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HEAT TRANSFER MECHANISMS BETWEEN WALL SURFACE AND FLUIDIZED BED

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Abstract—Following the film-penetration theory for mass transfer, originally developed by Toor and Marchello, a mechanism of heat transfer between fluidized beds and wall surfaces is proposed which includes both steady state conduction of heat through an emulsion layer at the wall and the unsteady-state absorption of heat by emulsion elements. A criterion is developed to suggest which mechanism controls, and the measured data are then compared with predictions of the bubbling bed model.

NOMENCLATURE

a, thermal diffusivity of emulsion layer [cm^2/s];
C_{ps}, heat capacity of solid [cal/g degC];
d_b, bubble diamter [cm];
d_p, diameter of particle [cm];
d_t, diameter of tower [cm];
f_b, time fraction that the surface is exposed to bubbles;
h_W, time-averaged heat-transfer coefficient [cal/cm^{2}s degC];
h_{wi}, local instantaneous heat-transfer coefficient [cal/cm^{2}s degC];
$I(t)$, age distribution function of emulsion elements on the surface;
k_e, effective thermal conductivity of emulsion layer [cal/cms degC];
l_e, effective thickness of emulsion layer [cm];
n, bubble frequency [s^{-1}];
t, time [s];
$\bar{t}$, mean age of the elements leaving the surface [s];
T_b, bed temperature [°C];
T_W, wall temperature [°C];
u_b, rise velocity of a bubble [cm/s];
u_0, superficial gas velocity [cm/s];
u_s, downward velocity of solids in dense phase [cm/s];
u_{mf}, superficial gas velocity at incipient fluidization [cm/s];
v_p, descending velocity of solid layer along the wall in Wicke's model [cm/s];
x, distance from heat exchange surface [cm];
z, see equations (14, 23, 25).

Greek symbols
α, ratio of wake volume to gas bubble volume;
δ, volume fraction of bubbles in fluidized bed;
ε_{mf}, void fraction of bed at minimum fluidization conditions;
ρ_e, apparent density of emulsion [g/cm^3];
ρ_s, density of solid [g/cm^3];
τ, see equation (17) [s];

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EXPERIMENTAL heat-transfer coefficients between exchanger surfaces and fluidized beds have been explained in terms of various mechanisms which may be classified as follows:

(1) Steady-state conduction of heat across the gas film, which is scoured by solids descending along the heat exchange surface: Leva *et al.* [1, 2], Dow and Jakob [3] and Levenspiel and Walton [4].

(2) Unsteady-state thermal conduction by single particles in direct contact with heat exchange surface: Botterill and Williams [5] and Ziegler *et al.* [6].

(3) Unsteady-state absorption of heat by fresh emulsion elements which are renewed intermittently by the violent disturbances in the core portion of the fluidized bed: Mickley *et al.* [7, 8].

(4) Steady-state conduction through the emulsion layer which is not often swept away: van Heerden *et al.* [9, 10] and Wicke and Fetting [11].

At the present state of knowledge, it would be useful to develop criteria to suggest which mechanism controls and which type of model should be used to represent a particular situation. In part this is done here. This paper presents a unification of mechanisms (3) and (4) and develops criteria to suggest when one or other of these mechanisms may be expected to apply. For a discussion of mechanisms (1) and (2) in relation with mechanisms (3) and (4) see Kunii and Levenspiel [12]. The consistency of the measured data with the bubbling bed model of Kunii and Levenspiel [13] is then considered.

THE BUBBLING BED MODEL

The bubbling bed model, developed recently, gives a simple representation of the bubble flow, the emulsion flow and the interaction of these streams in a fluidized bed.

When a bed is fluidized by gas at a superficial velocity, u_0, in excess of the minimum needed, u_{mf}, then gas voids, called bubbles, are seen to rise through a denser continuous region, called the emulsion. Assuming a constant bubble size in the bed or section of bed under consideration, the bubbling bed model then describes the flow in terms of one parameter, the effective bubble size d_b.

The velocity of rise of a bubble is given by

$$u_b = \frac{u_0 - u_{mf}}{\delta} = u_0 - u_{mf} + 0{\cdot}711\,(gd_b)^{\frac{1}{2}} \qquad (1)$$

where δ is the fraction of the bed consisting of bubbles.

The relation between the bubble frequency and the bubble diameter is given by

$$d_b = \frac{1{\cdot}5}{n}(u_b - u_{mf}), \qquad (2)$$

Rising bubbles are observed to drag behind them a wake of solids, consequently setting up a circulation pattern in the bed with downward solid flow in the emulsion. This downward velocity of solid is given by

$$u_s = \frac{\alpha\delta u_b}{1 - \delta - \alpha\delta}, \qquad (3)$$

where α is defined as

$$\alpha = \frac{\begin{pmatrix}\text{volume of wake dragged up the bed}\\ \text{behind a rising bubble}\end{pmatrix}}{(\text{volume of bubble})}.$$

This value can be estimated from the experimental data by Rowe and Partridge [14].

PROPOSED MODEL OF HEAT TRANSFER

In the core portion of a bubbling fluidized bed solids are mixed violently by gas bubbles and heat is transported by the solids at a high rate. On the other hand, at a vessel wall solids mostly move downward and form a solid sublayer which is replaced partially by fresh emulsion elements from the core portion of beds.

When a small heater such as a sphere or cylinder is immersed in a fluidized bed, rising bubbles hit the heater frequently, consequently the contact time of emulsion elements with the heat exchange surface is short. Therefore, we may expect solid elements to absorb heat by unsteady-state conduction. On the contrary, when a long wall surface is used as the heat exchanger, a steady temperature gradient will be set up across the emulsion layer, because the frequency of passing bubbles is small near the wall, and the contact time of solids is much longer. In the case of intermediate contacting time, both mechanisms should be taken into account.

The above situation is analogous to and can be explained by the film-penetration theory for gas absorption into liquids proposed by Toor and Marchello [15]. Consider a thin layer of emulsion of thickness l_e which suddenly contacts a heat exchange surface, and after a time t is suddenly replaced by a fresh element of emulsion from the core portion of the bed. The equation which represents this phenomenon is

$$\frac{\partial T}{\partial t} = a\frac{\partial^2 T}{\partial x^2} \qquad 0 \leqslant x \leqslant l_e, \tag{4}$$

where

$$a = \frac{k_e}{\rho_e C_{ps}} = \frac{k_e}{\rho_s C_{ps}(1 - \varepsilon_{mf})}.$$

Boundary conditions are

$$\begin{aligned} T &= T_b, \quad &&\text{at} \quad t = 0, \\ T &= T_W, \quad &&\text{at} \quad x = 0, \\ T &= T_b, \quad &&\text{at} \quad x = l_e. \end{aligned} \tag{5}$$

The bed temperature T_b and the wall temperature T_W are assumed to be independent of time.

The solution of equation (4) is

$$\frac{T_W - T}{T_W - T_b} = 2\sum_{i=1}^{\infty}(-1)^{i+1}\frac{\sin i\pi(1 - x/l_e)}{i\pi} \times \exp\left[-i^2\pi^2\frac{at}{l_e^2}\right]. \tag{6}$$

From the above solution, the instantaneous local heat-transfer coefficient, h_{Wi} is found to be

$$h_{Wi} = \left[\frac{k_e\rho_e C_{ps}}{\pi t}\right]^{\frac{1}{2}}\left[1 + 2\sum_{i=1}^{\infty} \times \exp\left\{-\frac{i^2 l_e^2}{at}\right\}\right] \tag{7a}$$

$$= \frac{k_e}{l_e}\left[1 + 2\sum_{i=1}^{\infty}\exp\left\{i^2\pi^2\frac{at}{l_e^2}\right\}\right]. \tag{7b}$$

Equations (7a) and (7b) are equivalent, but the former converges rapidly and is useful for short residence times on the surface, whereas the latter is useful for long residence times.

The observed coefficient of heat transfer h_W is the time averaged value of the instantaneous coefficient, and this is given by

$$h_W = \int_0^{\infty} h_{Wi} I(t)\,\mathrm{d}t, \tag{8}$$

where $I(t)$ is the age distribution function of emulsion elements on the surface. It is convenient to use one of two types of age distribution functions: first a random surface renewal, probably most representative of a surface in the main body of the bed continually contacted by rising bubbles, and secondly a uniform surface renewal, probably most representative of an emulsion flowing smoothly past the heating surface.

Case 1. *Random surface renewal*

For this case the age distribution of elements is given by

$$I(t) = \frac{1}{\bar{t}}\,\mathrm{e}^{-t/\bar{t}}, \tag{9}$$

where $\bar{t}$ is the mean age of the elements leaving the surface.

Replacing equations (7a, 7b, 9) in equation (8) gives the following expressions:

For rapid replacement, $(a\bar{t})^{\frac{1}{2}}/l_e < 1$,

$$h_W = \frac{k_e\rho_e C_{ps}}{\bar{t}}\left[1 + 2\sum_{i=1}^{\infty}\exp\left\{-\frac{2il_e}{\sqrt{a\bar{t}}}\right\}\right], \tag{10}$$

and for $(a\bar{t})^{\frac{1}{2}}/l_e < 0{\cdot}8$, within 20 per cent error,

$$h_W = \sqrt{\left(\frac{k_e\rho_e C_{ps}}{\bar{t}}\right)}. \tag{11}$$

For slow replacement, $(a\bar{t})^{\frac{1}{2}}/l_e > 1$,

$$h_W = \frac{k_e}{l_e}\left[1 + 2\sum_{i=1}^{\infty}\frac{1}{1 + i^2\pi^2 a\bar{t}/l_e^2}\right], \tag{12}$$

and for $(a\bar{t})^{\frac{1}{2}}/l_e > 1{\cdot}2$, within 20 per cent error,

$$h_W = \frac{k_e}{l_e}. \tag{13}$$

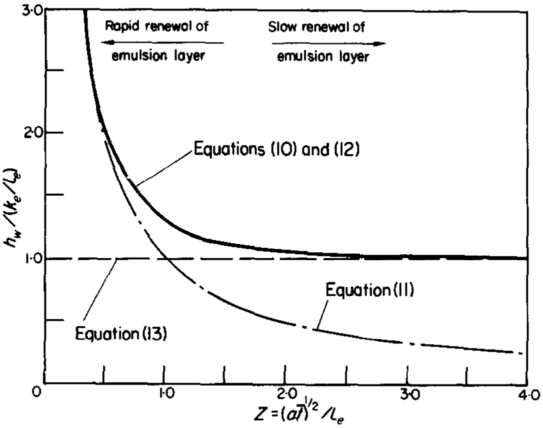

FIG. 1. Relationship between equations (10–13).

Toor and Marchello gave Fig. 1, illustrating the relationship between these equations. A criterion for estimating the controlling step is given by

$$Z = \frac{(a\bar{t})^{\frac{1}{2}}}{l_e} = \left[\frac{k_e\bar{t}}{\rho_s C_{ps}(1-\varepsilon_{mf})}\right]^{\frac{1}{2}}\Big/ l_e. \tag{14}$$

Thus:

$$\begin{pmatrix}\text{unsteady absorption into}\\ \text{emulsion elements controls}\end{pmatrix} \text{ when } Z \ll 1,$$
$$\begin{pmatrix}\text{steady-state transfer across}\\ \text{the emulsion layer controls}\end{pmatrix} \text{ when } Z \gg 1, \tag{15}$$

or in terms of the mean contact time

$$\begin{pmatrix}\text{short contact time and unsteady}\\ \text{state absorption controls}\end{pmatrix} \text{ when } \bar{t} \ll \tau,$$
$$\begin{pmatrix}\text{long contact time and steady}\\ \text{state transfer controls}\end{pmatrix} \text{ when } \bar{t} \gg \tau, \tag{16}$$

where

$$\tau = \frac{\rho_s C_{ps}(1-\varepsilon_{mf})l_e^2}{k_e}. \tag{17}$$

Case 2. Uniform surface renewal

For this case the age distribution of elements is given by

$$I(t) = \frac{1}{\bar{t}} \quad \text{for} \quad 0 < t < \bar{t},$$
$$I(t) = 0 \quad \text{for} \quad t > \bar{t}. \tag{18}$$

Replacing this function with equation (7) in equation (8) gives expressions similar to those for case 1:

For short contact times, or unsteady-state absorption controlling

$$h_W = \frac{2}{\sqrt{\pi}}\sqrt{\left(\frac{k_e\rho_e C_{ps}}{\bar{t}}\right)}, \tag{19}$$

and for long contact times, or steady-state transfer controlling

$$h_W = \frac{k_e}{l_e}. \tag{20}$$

Consider a surface immersed in a bubbling fluidized bed where the bubble frequency at the surface is n, and the time fraction that the surface is exposed to bubbles is f_b. If an element of emulsion stays in contact with the surface until it is swept away by a rising bubble, then the mean residence time of emulsion elements on the surface is

$$\bar{t} = \frac{1-f_b}{n}, \tag{21}$$

where f_b is given by the bubble fraction δ.

Assuming that heat transfer is negligible when the surface is bathed by bubbles, then equation (11) and equation (19) for rapid replacement of elements are modified to give the following expressions, respectively:

For random surface renewal

$$h_W = (1-\delta)\left(k_e\rho_e C_{ps}\frac{n}{1-\delta}\right)^{\frac{1}{2}}. \tag{22}$$

Criterion Z becomes

$$Z = \left(\frac{k_e\bar{t}}{\rho_s C_{ps}(1-\varepsilon_{mf})}\right)^{\frac{1}{2}}\Big/ l_e = \left(\frac{k_e(1-\delta)}{\rho_s C_{ps}(1-\varepsilon_{mf})n}\right)^{\frac{1}{2}}\Big/ l_e. \tag{23}$$

For uniform surface renewal

$$h_W = \frac{2}{[\sqrt{(\pi)}]}(1-\delta)\left(k_e\rho_e C_{ps}\frac{n}{1-\delta}\right)^{\frac{1}{2}}. \quad (24)$$

Criterion Z becomes

$$Z = \left(\frac{k_e(1-\delta)}{(2/\sqrt{\pi})^2\,\rho_s C_{ps}(1-\varepsilon_{mf})n}\right)^{\frac{1}{2}}\Big/ l_e. \quad (25)$$

Equation (24) is identical to the expression derived by Mickley *et al.* [8].

For slow replacement equation (13) or (20) gives

$$h_W = (1-\delta)\frac{k_e}{l_e}, \quad (26)$$

Hence, for a given physical situation and imposed flow rate u_0, this model gives the criteria of the transfer mechanism and the theoretical coefficients of heat transfer in terms of one measured quantity, the bubble frequency n or the bubble size d_b. Both parameters are connected by equation (2).

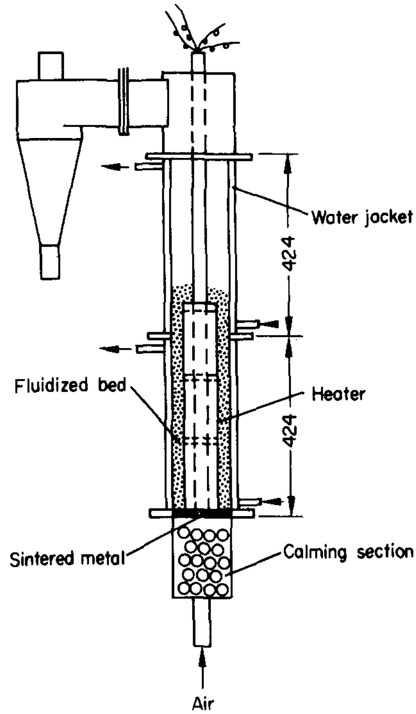

FIG. 2. Experimental equipment.

EXPERIMENTAL

An outline of the experimental equipment is illustrated in Fig. 2. Solid particles were fluidized by air in the annular space between the water jacket and the cylindrical electric heater. The heater was composed of three parts. The upper and lower sections were controlled by adjusting the electric inputs so as to avoid any axial loss from the middle section. Details of construction of the heater are shown in Fig. 3. From the measurement of temperature differences between the fluidized bed and the wall surface of the middle part of the heaters, heat transfer coefficients between the bed and the heater were calculated.

In addition, the frequency of bubbles was counted at the middle section of the heater at the same fluidizing conditions, but by using a transparent acrylic resin tube instead of the water jacket.

Properties of the fluidized solids are listed in Table 1.

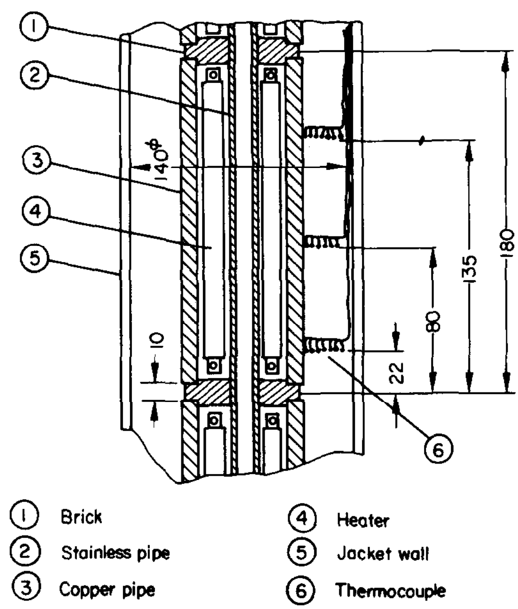

FIG. 3. Details of heater.

Table 1. Properties of particles

Solid	microspherical catalyst
Size distribution	100–250 μ
Mean size	152 μ
Density	1·54 g/cm^3
Heat capacity	0·22 cal/g deg C
Thermal conductivity	(3·33)(10)$^{-4}$ cal/cm s deg C
Minimum fluidization Velocity	2·0 cm/s
Void fraction at Incipient fluidization Conditions	0·505

COMPARISON OF THEORY AND EXPERIMENT

Obtained heat-transfer coefficients and bubble frequencies are shown in Figs. 4 and 5.

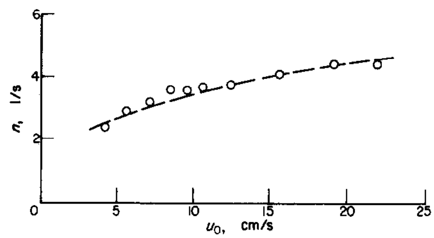

FIG. 4. Measured bubble frequencies.

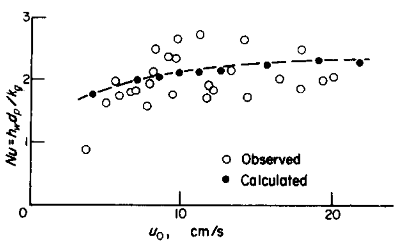

FIG. 5. Comparison of calculated results with observed data.

The criterion Z of equation (23) was calculated from these experimental data, where the value of l_e is assumed to be 2 mm. This gave

$$Z \leqslant 0{\cdot}1, \qquad (27)$$

which indicates a short contact time with unsteady state heating of the emulsion elements contacting the surface.

Heat-transfer coefficients were then calculated according to equation (22) and plotted in Fig. 5, where they can be compared with corresponding experimental values. The agreement is reasonable.

ANALYSIS OF PREVIOUS EXPERIMENTAL DATA

In order to calculate criterion Z or τ, it is necessary to know the thickness of emulsion layer, l_e which depends on the fluidizing conditions. Based on mechanism (4), Wicke and Fetting tabulated values for γl_e, where γ is the coefficient relating the downflow velocity of solids at the surface to the fluidizing velocity by $v_p = \gamma(u_0 - u_{mf})$. According to the bubbling bed model, the descending velocity of solid in the bed is given by equation (3). Assuming that v_p is approximated by u_s, we find that

$$\gamma = \frac{\alpha}{1 - \delta - \alpha\delta}. \qquad (28)$$

In Table 2 are shown the results of calculation of l_e from Wicke and Fetting's data by using equation (28), and assuming that δ is approximately 0·45 for all conditions. Even though the calculated l_e are scattered somewhat, these values may be useful guides for approximating l_e in these environments.

Table 2. Thickness of emulsion layer, based on Wicke and Fitting's data [11] and the predictions of the bubbling bed model

fluidization systems		particle diameter [cm]	superficial gas velocity [cm/s]	calculated by Eq. (28): l_e [cm]
SiC	Air	0·0065	0·7	0·047
SiC	CO_2	0·0065	1	0·032
SiC	Air	0·015	3	0·130
Sand	Air	0·0315	8	0·210
Sand	H_2	0·0315	16	0·130
Sand	CO_2	0·0315	12	0·080
Sand	Air	0·075	26	0·750
Al	Air	0·031	10	0·180
Al	H	0·031	21	0·10
Al	CO_2	0·031	17	0·07
Al	Air	0·045	16	0·23
Al	Air	0·075	24	1·23
Pb	Air	0·0125	18	0·041
Glass	Air	0·09	44	1·19

Substituting equations (1) and (2) into (22), gives

$$h_W = \sqrt{k_e \rho_e C_{ps}} \left[\frac{0{\cdot}711(g d_b)^{\frac{1}{2}}}{u_0 - u_{mf} + 0{\cdot}711(g d_b)^{\frac{1}{2}}} \times \frac{1{\cdot}5(u_0 - u_{mf})}{d^b} \right]^{\frac{1}{2}}. \qquad (29)$$

From the experimental data on heat transfer by Mickley and Fairbanks [7], Olin and Dean [16] and Wicke and Hedden [17] the bubble sizes d_b which are satisfied equation (29) can be estimated. Table 3 shows the computed results and they are consistent with the previous data on bubble diameter reported in the bubbling beds [18, 19]. In Table 3, the values of Z in equation (23) are shown together, by using Table 2 and the calculated bubble size.

In the above analysis, necessary numerical values of void fraction, apparent density and heat capacity of the emulsion layer have been assumed to be equal to those of the incipient fluidized bed, whereas the effective thermal conductivity of the packed bed [20] is used here instead of the value for incipient conditions, because there has been no reported data for the latter state.

In the experiments of Levenspiel and Walton [4], Campbell and Rumford [21] and Matsuyama *et al.* [22], the container wall was used as the heat exchange surface. For these conditions relatively long contact times of emulsion elements with the wall container as the heat exchange surface may be expected, even approaching that of steady-state conduction controlling. From the analysed values of l_e of Table 2 we find that the conditions for steady-state conduction are satisfied, i.e. $Z > 1{\cdot}0$ in equation (25), when the mean residence time of emulsion elements at the transfer surface is not less than 30 s. Thus the bed wall experiments probably lay in the intermediate region between these two mechanisms.

CONCLUSIONS

(1) Following the results reported by Toor and Marchello for mass transfer, a mechanism of heat transfer in the fluidized bed is proposed which includes both the mechanisms of thermal conduction across the emulsion layer along the surface and the intermittent renewal of emulsion elements.

(2) Criteria to determine the controlling step in the heat-transfer process are given by equations (14), (15), (23) and (25).

(3) Theoretical estimation of heat-transfer coefficients roughly coincide with experiment.

Table 3. Computed results from experimental data by equations (29) and (23)

Investigators	Mickley and Fairbanks [7]						Olin and Dean [16]		Wicke and Fetting [17]	
Experimental systems	glass beads–He, $d_t = 10{\cdot}1$ cm, $d_p = 0{\cdot}006$ cm, $u_{mf} = 3{\cdot}05$ cm/s		glass beads–CH_4, $d_t = 10{\cdot}1$ cm, $d_p = 0{\cdot}006$ cm, $u_{mf} = 1{\cdot}83$ cm/s		glass beads–air, $d_t = 10{\cdot}1$ cm, $d_p = 0{\cdot}006$ cm, $u_{mf} = 0{\cdot}4$ cm/s		sand–air, $d_t = 10{\cdot}1$ cm, $d_p = 0{\cdot}0137$ cm, $u_{mf} = 2{\cdot}0$ cm/s		glass beads–air, $d_t = 10$ cm, $d_p = 0{\cdot}081$ cm, $u_{mf} = 46$ cm/s	
$\sqrt{(k_e \rho_e C_{ps})}$ estimated [cal/cm^2 s$^{1/2}$ degC]	1·02		0·678		0·605		0·532		0·213	
Criterion equation (23)	$Z \leqslant 0{\cdot}25$		$Z \leqslant 0{\cdot}1$		$Z \leqslant 0{\cdot}2$		$Z \leqslant 0{\cdot}1$		$Z \leqslant 0{\cdot}1$	
	u_0 [cm/s]	d_b [cm]	u_0 [cm/s]	d_b [cm]	u_0 [cm/s]	d_b [cm]	u_0 [cm/s]	d_b [cm]	u_0 [cm/s]	d_b [cm]
Computed results	3·35	0·4	3·05	0·85	1·52	1·20	2·49	1·02	4	3·04
	5·80	1·52	7·62	3·52	5·49	3·07	6·30	2·62	14	4·60
	7·62	2·31	10·7	5·20	8·54	4·38	11·6	5·60	24	5·10
	9·15	3·02	16·8	6·85	12·2	5·86	15·3	6·50	29	5·10
	12·2	4·38	21·3	7·76	18·3	8·0				
	18·3	6·72	30·5	8·86	24·4	8·86				

Also, some previously reported data were analysed and were found to be consistent with the bubbling bed model proposed by Kunii and Levenspiel.

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Résumé—En accord avec la théorie de la pénétration de film pour le transport de masse, développée d'abord par Toor et Marchello, on propose un mécanisme du transport de chaleur entre les lits fluidisés et les surfaces des parois, qui contient à la fois la conduction de la chaleur en régime permanent à travers une couche d'émulsion à la paroi et l'absorption de chaleur en régime instationnaire par des éléments de l'émulsion. On développe un critère pour suggèrer ce que le mécanisme contrôle, et les résultats mesurées sont alors comparés avec les prédictions du modèle d'un lit bouillonnant.

Zusammenfassung—Nach der ursprünglich von Toor und Marchello entwickelten Filmdurchdringungstheorie wird ein Modell für den Wärmeübergang zwischen Flissbett und Wandoberflächen vorgeschlagen, dass sowohl stationäre Wärmeleitung durch eine Emulsionsschicht an der Wand, als auch instationäre Wärmeaufnahme durch Emulsionselemente umfasst. Ein Kriterium für den vorherrschenden Einfluss wurde entwickelt und die gemessenen Werte wurden mit den Berechnungen für das "bubbling-bed" Modell verglichen.

Аннотация—На основании теори пленочной проницаемости для массообмена, впервые разработанной Тооромм и Марчелло, предлагается механизм переноса тепла между кипящими слоями и поверхностями стенок, слагающийся из стационарного переноса тепла через эмульсионный слой на стенке и нестационарной абсорбции тепла элементами эмульсии. Разработан критерий для описания предложенного механизма. Данные измерений сравниваются с теоретическими расчётами согласно двухфазной модели кипящего слоя.