



The determination of bubble velocity in fluidized beds
by Jeffrey Edward Surma

A thesis submitted in partial fulfillment of the requirements for the degree of Master of Science in
Chemical Engineering
Montana State University
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Abstract:

Bubble velocity in a fluidized bed was determined with the use of an assembly of optical probes. Two probe assemblies were developed, one employed the use of three optical probes, the second used four. The four probe assembly proved most adequate in use for local bubble velocity determination. The volumes of bubbles detected by the four-probe assembly were also determinable from the probe outputs.

The four probe assembly was used to determine bubble velocities in a fluidized bed of sand. The sand was irregular in shape and had an average particle diameter of $515 \pm 68 \mu\text{m}$. Experimental results were compared to the Davidson and Harrison equation and a modification of this equation. The modified equation, $U_b = C^*0(U_0 - U_{mf}) + U_{br}$, incorporated a distribution coefficient C^*0 which compared local bubble flow to the cross-sectional average.

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APPROVAL

of a thesis submitted by

Jeffrey Edward Surma

This thesis has been read by each member of the thesis committee and has been found to be satisfactory regarding content, English usage, format, citation, bibliographic style, and consistency, and is ready for submission to the College of Graduate Studies.

Dec. 2, 1985
Date

John T. Sears
Chairperson, Graduate Committee

Approved for the Major Department

Dec. 2, 1985
Date

John T. Sears
Head, Major Department

Approved for the College of Graduate Studies

12-10-85
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MO Malve
Graduate Dean

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NOMENCLATURE

A_T	Cross-sectional area of the bed
C_o	Distribution coefficient defined by Equation 5
C_o^*	Distribution coefficient defined in Table 3
D	Diameter of the bed
d_b	Equivalent spherical diameter of bubble
f_w	Bubble wake fraction
g	Gravitational constant
j	Volumetric flux density of mixture
K_b	Bubble coefficient used in Equation 1
L_c	Defined by Equation 8
r	Radial distance
r_{max}	Radius of bed
R_n	Frontal radius of bubble
T	Total time
t_{bi}	Time of bubble contact with probe
t_d	Time of bubble contact with probe A
t_p	Time for bubble to rise from probe A to probe D
U_b	Absolute velocity of bubble
U_{bo}	Superficial velocity of bubble phase
U_{br}	Bubble natural rise velocity
U_{mf}	Minimum fluidization velocity
U_o	Superficial gas velocity

X	Horizontal dimension of bubble used in Equation 8
X_p	Distance from bed center of probes B and C
Y	Vertical dimension of bubble used in Equation 8
Y_p	Vertical separation of probes A and D
δ_b	Volume fraction of bubbles
δ_{bi}	Local volume fraction of bubbles cross-sectionally
ϵ_f	Void fraction of bubbling bed as a whole
ϵ_{mf}	Void fraction at minimum fluidization
ρ_b	Density of bubble phase
ρ_d	Density of dense phase
$\langle \rangle$	Averaged over bed cross-sectional area
—	Arithmetic mean value or a weighted mean value

ABSTRACT

Bubble velocity in a fluidized bed was determined with the use of an assembly of optical probes. Two probe assemblies were developed, one employed the use of three optical probes, the second used four. The four probe assembly proved most adequate in use for local bubble velocity determination. The volumes of bubbles detected by the four-probe assembly were also determinable from the probe outputs.

The four probe assembly was used to determine bubble velocities in a fluidized bed of sand. The sand was irregular in shape and had an average particle diameter of $515 \pm 68 \mu\text{m}$. Experimental results were compared to the Davidson and Harrison equation and a modification of this equation. The modified equation,

$$U_b = C_o^* (U_o - U_{mf}) + U_{br} ,$$

incorporated a distribution coefficient C_o^* , which compared local bubble flow to the cross-sectional average.

INTRODUCTION

A fluidized bed is a column in which solid particles are suspended by an upward fluid flow. The phenomenon of fluidization occurs when there exists a balance between the upward drag force exerted on the particles by the fluidizing fluid and the downward gravitational force on the particles.

Fluidization was first successfully employed for industrial purposes in 1922 when Fritz Winkler demonstrated the use of fluidized beds for the gasification of coal. Since Winkler's gasification process, fluidization technology has had many successful breakthroughs. Although there are many successful fluidized bed operations in a wide variety of applications in industry, there have been numerous costly failures in the development of new fluidization processes. These failures can be attributed to the lack of predictable knowledge of what is happening within the fluidized bed. For this reason much of the recent research and development efforts in fluidization technology have been in the area of fluidization fundamentals such as fluid-particle interactions, and the hydrodynamic properties of fluidized beds.

Regimes of Fluidization

Gas-solid systems comprise to a large extent the majority of fluidization processes. In gas-solid systems the state of fluidization varies widely depending on gas velocity and particle properties. Except for a limited range of conditions under which individual particles can be said to be uniformly dispersed, particles in gas-solid systems aggregate, giving rise to several distinct flow regimes. There are at least four distinct regimes that have been observed [1]. They are:

1. Particulate regime
2. Bubbling/slugging regime
3. Fast fluidization
4. Dilute-phase flow

The particulate regime is that in which the gas velocity is above the minimum velocity required to incipiently fluidize the solid particles but below that at which bubbles are formed. The particulate regime is characterized by a smooth expansion of the bed with the particles evenly spaced and the fluid smoothly passing through the interstices without the formation of bubbles. In most gas-solid systems particulate fluidization is only observed for a narrow range of gas velocities.

As the velocity of the gas is increased beyond the range of particulate fluidization, the particles aggregate and voids or bubbles are formed. The phenomenon of bubbling in a fluidized bed is thought to be the result of instabilities in the lower regions of the bed [2]. Bubbles grow in size as the velocity of gas is further increased beyond that at which bubbles first form. If the bed is of sufficient height, the bubbles will coalesce and the formation of a single slug results. The diameter of the slug approaches the radial dimension of the bed with further increases in gas velocity.

The regime termed fast fluidization refers to that state at which bubble or slug stability diminishes. The bed becomes turbulent and considerable entrainment of solids results. The bed is described as a "dense entrained suspension characterized by an aggregative state in which much of the solid is, at any given moment, segregated in relatively large densely packed strands and clusters" [1].

Dilute-phase flow results when the concentration of solids in the bed is low and the gas velocity is well above the terminal velocity of the solid particles. All solids are entrained and carried out of the bed in the exiting gas.

Motivation for Fluidization Research

Many of the industrial applications of fluidized beds and those of practical interest require that the operation of the bed be in the bubbling/slugging regime. Bubbling in gas fluidized beds is important to mass and heat transfer as well as mixing in the bed. It has been determined one of the most important factors governing the extent of chemical conversion in a fluidized bed reactor operating in the bubbling regime is the diameter and velocity of bubbles in the bed. Generally, the smaller the diameter bubble, the greater is the extent of chemical reaction. Much interest therefore lies in the determination of the characteristics of bubbles within fluidized beds.

Research Objectives

It is the objective of this research to develop an optical probe assembly for use in the determination of local bubble properties in a freely bubbling gas-solid fluidized bed.

PREVIOUS RELATED RESEARCH AND BACKGROUND

Interest in the bubble phase in fluidized beds has led to the development of various techniques with which to investigate bubble characteristics and properties. Considerable stores of information pertaining to bubbles in fluidized media are present in the literature, yet to date there exists no adequately accurate direct measurements available on the size and corresponding velocity of bubbles in three-dimensional beds. Methods used to determine bubble velocities indirectly or by analogy are given in the following.

Two Dimensional Fluidized Beds

The use of two-dimensional fluidized beds and cine photography proved to be a viable means of qualitatively measuring and observing bubble properties and characteristics. The two-dimensional bed consists of two transparent parallel plates in close proximity allowing the observation of bubbles as they contact the transparent surfaces. Much insight into the mechanisms of bubble growth and coalescence, as well as information on the spatial distribution and shape of bubbles within fluidized beds, has been gained using two-dimensional beds. However, many differences between two and three-dimensional beds exist, and no accurate analogy can always be made between the two types of beds. Geldart [3] has proposed such an analogy, but with insufficient experimental data there is doubt in its validity.

X-ray Techniques

Other investigators used X-ray techniques to observe bubbles within three-dimensional beds [4,5,6]. Although the X-ray method of bubble observation has had wide application, the method is of limited value, due to the difficulty of identifying a particular bubble

when high concentrations of bubbles are present. The bed dimension is also limited in the direction of ray transmission to about 30 centimeters.

Probe Techniques

The technique most commonly employed for the detection of bubbles within fluidized beds is the use of intrusive probes. Under normal operating conditions bubbles are well distributed over the cross section in the lower regions of the fluidized bed. As bubbles rise, they coalesce with adjacent bubbles and grow. At sufficient heights above the gas distributor a single train of bubbles rises along the bed center line. Techniques limited to the detection of undisturbed, single rising bubbles such as X-ray, and the early probes which were large in size, were of limited use in the lower regions of the bed. With the development of smaller probes, local bubble measurements within groups of bubbles in the lower regions of the bed have been made possible. Probes have been used extensively in the measurements of fluidization hydrodynamics. Various types of probes have been developed; the resistivity probe, the inductance probe, the capacitance probe, the pneumatic probe, and the optical probe are the typical probes used. Of these, the capacitance, pneumatic, and the optical probes have had the highest degree of success in bubble detection.

Capacitance Probes

The early capacitance probes consisted of two opposing plates. Plate-type capacitance probes never gained acceptance as a means of bubble detection, due largely to the shape of the probes which caused destruction of the rising bubbles rather than the determination of the local state of bubble hydrodynamics. However, Werther and Molerus [7] developed an inobtrusive needle capacitance probe which proved adequate in its utility as a means of bubble detection. Though capacitance probes have had extensive use in the study of bubble phenomena, they suffer three major shortcomings: (1) the signal-to-noise ratio becomes

unacceptably low when probes and cables are long, (2) the dependence of probe calibration on bed material and bed operating conditions, and (3) the need for sophisticated and expensive equipment for signal analysis.

Pneumatic Probes

Pneumatic techniques have had success in the measurement of bubble properties. One major advantage of this technique is the capability of bubble detection at high temperatures and pressures. The determination of gas velocity in the free board (i.e., region of bed directly above fluidized media) and within bubbles is possible when the probe is used as an anemometer. A probe of small dimension developed by Flemmer [8] has proven adequate as a bubble measurement device. The pneumatic probes have the drawback of slow response time due to the dead volume of the system (pressure transducer and piping), limiting the detection of bubbles with frequencies ranging from 0-4 Hz.

Optical Probes

The design of the optical probe has forgone various changes since the technique was first used by Yasui and Johnson [9]. Opto-electronic components have been reduced in size to where the implementation of these small electronic devices are well suited to the optical probe. Such a probe was developed by Wen and Dutta [10]. The optical probes suffer none of the shortcomings that exist for the capacitance probe, and unlike pneumatic probes are capable of detecting bubbles with frequencies much greater than 4 Hz. Optical probes are simple to implement in a bubble detection scheme and require no expensive equipment for their use. The small size leads to the use of multi-probe configurations from which many bubble characteristics are determinable.

Table 1 lists some of the probes that have been implemented in the study of fluidization characteristics in gas-solid systems along with the individual investigators.

Table 1. Probes Used in Bubble Investigation.

Probe Type	Investigators
Capacitance	Toei et al. [4] Geldart and Kelsey [11] Watkins and Greasy [12] Harrison and Leung [13] Davidson et al. [14] Werther and Molerus [7]
Pneumatic	Svoboda et al. [15] Flemmer [8] Ran [16] Viswanathan [17]
Optical	Yasui and Johnson [9] Winter [18] Wen and Dutta [10] Whitehead et al. [19]
Resistivity	Burgess [20]

Measurement of Bubble Velocity

At normal operating conditions there are groups or clouds of bubbles rising in the bed. With the use of external devices such as X-ray, capacitance plates, etc., the capability of bubble discrimination is greatly diminished. Thus, bubble velocity measurements in three-dimensional beds are most successfully accomplished with the use of internal probes.

A bubble can be detected by two probes of known vertical separation. The delay in detection is utilized in the determination of the bubble velocity. Bubble size can also be approximated from the probe signal duration. Most bubble velocity data reported in the literature were determined via this method.

The method requires assumptions about the path of the bubble center in relation to the probe end. It is generally assumed that the bubble moves vertically along a line. The probe may in fact see an edge of the bubble, and the probe signal registered results in the length of a chord rather than the diameter.

An effort to determine bubble orientation with respect to the probe requires a matrix of probes which may in fact destroy the integrity of bubble shape. If a minimum number of miniature probes are used, it may be possible to develop a procedure to accurately determine bubble velocity, and bubble orientation with respect to the probes.

One such multi-probe system was developed by Burgess [20]. It consisted of five resistivity probes situated in such a manner that velocity, and bubble shape, as well as bubble orientation with respect to the probes could be determined. No other efforts in this area have been reported in the literature.

General Characteristics of Bubbles

Bubbles in fluidized media are in many respects similar to gas bubbles in a liquid. The bubbles have a precise edge or boundary and are essentially free of particles, except for a trailing wake. The bubbles are approximately spherical in shape but the lower wake region is indented as depicted in Figure 1. Perturbations about the stable or idealized shape are common due to the effects of neighboring bubbles and the vessel wall.

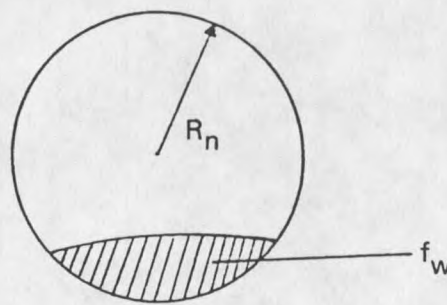


Figure 1. Idealized bubble shape.

The effect of bringing bubbles in close proximity to each other (i.e., increasing the concentration of bubbles) is to promote coalescence during which considerable distortion

of the bubble shape occurs. When the bed is operated at reasonably high superficial velocities the stable bubble shape may never exist. The bubble shape appears to be a characteristic of the fluidized medium. The discernible feature that characterizes a bubble in one material from that in another is the wake fraction, f_w , of the bubble. There exists no information for the reliable prediction of the wake fraction from particle properties, but the bed voidage, ϵ_f , has been noted to affect the wake fraction. The greater the voidage at minimum fluidization, ϵ_{mf} , the smaller the bubble wake fraction.

Rise Velocity of Bubbles

Bubbles in fluidized beds rise with a velocity that is dependent on the bubble size, bubble concentration (i.e., bubble volume fraction, δ_b), and the properties of the fluidized material.

The principal parameter affecting bubble velocity is that of size. Larger bubbles rise with a greater velocity than do smaller bubbles. Bubbles whose diameter is much smaller than the diameter of the column ($d_b/D < 0.1$) rise with a velocity that is independent of vessel diameter. When this situation exists the bed is said to be a bubbling bed. When the relative size of a bubble is increased its rise velocity becomes increasingly influenced by the vessel walls. As the bubble diameter is yet further increased in size and approaches that of the column, its velocity becomes independent of its volume and is solely a function of column diameter. When this situation exists the bed is referred to as a slugging bed, refer to Figure 2. There exists a transition region between the bubbling and slugging beds where the velocity of a bubble is affected both by the bubble and column diameters.

The concentration of bubbles can have a marked effect on the velocity of bubbles. As the concentration is increased, the bubble velocity increases. This is due to an upward flow of solids which augments the bubble rise velocity.

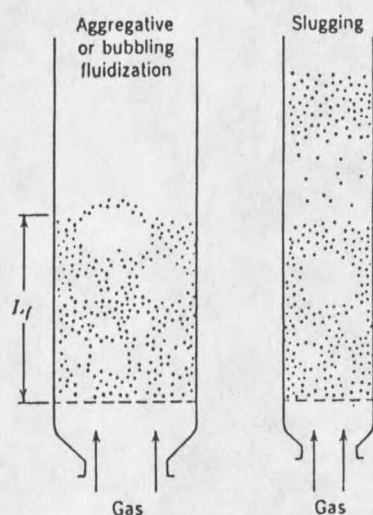


Figure 2. Bubbling and slugging beds.

Bubble velocity can also be affected by the presence of other bubbles at a higher axial position in the bed. As a bubble rises, it pulls with it a wake of solids with an upward velocity approximately equal to that of the bubble. When another bubble is drawn into this wake it will quickly catch and coalesce with the preceding bubble. In almost all cases the mechanism by which two bubbles coalesce is by the lower bubble being absorbed through the base of a preceding bubble. In general the preceding bubble suffers little distortion and rises at its normal velocity.

Properties of the solid have been shown to have only slight effects on bubble velocity [21]. The only observed effect is the slowing of bubbles with increasing wake fraction. This probably is the effect of energy dissipation in the wake of the bubble.

Isolated bubbles in a quiescent bed with a superficial gas velocity just beyond that of minimum fluidization have been observed to rise with a relatively constant velocity. This constant velocity is termed the natural rising velocity of the bubble, U_{br} . The form of the equation which predicts this velocity is,

$$U_{br} = K_b (gR_n)^{1/2} \quad (1)$$

which results from application of the steady-state mechanical energy balance to describe the rate of rise of a single spherical bubble in a low viscosity medium. R_n is the frontal radius of curvature as indicated in Figure 1. Davies and Taylor [22] developed the equation for the rise of gas bubbles in a liquid, but it was also shown to be applicable to gas-solid fluidized beds.

Application of theory to fluidized systems necessitated the postulation of the two-phase theory of fluidization. A model of aggregative fluidization may be set up by considering a fluidized bed as a two-phase system consisting of (1) a particulate phase in which the fluid flow-rate is equal to the flow-rate required to incipiently fluidize the bed, and (2) a bubble phase which carries the additional flow of fluidizing fluid.

There is evidence both supporting and negating the validity of the two-phase theory, but improved experimental determination of bubble properties are needed to test the theory.

Starting with the two-phase model, Davidson's [23] analysis of bubble motion in a fluidized bed led to the same form equation as the Davies and Taylor equation

$$U_{br} = 0.711(gd_b)^{1/2} \quad (2)$$

where d_b is the diameter of a sphere having the same volume as the spherical cap bubble. The constant K_b was determined experimentally to be 0.71 by many investigators, but values have been reported in the literature ranging from 0.57 to 0.85 for a bubbling bed. A value of 0.35 for K_b has been reported by most investigators of slugging beds where d_b is the bed diameter.

Davidson's model successfully accounted for the movement of both gas and solid as well as the pressure distribution about rising bubbles. Models of fluidization have also been developed by Jackson [24] and Murray [25], but will not be elaborated here.

The next consideration was that of the rise velocity of a crowd of bubbles. Nicklin [26] developed the first successful approximation for gas-liquid systems, which was later applied to fluidized beds by Davidson and Harrison [21]. It was assumed the relative velocity between bubble and emulsion is unaffected, but that the emulsion has an upward velocity of $U_o - U_{mf}$, from the two-phase theory. With this approximation the absolute rise velocity of bubbles in a freely bubbling bed is given by

$$U_b = U_o - U_{mf} + U_{br} \quad (3)$$

where

$$U_{br} = 0.711(gd_b)^{1/2} \quad (2)$$

Equation 3 was derived based on the concept of a single bubble rising in an infinite medium and from analogy with slug flow. Hence, the equation cannot account for interference of other bubbles, or the effects of nonuniform bubble flow. An equation developed by Weimer and Clough [27] has attempted to account for those nonidealities. The equation was derived from concepts quite different from those of previous investigators, yet the result is similar to that of Equation 3. The development of the equation was obtained assuming churn-turbulent flow throughout the fluidized bed and by considering a multiple-particle drag coefficient. The improved bubble velocity equation has the form

$$\bar{u}_b = C_o \langle u_{bo} \rangle + 0.71 \left[\frac{gd_b(\rho_d - \rho_b)(1 - \langle \delta_b \rangle)}{\rho_d} \right]^{1/2} \quad (4)$$

where C_o is a distribution coefficient. This coefficient represents the effect of non-uniform bubble flow and volume fraction. A value of C_o equal to 1 is indicative of uniform bubble flow cross-sectionally. Values greater than 1 represent bubble flow tending toward the bed's axial center, and values less than 1 represent bubble flow near the wall. The coefficient is determined from

$$C_o = \frac{\langle \delta_{bj} \rangle}{\langle \delta_b \rangle \langle j \rangle} = \frac{\frac{1}{A_T} \int A_T \delta_{bj} dA_T}{\left[\frac{1}{A_T} \int A_T \delta_b dA_T \right] \left[\frac{1}{A_T} \int A_T j dA_T \right]} \quad (5)$$

where each integral representing an average over the bed cross-section, can be determined from

$$\langle F \rangle = \frac{1}{A_T} \int A_T F dA_T = \frac{2\pi}{\pi r_{\max}^2} \int_0^{r_{\max}} (rF) dr \quad (6)$$

for $F = \delta_{bj}$, and δ_{bj} .

These integrals are evaluated numerically using data from an experimental investigation.

No other theoretical treatment of bubbles in fluidized media have been developed to date.

EXPERIMENTAL EQUIPMENT

The equipment used in the development of the bubble velocity determination procedure was of three types: the fluidizing system, the optical probes, and the Apple II micro-computer/Cyborg data acquisition system. A schematic drawing of the overall experimental system is shown in Figure 3. A two-dimensional fluidized bed was used for determination of bubble shape characteristics.

Fluidizing System

The fluidization column was constructed of 6.35 mm thick, clear plexiglass with an inside diameter of 34.3 cm and a height of 1.5 m above the distributor plate. Flanges, 1.9 cm thick, were attached to the top and bottom of the column to enable the sections of the fluidization system to bolt together. The column was supported by an angle-iron frame, anchored to the floor.

A 15 cm high galvanized steel funnel, 34.3 cm diameter at the bottom and 48.3 cm diameter at the top, was fitted with a rubber gasket and bolted to the top flange of the column. This funnel served as a disengagement section. A steel perforated plate sandwiched between two 6.4 mm thick, plexiglass plates was attached to the top of the funnel, which allowed for the filling of the column. The exiting air was vented outside through a 20 cm stove pipe.

The gas distributor consisted of a 20 gauge steel plate with 1/8 inch drilled orifices at a 3/4 inch pitch. A layer of 100 mesh wire fabric was placed directly under the gas distributor. The distributor was fitted with rubber gaskets, and bolted between the lower bed flange and the plenum.

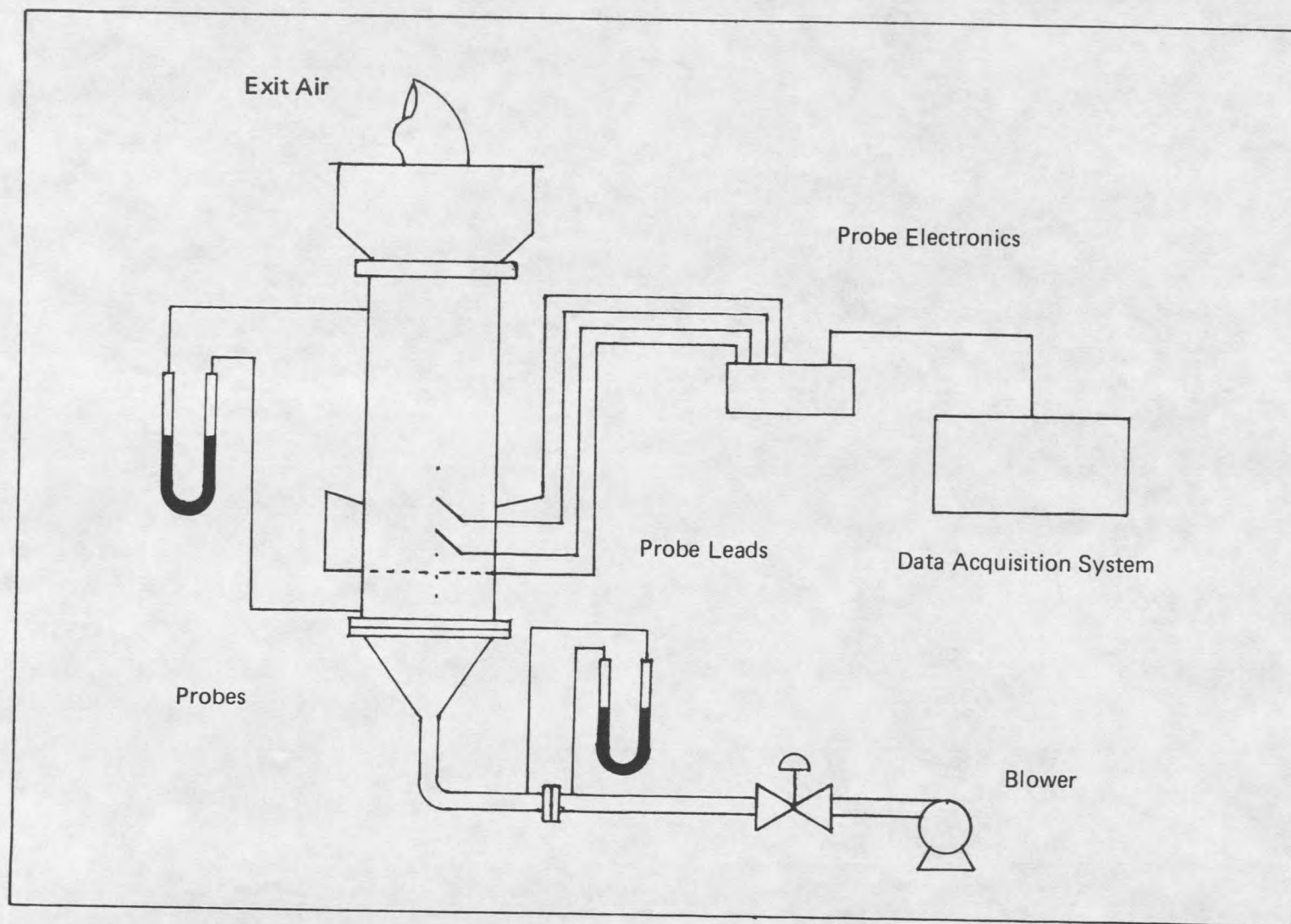


Figure 3. Overall experimental system.

The plenum fabricated from galvanized sheet metal was a funnel 15 cm high, 34.3 cm diameter at the top, and 5.5 cm diameter at the bottom. The plenum was filled with 1/2 inch Berl saddles to assist better distribution of the gas.

Air was supplied to the bed by a size 4L Sutorbilt blower driven by a 2.24 KW electric motor. A flexible rubber hose connected a 5.1 cm nominal diameter schedule 40 pipe to the plenum. Air flow rates were measured utilizing a 1.59 cm orifice with vena contracta taps in the feed line. The pressure difference was read from water-filled manometers. A gate valve located in the feed line, along with a bypass valve, was used to control the air flow rates.

Optical Probes

Optical probes used in this work were developed by Wen and Dutta [10]. In this investigation two different sized probes were utilized.

The first size probe was constructed from 0.635 cm outer diameter stainless steel tubing. The probe was 30 cm in length with an 8 mm by 5 mm window machined out at one end; refer to Figure 4. A photoemitter (Archer TIL906-1) opposing a photodetector (Archer TIL414) was located at each end of the window.

The second size probe was constructed from 0.476 cm outer diameter stainless steel tubing and was 30 cm in length. A photoemitter (Motorola MLED15) opposed the detector (Motorola MRD150) in a 6 mm by 4 mm window in the probe end.

Bed access by probes was accomplished utilizing drilled ports. Each port was fitted with a 1/4 inch bushing. When not in use they were plugged with a 1/8 inch pipe plug. Three vertical rows of ports were located at 120° circumferential separation around the bed. Each vertical row had nine ports, whose centers were located 9.2, 12.2, 15.2, 18.2, 22.2, 26.2, 31.2, 36.2, and 41.2 cm above the gas distributor. Located on each probe was a

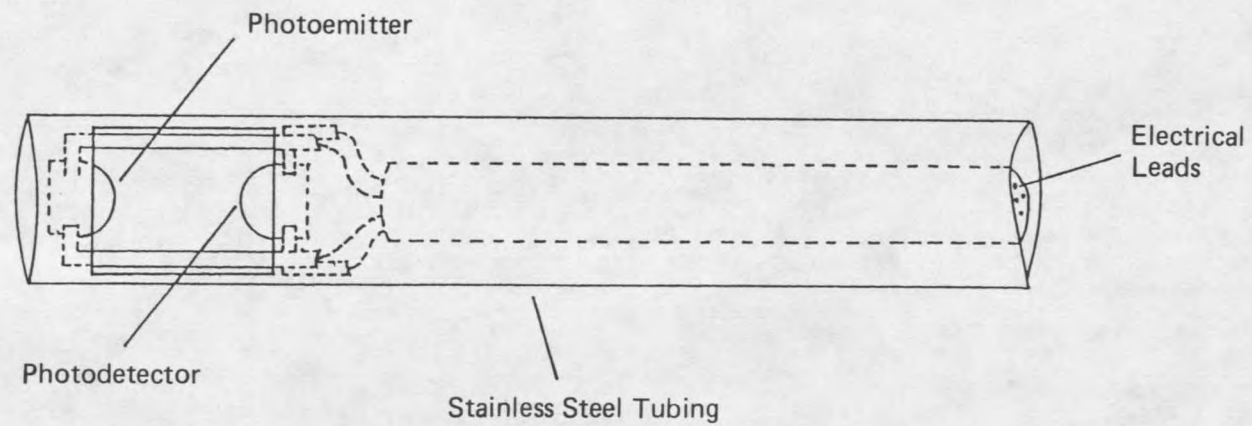


Figure 4. Optical probe construction.

compression fitting with 1/8 inch threaded end which could be coupled with a bushing in a probe port. The compression fitting also allowed for radial adjustment of the probe.

Data Acquisition System

The data acquisition from probe outputs was accomplished with the Apple II micro-computer interfaced with a Cyborg/Isaac automated acquisition and control system. The analog probe output ranged from 0.0 volts to -1.0 volts. The signal was converted from analog to digital using any of the four Schmitt Triggers on the Cyborg system. The result of analog to digital conversion is a four-bit binary number which can be stored in the Apple II memory for further processing. The Cyborg system is capable of sample intervals of 1 millisecond with a resolution of 1.00352 milliseconds. This was the sample rate used in data acquisition from probe outputs.

Two Dimensional Fluidized Bed

For purposes of bubble shape determination a two dimensional fluidized bed was constructed. Two parallel 0.635 cm thick plexiglass plates 47 cm by 122 cm with 1.9 cm separation were supported by an angle iron frame. The gas distributor was porous steel plate. Air flow was supplied to the bed by house air in the laboratory.

DEVELOPMENT OF PROCEDURE

In the development of the procedure to determine bubble velocity, two probe assemblies were tested. Figure 5 shows the probe configuration in the fluidized bed. The first probe assembly utilized three probes, probes A, B, and C as depicted in Figure 5. In the second probe assembly probe D was added.

During the developmental stages, the output of the probes were processed by the Apple II/Cyborg system and viewed on the computer terminal. The 0.635 cm diameter probes were used, and fluidized particles were glass spheres, 439.0 μm in diameter.

A computer algorithm was developed to effectively discern between useful and non-useful data. The algorithm also determined information regarding bubbles in close proximity to the detected bubble.

Application of the four probe assembly in a bubbling/slugging bed of sand followed the assembly development. Bubble hydrodynamics were studied in a bed of sand particles, $515.0 \pm 68.0 \mu\text{m}$ in diameter.

Three Probe Assembly

It was desired to keep bubble disturbances resulting from probe contact to a minimum. Therefore, the first probe configuration used three optical probes. This was the minimum number of probes required to adequately determine bubble velocity, volume, and orientation with respect to the probes. The three probes were situated in a horizontal plane at a given distance above the gas distributor, and were secured with the compression fittings in the probe ports. The end of probe A was at the axial center line of the bed, and ends of probes B and C were set at a distance, X_p , from center which corresponded to the

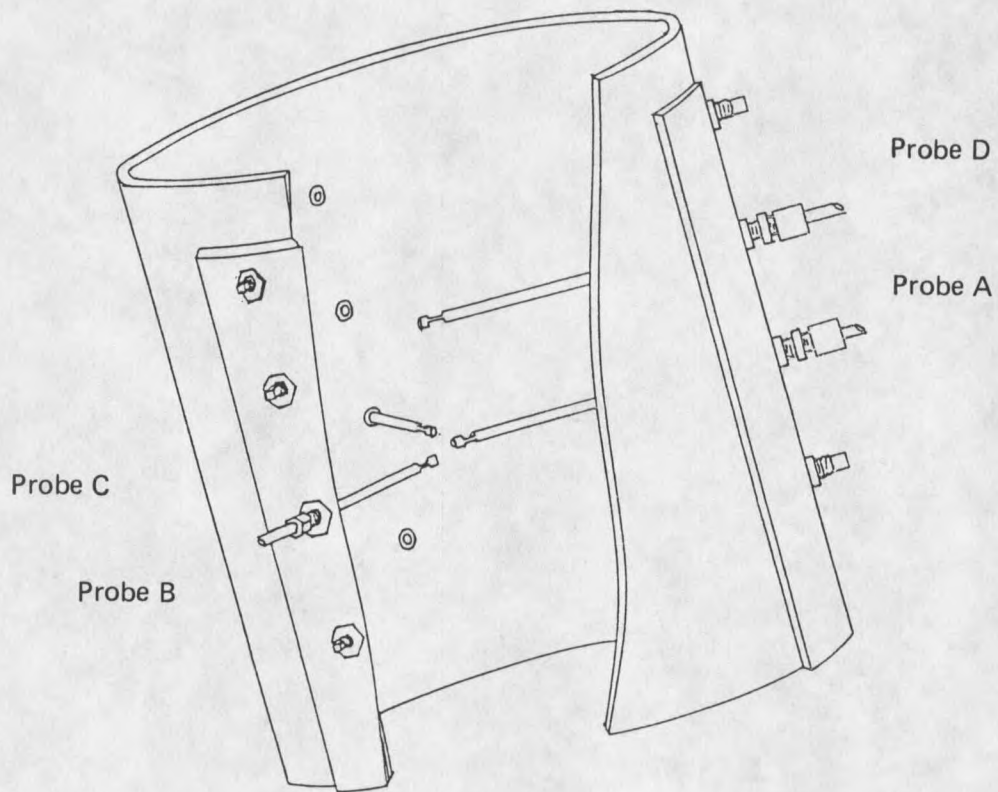


Figure 5. Optical probe configuration in fluidized bed.

dimensions of the bubble to be detected. Figure 6 illustrates the required bubble-to-probe orientation to enable the determination of bubble velocity.

Velocity Determination

To determine the bubble velocity, the bubble must contact the probes with the required orientation, and bubble shape must be assumed or known.

The shape of the bubble was approximated by observation of bubbles in a two dimensional bed with the fluidized medium of interest.

The velocity can then be determined from

$$U_b = L_c/t_d \quad (7)$$

where L_c is determined from the ratio of the maximum horizontal dimension, X , to the central vertical chord, Y , of the bubble as shown in Figure 6, and t_d is the duration of response for probe A. Then

$$L_c = 2X_p(Y/X) \quad (8)$$

where X_p is the distance from bed center to ends of probes B and C.

Results

The three probe method proved inadequate in its determination of bubble velocity. This was largely due to the method's requirement of assuming a bubble shape.

There was no way to determine if the shape of the detected bubble was indeed the assumed shape.

The fact that the horizontal dimension of the bubble was required to correspond to probe settings also limited the method's utility. To determine the velocity of another size bubble the settings of probes B and C had to be changed to correspond to the new bubble size.

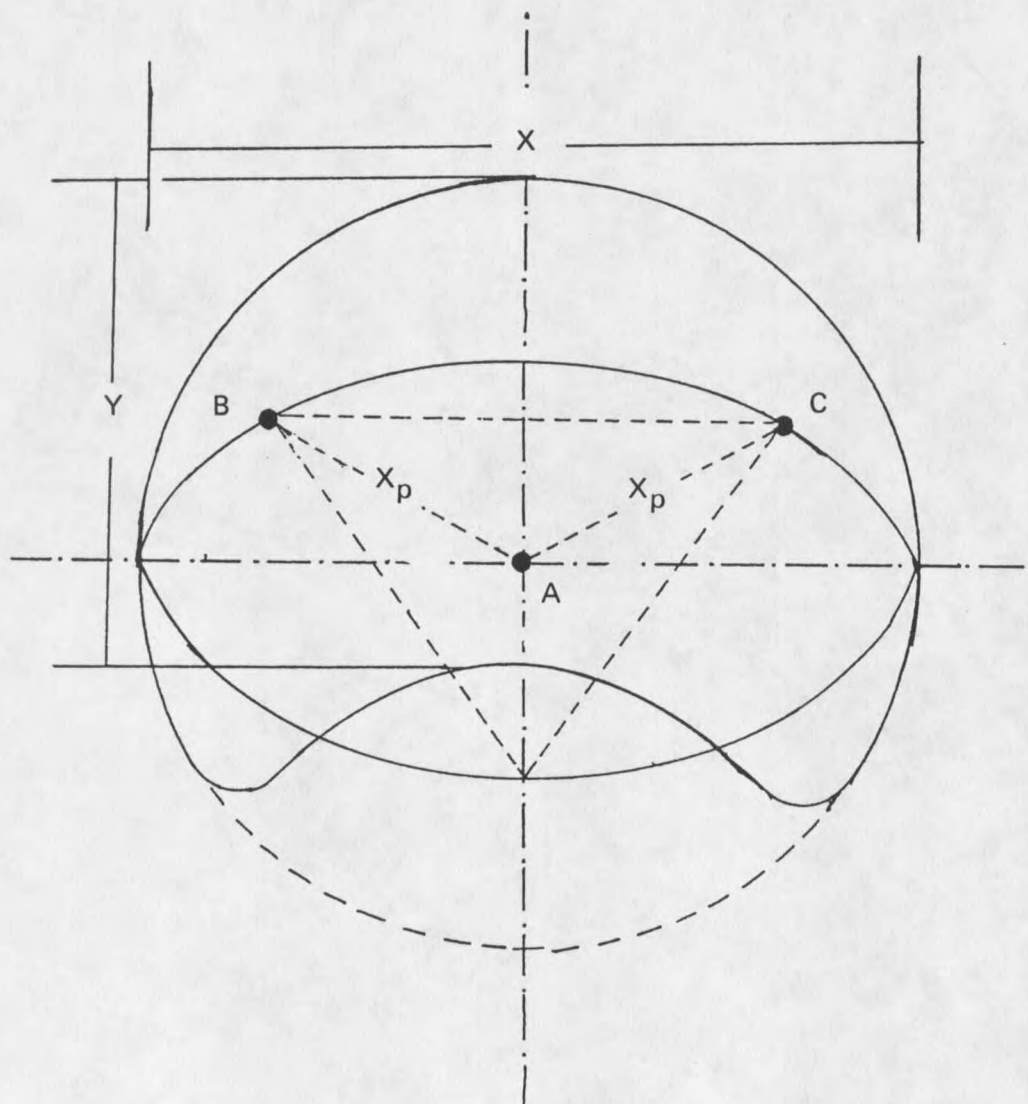


Figure 6. Bubble-to-probe orientation for three probe assembly.

Four Probe Assembly

The inadequacies of the three probe assembly prompted the addition of a fourth probe, probe D depicted in Figure 5. Figure 7 illustrates the required bubble-to-probe orientation to determine bubble velocity and volume with the four probe assembly. The utility of the four probe assembly was greatly enhanced over that of the three probe assembly. Unlike the indirect determination of bubble velocity with the three probe method, velocity with the four probe method can be directly determined.

Velocity Determination

The bubble velocity is directly determined from

$$U_b = Y_p/t_p \quad (9)$$

where Y_p is the vertical separation of probes A and D, refer to Figure 7, and t_p is the time elapsed for the bubble to rise from probe A to D.

Bubble velocities were determined for those bubbles that contacted the probes properly. The proper orientation was determined by the response of probes B and C. If the initiation and duration of probe responses for probes B and C are nearly the same, and probe A detected the bubble some time prior to probes B and C, the bubble had to have been oriented as depicted in Figure 7. Velocity and volume can be determined for all bubbles with correct orientation to the probes, as long as the bubble size is larger than the settings, X_p , of probes B and C. Therefore, the settings, X_p , of probes B and C are set somewhat less than the dimension of bubbles expected at the level at which measurements are made.

Volume Determination

The volume of a bubble can be determined from the probe response of the four probe assembly. This was accomplished by assuming the detected bubbles are spherical cap in

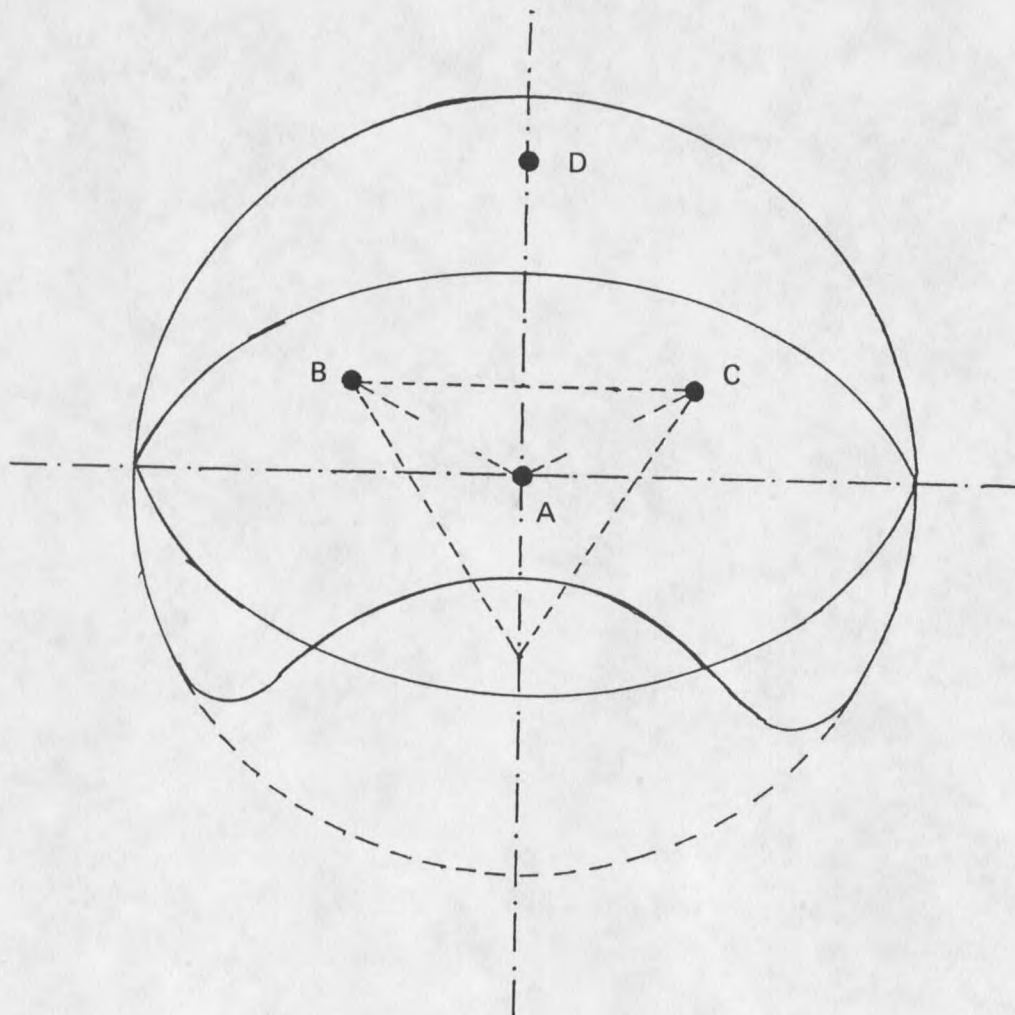


Figure 7. Bubble-to-probe orientation for four probe assembly.

shape. This assumption was validated by observing bubbles in the two dimensional bed. The bubble volume was then determined by equations given in Appendix A.

Bubble volume could also be approximated from assumed bubble shapes conceived with the aid of probe signals. This was possible even when the detected bubble was not oriented exactly in the correct manner.

Results

The four probe assembly proved to not only give an accurate direct measurement of bubble velocity, but bubble volume was also determinable. Bubble-to-probe orientation was easily determined from the response of the four probe assembly. The method was greatly enhanced from that of the three probe method in that a range of bubble sizes could be detected with a single setting of the probe assembly.

Figure 8 compares results of bubble velocities determined with the three and the four probe assemblies. The data is for 9.0 cm diameter bubbles in the glass spheres, at a superficial gas velocity of $2.4U_{mf}$. The minimum fluidization velocity for the glass spheres was 16.5 cm/sec.

Development of Algorithm

Bubbles rise in a somewhat random manner in the fluidized bed. Therefore, the time between the detection of properly oriented bubbles varies. A computer algorithm was developed to monitor data acquisition, and to analyze for bubbles of correct orientation to the probes. When such a bubble was detected subsequent storage of that data was desired.

The result of the input to the Cyborg system from the probes was a four-bit binary number. The decimal equivalent to that number was stored in the Apple II memory. The sampling interval was chosen as 1 millisecond and the duration of sampling of the probe outputs was limited to 4.5 seconds due to memory limitations of the Apple II computer.

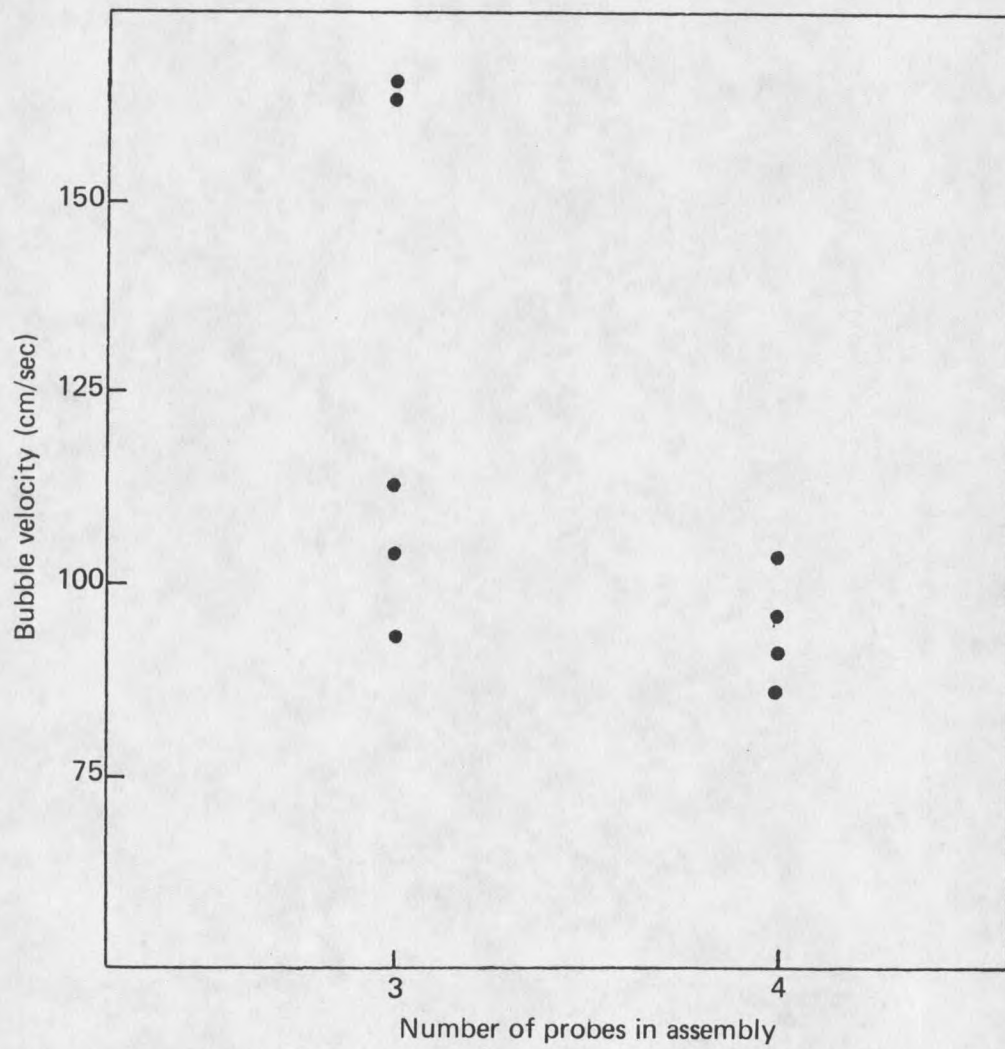


Figure 8. Comparison of three and four probe assemblies.

The sequence of events was then (1) data acquisition from the probe assembly for a period of 4.5 seconds, (2) analysis of the 4500 data points to determine if the desired event took place (i.e., a bubble contacting the probes with the desired orientation), and (3) if the desired event occurred, output of that data in its raw form resulted.

It was desired to gain information on bubbles in close proximity to the detected bubble. A subroutine was therefore added to the computer program which did this analysis.

Procedural Methods

The four probe assembly was used to measure bubble velocity and size relationships in a fluidized bed operating in the bubbling/slugging regime.

The particulate material in the bed was screened sand of irregular shape. The sand density was 2.54 ± 0.15 grams/cm³. The average particle diameter was 515.0 ± 68.0 μ m determined by the use of an image analyzer (Cambridge/Olympus Quantinent 10). The complete analysis of the particulate material is given in Appendix B. The static bed depth was 45 cm.

The fluidizing fluid was air at 30.0 ± 2.0 °C. The minimum fluidization velocity was measured using the bed pressure drop technique, U_{mf} being defined by the intersection of the graphical asymptotes of the two sections of pressure drop against gas flow curve during decrease in gas flow. The measured minimum fluidization velocity of the material was 5.60 cm/sec. The operating superficial gas velocities were $1.6U_{mf}$, $2.0U_{mf}$, and $2.4U_{mf}$. The corresponding expanded bed depths were approximately 47, 49, and 54 cm, respectively.

The 0.476 cm diameter probes were employed during the investigation of bubble velocity in fluidized sand. The smaller probes were required because of detected bubble disturbances with the 0.635 cm diameter probes.

The four probe assembly was placed at three different levels in the bed 15.2, 26.2, and 31.2 cm above the distributor plate. This was to enable a wider size range of bubbles to be detected.

Operational Procedures

The first step was to secure the four probe assembly in the probe ports at one of the three levels in the bed. Probes B and C as depicted in Figure 7 of this text were then set at a distance that would allow for the detection of the expected size range of bubbles at that level in the bed.

The flow rate of the gas was then set at the desired velocity, and was measured with the water manometer of the orifice meter.

The Apple II/Cyborg system was turned on with the probe outputs connected to the Schmitt triggers on the Cyborg. The power to the probe assembly was then turned on. The program controlling the data acquisition was then executed. The system was allowed to operate until the desired number of bubbles were detected.

The local bubble volume fraction across the bed cross section was detected for each level in the bed with the use of one of the optical probes set at several radial positions. A volume element at the point of the probe contains on a time-averaged basis a fraction δ_b of gas bubbles. This fraction theoretically equals the ratio of the time in which the probe registers bubbles to the total time for which measurements have been made T:

$$\delta_b = \lim_{T \rightarrow \infty} \frac{1}{T} \sum t_{bi} \quad (10)$$

Data Reduction

The output of the system was a hard copy of the raw data in decimal form. The series of numbers which corresponds to a bubble, oriented in the correct manner, were converted to a graphical representation of the individual probe outputs as shown in Figure 9. The

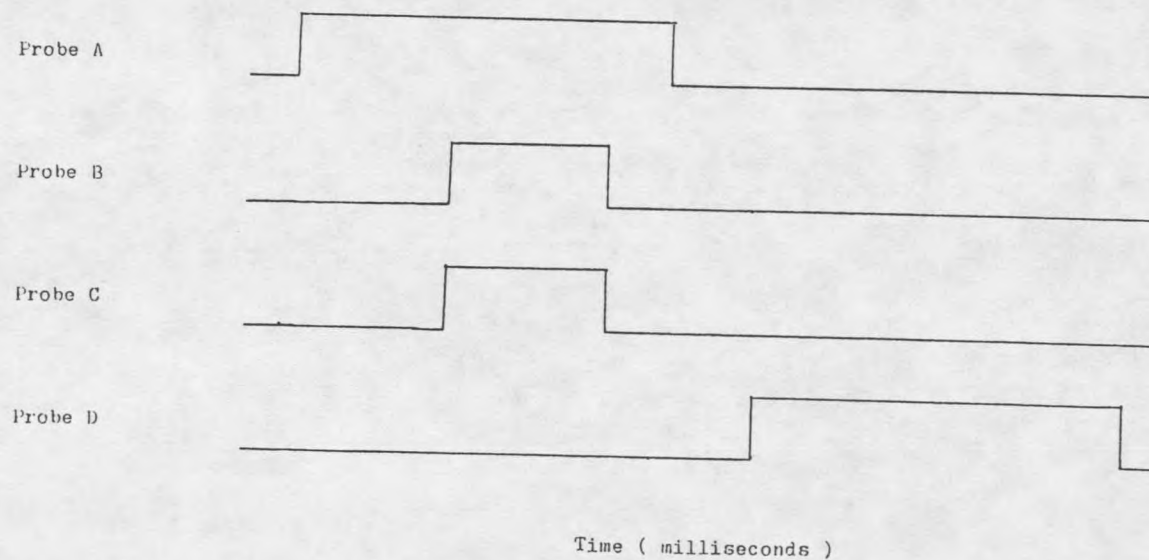


Figure 9. Probe assembly output for correctly oriented bubble.

computer algorithm did not analyze only for bubbles specifically oriented in the correct manner but also for those that deviated slightly from the desired bubble-to-probe orientation. The final analysis was done by graphing the raw data as in Figure 9. From this it was determined if the probe responses were usable for the determination of bubble velocity as well as volume. If so, the velocity was directly determined using Equation 9. The volume was then determined using the equations in Appendix A.

RESULTS AND DISCUSSION

Probe Signal Reliability

The effect of the probe on bubble integrity was analyzed by two methods. In the first method two probes were placed in the bed at a given position, one directly above the other. The signals from both probes were compared. In the sand the signal of the upper probe was disjointed when the 0.635 cm diameter probes were used. However, when the smaller 0.476 cm diameter probes were used the signals of both probes were nearly the same.

The second method was to place a section of tubing, which was of the same dimension as the probe, into the two-dimensional bed and observe the effect of the tube's presence on the bubble. This was only done with the 0.476 cm diameter probes because the larger probes failed the first test in the sand as stated above. Observation of bubbles contacting the tubing indicated the probes had varying effects on bubbles depending on the circumstances. If the bubble was large (i.e., greater than 4 cm diameter), the probe appeared to have little or no effect on the bubble. Smaller bubbles were observed to occasionally split into two bubbles and coalesce a short distance above the probe.

Bubble Distribution

Bubble velocity measurements were made at the bed center; it was therefore desired to confirm that the bubble distribution was symmetrical about the bed axial center.

Verification of uniform dispersion of gas from the distributor plate was accomplished by the measurement of local bubble volume fraction at various radial and circumferential positions. At an axial position of 15.2 cm above the distributor plate, bubble voidage was measured at the three circumferential positions of the probe ports. It was determined that

uniform dispersion of gas existed because the bubble volume fraction curves were symmetrical at the three probe positions. It was assumed that the bubble flow was consistently symmetrical about the axial centerline at all bed heights. Measurements were therefore obtained at one circumferential position for bubble volume fraction at axial heights of 26.2 and 31.2 cm.

Figures 10-12 show that the bubble flow tends toward the bed center line with increasing height in the bed. This is what is to be expected as the center of the bed is the path of least resistance to gas flow.

The flow pattern observed at an axial position of 15.2 cm shows a greater bubble phase flow near the bed walls than in the center, as shown in Figure 10. This has been observed by other investigators [7], but there exists no explanation for why this might occur.

Figures 11 and 12 are bubble volume fraction data for axial positions of 26.2 and 31.2 cm, respectively. The general shape of these two plots indicates bubble flow tending to the center with increasing bed height. All plots at the three axial positions depict a lower bubble flow at bed center than the maximum at that bed height. This is due to the initiation of flow at the distributor near the bed wall and would not be expected to exist at higher bed heights.

Bubble Shape

To estimate the volume of a bubble an approximate bubble shape must be known. With the use of the two-dimensional bed and a 35 mm camera a number of photographs of bubbles were obtained. Bubble shape was observed to vary considerably from bubble to bubble.

When a bubble was detected by the four probe assembly the criterion for keeping that data was that the bubble be properly aligned with the probes. For this to occur the bubble would have to be fairly symmetrical in shape, thus undisturbed by neighboring bubbles. A

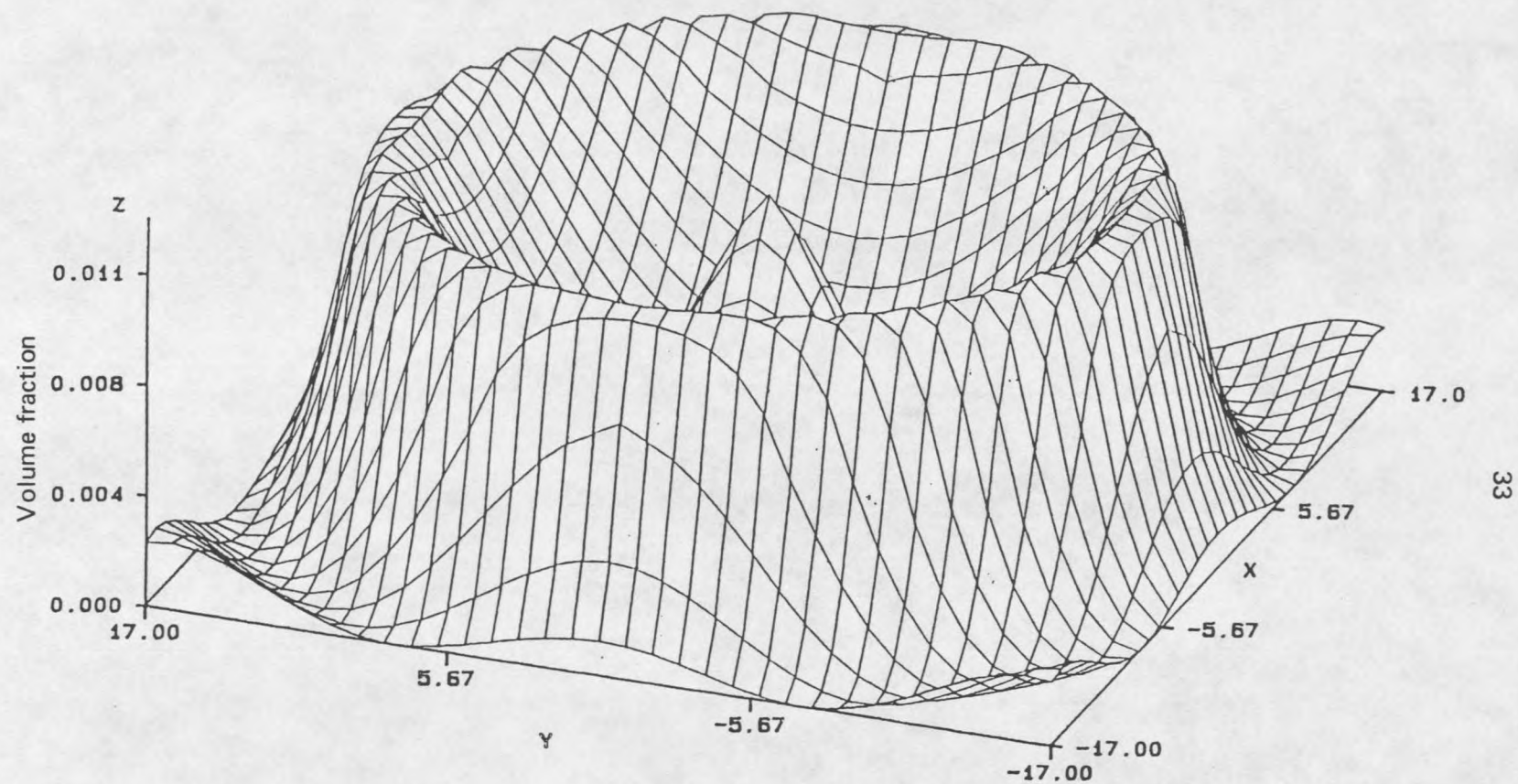


Figure 10. Bubble volume fraction at 15.2 cm bed height.

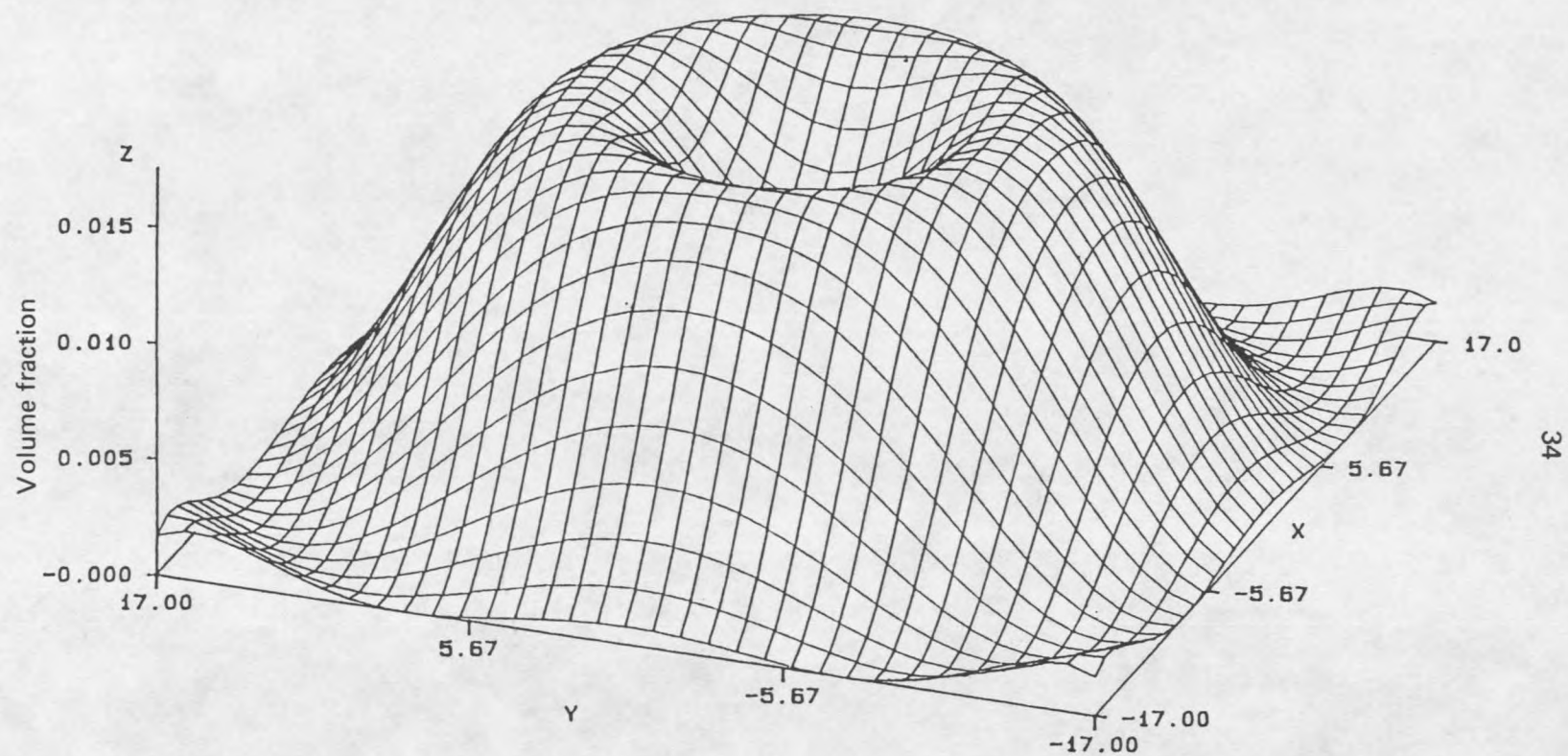


Figure 11. Bubble volume fraction at 26.2 cm bed height.

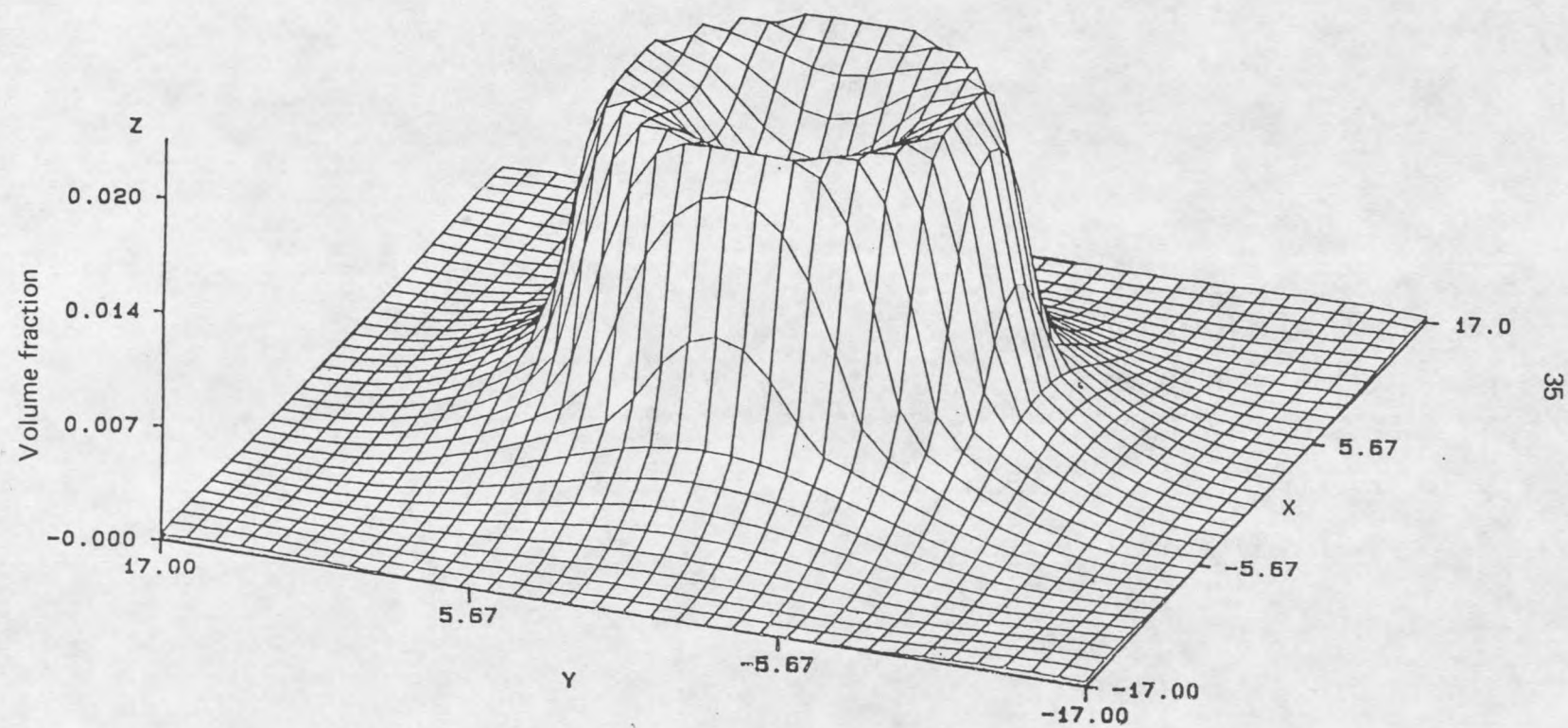


Figure 12. Bubble volume fraction at 31.2 cm bed height.

photograph of an undisturbed bubble in the two-dimensional bed was therefore used as an estimation of bubble shape in the bed, see Figure 13. From the photograph an estimation of bubble wake fraction and the shape of the bubble were obtained. This aided in the bubble volume calculations.

The undefined boundary between the bubble phase and the dense phase at the bubble top was thought to result from instabilities that have been observed with fluidization of large particles [23]. This could cause problems with probe response and velocity determination in the three-dimensional bed. It was, however, observed that when bubbles occasionally contacted the bed wall in the three-dimensional bed, the bubble interface was much more distinct than that shown in Figure 13. The walls of the two-dimensional bed probably initiated much of the particle draining within bubbles.

Bubble Velocity

Figure 14 shows a plot of absolute bubble velocity versus equivalent spherical diameter for a superficial gas velocity of $2U_{mf}$ in the bed of sand. The individual data points are differentiated among bubbles with preceding, trailing, both preceding and trailing, or no bubbles in close proximity along the rise path of the detected bubble. The solid line is a second-order least squares fit of the data.

Occasionally bubbles with unexpectedly high absolute velocities were detected. All bubbles that exhibited higher than average velocities were found to have bubbles preceding them. This was expected in that the bubble in question would have an increased velocity due to the presence of the rising wake of the preceding bubble. No conclusive statements can be made as to the effect of bubbles trailing the detected bubble.

For a freely bubbling fluidized bed Equation 3 is usually used for purposes of comparison to experimental results. The dashed line on Figure 14 is a plot of Equation 3. Experimental bubble velocities corresponding to equivalent spherical diameters greater

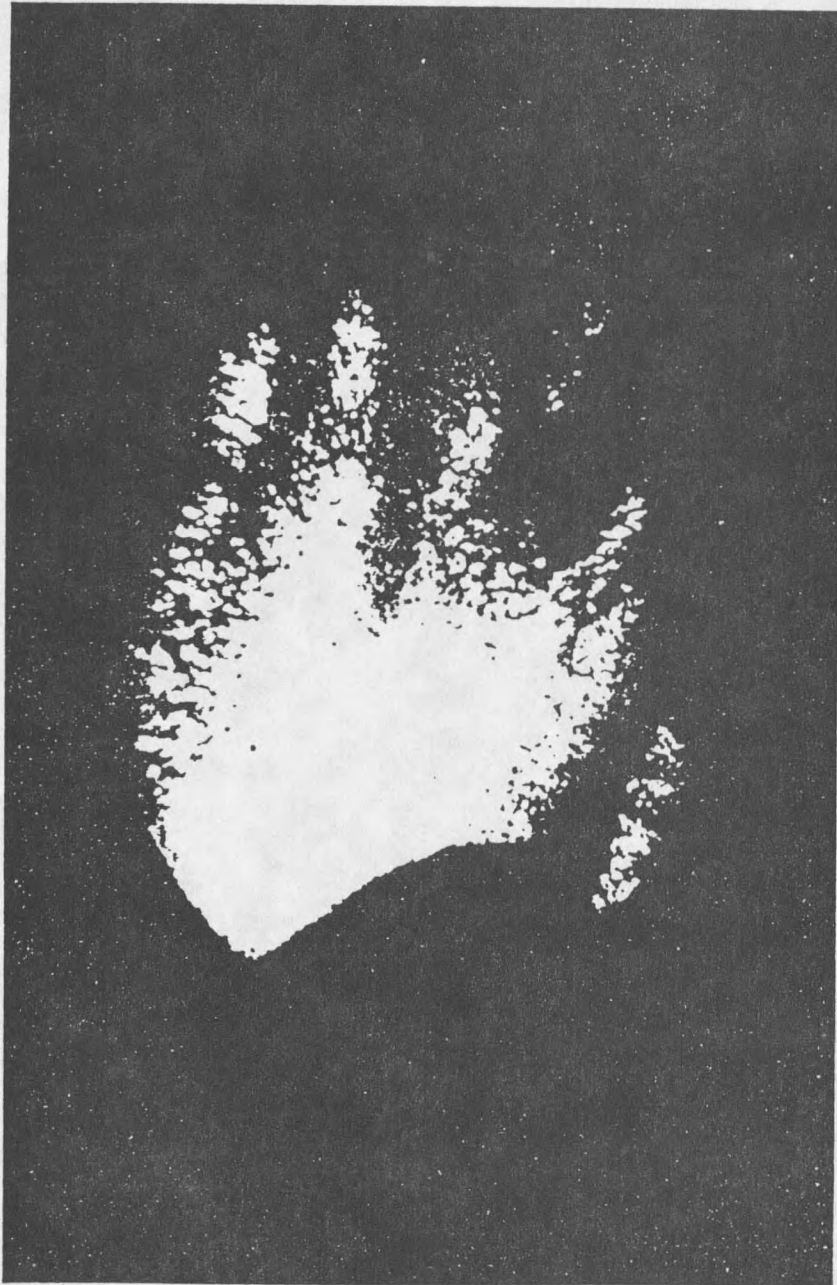


Figure 13. Photograph of bubble from two-dimensional bed.

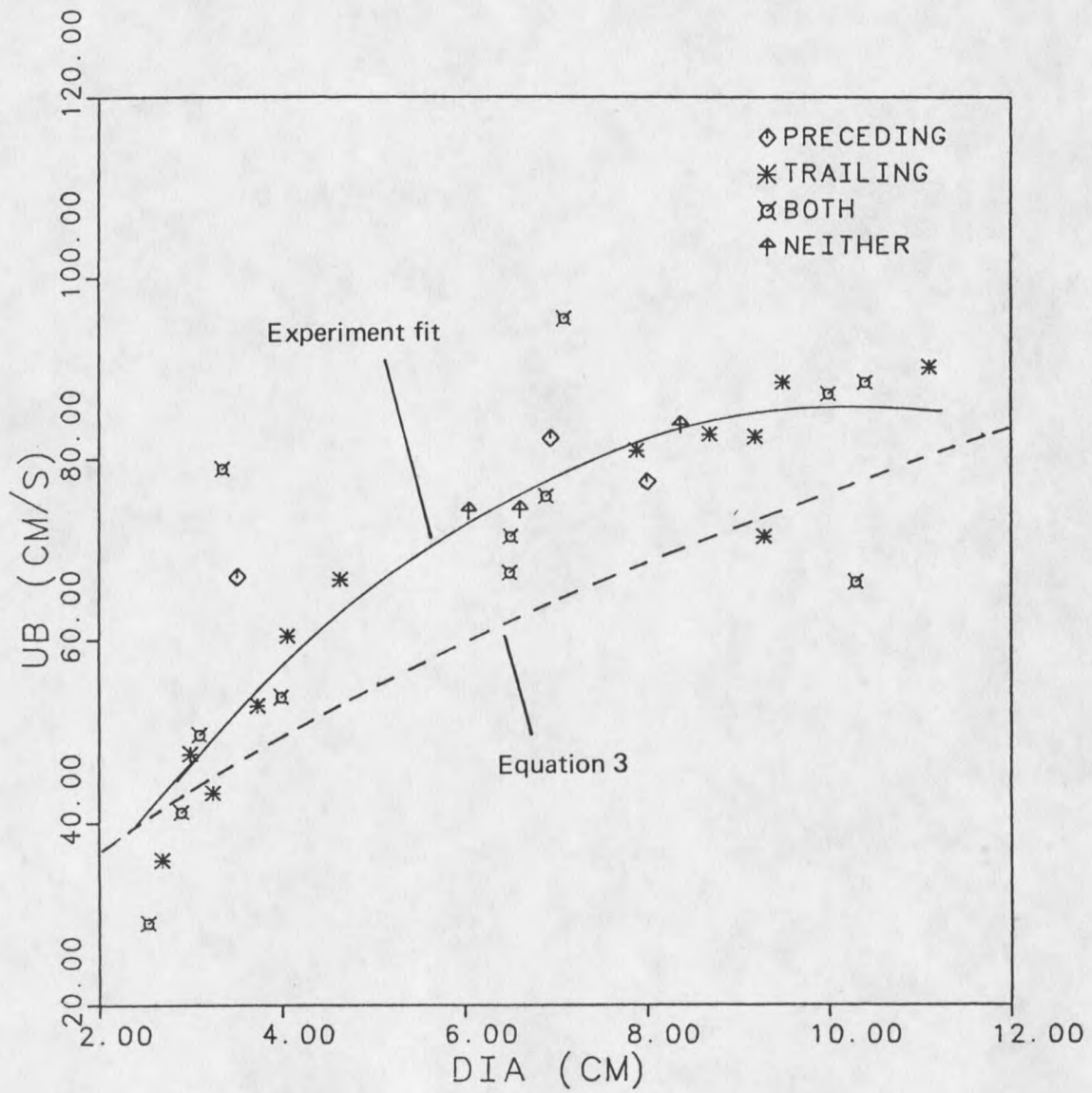


Figure 14. Bubble velocity data for $2U_{mf}$.

than 3 cm exhibit greater velocities than predicted by Equation 3. It is not surprising that Equation 3 is not predictive for bubble velocity in a freely bubbling bed. The equation was developed for a single bubble rising in an infinite medium and from analogy with slug flow.

Intuitively, from inspection of bubble volume fraction data the experimental results can be explained. For velocity data corresponding to bubbles with diameters less than 3 cm, experimental results are lower than that predicted by Equation 3. This can be explained by inspection of Figure 10, which is bubble volume fraction data at an axial position of 15.2 cm. The average bubble diameter at 15.2 cm depth in the bed was found to be 3.4 cm. Therefore, bubbles of this diameter and less would be present at that level of the bed. As shown in Figure 10, more or larger bubbles rise in a ring near the bed wall. The net flow of particulate matter must then be down in the bed center. The center of the bed is where the velocity measurements were made. Bubble velocities, which are lower than predicted, are thus explained because bubbles rising in the center of the bed meet a downward flow of solids.

At axial positions of 26.2 and 31.2 cm the net flow of bubble volume tends toward the bed center, as depicted in Figures 11 and 12. Bubbles rising at bed center should then exhibit higher velocities than predicted by Equation 3. The flow of bubbles in the bed center carry particles in a net upward direction at bed center. Bubbles rising at bed center will have their rise velocity augmented by the upward flow of solids. The experimental results are therefore explained from an intuitive standpoint.

The solids flow expected from inspection of bubble volume data was substantiated by observation of solids movement at the bed wall. The particulate matter was observed to move downward along the wall at bed heights greater than 20 cm. Below 20 cm the solids along the wall were not observed to move.

The equation developed by Weimer and Clough [27], Equation 4, tried to account for nonuniform bubble flow as well as the effect of local relative velocity between phases. An attempt was made to use this equation as a comparison to that of Equation 3 and the experimental results. Table 2 gives the values, calculated from bubble volume fraction data, required for use in Equation 4.

Table 2. Calculations for Determining Distribution Coefficient, C_o .

Bed Height (cm)	15.2	26.2	31.2
$\langle j \rangle$	0.3712	0.6222	0.4299
$\langle \delta_b \rangle$	0.0070	0.0079	0.0062
$\langle \delta_{bj} \rangle$	0.0032	0.0072	0.0075
$C_o = \frac{\langle \delta_{bj} \rangle}{\langle \delta_b \rangle \langle j \rangle}$	1.22	1.46	2.81

The bubble velocities calculated from Equation 4 were lower than those calculated via Equation 3. It would be expected that the results of applying Equation 4 would yield greater values for bubble velocity due to the values of the distribution coefficient indicating bubble flow tending toward the bed center. From comparison of other work reported in the literature [2,27] it appears that the values of experimental bubble volume fraction determined in this work were lower than required. This can probably be attributed to the optical probes inherent limitation of detecting only distinct bubbles. Gas passing through the bed in a lean dense phase will not be detected. This gas flow results from short circuiting of the gas from bubble to bubble. Therefore, the experimentally determined values of bubble volume fraction data are low. Equation 4 was not used as a comparison of experimental data because of the problems encountered with the bubble volume fraction data.

One way to utilize the bubble volume fraction data is in the form of a ratio, thus absolute values are not required and only relative magnitudes of the values are of importance. If the point value of bubble volume fraction is compared to the average cross-sectional value, a new distribution coefficient is the result. This distribution coefficient represents the increased or decreased bubble flow at a point relative to the average across the bed.

Incorporation of the new distribution coefficient into an equation of the same form as Equation 3 results in an equation that can predict local values of bubble velocity.

$$U_b = C_o^*(U_o - U_{mf}) + U_{br} \quad (11)$$

This equation can be used to predict bubble velocities at a given radial position in the bed.

The values of C_o^* calculated for the bed center are given in Table 3.

Table 3. Calculations for C_o^* .

Bed Height (cm)	15.2	26.2	31.2
$\langle \delta_b \rangle$	0.0070	0.0079	0.0062
δ_i	0.0100	0.0100	0.0150
$C_o^* = \frac{\delta_i}{\langle \delta_b \rangle}$	1.43	1.27	2.42

Table 4 compares the experimental data with Equation 11. The values of the distribution coefficients used in Equation 11 are those given in Table 3. Values calculated using Equation 4 are also given as a comparison. Equation 11 can better predict the local cross-sectional bubble velocity than the other equations used.

Bubble velocity data was also taken at operational gas velocities of $1.6U_{mf}$ and $2.4U_{mf}$. Figure 15 shows the results in the form of second-order fits to the data. The curve for $1.6U_{mf}$, when compared to that for $2.0U_{mf}$, give velocity values lower than if the two-phase theory strictly held. If the two-phase theory held, a 40 percent difference in the

Table 4. Comparison of Equations 4 and 11 With Experimental Data. Data for $2U_{mf}$.

Height at Which Measurements Were Made (cm)	Diameter (cm)	Velocity (cm)	Velocity Calculated by Equation 4 (cm/s)	Velocity Calculated by Equation 11 (cm/s)
15.2	2.5	29.2	40.8	43.2
	2.7	36.0	42.2	44.6
	2.9	41.3	43.5	45.9
	3.0	47.7	44.2	46.6
	3.3	43.4	46.0	48.4
	3.4	79.0	46.6	49.0
	3.5	66.4	47.2	49.6
	3.7	52.9	48.4	50.8
	4.0	53.8	50.1	52.5
	4.1	60.5	50.7	53.1
	4.7	66.7	53.9	56.3
26.2	3.1	49.8	44.8	46.3
	6.1	74.7	60.6	62.1
	6.5	67.4	62.4	63.9
	6.5	71.4	62.4	63.9
	6.6	74.7	62.8	64.3
	6.9	75.9	64.1	65.6
	7.0	82.5	64.5	66.0
	7.1	95.7	64.9	66.4
	7.9	81.0	68.2	69.7
	8.0	77.4	68.6	70.1
	8.4	83.9	70.1	71.6
	9.2	82.5	73.1	74.6
	9.3	71.4	73.5	75.0
	10.3	66.4	77.0	78.6
31.2	10.4	88.6	77.4	78.9
	8.7	82.8	71.3	79.2
	9.5	88.6	74.2	82.2
	10.0	87.3	76.0	83.9
	11.1	90.3	79.8	87.7

excess velocity beyond minimum fluidization would be expected. This, for example, would result in a 4.0 cm/sec difference between values for $1.6U_{mf}$ and $2.0U_{mf}$ for a diameter of 8.0 cm. The deviation from theory arises because the bubble distribution has been observed to change with changing superficial gas velocity [27]. At lower gas velocities flow profiles as shown in Figure 10 would be expected at greater axial positions. With higher gas velocities fully developed flow is attained at lower positions in the bed; this would correspond

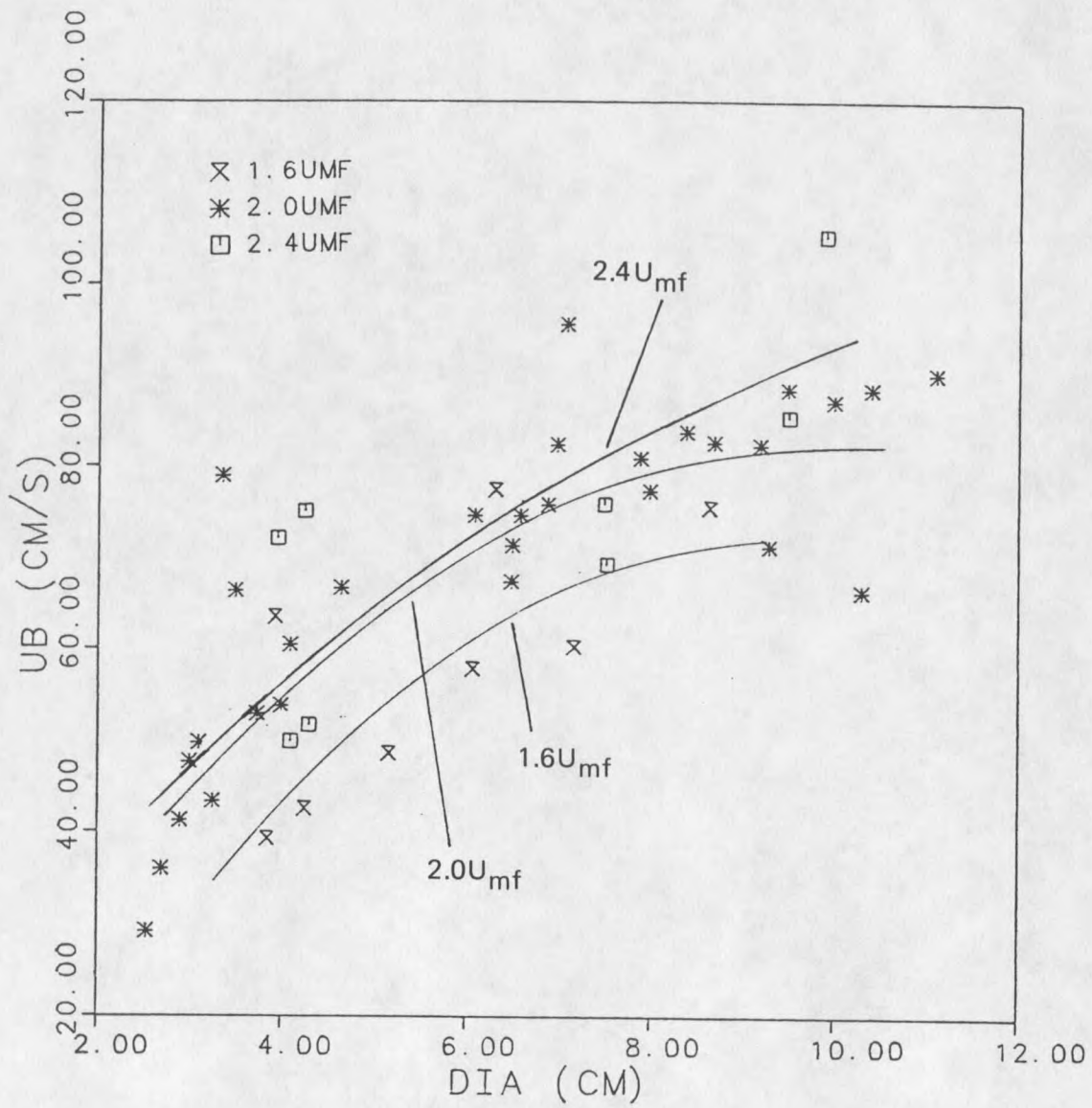


Figure 15. Comparison of experimental results for $1.6U_{mf}$, $2.0U_{mf}$, and $2.4U_{mf}$. Data taken at 15.2, 26.2, and 31.2 cm bed height.

closely to the flow profile as shown in Figure 12. Therefore the results can be explained. The close proximity of the $2.4U_{mf}$ curve to the $2.0U_{mf}$ curve probably is due to insufficient amounts of good data at a velocity of $2.4U_{mf}$.

Slug velocity was calculated to be 73.3, 75.5, and 77.5 cm/sec for $1.6U_{mf}$, $2.0U_{mf}$, and $2.4U_{mf}$, respectively. Experimental bubble velocities were found to exceed the slug velocities predicted from

$$U_s = k_1 (U_o - U_{mf}) + 0.35(gD)^{1/2} \quad (12)$$

from Ormiston et al. [28]. Bubble velocity can surpass the slug velocity in the lower regions of the bed due to coalescence taking place in this region. Work by Kehoe and Davidson [29] showed that U_s is considerably larger than predicted by Equation 12, with k_1 equal to 1, within a meter of the distributor. All bubble velocity measurements in this work were done at heights 31.2 cm and below, so bubble velocities greater than slug velocity were expected.

SUMMARY

1. The four-probe assembly developed in the course of this research was more successful than the three-probe assembly suggested by Dutta and Wen.
2. The four-probe assembly can apparently be used to determine local bubble velocities in a fluidized bed. This method compares favorably to literature methods previously used, and may offer advantages in determining individual bubble velocities.
3. Bubble volume as well as an approximation of bubble shape can be determined from the probe assembly response.
4. Experimentally determined results can be correlated by use of an equation incorporating a distribution coefficient which compares local bubble flow to the cross-sectional average.
5. Bubble volume fraction cannot be successfully determined utilizing the optical probes.

RECOMMENDATIONS FOR FUTURE RESEARCH

1. With the two-dimensional bed and cine photography and an optical probe mounted in the two-dimensional bed, comparison of observed bubbles and bubbles detected by the probes could be accomplished.
2. Modification of the probe assembly so radial movement of the entire assembly could be accomplished is suggested, then the radial effects on bubble velocity could be studied.
3. The utilization of the four-probe assembly along with the two-dimensional bed could be used to determine the effects of differing materials on bubble velocity.

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APPENDICES

APPENDIX A

BUBBLE VOLUME CALCULATIONS

BUBBLE VOLUME

Volume of a Spherical Cap Bubble

The volume of a spherical bubble can be determined from the probe responses as shown in Figure 16. Dimension L_c is given by probe A contact and dimensions X_p and L_D or given by probe B and C contact.

Thus,

$$r_c = \frac{X_p^2 + L_D^2}{2 L_D} \quad (A1)$$

For $r_c < L_c$,

$$b'_c = \frac{1}{2} (2 r_c - L_c) \quad (A2)$$

and

$$a'_c = [4 r_c b'_c - (2 b'_c)^2]^{1/2} \quad (A3)$$

then,

$$V = \frac{4}{3} \pi r_c^3 - \frac{\pi}{3} b'_c [3(a'_c)^2 + 4(b'_c)^2] \quad (A4)$$

For $r_c > L_c$,

$$a_c = [2 r_c L_c - L_c^2]^{1/2} \quad (A5)$$

$$b_c = L_c/2 \quad (A6)$$

then,

$$V = \frac{\pi}{3} b_c [3 a_c^2 + 4 b_c^2] \quad (A7)$$

the equivalent spherical diameter is then given by,

$$d_c = \left(\frac{6V}{\pi} \right)^{1/3} \quad (A8)$$

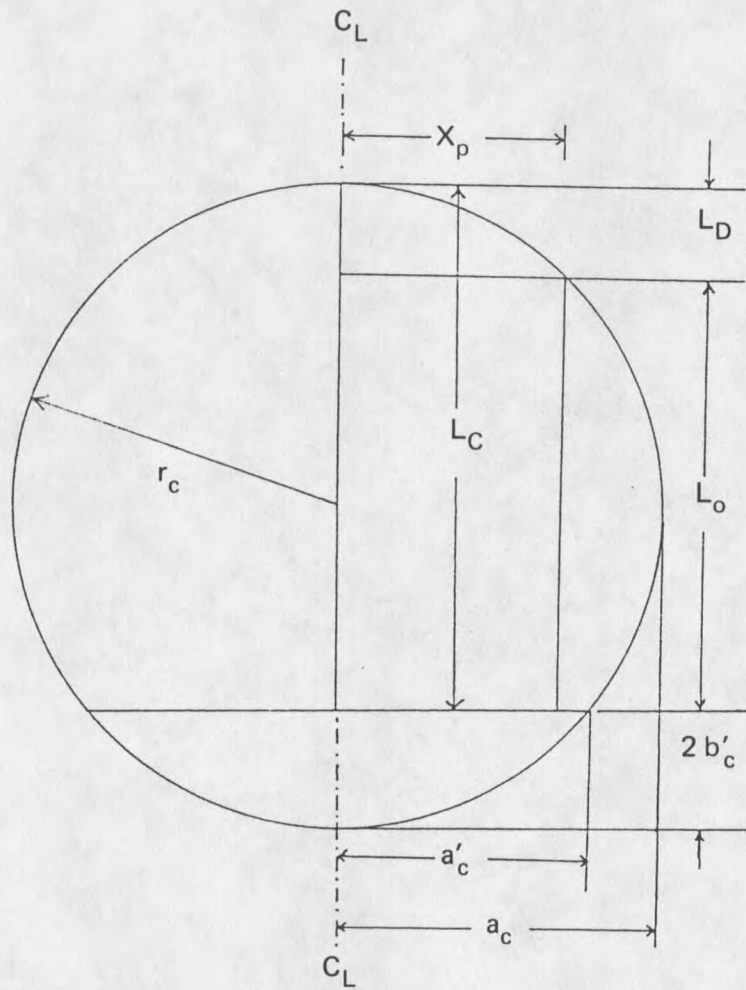


Figure 16. Spherical cap bubble.

Numerical Volume Calculation

If the bubble contacting the probe assembly is of some shape other than spherical cap, the volume can still be estimated. The approximate shape of the bubble is given by responses of probes A, B, and C as shown in Figure 17.

Irregular volumes are calculated by the following. Let $A_0, A_1 \dots A_n$ be the cross-sectional areas at equally spaced parallel chords and h be their distance apart. The volume can then be approximated by,

$$V_s = (b/3)[(A_0 + A_n) + 4(A_1 + A_3 + A_5 + \dots + A_{n-1}) + 2(A_2 + A_4 + A_6 + \dots + A_{n-2})] \quad (A9)$$

where n is even. This is Simpson's rule for numerical integration.

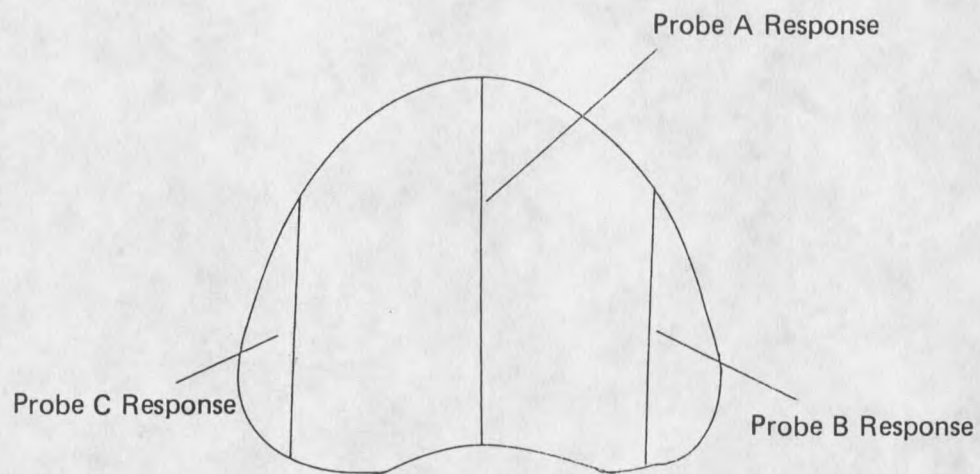


Figure 17. Irregular bubble shape.

APPENDIX B
ANALYSIS OF SAND

	UNITY	AREA
Total :	217.000000	33.253820
Mean :	1.000000	.153243
Std. Dev. :	0	.154230
Samples :	217.000000	
Undersize :	0	
Oversize :	0	

AREA	UNITY	
	0	undersize
0	185.000000	*****
.240000	22.000000	*****
.480000	7.000000	***
.720000	2.000000	
.960000	0	
1.200000	1.000000	
1.440000	0	
1.680000	0	
1.920000	0	
2.160000	0	
2.400000	0	
2.640000	0	
2.880000	0	
3.120001	0	
3.360000	0	
3.600000	0	
3.840001	0	
4.080000	0	
4.320000	0	
4.560000	0	
4.799999	0	
5.039999	0	
5.279999	0	
5.519999	0	
5.759999	0	
6.000000	0	oversize

Figure 18. Linear distribution of unity by area, units of area in millimeters.

	<u>UNITY</u>		<u>AREA</u>
Total	:	217.000000	33.253820
Mean	:	1.000000	.153243
Std. Dev.	:	0	.154230
Samples	:	217.000000	
Undersize	:	0	
Oversize	:	0	

AREA	UNITY	
	0	undersize
1.000000	2.000000	***
1.080946	0	
1.168444	0	
1.263024	0	
1.365261	0	
1.475773	1.000000	*
1.595231	0	
1.724359	0	
1.863938	0	
2.014817	1.000000	*
2.177907	0	
2.354200	0	
2.544763	0	
2.750751	5.000000	*****
2.973412	15.000000	*****
3.214098	38.000000	*****
3.474266	53.000000	*****
3.755493	49.000000	*****
4.059484	32.000000	*****
4.388084	13.000000	*****
4.743280	6.000000	*****
5.127229	2.000000	***
5.542257	0	
5.990880	0	
6.475817	0	
7.000007	0	oversize

Figure 19. Log distribution of unity by area, units of area in millimeters.



Figure 20. Linear distribution of unity by length, units of length in millimeters.

	UNITY	CONVEX PERIM.
Total :	217.000000	307.495000
Mean :	1.000000	1.417028
Std. Dev. :	0	.641618
Samples :	216.000000	
Undersize :	0	
Oversize :	0	

CONVEX PERIM.	UNITY	
0	4.000000	undersize
.280000	0	*****
.560000	23.000000	*****
.840000	53.000000	*****
1.120000	41.000000	*****
1.400000	39.000000	*****
1.680000	27.000000	*****
1.960000	12.000000	*****
2.240000	5.000000	*****
2.520000	4.000000	*****
2.800000	3.000001	*****
3.080001	3.000001	*****
3.360000	0	
3.640001	2.000000	***
3.920001	0	
4.200000	1.000000	*
4.480000	0	
4.760000	0	
5.040001	0	
5.320002	0	
5.600002	0	
5.880002	0	
6.160002	0	
6.440002	0	
6.720002	0	
7.000000	0	oversize

Figure 21. Linear distribution of unity by convex perimeter, units in millimeters.

	<u>UNITY</u>	<u>PERIMETER</u>
Total :	217.000000	329.135000
Mean :	1.000000	1.516751
Std. Dev. :	0	.704096
Samples :	217.000000	
Undersize :	0	
Oversize :	0	

PERIMETER	UNITY	
	0	undersize
0	4.000000	*****
.320000	3.000001	*****
.640000	35.000000	*****
.960000	50.000000	*****
1.280000	38.000000	*****
1.600000	42.000000	*****
1.920000	18.000000	*****
2.240000	13.000000	*****
2.560000	4.000000	*****
2.880000	1.000000	*
3.200000	5.000000	*****
3.520000	0	
3.840000	3.000001	*****
4.160000	0	
4.480000	1.000000	*
4.800001	0	
5.120000	0	
5.440002	0	
5.760002	0	
6.080002	0	
6.400002	0	
6.720002	0	
7.040002	0	
7.360002	0	
7.680002	0	
8.000000	0	oversize

Figure 22. Linear distribution of unity by perimeter, units in millimeters.

	<u>UNITY</u>	<u>BREADTH</u>
Total :	217.000000	78.640000
Mean :	1.000000	.362396
Std. Dev. :	0	.158581
Samples :	217.000000	
Undersize :	0	
Oversize :	0	

BREADTH	UNITY	
	0	undersize
0	4.000000	*****
.120000	43.000000	*****
.240000	71.000000	*****
.360000	58.000000	*****
.480000	26.000000	*****
.600000	10.000000	*****
.720000	2.000000	**
.840000	2.000000	**
.960000	1.000000	*
1.080000	0	
1.200000	0	
1.320000	0	
1.440000	0	
1.560000	0	
1.680000	0	
1.800000	0	
1.920000	0	
2.040000	0	
2.160000	0	
2.280000	0	
2.400000	0	
2.520000	0	
2.640000	0	
2.760000	0	
2.879999	0	
3.000001	0	oversize

Figure 23. Linear distribution of unity by breadth.

	UNITY	ROUNDNESS
Total	: 216.000000	3.079e305
Mean	: 1.000000	1.426e+03
Std. Dev.	: 0	193.016100
Samples	: 216.000000	
Undersize	: 0	
Oversize	: 1.000000	

ROUNDNESS	UNITY	
	0	undersize
1.080e+03	0	
1.080e+03	0	
1.160e+03	13.000000	*****
1.240e+03	47.000000	*****
1.320e+03	64.000000	*****
1.400e+03	37.000000	*****
1.480e+03	25.000000	*****
1.560e+03	16.000000	*****
1.640e+03	3.000001	****
1.720e+03	2.000000	**
1.800e+03	3.000001	****
1.880e+03	2.000000	**
1.960e+03	1.000000	*
2.040e+03	1.000000	*
2.120e+03	0	
2.200e+03	0	
2.280e+03	0	
2.360e+03	1.000000	*
2.440e+03	0	
2.520e+03	0	
2.600e+03	0	
2.680e+03	0	
2.760e+03	0	
2.840e+03	0	
2.920e+03	1.000000	*
3.000e+03	1.000000	oversize

Figure 24. Linear distribution of unity by roundness.

APPENDIX C

COMPUTER ALGORITHMS

THIS PROGRAM IS FOR DATA ACQUISITION IN A FLUIDIZED BED
WITH THE FOUR PROBE ASSEMBLY

```

20  DIM AR(4501)
30  I = 0:N = 0
31  & ARRAYCLR,AR
32  PRINT "NOW TAKING DATA"
50  & OTRIGIN,(AV) = AR,(RT) = 1
51  HOME
52  PRINT "ANALYZING DATA"
61  FOR I = 0 to 4500
63  IF AR(I) = 8 OR 0 THEN GOTO
    80
65  NEXT I
66  IF I >= 4500 THEN GOTO 30
80  FOR N = (I - 60) TO I
81  IF N >= 4500 THEN GOTO 30
82  IF N < 1 THEN N = 1
83  IF AR(N) = 14 THEN GOTO 86
84  NEXT N
85  GOTO 65
86  PR# 1
87  FOR N = (I - 60) TO (I + 200)

88  IF N >= 4500 THEN GOTO 30
89  IF N < 1 THEN N = 1
90  PRINT AR(N); SPC( 1);
91  NEXT N
92  GOSUB 500
93  PRINT : PRINT : PRINT
94  HOME
95  PR# 0
96  I = I + 200
100  IF I >= 4500 THEN GOTO 30

110  GOTO 65
120  END
500  REM SUBROUTINE TO DETERMINE

501  REM NEIGHBORING BUBBLE LOCATIONS
502  PRINT
505  IF I < 500 THEN RETURN
506  C1 = 0:C2 = 0:C3 = 0
507  T1 = 0:T2 = 0
510  CNT = 0
520  FOR K = (I - 500) TO I
530  CNT = CNT + 1
535  IF C1 < K 0 THEN GOTO 550
540  IF AR(K) < K 15 THEN C1 = CNT

```

```
550     IF AR(K) = 14 THEN C2 = CNT:
        GOTO 1000
560     NEXT K
570     PRINT : PRINT "NO PRECEDING BUBBLE"
580     FOR K = (I + 200) TO (I + 500)
585     IF K > = 4501 THEN RETURN

590     CNT = CNT + 1
600     IF AR(K) < > 15 THEN C3 = CNT:
        GOTO 1050
610     NEXT K
620     PRINT : PRINT "NO TRAILING BUBBLE"
630     RETURN
1000    T1 = C2 - C1
1010    PRINT : PRINT "PRECEDING BUBBLE";
        T1;" MSEC"
1020    GOTO 580
1050    T2 = (C3 -C2) + 200
1060    PRINT : PRINT "TRAILING BUBBLE
        ";T2;" MSEC"
1070    RETURN
```

THIS PROGRAM OBTAINS DATA FOR THE CALCULATION
OF BUBBLE VOLUME FRACTION

```
14   K = 1
15   COUNT = 0
20   DIM AR(4501)
21   N = 0
22   COUNT = COUNT + 1
30   & ARRAYCLR,AR
40   & OTRIGIN, (AV) = AR,(RT) = 5
50   FOR I = 0 to 4500
60   IF AR(K) = K GOTO 80
70   N = N + 1
80   NEXT I
90   VOIDFR = N / 4501
91   VS = VS + VOIDFR
92   VA = VS / COUNT
95   PR# 1
96   PRINT : PRINT : PRINT
100  PRINT "THE VOID VOLUME IS ";
      VOIDFR
105  PRINT "THE AVERAGE VOID VOLUME
      IS ";VA
106  PRINT "ANALYZED FOR TOTAL TIME
      ";COUNT * 20.58;" SECS"
107  PR# 0
108  HOME
110  GOTO 21
120  END
```

APPENDIX D

BUBBLE VOLUME FRACTION DATA

Table 5. Bubble Volume Fraction Data.

Bed Height (cm)	15.2		26.2		31.2	
	Radial Position (cm)	δ_b	Radial Position (cm)	δ_b	Radial Position (cm)	δ_b
	0	0.0104	0	0.0102	0	0.0150
	3	0.0069	2.2	0.0113	3	0.0165
	5	0.0039	4.2	0.0112	5	0.0195
	7	0.0056	6.2	0.0134	7	0.0226
	9	0.0062	8.2	0.0146	9	0.0037
	11	0.0075	10.2	0.0148	11	0.0021
	13	0.0114	12.2	0.0083	13	0.0012
	15	0.0100	14.2	.0036	15	0.0014
			16.2	0.0015		

APPENDIX E
BUBBLE VELOCITY DATA

Table 6. Bubble Velocity Data. Data at $2U_{mf}$.

Equivalent Bubble Diameter (cm)	Bubble Velocity (cm/sec)	Height in Bed of Measurement (cm)
2.5	29.2	15.2
2.7	36.0	15.2
2.9	41.3	15.2
3.0	47.7	15.2
3.1	49.8	26.2
3.3	43.4	15.2
3.4	79.0	15.2
3.5	66.4	15.2
3.7	52.9	15.2
4.0	53.8	15.2
4.1	60.5	15.2
4.7	66.7	15.2
6.1	74.7	26.2
6.5	67.4	26.2
6.5	71.4	26.2
6.6	74.7	26.2
6.9	75.9	26.2
7.0	82.5	26.2
7.1	95.7	26.2
7.9	81.0	26.2
8.0	77.4	26.2
8.4	83.9	26.2
8.7	82.8	31.2
9.2	82.5	26.2
9.3	71.4	26.2
9.5	88.6	31.2
10.0	87.3	31.2
10.3	66.4	26.2
10.4	88.6	26.2
11.1	90.3	31.2

APPENDIX F

PROBE ELECTRONICS

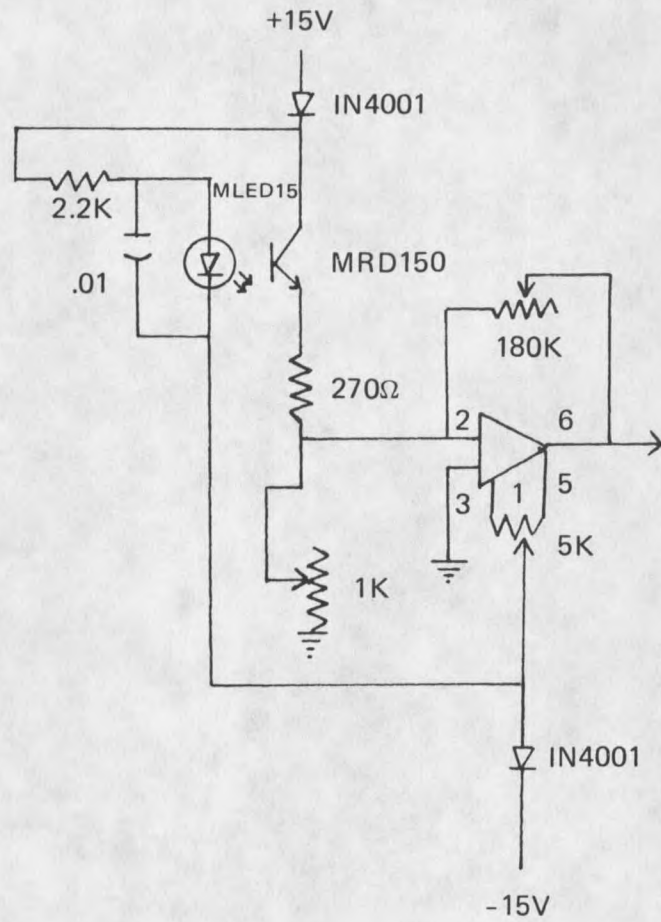


Figure 25. Electronic circuit for probe.

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