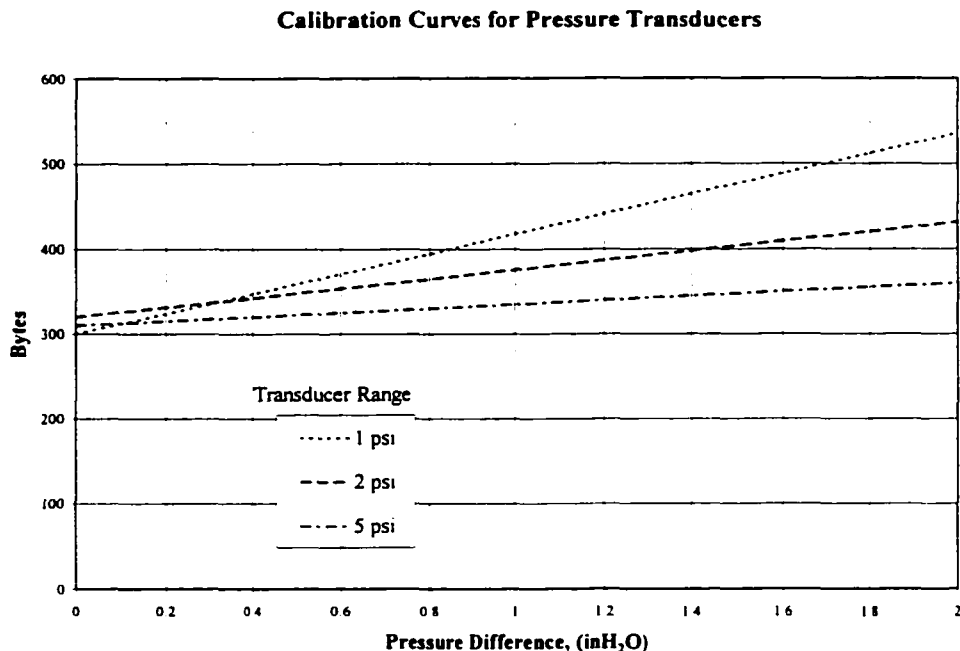


Figure 4.9: Calibration Curve for Three Different Range Pressure Transducers.

The purge gas velocity in the probes is a critical parameter for this measurement. The air velocity in the probe can not be too low because solid will enter the probe and then the transducer. Figure 4.10 shows the pressure difference between two elevations as a function of the purge gas velocity and the transducer range. It can be seen that the air velocity in the probe must be above 0.35m/s and that the variation from transducer to transducer is small compared to the actual measured pressure difference.

Another measurement parameter is the total measuring time or record length which will give a time independent mean pressure difference. Figure 4.11 shows the r.m.s value as a function of measuring time. It can be seen that for record lengths of 60s and longer, the variation of the calculated mean pressure difference is much less than 1%. Axial and radial pressure differences in this investigation are obtained from a signal of 8,000 points acquired at 100 Hz frequency.

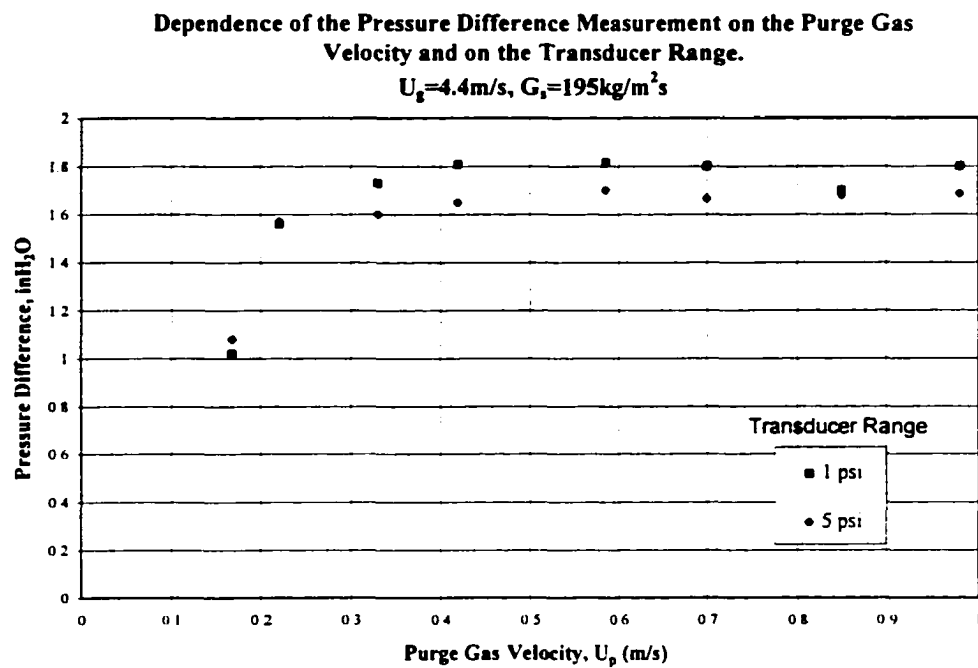


Figure 4.10: Pressure Difference as a Function of the Purge Gas Velocity and the Transducer Range.

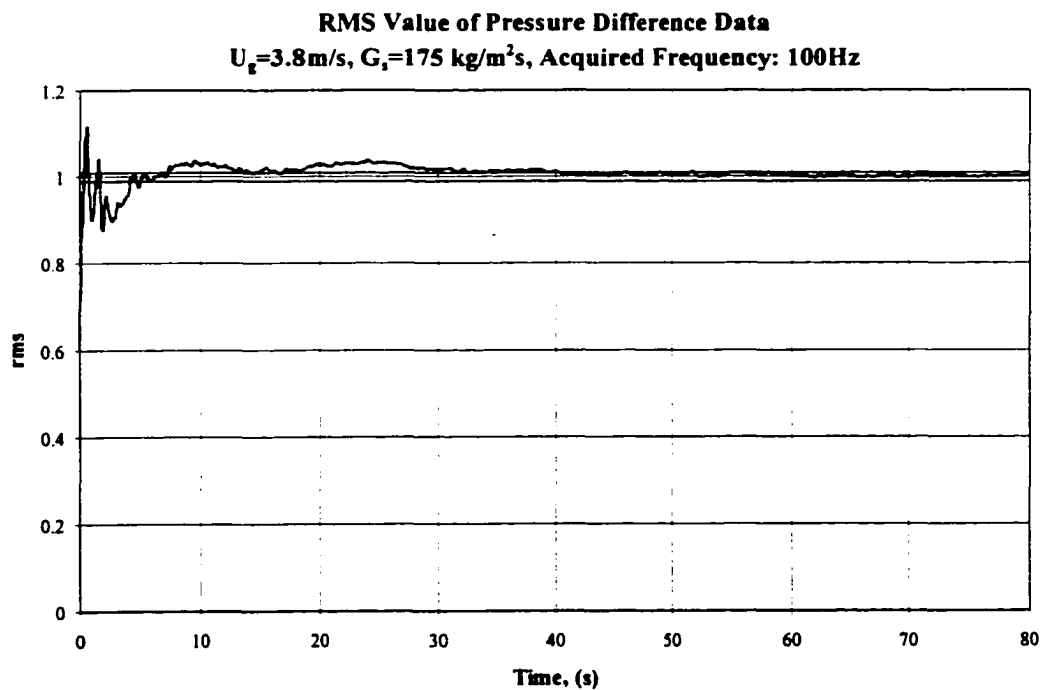


Figure 4.11: Dependence of the Mean Pressure Difference on the Record Length.

4.4.2. RADIAL SOLID FRACTION MEASUREMENTS

The radial solid fraction at different elevations in the riser is obtained using the x-ray system. The x-ray source provides a polychromatic beam of high energy. Filtering is used to cut the lower frequencies so that the final beam has a narrow energy spectrum and is considered to be monochromatic. The decrease in the intensity of a monochromatic beam is proportional to the thickness of the absorbing material and can be expressed by the Beer-Lambert law as:

where: I_0 is the initial intensity of the beam,

x is the length of the absorbed material (cm), and

$$I = I_0 e^{-\mu x} \quad (4.3)$$

μ is the attenuation coefficient, (cm^{-1}).

It has been shown that μ is a function of the material density. So:

where: k is the mass attenuation coefficient (cm^2/kg), and

ρ is material density, (kg/cm^3).

$$\mu = k\rho \quad (4.4)$$

As the beam travels through a material, energy losses are due to absorption and scattering. In order to eliminate the scattering a lead diffraction shield is used before the image intensifier screen (Figure 4.5). The shield is made out of 2.54 cm wide lead strips which are parallel to the x-ray path. The x-rays which pass straight through the absorbing

medium are not influenced by the lead shield. The scattered beams which would strike the intensifier screen at an angle are absorbed by the lead shield.

When the x-rays pass through a fluidized bed (Figure 4.12) applying the Beer-Lambert law gives:

On the other hand, when x-rays pass through an empty bed:

$$\ln \frac{I_{w+s}}{I_0} = -(k\rho)_w x_w - (k\rho)_s x_s \quad (4.5)$$

where: I_{w+s} is the intensity at the image intensifier when the beam passes through a fluidized bed,

$$\ln \frac{I_w}{I_0} = -(k\rho)_w x_w \quad (4.6)$$

I_w is the intensity when the beam passes through an empty bed.

Subtraction of the above equations (4.5) and (4.6) gives:

where ρ_p is the particle density.

$$\ln \frac{I_{w+s}}{I_0} - \ln \frac{I_w}{I_0} = -k_s \rho_s x_s \quad (4.7)$$

$$\rho_s = \rho_p (1 - \epsilon) \quad (4.8)$$

Knowing k , ρ_p and x and measuring the initial and final intensities, the local solid fraction averaged along the chord x_s can be determined. The mass attenuation coefficient, k for the solid was determined from the calibration. Six boxes with the same length of 15.2 cm and

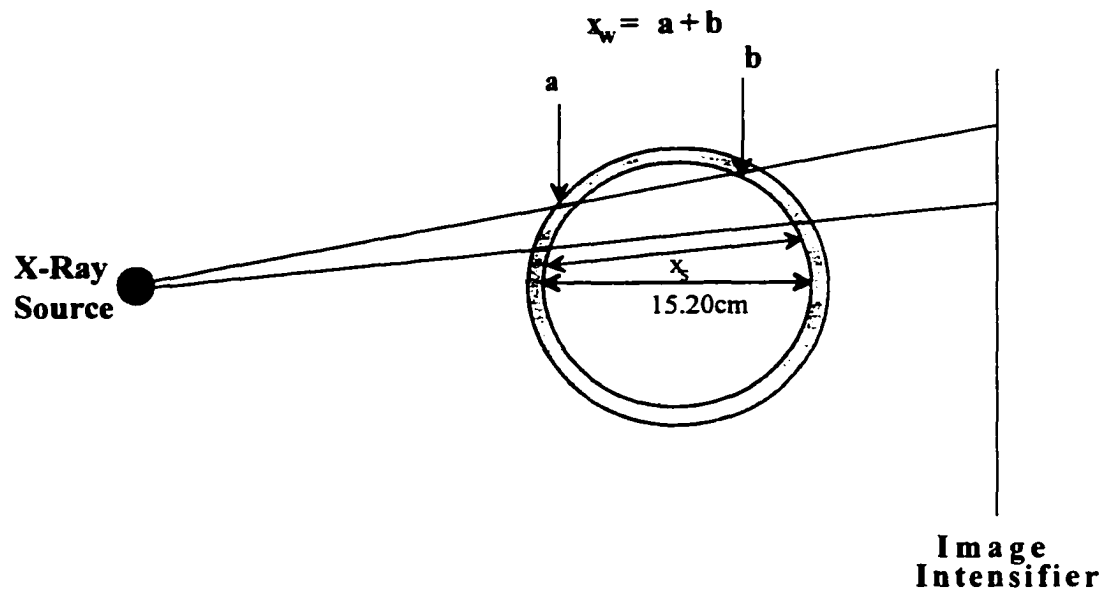


Figure 4.12: X-Ray Measurements of the Chord Average Solid Fraction.

different widths (2.54, 5.08, 7.64, 10.16, 12.7 and 15.2 cm) were used for the calibration. The wall thickness of these boxes is the same as the wall thickness of the riser (7 mm). Measurements with the x-ray were done first having each box empty, to give I_w , and then filled with solid (packed bed), to give I_{w+s} . The density of the solid in the boxes is the density of a packed bed and is determined by measuring the mass of the solid in each box and divided by the corresponding volume. Using the equation 4.7, since the distance is known, the mass attenuation, k_s , can be calculated. Measurements were done for each of the fifteen photodiodes and an average k_s was determined as it is shown in Figure 4.13.

A radial solid density profile can be calculated using the method of image reconstruction, from the chord average x-ray signals. When the beam passes through the bed the Beer-Lambert law gives:

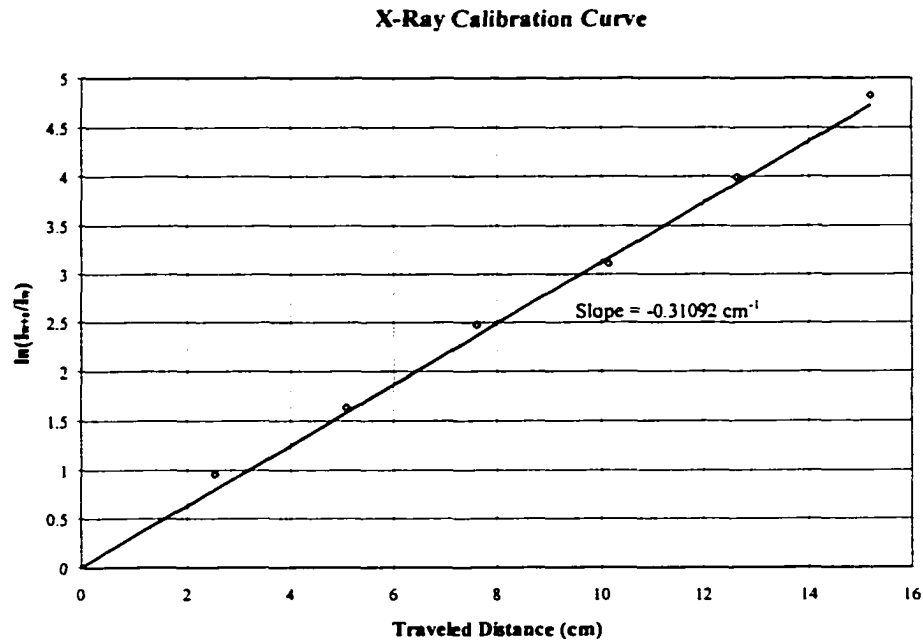


Figure 4.13: X-Ray Calibration Curve of the Equilibrium FCC Catalyst.

where $\rho(x,y)$ is the local solid density.

Assuming axial symmetry, (Figure 4.14):

the radial density profile can be expressed as a polynomial of nth order:

$$\ln \frac{I_{w+s}}{I_0} - \ln \frac{I_w}{I_0} = - \int_A^B k \rho(x, y) ds \quad (4.9)$$

$$\rho(x, y) = \rho_s [1 - \epsilon(r)] \quad (4.10)$$

But:

From equations 4.9-4.13, it follows:

$$r = \sqrt{x^2 + y^2} \quad (4.11)$$

where:

$$[1 - \epsilon(r)] = a_0 + a_1 r + a_2 r^2 + \dots + a_n r^n \quad (4.12)$$

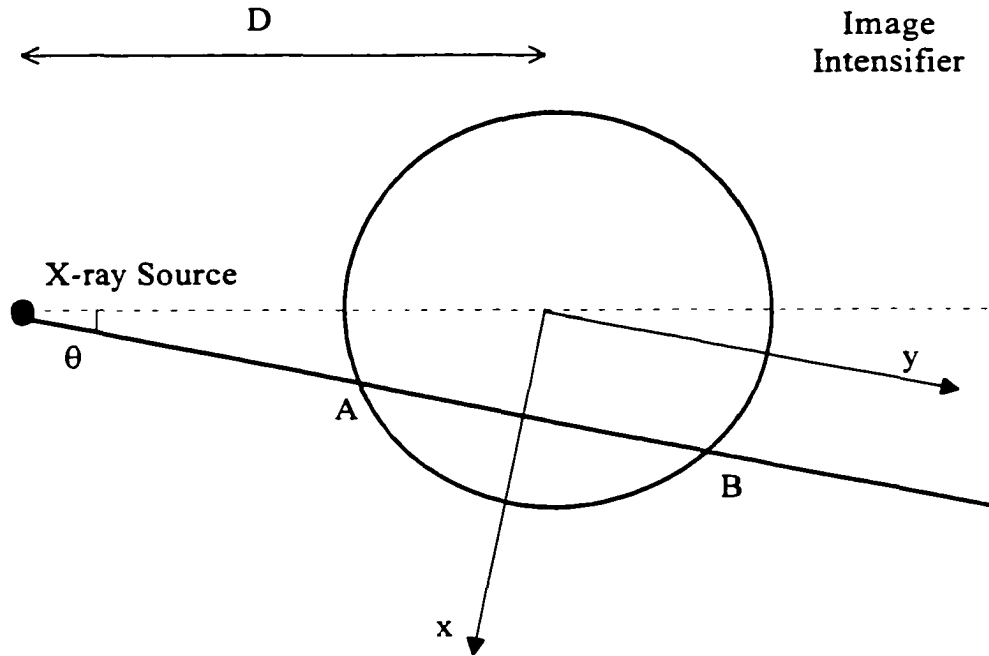


Figure 4.14: Geometry of Reconstruction of Radial Solid Profile.

The left hand side of equation 4.14 is determined from the signals. The right hand side is

$$\left(\frac{\partial \rho(r)}{\partial r} \right)_{r=0} = 0 \Rightarrow a_1 = 0 \quad (4.13)$$

a polynomial of some arbitrary order. For a m th ($m < 13$) order polynomial the above

$$\left[\ln \frac{I_{w+s}}{I_0} - \ln \frac{I_w}{I_0} \right] \left(-\frac{1}{k\rho_s} \right) = \int_A^B [a_0 + a_2 r^2 + \dots + a_n r^n] ds = a_0 P_0 + a_2 P_2 + \dots + a_n P_n \quad (4.14)$$

$$P_0 = \sqrt{R^2 - D^2 \sin^2 \theta} \quad (4.15a)$$

$$P_2 = \int_0^{P_0} [(D \sin \theta)^2 + y^2] dy \quad \dots \quad (4.15b)$$

$$P_n = \int_0^{P_0} [(D \sin \theta)^2 + y^2]^{\frac{n}{2}} dy \quad (4.15c)$$

system is a system with m unknowns and thirteen equations (thirteen photodiodes read along chords through the bed):

The mathematical solution of this system (least squares method) is given as follows:

$$a_0 P_{0,i} + a_2 P_{2,i} + \dots + a_m P_{m,i} = I_i \quad \text{for } i = 1 - 13 \quad (4.16)$$

Solving the system of Equation 4.17a with the physical constrains of the Equation 4.17b

$$\begin{aligned} f_i(a_0, a_2, \dots, a_m) &= a_0 P_{0,i} + a_2 P_{2,i} + \dots + a_m P_{m,i} - I_i = \text{minimum for } i = 1 - 13 \Rightarrow \\ &\Rightarrow \sum_{i=1}^{13} [f_i(a_0, a_2, \dots, a_m)]^2 = \text{minimum} \Rightarrow \\ &\Rightarrow \frac{\partial \left[\sum_{i=1}^{13} [f_i(a_0, a_2, \dots, a_m)]^2 \right]}{\partial a_0} = 0, \quad \dots, \quad \frac{\partial \left[\sum_{i=1}^{13} [f_i(a_0, a_2, \dots, a_m)]^2 \right]}{\partial a_m} = 0 \end{aligned} \quad (4.17a)$$

$$\varepsilon(r) > 0, \quad \varepsilon(r) < 0.63, \quad \varepsilon(\text{centerline}) \neq 0, \quad (4.17b)$$

gives the coefficients for a m th order polynomial of the radial density profile:

The last parameter that has to be selected in order to obtain a radial solid density profile, using the above algorithm is the polynomial order. Different polynomial orders

$$[1 - \varepsilon(r)] = a_0 + a_2 r^2 + \dots + a_m r^m \quad (4.18)$$

from three to six were tried to examine their influence into the radial solid density profile. Figure 4.15 show the radial solid density profile at 0.80m (5.3L/D) above the nozzle inlet in the riser. The superficial gas velocity is $U_g = 3.0\text{m/s}$ and the nozzles are located at the center of the riser. It may be observed that each polynomial gives a quite different radial

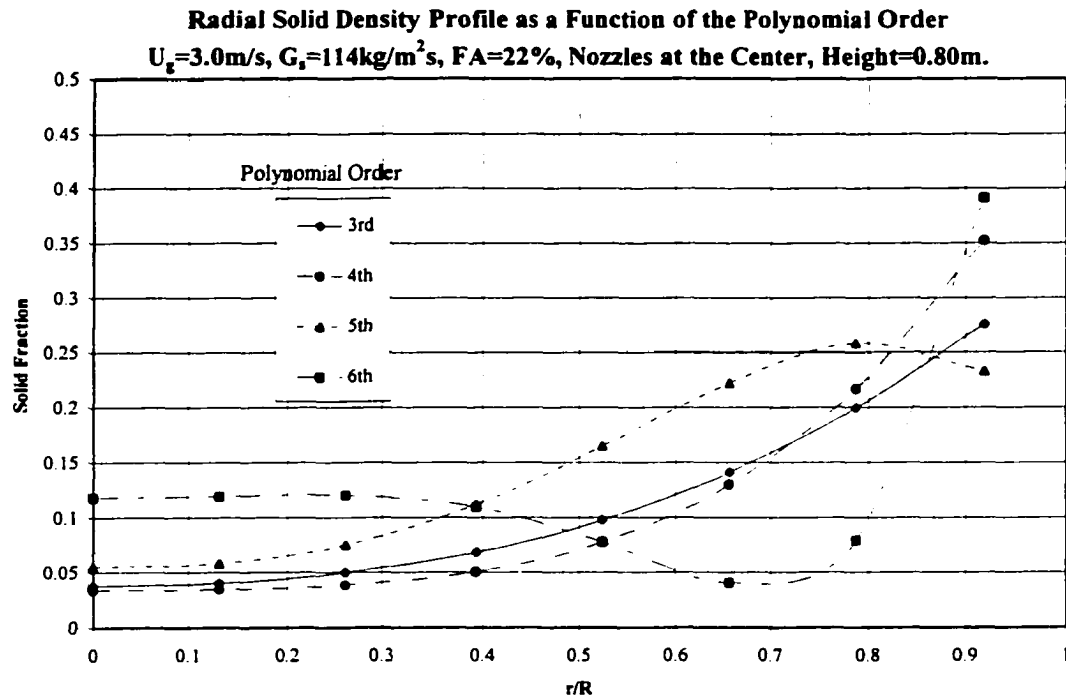


Figure 4.15: Radial Solid Profile as a Function of the Polynomial Order.

profile. For high order polynomial (fifth and sixth order) the shape of the profile does not correspond to the accepted distribution of solid in the riser. This is in agreement with previous studies with the x-ray system that showed that the polynomial order cannot be higher than four (Feindt, 1990). The difference between third and fourth order polynomials is that the third one gives higher solid fraction close to the riser center and lower close to the wall while the fourth order polynomial gives flatter distribution close to the center and higher values close to the wall. The profile obtained from the fourth order polynomial represents better the core-annulus flow that exists in the riser with the certain inlet configuration. Figure 4.16 shows how the cross sectional average solid fraction changes with the polynomial order. It may be observed that all polynomial orders give a cross sectional average very close to the average obtained from the pressure

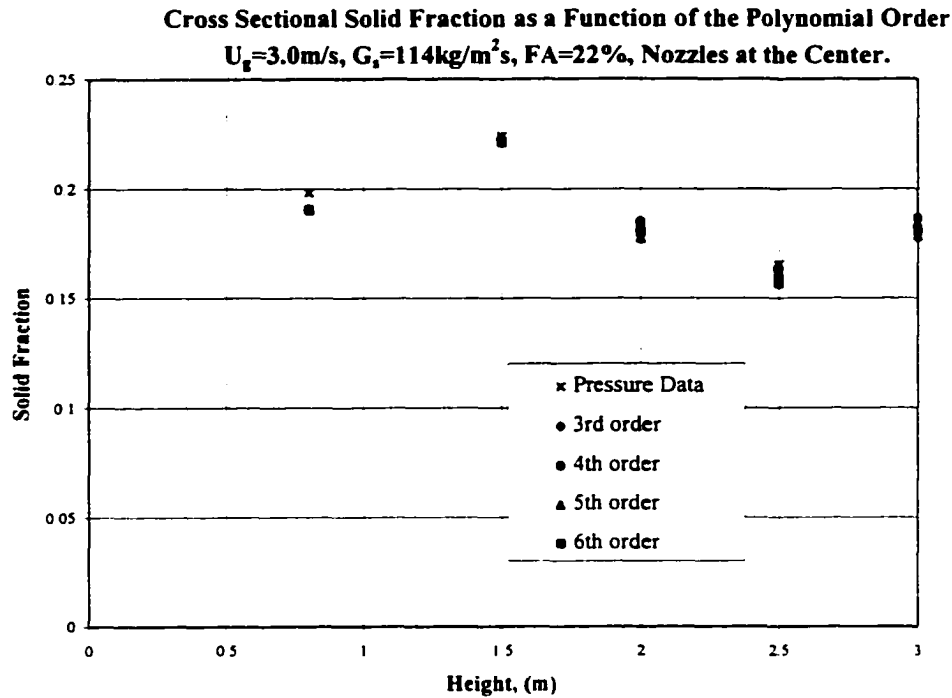


Figure 4.16: Cross Sectional Average Solid Fraction as a Function of the Polynomial Order.

measurements. On the other hand the fourth order polynomial is the one that gives average solid fractions closer to those obtained from the pressure measurements. The fourth order polynomial is used for the construction of all the radial solid density profiles in this work.

Feindt (1990), showed that a total measuring time of 1 min is enough to give a time independent radial solid density profile. The x-ray data in this investigation are obtained from a signal of 16,000 points acquired at 100 Hz frequency, a record of 160s.

4.4.3. LOCAL SOLID FLUX MEASUREMENTS

The local solid flux in different positions of the riser can be estimated using a non-isokinetic sampling probe. This kind of probe has been used by other investigators,

Monceaux et al. (1985) and Rhodes et al. (1988). They collected solid with the aid of a vacuum pump from a certain position inside the riser into a filter bag. Measuring the weight of solids that were collected over a certain time they estimated the solid flux.

An improved measurement technique was developed and a new device was installed at the City College fast fluidized bed. The probe sampling device is shown in Figure 4.17. The sampling probe is stainless steel of I.D. 7.54 mm and O.D. 9.54 mm. The objective of this probe is to provide the net flux of solid (upward-downward), at each position. Two different measurements must be made consecutively, first with the probe opening facing upward and then downward. The sampling probe of Figure 4.17 has the advantage that it measures the upward and downward solid fluxes at exactly the same point in the riser, since its edge does not change position when the probe is rotating 180°. The solids are aspirated with the aid of a vacuum pump from the sampling probe into a small fluidized bed which is maintained at minimum fluidization conditions. The bed has a total height of 1.22m and an I.D. of 7.62 cm. The top of the bed is a sintered plate in order to separate solid and gas. The leakage rate in the sampling probe is controlled by a valve. The fluidized bed is equipped with two pressure probes which measure the increase in the pressure drop with time in the bed due to the solid coming from the riser. This slope of the differential pressure gives the solid flux according the following mass balance equation:

where: G_{local} is the local solid flux ($\text{kg/m}^2\text{s}$),

A_{bed} is the cross sectional area of the sampling bed (m^2), and

A_{probe} is the cross sectional area of the sampling probe (m^2).

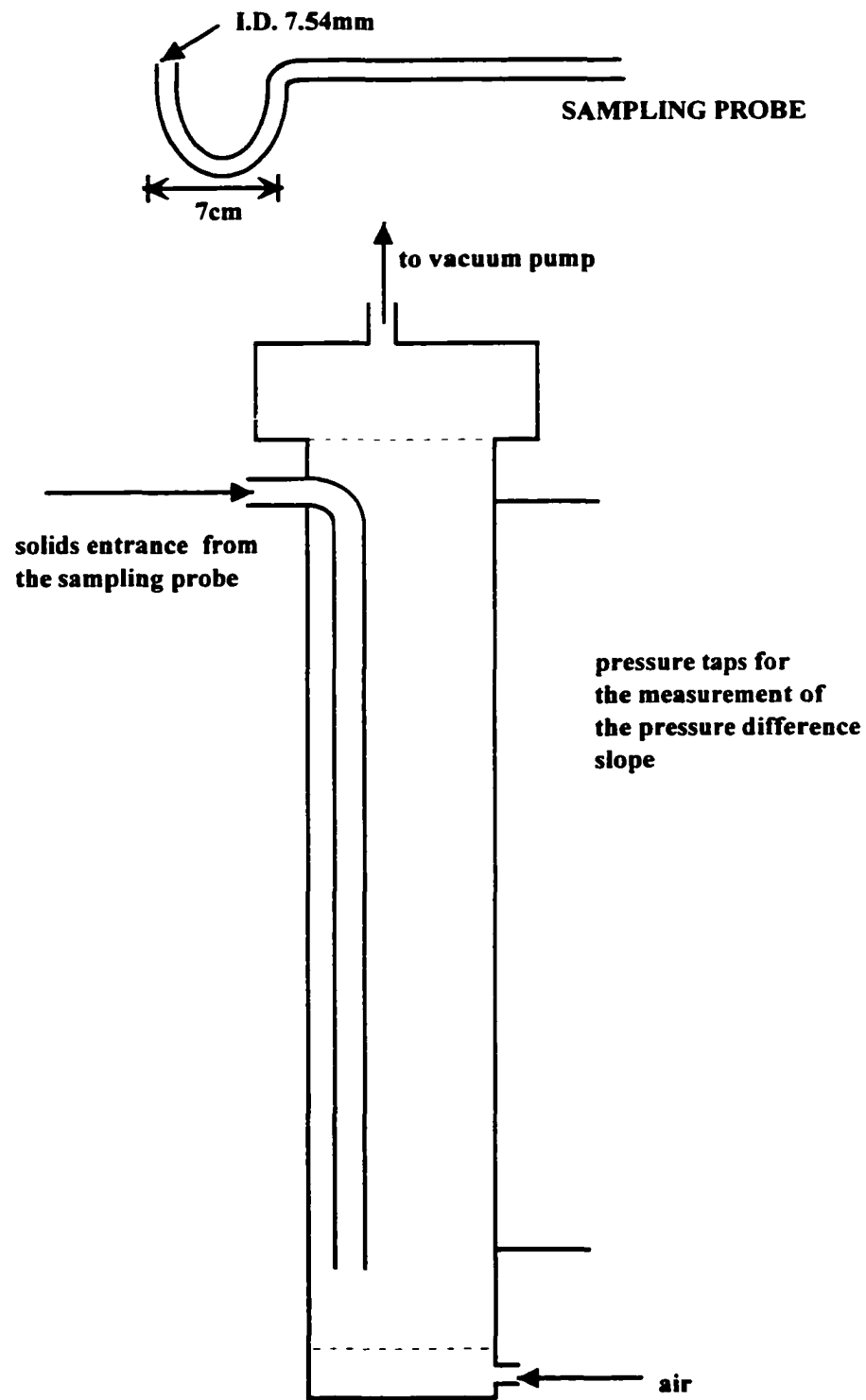


Figure 4.17: Local Solid Flux Measurement Device.

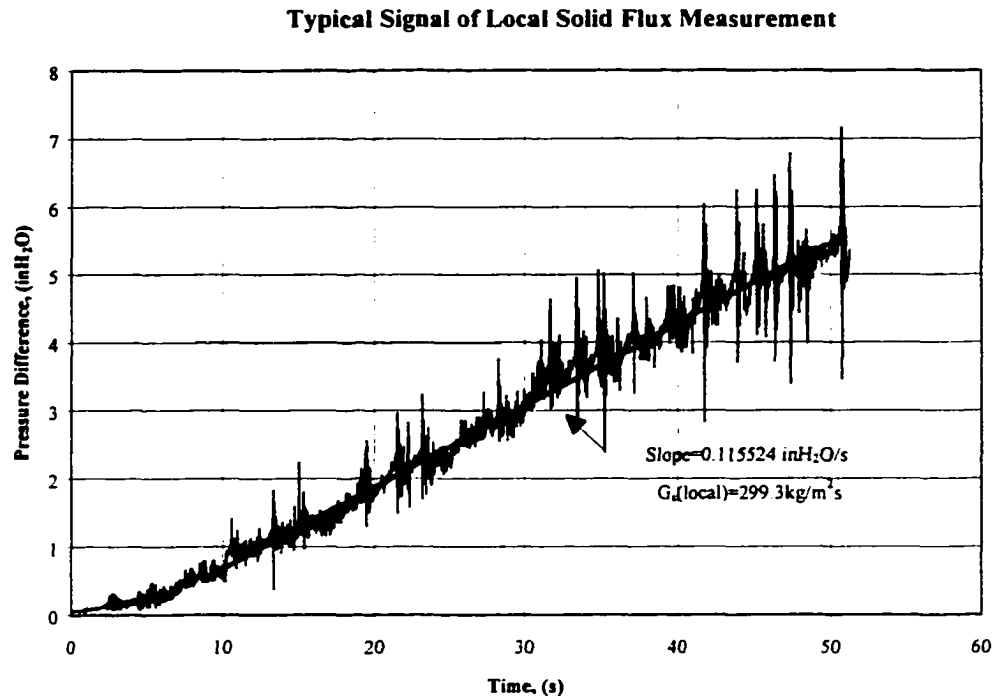


Figure 4.18: Typical Signal of Solid Flux Measurement.

$$G_{\text{local}} = \frac{A_{\text{bed}}}{A_{\text{probe}}} \frac{1}{g} \frac{d(\Delta P)}{dt} \quad (4.19)$$

A typical signal from this measurement is shown in Figure 4.18.

Different aspirating velocities were tried in order to examine their influence on the solid flux measurements. The results are shown in Figure 4.19. It may be observed that the net radial solid flux profile is not affected significantly by the aspirating velocity (2.5 to 9.5m/s). The flux values are almost independent of the aspirating velocity close to the riser wall while an increase of the measured flux with the aspirating velocity can be observed close to the riser axis. This happens because the downflow of the solid at the center of the riser is negligible.

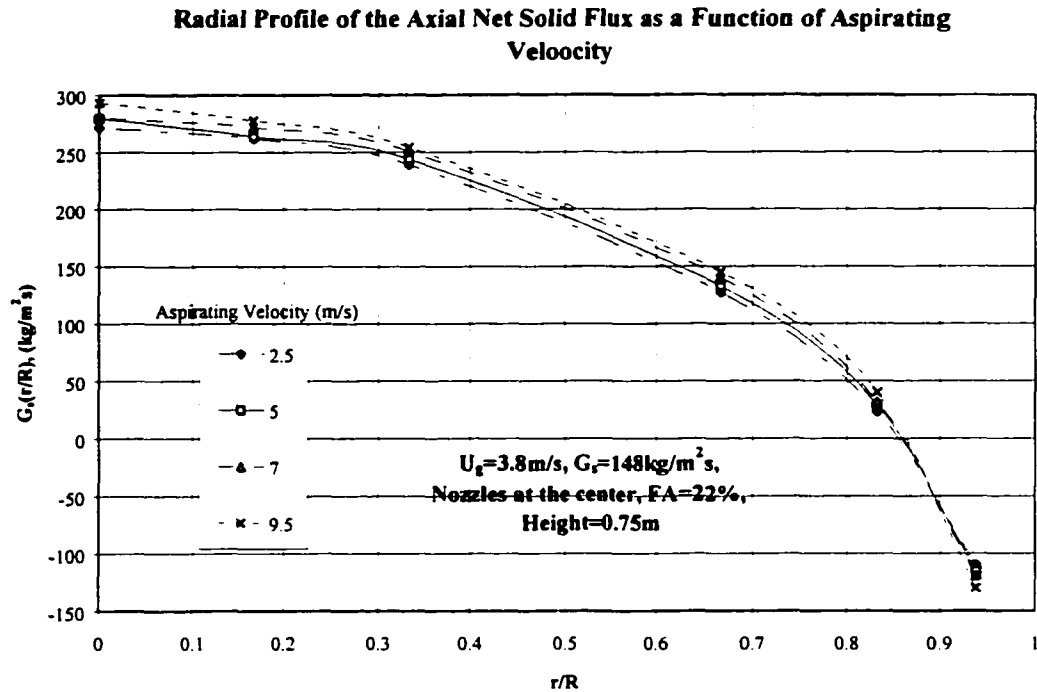


Figure 4.19: Radial Profile of the Axial Net Solid Flux as a Function of the Aspirating Velocity.

Table 4.2 shows the values of the solid circulation rates obtained from the radial solid flux measurements. It is seen that as the aspirating velocity increases the measured total solid circulating rate increases but for all the cases the values of the integral over the cross-section are very close to the independently measured circulation rate. An aspirating velocity of 2.5 m/s appears to give the closest value for the overall solid rate and this velocity is used for all of the local solid flux measurements in this investigation.

Aspirating Velocity, (m/s)	G_z , (kg/m ² s)
2.5	152.4
5	156.5
7.5	161.7

9.5	166.2
<i>Measured Circulation Rate</i>	<i>148</i>

Table 4.2: Solid Circulation Rates as a Function of the Aspirating Velocity.

Aguillón J. et al. (1995), examined the influence of the shape of the edge of the probe on the measurements. They used a circular edge probe and a square one with the same cross sectional area. They showed that this difference in the shape does not affect the measurements.

4.5. Operating Variable Range

The entrance section of the modified City College fast fluidized bed facility is a mockup of the entrance section of a FCC unit which consists of three major components. These are a catalyst upflow section through which the catalyst is carried by "fluffing" steam, an area expansion section in which the oil nozzles are located, and the initial region section of the straightsided riser in which most of the oil vaporization and cracking takes place. The City College riser entrance consists of a reduced diameter section which expands to the riser diameter. At this expansion there are four nozzles which are a mockup of the oil nozzles. This section simulates the expansion that takes place at a FCC unit entrance because of the oil vaporization. The air that enters the system below the expansion represents the "fluffing" steam and it will be called "fluffing" air.

The scope of this work is a detailed experimental study of the initial riser section. The modified City College riser gives the ability to study the entrance region by adding to

the system two more operating variables which are critical for this kind of study. These are the radial position of the four nozzles and the "fluffing" air flowrate.

All the operating parameters that are studied, are:

- Superficial Gas Velocity, $U_g(\text{m/s})$ - from 2.1 to 5.2m/s.
- Solid Circulating Rate, $G_s(\text{kg/m}^2\text{s})$ - three rates for each velocity from 42 to 180 $\text{kg/m}^2\text{s}$.
- "Fluffing" Air flowrate as a percentage of the total flowrate, $FA(\%)$ - from 22% to 38%.
- Three different nozzles positions- nozzles are located symmetrically close to the riser wall, at the middle of a radius and at the center of the riser touching each other.
- The solid inventory in the slow bed is kept constant at 2.50m.

The data obtained from the above range of operating conditions permit a detailed description of the behavior of the gas-solid flow in the initial riser section.

V. RESULTS

The focus of this work is to examine experimentally the influence of the entrance configuration (nozzle position and "fluffing" air flowrate) on the gas-solid flow in the riser. The experimental work includes axial and radial pressure gradient measurements and radial solid concentration and flux measurements.

5.1. Axial Solid Distribution

The cross-sectional average, time mean one-dimensional momentum equation for the gas-solid flow in the riser is:

$$-\frac{dP}{dx} = \rho_s(1 - \epsilon)g + \frac{d}{dx}[\rho_s(1 - \epsilon)U_s^2 + \rho_g\epsilon U_g^2] \quad (5.1)$$

Neglecting the gas momentum term because $\rho_g \ll \rho_s$, equation 5.1 becomes:

$$-\frac{dP}{dx} = \rho_s(1 - \epsilon)g + \frac{d}{dx}[\rho_s(1 - \epsilon)U_s^2] \quad (5.2)$$

An apparent solid fraction can be obtained from the axial pressure difference data if it is considered that:

$$\rho_s(1 - \epsilon)g + \frac{d}{dx}[\rho_s(1 - \epsilon)U_s^2] \equiv \rho_s(1 - \epsilon_{\text{apparent}})g \quad (5.3a)$$

$$\frac{(dP)_{\text{avg}}}{L} = \rho_s(1 - \epsilon_{\text{apparent}})g \quad (5.3b)$$

The apparent solid fraction is the same as the true solid fraction when the solid axial velocity is constant. When acceleration of solid takes place one part of the pressure gradient is due to this acceleration and therefore the apparent solid fraction is greater than

the true solid fraction. Expressed in terms of the void fraction, one minus the solid fraction, the apparent void fraction will be less than the true void fraction when acceleration of solid is occurring. However, acceleration of solid can also occur even if the apparent void fraction is about the same as the true void fraction. In that case according the equation 5.3a:

$$\frac{d}{dx} \int_0^R (1 - \epsilon) U_s^2 r dr = 0 \quad (5.4a)$$

Modifying the above equation becomes:

$$\int_0^R U_s^2 \frac{d\epsilon}{dx} r dr = 2 \int_0^R U_s (1 - \epsilon) \frac{dU_s}{dx} r dr \quad (5.4b)$$

When the void fraction increases from the dense region of the riser to the dilute region, the radial density profile data reported in previous works have shown that there is a strong increase of the void fraction close the riser wall and changes only slightly at the riser center. That means that the core region of the riser expands in diameter from the dense to dilute region. In that case, equation 5.4b shows that close the riser wall the solid velocity increases strongly since the void fraction increases strongly. On the other hand, close the riser center where void fraction changes slightly, the solid velocity also changes only slightly. The radial profile of the axial solid velocity, therefore, rearranges its shape with higher velocities close the riser wall. The cross sectional average of the solid velocity is however considerably higher after the profile rearrangement. This readjustment of the axial solid velocity profile is characteristic for this kind of solid acceleration where the apparent and the true void fractions remain about the same.

Figure 5.1 shows the axial voidage distribution for a superficial velocity, U_g of 3.8m/s and solid circulation rate, G_s of 111kg/m²s. In this case the "fluffing" air flowrate (FA) is 22% of the total flowrate in the riser and the four nozzles are located at the center of the riser touching each other. The cross sectional average true void fraction obtained from the x-ray data are also drawn in Figure 5.1. It is seen that when the voidage increases (from 1.50m to 3.0m) the x-ray void fraction data are slightly higher than the pressure difference data. This difference is due to the increasing axial solid velocity. The equation of continuity for solid can be written:

$$\overline{G}_s = \rho_s \overline{(1 - \epsilon)} U_s \quad (5.5a)$$

OR

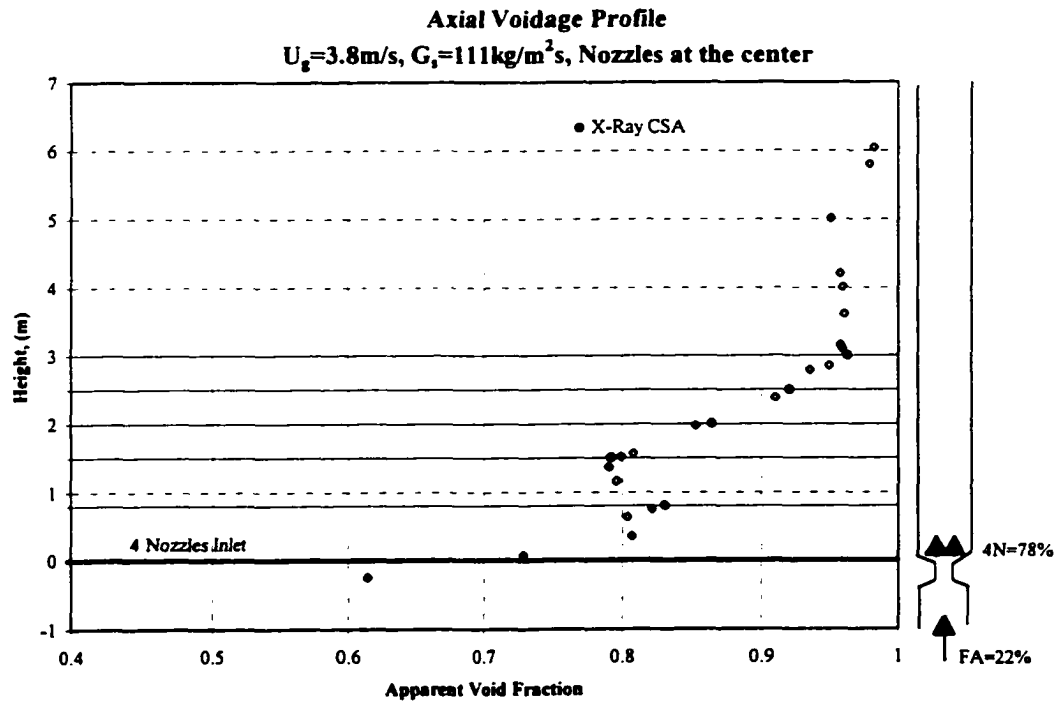


Figure 5.1: Axial Voidage Profile for Nozzles Located at the Center.

$$\overline{G_s} = \rho_s \overline{(1 - \epsilon)U_s} \quad (5.5b)$$

where the bars denote cross-sectional average.

Experience shows that in general:

$$\overline{(1 - \epsilon)U_s} \cong \overline{(1 - \epsilon)}\overline{U_s} \quad (5.5c)$$

The validity of the equation of 5.5c has been checked from the independent experimental data of the radial profile of solid density and radial profile of the axial solid velocity that are presented in the next chapters. The equation 5.5 shows that as the solid accelerates, the void fraction must increase in order to maintain a constant net solid flux, G_s . The solid accelerates from above the nozzles up to a height of 2.5 L/D. There is a small section where solid density is constant and then a small acceleration up to a height of 3.5 L/D (about 0.53m). After this height and up to a height of 10 L/D (1.50m) solid decelerates. Above 10 L/D, the void fraction increases uniformly and the flow becomes dilute. In this dilute flow region of the riser, the upflow area is greater than in the dense flow region. This results in the decrease of the solid and gas velocity at the riser axis since there is an expansion of the upflow area. Figure 5.2 shows the axial solid hold up for the same operating conditions but with the four nozzle located in the middle of the riser (between wall and axis). The axial voidage profile shows different behavior. The solid accelerates up to a height of 3.5 L/D (0.53m) and then decelerates up to 7 L/D (about 1.0m). From that height and up to 20 L/D (3.0m), the solid density decreases slightly except for a small increase that appears at 19 L/D. After that solid fraction decreases rapidly to the dilute flow region. The axial voidage profile for the third case where the nozzles are located as close as possible to the wall of the riser is shown in Figure 5.3. The solid immediately above the four nozzles and up to 7 L/D decelerates. After that solid accelerates slowly up

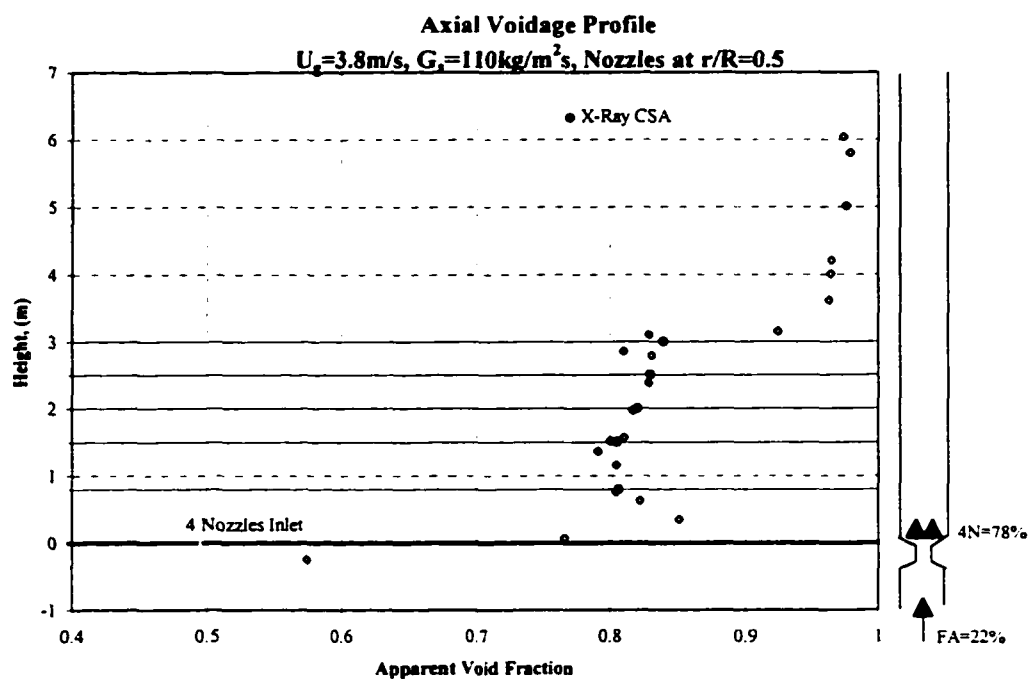


Figure 5.2: Axial Voidage Profile for Nozzles Located at the Middle.

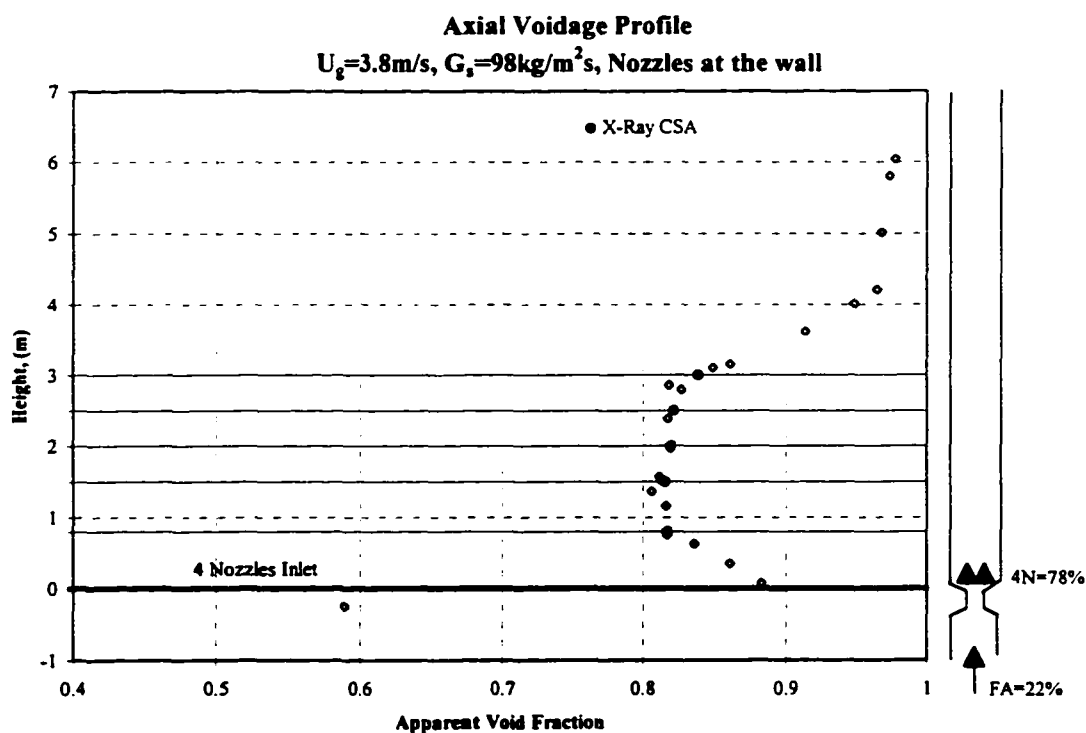


Figure 5.3: Axial Voidage Profile for Nozzles Located at the Wall.

to 20 L/D and then smoothly accelerates up to a height of 28 L/D where flow becomes dilute.

The apparent void fraction data are compared with the cross sectional average void fraction obtained from x-ray data. As can be seen from Figures 5.1, 5.2 and 5.3, the cross sectional void fractions obtained from x-ray data are in good agreement with the void fraction data obtained from the pressure difference readings.

The dependence of the apparent void fraction on the pressure tap spacing is also examined and the results are shown in Figures 5.4, 5.5 and 5.6 for the above three runs. Nested pressure taps of different spacing (0.5, 2.7 and 8 L/D) are installed so that the apparent void fraction at the same elevation can be obtained. This elevation is at 0.75m (5 L/D) above the nozzle inlet. The apparent void fraction for pressure tap spacing smaller than 1 L/D is the lowest for all the three cases. This difference that always appears for the 0.5 L/D spacing appears to be the result not only of the acceleration of solid in this section but mainly of the local effects of turbulence. For spacing larger than 1 L/D the apparent void fraction decreases with increasing pressure tap spacing. Figure 5.4 shows that the pressure reading with the nozzles located at the center, appear to have the largest difference for different tap spacings. Figures 5.5 and 5.6 show that for the nozzles located at the middle and at the wall the apparent void fraction is slightly decreasing with the pressure tap spacing. This is in agreement with the axial voidage profiles. Figure 5.1 shows that the initial section is a complex entrance section characterized by local rapid accelerations and decelerations of solid. On the other hand, Figures 5.2 and 5.3 show that when the nozzles are located at the middle or close to the wall of the riser, there is a

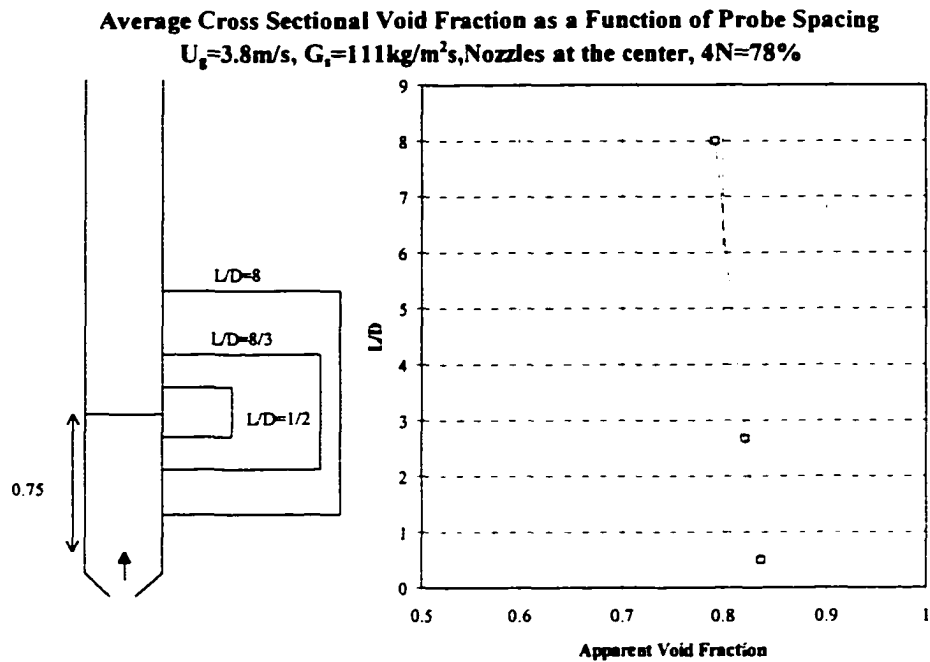


Figure 5.4: Apparent Void Fraction as a Function of the Pressure Taps Spacing for Nozzles Located at the Center.

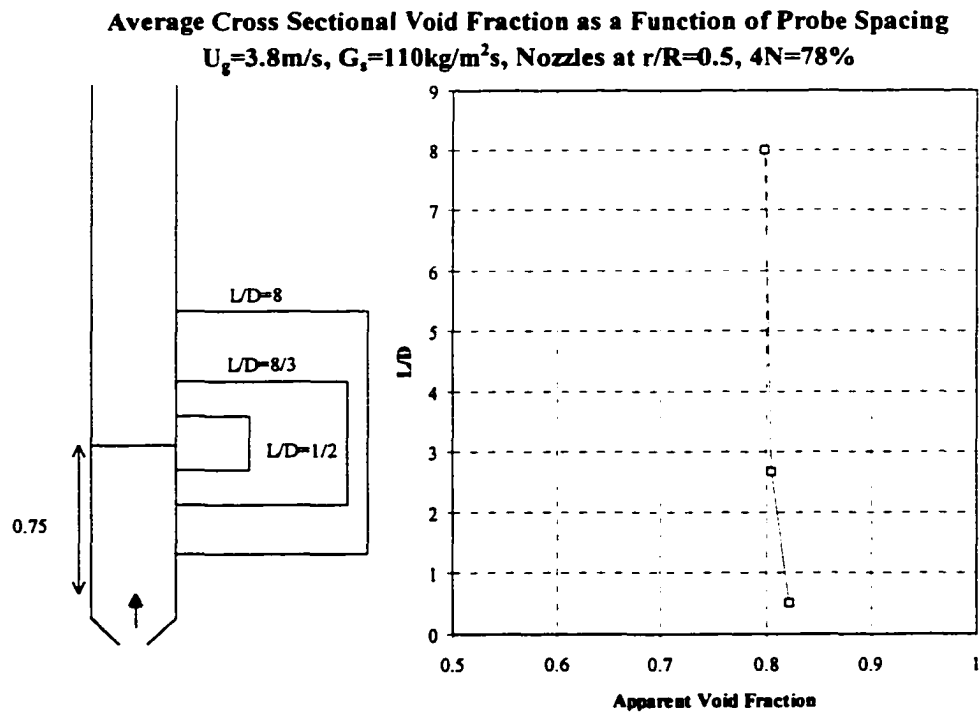


Figure 5.5: Apparent Void Fraction as a Function of the Pressure Taps Spacing for Nozzles Located at the Middle.

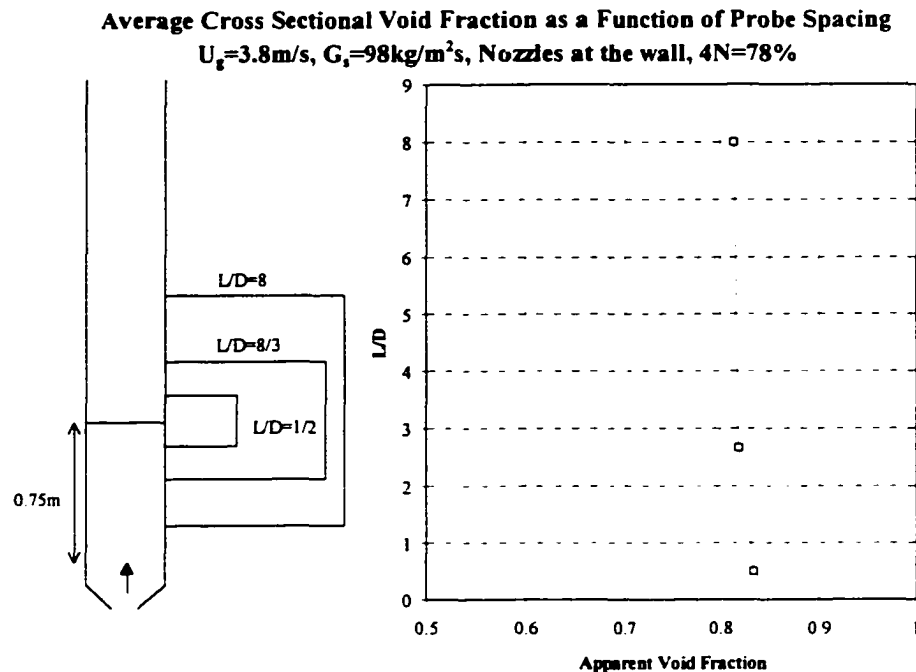


Figure 5.6: Apparent Void Fraction as a Function of the Pressure Taps Spacing for Nozzles Located at the Wall.

smoother deceleration of solid which is the cause of the small increase of the apparent void fraction with the pressure tap spacing.

Figure 5.7 shows the axial pressure difference along the riser for the three different nozzle positions. The axial pressure profile for the three cases is similar. After the initial section (0-1.0 m), all profiles appear linear until the inflection point. The different axial pressure profile initial slopes is another indication of the complexity of the flow in the entrance section. The difference in the height of the inflection point is shown clearly in Figure 5.7. When the nozzles are at the center the inflection point is much lower than with the other two nozzle positions. It is also shown that the total pressure drop in the riser is smaller when the nozzles are located at the center. This means that the

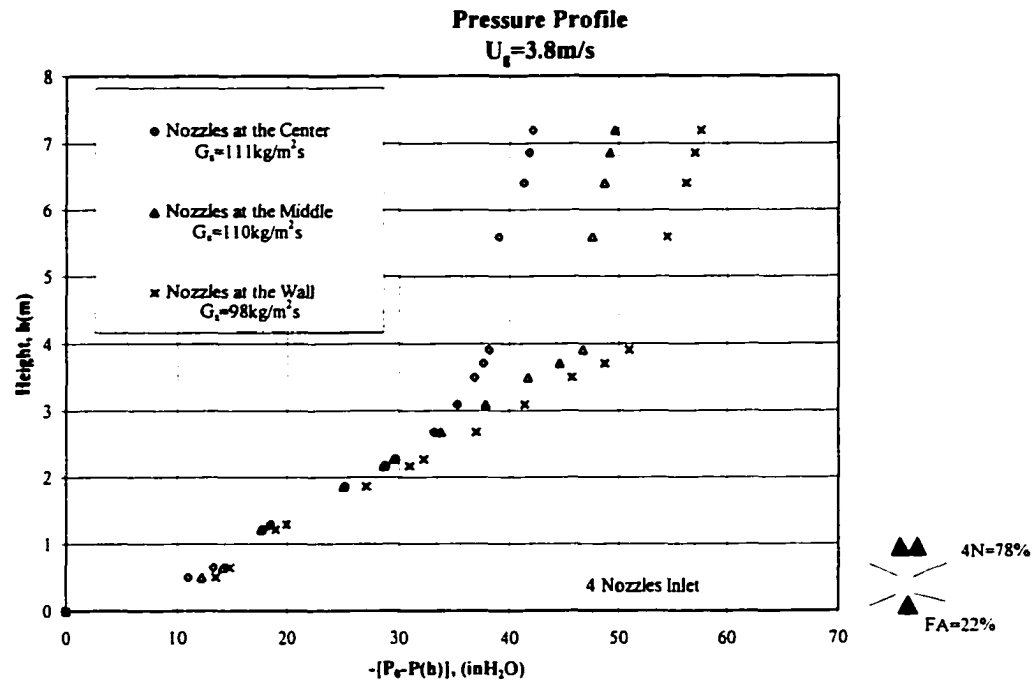


Figure 5.7: Axial Pressure Profile for the Three Nozzle Positions.

pressure drop in the contraction-expansion section and below to the U-tube connecting the slow bed to the fast bed, must be greater when the nozzles are located close to the riser axis in order to satisfy the pressure balance. This additional pressure drop is due to the difference in area between the two positions. When the nozzles are close the riser axis they cover 50% more of the cross sectional area than when the nozzles are located close to the riser wall due to the horizontal tubes of the nozzles.

In order to examine the influence of the nozzle configurations on the riser flow for a range of operating conditions, different solid circulation rates and gas superficial velocities must be investigated. Figure 5.8 shows the axial voidage profile for the same superficial velocity as in the previous set of cases, $U_g = 3.8 \text{ m/s}$, but for a solid circulation

rate of $G_s=73\text{kg/m}^2\text{s}$. The nozzles are located at the center of the riser and the "fluffing" air flowrate is $FA=22\%$. It can be seen again that the solid accelerates from above the nozzles up to about 0.65m (4.3 L/D). Then there is a section where local acceleration and deceleration of solid take place until a height of 1.3m (8.6 L/D) where the solid smoothly accelerates to the dilute flow region. The bed is more dilute compared with that shown in Figure 5.1 where the solid circulation rate is about 50% higher. When the nozzles are located in the middle, Figure 5.9, the solid accelerates up to 2 L/D and then decelerates up to 0.70m (4.7 L/D). A constant density section follows and then an acceleration section which consists of two parts. A weak acceleration followed by a strong acceleration which carries to the upper dilute zone of the riser. With the nozzles located at the wall, Figure

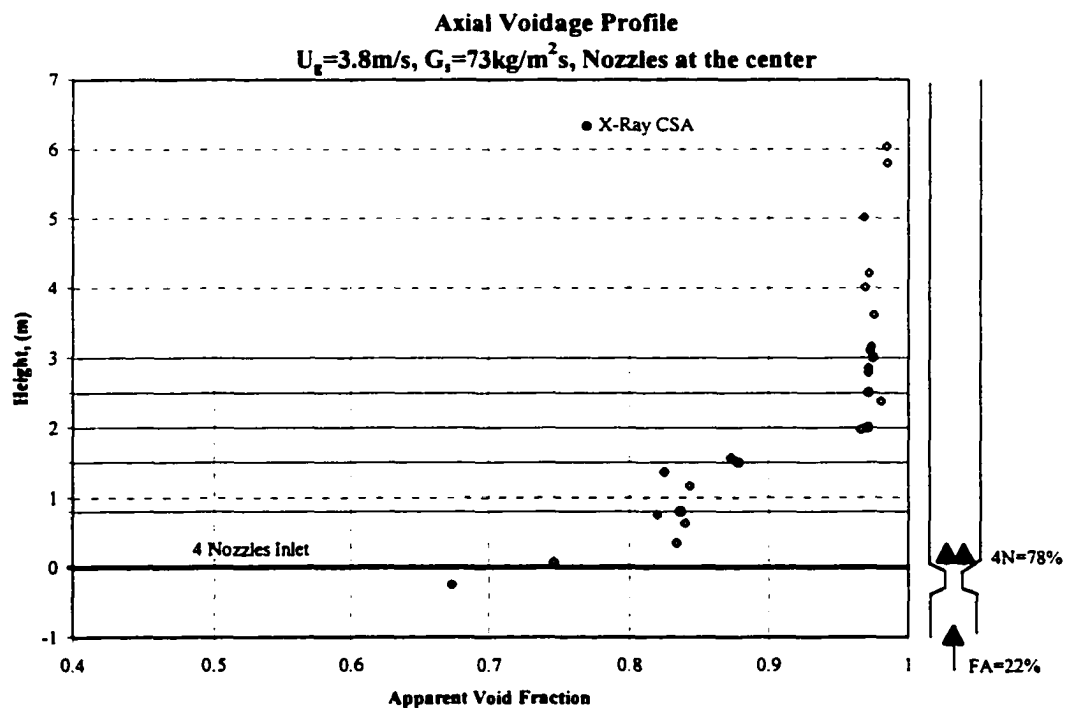


Figure 5.8: Axial Voidage Profile for Nozzles Located at the Center.

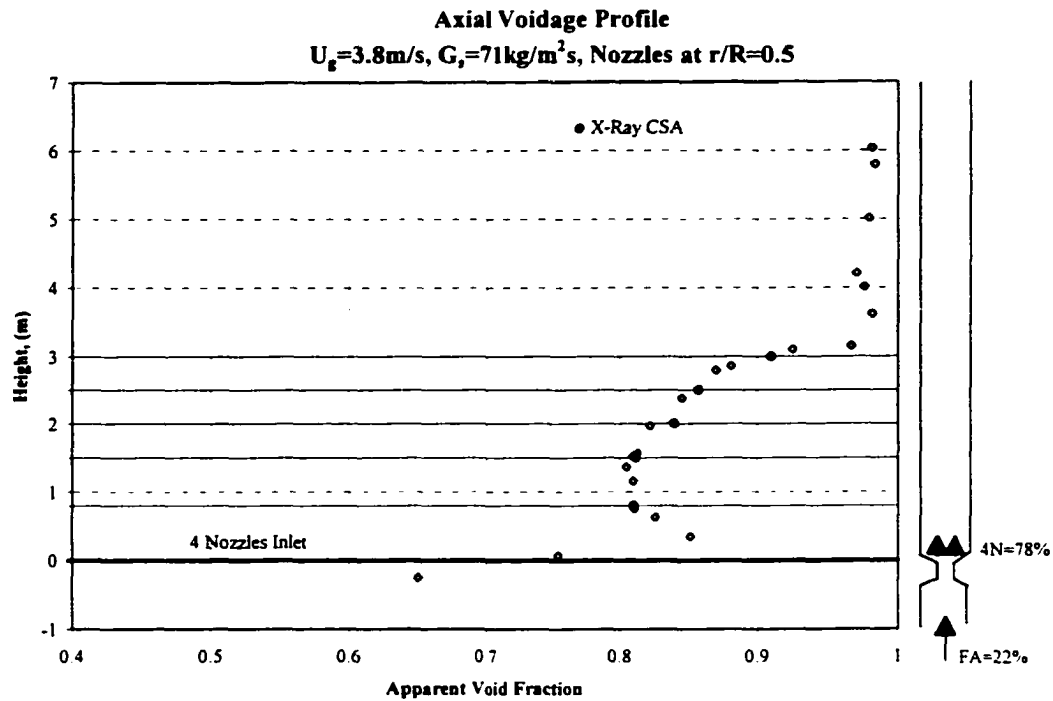


Figure 5.9: Axial Voidage Profile for Nozzles Located at the Middle.

5.10, the solid decelerates up to 0.70m (4.3 L/D). The axial voidage profile does not differ significantly when nozzles are in the middle or close to the wall. This is because with a nozzle O.D. of 38 mm and a wall thickness of 7 mm, when the nozzles are touching the wall, there is still a gap of about 7 mm between air inlet and riser wall. The comparison between the apparent void fraction and the true void fraction from x-ray data are in good agreement. Small differences appear in the acceleration sections where the true void fraction is higher than the apparent value.

Figure 5.11 shows the effect of the pressure tap spacing for the cases with the low solid circulation rate. For all the nozzle positions, the elevation of 0.75m (5 L/D), where the apparent void fraction is obtained from the nested pressure probes, is at the end of a deceleration zone and the beginning of an almost constant density section. This is the

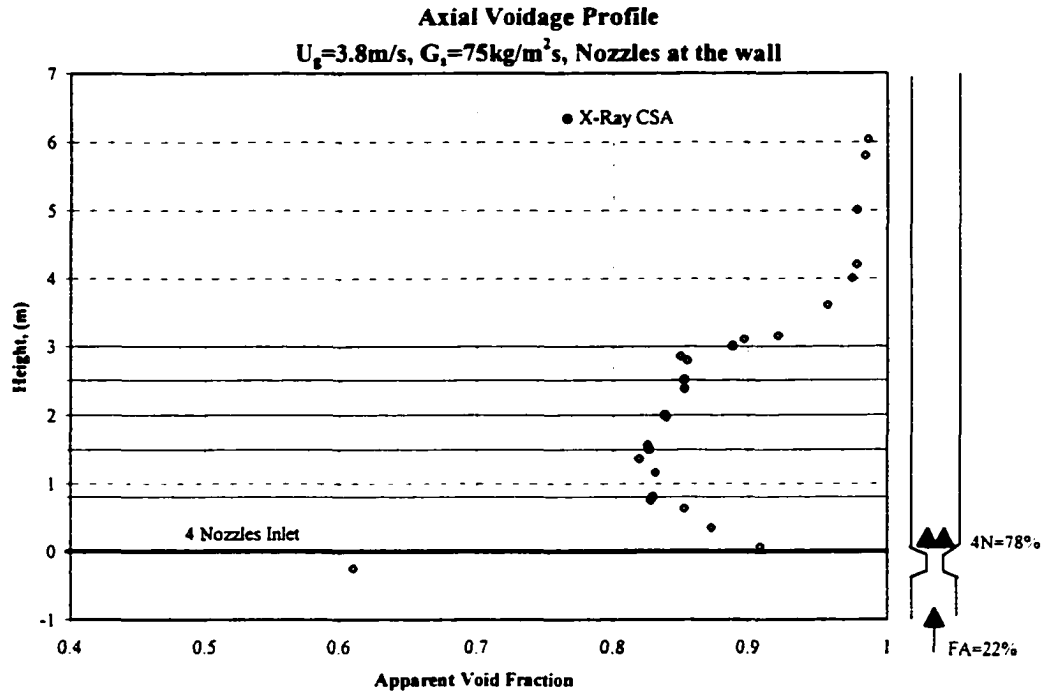


Figure 5.10: Axial Voidage Profile for Nozzles Located at the Wall.

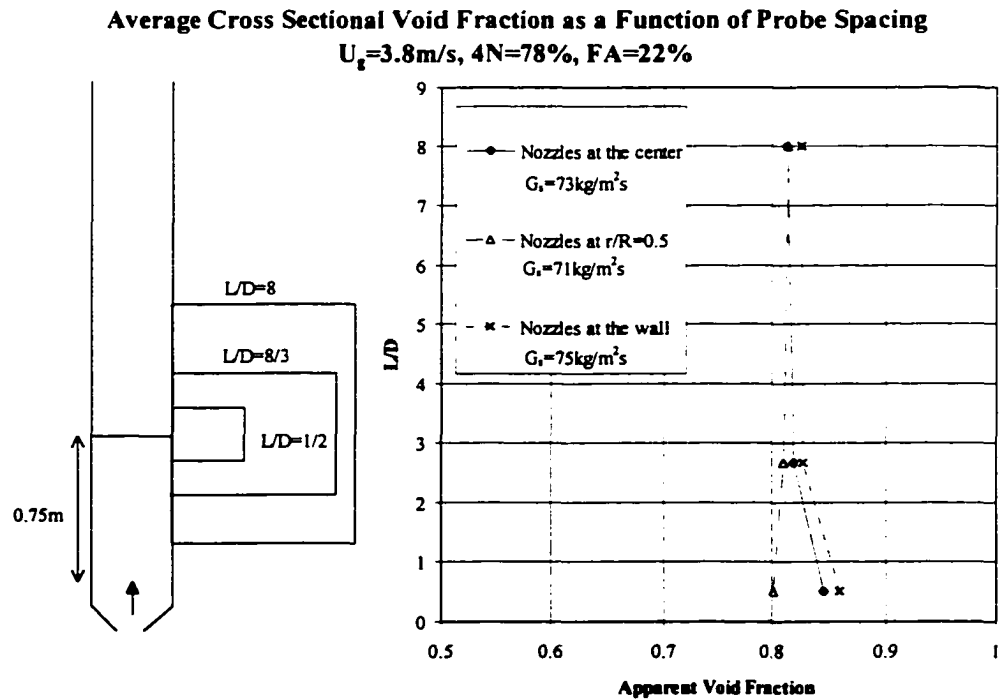


Figure 5.11: Apparent Void Fraction as a Function of the Pressure Taps Spacing for Three Nozzle Positions.

reason that in Figure 5.11 the difference between apparent void fraction at 2.7 and at 8 L/D spacing is negligible. The 0.5 L/D measurement appears to be higher than with the other spacings. When the nozzles are located at the center or at the wall the 0.5 L/D spacing gives the highest apparent void fraction value. When the nozzles are located in the middle of the riser the smallest spacing gives the lowest apparent void fraction because the axial voidage profile in this case is the most complex.

The next set of data refers to experiments with a higher gas superficial velocity. Figure 5.12 shows the axial voidage profile for a superficial gas velocity U_g of 5.1 m/s and solid circulation rate G_s of 180 kg/m²s. The nozzles are located at the center of the riser and the "fluffing" air flowrate is 22% of the total flowrate. The solid in this case

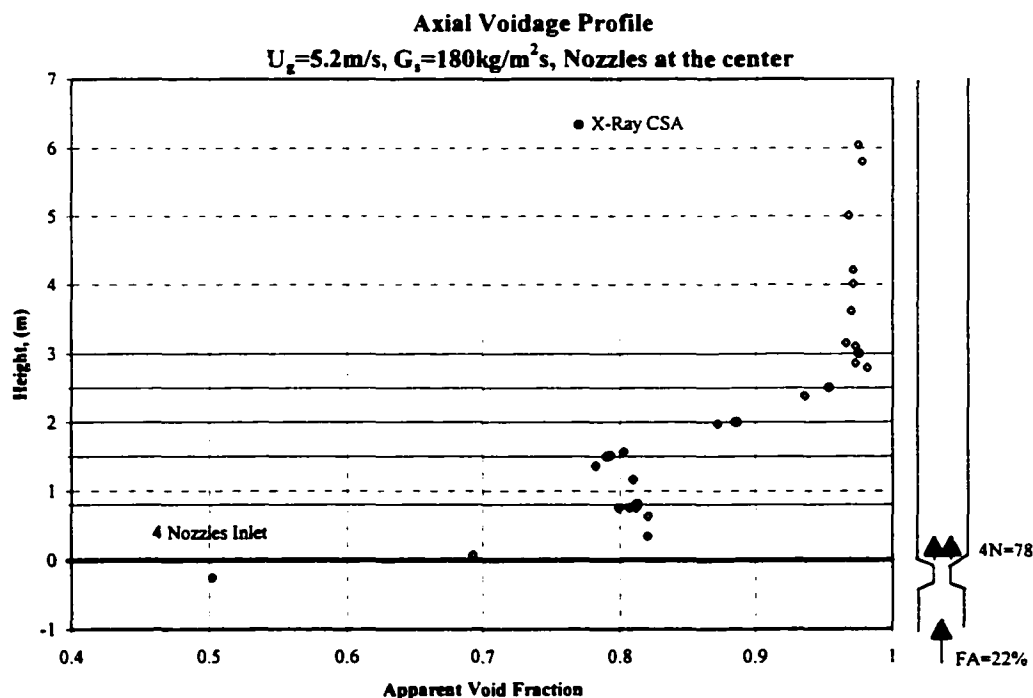


Figure 5.12: Axial Voidage Profile for the Nozzles Located at the Center.

accelerates from above the nozzle inlet up to a height of 0.3m (2 L/D). After this elevation and up to 1.5m (10 L/D) there is a section of complex flow. At this section the void fraction increases and decreases locally, but the final voidage at 1.5m (10 L/D) is lower than in 0.3m (2 L/D) which means that there is a net deceleration of the solid from 0.3m to 1.5m. Above 1.5m the void fraction continuously increases up to the upper dilute region of the riser.

Different behavior appears when the nozzles are located in the middle between riser wall and axis, keeping all the other operating conditions almost constant. Figure 5.13 shows that the void fraction increases again up to 0.3m but now the actual void fraction values are higher than in the previous case. From 0.3m to 1.50m, the void fraction decreases so according Equation 5.4 solid deceleration takes place. Above 1.50m an

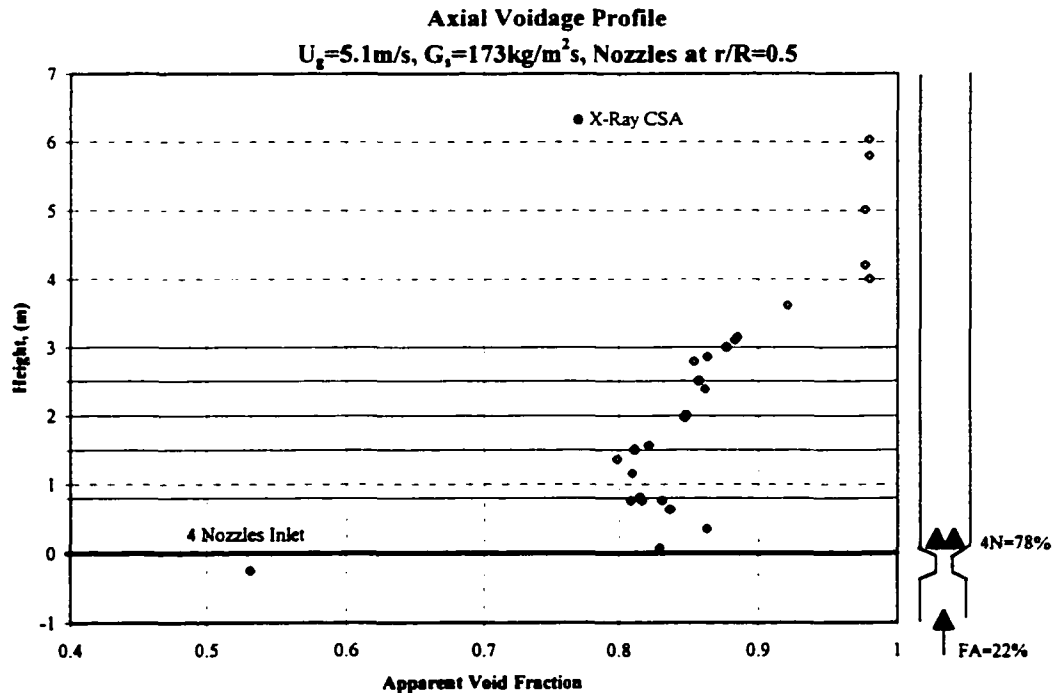


Figure 5.13: Axial Voidage Profile for the Nozzles Located at the Middle.

acceleration of solid starts and continues up to the dilute region with a small complex section between 2.5 and 3m where void fraction increases and decreases locally. The height of the inflection point is higher than the case where the nozzle are located at the center of the riser.

The last case when the nozzles are located close to the riser wall is shown in Figure 5.14. With this nozzle configuration, the void fraction decreases from immediately above the nozzle inlet up to 1.5m (10 L/D). Then there is a small increase and an almost constant density section followed by the inflection point and the dilute region. The height of the inflection point is slightly greater compared with the case where the nozzles are located at the middle. It can be seen, comparing Figures 5.12, 5.13 and 5.14 that when the

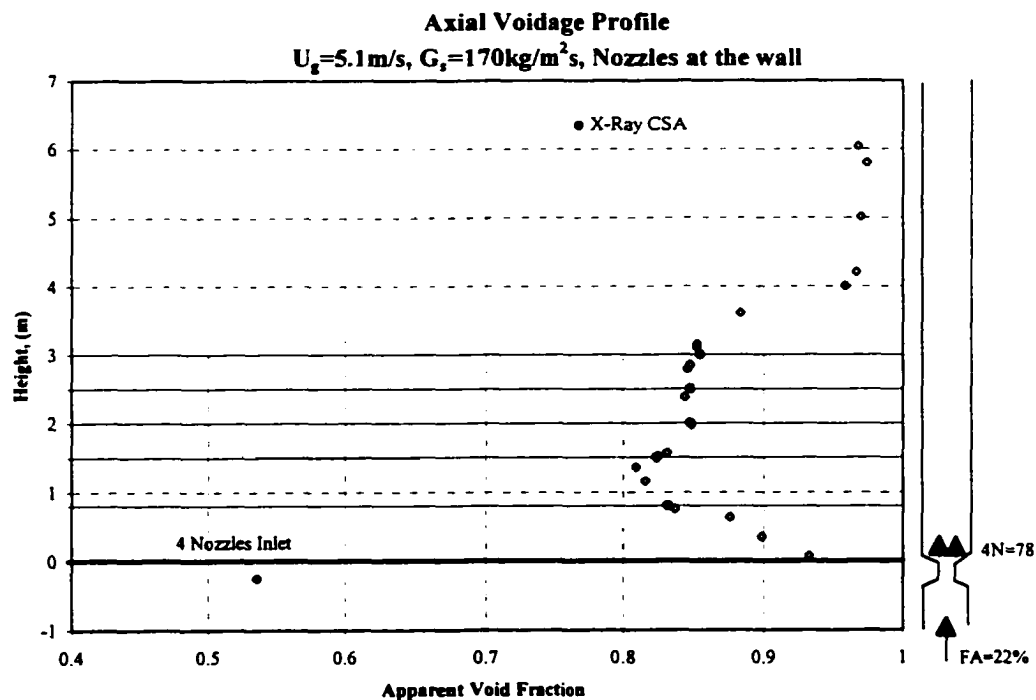


Figure 5.14: Axial Voidage Profile for the Nozzles Located at the Wall.

nozzles are located close to the riser wall, the axial voidage profile is smoother, the local acceleration and deceleration sections are eliminated. These observations are similar with them obtained for the case of 3.8m/s superficial gas velocity (Figures 5.1, 5.2 and 5.3). It is also seen from Figures 5.12-5.14 that the void fraction data obtained from the x-ray measurements are in good agreement with the apparent void fraction data.

Figure 5.15 shows the effect of the pressure tap spacing on the axial pressure reading measurements. It is seen that the difference between the 2.7 and 8L/D spacings is negligible and at the limits of the experimental error. The 0.5L/D tap spacing gives an apparent void fraction much higher especially for the cases where the nozzles are located at the middle or close to the riser wall. This happens because for these two cases the

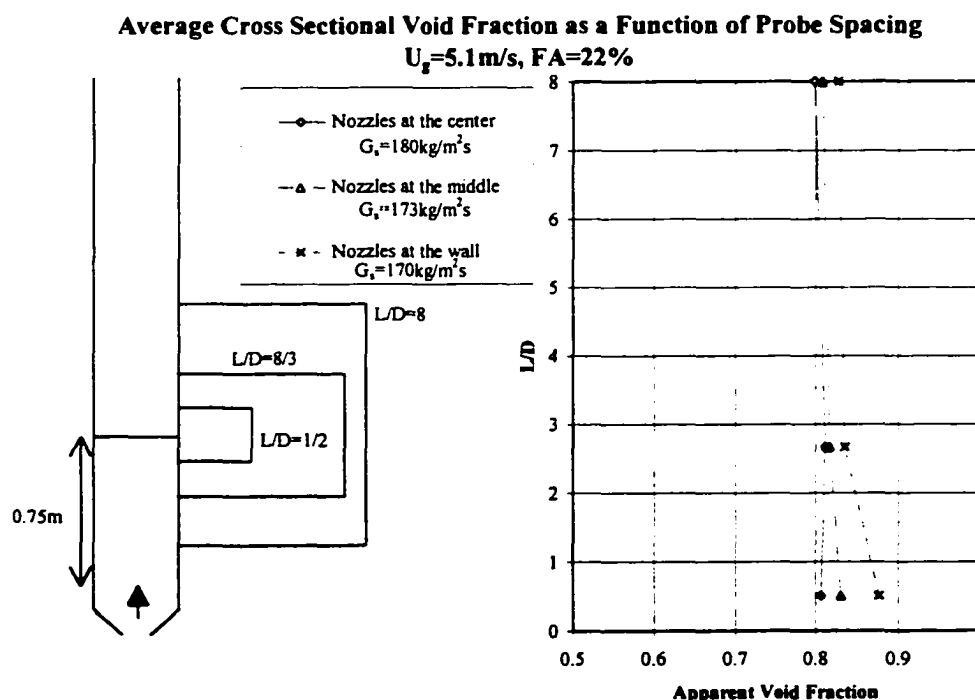


Figure 5.15: Apparent Void Fraction as a Function of the Pressure Taps Spacing for Three Nozzles Positions.

elevation of 0.75m where the apparent void fraction is obtained, is in the middle of a strong deceleration section which results lower apparent void fraction to be obtained for larger tap spacing. Also pressure data measured with a tap spacing smaller than $1L/D$ are strongly affected by the turbulence.

Figure 5.16 shows the axial pressure profile for the above operating conditions for the three nozzles positions. It is seen that the profiles are generally similar. The total pressure drop in the riser is lower when the nozzles are located at the center and this is because the additional pressure drop in the contraction-expansion section and below to the U-tube connecting the slow bed to the fast bed, since the entrance area in the riser is

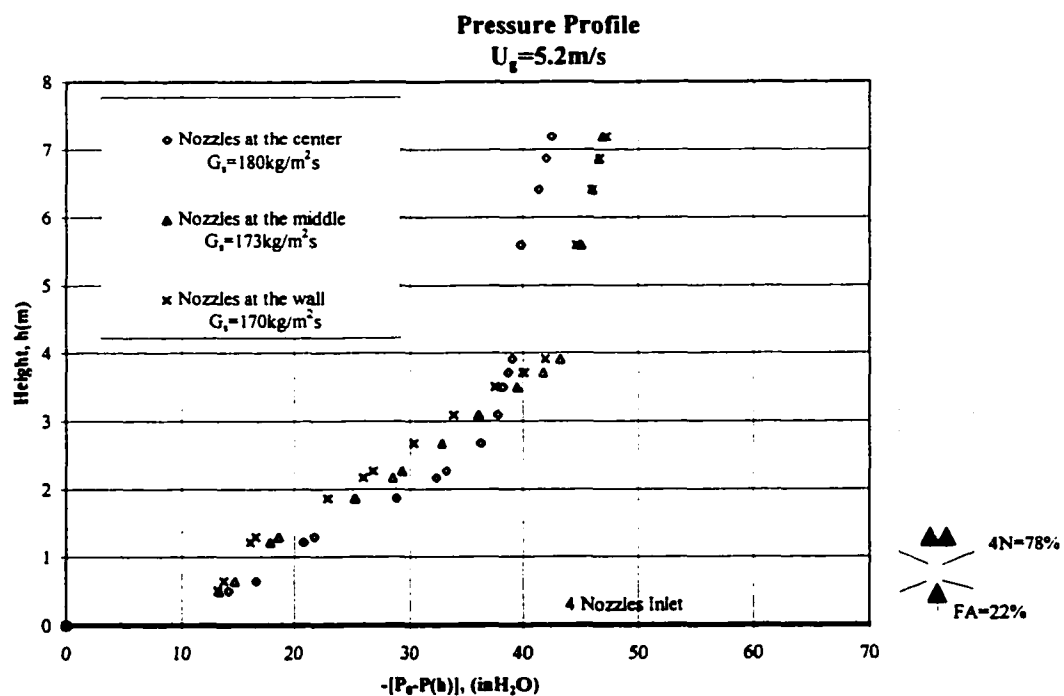


Figure 5.16: Axial Pressure Profile for the Three Nozzle Positions.

50% smaller when the nozzles are in the center than when the nozzles are located at the riser wall.

The next set of data are for the same superficial gas velocity but for a reduced solid circulation rate of $111\text{kg/m}^2\text{s}$. Figure 5.17 shows the axial voidage profile for "fluffing" air flowrate 22% and for three different nozzle positions. When the nozzles are located close to the riser axis the apparent void fraction increases up to 0.3m (2 L/D) and then decreases up to 1.50m (10L/D). It follows a solid acceleration up to the dilute flow region of the riser. When the nozzles are located at the middle the apparent void fraction increases up to 0.3m as in the previous case but now the bed is more dilute at this section. After this elevation solid decelerates up to 1.5m and then accelerates rapidly to reach the upper region of the riser. When the nozzles are located close to the riser wall the solid

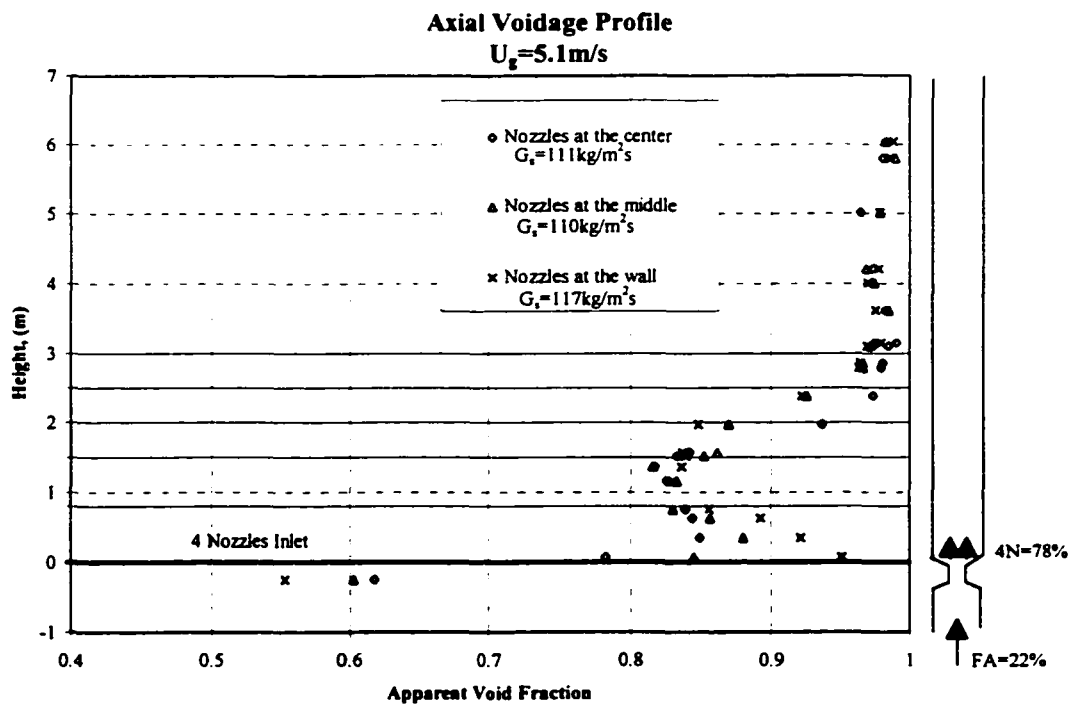


Figure 5.17: Axial Voidage Profile for the Three Nozzle Positions.

decelerates from 0 to 1.5m (10 L/D) and then accelerates to the upper dilute region. In this set of data the height of the inflection point increases as the nozzles are moved from the center to the wall. It may also be observed in comparison with the high solid circulation rate cases (Figures 5.12, 5.13, 5.14), that the constant density section does not exist and that the bed is generally more dilute. The case of an even smaller solid circulation rate is shown in Figure 5.18, where the solid circulation rate is $G_s=65\text{kg/m}^2\text{s}$. In this case the flow in the riser is a dilute flow but in the entrance section from 0 to 1.5m the effect of the nozzle configuration on the axial voidage profile is generally the same as with the cases discussed before.

Experiments were carried out with two more superficial gas velocities. Figures 5.19 and 5.20 show some representative results of these experiments. The full sets of data

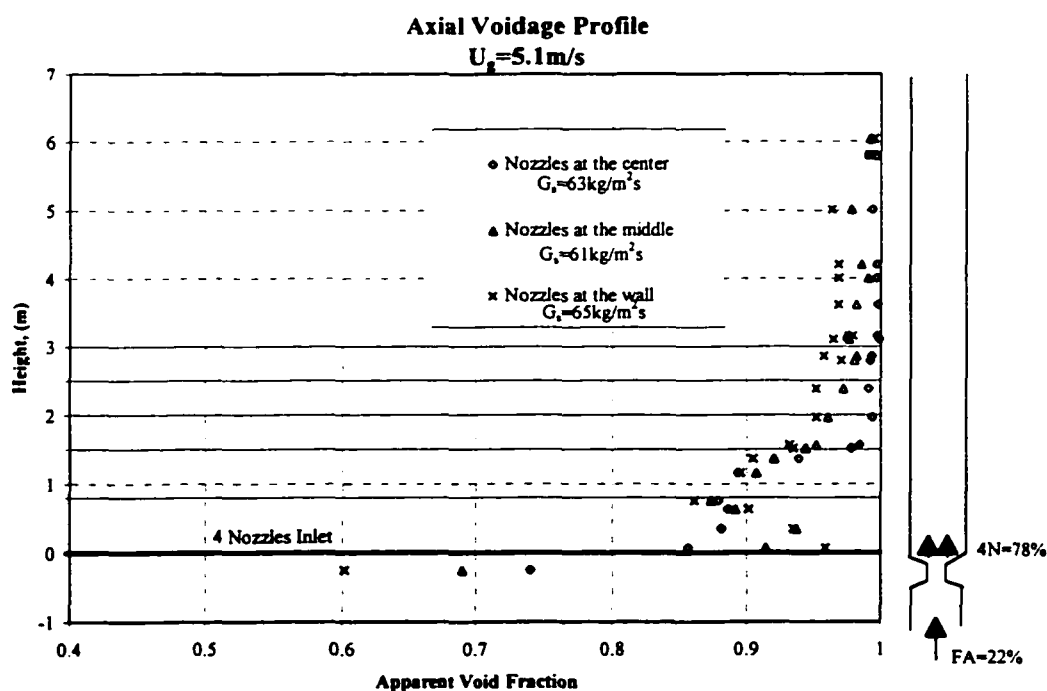


Figure 5.18: Axial Voidage Profile for the Three Nozzle Positions.

for each superficial gas velocity are presented in Appendixes A to C. Figure 5.19 shows the axial voidage profile for a superficial gas velocity U_g of 3.0m/s, for the three different nozzle configurations. The solid circulation rate was kept as constant as possible ($G_s=107\text{--}114\text{ kg/m}^2\text{s}$) and the "fluffing" air flowrate is $FA=22\%$ of the total flowrate. It is seen in this low superficial gas case that for every nozzle position the riser is denser than in the previous higher superficial gas velocity cases and the height of the inflection point has increased. The effect of the nozzle position on the solid distribution in the initial section from 0 to 1.50m (10 L/D) is the same with the last cases. Different behavior appears at the section from 1.5 to 3.0m (10-20 L/D). When the nozzles are at the center, from 1.5 to 3.0m the apparent void fraction increases smoothly and the dilute flow region

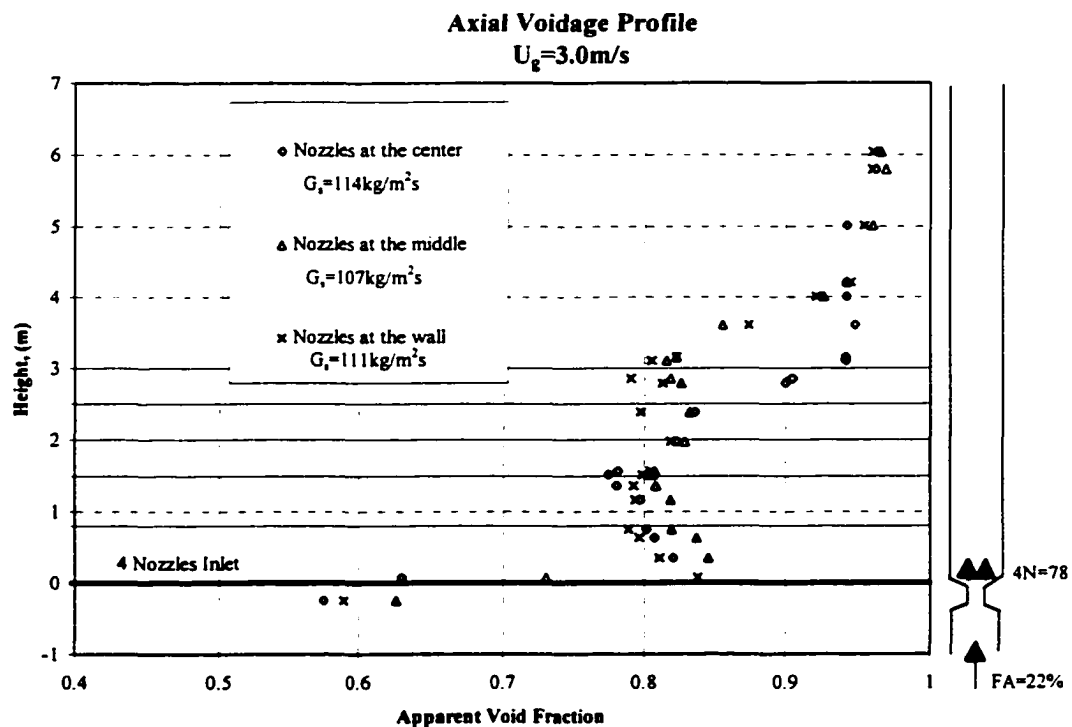


Figure 5.19: Axial Voidage Profile for the Three Nozzle Positions.

is reached. When the nozzles are located at the middle or close to the riser wall this section from 1.5 to 3.0m is a section of strong complex flow. Local acceleration and deceleration sections appear mainly for the case with the nozzles at the riser wall. The height of the inflection point is the same for the wall and middle positions and greater than the height of the inflection point when the nozzles are in the center of the riser. The same behavior is observed for even lower superficial gas velocity. Figure 5.20 shows the axial apparent void fraction distribution for a superficial gas velocity, U_g of 2.1m/s. It is seen that the apparent void fraction is smaller for each nozzle position and the height of the inflection point is increased due to the decrease of the superficial gas velocity. A

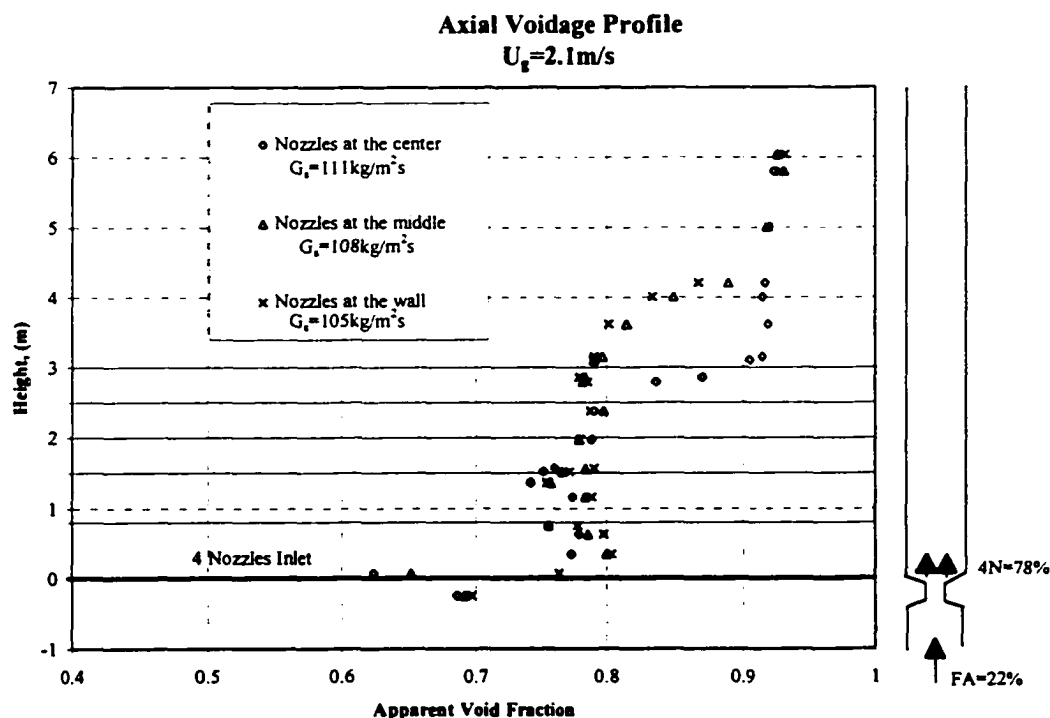


Figure 5.20: Axial Voidage Profile for the Three Nozzle Positions.

complex flow structure occurs at the section from 0 to 2.0m for all nozzle positions. At this section the apparent void fraction increases and decreases rapidly locally.

Another operating parameter in riser flow is the effect of changing the "fluffing" air flowrate at constant total flowrate. Figure 5.21 shows the axial voidage profile for three different "fluffing" air flowrates for a single total superficial gas velocity ($U_g=3.8\text{m/s}$). The nozzles are located at the center of the riser and the solid circulation rate is $G_s=111\text{kg/m}^2\text{s}$. It can be seen that the axial voidage profile appears to have somewhat different behavior for different "fluffing" air flowrate percentages. The most interesting difference in terms of riser performance as a chemical reactor is at the section from above the nozzle up to an elevation of 2.5 L/D. When "fluffing" air percentage is

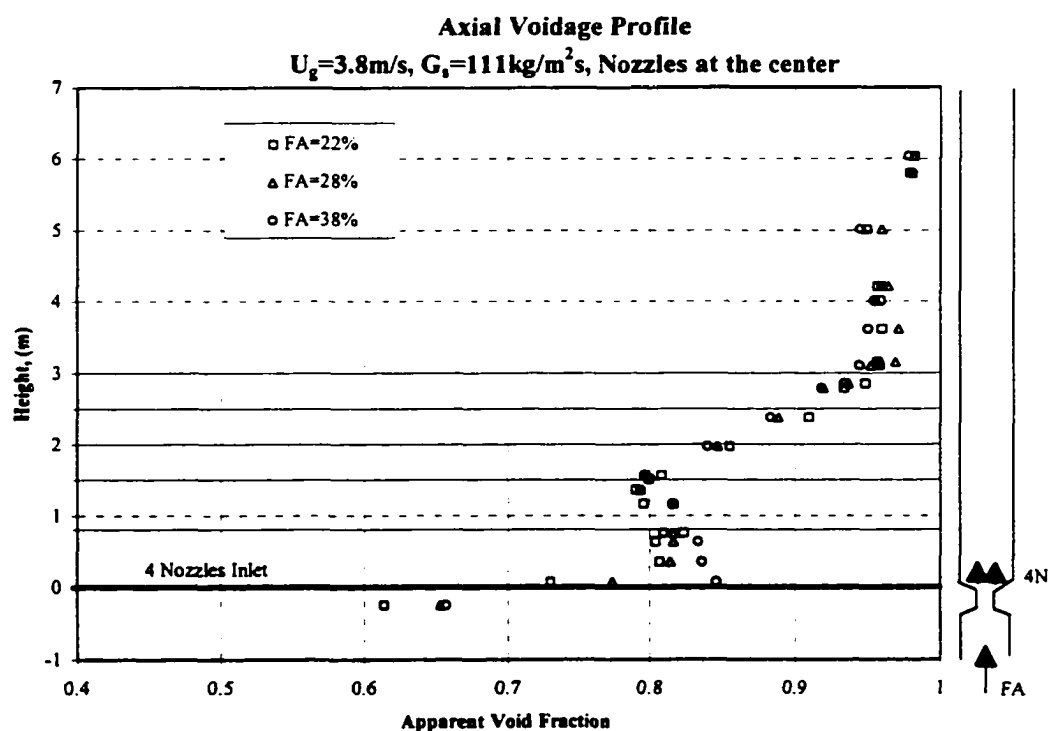


Figure 5.21: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Center.

relative low the bed is very dense at the beginning (height=0) and the profile shows a rapid acceleration section, from apparent void fraction about 0.59 at zero height to an apparent void fraction of about 0.75 at 2.5 L/D. This has, as result, very good gas-solid contact and mixing at this case. On the other hand when the "fluffing" air is relative high, this acceleration almost disappears which results in poorer gas-solid mixing and therefore a lower level of reactor performance.

Figure 5.22 shows the influence of the "fluffing" air flowrate, for the same superficial gas velocity and solid circulation rate, with the four nozzles located at the middle of the riser. It is observed that the difference in the profiles is not significant at the initial section for the changing values of the "fluffing" air flowrates. On the other hand,

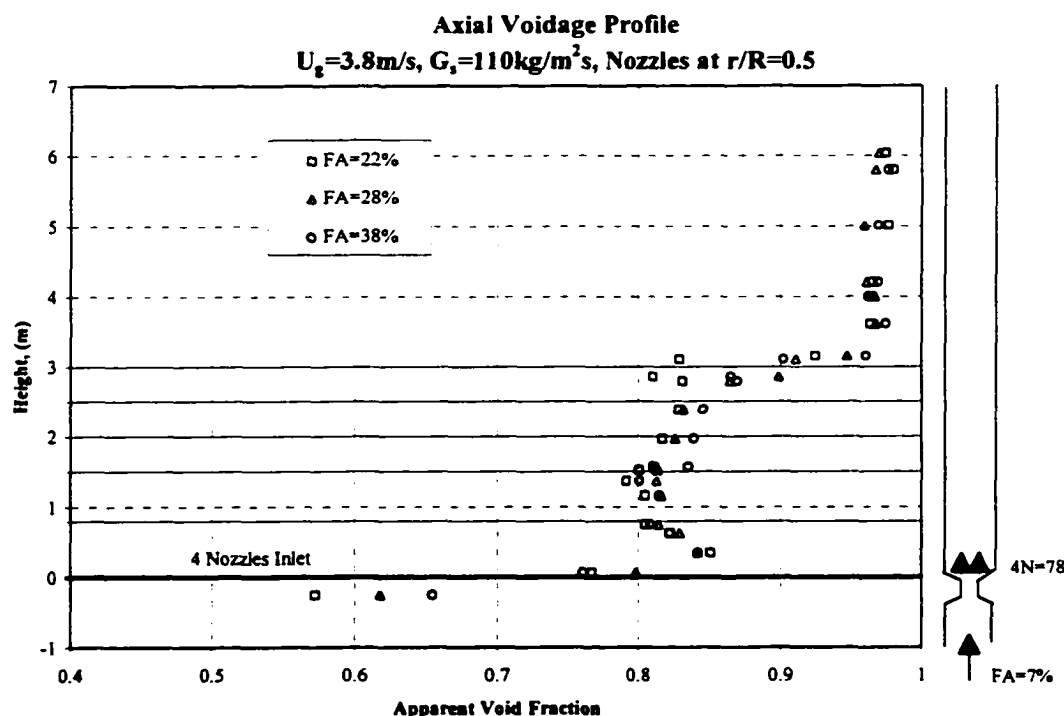


Figure 5.22: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Middle.

different behavior appears at higher elevations, from 10 to 20 L/D. In this section when the "fluffing" air is relative high, a smooth acceleration of solid takes place. When the "fluffing" air is low, the axial voidage profile shows a small deceleration at 20L/D and then a rapid approach to the dilute region.

Finally when the four nozzles are close to the riser wall the results do not show any significant difference. Figure 5.23 shows the axial voidage profile for the nozzles located at the wall. The three axial profiles are almost identical over the entire height of the riser. In the initial section 0-0.75m (0-5 L/D) solid decelerates in the same manner for all the cases but when "fluffing" air is relatively low the apparent void fraction appears to be slightly higher compared with that at the high "fluffing" air flowrates.

The next set of data present the effect of the "fluffing" air flowrate on the axial solid distribution for a superficial gas velocity U_g of 5.1m/s. Figure 5.24 show the axial voidage distribution for a solid circulation rate G_s of 180kg/m²s, with the nozzles located close to the riser axis. It is seen again that the effect of the "fluffing" air flowrate on the axial voidage profile is significant at the initial section from 0 to 1.50m (10L/D). When the "fluffing" air flowrate is relatively low, 22% or 28% of the total flowrate, the apparent void fraction increases from 0 to 0.3m (2 L/D) and a complex behavior follows where the void fraction increased and decreases locally. For the high "fluffing" air flowrate, 38%, the apparent void fraction decreases continuously from 0 to 1.50m, keeping almost the same slope. Above this elevation of 1.5m (10L/D) higher "fluffing" air flowrate causes a slightly increase of the apparent void fraction. The same behavior is observed for lower

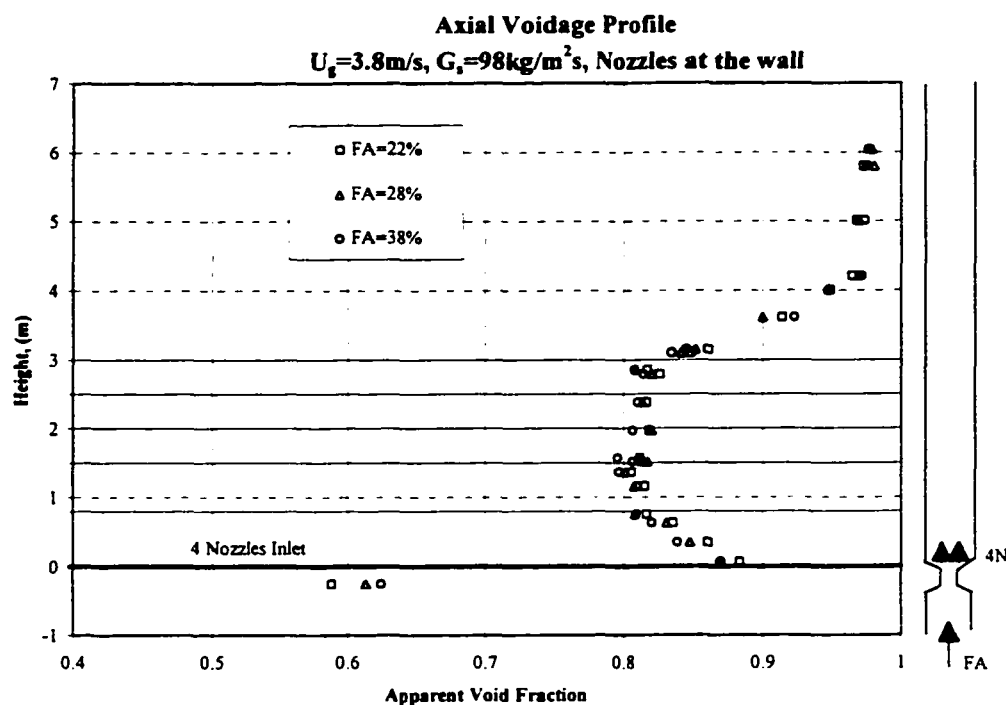


Figure 5.23: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Wall.

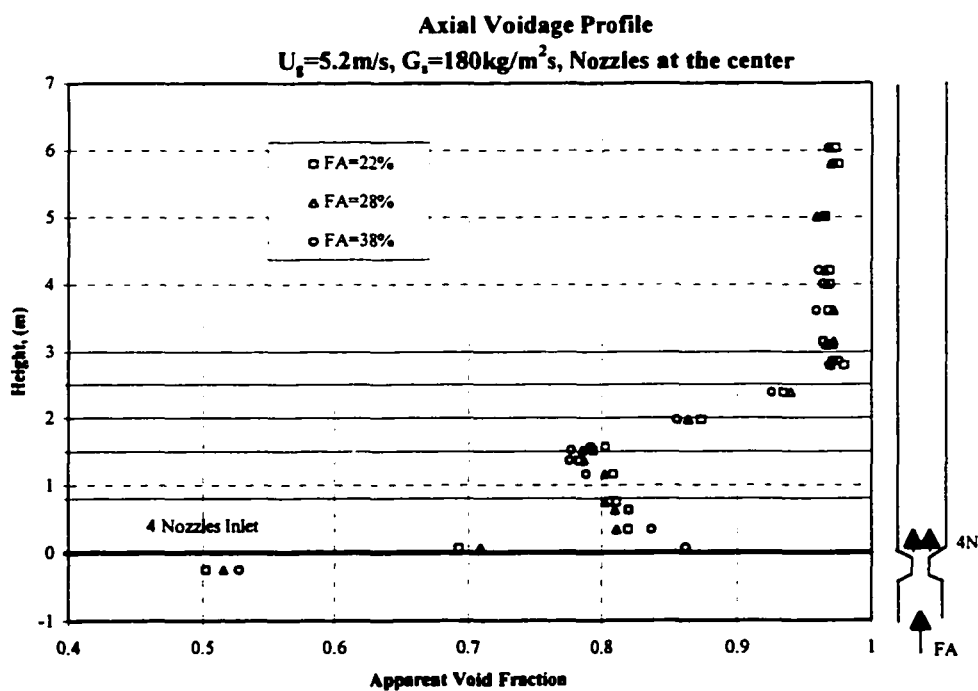


Figure 5.24: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Center.

solid circulation rates ($G_s=110\text{kg/m}^2\text{s}$ and $66\text{kg/m}^2\text{s}$) keeping all the other operating conditions constant, as Figures 5.25 and 5.26 show.

Figures 5.27, 5.28 and 5.29 show the effect of the "fluffing" air flowrate on the axial solid profile when the nozzles are located at the middle of the riser for three different solid circulation rates, and all the other operating conditions the same. It is seen that for this nozzle position the effect of the "fluffing" air is not very significant at the entrance section (0-1.50m). An increase of the "fluffing" air causes a slightly decrease of the apparent void fraction, but this change in the solid behavior is much smaller compared with the case with the nozzles at the center.

The results of the last case with the nozzles located as close as possible to the riser wall, are presented in Figures 5.30, 5.31 and 5.32. It is seen that the effect of the "fluffing" air flowrate is negligible along the whole riser height.

The same general observations can be obtained from experimental results for smaller superficial gas velocities. Figures 5.33, 5.34 and 5.35 show the corresponding results for a gas superficial velocity of 3.0m/s and Figures 5.36, 5.37 and 5.38 for 2.1m/s . It may be seen that for the low superficial gas velocities cases the effect of the "fluffing" air flowrate on the solid distribution at the initial section (0 to 1.50m), is not so strong even when the nozzles are located close to the riser axis.

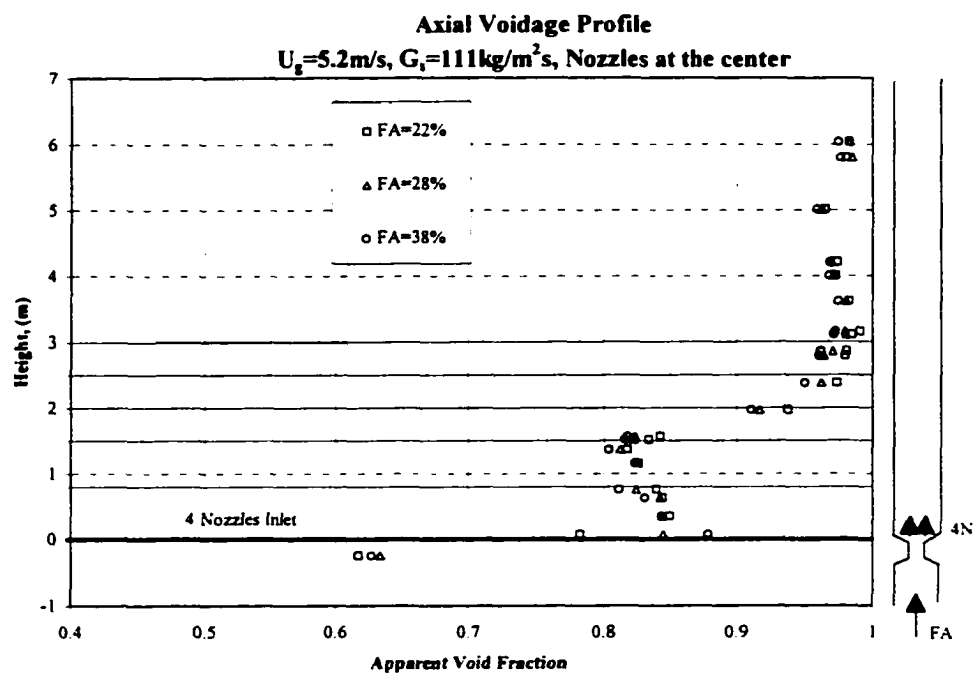


Figure 5.25: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Center.

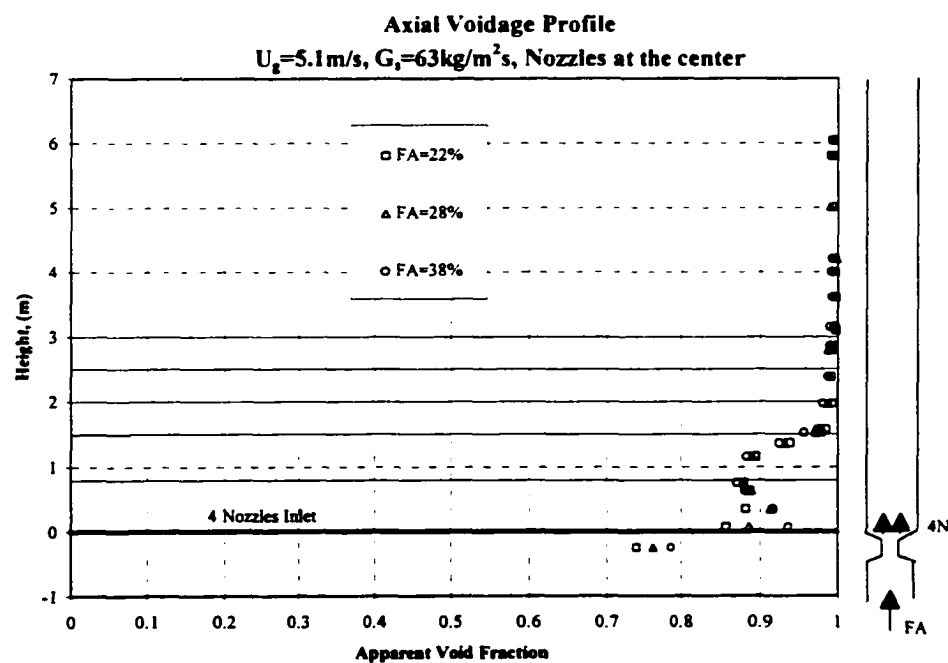


Figure 5.26: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Center.

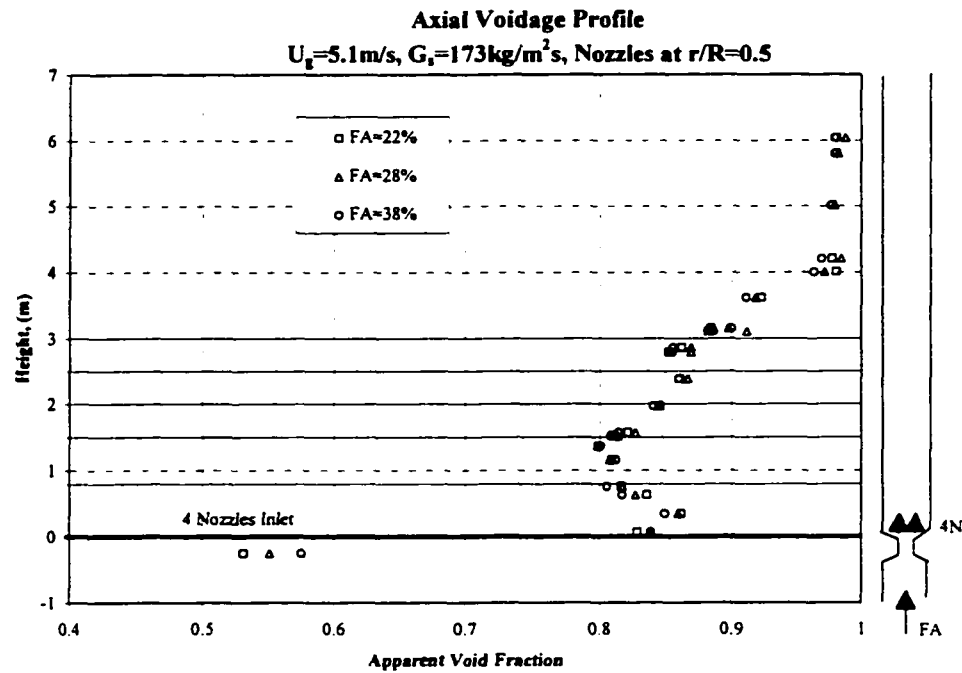


Figure 5.27: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Middle.

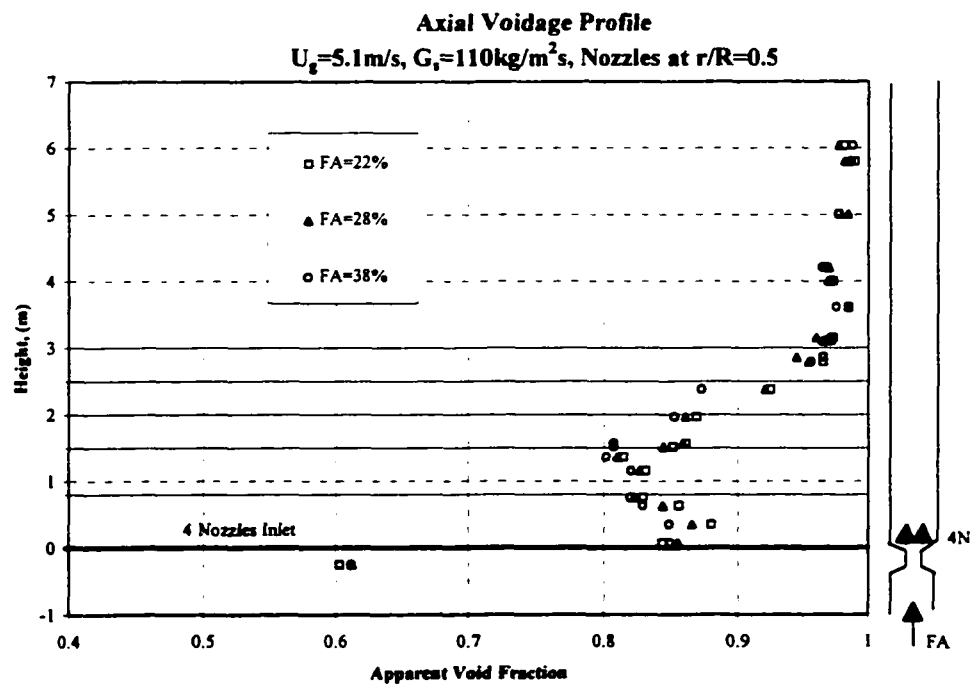


Figure 5.28: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Middle.

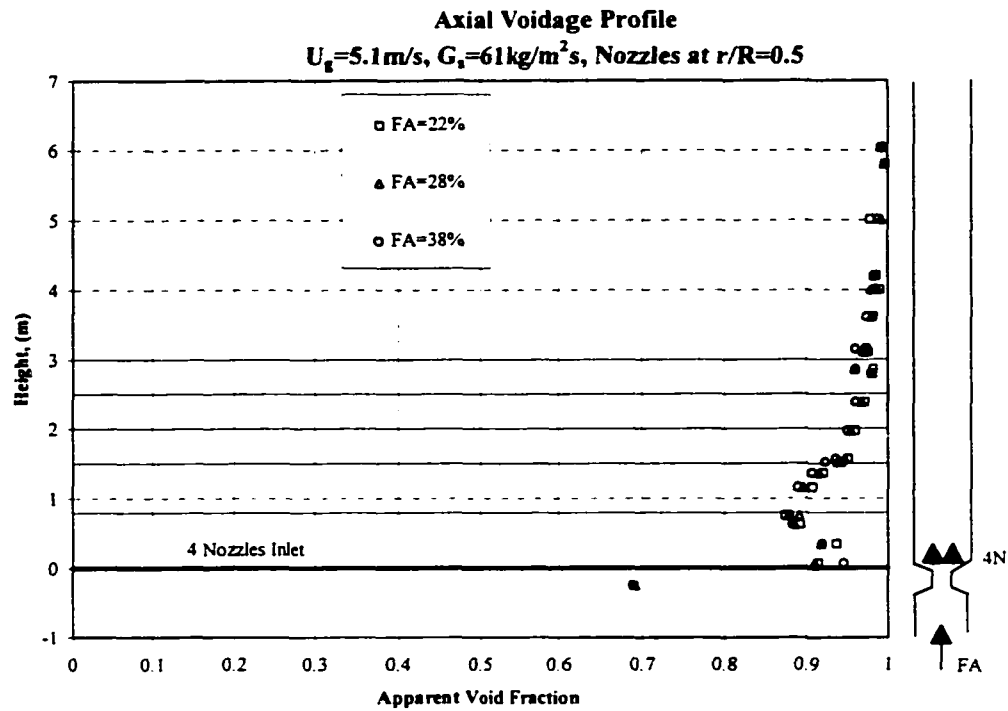


Figure 5.29: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Middle.

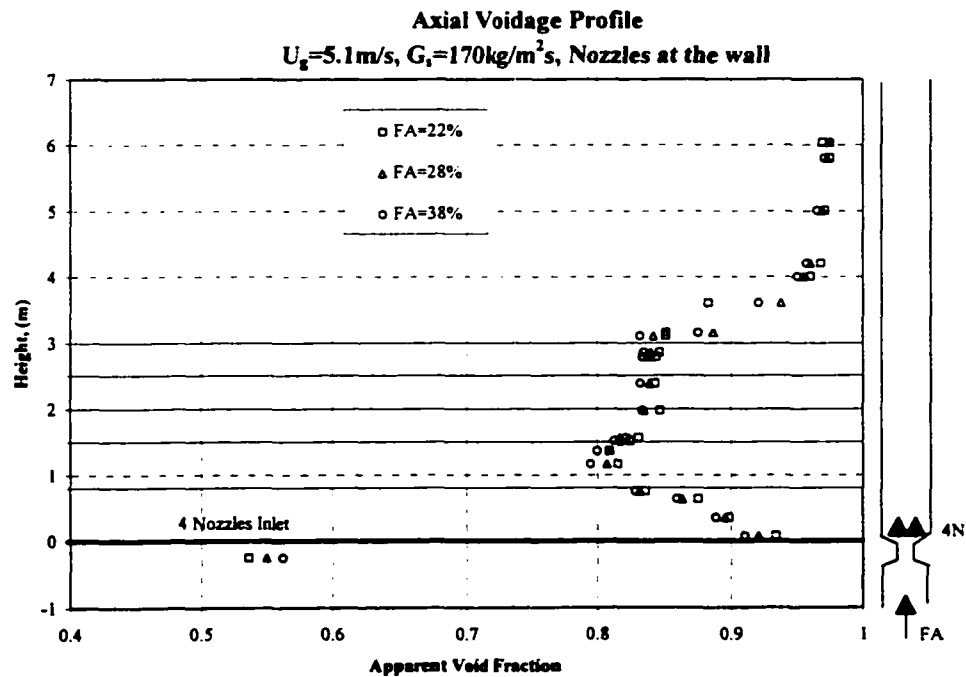


Figure 5.30: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Wall.

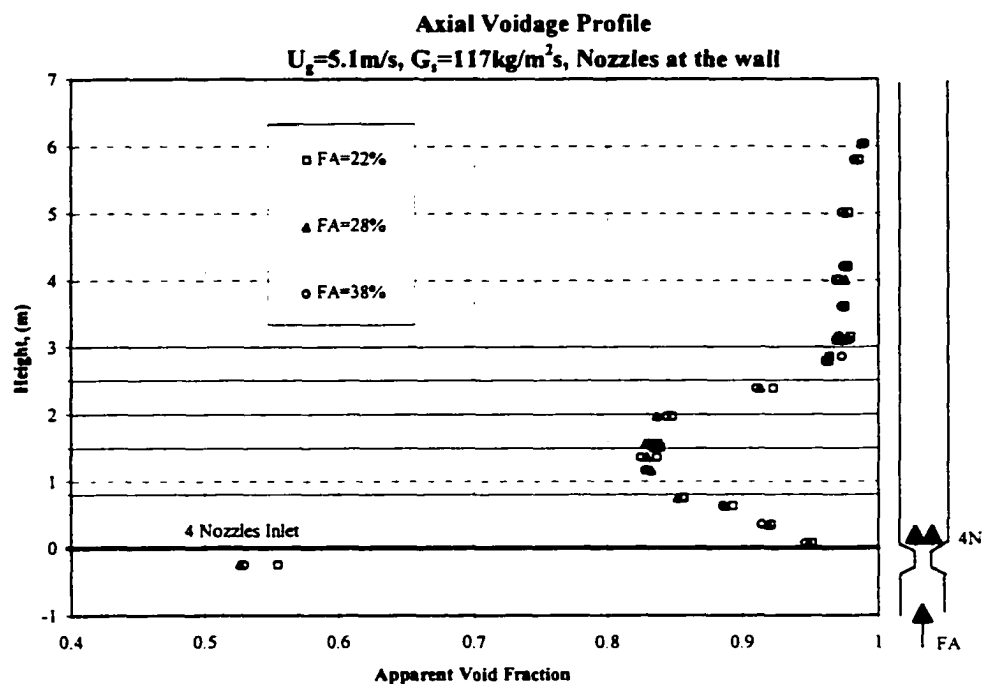


Figure 5.31: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Wall.

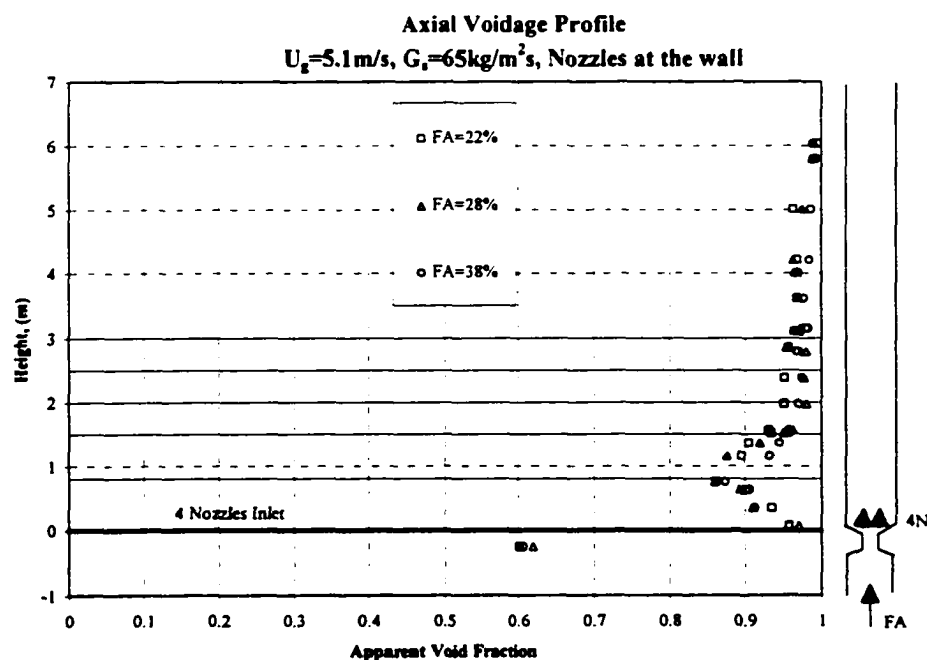


Figure 5.32: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Wall.

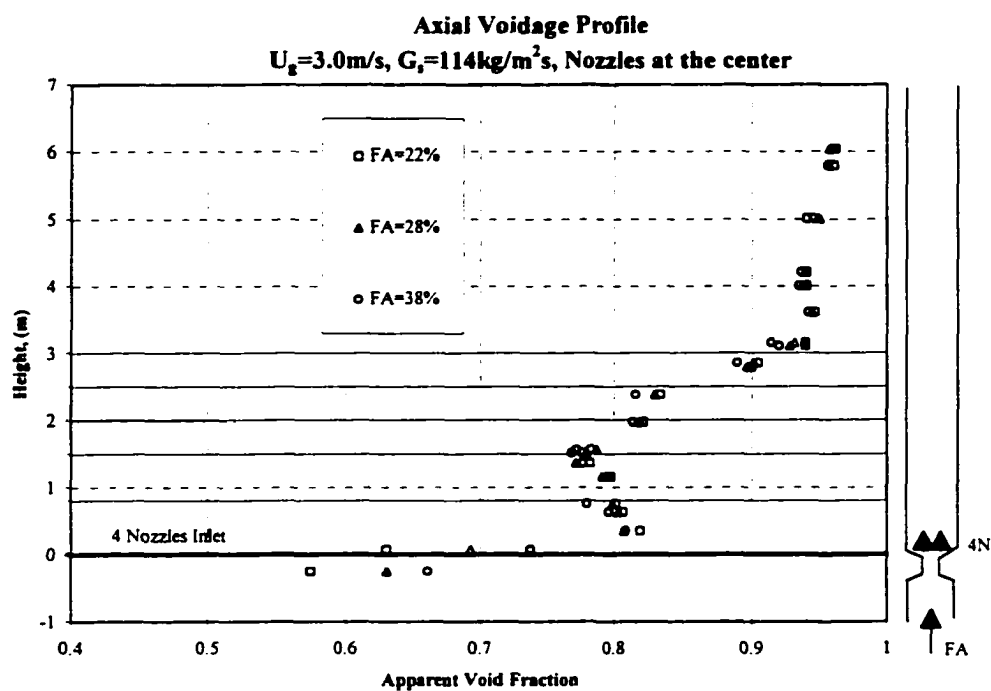


Figure 5.33: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Center.

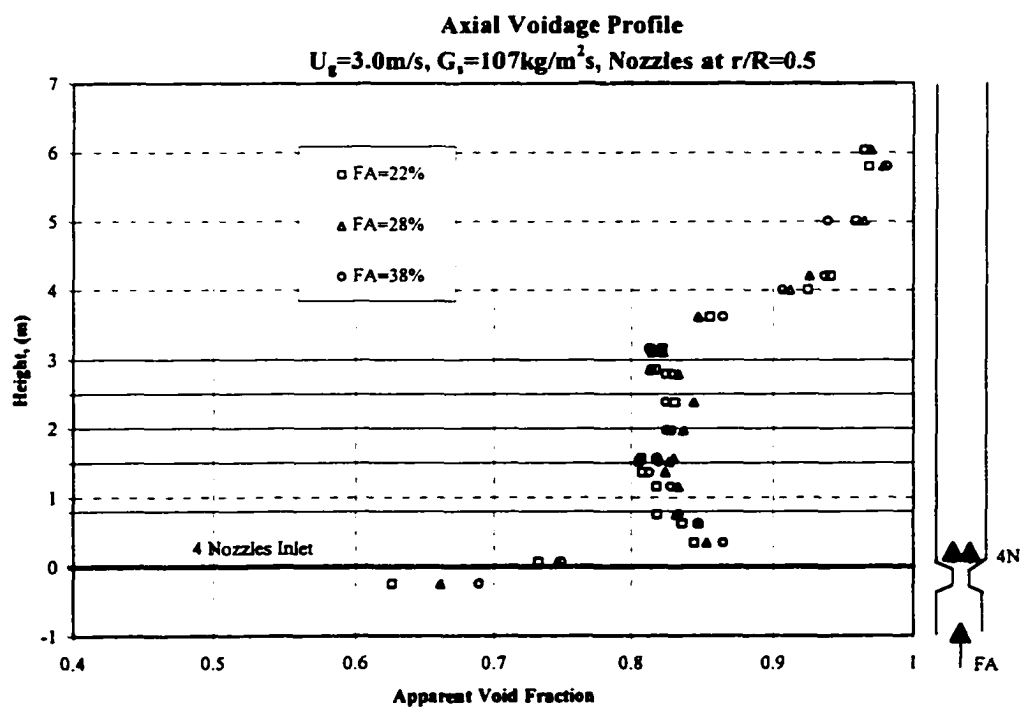


Figure 5.34: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Middle.

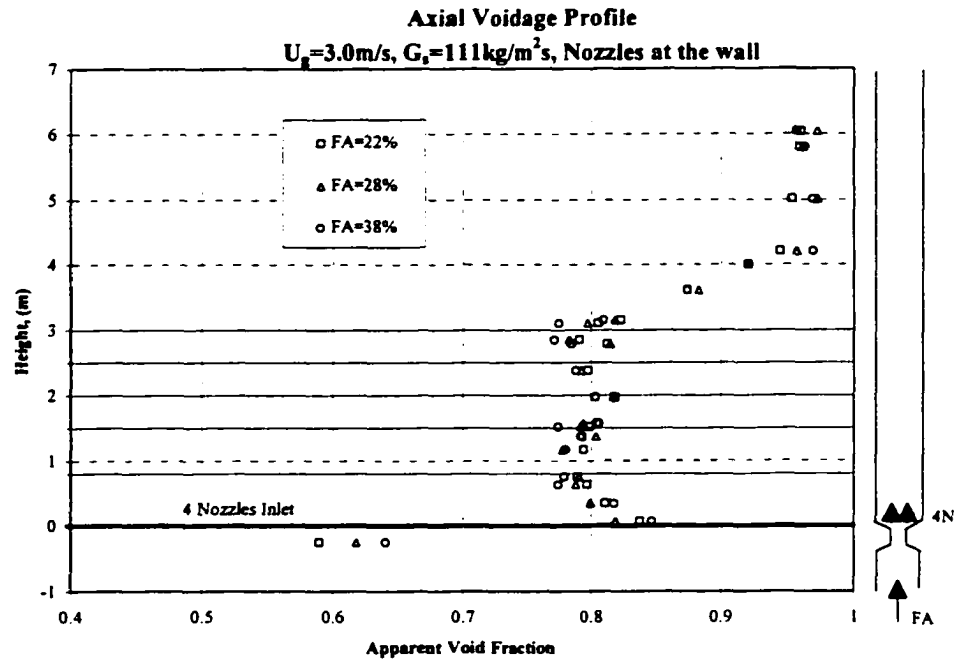


Figure 5.35: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Wall.

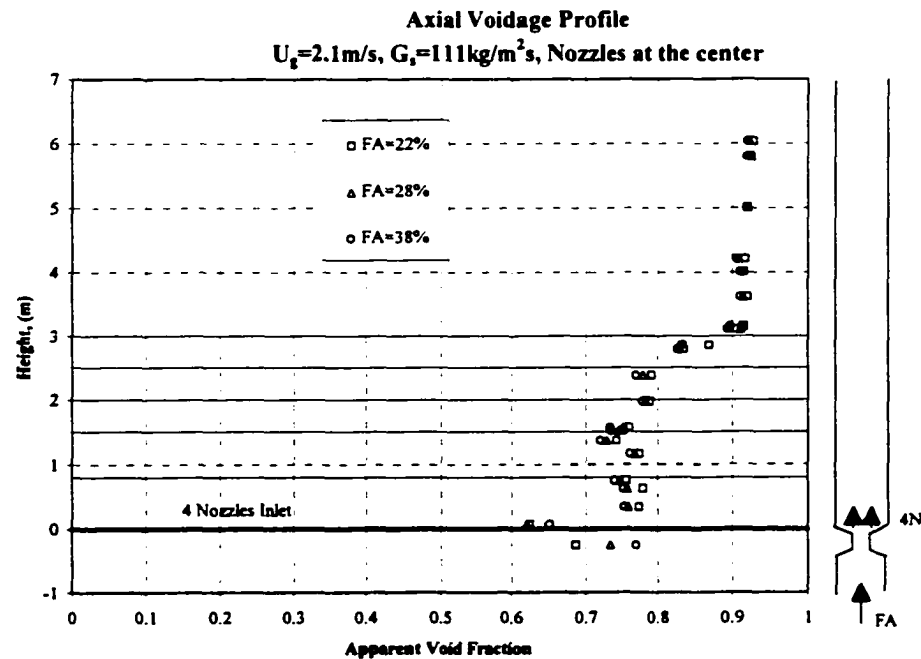


Figure 5.36: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Center.

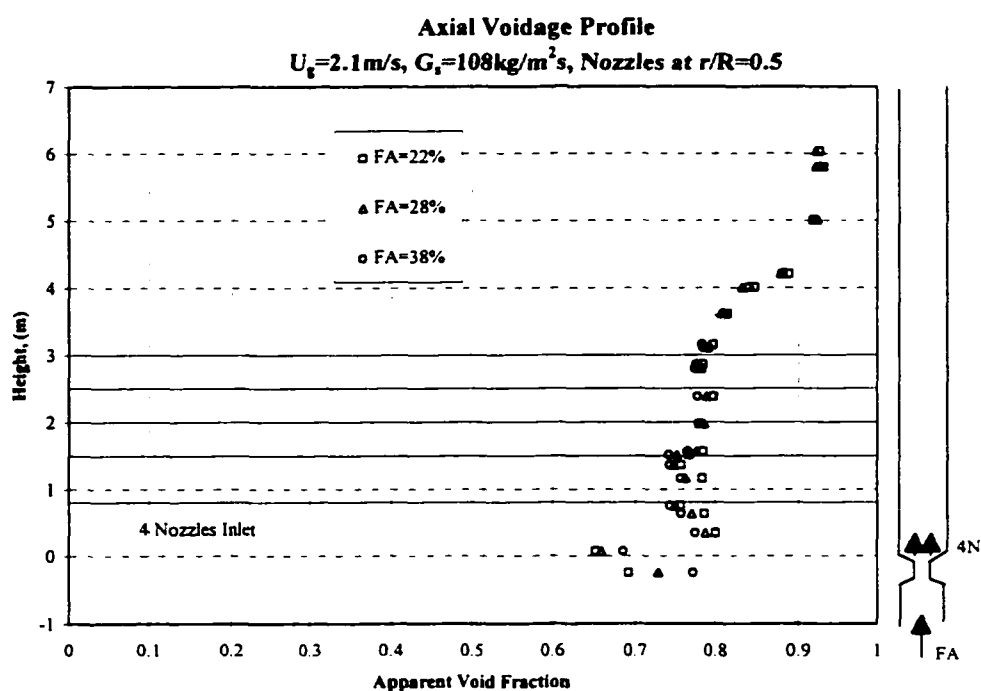


Figure 5.37: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Middle.

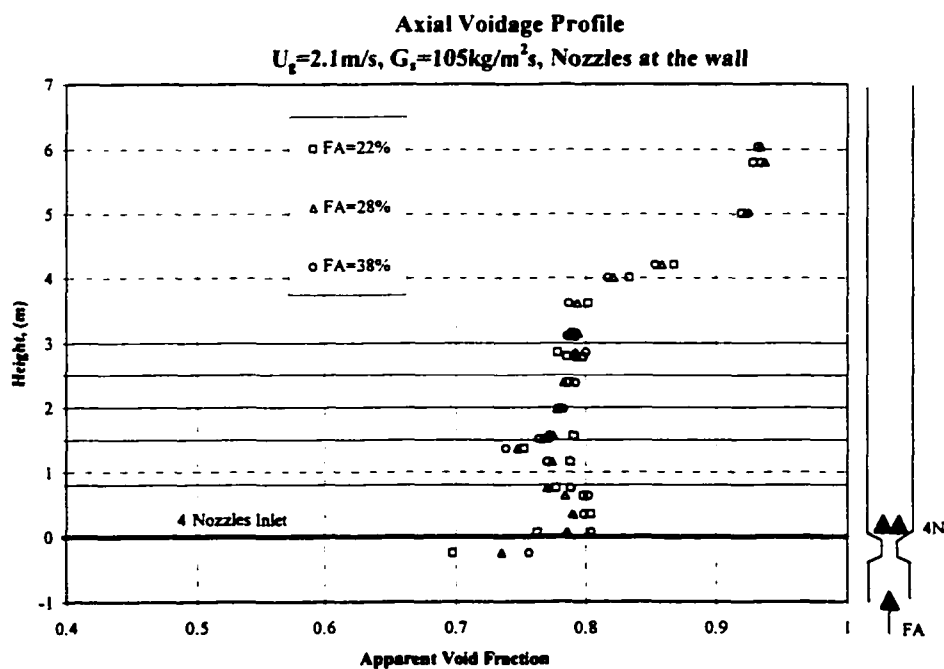


Figure 5.38: Axial Void Fraction as a Function of the "Fluffing" Air Percentage for Nozzles Located at the Wall.

5.2 Radial Solid Concentration Profile

The radial solid density profiles are obtained from the fifteen measured chordal average solid fractions using the x-ray system as described in section 4.4.2. The x-ray measurements were taken at five different elevations in the riser, from 5.3 L/D to 20 L/D above the four nozzle outlets.

Figures 5.39a, 5.39b and 5.39c show the radial solid density profiles at the five elevations for the three different nozzle positions. The superficial gas velocity is $U_g=3.8\text{m/s}$ and the "fluffing" air flowrate is $FA=22\%$ of the total flowrate. The solid circulation rate could not be held absolutely constant and varied from $G_s=98$ to $111\text{kg/m}^2\text{s}$. Figure 5.39a shows the radial solid concentration profiles at five different

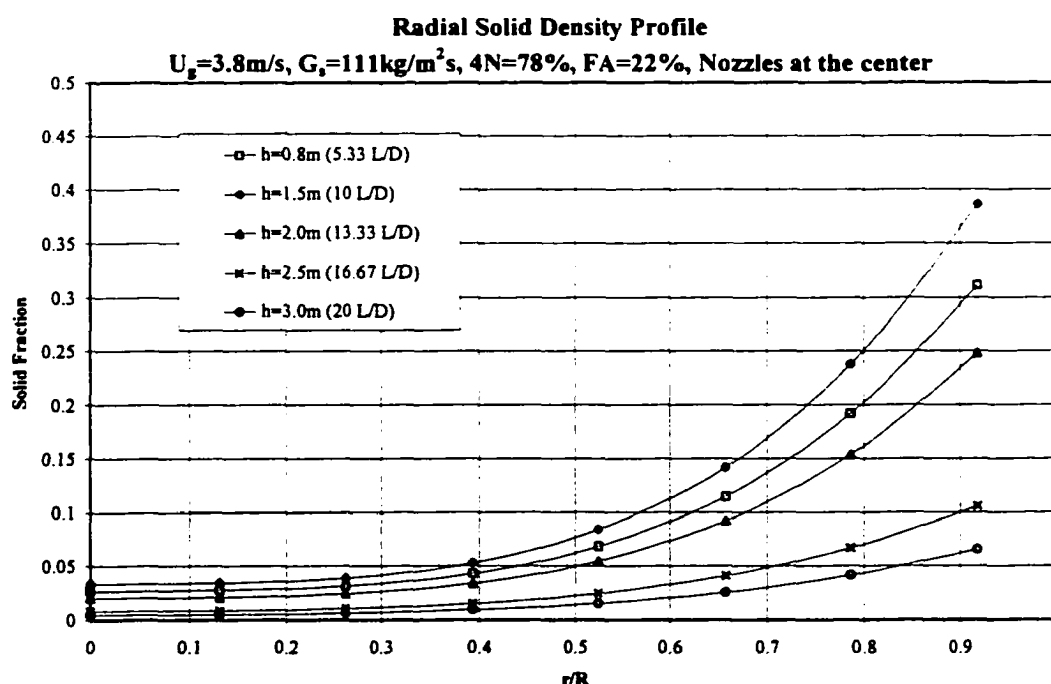


Figure 5.39a: Radial Solid Density Profile at Five Different Elevations for the Nozzles Located at the Center.

elevations when the nozzles are located at the center of the riser. It may be observed that there is a strong radial solid concentration gradient for all the elevations. Catalyst fraction is very high close to the riser wall since the center of the riser is very dilute in catalyst. This gradient indicates poor gas-solid mixing which should result in low performance of the riser as a chemical reactor. It is also clear that the solid fraction at 1.50m (10 L/D) is higher than that at 0.80m (5.3 L/D) along the whole riser radius which is proof of the deceleration that is observed at this section (5.3 L/D to 10 L/D) from the axial voidage profile (Figure 5.1). Above that elevation, the radial solid profile decreases with height since acceleration of solid takes place. Figure 5.39b shows the radial density profile for the same operating conditions but with the four nozzles located at the middle of the riser.

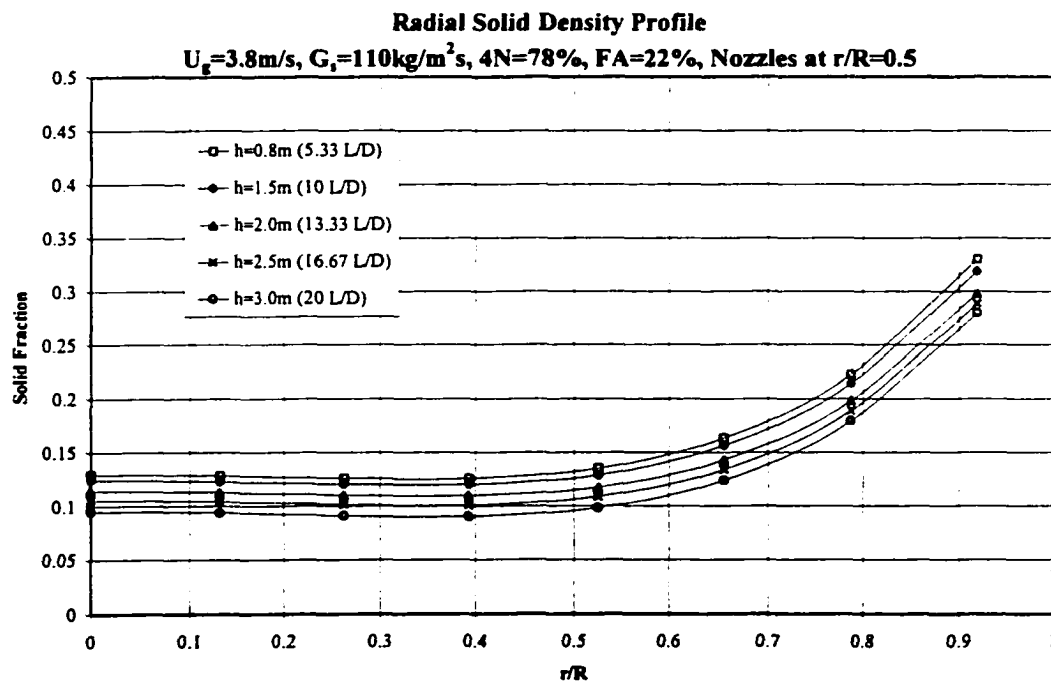


Figure 5.39b: Radial Solid Density Profile at Five Different Elevations for the Nozzles Located at the Middle.

There is also in this case, a solid fraction gradient in the radial direction but it is not as strong. The solid fraction at the center of the riser is much higher than it is with the nozzles located at the center, and is almost constant for a distance up to $r/R=0.55$. The solid fraction increases near the wall but never reaches the high values that occur when the nozzles are in the center. This nozzle position exhibits much better gas-solid mixing. It can also be seen that the radial solid density profile at 0.80m (5.3 L/D) is now slightly higher than the profile at 1.50m (10 L/D). The axial voidage profile (Figure 5.2) shows that this section is approximately at constant solid density. Above this elevation and up to 2.5m (1.67 L/D) a slight decrease of the radial density profile appears. This decrease has the same order of magnitude with the experimental error. At 3.0m (20 L/D) the radial

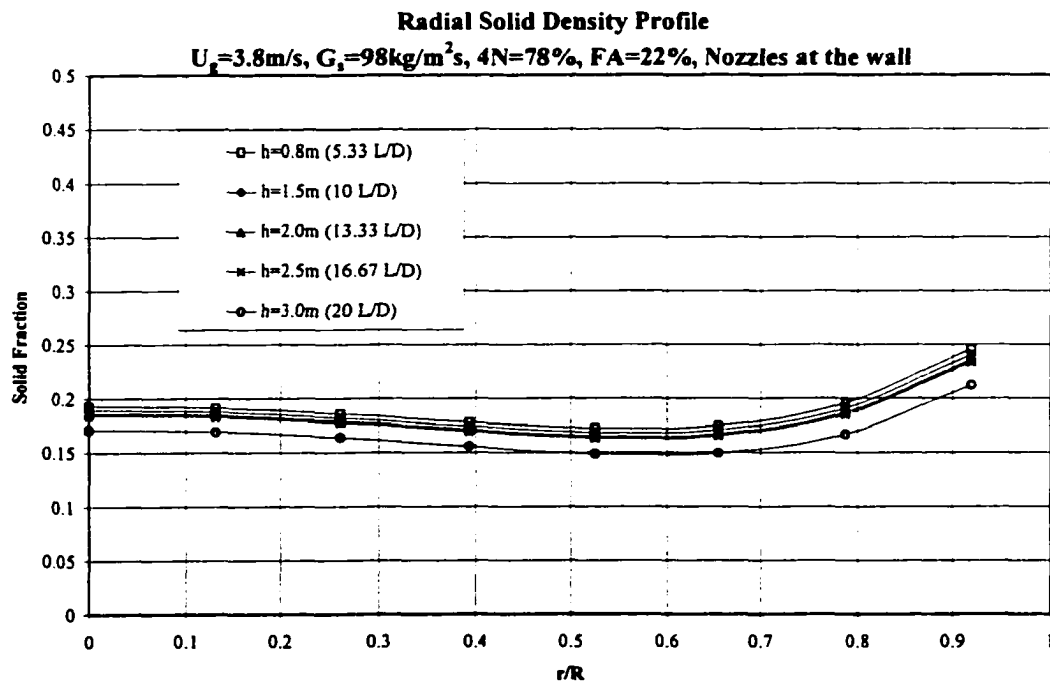


Figure 5.39c: Radial Solid Density Profile at Five Different Elevations for the Nozzles Located at the Wall.

solid fraction decreases which means that an acceleration of solid takes place from 2.5m to 3.0m. For the third case when the four nozzles are located as close as possible to the riser wall, the radial solid density profiles show a much flatter behavior as can be seen from Figure 5.39c. The solid fraction at the centerline has the highest value of the three cases and maintains this value with small variations up to a distance of $r/R=0.8$ from the riser centerline. The solid concentration increases close to the wall but still this increase is much smaller even than that with the nozzles located at the middle. The radial solid profiles are almost the same for the four first elevations (5.3 L/D to 16.7 L/D) and the last at 20L/D is somewhat lower. This is in agreement with the axial voidage profile (Figure 5.3) which show an almost constant density section from 5.3 L/D to 18 L/D where acceleration takes place. This solid distribution in the radial direction is an indication of very good gas-solid mixing and a low degree of gas backmixing which means high performance. Weinstein et al. (1995) used three different nozzle designs to examine the influence in the radial solid density profile. Their results are similar with the results of the present investigation. They showed that when the air was inserted close to the riser wall the radial solid density profile was relatively flat even close to the riser wall. The radial solid density profile in the current work appears to be flat when the nozzles are close to the wall but there still is an increase close to the riser wall. This difference is due to the different nozzle design. The nozzle that was used in the previous studies was designed such as to insert the air into the riser with a velocity vector to carry it to the wall. In this study even when the nozzles are touching the riser wall there is a gap of about 7 mm

between riser wall and air inlet. Both works show that the effect of the nozzle design can be felt up to a height of 20 L/D.

Figures 5.40a, 5.40b and 5.40c show the radial solid density profiles for the lower solid circulation rate case. Figure 5.40a is the case where the nozzles are located at the center of the riser. The radial solid profile shape is very similar to that with the high solid circulation rate. There is a strong gradient in solid concentration in the radial direction which indicates poor gas-solid contact and mixing. It can also be seen that the solid profile at 0.80m (5.3 L/D) is now higher than the profile at 1.50m (10 L/D) which means that the observed deceleration section in the higher solid circulation rate is not present. This is due, as is shown in Figure 5.8, to the decrease in the height of the inflection point.

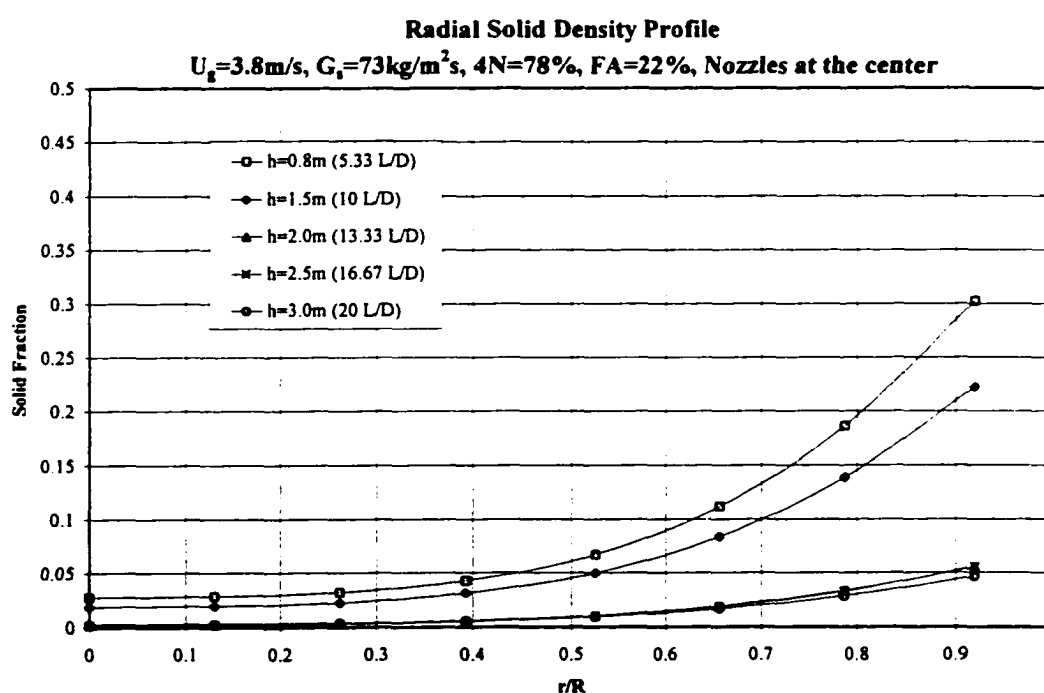


Figure 5.40a: Radial Solid Density Profile at Five Different Elevations for the Nozzles Located at the Center.

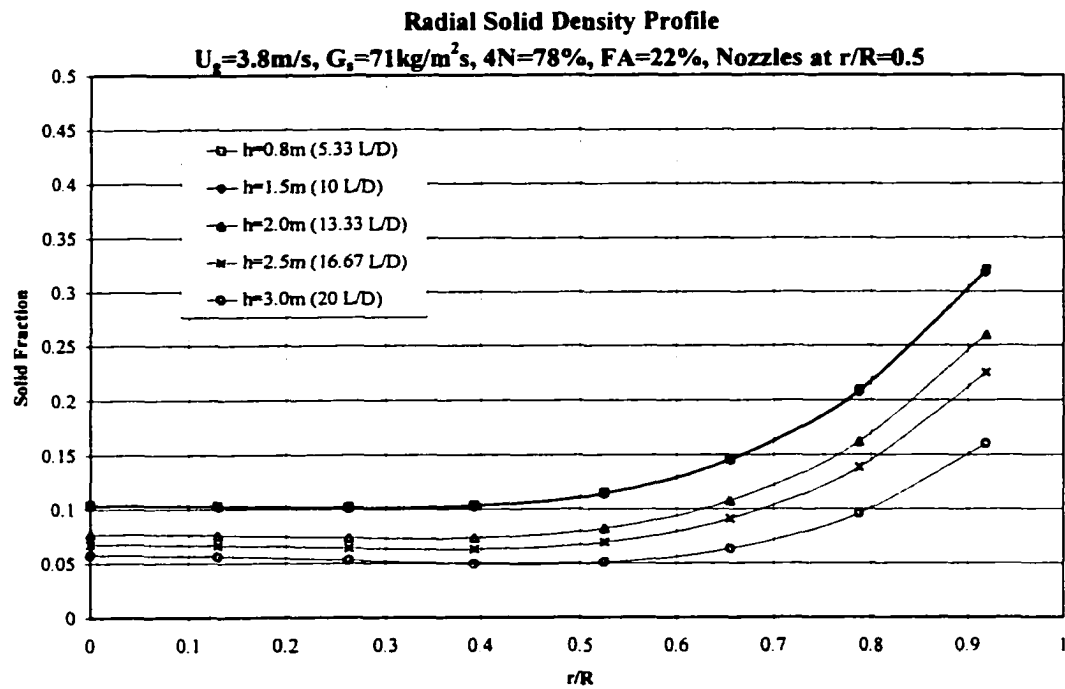


Figure 5.40b: Radial Solid Density Profile at Five Different Elevations for the Nozzles Located at the Middle.

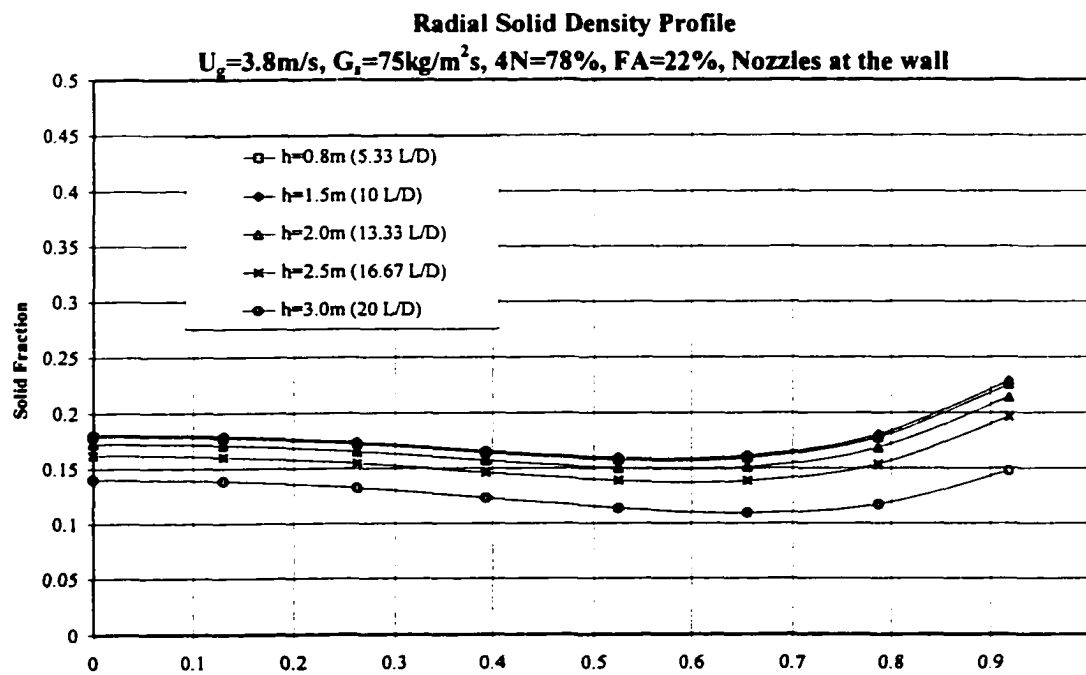


Figure 5.40c: Radial Solid Density Profile at Five Different Elevations for the Nozzles Located at the Wall.

The gas-solid flow above 1.50m (10 L/D) is dilute in solid and the radial solid density profiles at the three higher elevations are almost identical. It is interesting that even in this dilute flow there still is a strong gradient of solid density in the radial direction. The case with the four nozzles located in the middle of the riser is shown in Figure 5.40b. It can be observed that the radial solid profile at 0.80m (5.3 L/D) and 1.50m (10 L/D) are very similar which is an indication of a constant axial density profile in this section. Above 1.5 L/D, acceleration of solid takes place and radial solid density profiles decrease with the height. Figure 5.40c shows the third inlet configuration where the nozzles are as close as possible to the riser wall. It is seen again that this configuration gives the best radial solid distribution since the radial solid fraction is almost constant through the largest part of the riser diameter. The radial solid density profile remains constant up to an elevation of 10 L/D and then decreases with height since acceleration of solid takes place.

Similar are the results of experiments with different superficial gas velocity. Figures 5.41a-c, 5.42a-c and 5.43a-c show the corresponding results for superficial gas velocity 5.1m/s, 3.0m/s and 2.1m/s respectively. The main observation from all these data is the same. The strong radial concentration gradient that exists when the nozzles are at the center of the riser becomes an almost flat distribution when the nozzles are located close to the wall. The complete set of data for each superficial gas velocity is presented at Appendixes A-C. Comparison of Figures 5.39a-c and 5.43a-c show how strong the influence of the inlet configuration is on the radial solid distribution. It is also shown that this influence can be felt up to 20L/D.

Figures 5.44a to 5.44e show the influence of the "fluffing" air flowrate on the radial solid concentration profile for five different elevations of the riser and with the nozzles located at the center. At 0.80m (5.3 L/D) height the radial density profile remains mainly the same for different "fluffing" air percentages except for the section close to the riser wall where solid fraction increases as "fluffing" air decreases (Figure 5.44a). This behavior does not change at the height of 1.50m (10 L/D) (Figure 5.44b). When the acceleration zone section is reached Figure 5.44c shows that the radial density profile is higher when the "fluffing" air flowrate is relative high. That means that the elevation where the acceleration of solid begins is lower as "fluffing" air percentage decreases. At the dilute flow region of the riser the solid fraction at the center independent of the "fluffing" air flowrate but increases close to the wall as "fluffing" air increases (Figure 5.44d and 5.44e).

A different effect of the "fluffing" air on the radial solid distribution is shown when the four nozzles are located at the middle of the riser (Figures 5.45a to 5.45e). At the lowest elevation of 0.80m (5.3 L/D) the solid fraction at the center of the riser increases as the "fluffing" air flowrate decreases. On the other hand the "fluffing" air flowrate does not affect the density profile close to the riser wall. Figures 5.45b to 5.45e show that for higher elevations the solid fraction increases as "fluffing" air decreases along the whole diameter of the riser.

Finally when the nozzles are located close to the riser wall, an increase of the "fluffing" air does not affect the solid profile at the center of the riser but causes an increase of solid fraction close the wall at height of 0.80m (5.3 L/D) (Figure 5.46a). At

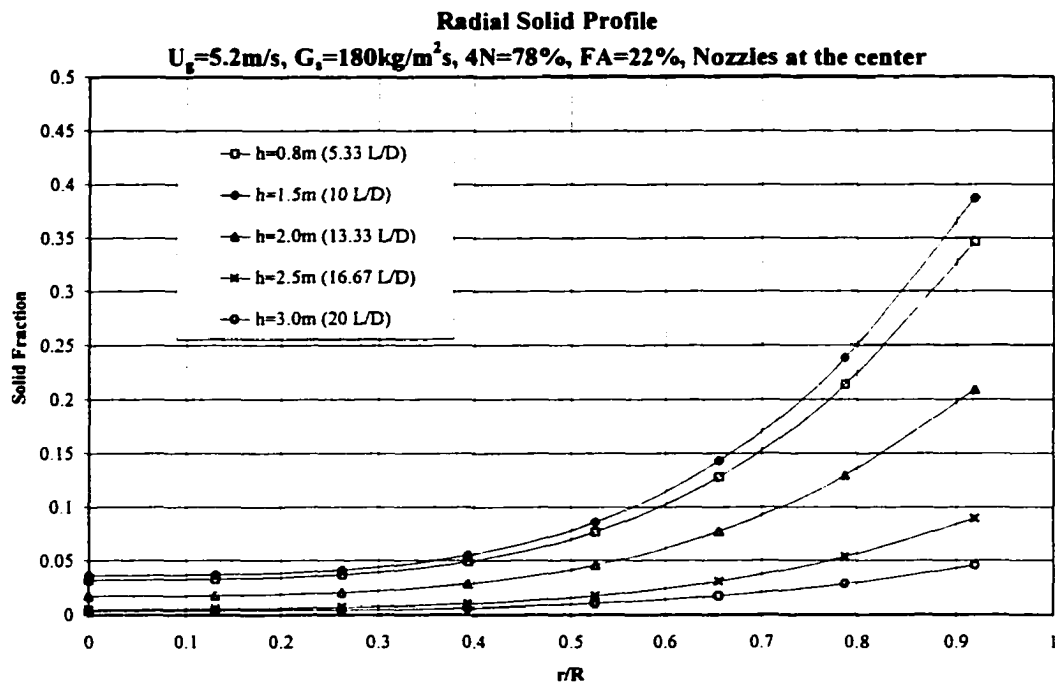


Figure 5.41a: Radial Solid Density Profile for Different Elevations for the Nozzles Located at the Center.

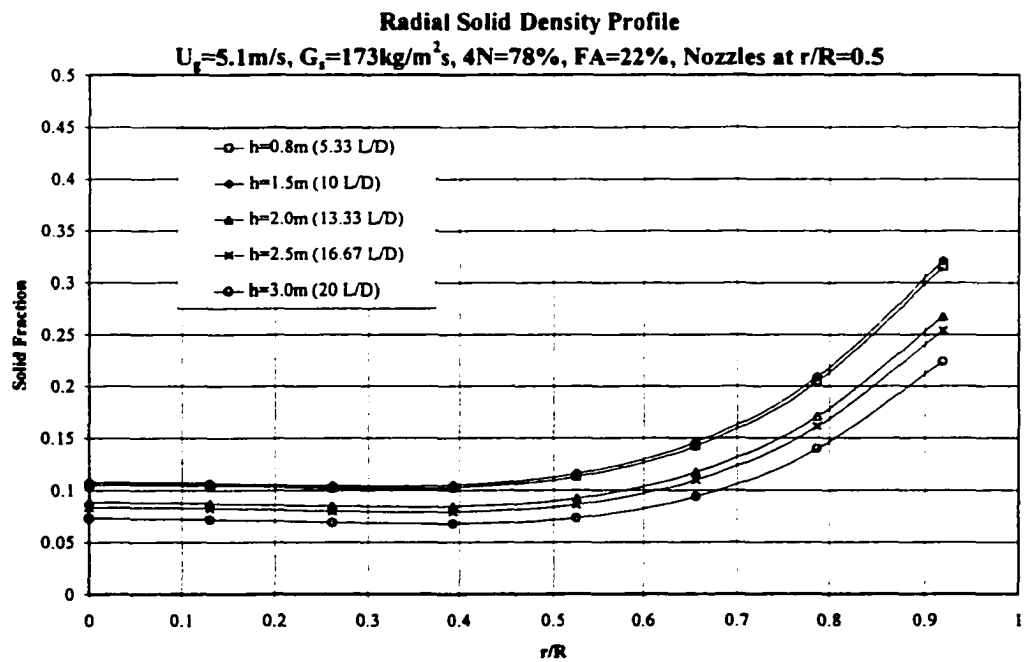


Figure 5.41b: Radial Solid Density Profile for Different Elevations for the Nozzles Located at the Middle.

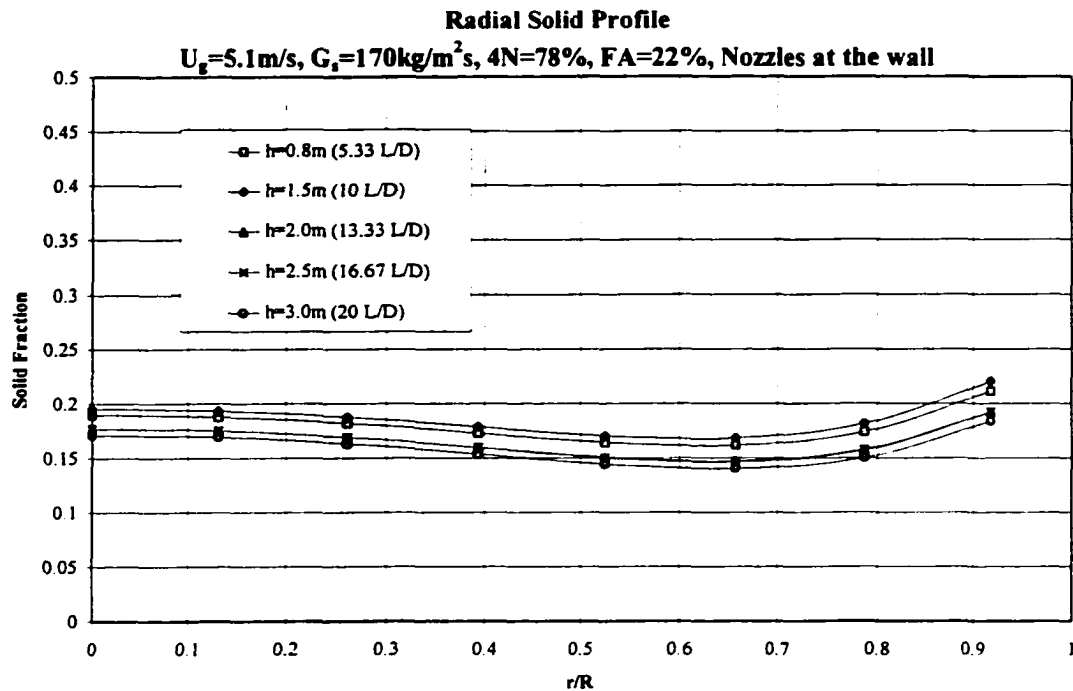


Figure 5.41c: Radial Solid Density Profile for Different Elevations for the Nozzles Located at the Wall.

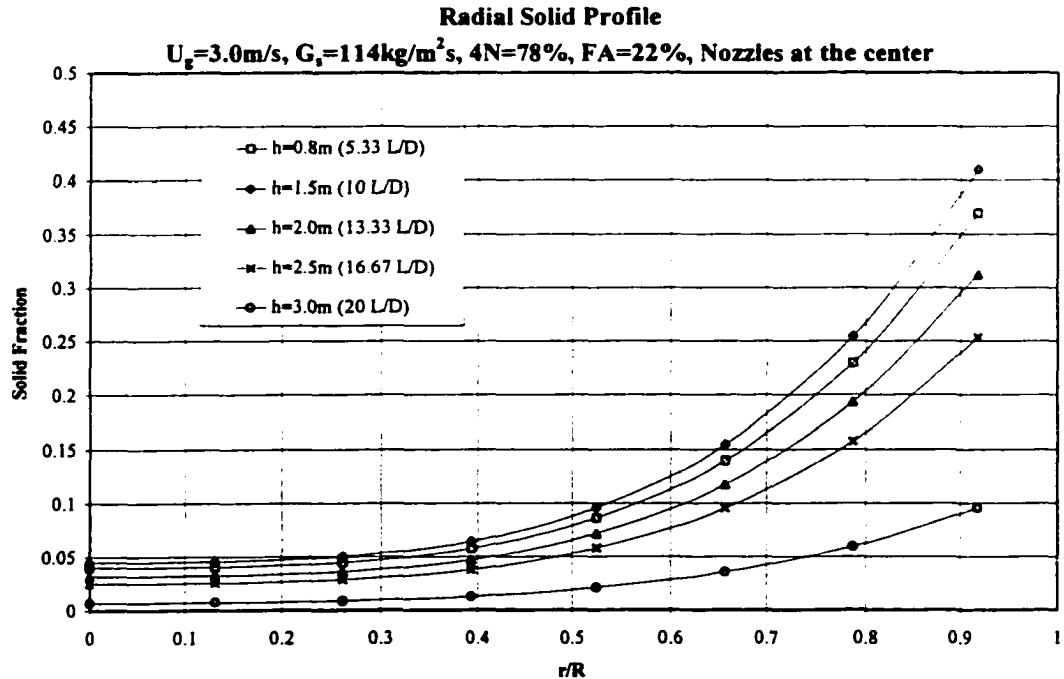


Figure 5.42a: Radial Solid Density Profile for Different Elevations for the Nozzles Located at the Center.

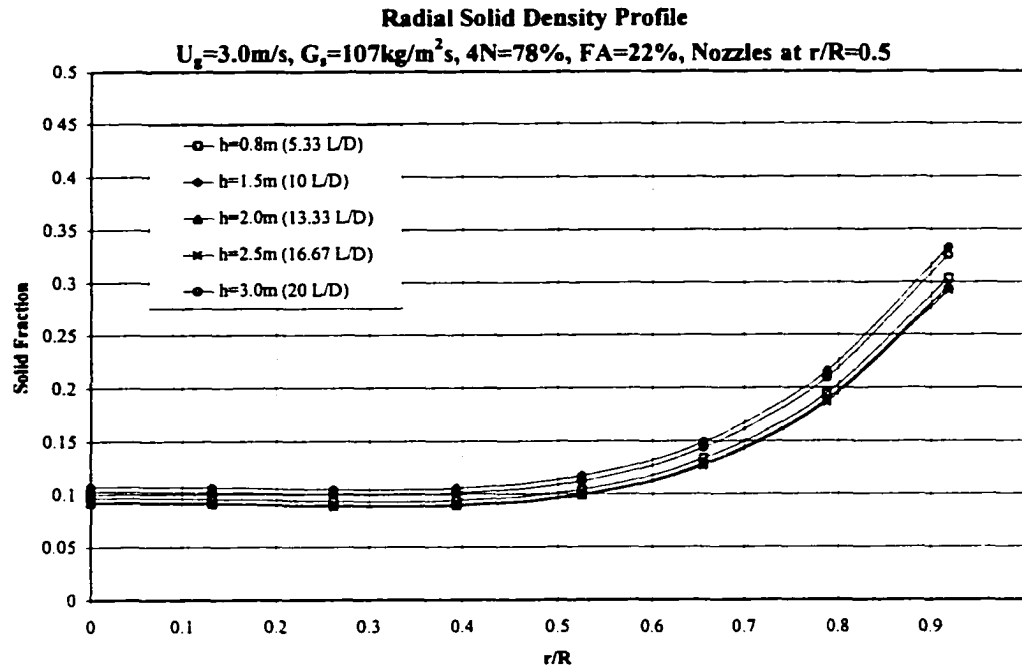


Figure 5.42b: Radial Solid Density Profile for Different Elevations for the Nozzles Located at the Middle.

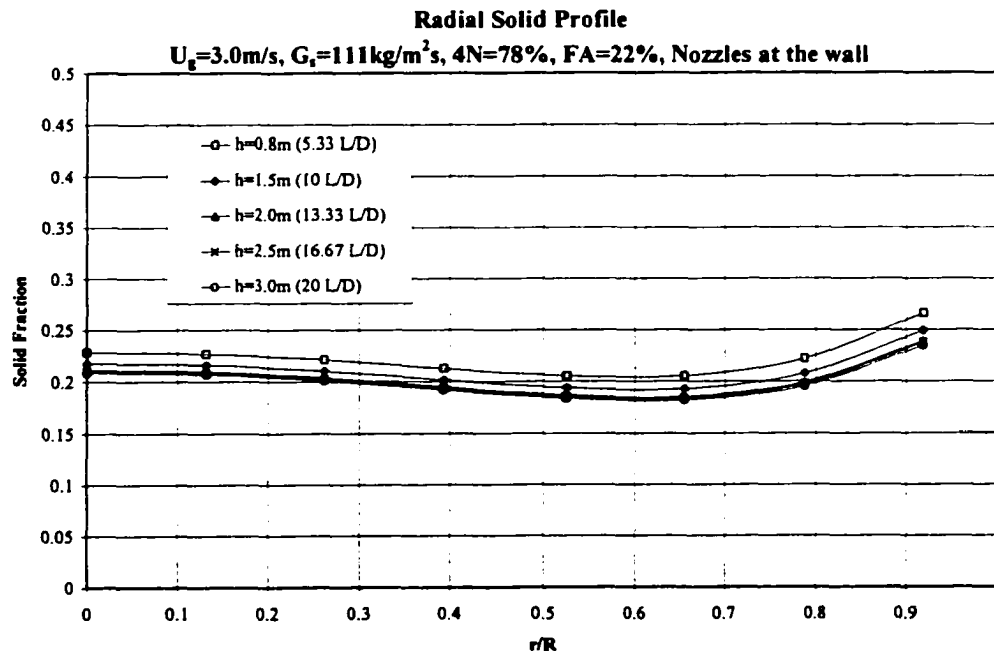


Figure 5.42c: Radial Solid Density Profile for Different Elevations for the Nozzles Located at the Wall.

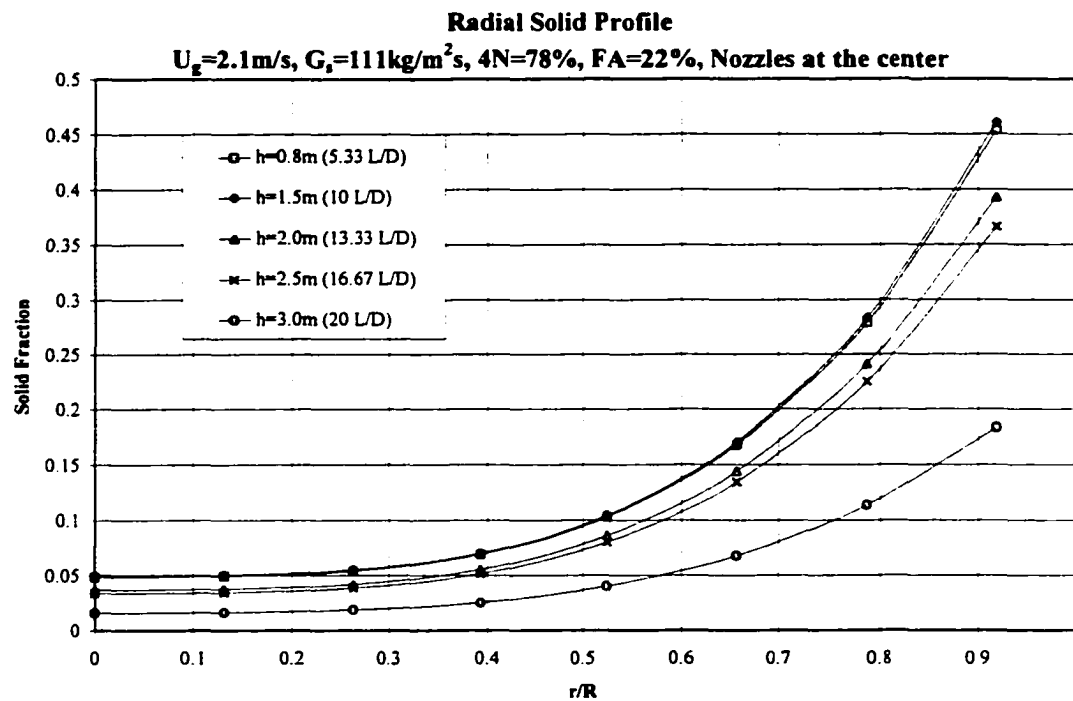


Figure 5.43a: Radial Solid Density Profile for Different Elevations for the Nozzles Located at the Center.

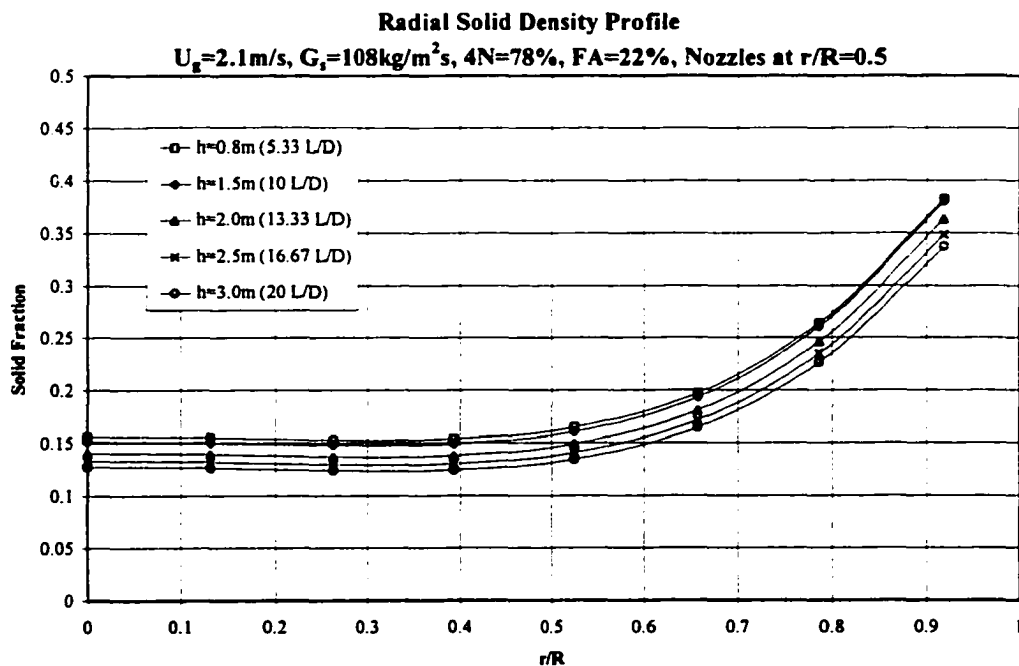


Figure 5.43b: Radial Solid Density Profile for Different Elevations for the Nozzles Located at the Middle.

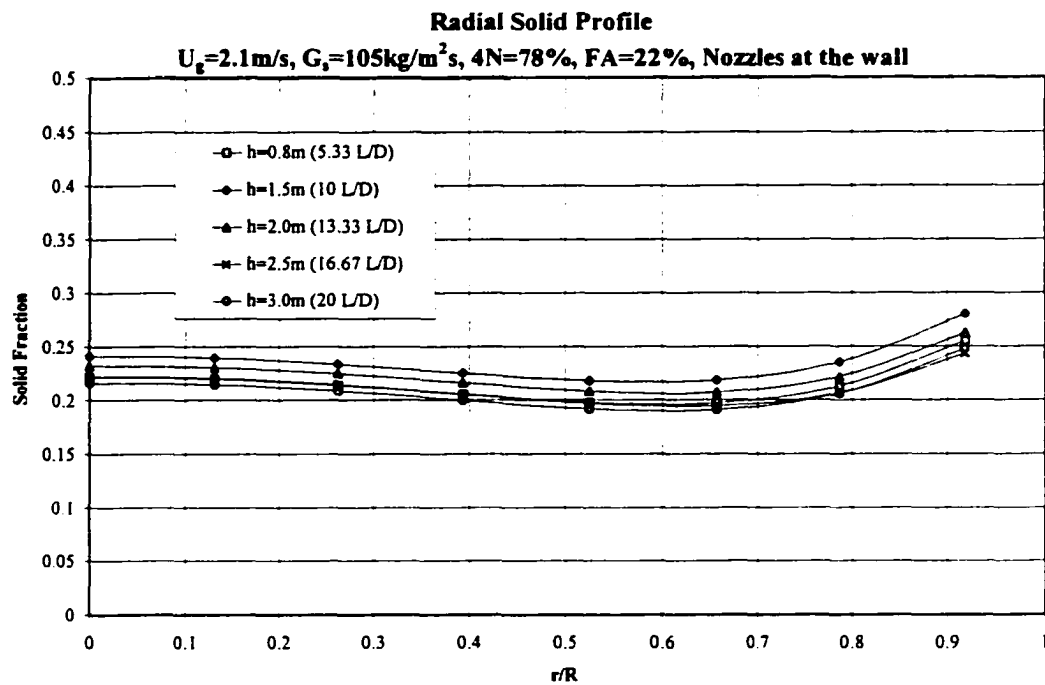


*Figure 5.43c: Radial Solid Density Profile for Different Elevations
for the Nozzles Located at the Wall.*

increasing elevation, the influence of the "fluffing" air becomes negligible. Figures 5.46b to 5.46e show that the solid density profiles appear to be slightly higher when the "fluffing" air is relatively high.

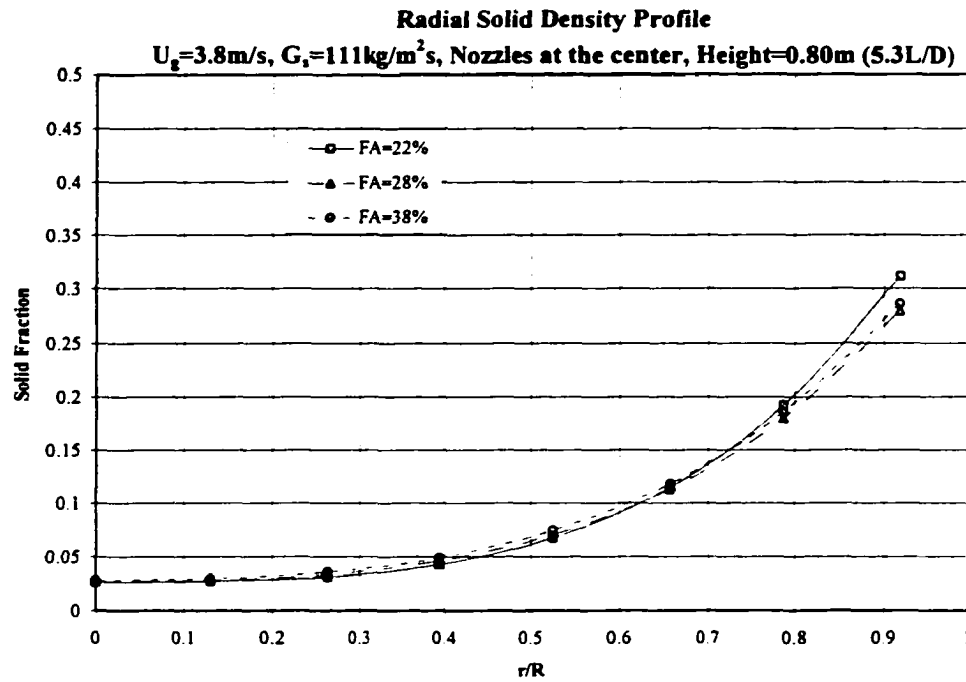


Figure 5.44a: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Center, Height=0.80m.

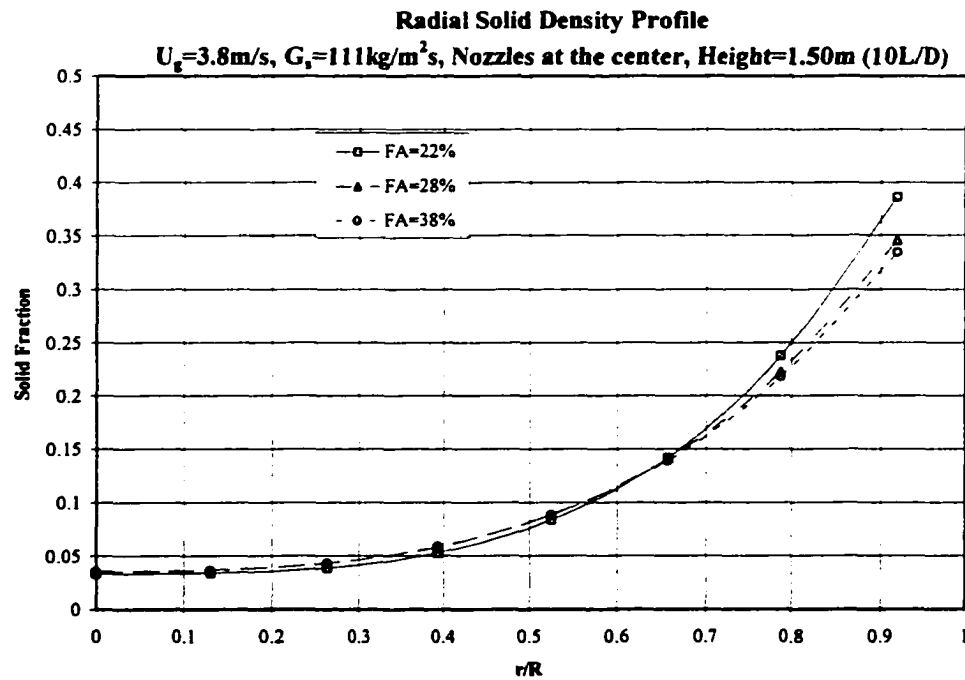


Figure 5.44b: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Center, Height=1.50m.

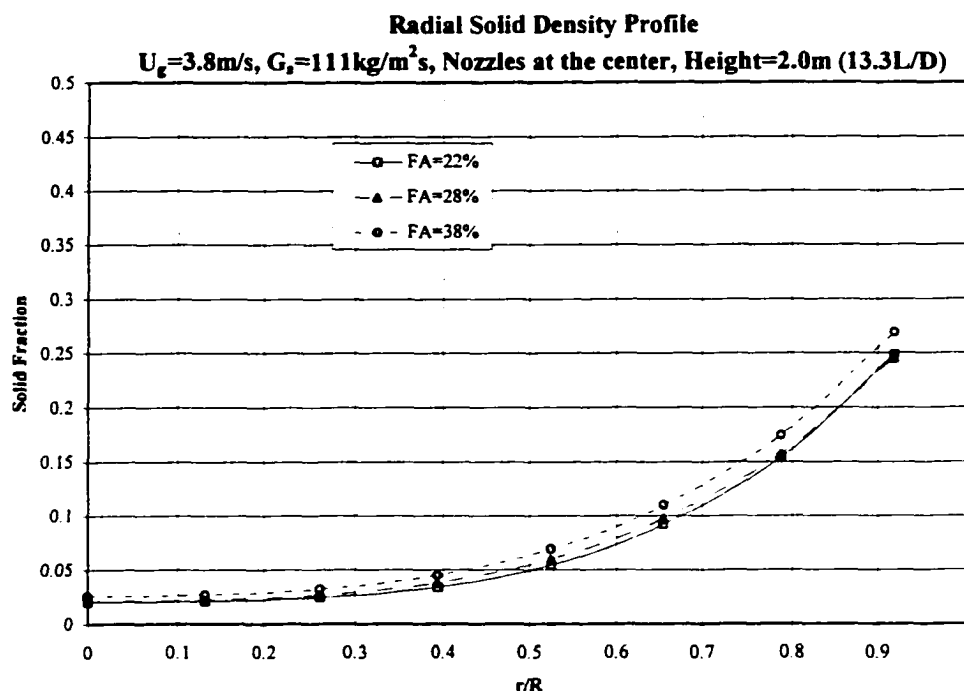


Figure 5.44c: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Center, Height=2.0m.

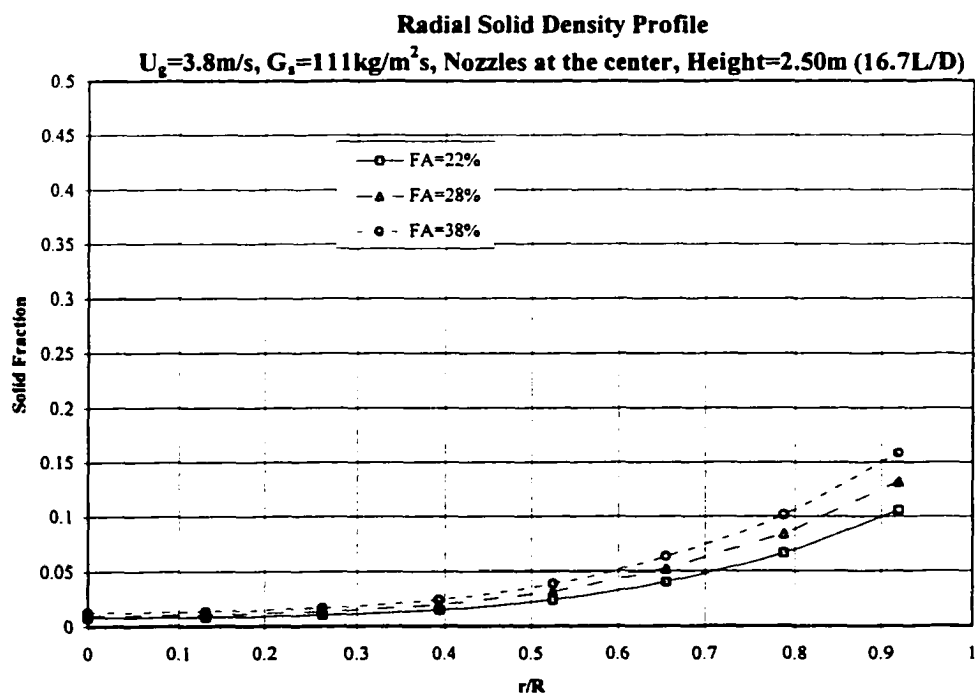


Figure 5.44d: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Center, Height=2.50m.

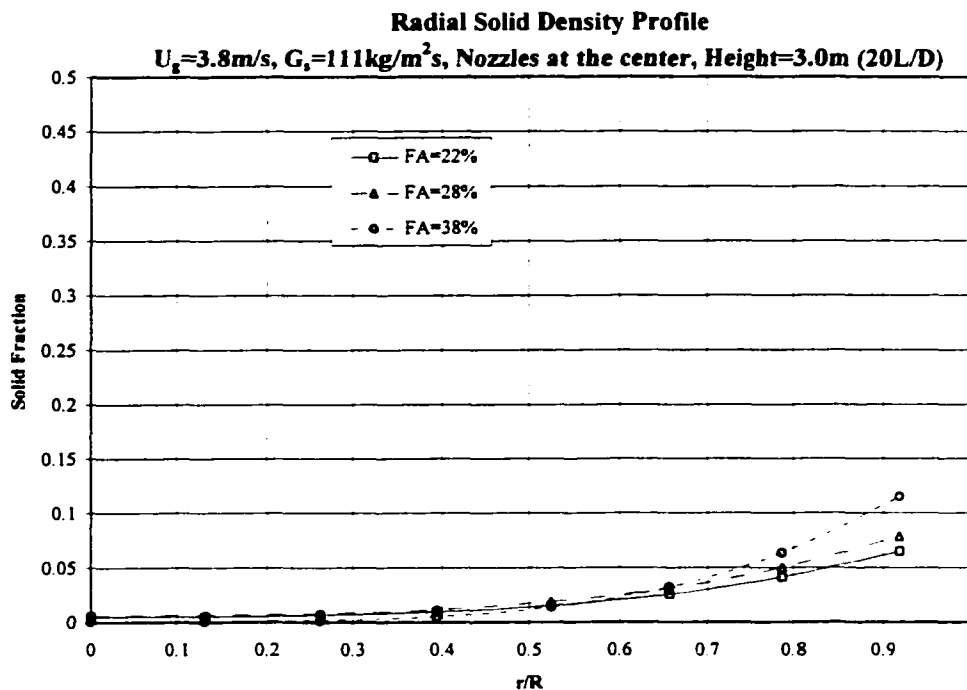


Figure 5.44e: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Center, Height=3.0m.

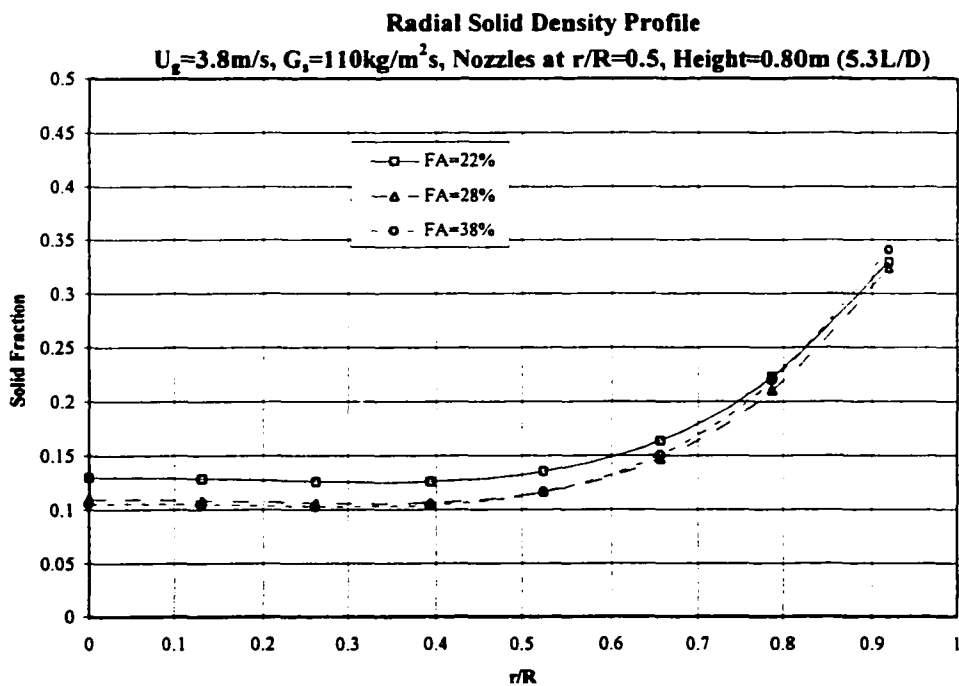


Figure 5.45a: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Middle, Height=0.80m.

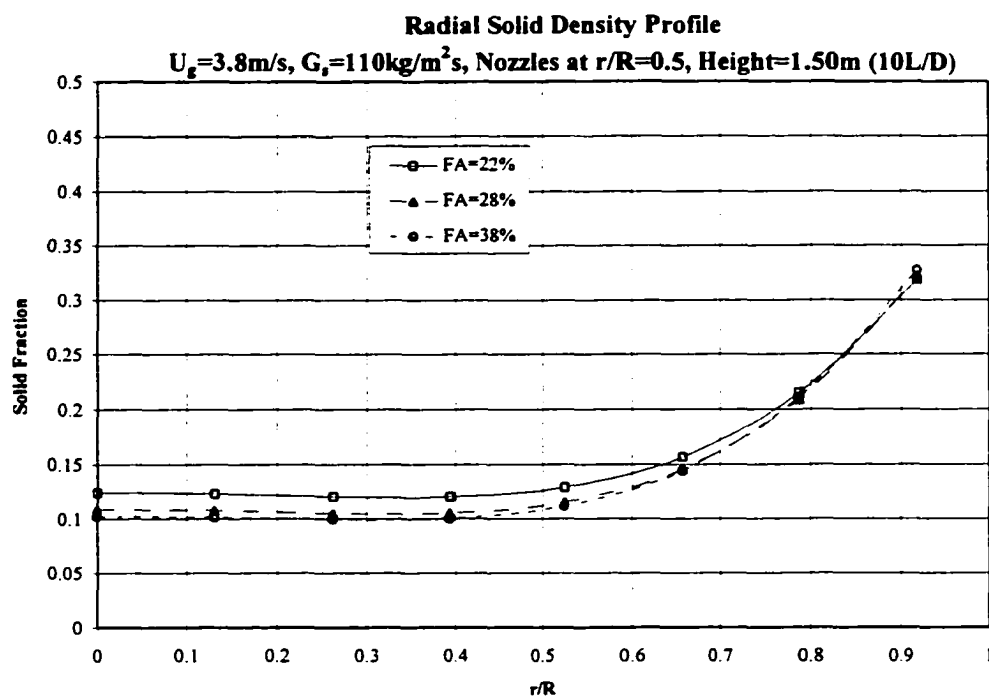


Figure 5.45b: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Middle, Height=1.50m.

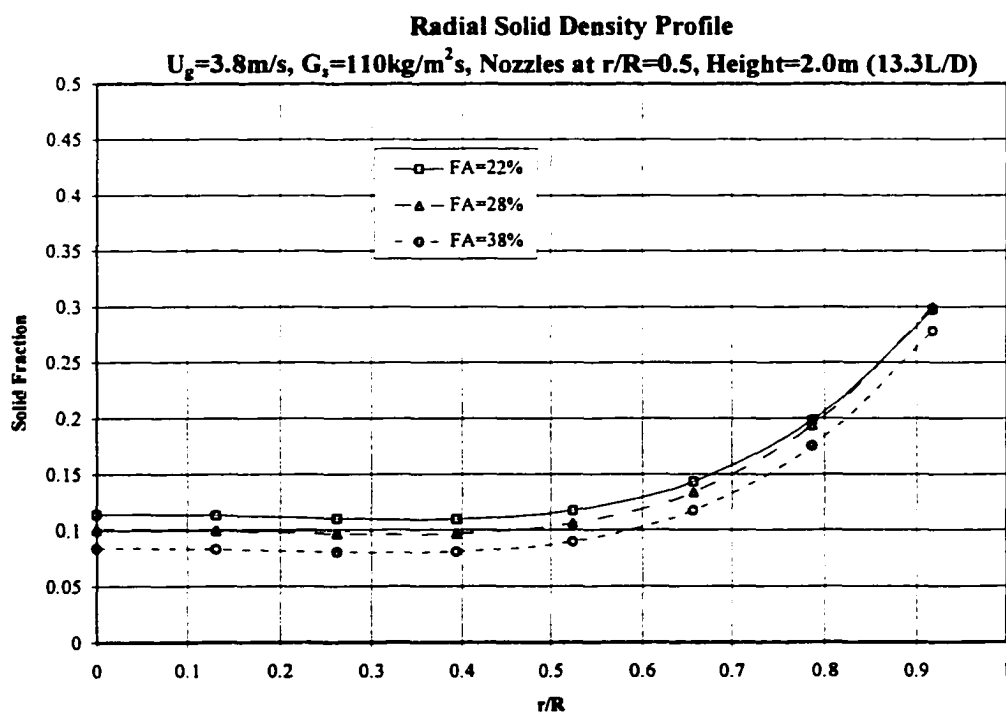


Figure 5.45c: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Middle, Height=2.0m.

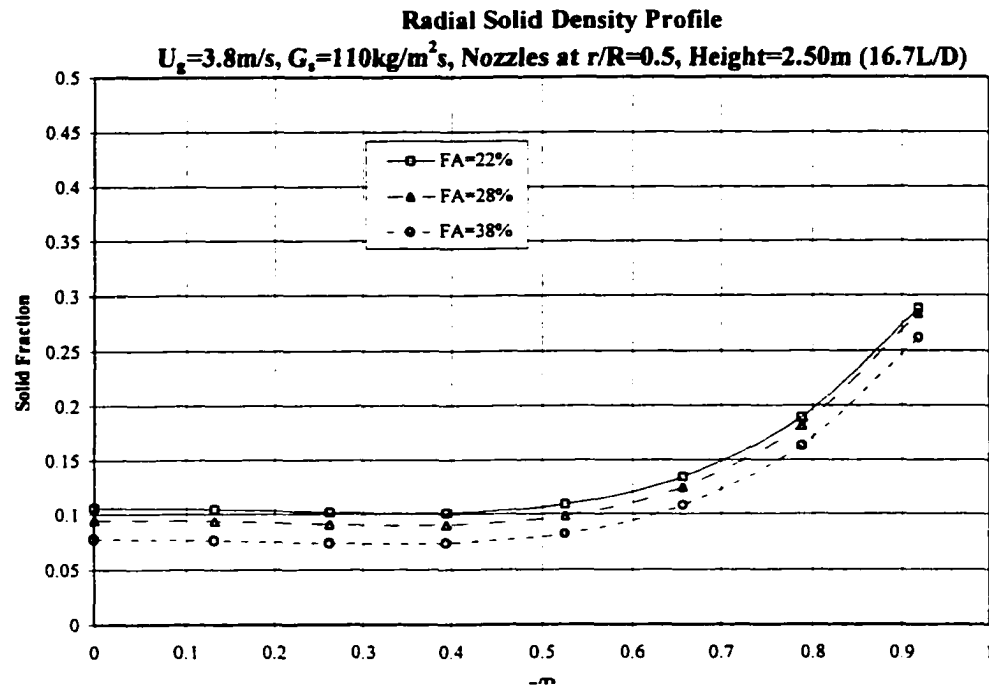


Figure 5.45d: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Middle, Height=2.50m.

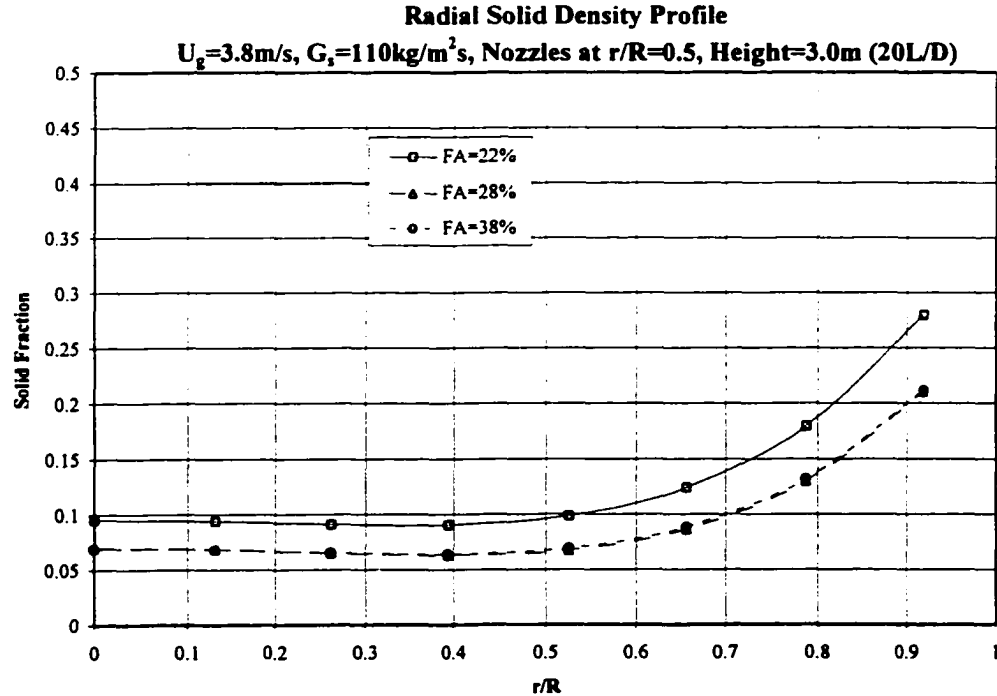


Figure 5.45e: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Middle, Height=3.0m.

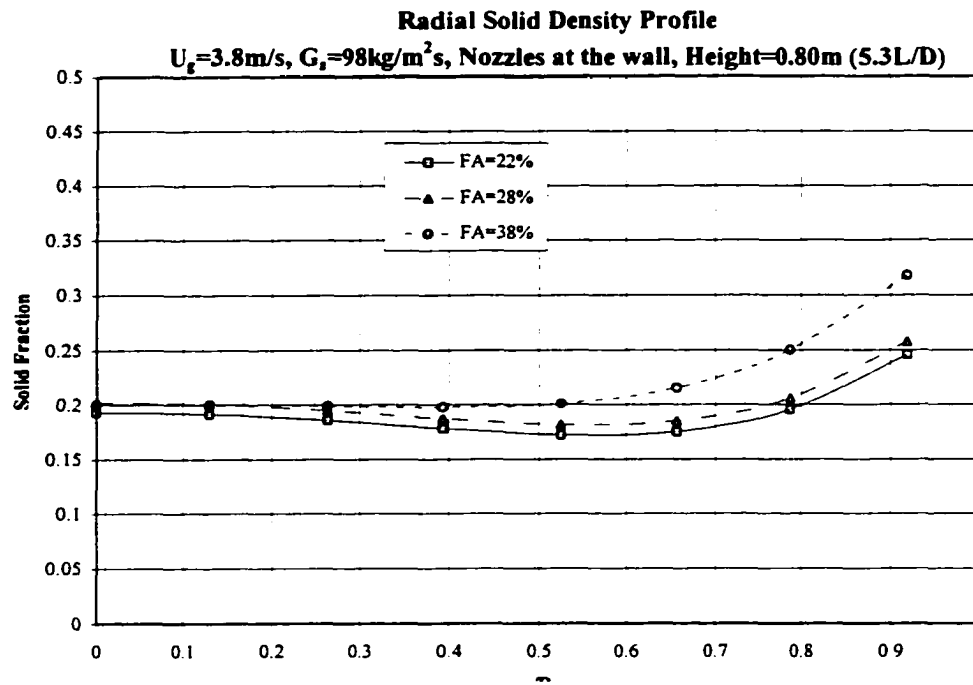


Figure 5.46a: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Wall, Height=0.80m.

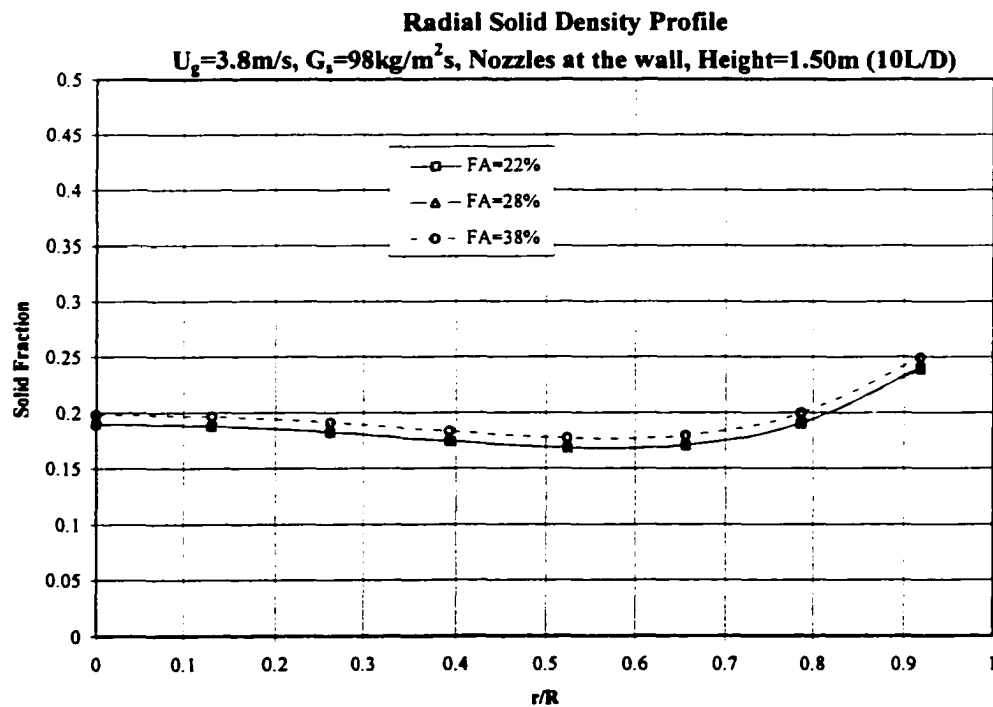


Figure 5.46b: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Wall, Height=1.50m.

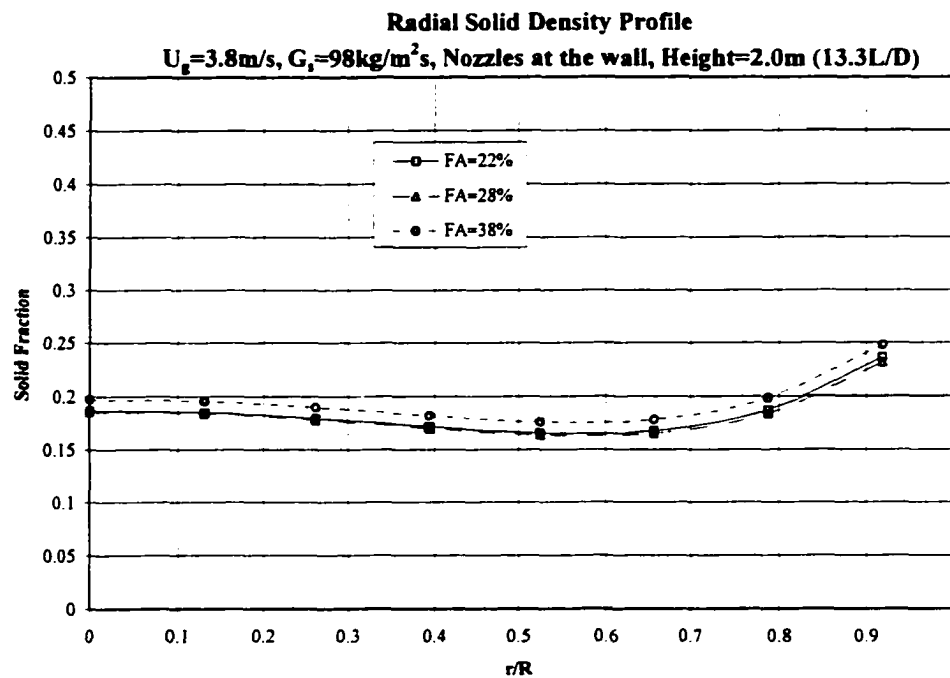


Figure 5.46c: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Wall, Height=2.0m.

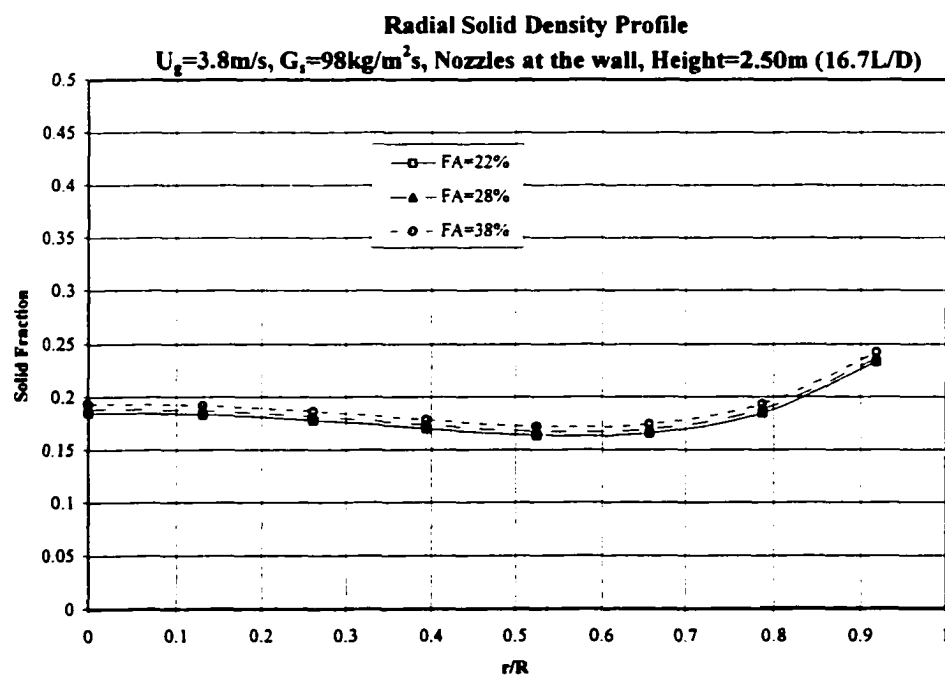


Figure 5.46d: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Wall, Height=2.50m.

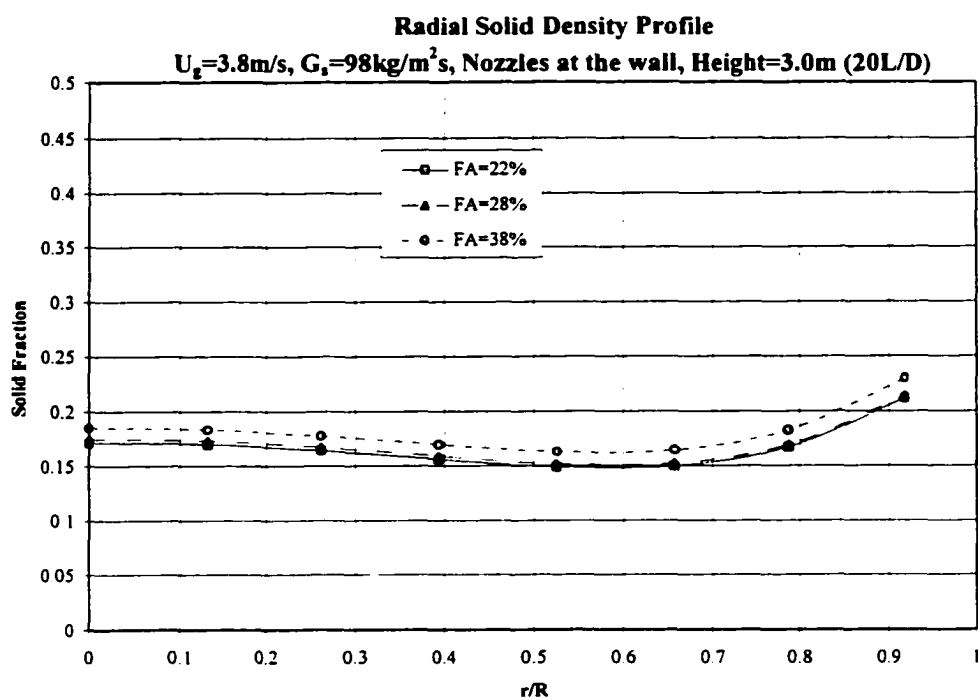


Figure 5.46e: Radial Solid Density Profile as a Function of the "Fluffing" Air for the Nozzles Located at the Wall, Height=3.0m.

5.3 Radial Pressure Profile

The radial pressure difference measurements can give important information about the physics of the gas-solid flow in a fluidized riser. In this chapter the effect of the inlet configuration on the radial pressure difference profile will be examined.

Figures 5.47 and 5.48 show the radial pressure difference 0.75m (5 L/D) and 1.70m (11.3 L/D), respectively above the nozzle inlet, obtained with the nozzles located at the center of the riser, for three different "fluffing" air flowrates. The superficial gas velocity is $U_g = 3.8 \text{ m/s}$ and the solid circulation rate is $G_s = 142 \text{ kg/m}^2 \text{ s}$. It may be observed that there is a pressure difference between the wall and the center with pressure lower in

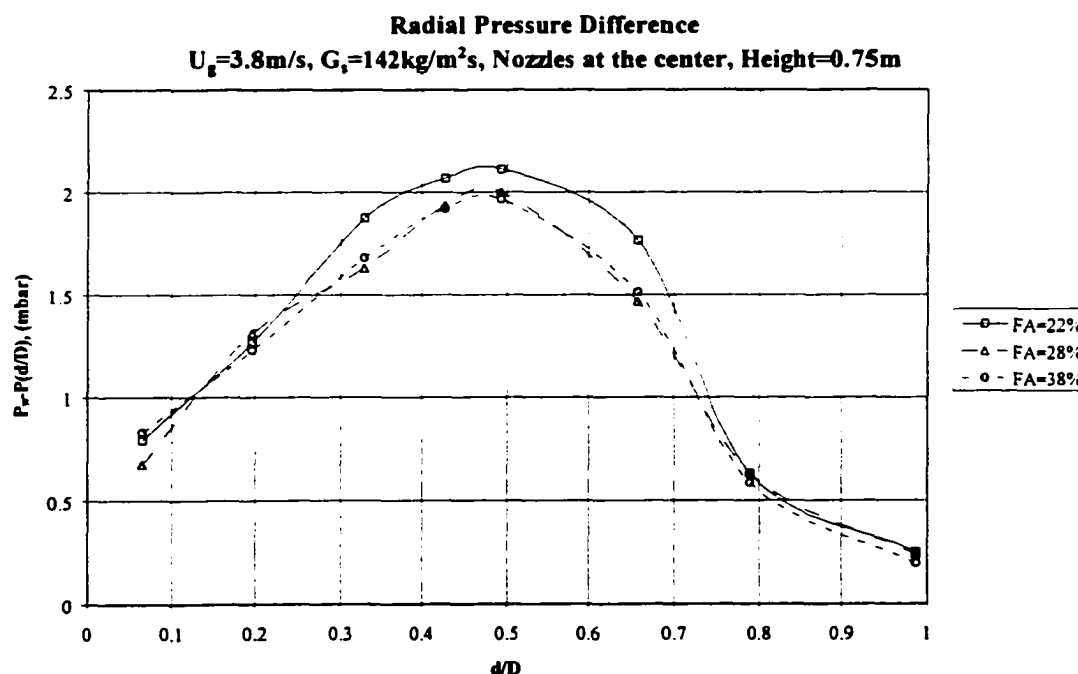


Figure 5.47: Radial Pressure Gradient as a Function of the "Fluffing" Air for the Nozzle Located at the Center, Height=0.75m.

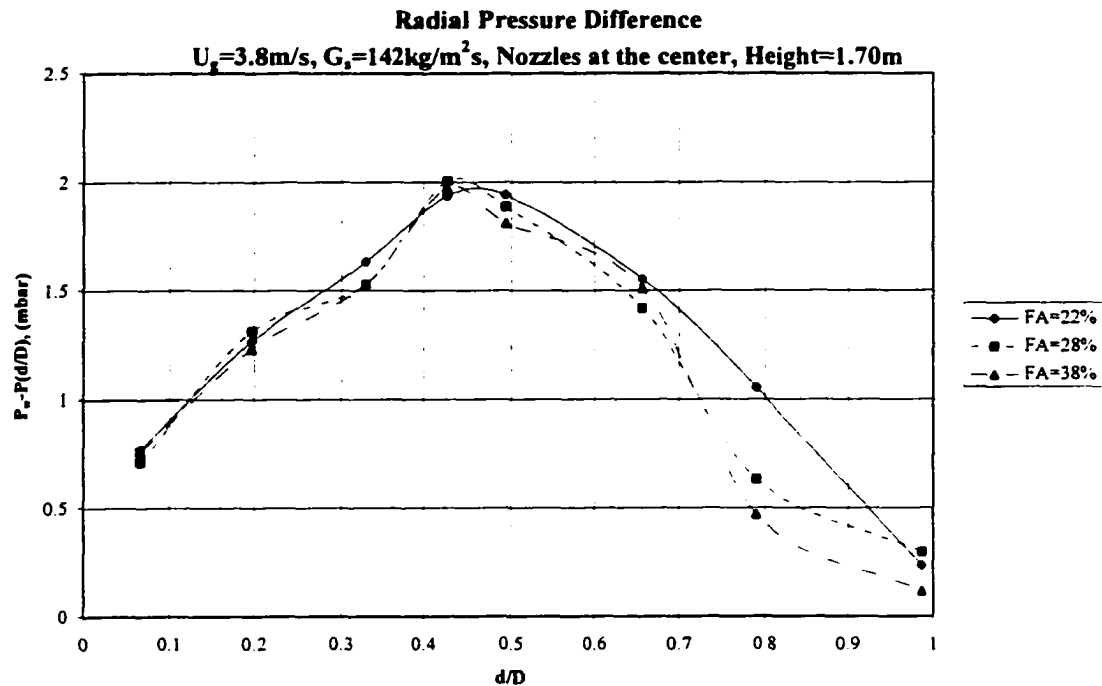


Figure 5.48: Radial Pressure Gradient as a Function of the "Fluffing" Air for the Nozzle Located at the Center, Height=1.70m.

the center for all the cases. The order of magnitude of the radial pressure difference is about the same as the axial pressure difference for change in elevation of one radius. The pressure difference profile has a generally parabolic shape as does the radial solid fraction profile. Changing the "fluffing" air flowrate does not significantly affect the radial pressure gradient profile. Radial pressure profiles are almost identical for "fluffing" air rates varying from 22% to 38% of the total flowrate.

Figure 5.49 shows the pressure difference profile as a function of the "fluffing" air percentage when the four nozzles are located at the riser wall. The operating conditions are the same as in the previous case. The radial pressure gradient profile is flatter since the radial solid concentration for this nozzle position is flatter, but still there is a

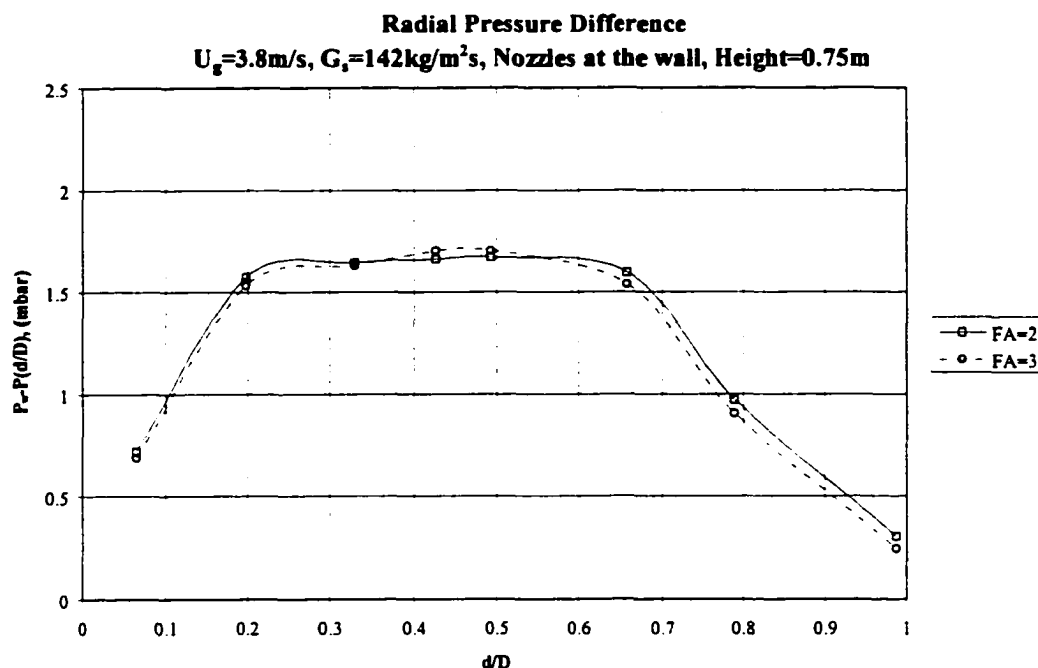


Figure 5.49: Radial Pressure Gradient as a Function of the "Fluffing" Air for the Nozzle Located at the Wall, Height=0.75m.

significant pressure difference between the wall and the centerline of the riser with pressure lower in the center. Changing the "fluffing" air percentage does not have any influence in these profiles.

Figures 5.50 and 5.51 show the effect of the nozzle configuration on the radial pressure difference profile at heights 0.75m (5 L/D) and 1.50m (11.3 L/D), above the nozzles inlet respectively. The superficial gas velocity is $U_g=3.8\text{m/s}$, the solid circulation rate $G_s=142\text{kg/m}^2\text{s}$ and the "fluffing" air flowrate 22% of the total flowrate. It may be observed that the parabolic pressure gradient that appears when the nozzles are in the center of the riser becomes flatter when the nozzles are moved to the riser wall, and this is

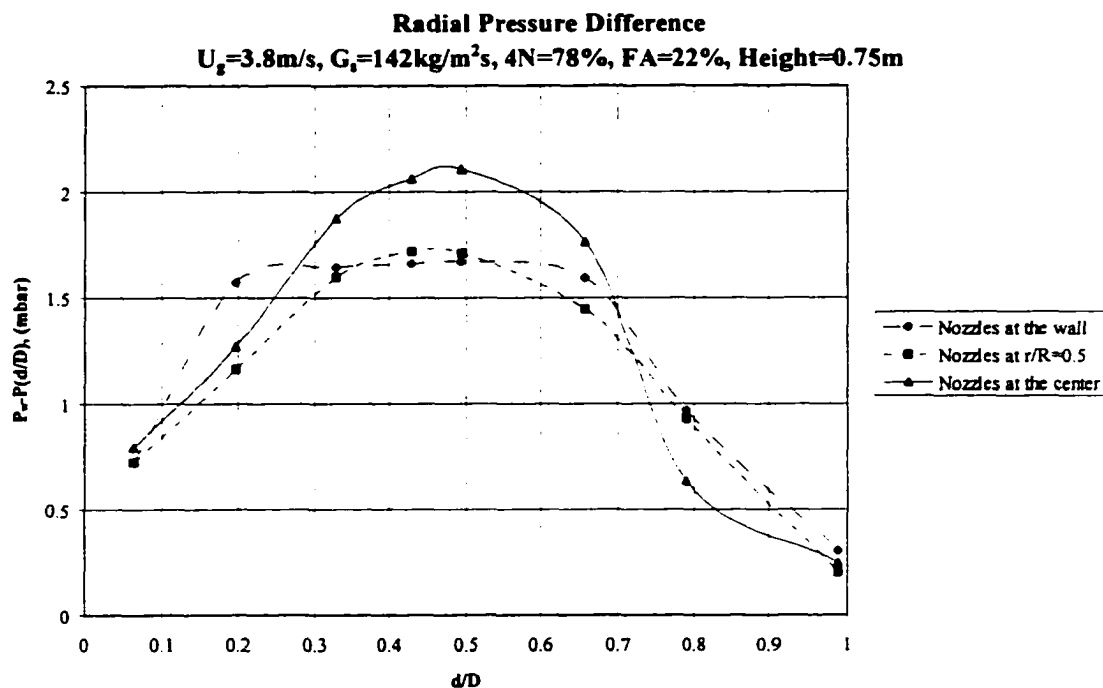


Figure 5.50: Radial Pressure Gradient as a Function of the Nozzle Configuration, Height=0.75m.

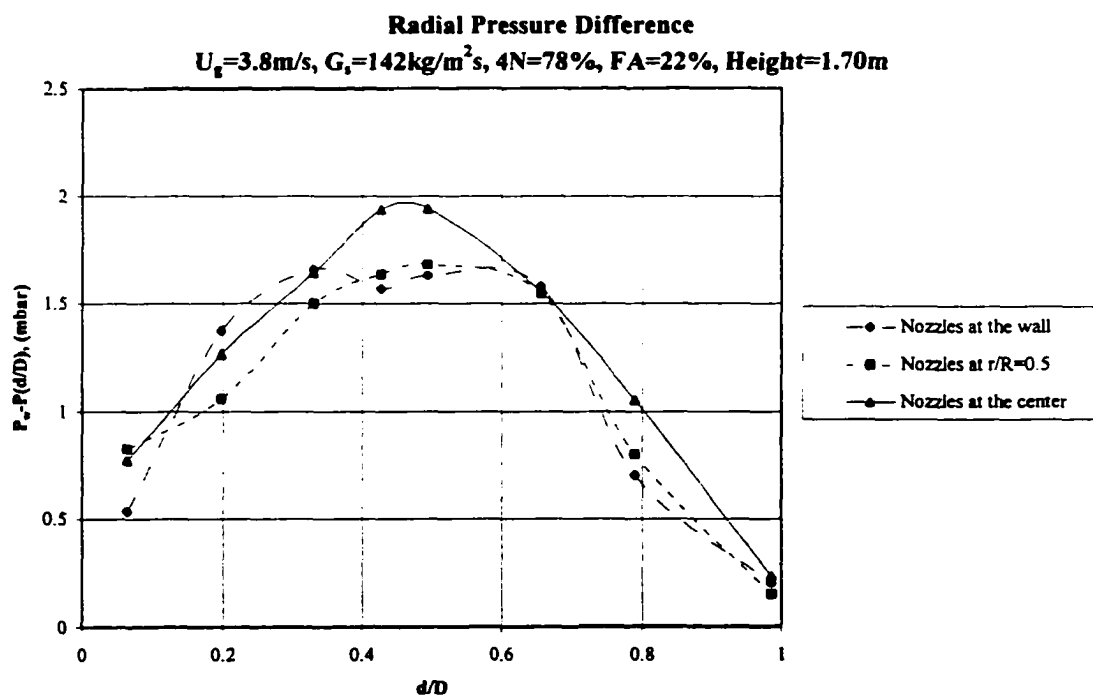


Figure 5.51: Radial Pressure Gradient as a Function of the Nozzle Configuration, Height=1.70m.

in agreement with the behavior of the radial solid density profile with the nozzle position change. The pressure difference between the wall and the centerline is higher when the nozzles are in the center because in this case the centerline flow is the most dilute.

Figures 5.52 and 5.53 show the radial pressure gradient for a lower solid circulation rate of $G_s=75\text{kg/m}^2\text{s}$, and all the other operating conditions constant. It is seen that at 0.75m (Figure 5.52), there is not a significant difference of the radial pressure difference profile compared with that of the high solid circulation rate case, for the three nozzle positions. It is seen again that the pressure difference profile is flatter when the nozzles are located close to the riser wall. At 1.70m (Figure 5.53), the radial pressure

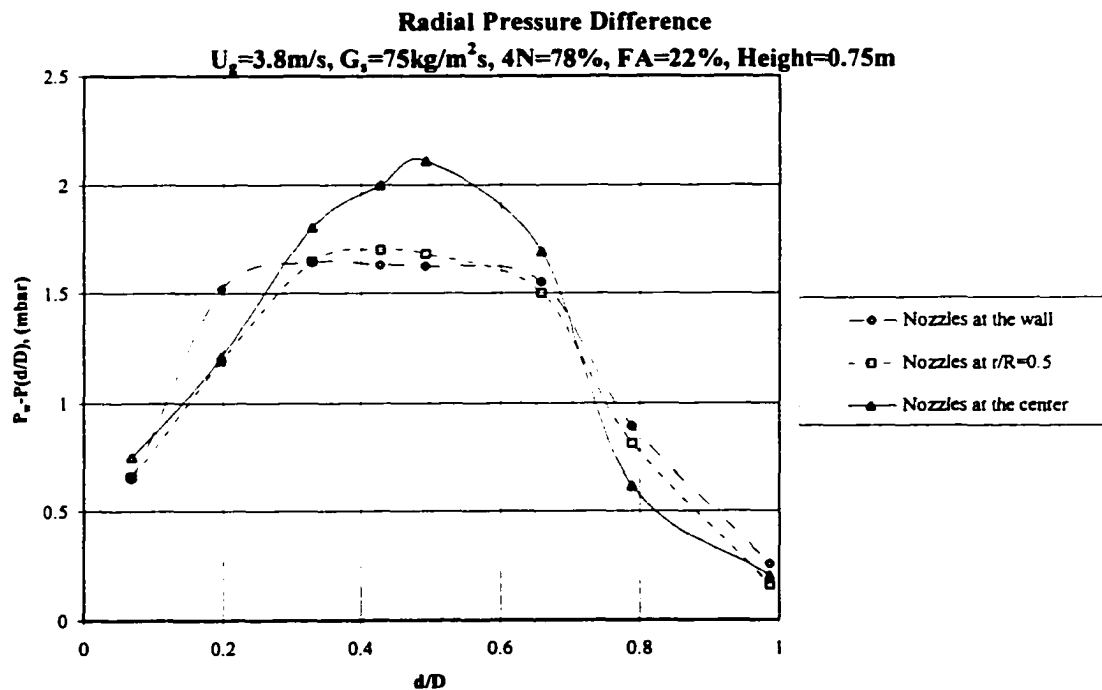


Figure 5.52: Radial Pressure Gradient as a Function of the Nozzle Configuration, Height=0.75m.

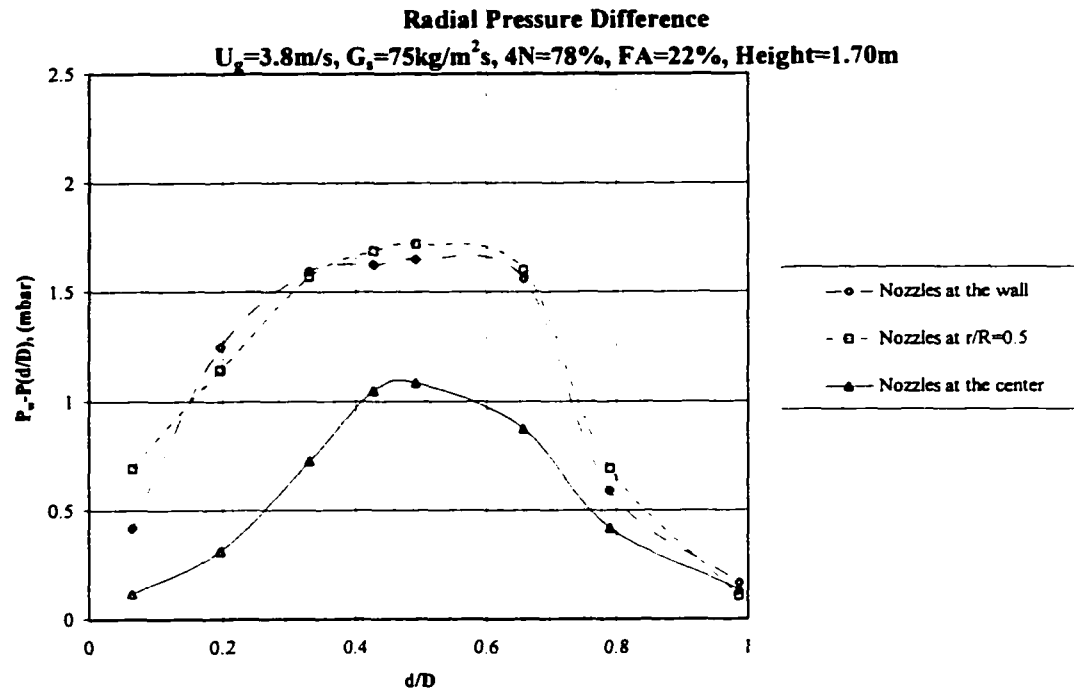


Figure 5.53: Radial Pressure Gradient as a Function of the Nozzle Configuration, Height=1.70m.

difference is much smaller when the nozzles are located at the center. This happens this elevation of 1.70m is the elevation of the inflection point as Figure 5.8 shows.

5.4 Radial Profiles of the Axial Solid Flux and the Axial Solid and Gas Velocity

The measurements of the radial distribution of the net axial solid flux also provide important information about the gas-solid flow in the riser. These experimental data combined with the radial concentration data and a correlation for the slip velocity can give the radial profiles of the axial solid velocity and of the axial gas velocity respectively. The effect of the nozzle configuration and the "fluffing" air flowrate on these profiles are examined.

Figure 5.54 shows the radial distribution of the axial net solid flux for a superficial gas velocity, $U_g=3.8\text{m/s}$ and a solid circulation rate, $G_s=148\text{kg/m}^2\text{s}$. The "fluffing" air flowrate is 22% of the total flowrate and the three different nozzle positions and the curves are for an elevation 0.75m (5 L/D) above the nozzles inlet. It is seen that

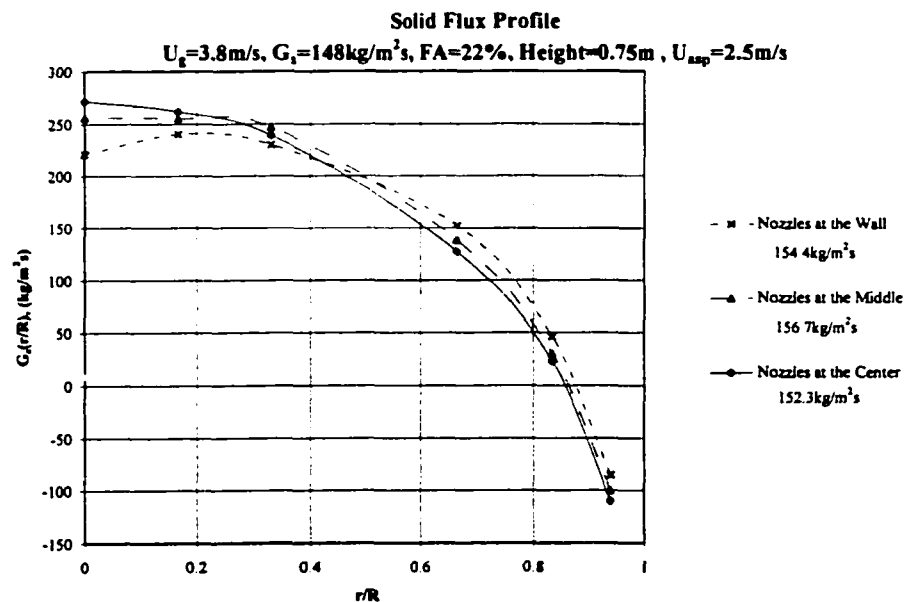


Figure 5.54: Radial Profile of the Net Axial Solid Flux as a Function of the Nozzle Configuration, Height=0.75m.

the effect of the nozzle position on the net solid flux profile is not very significant. When the nozzles are located close to the riser axis, the flux profile has an almost parabolic shape. The net solid flux is negative close to the riser wall and appears a maximum value at the riser axis. This profile remains almost the same when the nozzles are moved to the middle. The only significant difference is that the profile has a fairly constant section from $r/R=0$ to $r/R=0.3$. When the nozzles are located at the riser wall the net flux profile does not change significantly at the section from $r/R=0.35$ to $r/R=1$ but now a maximum appears at $r/R=0.18$ and the flux at the riser axis is slightly smaller. The same observations can be made from Figure 5.55 which shows the solid flux profile for the same operating conditions but at the elevation of 1.70m (11.3 L/D) above the nozzles

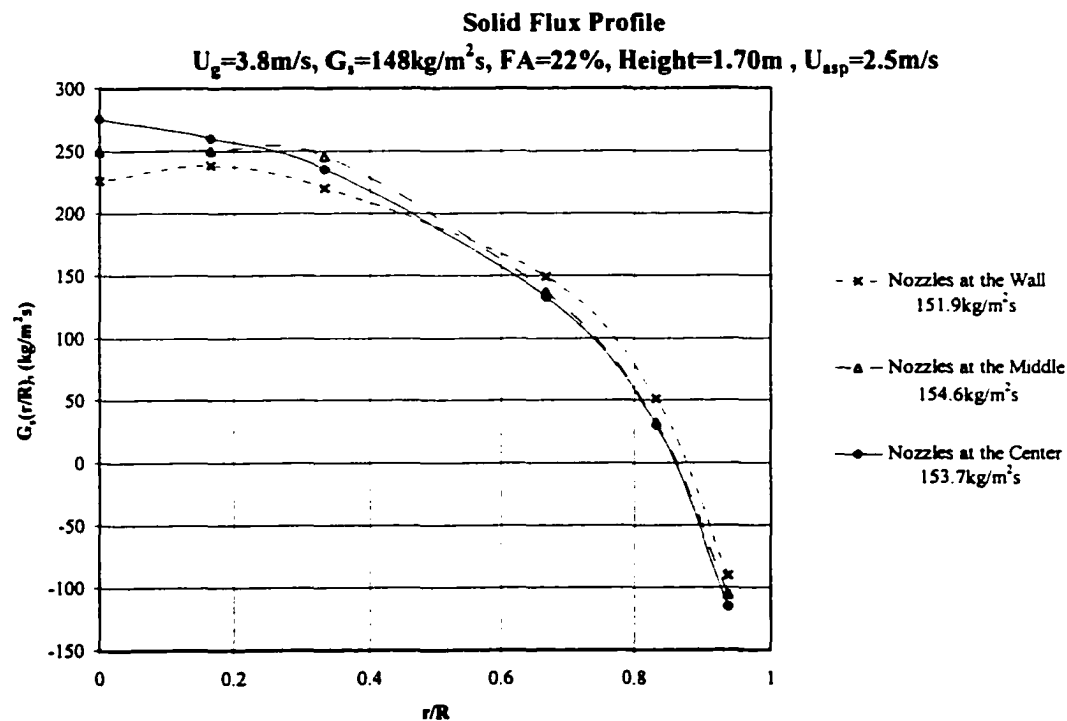


Figure 5.55: Radial Profile of the Net Axial Solid Flux as a Function of the Nozzle Configuration, Height=1.70m.

inlet. The three profiles, one for each nozzles position are very similar with the profiles at height 0.75m.

Figures 5.56 and 5.57 present the axial net solid flux radial distribution for a smaller solid circulation rate ($G_s=72\text{kg/m}^2\text{s}$), at 0.75m and 1.70m (5 and 11.3 L/D), above the nozzles inlet, respectively. At the height of 0.75m, (Figure 5.56), it can be seen that there are not significant differences of the profiles for the three different nozzle positions. At the elevation of 1.70m (Figure 5.57) the flux profile, with the nozzles located close to the riser wall and at the middle are not different from the corresponding profiles at 0.75m. The net solid flux profile when the nozzles are located at the riser center is almost flat along the 70% of the riser radius. This happens because for these operating conditions the inflection point is at the elevation of 1.70m.

Figures 5.58 and 5.59 show the net solid flux radial distribution for a higher superficial gas velocity, $U_g=5.2\text{m/s}$, again at two elevations and for the three nozzle configurations. In this case also, the effect of the nozzle position on the radial profile of the axial net solid flux is the same as with the lower superficial gas velocity case.

For all the cases the net solid circulation rate can be calculated by integrating the experimental data of the local net solid flux over the riser cross section. These results are shown in every figure from Figure 5.54 to 5.59 for each case separately. It is seen that the results of the integration are generally very close to the measured total solid circulation rate given in the Figure title. It can be seen that the difference between the measured and the intergrated solid circulation rate is $0\pm 5\%$ for the high circulation rate. For the lower circulation rate this difference reaches up to $\pm 10\%$. This increase of the difference for the

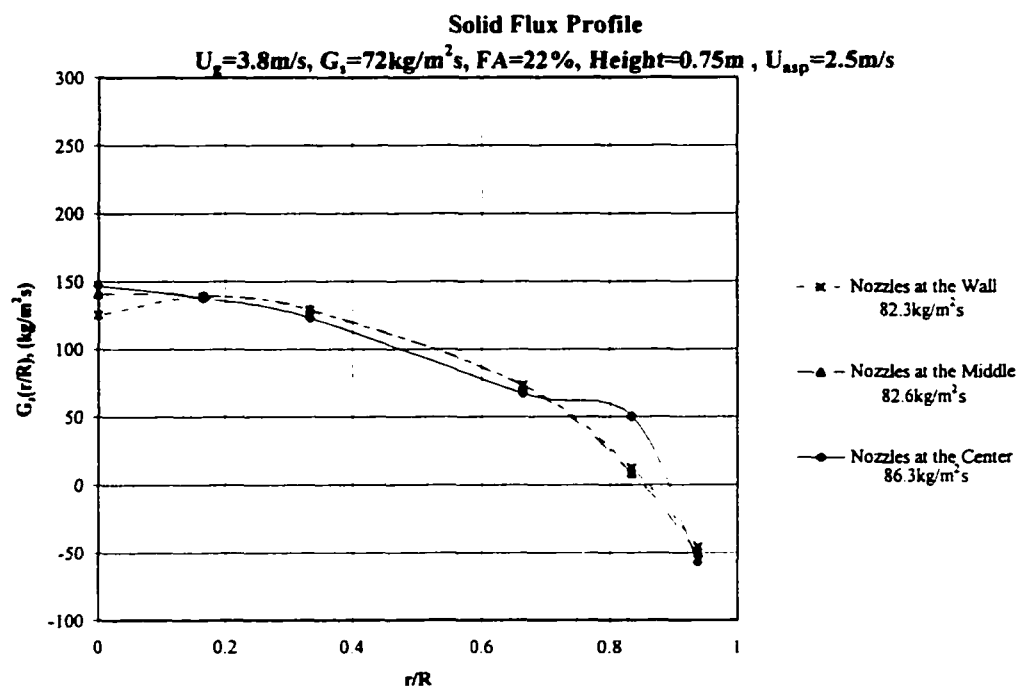


Figure 5.56: Radial Profile of the Net Axial Solid Flux as a Function of the Nozzle Configuration, Height=0.75m.

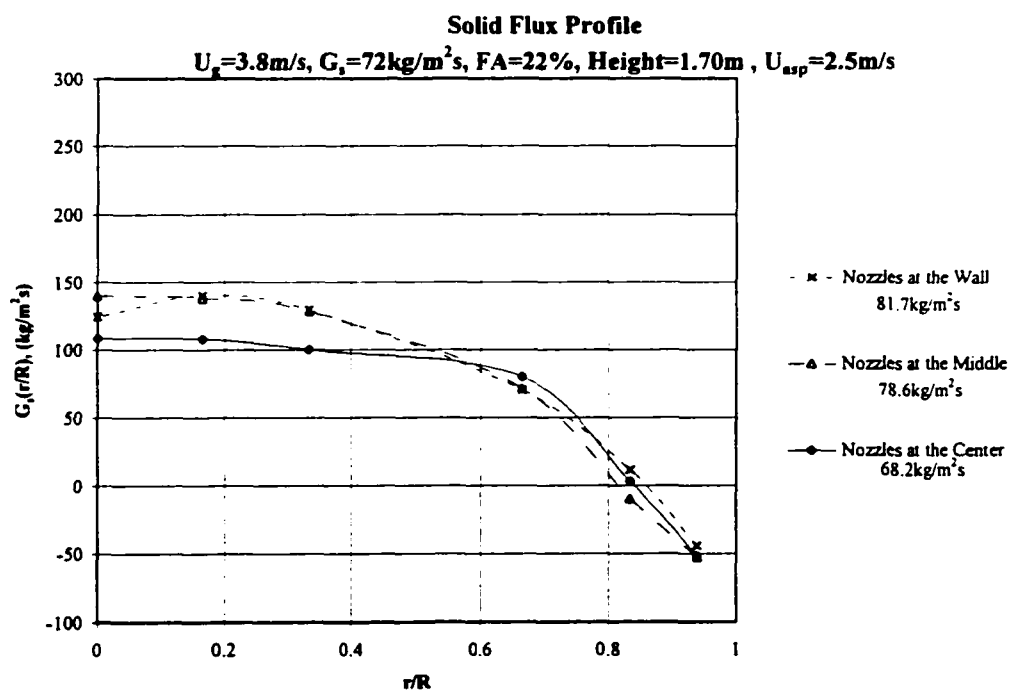


Figure 5.57: Radial Profile of the Net Axial Solid Flux as a Function of the Nozzle Configuration, Height=1.70m.

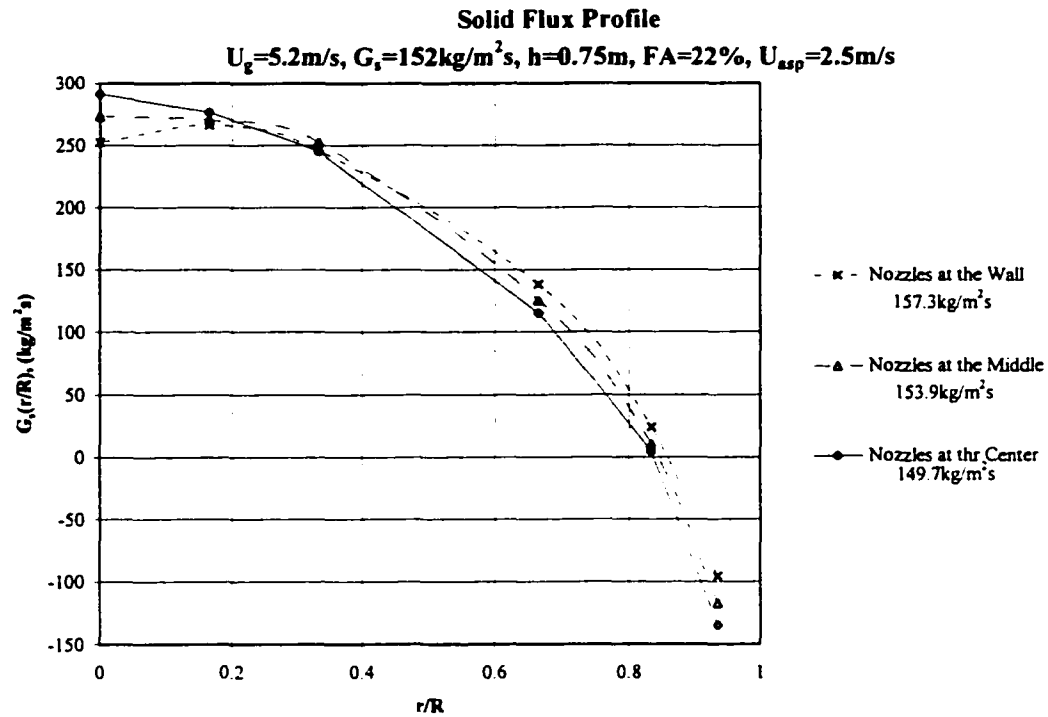


Figure 5.58: Radial Profile of the Net Axial Solid Flux as a Function of the Nozzle Configuration, Height=0.75m.

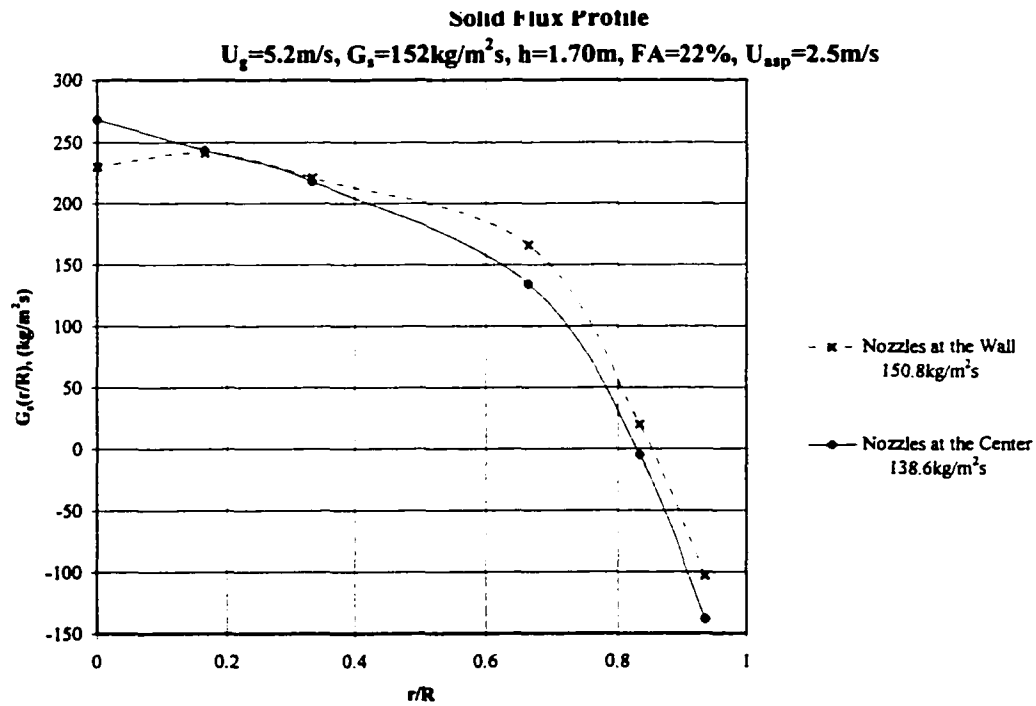


Figure 5.59: Radial Profile of the Net Axial Solid Flux as a Function of the Nozzle Configuration, Height=1.70m.

lower solid circulation rate is expected since in these measurements the experimental error increases.

The effect of the "fluffing" air flowrate on the radial profile of the axial net solid flux is presenting at the following set of data. Figures 5.60 and 5.61 show the solid flux profile for different "fluffing" air flowrates and at different elevations on the riser, while the nozzles are located at the center. Figures 5.62 and 5.63 show the same effect for the nozzles located at the wall and Figure 5.64 for the nozzles located at the middle. It can be observed from all these results that the "fluffing" air flowrate does not affect the solid flux distribution. The slight differences that appear are not important since they are of the

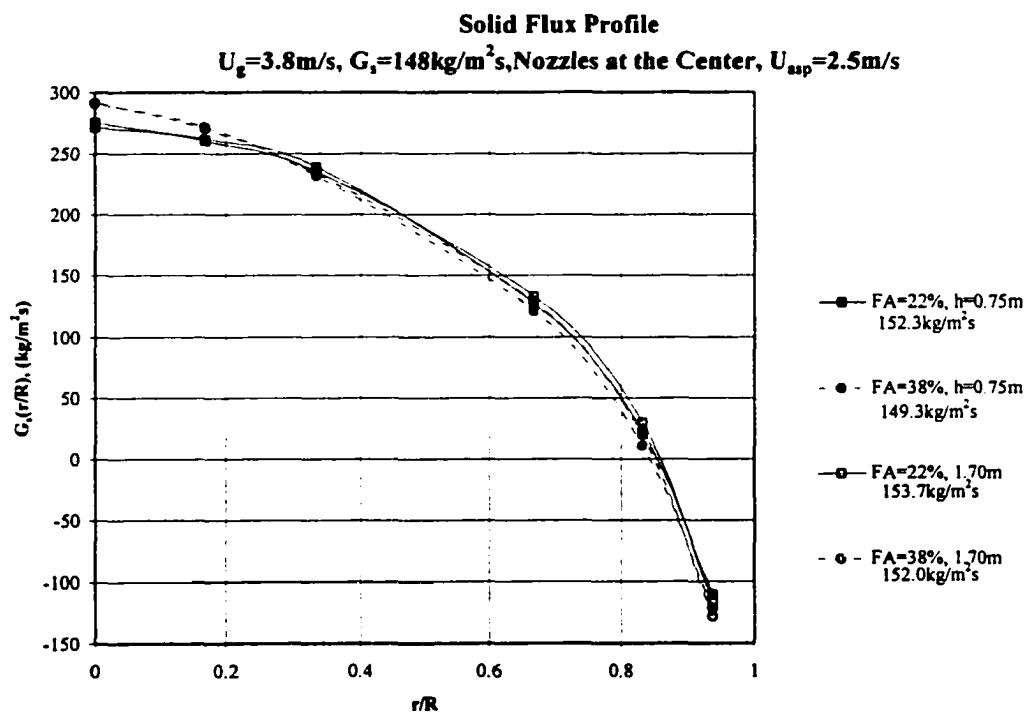


Figure 5.60: Net Solid Flux Radial Profile as a Function of the "Fluffing" Air Flowrate, High Solid Circulation Rate.

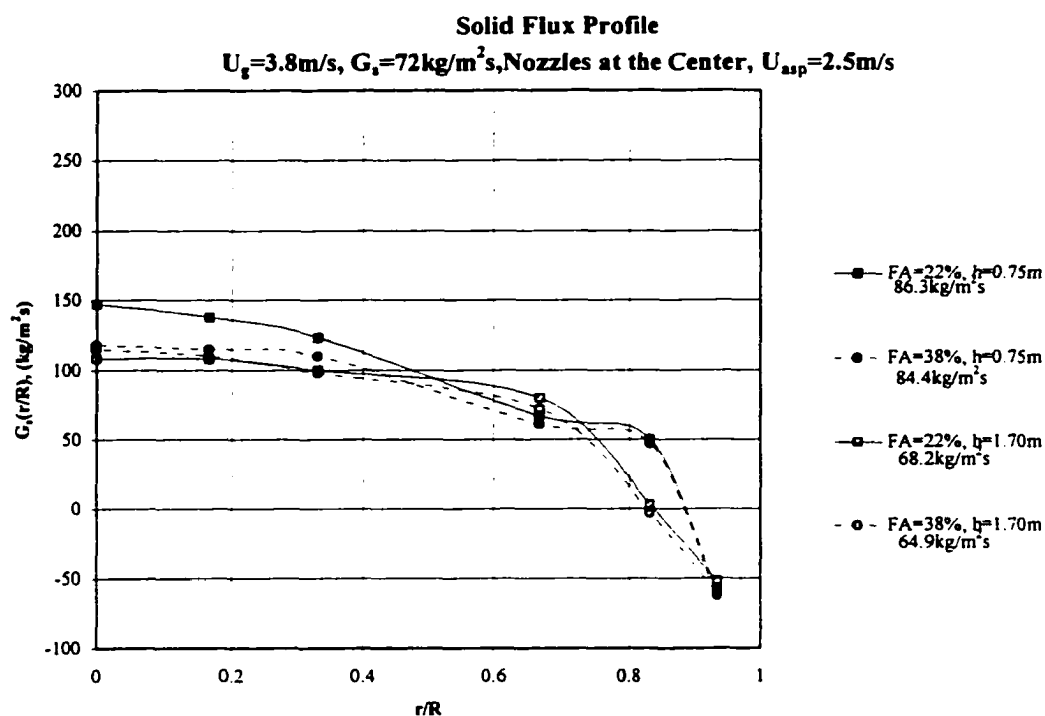


Figure 5.61: Net Solid Flux Radial Profile as a Function of the "Fluffing" Air Flowrate, Low Solid Circulation Rate.

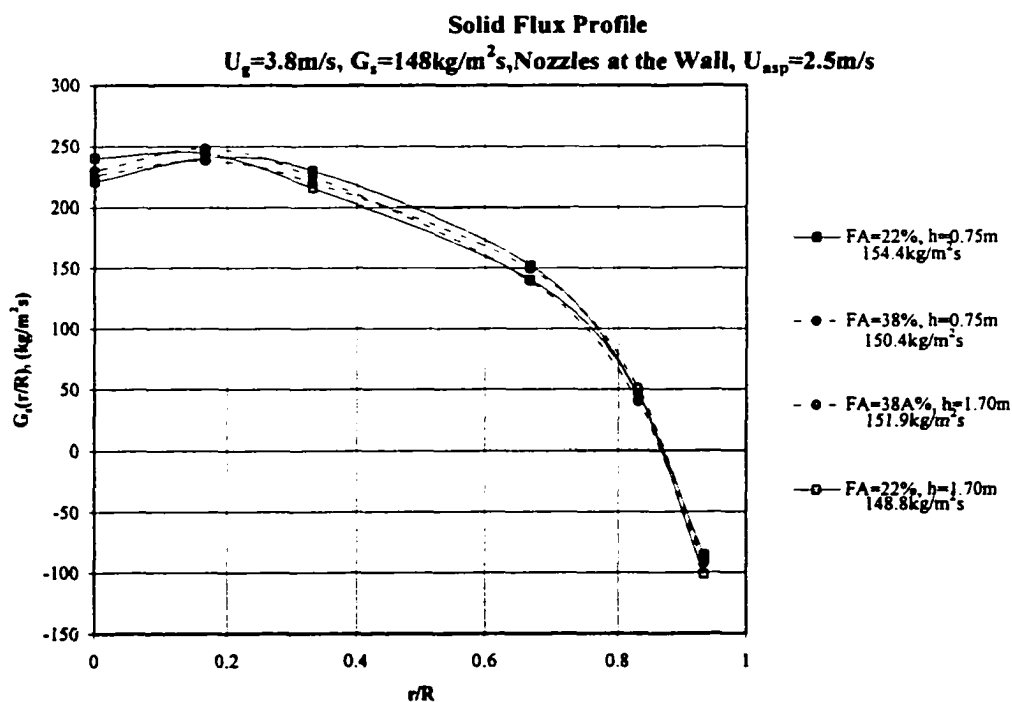


Figure 5.62: Net Solid Flux Radial Profile as a Function of the "Fluffing" Air Flowrate, High Solid Circulation Rate.

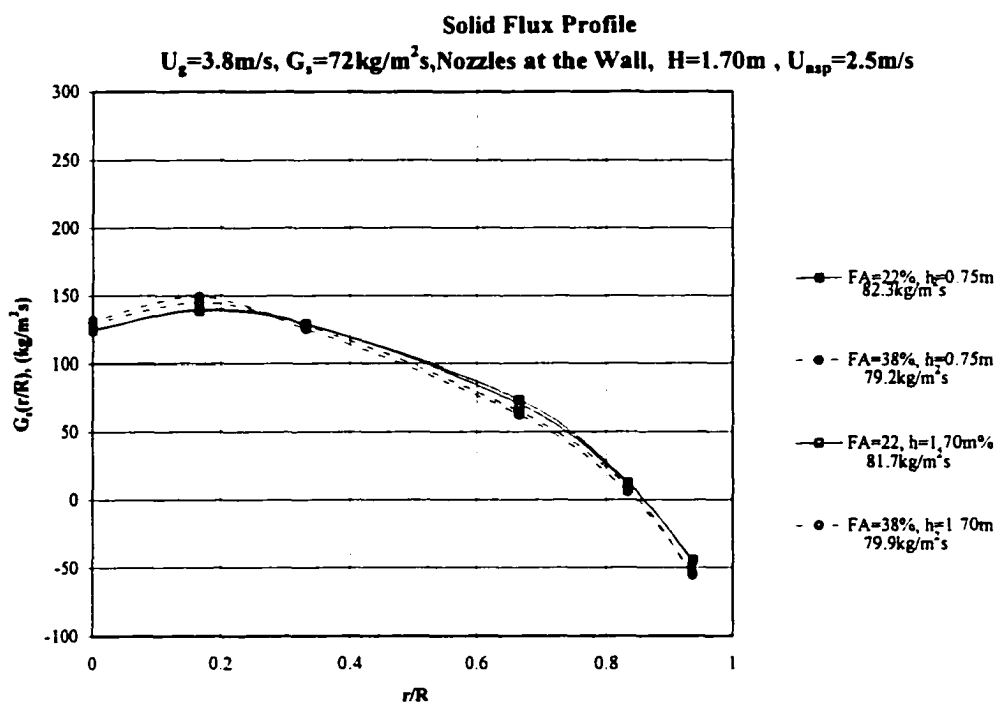


Figure 5.63: Net Solid Flux Radial Profile as a Function of the "Fluffing" Air Flowrate, Low Solid Circulation Rate.

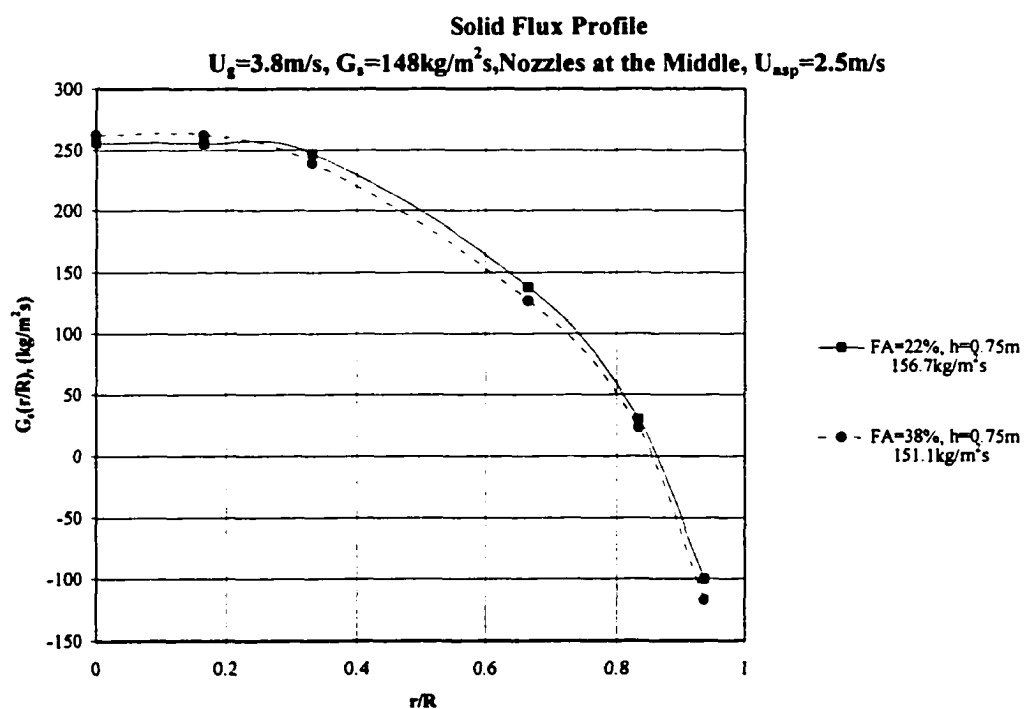


Figure 5.64: Radial Profile of the Net Axial Solid Flux as a Function of the "Fluffing" Air Flowrate, High Solid Circulation Rate.

same order of magnitude as the experimental error.

The radial distribution of the axial solid velocity can be calculated from the radial solid density and radial net flux distribution data, as follows:

$$G_s(r/R) = \rho_s[1 - \varepsilon(r/R)]U_s(r/R) \quad (5.6)$$

Figures 5.65 and 5.66 show the radial profile of the axial solid velocity obtained from the solid flux data of Figures 5.54 and 5.55 and from the corresponding radial concentration data. It is seen (Figure 5.65), that the effect of the nozzle configuration on the solid velocity is significant. When the nozzles are located at the riser center the velocity profile is very high at the center. The velocity profile close the riser wall (from $r/R=0.82$ to

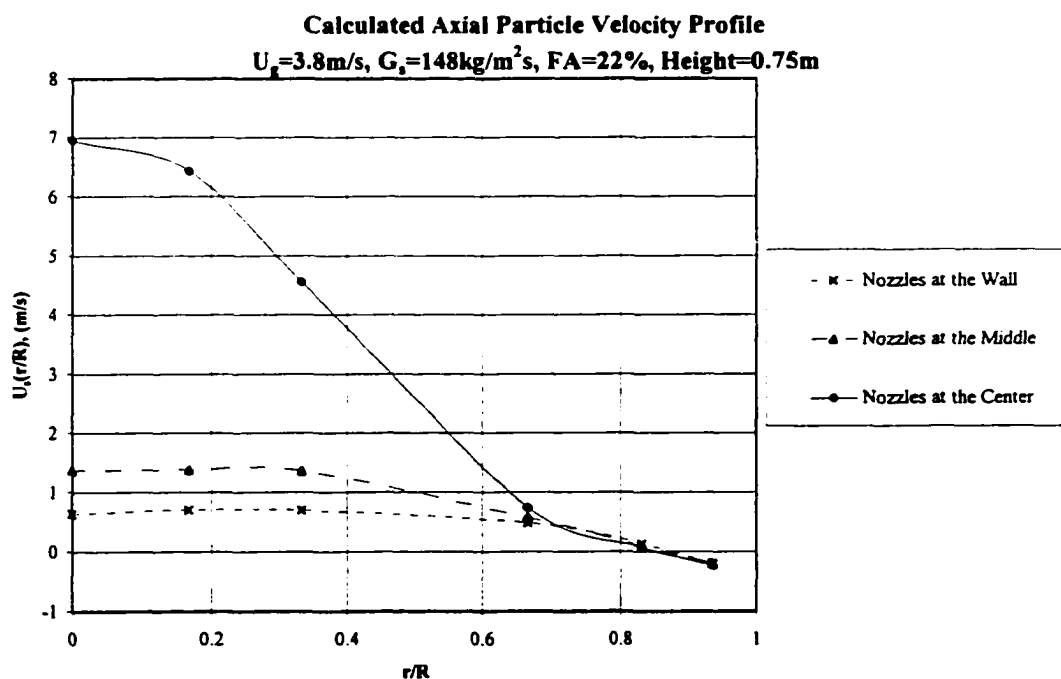


Figure 5.65: Calculated Radial Profile of the Axial Solid Velocity as a Function of the Nozzle Configuration, Height=0.75m.

$r/R=1$) is negative since in this region the solid is moving downward (Figure 5.54 and 5.55). The radial position where the velocity becomes zero differs slightly from the position where the solid flux becomes zero. This difference is due to the error that inserts after the smoothing of the curve. The velocity distribution has a maximum of about 7m/s at the riser axis where the solid fraction is very low (Figure 5.39a). It can be seen that the solid velocity at the axis is much higher than the superficial gas velocity ($U_g=3.8\text{m/s}$). This shows that when the nozzles are located at the center of the riser the gas-solid flow in the core is a very dilute flow in solid but with high solid velocity. It is also shown that for this nozzle configuration a strong radial gradient of the axial solid velocity exists as does solid concentration. When the nozzles are located at the middle the solid velocity profile changes in the core section from the riser axis up to $r/R=0.65$. In this section the solid velocity is much smaller and the distribution much flatter. The solid velocity has a value of 1.4m/s along almost the 40% of the riser radius. In annular section from $r/R=0.65$ to the riser wall ($r/R=1$) the two profiles are very similar without any important difference. The third velocity profile, when the nozzles are located as close as possible to the riser wall, is also similar in the annular section, ($r/R=0.70$ to $r/R=1$). For this nozzle configuration the velocity appears to be almost constant over the core 70% of the riser radius. The actual value of this velocity is about 0.8m/s, smaller than in the case where the nozzles are located at the middle. It is seen that the solid velocity at the riser axis decreases as the nozzles are moved to the riser wall since the solid fraction at the riser axis increases as the nozzles are moved to the riser wall.

Figure 5.66 shows the results for the same operating conditions but at height 1.70m (11.3 L/D) above the nozzle inlet. Comparing of these solid velocity profiles with those at 0.75m, it can be observed that deceleration of solid takes place from 0.75m to 1.70m when the nozzles are located at the center of the riser. This decrease in solid velocity is clearer at the riser axis where the velocity decreases from about 7m/s at 0.75m (5 L/D) to about 5.8m/s at 1.70m (11.3 L/D). This was also shown by the axial voidage profile (Figure 5.1). On the other hand, the velocity profiles for the nozzles at the middle or at the riser wall are the same for the height of 1.70m. This, in terms of performance of riser as chemical reactor, means that the nozzles located at the center configuration is a configuration with relatively high degree of gas backmixing, reduced selectivity and

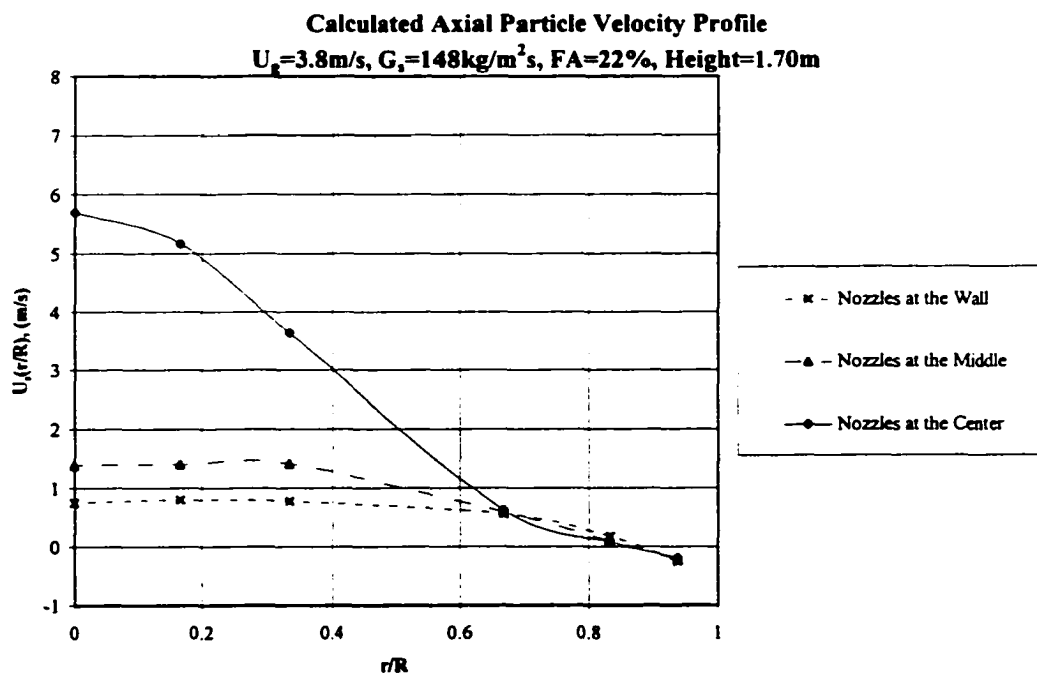


Figure 5.66: Calculated Radial Profile of the Axial Solid Velocity as a Function of the Nozzle Configuration, Height=1.70m.

generally low performance. Figures 5.67 and 5.68 show the axial solid velocity radial profile for the reduced solid circulation rate case, at 0.75m and 1.70m (5 and 11.3 L/D), respectively. It is seen again that when the nozzles are located at the riser center a strong radial gradient of the solid velocity exists. When the nozzles are located at the middle or as close as possible to the riser wall the axial solid velocity profile is much flatter. For all the cases, in the annular section from $r/R=0.82$ to $r/R=1$ (wall), the solid velocity is negative. Comparing Figure 5.67 and 5.68, it may be observed that the velocity profile does not change significantly between 0.75m and 1.70m when the nozzles are located at the middle or at the wall. When the nozzles are located at the center the solid velocity at 1.70m (11.3 L/D) is higher than at 0.75m (5 L/D) which means that an acceleration of solid takes place. This acceleration is in the inflection point region that exists at 1.70m for the certain operating conditions, (Figure 5.8).

Figures 5.69 and 5.70 show the results obtained with a higher superficial gas velocity ($U_g=5.2\text{m/s}$). In this case also the strong axial solid velocity gradient that exists when the nozzles are located at the center disappears when the nozzles are moved to the wall. A strong deceleration of solid takes place from 0.75m to 1.70m when the nozzles are located at the center. These axial solid velocity results show that the nozzles located at the center configuration gives strong radial gradients and strong decelerations for a variety of operating conditions.

Figure 5.71 show the effect of the "fluffing" air flowrate on the radial distribution of the axial solid velocity, when the nozzles are located at the center of the riser. The superficial gas velocity is $U_g=3.8\text{m/s}$, and the solid circulation rate, $G_s=148\text{kg/m}^2\text{s}$. At the

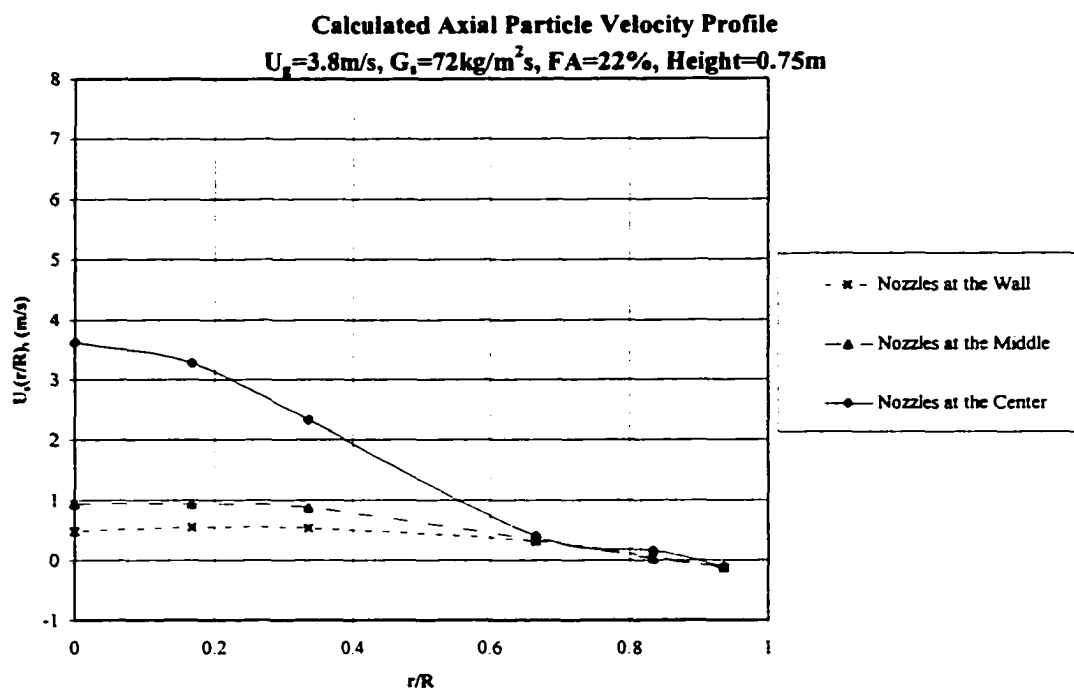


Figure 5.67: Calculated Radial Profile of the Axial Solid Velocity as a Function of the Nozzle Configuration, Height=0.75m.

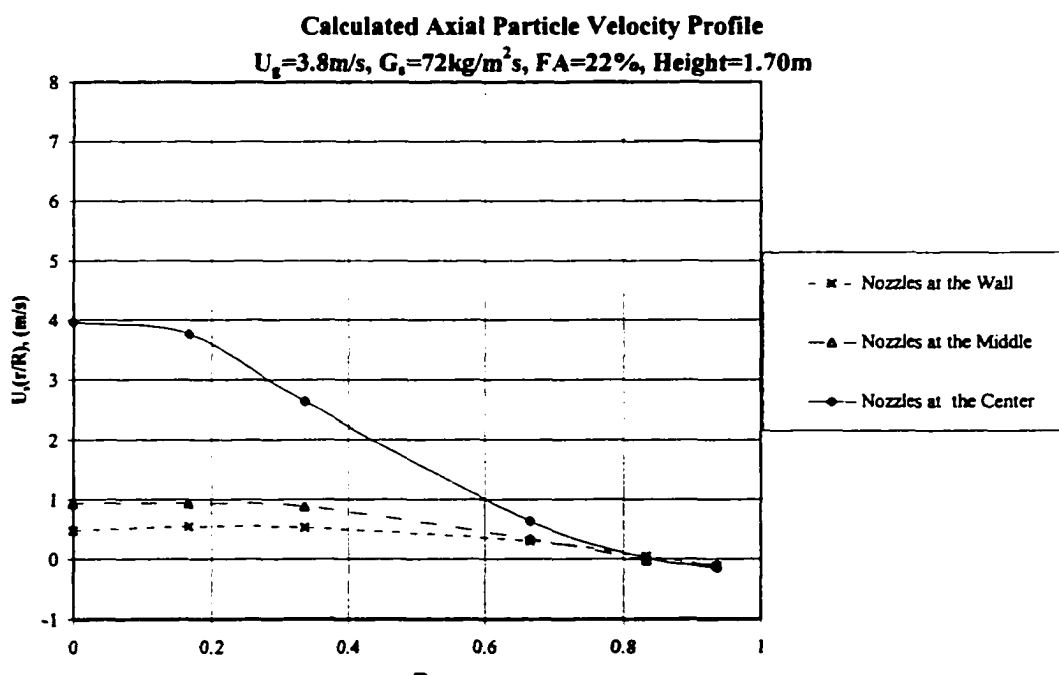


Figure 5.68: Calculated Radial Profile of the Axial Solid Velocity as a Function of the Nozzle Configuration, Height=1.70m.

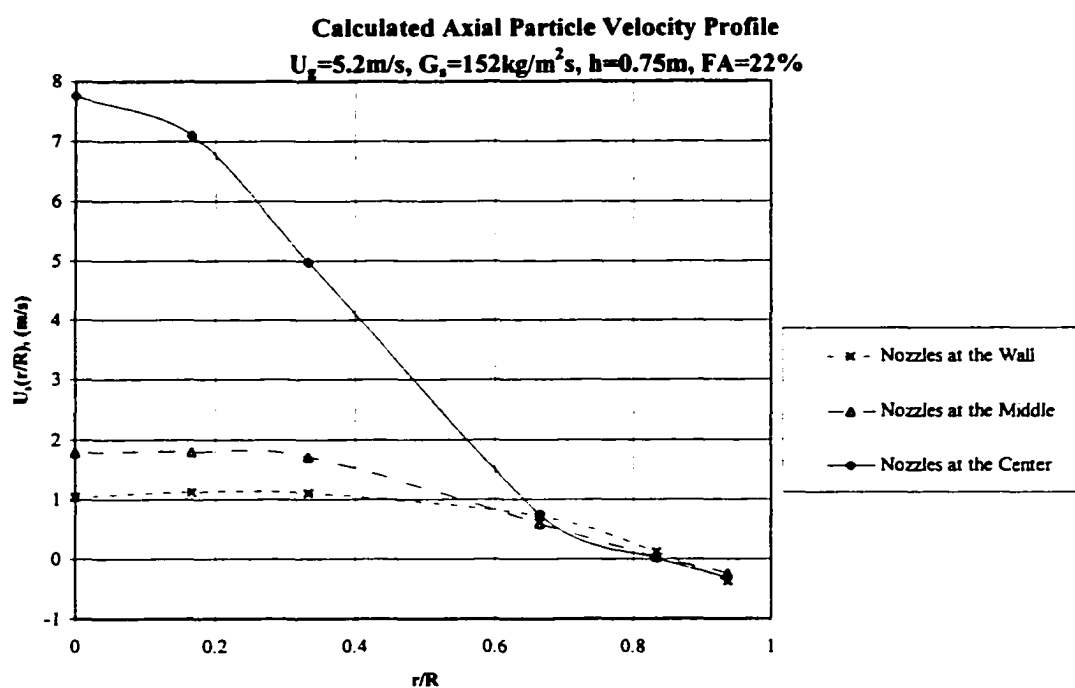


Figure 5.69: Calculated Radial Profile of the Axial Solid Velocity as a Function of the Nozzle Configuration, Height=0.75m.

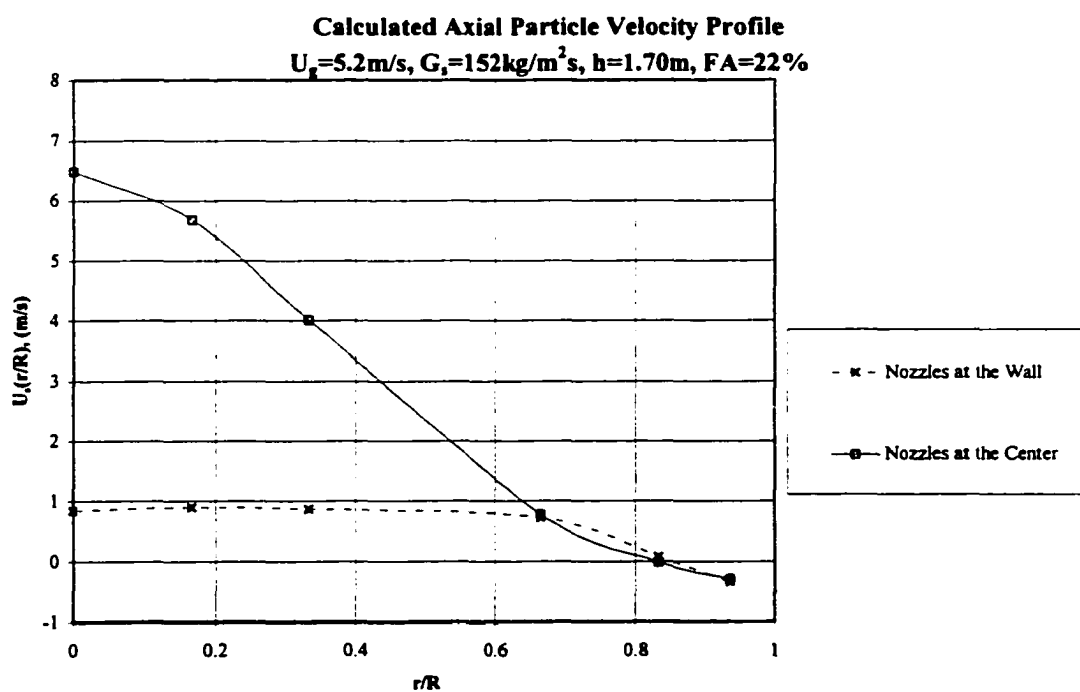


Figure 5.70: Calculated Radial Profile of the Axial Solid Velocity as a Function of the Nozzle Configuration, Height=1.70m.

elevation of 0.75m (5 L/D), it can be seen that for relative high "fluffing" air flowrate the solid velocity decreases. This appears to happen because with this inlet configuration an increase of the "fluffing" air flowrate causes a decrease of the flowrate of the four (located at the center) nozzles. Moving to higher elevations (1.70m, 11.3 L/D), it can be seen that the same behavior exists. The velocity profile for higher "fluffing" air flowrates is lower. It is shown that for any "fluffing" air flowrate there is a deceleration of solid moving from 0.75m to 1.70m (5 to 11.3 L/D). This deceleration is stronger for smaller "fluffing" air flowrates.

Figure 5.72 show the effect of the "fluffing" air flowrates for the reduced solid circulation rate case, $G_s=72\text{kg/m}^2\text{s}$. It is seen that for this case the effect of the "fluffing"

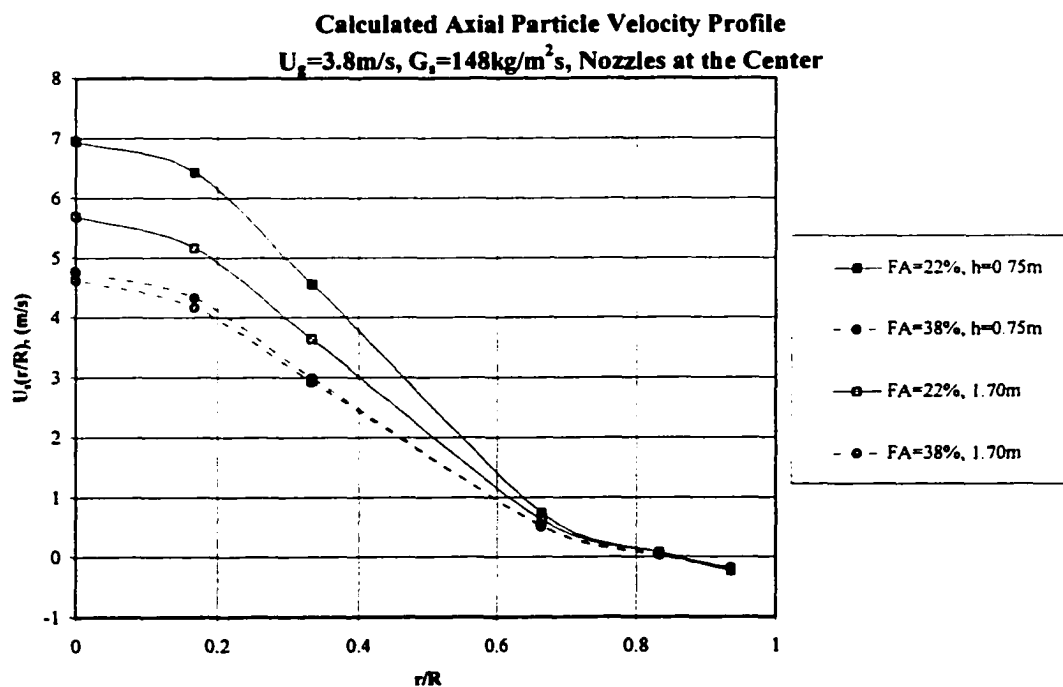


Figure 5.71: Calculated Radial Profile of the Axial Solid Velocity as a Function of the "Fluffing" Air Flowrate, High Solid Circulation Rate.

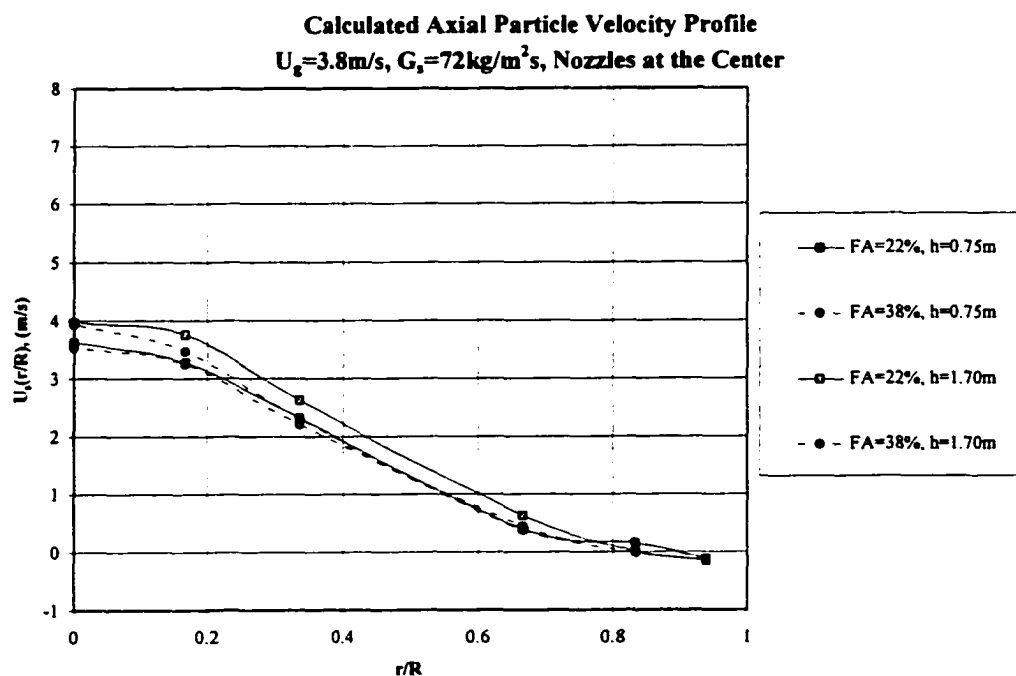


Figure 5.72: Calculated Radial Profile of the Axial Solid Velocity as a Function of the "Fluffing" Air Flowrate, Low Solid Circulation Rate.

air on the axial solid velocity profiles is not very significant. It may be noticed that the acceleration of solid that takes place from 0.75m to 1.70m (5 to 11.3 L/D) for the low "fluffing" air flowrate does not exist for the high "fluffing" air flowrate.

Figures 5.73 and 5.74 show the effect of the "fluffing" air flowrate on the solid velocity when the nozzles are located close to the riser wall. It is seen that "fluffing" air flowrate does not at all affect the solid velocity profile for all the operating conditions considered when the nozzles are located close to the riser wall. This is in agreement with the axial and radial density profiles which showed that "fluffing" air flowrate does not affect the gas-solid flow when the nozzles are as close as possible to the riser wall.

The experimental data already presented (the radial solid density profile, the radial profile of the axial net solid and the calculated radial profile of the axial solid velocity)

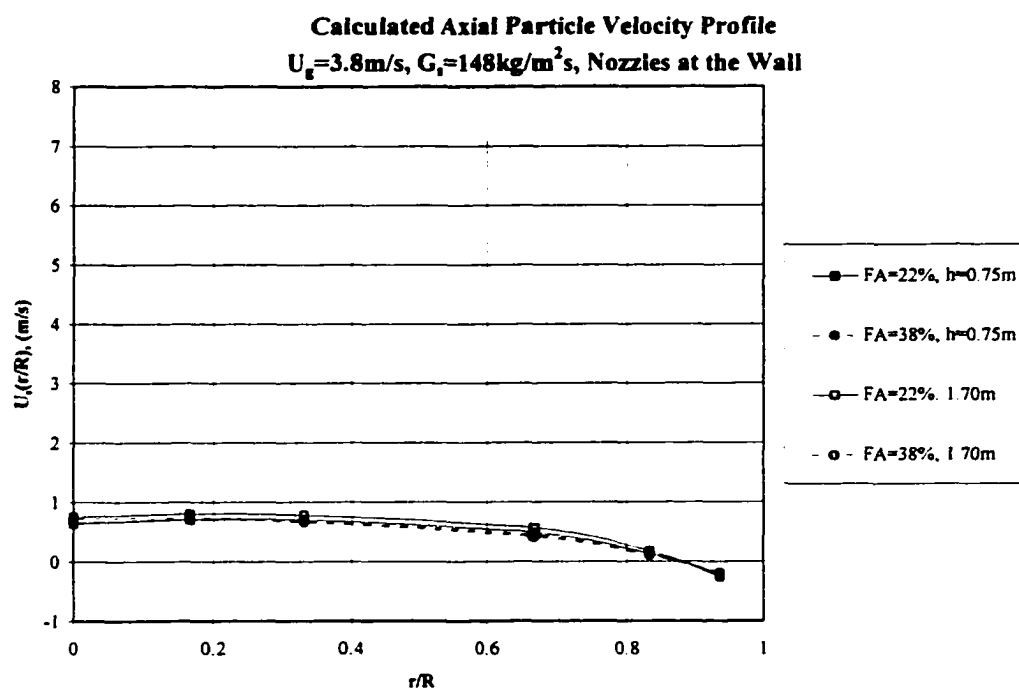


Figure 5.73: Calculated Radial Profile of the Axial Solid Velocity as a Function of the "Fluffing" Air Flowrate, High Solid Circulation Rate.

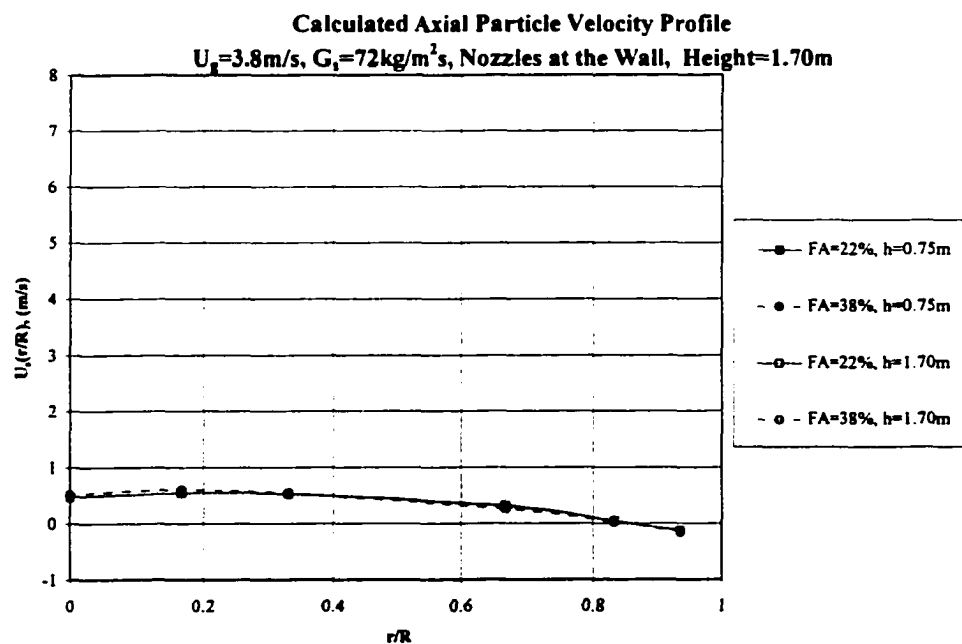


Figure 5.74: Calculated Radial Profile of the Axial Solid Velocity as a Function of the "Fluffing" Air Flowrate, Low Solid Circulation Rate.

can be combined with a correlation for the slip velocity in order to estimate the radial profile of the axial interstitial gas velocity. The correlation that is used in the present work is a modified form of Matsen equation:

$$\text{Slip velocity definition: } U_g(r/R) = U_s(r/R) + \beta[1 - \epsilon(r/R)]^{0.3} \quad (5.7a)$$

where β is a constant, primary depended on the terminal velocity of the solid.

A radial profile of the axial gas velocity can be estimated by solving the above equation with the global gas continuity equation:

$$\text{Global Gas Continuity: } U_g = \frac{1}{0} \int \{U_s(r/R) + \beta[1 - \epsilon(r/R)]^{0.3}\} \epsilon(r/R) \, r dr \quad (5.7b)$$

where U_g is the superficial gas velocity.

The application of the column-average slip velocity correlation of equation 5.7a gives only a crude estimation of the radial profile of the axial interstitial gas velocity. The most important result is the estimation of the magnitude of the gas velocity close the riser wall and axis and the shape of the profile for the different operating conditions.

Figure 5.75 shows the radial profile of the axial interstitial gas velocity for the three nozzle positions at 0.75m (5 L/D) above the nozzles. The superficial gas velocity is $U_g=3.8\text{m/s}$, the solid circulation rate $G_s=148\text{kg/m}^2\text{s}$, and the "fluffing" air flowrate is $FA=22\%$ of the total flowrate. It is shown that the nozzle position has a great effect on the gas velocity profile. This was expected since the nozzle position is also critical in determining the solid velocity profile. When the nozzles are located at the center of the riser, the radial profile of the axial gas velocity has a very peaked shape with a maximum value of about 7.7m/s at the riser axis, similar to the corresponding radial profile of the

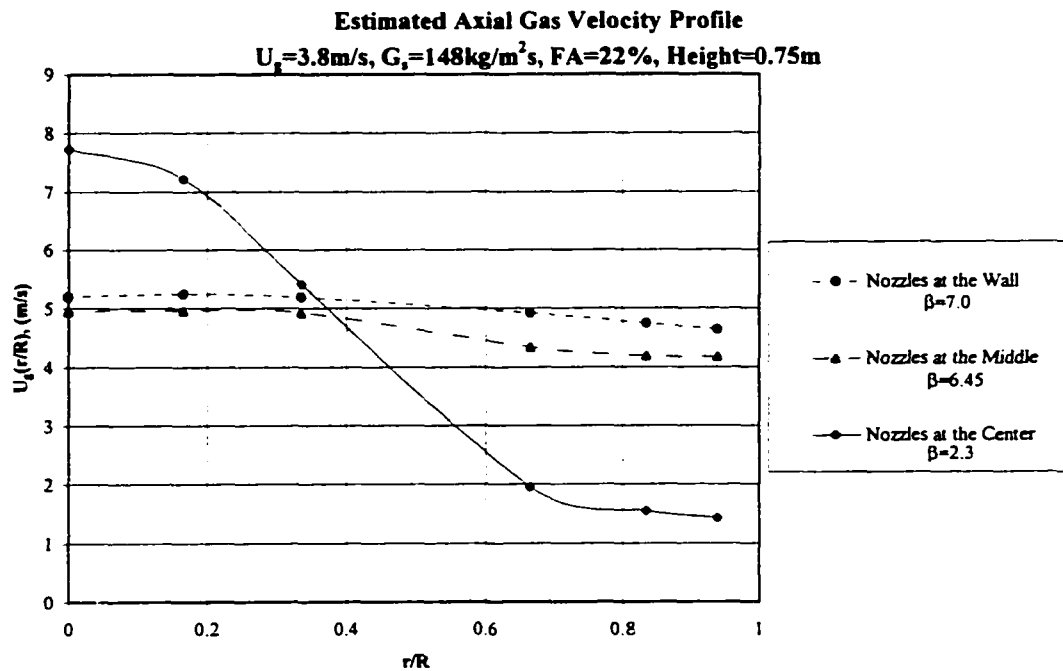


Figure 5.75: Estimated Radial Profile of the Axial Gas Velocity as a Function of the Nozzle Position, Height=0.75m.

axial solid velocity, (Figure 5.65). It can also be observed that the gas velocity close to the wall ($r/R \approx 1$) is positive although the net solid flux and the solid velocity in this section is negative. When the nozzles are located at the middle the axial gas velocity profile becomes almost flat. The velocity varies from 4.1 m/s at the wall to about 5.0 m/s at the axis. This axial velocity profile becomes slightly flatter when the nozzles are located at the wall, but the difference between these two nozzles positions is practically negligible.

Figure 5.76 shows the axial interstitial gas velocity profiles for the same operating conditions but at an elevation in the riser of 1.70 m (11.3 L/D). The general behavior of the gas velocity profile with the different nozzle positions remains the same as discussed for the 0.75 m (5 L/D) measurements. Comparing the two figures (Figures 5.75 and 5.76),

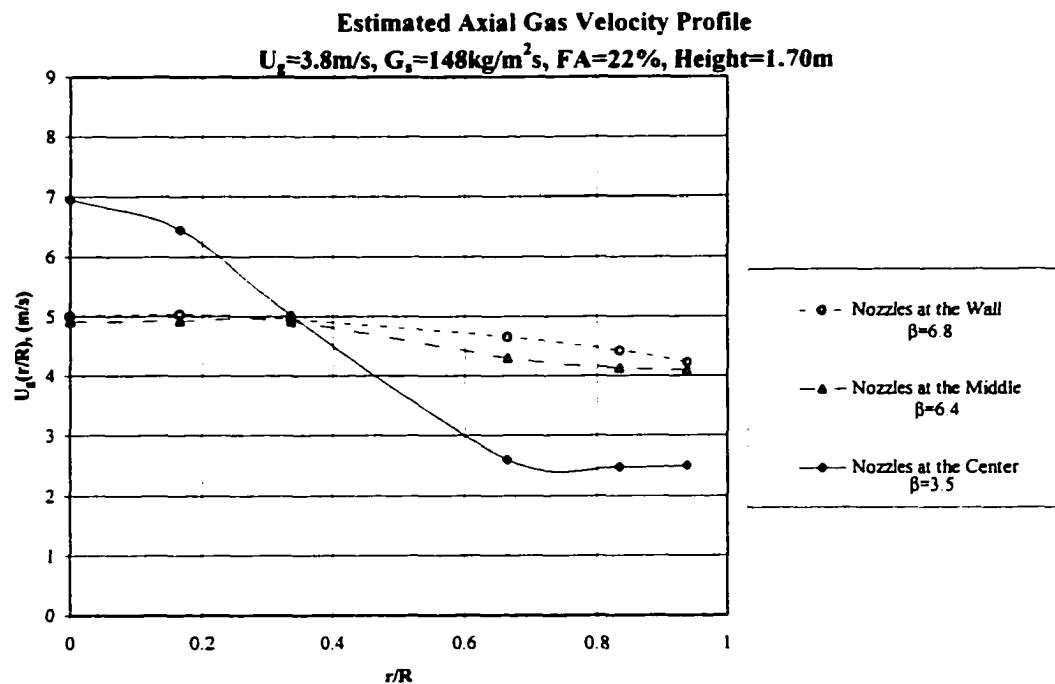


Figure 5.76: Estimated Radial Profile of the Axial Gas Velocity as a Function of the Nozzle Position, Height=1.70m.

it can be observed that the gas velocity at 1.70m (11.3 L/D) is smaller close the riser axis and higher close the riser wall than at 0.75m (5 L/D), when the nozzles are located at the center of the riser. For the same operating conditions a deceleration of solid was also observed, (Figures 5.65 and 5.66). There is not any significant change on the velocity profiles of 0.75m and 1.70m when the nozzles are located either at the middle or as close as possible to the riser wall.

Figures 5.75 and 5.76 also show the value of the constant β for each case. It may be seen that β varies with nozzle position. When the nozzles are located at the center β has a value of about 7, which becomes about 3 when the nozzles are located as close as possible to the riser wall. This demonstrate the crudeness of the method used for

calculating the slip velocity which is based on a cross-sectional average properties correlation.

Figures 5.77 and 5.78 show the radial distribution of the axial gas velocity for the reduced solid circulation rate case, ($G_s=72\text{kg/m}^2\text{s}$) and keeping all the other operating conditions constant. It is seen that the velocity gradient when the nozzles are located at the center is not as strong as in the high solid circulation rate case. The profiles for the other two nozzle positions are flat. Another observation made from these velocity profiles is that the velocity is higher close to the wall and then decreases up to a distance $r/R \approx 0.7$. This probably is another error results from the slip velocity correlation and is more evident when the nozzles are located at the center since this nozzle position gives the

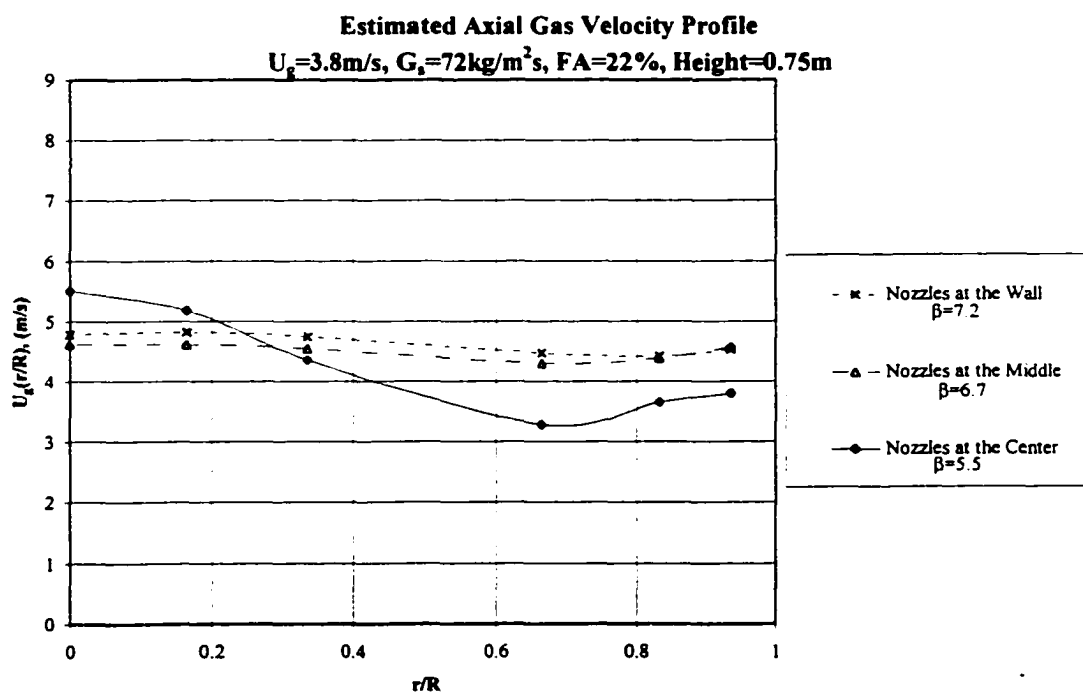


Figure 5.77: Estimated Radial Profile of the Axial Gas Velocity as a Function of the Nozzle Position, Height=0.75m.

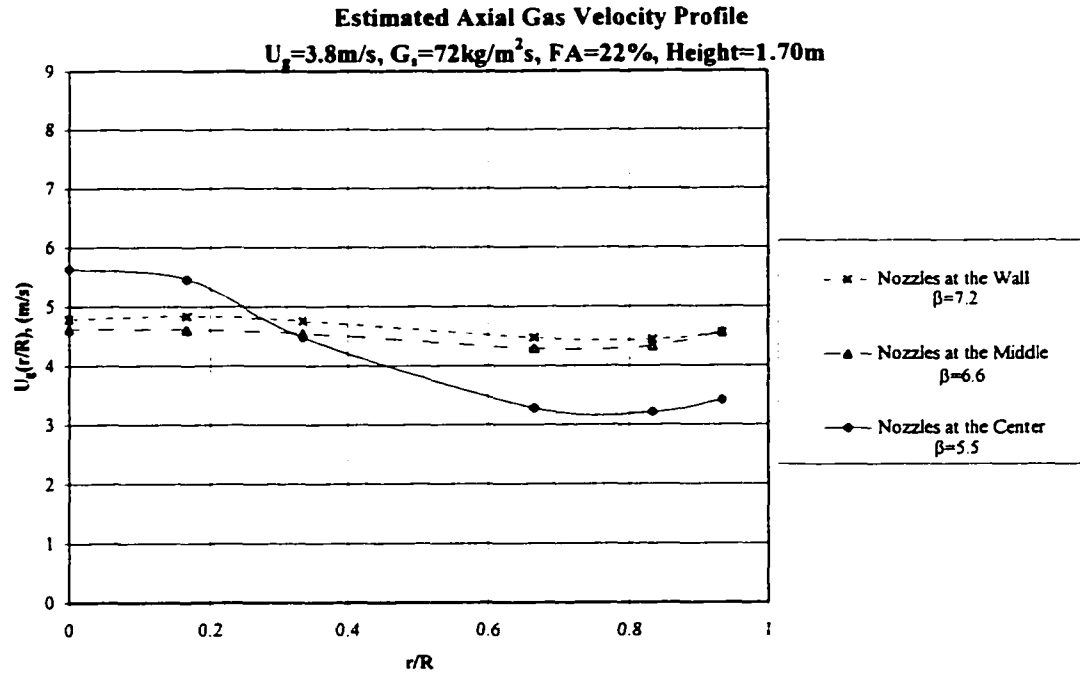


Figure 5.78: Estimated Radial Profile of the Axial Gas Velocity as a Function of the Nozzle Position, Height=1.70m.

highest solid fraction close the riser wall. The same decrease appears for the other two nozzle positions but it is much smaller since for these nozzle positions the solid fraction at the wall is much smaller.

Figures 5.79 and 5.80 show the axial gas velocity profile for a higher superficial gas velocity, $U_g=5.2\text{m/s}$. The solid circulation rate is $G_s=152\text{kg/m}^2\text{s}$ and all the other operating parameters are kept constant. It may be observed again that the strong velocity gradient that exists when the nozzles are located at the center of the riser becomes a flat profile as the nozzles are moved to the riser wall. When the nozzles are located at the center of the riser the gas velocity at 1.70m (Figure 5.80) is lower close the riser axis and higher close the riser wall than at 0.75m (Figure 5.79). On the other hand there is not any

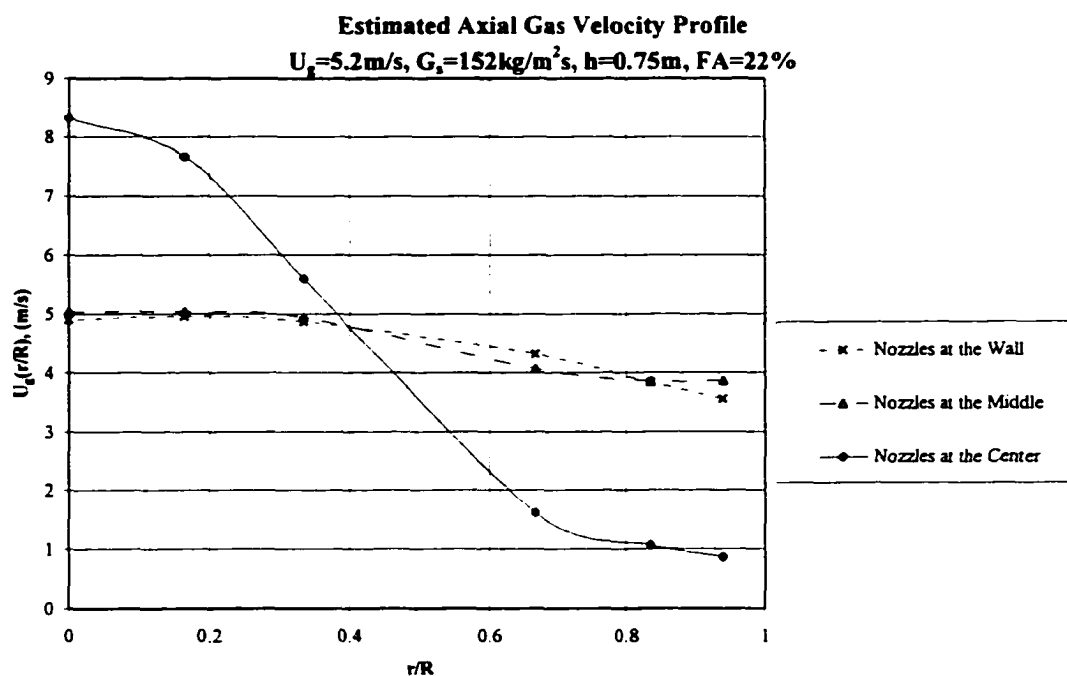


Figure 5.79: Estimated Radial Profile of the Axial Gas Velocity as a Function of the Nozzle Position, Height=0.75.

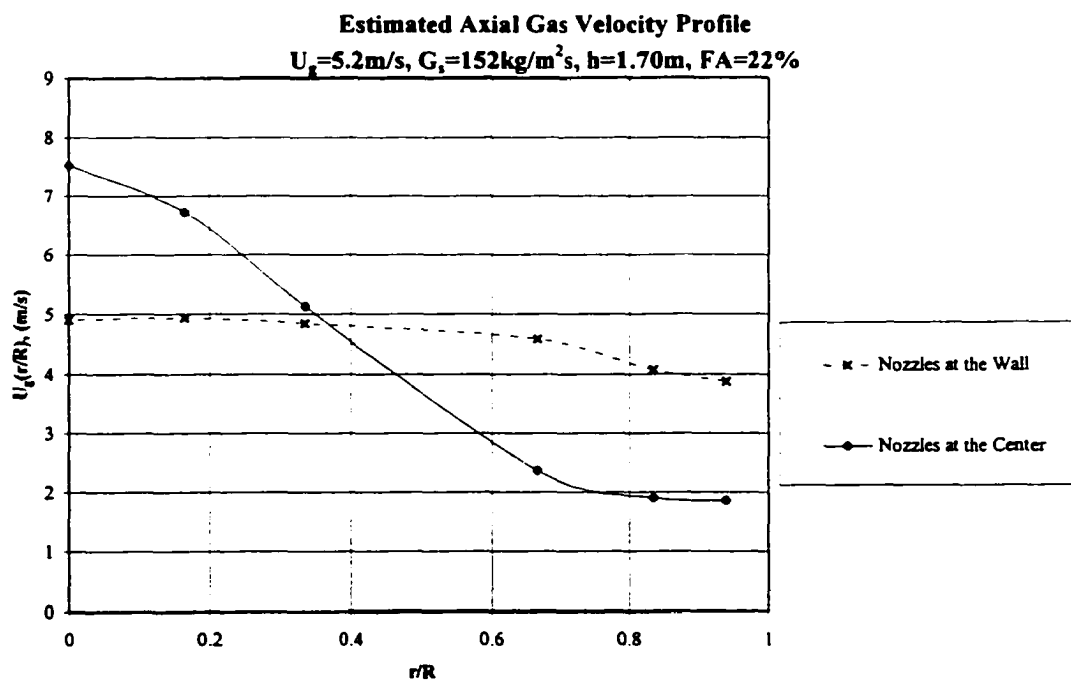


Figure 5.80: Estimated Radial Profile of the Axial Gas Velocity as a Function of the Nozzle Position, Height=1.70m.

deceleration or acceleration from 0.75m to 1.70m (5 to 11.3 L/D) for the other two nozzle positions.

Figures 5.81 and 5.82 show the effect of the "fluffing" air flowrate on the radial profile of the axial gas velocity for two solid circulation rates, $G_s=148\text{kg/m}^2\text{s}$ and $72\text{kg/m}^2\text{s}$ respectively. The superficial gas velocity is $U_g=3.8\text{m/s}$, the nozzles are located at the center of the riser and the velocity profiles are estimated at two different elevations 0.75m and 1.70m (5 and 11.3 L/D), above the nozzle inlet. It is seen (Figure 5.81) that the gas velocity close the riser axis decreases when "fluffing" air increases and the opposite takes place close the riser wall. The same behavior appears at both elevations. This happens because an increase of the "fluffing" air flowrate means a decrease of the four nozzles flowrate and since the nozzles are located at the center of the riser this results a decrease of the gas velocity close the riser axis and a increase close the riser wall. The same behavior appears for the low solid circulation rate (Figure 5.82) but in this case the increase or decrease of the gas velocity is very small.

When the nozzles are located as close as possible to the riser wall the effect of the "fluffing" air flowrate on the radial profile of the axial gas velocity is negligible for the operating conditions examined. Figures 5.83 and 5.84 show this effect for exactly the same operating conditions with the nozzles located at the center. It can be seen that the velocity profiles are almost identical for the two "fluffing" air percentages, (22% and 38%).

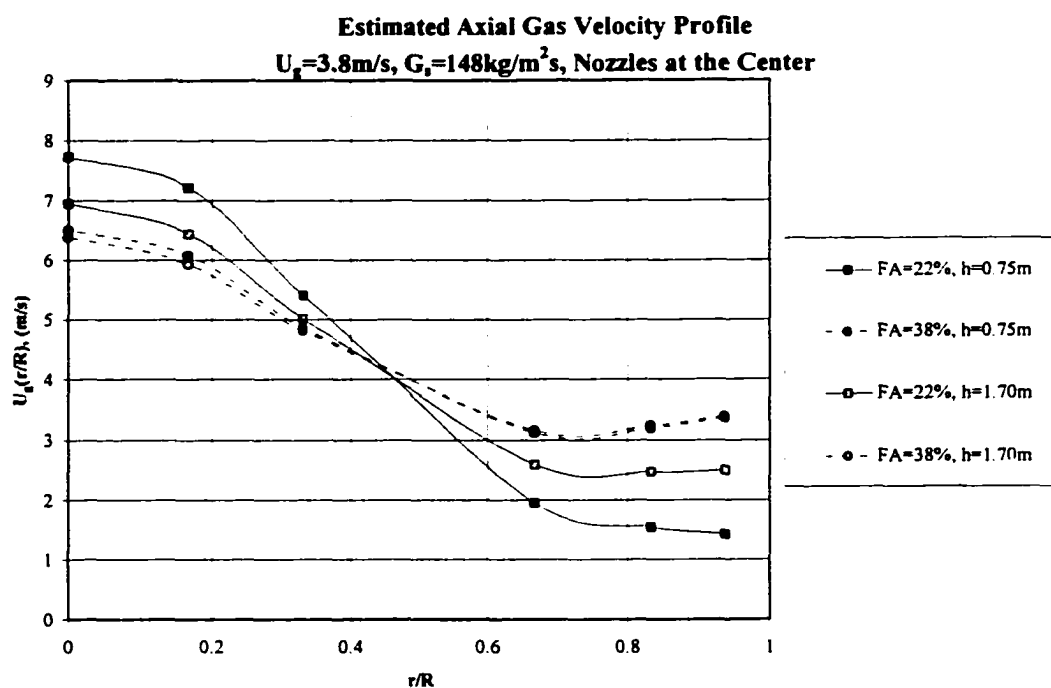


Figure 5.81: Estimated Radial Profile of the Axial Gas Velocity as a Function of the "Fluffing" Air, High Solid Circulation Rate.

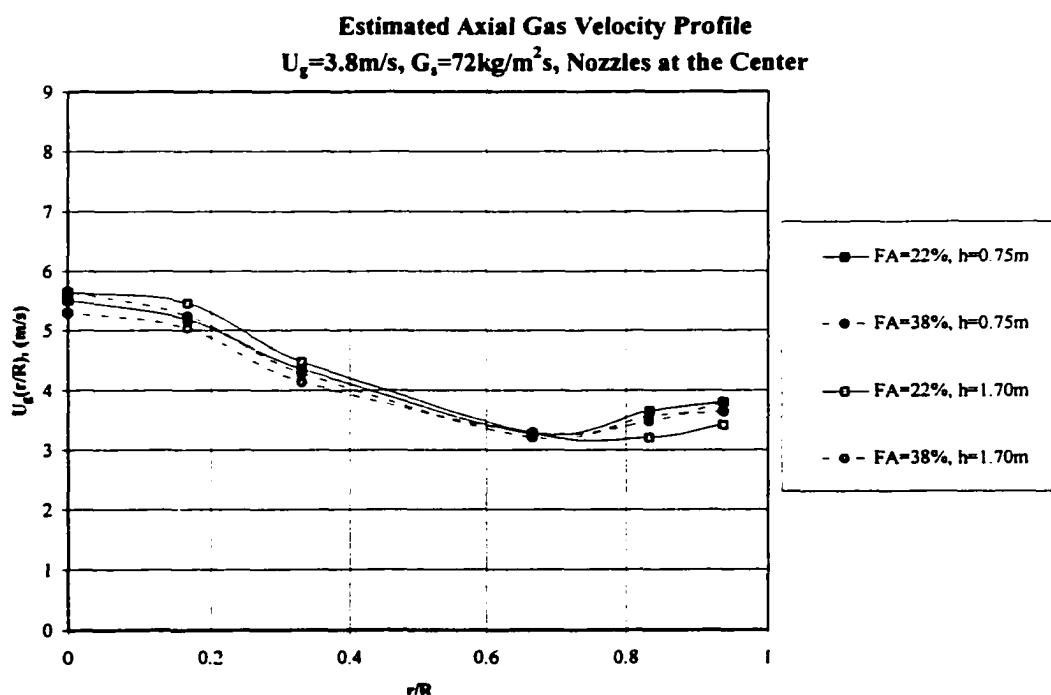


Figure 5.82: Estimated Radial Profile of the Axial Gas Velocity as a Function of the "Fluffing" Air, Low Solid Circulation Rate.

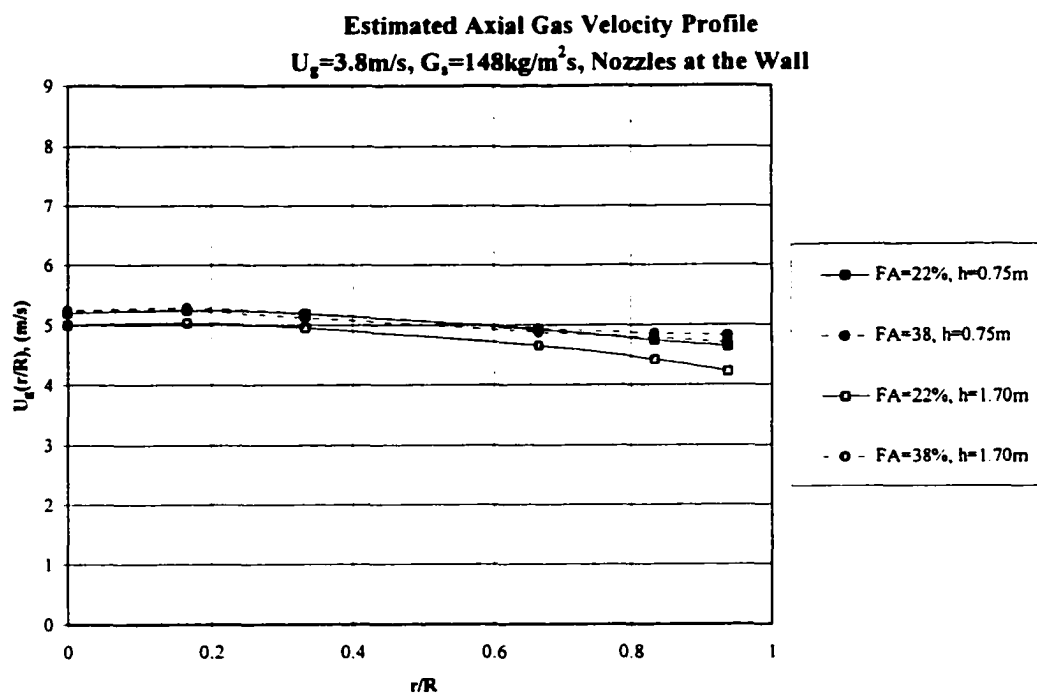


Figure 5.83: Estimated Radial Profile of the Axial Gas Velocity as a Function of the "Fluffing" Air, High Solid Circulation Rate.

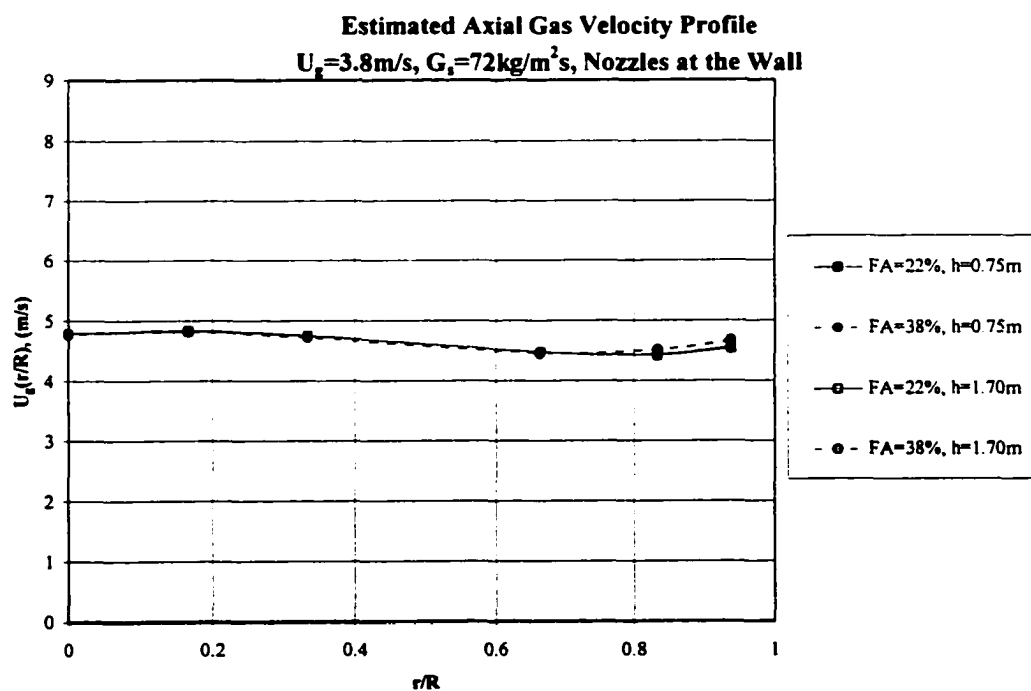


Figure 5.84: Estimated Radial Profile of the Axial Gas Velocity as a Function of the "Fluffing" Air, Low Solid Circulation Rate.

VI. CONCLUSIONS

The main conclusions of this detailed experimental work are summarized as follows:

1. Different axial solid density profiles result for different nozzle positions especially over the critical inlet section from 0 to 20 L/D. The axial voidage profile appears to show more complex acceleration and deceleration patterns when the nozzles are close to the riser axis than when they are close to the riser wall.
2. The apparent solid fraction is a function of the pressure tap spacing when this is smaller than one L/D. For larger spacings, the apparent solid fraction is not affected by the spacing if the axial solid density is constant over the section where measurements take place. When there is solid acceleration the apparent solid fraction decreases with tap spacing and when the solid decelerates the apparent solid fraction increases with the pressure tap spacing.
3. The radial solid profile is strongly dependent on the nozzle configuration. A strong radial solid concentration gradient exists if the nozzles are close to the riser axis. This results in poor gas-solid mixing and a high degree of backmixing. On the other hand when the nozzles are located closer to the riser wall the radial solid density profile is much flatter which results in better mixing and a higher level of performance of the riser as a chemical reactor.
4. The "fluffing" air influence on the axial solid density profile depends on the nozzle position. When the nozzles are at the axis of the riser, lower "fluffing" air

flowrates give stronger solid acceleration sections, increasing the gas-solid contact and minimizing the backmixing in the section 0-10 L/D. As the "fluffing" air increases these sections become constant density sections or deceleration sections for very high "fluffing" air flowrates. As the nozzles are moved to the riser wall the effect of the "fluffing" air becomes negligible.

5. The "fluffing" air slightly affects the radial solid distribution. Different behavior is apparent for different nozzle positions and different elevations.
6. A new pair of pressure probes have been developed for the radial pressure measurements. The new probes are designed such that their influence on the flow field is small.
7. The radial pressure measurements confirm that a radial pressure gradient exists in the riser. The pressure is higher close the wall and the order of magnitude of the pressure difference between centerline and wall is the same as the order of magnitude of the axial pressure difference over an elevation change of one radius.
8. The shape of the radial pressure gradient is dependent on the nozzle positions. The parabolic profile that exists when the nozzles are at the center becomes flatter as the nozzles are moved towards the riser wall, which is in agreement with the corresponding radial solid density profile changes.
9. There is no significant effect of the "fluffing" air on the radial pressure distribution.
10. An improved technique was developed for the local solid flux measurements.

11. The nozzle radial position slightly affects the radial profile of the axial net solid flux. When the nozzles are located at the center the profile has a parabolic shape with maximum net flux at the riser axis. When the nozzles are at the middle the profile becomes a little flatter close the axis. When the nozzles are moved at the wall the profile appears to have a maximum at $r/R \approx 0.2$ and not at the axis.
12. There is no significant effect of the "fluffing" air flowrate on the net flux profile.
13. The radial profile of the axial solid velocity is strongly affected from the nozzle configuration. A strong radial velocity gradient exists when the nozzles are close to the riser axis. When the nozzles are at the middle this profile becomes flatter and finally with the nozzles as close as possible to the riser wall the radial distribution of the axial solid velocity becomes almost flat along the entire riser radius.
14. The estimated axial gas velocity profile exhibits the same behavior as does the axial solid velocity profile. The strong radial gradients that exist with the nozzles at the center disappear when the nozzles are moved at the wall. It is seen that the gas and solid velocity profiles as do the radial solid density and radial pressure difference profiles have strong gradients when the nozzles are located at the center. These gradients should have negative results in terms of reaction selectivity and yield. On the other hand the flat profiles that exist with the nozzles at the wall should result in higher performance of the riser as a chemical reactor.
15. The effect of the "fluffing" air flowrate on the axial solid velocity is significant when the nozzles are located at the center of the riser. Higher "fluffing" air

flowrate for this case causes a decrease of the solid velocity close the riser axis. The "fluffing" air effect on the solid velocity is negligible when the nozzles are located at middle or close the riser wall.

16. The effect of the "fluffing" air on the axial gas velocity is also negligible when the nozzles are located at the middle or at the wall. When the nozzles are located at the center of the riser an increase of the "fluffing" air causes a decrease of the gas velocity close the riser axis and an increase close to the riser wall.
17. A complete set of internally consistent data is presented herein. This database can provide the necessary information for an improved phenomenological design of riser reactors and also can be used for the development and calibration of two-dimensional CFD models.

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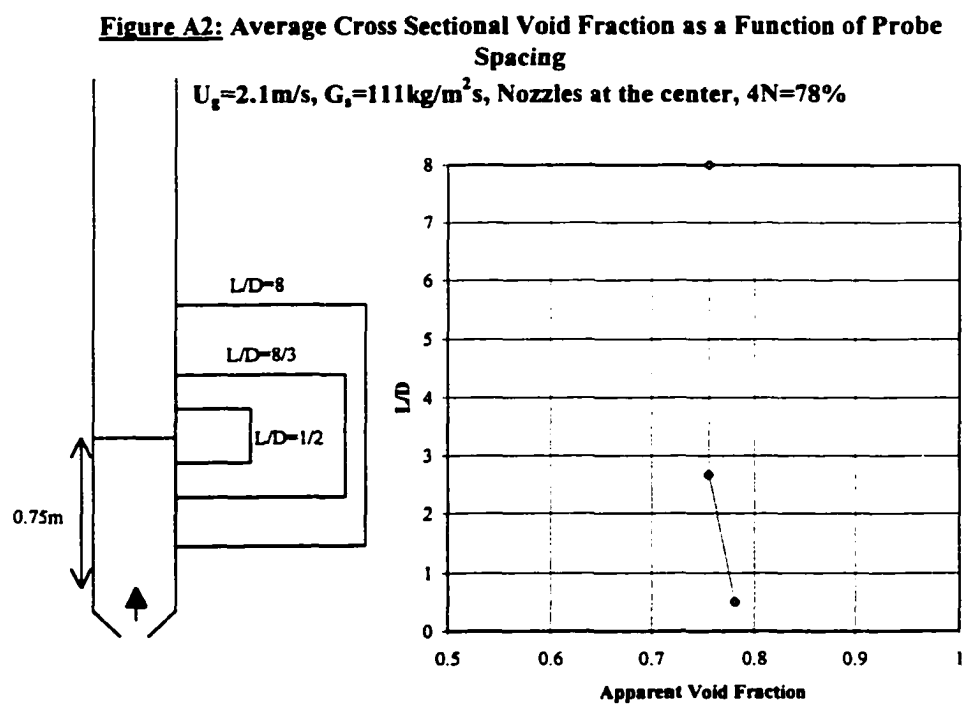
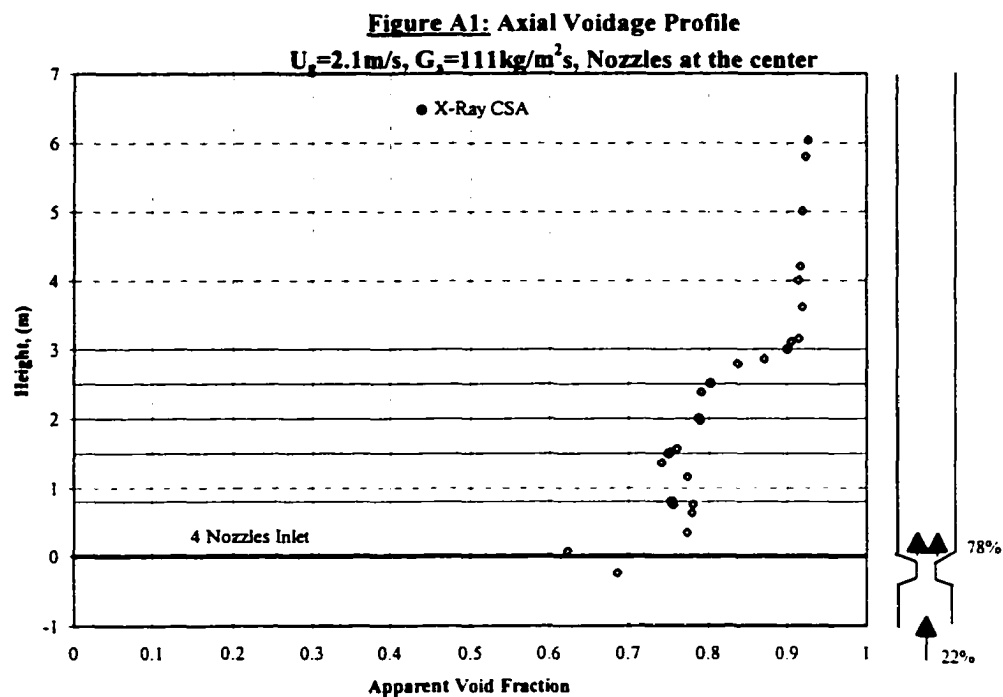
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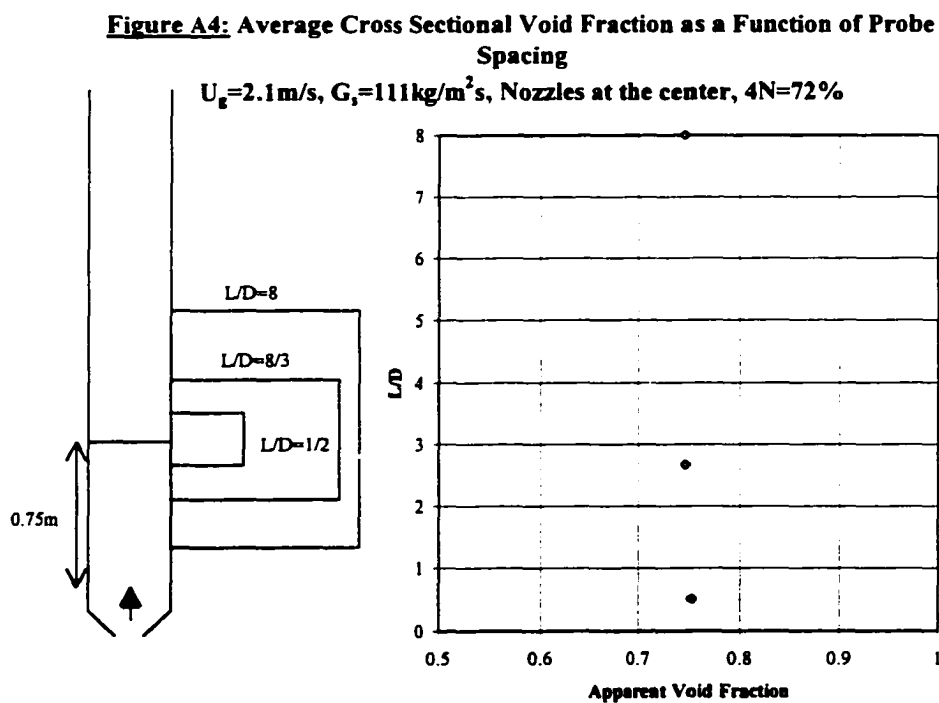
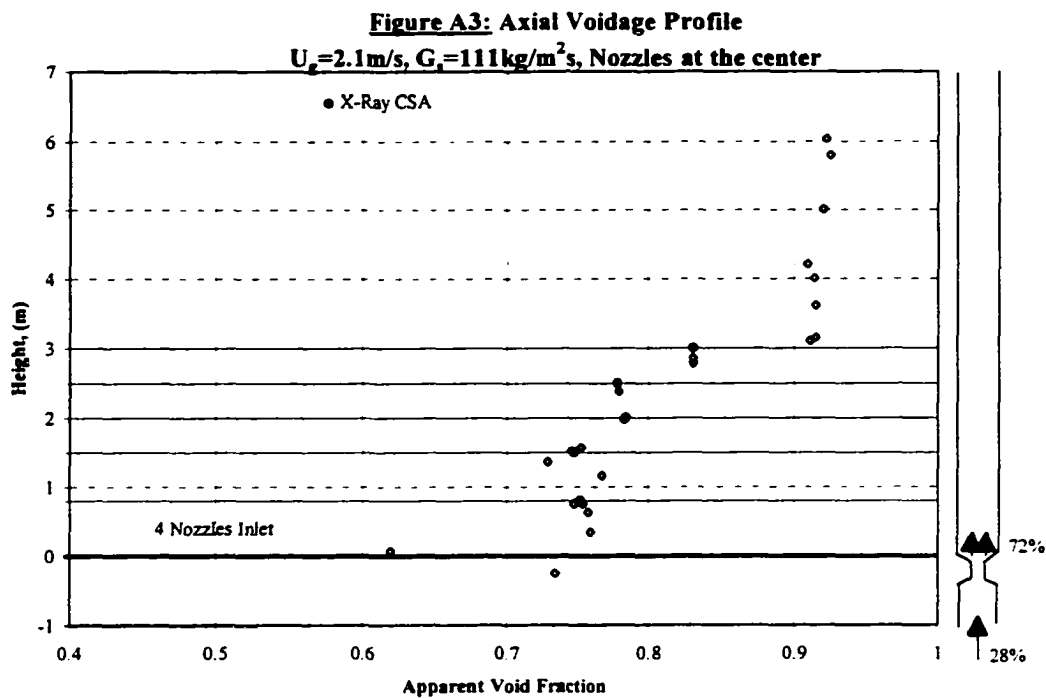
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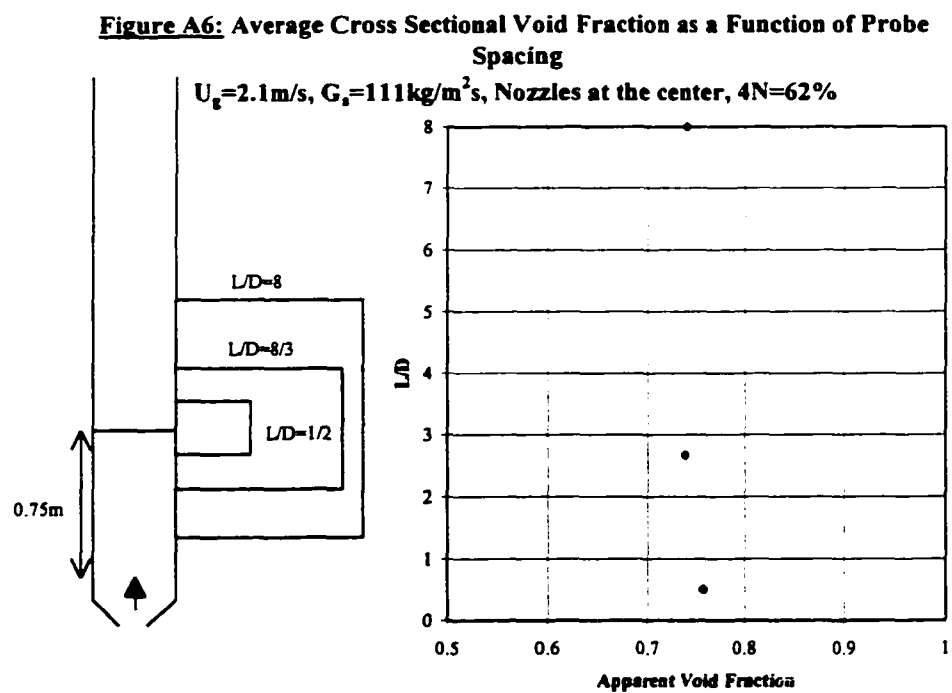
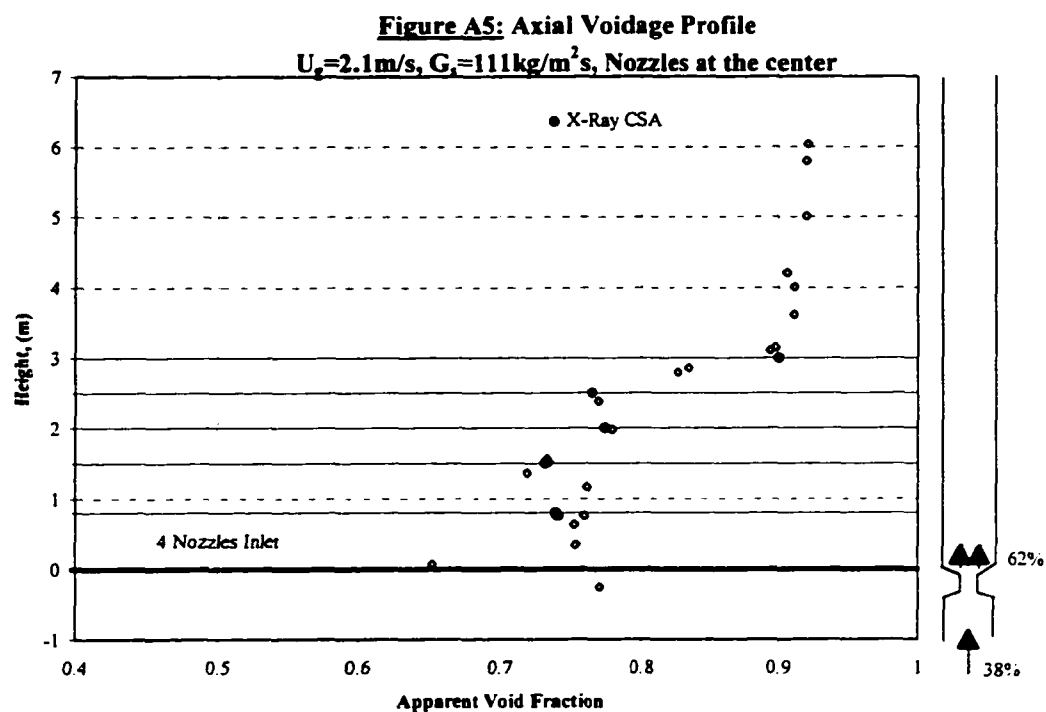
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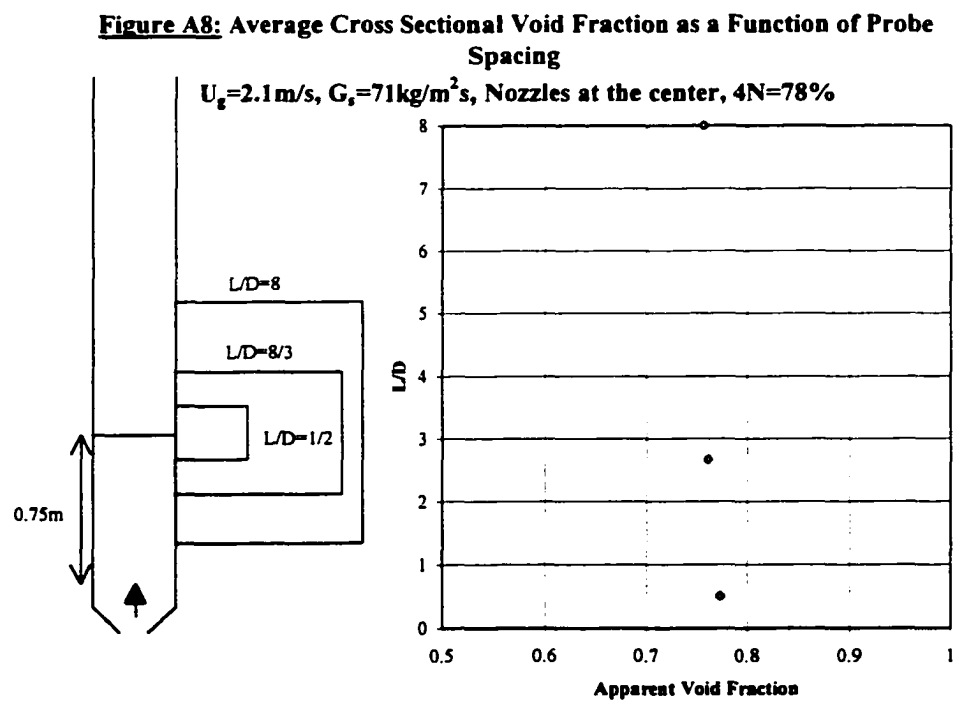
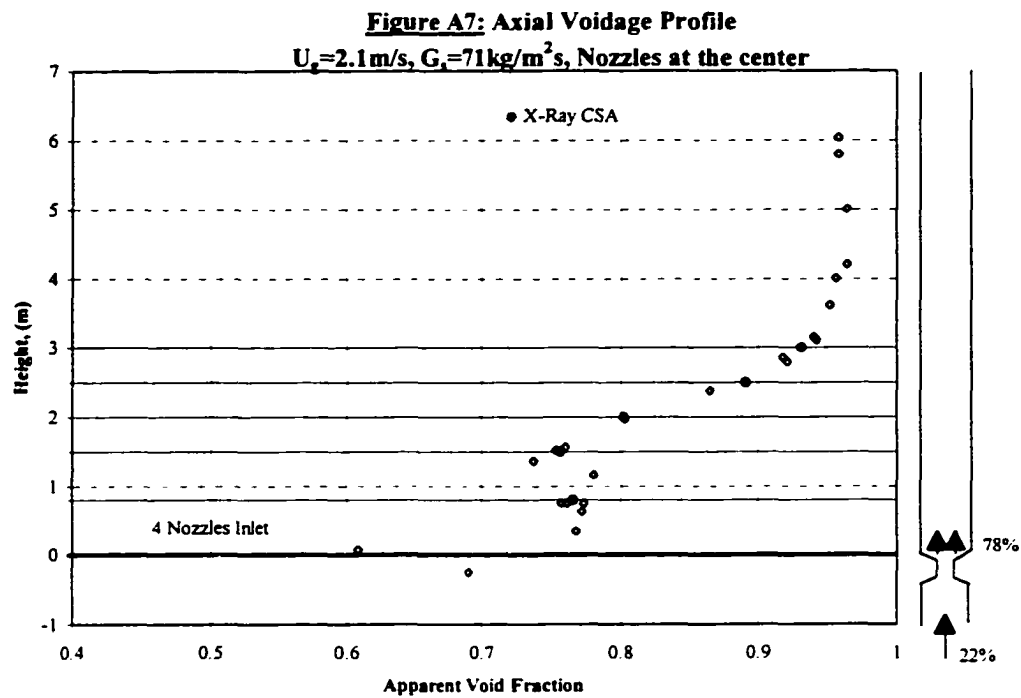
APPENDIX A

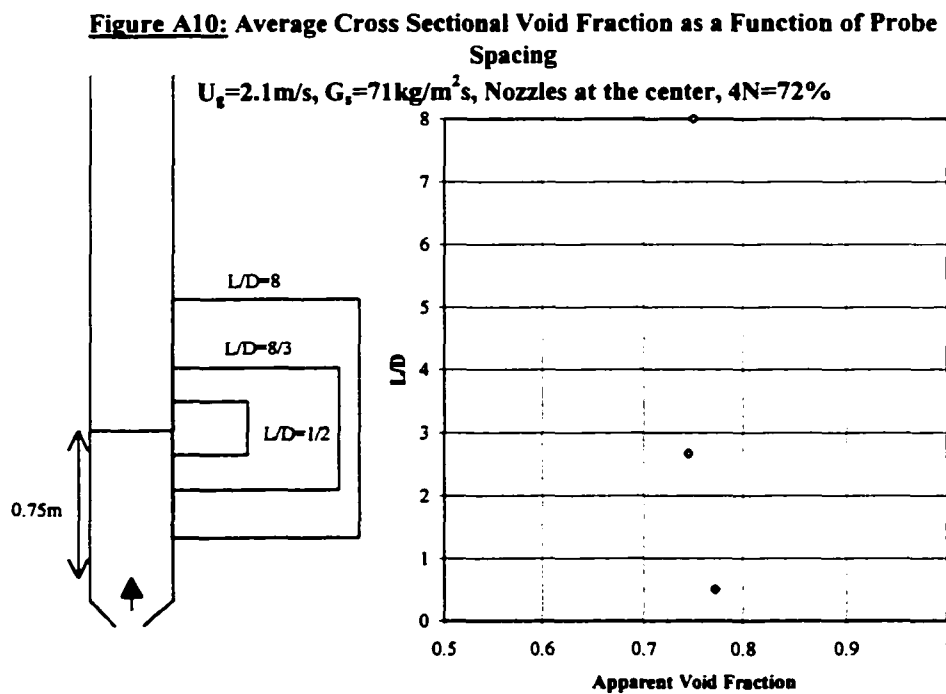
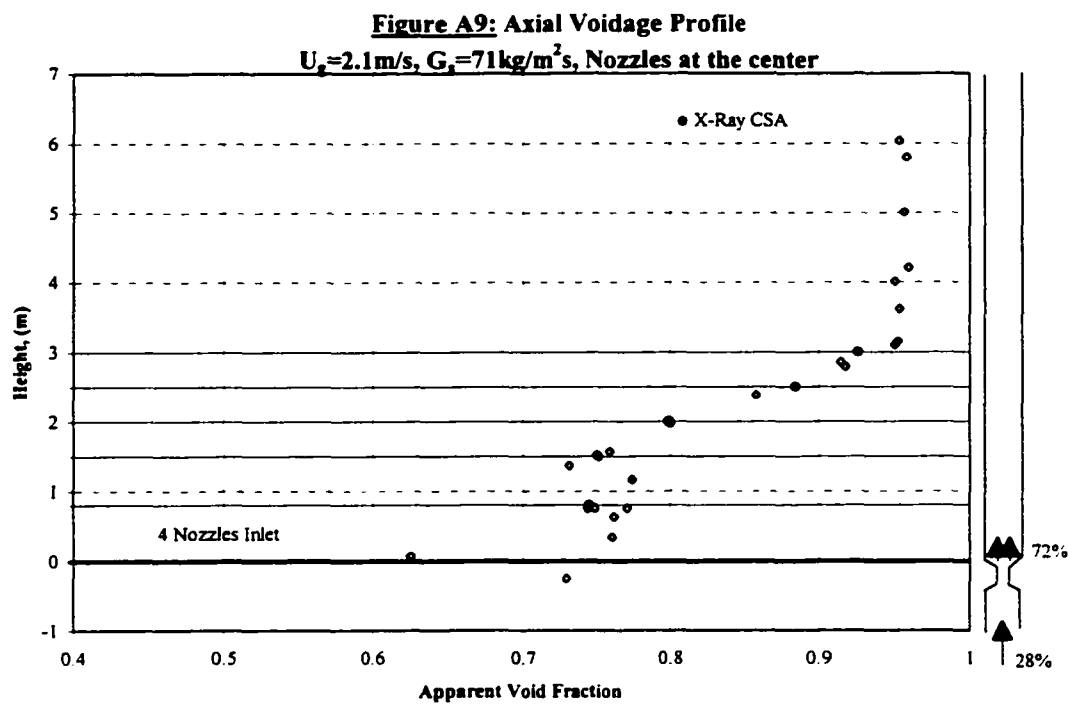
In Appendix A are presented all the experimental results with the nozzles located at the center of the riser. These results include axial and radial solid distribution and the effect of the pressure tap spacing on the axial pressure difference measurements for a variety of operating conditions, gas superficial velocity, solid circulation rate and "fluffing" air flowrate.











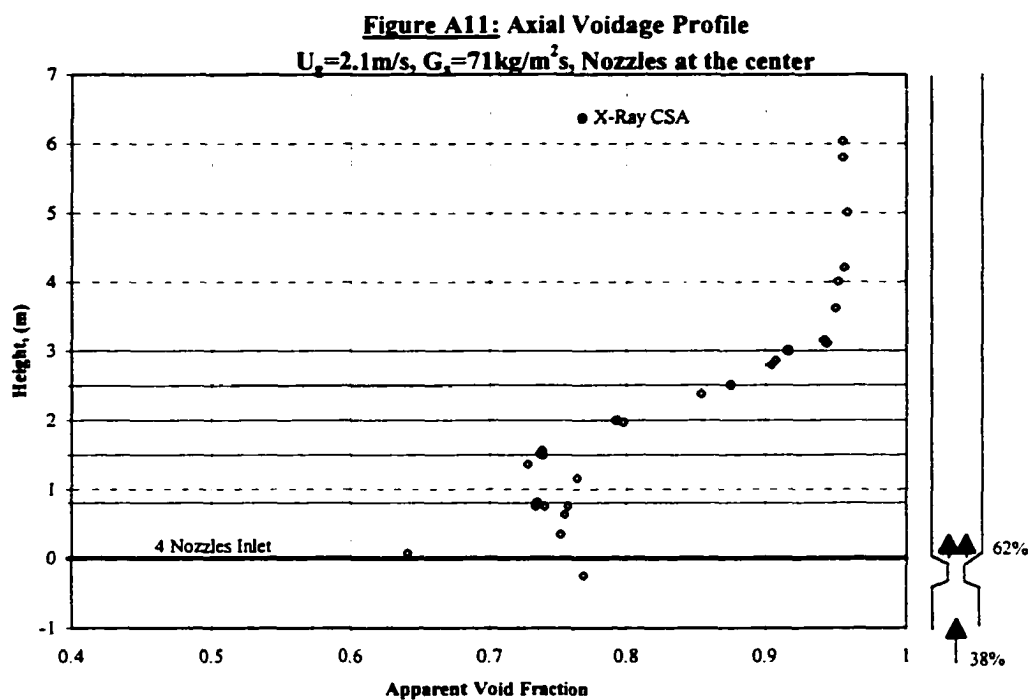
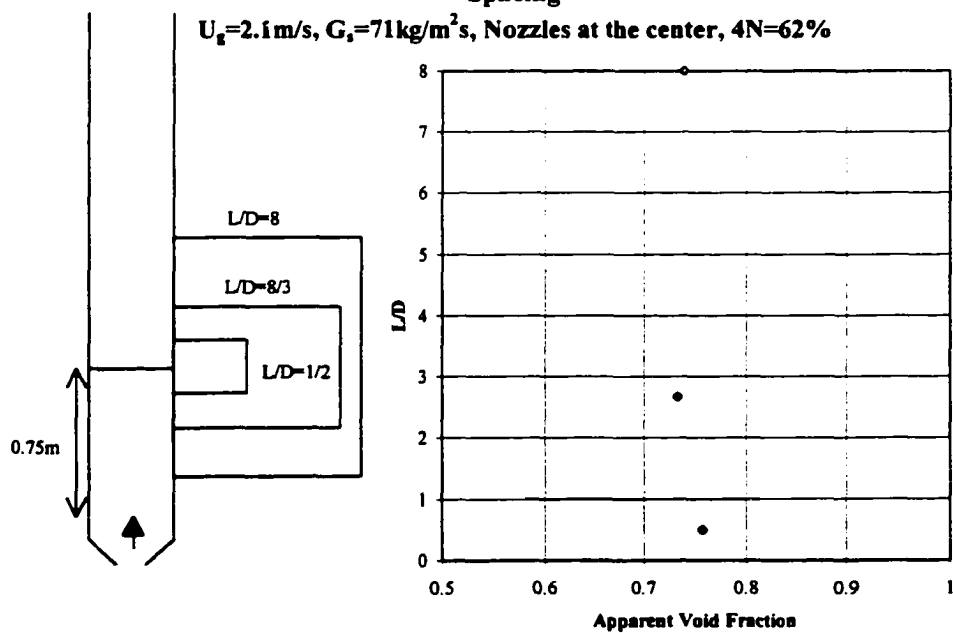
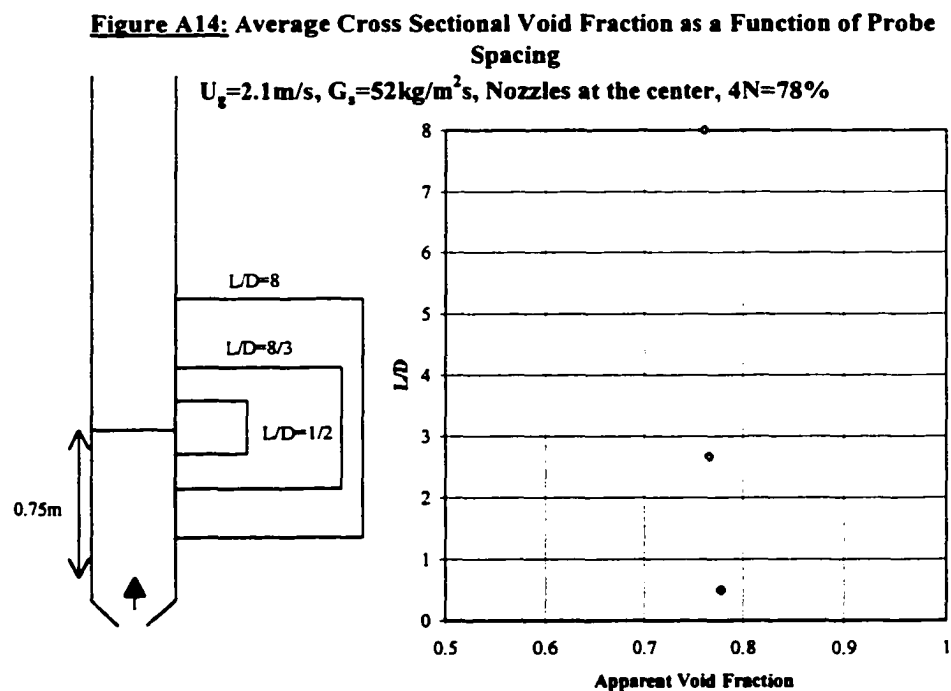
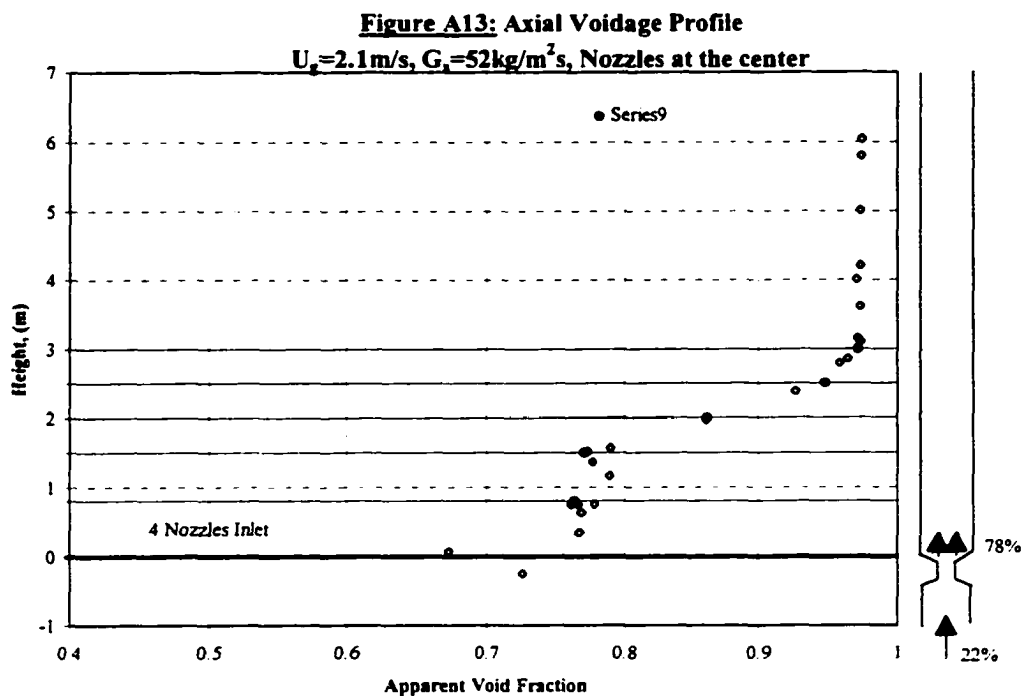


Figure A12: Average Cross Sectional Void Fraction as a Function of Probe Spacing





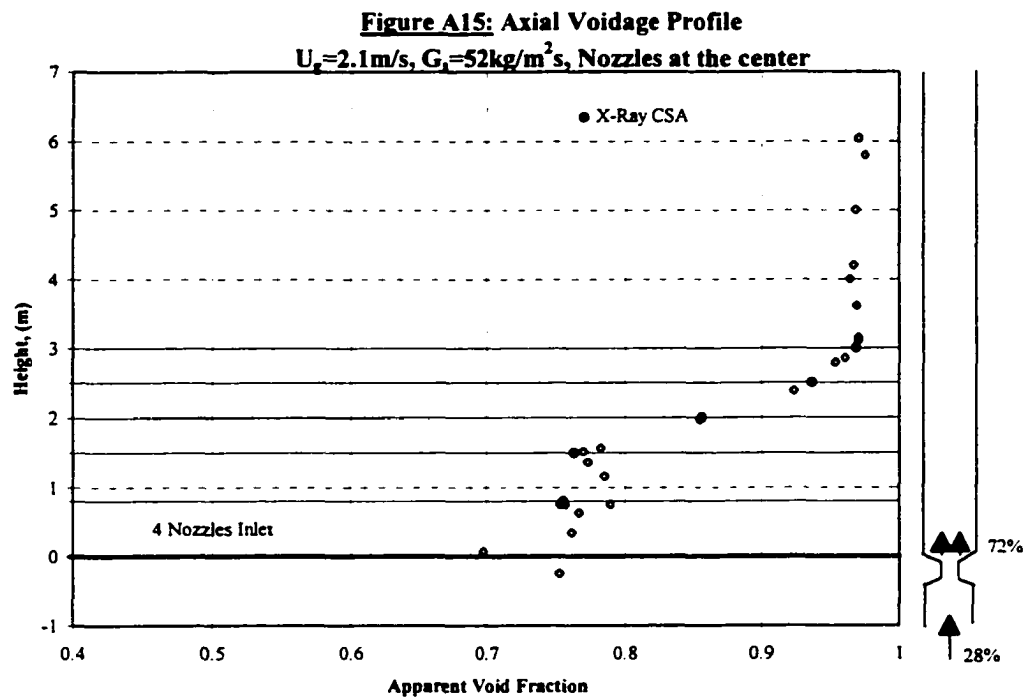
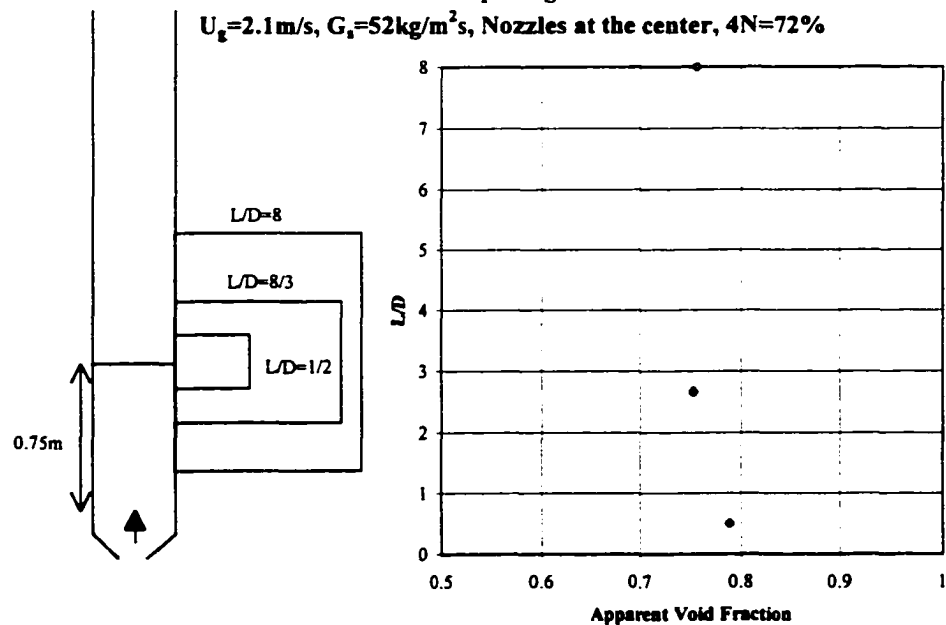


Figure A16: Average Cross Sectional Void Fraction as a Function of Probe Spacing



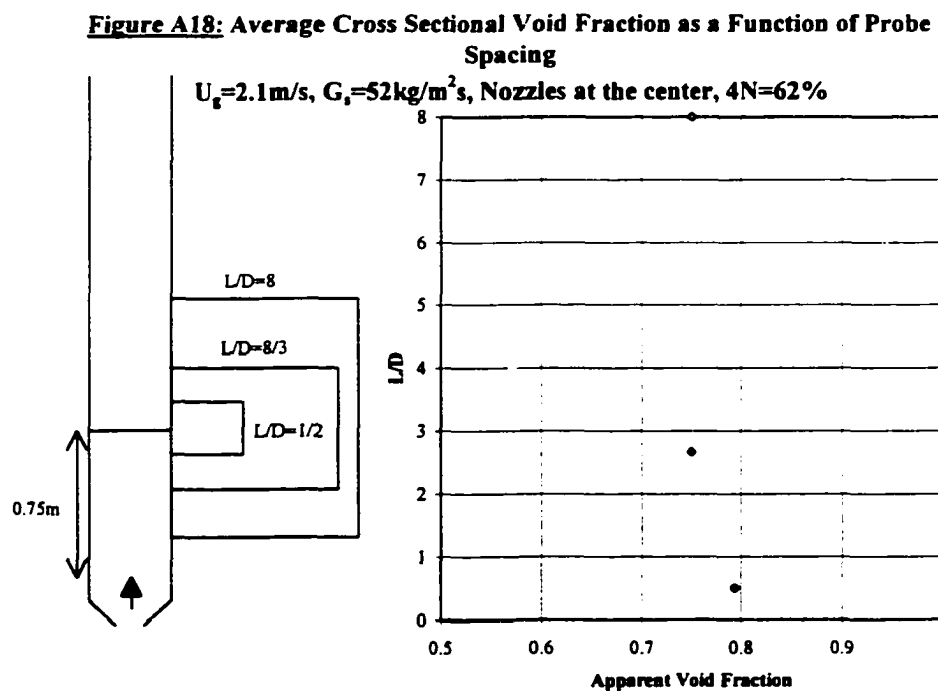
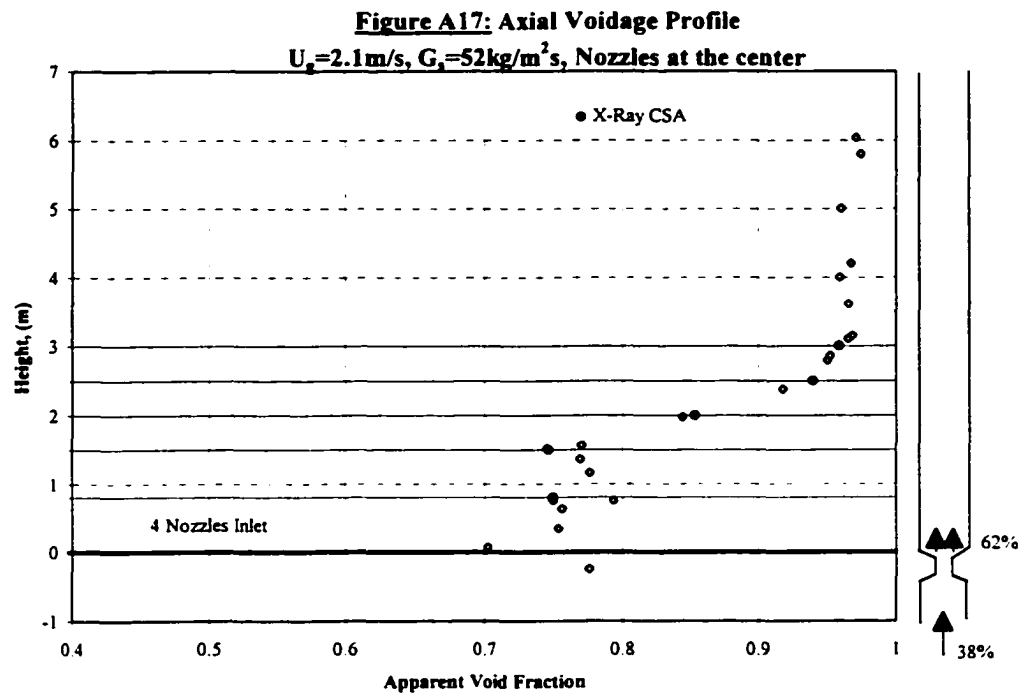


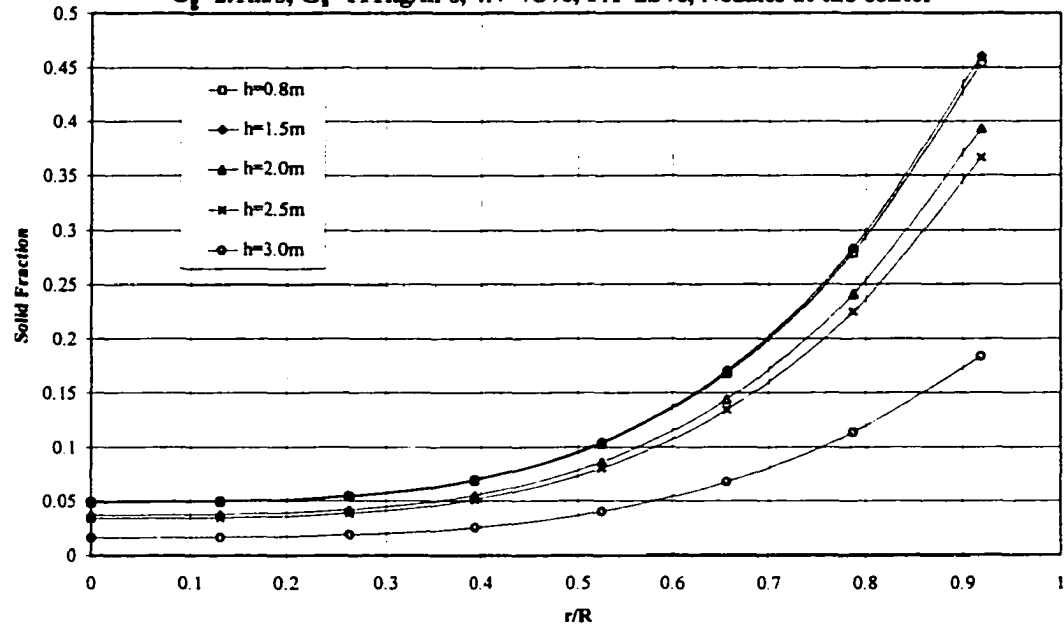
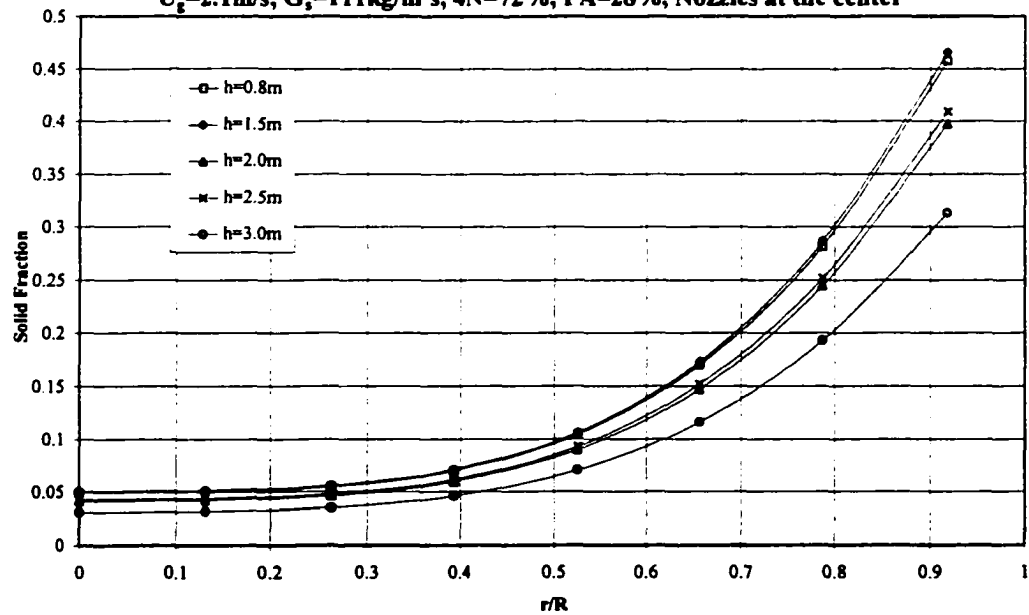
Figure A19: Radial Solid Profile $U_g=2.1\text{m/s}$, $G_s=111\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center**Figure A20: Radial Solid Profile** $U_g=2.1\text{m/s}$, $G_s=111\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center

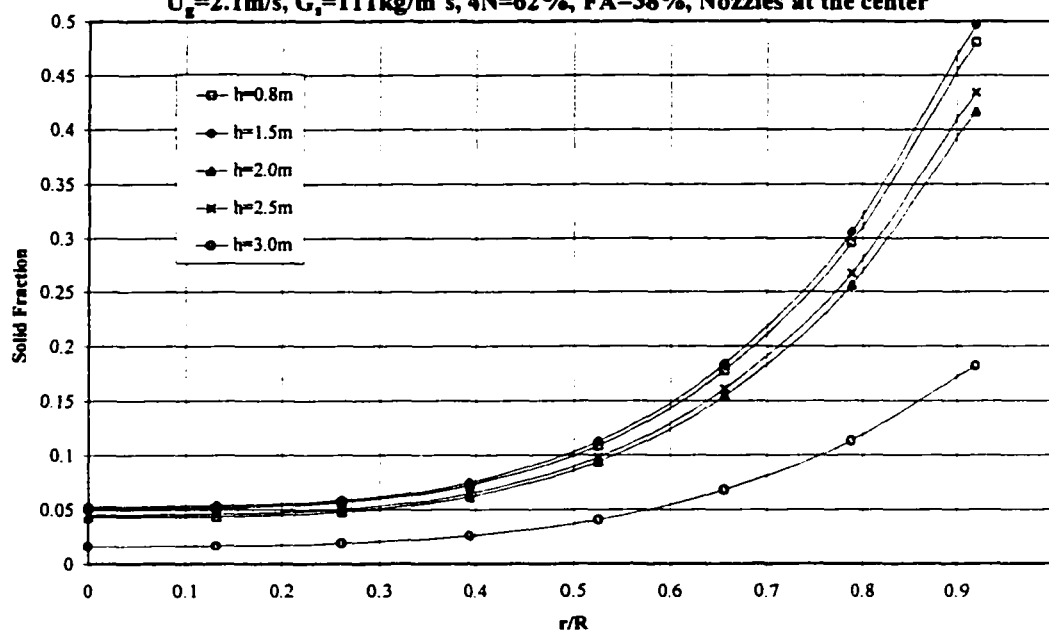
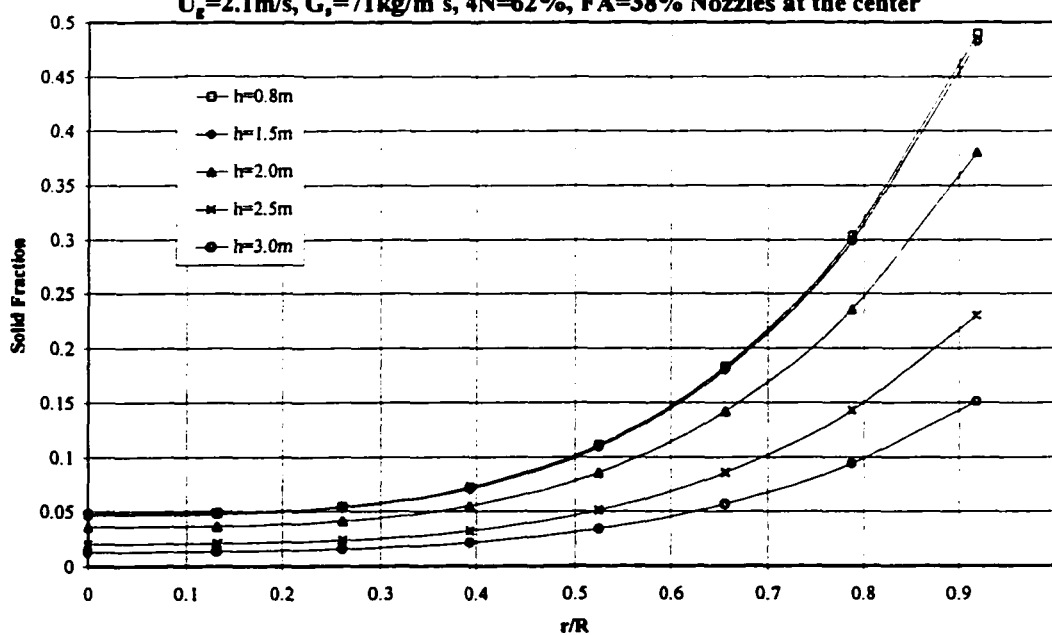
Figure A21: Radial Solid Profile $U_g=2.1\text{m/s}$, $G_s=111\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center**Figure A22: Radial Solid Profile** $U_g=2.1\text{m/s}$, $G_s=71\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$ Nozzles at the center

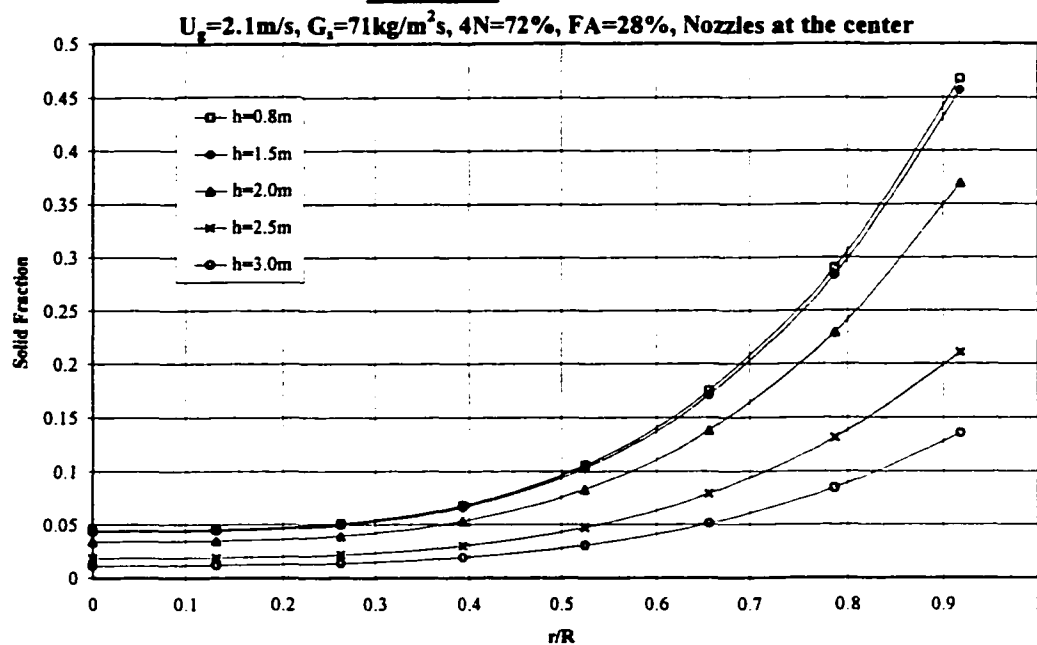
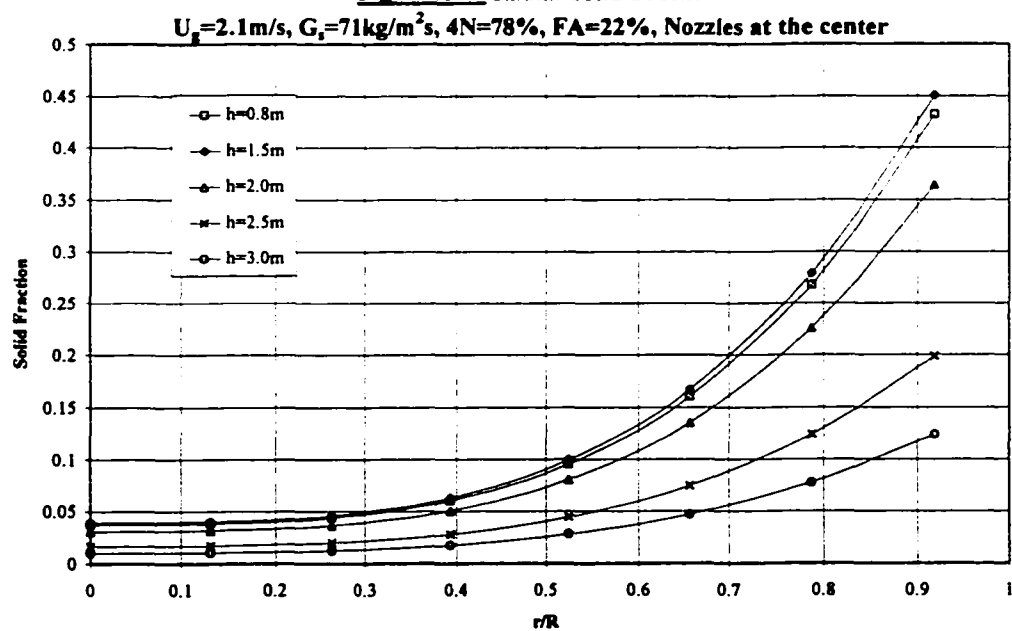
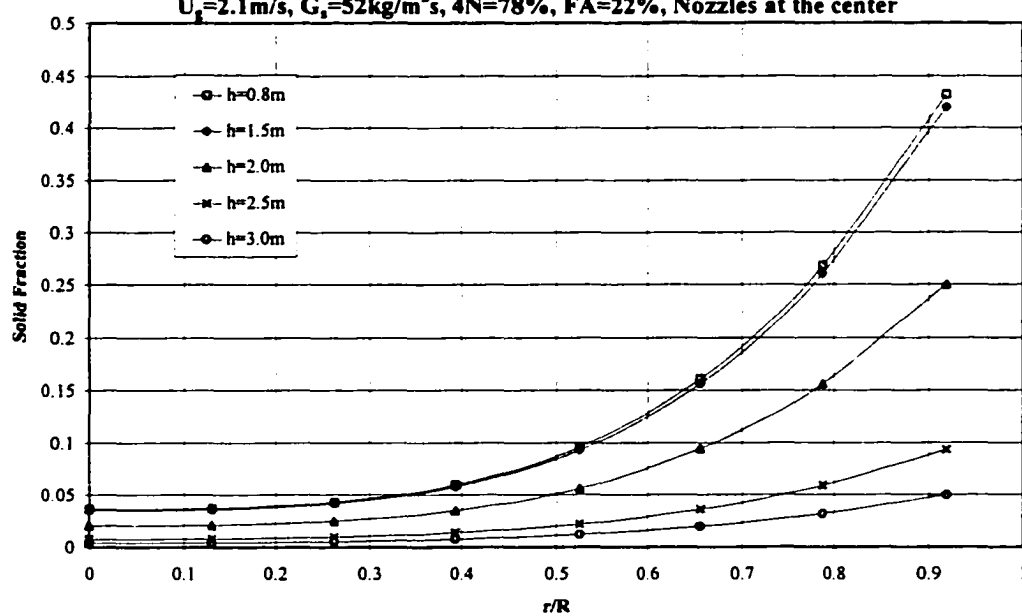
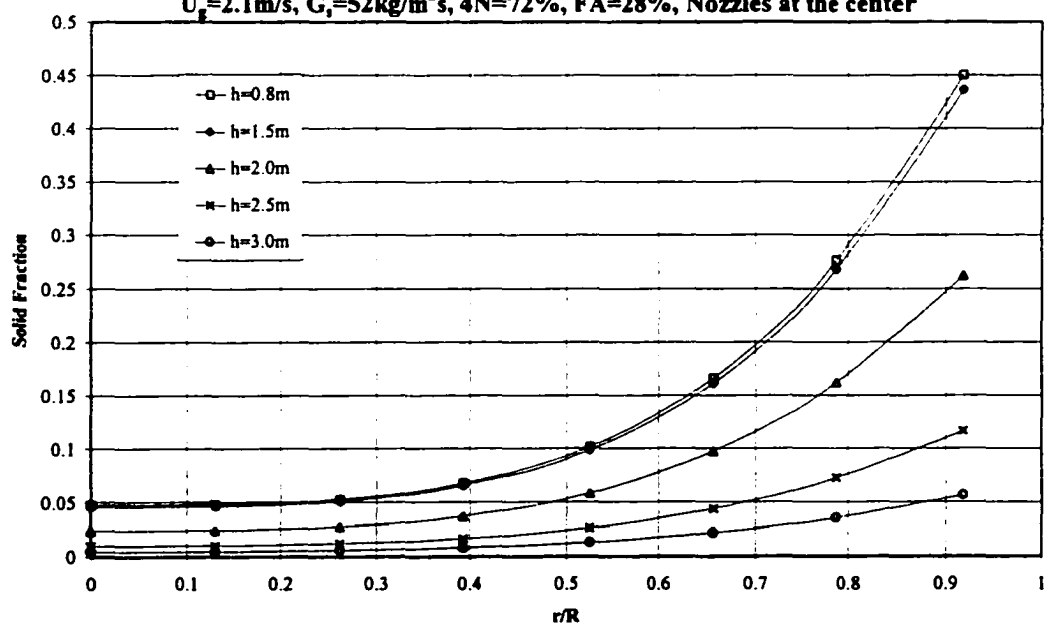
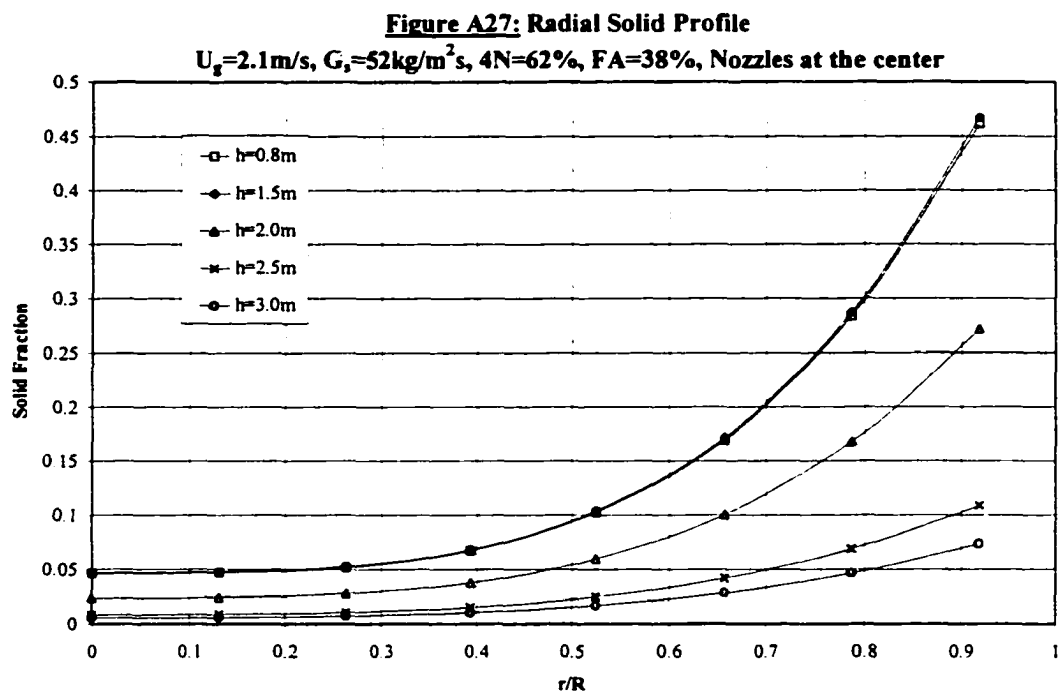
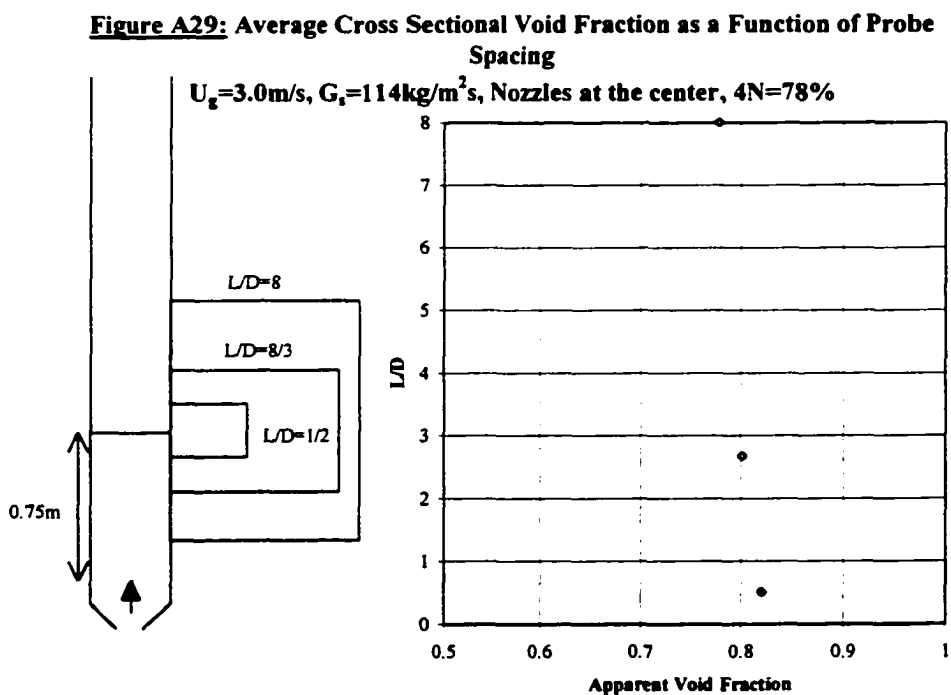
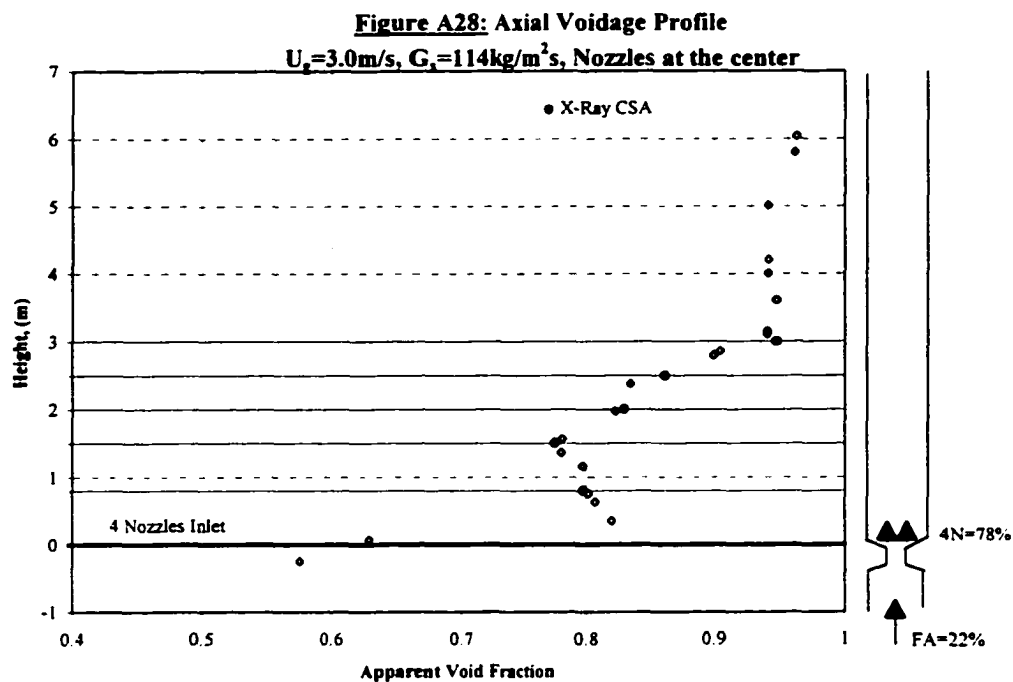
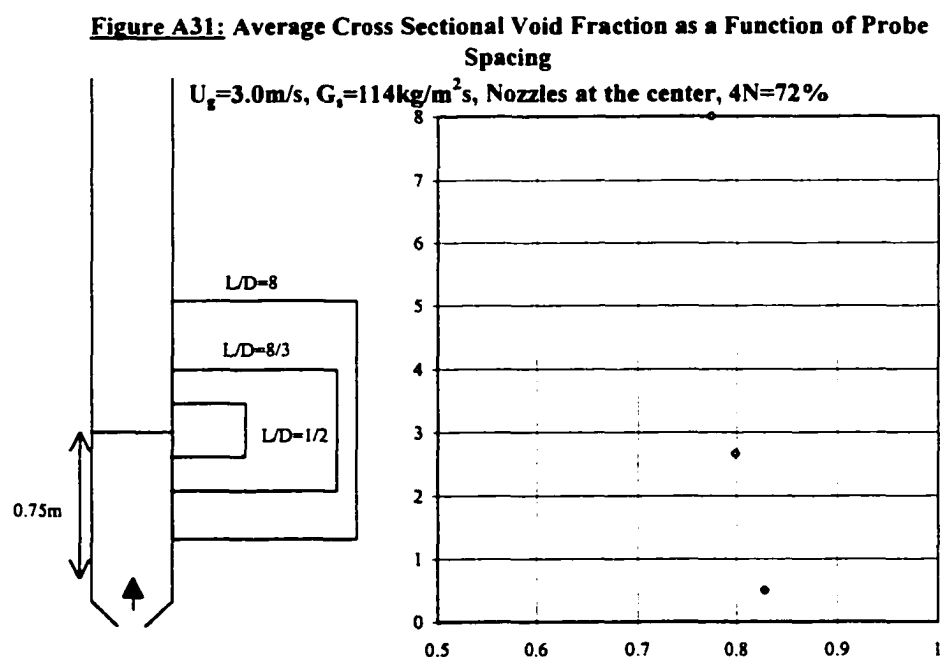
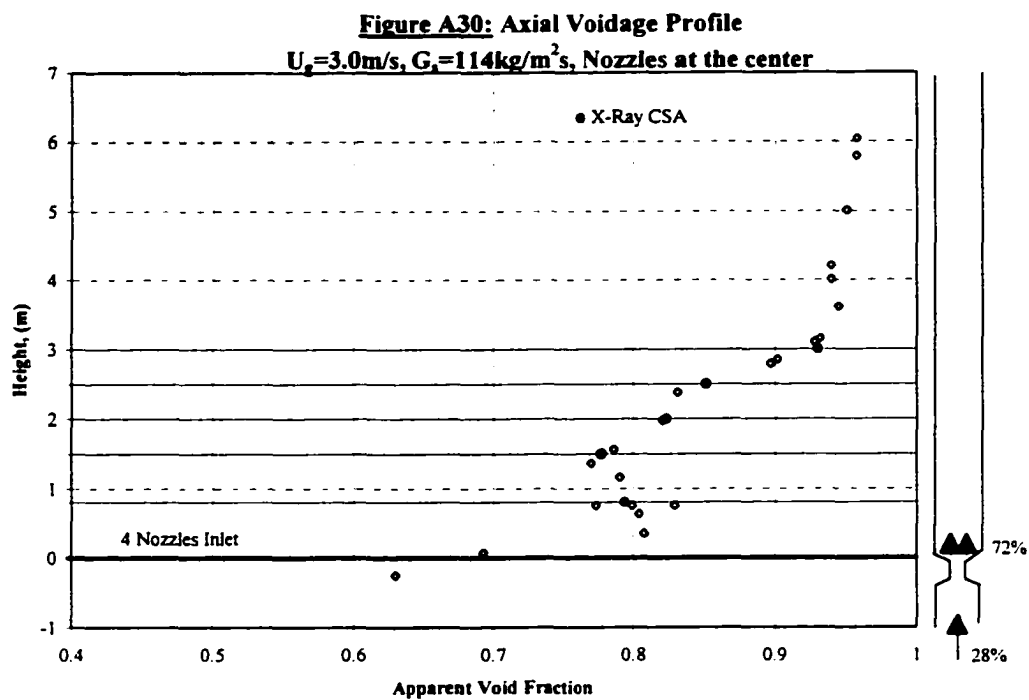
Figure A23: Radial Solid Profile**Figure A24: Radial Solid Profile**

Figure A25: Radial Solid Profile $U_g=2.1\text{m/s}$, $G_s=52\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center**Figure A26: Radial Solid Profile** $U_g=2.1\text{m/s}$, $G_s=52\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center







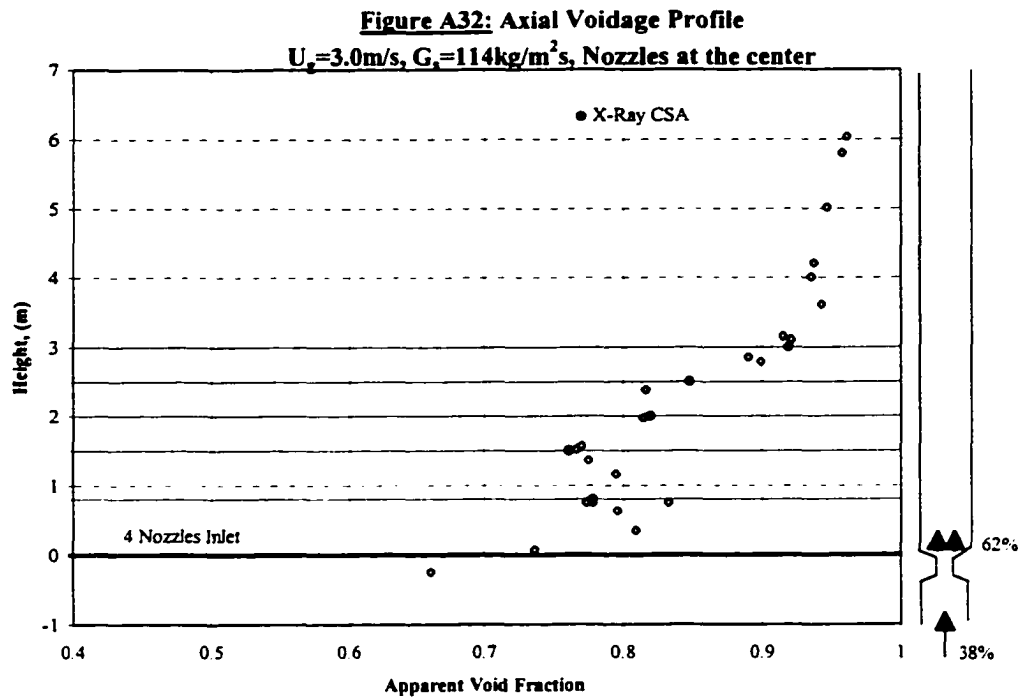
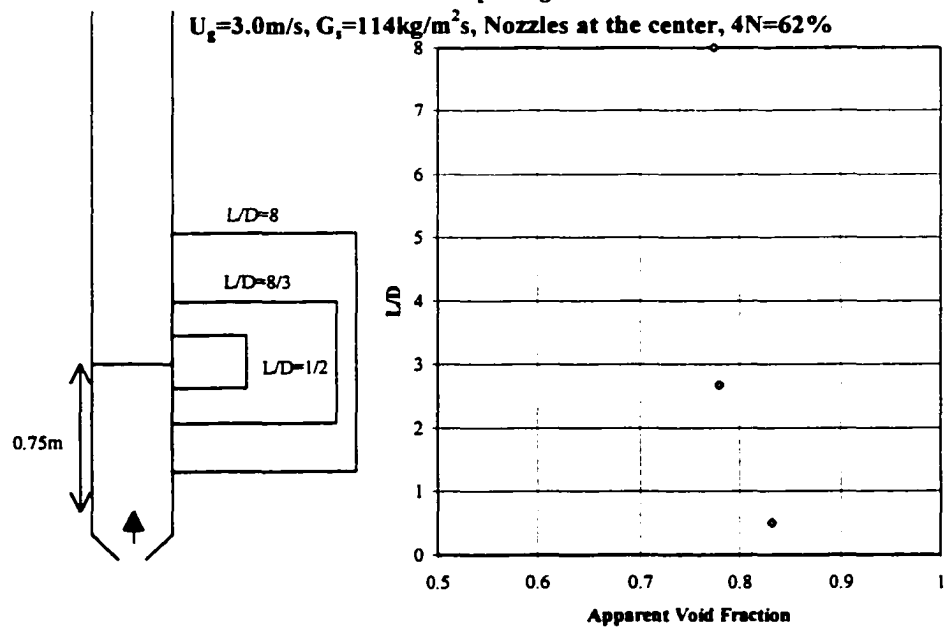
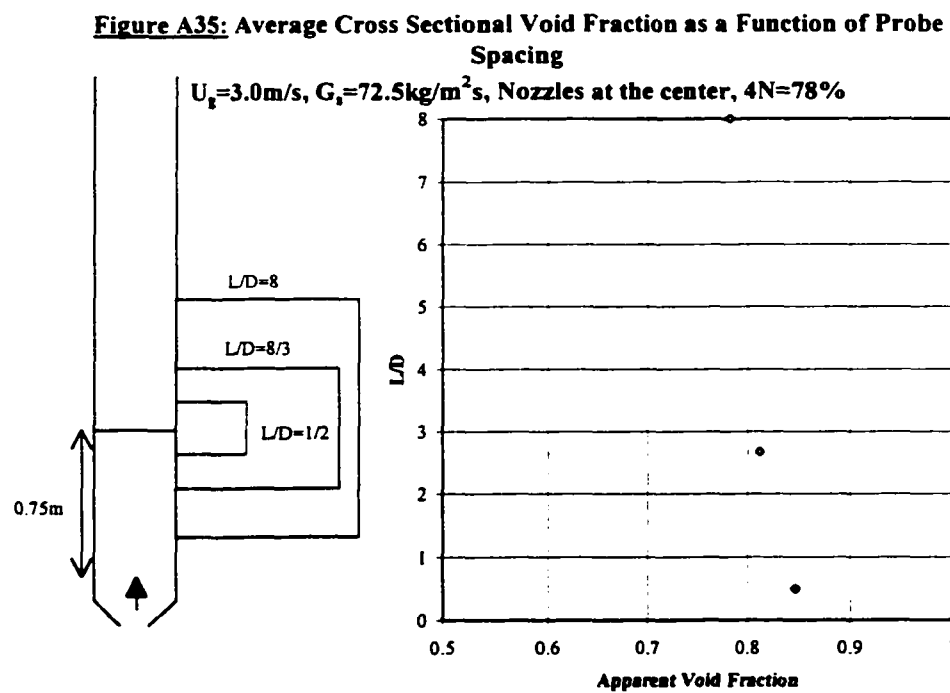
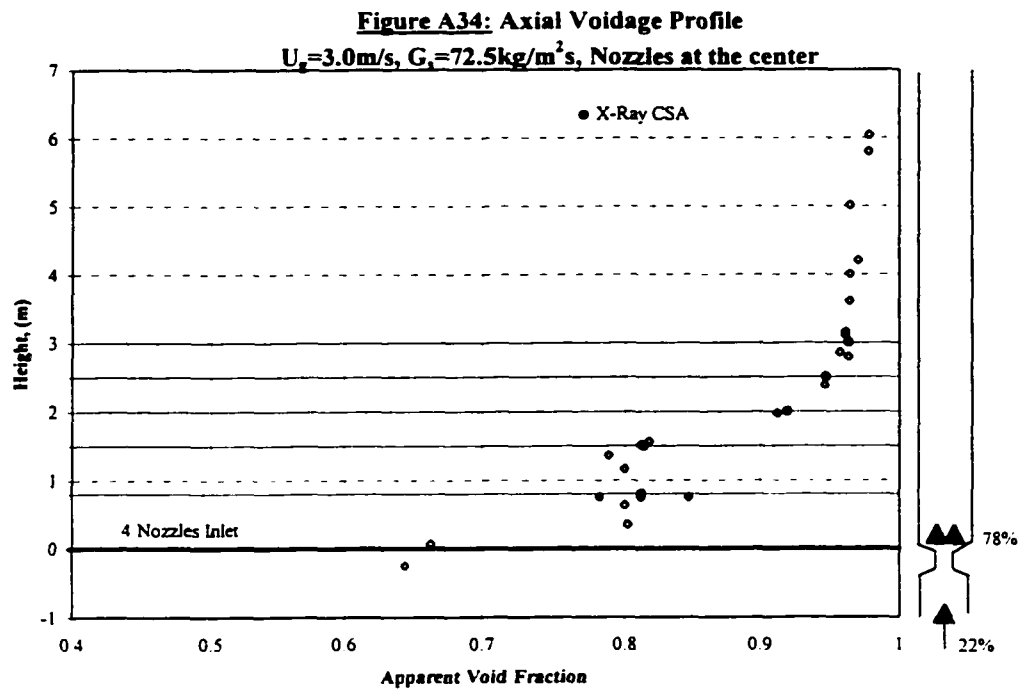
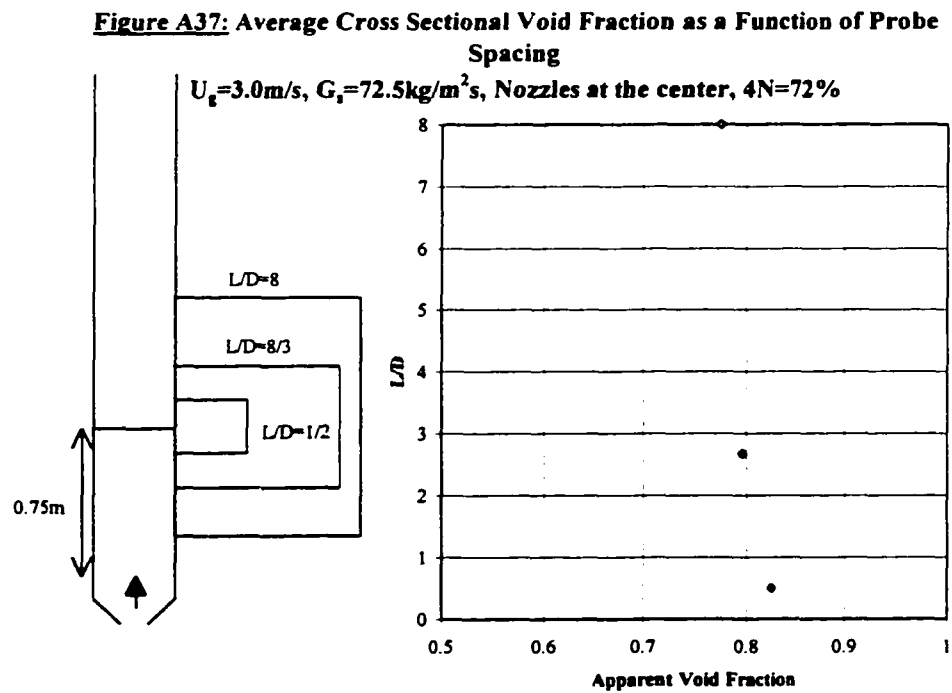
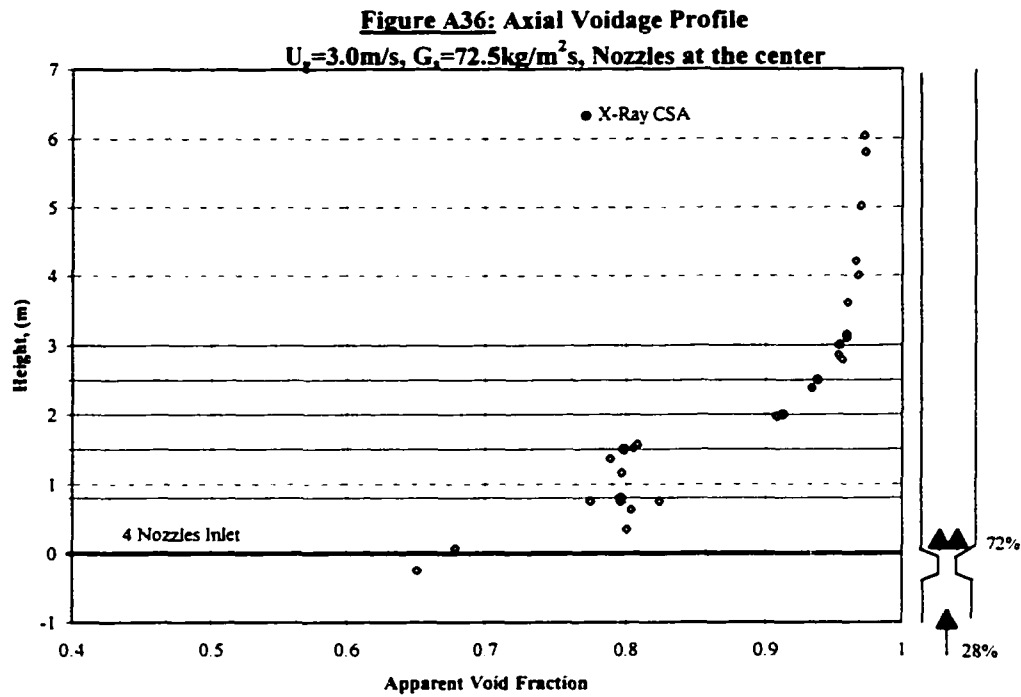


Figure A33: Average Cross Sectional Void Fraction as a Function of Probe Spacing







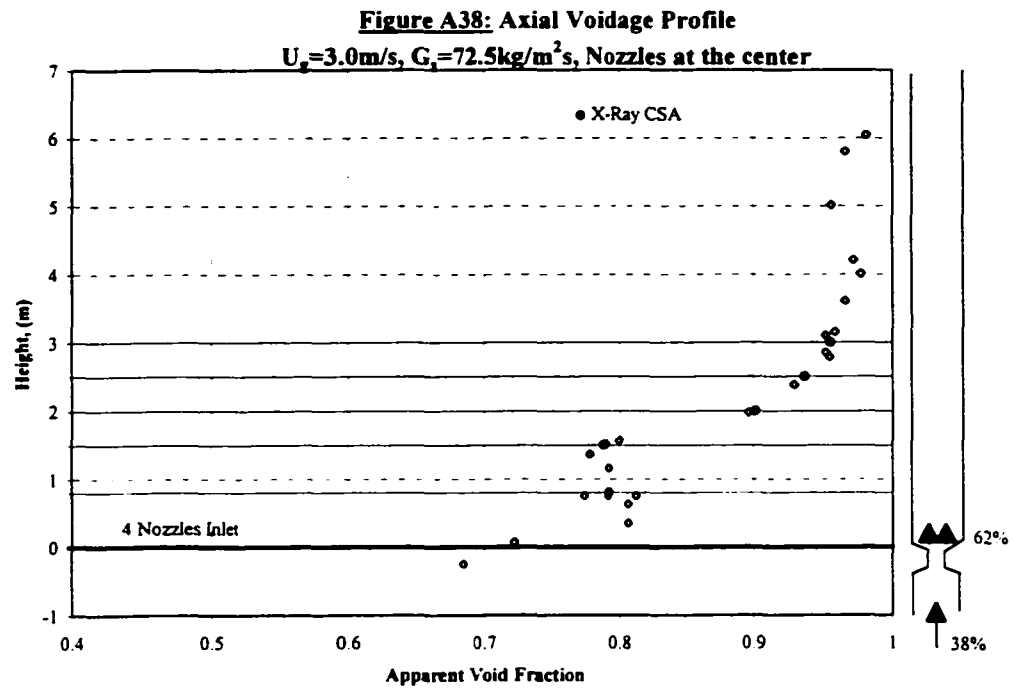
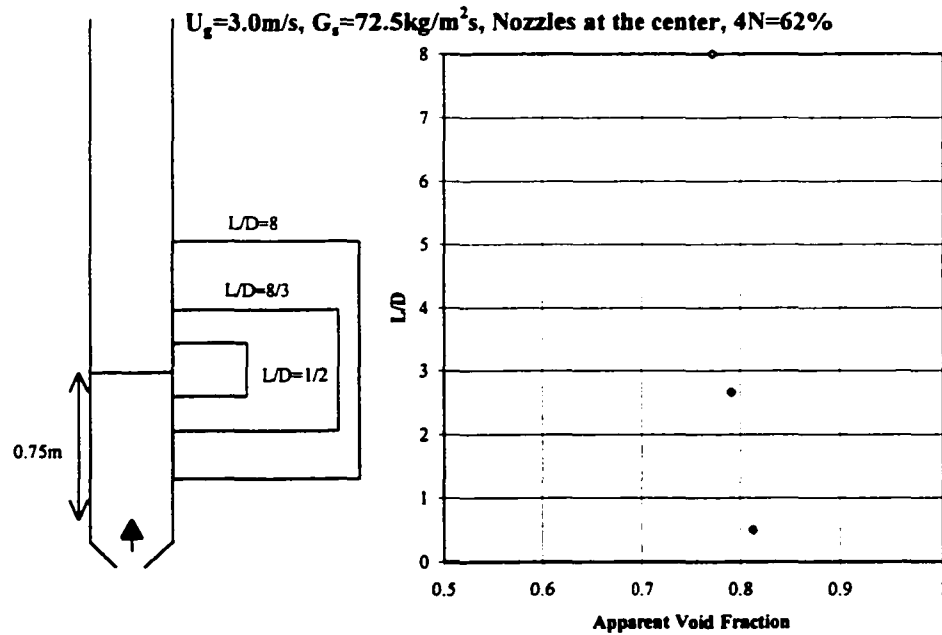


Figure A39: Average Cross Sectional Void Fraction as a Function of Probe Spacing



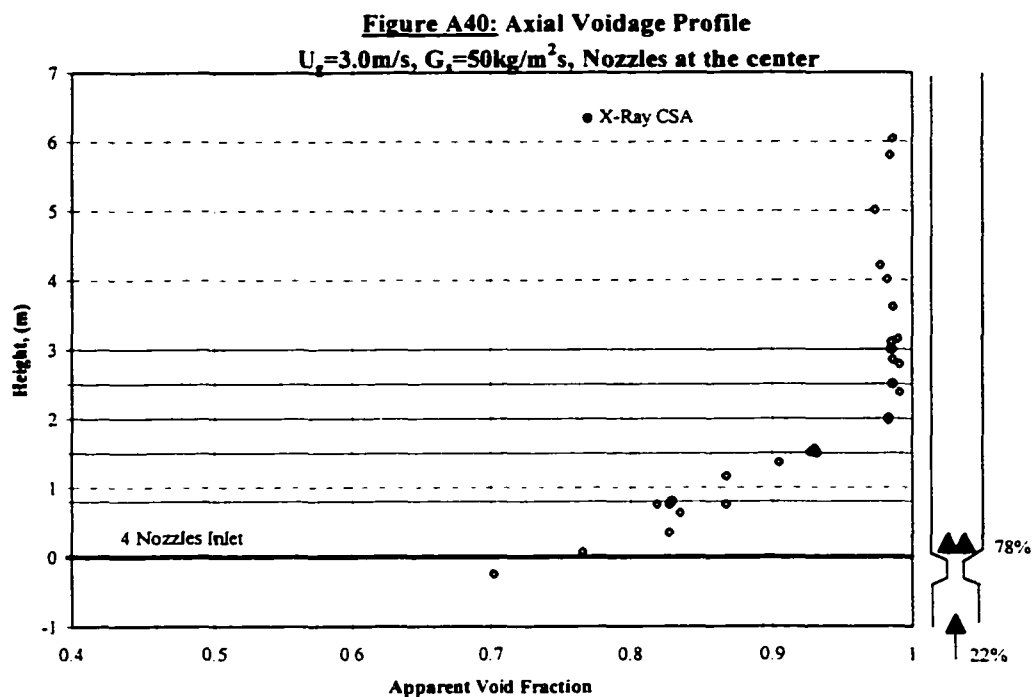
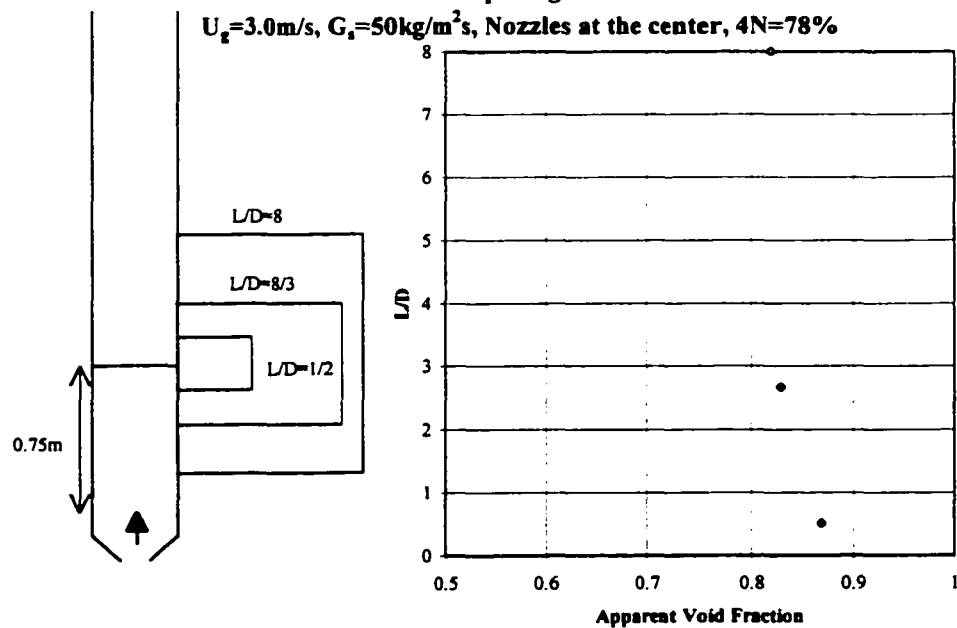
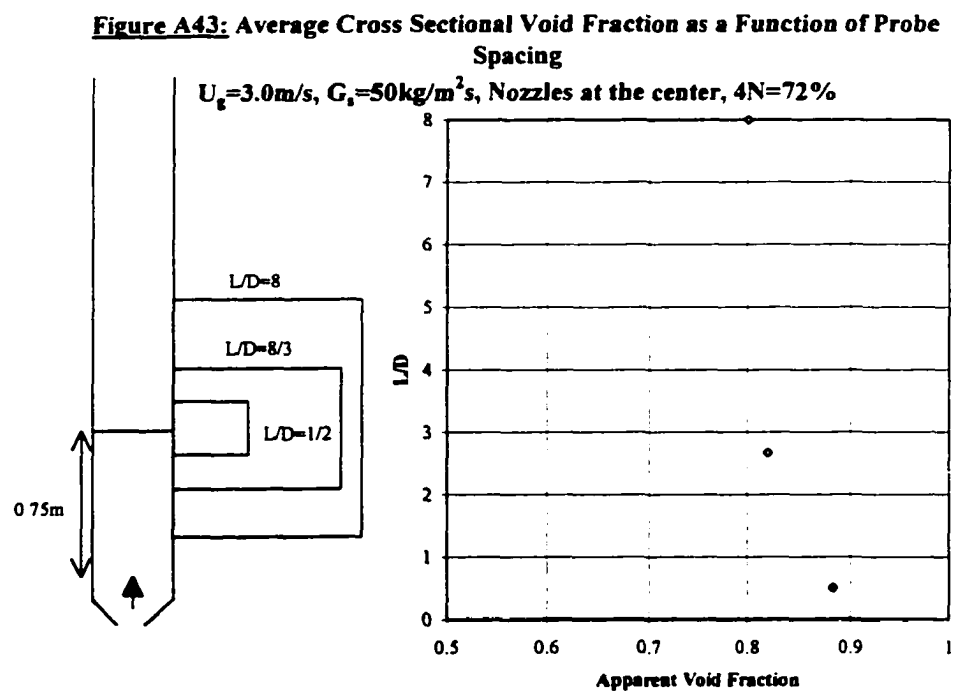
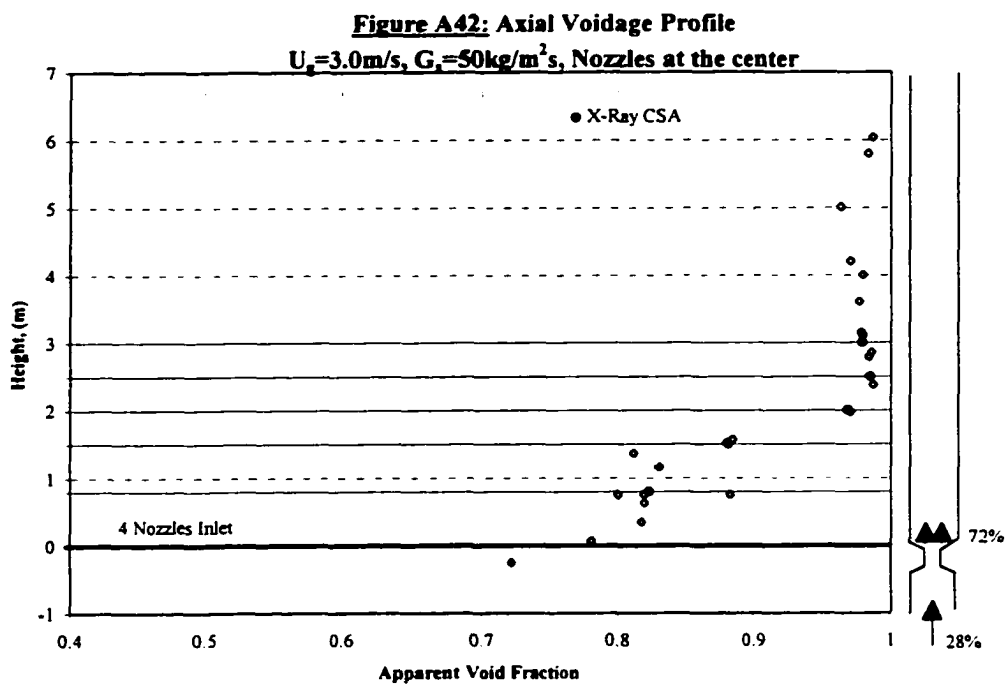


Figure A41: Average Cross Sectional Void Fraction as a Function of Probe Spacing





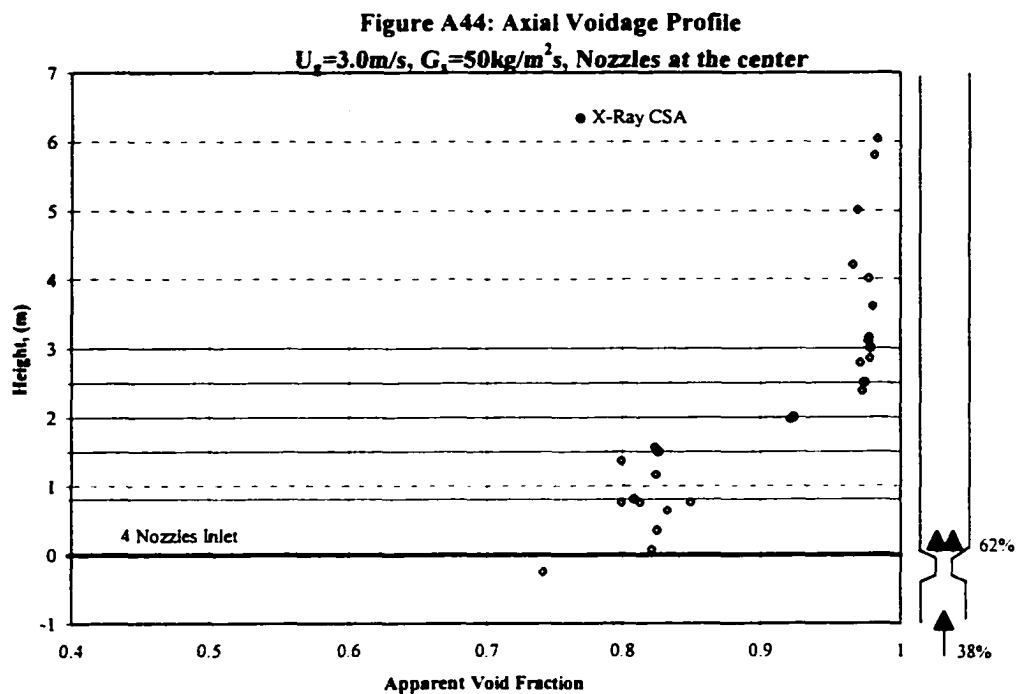


Figure A45: Average Cross Sectional Void Fraction as a Function of Probe Spacing

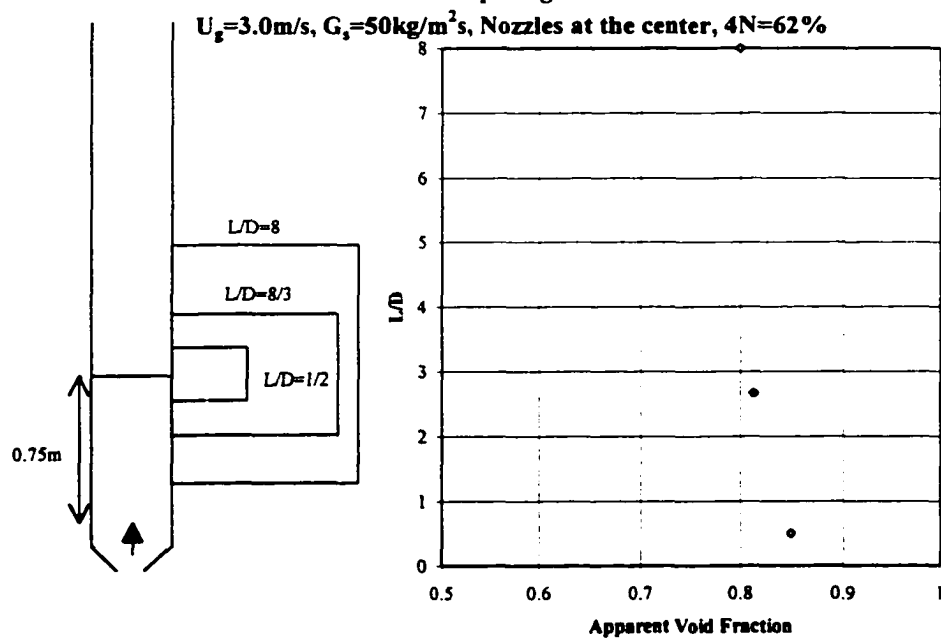


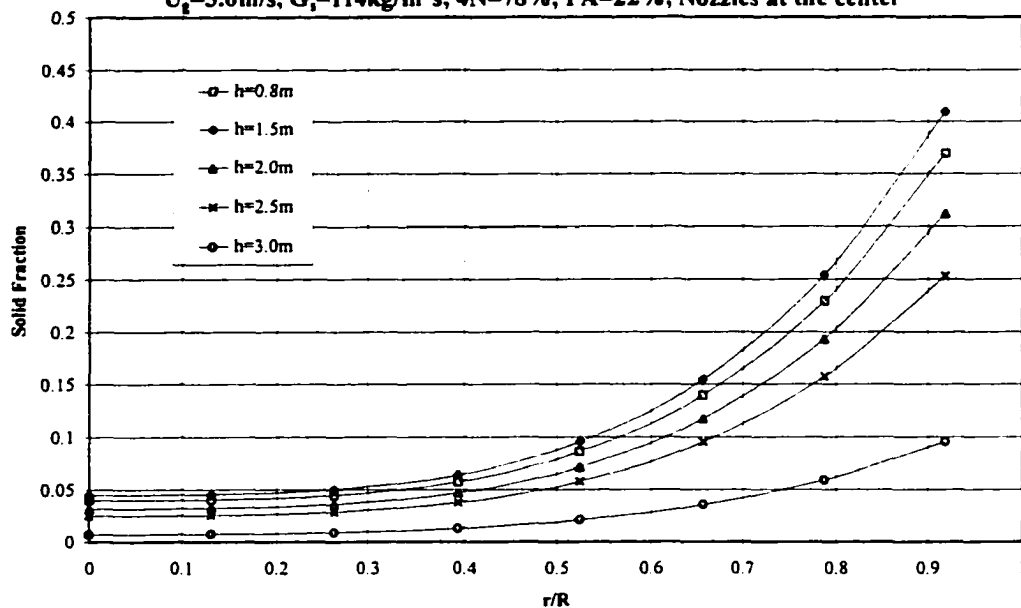
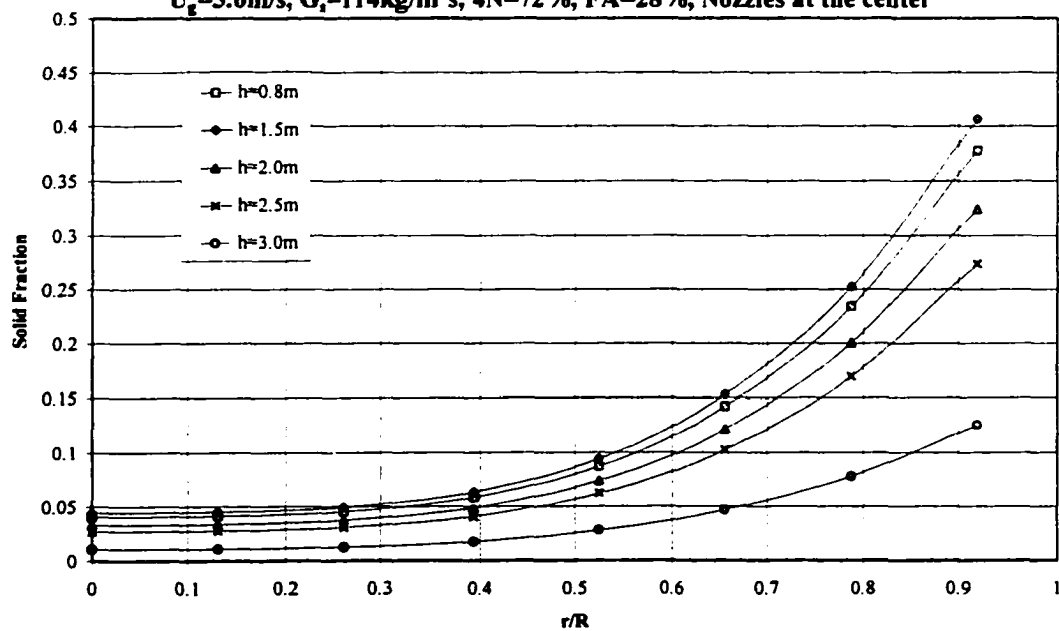
Figure A46: Radial Solid Profile $U_t=3.0\text{m/s}$, $G_t=114\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center**Figure A47: Radial Solid Profile** $U_t=3.0\text{m/s}$, $G_t=114\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center

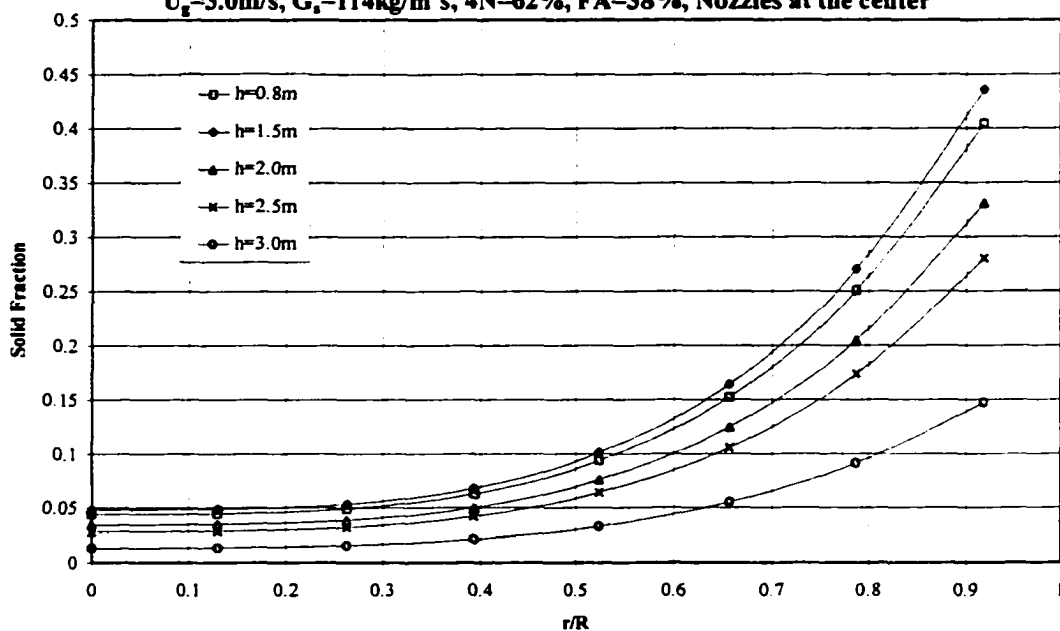
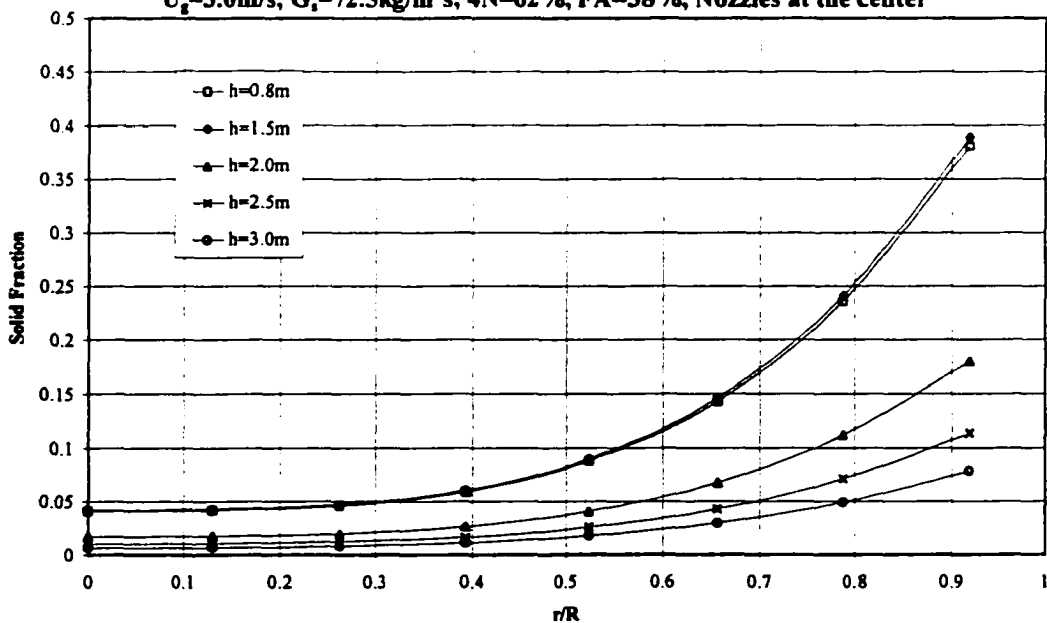
Figure A48: Radial Solid Profile $U_g=3.0\text{m/s}$, $G_s=114\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center**Figure A49: Radial Solid Profile** $U_g=3.0\text{m/s}$, $G_s=72.5\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center

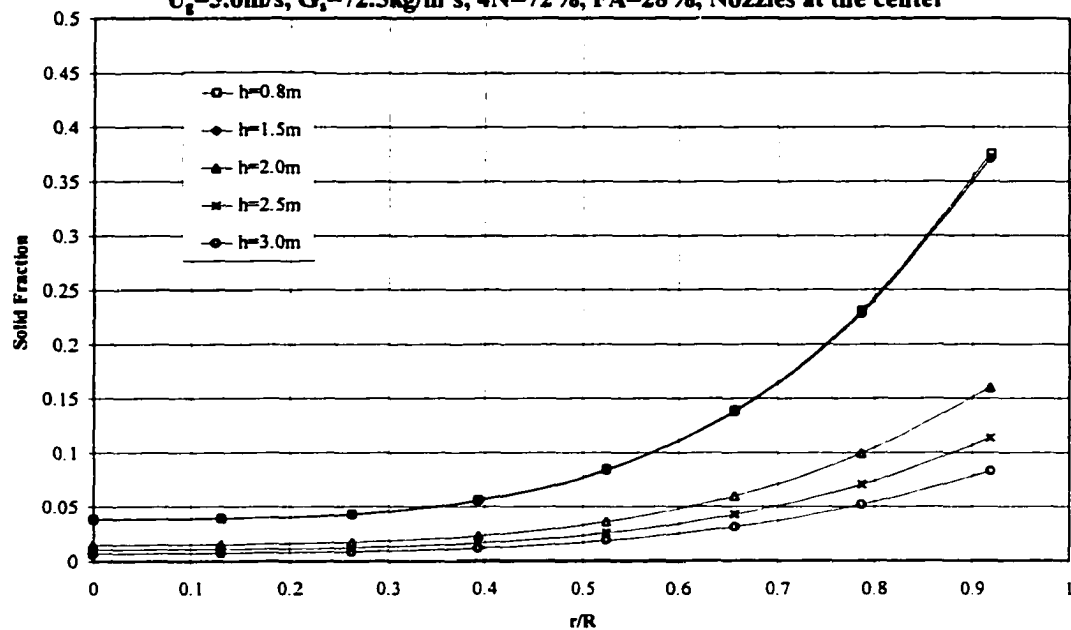
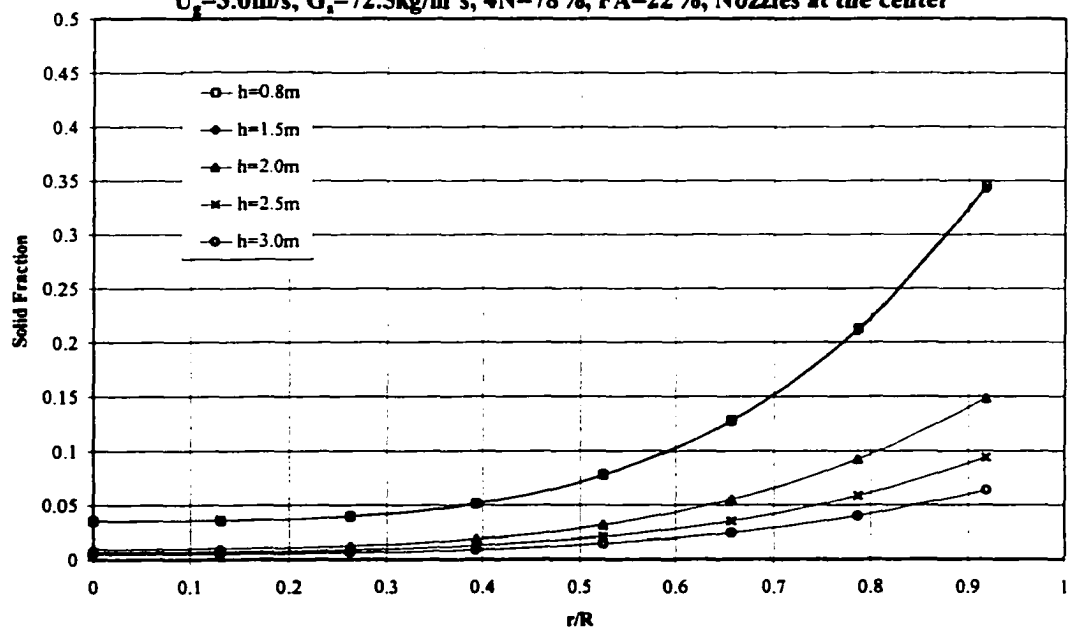
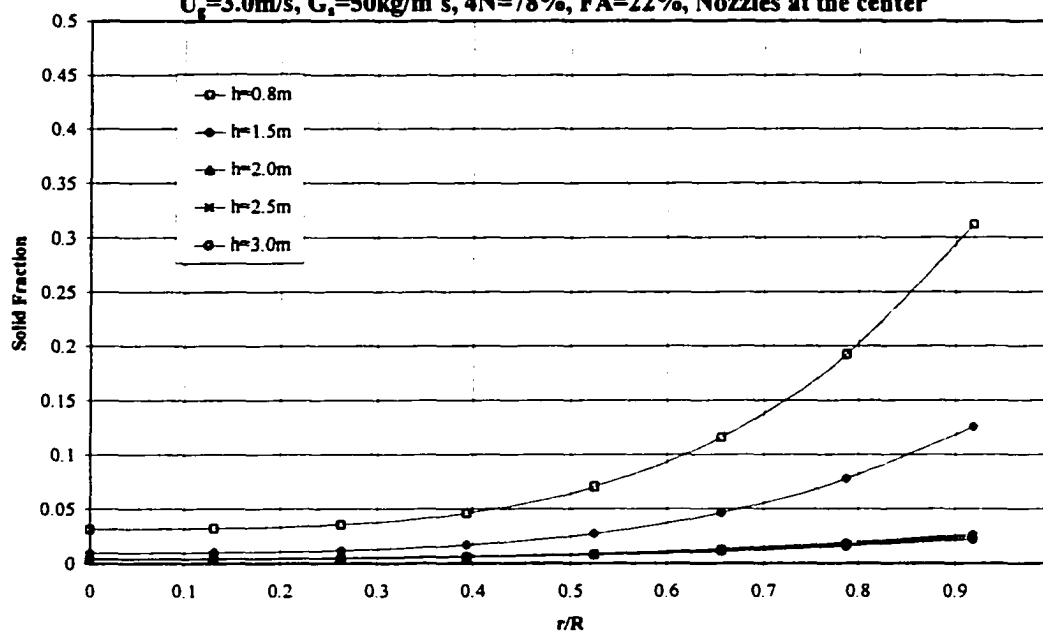
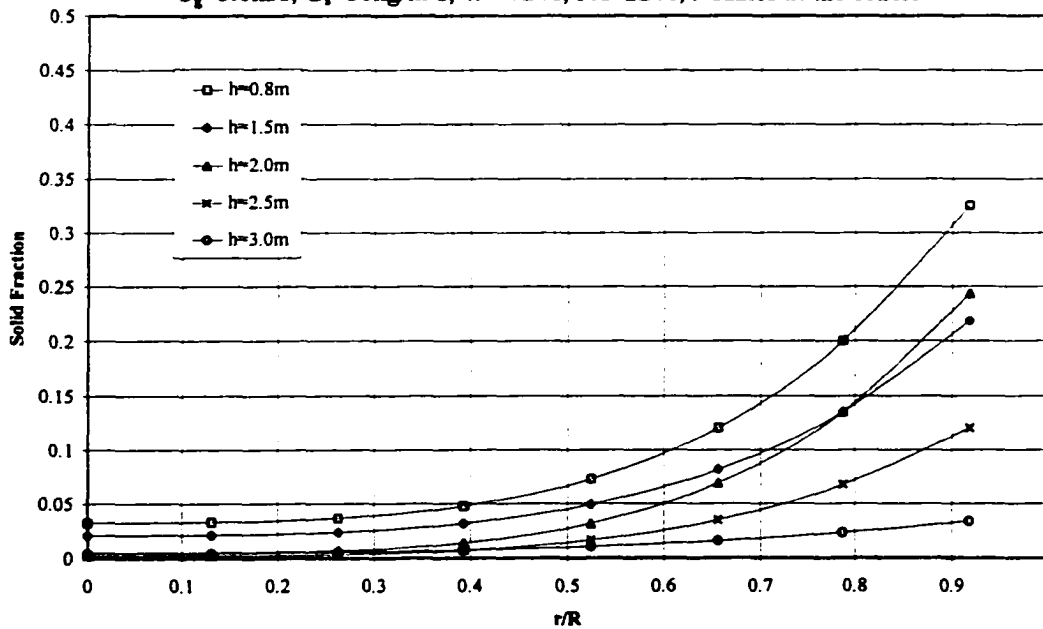
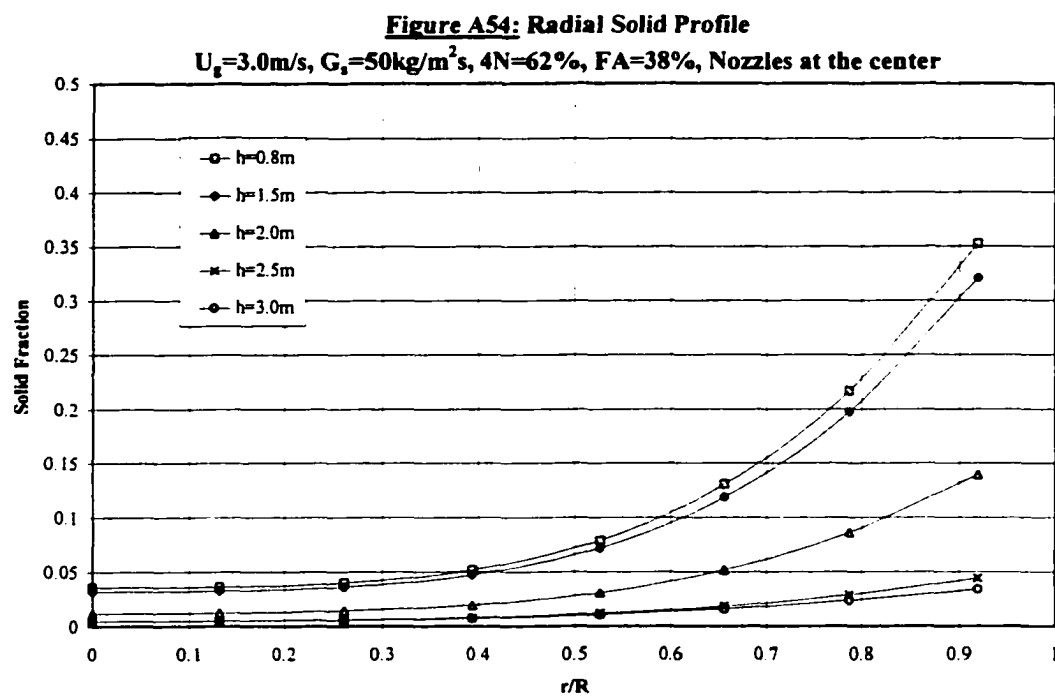
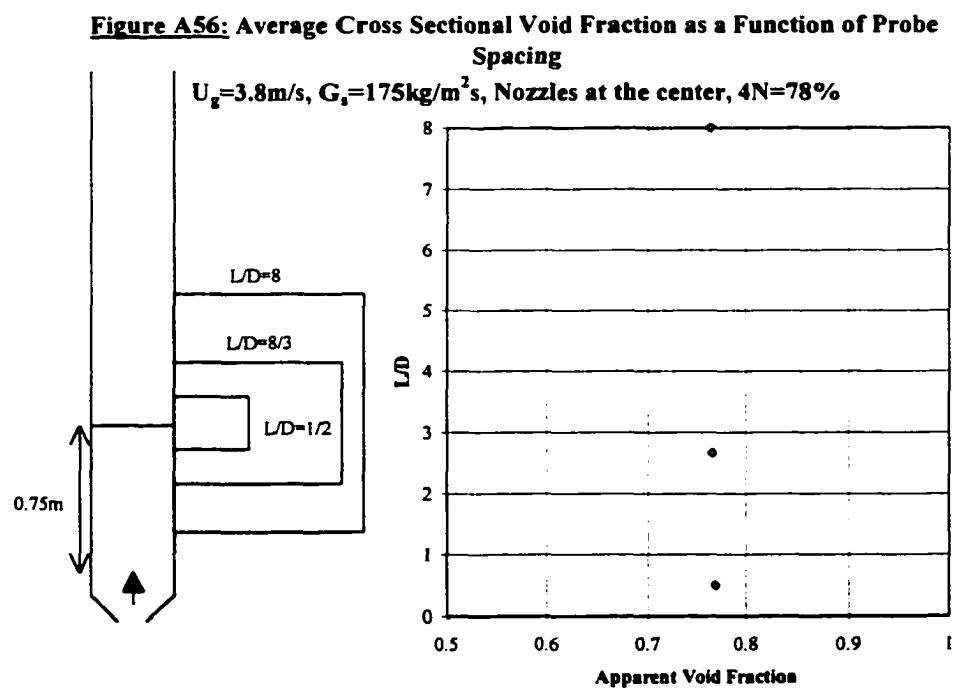
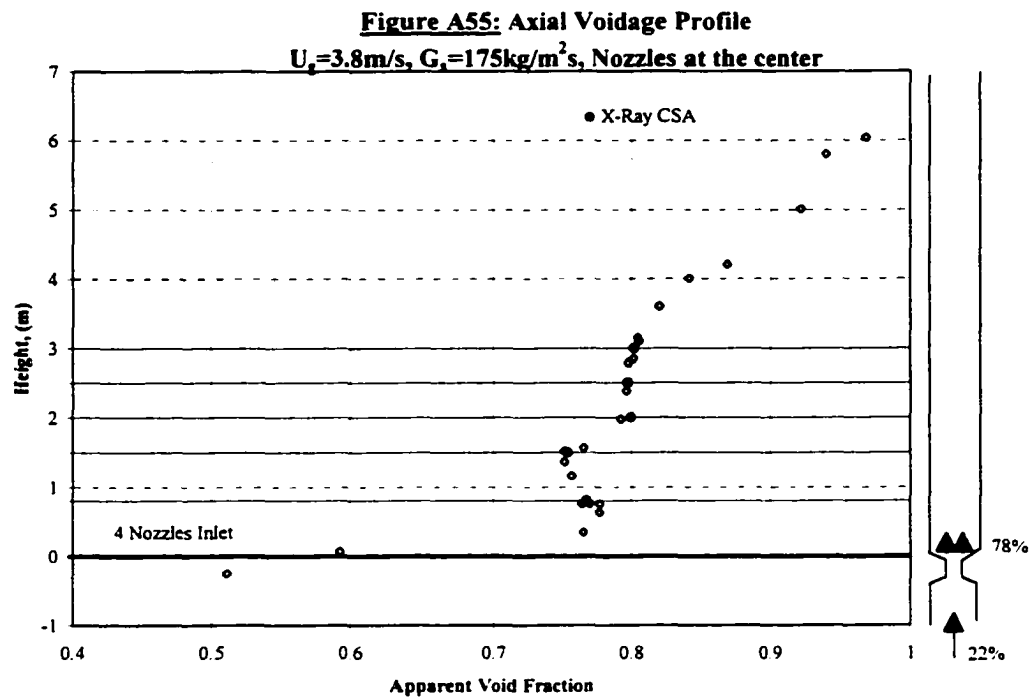
Figure A50: Radial Solid Profile $U_g=3.0\text{m/s}$, $G_s=72.5\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center**Figure A51: Radial Solid Profile** $U_g=3.0\text{m/s}$, $G_s=72.5\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center

Figure A52: Radial Solid Profile $U_t=3.0\text{m/s}$, $G_t=50\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center**Figure A53: Radial Solid Profile** $U_t=3.0\text{m/s}$, $G_t=50\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=22\%$, Nozzles at the center





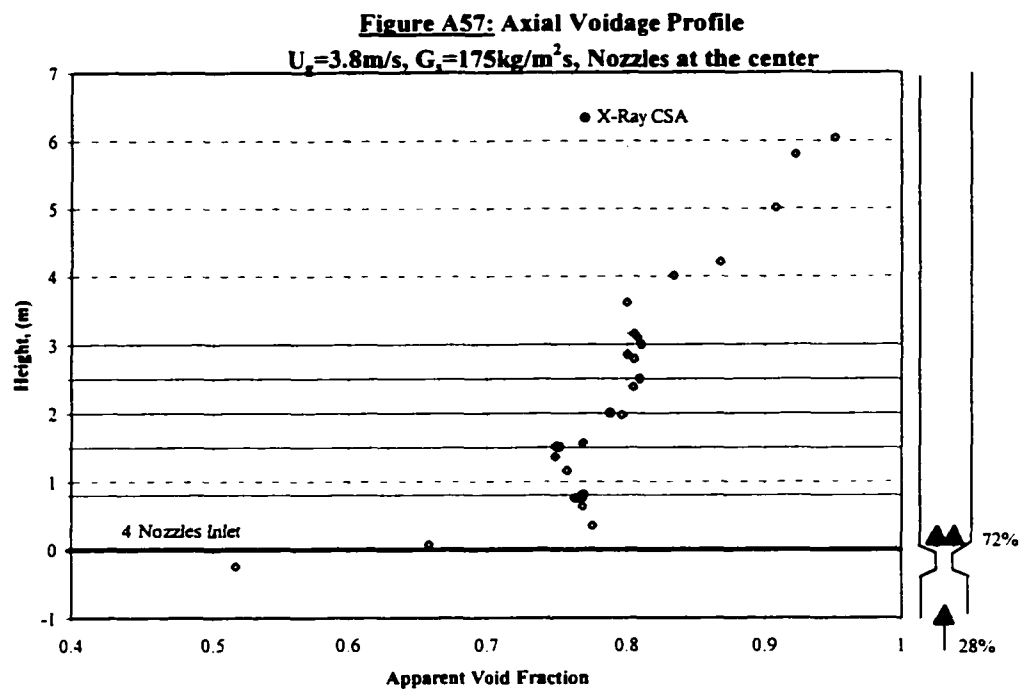
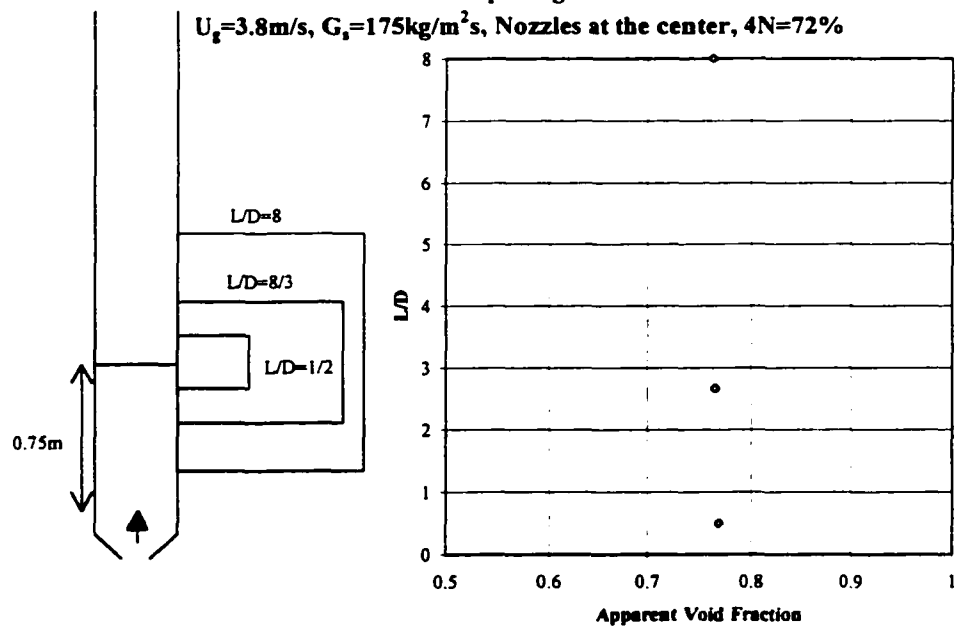
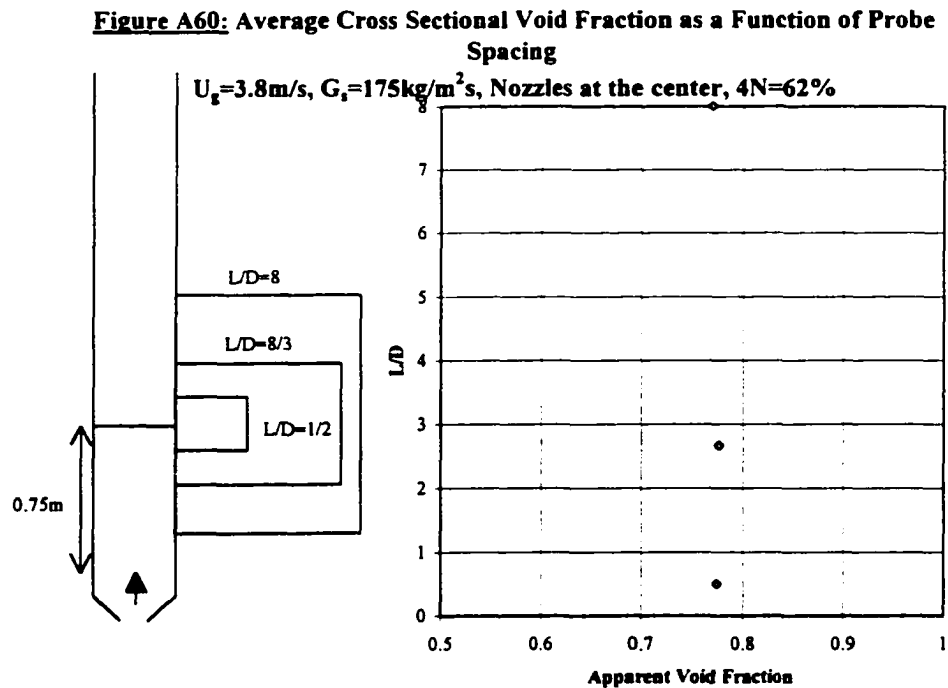
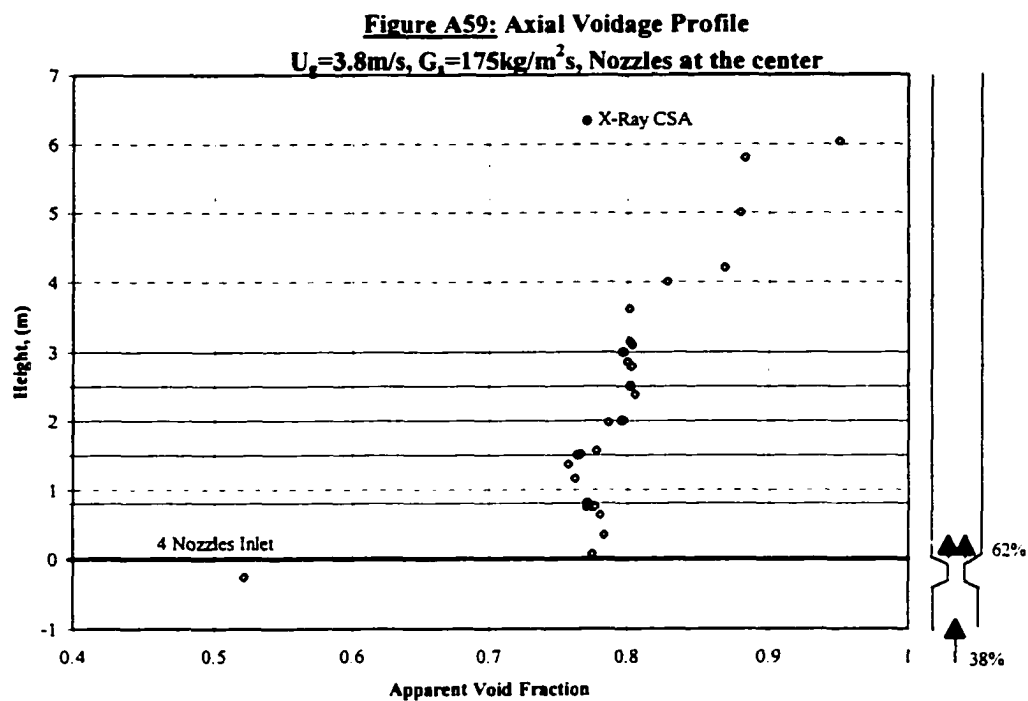
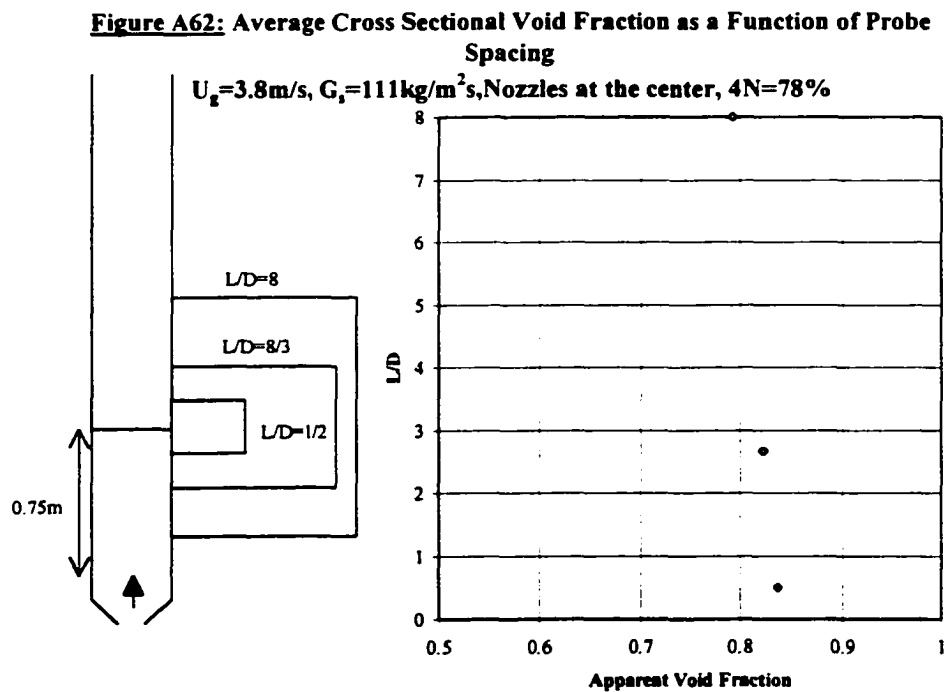
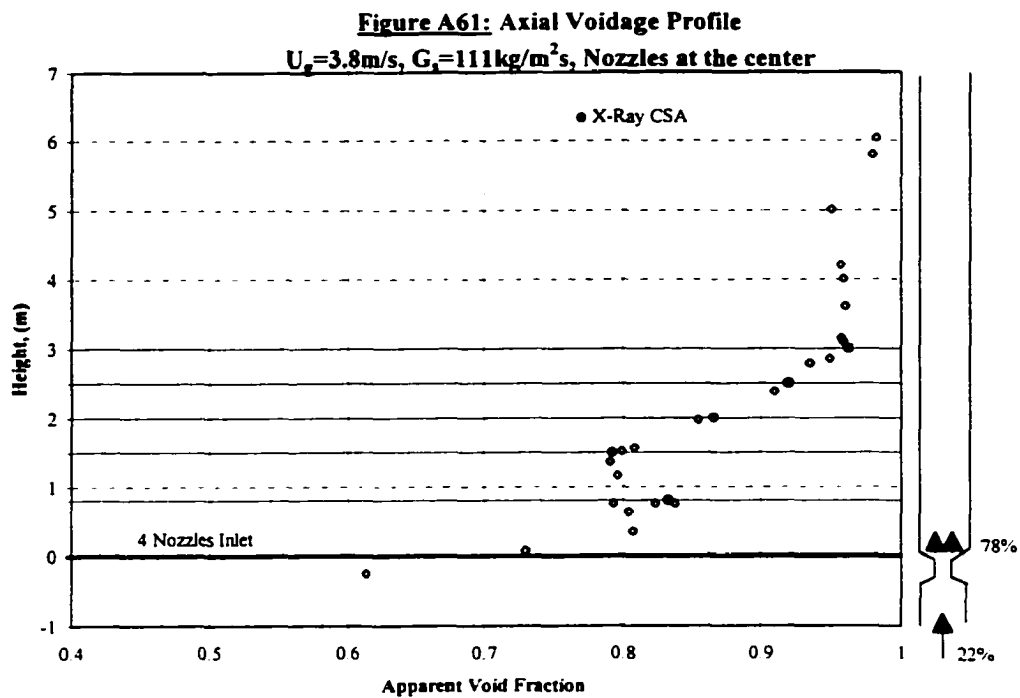
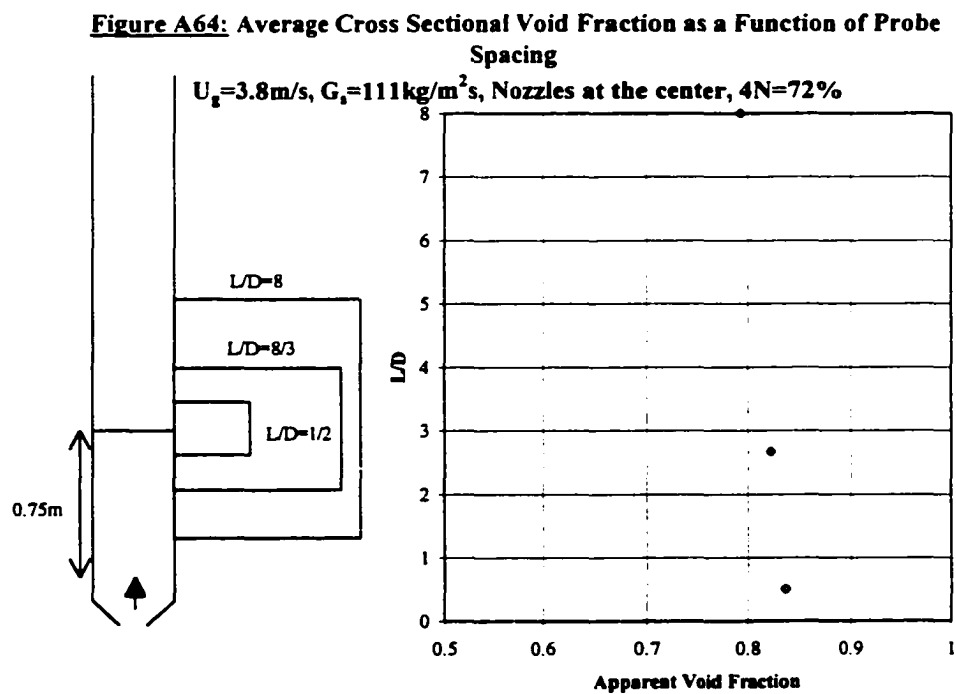
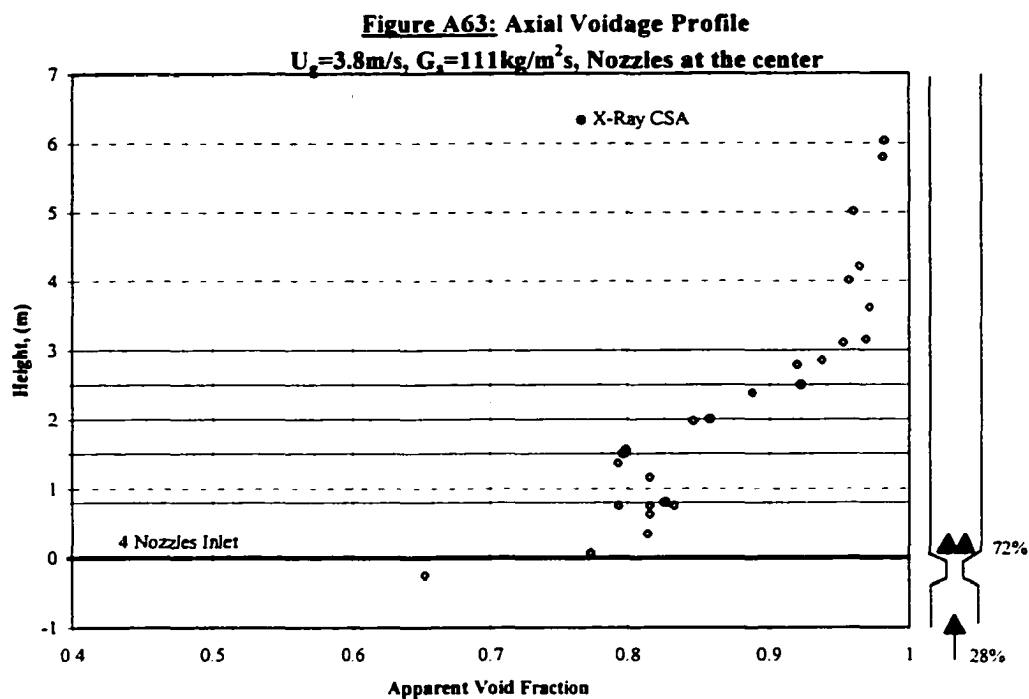


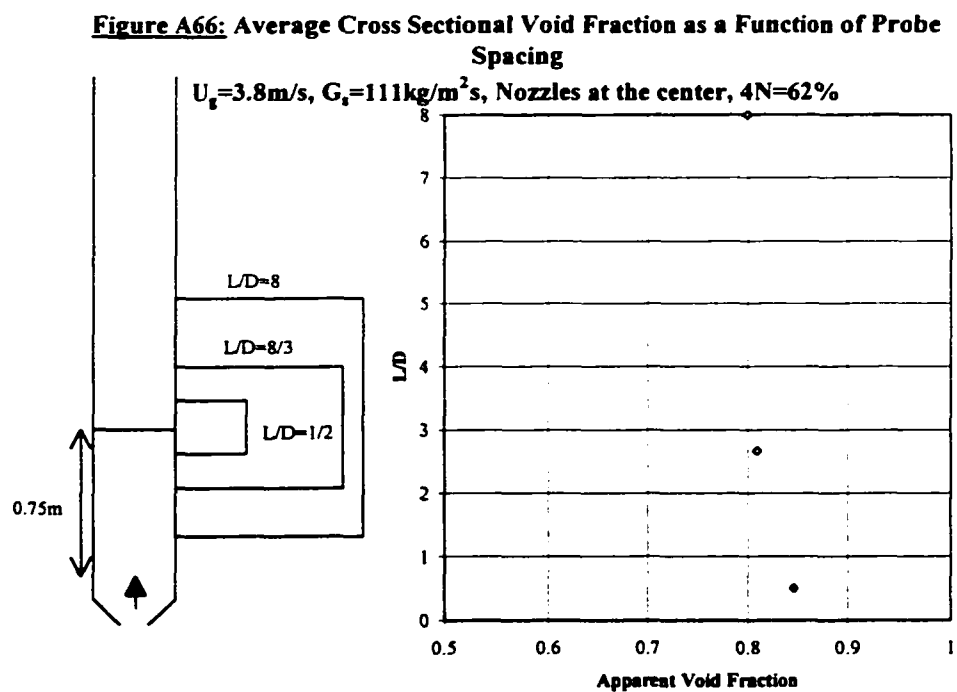
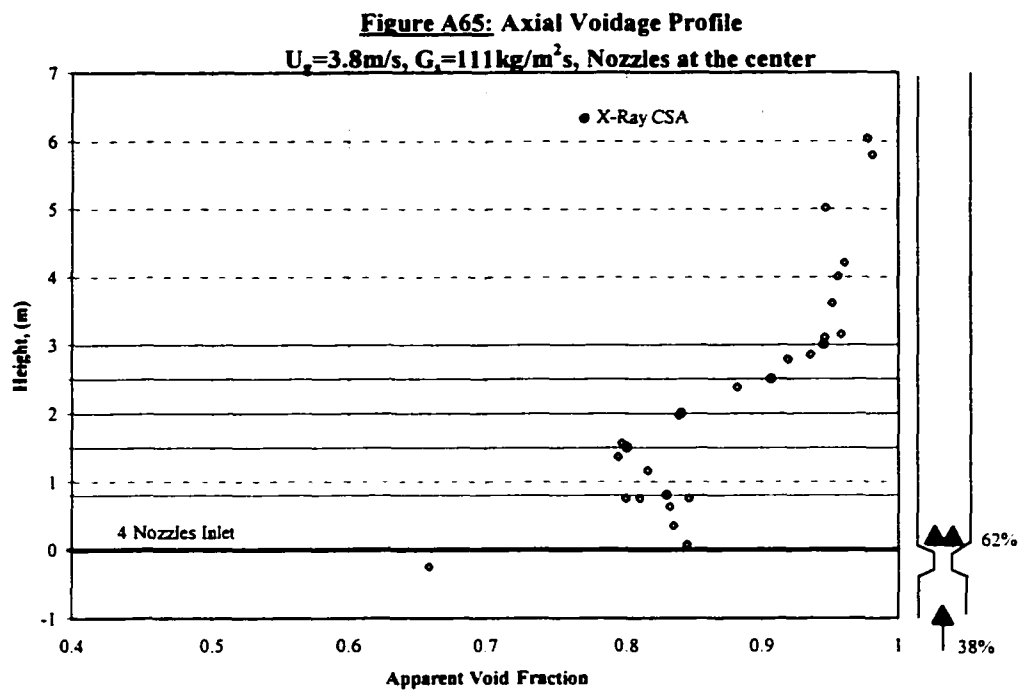
Figure A58: Average Cross Sectional Void Fraction as a Function of Probe Spacing

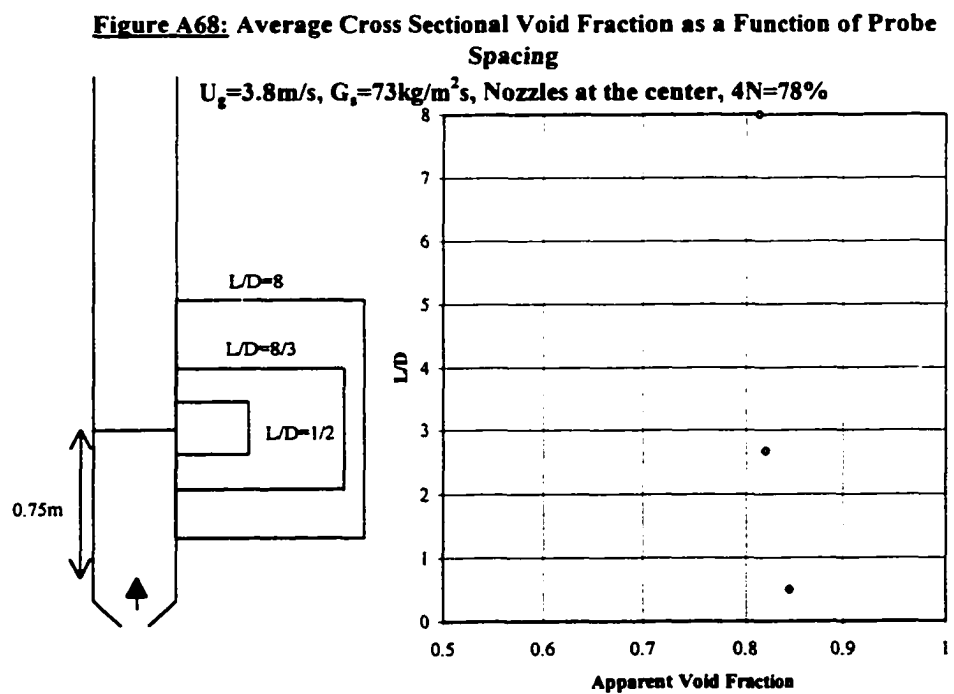
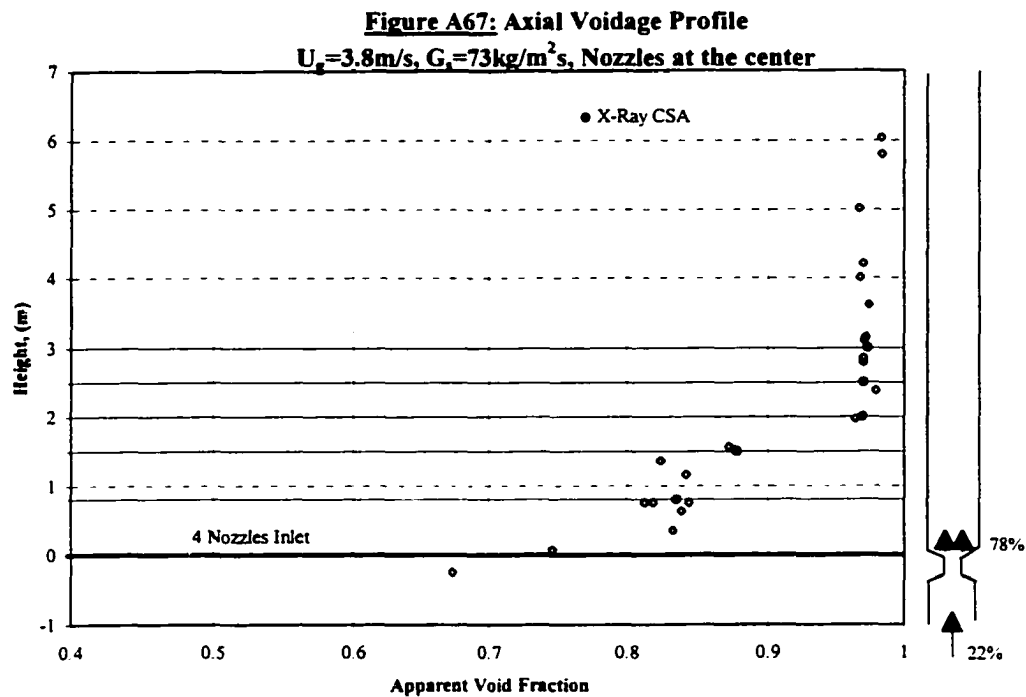


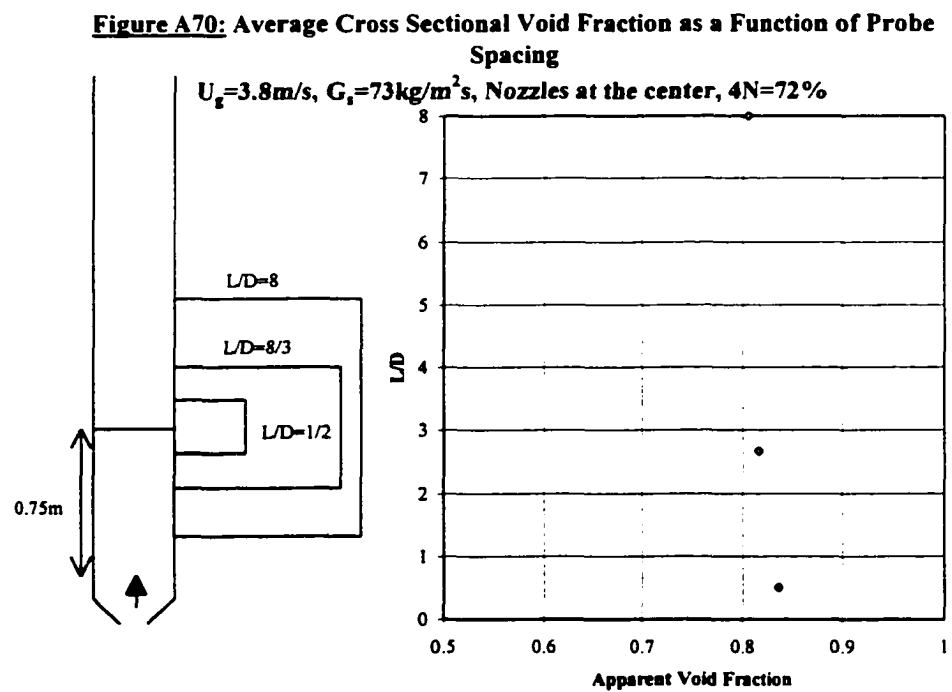
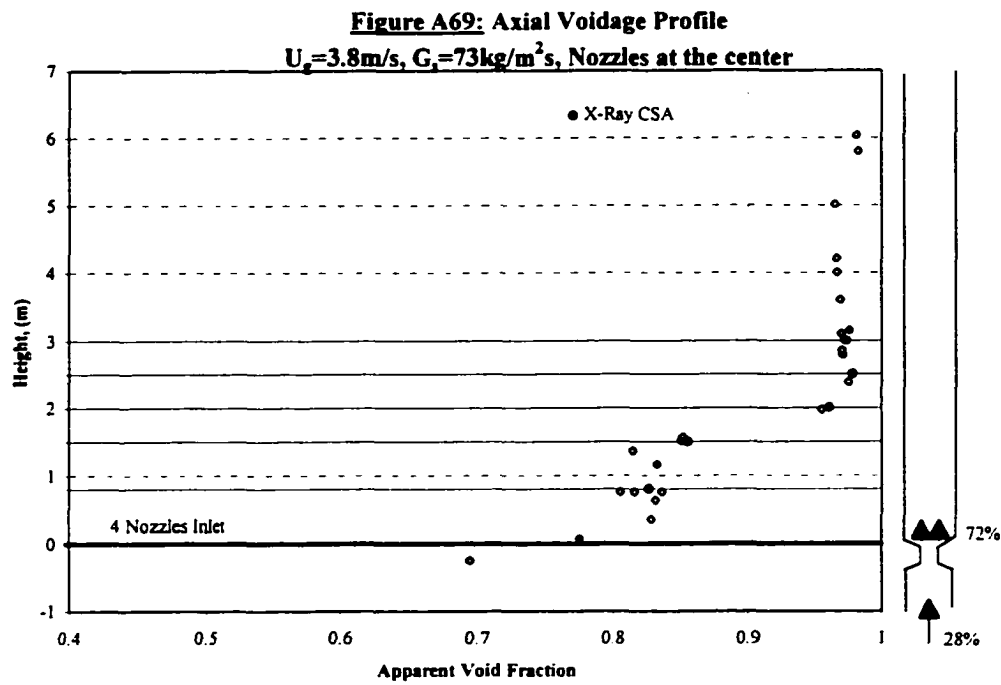












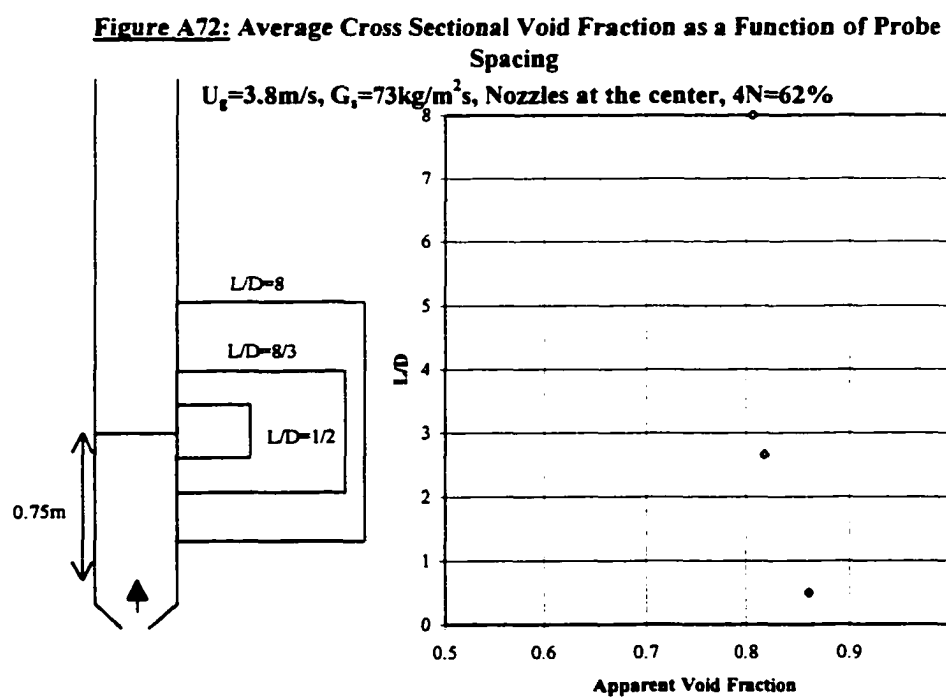
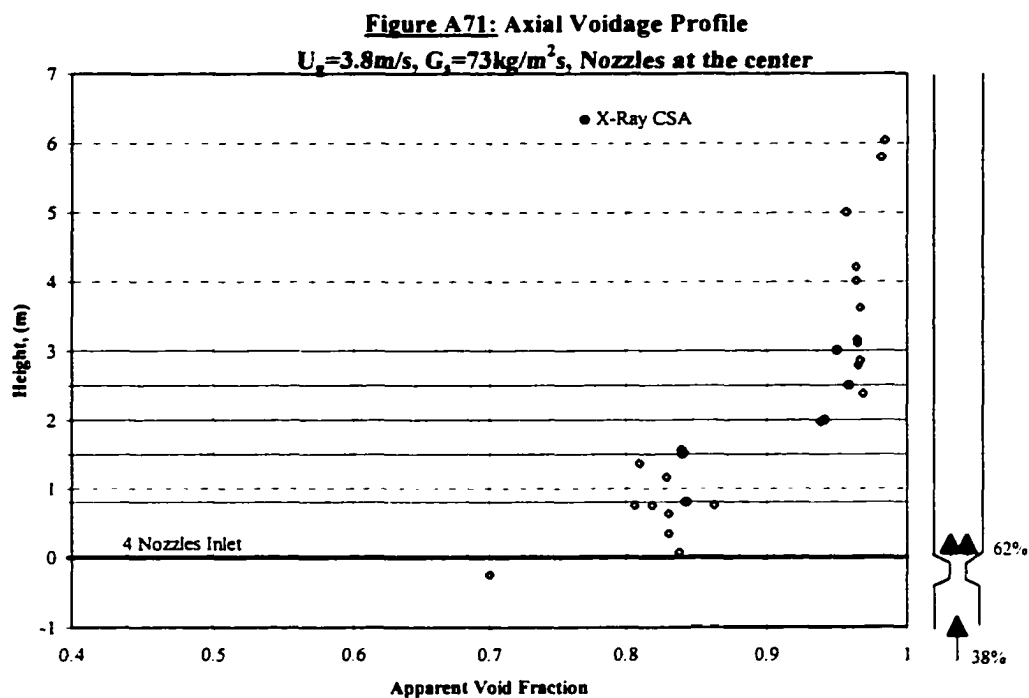


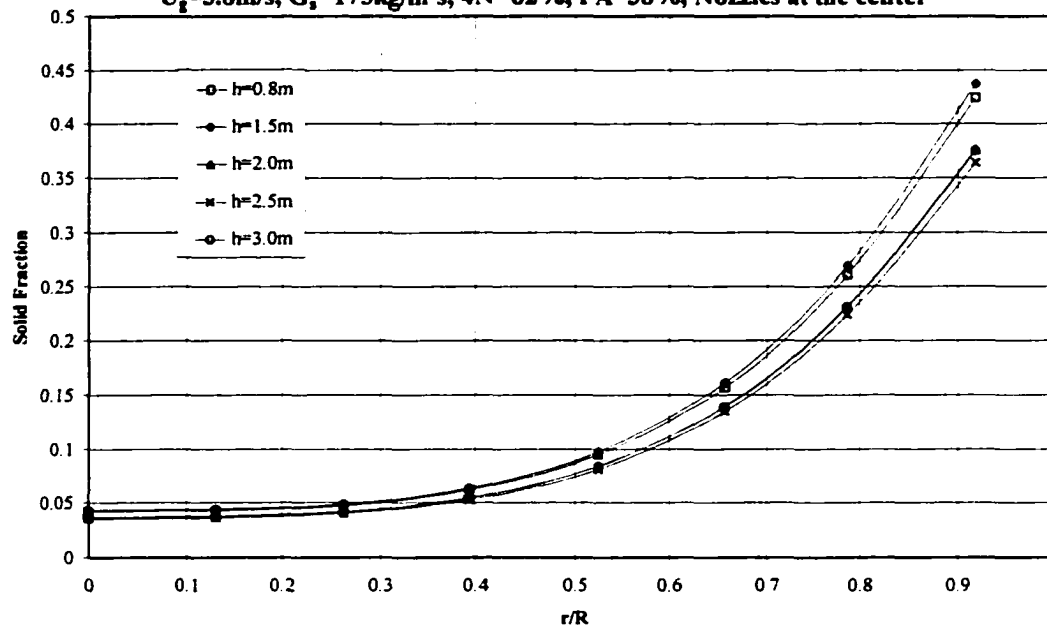
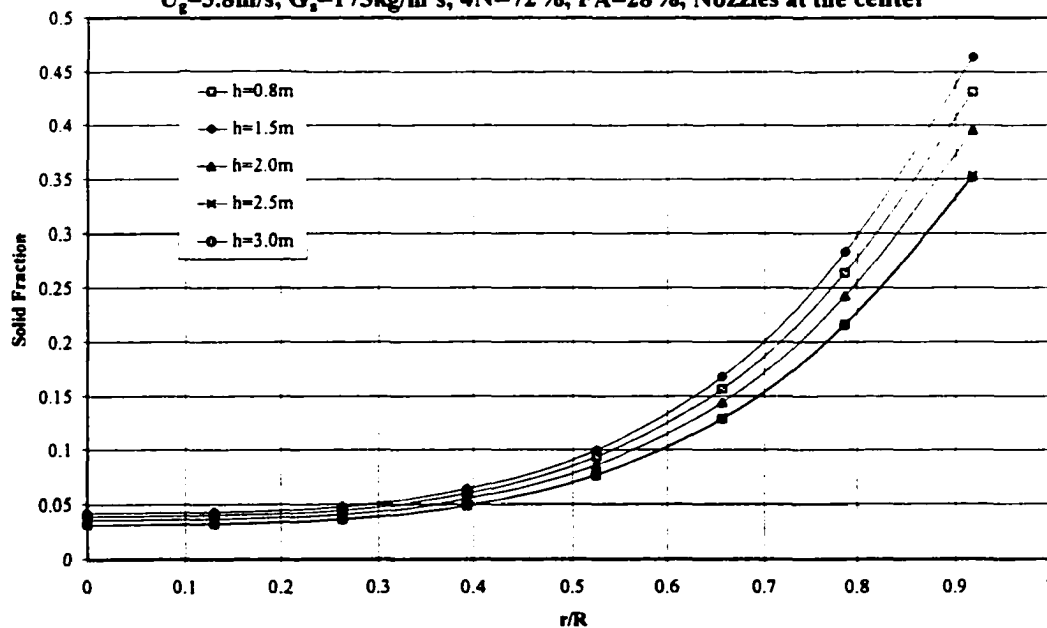
Figure A73: Radial Solid Profile $U_g=3.8\text{m/s}$, $G_s=175\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center**Figure A74: Radial Solid Profile** $U_g=3.8\text{m/s}$, $G_s=175\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center

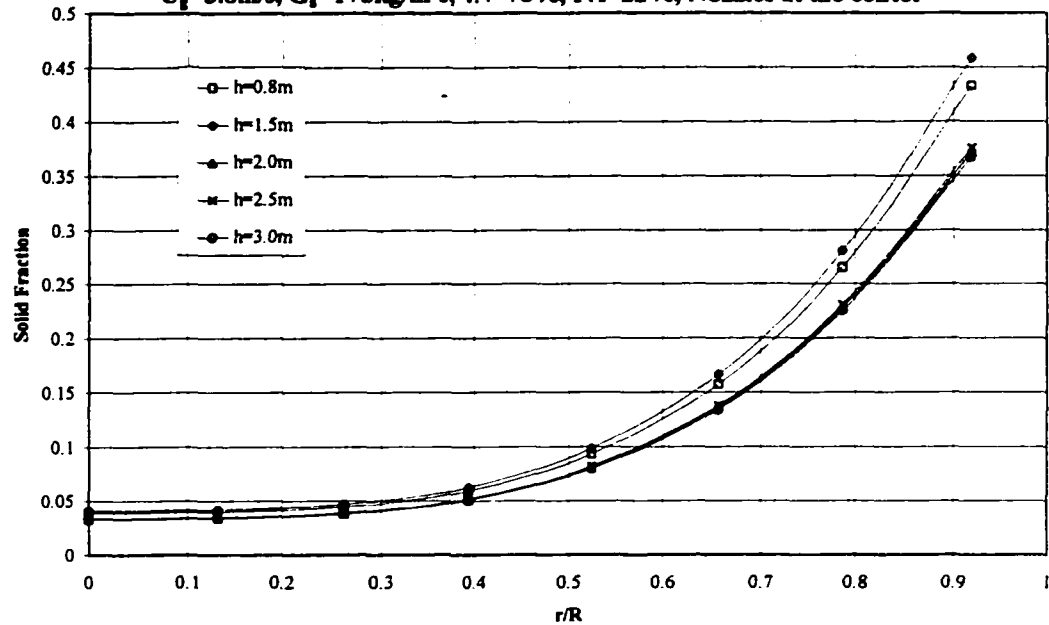
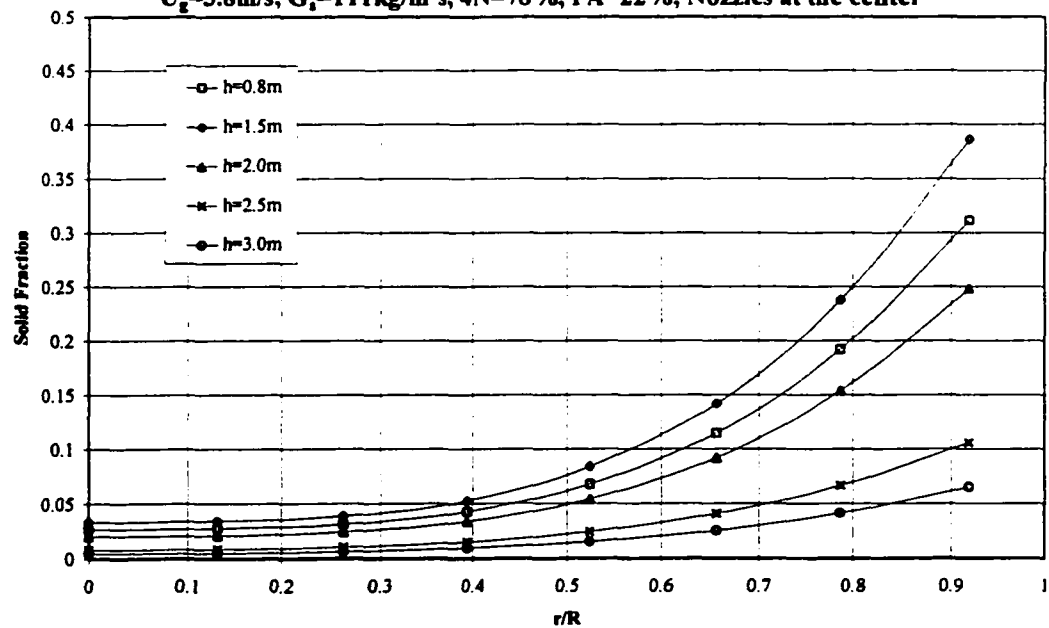
Figure A75: Radial Solid Profile $U_g=3.8\text{m/s}$, $G_s=175\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center**Figure A76: Radial Solid Density Profile** $U_g=3.8\text{m/s}$, $G_s=111\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center

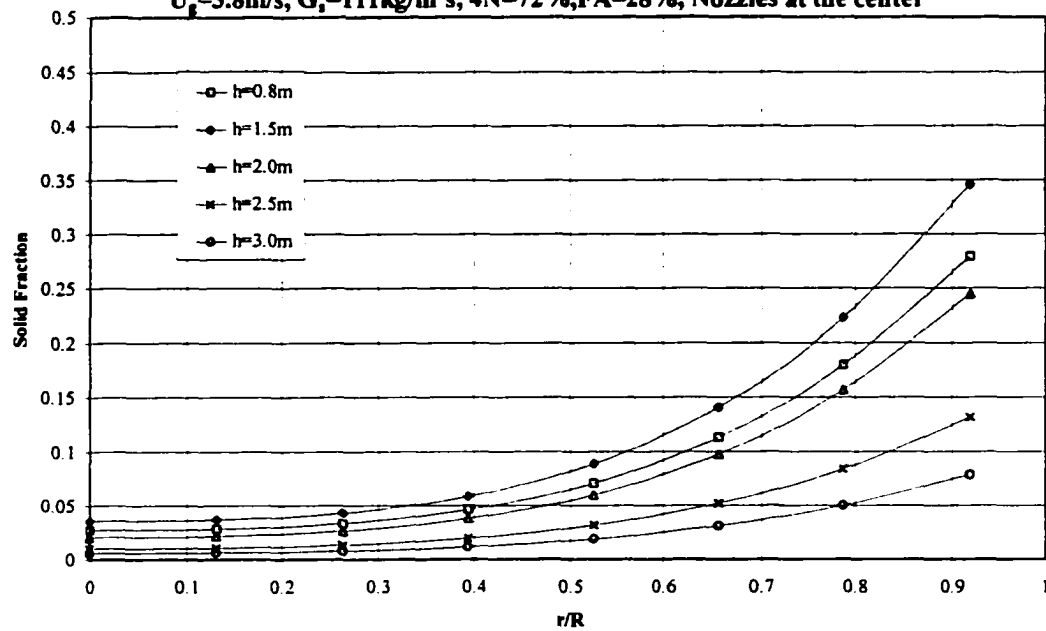
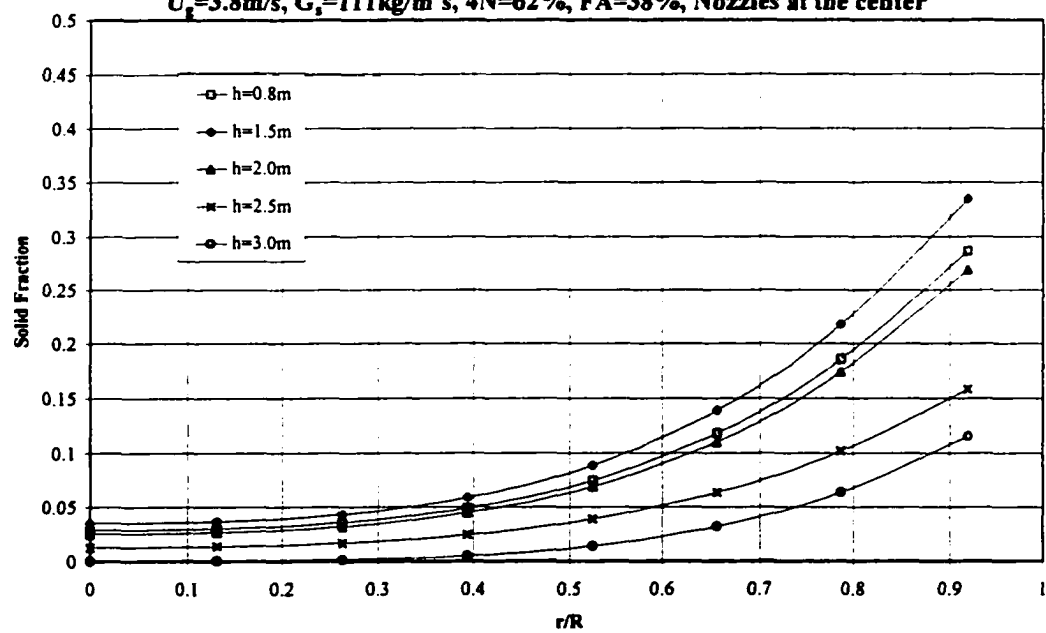
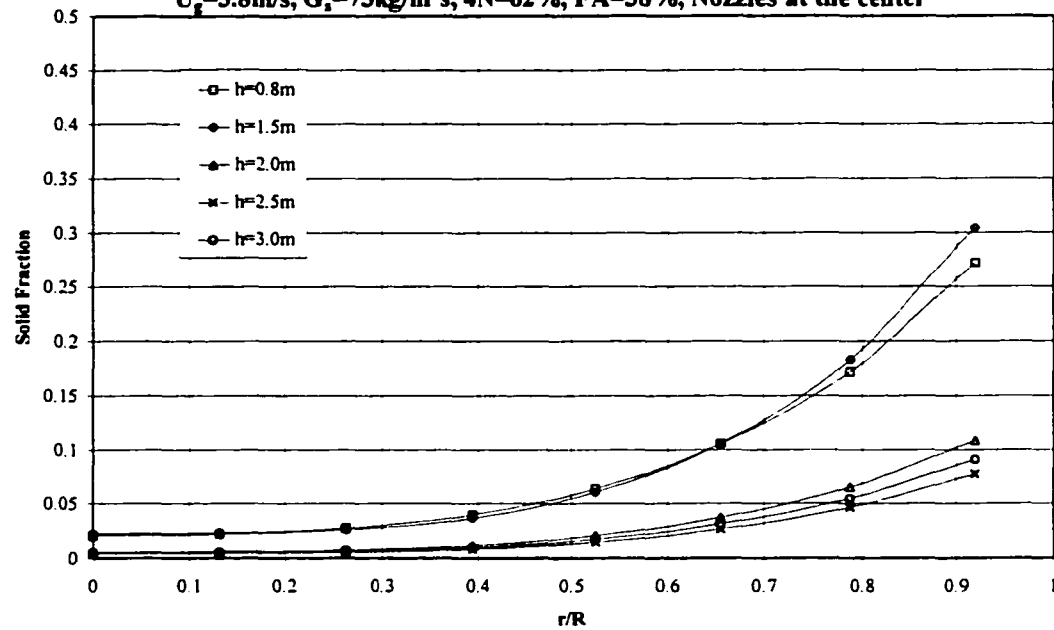
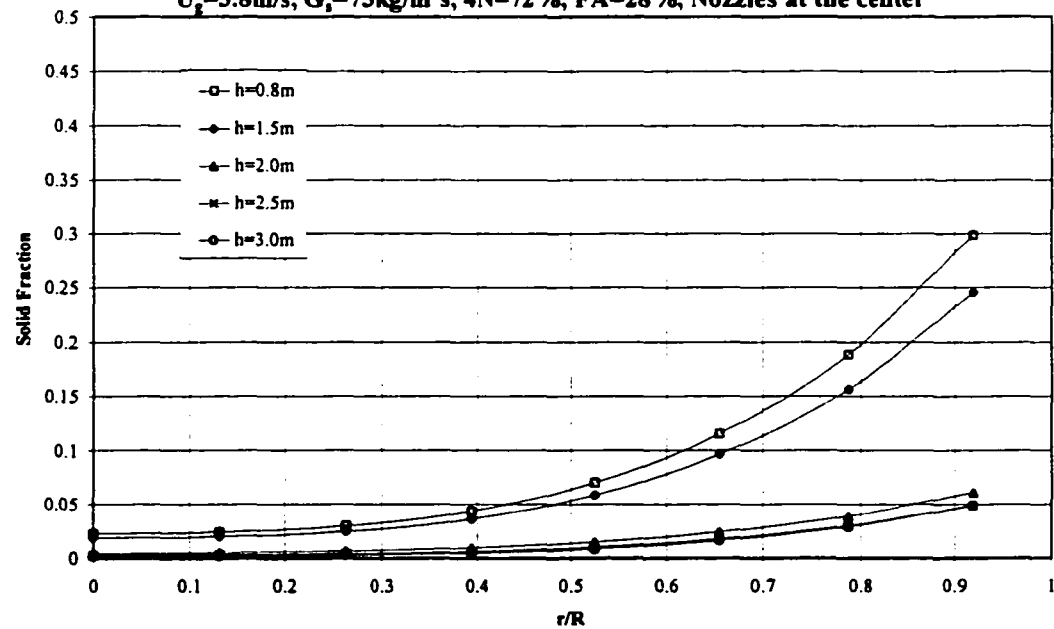
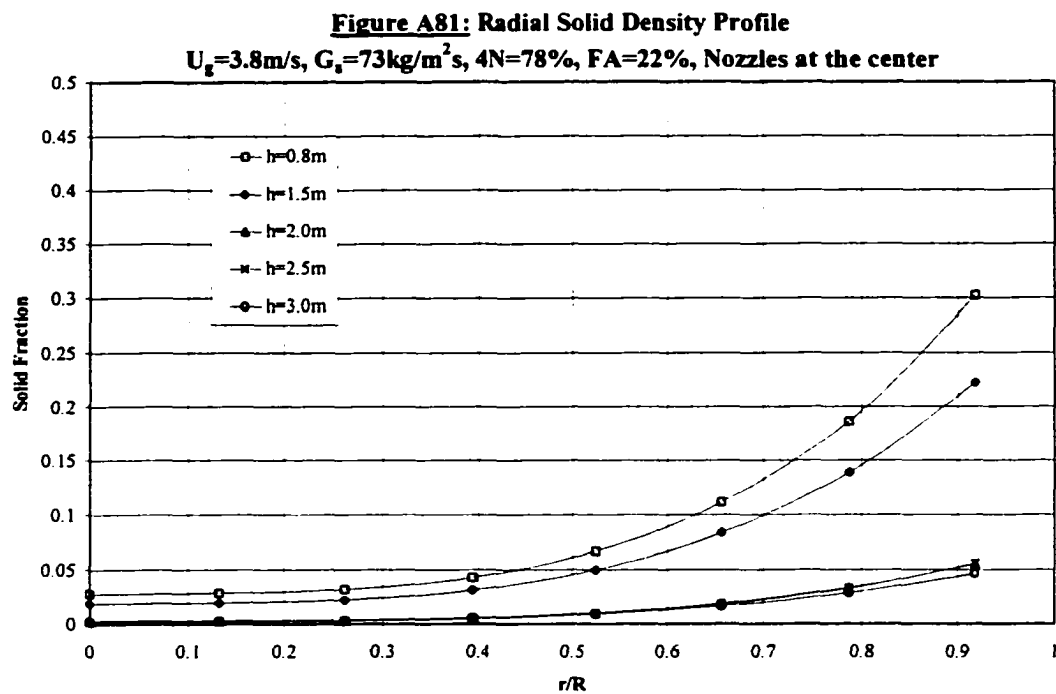
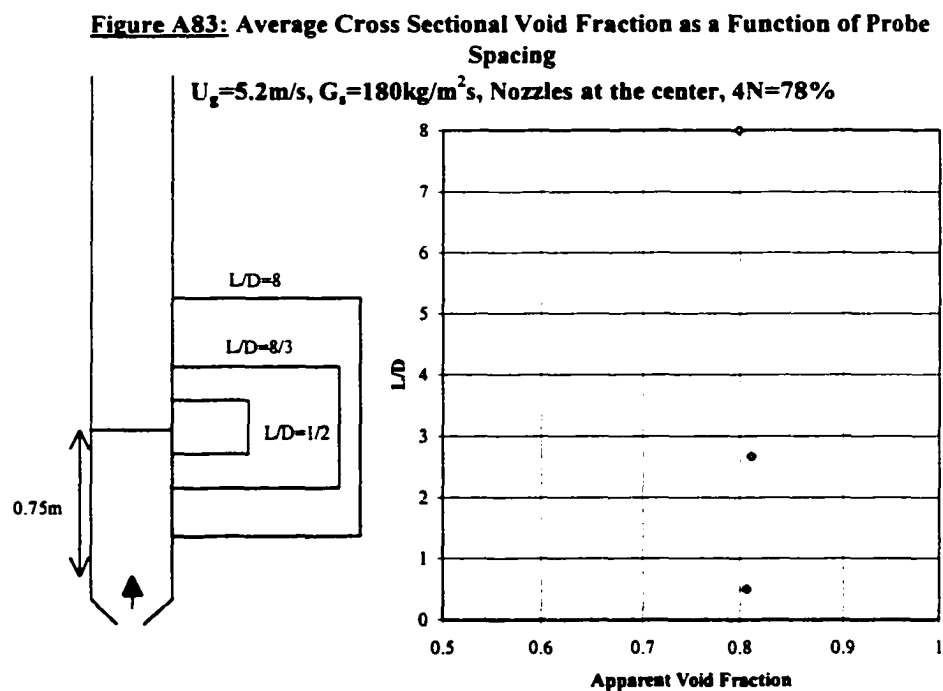
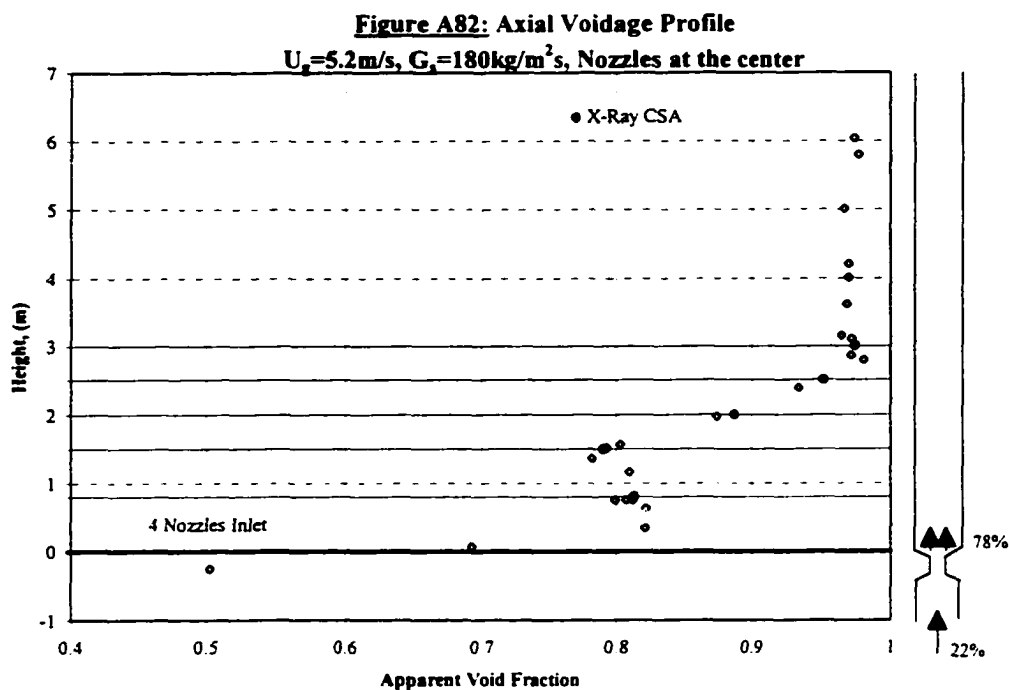
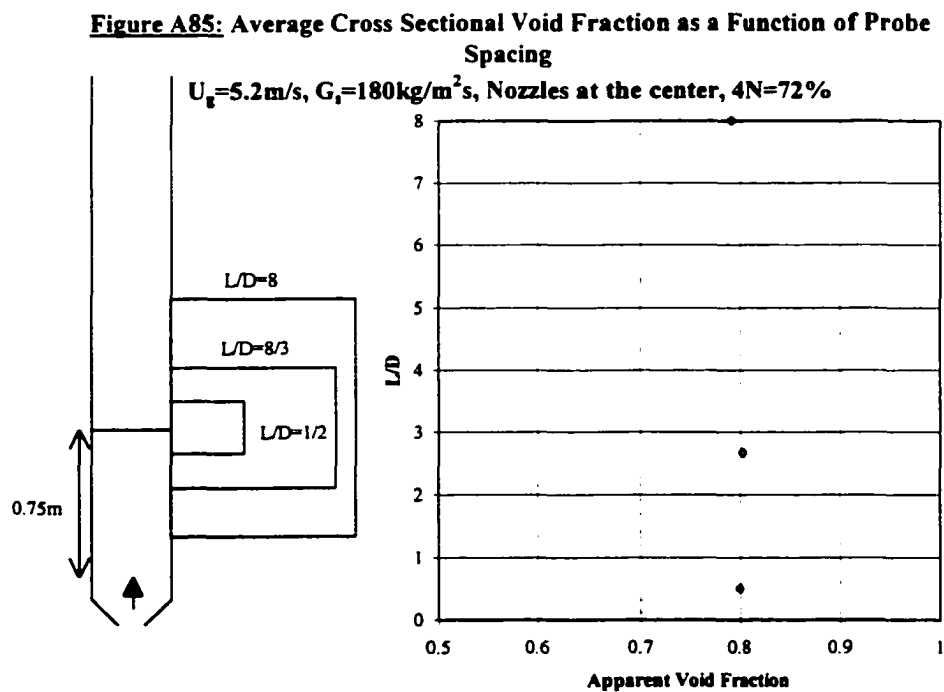
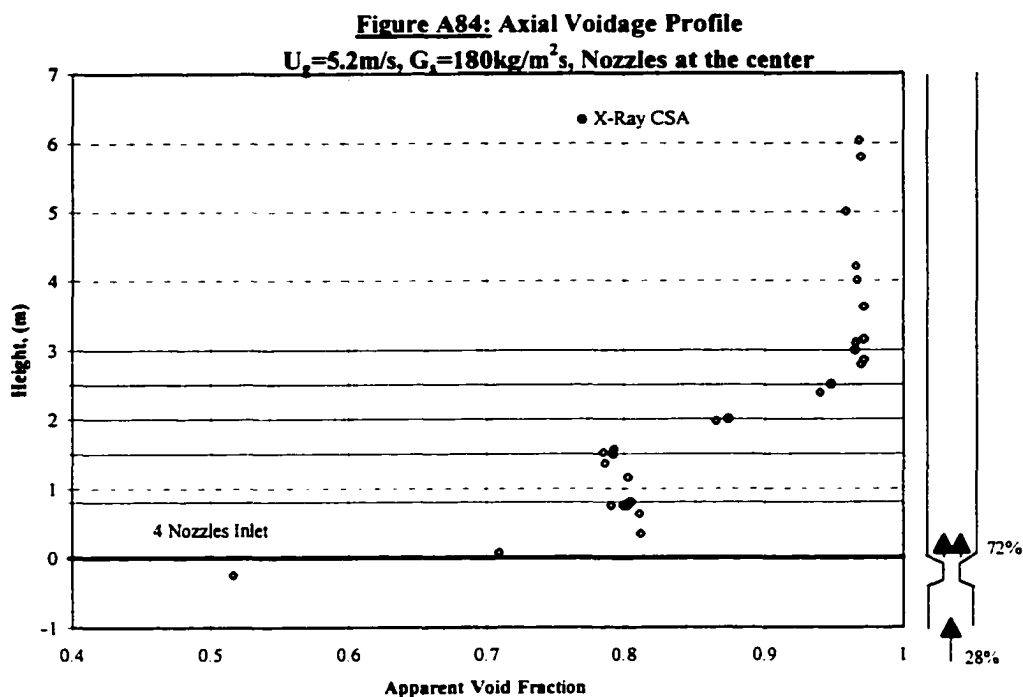
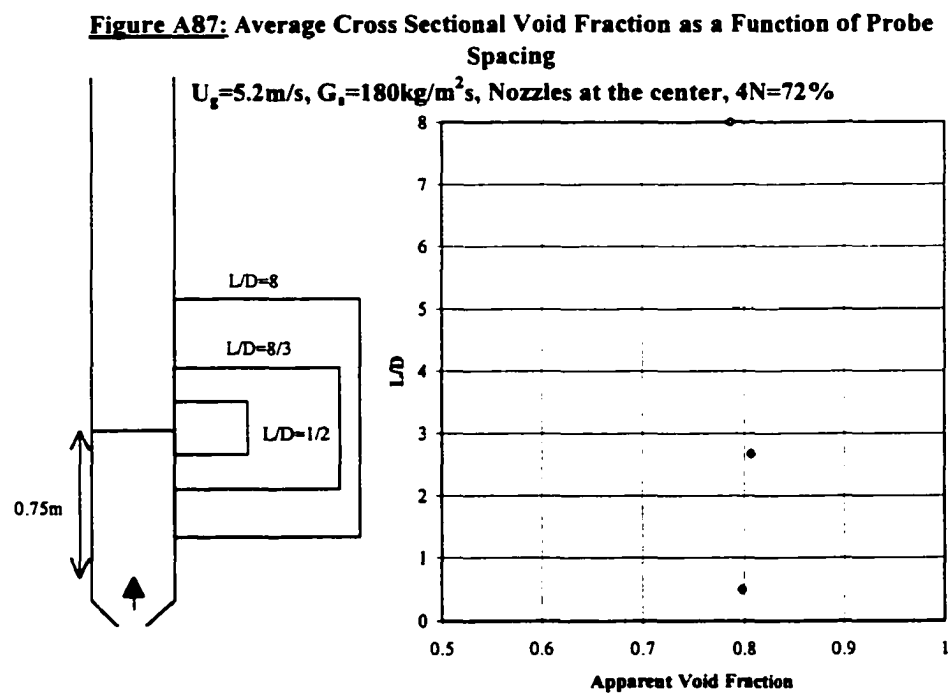
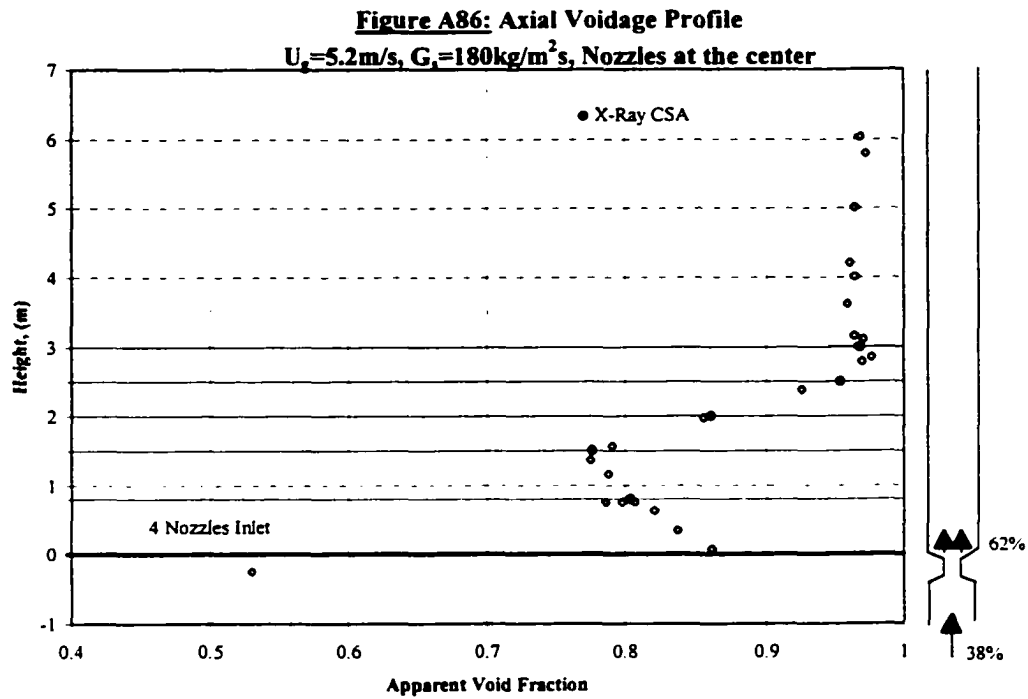
Figure A77: Radial Solid Profile $U_f=3.8\text{m/s}$, $G_s=111\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center**Figure A78: Radial Solid Profile** $U_f=3.8\text{m/s}$, $G_s=111\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center

Figure A79: Radial Solid Profile **$U_s=3.8\text{m/s}$, $G_s=73\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center****Figure A80: Radial Solid Profile** **$U_s=3.8\text{m/s}$, $G_s=73\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center**









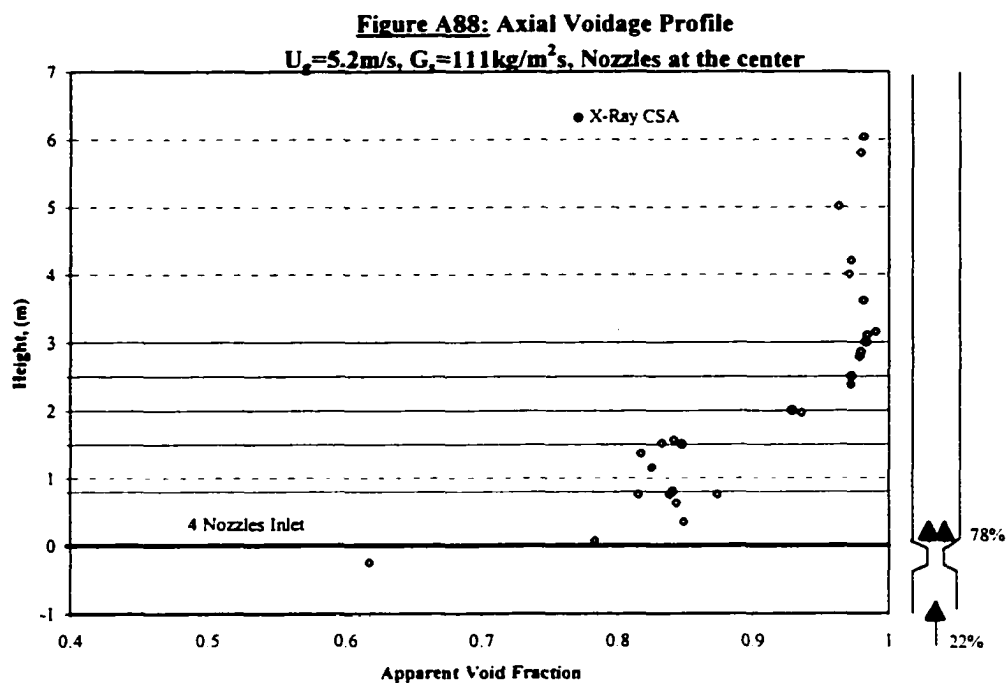
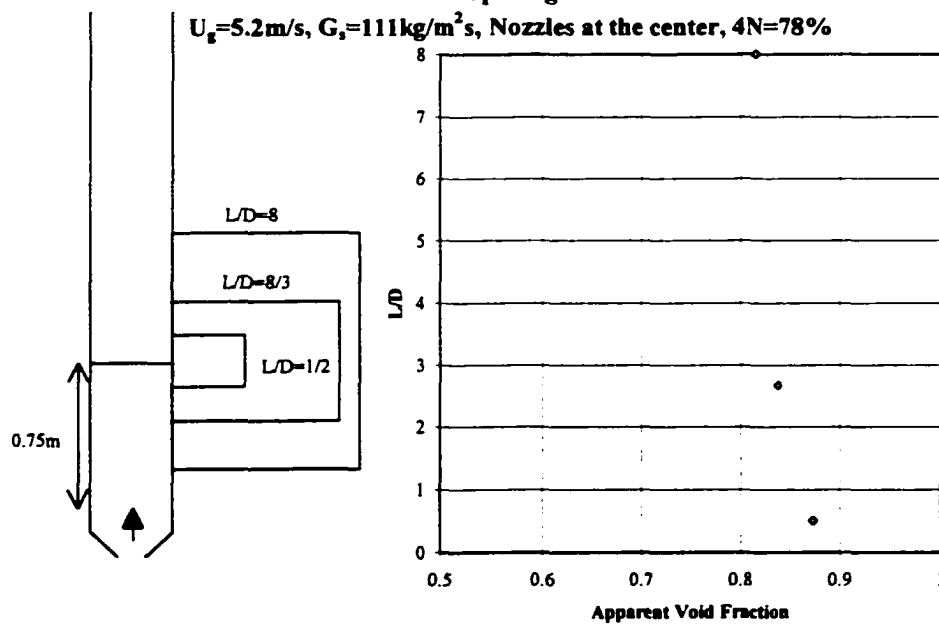


Figure A89: Average Cross Sectional Void Fraction as a Function of Probe Spacing



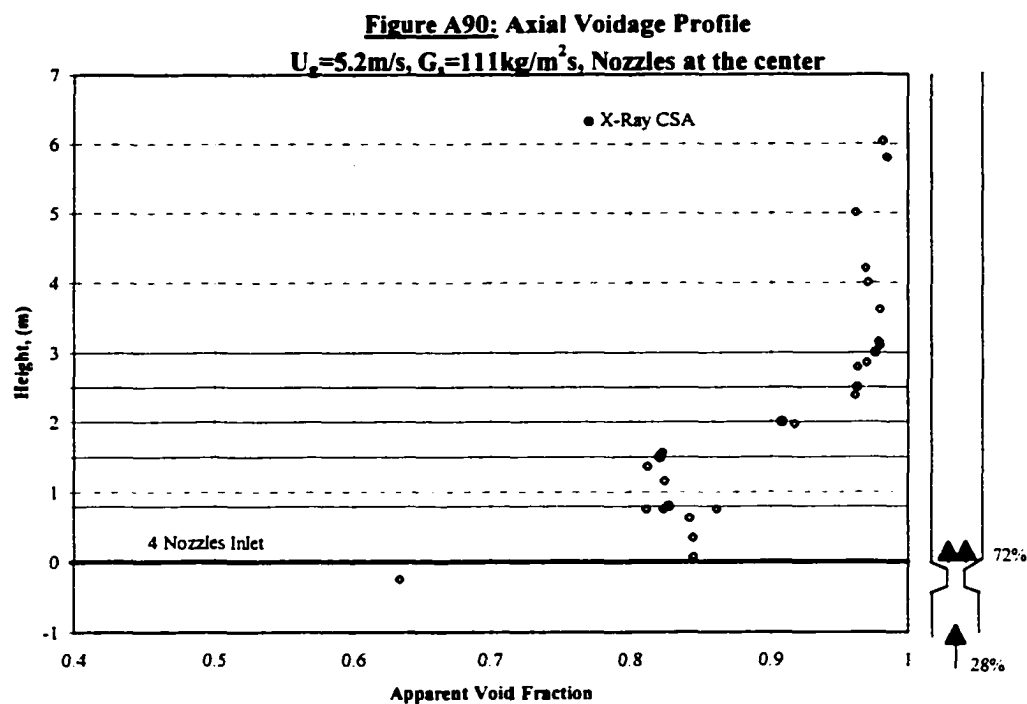
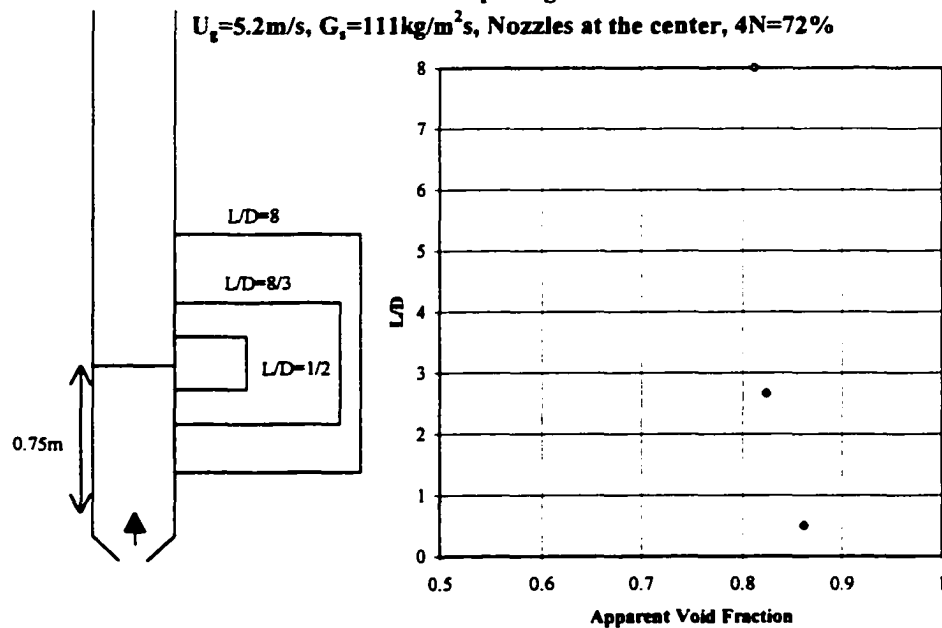
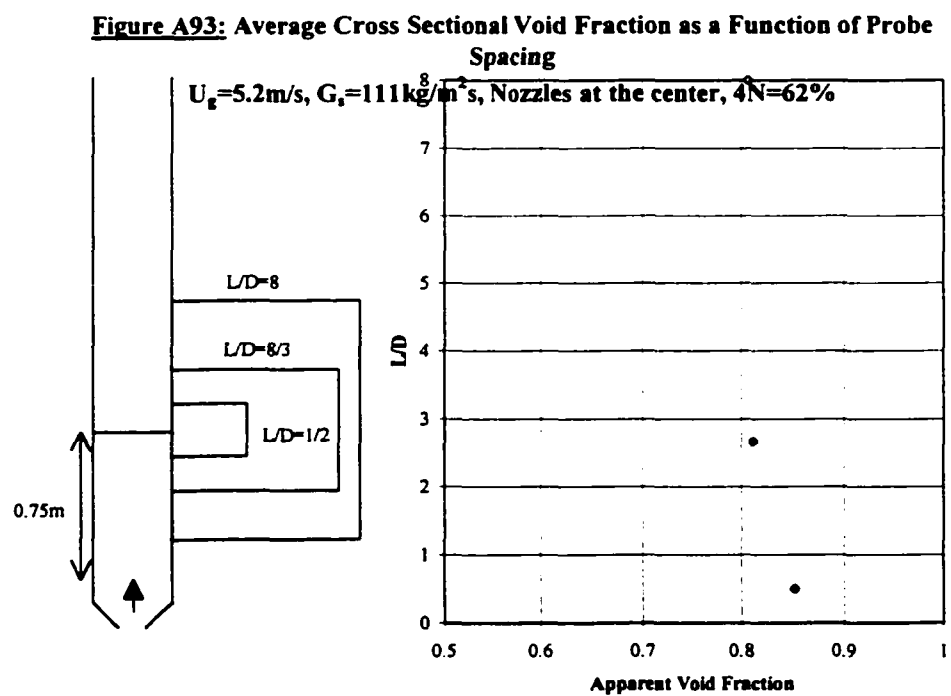
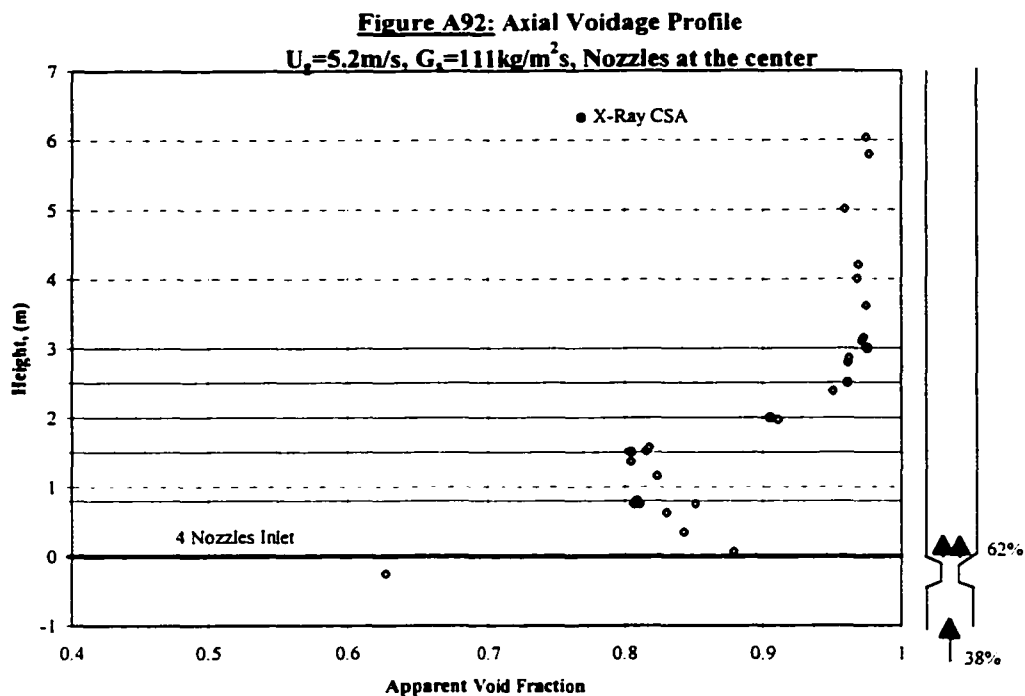


Figure A91: Average Cross Sectional Void Fraction as a Function of Probe Spacing





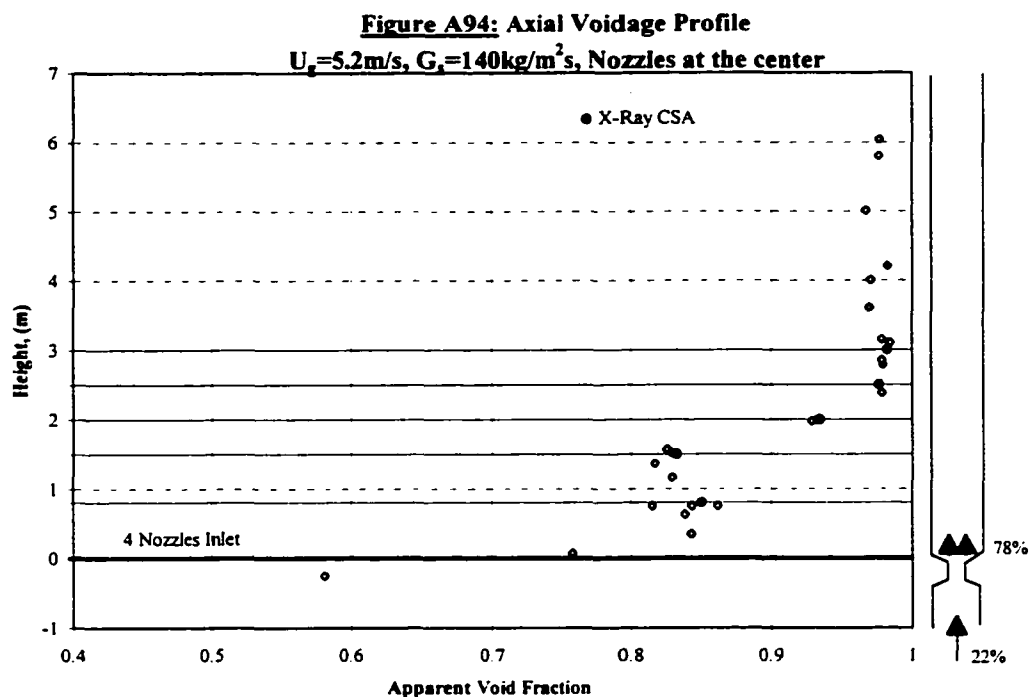
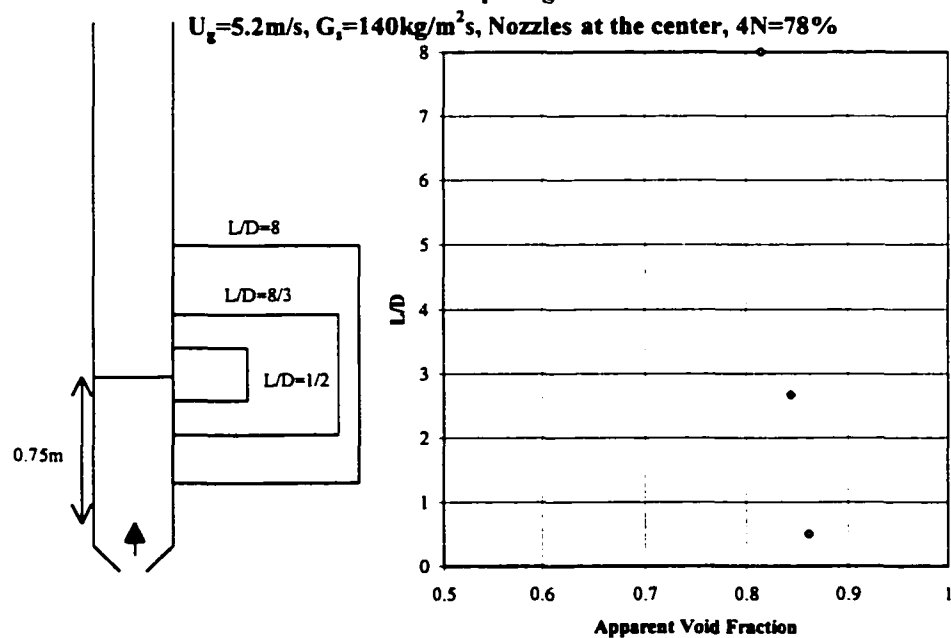


Figure A95: Average Cross Sectional Void Fraction as a Function of Probe Spacing



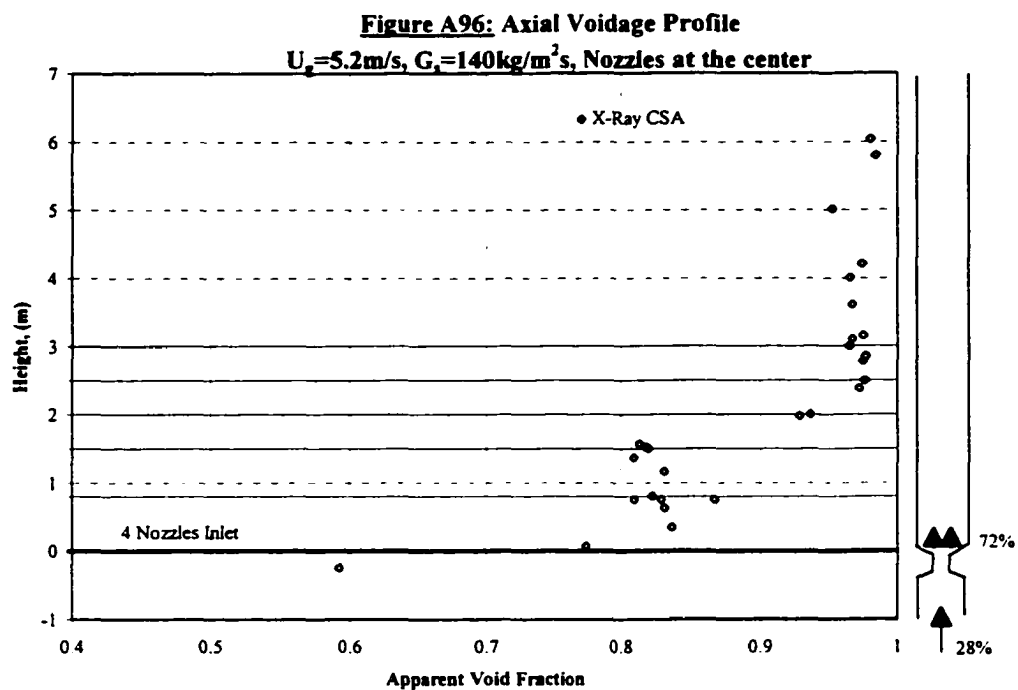
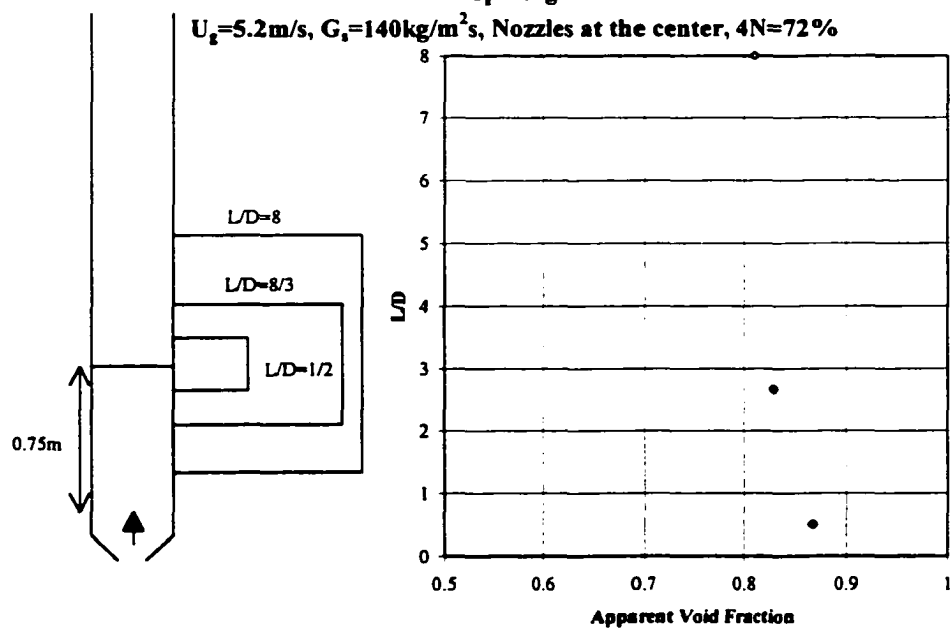
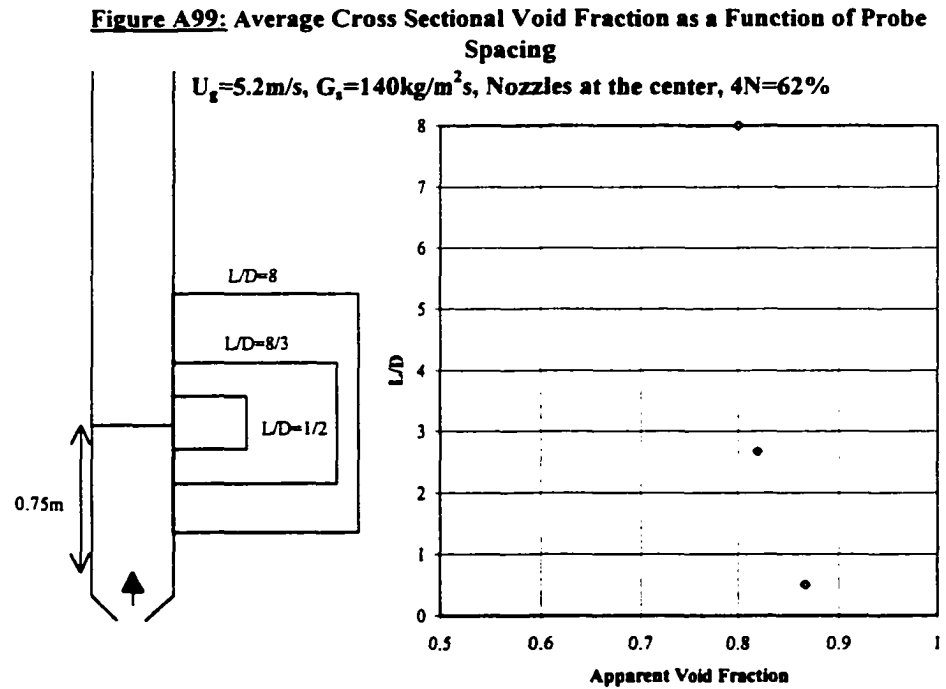
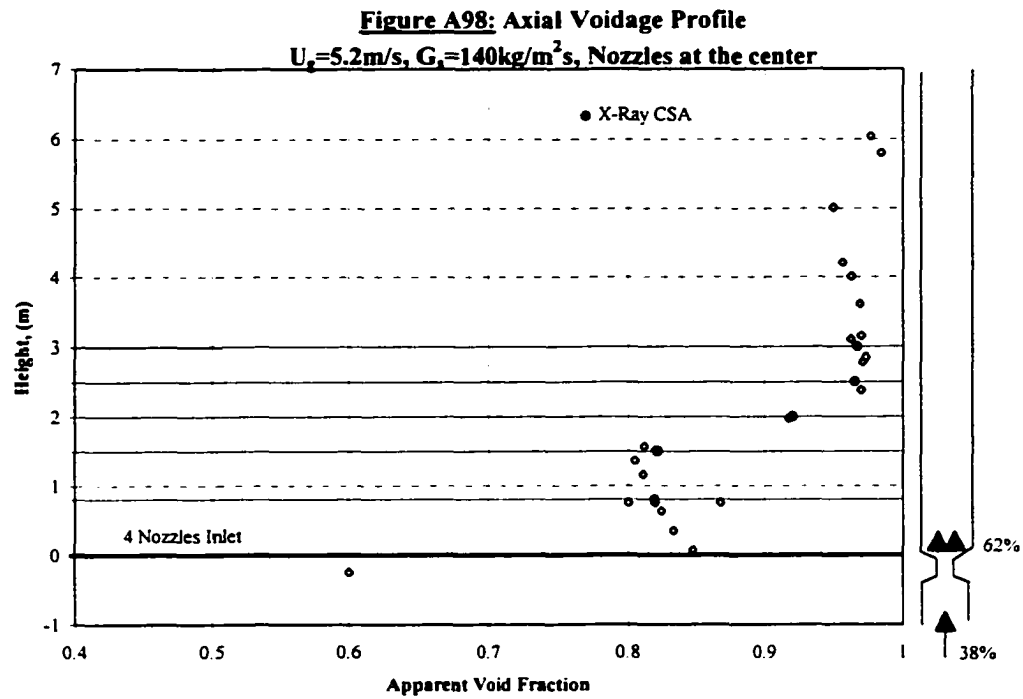


Figure A97: Average Cross Sectional Void Fraction as a Function of Probe Spacing





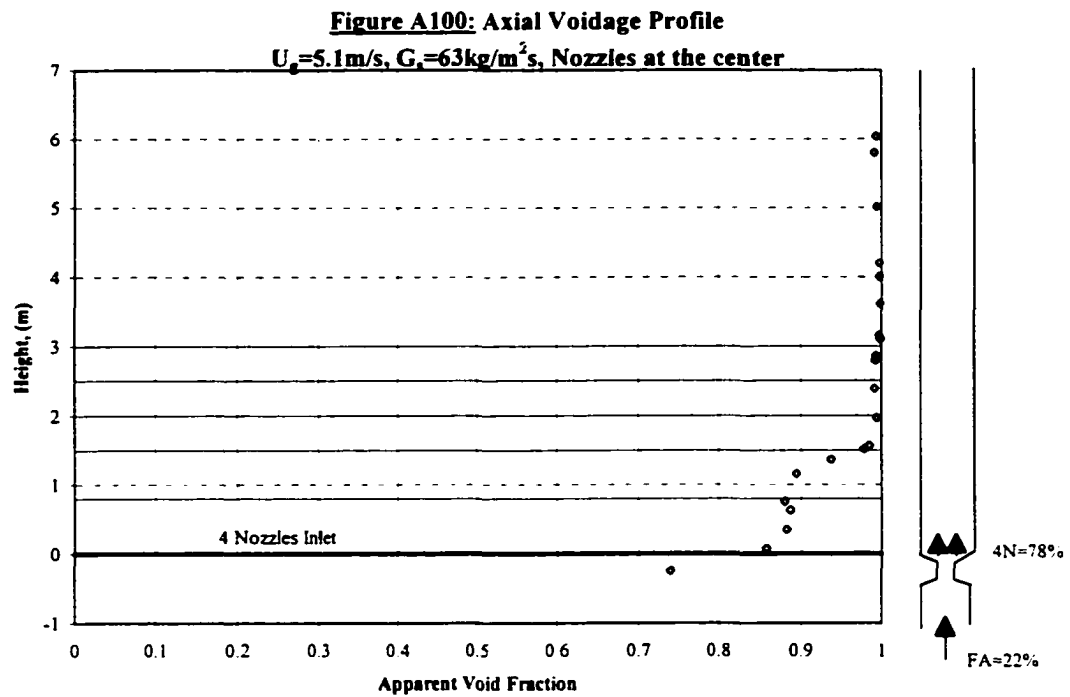
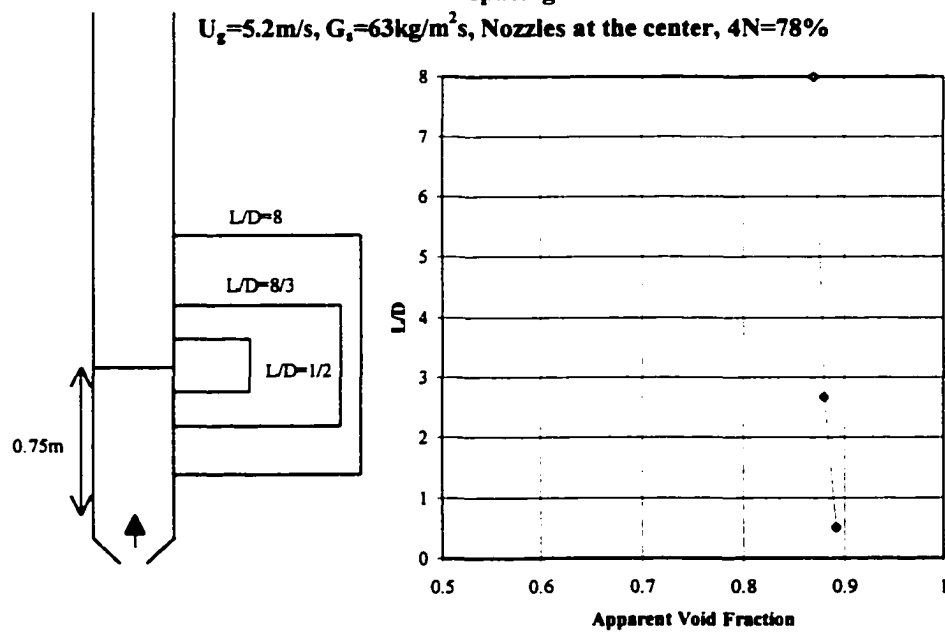


Figure A101: Average Cross Sectional Void Fraction as a Function of Probe Spacing



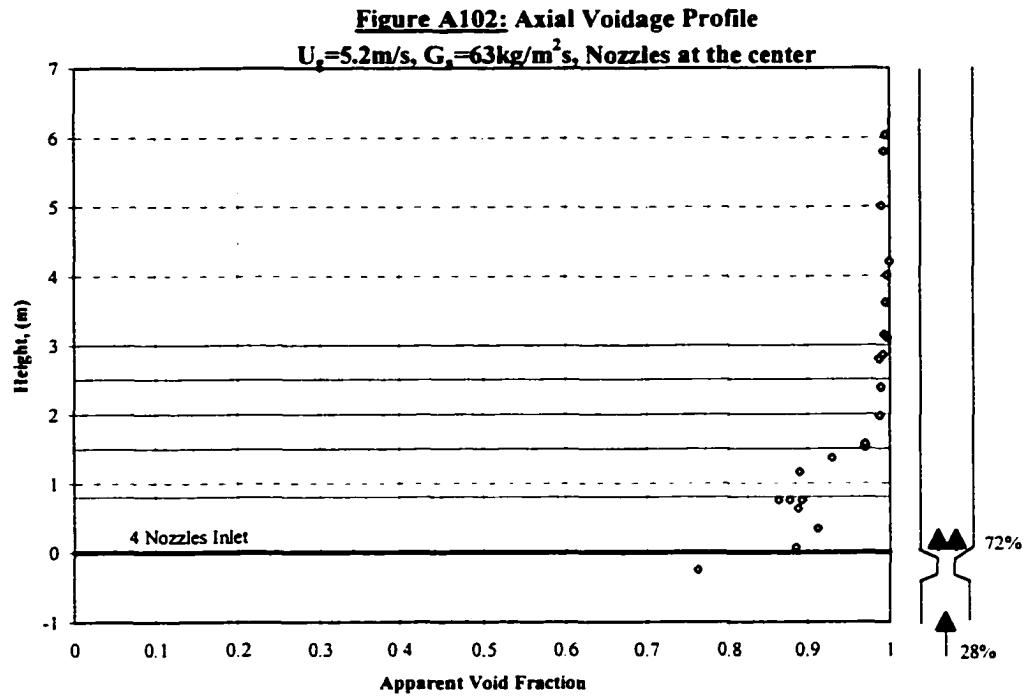
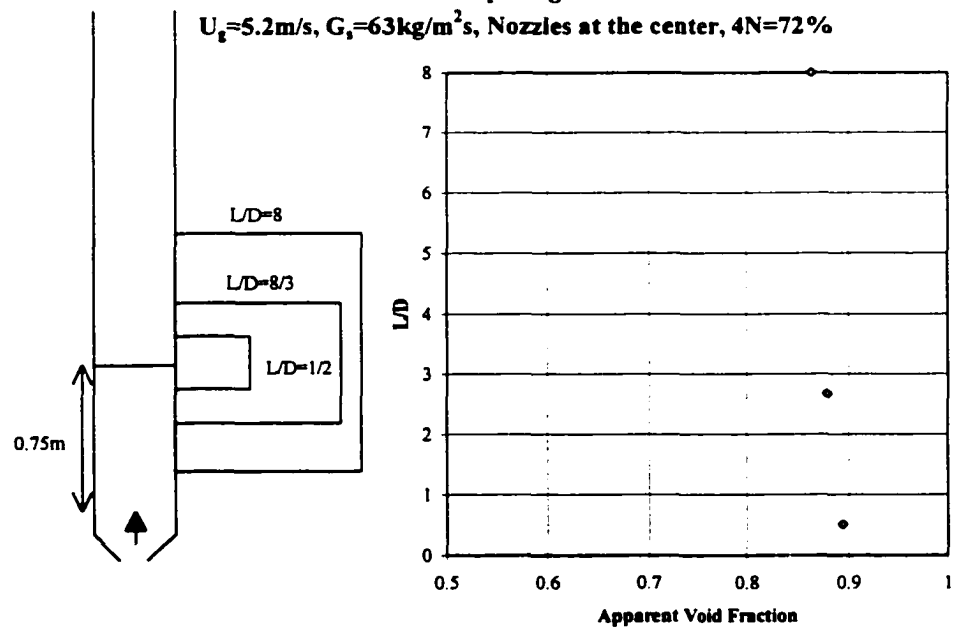


Figure A103: Average Cross Sectional Void Fraction as a Function of Probe Spacing



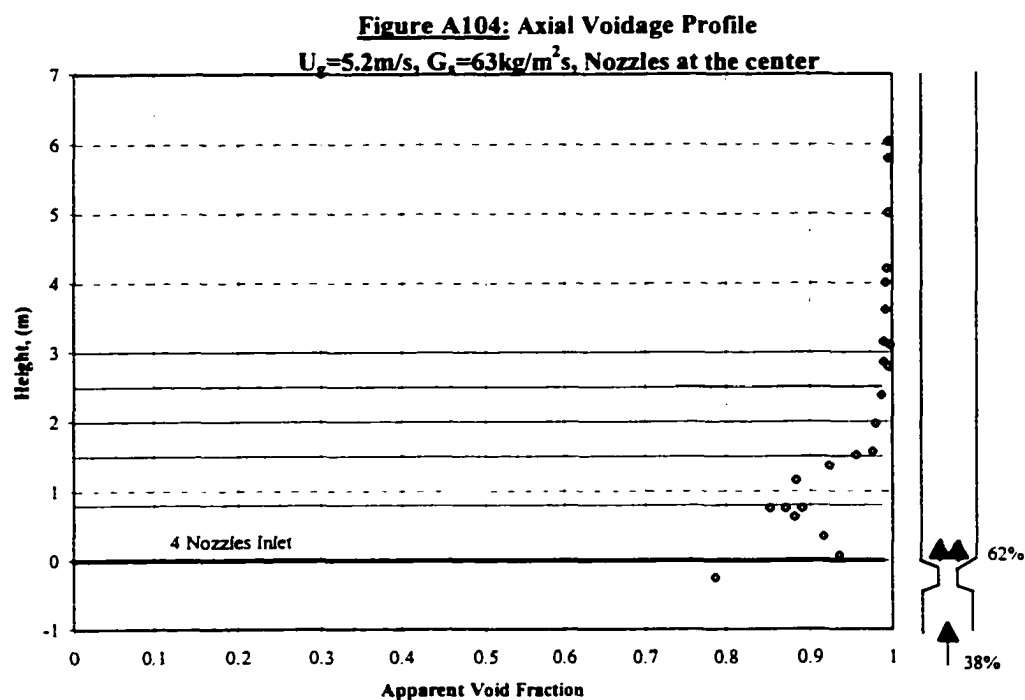


Figure A105: Average Cross Sectional Void Fraction as a Function of Probe Spacing

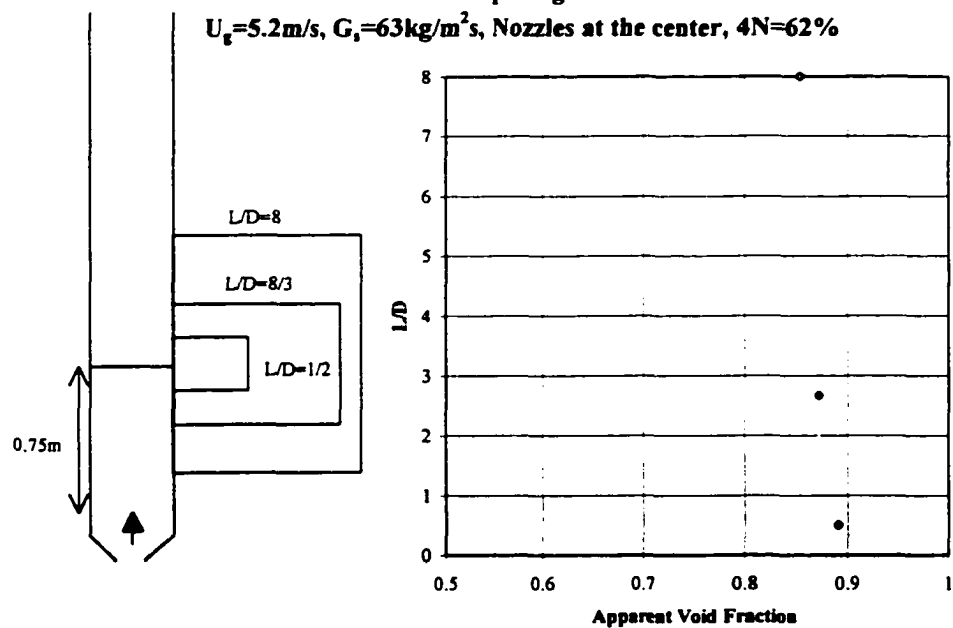


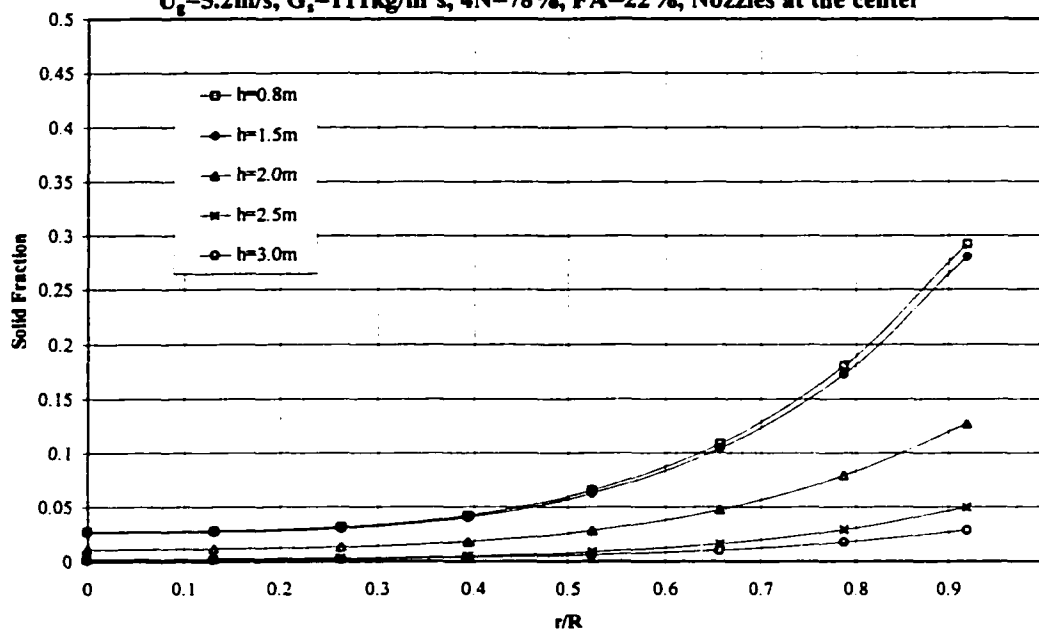
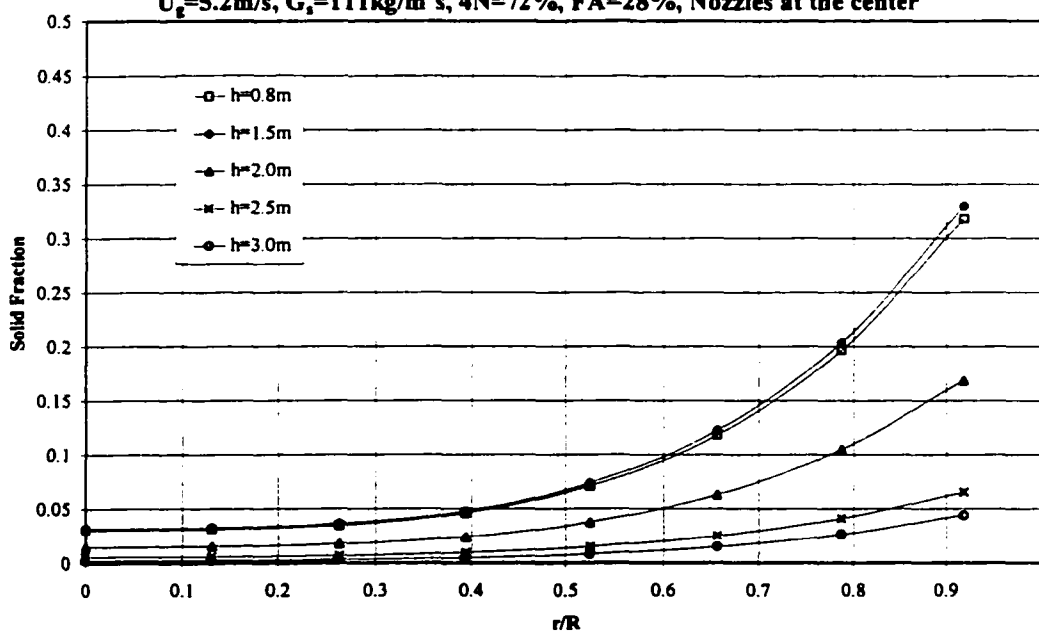
Figure A106: Radial Solid Profile $U_g=5.2\text{m/s}$, $G_s=111\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center**Figure A107: Radial Solid Profile** $U_g=5.2\text{m/s}$, $G_s=111\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center

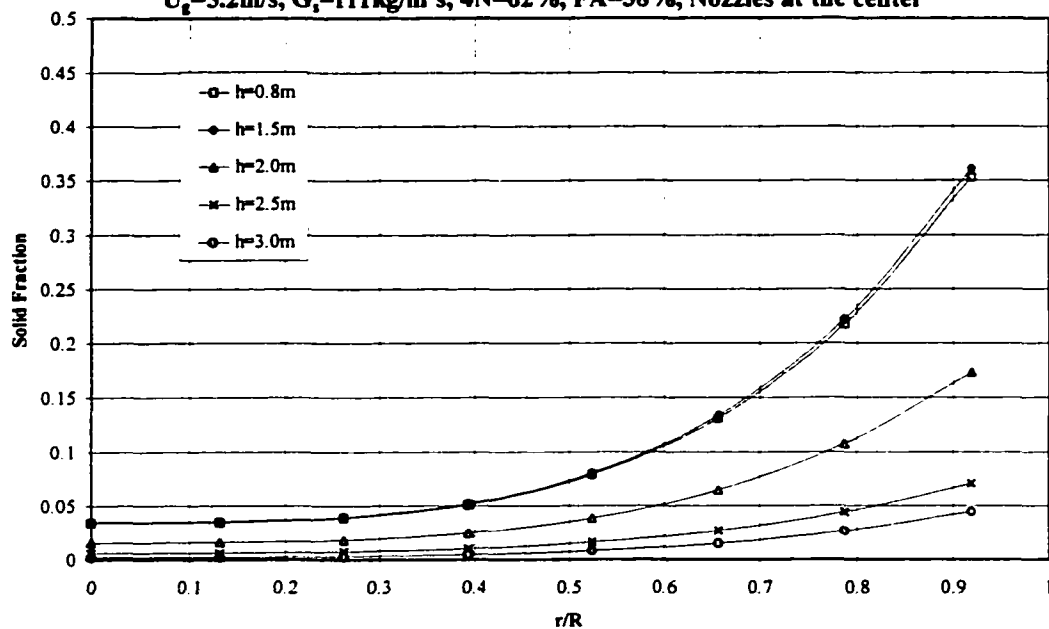
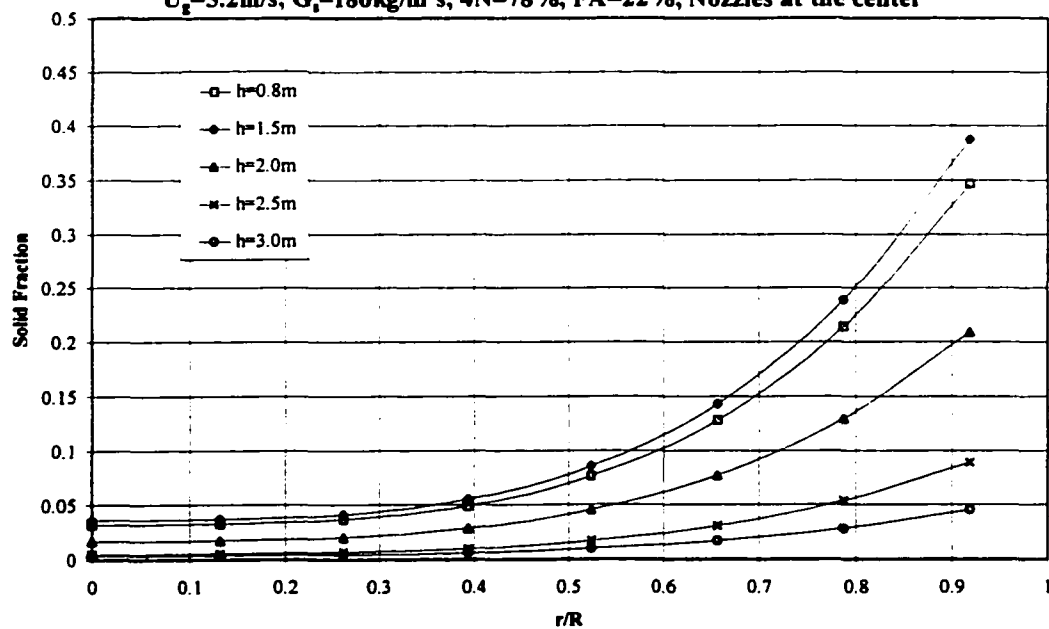
Figure A108: Radial Solid Profile **$U_t=5.2\text{m/s}$, $G_t=111\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center****Figure A109: Radial Solid Profile** **$U_t=5.2\text{m/s}$, $G_t=180\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center**

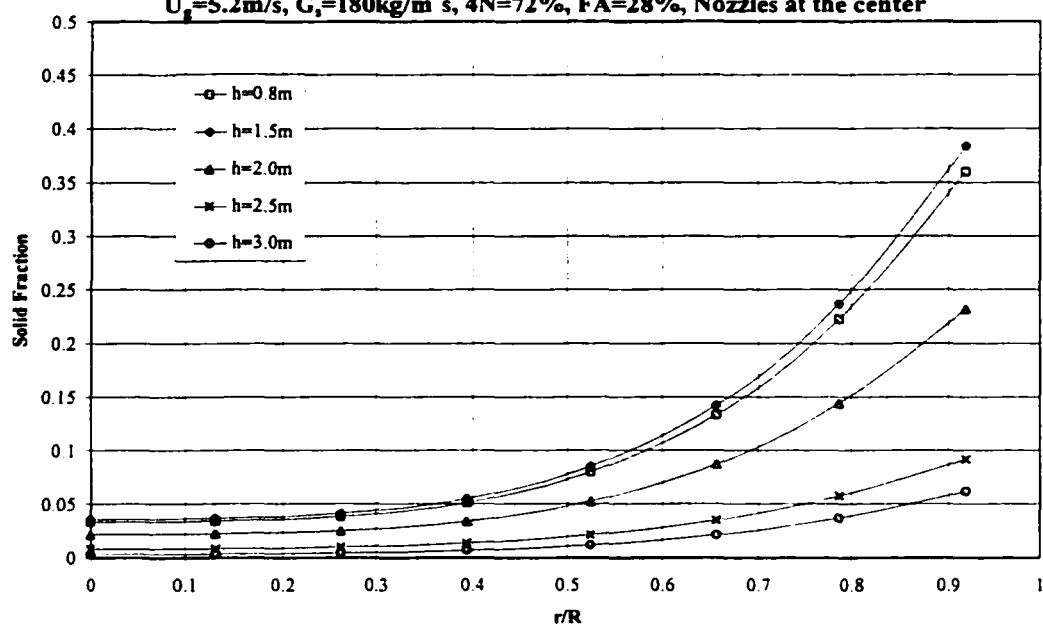
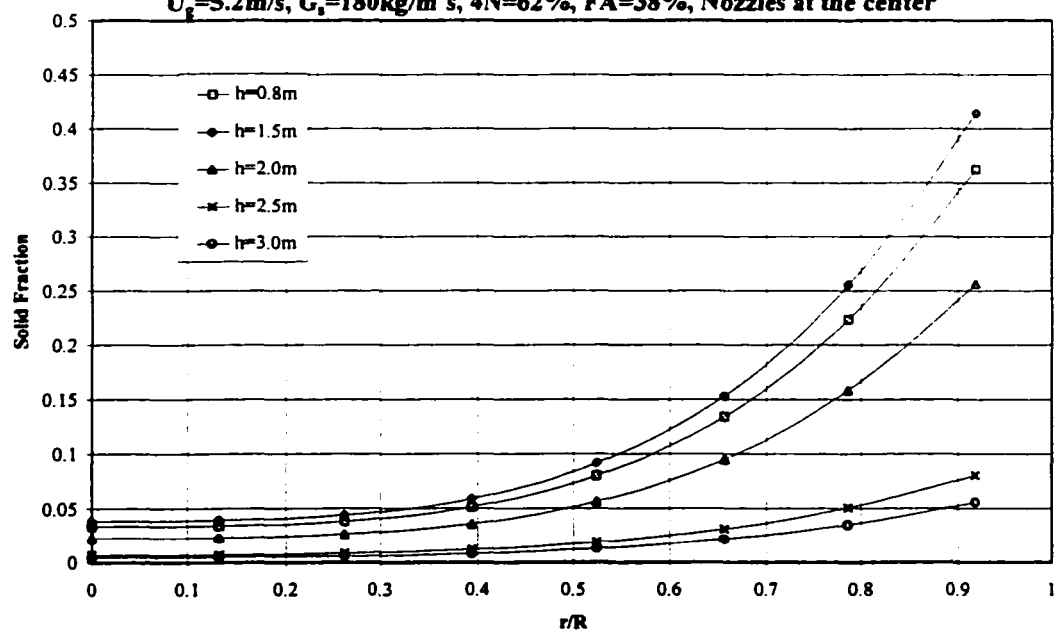
Figure A110: Radial Solid Profile $U_f=5.2\text{m/s}$, $G_s=180\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center**Figure A111: Radial Solid Profile** $U_f=5.2\text{m/s}$, $G_s=180\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center

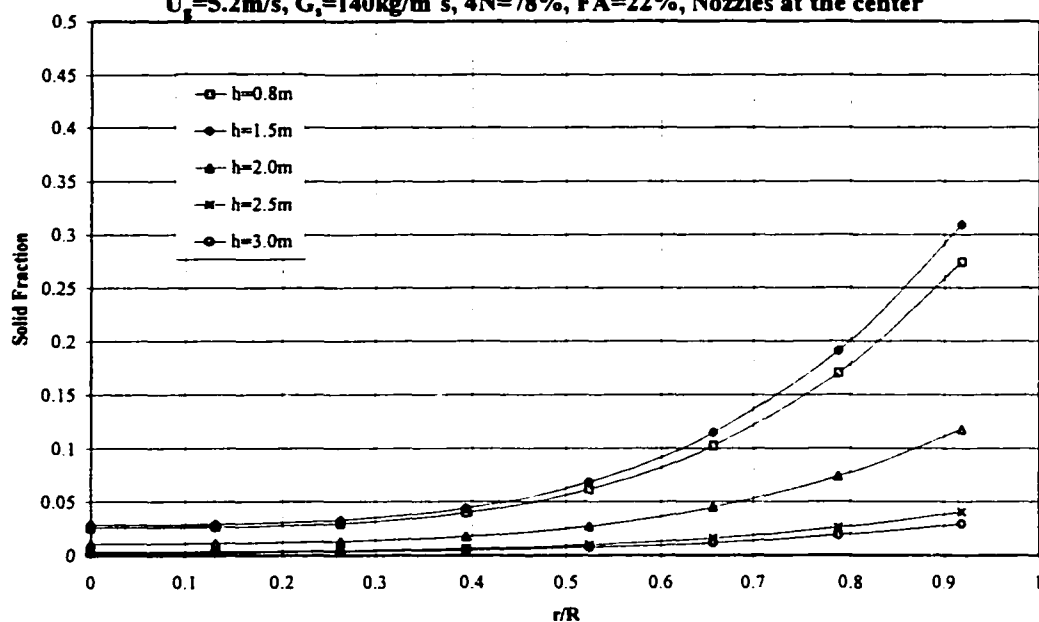
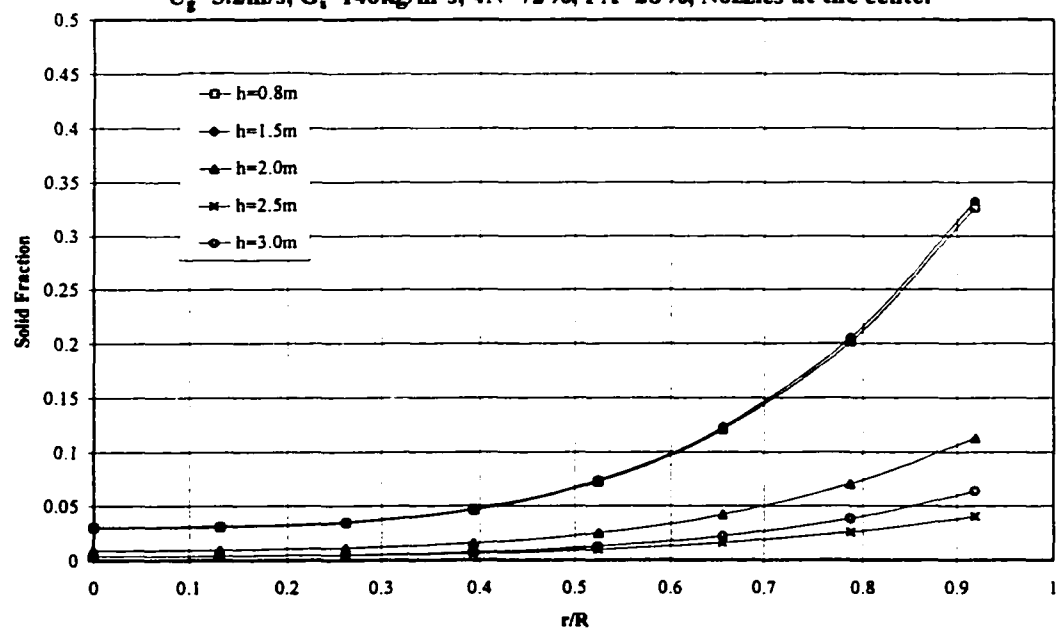
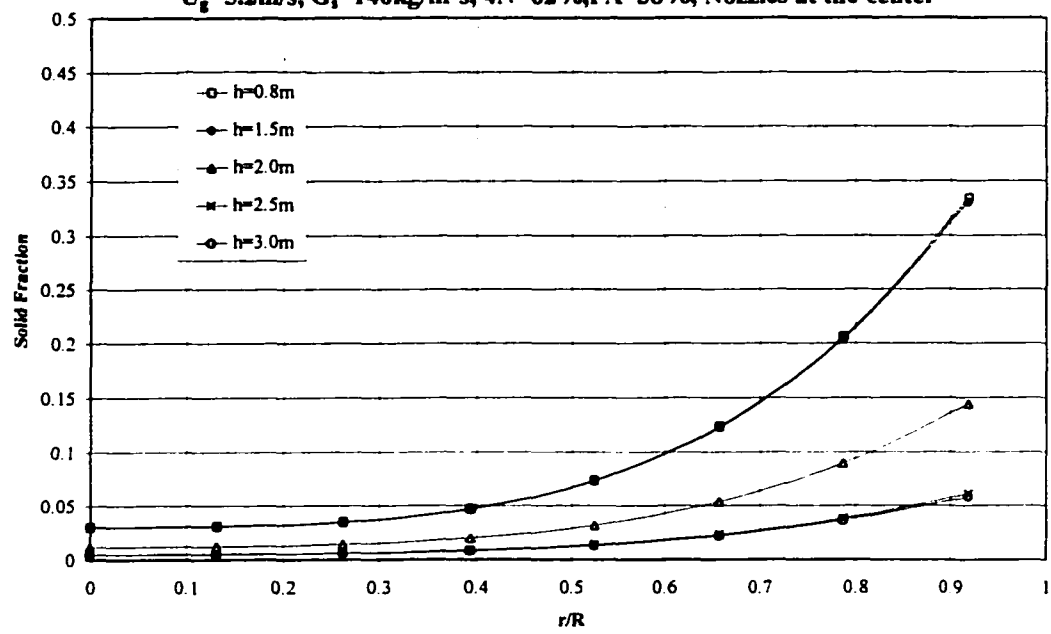
Figure A112: Radial Solid Profile **$U_g=5.2\text{m/s}$, $G_i=140\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the center****Figure A113: Radial Solid Profile** **$U_g=5.2\text{m/s}$, $G_i=140\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the center**

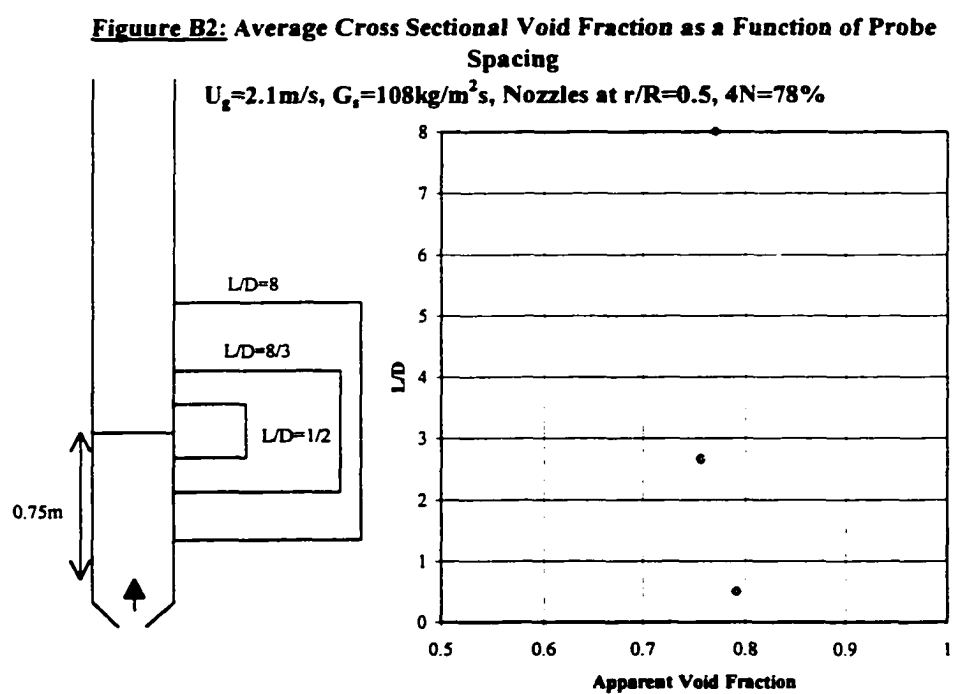
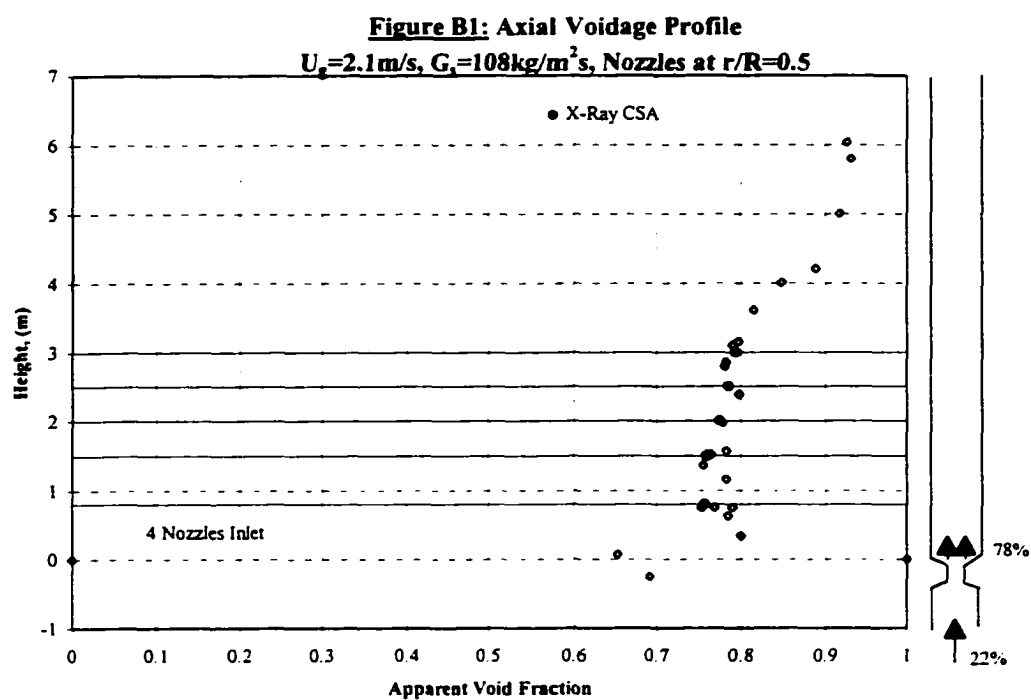
Figure A114: Radial Solid Profile

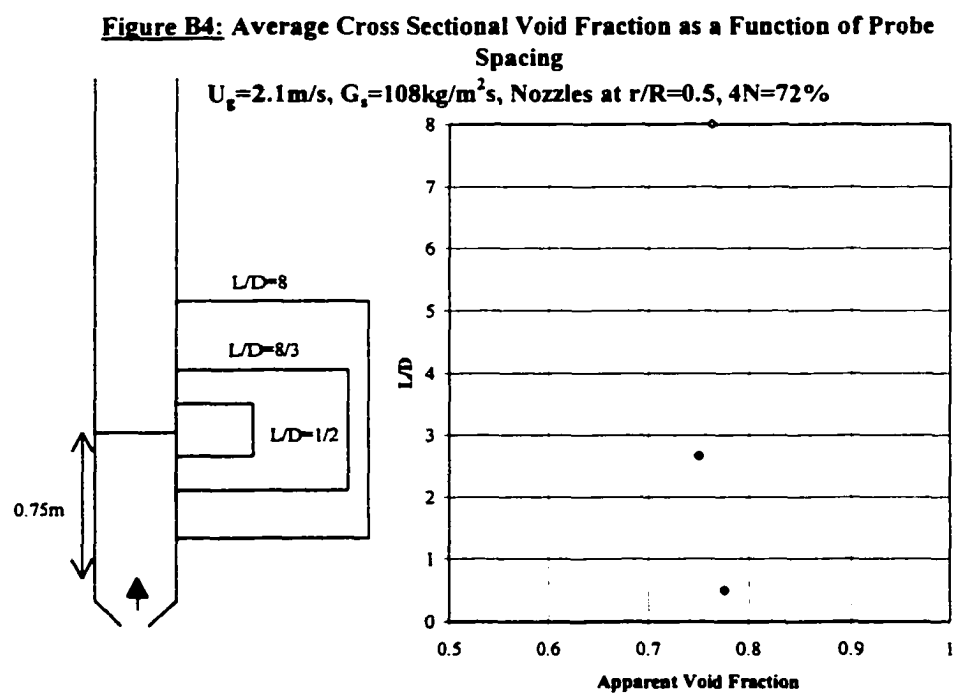
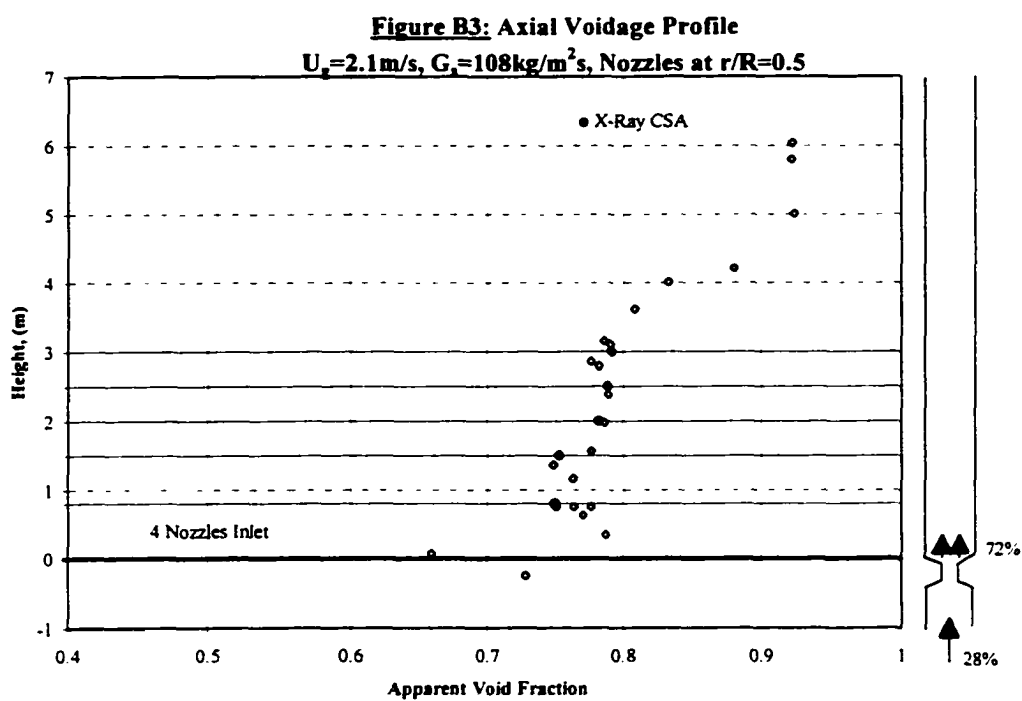
$U_i=5.2\text{m/s}$, $G_i=140\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the center

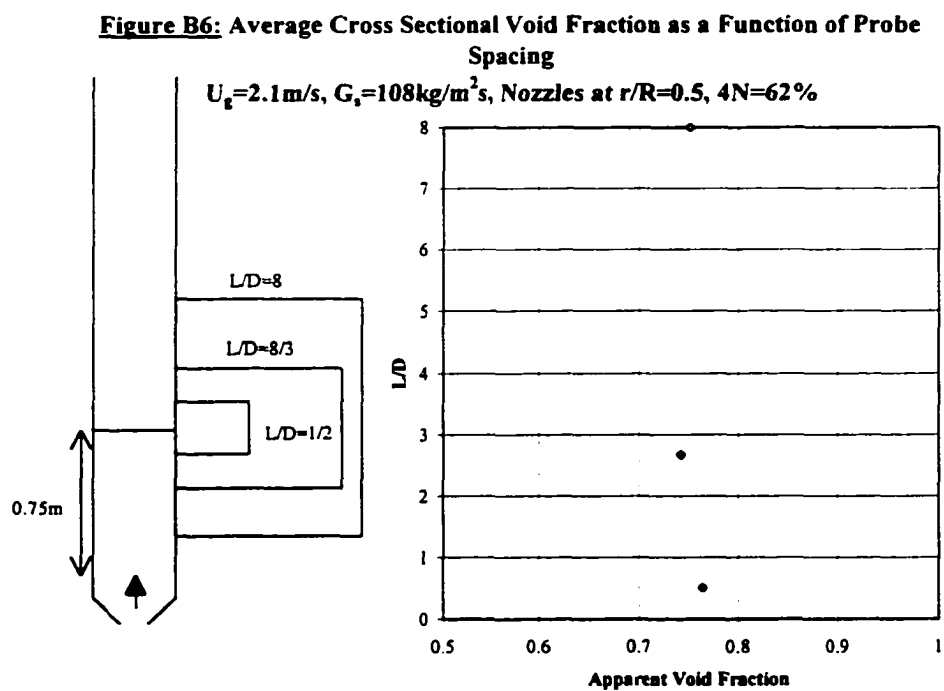
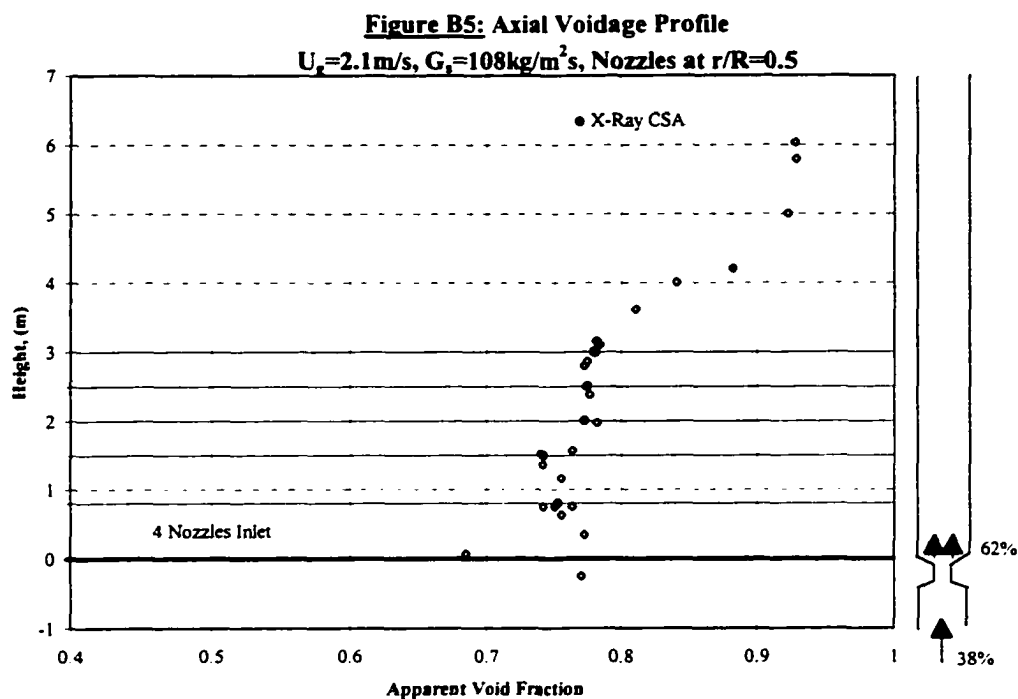


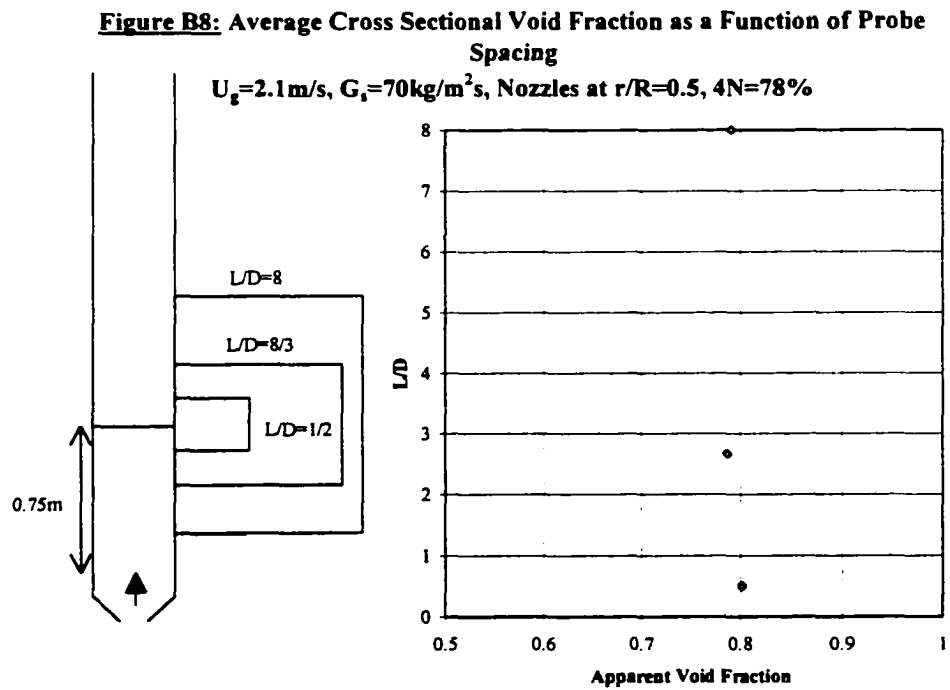
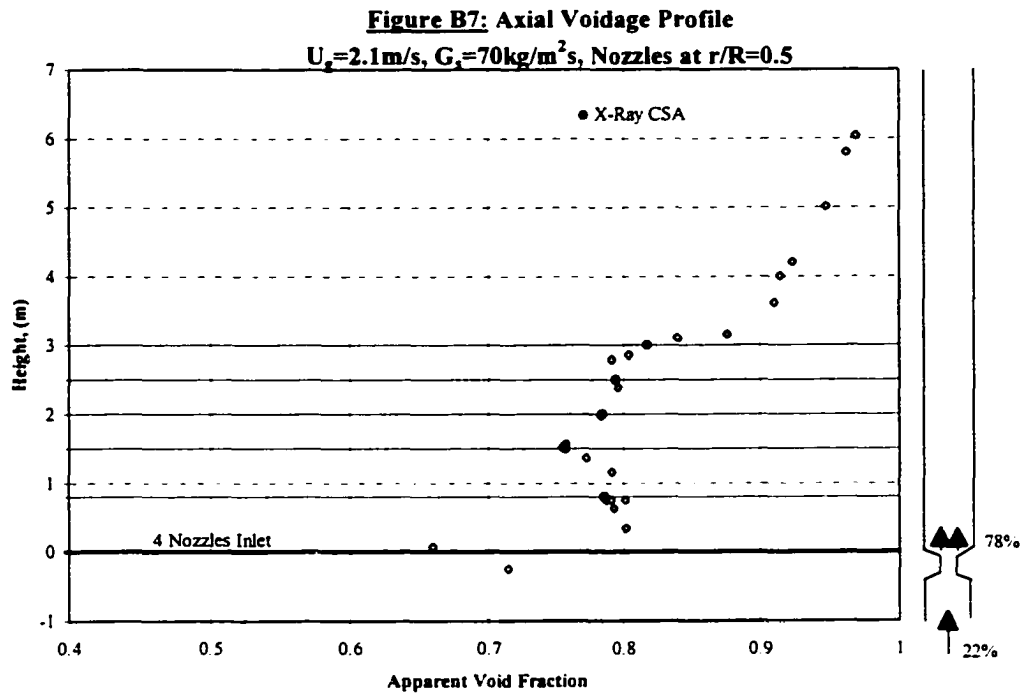
APPENDIX B

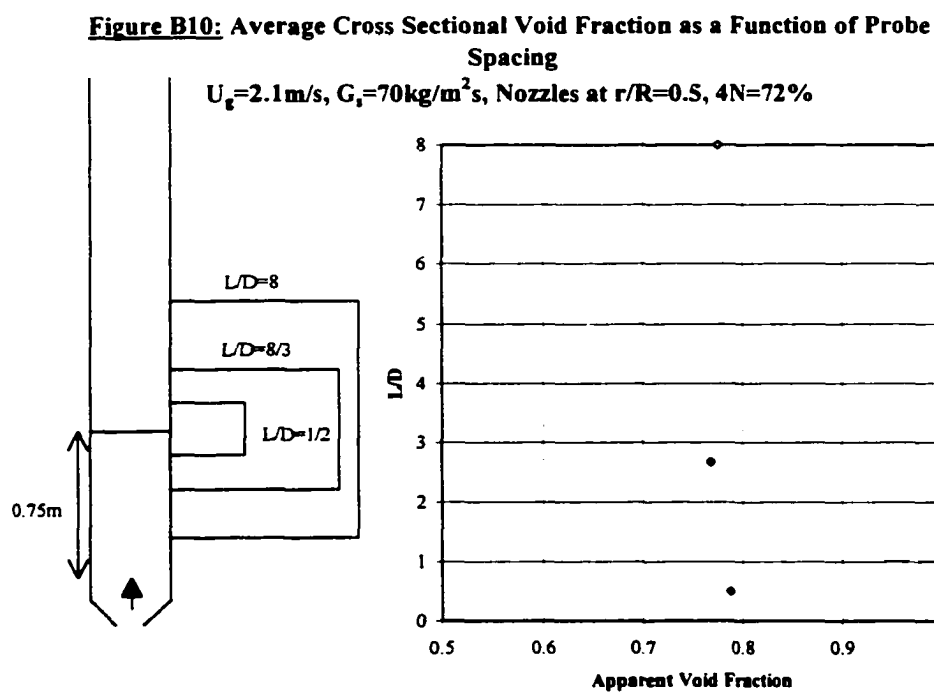
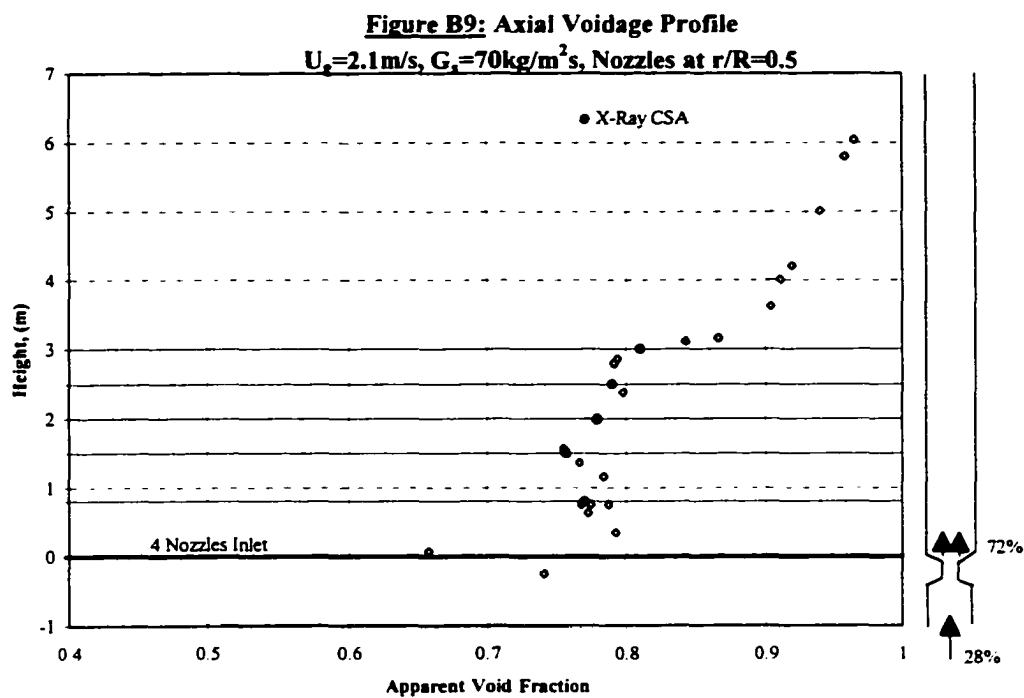
In Appendix B are presented all the experimental results with the nozzles located at the middle of the riser. These results include axial and radial solid distribution and the effect of the pressure tap spacing on the axial pressure difference measurements for a variety of operating conditions, gas superficial velocity, solid circulation rate and "fluffing" air flowrate.

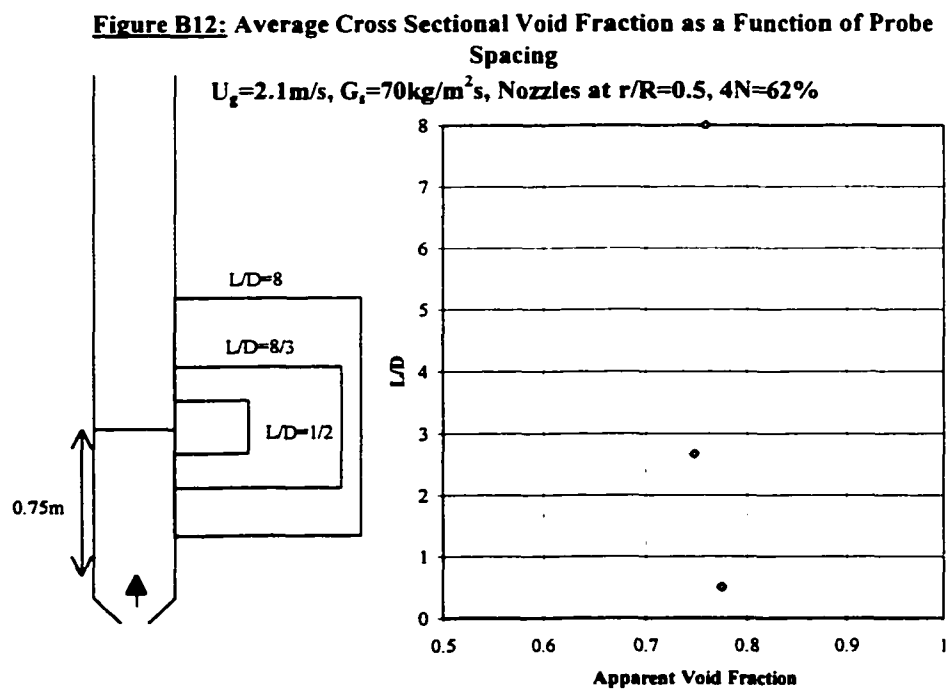
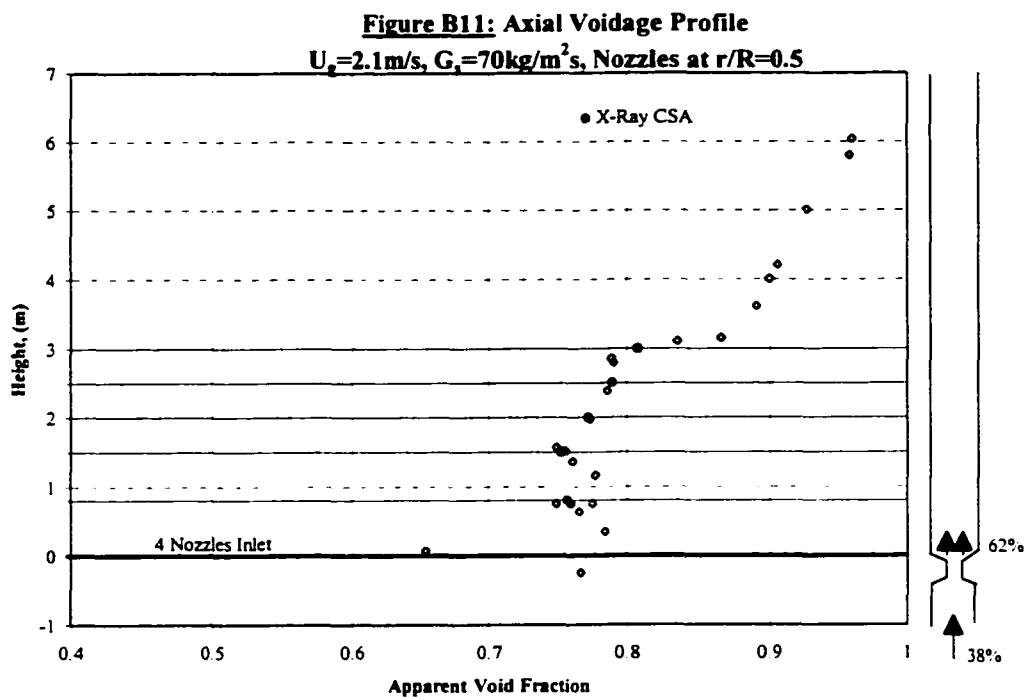












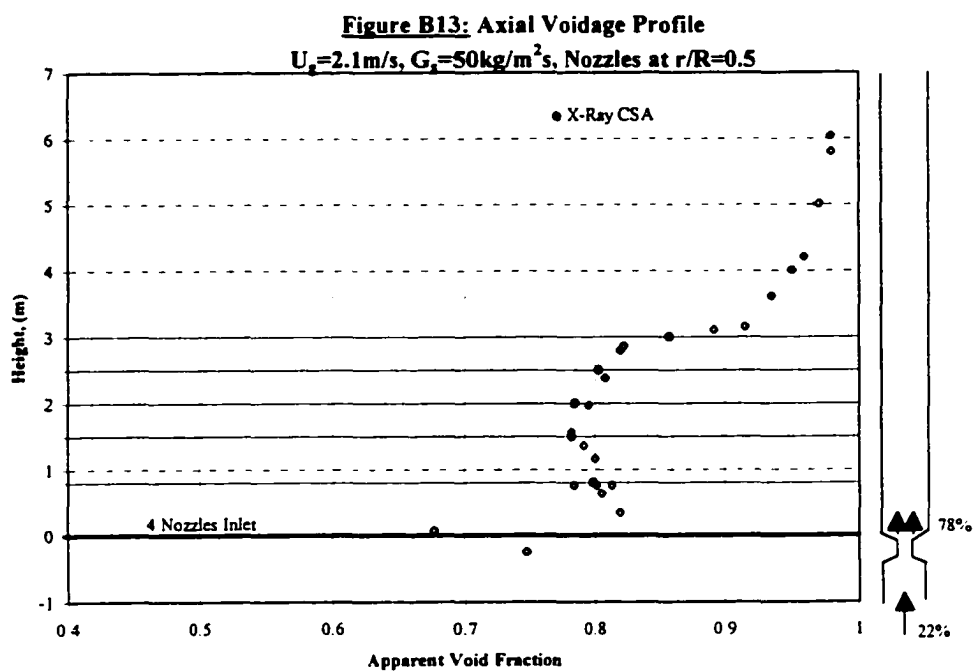
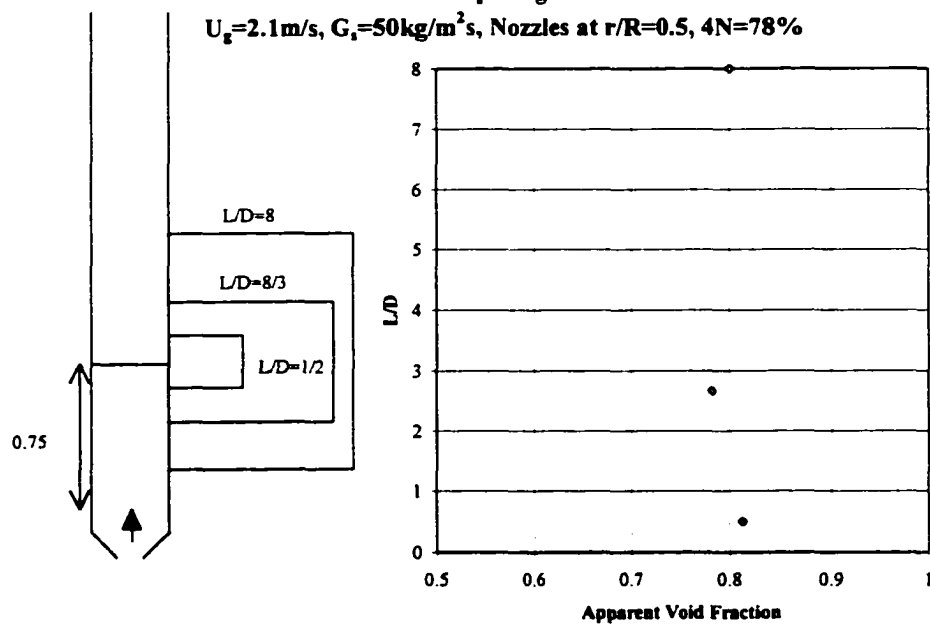


Figure B14: Average Cross Sectional Void Fraction as a Function of Probe Spacing



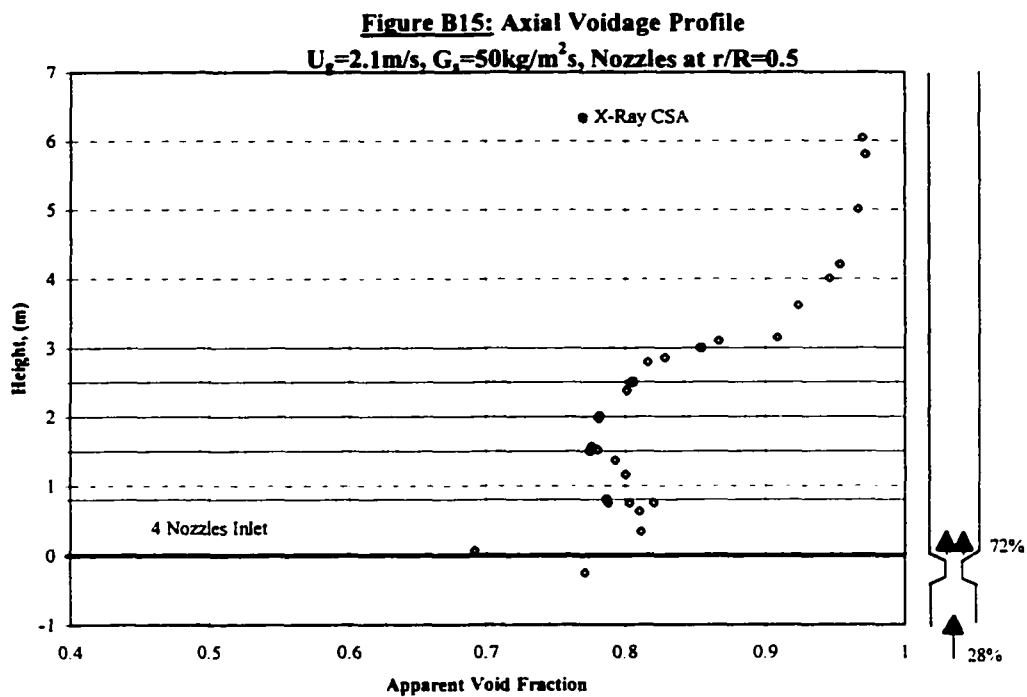
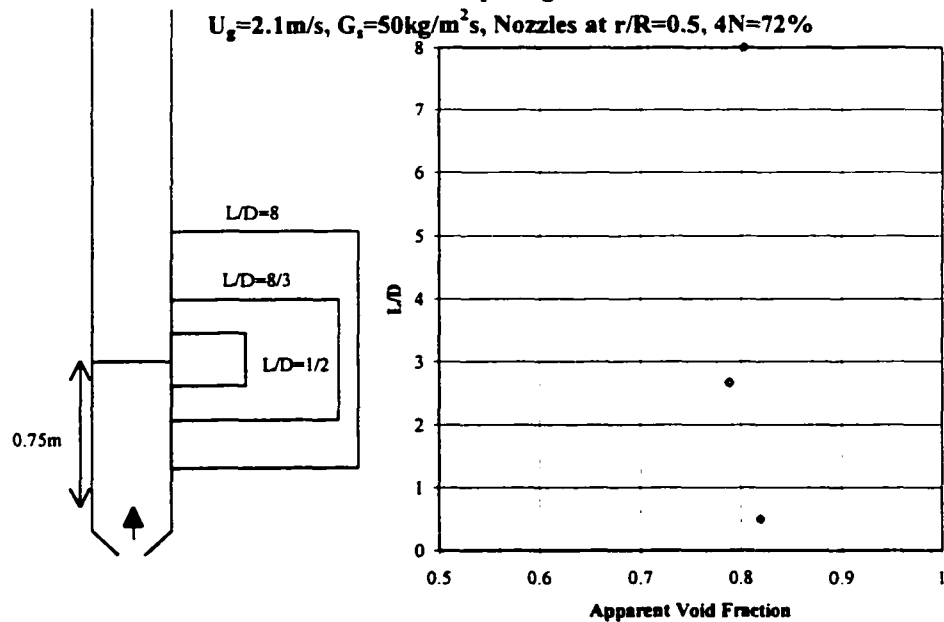


Figure B16: Average Cross Sectional Void Fraction as a Function of Probe Spacing



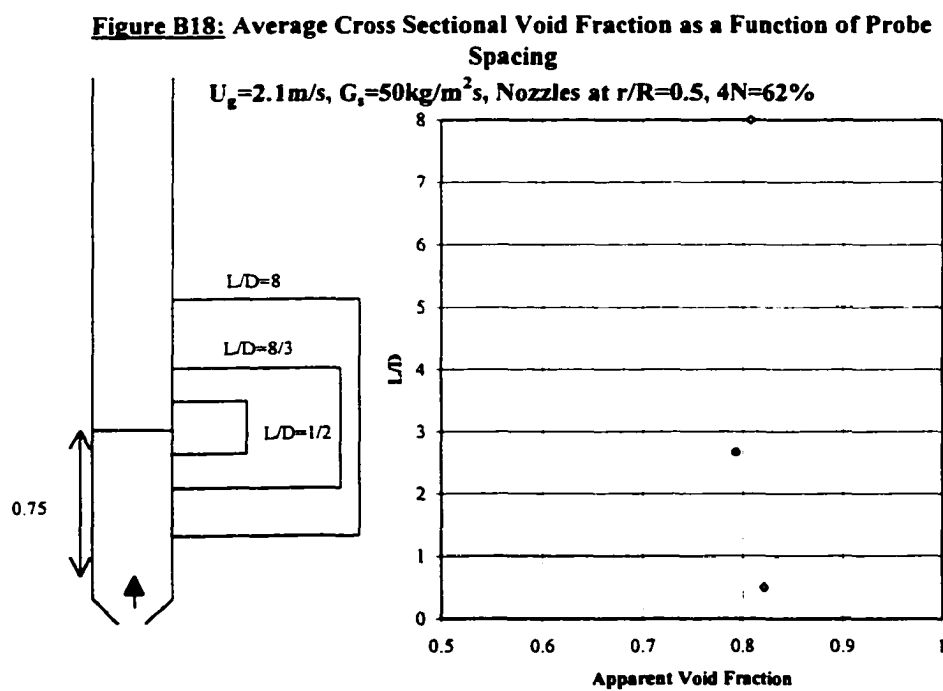
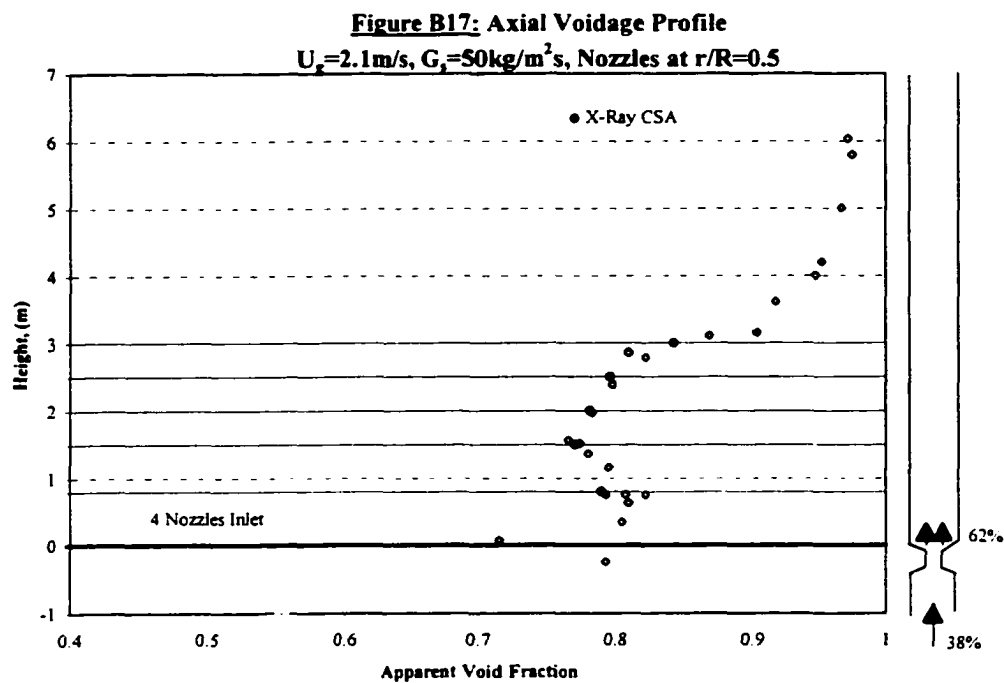


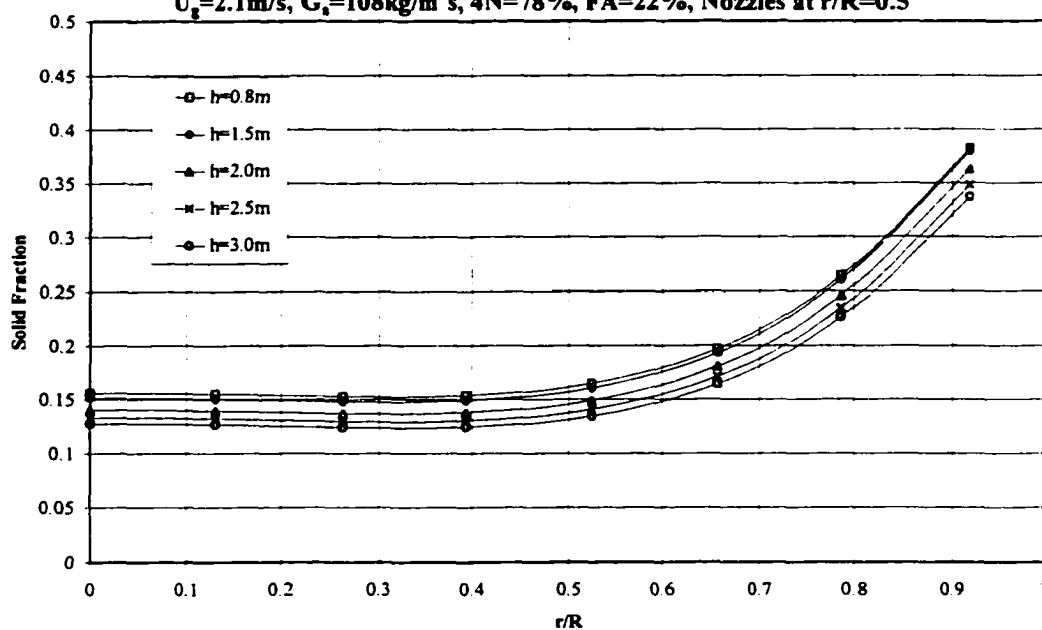
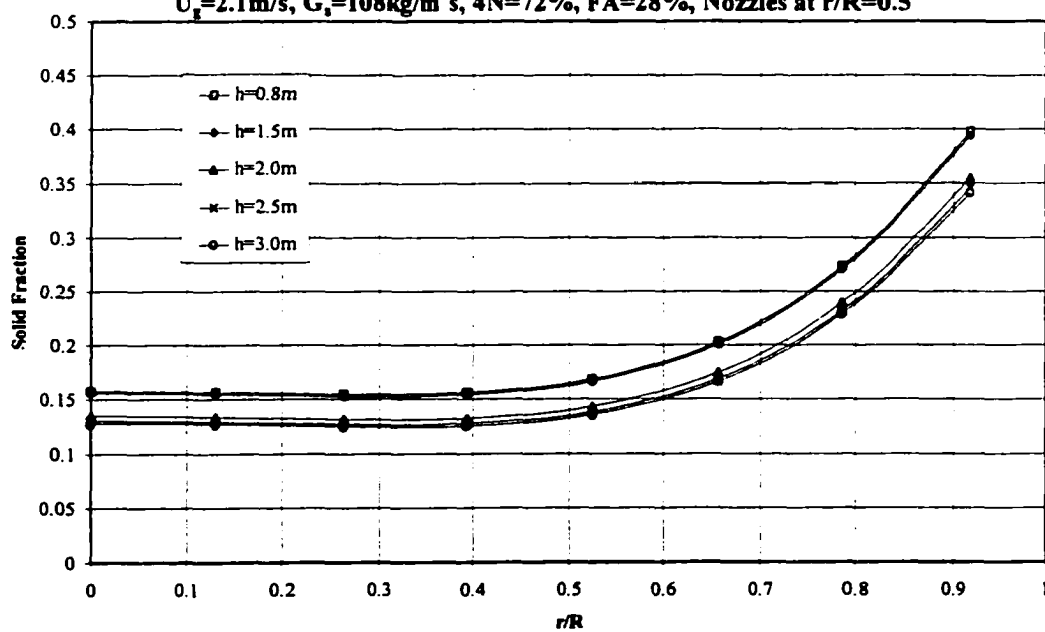
Figure B19: Radial Solid Density Profile $U_f=2.1\text{m/s}$, $G_s=108\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ **Figure B20: Radial Solid Profile** $U_f=2.1\text{m/s}$, $G_s=108\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ 

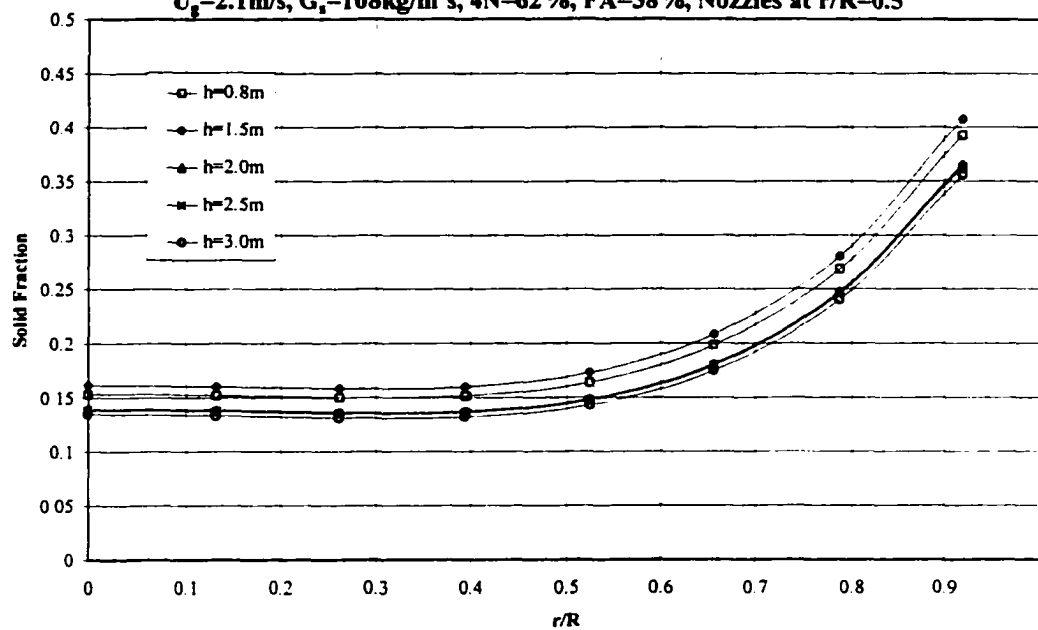
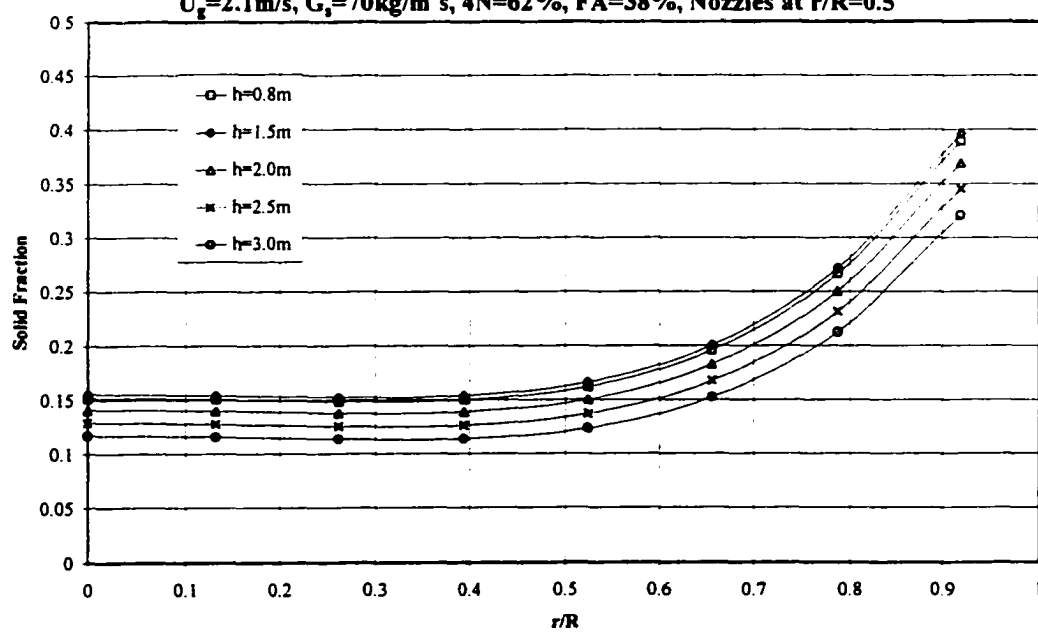
Figure B21: Radial Solid Profile $U_g=2.1\text{m/s}$, $G_s=108\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ **Figure B22: Radial Solid Profile** $U_g=2.1\text{m/s}$, $G_s=70\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ 

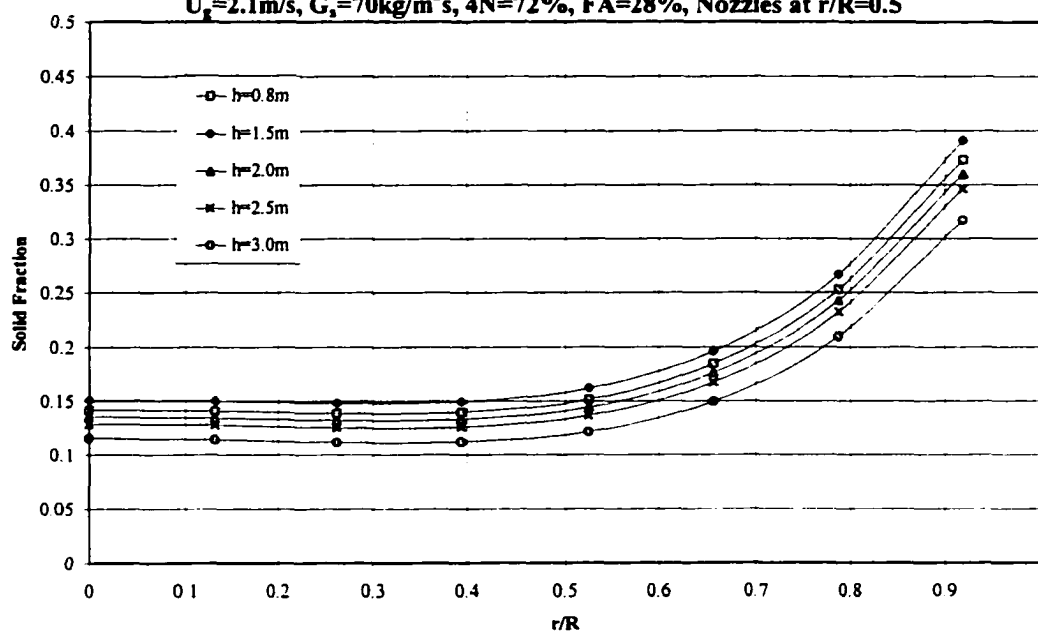
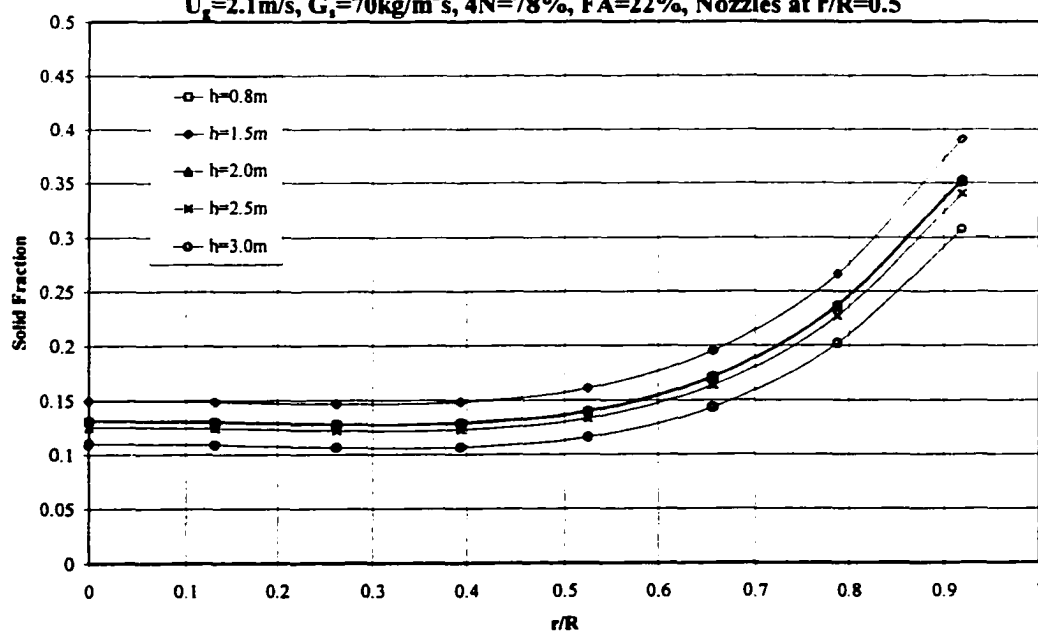
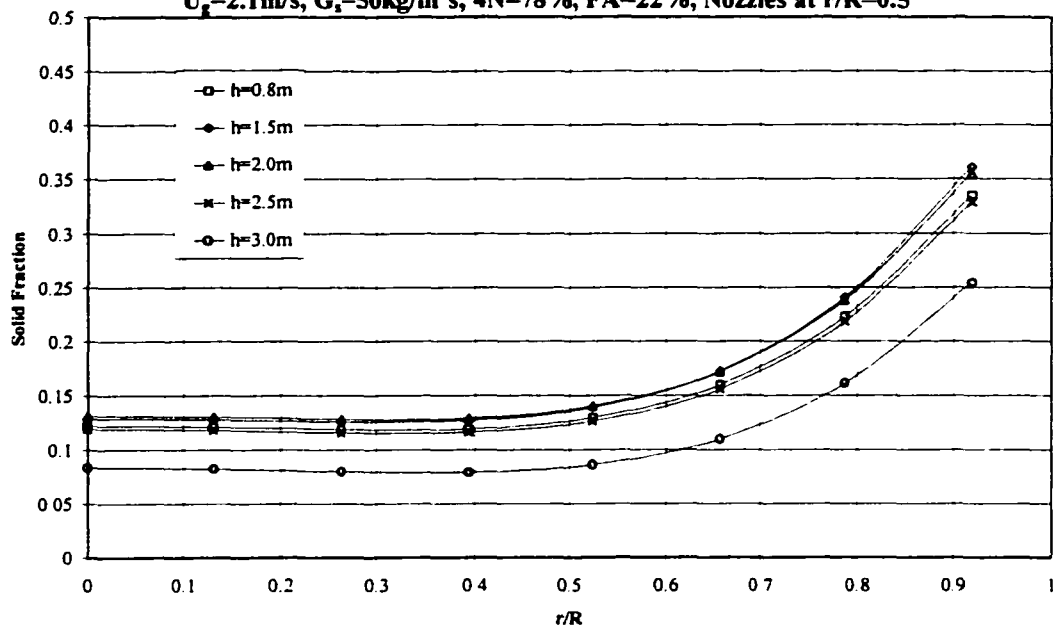
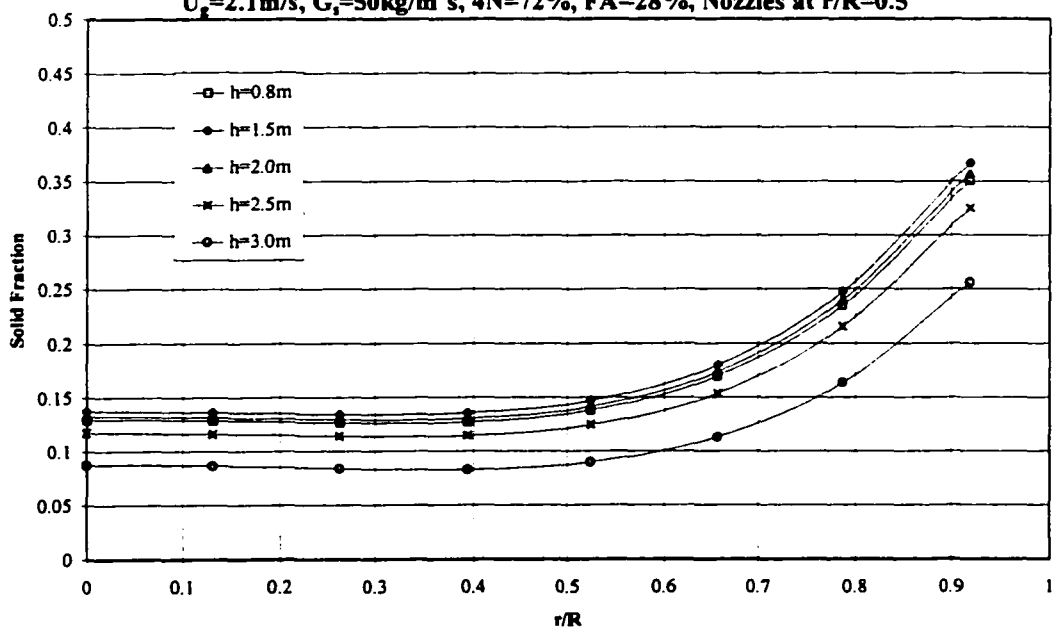
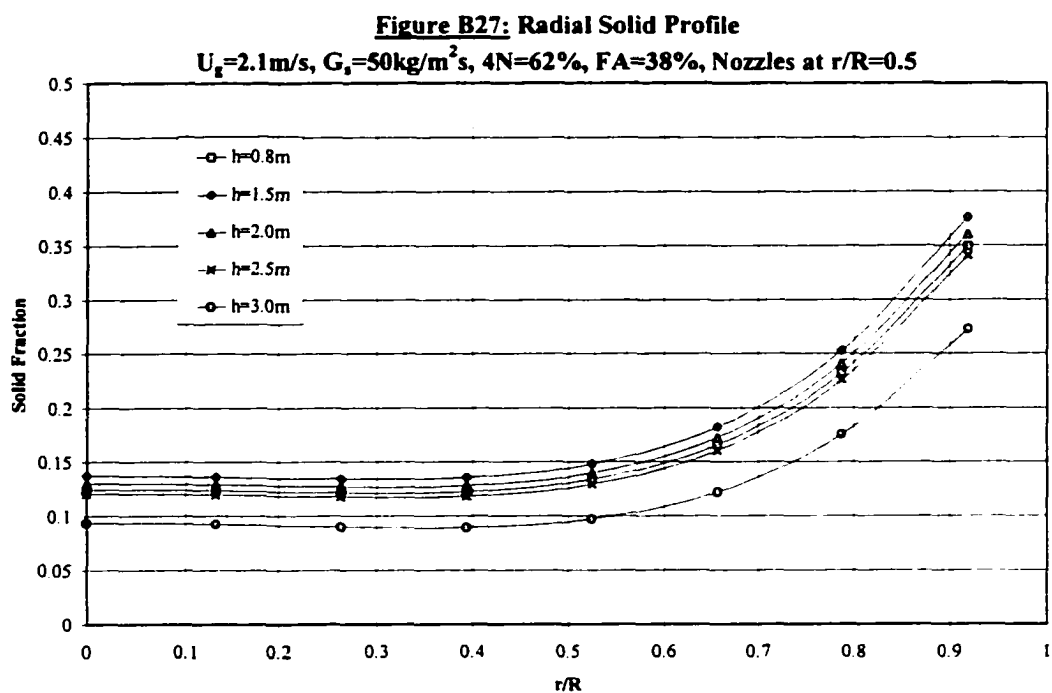
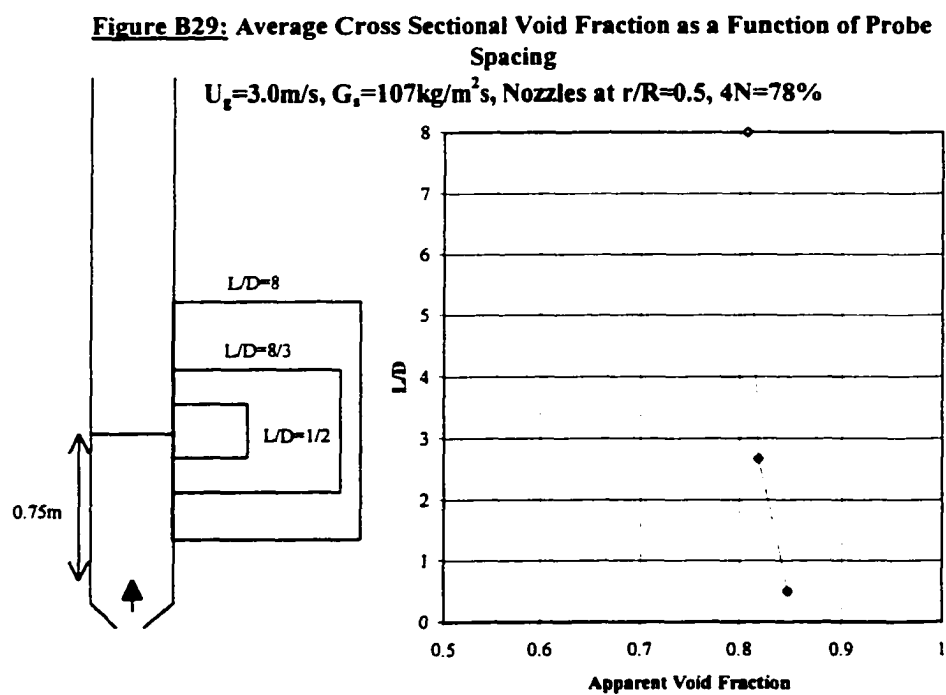
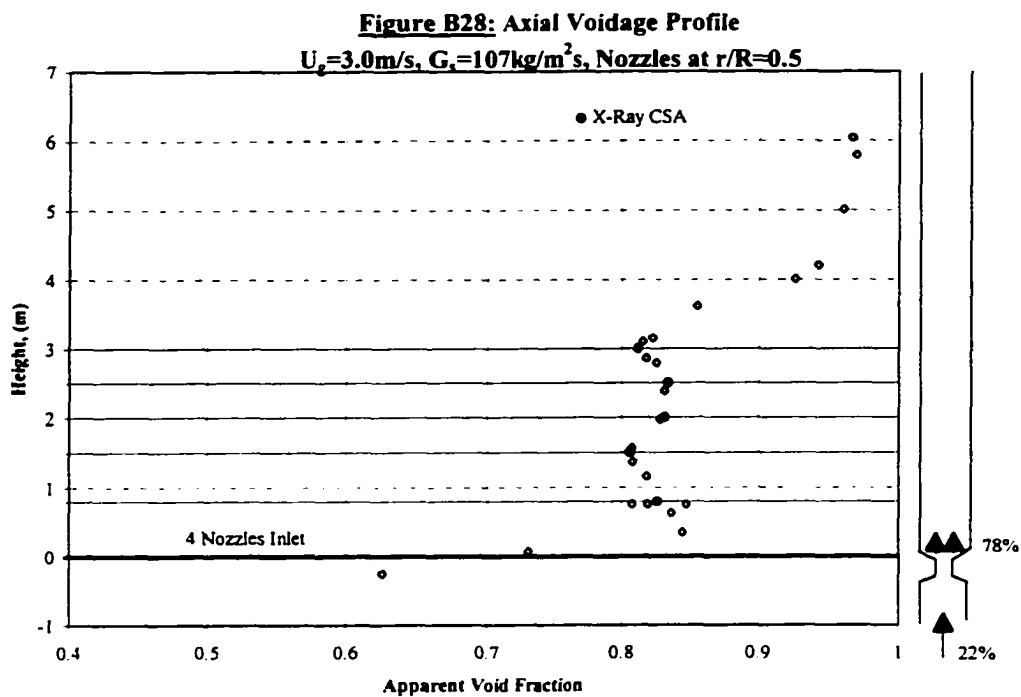
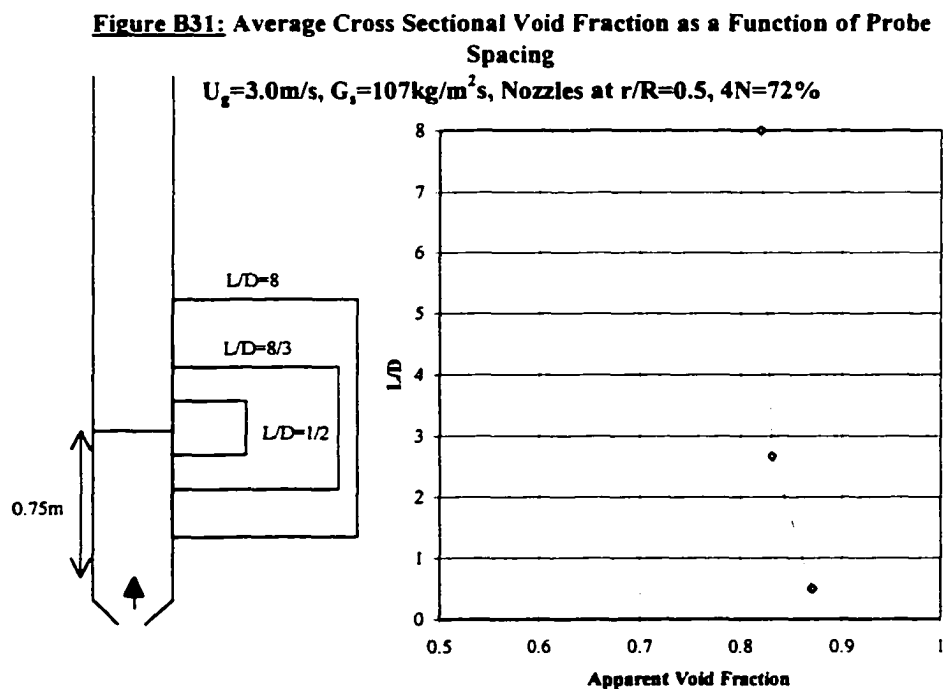
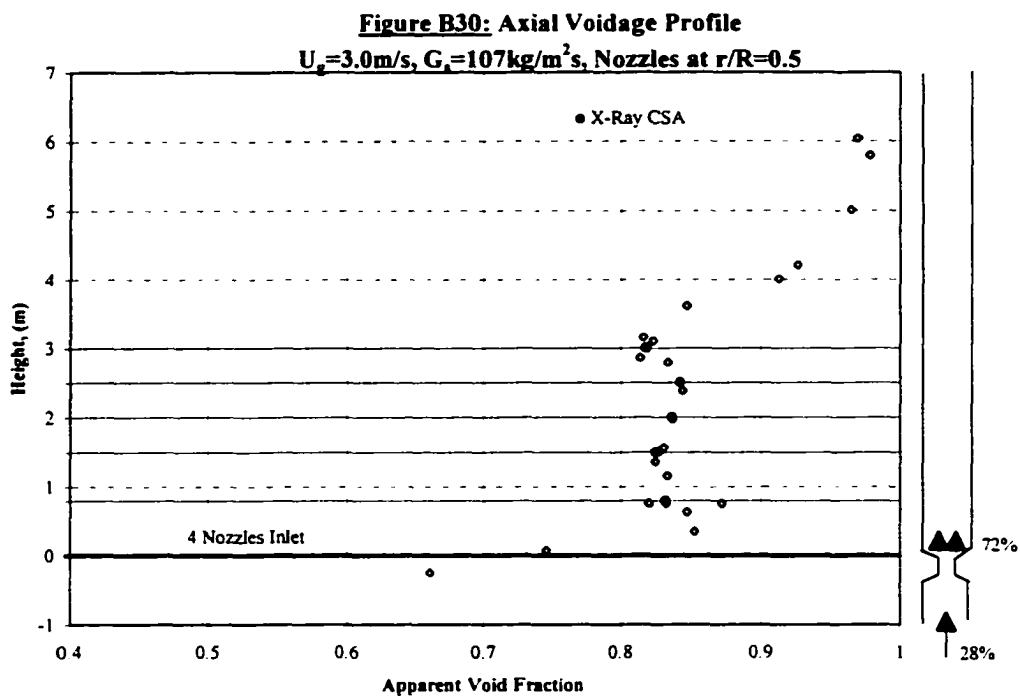
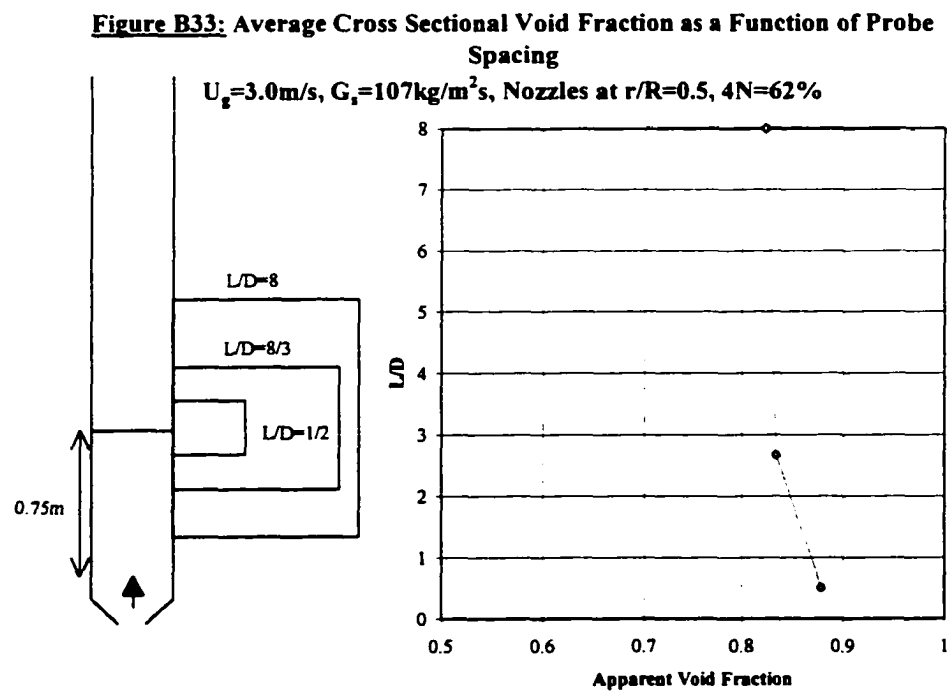
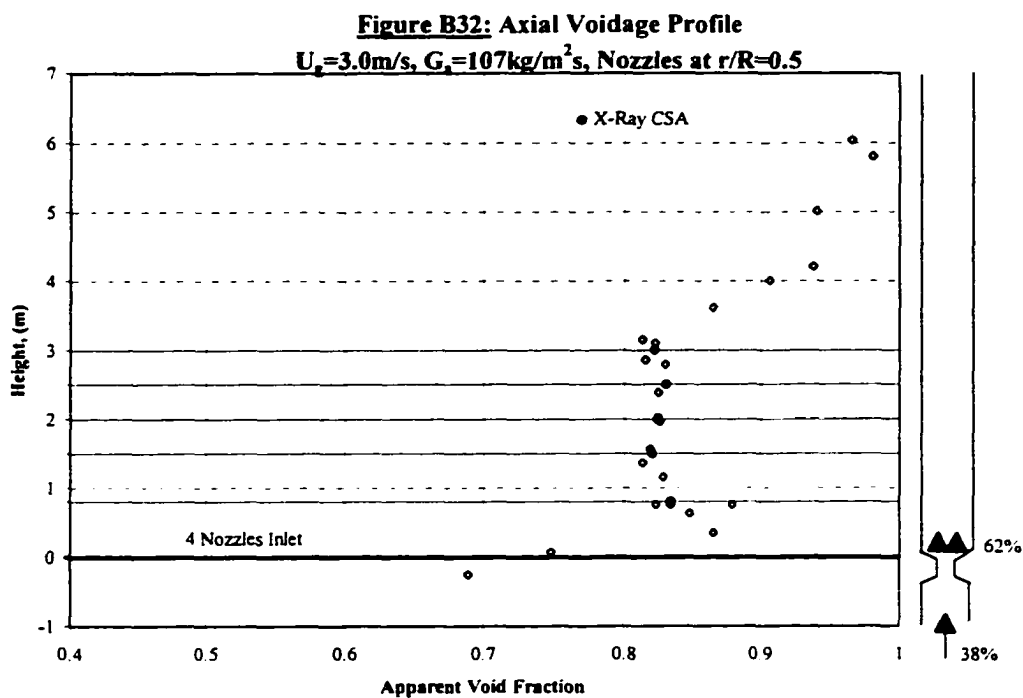
Figure 5.23: Radial Solid Profile $U_g=2.1\text{m/s}$, $G_s=70\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ **Figure B24: Radial Solid Density Profile** $U_g=2.1\text{m/s}$, $G_s=70\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ 

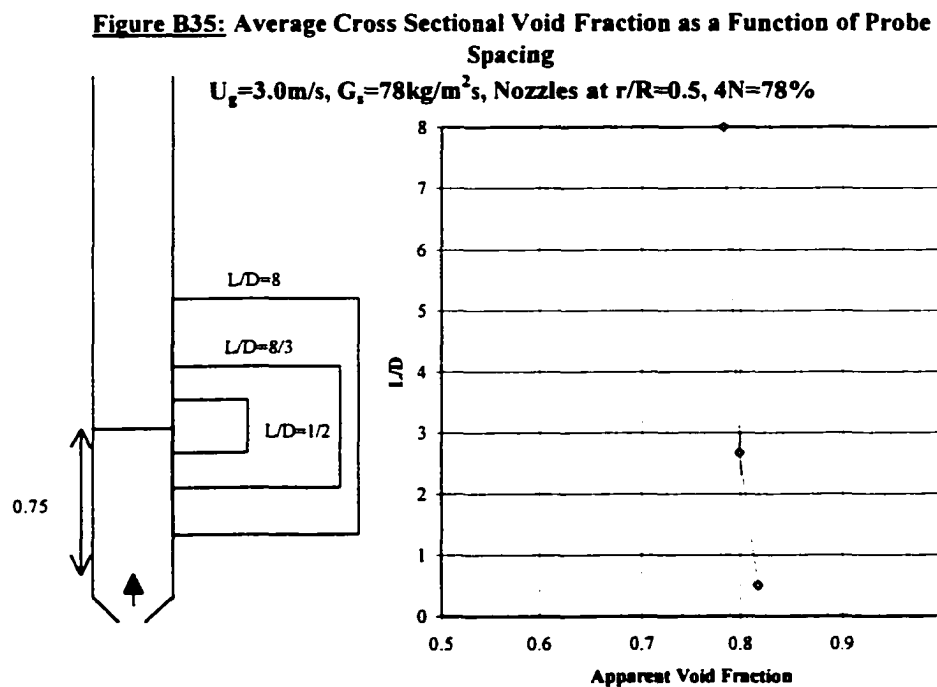
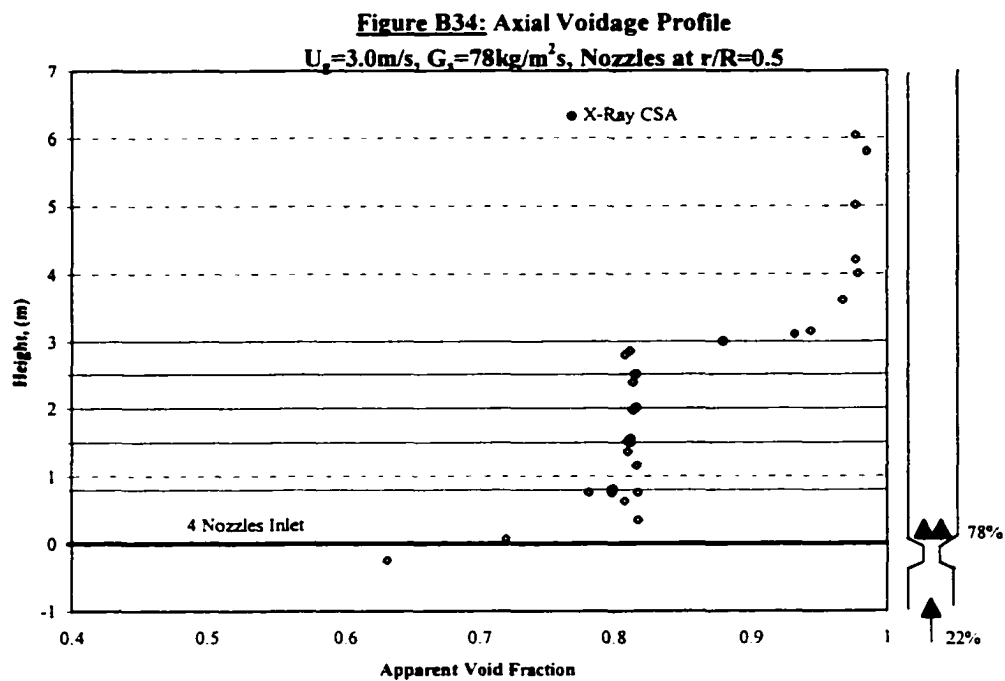
Figure B25: Radial Solid Profile $U_g=2.1\text{ m/s}$, $G_s=50\text{ kg/m}^2\text{ s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ **Figure B26: Radial Solid Profile** $U_g=2.1\text{ m/s}$, $G_s=50\text{ kg/m}^2\text{ s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ 











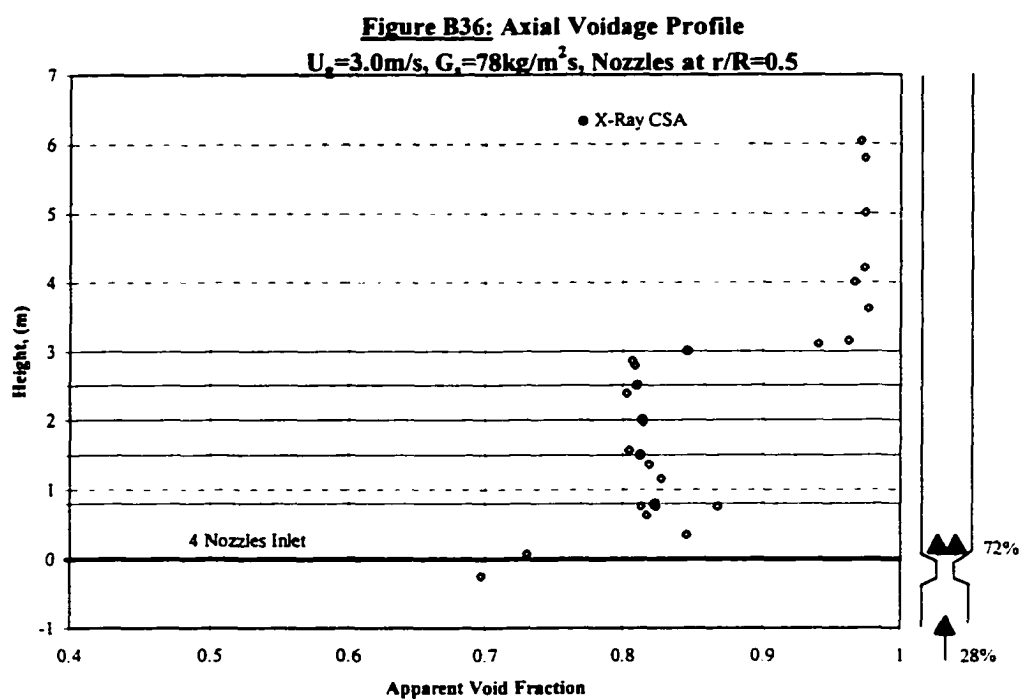
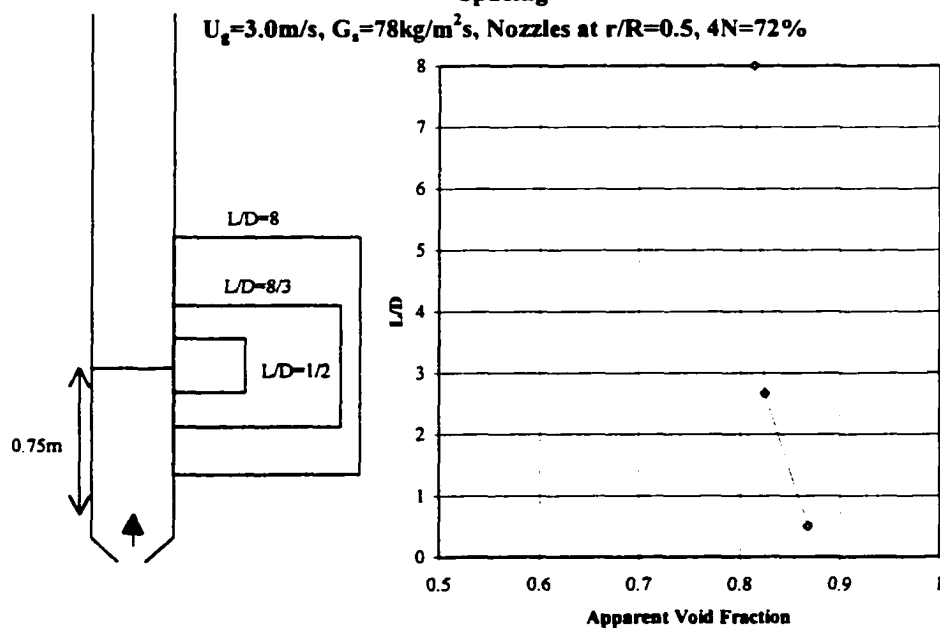


Figure B37: Average Cross Sectional Void Fraction as a Function of Probe Spacing



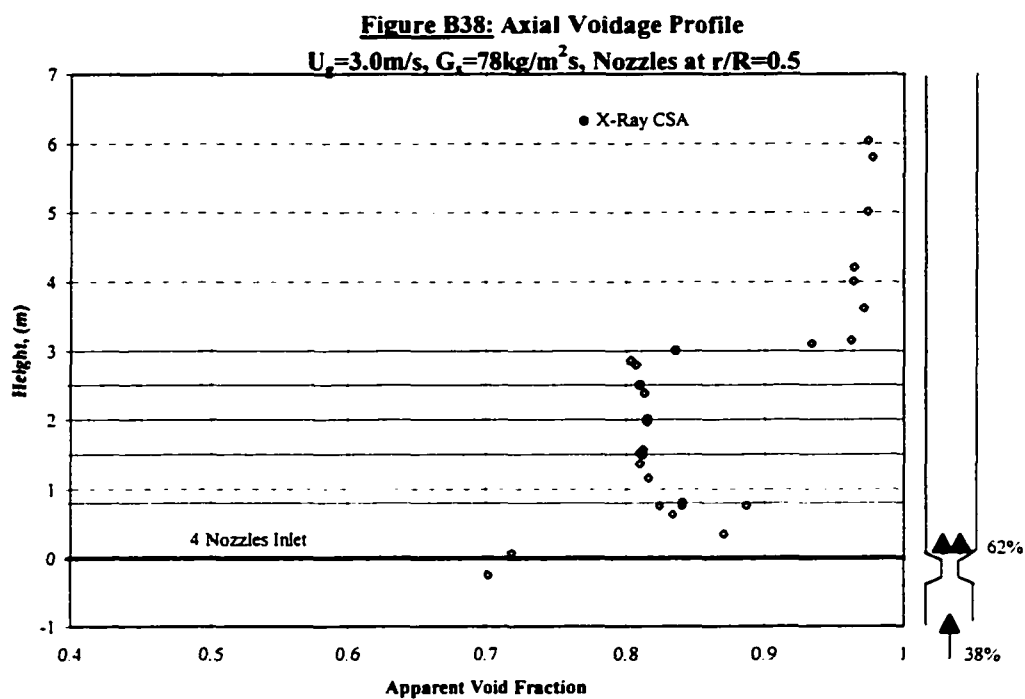
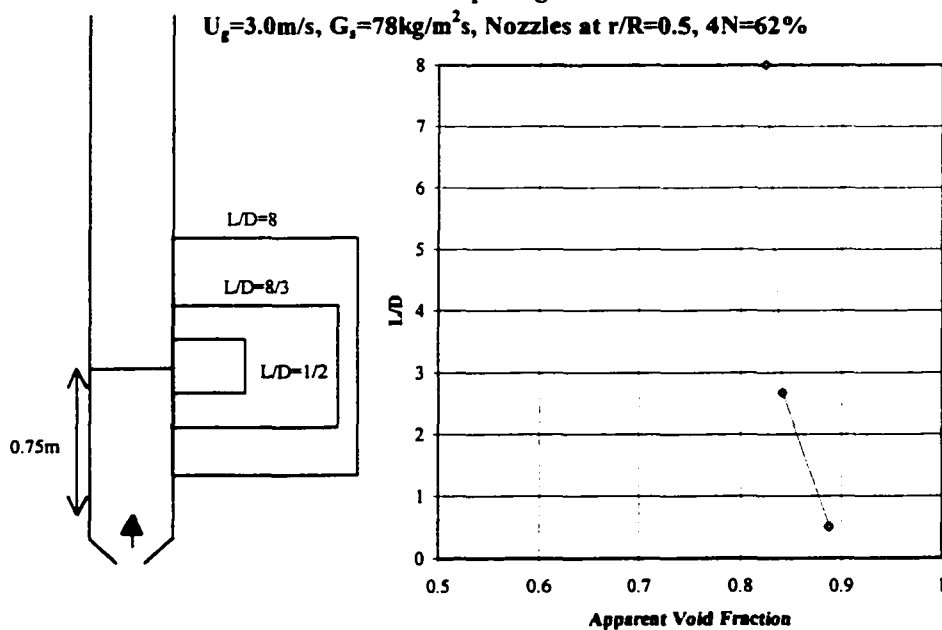
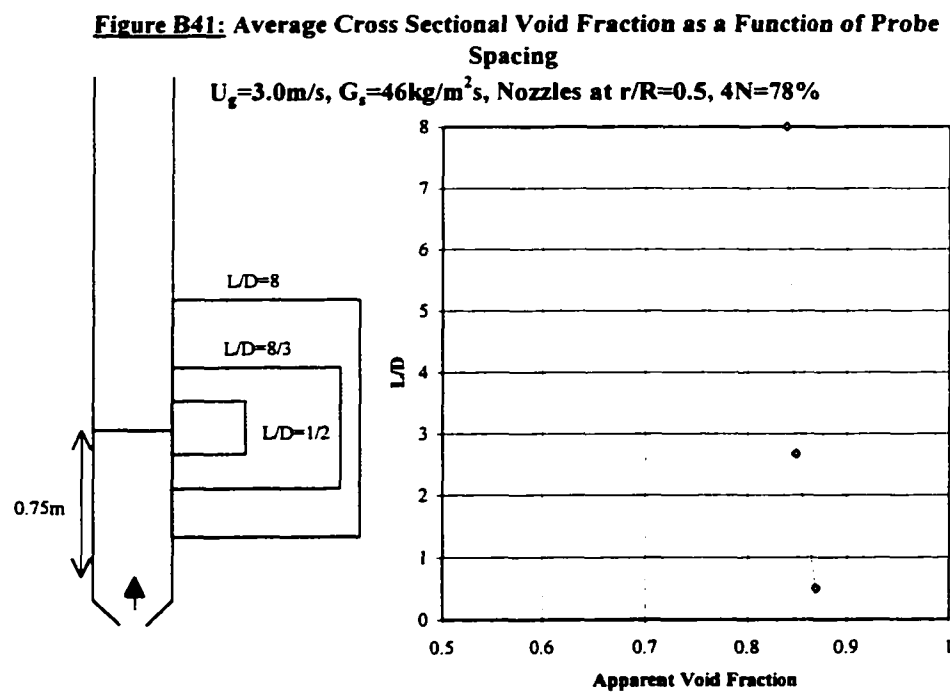
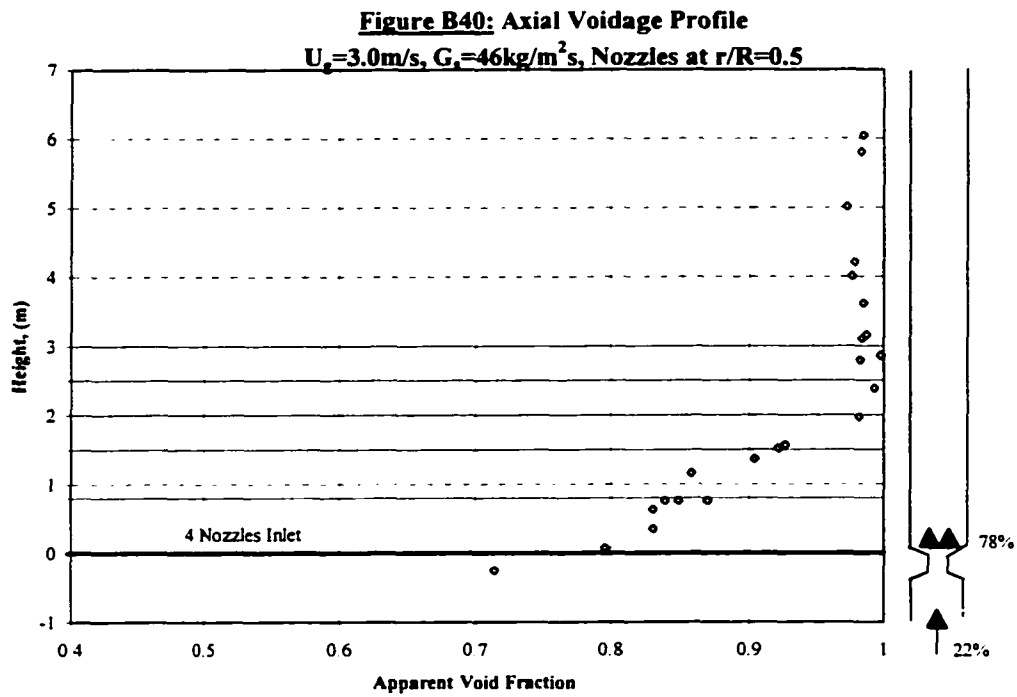
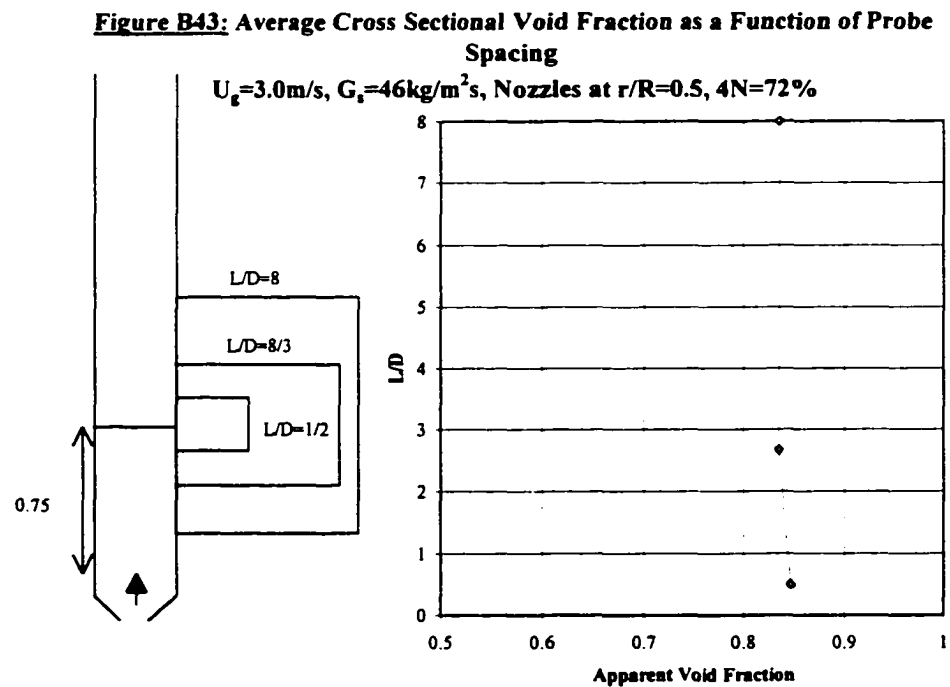
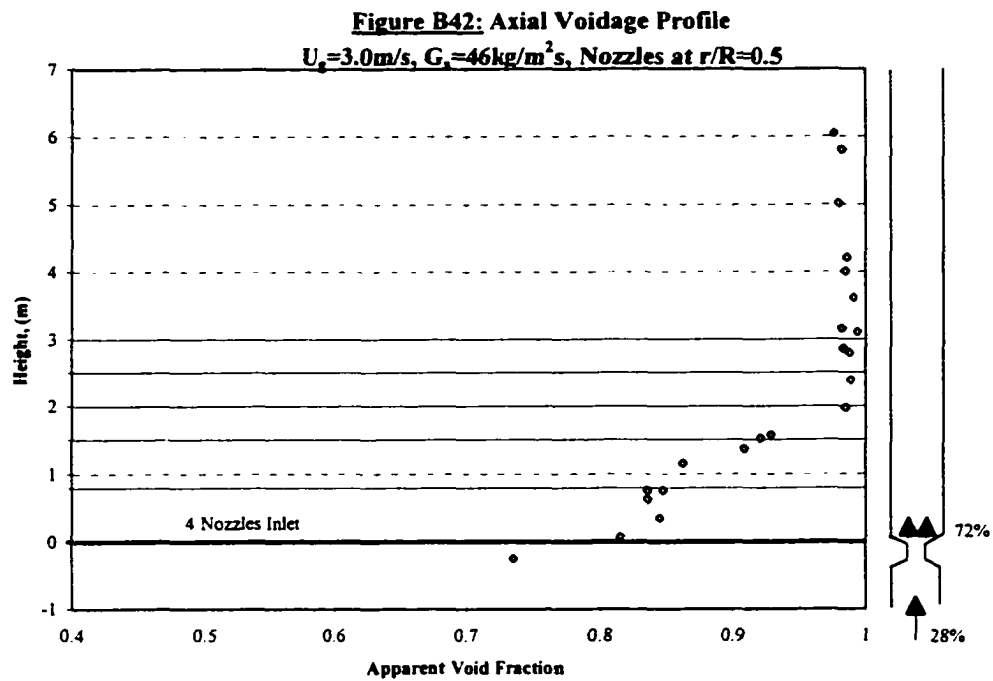


Figure B39: Average Cross Sectional Void Fraction as a Function of Probe Spacing







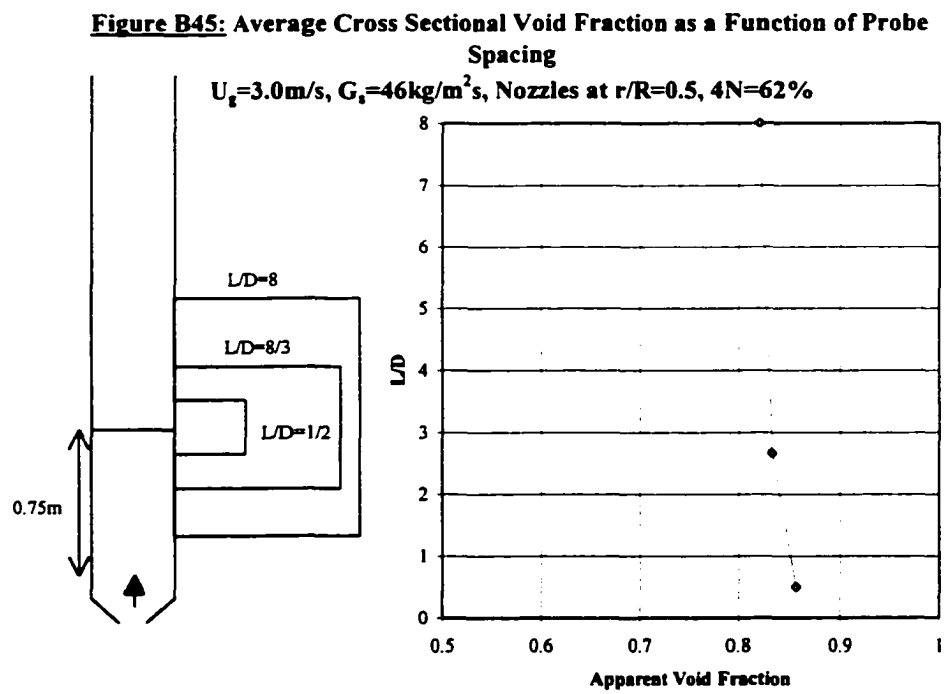
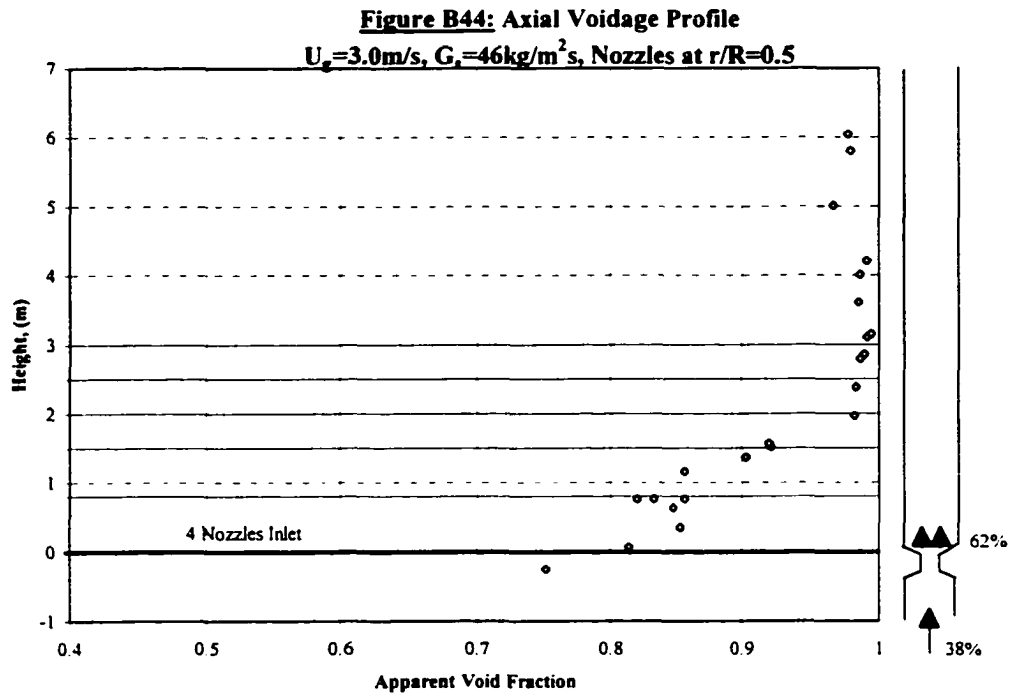


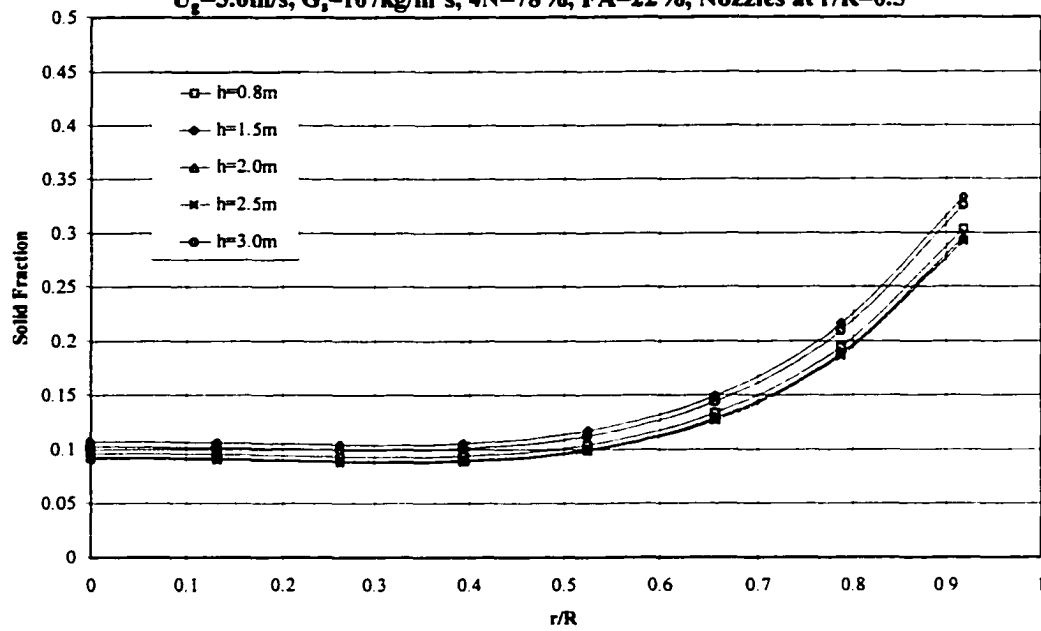
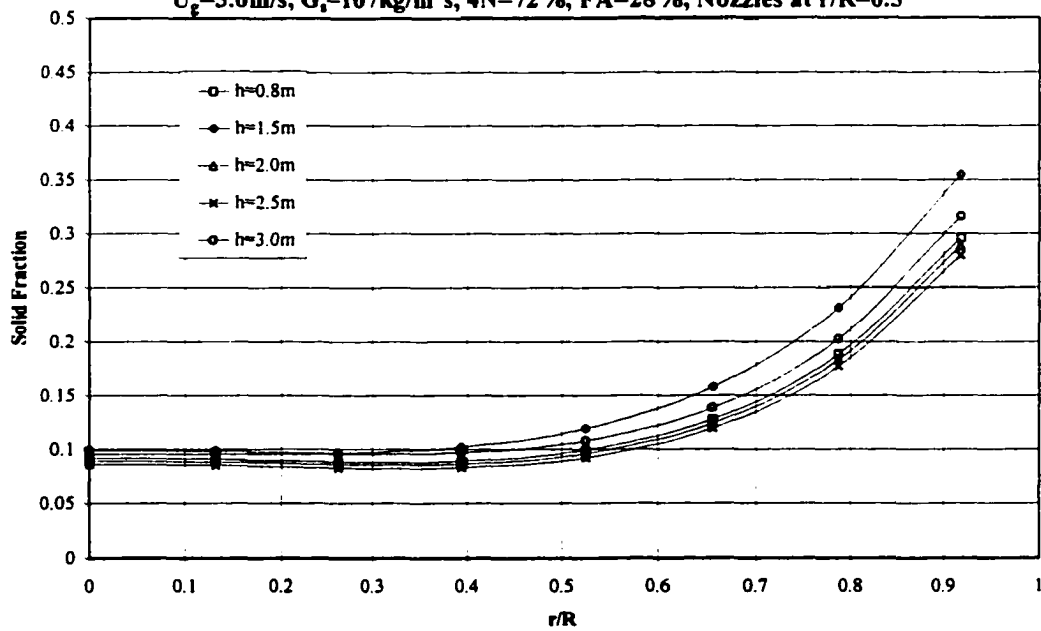
Figure B46: Radial Solid Density Profile $U_t=3.0\text{m/s}$, $G_t=107\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ **Figure B47: Radial Solid Profile** $U_t=3.0\text{m/s}$, $G_t=107\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ 

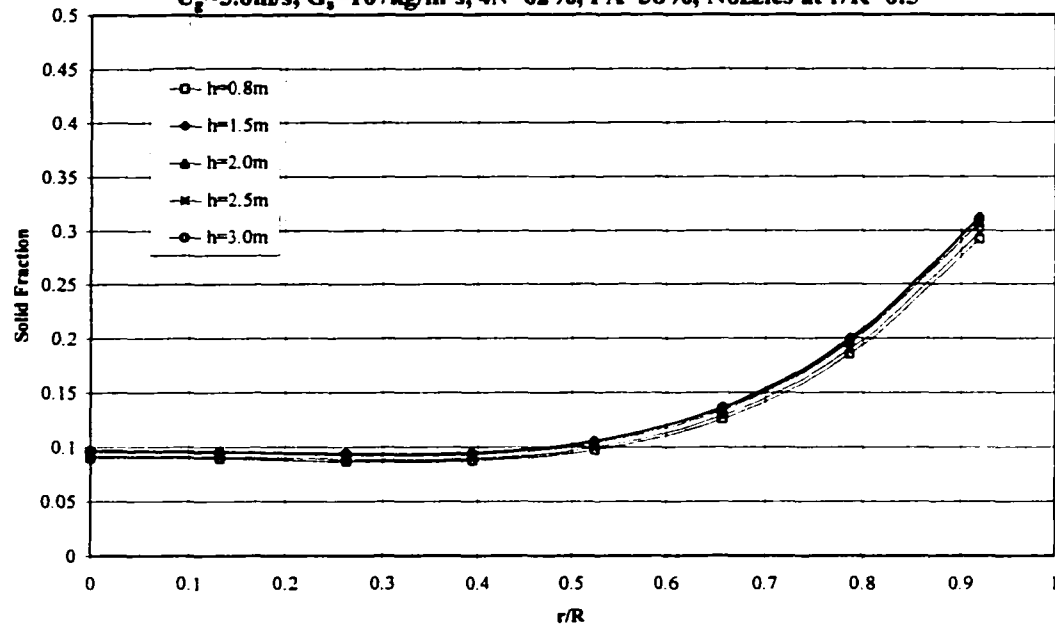
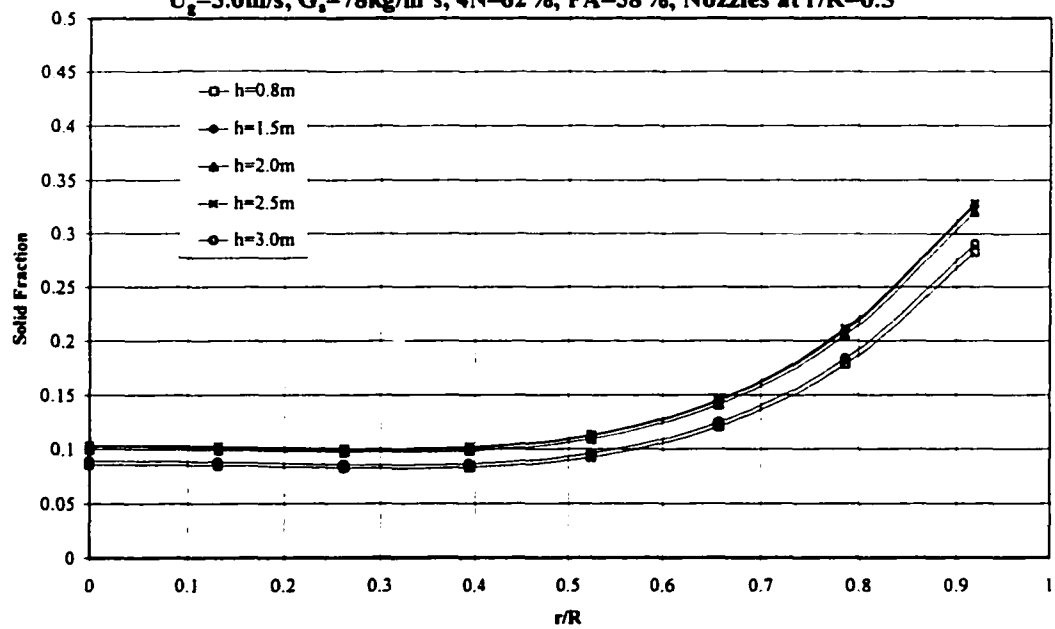
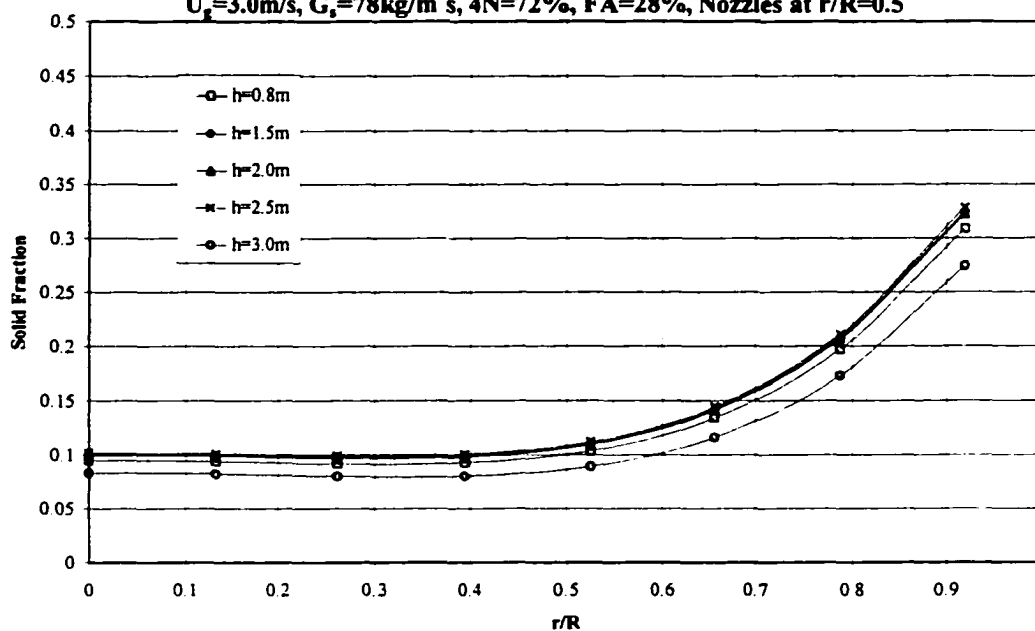
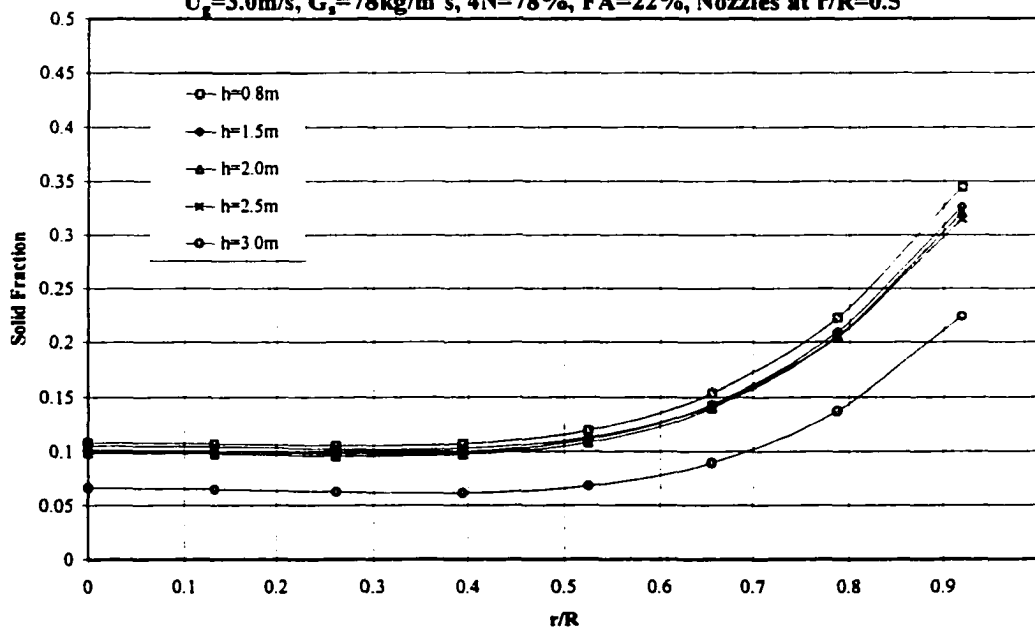
Figure B48: Radial Solid Profile $U_g=3.0\text{m/s}$, $G_s=107\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ **Figure B49: Radial Solid Profile** $U_g=3.0\text{m/s}$, $G_s=78\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ 

Figure B50: Radial Solid Profile $U_g=3.0\text{m/s}$, $G_s=78\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ **Figure B51: Radial Solid Density Profile** $U_g=3.0\text{m/s}$, $G_s=78\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ 

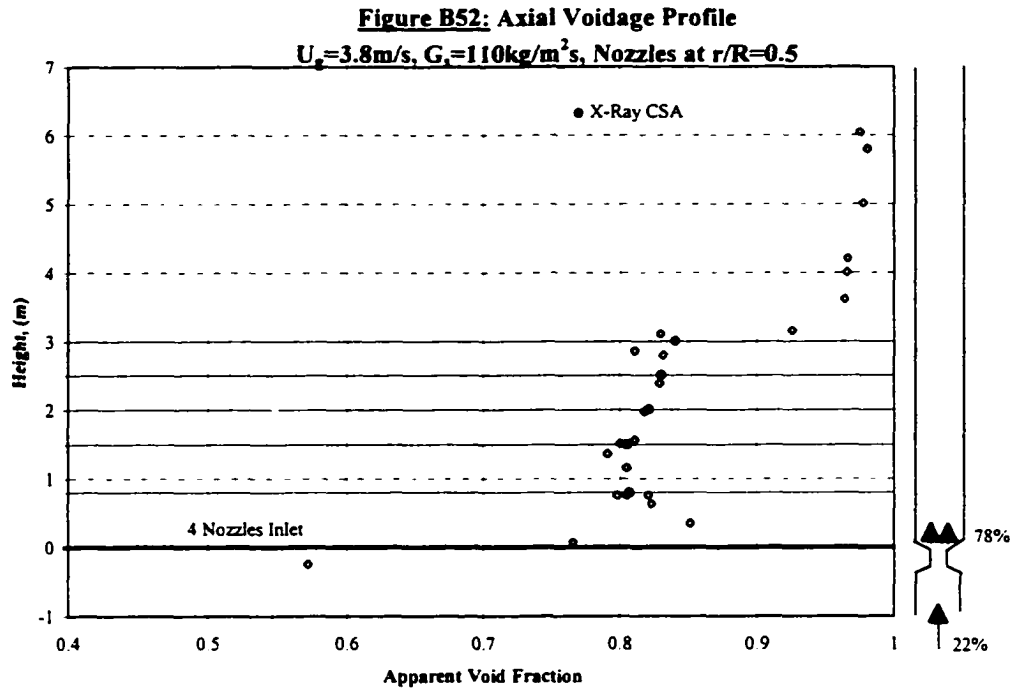
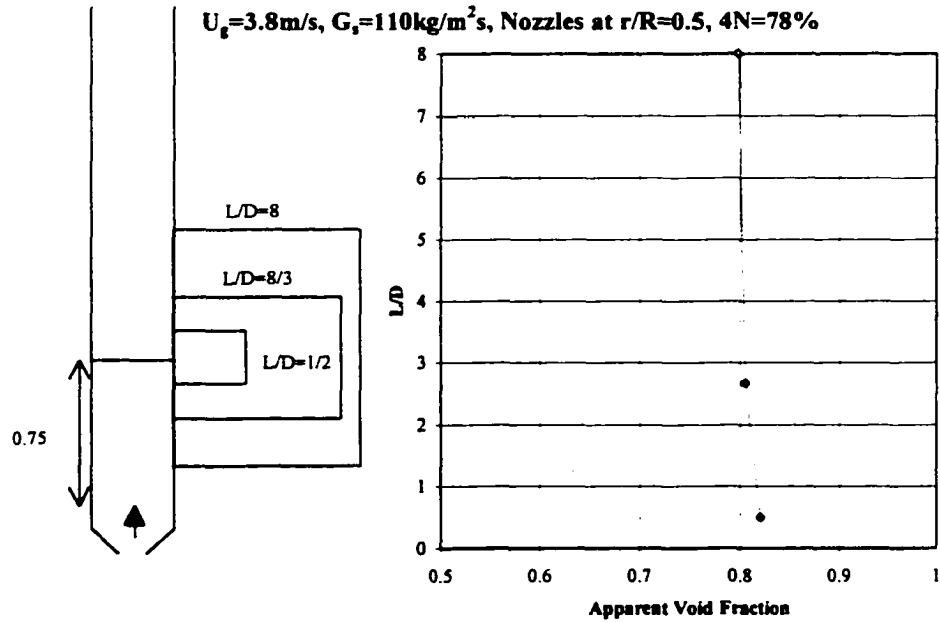
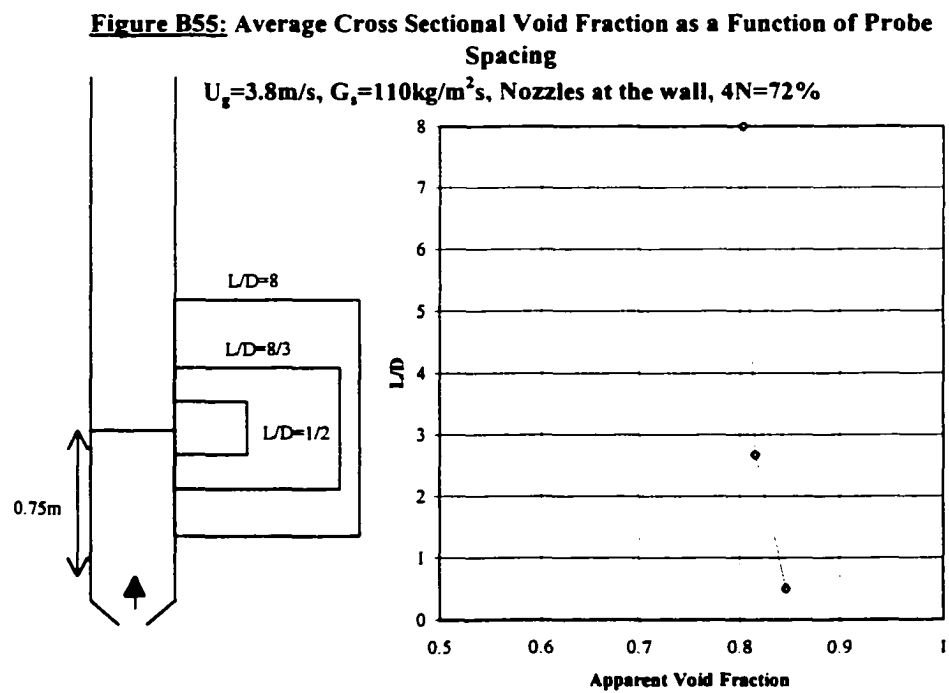
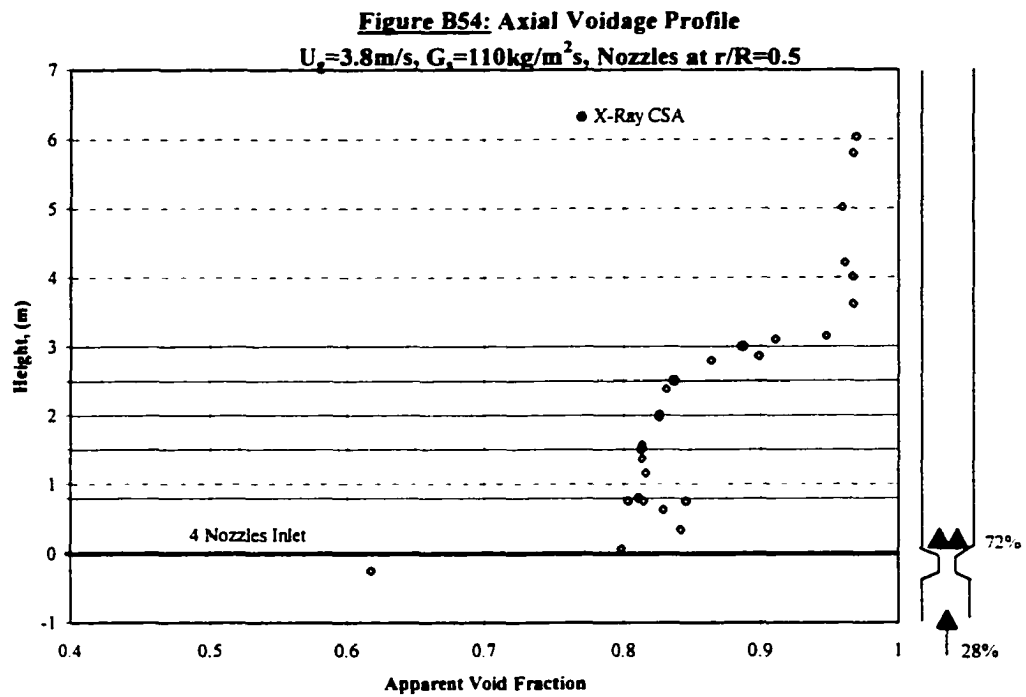


Figure B53: Average Cross Sectional Void Fraction as a Function of Probe Spacing





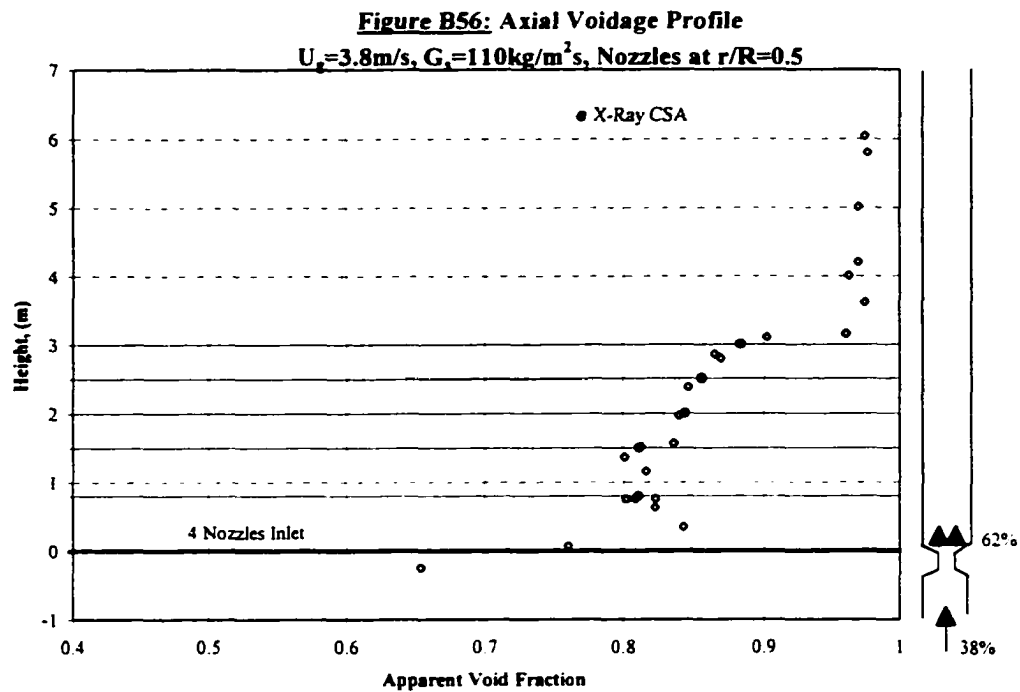
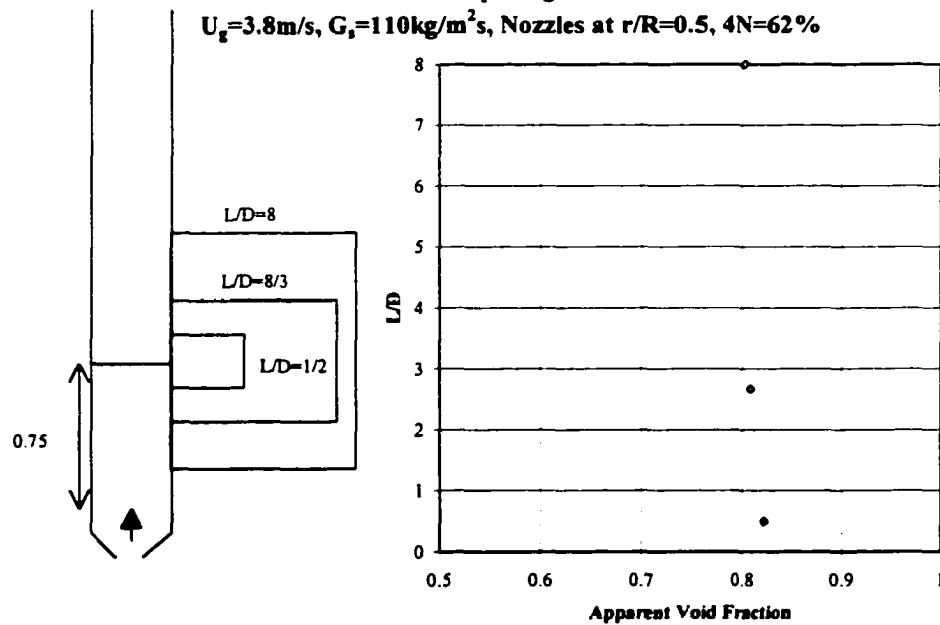
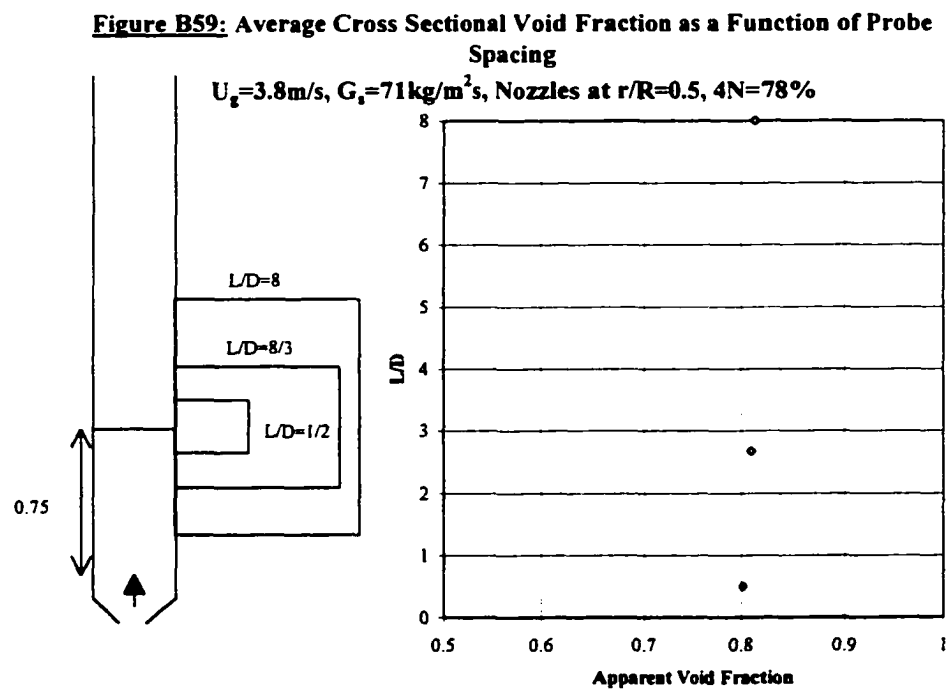
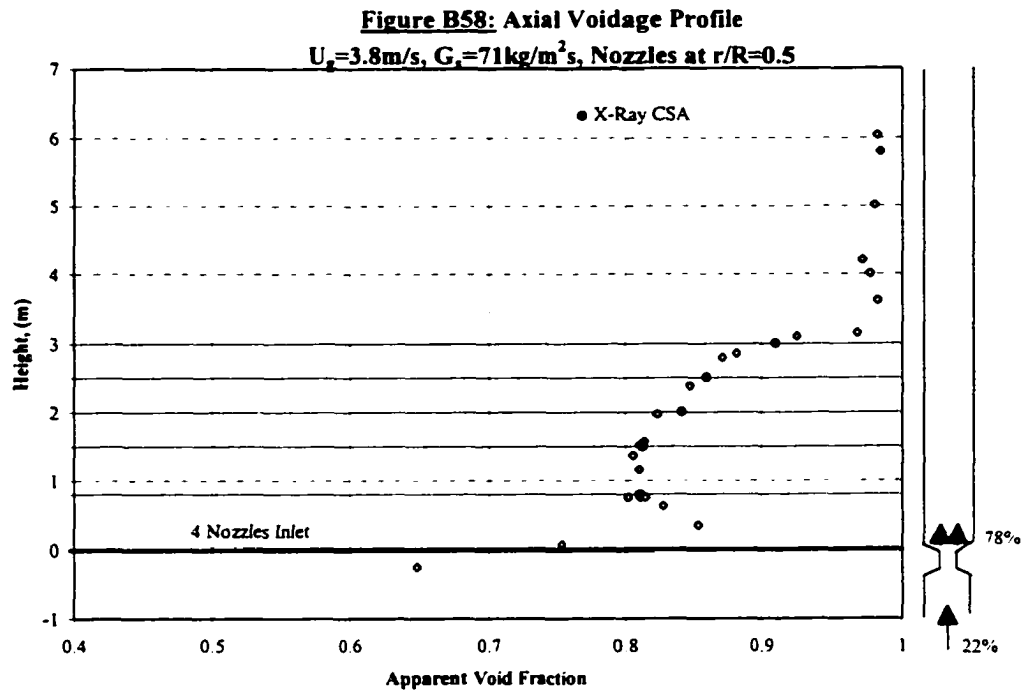


Figure B57: Average Cross Sectional Void Fraction as a Function of Probe Spacing





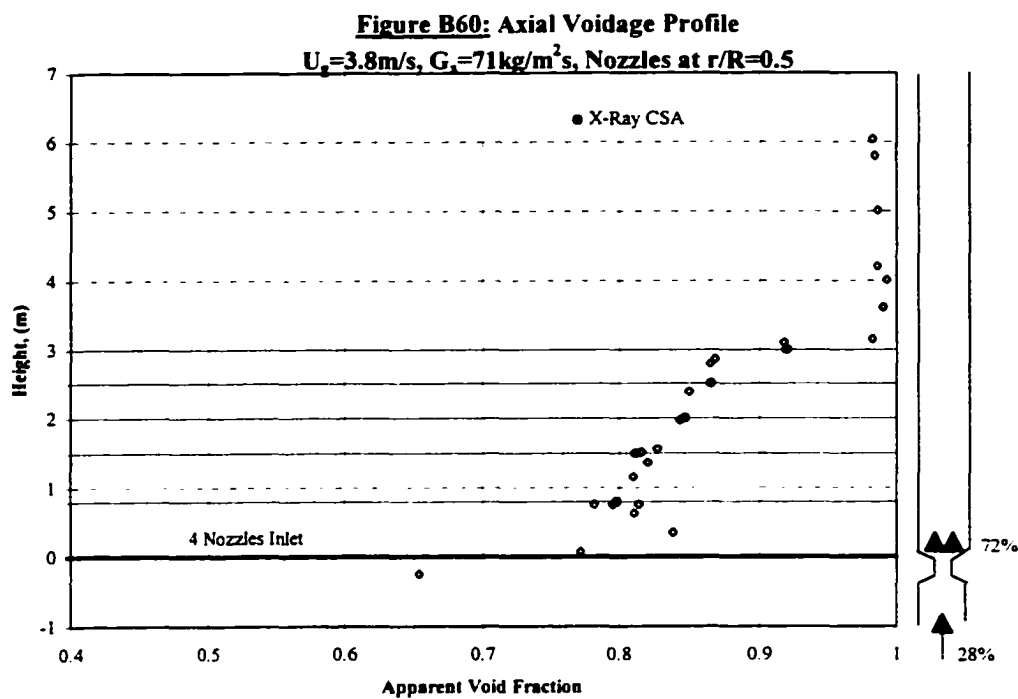
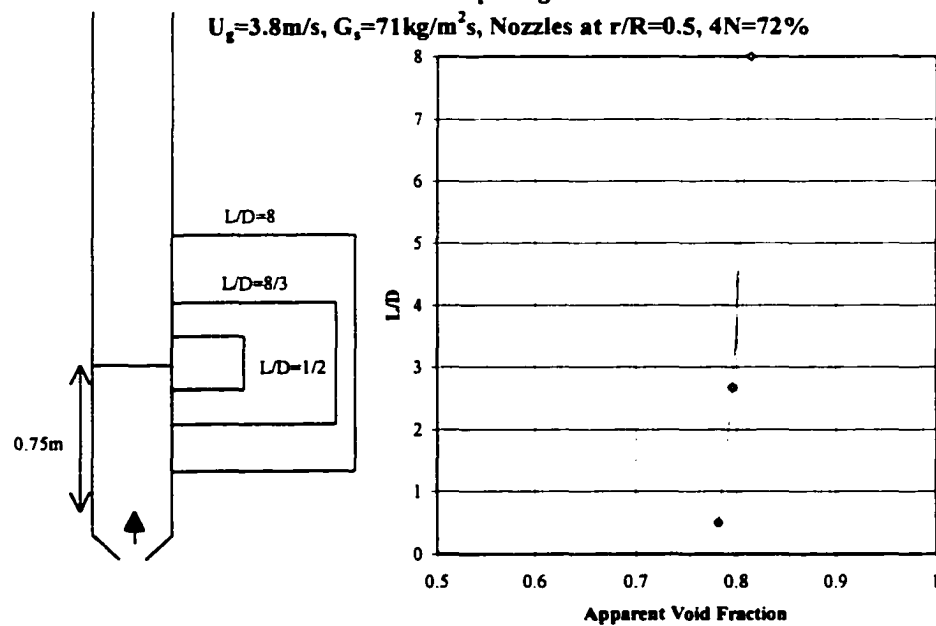
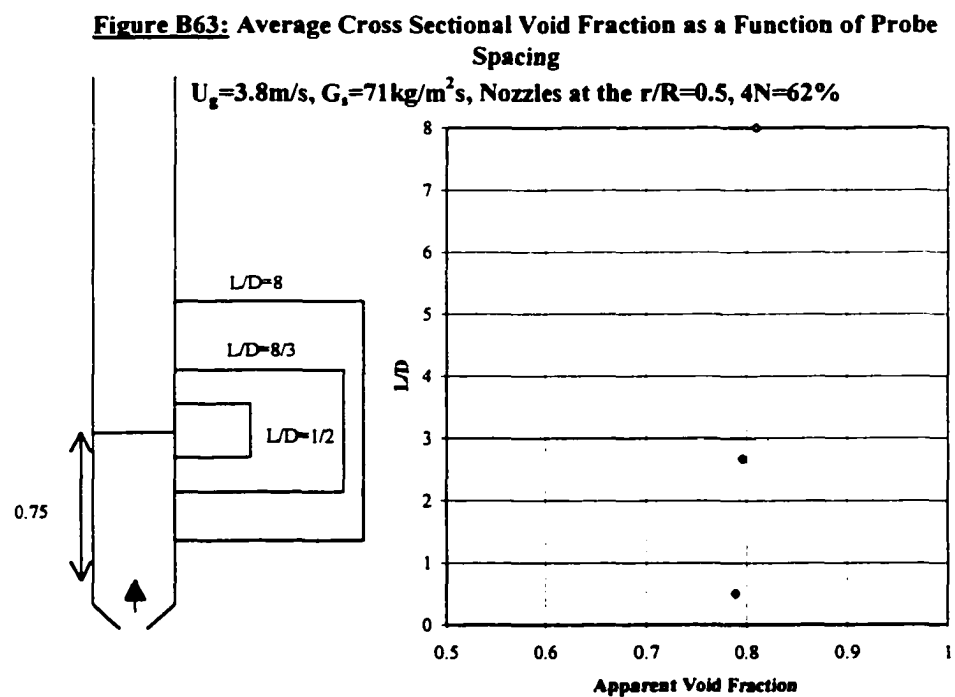
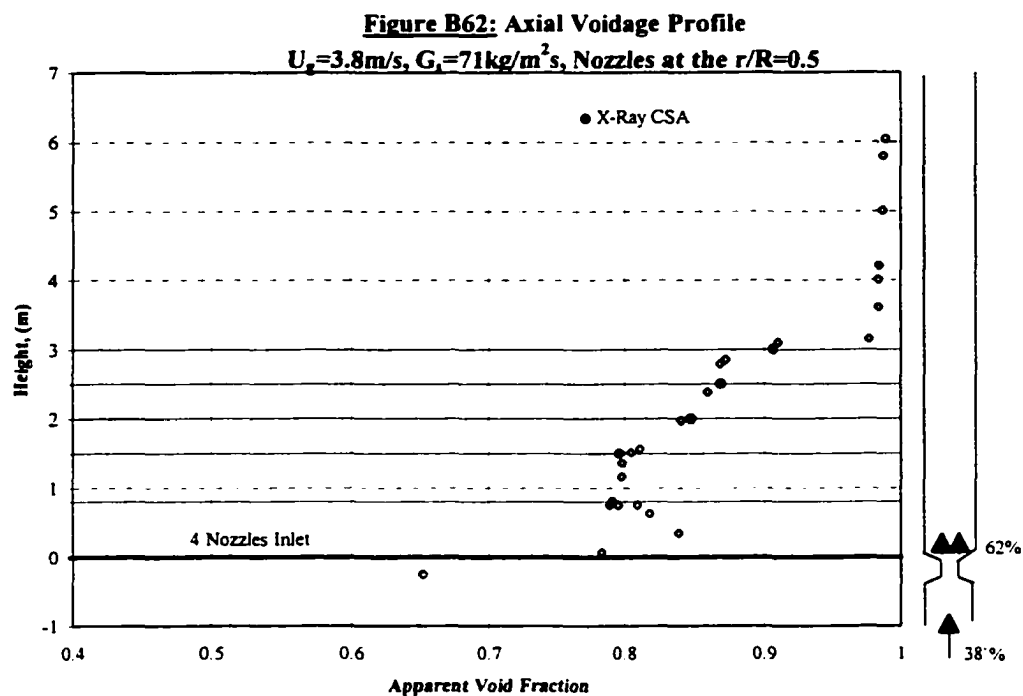


Figure B61: Average Cross Sectional Void Fraction as a Function of Probe Spacing





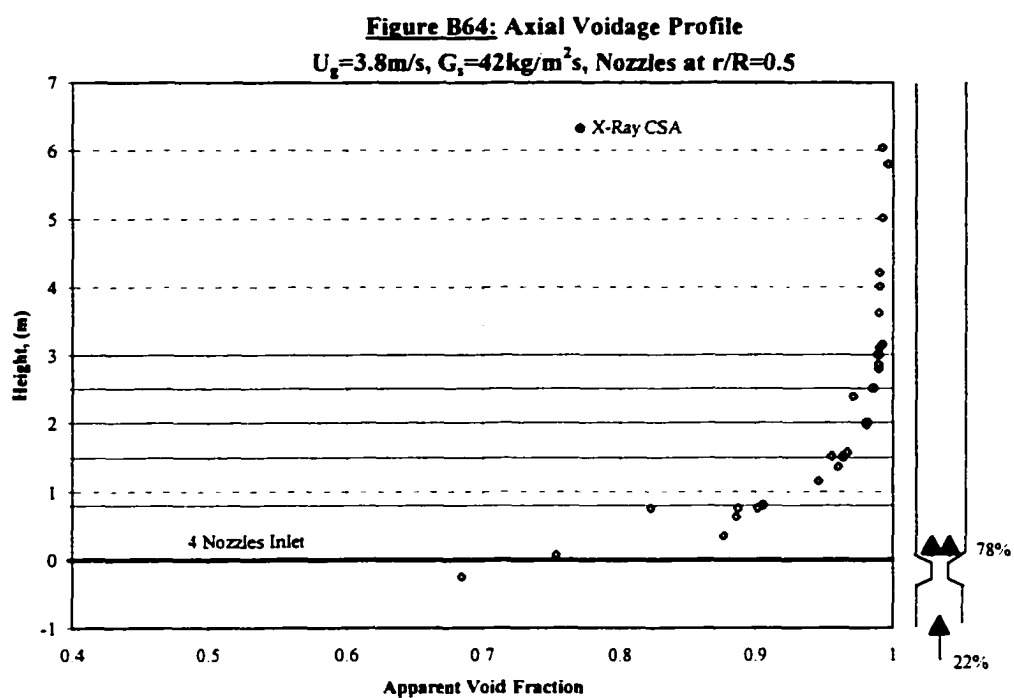
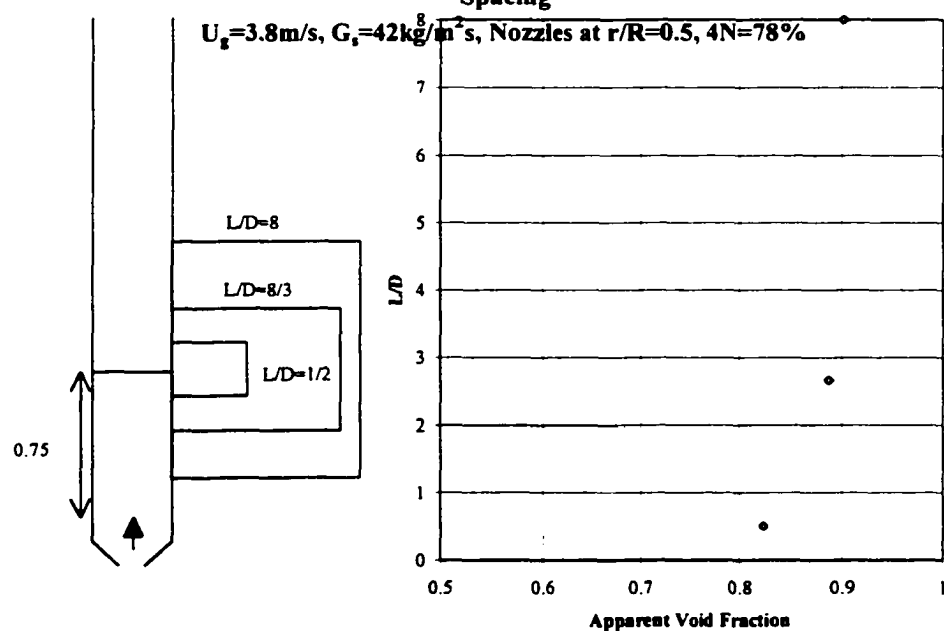


Figure B65: Average Cross Sectional Void Fraction as a Function of Probe Spacing



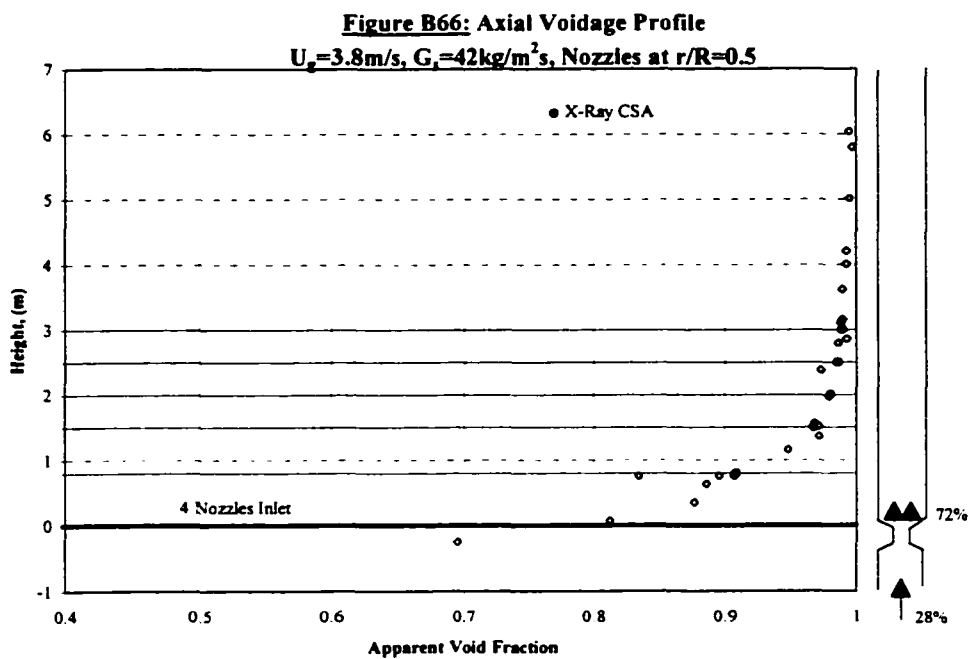
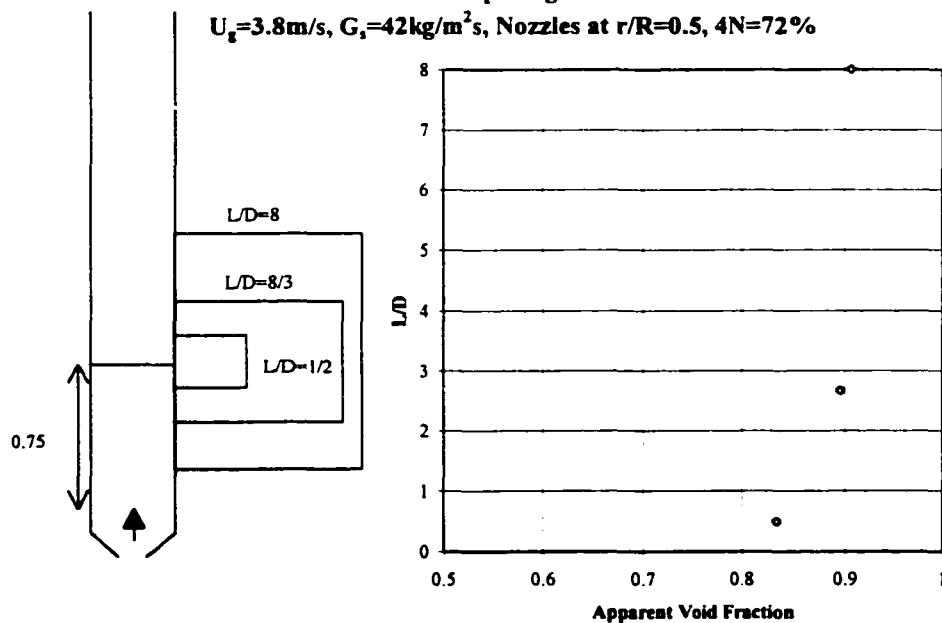


Figure B67: Average Cross Sectional Void Fraction as a Function of Probe Spacing



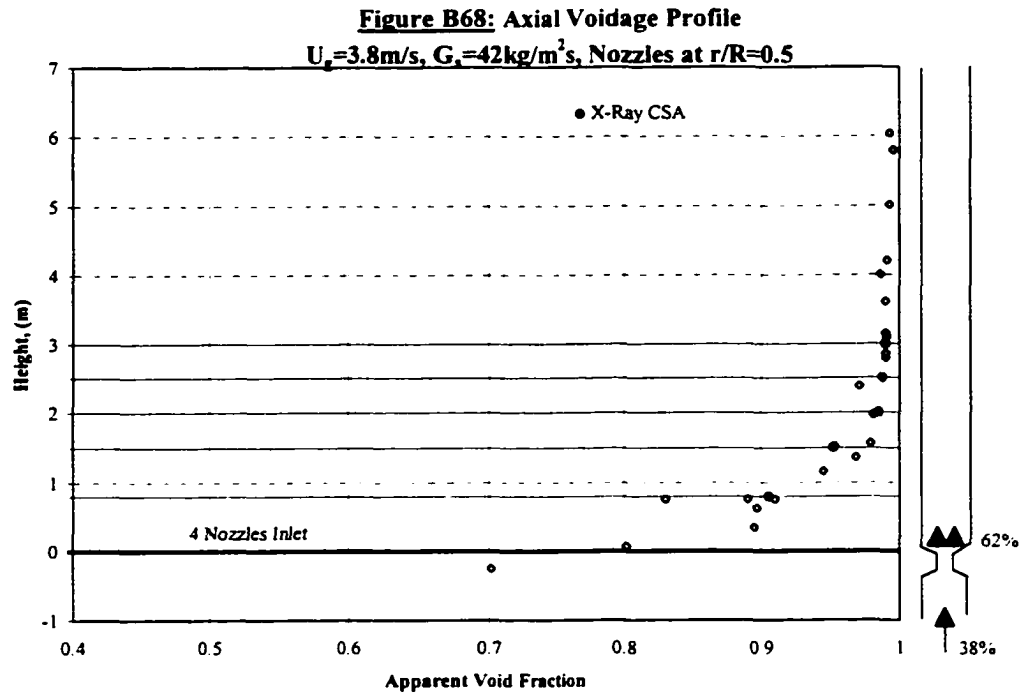


Figure B69: Average Cross Sectional Void Fraction as a Function of Probe Spacing

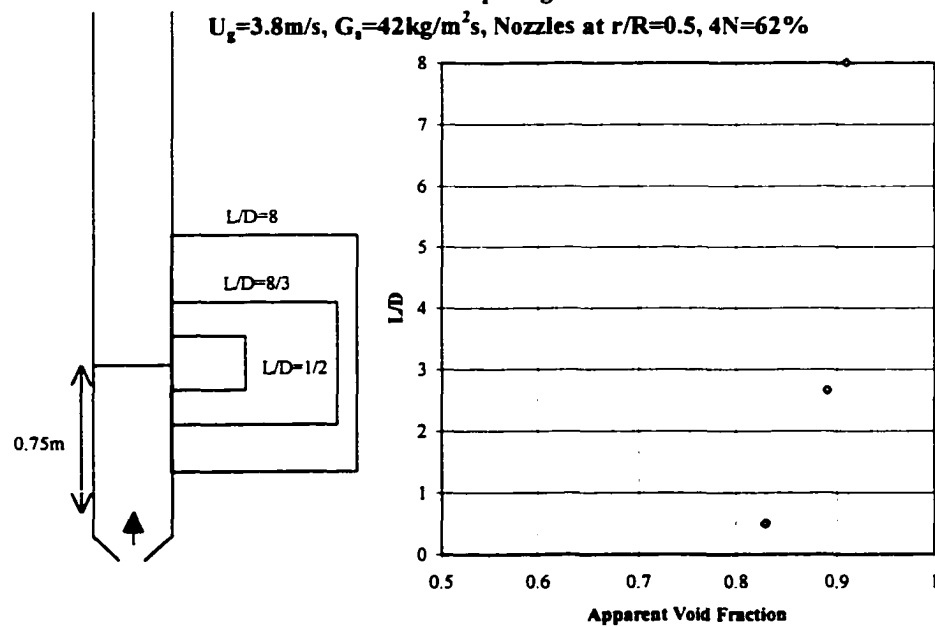


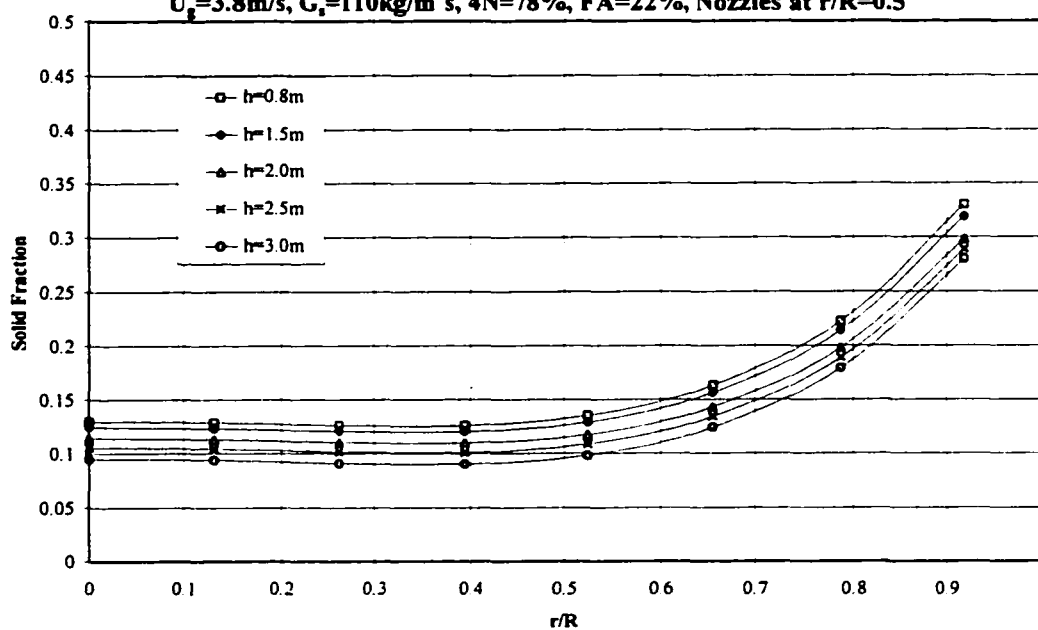
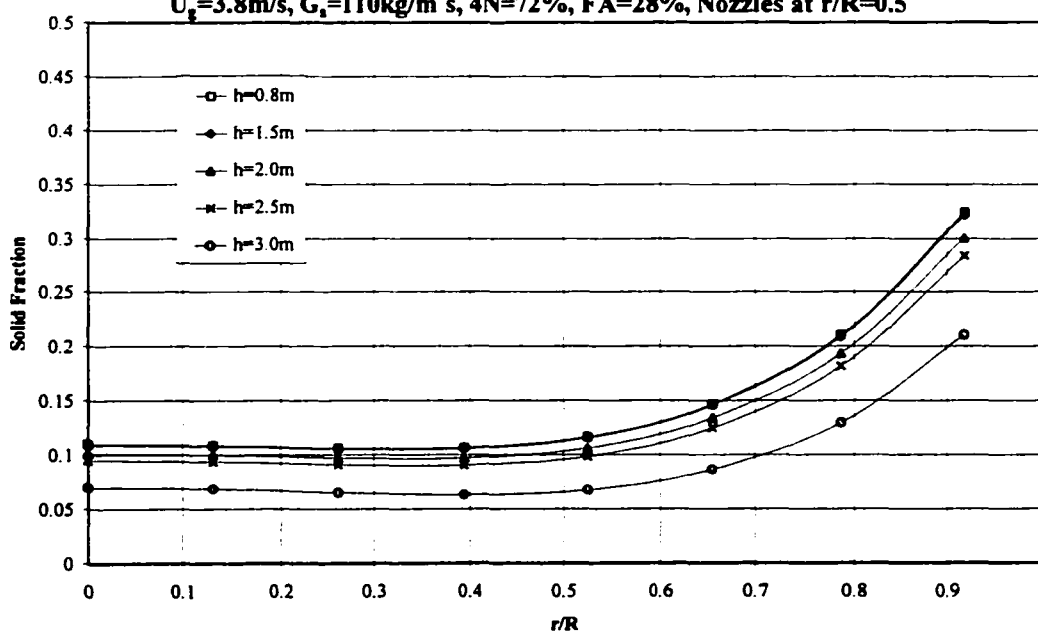
Figure B70: Radial Solid Density Profile $U_t=3.8\text{m/s}$, $G_t=110\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ **Figure B71: Radial Solid Profile** $U_t=3.8\text{m/s}$, $G_t=110\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ 

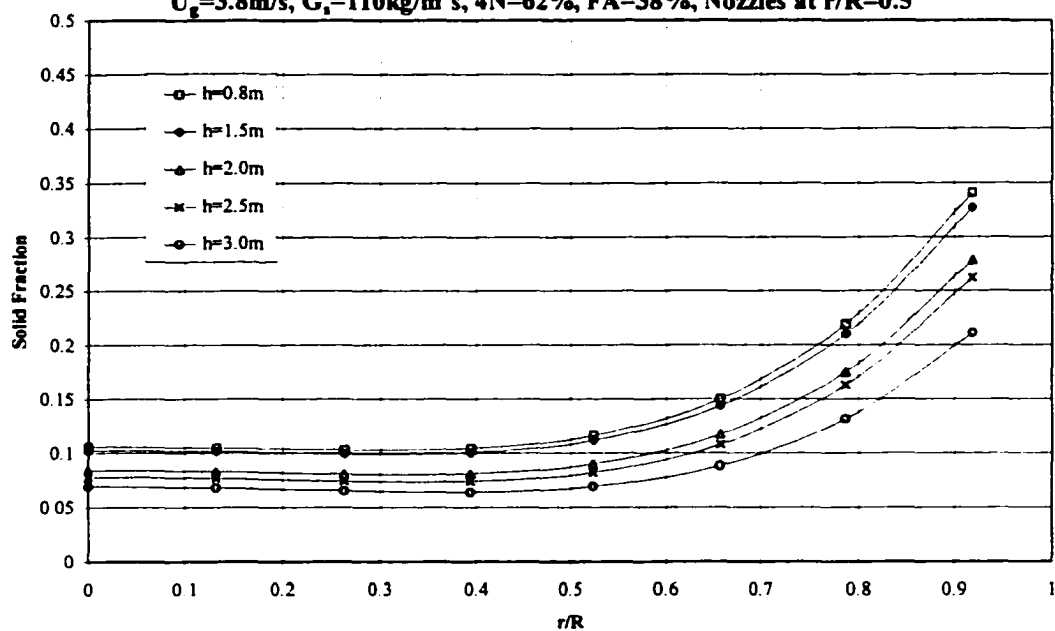
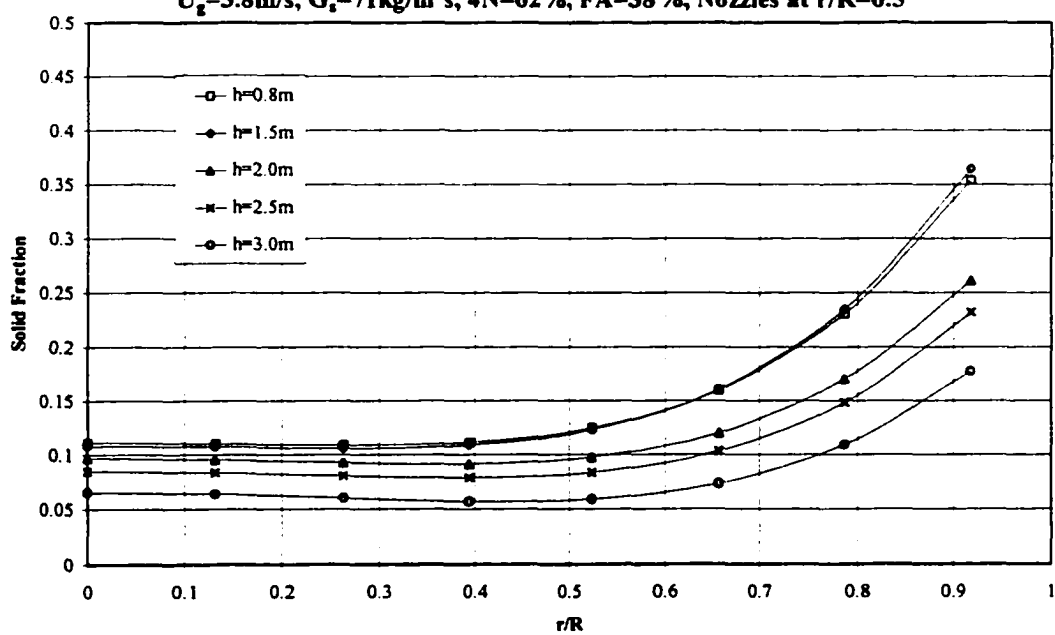
Figure B72: Radial Solid Profile $U_g=3.8\text{m/s}$, $G_s=110\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ **Figure B73: Radial Solid Profile** $U_g=3.8\text{m/s}$, $G_s=71\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ 

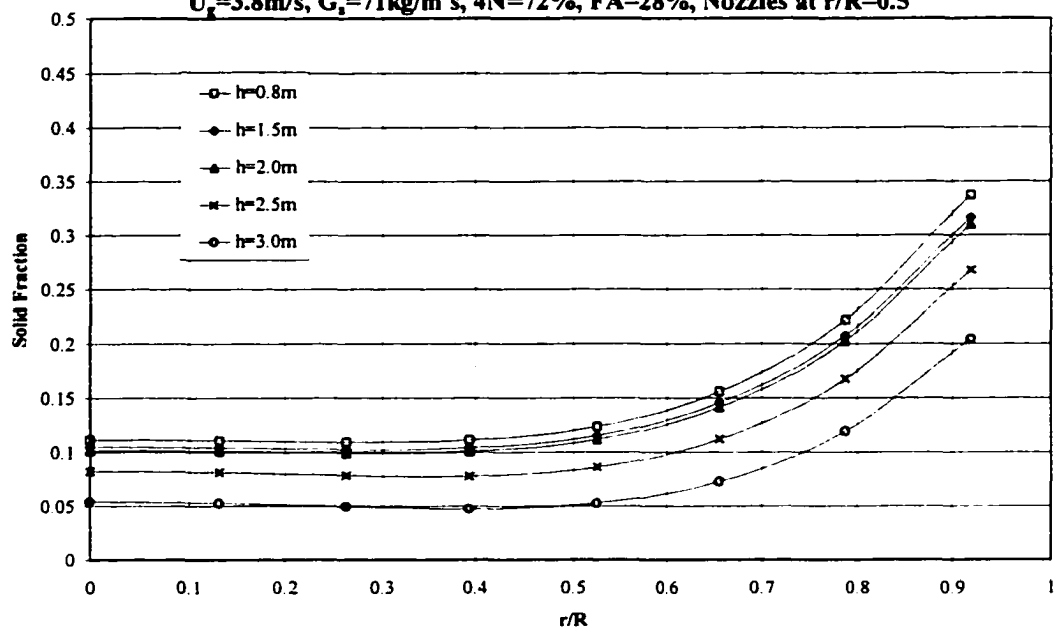
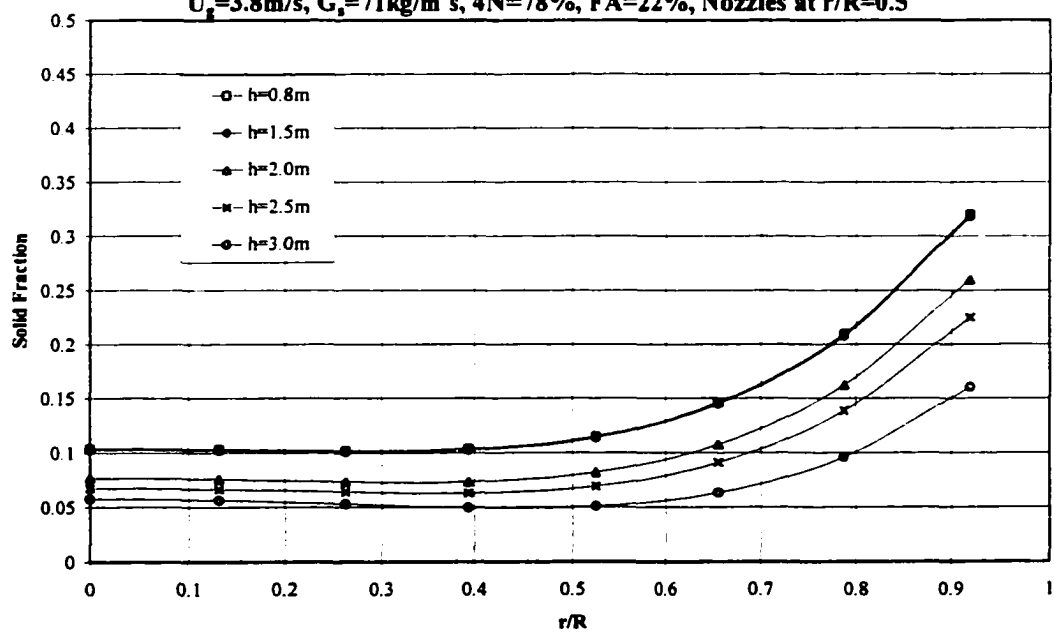
Figure B74: Radial Solid Profile $U_g=3.8\text{m/s}$, $G_s=71\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ **Figure B75: Radial Solid Density Profile** $U_g=3.8\text{m/s}$, $G_s=71\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ 

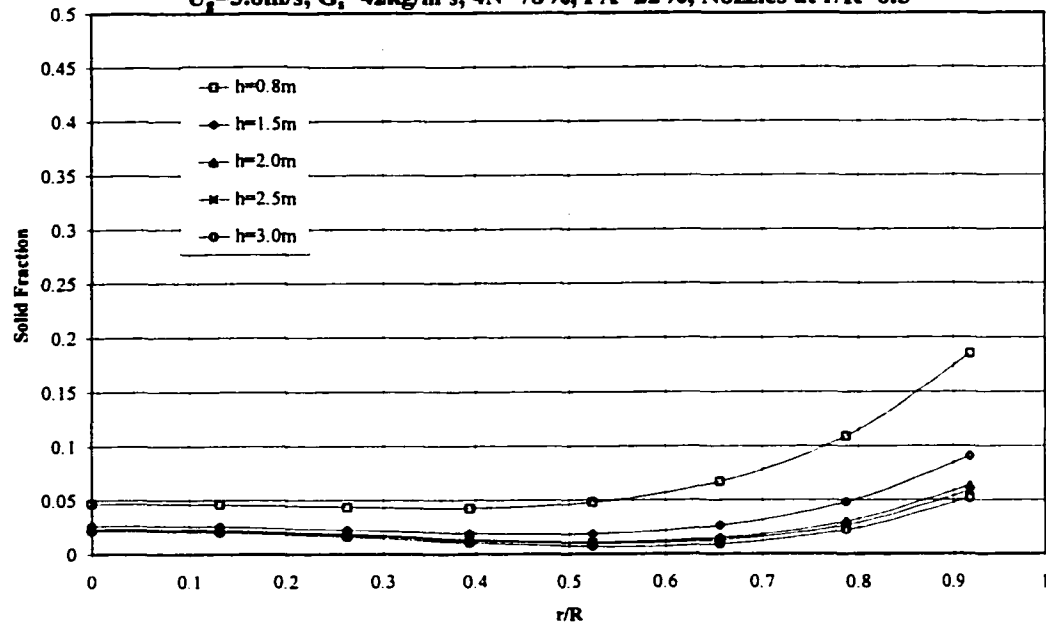
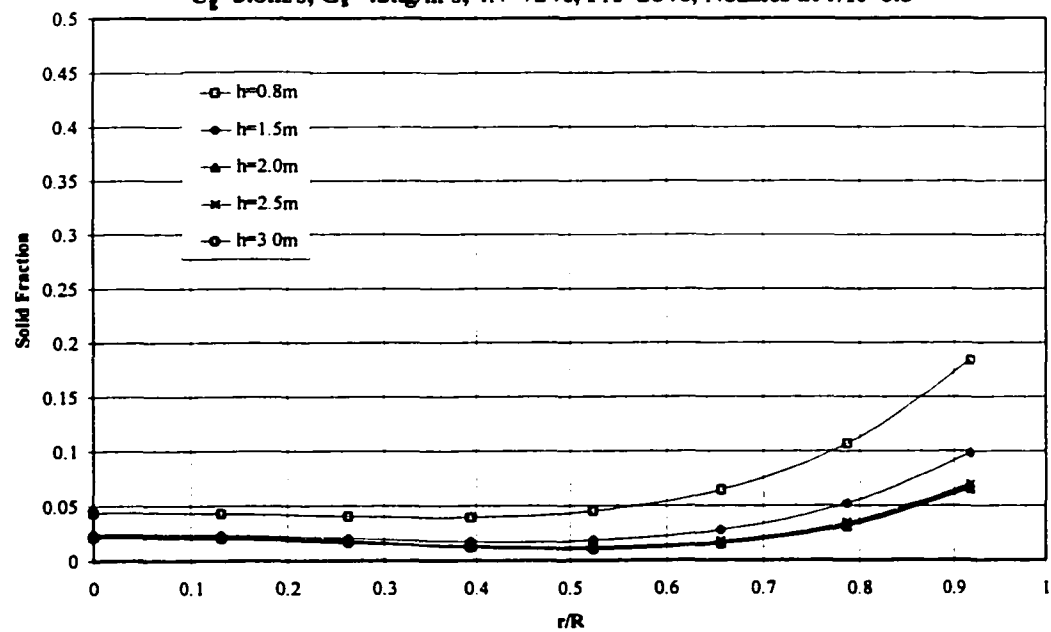
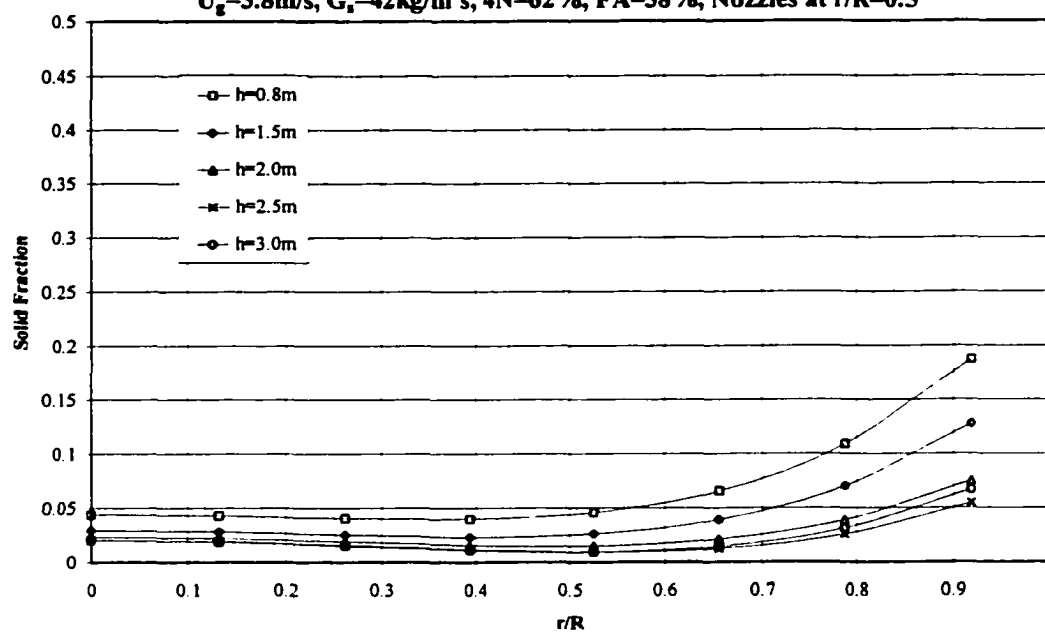
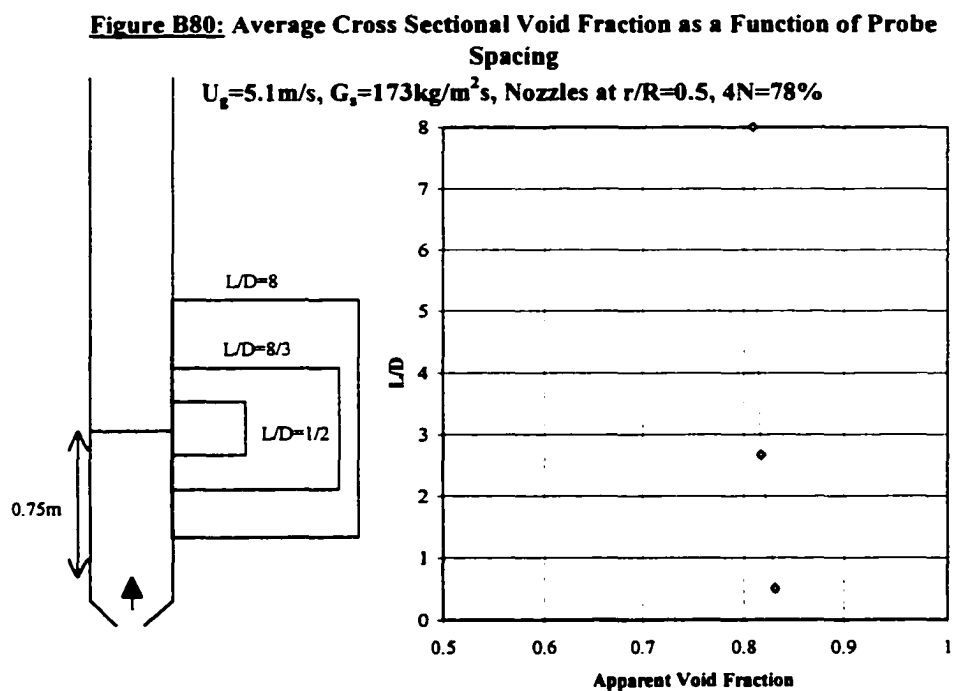
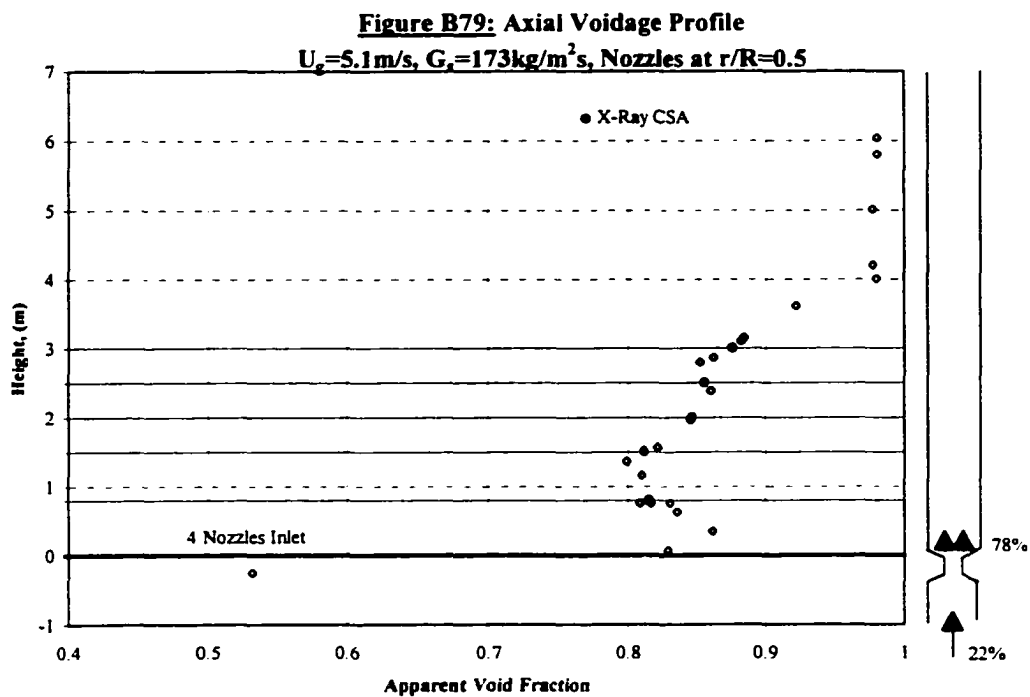
Figure B76: Radial Solid Profile $U_g=3.8\text{m/s}$, $G_s=42\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ **Figure B77: Radial Solid Profile** $U_g=3.8\text{m/s}$, $G_s=42\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ 

Figure B78: Radial Solid Profile $U_g=3.8\text{m/s}$, $G_s=42\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ 



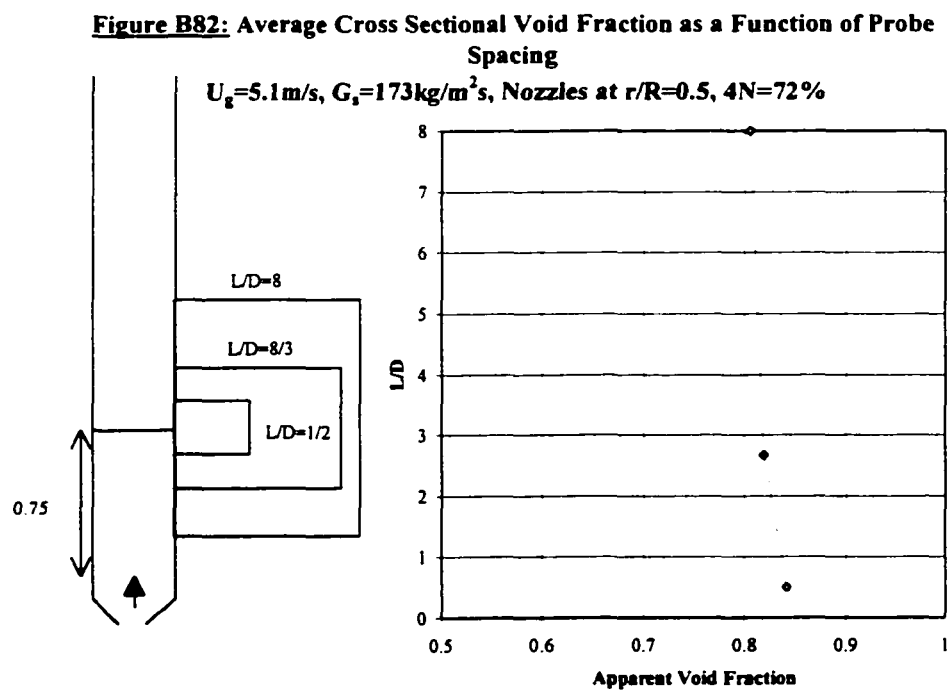
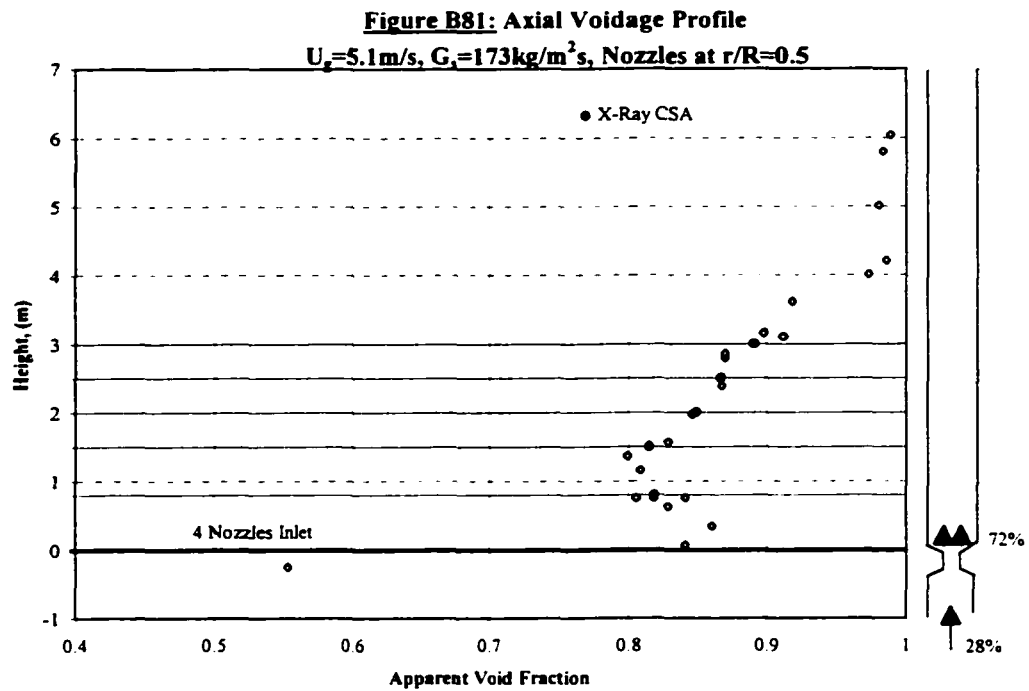


Figure B83: Axial Voidage Profile
 $U_g=5.1\text{ m/s}$, $G_g=173\text{ kg/m}^2\text{ s}$, Nozzles at $r/R=0.5$

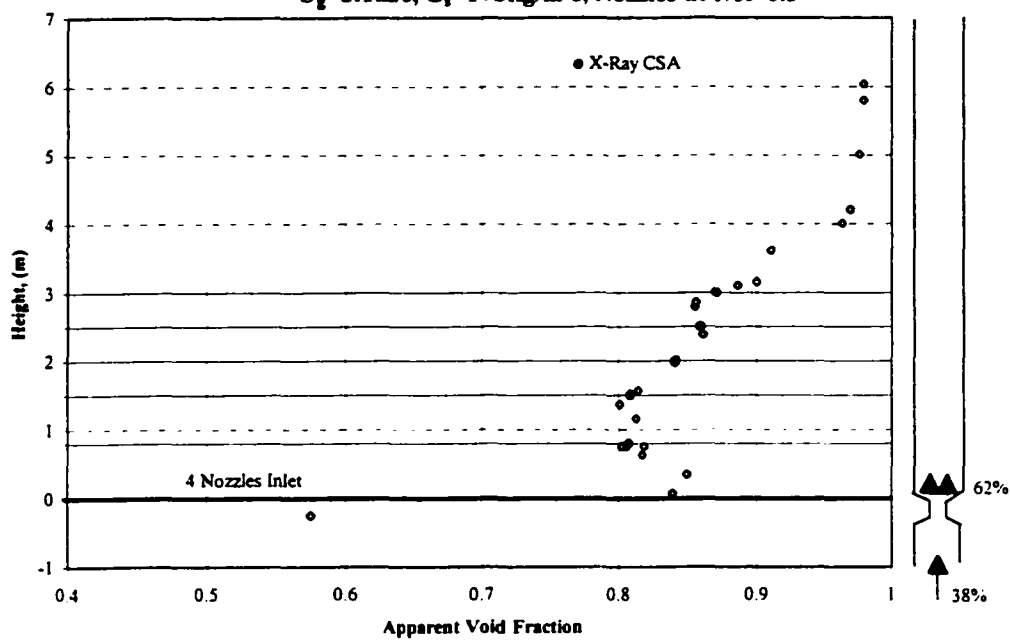
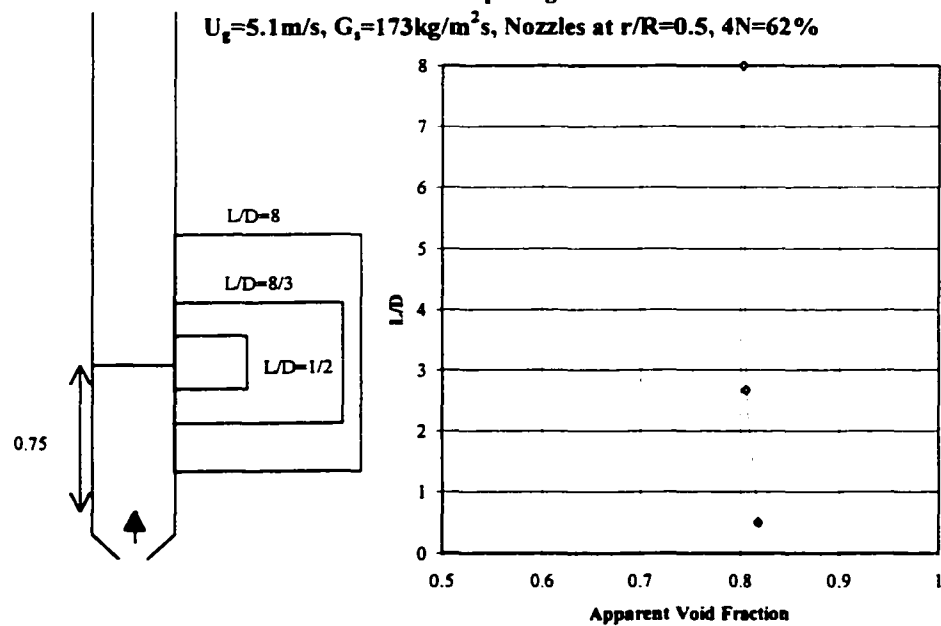
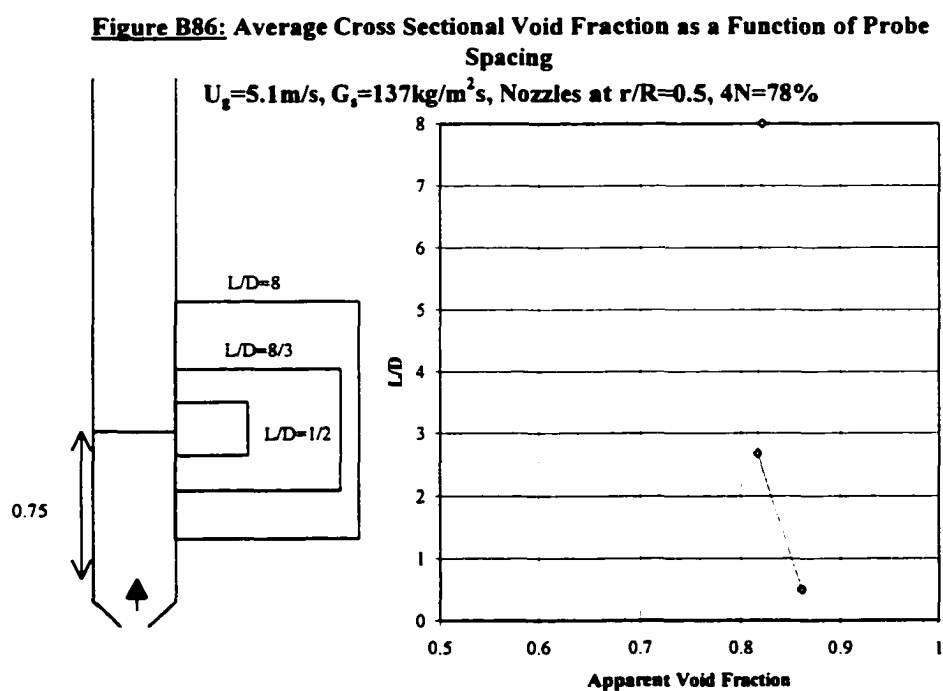
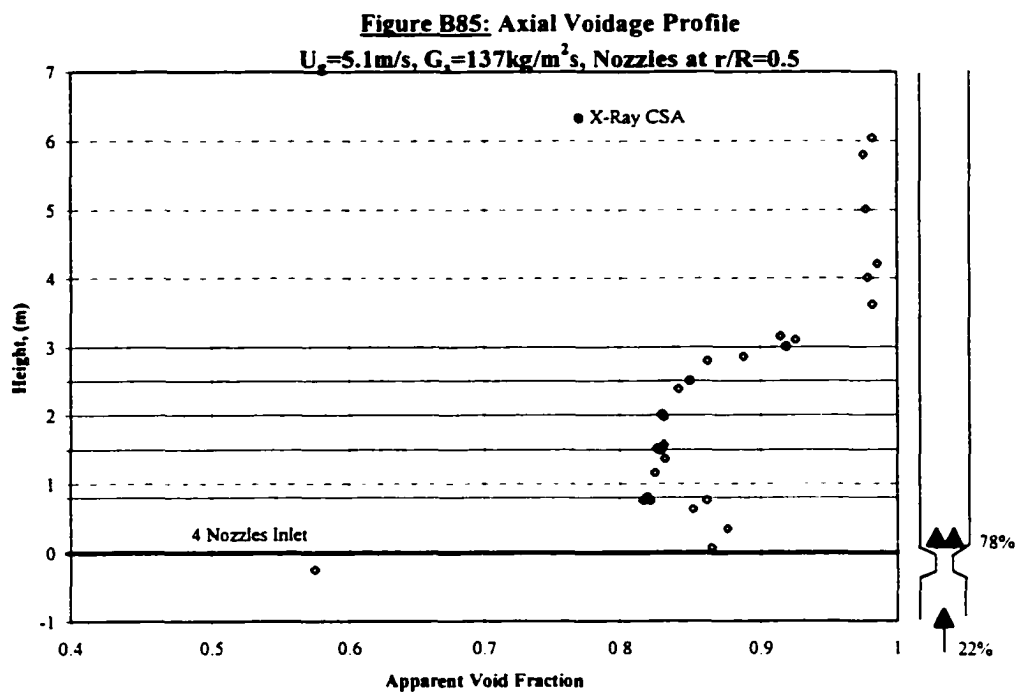
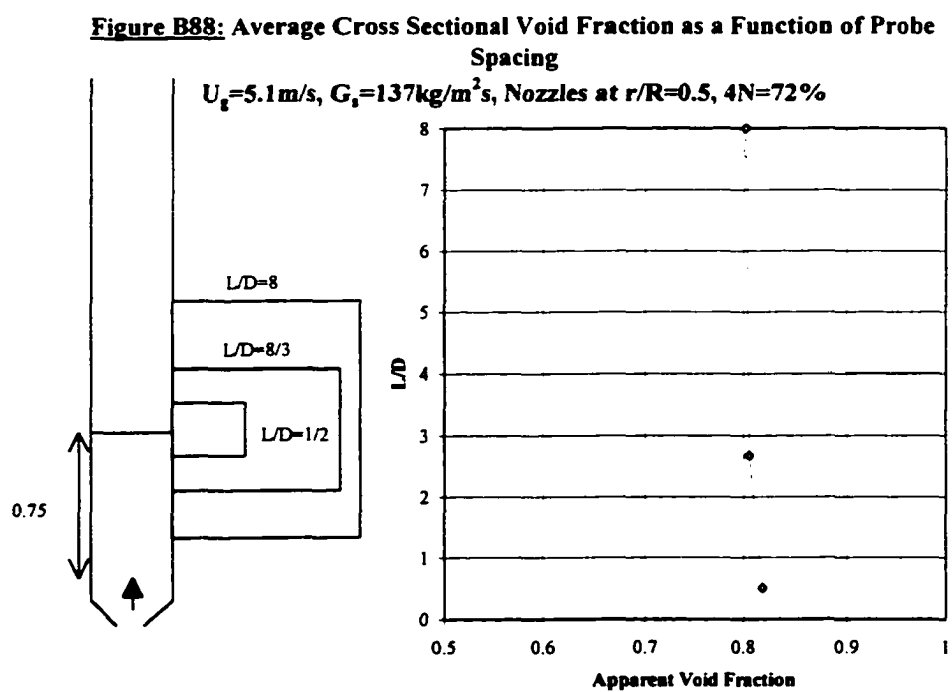
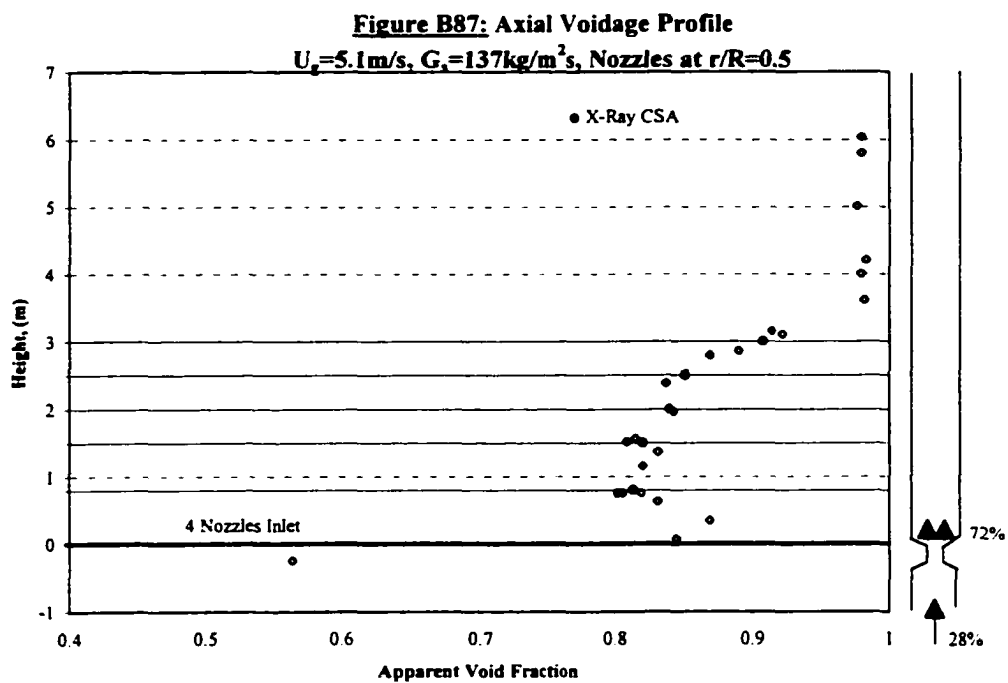
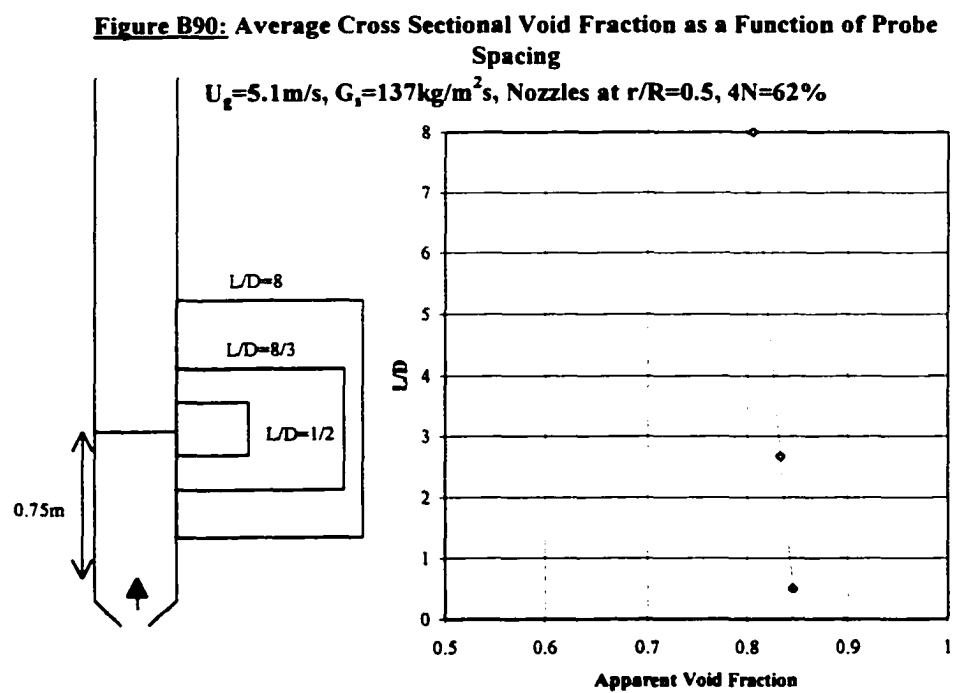
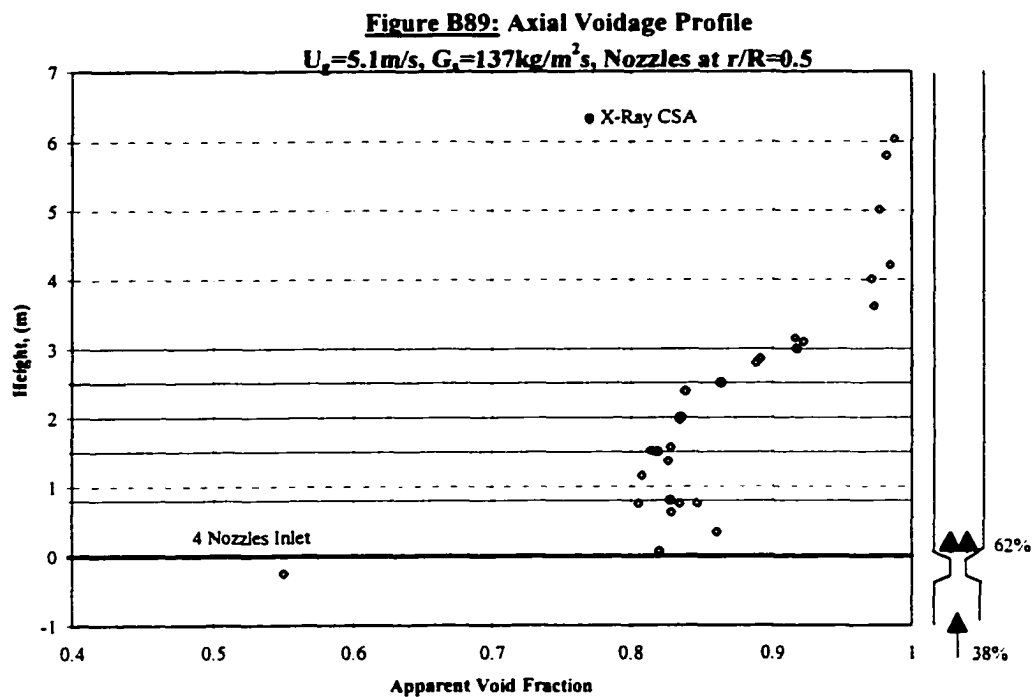


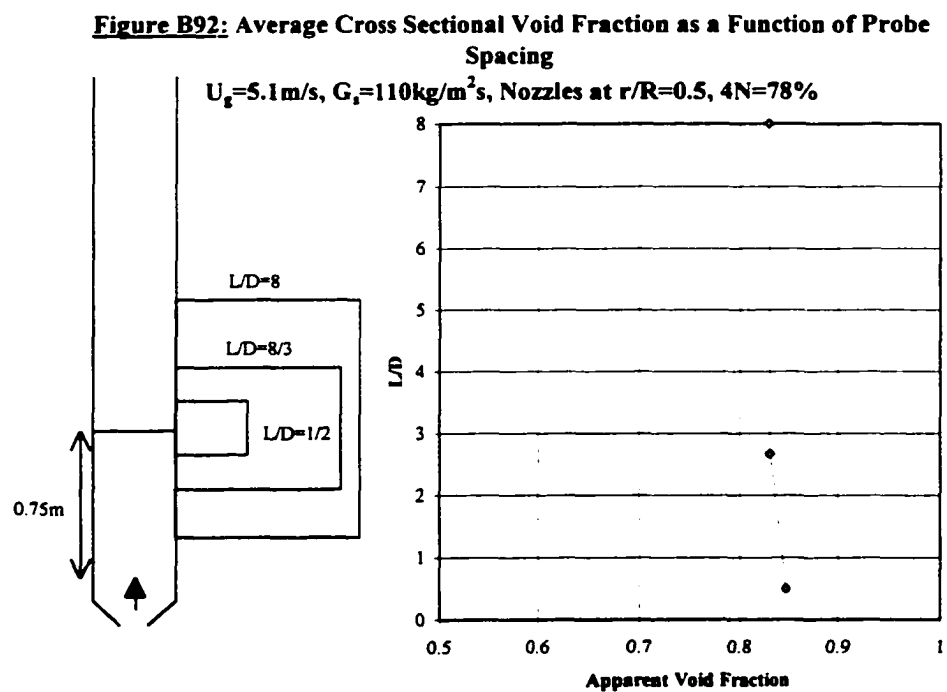
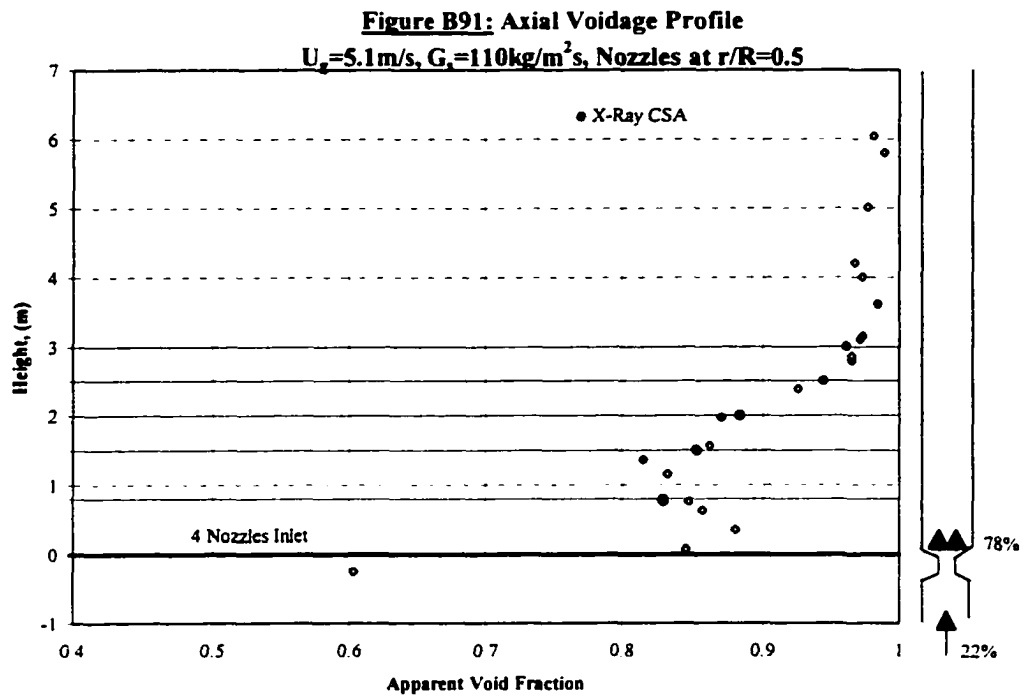
Figure B84: Average Cross Sectional Void Fraction as a Function of Probe Spacing

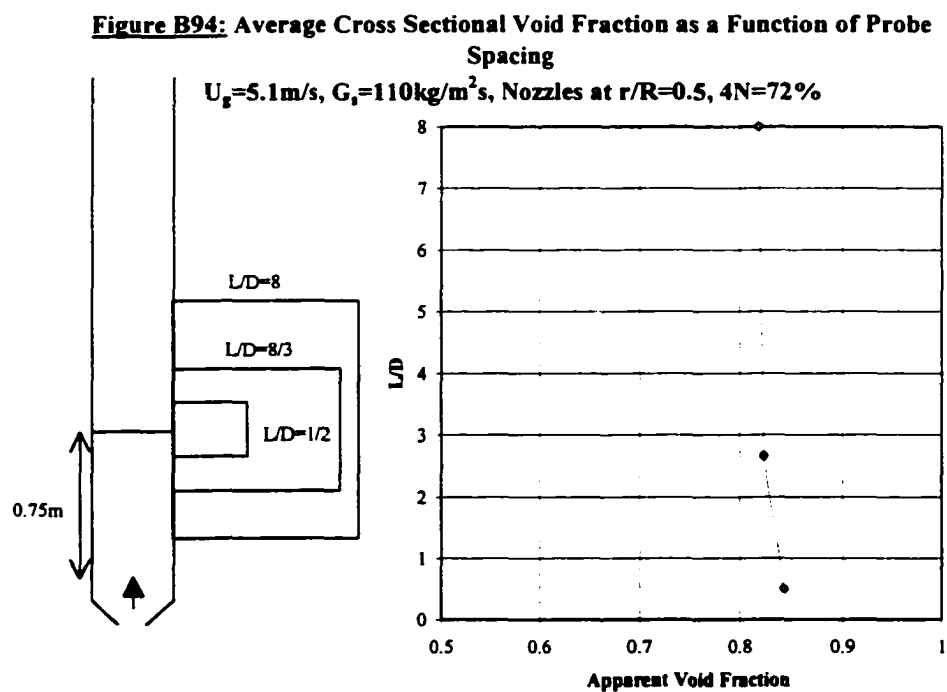
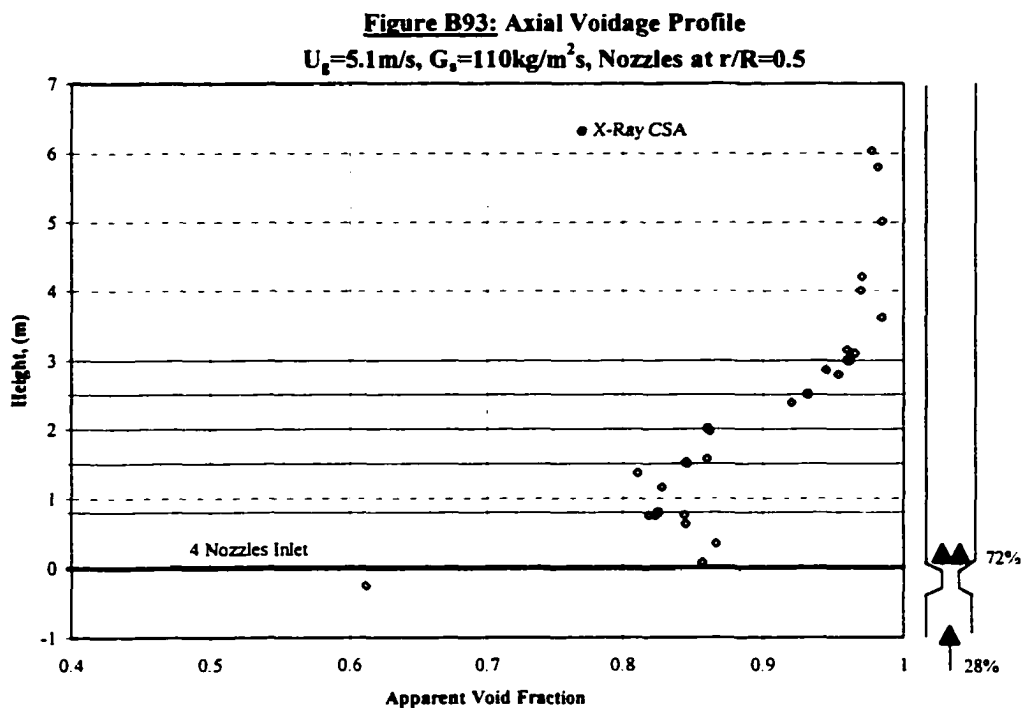












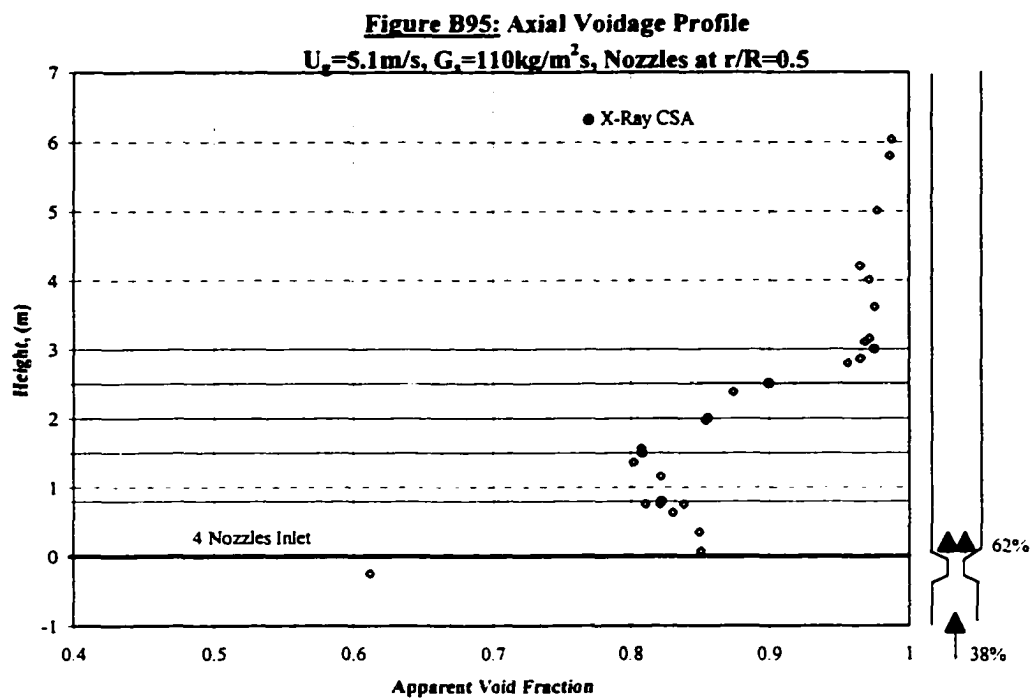
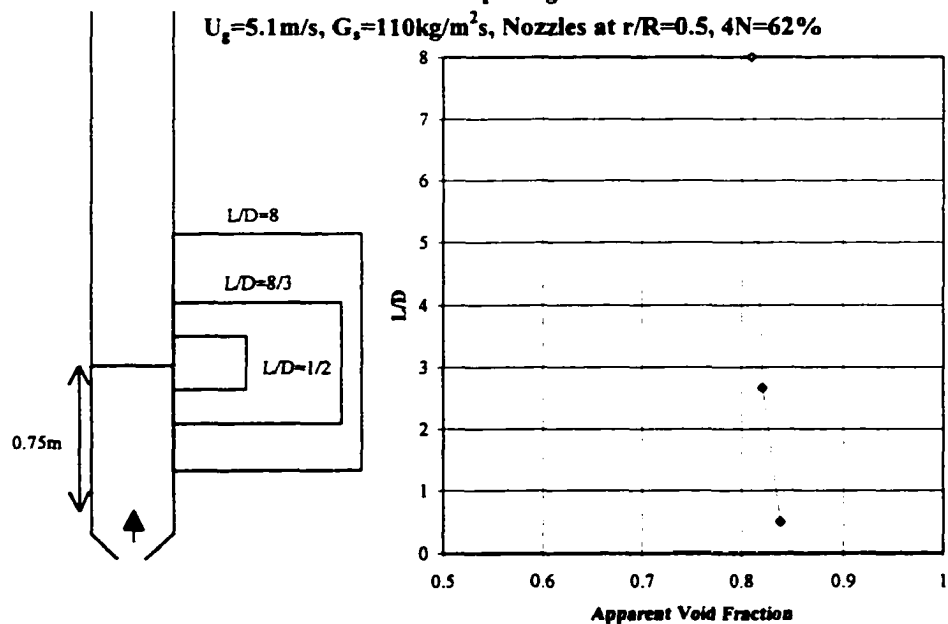
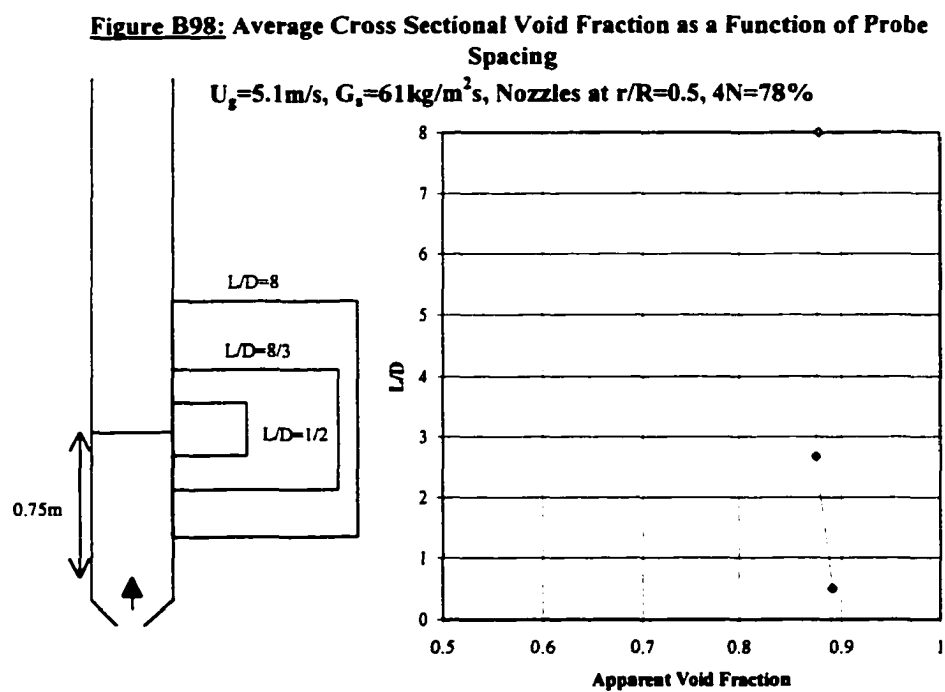
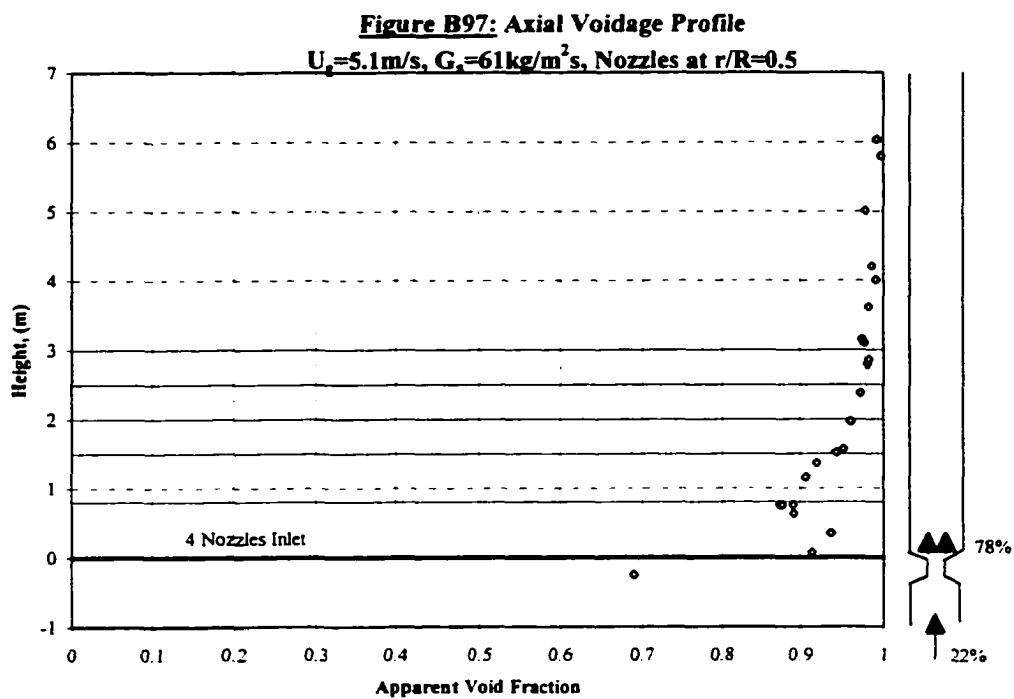


Figure B96: Average Cross Sectional Void Fraction as a Function of Probe Spacing





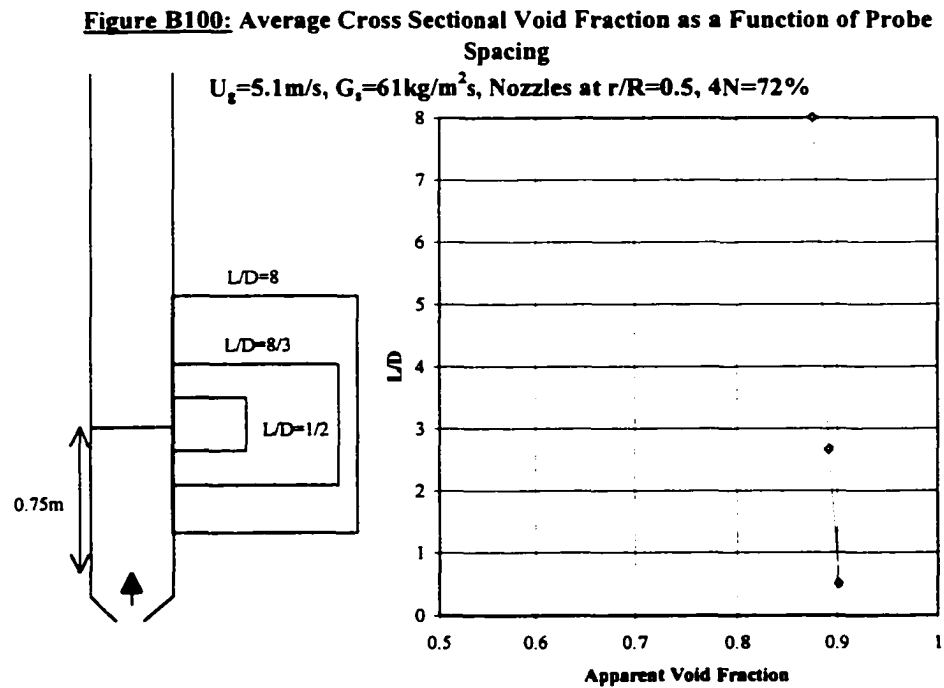
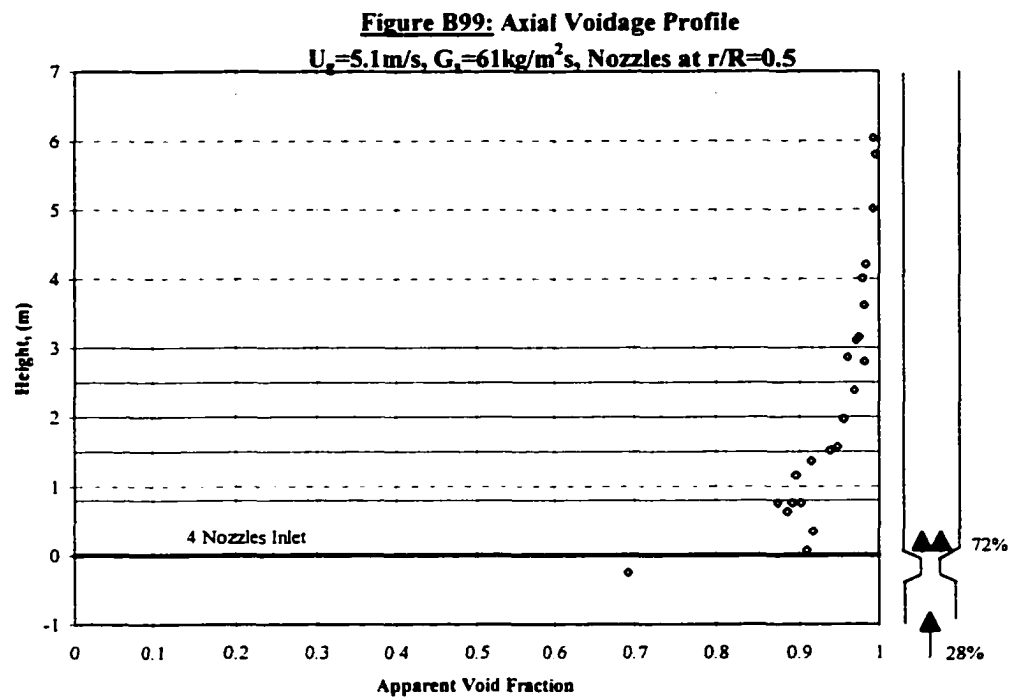


Figure B101: Axial Voidage Profile
 $U_g=5.1\text{m/s}$, $G_g=61\text{kg/m}^2\text{s}$, Nozzles at $r/R=0.5$

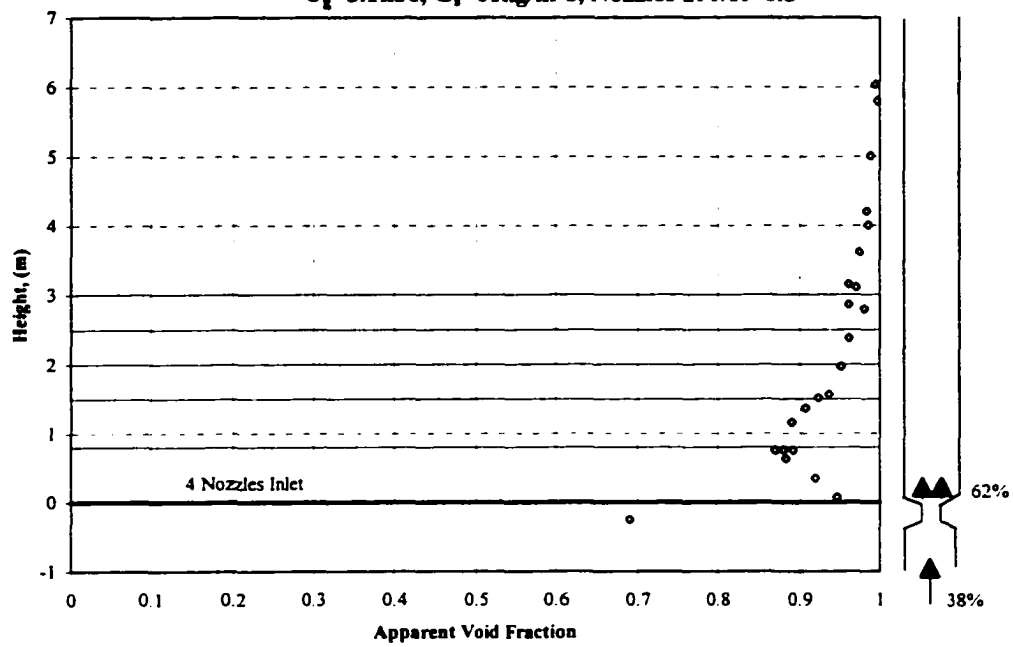


Figure B102: Average Cross Sectional Void Fraction as a Function of Probe Spacing

$U_g=5.1\text{m/s}$, $G_g=61\text{kg/m}^2\text{s}$, Nozzles at $r/R=0.5$, $4N=62\%$

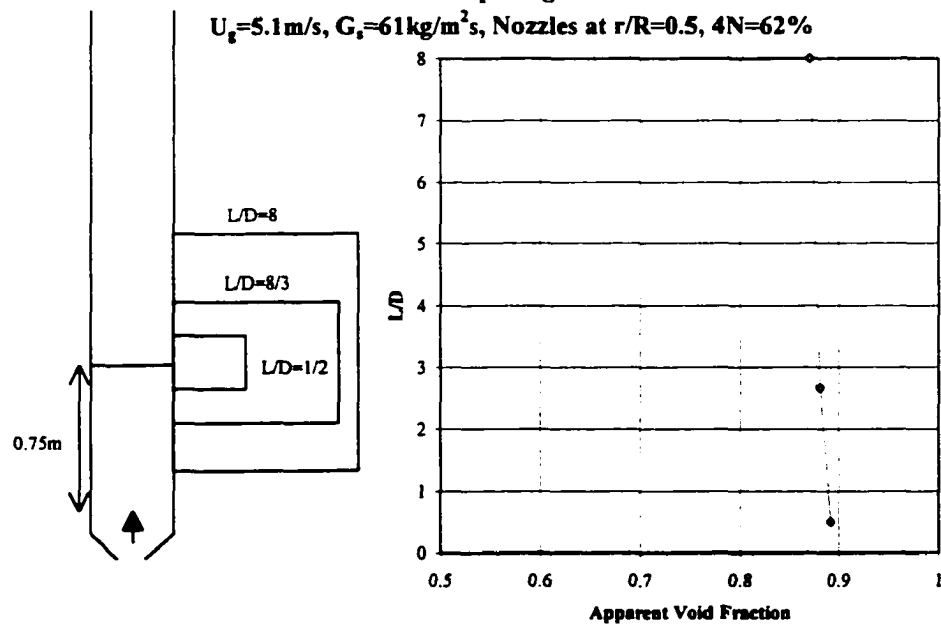


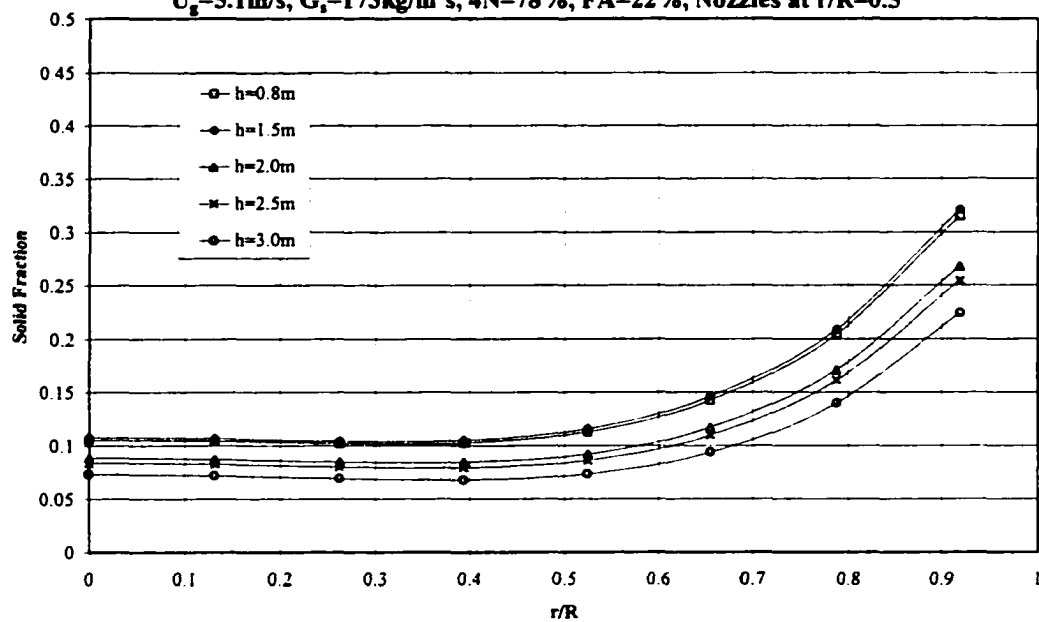
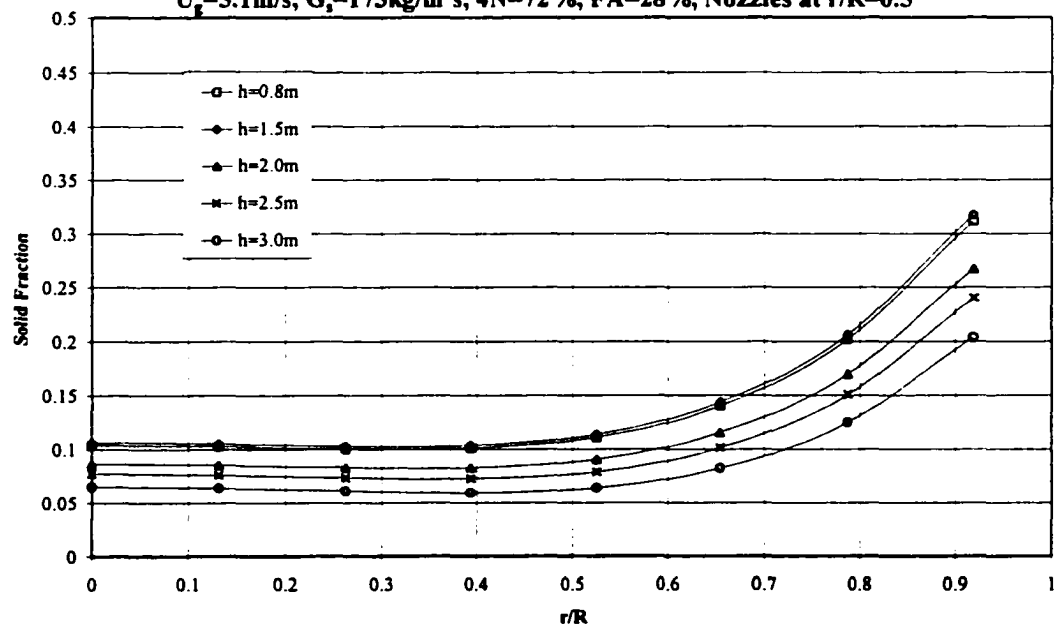
Figure B103: Radial Solid Density Profile $U_g=5.1\text{m/s}$, $G_s=173\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ **Figure B104: Radial Solid Profile** $U_g=5.1\text{m/s}$, $G_s=173\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ 

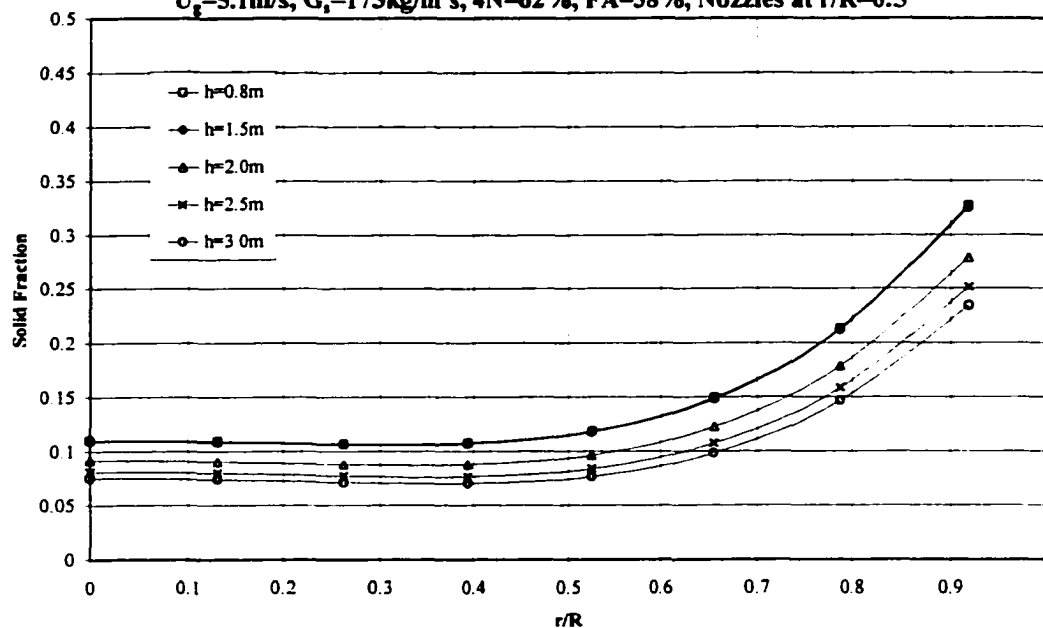
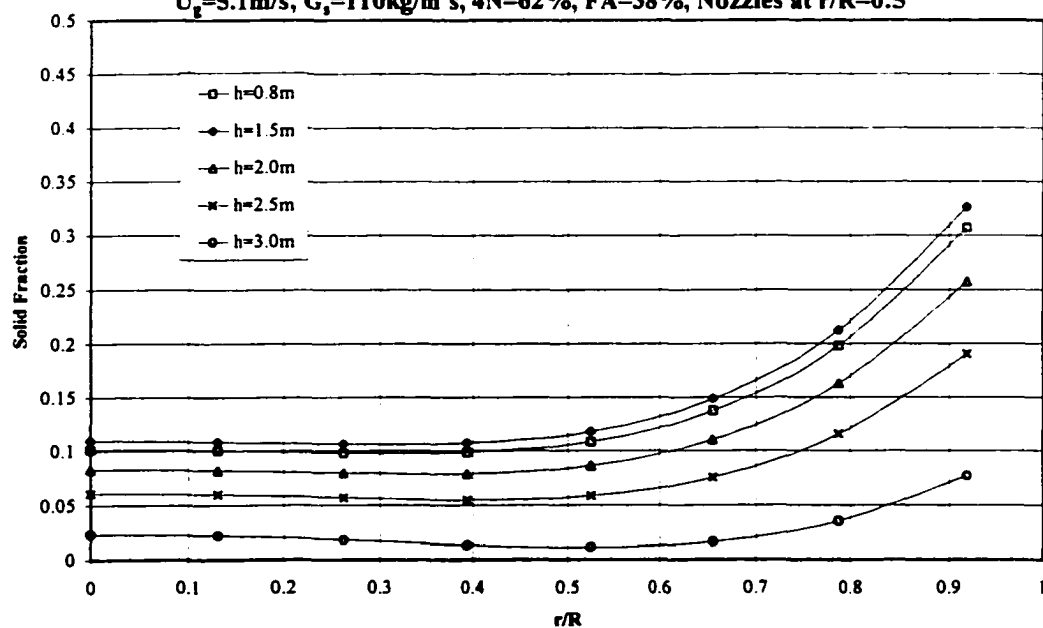
Figure B105: Radial Solid Profile $U_g=5.1\text{m/s}$, $G_s=173\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ **Figure B106: Radial Solid Profile** $U_g=5.1\text{m/s}$, $G_s=110\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ 

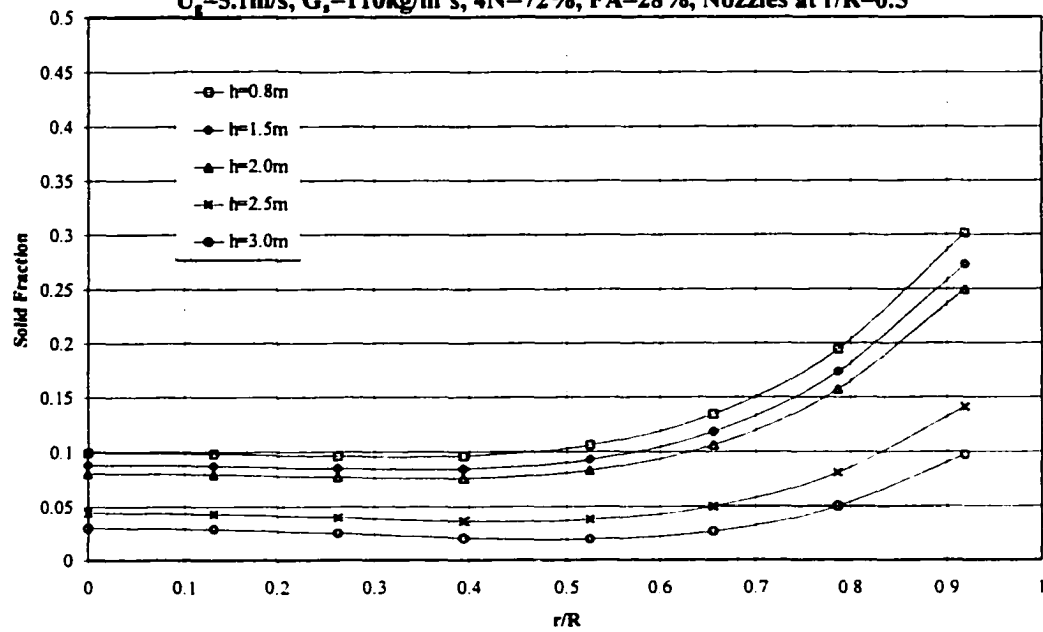
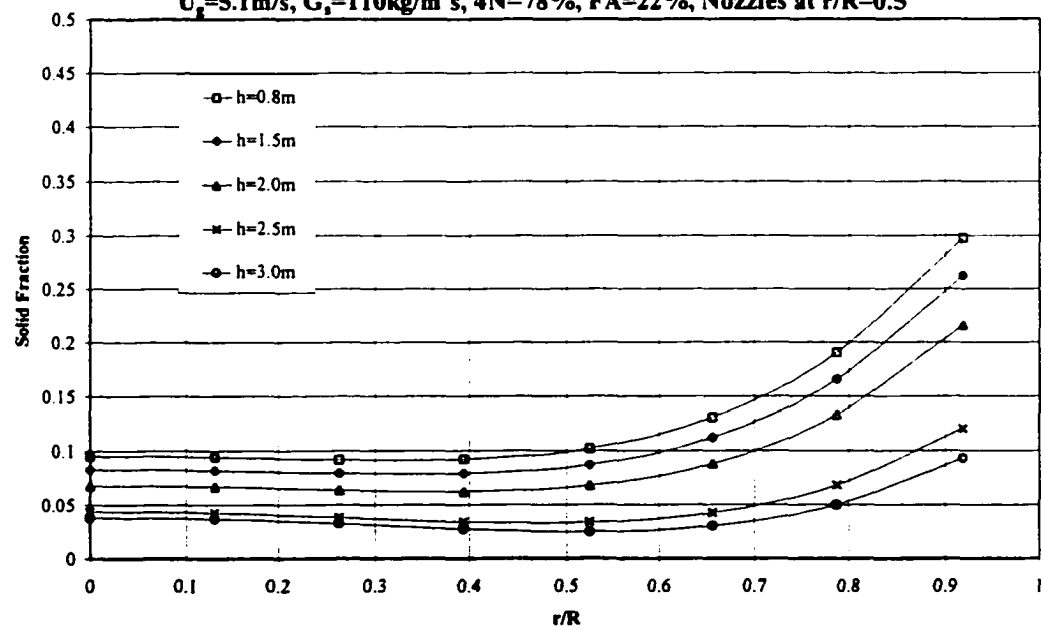
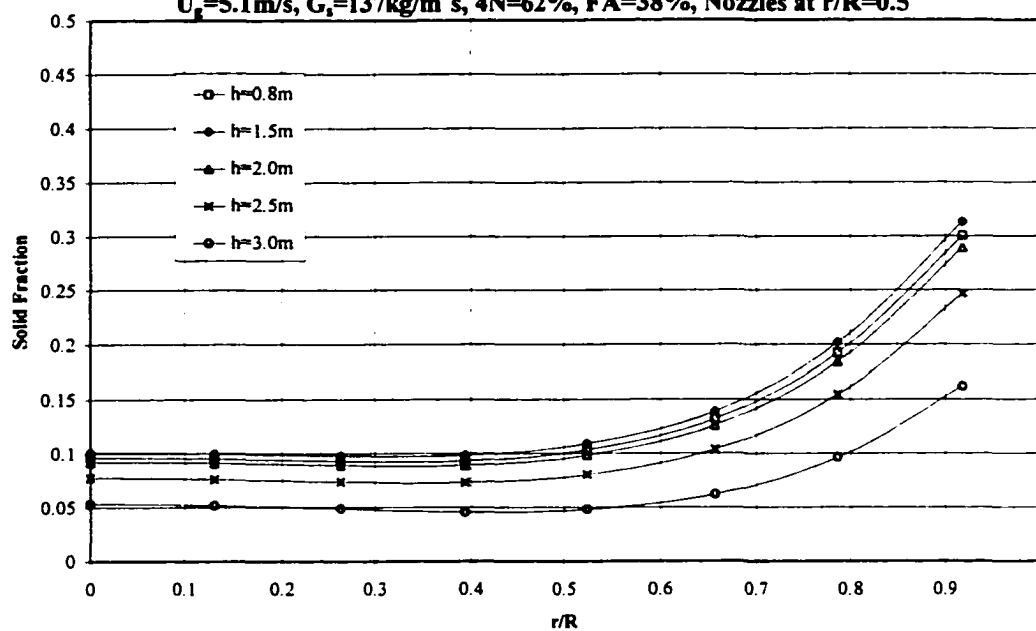
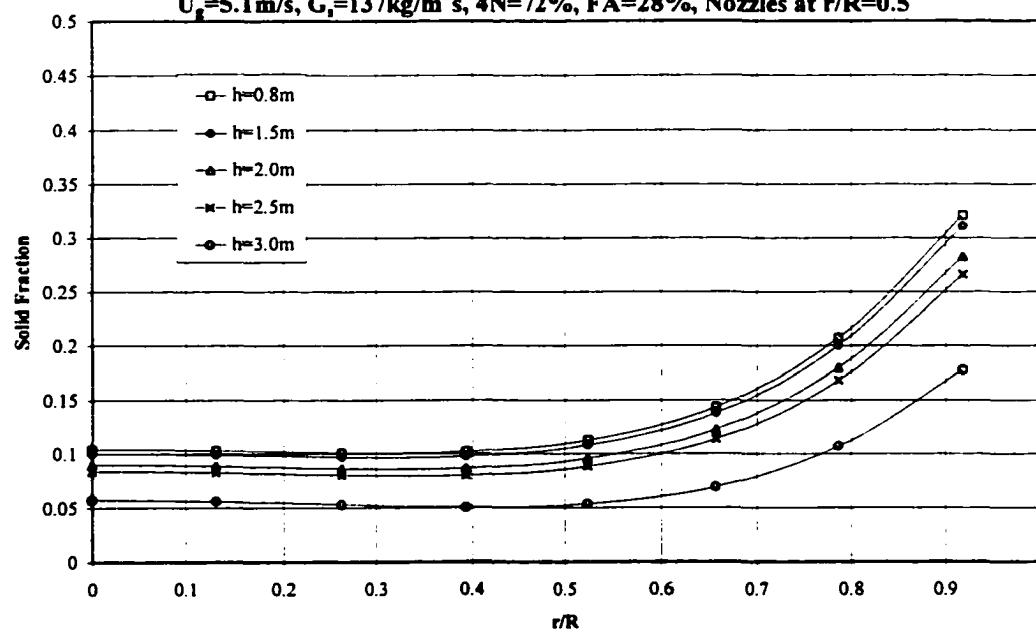
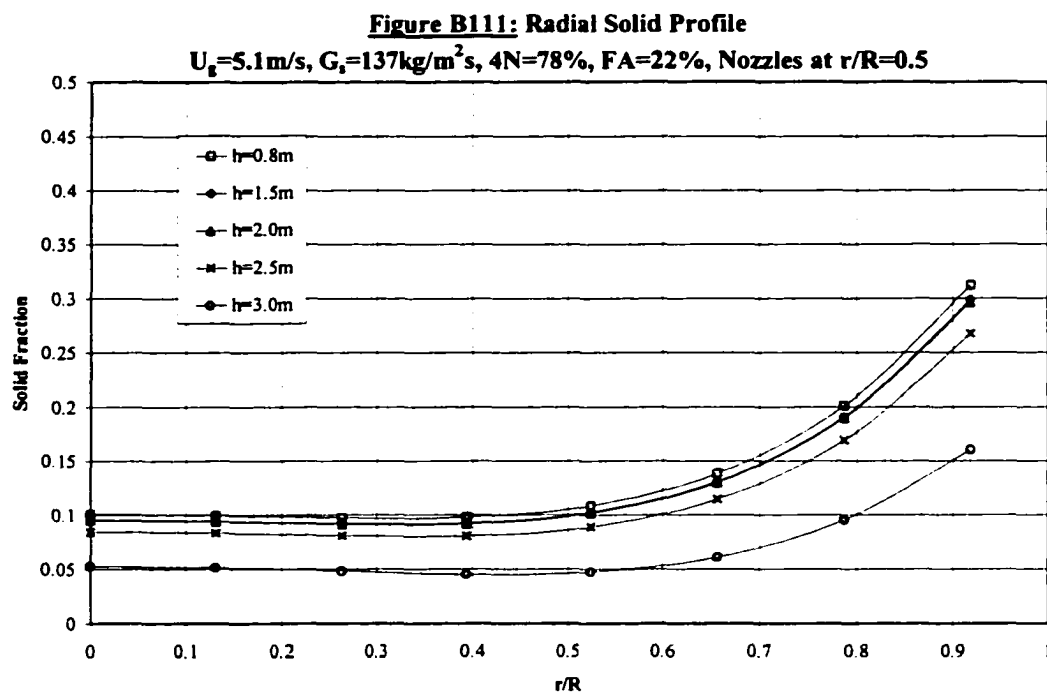
Figure B107: Radial Solid Profile $U_f=5.1\text{m/s}$, $G_s=110\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ **Figure B108: Radial Solid Density Profile** $U_f=5.1\text{m/s}$, $G_s=110\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at $r/R=0.5$ 

Figure B109: Radial Solid Profile $U_g=5.1\text{m/s}$, $G_t=137\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at $r/R=0.5$ **Figure B110: Radial Solid Profile** $U_g=5.1\text{m/s}$, $G_t=137\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at $r/R=0.5$ 



APPENDIX C

In Appendix C are presented all the experimental results with the nozzles located as close as possible to the riser wall. These results include axial and radial solid distribution and the effect of the pressure tap spacing on the axial pressure difference measurements for a variety of operating conditions, gas superficial velocity, solid circulation rate and "fluffing" air flowrate.

Figure C1: Axial Voidage Profile
 $U_g=2.1\text{m/s}$, $G_g=105\text{kg/m}^2\text{s}$, Nozzles at the wall

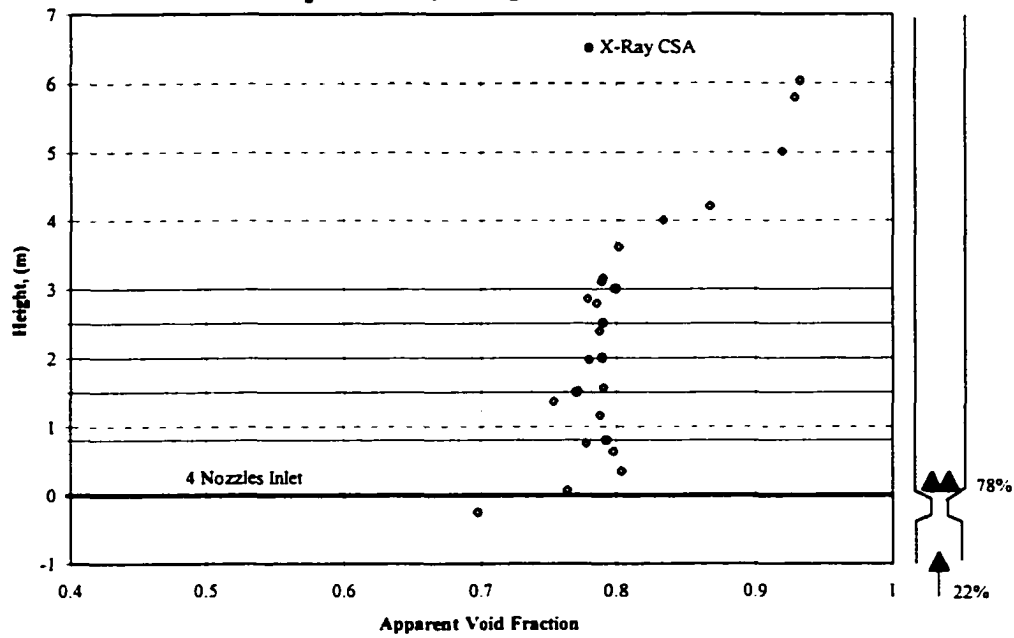
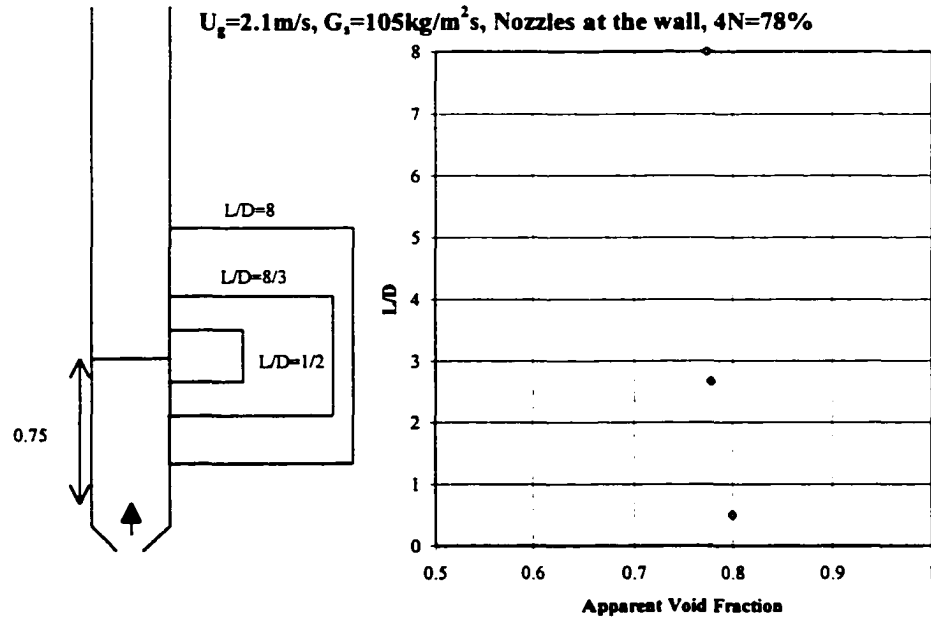
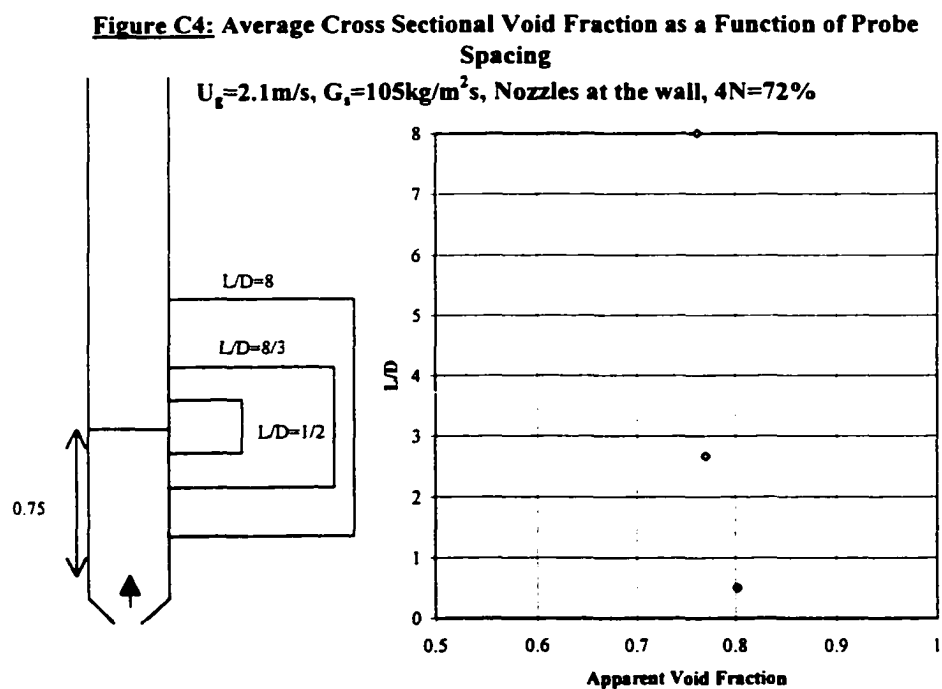
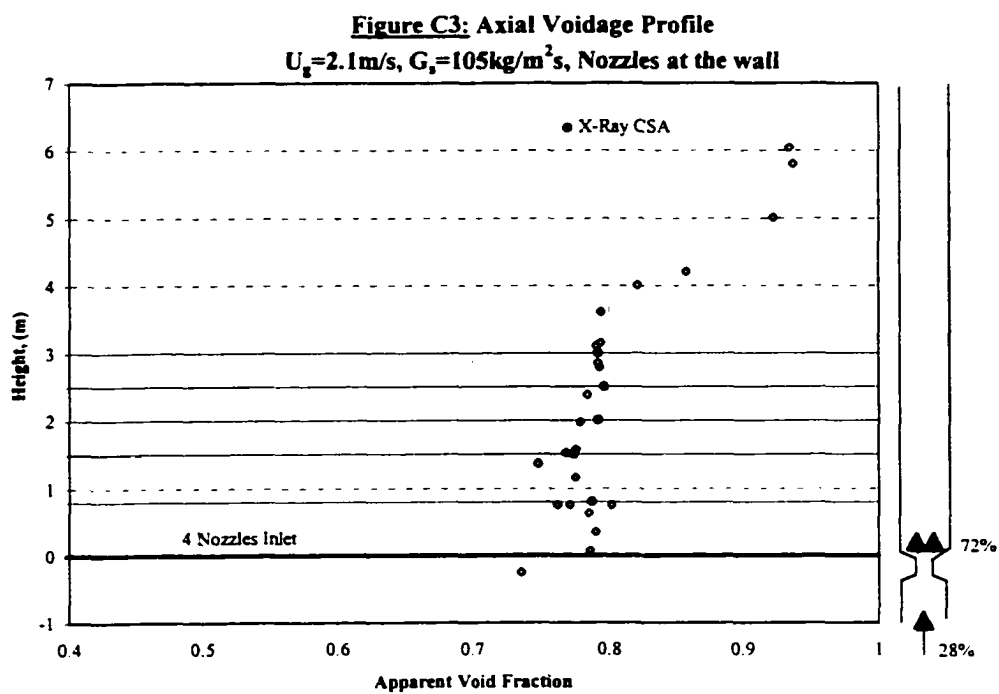


Figure C2: Average Cross Sectional Void Fraction as a Function of Probe Spacing





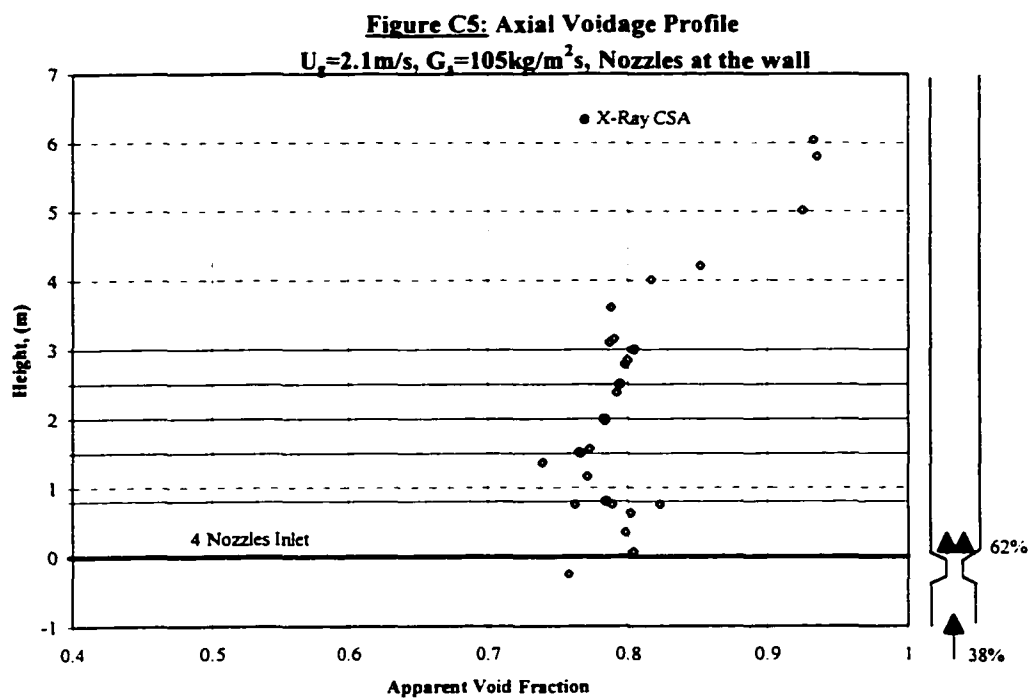
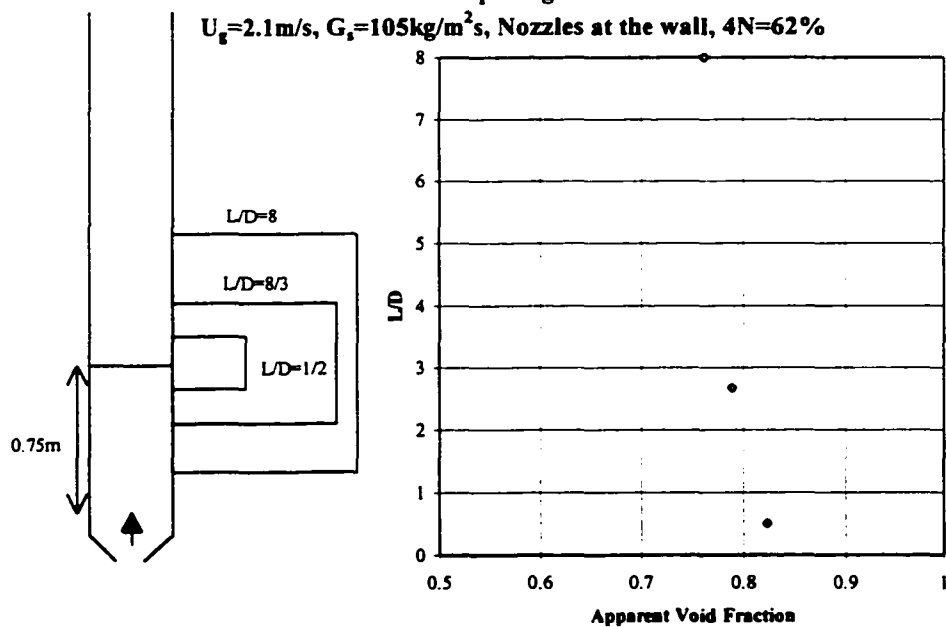
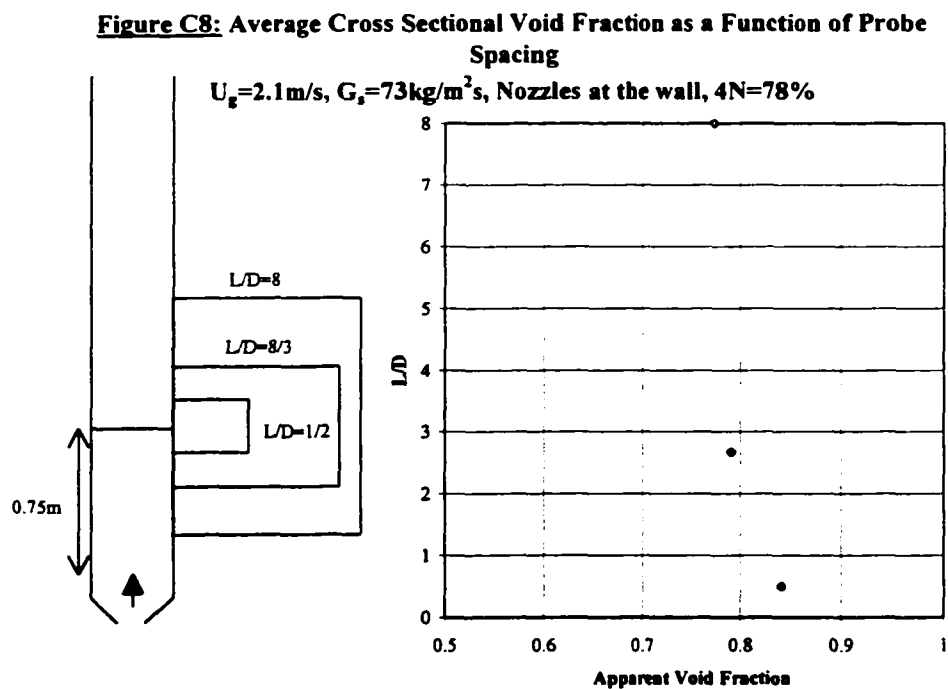
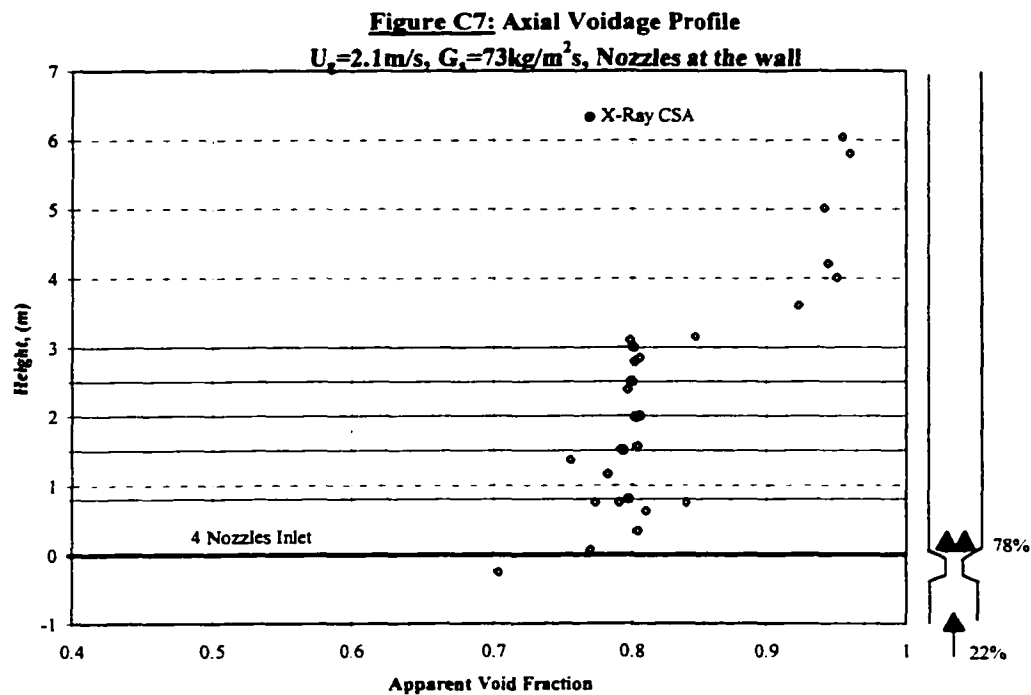


Figure C6: Average Cross Sectional Void Fraction as a Function of Probe Spacing





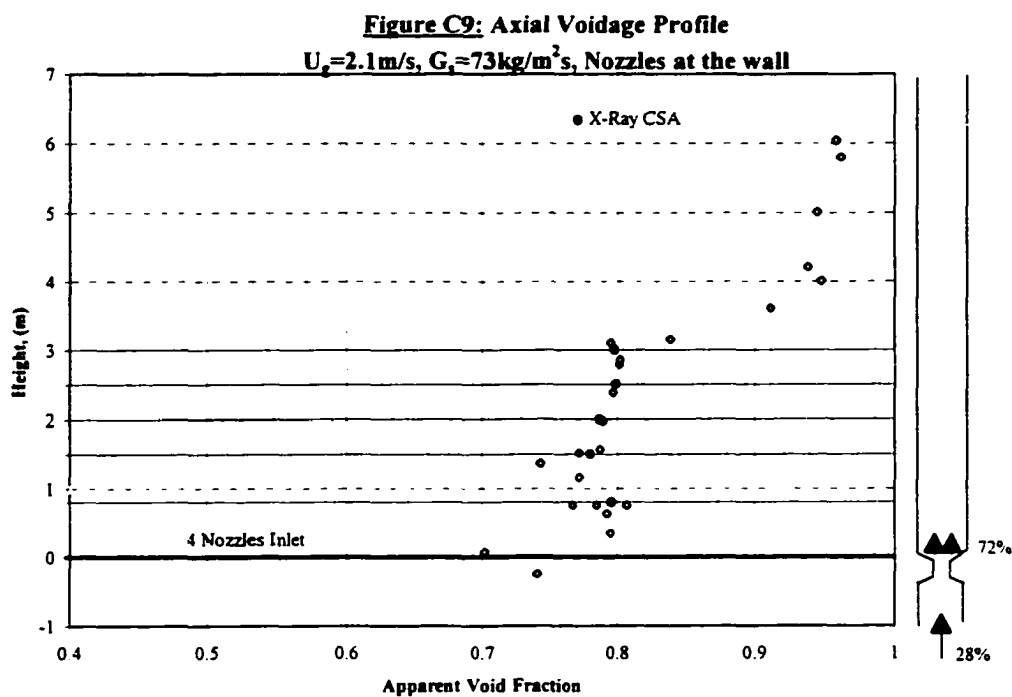
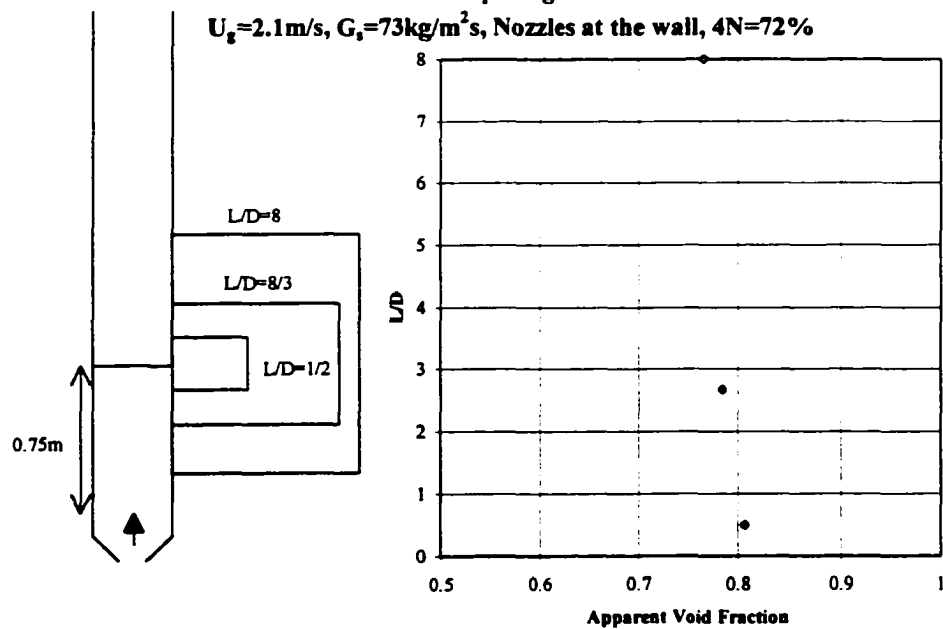
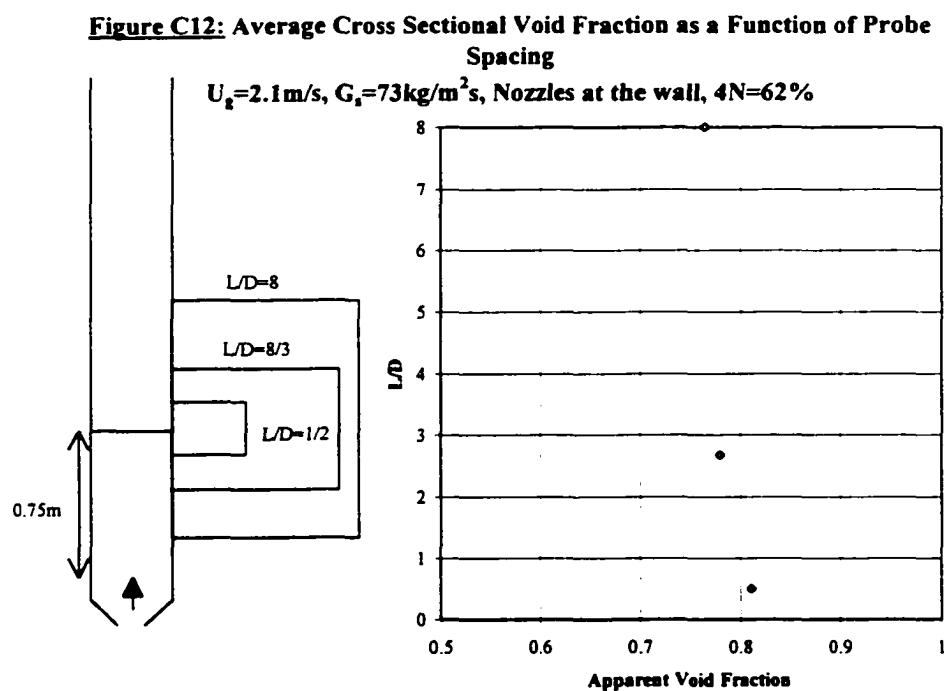
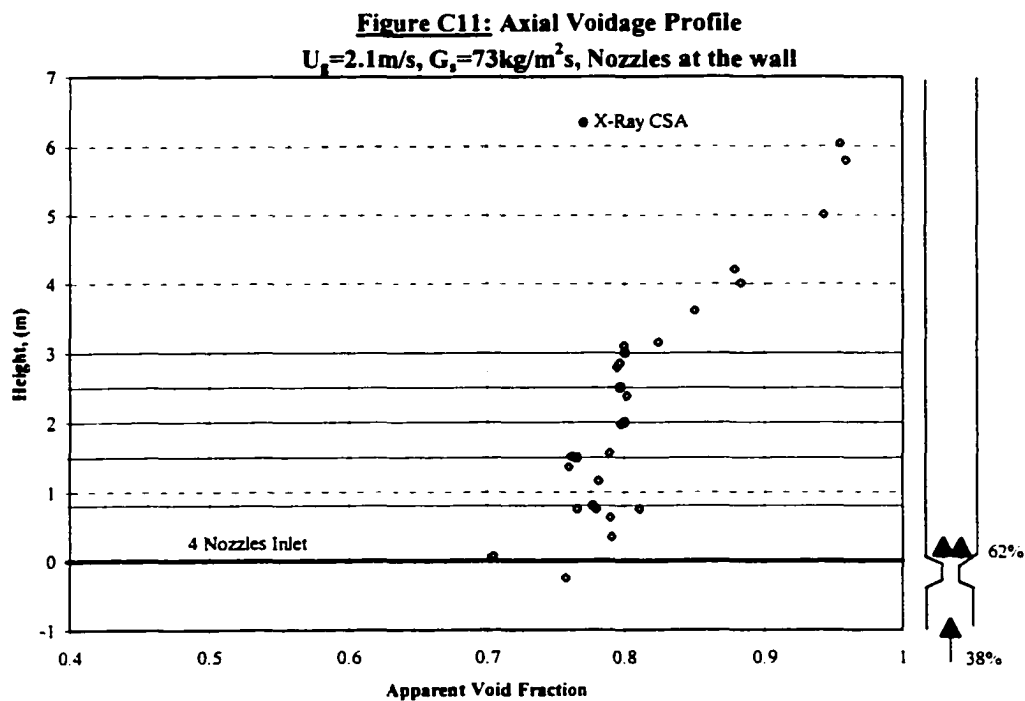
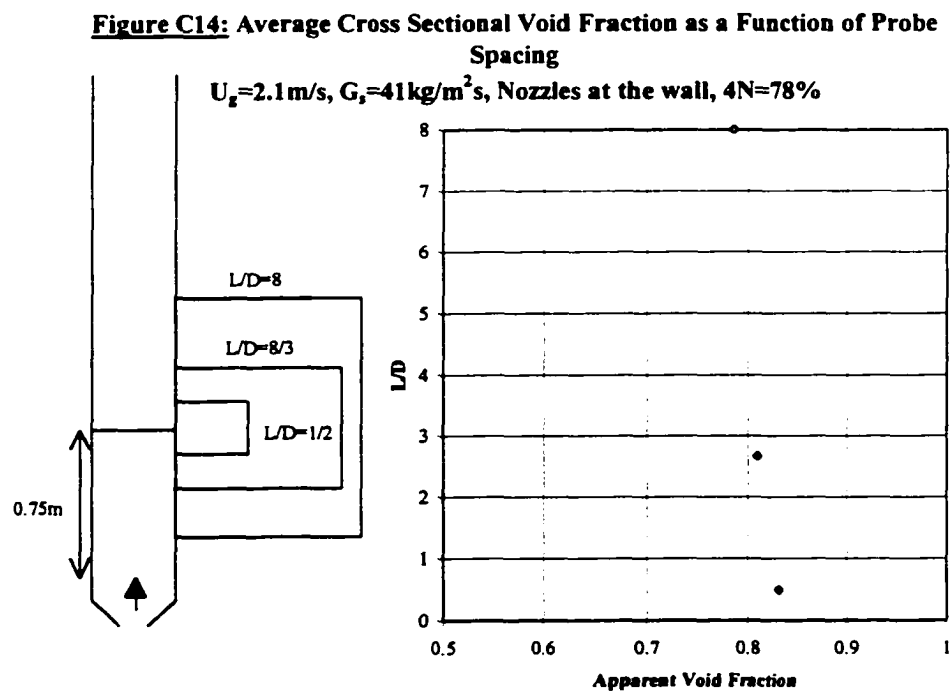
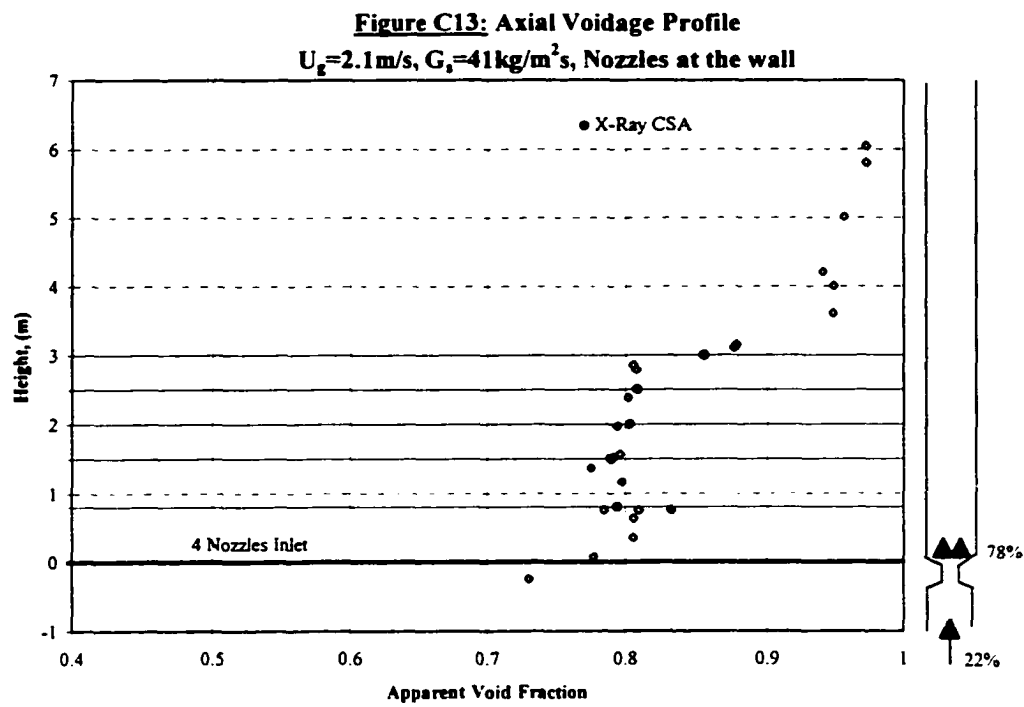
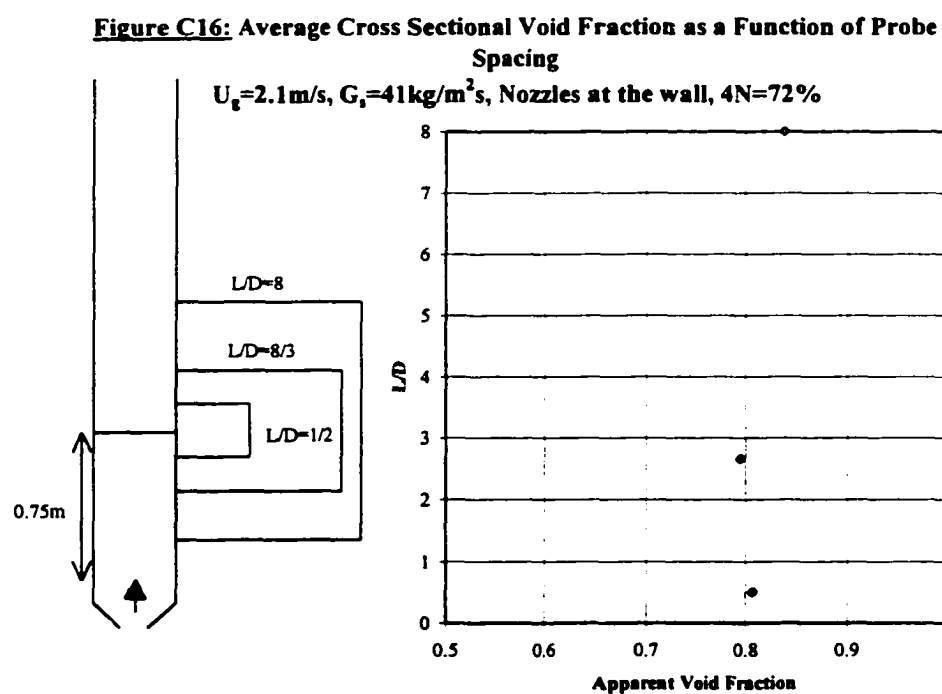
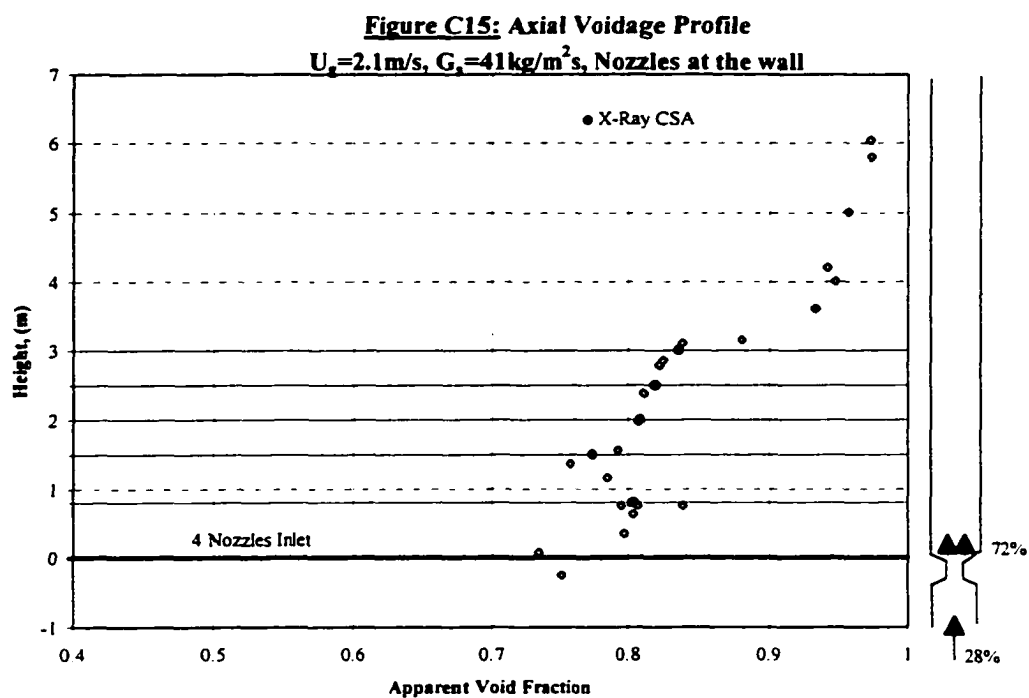


Figure C10: Average Cross Sectional Void Fraction as a Function of Probe Spacing









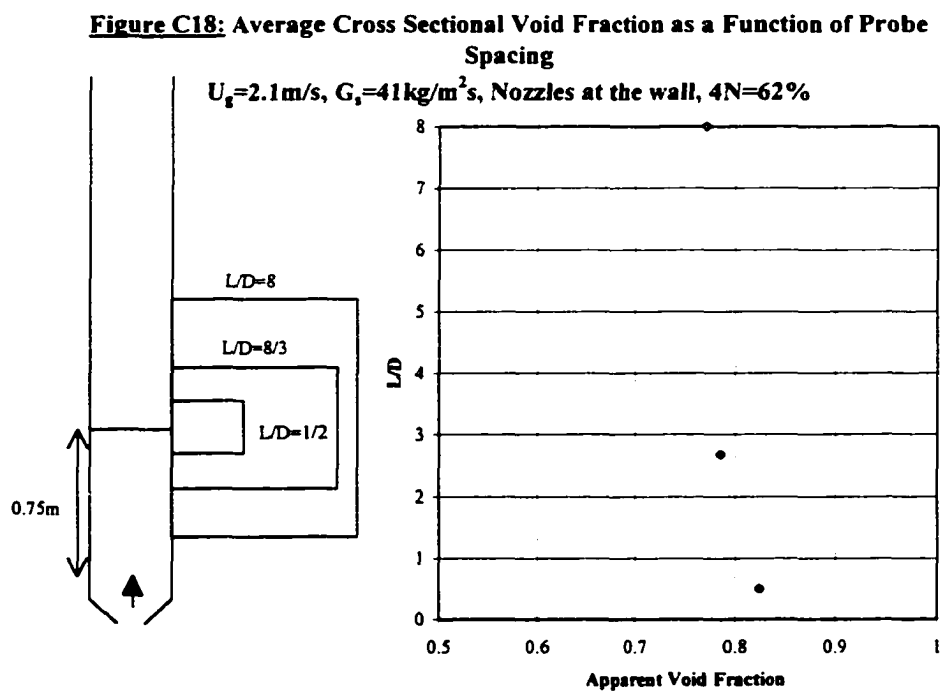
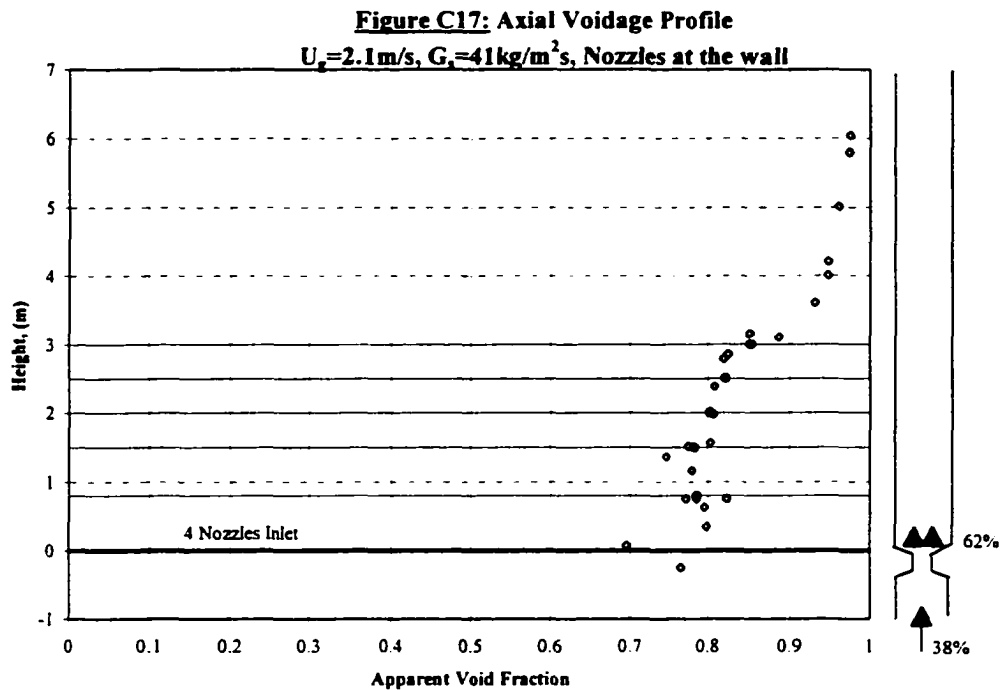


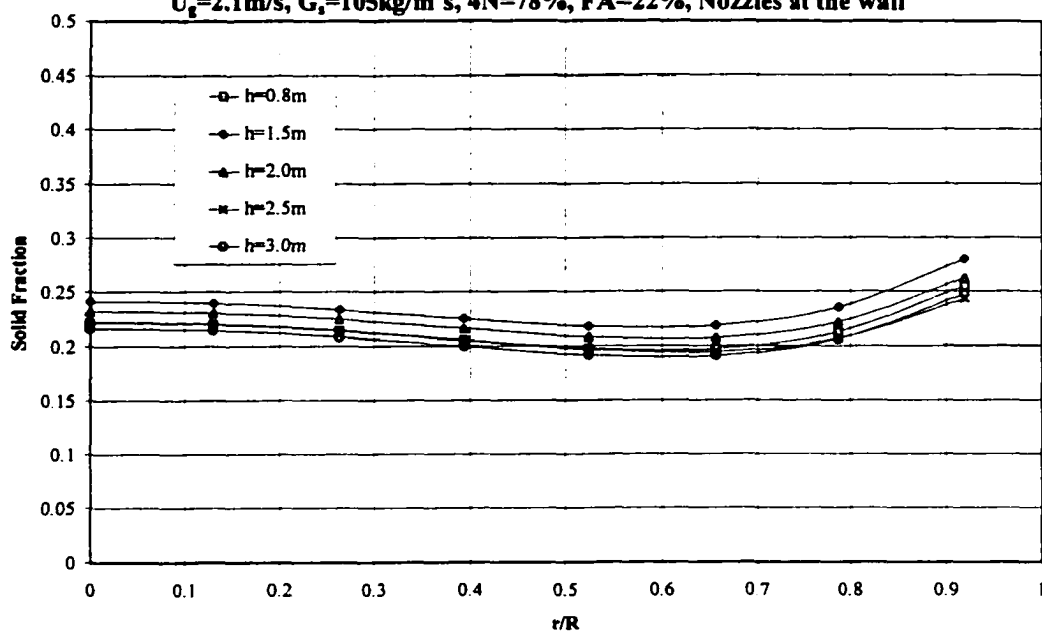
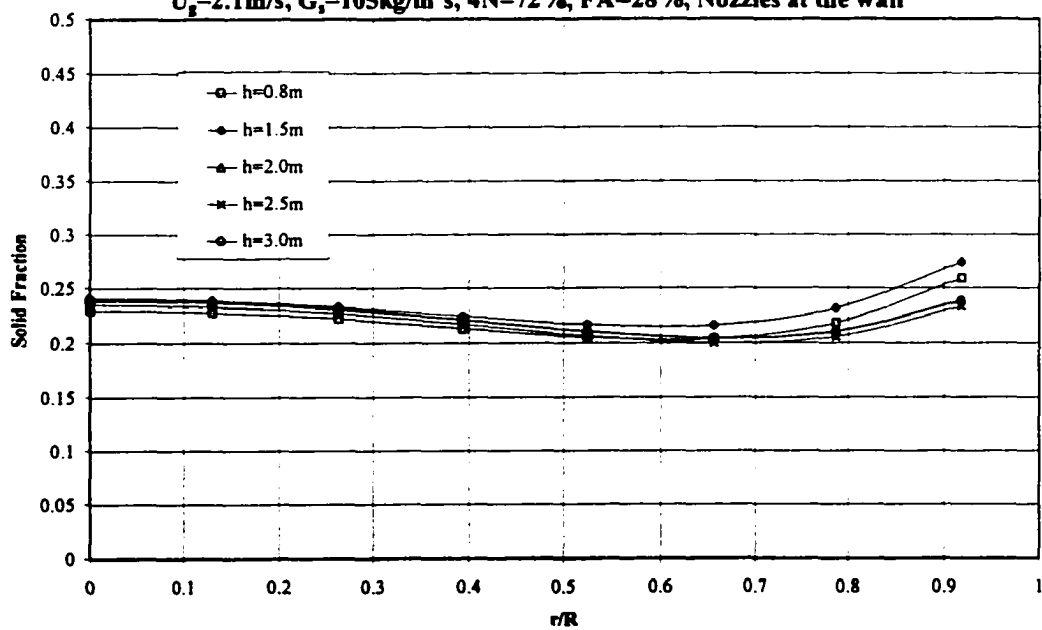
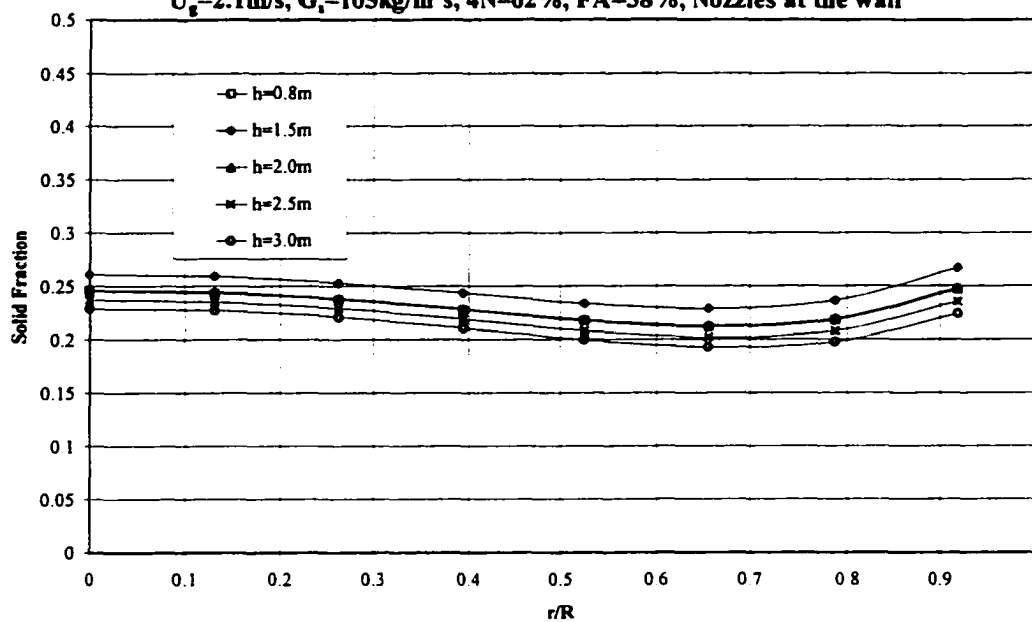
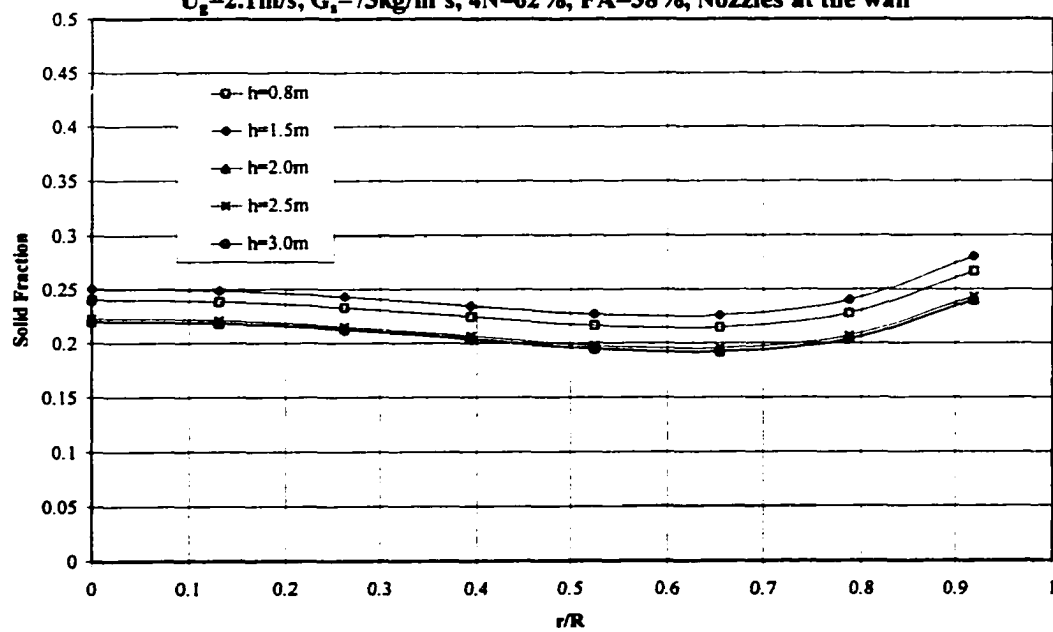
Figure C19: Radial Solid Profile $U_g=2.1\text{m/s}$, $G_s=105\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall**Figure C20: Radial Solid Profile** $U_g=2.1\text{m/s}$, $G_s=105\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall

Figure C21: Radial Solid Profile $U_g=2.1\text{m/s}$, $G_s=105\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the wall**Figure C22: Radial Solid Profile** $U_g=2.1\text{m/s}$, $G_s=73\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the wall

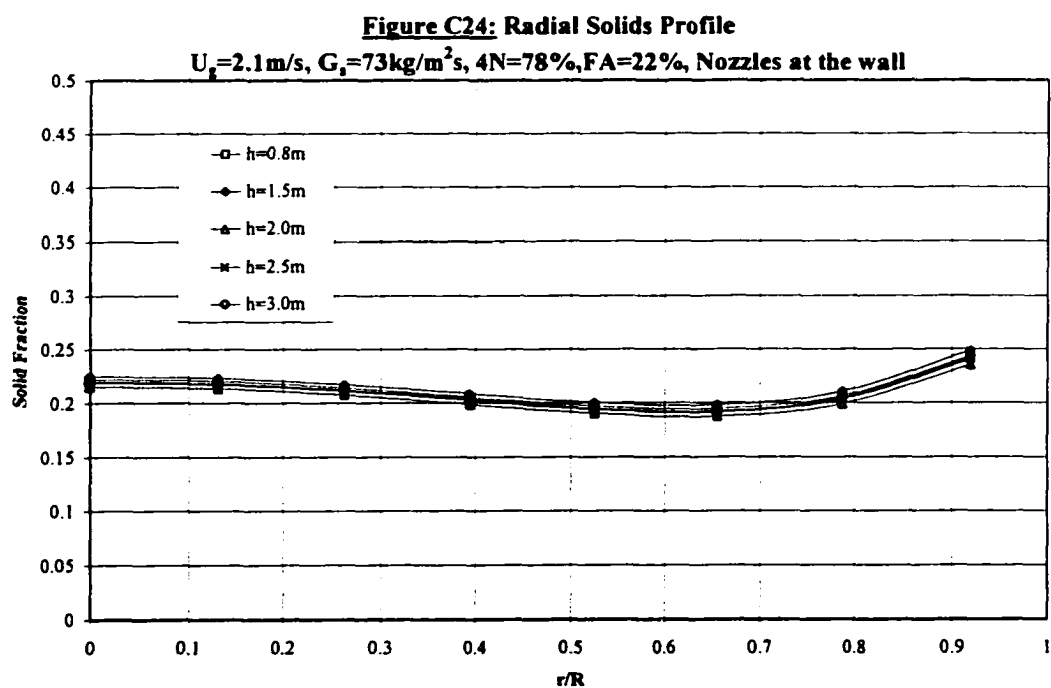
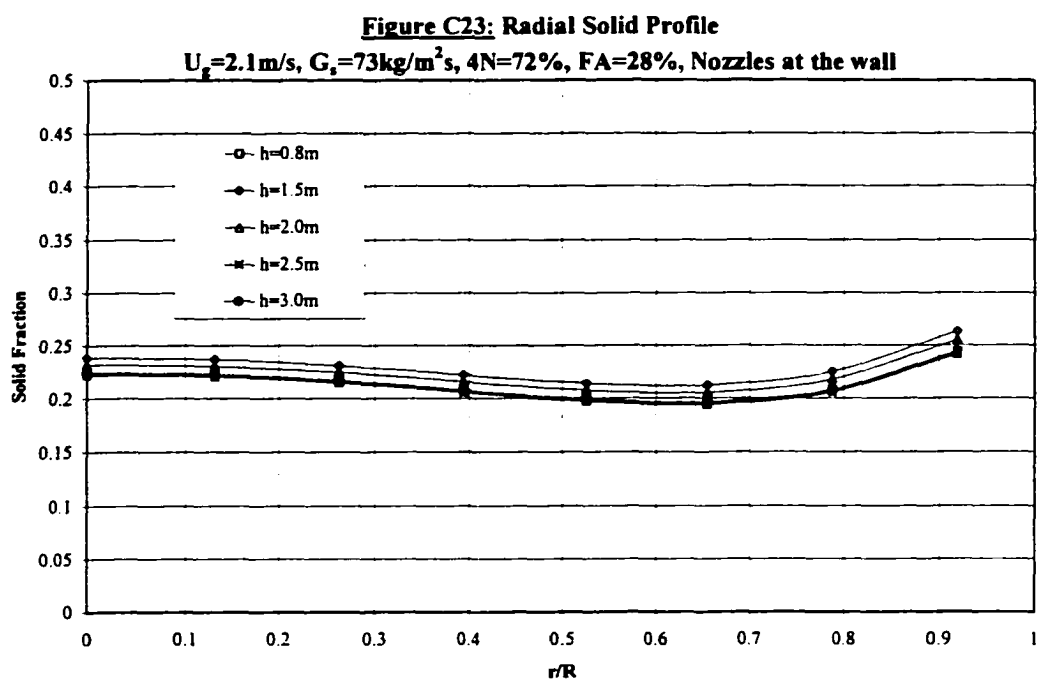
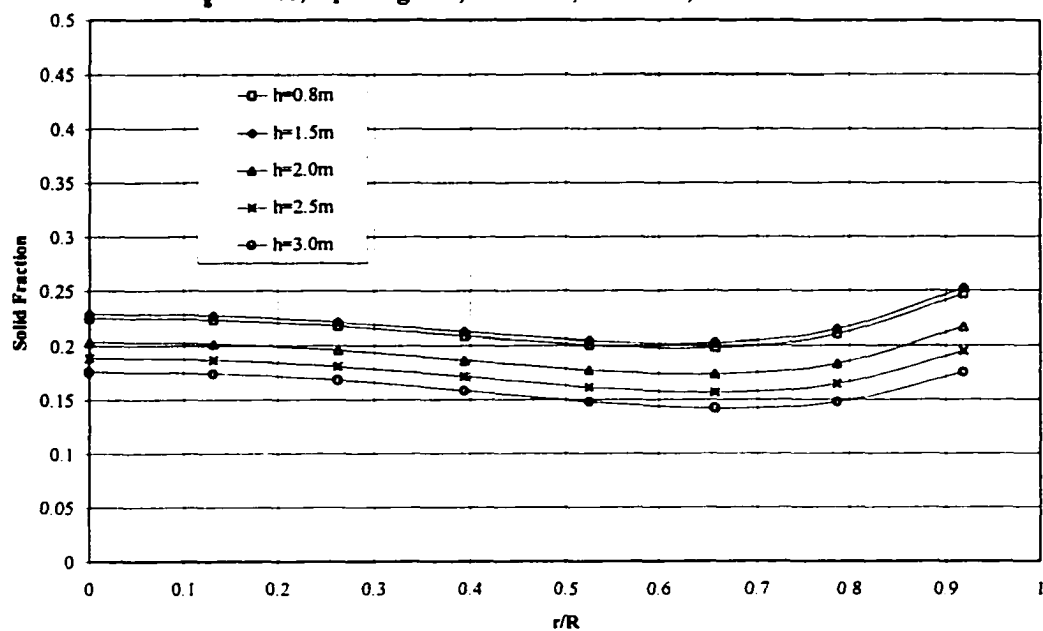
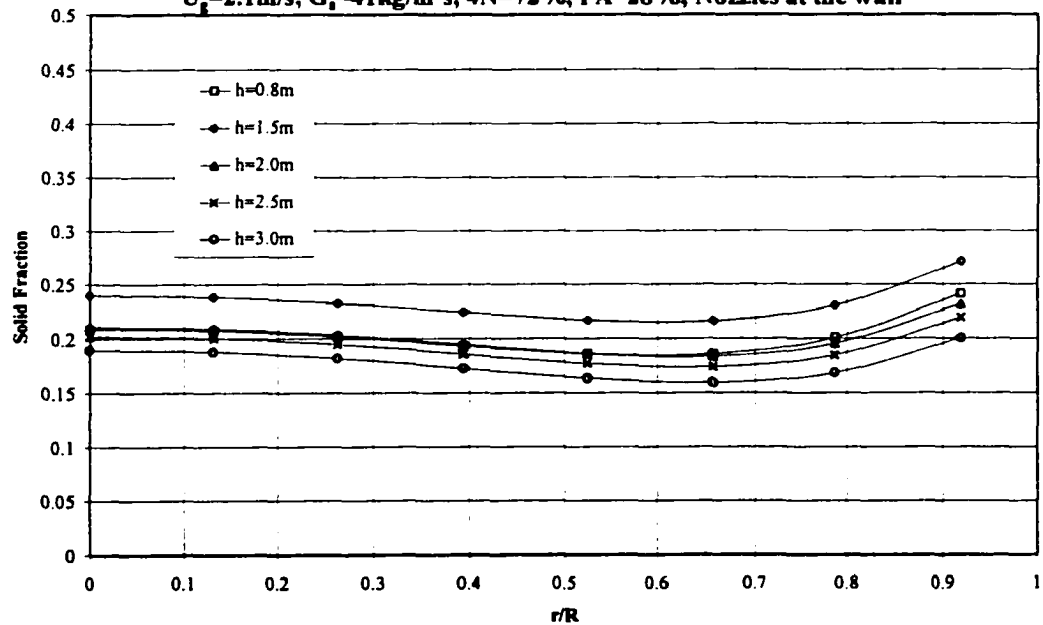
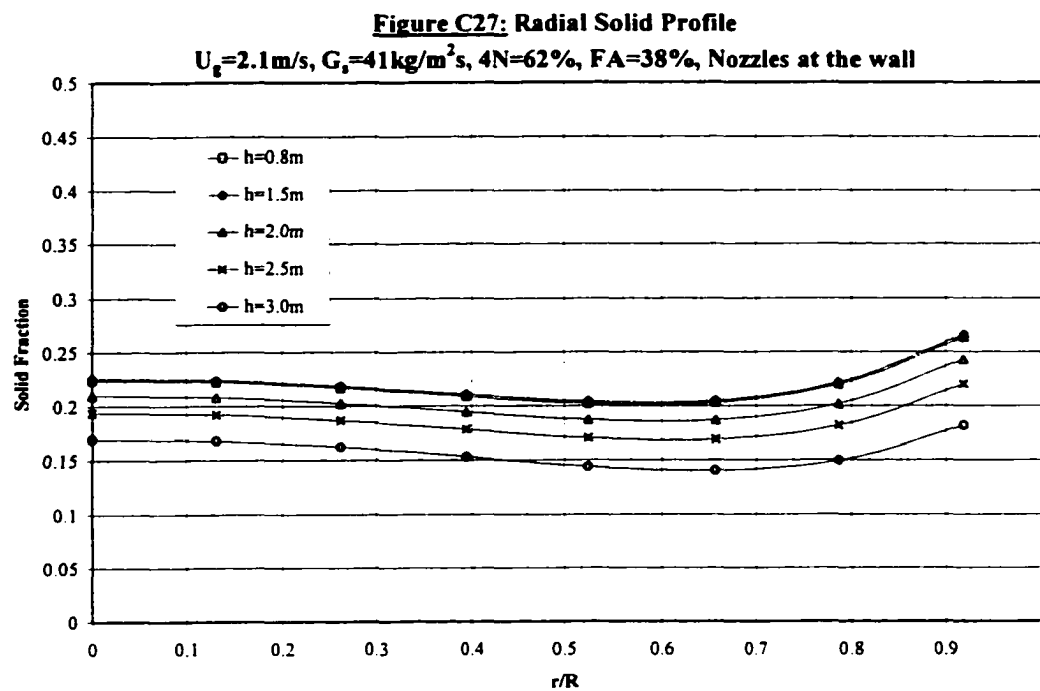
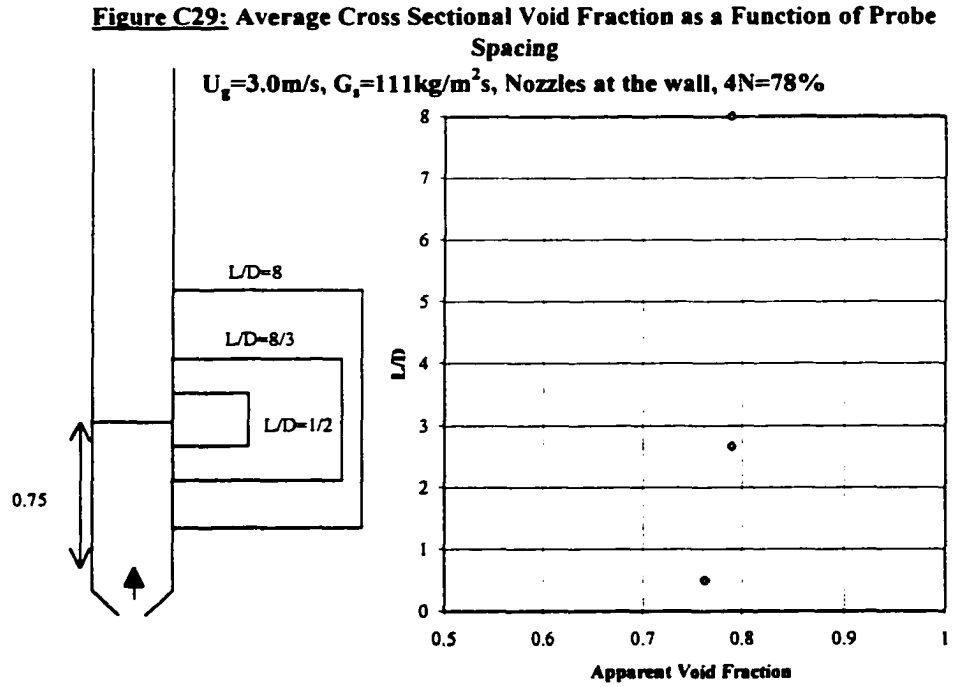
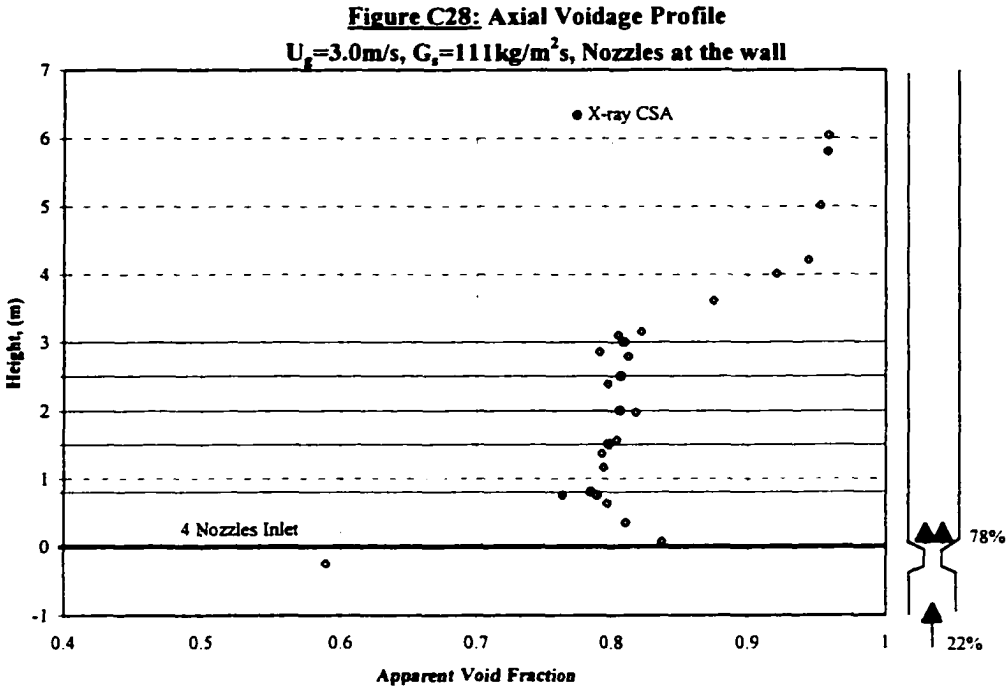
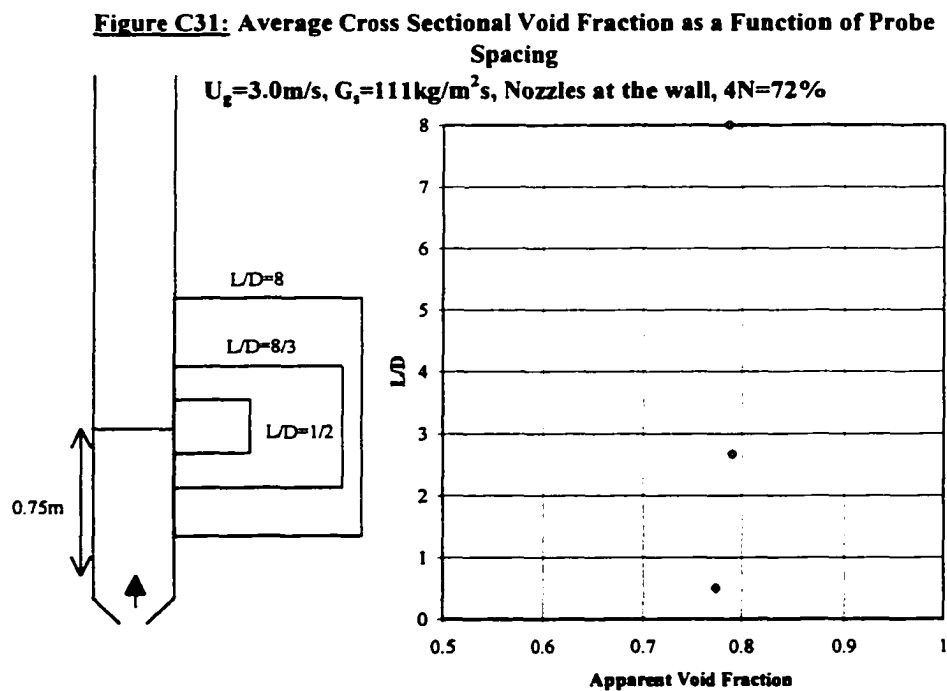
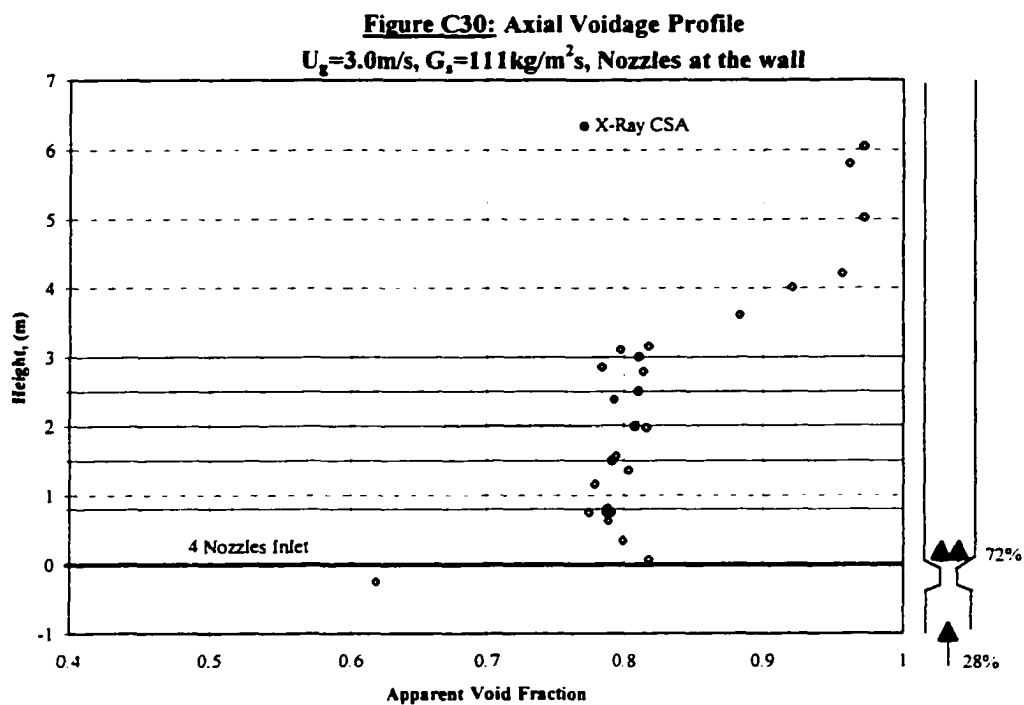
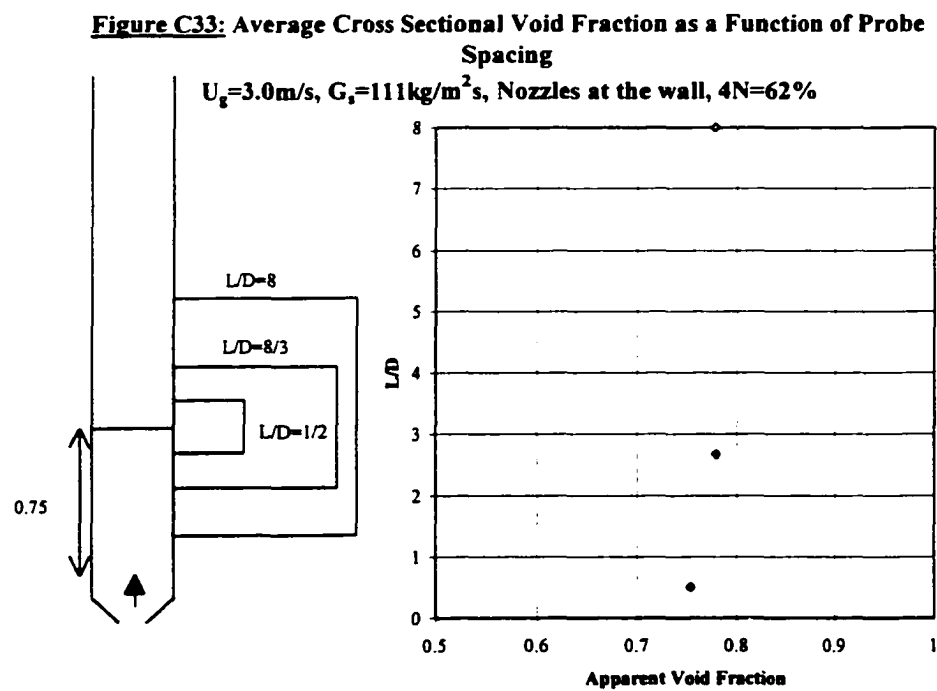
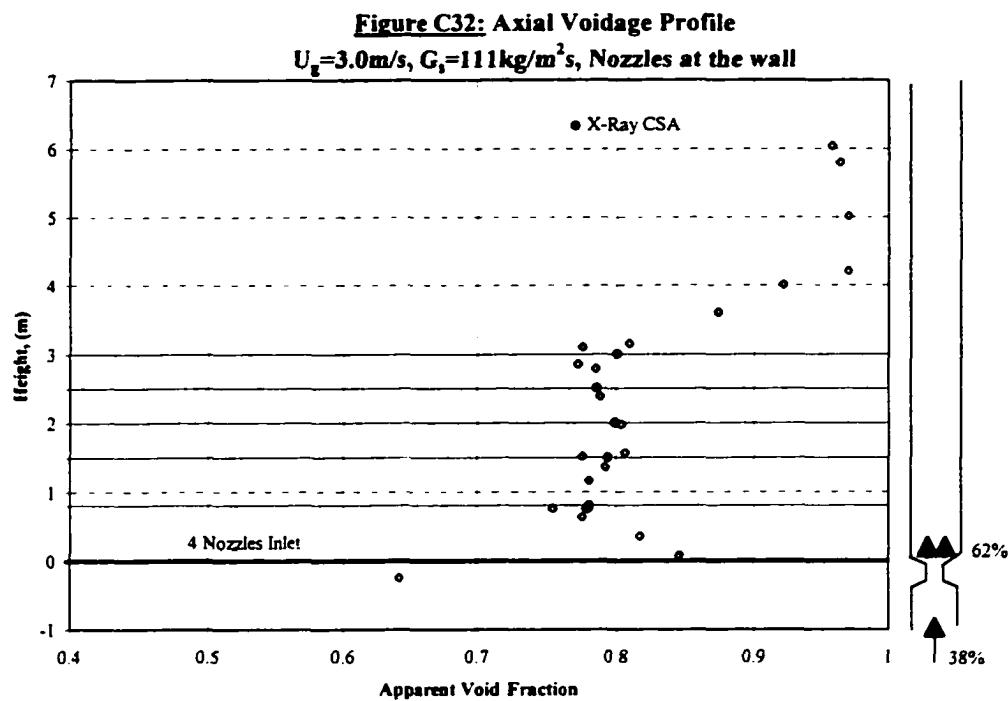


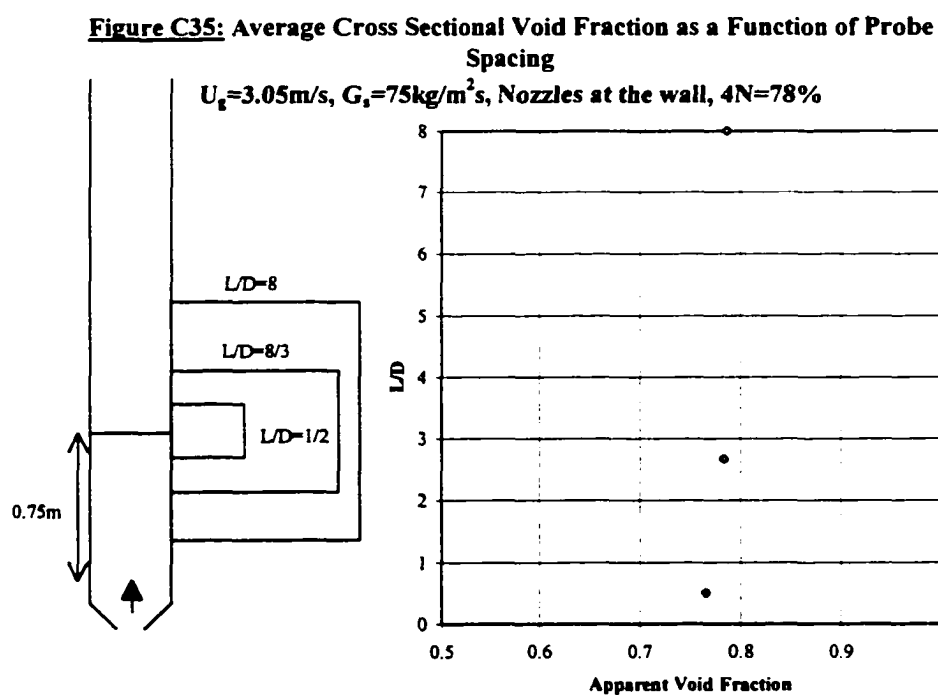
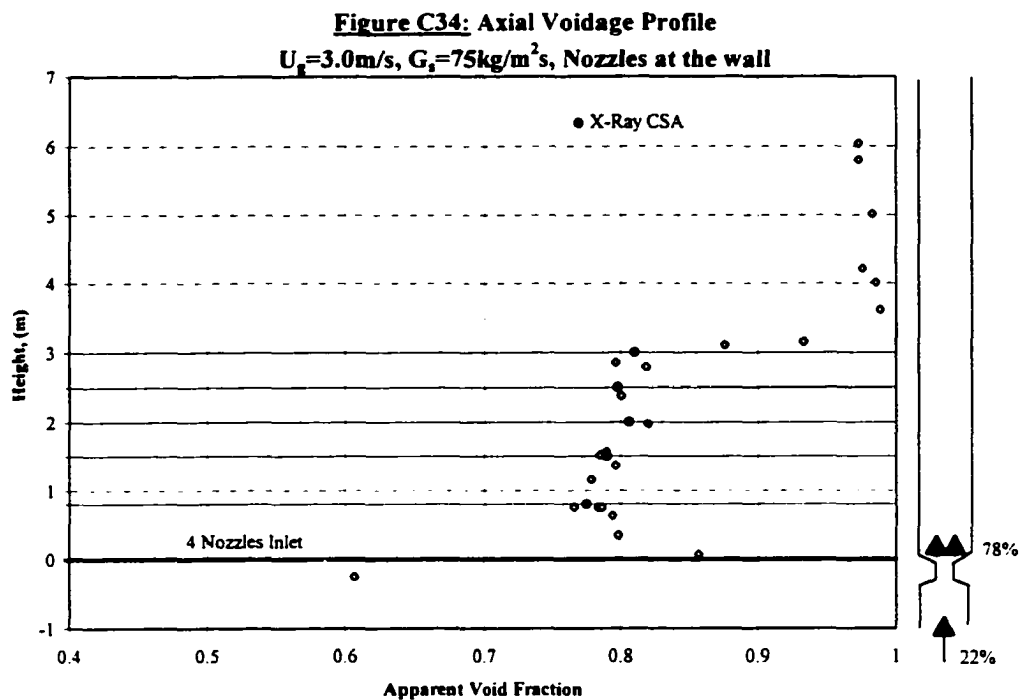
Figure C25: Radial Solid Profile $U_g=2.1\text{m/s}$, $G_s=41\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall**Figure C26: Radial Solid Profile** $U_g=2.1\text{m/s}$, $G_s=41\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall

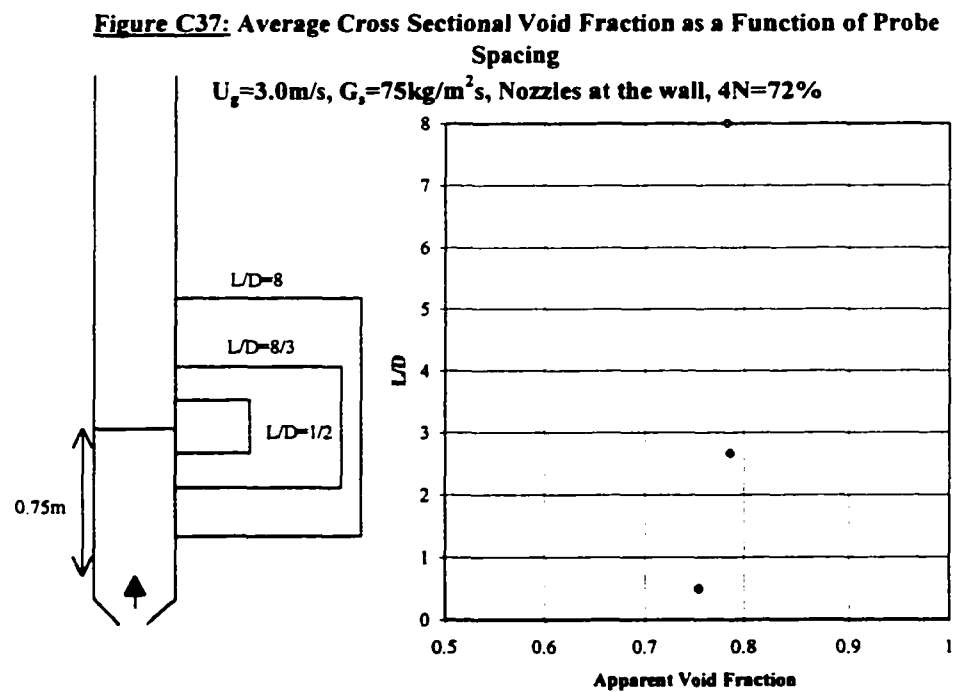
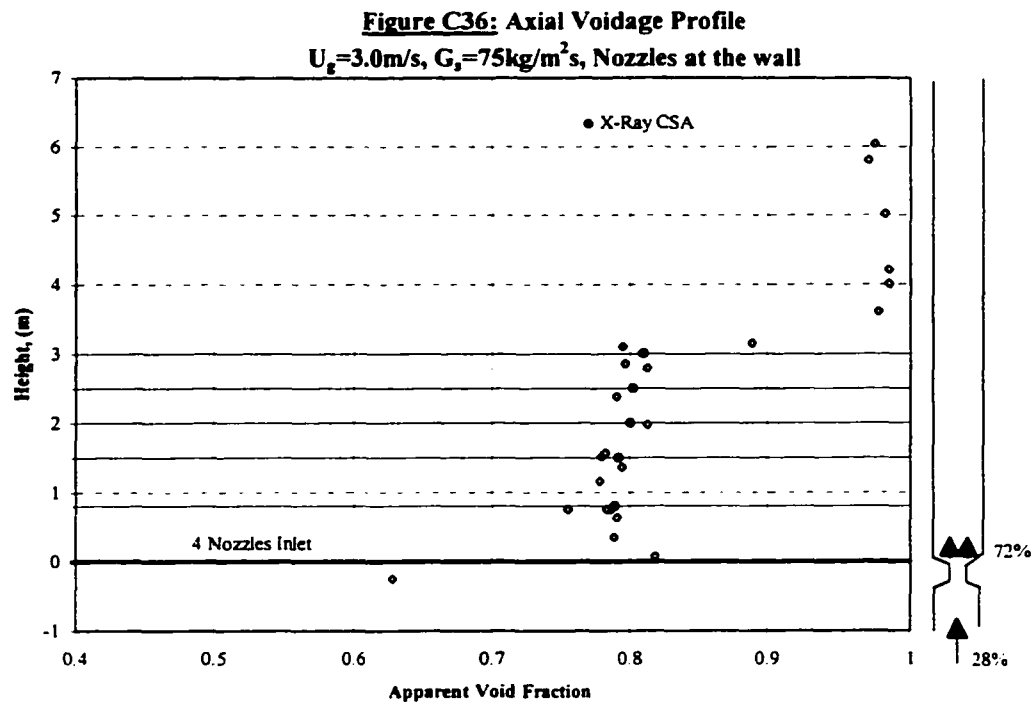


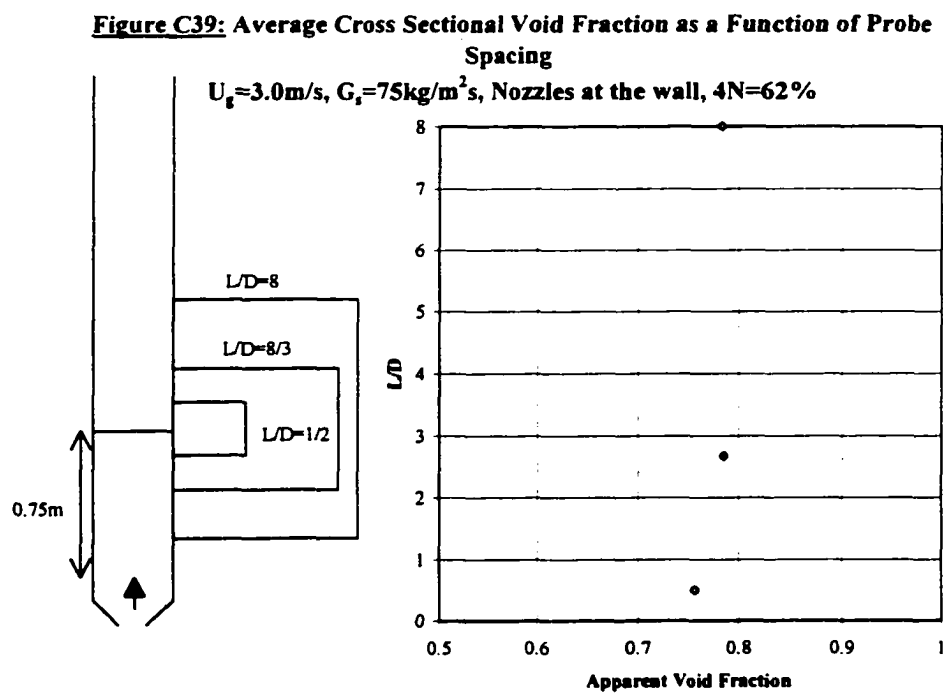
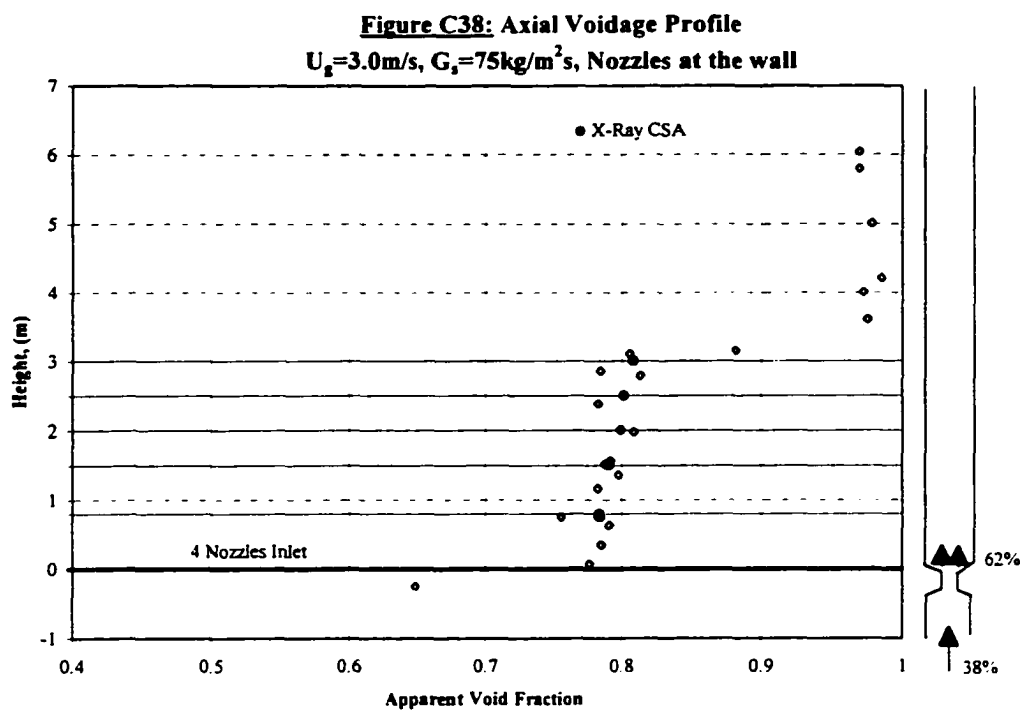


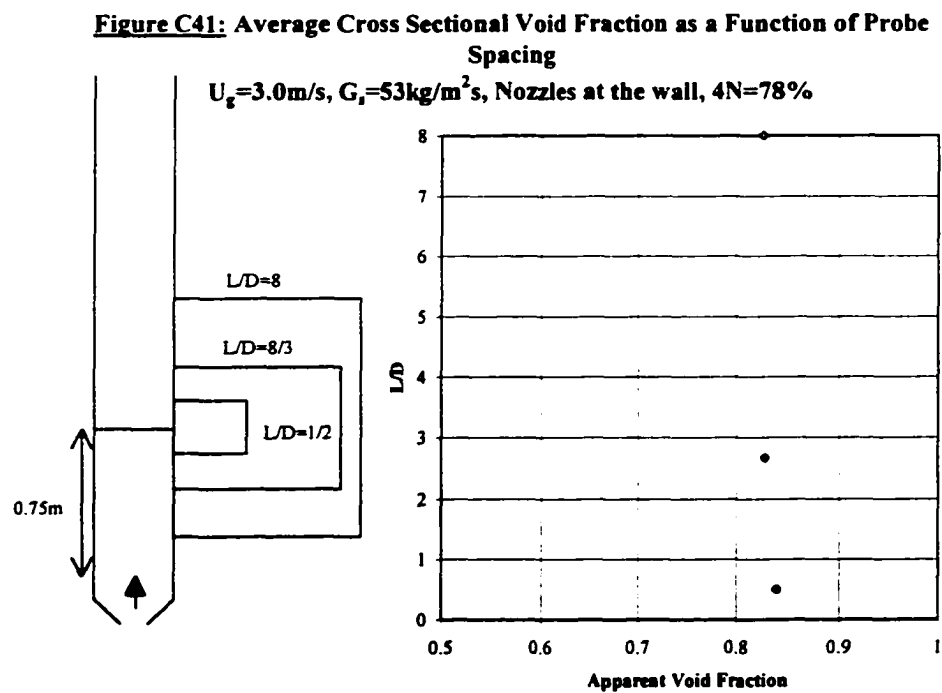
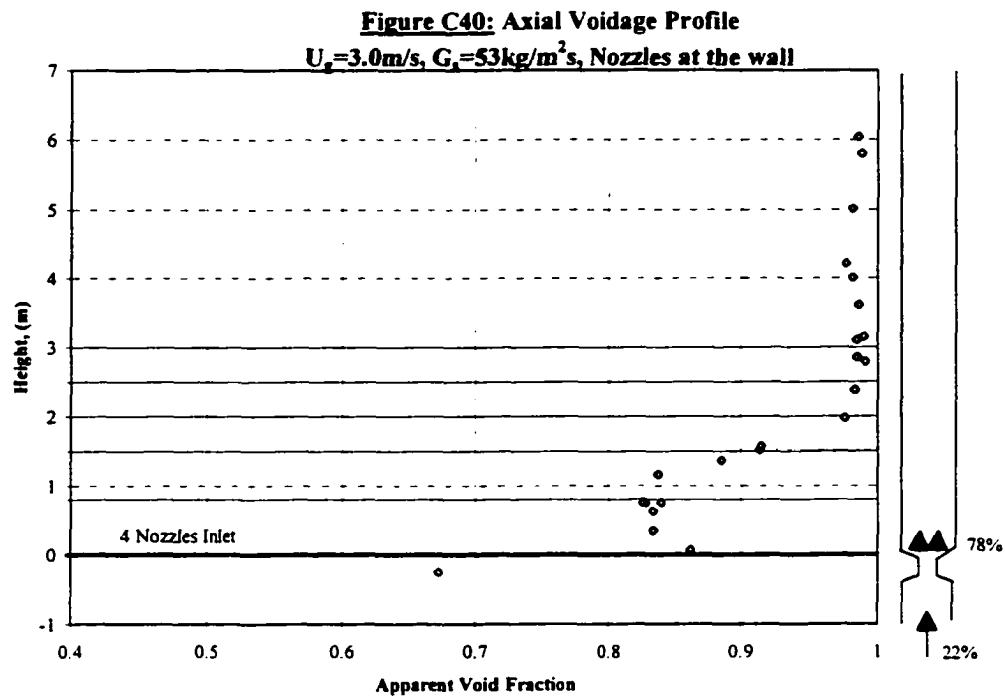












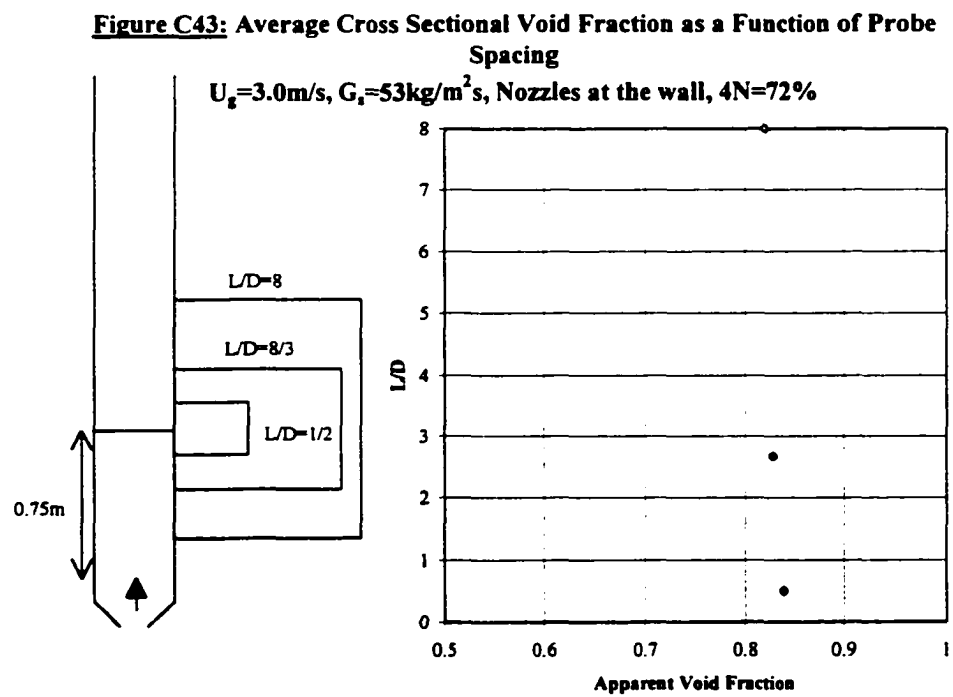
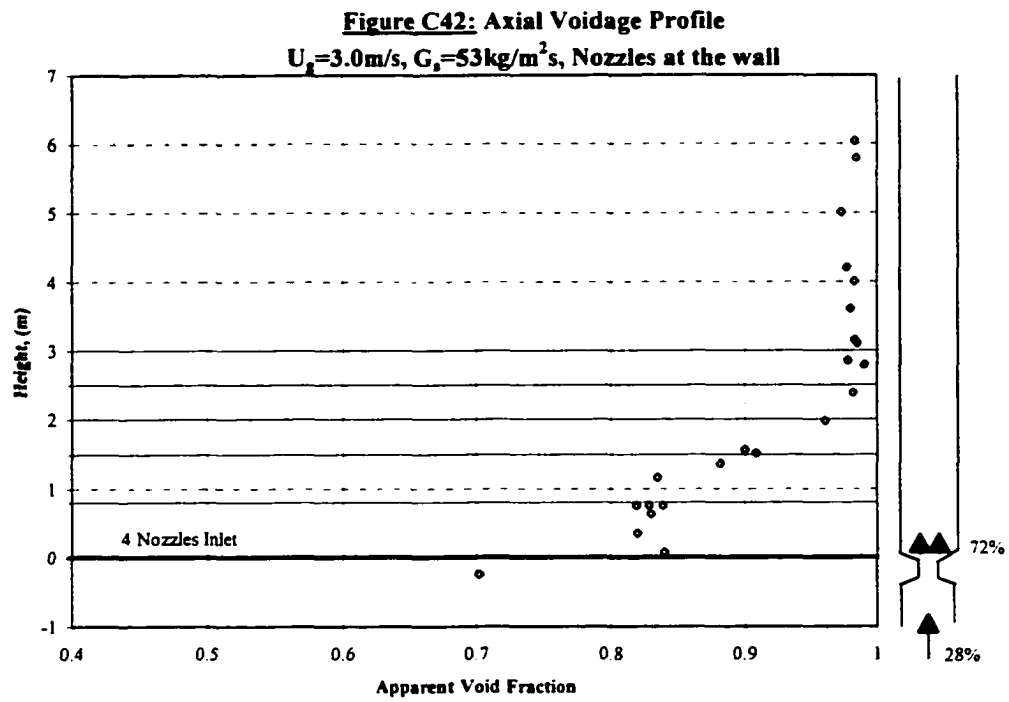


Figure C44: Axial Voidage Profile
 $U_g=3.0\text{m/s}$, $G_s=53\text{kg/m}^2\text{s}$, Nozzles at the wall

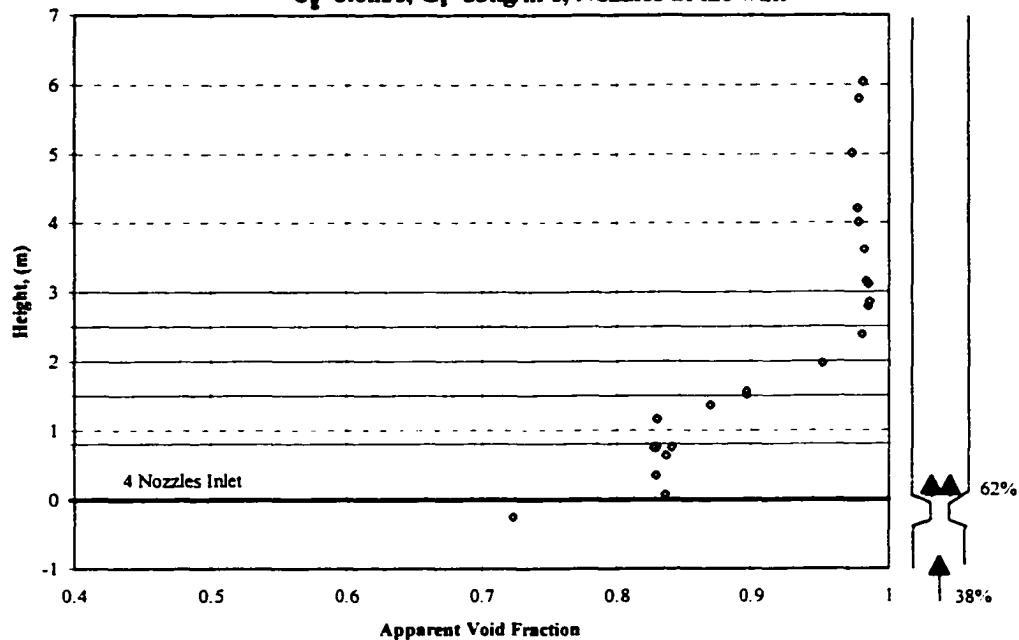


Figure C45: Average Cross Sectional Void Fraction as a Function of Probe Spacing

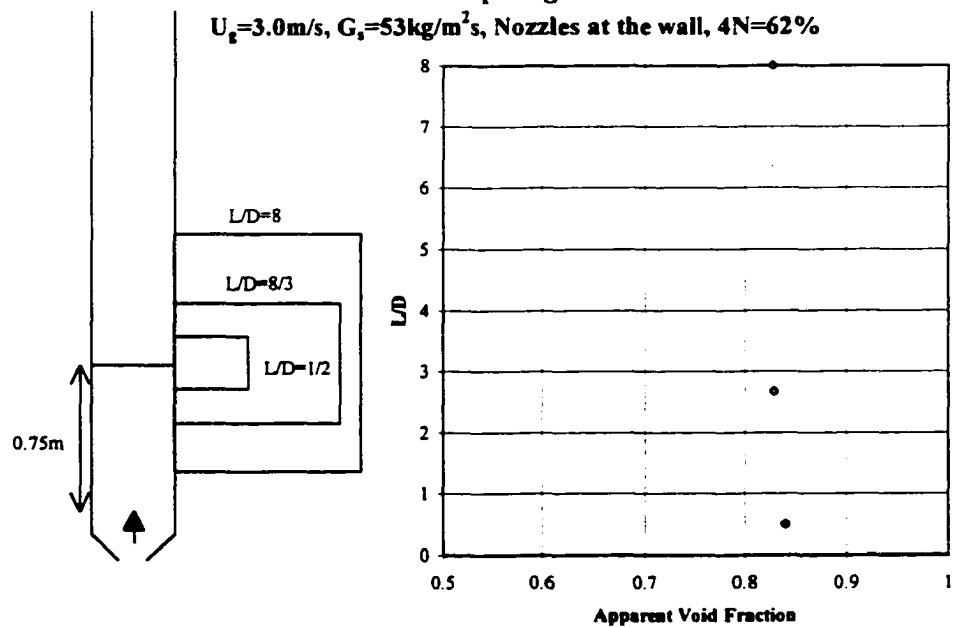


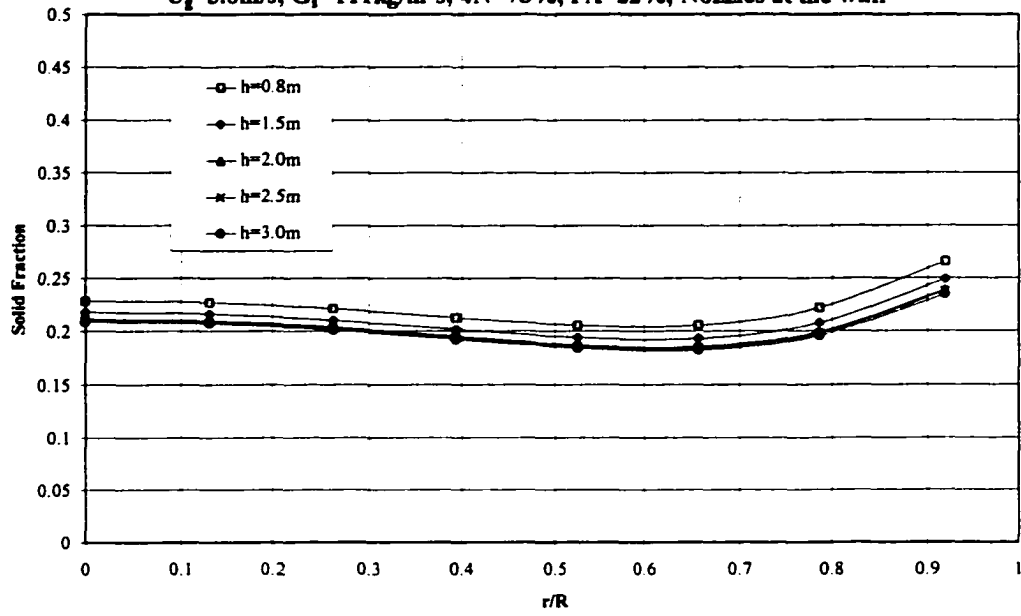
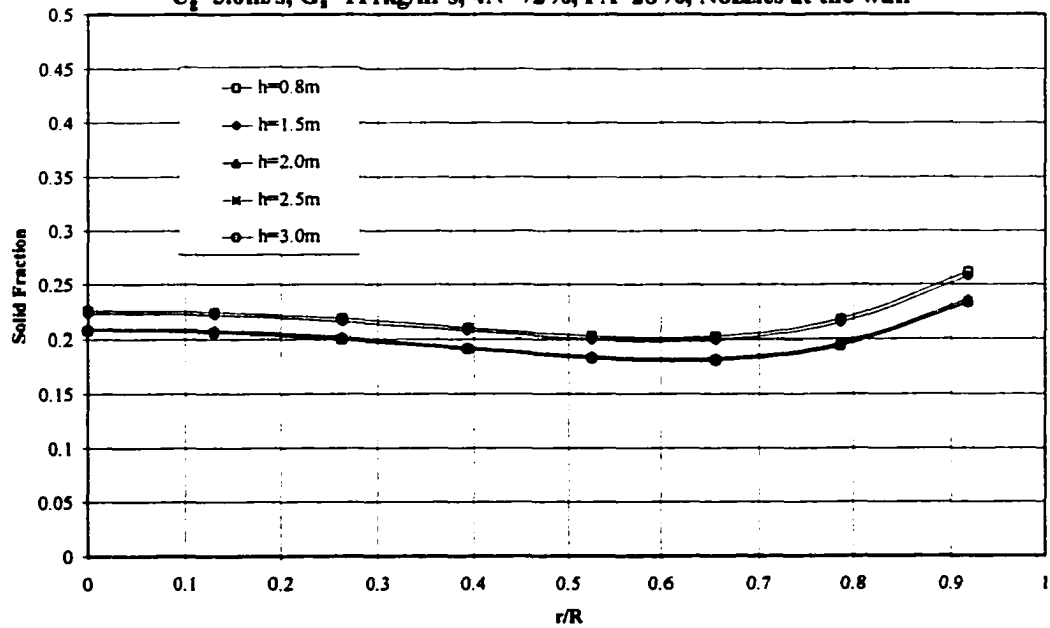
Figure C46: Radial Solid Profile $U_t=3.0\text{m/s}$, $G_t=111\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall**Figure C47: Radial Solid Profile** $U_t=3.0\text{m/s}$, $G_t=111\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall

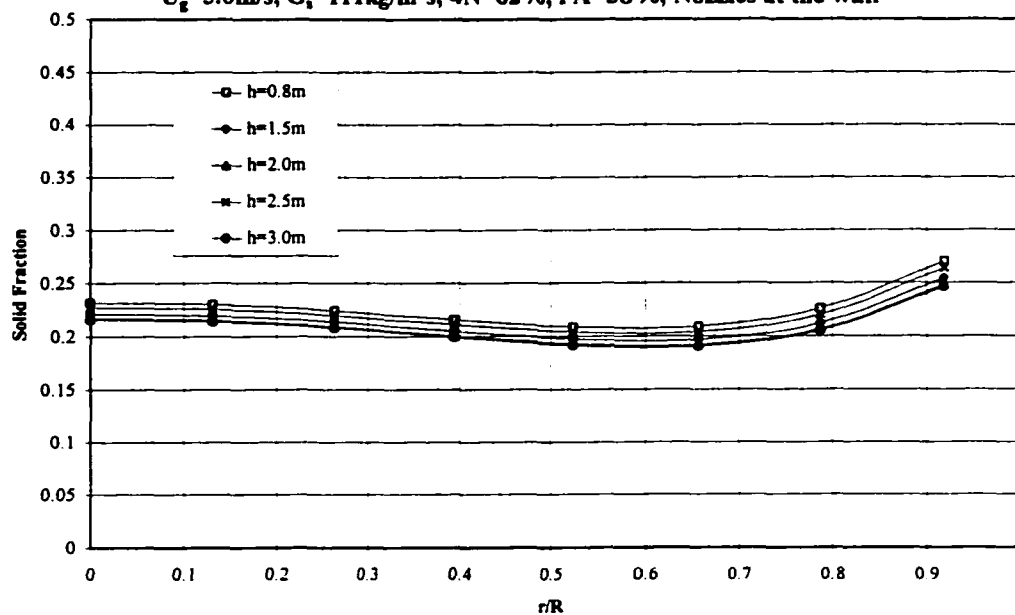
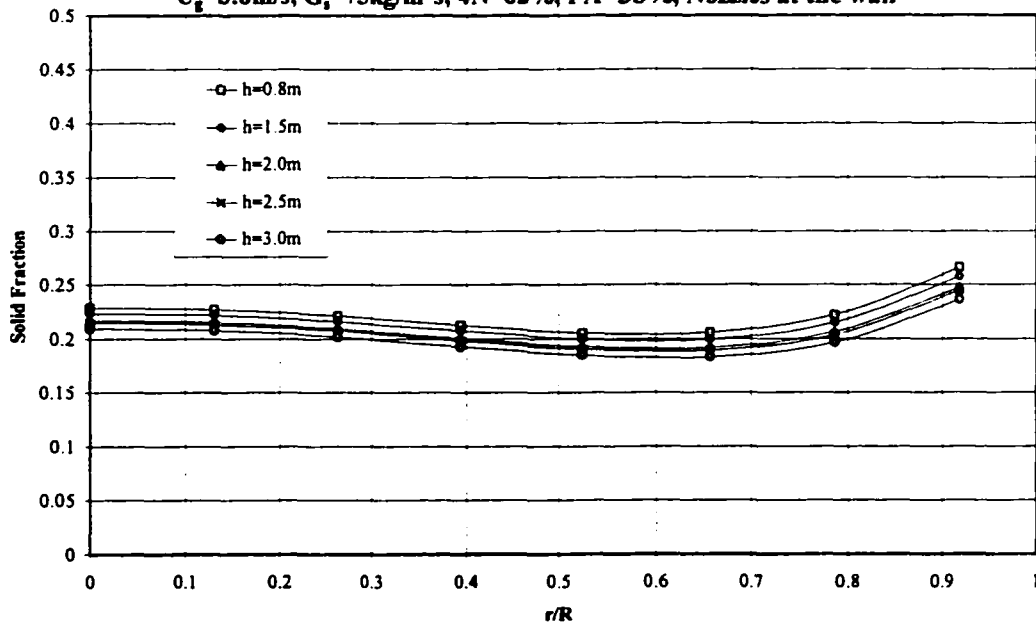
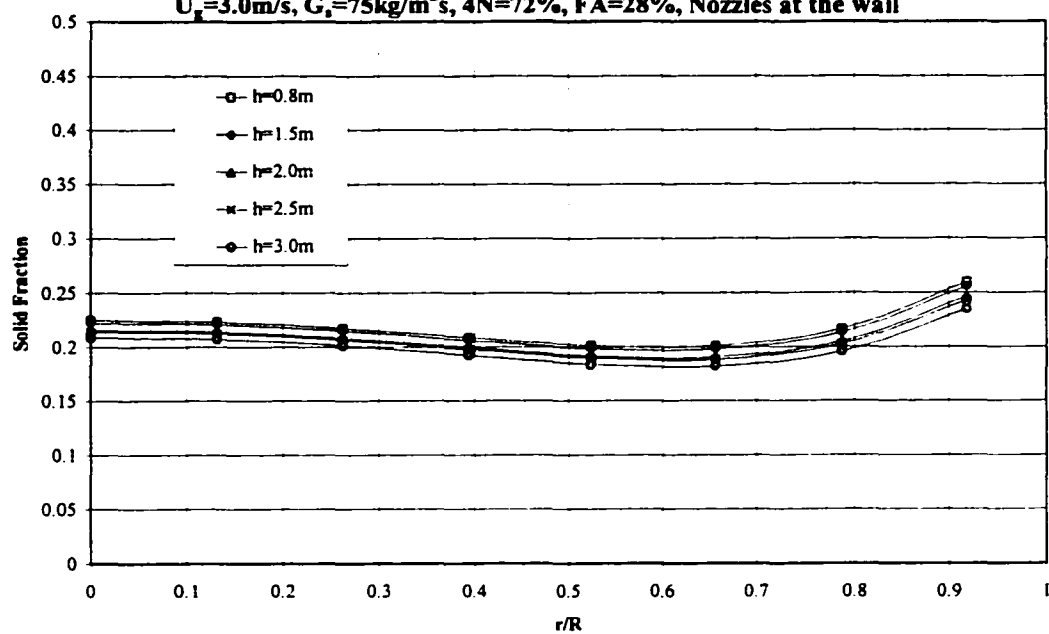
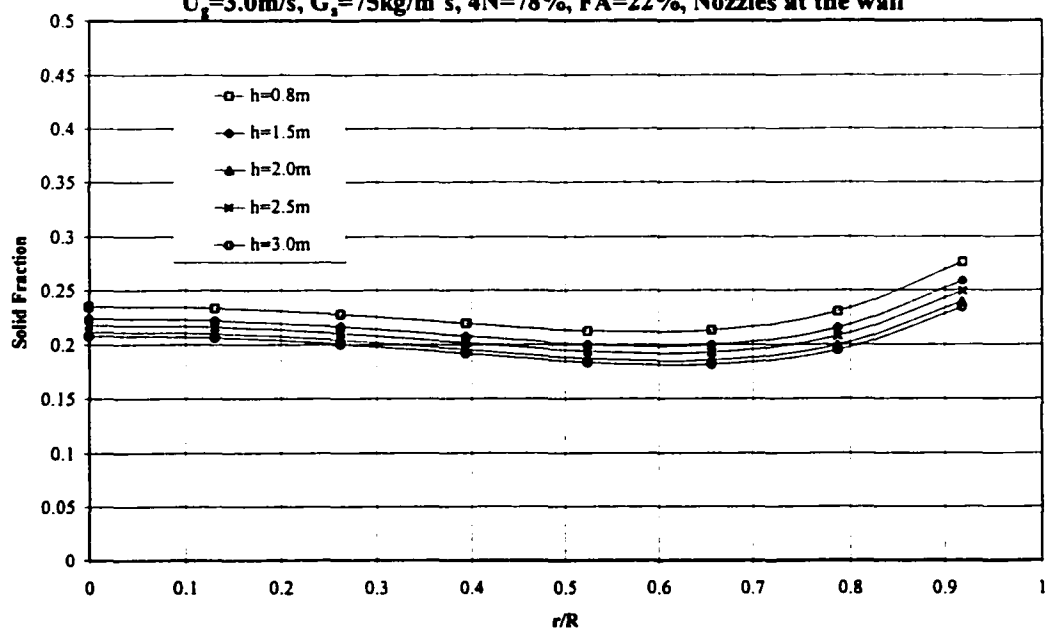
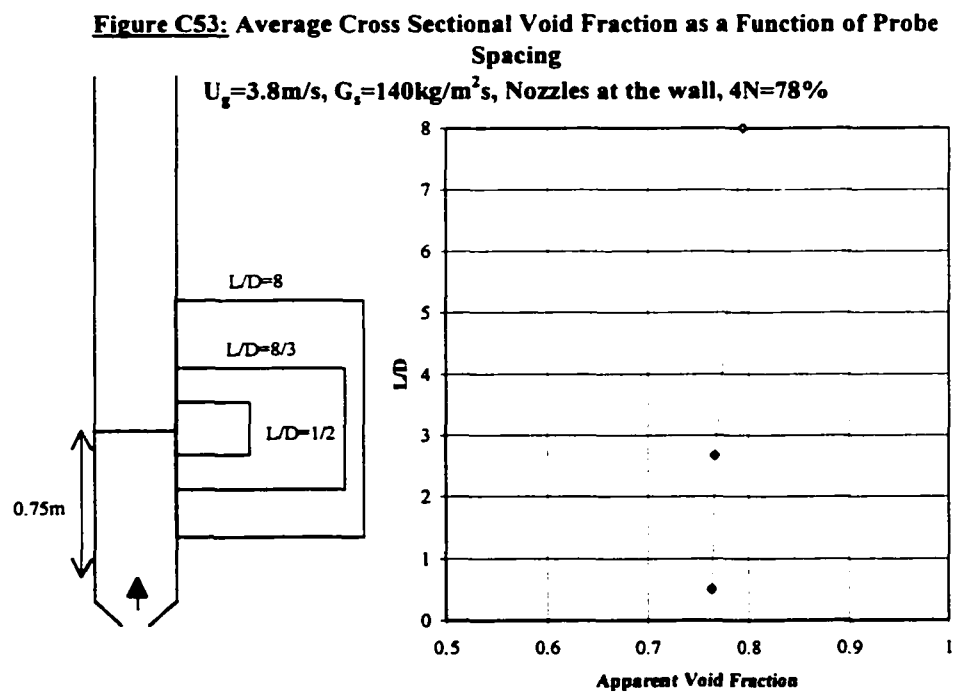
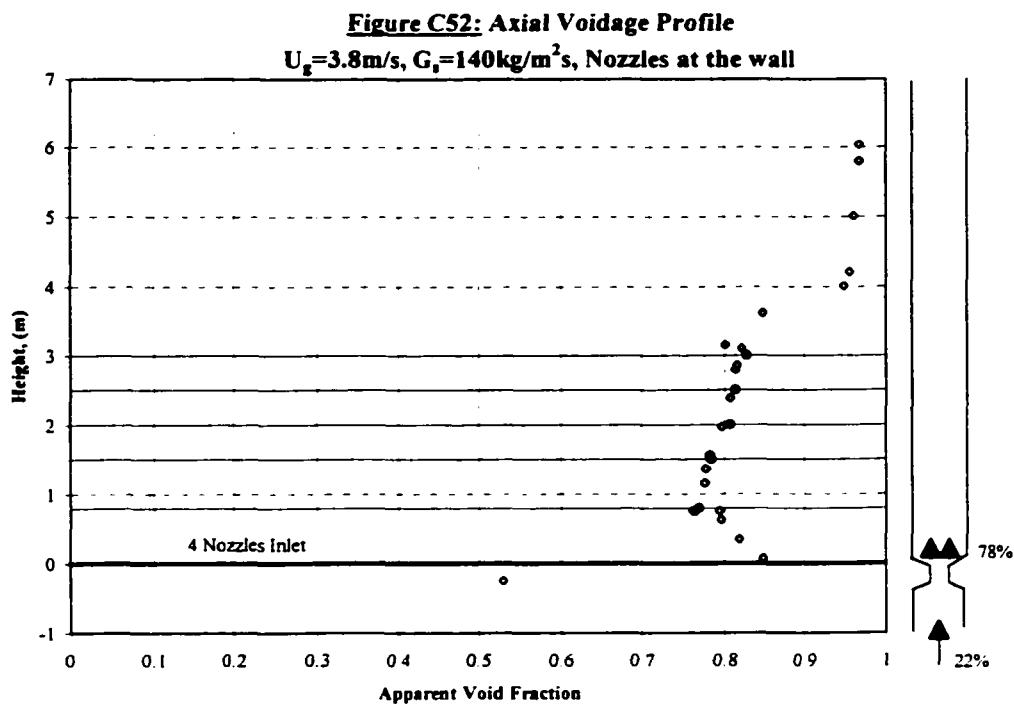
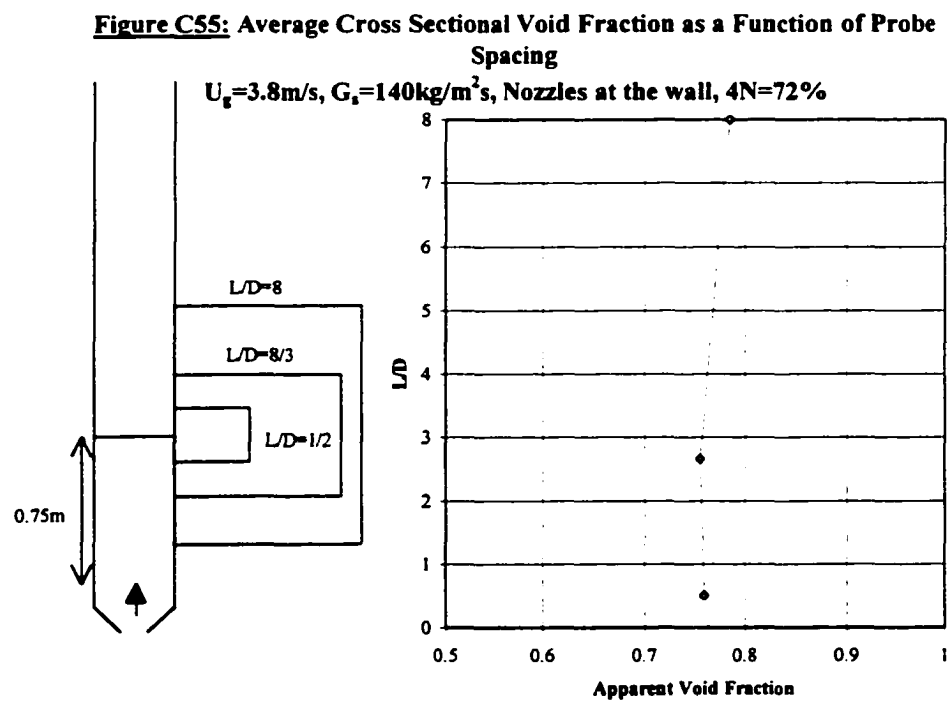
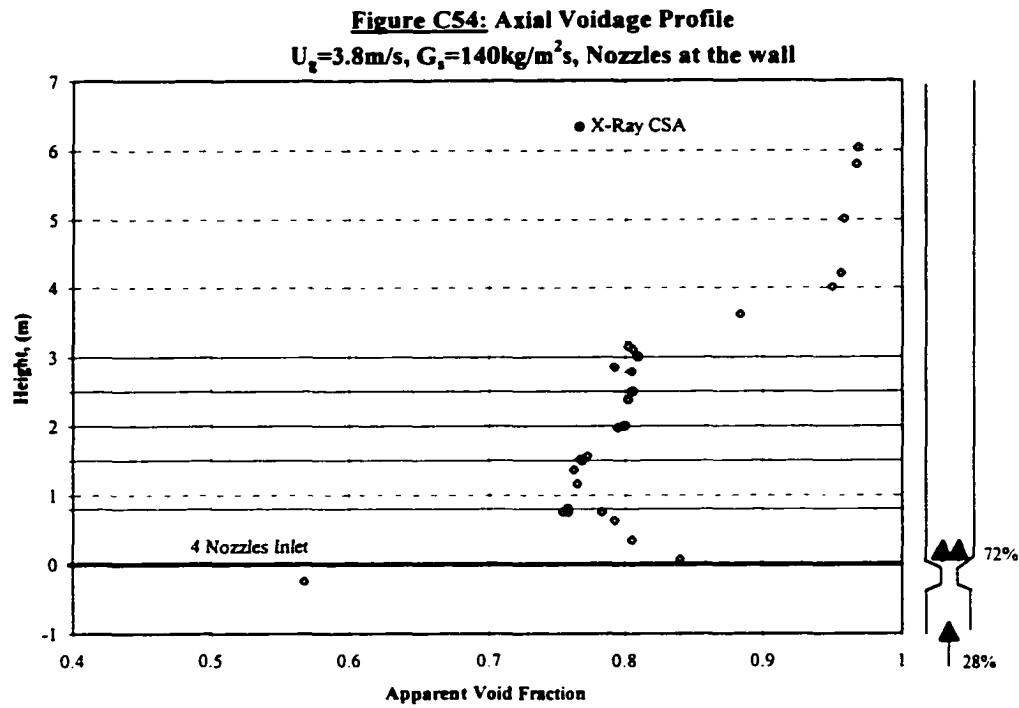
Figure C48: Radial Solid Profile $U_t=3.0\text{m/s}$, $G_t=111\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the wall**Figure C49: Radial Solid Profile** $U_t=3.0\text{m/s}$, $G_t=75\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the wall

Figure C50: Radial Solid Profile $U_s=3.0\text{m/s}$, $G_s=75\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall**Figure C51: Radial Solid Profile** $U_s=3.0\text{m/s}$, $G_s=75\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall





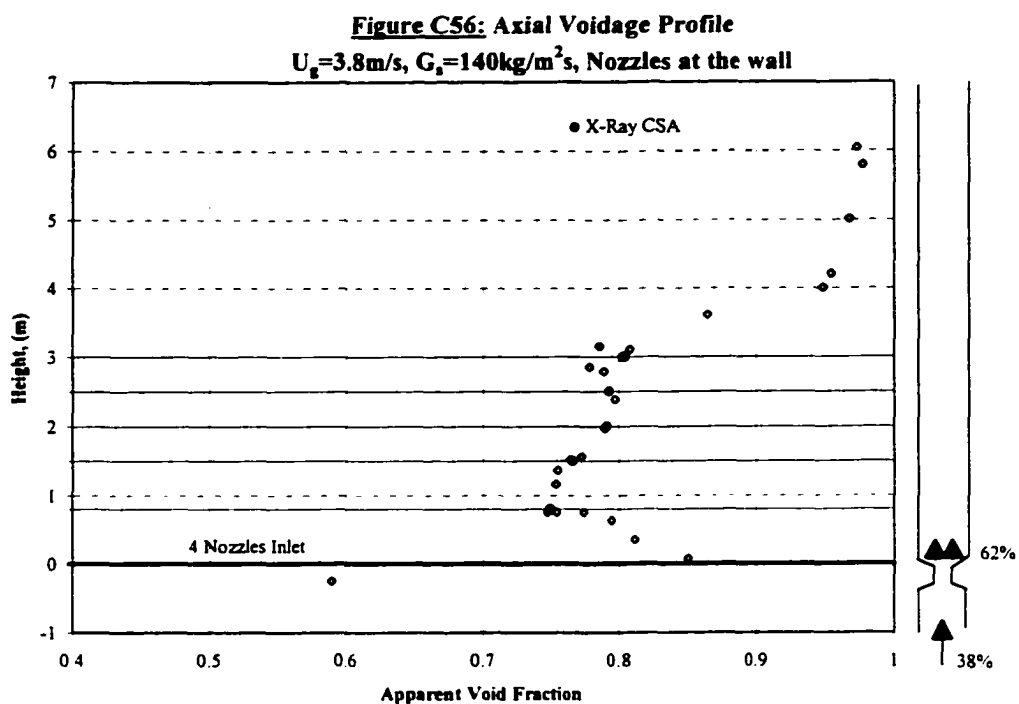


Figure C57: Average Cross Sectional Void Fraction as a Function of Probe Spacing

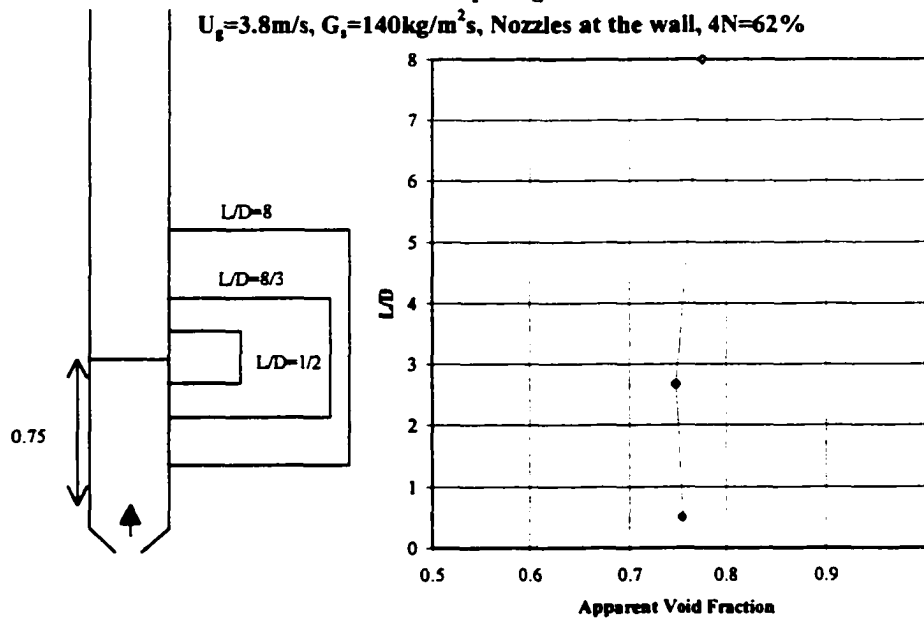


Figure C58: Axial Voidage Profile
 $U_g=3.8\text{m/s}$, $G_g=98\text{kg/m}^2\text{s}$, Nozzles at the wall

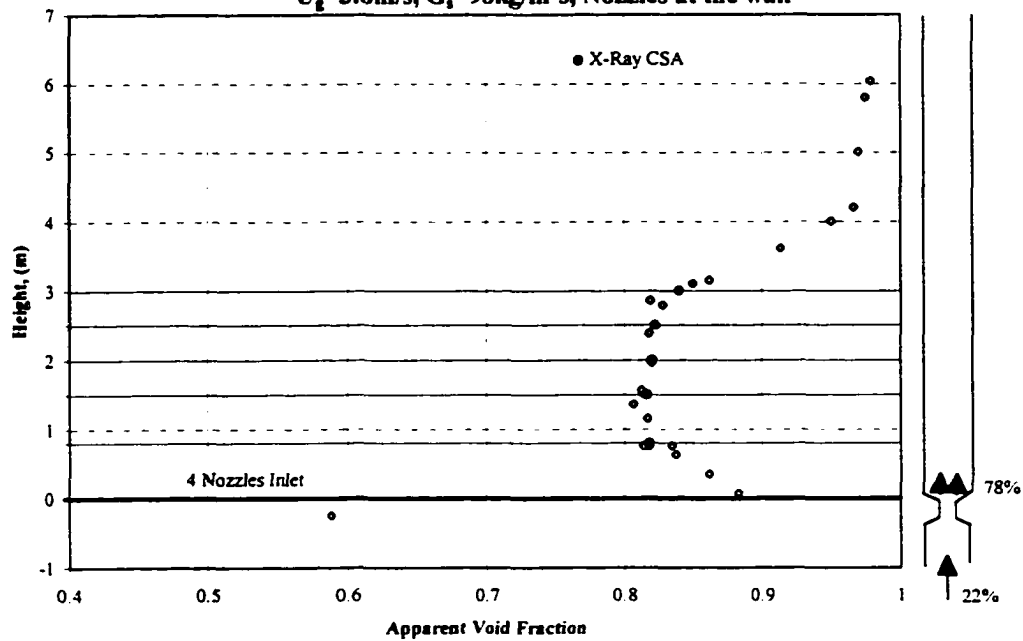
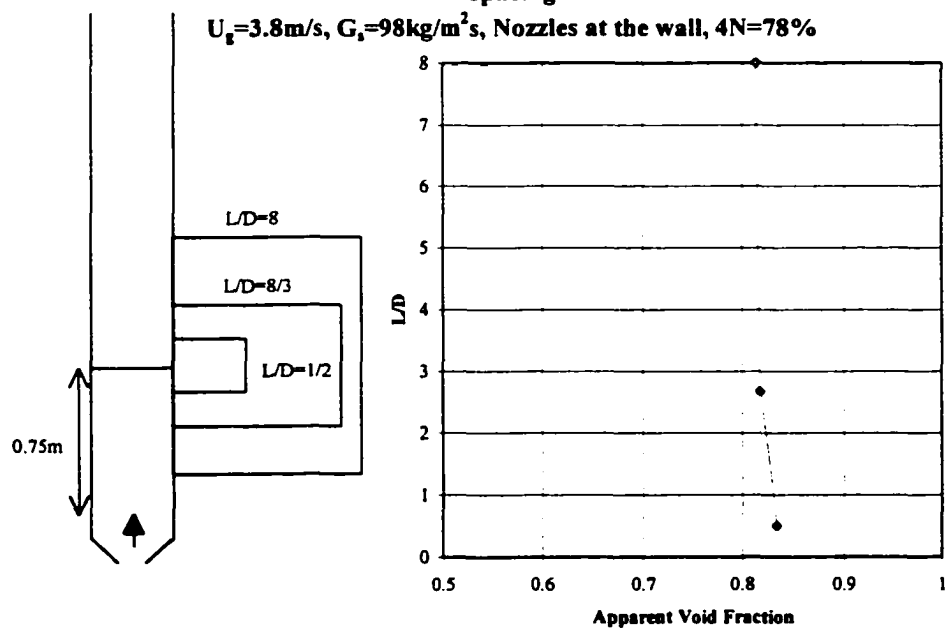
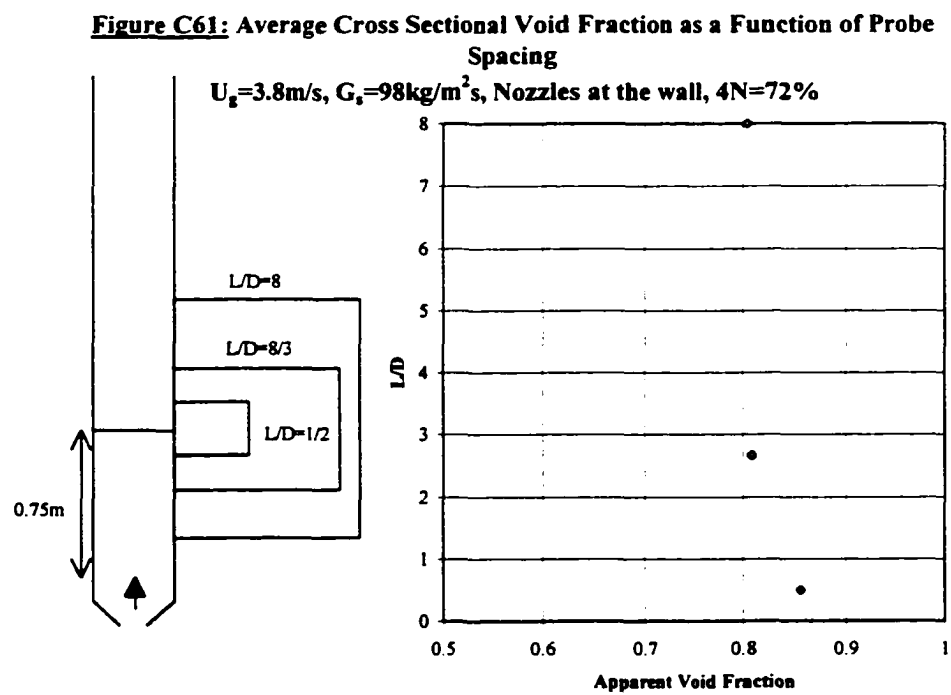
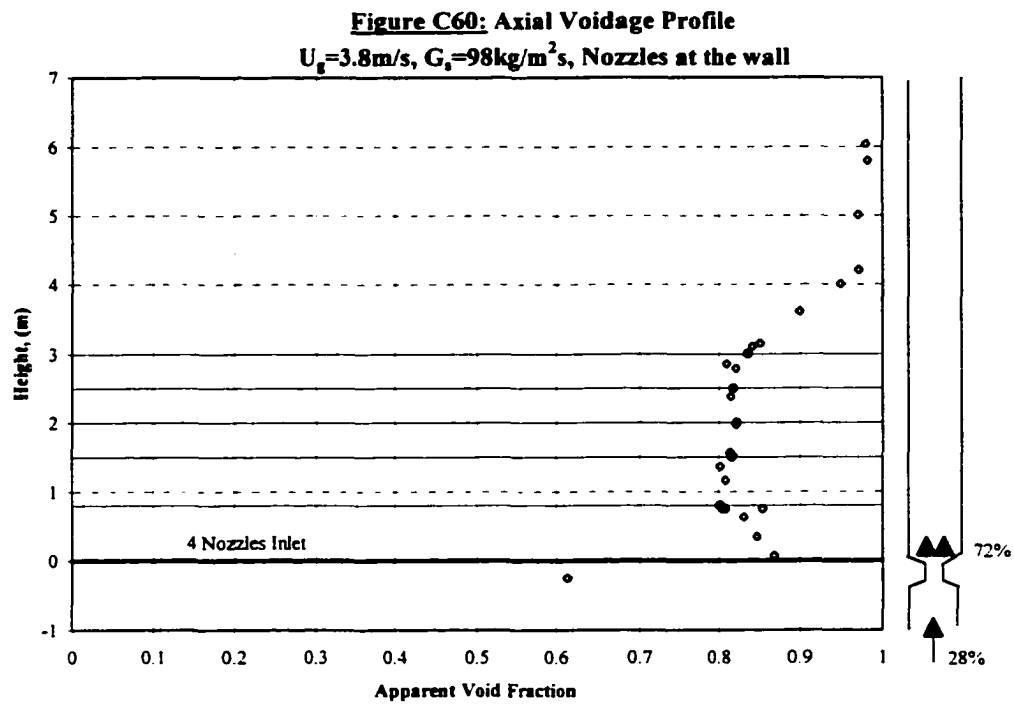
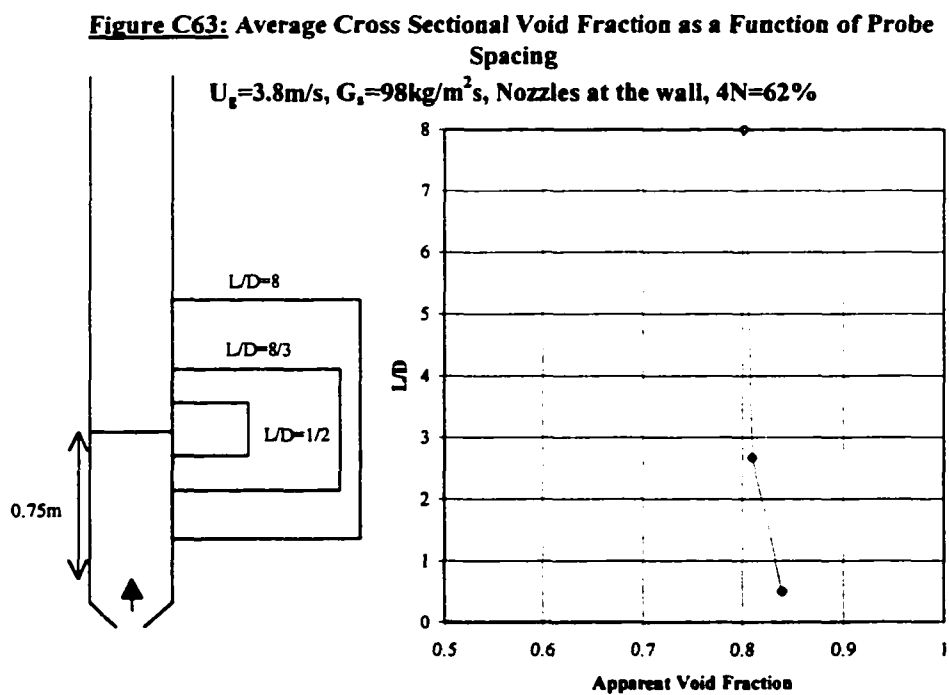
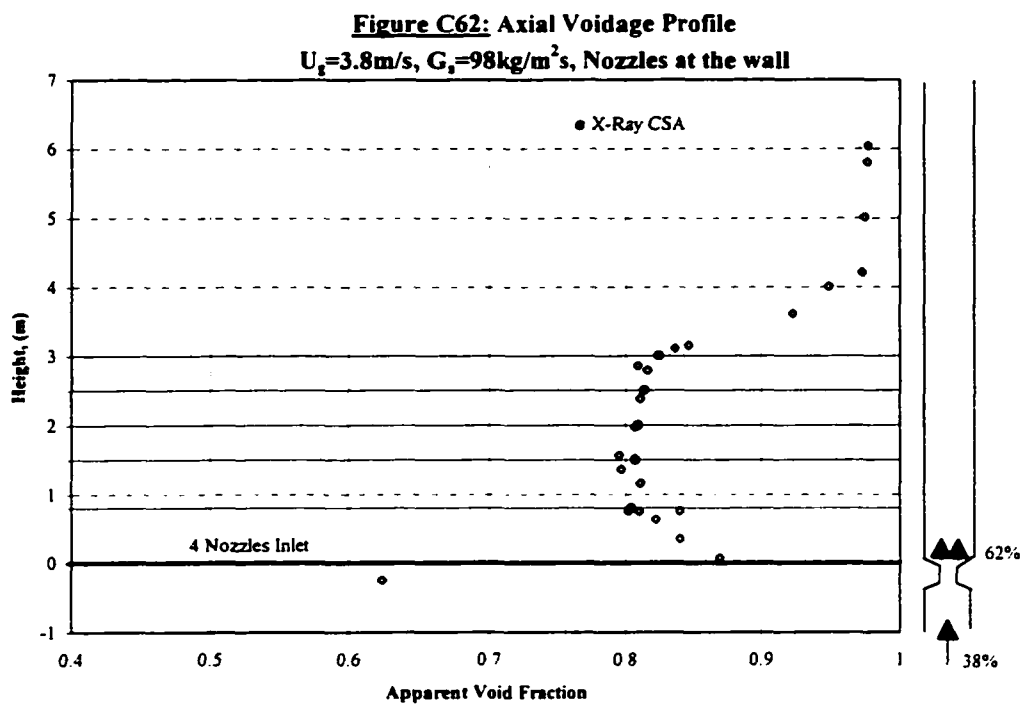


Figure C59: Average Cross Sectional Void Fraction as a Function of Probe Spacing







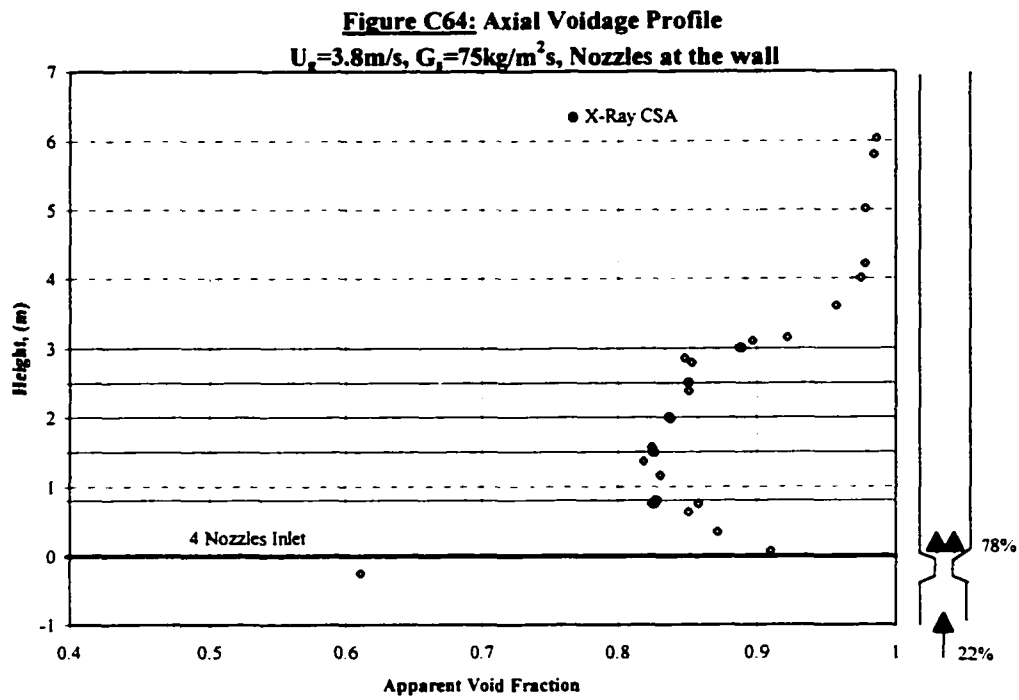
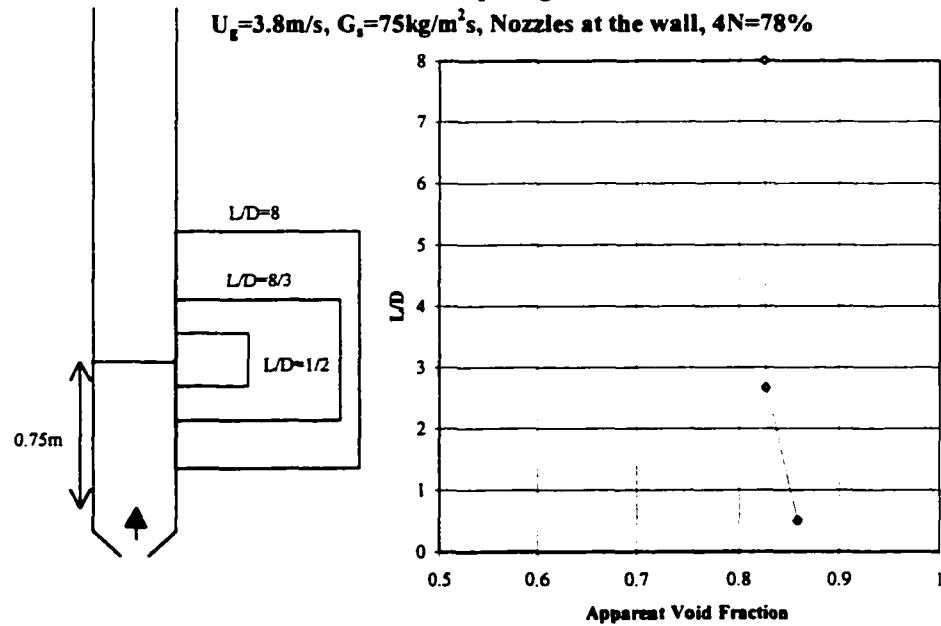
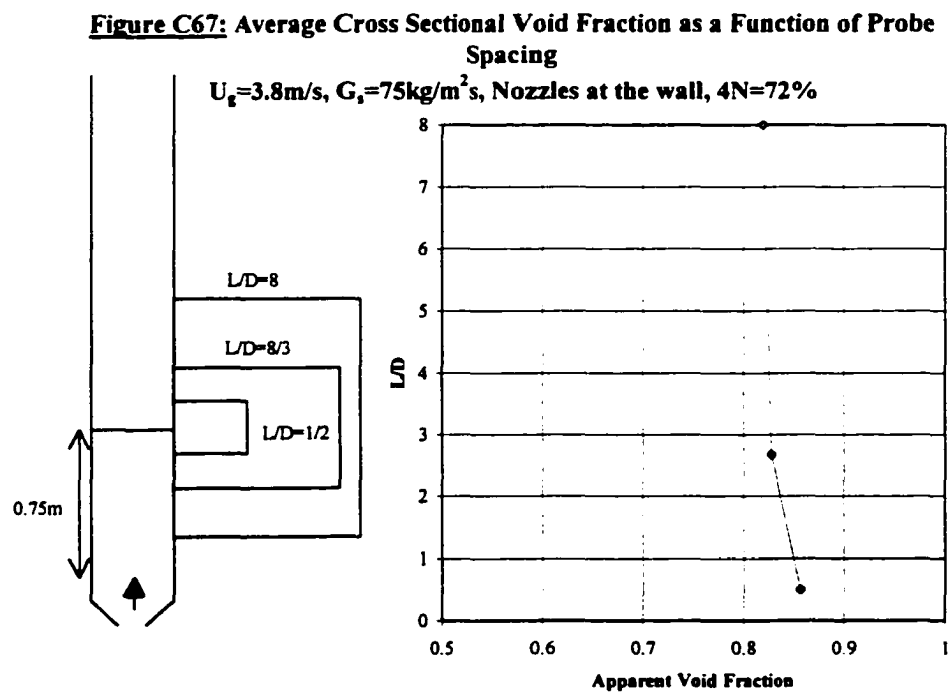
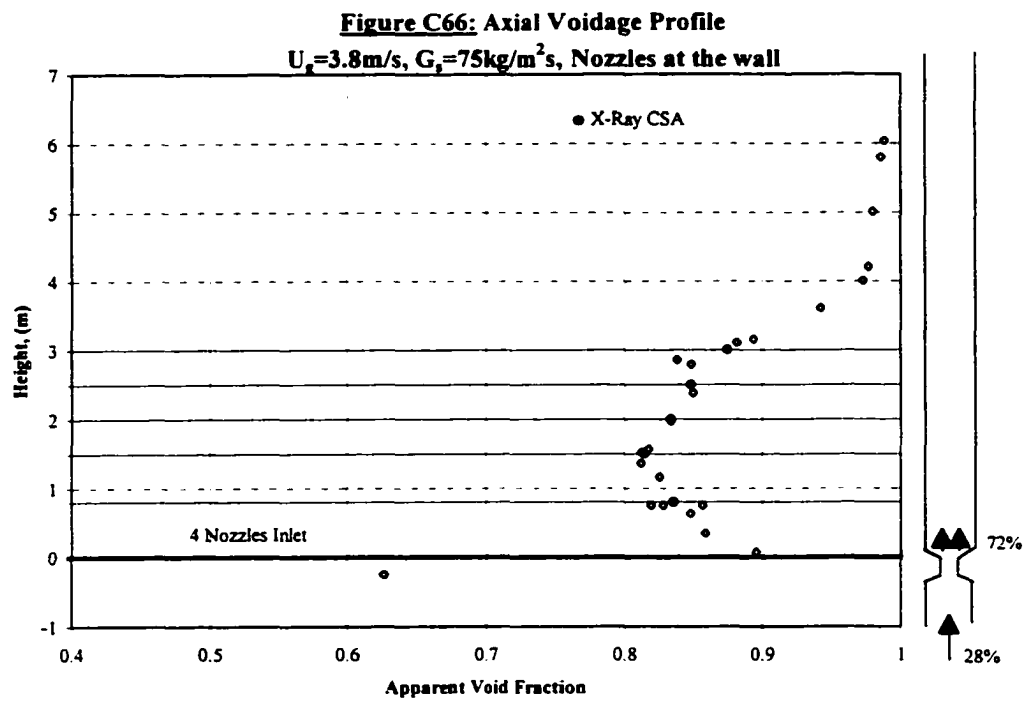


Figure C65: Average Cross Sectional Void Fraction as a Function of Probe Spacing





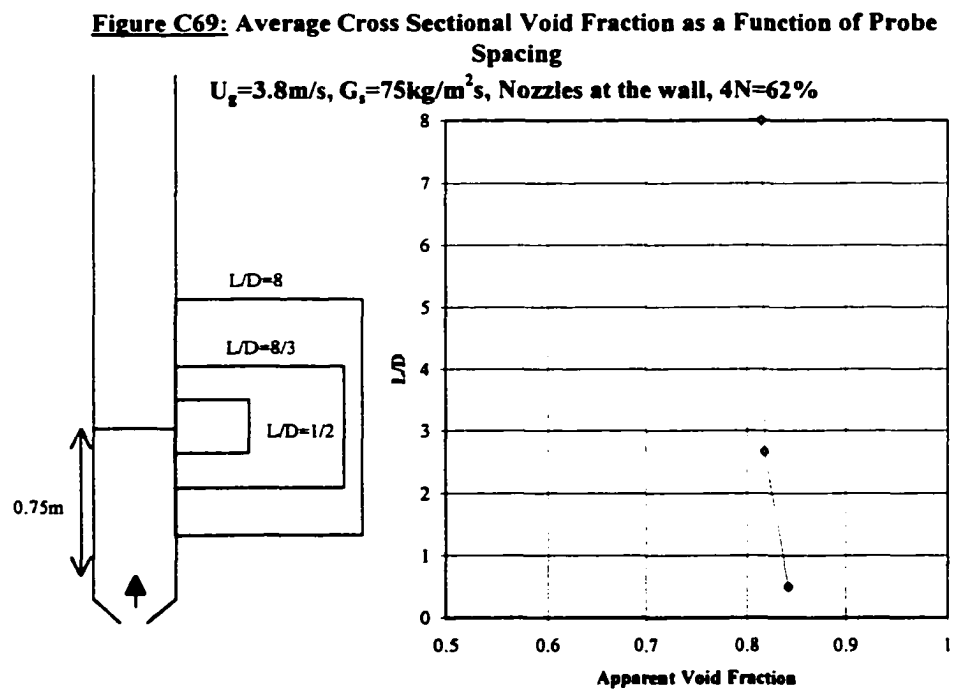
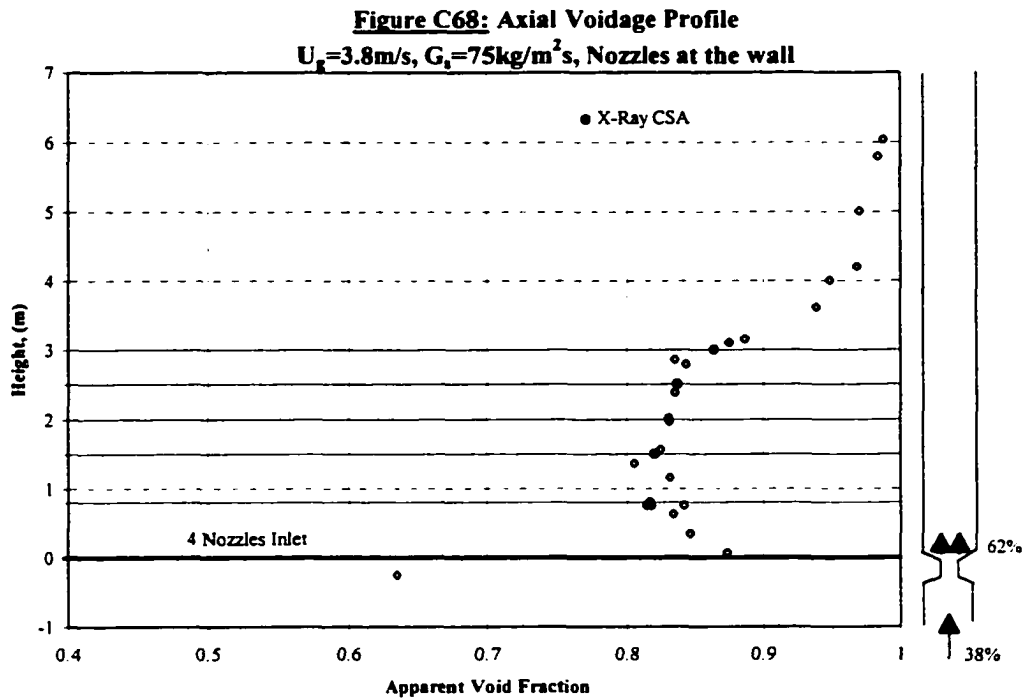


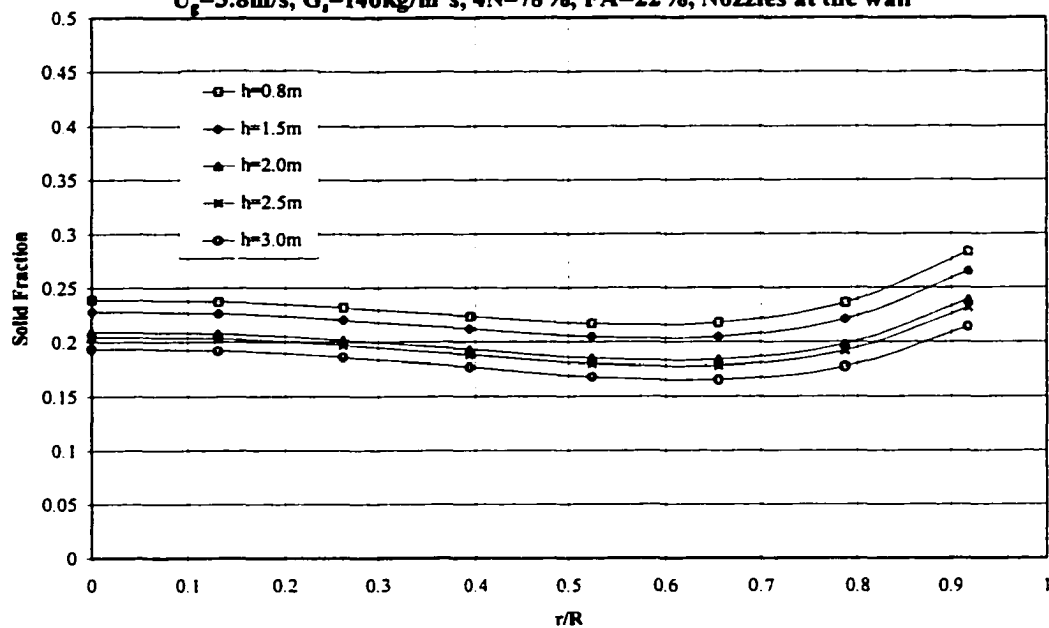
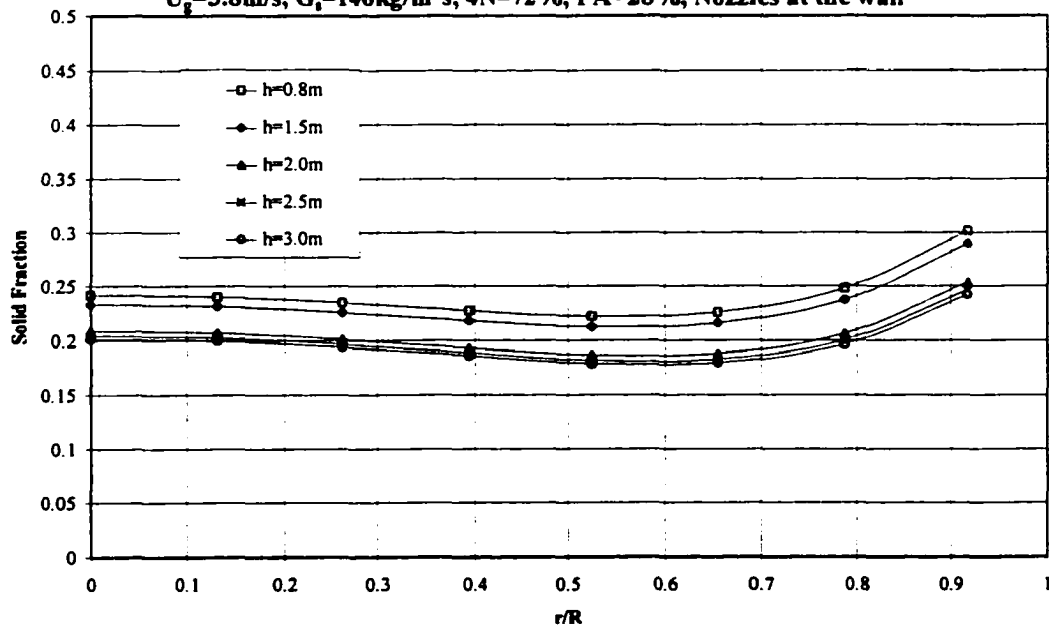
Figure C70: Radial Solid Profile $U_g=3.8\text{m/s}$, $G_t=140\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall**Figure C71: Radial Solid Profile** $U_g=3.8\text{m/s}$, $G_t=140\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall

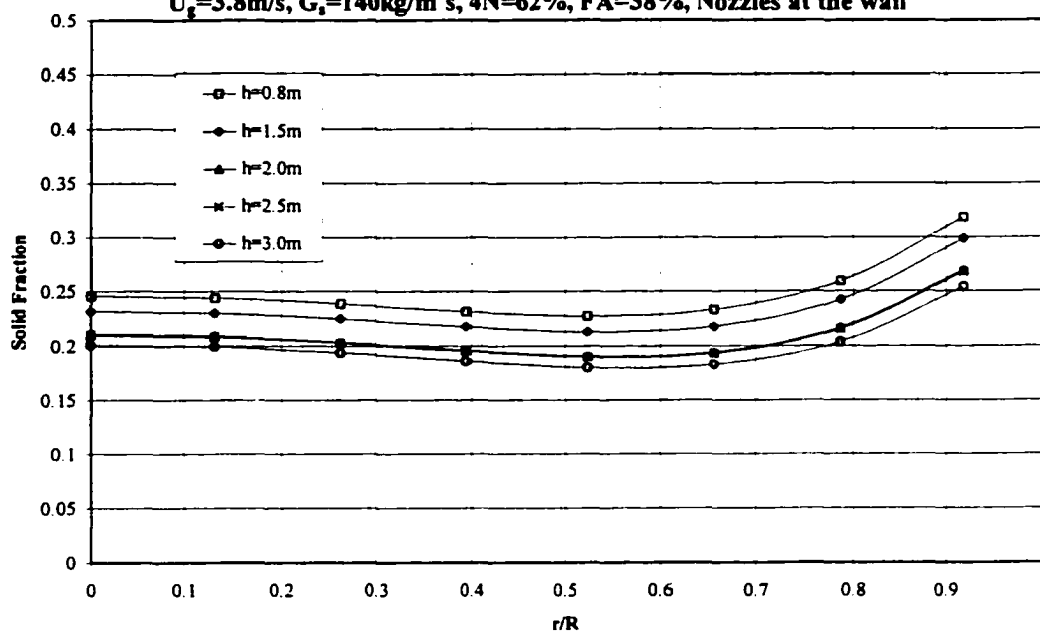
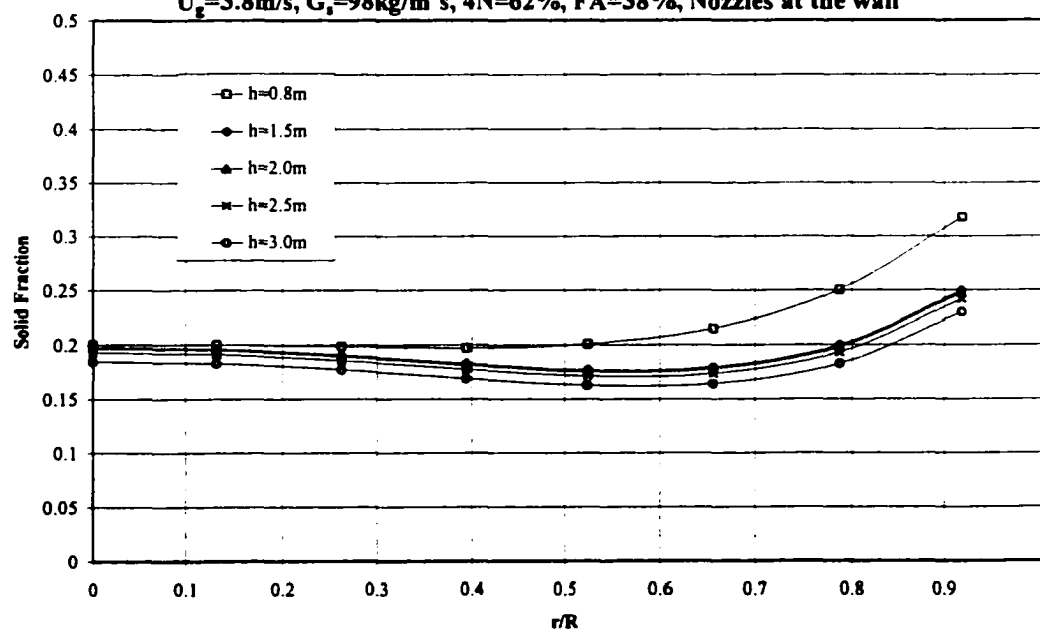
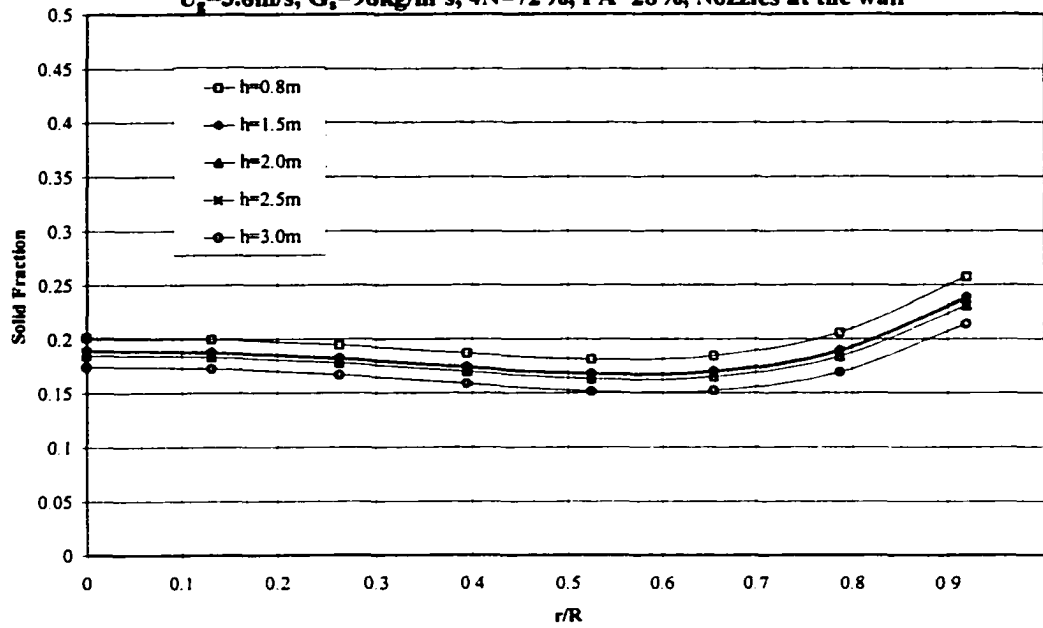
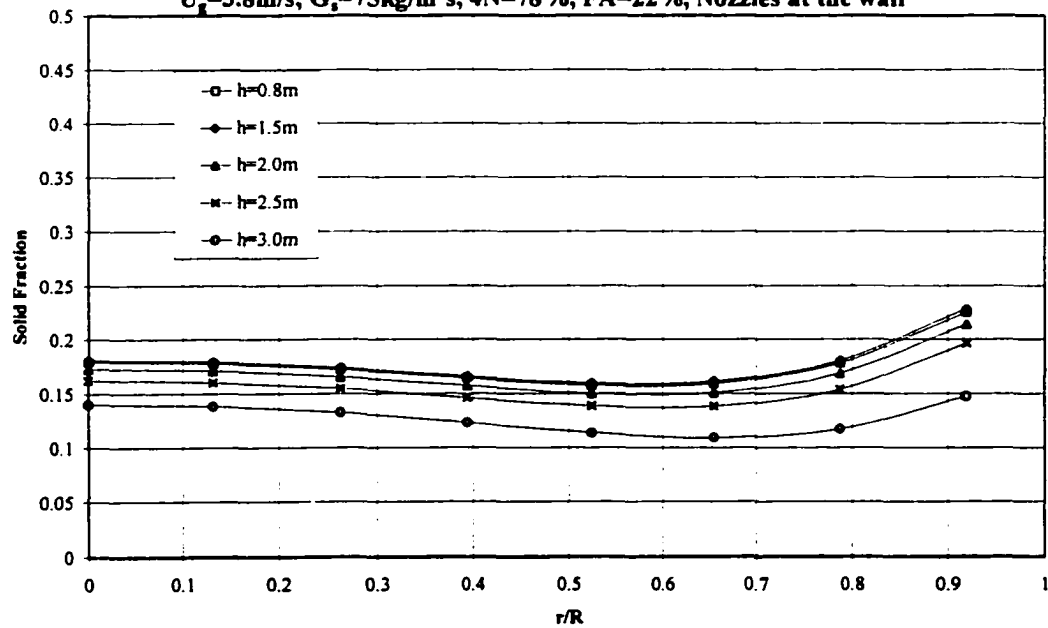
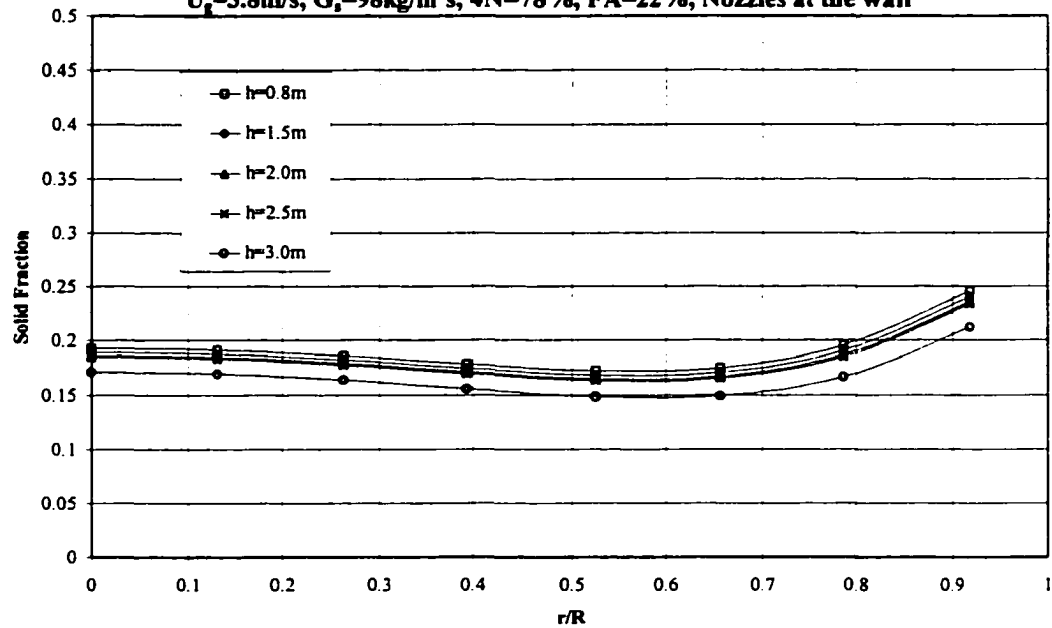
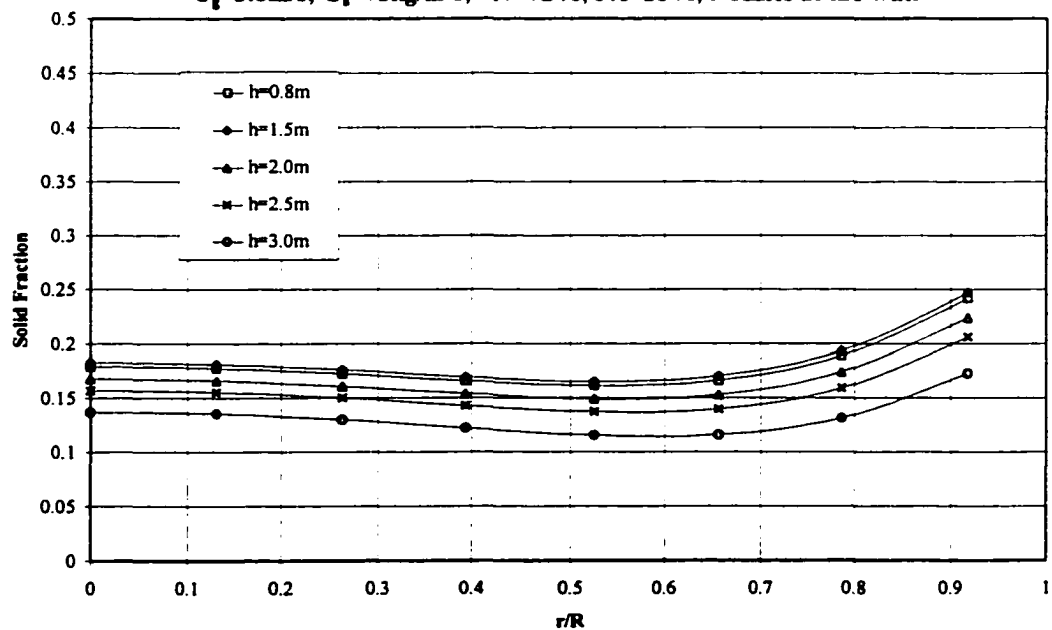
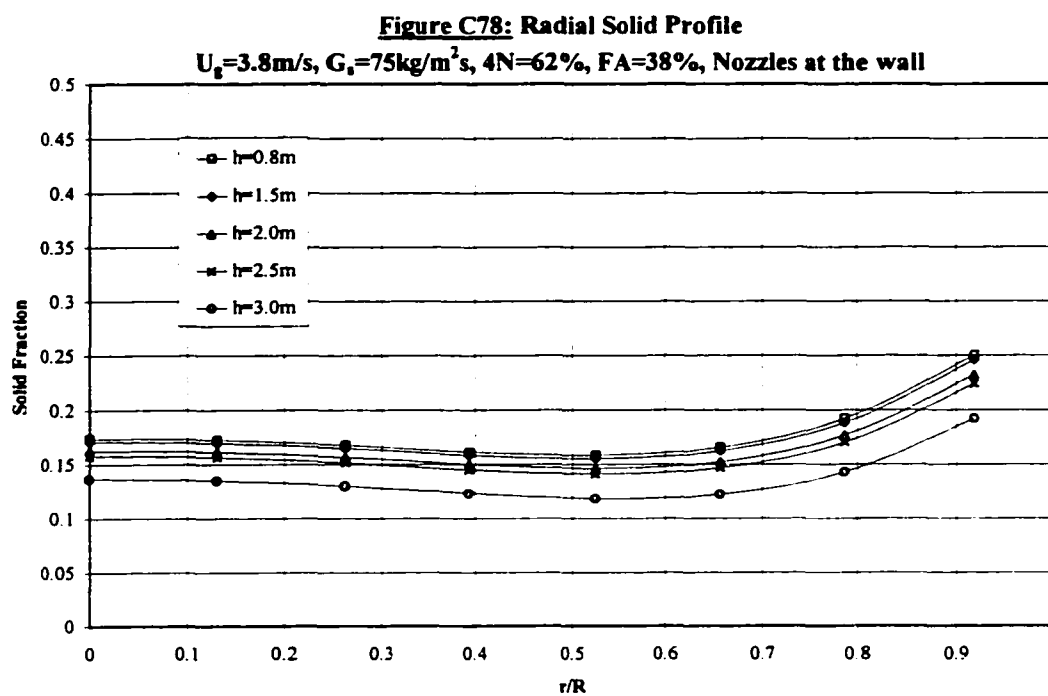
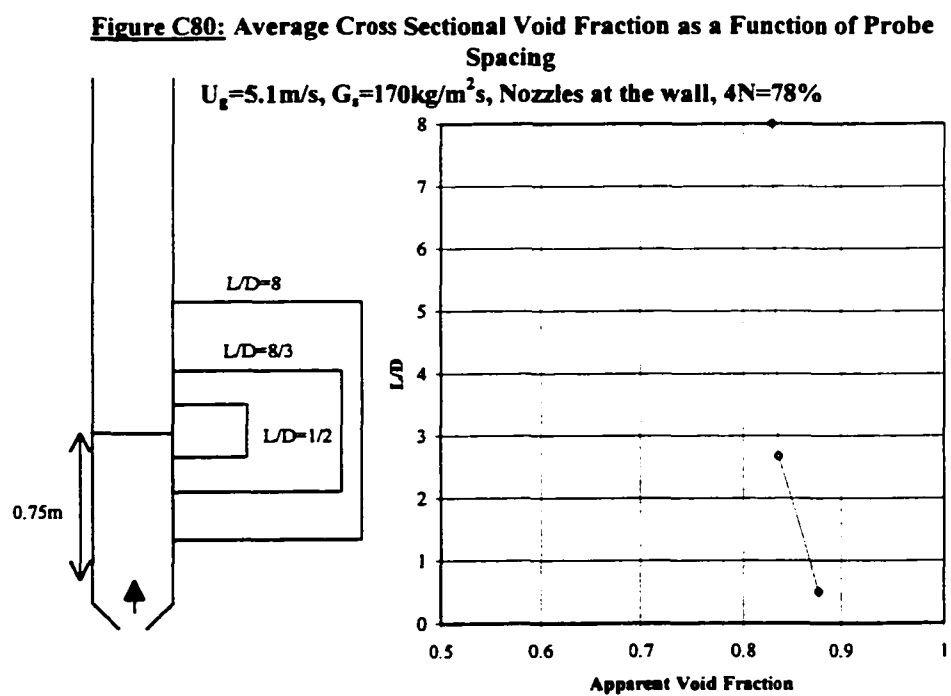
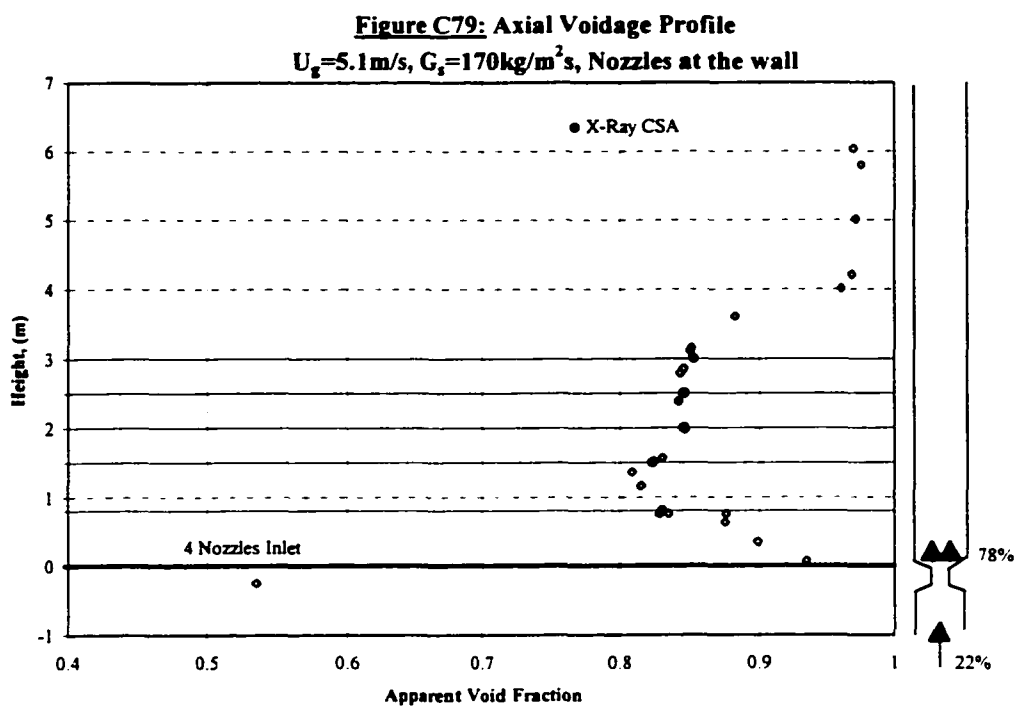
Figure C72: Radial Solid Profile $U_t=3.8\text{m/s}$, $G_s=140\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the wall**Figure C73: Radial Solid Profile** $U_t=3.8\text{m/s}$, $G_s=98\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the wall

Figure C74: Radial Solids Profile $U_s=3.8\text{m/s}$, $G_s=98\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall**Figure C75: Radial Solid Density Profile** $U_s=3.8\text{m/s}$, $G_s=75\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall

Figuer C76: Radial Solid Density Profile $U_g=3.8\text{m/s}$, $G_s=98\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall**Figure C77: Radial Solid Profile** $U_g=3.8\text{m/s}$, $G_s=75\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall





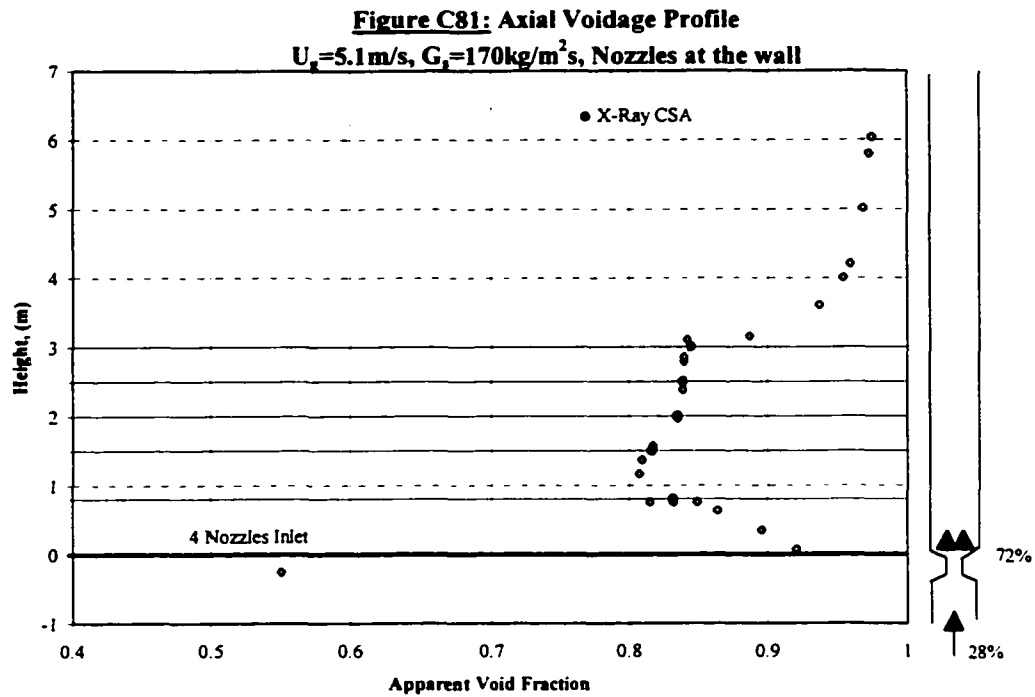
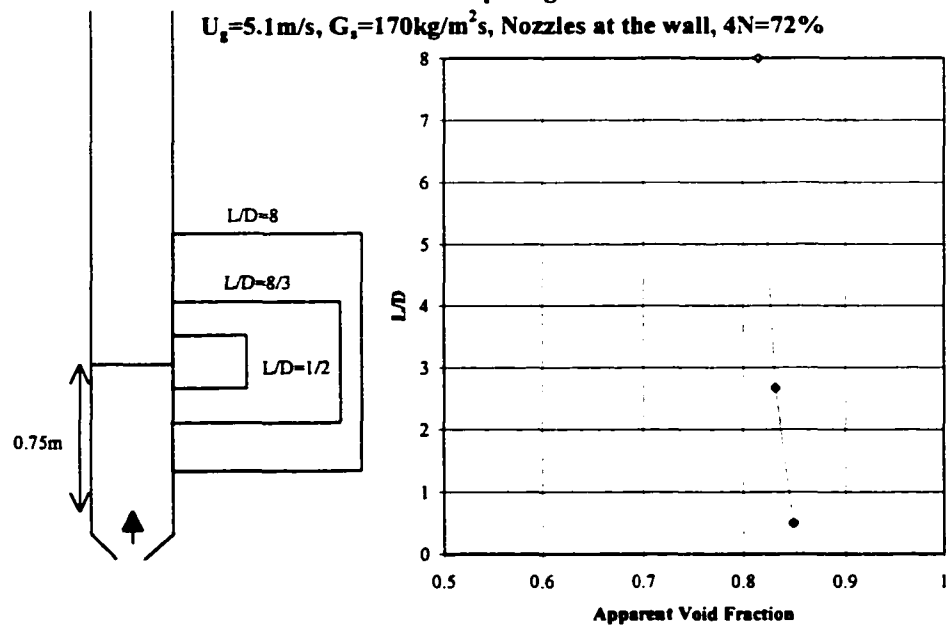
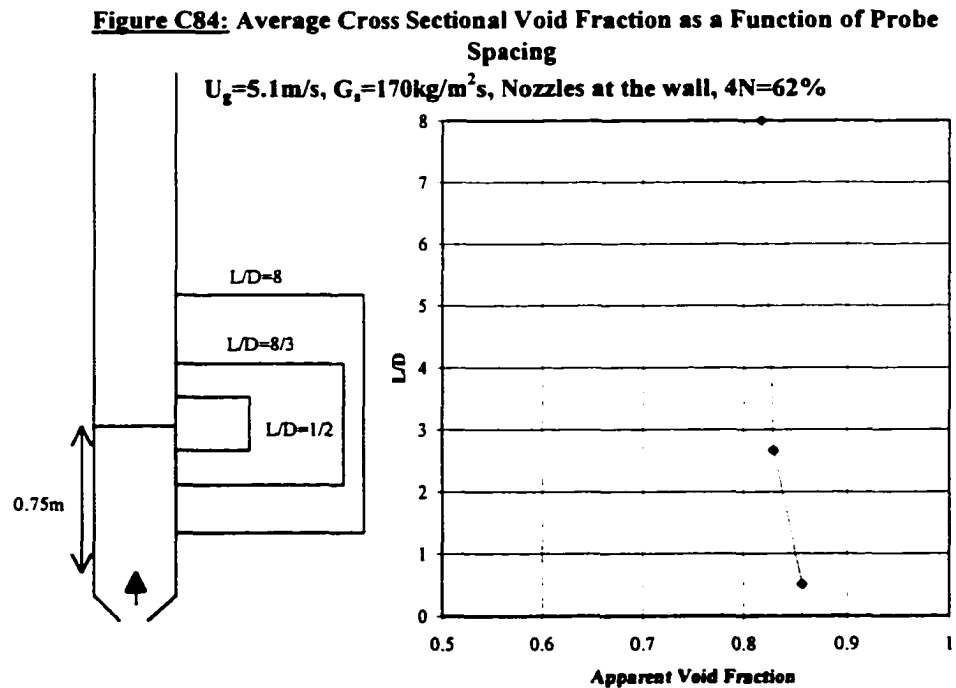
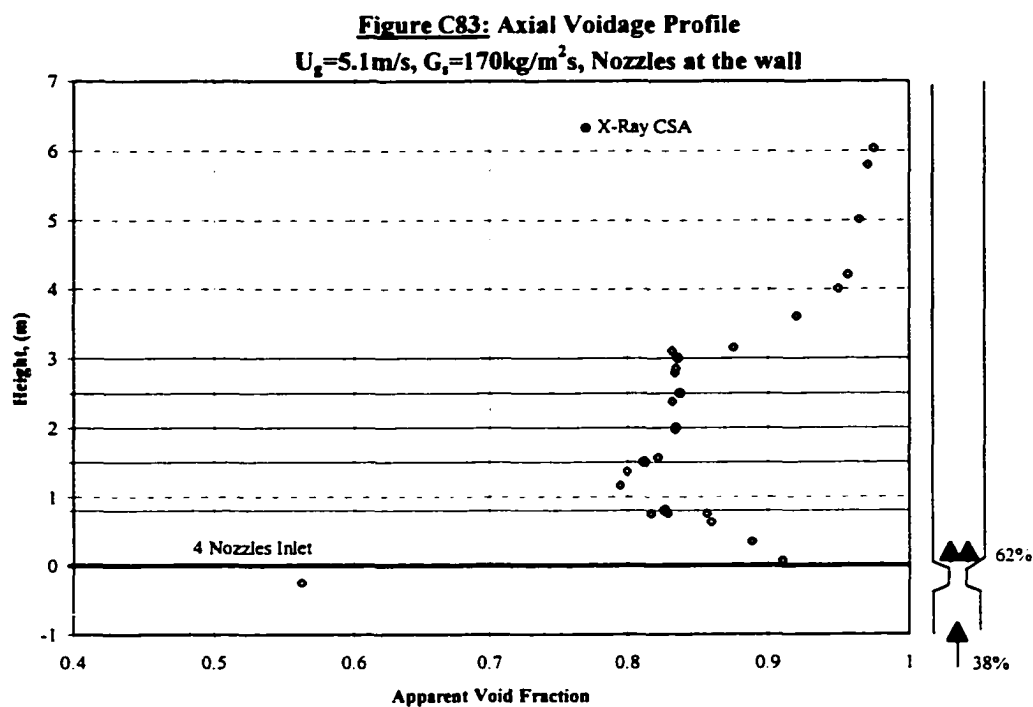
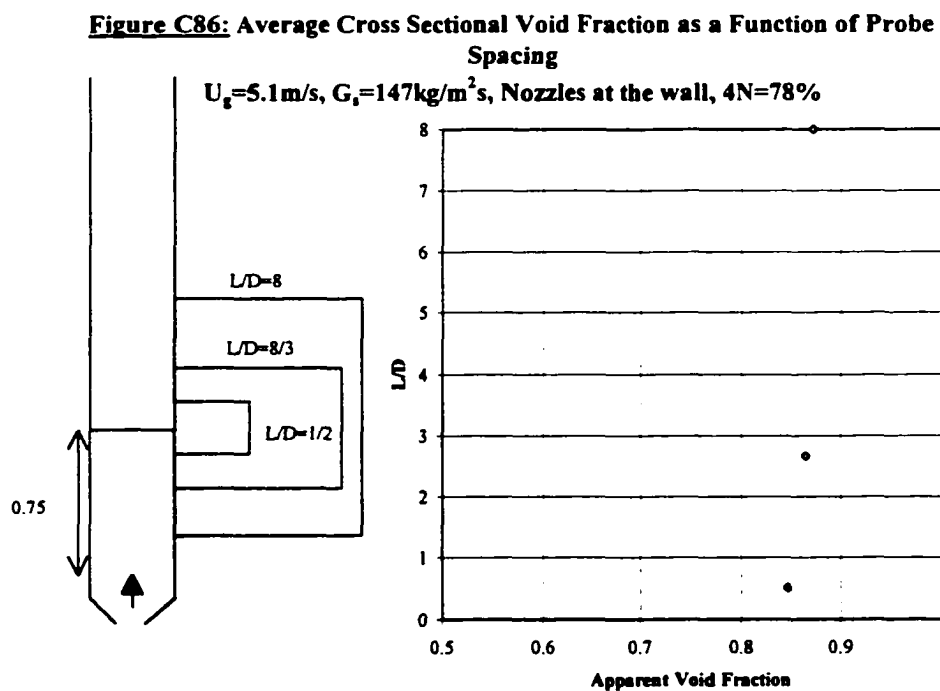
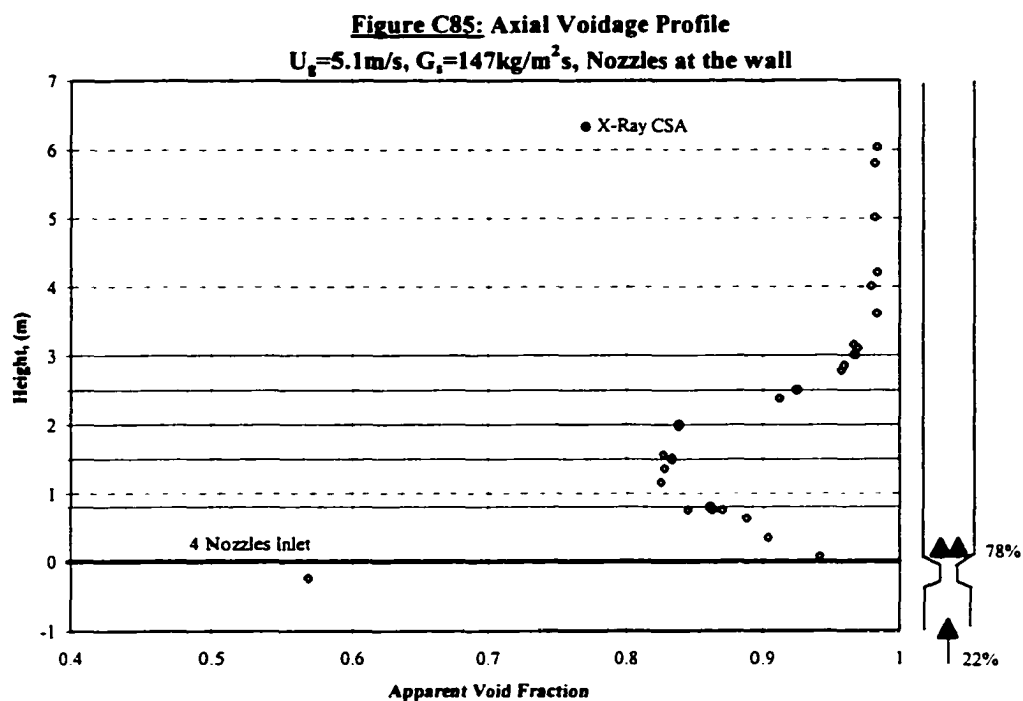
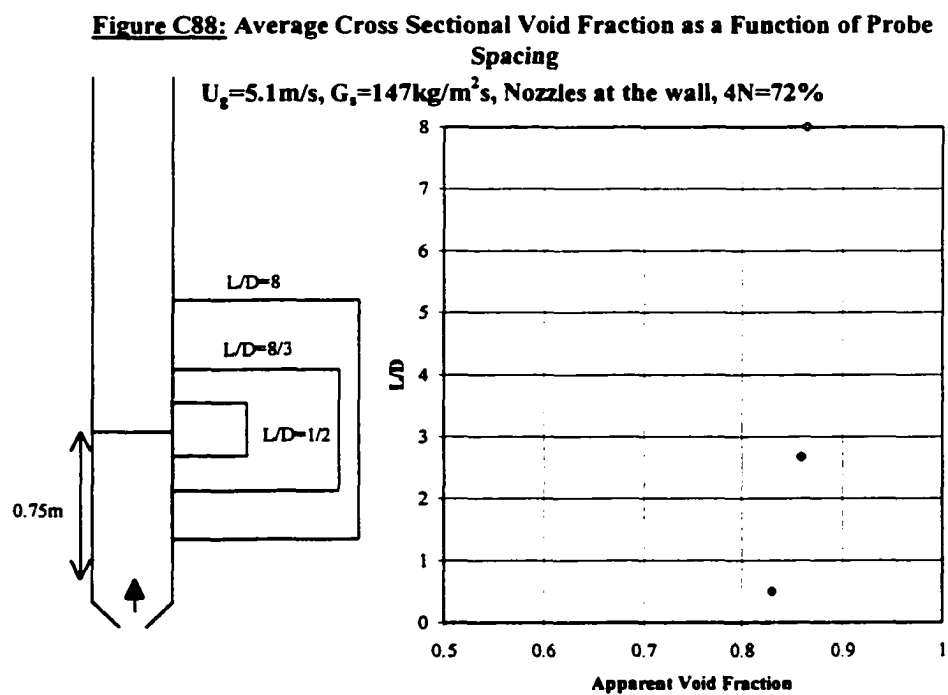
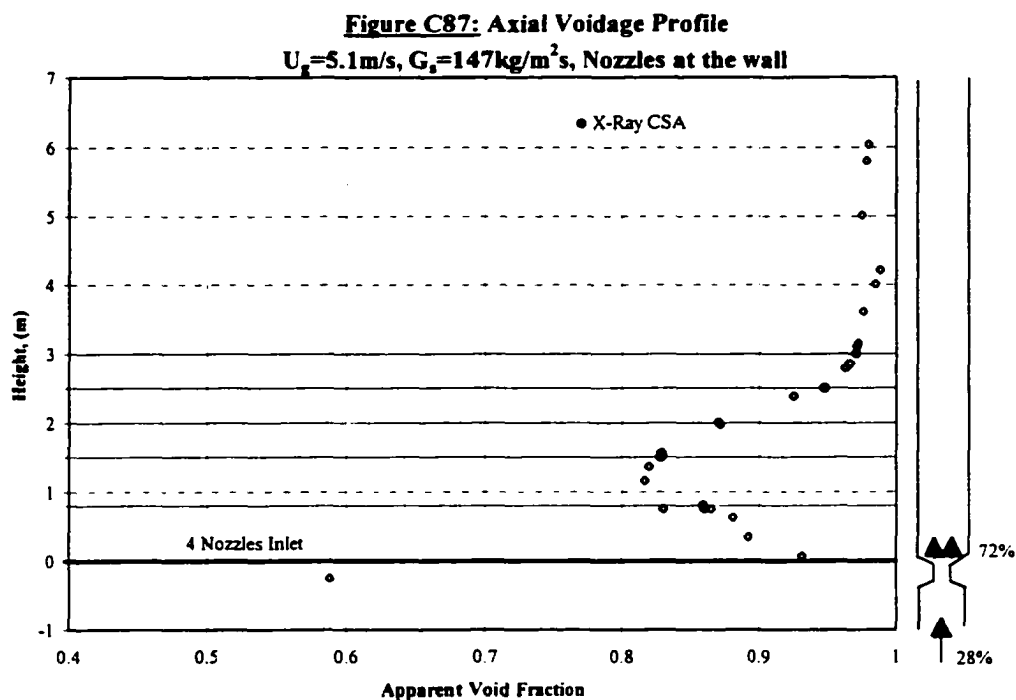


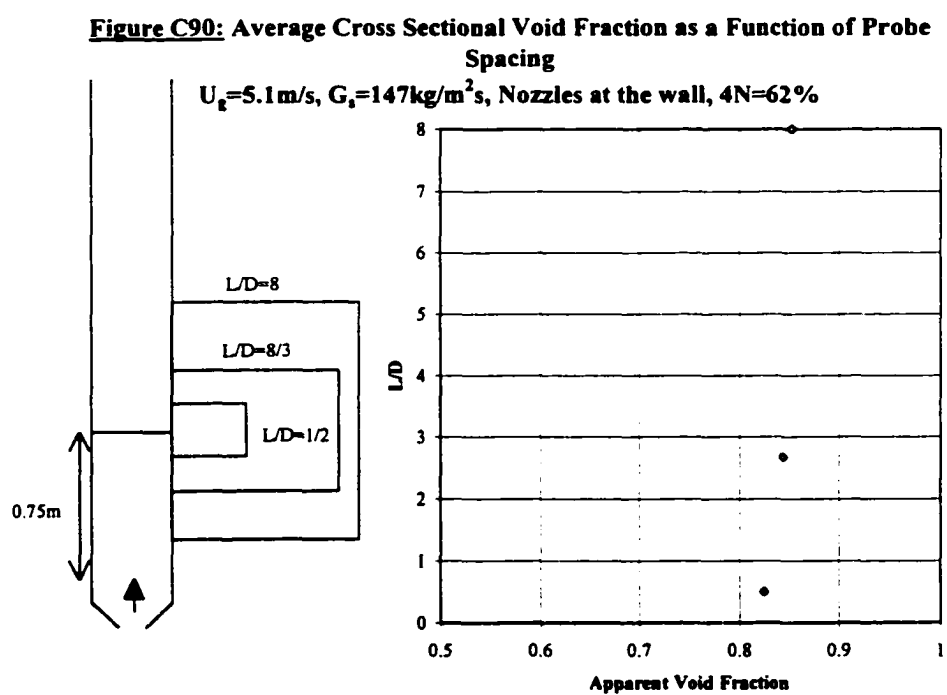
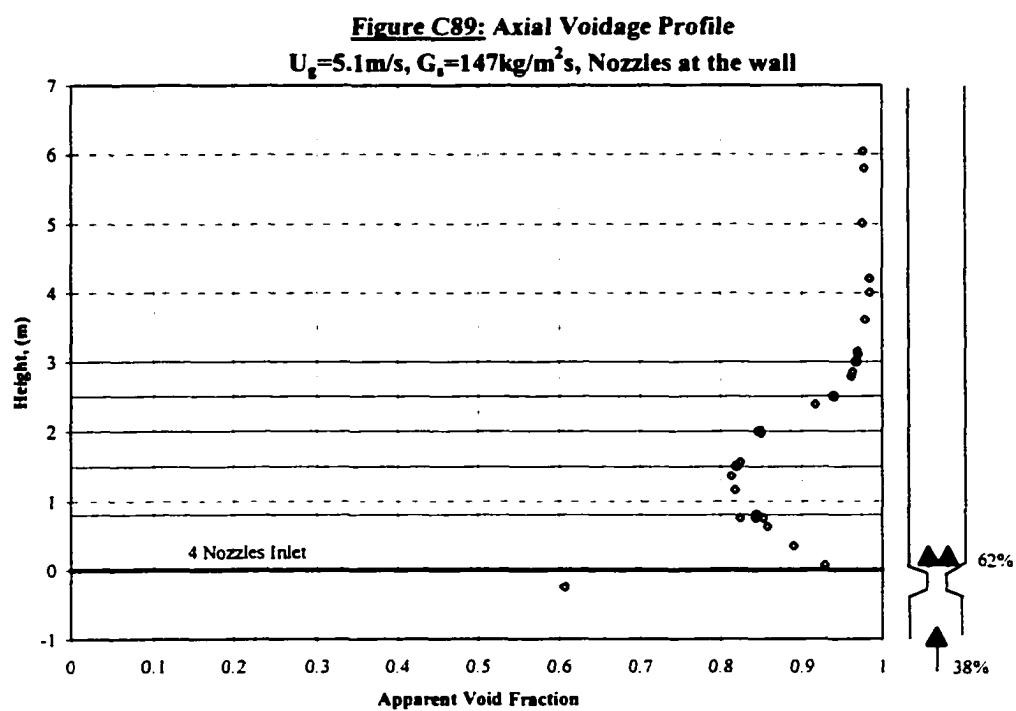
Figure C82: Average Cross Sectional Void Fraction as a Function of Probe Spacing

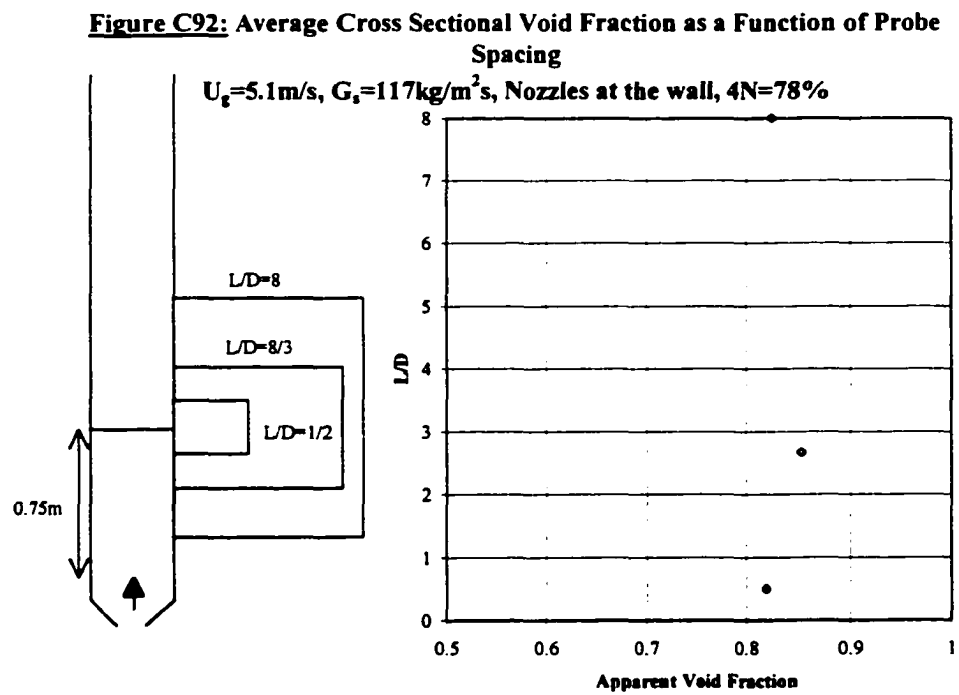
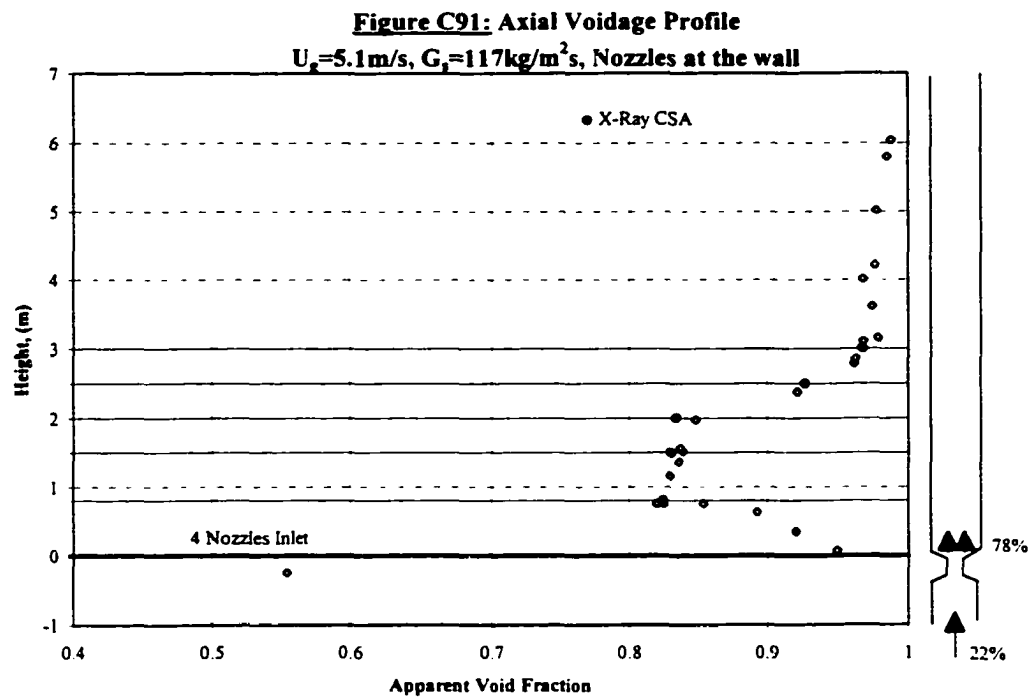












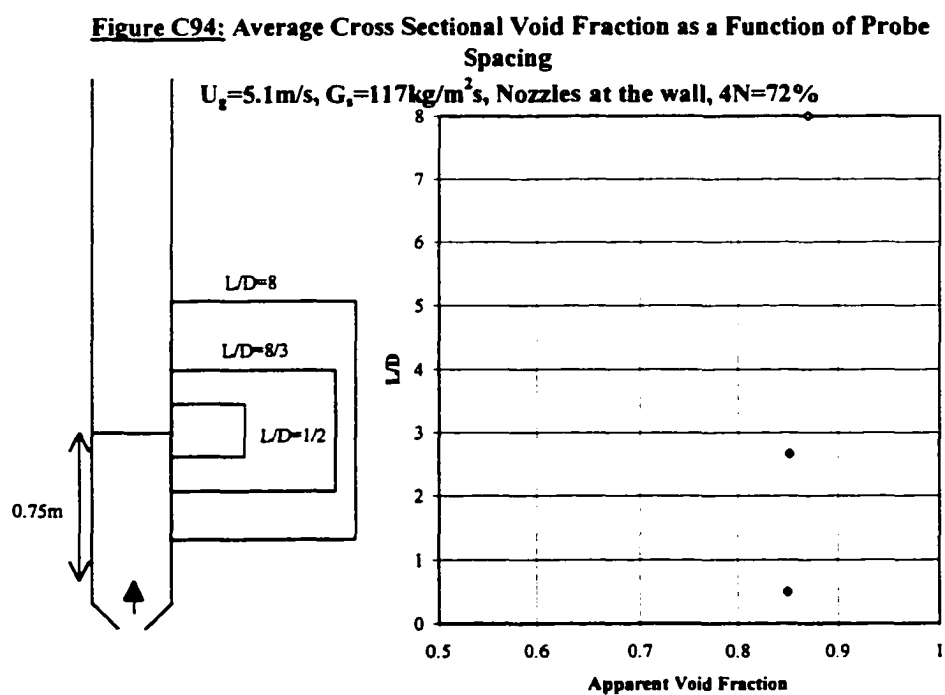
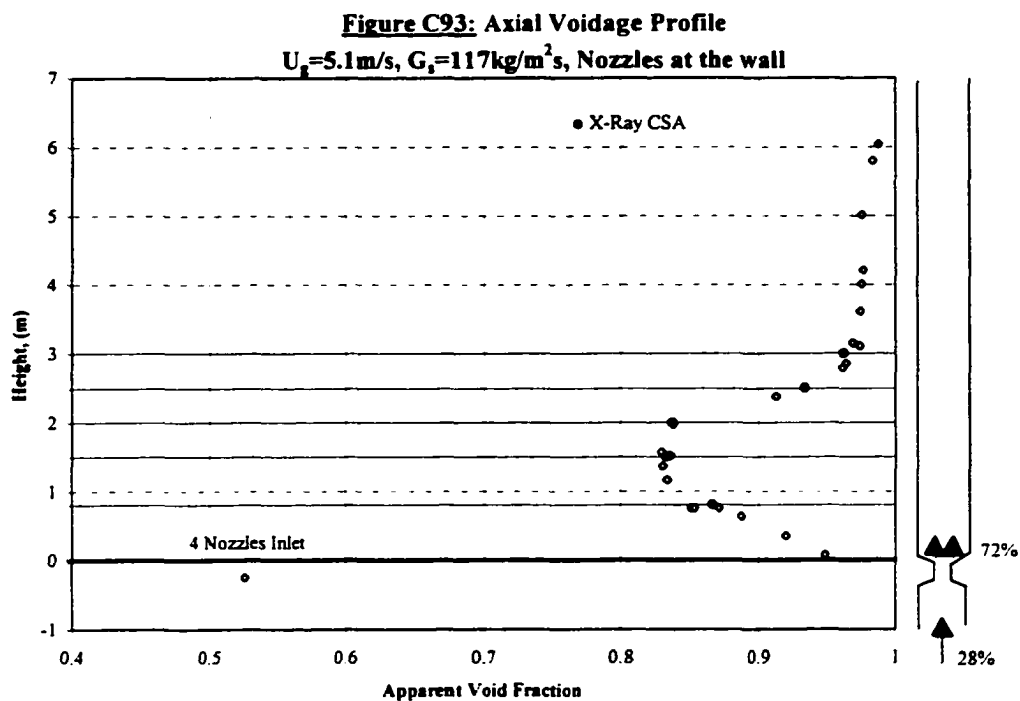


Figure C95: Axial Voidage Profile
 $U_g=5.1\text{m/s}$, $G_t=117\text{kg/m}^2\text{s}$, Nozzles at the wall

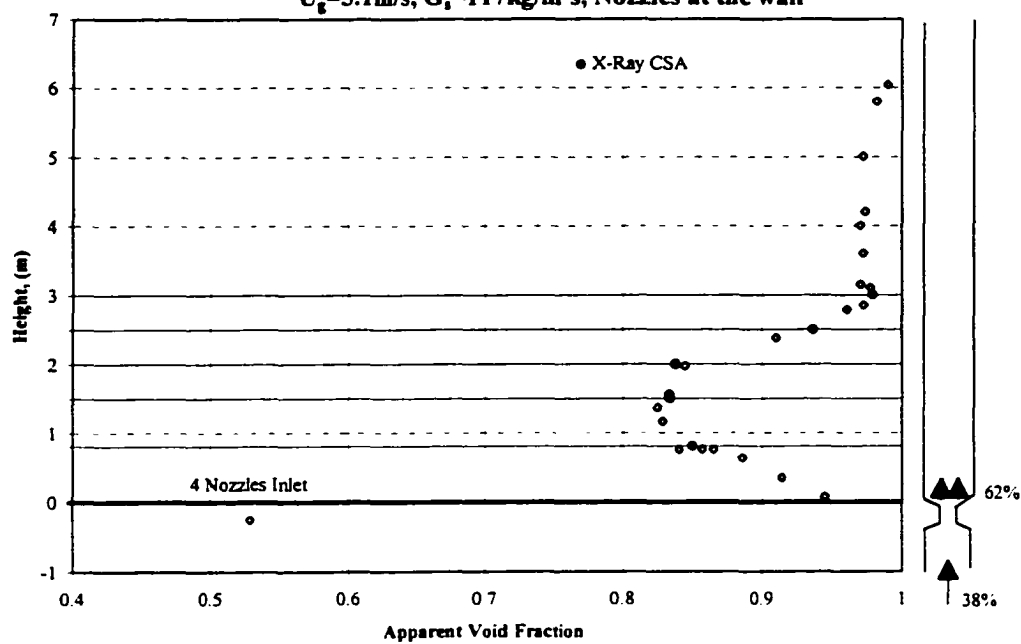
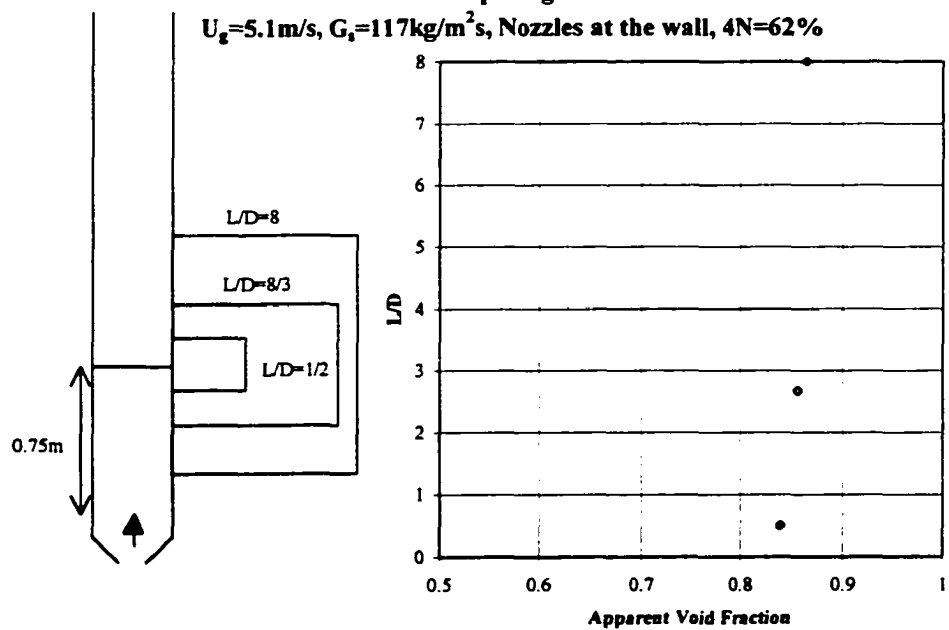


Figure C96: Average Cross Sectional Void Fraction as a Function of Probe Spacing



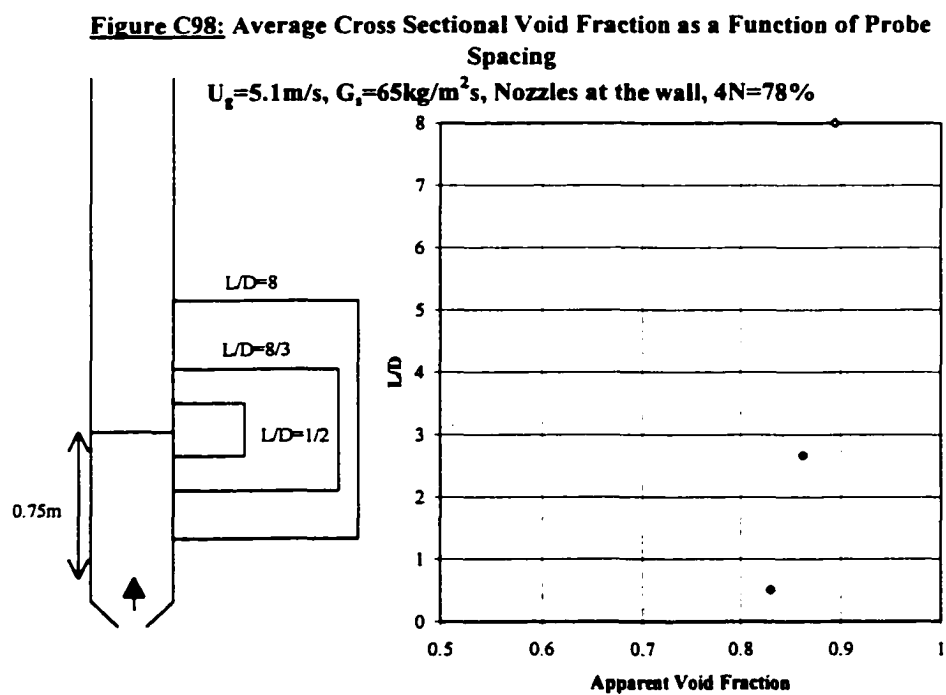
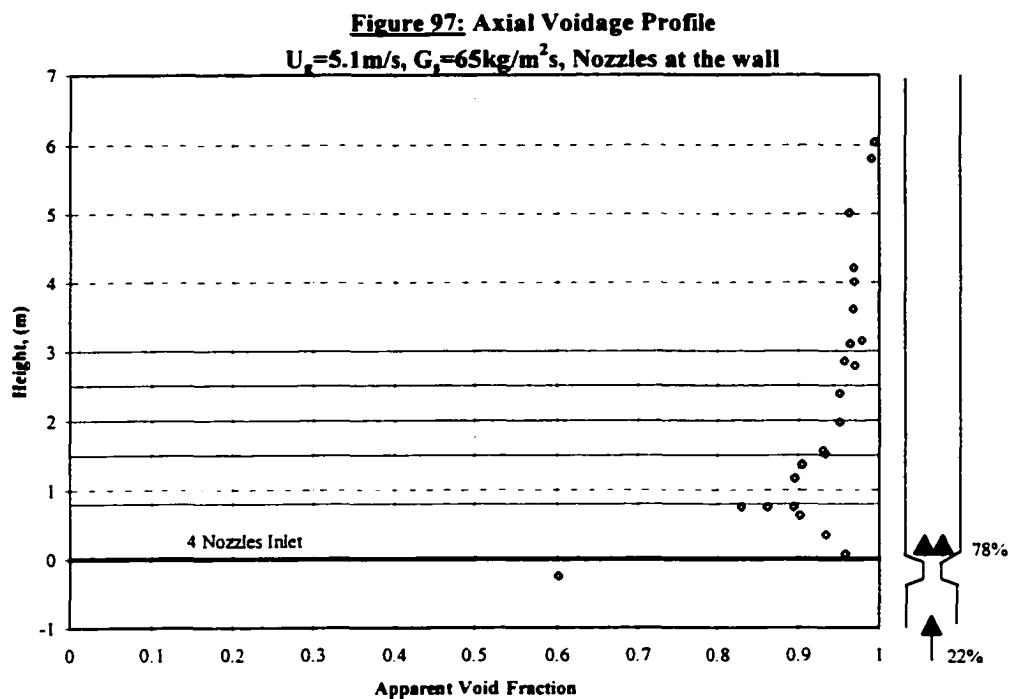
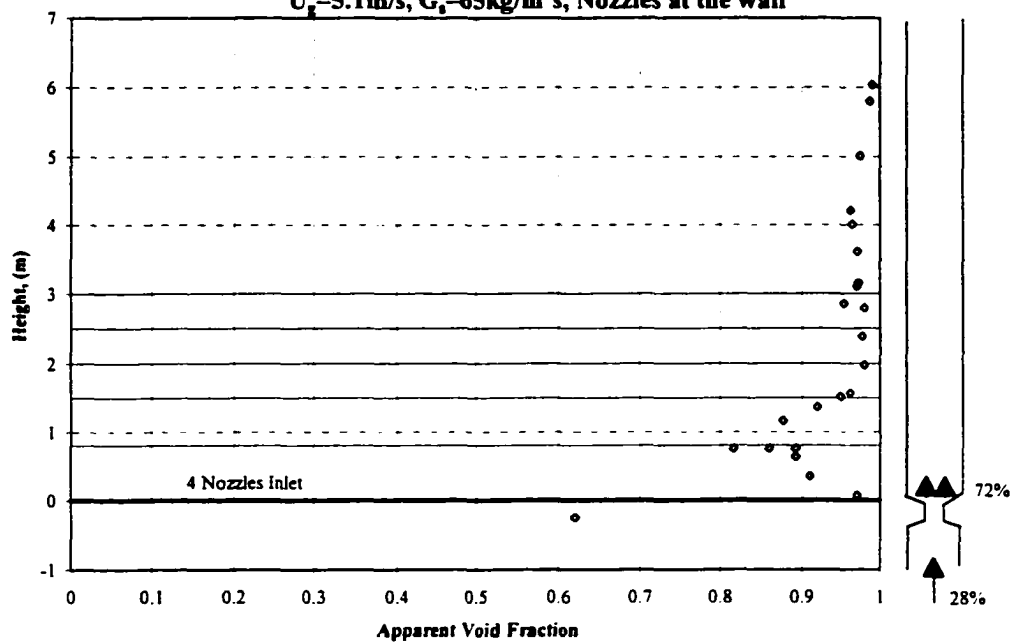
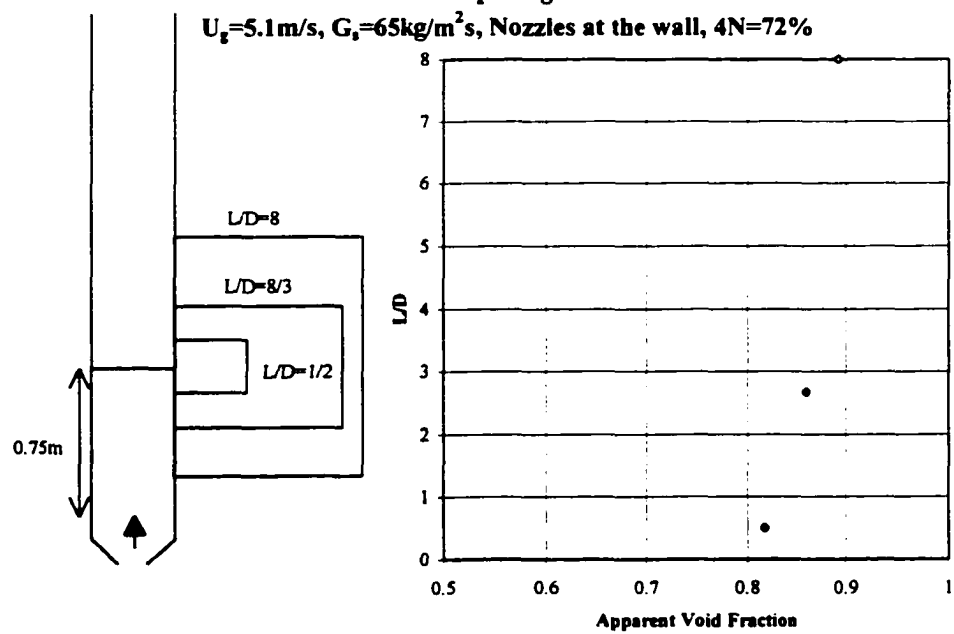


Figure C99: Axial Voidage Profile $U_g = 5.1 \text{ m/s}$, $G_g = 65 \text{ kg/m}^2\text{s}$, Nozzles at the wall**Figure C100: Average Cross Sectional Void Fraction as a Function of Probe Spacing** $U_g = 5.1 \text{ m/s}$, $G_g = 65 \text{ kg/m}^2\text{s}$, Nozzles at the wall, $4N=72\%$ 

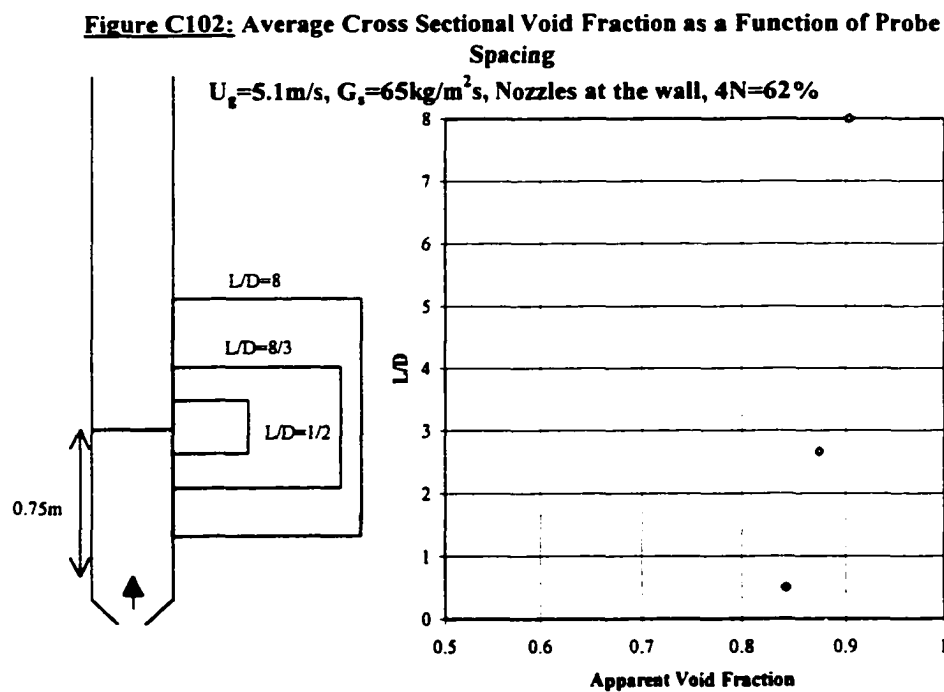
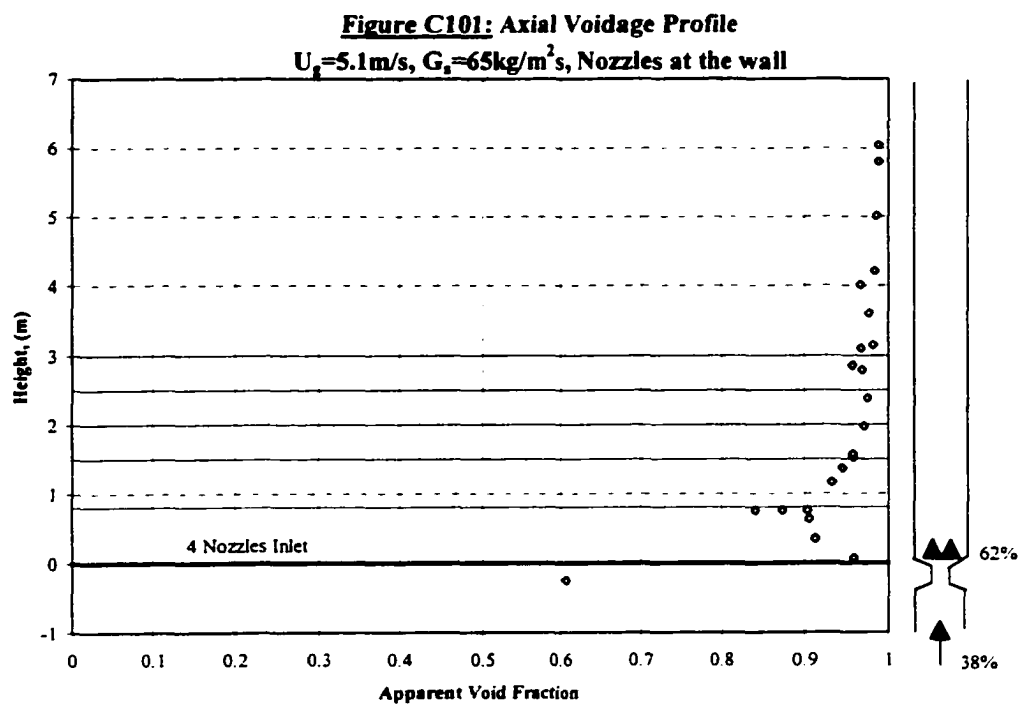


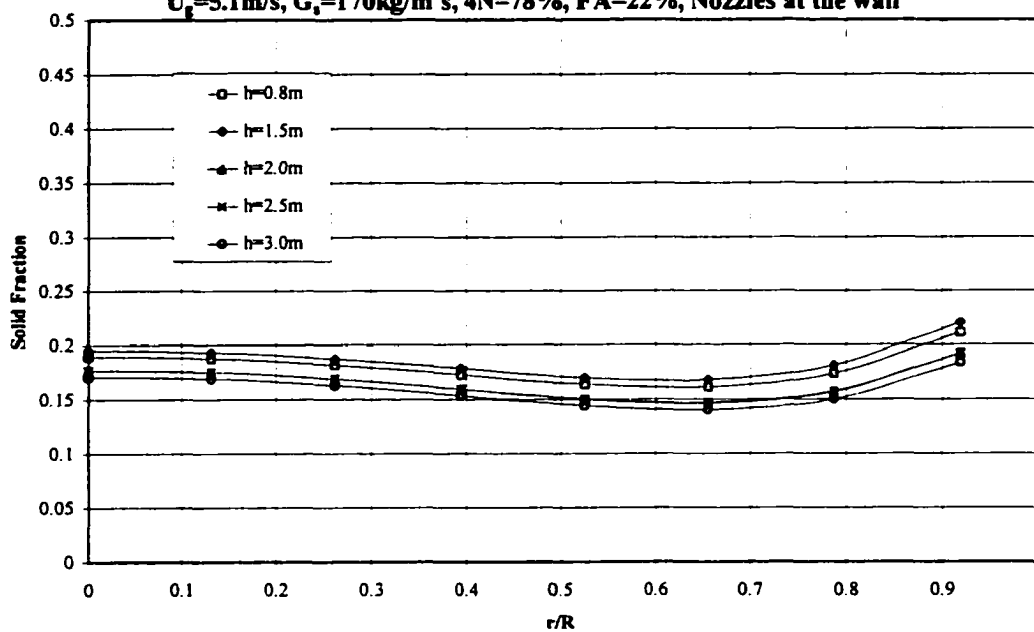
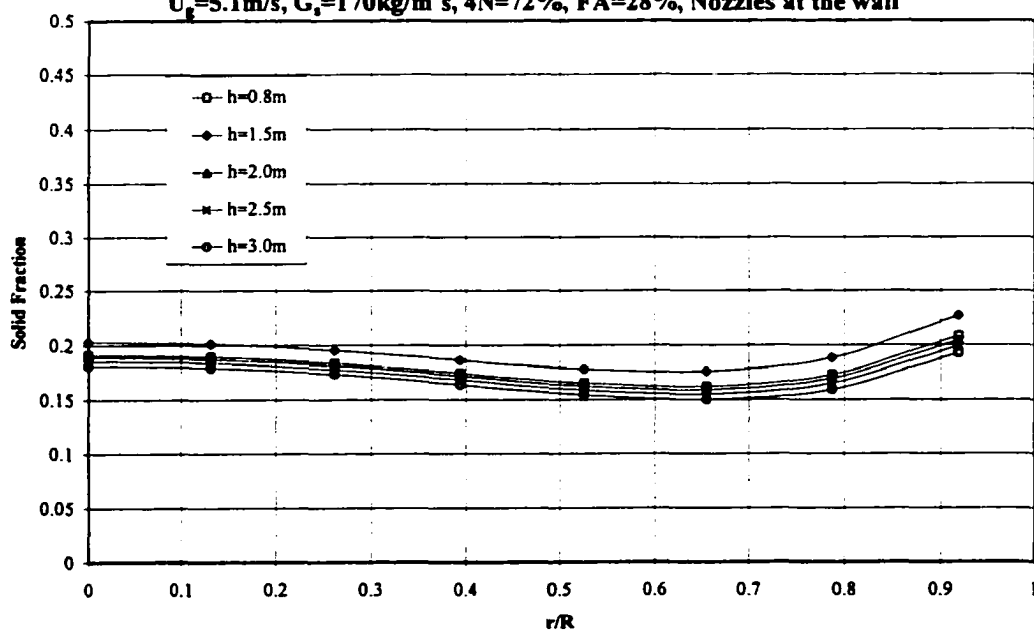
Figure C103: Radial Solid Profile **$U_f=5.1\text{m/s}$, $G_s=170\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall****Figure C104: Radial Solid Profile** **$U_f=5.1\text{m/s}$, $G_s=170\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall**

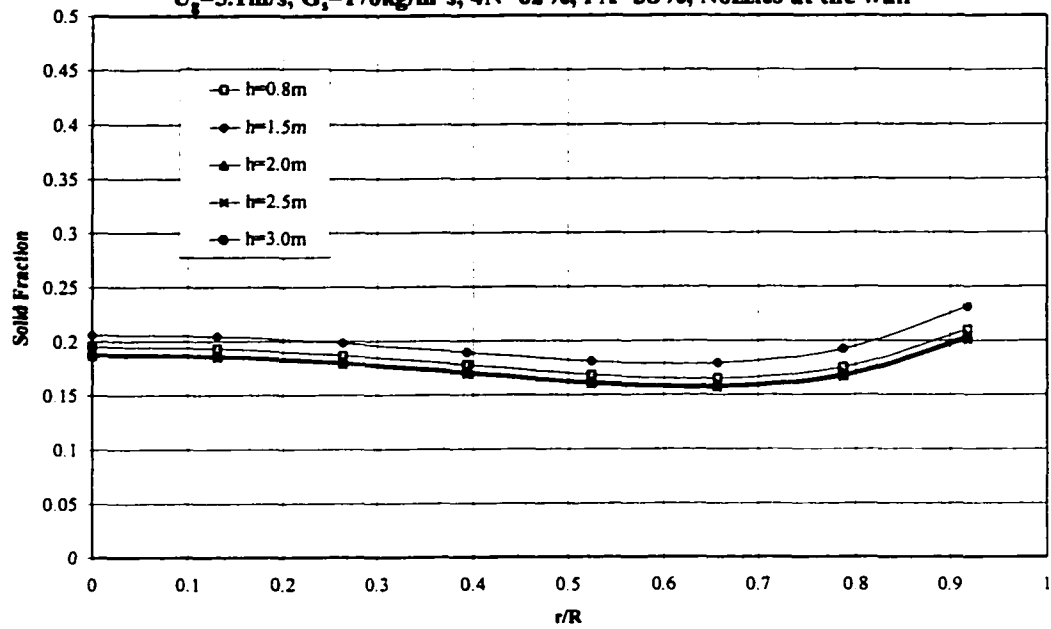
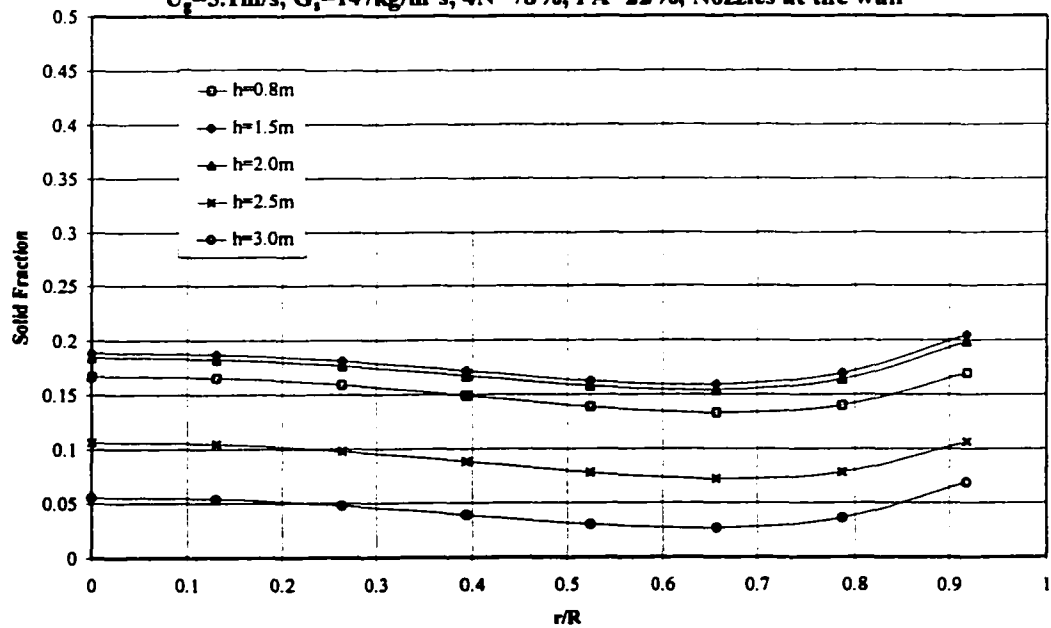
Figure C105: Radial Solid Profile $U_f=5.1\text{m/s}$, $G_s=170\text{kg/m}^2\text{s}$, $4N=62\%$, $FA=38\%$, Nozzles at the wall**Figure C106: Radial Solid Profile** $U_f=5.1\text{m/s}$, $G_s=147\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall

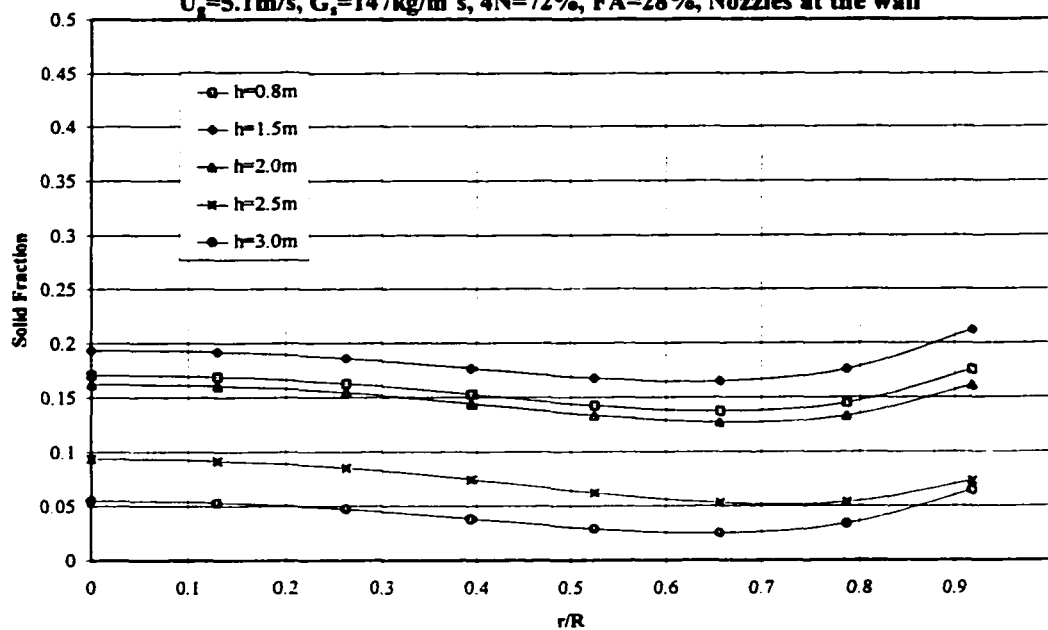
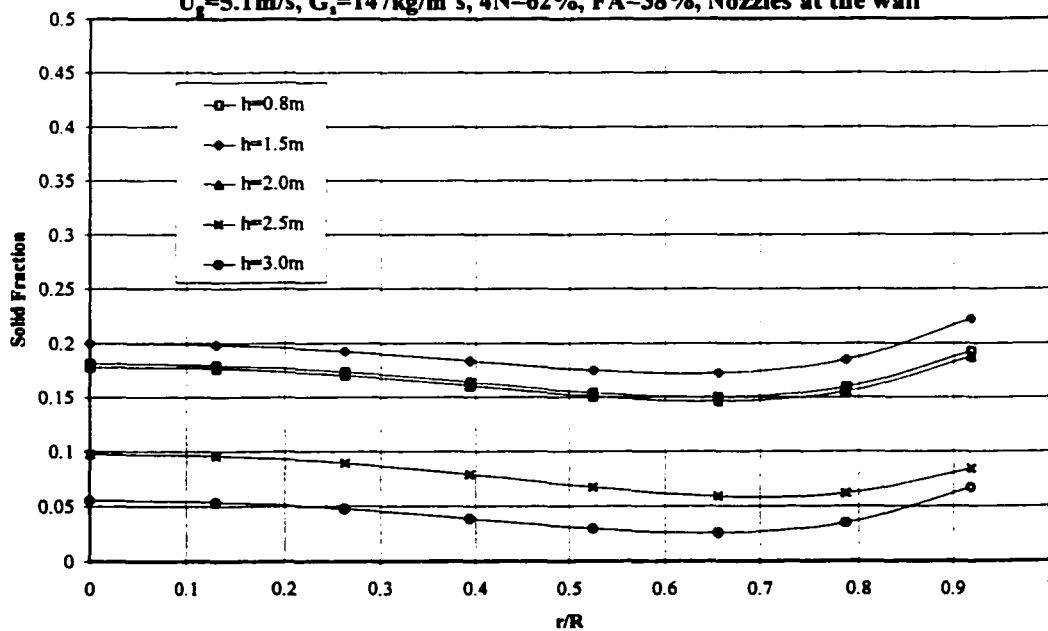
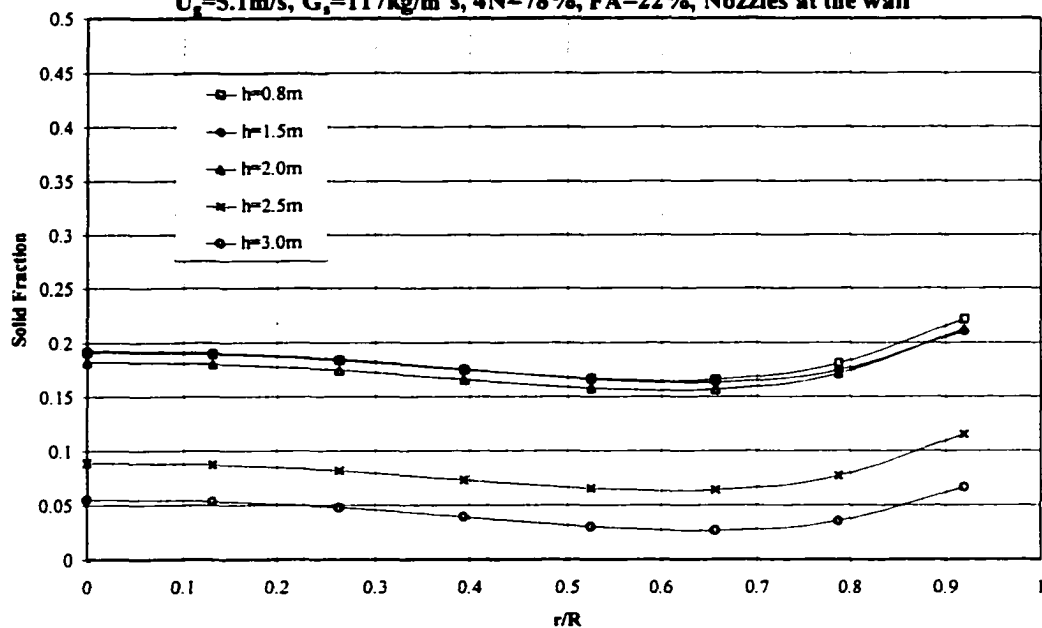
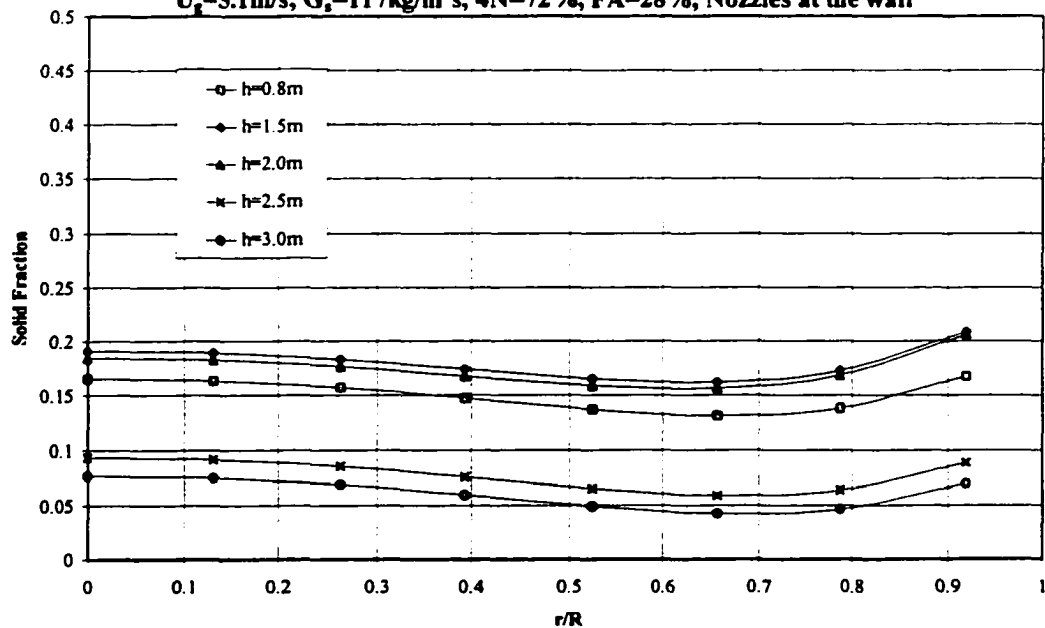
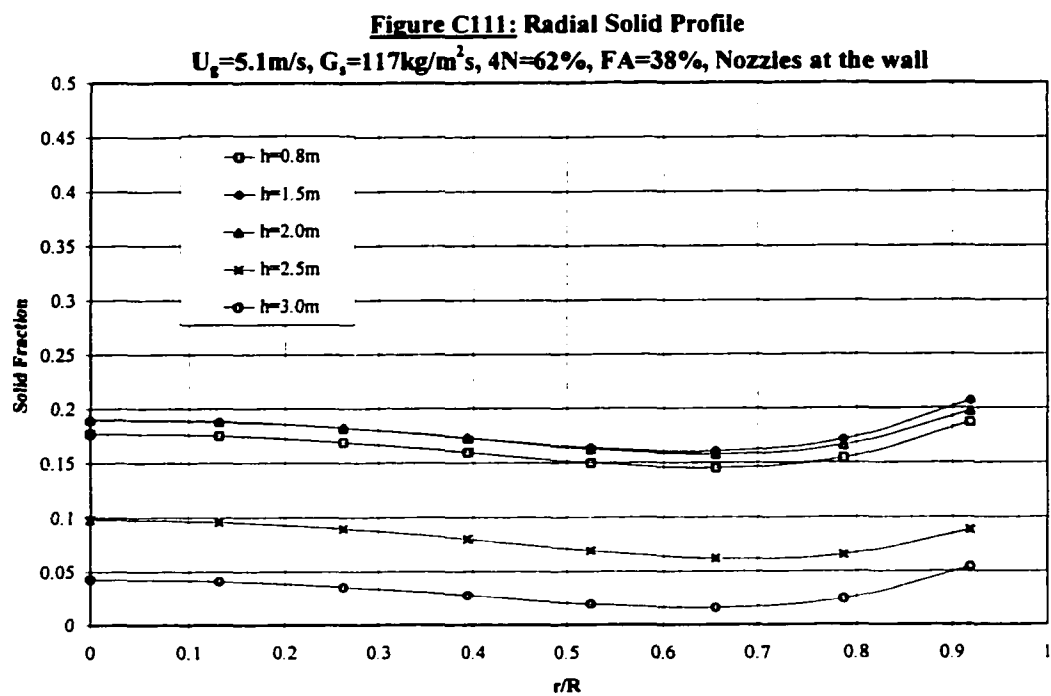
Figure C107: Radial Solid Profile $U_s=5.1\text{ m/s}$, $G_s=147\text{ kg/m}^2\text{ s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall**Figure C108: Radial Solid Density Profile** $U_s=5.1\text{ m/s}$, $G_s=147\text{ kg/m}^2\text{ s}$, $4N=62\%$, $FA=38\%$, Nozzles at the wall

Figure C109: Radial Solid Density Profile $U_s=5.1\text{m/s}$, $G_s=117\text{kg/m}^2\text{s}$, $4N=78\%$, $FA=22\%$, Nozzles at the wall**Figure C110: Radial Solid Profile** $U_s=5.1\text{m/s}$, $G_s=117\text{kg/m}^2\text{s}$, $4N=72\%$, $FA=28\%$, Nozzles at the wall



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