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# An investigation into heat transfer in circulating fluidized beds

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**Abstract**—A model was proposed to predict the heat transfer in a circulating fluidized bed. To verify the model, experiments were conducted in a 102 mm diameter 5.5 m high Plexiglas column, in which the heat transfer coefficient was measured for different superficial velocities and solid circulation rates and two particle sizes. Results were compared with the experimental data of Mickley and Trilling, Kiang *et al.*, Fraley *et al.*, and Kobro and Brereton.

## 1. INTRODUCTION

FOR PROPER design of circulating fluidized bed boilers it is important to know the effect of design and operating parameters on the bed to wall heat transfer coefficient. At present there is a dearth of mechanistic models for predicting this effect. Grace [1a] inferred that the bed density has a major influence on the heat transfer. This inference was based on experimental data of Fraley *et al.* [2], Kiang *et al.* [3], Stromberg [4b], Mickley and Trilling [5] and Wen and Miller [6] and the data computed from the theoretical model proposed by Martin [7]. Subbarao and Basu [8] suggested a theoretical model of heat transfer based on the packet theory of Mickley and Fairbanks [9]. Development of a mechanistic model is hindered by the lack of information on the residence time of clusters on the wall. In the absence of a more realistic hydrodynamic model, the cluster theory of Subbarao [10] was used to estimate the residence time. This model does not demonstrate the effect of particle size or the bed temperature. The present work improves upon this model to provide a more comprehensive, yet analytical, expression for predicting the heat transfer coefficient in a circulating fluidized bed. Here, the heat transfer coefficient is expressed in terms of cluster residence time. Appropriate expressions for residence time can be plugged into it. Experiments were carried out to verify the results predicted by the proposed model of heat transfer. Both experimental and predicted results have been compared with those of Kobro and Brereton [12], Fraley *et al.* [2], Mickley and Trilling [5], Martin [7], Kiang *et al.* [3] and Subbarao and Basu [8].

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## 2. MODEL

The circulating fluidized bed normally operates in the fast bed regime. Visual observations and video tapes [13] show fast beds to comprise of dense clusters or strands and a gas phase continuum with dispersed solids [14]. The latter (dispersed) phase is sometimes referred to as voids in the present work. The hydrodynamics of fast beds is far from complete. Though a number of works [24, 25] picture this as an upflowing dilute core and a generally down flowing denser solid agglomerates on the wall. Since these speculations are based on relatively small beds (0.1–0.2 m diameter) these are not necessarily a true representation of fast beds in large industrial units.

Let $\delta_c$ be the fraction of the heat transfer surface exposed to the cluster and $(1-\delta_c)$ the fraction to voids or the dispersed phase at any instant. When the cluster is in contact with the gas film, heat will flow to the wall surface by conduction through the gas film and by radiative interchange between the cluster and the wall during the period of contact. When the dispersed phase or void is in contact with the wall, heat will be transferred by convection to the wall surface and by radiative exchange between the wall and distant clusters through the intervening space. The average heat transfer coefficient can be written as

$$h = (h_c + h_{cr})\delta_c + (h_b + h_{br})(1-\delta_c). \quad (1)$$

If the cluster is approximated to be at the same temperature as the bed, then $h_{cr}$ and $h_{br}$ will have similar magnitude so that equation (1) can be written as

$$h = h_c\delta_c + h_b(1-\delta_c) + h_r \quad (2)$$

since $h_{cr} = h_{br} = h_r$. If heat transferring surfaces on opposite walls are so close that they can see each other, $h_{br}$ should be calculated based on radiation exchange through dust clouds between two cold walls. The cluster in contact with the wall cools as it transfers heat to the wall by conduction and radiation. However, the error due to the approximation of the

## NOMENCLATURE

| | |
|---|---|
| $B$ | constant in equation (4) |
| $c_c$ | specific heat of cluster [$kJ\,kg^{-1}\,K^{-1}$] |
| $c_g$ | specific heat of gas [$kJ\,kg^{-1}\,K^{-1}$] |
| $c_s$ | specific heat of solid [$kJ\,K^{-1}$] |
| $d_p$ | particle diameter [m] |
| $D_b$ | stable bubble diameter [m] |
| $D_{bed}$ | bed diameter [m] |
| $D_c$ | cluster diameter [m] |
| $e_c, e_p, e_w$ | emissivity of cluster, particle and wall, respectively |
| $f_{c-w}$ | cluster-to-wall view factor |
| $g$ | acceleration due to gravity, 9.806 m $s^{-2}$ |
| $h$ | average heat transfer coefficient [$kW\,m^{-2}\,K^{-1}$] |
| $h_b$ | heat transfer coefficient due to bubbles [$kW\,m^{-2}\,K^{-1}$] |
| $h_{br}$ | radiant heat transfer coefficient through bubbles [$kW\,m^{-2}\,K^{-1}$] |
| $h_c$ | heat transfer coefficient due to cluster [$kW\,m^{-2}\,K^{-1}$] |
| $h_{cr}$ | radiant heat transfer coefficient through gas film when a cluster is in contact with it [$kW\,m^{-2}\,K^{-1}$] |
| $h_r$ | total radiant heat transfer coefficient [$kW\,m^{-2}\,K^{-1}$] |
| $k_c$ | thermal conductivity of cluster [$kW\,m^{-1}\,K^{-1}$] |
| $k'_e, k_e$ | effective thermal conductivity [$kW\,m^{-1}\,K^{-1}$] |
| $k_{ew}$ | effective thermal conductivity of particle cluster near the wall [$kW\,m^{-1}\,K^{-1}$] |
| $k_g$ | thermal conductivity of gas [$kW\,m^{-1}\,K^{-1}$] |
| $k_p$ | thermal conductivity of particle [$kW\,m^{-1}\,K^{-1}$] |
| $n$ | fraction of particle diameter in equation (9) |
| $Pr$ | Prandtl number |
| $Q_r$ | radiant heat exchange [kW] |
| $Re_a$ | particle Reynolds number based on flow through cluster |
| $R_c$ | thermal resistance of cluster [$m^2\,K\,kW^{-1}$] |
| $R_w$ | wall contact resistance [$m^2\,K\,kW^{-1}$] |
| $t_c$ | residence time of a cluster along the wall [s] |
| $T_B$ | bed temperature [K] |
| $T_w$ | wall temperature [K] |
| $U$ | superficial gas velocity [$m\,s^{-1}$] |
| $U_c$ | average velocity of a cluster [$m\,s^{-1}$] |
| $U_{mf}$ | superficial velocity at incipient fluidization [$m\,s^{-1}$] |
| $U_T$ | terminal velocity of particles [$m\,s^{-1}$] |
| $W$ | solid circulation rate [$kg\,m^{-2}\,s^{-1}$] |
| $x$ | ratio of solid concentration near the wall to the average solid concentration in the bed cross section |
| $y$ | volumetric concentration of solid in the voids. |

Greek symbols

| | |
|---|---|
| $\delta_{av}$ | average volumetric concentration of solid in the bed |
| $\delta_c$ | fraction of wall surface covered by clusters |
| $\varepsilon$ | bed voidage |
| $\varepsilon_c$ | cluster voidage at minimum fluidization velocity |
| $\varepsilon_w$ | voidage near the wall |
| $\rho_c$ | density of cluster [$kg\,m^{-3}$] |
| $\rho_g$ | density of gas [$kg\,m^{-3}$] |
| $\rho_m$ | mean suspension density in the bed [$kg\,m^{-3}$] |
| $\rho_p$ | density of particles [$kg\,m^{-3}$] |
| $\rho_{sus}$ | suspension density in the voids [$kg\,m^{-3}$] |
| $\sigma$ | Stefan–Boltzmann constant, $5.67\times10^{-11}\,kW\,m^{-2}\,K^{-4}$. |

radiative term in equation (2) may not be significant because the fraction of the wall area covered by clusters, $\delta_c$, and their contact time with the wall are both small. During the brief time of contact the cluster does not cool sufficiently to introduce a major error. For comparison a numerical solution of the one-dimensional transient heat conduction equation for the cluster was carried out following the analysis of Vedamurthy and Sastri [15]. For a cluster at 1123 K contacting a wall at 353 K, the calculated value of $h_r$ was 0.0766 kW $m^{-2}$ $K^{-1}$ while from equation (15) under the assumption that $h_{cr} = h_{br} = h_r$ it was 0.077 kW $m^{-2}$ $K^{-1}$. Thus it justifies the approximation of equation (2).

The heat conduction from the cluster is similar to that from the emulsion packet in a bubbling fluidized bed. Thus the thermal resistance can be written as the sum of the resistance at the wall and that of the cluster [16]. A number of expressions for contact resistance are available in the literature [16–18]. Most workers chose a thickness of the contact resistance as a fraction of the particle diameter. However, Schlunder [20] and Boch and Molerus [18, 21] presented values of contact resistance without the aid of empiricism. Assuming sand particles as spherical, we take the contact resistance, $R_w$ as $d_p k_{ew}/10$. The thermal time constant for typical solids of a fast bed combustor calculated from Glicksman [19] is about 0.02 s, which is less than the typical residence time of clusters on the wall. A numerical solution of unsteady state heat transfer through a semi-infinite cluster showed that during the residence time the temperature gradient within the cluster does not extend beyond a few particle diameters. Thus one can assume the cluster to be semi-infinite for heat transfer calculations. So, using the analysis of Baskakov [16] and Mickley and Fairbanks [9] the heat transfer from the cluster may be written as

$$h_c = \frac{1}{R_w + R_c} = \frac{1}{\frac{d_p}{10k_{ew}} + \sqrt{\left(\frac{t_c \pi}{4k_c C_c \rho_c}\right)}}. \quad (3)$$

To determine the effective thermal conductivity of the semi-infinite phase of clusters, $k_c$ (when $\varepsilon = \varepsilon_c$) and that in the zone adjacent to the wall, $k_{ew}$ (when $\varepsilon = \varepsilon_w$), Gelperin and Einstein [17] recommended the general expression. $k_{ew}$ and $k_c$ can be obtained by substituting appropriate values of the voidage

$$\frac{k_c}{k_g} = \frac{k'_e}{k_g} + B\, Re_a\, Pr \quad (4)$$

where $B = 0.07$–$0.15$ depending on the grain shape.

The value of 0.15 for $B$ was taken for spherical particles. Bobkov and Gupalo's [22] theoretical calculations suggest gas percolation through clusters. Though they did not support their conclusion with experiments, continuous deformation, splitting and dissolution of clusters suggest that there is a gas percolation velocity of the order of $U_{mf}$ and the voidage of the cluster may be similar to that of a bubbling bed at minimum fluidization. The values of $k'_e$ were taken from the graphs of Baskakov for the voidages of $\varepsilon_c$ and $\varepsilon_w$ [17], respectively. The average voidage near the wall, $\varepsilon_w$, within a distance $nd_p$ from the wall (where $n$ is a fraction) was determined using the expression of the sector volume of a sphere [23]

$$\varepsilon_w = 1 - \pi n\left(\frac{1}{2} - \frac{n}{3}\right). \quad (5)$$

For a gap of one-tenth particle diameter $n = 0.1$. So, $\varepsilon_w$ is found to be 0.853. However, for design $k_{ew}$ may be approximated as $k_g$. The specific heat and density of the cluster were

$$c_c = c_s(1-\varepsilon_c) + c_g \varepsilon_c \quad (6)$$

$$\rho_c = \rho_p(1-\varepsilon_c). \quad (7)$$

If '$y$' is the volume fraction of solids in the dispersed phase of the fast bed the average volume fraction of the cluster, $\delta_{av}$, can be related to the average void fraction, $\varepsilon$, in the bed as

$$\delta_{av} = \frac{1-\varepsilon-y}{1-\varepsilon_c-y}. \quad (8)$$

Recent experiments by Weinstein *et al.* [24] and Monceaux *et al.* [25] showed that the solid concentration near the wall is much higher than that in the centre of the bed. If the solid concentration at the wall is $x$ times the average bed concentration

$$\delta_c = \frac{x(1-\varepsilon)-y}{1-\varepsilon_c-y}. \quad (9)$$

The average bed density is given by

$$\rho_m = \rho_p(1-\varepsilon). \quad (10)$$

To determine the heat transfer coefficient when voids (dispersed phase) are in contact with the wall, Subbarao and Basu [8] considered transient conduction of heat to the stagnant semi-infinite void space between the clusters. This yields a rather low value of the heat transfer coefficient. Closer examination suggests that in between contacts with two successive clusters, the wall is in contact with an upflowing dispersed phase with a continually developing boundary layer on the wall. In the absence of any appropriate correlation for this kind of situation, we use the correlation of Wen and Miller [6] for heat transfer from the dilute phase, which is given as

$$h_b = \frac{k_g}{d_p}\frac{c_s}{c_g}\left(\frac{\rho_{sus}}{\rho_p}\right)^{0.3}\left(\frac{U_T^2}{gd_p}\right)^{0.21} Pr \quad (11)$$

where $\rho_{sus}$ is the suspension density of gas–solid mixture.

If $y$ represents the volumetric concentration of solid dispersed in the voids or the dilute phase, then

$$\rho_{sus} = y\rho_p + (1-y)\rho_g. \quad (12)$$

At present no information on volumetric concentration of solids in the dispersed phase is available. So, we assume it to be 0.001 following the suggestion of Kunii and Levenspiel [26] for solids dispersed in bubbles.

For determination of radiative heat transfer, $h_r$, the clusters which are away from the wall or in contact with it are assumed to be at the bed temperature $T_B$. The radiant heat exchange between the cluster and the wall, both being considered gray, is given by

$$Q_r = \delta(\pi/4)D_c^2 \sigma f_{c\text{-}w}(T^4 - T_w^4) \quad (13)$$

$$Q_r = h_r(\pi/4)D_c^2(T - T_w) \quad (14)$$

where $f_{c\text{-}w}$ is the cluster-to-wall view factor. The intervening gas between the wall and the cluster is assumed to be transparent to thermal radiation. Therefore

$$h_r = f_{c\text{-}w}\sigma(T_w + T_B)(T_B^2 + T_w^2). \quad (15)$$

The view factor $f_{c\text{-}w}$ depends on the shape, disposition and emissivities of the two bodies. Since the gap between the cluster and the wall is very small compared to their lengths, they can be assumed to be parallel so that

$$f_{c\text{-}w} = \frac{1}{1/e_w + 1/e_c - 1}. \quad (16)$$

Since the particles in the cluster undergo multiple reflections, the emissivity of the cluster is more than that of the particles, and Grace [1b] suggests

$$e_c = 0.5(1 + e_p). \quad (17)$$

Substituting equations (9), (11) and (15) in equation (3), the average heat transfer coefficient in a circulating fluidized bed combustor is finally given by

$$h = \frac{\delta_c}{\dfrac{d_p}{10k_{ew}} + \sqrt{\left(\dfrac{\pi t_c}{4k_c c_c \rho_c}\right)}} + (1-\delta_c)\left[\frac{k_g}{d_p}\right]\left[\frac{c_s}{c_g}\right]\left[\frac{\rho_{sus}}{\rho_p}\right]^{0.3}\left[\frac{U_T^2}{gd_p}\right]^{0.21} + \sigma f_{c\text{-}w}(T_B+T_w)(T_B^2+T_w^2). \quad (18)$$

The present expression, equation (18), given in closed form, shows the effect of gas properties, solid properties, bed temperatures, wall temperatures, bed density and cluster residence time. All parameters except bed density and cluster residence time are measurable design or operating variables. Bed density and residence time are dictated by the bed hydrodynamics which is in turn, is affected by dimensions of the bed, location and geometry of the heat transfer surface. At present no comprehensive model of the hydrodynamics of fast bed relating bed density and cluster residence time is available. Thus in the absence of more detailed information we use the expression of cluster residence time derived by Subbarao [10] based on his cluster model. Subbarao [10] obtained the average cluster residence time on the wall as

$$t_c = \frac{\rho_m D_b}{\rho_p^{1/3}(1-\varepsilon_c)^{1/3} U^{1/3} W^{2/3}}. \quad (19)$$

It may be noted that the present model can also work with the cluster diameter calculated from any other hydrodynamic model.

The stable bubble diameter $D_b$ is given by

$$D_b = \frac{2U_T^2}{g}. \quad (20)$$

If $2U_T^2/g > D_{bed}$, then bubble diameter is taken to be equal to the bed diameter. This demonstrates a possible effect of the bed diameter on the heat transfer coefficient below the stable bubble size. The length of the heat transferring surface may have an effect on the heat transfer coefficient. If an isolated small heat transfer probe is located on an adiabatic wall, the cluster always reaches the probe with bed temperature, whereas in the case of a continuous heat transferring surface, such as in industrial boilers, clusters may reach the measuring section after having transferred some heat to the surfaces above or below it. Thus heat transfer coefficients, measured by small probe in bench scale units, may be larger than industrial boiler with long water wall panels. Wu *et al.* [27] reported some effect of the size of probe on the heat transfer coefficient.

With the above substitution one could predict the effect of important design conditions. For example by substituting the radial variation of the solid fraction [24] or $x$, the above equation predicts a continuous increase in the heat transfer coefficient from the centre towards the wall of the combustor. Furthermore, by incorporating the vertical voidage variation [14], equation (18) will predict a lower heat transfer coefficient at the top of the fast bed than at the bottom. Thus this model can successfully predict the physical behaviour of the heat transfer in a fast bed.

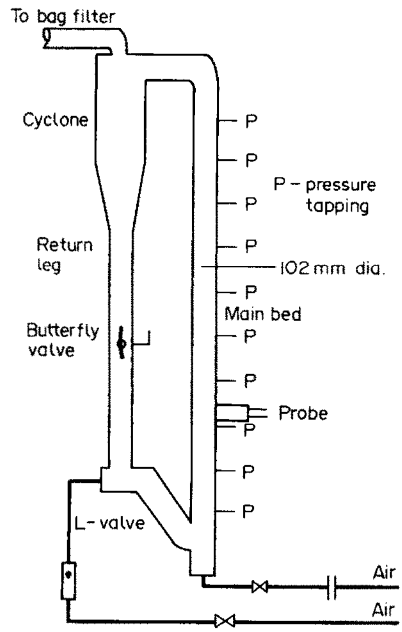

FIG. 1. Schematic of the 102 mm fluidized bed unit.

## 3. EXPERIMENTAL FACILITY

To compare above predictions with measured values, heat transfer coefficients were measured in a cold circulating fluidized bed. The fast bed comprised a 5.5 m tall column made up of Plexiglas tube sections of 102 mm diameter (Fig. 1). Entrained solids were recovered in a cyclone and returned to the bottom of the column with the help of a pneumatically operated L-valve. Air was supplied to the distributor plate by a high pressure blower and the flow rate was measured by a standard orifice meter. Compressed air was used to operate the L-valve in controlling the solid circulation through the bed and the flow rate was measured by a calibrated rotameter. A butterfly valve was located midway in the return leg to measure the solid circulation rate in the column by closing the valve and measuring the volume of solids collected above it over a certain period of time.

The bed to wall heat transfer was measured by a probe (Fig. 2) flush with the wall and installed 2.1 m above the distributor plate. It was a 100 mm long 25 mm diameter carbon steel rod with four iron–constantan thermocouples located at uniform intervals of 20 mm along its length. It was insulated against any heat loss so as to achieve one-dimensional heat conduction and thereby a linear temperature variation

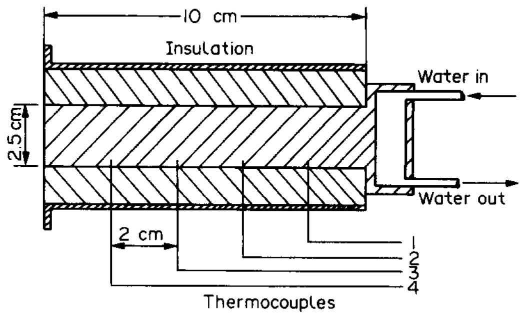


FIG. 2. Heat transfer probe.

Table 1. Properties of sand particles used

| (a) | | (b) | |
|---|---|---|---|
| 1. Mean size: 227 $\mu$m | | 1. Mean size: 87 $\mu$m | |
| 2. Density: 2650 kg m$^{-3}$ | | 2. Density: 2650 kg m$^{-3}$ | |
| 3. $U_{mf} = 0.054$ m s$^{-1}$ | | 3. $U_{mf} = 0.008$ m s$^{-1}$ | |
| 4. $U_T = 1.88$ m s$^{-1}$ | | 4. $U_T = 0.72$ m s$^{-1}$ | |
| Range ($\mu$m) | Weight fraction | Range ($\mu$m) | Weight fraction |
| 0–53 | 0.0005 | 0–53 | 0.1012 |
| 53–106 | 0.0138 | 53–106 | 0.3200 |
| 106–212 | 0.3404 | 106–212 | 0.5773 |
| 212–355 | 0.4509 | 212–355 | 0.0015 |
| 355–417 | 0.1161 | | |
| 417–500 | 0.0783 | | |

along the centre line. One face of the probe was flush with the bed wall, while the other face was in contact with a chamber through which boiling water from a constant head water tank was circulated. Two pressure tappings 313 mm across the probe measured the pressure drop and helped estimate the local voidage. The bed temperature was measured by another iron–constantan thermocouple installed inside the bed.

Sand particles of average sizes 227 and 87 $\mu$m and of density 2650 kg m$^{-3}$ were used in the investigation. The size distributions are given in Table 1. To avoid discharge by statical electricity, the whole apparatus was earthed. The fluidizing air was passed through the distributor plate at desired rates. By operating the L-valve, solids from the return leg were fed into the bed. These solids were then immediately entrained in the upflowing air. When equilibrium between solids entrained and recycled was reached and the thermocouple readings were steady, the pressure drop across the probe, the rotameter and the thermocouple readings were recorded. The solid circulation rate was measured by using the butterfly valve. Similar readings were noted for different values of air velocities, solid circulation rates and two sizes of sand particles. Particular care was taken to check that temperature distribution along the length of the probe was linear. If it was not so, the test was discarded and the probe was checked for any lateral heat loss.

## 4. RESULTS AND DISCUSSION

The heat transfer coefficient was calculated from the temperature at the tip of the probe (wall temperature) and the heat flux along the length of the probe. Both were determined from the measured steady-state temperature distribution along the length of the probe. Thus the heat transfer coefficient was computed for each operating condition. The local voidage or the suspension density of the bed were estimated directly from the measured pressure drop across the probe. Table 2 gives the operating parameters and the measured heat transfer coefficients.

### 4.1. *Effect of superficial velocity*

Measured heat transfer coefficients and voidages for both 227 and 87 $\mu$m sand particles are plotted against superficial velocity in Fig. 3. The bed voidage, as determined from the pressure drop across the 313 mm section of the bed containing the heat transfer probe, increased with velocity when the solid recycle rate is kept constant. This is in agreement with the empirical equation developed by Reying *et al.* [28] for a fast bed except that the empirical constant was found to be somewhat different. With the increase of fluidization velocity the solid concentration in the bed decreases as a result of which the heat transfer coefficient decreases. Thus the major contribution to heat transfer is made by particle convection.

### 4.2. *Effect of suspension density*

Figure 4 shows the effect of suspension density on the heat transfer coefficient. According to the present model, the fraction of cluster on the wall rather than the bulk suspension density affects the heat transfer. However, these two are closely related by equations (9) and (10). The data of the present experiment as well as those of Mickley and Trilling [5], Kiang *et al.* [3], Fraley *et al.* [2] and Kobro and Brereton [12] are plotted on Fig. 4. The suspension density, which was measured by measuring static pressure on the wall, is found to be a dominant factor influencing the heat transfer coefficient in a fast bed.

The heat conduction from clusters is much higher than that from the gas. Therefore, the heat transfer coefficient increases with the increase in suspension density. The drop in heat transfer coefficient with increasing superficial velocity (Fig. 3) may be attributed to the drop in suspension density. This effect is in agreement with Mickley and Trilling [5] and with Wu *et al.* [27].

### 4.3. *Effect of circulation rate*

Circulation rate indicates the turnover of solids through the bed. For a given bed density different circulation rates can have a different solid velocity [14]. Thus the heat transfer coefficient may be different for different circulation rates but at the same bed density.

Equation (18) suggests that the heat transfer

Table 2. Experimental data on heat transfer

| Run No. | $U$ (m s$^{-1}$) | $T_w$ (K) | $T_B$ (K) | $W$ (kg m$^{-2}$ s$^{-1}$) | $\rho_m$ (kg m$^{-3}$) | $h$ (kW m$^{-2}$ K$^{-1}$) |
|---|---|---|---|---|---|---|
| (a) Particle size: 227 $\mu$m | | | | | | |
| 1 | 4.2 | 353 | 313 | 67.21 | 40.95 | 0.175 |
| 2 | 4.5 | 355 | 316 | 60.91 | 28.35 | 0.150 |
| 3 | 5.3 | 357 | 321 | 67.70 | 23.63 | 0.141 |
| 4 | 3.7 | 351 | 314 | 60.55 | 50.40 | 0.216 |
| 5 | 3.3 | 345 | 321 | 60.44 | 72.45 | 0.265 |
| 6 | 3.0 | 342 | 311 | 44.02 | 96.79 | 0.286 |
| 7 | 4.2 | 351 | 305 | 63.86 | 37.80 | 0.172 |
| 8 | 3.7 | 348 | 303 | 59.56 | 39.38 | 0.196 |
| 9 | 3.5 | 343 | 305 | 64.40 | 59.85 | 0.250 |
| 10 | 4.6 | 343 | 307 | 92.20 | 37.80 | 0.170 |
| 11 | 4.3 | 343 | 307 | 90.10 | 47.25 | 0.175 |
| 12 | 4.0 | 346 | 307 | 76.50 | 56.70 | 0.226 |
| 13 | 3.6 | 346 | 307 | 65.10 | 63.00 | 0.238 |
| 14 | 3.2 | 344 | 305 | 53.60 | 72.40 | 0.279 |
| 15 | 5.0 | 343 | 308 | 70.60 | 26.65 | 0.146 |
| 16 | 4.5 | 348 | 313 | 67.30 | 31.50 | 0.166 |
| 17 | 4.7 | 348 | 313 | 79.90 | 22.05 | 0.167 |
| 18 | 4.3 | 348 | 313 | 75.40 | 33.00 | 0.180 |
| 19 | 4.0 | 348 | 313 | 76.20 | 45.68 | 0.209 |
| (b) Particle size: 87 $\mu$m | | | | | | |
| 20 | 4.6 | 348 | 313 | 59.42 | 34.65 | 0.168 |
| 21 | 4.2 | 345 | 307 | 57.34 | 40.95 | 0.218 |
| 22 | 5.0 | 347 | 303 | 39.50 | 25.20 | 0.193 |
| 23 | 4.0 | 340 | 305 | 96.23 | 44.10 | 0.320 |
| 24 | 3.7 | 336 | 307 | 117.00 | 58.63 | 0.364 |
| 25 | 4.3 | 341 | 307 | 56.67 | 36.59 | 0.254 |
| 26 | 4.5 | 343 | 307 | 57.55 | 32.80 | 0.209 |
| 27 | 4.7 | 345 | 307 | 30.00 | 29.30 | 0.219 |
| 28 | 4.6 | 348 | 313 | 35.86 | 21.50 | 0.163 |

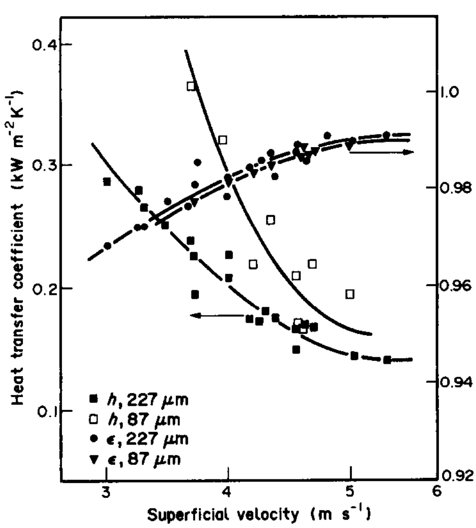


FIG. 3. Variation of heat transfer coefficient and voidage with superficial velocity.

coefficient would increase with an increase in circulation rate, but its effect is not significant. According to the proposed model used in conjunction with Subbarao's equation of cluster residence time, a 100% increase in the circulation rate (unchanged bed density) results in about a 10% enhancement of heat transfer rate. No major influence of the circulation rate except through its effect on the suspension density could be studied within the limited range of the present experiments. Thus one can justify the use of some empirical equations relating the recycle rate with suspension density and other parameters.

### 4.4. *Effect of particle size*

The heat transfer coefficient decreases with increasing particle size for all bed densities (Fig. 4) and fluidization velocities (Fig. 3). This observation is apparent from the results of 87 and 227 $\mu$m particles in Fig. 3 and is supported by the data of Mickley and Trilling [5] who performed experiments with five different particle sizes between 70 and 451 $\mu$m, and those of Kobro and Brereton [12] who experimented with two particle sizes. The particle convective component of heat transfer from the clusters to the wall is enhanced with the increase in surface area of particles per unit volume as the particle size decreases. The difference in heat transfer coefficients between 87 and 227 $\mu$m particles shows an apparent decrease with increasing velocity or decreasing bed density (Fig. 3). The difference in heat transfer coefficients is due to the difference in contact resistance of clusters. Since the fraction of wall surface covered by clusters is less at higher velocity, the impact of particle size difference is less prominent at lower bed density. This effect, however, was not very clear in the data of Mickley and Trilling (Fig. 4).

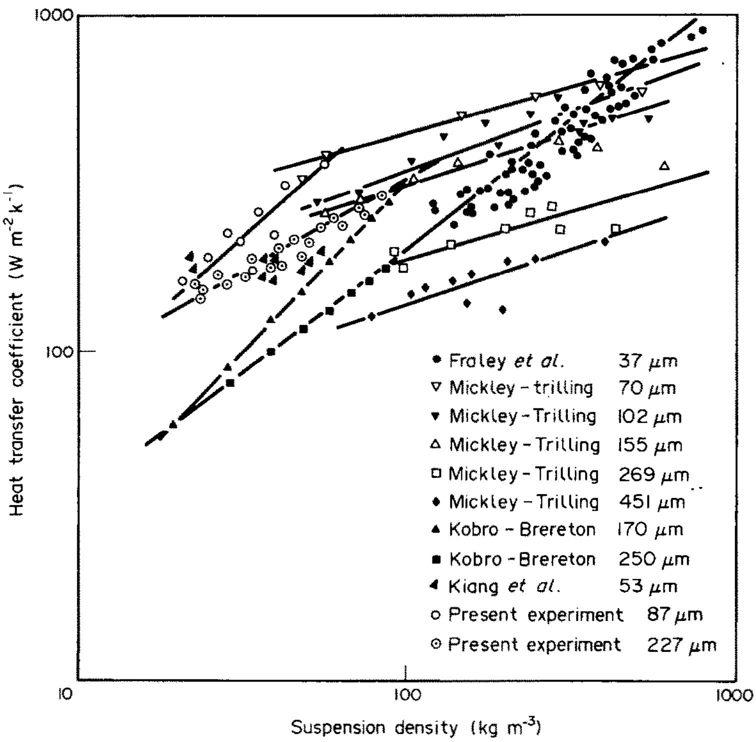


FIG. 4. Experimental data on heat transfer coefficient as a function of suspension density.

The above effect is observed for a narrow size range of particles. It is not known whether such an effect of mean particle size will be evident for a wide size distribution as observed in industrial units.

### 4.5. *Model predictions*

The present model (equation (18)) successfully predicts all the above effects of physical variables on heat transfer. This demonstrates the correctness of physical modelling of the process of heat transfer in a circulating fluidized bed combustor. The present model suffers from the same uncertainty as that of models of bubbling fluidized bed heat transfer, i.e. the residence time. No comprehensive expression of packet residence time in bubbling fluidized bed is available. However, to demonstrate the correctness of physical modelling most workers chose an empirical relation that explains their results best. In the present case of a fast bed, only one analytical expression of residence time [10] is available. The present model used in conjunction with this expression predicts the effect of all design and operating parameters reasonably well though the exact agreement is not obtained in all cases. Resolution of this disagreement must await further information on the hydrodynamics of a fast bed. A dearth of experimental data over a wide range of operating conditions prevented a comprehensive comparison of model predictions. Figure 5 shows one of the limited set of data. The heat transfer coefficient is plotted as a function of bed density at two different temperatures of 298 and 1123 K with bed density varying up to 90 kg m$^{-3}$ (Kobro and Brereton [12]). Heat transfer coefficients at these operating conditions were computed using the present model at both these temperatures, and the model of Martin [7] and Subbarao and Basu [8] at the lower temperature of 298 K. The solid circulation rates were taken from Stromberg [4b] as indicated on the $x$-axis. The values predicted from Martin's model (using constant $k = 2$ and 2.6) are an order of magnitude lower than the experimental ones. Thus it leaves the applicability of Martin's analogy of simulating random movements of fluidized particles with the motion of gas molecules open to question. The model proposed by Subbarao and Basu [8] did not consider radiation and underestimated the gas convective component. The net effect of their approximations is some underprediction of heat transfer rates at room temperature beds and a gross underestimation of heat transfer in hot combustors.

In Fig. 6 the present model shows an interesting effect of distribution of solid concentration across the cross-section of the bed. Predicted values of the heat transfer coefficient for 87 μm are plotted against bed density with $x$, the ratio of suspension density at the wall to the mean value, as a parameter. It is seen that as $x$ increases, values of the heat transfer coefficient increase significantly and the model prediction is now closer to the experimental curve. Such gradients of suspension density were observed by Weinstein *et al.* [24] and Monceaux *et al.* [25]. Weinstein *et al.* [24] observed that the ratio of solid concentration near the wall to the average solid concentration in the bed (i.e. $x$) varies from about 1.5 to 2.5. The value of $x$ depends

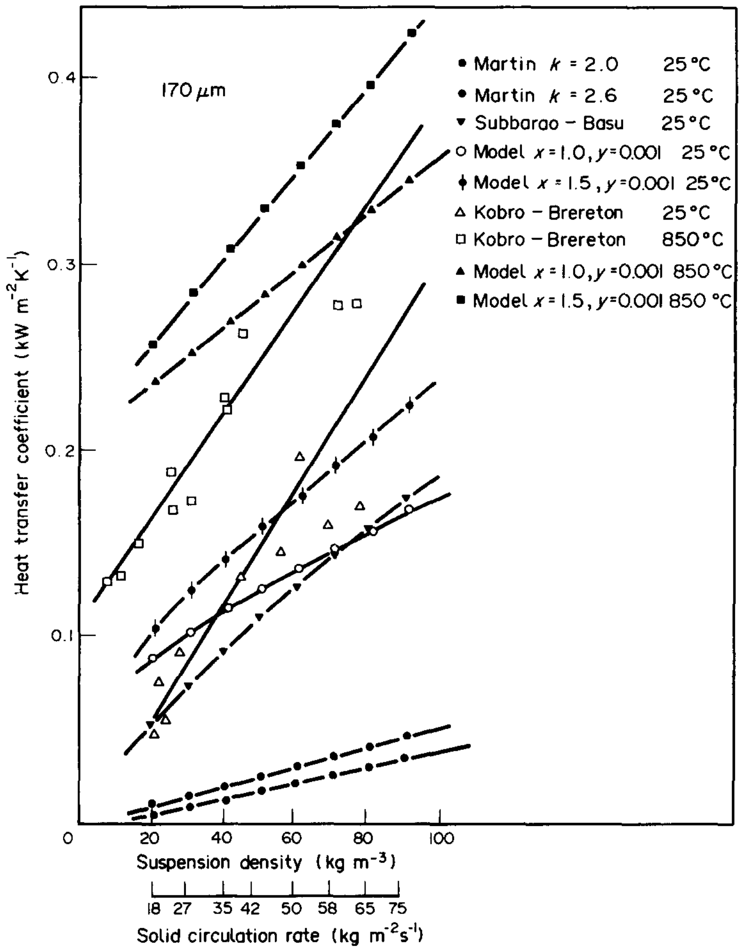


FIG. 5. Comparison of predicted values with experimental data of Kobro and Brereton [12] for 170 μm particles at 298 and 1123 K.

on equipment and the operating conditions. Present knowledge of hydrodynamics does not permit assigning any general value of $x$. However, in the present work the agreement between theory and experiment is best for $x \sim 2.0$. The existence of a higher heat transfer coefficient at the wall than at the centre is evident from the measurements of Monceaux *et al.* [25].

The effect of solid concentration in the dispersed phase, i.e. $y$, on the predicted heat transfer coefficient for the 227 μm particle is shown in Fig. 7. As the value of $y$ increases, heat transfer by convection from the gas–solid suspension in the voids to the wall increases resulting in an increased heat transfer coefficient, but its slope with suspension density remains unaltered. The curve with $x = 2.0$ and $y = 0.001$ yields the best agreement with the experimental results of 227 μm particles.

### 4.6. *Effect of temperature*

Kobro and Brereton's [12] data for 170 μm particles at 850°C are compared with model predictions in Fig. 5. The heat transfer coefficient at 1123 K is significantly higher than that at 298 K because the gas convective component is enhanced due to the increased thermal conductivity of gas film at the elevated temperature and the high radiative exchange between clusters and the wall surface. However, the difference in measured heat transfer coefficients between 298 and 1123 K is much lower than that predicted from the present theory. This may be attributed to the fact that the bed between the heat transfer probe and the opposite water-cooled wall of Kobro and Brereton was not opaque as assumed in the model. Intermittent radiation from the relatively cool opposite wall may be responsible for the observed low heat transfer coefficient.

### 4.7. *Comparison of theoretical predictions and experimental data*

The present experiments as well as the experiments of other workers were carried out over a wide range of velocity, recycle rate, bed density, particle size and bed temperature. In order to facilitate easy comparison of predicted results with experimental values, all data were plotted in Fig. 8, with the measured heat

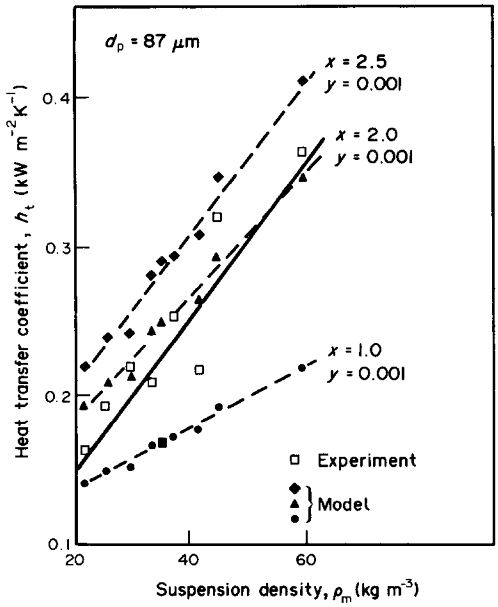


FIG. 6. The effect of local solid concentration near the wall on predicted values of heat transfer coefficients as compared to measured values for the 87 $\mu$m particles.

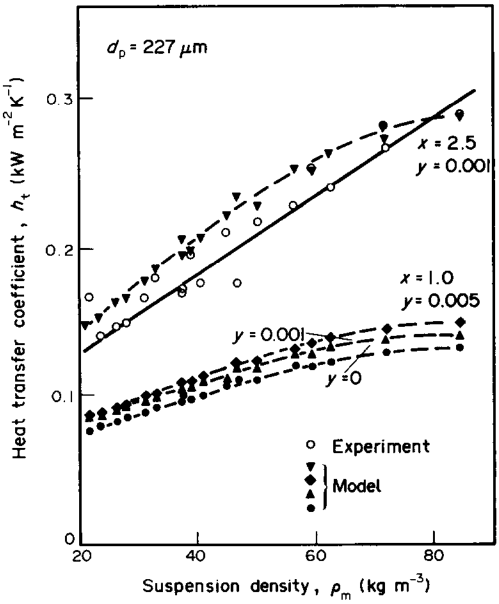


FIG. 7. The effect of solid concentration in the void or dilute phase on predicted heat transfer coefficients as compared to measured values for the 227 $\mu$m particle.

transfer coefficient and the theoretical prediction as the coordinates. The value computed from the present model corresponds to the operating condition in each case. Since the data on solid circulation rates were not provided by Mickley and Trilling [5] a uniform value of 40 kg m$^{-2}$ s$^{-1}$ was assumed as the circulation rate. The values of $x$ and $y$ were assumed to be 1.0 and 0.001, respectively. Although most of the values are seen to cluster around the 45° line which is the line of perfect agreement; predicted values of Kobro and Brereton [12] and the present experiment are generally above the line. As shown in Figs. 5–7 agreement can be improved by using appropriate values of $x$ and $y$.

Most experimental data were within 30% of the theoretical values. The data of Mickley and Trilling [5] at bed densities higher than 400 kg m$^{-3}$ were not

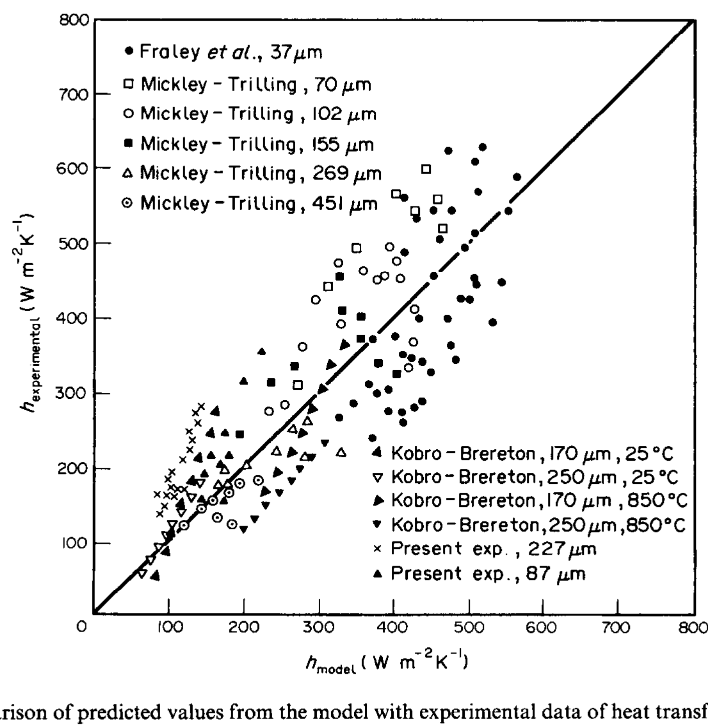


FIG. 8. Comparison of predicted values from the model with experimental data of heat transfer coefficient.

considered, since these corresponded to bubbling beds.

## 5. CONCLUSIONS

(1) Heat transfer in a circulating fluidized bed combustor can be predicted by a cluster renewal model proposed here.

(2) Bed-to-wall heat transfer increases with increasing suspension density; but it decreases if the fluidization velocity is increased while keeping the recycling rate constant.

(3) Within the present range of experimental conditions finer particles resulted in a higher heat transfer coefficient for a given suspension density.

(4) A higher heat transfer at a higher bed temperature is predicted by the proposed model due to the effect of radiation as well as increased thermal conductivity of the gas film at the elevated temperature.

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## RECHERCHE SUR LE TRANSFERT THERMIQUE DANS LES LITS FLUIDISES A CIRCULATION

**Résumé**—Un modèle est proposé pour prédire le transfert de chaleur dans un lit fluidisé à circulation. Pour vérifier le modèle, des expériences sont effectuées dans une colonne en plexiglas de 102 mm de diamètre et de 5,5 m de hauteur, dans laquelle le coefficient de transfert thermique est mesuré pour différentes vitesses, différents taux de circulation de solide et deux tailles de particules. Les résultats sont comparés avec les données expérimentales de Mickley et Trilling, Kiang *et al.*, Fraley *et al.*, Kobro et Brereton.

## UNTERSUCHUNG ZUR WÄRMEÜBERGUNG IM ZIRKULIERENDEN WIRBELBETT

**Zusammenfassung**—Es wurde ein Modell erstellt, um den Wärmeübergang in einem zirkulierenden Wirbelbett zu berechnen. Um das Modell zu überprüfen, wurden Versuche in einem 5,5 m hohen Plexiglaszylinder von 102 mm Durchmesser durchgeführt. Der Wärmeübergangskoeffizient wurde für verschiedene Oberflächengeschwindigkeiten und feste Zirkulationsverhältnisse sowie zwei Partikelgrößen gemessen. Die Ergebnisse wurden mit den entsprechenden experimentell ermittelten Werten von Mickley und Trilling, Kiang *et al.*, Fraley *et al.*, sowie Kobro und Brereton verglichen.

## ТЕПЛООБМЕН В ЦИРКУЛИРУЮЩИХ ПСЕВДООЖИЖЕННЫХ СЛОЯХ

**Аннотация**—Предложена модель для расчета теплообмена в циркулирующем псевдоожиженном слое. Модель проверялась экспериментально на колонне из органического стекла диаметром 102 мм и высотой 5,5 м. Коэффициент теплообмена в колонне измерялся для различных скоростей фильтрации, скоростей циркуляции твердых частиц и двух размеров частиц. Проведено сравнение результатов с экспериментальными данными Микли и Триллинга, Кианга и др., Фрейли и др., а также Кобро и Бреретона.