ABSTRACT

The major objective of this project is to understand the phenomenon of erosion of tube surfaces in fluidized bed combustors and its dependence on various operating and system parameters and hence on the nature of fluidization. It is, therefore, essential to quantify the quality of fluidization. To accomplish this, experiments were conducted in a variable thickness two-dimensional bed with the same charge as are planned to be used for conducting erosion studies. The series of experiments comprising of the three sand particles and for four bed thickness values is completed. These results are combined with similar experiments on a fixed width two-dimensional bed to develop a proper interpretation of data. The results are written in a form of research paper which is being published in Powder Technology.

The hydrodynamic features in a two-dimensional bed is related to bubble dynamics and resulting solids elutriation from the bed surface. Bubble velocities are measured in beds of three different silica sand particles. It is found that for the bubbles of the same size, the bubble rise velocity in a bed of larger particles will be less than in a bed of smaller particles of the same material and sphericity. Some additional experiments are planned in variable width bed and thereafter all these results will be written for publication. The film records are also analysed to obtain the mechanistic picture of the bubble burst and solids projection. A model for the initial solids projection rate is developed and its predictions are compared with the experimentally measured initial solids projection rates. A paper describing these results is communicated to a journal for publication.

The solids concentration around the periphery of a horizontal tube is successfully measured by a fiber optic probe. Tubes of 11, 28 and 50.8 mm diameter immersed in an air-fluidized bed of 1.4 mm average diameter glass beads are used and measurements are taken as a function of air fluidizing velocity. Two papers describing our results have been published, and a third paper will be presented at the Am. Inst. Chem. Engrs. meeting at New York.
The tube to bed heat transfer is measured for the 28.6 and 50.8 mm diameter horizontal tubes immersed in the bed as a function of fluidizing velocity.

A review of papers on erosion is under progress and limited summaries and assessments have been prepared. More detailed analysis and experimentation will be conducted during the second year of the program.

EXECUTIVE SUMMARY

Fluidized-bed combustion has emerged as a very successful technology for the combustion of coal in an efficient and environmentally acceptable mode. However one of the serious problems which has been encountered by the recent investigations all over the world is the erosion of metal of boiler tubes, employed to remove the heat of combustion from fluidized beds. Though the observed metal damage and degree of erosion has differed widely amongst the different investigations, all agree that the tube bundle erosion is the stumbling block in the practical realization of the technology. The work under this contract proposes to undertake a number of investigations which would help to understand the erosion phenomenon and its dependence on various operating and system parameters and hence on the nature of fluidization.

To accomplish this broad goal, we have at our disposal four fluidized beds: three operating at ambient temperature and pressure, and the fourth at high temperatures up to about 1300 K and with some modifications for pressures up to about four atmospheres. The first category includes a fixed thickness two-dimensional bed, a variable thickness two-dimensional bed and a 0.305 m square three-dimensional bed, and the second category includes a 0.153 m diameter cylindrical fluidized-bed. Further, to follow the bubble motion and solids movement a high speed cinephotography and automated film analysis technique has been developed. Some preliminary work has been done to develop and establish a fiber optic probe technique to determine the particle concentration on the surface of a simulated boiler tube in a fluidized-bed. It is planned to undertake a variety of investigations to help understand the complicated metal erosion phenomenon mentioned above during the first year of this contract and continue the same to completion in the subsequent years.

A variable thickness two-dimensional fluidized-bed was fabricated with unique facilities, to offer information on fluidization behavior of the bed with and without the presence of tubes of different diameter and different tube bundle configurations. The characteristic change with particle size and fluidizing velocity will enable us to understand the phenomenon of tube erosion in a systematic fashion leading to its quantitative and mechanistic description at the end.

We, therefore, performed experiments with narrow size range silica sand particles of average diameter 730, 1237 and 2356 μm. The minimum fluidization velocity was determined by measuring the bed pressure drop as a function of fluidizing velocity and employing the
well adopted method as detailed by Saxena and Vogel [1]. Such experiments were conducted for four different values of bed thickness which can be accurately obtained on this facility. These are 31.75, 44.45, 50.40 and 63.50 mm. The minimum fluidization velocity is also obtained for the same bed charges in a fixed thickness two-dimensional fluidized-bed [2] and in a three-dimensional fluidized-bed [3] after reducing the cross-section of the bed from 0.305 X 0.305 m to 0.152 X 0.152 m to enable us to conduct the experiments for the larger size particles also. Both these beds have been extensively used by our research group during the last few years in successfully conducting a variety of heat transfer, hydrodynamics and solids elutriation studies. Some of these are: Grewal and Saxena [4]; Coel, Saxena and Dolidovich [5]; Grewal, Cheung and Saxena [6]; Saxena, Mathur and Sharma [7]; Saxena and Mathur [8] and Varma and Saxena [9]. The average particle size was determined by analysis on a sonic sifter in conjunction with a sensitive analytical balance. It is found that the empty bed channel pressure drop is dependent on its thickness and decreases as the latter is increased for a given gas velocity. It was found that minimum fluidization velocity is dependent on the thickness of the bed and the wall effects are minimum for a bed of 73, 93 and 116 mm thicknesses for 730, 1237 and 2356 μm sand particles respectively. Our preliminary results are under publication and a detailed paper will be published describing the dependence of wall effect on particle size, fluidized-bed thickness and gas velocity in the near future. This will then establish the optimum dimensions of two-dimensional bed for detailed bed hydrodynamic studies and bubble dynamics. The latter is established by taking recourse to a high speed cine photographic technique, described below, which has been developed under this contract.

This series of experiments were conducted in a variable thickness two-dimensional bed of width 60.9 cm and thickness 6.35 cm, equipped with bubble cap distributor plate. The test section of the bed was provided with plexiglas window for visual observation and for photographing the bed activity. The bubble motion is photographed by a Nova high speed camera at a filming speed of 1024 pictures per second. The film records were analyzed on an automated film analysis system consisting of LW224A photo optical data analyzer and a LW10C photo-optical digitizer provided with a LW 1224 electronic graphic calculator interfaced to an IBM 3081 D computer through a teletype 43/Esprient II terminal. Beds of three different silica sand particles were analyzed and it was concluded that the bubble shape and growth depends upon the particle size, and the bubble velocity is correlated in terms of the bubble frontal area. Experimental data indicated an increase in bed viscosity with an increase in bed particle diameter. It therefore follows that for the same size of bubble, the bubble rise velocity in a bed of larger particles will be less than in a bed of smaller particles of the same material and sphericity. Our investigations have covered a much larger range of particle sizes than any of the previous investigations.

It is important to note the amount of solids projected from the surface of the fluidized beds and finally entrained in the gas stream
exiting from the combustor. These solids establish the tube metal erosion of the freeboard cooling coils as well as of the turbine blades in those cases where the hot combustion gas are expanded to drive a gas turbine for power generation. Rising bubbles in gas fluidized beds are found to create a bulge at the bed surface, which grows in size as the bubble approaches the surface. The solids contained in the bulge above the bubble are projected into the freeboard at the bubble burst, and this constitutes the upper limit of solids elutriation from the bed. Experiments were carried out in variable thickness two-dimensional bed with three different particles and the corresponding bed condition is recorded cinematically at a filming rate of 1024 frames per second. The film records have been analyzed to obtain the mechanistic picture of the burst and solids projection from the bulge of a single bubble. A model for the initial solids projection rate is developed and its predictions are compared with the experimentally measured initial solids projection rates from beds of three particles. In the theoretical model, the concepts of solids drainage and momentum of the wake solids are incorporated for the first time and this comprehensive model based predictions are in good agreement with the experimental values. A detailed experimental effort is completed and results have been communicated for publication.

The solids concentration around the periphery of a horizontal tube was successfully measured by a fiber optic probe. An image carrying fiber optic probe with its associated light source in conjunction with a close-circuit television camera, TV monitor and a video-cassette recorder has been used to measure the solids concentration profile at four angular positions around the periphery of horizontal tubes of 11 and 28 mm immersed in an air-fluidized bed of 1.4 mm average diameter glass beads as a function of fluidizing velocity. The light from the light source is transmitted from the proximal to the distal end through 30 µm diameter outer glass fibers and illuminate the test surface, 4 mm by 5.5 mm. The images of the illuminated particles on the glass tube surface are carried by the central glass fibers to the proximal end where these are magnified either for direct viewing or on a TV monitor through CCTV camera system. The images may also be recorded on a video-cassette recorder for detailed frame by frame analysis. In all cases it is found that the bulk bed voidage is considerably smaller than the voidage values at the tube surface. The latter varies with the location around the periphery and for the same position with tube diameter and air fluidizing velocity. In all cases, the smallest values are found for the downstream side of the tube, largest values for the equatorial sides and intermediate values for the upstream side. The variation of these values are discussed with fluidizing velocity and tube diameter. Estimates of the air film thickness at various locations between the tube and the particles are also obtained. These values are very useful in developing the mechanistic models for tube erosion and heat transfer by particle motion. Two papers describing our results have been published and a third will be presented in the Am. Inst. Chem. Engrs. meeting at New York.
Knowledge of heat transfer coefficient is very important for an adequate design of fluidized-bed reactors from which thermal energy must be removed to keep the temperature constant. The difficulty in establishing a reliable value for heat transfer coefficient stems from the fact that it depends on a large number of system and operating parameters and its interaction. The influence of the design and operating parameters on heat transfer coefficient for an immersed, horizontal, smooth heat transfer tube in a gas fluidized bed has not been resolved update because of its complicated hydrodynamics. This particular investigation is devoted to understand the variation of heat transfer coefficient for a horizontal tube as its diameter and bed particle sizes are varied. The tube to bed heat transfer is measured for the 28.6 and 50.8 mm diameter horizontal tubes immersed in the bed for the optimum conditions established by our earlier works. These datas will provide excellent base for testing the available correlations and mechanistic theories of heat transfer.

A review of papers on erosion is under progress and a brief summary is reported in two quarterly reports. A complete review of erosion literature will be developed during the second year of this contract work.

OBJECTIVES

The major over-all objective of the project is to generate information which may enable to understand the erosion of tube bundles in gas fluidized beds. To accomplish this following specific tasks have been planned for the current year.

Task 1: An image carrying fiber optic probe technique will be established to measure the particle concentration profiles at various angular positions at the tube surface of diameters about 11.28 and 50.8 mm as a function of fluidizing velocity in the 0.305 m square fluidized-bed facility.

Task 2: Experiments will be conducted in a variable thickness two-dimensional bed to establish the dependence of wall effects on bed thickness and particle size. The experimental results will be analyzed to establish the point of transition from two-dimensional behavior to three-dimensional for quantitatively characterized quality of bed fluidization.

Task 3: Experiments will be conducted in a two-dimensional bed in conjunction with cinemophotography to determine bubble velocity and solids elutriation from the bed surface for various specified operating conditions. This will enable characterization of the quality of fluidization of experiments conducted in Tasks 1 and 2, and later work of tube erosion.

Task 4: For the same bed charge, tubes and operating conditions heat
transfer coefficients will be measured between the tube and the 0.305 m square fluidized bed.

Task 5: The available literature on tube erosion will be analyzed and assessed to develop a state-of-the-art review. Experiments will be planned to investigate erosion loss in the 0.305 m square bed for bed charge and experimental conditions employed in Tasks 1 to 4.

INTRODUCTION AND BACKGROUND

Illinois high sulfur coal can be efficiently and economically combusted and desulfurized by adopting the fluidized-bed combustion technology. However, the worldwide exploitation of this technology has experienced a difficulty in successfully accomplishing this due to the severe and unexpected metal damage and erosion of the boiler tubes employed to remove the heat of combustion in the bed. Much effort in recent years is being directed to understand the phenomenon. Our program of research is also oriented in this direction. We are adequately equipped to initiate such a research program and develop adequate techniques to generate data which will enable to understand the causes which result in the metal loss and the mechanisms which govern it for different fluidizing conditions. We have planned a set of experimental tasks for the first year of the program and these are outlined in previous sections. Some experimental work has been completed and this is being reported in the next two sections.

EXPERIMENTAL PROCEDURES

The bubble dynamics and solids projection from the bed surface studies are conducted in the variable thickness two-dimensional bed. The variable thickness fluidized bed is fabricated from plexiglas, which enables and facilitates visual observation of any section of the bed and for its photorecording. The fluidizing air is supplied by two 18.65 k.w., two cylinder, two-stage air cooled compressors connected in parallel and capable of supplying a maximum of 0.1 m/s at 375 kPa. The compressed air is dried to a dew point temperature of 283k by passing air from one compressor through a column of silica-gel and that from the second compressor through a freon cooling unit. The compressed and dried air is passed through two filters for the removal of oil or any other vapors. The air flow rate is measured by two calibrated rotameters. A thermocouple is installed in the air supply line just before the fluidized bed to measure the air temperature. It is connected to a Leeds and Northrup 914 Numatron temperature recorder which has a measuring range of 235-1116 k with a resolution of 0.1k. The air is transported to the bottom of the bed through a 6.35 mm schedule 40 pipe and enters into the calming section of the bed via bubble caps on the distributor plate. The air leaving the bed column passes through a cyclone and a fabric filter before releasing into the atmosphere via an exhaust duct system.

The bed activity is photographed by a Nova high-speed camera, capable of filming 10 to 10,000 frames per second in the 16 mm format. The bed is illuminated by two mounted spot lights, each having eight 500W tungsten bulbs. A light meter with a 1° angle of incidence is focused on the bubbles so as to obtain the appropriate
film speed-lens opening combination that would maximize the contrast between the bubbles and the emulsion in the bed.

The solids concentration profile around the periphery of horizontal tubes using fiber optic probe is conducted in the three dimensional 0.305 m square fluidized bed. The fluidized-bed facility comprising of several independent units is employed in this work and its schematic is shown in Figure 1. These units are: air supply system, fluidized-bed system and an off-gas system which removes the entrained dust particles in the gas. A brief description of these units are given in the following. The fluidizing air is supplied by two, two-cylinder, two-stage air-cooled reciprocating compressors connected in parallel. The compressed air is passed from one compressor through a column of silica gel and that from the second compressor through a freon cooling unit and thereby its humidity level is brought to a dew point temperature of 283 K. The compressed and dried air is passed thereafter through two filters to remove oil vapors. A muffler is installed on the bypass line to reduce the noise level. The air flow is metered on calibrated rotameters with a precision of one percent.

The fluidization column is of 0.305 m square cross-section and consists of a plenum chamber, a bubble-cap distributor plate, test and freeboard sections. The free board section is 1.51 m height and has seven solids sampling and seven pressure probe ports. The middle test section is 0.61 m tall and is provided with three temperature and nine pressure probes. A front plexiglas window permits the visual observation of the quality of fluidization in the test section. The bottom section, wind box or plenum chamber, is about 0.48 m in height and here the fluidization air rearranges itself for a uniform flow through the bed. The air first flows through an air jet breaker plate in the plenum chamber and then through the fluidized bed distributor plate. Both these plates employ flat top bubble caps screwed into a 0.318 cm thick steel plate with 0.635 cm diameter holes on 2.8 cm center-to-center spacing.

The off-gas system comprises of primary and secondary cyclones and a fabric filter, and is quite similar to that used in our earlier works, Grewal and Saxena [4]. The cyclones have a 100 percent particle removal efficiency for particles greater than 25 μm and a flow capacity of 60 l/s. Particles collected in the cyclones fall in glass jars which are connected at the bottom end.

The pressure probes are connected to glass manometers with flexible tygon tubings. The manometers are filled with appropriate fluids of two different densities. The air temperature is measured by a copper-constantan thermocouple in conjunction with a direct reading digital indicator. For measuring the bed temperature, iron-constantan thermocouples are provided and an Omega trendicator is used as a part of the data acquisition system to establish temperatures directly in degrees with a resolution of 0.1 degree.

Two plexiglas side plates are used to mount a 11 mm and a 28 mm outside diameter glass tubes of 0.5 mm and 1.3 mm wall thickness respectively. The image carrying fiber optic probe supplied by A. O. Scientific Instruments and referred to as borescope is inserted in these tubes and its details are given below. Spherical glass beads of a narrow size range, predominantly 1.2 to 1.7 mm, with an average
Figure 1. The UIC 0.305 m square fluidized-bed facility. (1) Wayne air compressor, (2) Curtis air compressor, (3) dessicant dryer, (4) surge tank (5) refrigerator dryer, (6) oilscr filter, (7) carbon filter, (8) pressure regulator, (9) bypass exhaust, (10) rotameters, (11) predistributor plate, (12) distributor plate, (13) pressure probes, (14) manometers, (15) cyclones, (16) drum, (17) optic probe, (18) light source, (19) T.V. camera, (20) monitor, and (21) video recorder.
diameter of 1.4 mm are used as bed material. The equipment employed to establish the particle size distribution consists of an analytical balance, riffle sampler and a sonic sifter with a complete set of sieves.

The bed voidage, ε, is computed by measuring the bed pressure drop, ΔP, across two points which are separated by H from the following relation:

\[ 1 - \varepsilon = \frac{\Delta P}{H (\rho_s - \rho_f)} \]

\( \rho_s \) and \( \rho_f \) are the densities of the solid particles and air respectively. The glass tubes are mounted at a distance of 28 cm above the distributor plate, while the pressure probes are located 15 and 25 cm above the distributor plate. The measured \( \rho_s \) value is 2.49 g/cm\(^3\) and that of air is taken as 0.001 g/cm\(^3\).

The schematic of the borescope is shown in Figure 2, and it consists of an objective lens (a), a transmitting part (b), and an eye piece assembly (c). The transmitting part comprises of an inner image guide (b\(_1\)) prepared by bundling together parallel glass fibers of 30 μm diameter, and an outer light guide (b\(_2\)) made from similar fibers to illuminate the object to be viewed. The light guide is enclosed in a 8 mm stainless steel tube. The right end of the transmitting part is referred to as the proximal end and here the light carrying fibers communicate to the external light source (e) through a light guide input adaptor (d). The left end of the transmitting part is called as the distal end and here the objective lens (a) with a field of view of 45° is attached to the image guide. At the proximal end of the image guide is the eye piece assembly (c) comprising of the relay lens (c\(_1\)) and the slide focus control (c\(_2\)). The overall length of the borescope is 30 cm and has a depth of focus varying from 1 mm to infinity. The forward viewing is conveniently converted for right angle observation by slipping the stainless steel tube (f) with right angle mirror (f\(_1\)) over the transmitting part (b). 360° rotation of the tube (f) containing the right angle mirror (f\(_1\)) allows quick and effortless viewing around the circumference of the glass tube. The external light source (e) with variable light intensity capability (50, 100, 150W) is connected to the borescope by flexible light guide.

The borescope is inserted inside the glass tube fixed horizontally in the fluidized bed. The light from the light source is transmitted from the proximal to the distal end and illuminates the surface after being reflected from the mirror. The images of the illuminated particles on the glass tube surface and of the illuminated light guide ring are focussed by the objective lens (a) on the input face of the image guide (b\(_1\)). The latter transports it to the proximal end where the relay lens (c\(_1\)) in the eye piece assembly (c) magnifies it for either direct viewing or on a TV monitor (h) through a closed circuit TV camera (g). The projected image is also recorded on the video cassette (i) for detailed analysis. (g) is a black and white TV camera with C-mount lens (g\(_1\)), having f/1.8. The vidicon in the camera converts optical images into synchronized electrical signals to be fed to the TV monitor (h) with resolution power of 500
lines. Here the cathode-ray tube converts the synchronized electrical signal into the optical image at the rate of 30 fps. An industrial Panasonic video cassette recorder (i) records this synchronized signal on a high density Scotch color-plus video cassette. There are four video heads in this VCR, two are used for normal play back and the other two are used for slow and still mode operation.

The borescope is used to view and record the surface particle population around the glass tube surface at various angular positions. For a 11 mm tube, the four angular positions employed are 0°, 90°, 180° and 270° as shown in Figure 9. On the other hand for a 28 mm tube these positions as shown in Figure 10 are 0°, 60°, 180° and 300°. The recordings are taped for the fixed bed and for the fluidized bed at various fluidizing air velocities in the range up to 1.1 $U_{mf}$. Here, $U_{mf}$ is the minimum fluidization velocity.

The heat transfer experiments are conducted in a 152 x 152 mm square fluidized bed. The heat transfer tubes are of aluminum tubes of diameter 28.6 and 50.8 mm. The tubes are machine finished, smooth surfaces and are energized by Watlow 120V, 1000W Fire rod calrod heaters. Teflon fittings and O-rings at the end of the tubes reduces gas leakage and heat loss from the ends of the tube to less than one percent. The design details of the heat transfer tube is given in Figure 3. The tubes have four copper-constantan thermocouples of length 3.7, 6.7, 9.8, and 12.8 cm at 90° separation. The thermocouples are cemented with copper cement in milled grooves along the surface of the probe enabling the Omega Trendicator to determine the surface temperature of the tube. The latter display the temperature digitally with a resolution of 0.1k. Power to the heater is supplied by a HP 6274 B DC power supply rated for 0-60 volts, 0-15 amhers with a load regulation for constant current of less than 0.02% of output plus 500A for a load change equal to the voltage rating of the supply. A voltmetre and ammeter with an accuracy of 0.25 volts and 0.5 amphere respectively are used for measurement. Higher gas mass fluidizing velocities are achieved by two 18.5 k.w. reciprocating compressors, capable of delivering a maximum mass velocity of 1.2 kg/m$^2$. The air is adequately filtered and dried prior to its entrance to the bed. Silica sand of average diameter 730, 1237 and 2356 μm are used as bed material. The mass fluidizing velocity is varied over a wide range, with a maximum value of 1.12 kg/m$^2$.s.

RESULTS AND DISCUSSION

Rising bubbles in gas-fluidized beds are found to create a bulge at the bed surface, which grows in size as the bubble are projected into the freeboard at the bubble burst, and this constitutes the upper limit of solids elutriation from the bed. Experiments have been carried out in a variable thickness two-dimensional beds of three sand particles of average diameter 730, 1237 and 2356 μm. The corresponding bed condition is recorded cinemographically at a filling rate of 1024 frames per second. The film records have been analyzed to obtain the mechanistic picture of the burst and solids projection from the bulge of a single bubble. The film image from the
Figure 2. A detailed schematic of the borescope and associated measuring equipment. (a) objective lens, (b) transmitting part, (b₁) image guide, (b₂) light guide, (c) eyepiece assembly, (c₁) relay lens, (c₂) slide focus control, (d) light guide input adaptor, (e) light source, (f) stainless steel tube, (f₁) right angle mirror, (g) T.V. camera, (g₁) C-mount lens, (h) T.V. monitor, (i) industrial VCR.

Figure 3. Design details of heat transfer tube assembly: (1) Aluminum tube, (2) O-ring, (3) calrod heater, (4) Teflon cap. All dimensions are in millimeters.
The analyzer is projected onto the screen of the digitizer on which continuous measurement of length and area are possible through the use of a cursor connected to the electronic graphics calculator. The properties of the particles and their beds at minimum fluidization, as well as the operating conditions at which the bed activity is photographed are listed in Table I.

Two typical sequences of bubble rise and burst are shown in Figures 4 and 5 which illustrates the bulge-burst phenomenon. Figure 4 refers to the bed of silica sand \((d_p = 2356 \, \mu m)\), and shows bubble position at four instants from bubble birth to burst at the bed surface. In Figure 4A, the bubble is approximately at the bed center, and the bulge at the bed surface due to the presence of this bubble is clearly visible. The bubble nose is at the bed surface level in Figure 4B, and the large bulge volume above the bed surface can be seen. The bubble center is almost at the bed surface level in Figure 4C, and the bulge thinning-out is evident. Finally, bubble burst is seen in Figure 4D, and the bulge volume at this instant almost corresponds to the solids projected into the freeboard at bubble burst.

A similar trend of bulge growth and decay in the bed of glass beads \((d_p = 3050 \, \mu m)\) is seen in Figure 5. A bulge makes an appearance at the top of the bed almost simultaneously as the bubble appears at the distributor plate. The bulge becomes more pronounced as the bubble rises in the bed, and it reaches a maximum thickness when the bubble roof is at the bed surface, Figure 5C. Almost the entire bubble emerges out of the bed before bursting while its wake continues to retain its integrity in being a part of the bed surface, Figure 5D. All the solids projected into the freeboard are seen to originate from the bubble bulge. Since bubble bursts in beds of the other particles are found to exhibit the same mechanistic details. These observed details are now utilized in constructing a model for the initial solids projection rate.

Figure 6 (a–f) shows that the experimental \(E_0\) values are greater than those predicted by the modified theory of Chen and Saxena for beds of all the particles. The systematic underprediction is attributed to the lack of consideration of the contribution of transfer of momentum from the bubble and its wake to the bulge. Pemberton and Davidson [12] have shown that the forces exerted on the particles upstream of a rising bubble due to the wake momentum of the bubble causes a bulge to be formed having a weight of \(0.5 \, m_W\). Here the wake volume is considered to be one fourth of the total bubble volume. Consequently, the bulge volume is greater than that given by the model of Chen and Saxena, and the total bulge volume is considered, here to be the sum of the volumes due to the potential-flow propagation and the wake-momentum transfer.

As mentioned earlier it has been experimentally observed that the bulge increases in size during the rise of the bubble through the bed and attains its maximum value when the bubble nose is at the bed surface level. Subsequently, as the bubble rises further, and till it is completely out of the bed, the bulge solids drain back into the
Figure 4. Film sequence of bubble growth and burst in a fluidized bed of silica sand ($d_p = 1237 \mu m$).

Figure 5. Film sequence of bubble growth and burst in a fluidized bed of glass bead ($d_p = 3050 \mu m$).
Figure 6. Comparison of experimental and predicted values of $E_o$ for (a) 265 μm glass beads, (b) 1527 μm glass beads, (c) 3050 μm glass beads, (d) 734 μm sand, (e) 1237 μm sand, (f) 2356 μm sand.
Figure 7. The bulge decay process details.
bed. The proposed bulge growth model is therefore composed of two parts: the first corresponding to bulge growth while the bubble rises through the bed till its nose is at the bed surface level, and the second corresponding to bulge decay while the bubble rises through one bubble-diameter above the bed surface.

Consideration of the two forces affecting the bulge growth (bubble velocity propagation through the bed, and wake momentum being transferred across the bubble) shows that the contribution of wake momentum to bulge size will be constant during the course of bubble rise, while the contribution of bubble velocity changes with change in the distance of the bubble from the surface. Figure 7 illustrates the bulge decay process. Figure 7A corresponds to the instant when the bubble nose is at the bed surface, and the bulge is at its maximum growth and before drainage starts.

Following this instant, as the bubble rises above the bed surface, the bulge solids fall back relative to the bubble, and the bulge starts thinning out, Figure 7B, leading to bubble burst when finally the entire bubble has arisen out of the bed, Figure 7C. The bulge decay process completes in $2R_b/V_b$ seconds. This process is modelled by conceptualizing the bulge and bubble to form concentric circles during the entire bulge decay process, Figure 7D and 7E. The bulge area in Figure 7D is made equal to the bulge area in Figure 7A, and in the idealized model, the bulge apex-bubble center distance is assumed to decrease from $Z_0$ at $t = 0$ to $Z_f$ at $t = 2R_b/V_b$. $E_0$ values calculated using modified $V_{bul}$ values are shown by the curves marked 2 in Figure 6. These curves over predict the experimental data in all cases.

The bulge decay is modelled in a manner similar to that of Pemberton and Davidson [13]. $E_0$ values, thus obtained are shown in Figure 6 as the curves marked 3. These values on the average predict our two-dimensional bed data much better than curves 1 or 2. It is clear also that wake momentum-solids drainage model of Pemberton and Davidson [13] alone under predicts the data. Values corresponding to this model are given by difference of curves 3 and 1.

The proposed model overpredicts all the data for the smallest particles, glass beads ($d_p = 265 \mu m$). One possible reason could be that the greater inter-particle forces in smaller particles will tend to inhibit the drainage of solids from the bulge, leading to larger bulges than otherwise expected. Another conclusion that can be drawn from these data is that, in general, the data for large bubbles is better predicted by the potential flow model of Chen and Saxena [11] without the inclusion of drainage. The reason for this could be non-zero value of the bed viscosity which would tend to inhibit the motion of the particles and thus cause the bulge volume to be greater than that for potential flow. The bulge volumes predicted by potential flow and creeping flow tend towards each other as the value of $R_b/H_b$ increases, Mathur et.al. [14]. Finally, the effect of bed walls on the parameters measured in this study make this comparison with 2-D theories only approximate. A more conclusive
Figure 8. Bubble rise velocity as a function of bubble frontal area.
Figure 9. Variation of $\varepsilon_b$ at four peripheral positions ($0^\circ$, $90^\circ$, $180^\circ$ and $270^\circ$) of the 11 mm outside diameter glass tube, and $\varepsilon_b$ as a function of $U/U_{mf}$. 
Figure 10. Variation of $\varepsilon_b$ at four peripheral positions ($0^\circ$, $60^\circ$, $180^\circ$ and $300^\circ$) of the 28 mm outside diameter glass tube, and $\varepsilon_b$ as a function of $U/U_{mf}$. 
check of the model proposed here is possible for three-dimensional beds, and the trends indicated by the present results may be tested more unambiguously for 3-D beds.

A number of phenomena such as solids mixing, solids elutriation, gas-solid contacting and heat transfer characteristics of immersed surfaces depend in a sensitive manner on the bubbling features of gas fluidized beds. We have been investigating the hydrodynamic features in a two-dimensional bed [2] as related to bubble dynamics [7, 14] and solids elutriation from the bed surface [7, 8, 14]. The present experiments are conducted in a variable thickness two-dimensional fluidized bed of width 50.8 cm and thickness 6.35 cm equipped with bubble cap distributor plate. Silica sand of average particle diameter 730, 1237, 2356 \( \mu \text{m} \) are used as bed material. Some of the properties of this particle are listed in Table I as well as the air superficial fluidizing velocity at which the bubble velocities are measured.

It is generally found that the bubble shape is almost spherical especially in the lower portion of the bed and the bubble grows by coalescence as it rises up in the bed. The measured bubble velocities, \( U_B \), are plotted as a function of the bubble frontal area, \( A_B \), in Figure 8 as a log-log plot. These data are analyzed in terms of the theory for bubble size.

Experimental data suggest an increase in \( \mu_B \) with an increase in \( d_p \) consequently for the same size of bubble, the bubble rise velocity in a bed of larger particles will be less than in a bed of smaller particles of the same material and sphericity. This is evident from an examination of the data for the three silica sand particles in Figure 8. The measured \( U_B \) values for a bubble with a frontal area of 14.3 cm\(^2\), is 34.26, 28.33 and 15.3 cm/s for beds of mean particle size of 730, 1237 and 2356 \( \mu \text{m} \) respectively.

In Figures 9 and 10, the experimental results are presented in terms of reduced air fluidizing velocity, viz., \( U/U_{mf} \). In these figures the bulk bed voidage, \( \varepsilon_b \), as computed from the relation of equation 18 is also shown plotted at different fluidizing velocities. The average tube surface voidage, \( \varepsilon_s \), as obtained by taking the mean of four \( \varepsilon_s \) values at a particular value of \( U \) is also plotted as a function of \( U \) for comparison purposes.

In Figure 11, is shown a part of the curved surface PQRS of the glass tube immersed in the fluidized bed. ABCD is the portion of this curved surface (4mm by 5.5 mm) which is illuminated and hence is seen by the probe which is located at the tube axis 00'. A horizontal plane passing through the tube axis is taken as a reference and the location of the particle is specified by its distance from this plane. PMQR represents the circular side cross-section of the tube. Lengths of the chords AB and CD are same and they subtend the same cone angle \( \theta \) at the tube axis 00'. Further, particles at positions A and M are specified by the distance E and E + H (= outside tube radius) respectively. The portion ABCD of the tube surface is viewed
Figure 11. A section of the immersed tube emphasizing the curved portion (4 mm by 5.5 mm) illuminated and photographed by the image carrying fiber optic probe.
on the TV monitor after amplification by a factor \( A \). The value of \( A \) is the same for any point along AD but is different for every point along the arc AB. In other words, the magnification factor is uniquely defined by the position of the point relative to the reference plane and remains the same for all points which are located at the same vertical distance, \( Z \). This circumstance enable us to establish uniquely \( A \) as a function of \( Z \). Experiments are conducted to determine \( A \) for two particles of diameter 1.4 and 1.7 mm. The values so obtained will establish the sensitivity of \( A \) to small increments in the value of \( Z \). Particles are kept at various positions parallel to the arc AB or CD at the tube surface. To vary \( Z \) somewhat significantly a tygon sleeve is slipped over the tube and particles are held on the point vertically above 0 only in this work. Further, \( A \) is computed by the ratio of the diameter of the particle image on the monitor to its actual value. Figure 12 graphs the experimental results for the two particles and smooth curve is drawn through them and used subsequently for the calibration of \( F \). To this extent, it is reasonable to assume that \( A \) is independent of particle diameter as long as particle belong to a narrow size range and depends only on \( Z \).

Typical photographs of particle population at four angular positions at the outer periphery of the glass tubes as seen on the monitor at four different reduced fluidizing velocities are shown in Figure 14 for 11 mm tube and in Figure 15 for 28 mm tube. The four angular positions are: top \((0^\circ)\), right side \((90^\circ)\), bottom \((180^\circ)\) and left side \((270^\circ)\) for 11 mm tube as shown in Figure 14, and \(0^\circ, 60^\circ, 180^\circ\) and \(300^\circ\) for 28 mm as shown in Figure 15. Adopting the procedure outlined above, \( s \) values as computed from eq. \( 19 \) are plotted in Figures 9 and 10 for the tubes. Also shown in these figures are the bulk bed voidage as calculated from eq. 1. The values of the surface voidage in the fixed-bed configuration, \( \varepsilon \), at various angular positions for the two tubes are listed in Table III.

The results of Figure 9 conclusively show that tube surface voidage or particle population is the same for \( 90^\circ \) and \( 270^\circ \) angular positions for the entire range of fluidizing velocity. This of course is understandable on the basis of symmetry and the fact that gas bubbles are responsible for particle motion. The upstream side of the tube \((180^\circ)\) has lower bed voidage or greater particle population than the equatorial sides of the tube \((90^\circ \) and \(270^\circ)\). On the other hand, the maximum particle concentration is on the downstream side \((0^\circ)\) of the tube. All these results agree with the conjectured particle concentration profile around the periphery of the tube to explain the variation of local heat transfer coefficient between the tube and the bed, Saxena et. al. [15]. It should be noted that at each angular position, the value of \( s \) increases with \( U \). The increase is most pronounced at the equatorial sides, and is least at the downstream side of the tube.

Comparison of the lowest velocity \( \varepsilon \) values from Figure 9 with those given in Table III also leads to some interesting information. For example at the upstream side of the tube, the particle population remains unaffected for gas flow velocities upto
Figure 12. Variation of magnification factor, $A$, as a function of distance of the particle from the reference plane, $z$.

Figure 13. Relative position of the optic probe in the two glass tubes. All dimensions are in mm.
Figure 14: Typical records of particle population at various peripheral positions (P.P.) of the 11 mm tube at different air velocities.
<table>
<thead>
<tr>
<th>$U/U_m$</th>
<th>$0^\circ$</th>
<th>$60^\circ$</th>
<th>$180^\circ$</th>
<th>$300^\circ$</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.99</td>
<td>![Image]</td>
<td>![Image]</td>
<td>![Image]</td>
<td>![Image]</td>
</tr>
<tr>
<td>1.04</td>
<td>![Image]</td>
<td>![Image]</td>
<td>![Image]</td>
<td>![Image]</td>
</tr>
<tr>
<td>1.06</td>
<td>![Image]</td>
<td>![Image]</td>
<td>![Image]</td>
<td>![Image]</td>
</tr>
<tr>
<td>1.09</td>
<td>![Image]</td>
<td>![Image]</td>
<td>![Image]</td>
<td>![Image]</td>
</tr>
</tbody>
</table>

Figure 15  Typical records of particle population at various peripheral positions (p.p.) of the 28 mm tube at different air fluidizing velocities.
Table I. Particle and fluidized bed properties and salient operating conditions

<table>
<thead>
<tr>
<th></th>
<th>Glass Beads</th>
<th>Silica Sand</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mean particle diameter $d_p$ (um)</td>
<td>265 1527 3050</td>
<td>730 1237 2356</td>
</tr>
<tr>
<td>Particle density, $\rho_s$ (Kg/m$^3$)</td>
<td>2970 2940 2490</td>
<td>2565 2580 2650</td>
</tr>
<tr>
<td>Minimum fluidizing velocity, $U_{mf}$ (m/s)</td>
<td>0.06 0.55 1.90</td>
<td>0.22 0.48 0.69</td>
</tr>
<tr>
<td>Voidage at minimum fluidization, $E_{mf}$</td>
<td>0.40 0.40 0.42</td>
<td>0.38 0.40 0.40</td>
</tr>
<tr>
<td>Slumped bed height, $H_{mf}$ (m)</td>
<td>0.47 0.27 0.185</td>
<td>0.17 0.19 0.19</td>
</tr>
<tr>
<td>Superficial fluidizing velocity, $U$ (m/s)</td>
<td>0.65 0.75 2.21</td>
<td>0.45 0.88 1.019</td>
</tr>
<tr>
<td>Bed temperature, $T_b$ (K)</td>
<td>294 308 298</td>
<td>299.9 301.3 301.2</td>
</tr>
<tr>
<td>Bed pressure, $P_b$ (KPa)</td>
<td>16.3 165 165</td>
<td>164.33 126.4 119.5</td>
</tr>
</tbody>
</table>
TABLE II. Values of the correction factor ($F$) and average particle image area ($a$) as a function of particle vertical distance from the tube surface ($r$) for particles of 1.4 mm average diameter.

<table>
<thead>
<tr>
<th>$r$ (mm)</th>
<th>$a_{(11 \text{ mm tube})}$</th>
<th>$a_{(28 \text{ mm tube})}$</th>
<th>$F$</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.00</td>
<td>1458</td>
<td>1239</td>
<td>0.666</td>
</tr>
<tr>
<td>0.03</td>
<td>1448</td>
<td>1233</td>
<td>0.637</td>
</tr>
<tr>
<td>0.05</td>
<td>1439</td>
<td>1225</td>
<td>0.619</td>
</tr>
<tr>
<td>0.07</td>
<td>1433</td>
<td>1217</td>
<td>0.600</td>
</tr>
<tr>
<td>0.10</td>
<td>1423</td>
<td>1212</td>
<td>0.571</td>
</tr>
<tr>
<td>0.20</td>
<td>1388</td>
<td>1190</td>
<td>0.475</td>
</tr>
<tr>
<td>0.30</td>
<td>1354</td>
<td>1160</td>
<td>0.380</td>
</tr>
<tr>
<td>0.40</td>
<td>1330</td>
<td>1137</td>
<td>0.286</td>
</tr>
<tr>
<td>0.50</td>
<td>1295</td>
<td>1116</td>
<td>0.190</td>
</tr>
<tr>
<td>0.60</td>
<td>1268</td>
<td>1096</td>
<td>0.095</td>
</tr>
<tr>
<td>0.70</td>
<td>1236</td>
<td>1080</td>
<td>0.000</td>
</tr>
</tbody>
</table>

TABLE III. Fixed bed voidage at the tube surface, $\varepsilon_s$ at various angular positions for the two glass tubes.

<table>
<thead>
<tr>
<th>Angular Position</th>
<th>$\varepsilon_s$ (11 mm)</th>
<th>$\varepsilon_s$ (28 mm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0°</td>
<td>0.616</td>
<td>0.635</td>
</tr>
<tr>
<td>90°</td>
<td>0.632</td>
<td>0.646</td>
</tr>
<tr>
<td>180°</td>
<td>0.625</td>
<td>0.638</td>
</tr>
<tr>
<td>270°</td>
<td>0.640</td>
<td>0.643</td>
</tr>
</tbody>
</table>
0.99 \( U_{mf} \), while for the 90° and 270° positions the particle population is decreased for the same gas flow. The reduction in particle population for 180° position is intermediate to those for 90° (or 270°) and 0° position. These results also show that in general there is some particle movement in the bed with gas flow even for velocities below minimum fluidization velocity.

The mean of the four \( \bar{e}_s \) values at a particular gas velocity is interpreted as mean tube surface voidage and is designated as \( \bar{e}_s \). In Figure 9, variation of \( \bar{e}_s \) is shown with reduced gas velocity and it is interesting to note that its values are significantly larger than the bulk bed voidage values, \( e_b \), also graphed in this figure. The pronounced difference in the two sets of values, emphasizes the fact that gas flow around an immersed surface in a granular bed is much larger than in the bulk of the bed.

Results similar to those of Figure 9 are displayed for 28 mm tube in Figure 10, and these confirm qualitatively the findings from the 11 mm tube. However, there are some quantitative differences in the results for the tubes. For example, at each angular position, the variation of \( \bar{e}_s \) with gas velocity for the wider tube (28 mm) is less than that for narrowed tube (11 mm). The downstream side of the tube exhibits a constant value for a wide range of gas velocities up to 1.06 \( U_{mf} \) and it is the same as obtained for the fixed bed mode, Table III.

This emphasizes the presence of a defluidized cap on the downstream side of the tube as has been conjectured earlier, Saxena et. al. [15]. The bulk bed voidage values are again found to be significantly smaller than the mean tube surface voidage values. Our measurement suggest a similar variation and value for the gas film thickness for this 28 mm tube as that for 11 mm tube.

The heat transfer experiments are conducted with 28.6 and 50.8 mm diameter tubes, kept at a distance of 28 cm from the distributor plate. The slumped bed height is kept about the same in all experiments at approximately 35 cm. The total heat transfer coefficient, \( h_w \) is calculated from:

\[
h_w = \frac{Q}{A (T_w - T_b)}
\]

The maximum absolute error in the experimental values of \( h_w \) is estimated to be ± 8%. The precision of our measurement as judged from the reproducibility of data points is about ± 2%.

The values of \( h_w \) are plotted in Figure 16 as a function of \( G \). It is to be noted that \( h_w \) in all cases increases with increase in \( G \) until a maximum value of \( h_{w max} \) is attained for a particular value of \( G \), \( G_{opt} \), and decreases slowly thereafter with further increase in \( G \). These figures also indicate that \( h_w \) decreases with an increase in the solid particle diameter at the same fluidizing velocity. Both these qualitative trends are in complete agreement with the reported findings in the literature.
Figure 16. Variation of $h_w$ with $G$ for tubes of diameter 28.6 mm and 50.8 mm.
REFERENCES