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Heat transfer from vertical surfaces to dense gas-fluidized beds

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Abstract—Heat transfer to beds of solids fluidized by gas has been measured from bed walls and from small diameter heated cylinders immersed in the fluidized solid. The experiments have been carried out in beds of 90 mm and 290 mm dia., for different bed depths, for different locations of the heating surface within the bed and at the walls. The size of particles was varied from 100 μm to 1 mm in diameter, in three materials, glass, Diakon, and nickel. Four different distributors were examined. By varying the position of the heating surface at the walls it has been found that there is a significant variation in heat transfer with distance from the distributor. By varying the diameter of the bed it has been found that peak heat transfer coefficients are greater in the smaller diameter bed, but the dependence of heat transfer coefficient upon velocity peaks sharply in the smaller diameter bed with a more gradual change in the larger bed. The experimental measurements have been expressed in the form of a correlation of Nusselt upon Archimedes group, and upon a thermal capacity group. The dependence of Nusselt upon the thermal capacity group is much greater than previously reported.

INTRODUCTION

THERE HAVE been several experimental investigations into heat transfer from surfaces to gas-fluidized beds, but few have covered a wide range of variables. Correlations that have been proposed to describe the experiments have included only a small number of the significant variables, and describe only a small fraction of the published experimentation. The fundamental difficulty is the large number of physical variables that might be expected to affect the rate of heat transfer. The rate of heat transfer from a vertical surface to circulating solids might be expected to depend upon the temperature difference between the surface and solids in the immediate neighborhood, upon the position of the surface in the bed, upon gas velocity, density, heat capacity and thermal conductivity, upon particle size, shape, density, heat capacity, thermal conductivity and particle size distribution, upon bed depth, diameter, and shape if not cylindrical, and upon distributor type and geometrical characteristics.

Even if we consider only solids that can be readily fluidized it is not perhaps surprising that a clear picture of the effect of even a few of the variables has not emerged although some aspects of the general behavior have become clear. Thus it is generally found that above the point of incipient fluidization the rate of heat transfer increases with gas velocity to a maximum value, followed by a decrease in heat transfer as the gas velocity is further increased. The maximum rate of heat transfer is the most frequently quoted result of experimental investigations and the most frequent object of correlation, with a smaller emphasis on the gas velocity at which the peak rate occurs.

Early investigators such as Dow and Jakob [1], Leva *et al.* [2], and Levenspiel and Walton [3], fluidized a variety of solids in beds of small diameter, and reached a variety of conclusions. Dow and Jakob presented a dimensionless correlation that showed a relatively large increase of heat transfer with diameter of bed from experiments in a 2″ and 3″ dia. tube, and a strong inverse correlation with bed depth. Leva *et al.*, who examined beds of 2″ and 4″ dia., reported that the concentration of solids, density of solids and bed diameter had no effect upon heat transfer, and the core of the fluidized bed was isothermal. Levenspiel and Walton reported that heat transfer increased with solid density, and with a reduction in particle size. Baerg *et al.* [4] reported the occurrence of a maximum in the rate of heat transfer.

Although the earlier workers examined heat transfer from bed walls to solids in small diameter beds, later workers examined heat transfer from electrically heated coils to fluidized particles. Kharchenko and Makhorin [5] studied heat transfer between a copper calorimeter and fluidized quartz. They concluded that local heat transfer was greater in the middle of the bed than near the walls or distributor. Lefroy [6] carried out similar experiments. Kim *et al.* [7], Wunder and Mersmann [8] and Yamazaki and Jimbo [9], have

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NOMENCLATURE

Ar	Archimedes group, $d^3\rho(\rho_s-\rho)g/\mu^2$
c	gas specific heat capacity [kJ kg K^{-1}]
c_s	solid specific heat capacity [kJ kg K^{-1}]
d	particle diameter [m]
D	bed diameter [m]
D_s	bed reference diameter, 0.09 m
g	acceleration due to gravity [m s^{-2}]
h_m	mean surface heat transfer coefficient [W m^{-2} K^{-1}]
L	bed depth [m]
Nu_m	mean surface Nusselt number, $h_m d/\lambda$
Rc_s	thermal capacity group, $\rho_s c_s/\rho c$
U	gas superficial velocity [m s^{-1}]
U_{mf}	gas superficial velocity at minimum fluidization [m s^{-1}].

Greek symbols

λ	fluid thermal conductivity [W m K^{-1}]
ρ	fluid density [kg m^{-3}]
ρ_s	solid density [kg m^{-3}]
μ	fluid viscosity [kg m^{-1} s^{-1}].

also measured heat transfer to immersed vertical surfaces.

Of the significant variables, the only clear-cut experimental dependence to be established is that of particle diameter, where the consensus is that the peak rate of heat transfer increases as the particle diameter is reduced. Experimental support for this finding is given amongst others by Ozkaynak and Chen [10], Zhang and Ouyang [11], and Levenspiel and Walton [3], with little evidence of the effect of other properties.

A study of reviews on heat transfer to fluidized systems does little to dispel the impression of fragmented understanding. Thus neither Botterill [12] nor Ainshtein and Gelperin [13], were able to recommend correlations of wide application. Ainshtein and Gelperin comment upon the lack of understanding of heat transfer mechanisms at the time of their review, although later reviews such as that of Botterill describe theories based upon transitory contact of pockets of particles at the heating surface, while Denloye and Botterill [14] distinguish between particle convective and gas convective modes of heat transfer in their correlating equation.

In spite of the development of the understanding of mechanisms of heat transfer to surfaces within fluidized beds, the effect of many factors such as bed diameter, distributor type, particle density and heat capacity, are as yet uncertain. Normal experimental variability, differences in particles, the difficulty of comparing experimental results from different experimental arrangements including distributors, are major reasons for the difficulty in establishing the dependence of fluidized bed heat transfer upon many of the significant variables.

In this paper we describe an experimental study that concentrates upon some aspects of fluidized bed heat transfer that are not clear. Thus we examine the change of heat transfer at the wall in beds of different diameter in experiments that allow heat transfer coefficients to be measured at different vertical positions at the wall. These experimental results are then compared with measurements of heat transfer to small immersed vertical surfaces in beds of different diameter with different distributors. This study has been carried out because the majority of experimental studies have been performed with the immersed surfaces of heated probes rather than heated walls. The experiments have been carried out with regular isometric particles over a size range from 100 μm to 1 mm, over a range in particle density from 1100 kg m^{-3} to 8900 kg m^{-3} and for a range of thermal capacities. The effect of four different distributors was studied.

EXPERIMENTAL EQUIPMENT AND PROCEDURES

Fluidization was carried out in beds of 90 mm and 290 mm cylinder dia.

Two 290 mm dia. beds were used. The jacketted bed was 600 mm deep and was fitted with a perforated plate distributor of 0.34% free area. Five annular section chests were folded around the bed each provided with a steam feed from a 100 psig main reduced to 8 psig and condensate drain to each annulus. This arrangement gave five annular sections, each 120 mm high, that extended from the base of the bed to the upper surface. The arrangement is shown in Fig. 1. The gas temperature leaving the bed was measured by means of concentric arrays of thermocouples at four different radial positions with cold junctions in the air inlet, so that radial temperature gradients in the exit air could be measured. The gas flowrate to the bed was measured by rotameters, and the pressure loss across the distributor was measured by the manometer arrangement illustrated in Fig. 1.

The bed of 90 mm dia. was also jacketted, and was provided as two bed depths of 95 mm and 192 mm. The particles were fluidized within an inner aluminum cylinder of 90 mm dia. that was surrounded by a steam jacket to which steam was fed from a low pressure boiler fired by gas. The exit temperature of the air was measured by a central and three annular sets of thermocouples with cold junctions embedded in the air entering the bed. The gas flowrate to the bed was measured by rotameters and pressure tappings that allowed measurements of pressure losses across both the distributor and the bed.

The second 290 mm dia. bed consisted of a flanged length of clear Perspex with a detachable distributor.

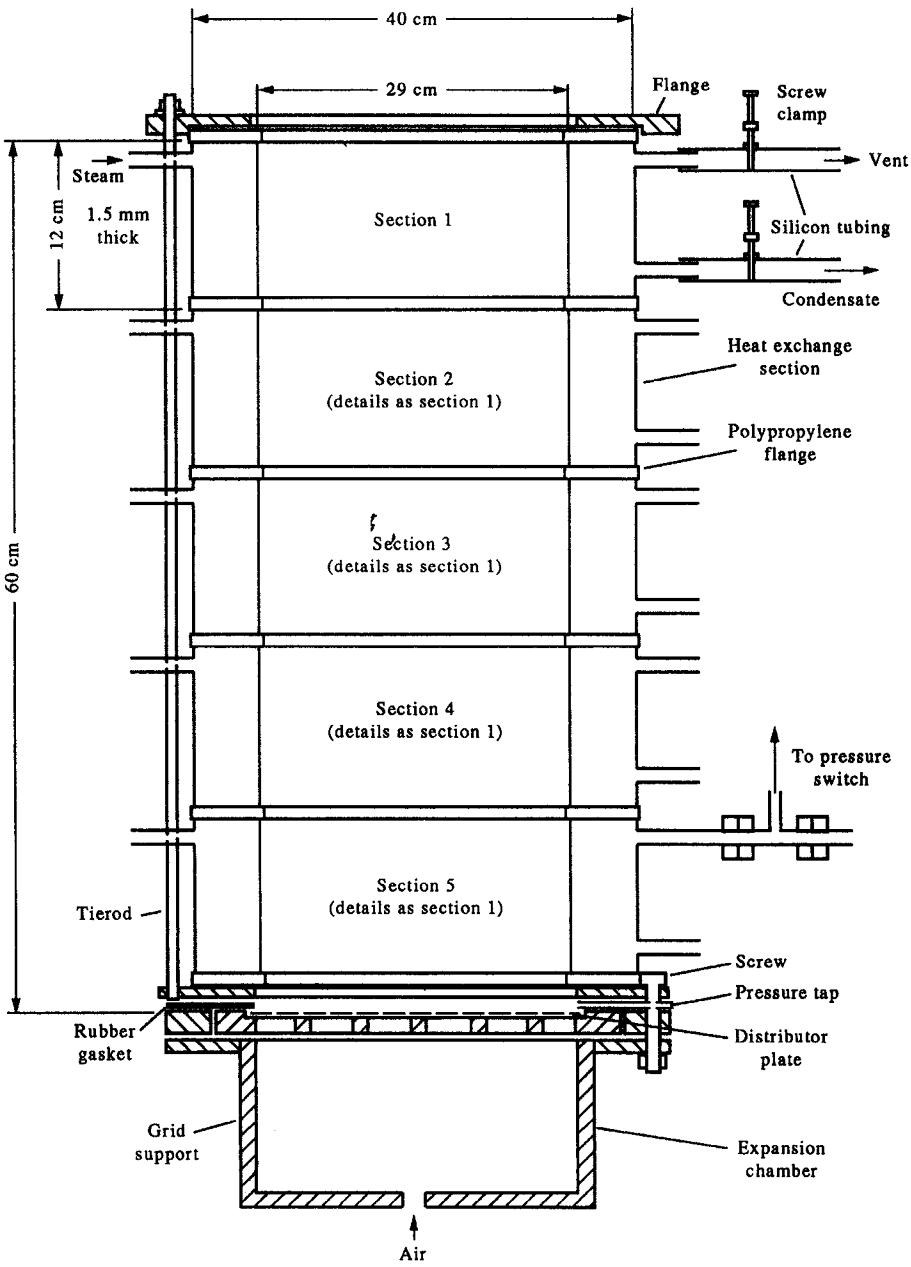

FIG. 1. Arrangement of equipment.

Four distributors were used, a porous plastic Vyon distributor, and three perforated plate distributors of the same hole size of 0.8 mm dia., but different hole patterns at 12 mm square pitch, 9 mm square pitch and 7 mm triangular pitch. The heating element in this bed was a tungsten wire wrapped around an insulating former of length 50 mm and 19 mm dia., and attached to a copper tube of the same diameter. There were three coils in all.

The coil heat transfer elements were calibrated by wrapping each in a plastic film and placing in a water bath at known temperatures. The resistance of each coil was measured at a number of temperatures giving calibration of resistance against temperature for each coil.

During the experiments each coil was placed within the bed and in operation each was connected to a resistance bridge so that the resistance could be measured to give the coil temperatures. The voltage across the coil was measured by a digital voltmeter. On measuring the bed temperature at the same time by an array of thermocouples, the temperature difference between coil and bed was found, and the heat transmitted to the bed was calculated as the square of the coil voltage divided by coil resistance. The coefficient of heat transfer from coil to bed was calculated as the

Table 1. Characteristics of distributors

Distributor type	Hole diameter [mm]	Free area %	Pitch
Perforated 1	0.794	0.34	12 mm square
Perforated 2	0.794	0.61	9 mm square
Perforated 3	0.794	1.17	7 mm equilateral triangle
"Vyon" porous plastic	Permeability of 12.35 Darcy and thickness of 17.1 mm		

heat transmitted through unit surface area of the coil wrapped former for unit temperature difference. The spacing between wire centers on adjacent coils was less than one wire dia., and therefore the coil former was considered as a uniformly heated cylindrical surface.

The resistance bridge was activated by switching on the bridge power supply once fluidization conditions had been established, and the coil resistance and voltage were measured after attaining steady state, about 10–15 min after switching on the bridge.

The heat transfer coefficient to the jacketted bed was found as the heat transferred to the air measured from the temperature rise of air passing through the bed divided by the heat transfer area, and by the temperature difference between the air and fluidized particles again measured by thermocouples. Since large sections of vertical wall were heated, the entire bed of fluidized particles was heated to a significant extent, and therefore a period of about 1 h was required to reach steady state after turning on the steam and fluidization had been established.

A similar procedure was used for a set of experimental conditions on each of the beds. The bed was filled with particles of the required grade to the level of a fine mesh at the top of the bed. The bed was fluidized at the minimum flowrate required for fluidization and particles were removed from the bed until there was no accumulation of particles at the retaining mesh. When the gas flowrate was increased, particles accumulated at the mesh and these particles were removed until there was no accumulation, and no heat transfer measurements were made until this condition had been established. Heat transfer measurements were made for the full range of available velocities for each particle grade and set of conditions. The characteristics of the distributors are given in Table 1.

The physical properties of particles are given in Table 2.

EXPERIMENTAL RESULTS

Wall-to-bed heat transfer

We distinguish here between heat transfer to the walls of a fluidized bed, and heat transfer to surfaces immersed in the bulk of the bed at some distances from the walls. Apart from early work on wall-to-bed heat transfer in narrow diameter fluidized beds, the bulk of published studies have been concerned with heat transfer to surfaces immersed in the bulk of the bed. Experiments on wall-to-bed heat transfer described here were carried out in beds of 90 and 290 mm dia., at different bed depths, and with the grades of particles detailed in Table 2. The distributor was the perforated plate of 0.34% of free area, with holes on a square pitch of 12 mm. The bed was filled with the required amount of 500 μm dia. ballotini for free fluidization to a depth of 600 mm, and the thermocouples arrays were immersed to a depth of 100 mm from the surface of the bed. The temperatures registered by the thermocouples were determined from the thermoelectric voltages measured by a digital voltmeter. Under the condition of experimental fluidization no significant radial or axial gradients were registered by the thermocouples when immersed in the bed. The wall heat transfer coefficients were calculated from the gain in enthalpy of the fluidizing air, divided by the temperature difference between the heating surface registered by a thermocouple attached to the surface and the temperature within the bed, and by the heat transfer area.

Table 2. Physical properties of particles

Solid material	Mean size [μm]	Size distribution [μm]	Density [kg m^{-3}]	Heat capacity [kJ kg^{-1} K^{-1}]
Glass ballotini	100	80–115	2950	0.69
	275	230–320	2950	0.69
	325	290–340	2950	0.69
	500	480–520	2950	0.69
	1000	850–1230	2950	0.69
Diakon	325	130–460	1228	1.46
Nickel shot	275	210–320	8900	0.46
	325	260–380	8900	0.46

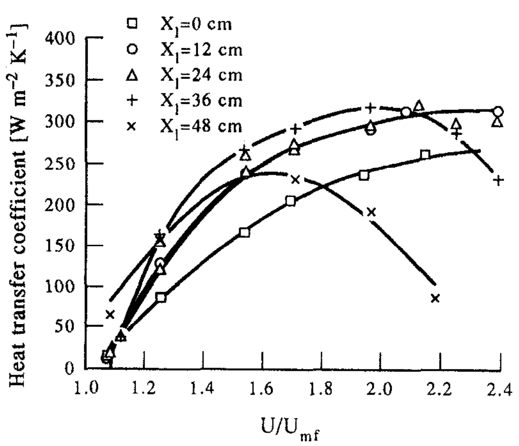

FIG. 2. Variation of wall heat transfer coefficient with position at the wall in bed of 290 mm dia. (glass ballotini of 500 μm; X_l is distance of heated surface from distributor).

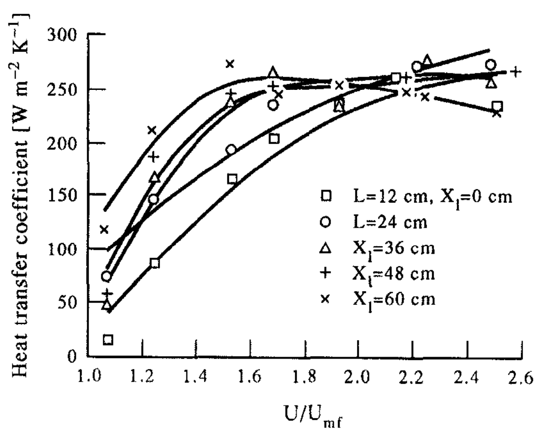

FIG. 3. Variation of wall heat transfer coefficient with height of heat transfer surface, l (glass ballotini of 500 μm).

The experimental dependence of heat transfer coefficient upon the depth of the heating surface at the wall is shown in Fig. 2 as a function of the ratio of U/U_{mf} where U_{mf} is the velocity at incipient fluidization. The height shown on the legend corresponds to the base of a single section of height 120 mm. The section nearest to the base of the distributor showed the smallest heat transfer coefficient, although the heat transfer coefficient steadily improved as the ratio U/U_{mf} was increased. At smaller values of U/U_{mf} the largest heat transfer coefficients were given by the two sections at the top of the bed, but as U/U_{mf} was increased, a maximum value followed by a rapid decrease was shown when heat was transferred from the uppermost section only, while the section below the uppermost showed a smaller fall at higher values of U/U_{mf}. Heat transfer coefficients at the lowest bed levels showed no apparent decrease over the range of flow rates investigated. We consider the most probable reason for the relatively sharp decrease of heat transfer at the upper levels in the bed is the enhanced oscillations in bed level reducing the utilization of heat transfer surface with increasing gas velocity. The most probable reason for lower heat transfer coefficients at the lowest section is the small velocity of solids circulation at the base of the wall.

In a second related series of experiments the height of the heat transfer surface at the wall was increased by admitting steam to the lowest section, lowest two sections, and so on until all five sections were heated. The experimental measurements of heat transfer coefficient are shown in Fig. 3, and consistent with the previous figure, the heat transfer coefficient is the lowest at the base of the wall. The effect of level fluctuation at the upper surface of the bed is also marked, but the reduction above the peak heat transfer coefficients is smaller, because the heat transfer coefficients are mean values for the whole of the wall region, while coefficients at the upper level in the previous figure represent the top section only.

The effect of bed height and diameter upon wall heat transfer

The effect of ratio of bed height to bed diameter was investigated in the 90 mm dia. bed using glass ballotini particles of 100, 275, 325, 500 and 1000 μm, and nickel shot particles of 275 and 325 μm dia. Two bed depths were examined 95 mm and 192 mm, and the results of a typical experiment are shown in Fig. 4 for 325 μm ballotini. The figure shows no effect of bed depth, but there is clearly a maximum in wall heat transfer at $U/U_{mf} = 1.75$. For other particle sizes examined in this bed, the values of U/U_{mf} at which the maximum was found increased as the particle size was reduced. This point is illustrated in Fig. 5 that shows the dependence of heat transfer coefficient upon U/U_{mf} for 100 μm and 500 μm glass ballotini in the 90 mm depth. The effect of U/U_{mf} upon h_{max} is very strong.

Experimental measurements of wall heat transfer for the 290 mm bed are shown in Fig. 6 for five different ratios of height to diameter. Again there was no effect of the ratio upon the experimental measurements. However an important difference is that there is no clear cut maximum in heat transfer coefficient.

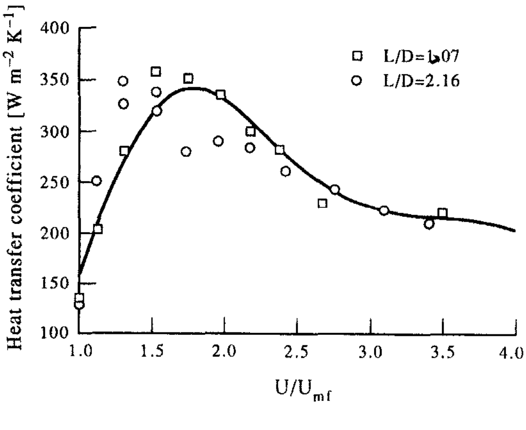

FIG. 4. Effect of L/D ratio on wall heat transfer coefficient for glass ballotini of 325 μm in bed of 90 mm dia.

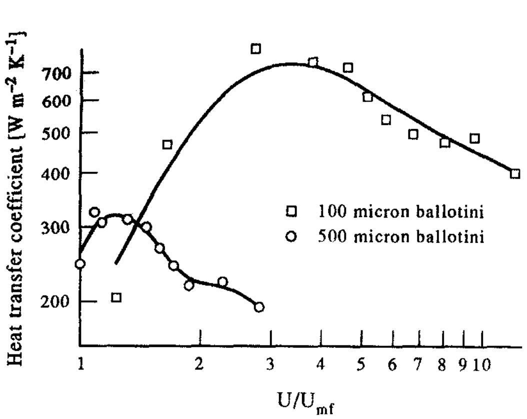

FIG. 5. Effect of particle size on wall heat transfer coefficient in bed of 90 mm dia.

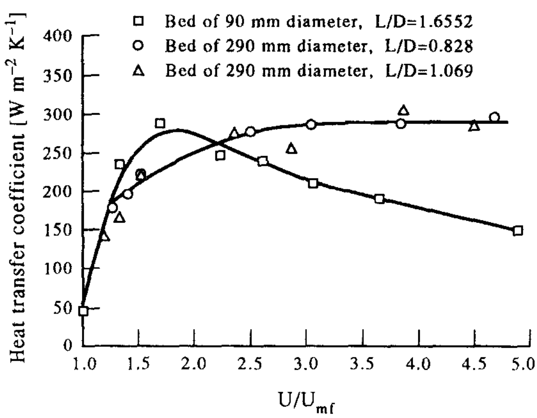

FIG. 7. Effect of bed diameter on wall heat transfer coefficient for Diakon particles of 325 μm.

This dependence was similar for all particles in that no maximum in heat transfer coefficient was found in the 290 mm dia. bed with the heat transfer coefficient always increasing with U/U_{mf} until a steady value was attained.

The characteristics of heat transfer to beds of different diameters when fluidizing the same particles were examined for 100 μm and 500 μm ballotini and 325 μm Diakon particles in beds of 90 mm and 290 mm dia. A typical comparison is shown in Fig. 7, where the particles are 325 μm Diakon. The effect of bed diameter is considerable. Here, as in all other experiments in the 290 mm bed, the heat transfer coefficient increased beyond the point of incipient fluidization until a maximum value was attained, that then remained constant for further increases in gas velocity. In the 90 mm bed on the other hand, the maximum heat transfer coefficient was attained at much smaller ratios of U/U_{mf}, but as the velocity was further increased a significant reduction in the heat transfer coefficient was found. The same pattern was shown for all three particle grades. The peak heat transfer coefficient was always smaller in the larger bed, but the maximum value was maintained as the velocity was increased. The differences between the velocities at which peak heat transfer occurred in the small and in the large diameter beds were considerable, but no decline from the maximum value was observed in the 290 mm bed.

The quantity of solids in the beds was measured by weighing, and the dependence of the ratio bed density to bed density at incipient fluidization upon the ratio of U/U_{mf} is shown in Fig. 8 for 325 μm Diakon in 90 mm and 290 mm dia. beds. For much of the range shown in this figure, heat transfer to the 290 mm bed was significantly greater than the 90 mm bed even though bed density in the larger bed was consistently smaller.

These figures illustrate the complex effect of bed diameter, as over most of the experimental range of velocity heat transfer in the larger bed was significantly greater than the smaller bed, but, at the peak, heat transfer in the smaller bed was greater. Based upon observations of the bed during the experiments, and upon the observations on slugging of Stewart and Davidson [15], we consider that the principal

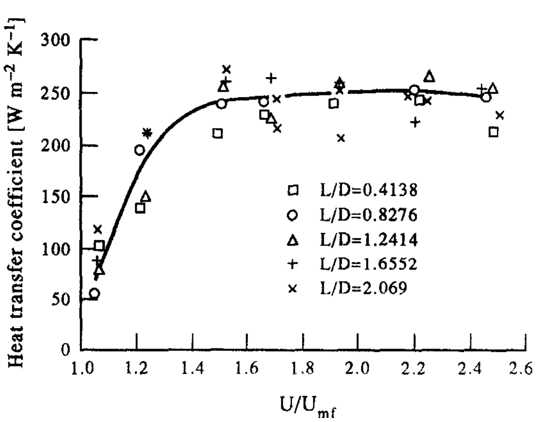

FIG. 6. Effect of L/D ratio on wall heat transfer coefficient for glass ballotini of 500 μm in bed of 290 mm dia.

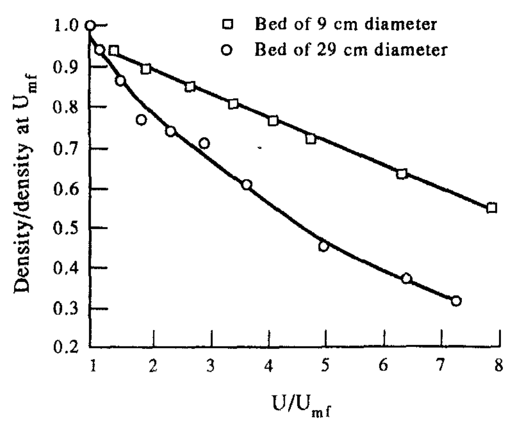

FIG. 8. Effect of bed diameter on the ratio bed density to bed densiy at incipient fluidization.

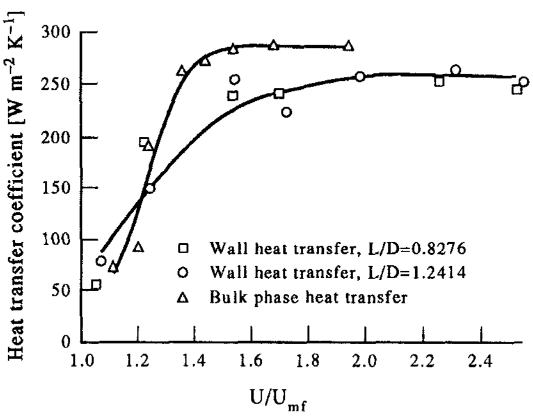

FIG. 9. Comparison between bulk-phase and wall heat transfer coefficients for glass ballotini 500 μm, in a bed of 290 mm dia. using perforated plate distributor of 12 mm pitch.

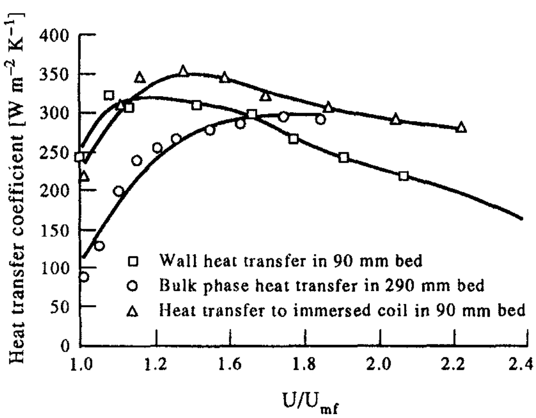

FIG. 10. Comparison between bulk-phase and wall heat transfer coefficient for glass ballotini 500 μm, in a bed of 90 mm and 290 mm dia., using porous plastic distributor.

reason for the sharp reduction in heat transfer in the smaller diameter bed was the onset of slugging in changing the pattern of solids circulation and reducing bubble velocity. In the 290 mm bed on the other hand, the intensity of particle movement increased with U/U_{mf}, and, even though solids hold up was reduced, heat transfer to the bed was maintained over the experimental range of gas flowrates once the peak had been attained.

Bulk phase heat transfer

In the next set of experiments, heat transfer to immersed coils was measured in the 290 mm dia. bed for four different distributors, and in the 90 mm bed fitted with the porous plastic distributor. The coils were supported away from the walls and immersed in the bed. Three coils were placed in the 290 mm bed, 35 mm from the axis of the bed and 150 mm above the distributor, 65 mm from the axis of the bed and 240 mm above the distributor, and 95 mm from the axis of the bed and 360 mm above the distributor; no significant differences in heat transfer coefficient were found at the locations measured by the coils.

Bulk phase and wall heat transfer

Figure 9 shows the comparison of mean wall and bulk phase heat transfer for the 290 mm bed when employing the same perforated plate distributor of 12 mm square pitch and fluidizing 500 μm ballotini. At low ratios of U/U_{mf} heat transfer to the bulk phase was greater, but coefficients approached the same values as U/U_{mf} was increased, although bulk phase values remained greater because the coils were always immersed. It is most probable that the lower heat transfer coefficients at the wall are due to low circulating velocities of solids at the base of the wall, combined with reduced wall coverage due to bed oscillations.

Similar dependences were found for the other solids, 325 μm Diakon and 100 μm glass ballotini.

The comparison of bulk phase and wall heat transfer when fluidizing 500 μm ballotini in the 90 mm dia. bed is shown in Fig. 10. Again, bulk phase heat transfer was greater, but the rate of decrease beyond the peak rate was significantly greater for wall heat transfer. We consider the differences in declination of the heat transfer coefficients above the peak to be due to the physical effect of the probe upon the pattern of fluidization. The probe was 22 mm dia. in a bed of 90 mm dia., and 54 mm long in a bed depth of 192 mm so that the development of gas slugs considered to be responsible for reduction in heat transfer beyond the peak, was interrupted on being pierced by the probe. To illustrate this point we have included bulk phase measurements for the 290 mm dia. bed showing that the bulk phase heat transfer coefficients for both beds become similar at higher values of the ratio U/U_{mf}.

The effect of distributor

The four types of distributor described in Table 2 were each placed in the 290 mm dia. bed and bulk phase heat transfer measurements for 500 μm ballotini are shown in Fig. 11. The differences between the distributors mainly take effect at low ratio of U/U_{mf} where the order of effectiveness in promoting heat transfer is, porous plate >7 mm hole pitch >9 mm hole pitch >12 mm hole pitch. Once the peak heat transfer coefficient was attained, the differences between the distributors had no apparent effect upon heat transfer. A similar pattern was found for both 100 μm ballotini and 325 μm Diakon, although the pattern of dependence for the 100 μm ballotini was not so clear cut as that shown in Fig. 11. Apparently for the isometric solids examined in this investigation the effect of distributor upon heat transfer is second order provided that a minimum standard of distributor effectiveness is maintained.

DISCUSSION

As we have pointed out the differences between bulk phase and wall heat transfer have not been sys-

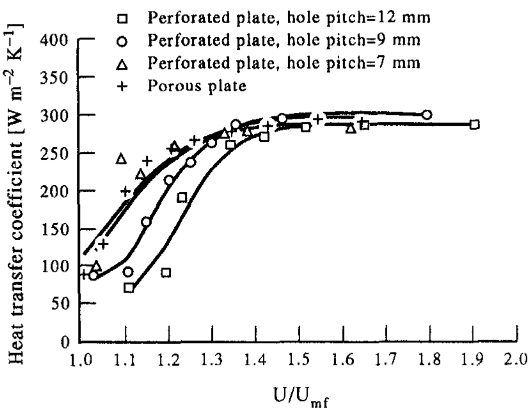

FIG. 11. The effect of distributor type and geometry on bulk-phase heat transfer coefficient for 500 μm glass ballotini in bed of 290 mm dia.

tematically examined by earlier workers. However, our comparison has shown that differences between bulk heat transfer measured away from the walls, and heat transfer at the walls are of the order 10% greater for bulk heat transfer, with lower heat transfer at the walls due to low circulation velocities of solids at the base and reduced particle contact at the upper bed surfaces because of oscillations in bed level. We did not examine differences in wall heat transfer due to the gas distributor, but we are confident that the differences are reflected by the relatively small effect of distributor shown in the experiments on bulk heat transfer.

The effect of bed diameter is more important, and, although attention in the literature has been concentrated on the maximum heat transfer coefficient, we believe that the differences in heat transfer caused by bed diameter are significant, because the sharp decline in the heat transfer coefficient above the peak experienced in small diameter beds was not found in the 290 mm dia. bed. We have concluded from our experimental programme that the study of peak heat transfer alone is likely to be only a partial guide in applications of heat transfer, and that the occurrence of slugging conditions in gas-fluidized beds will cause a reduction in heat transfer that will not be predicted in studies divorced from the geometry of the fluidized system.

With these reservations in mind we have expressed our results in a form that includes the physical quantities that have been varied in our experiments expressed with reference to the maximum heat transfer coefficient. We have preferred a fully dimensionless correlation because limitations on the correlation can be expressed particularly in reference to the Archimedes number, a most important variable in characterising fluidization.

For the entire set of experiments, non-linear least squares estimation was used to find the parameters in the following correlation:

$$Nu_m = 0.00515 Ar^{0.184} Rc_s^{0.6} (D/D_s)^{-0.12},$$
$$100 < Ar < 10^5, \quad SE = 12.8\%, \qquad (1)$$

where the Nusselt group is $Nu_m = h_m d/\lambda$, the Archimedes group is $Ar = d^3\rho(\rho_s - \rho)g/\mu^2$ the thermal capacity group is $Rc_s = \rho_s c_s/\rho c$, and the diameter group D/D_s is defined as D the bed diameter with D_s the reference diameter of 90 mm.

A graph of the dependence of the Nusselt upon the Archimedes group at parameters of the group Rc_s is shown in Fig. 12 with the diameter group incorporated into the Nusselt group as $Nu_m(D/D_s)^{0.12}$. Both the graph and equation (1) illustrate the importance of the thermal capacity group, a matter of interest since previous correlations follow the form of dependence of the Nusselt upon the Archimedes group, but none give such prominence to the group Rc_s [12, 13].

It is clearly important that the correlation should be compared with the experimental results of other workers. However, a comparison with the results of the experimentalists listed in the bibliography and of others, has shown that the effect of slugging on heat transfer has not been appreciated. The particular feature of this type of heat transfer that invalidates a general comparison is that once slugging sets in there is a marked reduction in heat transfer. A correlation, dimensionless or otherwise, that is not concerned with such a distinction will exhibit an inordinate scatter that will hide important relationships in heat transfer.

It is also clear that the correlation of equation (1) will not hold above a value of Archimedes number in the range 10^5–10^6. The reason is that above this Archimedes number, with turbulence established in the fluid and in the bed, heat transfer by convection enhanced by turbulence becomes the dominant mode, and the contribution of convection driven by circulating streams of solid is relatively reduced. This distinction, in a different form, has been made by Denloye and Botterill [14], although we consider that the implementation of this distinction as separate heat transfer coefficients is not a clear way of interpreting experimental measurements. The distinction is not expressed in equation (1), and, as we acknowledge its importance, some modification to this equation is necessary if heat transfer is to be correlated over a wider range of Archimedes number.

The principal difference between our experimental results and those of others in this range of Archimedes number is the importance of the group $Rc_s = \rho_s c_s/\rho c$. Other workers have considered this effect, but normally as the separate effect of density in particular, and heat capacity. For example Ozkaynak and Chen [10] examined the effect of particle density on heat transfer by choosing glass beads of two different densities, 2470 and 4490 kg m^{-3}. They found that density had no effect on heat transfer, an unlikely finding in isolation. However, the heat capacities of the two grades of particles were 0.75 and 0.44 kJ kg^{-1} K^{-1}; again different, but the products $\rho_s c_s$ are 1863 and 1967 kJ m^{-3} K^{-1}. Clearly the correct interpretation

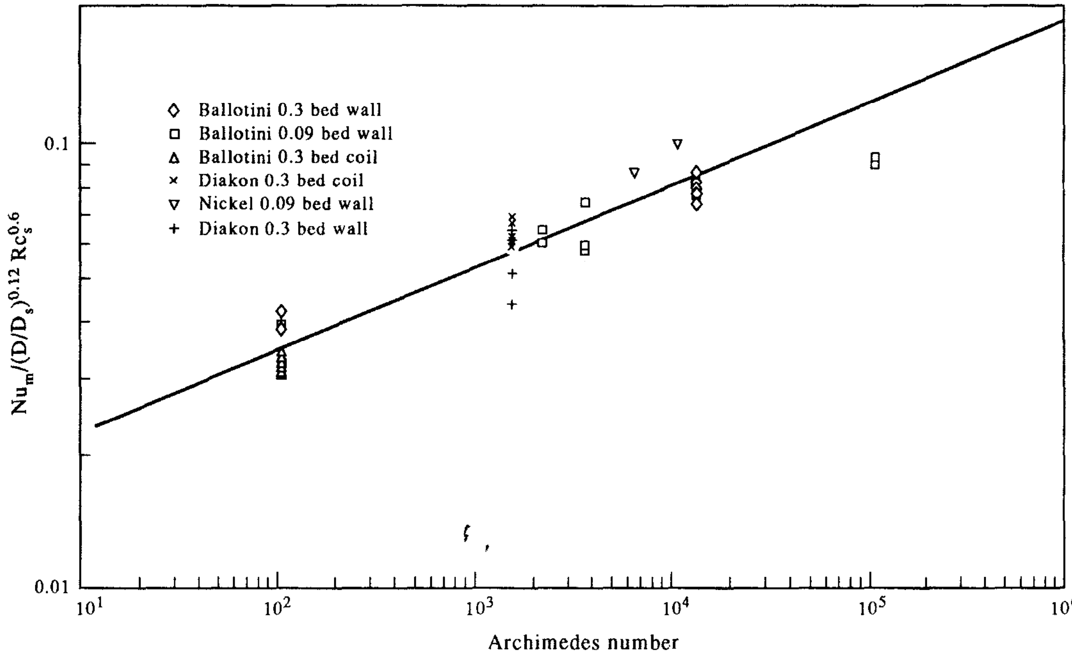

FIG. 12. The dependence of $Nu_m/(D/D_s)^{0.12}$ $Rc_s^{0.6}$ upon the Archimedes group.

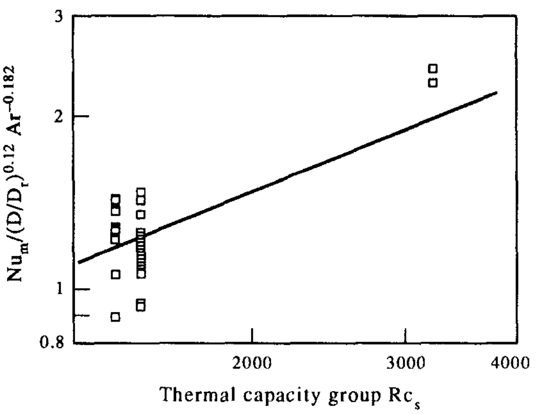

FIG. 13. The dependence of $Nu_m/(D/D_r)^{0.12}$ $Ar^{-0.182}$ upon Rc_s group.

of this experiment is that particles of similar product $\rho_s c_s$ give similar heat transfer coefficients.

The dependence of Nusselt group upon the group Rc_s is shown in Fig. 13 in which equation (1) has been arranged with the dependence of $Nu_m(D/D_r)^{0.12}Ar^{-0.182}$ shown as a function of the thermal capacity group Rc_s. It appears that the experimental evidence for the dependence of the Nusselt group upon Rc_s is not as well-defined as the dependence upon Archimedes number, a question we will consider further in relating the work described in this paper to other studies published in the literature.

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