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Heat transfer from vertical inserts in gas-fluidized beds

D. J. GUNN and N. HILAL
Department of Chemical Engineering, University of Wales Swansea, Swansea SA2 8PP, U.K.

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Abstract—For a chemical reaction of large exothermic or endothermicity when carried out in fluidized beds, an internal heat transfer surface additional to the walls of the bed is often necessary. The heat transfer characteristics of vertical tubular inserts placed in the fluidized bed were examined by measuring the heat transfer from coils integral in the inserts to the fluidized bed. It was found that the heat transfer coefficients measured at the coils corresponded closely to the heat transfer coefficient for bulk heat transfer measured in open fluidized beds under non-slugging conditions. Variations in heat transfer over the array of inserts were small, and distributor effects were significant only at the onset of fluidization. The experimental results were successfully related to an earlier correlation given for heat transfer in open beds. Copyright © 1996 Elsevier Science Ltd.

INTRODUCTION

In a recent investigation into heat transfer to beds of solids fluidized by gas, the principal concern was to characterize the heat transfer to vertical surfaces in open fluidized beds [1]. In that study the effect of bed diameter, bed depth, location at the wall and within the bed, particle diameter and material, and types of distributor were examined. The range of Archimedes number covered in the investigation was 10^2–10^5, a range in which bubbles and solid circulations generated by bubbles strongly influence heat transfer.

In open beds of small diameter, the bubbles quickly grow until constrained by the surrounding walls, and in this condition of slugging fluidization, the consequent reduction in bubble velocity and particle velocity significantly reduces heat transfer. In open beds of larger diameter, on the other hand, the development of large bubble flows may have an adverse effect upon some applications reducing heat transfer in fluidized beds. To alleviate the difficulties of fluidization in large open beds without encountering heat transfer difficulties due to slugging, the use of baffles or other mechanical devices has been considered.

The two principal reasons for the use of baffles and other mechanical devices are connected with the influence of bubbles. Particles entrained in the wake of large bubbles can cause severe mechanical disturbance of the fluidized bed and ancillary equipment, possibly damaging internal fixtures. The addition of vertical plates or horizontal or vertical wire mesh restricts the size of bubbles that can be formed and the mechanical disturbance is thereby lessened.

A second reason for the use of baffles may be found in this consequence of developed bubble flow in fluidized beds of fine particles. In wide and deep fluidized beds the bubbles coalesce, grow in size, and channel up the middle of the fluidized bed. This effect is sensitive to the diameter of the vessel because the walls hinder bubble development and the concentration of gas flow in the centre is less severe in narrow beds than in wide ones. Because of large bubble size the reduction in gas–solids contacting in wide and deep reactors may be very serious. The effect of bed diameter upon gas-phase mixing has been reported by May [2] and de Groot [3]. The use of baffles to minimize bubble growth may therefore also have a beneficial effect upon gas–solid contacting.

The published reports of the effect of baffles upon fluidized-bed reactors are conflicting. Volk *et al.* [4] report that baffles could provide defluidized regions in the fluidized bed, and required care in design if adequate radial mixing of gas and solids was to be obtained. Romero and Johanson [5] report a considerable improvement in the smoothness of fluidization by the use of baffles in a slugging fluidized bed. Overcashier *et al.* [6] found a narrowing of the reaction-time spectrum for the air fluidization of a microspheroidal catalyst in a slugging fluidized bed. Agrawal and Davies [7] reported an increase in the smoothness of a bed fluidizing coarse particles, but observed that for some applications the introduction of screens was a disadvantage, particulary for sticky particles and for reactions of high thermicity.

The experimental emphasis in studies on the effect of baffles has been mainly directed to horizontal tubes of various types. Thus Jodra *et al.* [8] studied the effect of several horizontal tube screens and found significant increases in the conversion of butanol and ammonia to form butyronitrile in a fluidized bed. In a later study Jodra and Aragon [9] proposed a correlation for bubble size distribution formed after the passage of gas through a row of baffles, while Yates and Ruiz-Matinez [10] studied the interaction between

NOMENCLATURE

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

horizontal tubes and gas bubbles in a fluidized bed. More recently, Dutta and Suciu [11] studied the effect of several different perforated horizontal baffles in reducing bubble size and found that all of the 29 baffles studied were effective, but to varying degrees. However, the introduction of complete horizontal baffles, and to a smaller extent, the introduction of horizontal tubes over the cross-section of the bed also reduces the mobility of streams of fluidized solids in the bed, so enhancing the probability of defluidization and reduction of heat transfer.

Agrawal and Davies [7] and Volk *et al.* [4] investigated the reduction of iron oxides by hydrogen. Agrawal and Davies fluidized a coarse particle size distribution converting the hydrogen without difficulty, and found no significant increase in conversion when baffles were inserted: the bubble velocity under these conditions was slower than the interstitial gas velocity. Volk *et al.* fluidized finer material and had considerable trouble with the conversion of hydrogen in the reactor, because of the high relative velocity of bubbles and eventually suspended vertical rods in the reactor. The rods of a circular or semi circular cross-section were suspended in a closely spaced array.

A great improvement in the smoothness of fluidization was found, and there was some improvement in the conversion of hydrogen. Volk *et al.* concluded that the reactor could be characterized by an equivalent diameter of 4 × flow area/(flow parameter of rods + reactor). They were able to operate a fluidized bed of diameter 23 in, with vertical inserts of equivalent diameter in the range 4–8 in and found the same conversion as later given by a 5'6" diameter reactor equipped with vertical inserts to give the same equivalent diameter as the smaller reactor.

For some special applications baffles can be arranged to give countercurrent flow of solids and gas during fluidization [6, 12]. But if such special applications are not of importance it appears that the suspension of vertical rods in a fluidized bed gives better fluidization than mesh or vertical plate baffles, because the latter arrangements require more careful design and are prone to cause defluidization. In addition, the mixing of solids and gas is impeded by baffles of large area, while vertical rods do not suppress mixing to the same extent and therefore, the serious consequences of defluidization for heat transfer are avoided.

For reactions that are significantly exothermic or endothermic, removal or addition of heat at the walls of the fluidized bed is less effective as the diameter of the bed is increased and therefore, internal heat exchangers become necessary in such circumstances. The main choices for heat transfer are either a tubular exchanger with horizontal tubes in the bed or vertical tubes. Heat transfer studies on internal heat exchangers are devoted entirely to heat transfer between fluidized beds and horizontal tubes or arrays of horizontal tubes [13–18]. Yet it is apparent that exchangers with horizontal tubes in fluidized beds may easily cause defluidization and impede the lateral mixing of solids and gas by unavoidable changes in the pattern of gas and solid flow within the bed. Vertical inserts, on the other hand, constitute an effective heat transfer surface by arranging that each insert is an annular pipe with down flow in the core, while retaining the benefits of improved fluidization as just described.

In this paper we describe an experimental investigation into heat transfer to vertical inserts in a gas-fluidized bed of diameter 0.29 m, and link the experimental results to a companion investigation into heat transfer to gas-fluidized beds of the same diameter and the same particles [1]. A comparison with heat transfer under conditions of slugging flow has not been made, because the earlier study was devoted to heat transfer under non-slugging conditions.

EXPERIMENTAL EQUIPMENT AND PROCEDURES

Fluidization was carried out in a bed of 290 mm diameter consisting of a flanged length of clear perspex pipe 800 mm deep with a detachable distributor. Three distributors were used, a porous plastic Vyon distributor, and two perforated plate distributors of the same hole size of 0.8 mm diameter, but different densities at 9 mm and 12 mm square pitch.

The inserts were fabricated from copper tubes, 20 mm in external diameter, held vertical as arrays on a

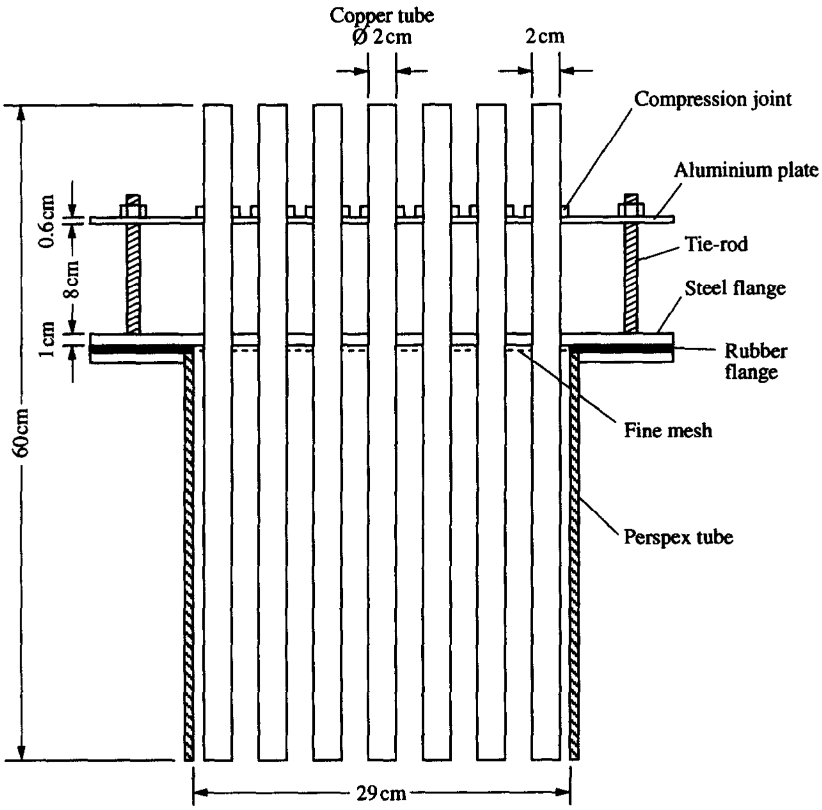

Fig. 1. Diagram of tube bundle in a fluidized bed.

40 mm square pitch. With this arrangement the full set of inserts consisted of 28 tubes held in place by an aluminium plate perforated with a similar number of holes for the flow of gas; the arrangement of the plate and tubes is shown in Figs. 1 and 2. The tubes could be withdrawn and inserted into the bed either as a complete array of 28 tubes or as a partial array.

Heat transfer was measured for three elements in the bed. Each heating element was formed from a tungsten wire wrapped around an insulating former of length 50 mm and 19 mm diameter attached to a copper tube within the array. 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 a calibration of resistance against temperature for each coil.

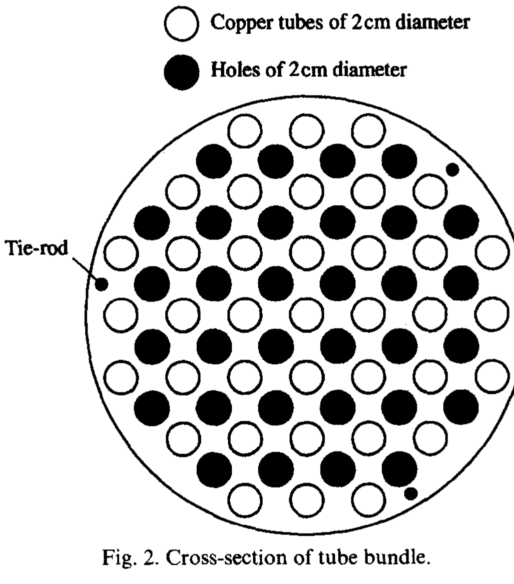

Fig. 2. Cross-section of tube bundle.

The three coils were placed at different positions on three inserts in the fluidized bed. During the operation of the bed, each coil was connected to a resistance bridge with the voltage across the bridge measured by a digital voltmeter. The bridge was balanced at a low bridge voltage and ambient temperature. In the operation, a higher bridge voltage was applied and the bridge output was used to calibrate both the voltage across the coil and the coil resistance. The bed temperature was measured at the same time by an array of thermocouples, so giving the temperature difference between coil and bed. The heat transmitted to the bed was calculated as the square of the coil voltage divided by the coil resistance. The coefficient of heat transfer from coil to bed was calculated as heat transmitted through unit surface area of the coil wrapped former for unit temperature difference. The spacing between centres of adjacent wires on each coil was less than one wire diameter, and therefore the coil former was considered as a uniformly heated cylindrical surface.

Table 1. Physical properties of particles

Material	Mean diameter μm	Size distribution μm	Density $kg\ m^{-3}$	Heat capacity $kJ\ kg^{-1}.K^{-1}$
Glass ballotini	100	80–115	2950	0.69
Glass ballotini	500	480–520	2950	0.69
Diakon	325	130–460	1228	1.46

In the basic experiment, spillage of particles from the bed was avoided by placing a constraining mesh at the upper level of the bed. Heat transfer experiments were carried out first at low velocities, but if the gas velocity was increased without removing particles from the bed, the heat transfer coefficient was found to decrease sharply because of accumulation of particles at the mesh. Therefore, for each experiment a gas velocity was set and particles that accumulated at the mesh were removed. At the condition of no accumulation, the density within the bed was the same with or without the mesh under condition of free fluidization. No heat transfer measurements were made until the correct bed density was obtained with the mesh in place and without particle accumulation. Heat transfer measurements were made with three distributors, and with glass ballotini and Diakon particles of the physical properties shown in Table 1.

EXPERIMENTAL RESULTS

In each experiment the tubes were suspended in the bed so that the bottom of each tube was 80 mm above the distributor. The arrangement is shown in Figs. 1 and 2. Three coils of tungsten wire, as previously described, were attached to three vertical tubes at different locations within the bed, and the effect of changing the axial and radial locations of each coil on heat transfer was studied. The effect of coil location at different gas velocities when fluidizing 100 μm glass ballotini is shown in Fig. 3, where X_R is the distance from the centreline of the bed, and X_L is the distance from the distributor. At a given gas velocity, the difference between heat transfer measurements obtained from each coil is small, and the variation in pattern over the range of velocities is random. The distributor in this case was the perforated plate with 12 mm square pitch perforations. Similar variations were obtained with different distributors when fluidizing 100 μm and 500 μm ballotini and 325 μm Diakon.

In subsequent work, heat transfer coefficients were calculated as the mean value from the three coils for each condition of distributor, particle grade and gas velocity.

The effect of the tube number

The experimental arrangement allowed variations in the number of tubes inserted into the bed.

In the next set of experiments the effect of changing the number of tubes on heat transfer was examined. Four tube patterns were examined, a single tube, three tubes, 15 tubes and 28 tubes. The experimental variation in heat transfer coefficient at different ratios of gas velocity to that at minimum fluidization is shown in Fig. 4, when Diakon was fluidized in the bed with the 12 mm pitch perforated plate distributor. It is apparent that the variation in heat transfer coefficient for the different patterns of tube placement is small and random. It may therefore be concluded that the measured heat transfer coefficients do not depend upon the tube pattern or density, and that the heat transfer coefficients so measured remain the same when a tube bundle is introduced, provided that the gas velocity is referred to the area of flow remaining between the tubes.

The same independence of heat transfer coefficient when the tube arrangement is changed is shown in Fig. 5. In this case only a single tube and a full tube set of 28 tubes were examined when 100 μm ballotini were fluidized using the perforated plate distributor of 12 mm square pitch. Again the measured heat transfer coefficients, functions of U/U_{mf}, were independent of tube arrangement. It is also evident that the heat transfer coefficients measured in the bulk of an open fluidized bed remain unchanged at the same value of U/U_{mf} when tube inserts are placed in the bed.

The same independence of heat transfer coefficient upon the tube arrangement was shown when fluidizing 500 μm ballotini.

The effect of distributor

The third set of experiments was designed to examine the effect of the distributor upon heat transfer for each of the three grades of particles. Figure 6 shows the dependence of heat transfer coefficient when fluidizing 100 μm glass ballotini for each of the three distributors. There is no apparent effect of the distributor, even at low values of U/U_{mf}, since the pattern of dependence upon distributor type is random throughout the range of U/U_{mf}.

The dependence of heat transfer coefficient upon U/U_{mf} for the porous plate and a perforated plate when fluidizing 500 μm ballotini is shown in Fig. 7, where it is evident that throughout the range of U/U_{mf}, the heat transfer coefficient is also independent of distributor type. In this case the vertical inserts have improved the performance of the fluidized bed in heat transfer, since similar experiments in the open fluidized bed showed a significant dependence upon the

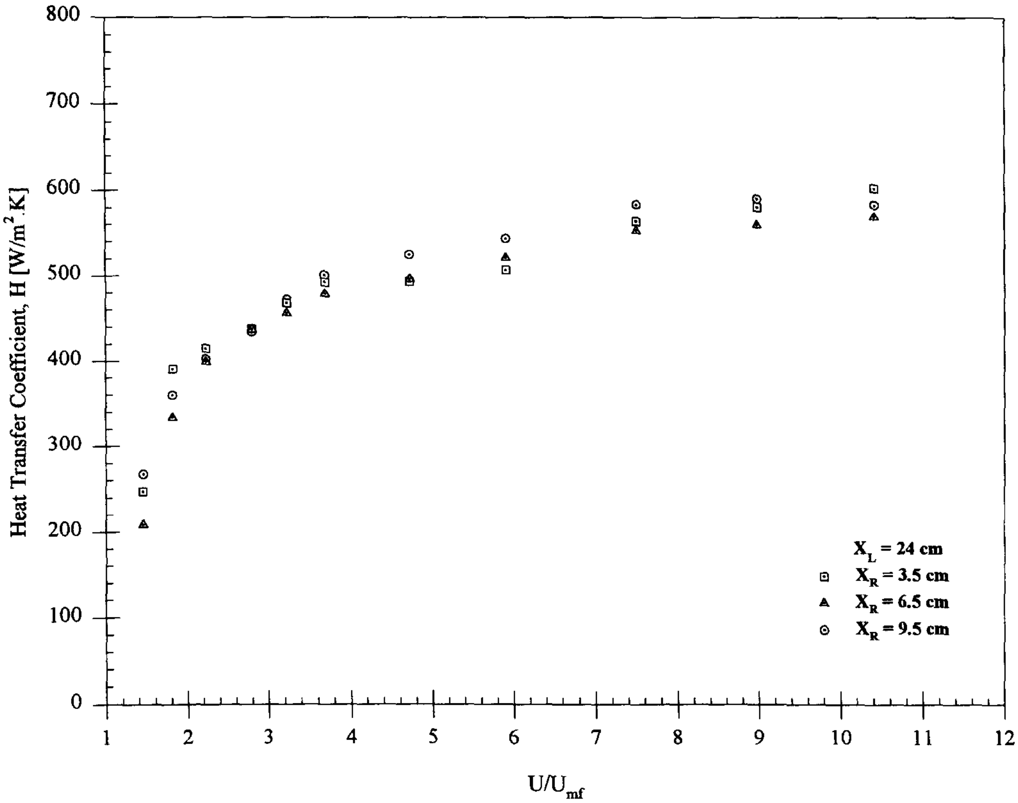

Fig. 3. Effect of radial location of a heat transfer element from 28 vertical inserts, using a perforated distributor of 12 mm square pitch and 100 μm glass ballotini, $X = 24$ cm.

distributor at low U/U_{mf} when fluidizing 500 μm ballotini [1], although a common asymptote was approached at higher values of U/U_{mf}.

The effect of distributor type when fluidizing 325 μm Diakon is shown in Fig. 8. Only for this case is there a dependence of heat transfer coefficient upon distributor type at low values of U/U_{mf}, with the greatest heat transfer coefficients obtained with the porous plate distributor and the smallest for the perforated plate distributor of 12 mm pitch. The same effect of distributor plate was found for the fluidization of 325 μm Diakon in an open bed at low values of U/U_{mf} with a common asymptote at higher values [19].

DISCUSSION

In comparing the experimental results of this paper with the companion study [1], it should be borne in mind that the experimental measurements in the fluidized bed taken with a single tubular insert corresponded to measurements of bulk heat transfer in an open fluidized bed, since the single tube has a negligible effect on bulk heat transfer. It is clearly an important finding in this paper that within the experimental limits of this investigation for non-slugging fluidization, the heat transfer to the vertical arrays of tubes in a fluidized bed corresponds closely to the measurements of bulk heat transfer in an open fluidized bed. Therefore, if there is a design requirement for removal of heat as for an exothermic reaction, or addition of heat as for an endothermic reaction, additional heat transfer area as vertical inserts may be placed in the fluidized bed with heat transfer coefficient corresponding closely to that measured in an open fluidized bed. The additional heat transfer area will be provided under improved conditions of fluidization, since the vertical inserts restrict the size of bubbles and the momentum of particles carried in the wake of bubbles. When the particles are isotropic and less than 0.5 mm in diameter, the bubble velocity in an open bed usually exceeds the interstitial gas velocity, but by the addition of vertical inserts bubble size is constrained so improving contact between gas and solid phase. For larger particles, the phase contact is not improved, since interstitial gas passes through the bubbles, but by reducing bubble size and particle momentum smoother fluidization is obtained.

By comparison with horizontal inserts or heat transfer tubes, the problem of defluidization is reduced and possibly avoided altogether.

As it is a clear finding of these experiments that heat transfer coefficients in open fluidized beds are maintained with improved fluidization, a correlation of heat transfer found for open fluidized beds should be valid. This correlation for open bed bulk heat transfer

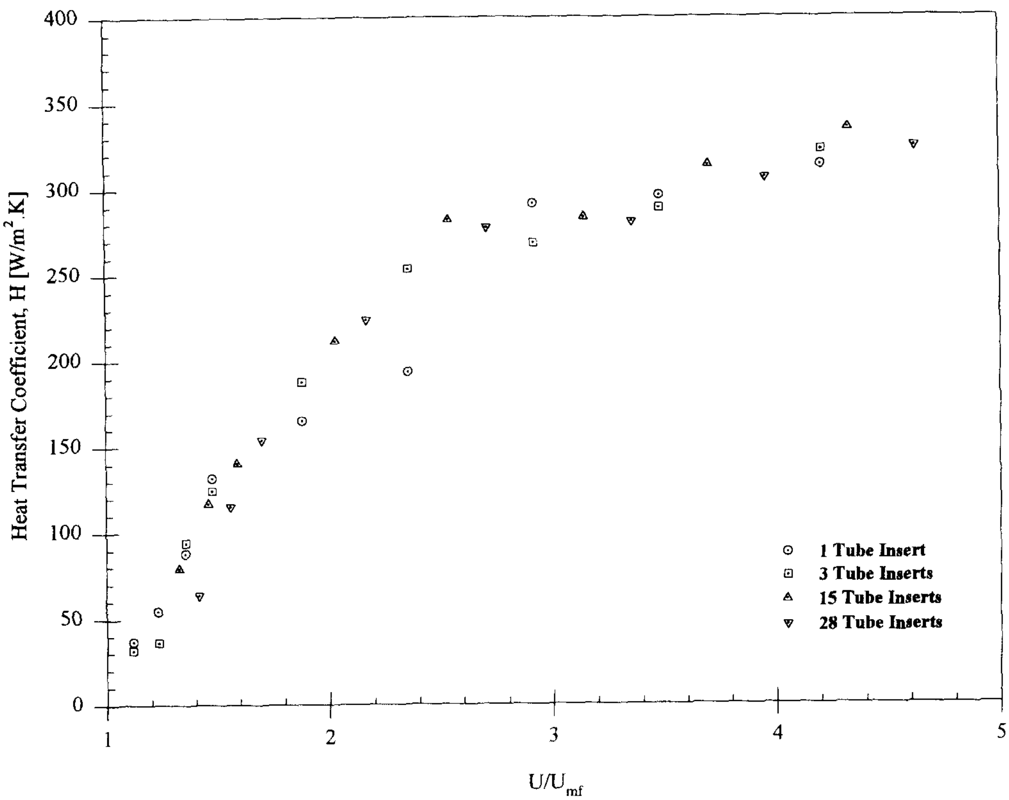

Fig. 4. Effect of the number of tube inserts using a perforated distributor of 12 mm square pitch and 325 μm Diakon.

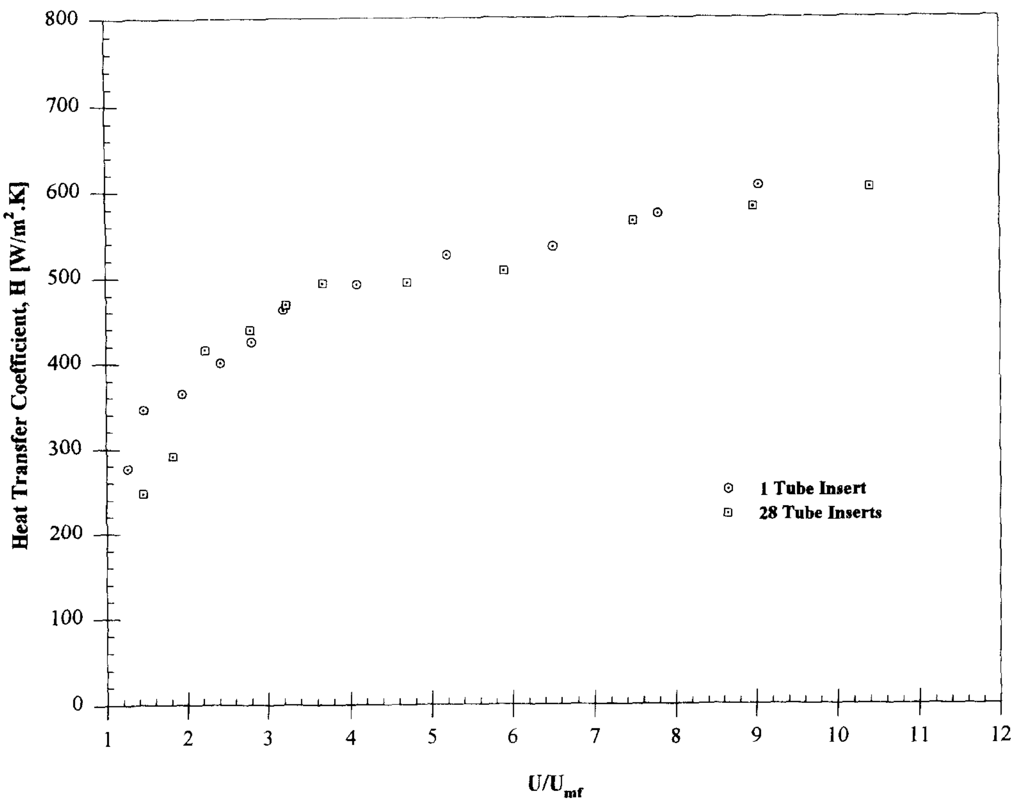

Fig. 5. Effect of tube inserts using a perforated distributor of 12 mm square pitch and 100 μm glass ballotini.

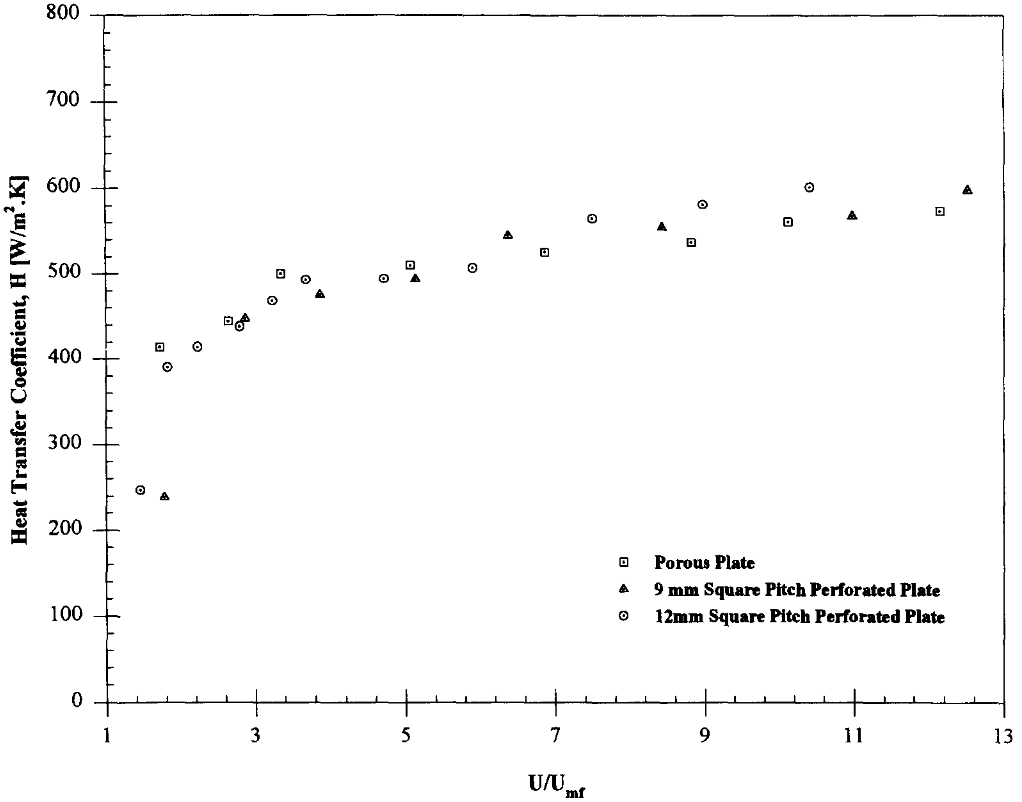

Fig. 6. Effect of distributor type and geometry on a heat transfer coefficient, using 100 μm glass ballotini.

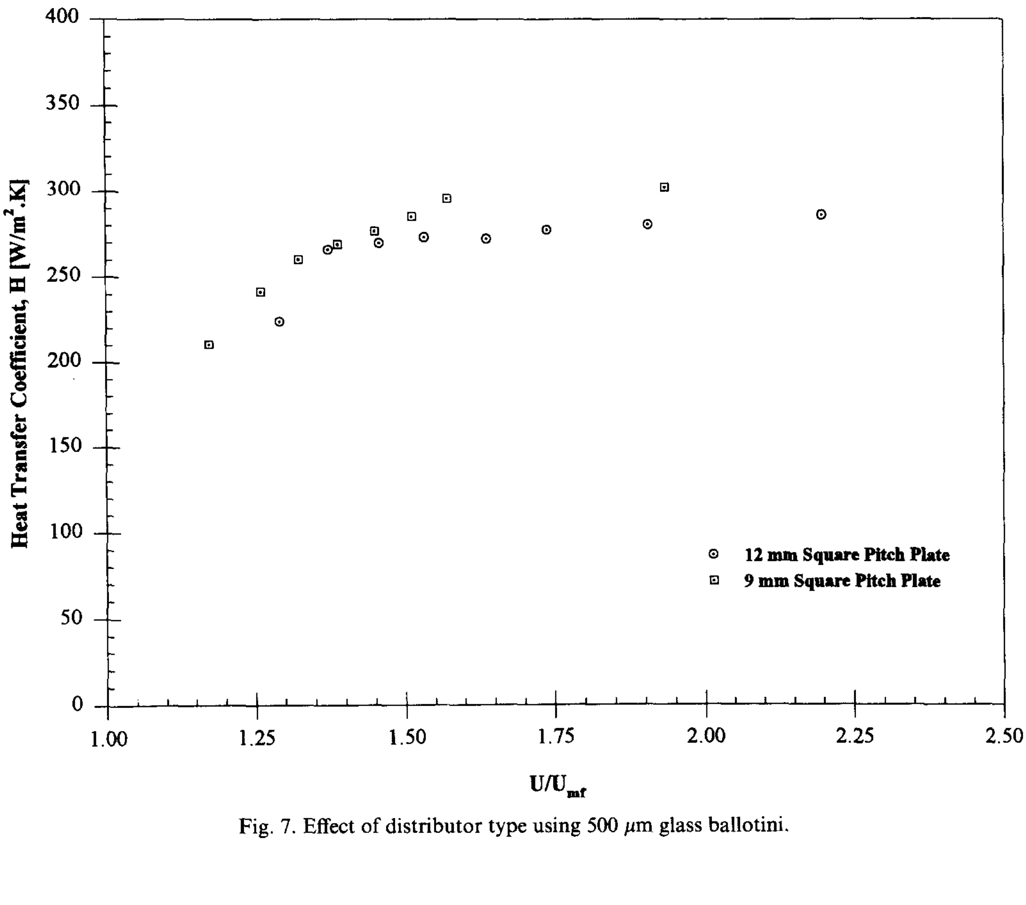

Fig. 7. Effect of distributor type using 500 μm glass ballotini.

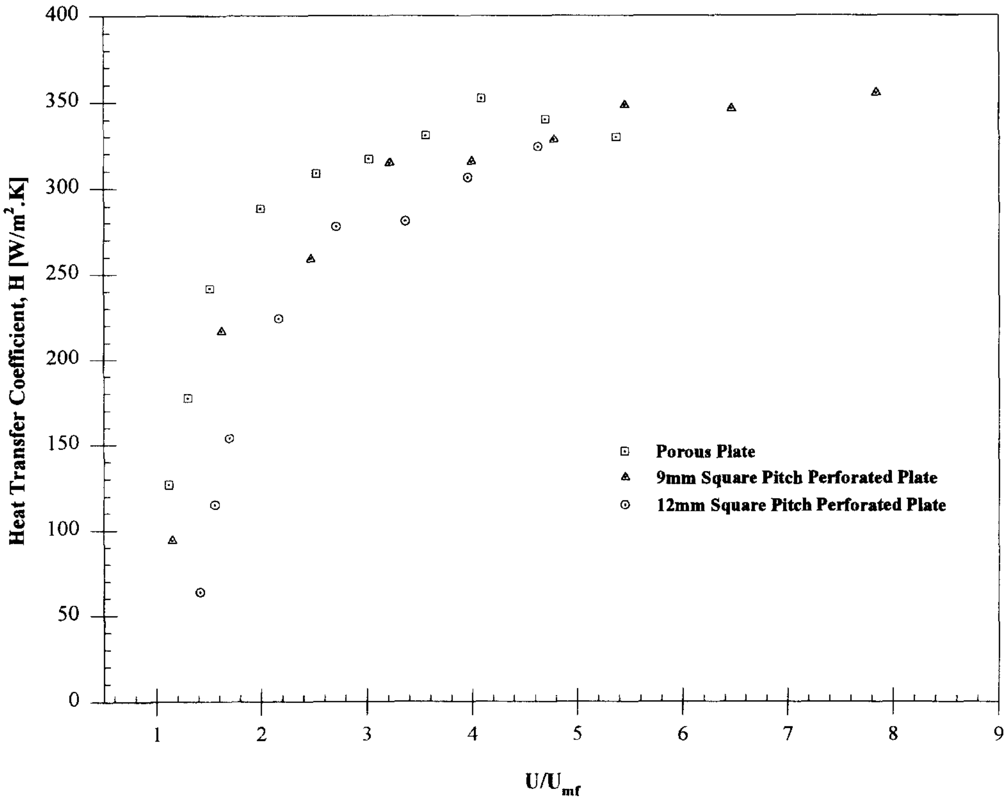

Fig. 8. Effect of distributor type using 325 μm Diakon.

is

$$Nu_m = 0.00515 \cdot Ar^{0.184} \cdot Rc_s^{0.6} \cdot \left(\frac{D}{D_s}\right)^{-0.12} \qquad (1)$$

for the range of Archimedes number $100 < Ar < 10^5$, where the dimensionless groups are the maximum Nusselt number $h_m \cdot d/\lambda$, the Archimedes number $d^3 \cdot \rho(\rho_s - \rho)g/\mu^2$, the thermal capacity group $Rc_s = \rho_s c_s/\rho c$ and the diameter ratio D/D_s, where D_s is the reference diameter 0.09 m. In applying equation (1) to determine the Nusselt number for the inserts, the geometry of the inserts is not considered, only the characteristics of the particles and the bed. The experimental evidence of this paper is that the geometry of the vertical inserts is not important.

The comparison between the experimental results of heat transfer with vertical inserts and heat transfer coefficients in the open fluidized bed is shown in Fig. 9, where the agreement is clearly satisfactory.

In considering the effect of internals Volk *et al.* [4] suggested that reactor performance could be maintained with scale by choosing the inserts to match the equivalent diameter of 4 × flow area/(perimeter of rods + bed). Their proposal was made in the light of their experimental findings on the reduction of iron oxide by hydrogen, but there is an implication that this scaling procedure would apply to heat transfer performance also. In Fig. 9, the value of D has been taken as the bed diameter, 0.29 m, and therefore the scaling procedure of Volk *et al.* does not hold, because a choice of D as that given by the equivalent diameter will reduce the values of the ordinate well below the correlating line. Thus, in contrast to the suggestion of Volk *et al.* the addition of inserts to the bed does not affect the heat transfer coefficient, although the quality of fluidization is improved.

The experimental results show that the effect of the distributor is restricted to low values of U/U_{mf} when the particle size distribution is not narrow, as for the Diakon particles. Otherwise, the effect of distributor type is apparently not important, provided that the distributor is effective with a sufficiently high pressure drop to provide even gas flow over the cross-section.

At higher values of U/U_{mf} the heat transfer coefficient of equation (1) is maintained, as shown in the figures, without a significant maximum in the heat transfer coefficient unless the bed diameter is small (<0.1–0.15 m), when a pronounced maximum may be obtained. The experimental results show that the heat transfer coefficient (1) is attained under the following conditions:

(a) for 100 μm glass ballotini when $U/U_{mf} > 4$;
(b) for 325 μm Diakon particles when $U/U_{mf} > 2.5$;
(c) for 500 μm glass ballontini when $U/U_{mf} > 1.3$.

with the heat transfer coefficient showing a sharp rise to the value of equation (1) from a point just above the minimum velocity for fluidization U_{mf}. The maximum

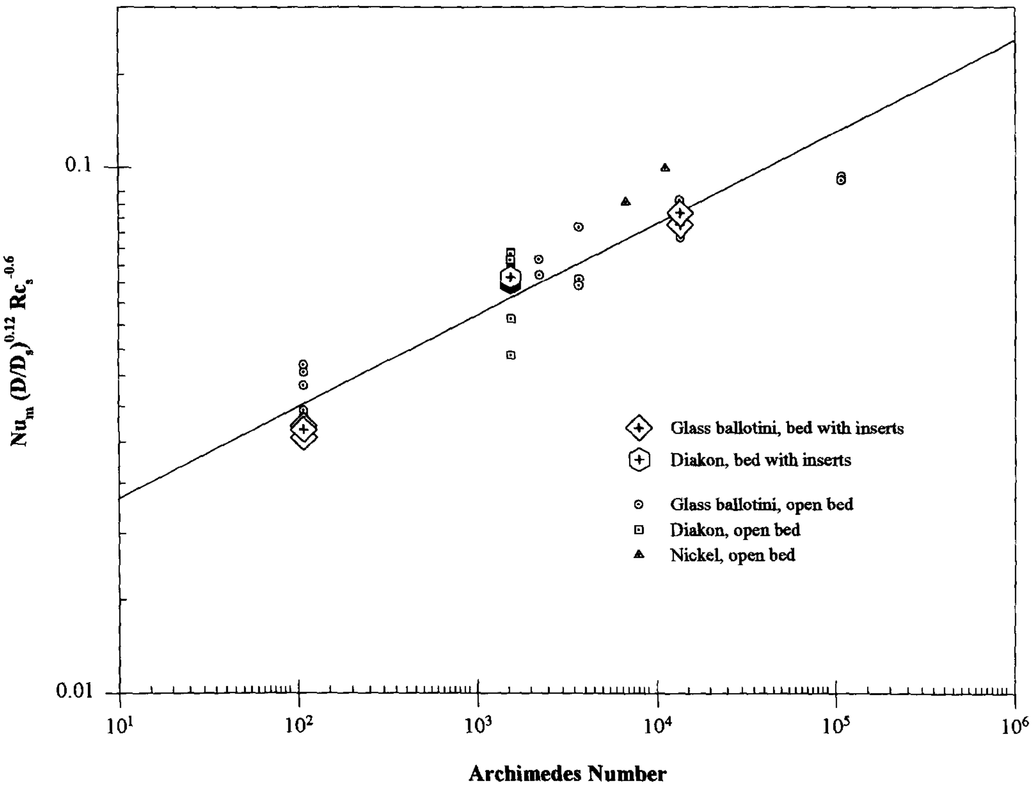

Fig. 9. Comparison of open fluidized bed heat transfer and heat transfer to inserts.

value is attained at smaller values of U/U_{mf} for larger particles, but it should be borne in mind that the operating range of U/U_{mf} is also reduced for larger particles.

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