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Transient heat transfer and bubble dynamics in a pressurized fluidized bed

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Abstract—A flow visualization technique for studying the gas bubble dynamics in a pressurized fluidized bed was developed and used to quantify these dynamics at the surface of a vertical tube submerged in the bed. Transient heat flux measurements were made and correlated with bubble motion. As a result, it is concluded that the heat transfer process is strongly affected by bubble dynamics and is much more complex than any of the generally accepted models can predict. It is also shown that the overall bed operating conditions are the primary driver for local bubble/particle motion around the tube which significantly affects the time-dependent fluctuation in local heat transfer.

1. INTRODUCTION

THE IMPLEMENTATION of fluidized bed technology for the combustion of coal to produce electrical power is a very promising method of making this an environmentally acceptable energy conversion technique. One of the limitations to this technology, however, is the ability to scale laboratory test data to commercial size systems. This inability is caused by a lack of understanding of the local heat transfer to boiler tubes submerged in the fluidized bed. The current techniques for predicting local heat transfer rates are extremely empirical; some even require that model coefficients be measured at full-scale geometries and conditions. These techniques attempt to predict an 'average' coefficient predicated on a steady-state type of analysis. It is well known, however, that the process is extremely transient as a result of transient and localized effects of the gas bubbles that percolate up through the bed. The result is that there is no consensus on predictive techniques for sizing heat transfer surface area for boiler tubes submerged in a fluidized bed combustor. What is drastically needed is an understanding of the complex transient heat transfer process at the surface of in-bed boiler tubes as a result of bubble dynamics at these surfaces.

A fluidized bed combustor is a device that consists of a vessel filled with small particles of coal (10% by volume) and limestone or sorbent (90% by volume). The bottom of the vessel consists of a distributor plate that allows air to be blown upward through the bed. As the flow velocity is increased the packed bed of stationary particles will begin to separate and the particles will be suspended in the air stream at a point where the drag force on the particle is equal to the weight of the particles. This condition is defined as minimum fluidization. As the flow is further increased, the movement of particles will resemble that of a liquid and at even higher velocities voids or bubbles percolate through the bed and the overall appearance is similar to a boiling liquid.

The technical problem of predicting flow characteristics and heat transfer in a fluidized bed has been studied extensively. The body of literature is so vast that it is not practical to attempt a complete review here, but even with this body of work no generally accepted procedure exists for predicting local heat transfer coefficients. The following brief review will summarize the primary modeling efforts in this field and point out what work is needed to provide the missing information for providing predictions of local heat transfer coefficients at submerged tube surfaces.

The bulk of the previous modeling efforts has been for horizontal tubes in large particle systems. Large particle systems are defined as those with thermal time constants substantially greater than their residence time at a heat transfer surface. At typical operating conditions of a fluidized bed combustor, particles of 1 mm or larger are 'large' by this definition. An initial attempt at modeling the heat transfer process in a fluidized bed was made by Mickley and Fairbanks [1]. This model considered the process in terms of a packet of particles that move from the interior of the bed to the heat transfer surface and then return to the interior of the bed. Heat was assumed to be transferred between the packet and the surface by a conduction process that is proportional to the net conductance of a quiescent bed, which must be measured experimentally. The heat transfer coefficient between the bed material and the heat transfer surface was assumed to be a function of packet residence time, and the heat transfer during bubble contact was assumed to be negligible. Therefore, the heat transfer coefficient was proportional to the square root of the product of particle thermal conductivity, particle density, particle specific heat, and a stirring factor. The stirring factor is a function of the frequency at which stationary packets contact the surface and the residence time of these packets at that location. Packets were assumed to be stationary during contact with the surface. This stirring factor must correlate the bulk bed operating conditions with the local dynamics of the particles at the surface. They were not able, however, to either produce an analytical expression for this factor or measure this factor experimentally, but suggested that future experimental work could accomplish this critical step by measuring heat transfer coefficients and bulk bed operating conditions.

This technique was extended by Glicksman and Decker [2] to include the effect of the bubble phase and a more usable expression of the stirring factor. This model not only included particle thermal properties, but also bubble residence time fraction, solids residence time, and bulk bed void fraction at minimum fluidization. They concluded that while this was a definite improvement and was a physically plausible model it was still difficult to apply because of its inherent dependence on two quantities difficult to measure, let alone predict: bubble residence time fraction and solids residence time. To resolve this dilemma, they adopted an empirical approach by modifying a general correlation used for packed beds which obviously does not include provisions for bubbles. This empirical model was developed for large particles and horizontal tubes and predicted Nusselt number as a function of bubble fraction, Reynolds number and Prandtl number. This model still requires knowledge of the bubble fraction which was estimated by experimentally determining bulk bed void fraction. This approach, therefore, assumes that the bed is uniform throughout which is not realistic and that the particles are stationary when in contact with the surface.

Catipovic *et al.* [3] conducted an experimental study of large particles around horizontal tubes from which they proposed a Nusselt number relation which included a residence time for tube/bubble contact, Archimedes number, Prandtl number, Reynolds number at minimum fluidization, and a ratio of particle diameter to tube diameter. They measured instantaneous heat transfer coefficients and obtained maximum and minimum values which were assumed to correspond to emulsion and bubble phase coefficients. They concluded that the local value of bubble residence time, obtained from capacitance measurements, differed significantly from the mean value for the bed as a whole. Here again the assumption was made that only two stable states occur at the tube surface, one with a stationary emulsion and one with a bubble.

As observed by Zabrodsky *et al.* [4], in large particle systems, the bubbles move slower than the interstitial gas, whereas in small particle systems the bubbles move faster than the interstitial gas. For horizontal tubes and large particles, the heat transfer coefficient was presented as an empirical function of gas thermal properties, bulk bed void fraction, particle diameter and superficial gas velocity. The value of bed void fraction was determined experimentally.

A mechanistic theory for large particle systems was developed by Ganzha *et al.* [5]. They assumed that in the absence of radiation the total heat transfer was composed of a conduction term and a convection term. The conduction term was calculated by (1) assuming that the stationary solid particles are distributed around the tube in an arrangement of unit orthorhombic cells and then (2) considering composite infinite layers of gas and solid. The unsteady-state heat conduction equations were then solved under these well-defined boundary and initial conditions. The convective term was evaluated by assuming that the boundary layer on the tube was disrupted at the front half of the particle and was reformed in its wake. The resulting model for large particle systems with horizontal tubes was in the form of Nusselt number as a function of Reynolds number at operating conditions and at minimum fluidization, Prandtl number, bed void fraction near the tube for operating condition and at minimum fluidization, Archimedes number, and ratio of particle diameter to tube diameter. It was determined that the bed void fraction near the surface was larger than the bulk bed void fraction. While this model was proposed as a mechanistic model it is still somewhat empirical and therefore limited to the conditions used for developing the constants. The authors concluded that a body of accurate data covering a wide range of conditions is needed to broaden the scope of this theory and that the specification of surface void fraction needs special attention. This model also provided the insight that local bed conditions were different from bulk bed conditions, which is extremely important.

Decker and Glicksman [6] extended their earlier model by including the effect of bubbles or voids. Their physically based model was proposed for large particle systems. They observed that in many instances the heat transfer behavior of large and small particles are contradictory. That is, heat transfer is proportional to particle size for large particles but is inversely proportional for small particles. For small particles, heat transfer is independent of bed pressure, but large particle heat transfer increases with pressure level. The modified model predicts total Nusselt number as a function of two separate Nusselt numbers and the bubble void fraction. One Nusselt number corresponds to the emulsion phase and the other to the void phase. This model requires knowledge of local bubble void fraction at the surface, bubble rise velocity at the surface and bubble length scale at the surface. None of these quantities were measured directly and therefore this technique is severely limited at this time. Even though authors realized the importance of local bed conditions, they assumed that two steady-state conditions occurred (i.e. stationary emulsion and bubbles).

A much smaller body of literature exists for fluidized beds with vertical tubes. In one of the more comprehensive studies, Boradulya *et al.* [7] conducted an experimental study of heat transfer in a high pressure fluidized bed with large particles and vertical tubes. When they tried to use various models they concluded that a major need exists to directly measure the values of bulk and local bed void fraction. In addition, there is a need for modifying existing ambient pressure horizontal-tube empirical models for conditions of high pressure vertical tube systems.

A model developed by Martin [8–10] makes use of some basic ideas adopted from the kinetic theory of gases to describe the mechanism of energy transfer through the moving particles. The heat transfer coefficient relation comprises three components: particle conduction, gas convection and radiation. The particle conduction is a transfer mechanism in which the particles transfer energy, which they have acquired in the bulk bed, to the tube surface. The gas convective and radiative components represent the direct heat transfer from the carrier gas to the tube surface. For very large particles (diameter greater than 3 mm), which require corresponding large gas velocities for fluidization, this component can play the dominant role. The model is somewhat more difficult to implement than the other models discussed here and does assume that the local bed void fraction is equal to the uniform bulk bed void fraction. The author observes that the model is capable of describing local dependencies provided the bulk void fraction is replaced by the corresponding local value.

An approach developed by Bock [11–13], which is an extension of Mickley and Fairbanks' packet theory, results in a simpler expression for the value of heat transfer coefficient as a function of bubble fraction and residence time. The void fraction can be either the bulk value or local value at the wall if known. Bock used a capacitance probe to determine local values of bubble void fraction.

As can be seen from these modeling efforts, most current techniques attempt to include the effect of bubble dynamics by some type of local or bulk 'average' void fraction or 'average' residence time. The result is a steady-state analysis assuming two exclusive processes: one for the condition when the bubble is in contact with the surface, and another when the emulsion, assumed to be stationary, is in contact with the surface. Since these techniques have proven to be inadequate for other than the specific geometries and test conditions used as the empirical basis for the model, a new approach is warranted. This approach is to study the process as a transient phenomenon using flow visualization and to investigate the local fluctuation in heat transfer at the tube surface with the variations in simultaneous local bubble and emulsion dynamics.

2. EXPERIMENTAL FACILITY

A pressurized fluidized bed test facility was fabricated for this study. A schematic of this facility is shown in Fig. 1. A 17 700 liter (625 ft^3), 300 atm nitrogen blowdown tank provides the source of fluidizing gas, which is pressurized by a liquid nitrogen pumping system. Nitrogen flow from the blowdown tank is metered to the fluidized bed by a series of flow and pressure regulators to independently control the pressure and volume flow rate to the bed within the operating conditions of 0.76 m s^{-1} (2.5 ft s^{-1}) to 49 m s^{-1} (15 ft s^{-1}), 1–20 atm. An orifice meter is used to measure flow rate upstream of the bed and along with various temperature and pressure sensors provides the quantification of bed operating conditions. The fluidized bed is a 30.5 cm (1 ft) diameter by 305 cm (10 ft) high cylindrical vessel with five side access ports. The distributor plate is interchangeable and currently comprises a 900 × 1200 mesh stainless steel screen sandwiched between perforated steel plates which provides a uniform distribution of fluidizing nitrogen across the bed cross section. A vertical 5 cm (2 in) diameter by 30.5 cm (12 in) long steel tube test section is installed within the bed, as shown in Fig. 2. The instrumented section of this tube includes a pyrex window for visualization, a fast response heat flux gage and a fast response thermocouple both mounted directly above the window. Inside the tube two high intensity lights are installed to provide both illumination for visualization through the window and a heat source. An air-cooling system is included to control the temperature at the instrumented section.

Flow visualization is recorded by a Spin Physics SP2000 system, a high-speed videographic motion analyzer. Figure 3 shows the SP2000 positioned at the viewing port of the fluidized bed. The motion analyzer utilizes a solid-state video sensor to provide video records from 60 to 2000 full frames per second or up to 12 000 partial frames per second. Playback speeds from 1 to 60 frames per second provide a slow down factor from 1 to 12 000 times. In the playback mode X and Y reticles can be activated to accurately locate the position of an image. Thus by knowing the frame rate, the exact time of the recording and position in each frame, displacements, velocities and accelerations can be accurately determined. Since the information processing is done digitally, the data can be transmitted directly to a computer for off-line analysis.

The heat flux gage (RdF Model No. 20466-3) is a fast response element photodeposited on a thin foil. The output from this and a similarly constructed thermocouple is recorded on a high-speed oscillograph along with a timing signal from the motion analyzer. The combination of these instruments provides the capability to accurately monitor the high-speed transient heat transfer and bubble dynamics necessary for this study.

3. DISCUSSION OF RESULTS

3.1. *Instantaneous velocity and heat flux data*

Heat flux measurements using the fast-response heat flux gage and flow visualization using the SP2000 high-speed videographic recording system were combined to provide insight into the transient effects of bubble and particle dynamics on the heat transfer

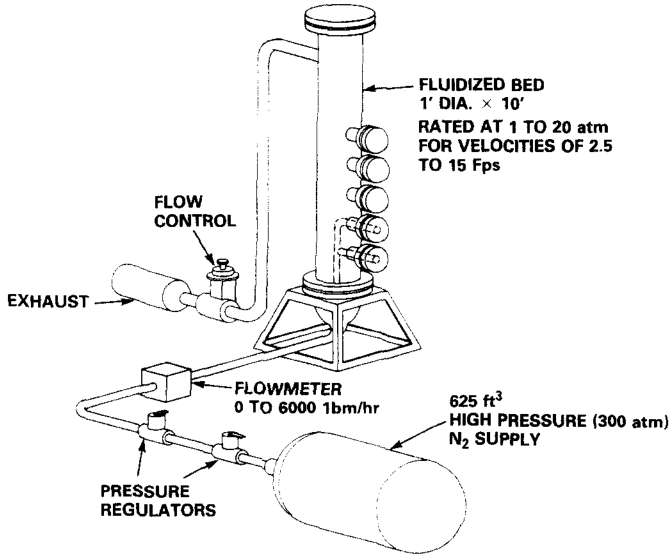

FIG. 1. Pressurized fluidized bed test facility.

from fluidized bed heat exchanger tubes. An example of the output from this videographic system is shown in the series of recordings illustrated in Fig. 4. The scenes shown in this figure are four frames of a video sequence recording the passing of two bubbles in a fluidized bed. In this test, glass beads with an average diameter of 800 μm (± 50 μm) were fluidized at ambient temperature and atmospheric pressure. These four photographs were taken from the monitor of the SP2000 motion analysis system. The video camera focused on the outside surface of the transparent window submerged in the fluidized bed by sighting down the length of an access port placed at a right angle to the tube. One can see that the elapsed time (shown in the top right-hand corner of the photograph) from scene (a) to scene (d) is 10.5 ms. Twenty-one frames (frame numbers appear in the lower right-hand corner of each photograph), of which these are four, make up this sequence, corresponding to a frame rate of 2000 frames s^{-1}. The reticles (i.e. white cross-hairs intersect at the particle location) track a single particle as it falls through the first bubble (the dark semi-circular area in the top left-hand corner) and lands on the bubble's lower surface. The second bubble enters the frame in scene (c) (frame No. 2203) at the lower center portion of view and by scene (d) (frame No. 2210) has moved a distance of more than half its diameter into the frame. Along with simply recording the overall flow field and aiding in the use of other instruments, the flow visualization method used here provides high-speed motion analyses for events such as bubble passage or determining emulsion velocity. Referring again to Fig. 4, the coordinates of the tracked particle can be combined with the recorded time to provide the average particle velocity in the focal plane of the image. In this test, the geometric calibration factor is 59.1 μm pixel^{-1}; and the particle travelled 27.7 pixels; so the average velocity of the particle was 155.9 mm s^{-1}. This example portrays the capabilities of the motion analysis system. The same technique was also applied to bubble leading and trailing edges in the same manner it was applied to the particle. The results of one of these tests are described and discussed in detail here.

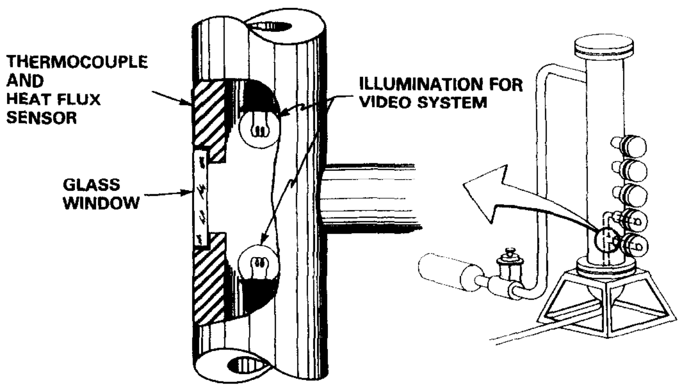

FIG. 2. Flow visualization access window.

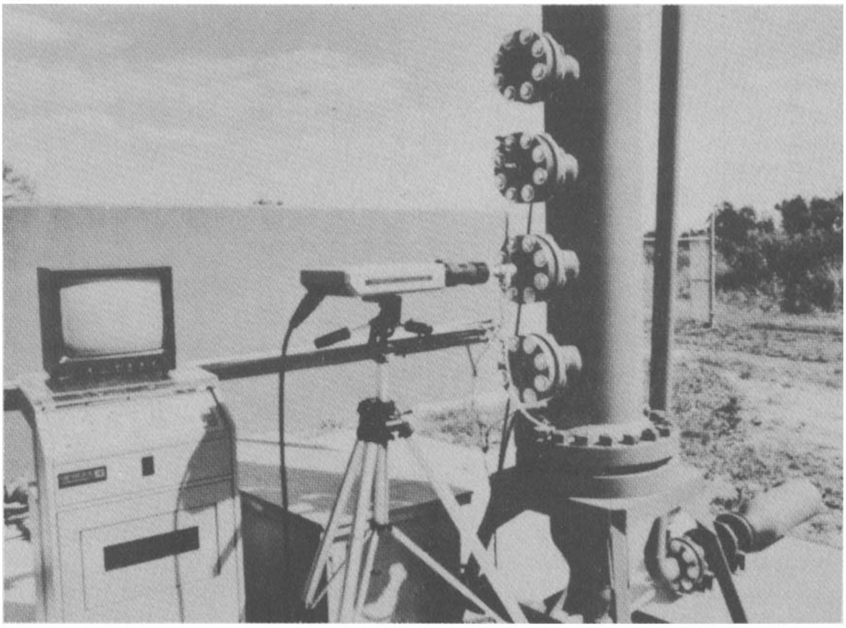

FIG. 3. High speed video system.

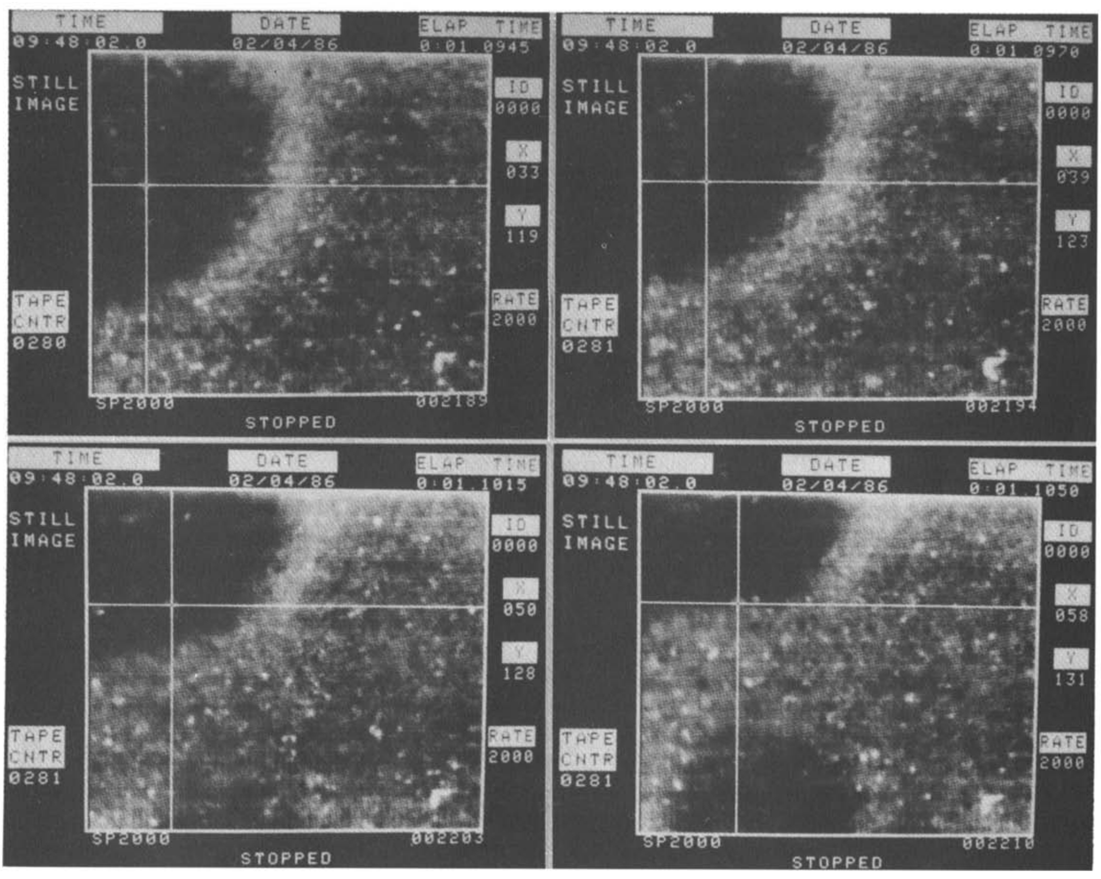

FIG. 4. High speed video of bubble.

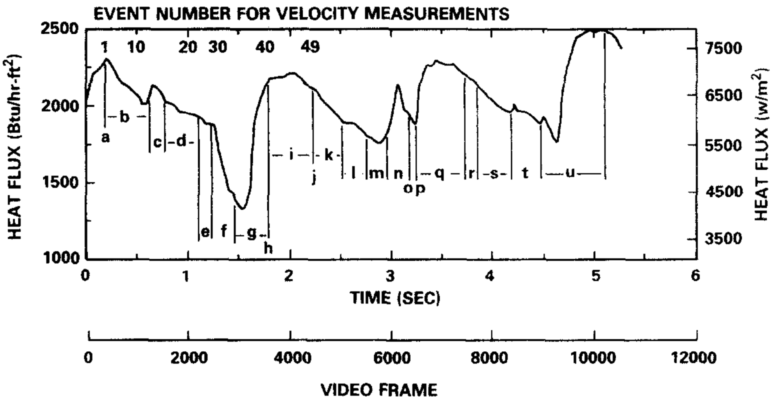

FIG. 5. Fluidized bed heat transfer.

The signal from the heat flux gage is typically less than 10 mV during these tests; so, the signal requires amplification. Because the environment surrounding the fluidized bed facility is electrically noisy, the signal must also be filtered to eliminate 60-cycle noise. Fortunately, the heat flux variations measured by the sensor are well below 60 Hz.

The video recordings for this particular test were made at 2000 frames s^{-1} with the SP2000 unit. A 200 mm lens set at f/5.6 was used to sight into the access tube at the surface of the glass viewport window inside the bed. The 'RECORD' signal of the SP2000 was recorded alongside the heat flux sensor output on the oscillograph. This provided a timing mark for synchronizing the recorded video images with the heat flux sensor oscillations.

Figure 5 shows a facsimile of the heat flux output during the time video recordings were made. This graph shows heat flux from the gage over a ~ 5.25 s period. It is seen that the maximum heat flux is approximately twice the minimum and the extremes are $\pm 30\%$ about the mean; so heat flux variations are not insignificant. The letters on the graph provide a key to a summary of the video sequence, while the numbers correspond to points at which particle velocity measurements were made.

A review of the video sequence reveals that several significant events occur during this test sequence. A commentary of the recorded video images is provided in Table 1 for the entire $\sim 10\,250$ frames (5.125 s). It can be seen that peaks in heat flux at the tube surface occur immediately after a bubble passes.

Table 1. Bubble dynamics observations

Key	Frame	Comment
a	450–550	Bed still; no particle movement
b	550–1280	Bubble passing at right edge of field of view. General movement of the particles is counterclockwise
c	1280–1600	Bed still; no particle movement for ~ 300 frames (~ 0.15 s)
d	1600–2250	General particle movement is counterclockwise. No bubble visible in field of view
e	2250–2360	Bed still; no particle movement for ~ 100 frames (~ 0.05 s)
f	2360–2941	Top of bubble entering field of view in Frame 2360
g	2941–3776	Bottom of bubble entering field of view in Frame 2941. Particle movement is upward and decelerates to near zero
h	3776	Bed paused briefly
i	3776–4505	Particles move counterclockwise
j	4505	Bed still
k	4505–5010	Particle movement upward, slow
l	5010–5678	Bubble; chaotic particle movement in dilute gas/solid mixture of inner bubble
m	5678–5920	Particle movement upward, slow
n	5930–6307	Bubble
o	6307–6460	Rapid particle movement upward in wake of bubble
p	6460–6517	Bed still
q	6577–7460	Two cycles of counterclockwise movement
r	7460–7700	Bed still
s	7700–8208	Particle movement upward
t	8208–9007	Bubble; chaotic particle movement in dilute gas/solid mixture
u	9007–10 250	Decelerating upward movement of particles

Heat flux slowly decreases when there is little or no particle movement at the tube surface. These phenomena may be explained as follows.

Just after a bubble passes, fresh particles come into contact with the tube. These particles originate from the bulk of the bed where the temperature is quite different from that of the tube surface. As the particles remain at the tube surface, the temperature difference between the tube and the particle decreases, thereby decreasing the rate of heat transfer by conduction. As a new bubble passes, particles are removed from the surface so it is now exposed only to the gas phase, or, at most, a very dilute mixture of gas and solid particles. The heat transfer by convection is less than that by conduction. As the bubble passes, new particles contact the surface to start the cycle over. The intermediate peaks occur as a result of intermediate velocities caused by bubbles that pass in the vicinity but not directly at the heat flux gage. Other intermediate results occur when already heated particles are moved along the surface but are not replenished by fresh lower temperature particles from the bed interior.

Particle velocity measurements were made at various times during the first 4500 frames (2.25 s) of the video sequence. This was accomplished with on-board digitizing capability of the SP2000. Particle position was tracked through several frames. These position/time measurements thereby provide an average velocity over that particular set of frames.

The magnitude of the particle velocity is shown as a function of elapsed test time for various event numbers in Fig. 6. It is seen that the major peak in velocity occurs in the wake of a bubble. The three minor peaks occur at times when the general particle movement is left-down-right, or counterclockwise.

Polar plots of velocity reveal more detail about particle movement. Figure 7(a) shows particle velocities during the time period 'b' of the test (see Fig. 5 and Table 1). Particle movement is clearly counterclockwise as a bubble passes on the extreme right of the field of view. Particles do not move upward at any time in this portion of the sequence. During period 'd' of the sequence, Fig. 7(b), particle movement is again counterclockwise with no movement upward. No bubble was visible in the field of view, but the overall movement is very similar to that during period 'b'.

Particle velocities just prior to and after the passage of the large bubble are shown in Fig. 7(c). The acceleration of the particles at the leading edge of the bubbles (points 28–30) the subsequent deceleration of the particles in the wake of the bubble (points 31–40) are clearly shown in which the particle movement is generally upward. Finally, the sequence closes with another series of counterclockwise particle movements (points 42–49).

It is quite clear from these data and visual observations, that the heat transfer process is much more complex than any of the previous models indicated and is a function of the complex local bubble dynamics at the surface. Even when a bubble does not directly pass a given heat transfer location, its motion can cause the emulsion at the location to accelerate or decelerate. Therefore, conditions at a localized tube location range from a stationary emulsion at the tube, an accelerating emulsion, a decelerating emulsion, and a fast moving bubble with or without particles raining through the bubble. The direction of emulsion flow also has a bearing on heat transfer due to the local temperature of the emulsion. Fresh emulsion from the bed interior has a different effect than emulsion with the same velocity that has been in contact with the tube. The origin of the emulsion and its velocity seems to be governed by the distance from the heat flux gage to the passing bubble.

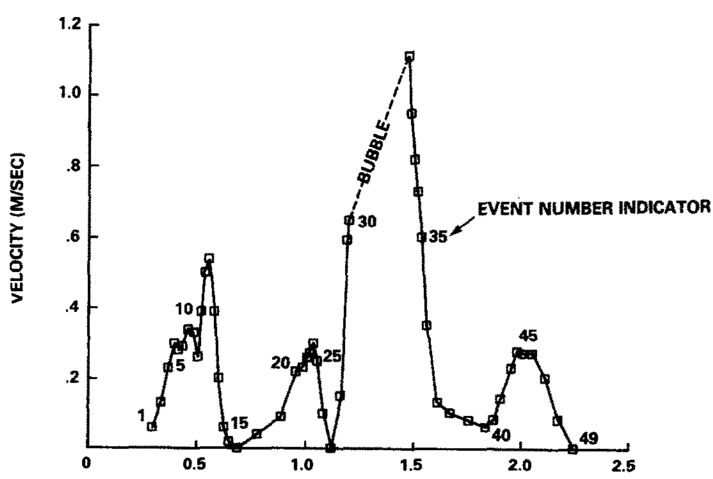

FIG. 6. Particle velocity.

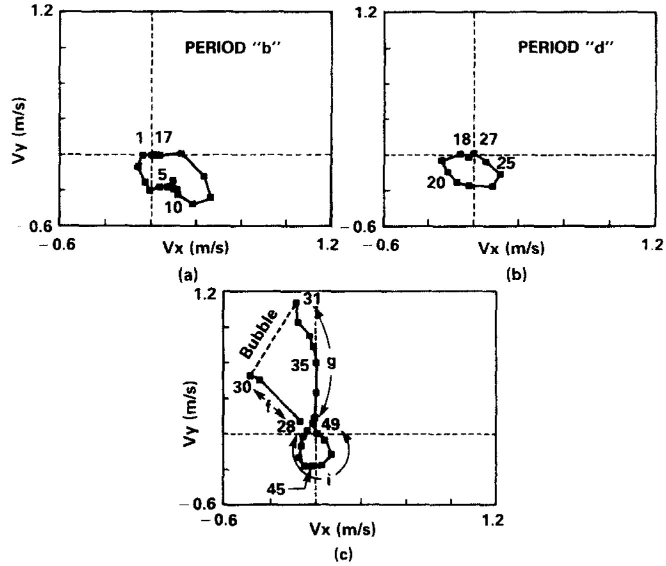

FIG. 7. Particle velocity components.

3.2. *Periodic heat transfer data*

In order to provide a qualitative understanding of the effect of various bed operating conditions on heat transfer, heat flux measurements were recorded for longer periods of time, shown in Figs. 8 and 9. Figure 8 shows three traces of heat flux with constant non-dimensional mass flow rate (mass flow rate divided by mass flow rate at minimum fluidization) at three bed pressures. As bed pressure increases, both the mean heat flux and the amplitude about the mean increases. The frequency of heat flux oscillations, however, remains relatively constant. Figure 9 shows three traces of heat flux for constant bed pressure at three non-dimensional mass flow rates. As mass flow rate increases the mean value of heat transfer, the amplitude of fluctuation and the frequency of fluctuation all increase. For the two lower values of mass flow rate the mean and amplitude increase only slightly, but the frequency exhibits a significant increase. The highest value of mass flow rate exhibits a significant

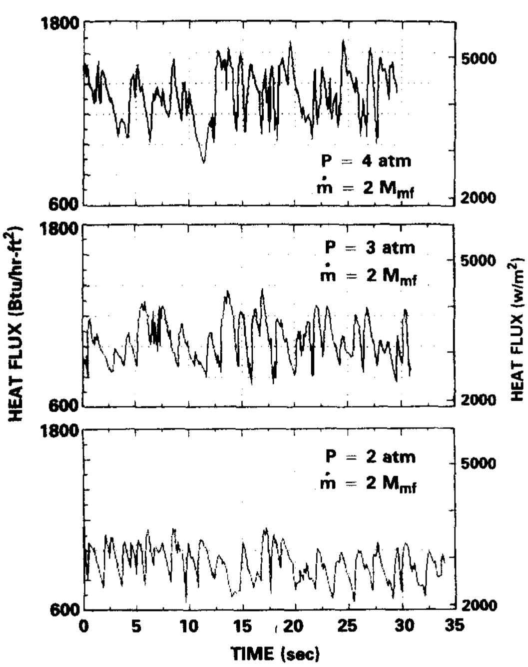

FIG. 8. Bed pressure effects on heat flux.

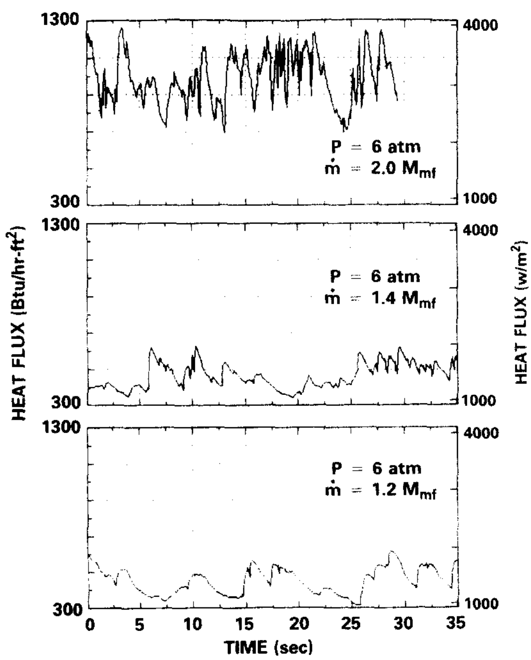

FIG. 9. Flow effects on heat flux.

increase in all three indicators of heat transfer. The importance of these three parameters is that the total heat transferred during a given period is not accurately represented by the mean value alone, but is a function of all three parameters.

4. CONCLUSIONS AND RECOMMENDATIONS

The major accomplishment of this study is the development of a flow visualization technique to document the local bubble dynamics at the surface of tubes submerged in a pressurized fluidized bed. Using this technique, it has become quite clear that the heat transfer process is directly affected by local bubble dynamics and that the local bubble dynamics are far more complex than any of the current models can predict. From detailed analysis of the heat flux data and visualization of local particle/bubble dynamics, a significant improvement in the understanding of the complex process has been gained. The analytical basis for most current models is that the heat transfer surface is exposed to one of two stable states. One state is a stationary emulsion of particles in contact with the surface that transfers heat by a conduction process. The second state is a convective state as a result of the surface being exposed to a moving bubble. This state is either assumed to have negligible heat transfer or at a level significantly lower than the emulsion state. The visualization of this process shows that the motion of the bubbles provides a range of emulsion velocities from a maximum at the trailing edge of a bubble to a stationary emulsion in the complete absence of a bubble. The velocities in between the extremes are caused by bubble motion in the vicinity of the heat flux gage. The closer the bubble is to the gage the higher the velocity. In addition, the direction in which the particles are moving is important. If the particle motion results in pulling fresh particles from the interior of the bed the heat transfer is greater due to a maximum temperature difference between the tube and particles than if the particles have had previous contact with the tube surface at another location. At a given temperature difference, maximum heat transfer occurs by pure conduction for a stationary emulsion, but with a stationary emulsion the heat transferred to the particle results in a gradual decrease in temperature difference and a gradual decrease in heat transfer. A steady replenishment of fresh particles from the bed interior maintains a high temperature difference, but does not provide a pure conduction process. As particle velocities increase the contact time decreases and the heat transfer decreases. In the limit when a bubble is in contact with the surface pure convection occurs. The minimum heat transfer results from a bubble in contact with the surface that has moved along the tube surface and has already exchanged heat with the tube thereby minimizing surface to bubble temperature difference. The net conclusion, therefore, is that the interaction between global bed conditions should affect the temperature history of the particles as well as the velocity and frequency of bubble motion. By reviewing the resulting heat transfer as a function of bed operating pressure and overall gas velocity, the mean heat flux value, the amplitude of variation about the mean and frequency of heat flux oscillation show an increase with both bed operating conditions.

The results presented here provide a 'qualitative' indication of the transient heat transfer process and the need to modify the basic assumptions used in developing most current models. This 'qualitative' approach should be expanded to provide the quantitative data necessary for model development. Recommended future research should include an extension of bed operating conditions to include a full range of bed pressures, bed temperatures, flow rates, particle sizes, particle compositions, tube orientations (horizontal and vertical), tube geometries (single tube and tube arrays) and tube sizes. This matrix of data would be the foundation for developing a quantitative prediction capability. In addition to the techniques used to study particle and bubble dynamics at the surface of the tube, a technique to track bubble motion from the interior of the bed to the heat transfer surface should be employed. The net result would be a model for predicting bubble dynamics in the interior of the bed as a function of overall bed operating conditions, a predictive model for local particle and bubble dynamics at the heat transfer surface as a function of bed bubble dynamics, and a model for predicting the heat transfer at the surface as a function of particle and bubble dynamics at the surface. The combined model would provide a method for predicting not only instantaneous values of heat transfer, but also an accurate method for predicting total heat transfer and consequently a procedure for sizing required heat transfer surface area for a given application.

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TRANSFERT THERMIQUE VARIABLE ET DYNAMIQUE DES BULLES DANS UN LIT FLUIDISE SOUS PRESSION

Résumé—Une technique de visualisation d'écoulement pour l'étude de la dynamique des bulles de gaz dans un lit fluidisé pressurisé est développée et utilisée pour quantifier cette dynamique à la surface d'un tube vertical immergé dans le lit. Des mesures de flux thermique variable sont effectuées et reliées au mouvement de bulle. On conclut que le mécanisme de transfert thermique est fortement affecté par la dynamique et qu'il est beaucoup plus compliqué que ce que peut prévoir n'importe quel modèle. On montre aussi que les conditions opératoires du lit déterminent le mouvement local bulle/particule autour du tube qui affecte significativement la fluctuation du transfert thermique local.

TRANSIENTE WÄRMEÜBERTRAGUNG UND BLASENBEWEGUNG IN EINEM UNTER DRUCK GESETZTEN FLIESSBETT

Zusammenfassung—Es wurde eine Technik zur Sichtbarmachung der Gasblasenbewegung in einem unter Druck gesetzten Fließbett entwickelt und zur Messung dieser Bewegungen an der Oberfläche eines vertikalen, in das Fließbett eingetauchten Rohres benutzt. Messungen transienter Wärmestromdichten wurden durchgeführt und mit der Blasen-Bewegung korreliert. Als Ergebnis folgte, daß der Wärmeübertragungsvorgang durch die Blasenbewegung stark beeinflußt wird, und daß er wesentlich komplexer ist als dies irgendeines der allgemein anerkannten Modelle vorhersagen kann. Es wird ebenso gezeigt, daß die Fließbettbetriebsbedingungen der primäre Verursacher für die lokale Blasen/Partikel-Bewegung in der Umgebung des Rohres sind, was die zeitabhängigen Schwankungen beim örtlichen Wärmeübergang bedeutend beinflußt.

НЕСТАЦИОНАРНЫЙ ТЕПЛООБМЕН И ДИНАМИКА ПУЗЫРЕЙ В ПСЕВДООЖИЖЕННОМ СЛОЕ ПОД ДАВЛЕНИЕМ

Аннотация—Разработана методика визуализации потока для изучения динамики пузырей газа в псевдоожиженном слое под давлением и получены количественные характеристики процесса у поверхности вертикальной трубы, погруженной в слой. Проведены измерения нестационарного теплового потока и найдена зависимость от движения пузырей. Сделан вывод о том, что на процесс переноса тепла большое влияние оказывает динамика пузырей и сам процесс в данном случае гораздо сложнее, чем это предсказывают общепринятые модели. Кроме того показано, что общие режимные параметры слоя обуславливают локальные движения пузырей/частиц около трубы, что в значительной степени сказывается на временных флуктуациях локального теплопереноса.