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Measurement of Heat Transfer Coefficient between Bubble Phase and Emulsion Phase

by
Shik Namkoong and Sei Ki Moon

Dept. of Chem. Eng., School of Eng., Hanyang University

氣泡相과 에멀존相間의 傳熱係數 測定

南宮寔* · 文世基**

ABSTRACT

Heat transfer coefficient between bubble phase and emulsion phase studied in the system of sand and γ-alumina particles fluidized by dry air. Gas velocity ratio to minimum fluidization velocity of the particles (U_0/U_{mf}) were taken in the range of 1.05 to 8.6, and the settled height of the particle to diameter of the bed (L_o/D_t) was fixed to 2, and the solid bed was heated to about 850°C. And heat transfer coefficients were measured by a steady state method analyzing the data of the temperature differences between bubble phase and emulsion phase occupying a high sensibility thermocouple probe. It was found that there is a linear correlation between $U_{mf}/\overline{D_b}$ and $\overline{H_b}$, here $\overline{D_b}$ and $\overline{H_b}$ indicate mean gas bubble diameter of the bed and mean over all internal heat transfer coefficient of bubble phase respectively, and that $\overline{H_b}$ are most available as a design data of the fluidized bed.

Of the behavior of gas solid fluidized bed a certain conclusion is supported by many research works[1] that the gas flows two paths; a rather small portion of the flowing gas works for the supporting of the particles, forming a dense phase, and the remaining major portion flows through the bed as rising gas bubbles reffered to as dilute phase or gas bubble phase. Appreciable interchange of flowing gas might occur between the two phases. Heat transfer between solid particles and the gas phase then must occur in more than one region. In this work, rate of heat transfer was analyzed based on rather simple two mechanisms. First the direct heat exchange between cycling gas and emulsion phase particles contacting in the region of "cloud zone"[2], and second the heat exchange between solids of cloud zone and solids of bulk emulsion zone[3][4] accompanied by random solid mixing in the quaking emulsion phase caused by the by-passing gas bubble. Therefore a heat transfer coefficient defined in terms of the temperature difference between emulsion and bubble phases as the driving force. In spite of importance of measurement of over all heat transfer coefficient of fluidized bed, the experimental result obtained by many investigations has a failure of large varieties. This fact is understood as a prime difficulties of fluidized bed reactor design because of lack of exact value of heat

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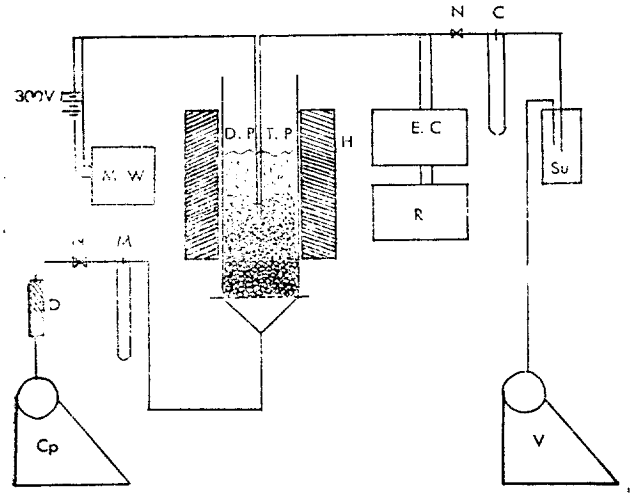
transfer coefficient at bottom of the reactor where we have unknown temperature distribution profile depending on complex flow pattern of gas and solid which is mainly affected by gas distribution device or mechanism. Here an exact device of measuring temperature of both gas bubble and contacting particles are sought. Yoshidome[5] and Yoo[6] have measured the temperature of fluidized bed placing a thermocouple device directly in the bed which has larger heat capacity by itself such that temperatures of solid and gas bubbles are not distinguishable. We have developed a high sensibility thermocouple device for this work and both temperatures were measured separately.



C; capillary flowmeter
Cp; compressor
D; dryer tube
$D. P.$; bubble frequency detector probe
$E. C.$; electric circuit
H; heater
M; orifice meter
$M. W.$: mini-writer
R; recorder
Su: suction tube
T. P; temperature detection probe
V; vacuum pump

Fig. 1. Schematic drawing of experimental arrangement

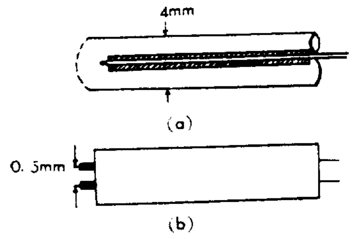


Fig. 2. High speed thermocouple probe(a) and leak-current detection probe(b)

Table 1. Operation condition

Fluidized bed	Cylndrical bed (5cmϕ×100cm)
Solid particles	SAND Size: 30—60 mesh D_p=4.23×10^{-2}cm ρ_p=2.63 U_{mf}=5.5 cm/sec α=0.039 ε_{mf}=0.54 Lo=9.4cm
	SAND Size: 80—150 mesh D_p=1.41×10^{-2}cm ρ_p=2.63 U_{mf}=2.5cm/sec α=0.21 ε_{mf}=0.45 L_o=11cm
	γ-ALUMINA(Neobead) Size: 30—60 mesh. D_p=4.23×10cm ρ_p=1.89 U_{mf}=2.8cm/sec α=0.033 ε_{mf}=0.24 Lo=11.1cm
Superficial gas velocities	5.8—25cm/sec
Fluid	air(15—20°C)

Experimental Apparatus and Procedure

A schematic drawing of experimental apparatus is shown in Fig. 1 and operation conditions are given in Table 1.

The cylindrical fludized bed, dimension 5cmϕ×100 cm stainless steel pipe and distributor with 0.5mmϕ, about 150 drilled holes were used. The used compressed dry air supplied through a needle valve and a calibrated oriffice manometer. After passing through the heated particles, the gas was released directly into the room. Part of it was drawn through the high speed thermocouple, with measured the gas temperature differences between emulsion phase and bubble phase gas from the bed. The voltage signal from the thermocouple was fed to a Hitachi mili-volt recorder calibrated to read temperature in °C. The thermocouple probe used for sampling the gas from the bed is shown in Fig. 2-a. The thermocouple was made with Chromel and Alumel wire with the junction located inside the probe tip about 4mm from the gas inlet. The probe body was made of stainless steel pipe with joints with sealed by asbestos. The tip was a 4mm o. d., 1mm wall stainless steel pipe, the closed end of which was perforated with 0.85×10^{-2} cm distance between the pushed end of the tip to allow sampling the gas. The hot gas was drawn through the tip in the end of the probe and past the thermocouple junction at high speed by a mechanical vaccum pump. The leak-current detection probe used for determing the bubble frepuency in the bed is shown in Fig. 2-b.

Theory

To derive a model for predicting heat transfer coefficients between emulsion phase and bubble phase in the gas-solid fluidized beds, the following assumptions are made;

1. The rate of heat transfer between the beds and the fluidized medium is very high.
2. The gas temperature in the emulsion phase is epual to the fluidized particle temperature.
3. The fluidized particle temperature in the bed is constant and homogeneous.

The heat balance epuation under these assumptions is

$$-V_b C_{pg}\rho_g \frac{dT_g}{d_t} = -V_b U_b C_{pg}\rho_g \frac{dT_g}{dl}$$

$$= \{(q_g C_{pg}\rho_g + h_{gi} S_b) + h_p \gamma_b \eta \pi \overline{D_p^2}\}(T_g - T_s)$$

$$-U_b C_{pg}\rho_g \frac{dT_g}{dl}$$

$$= \frac{\{(q_g C_{pg}\rho_g + h_{gi} S_b) + h_p \gamma_b \eta \pi \overline{D_p^2}\}(T_g - T_s)}{V_b} \quad (1)$$

where bubble surface area S_b is

$$S_b = \pi \overline{D_b^2}$$

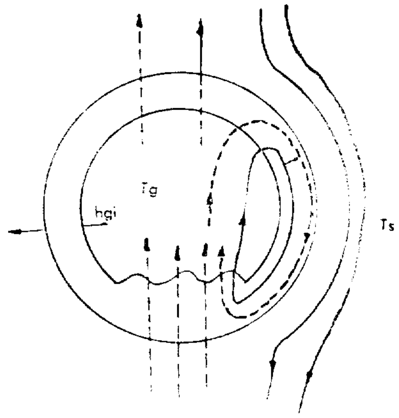


Fig. 3. Bubble model

bubble volume V_b is

$$V_b = \frac{4\pi}{3}\left(\frac{\overline{D_b}}{2}\right)^3 = \frac{\pi \overline{D_b^3}}{6},$$

$$h_{gi} = 0.975(k_g C_{pg}\rho_g)^{1/2}\left(\frac{g}{\overline{D_b}}\right)^{1/4}$$

gas circulation rate q_g is

$$q_g = \frac{270}{360}\pi \overline{D_b^2} U_{mf} = \frac{4}{3}\pi \overline{D_b^2}\ U_{mf}$$

Therefore, equation(1) is

$$-\frac{dT_g}{dl}$$

$$= \left\{\frac{4.5(U_{mf}/\overline{D_b})C_{pg}\rho_g + 5.85(k_g C_{pg}\rho_g)^{1/2} g^{1/4}/\overline{D_b}^{5/4}}{U_b\ C_{pg}\ \rho_g} \right.$$

$$\left. \frac{+ h_p \gamma_b \eta \dfrac{\overline{D_p^2}}{\overline{D_b^3}}}{} \right\} \times (T_g - T_s)$$

$$= \frac{\overline{H_b}}{U_b C_{pg}\rho_g}(T_g - T_s) \quad (2)$$

with boundary conditions, at $L=0$, $T_g = T_{gi}$, T_s=constant

where

$$\overline{H_b} = 4.5(U_{mf}/\overline{D_b})C_{pg}\rho_g$$

$$+ 5.85(k_g C_{pg}\rho_g)^{1/2} g^{1/4}/\overline{D_b}^{5/4} + h_p \gamma_b \eta \frac{\overline{D^2{}_p}}{\overline{D^3{}_b}}$$

If the gas temperature differences between emulsion phase and bubble phase are detected at one point, then,

$$\varDelta T = T_s - \left\{\left(\frac{\overline{D_b n}}{U_b}\right)T_g + \left(1 - \frac{\overline{D_b n}}{U_b}\right)T_s\right\}$$

$$\varDelta T = \left(\frac{\overline{D_b n}}{U_b}\right)(T_s - T_g) \quad (3)$$

from the bubble model, $(\overline{D_b n}/U_b)$ is expressed as follows,

$$\frac{A_t n}{(\pi/4)(\overline{D_b})^2} \cdot \frac{\pi}{6}\overline{D_b^3} = A_t \delta U_b,\quad \frac{\overline{D_b n}}{U_b} = \frac{3}{2}\delta$$

equation (3) takes the form

$$\Delta T = 1.5\delta(T_s - T_g) \qquad (4)$$

One can integrate equation(2) and substitute equation (4) to obtain

$$\ln\frac{(T_s - T_g)}{(T_s - T_{gi})} = \ln(\Delta T/\Delta T_o) = \frac{\overline{H_b}}{U_b C_{pg}\rho_g}\cdot l \qquad (5)$$

Experimental Results and Discussion

For each depth, particle size and flow rate, detected gas temperature differences between emulsion phase and bubble phase. ln $(\Delta T/\Delta T_0)$ are plotted in Figs. 4, 5 and 6.

From Figs. 4, 5 and 6, one can calculate the slope as given in Table 1, 2 and 3, and it is plotted in Fig. 7.

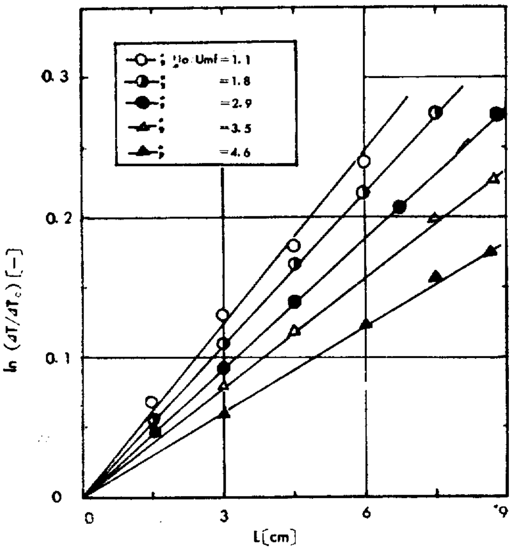


Fig. 4. ln$(\Delta T/\Delta T_0)$ vs l

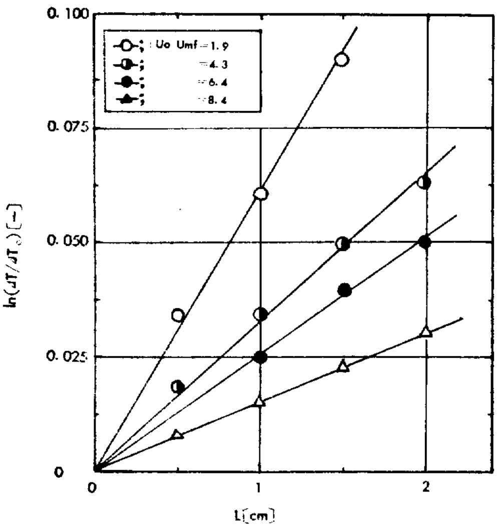


Fig. 5. ln $(\Delta T/\Delta T_0)$ vs l

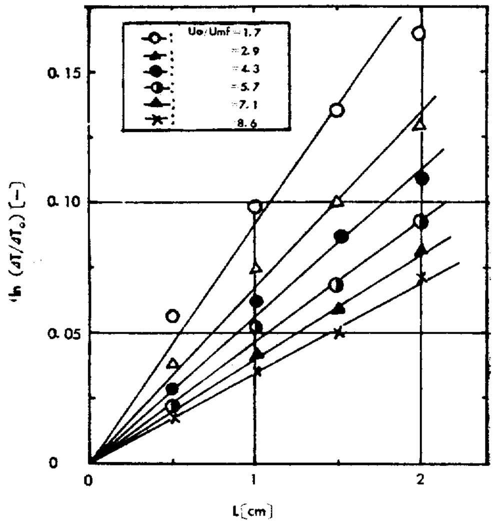


Fig. 6. ln$(\Delta T/\Delta T_0)$ vs l

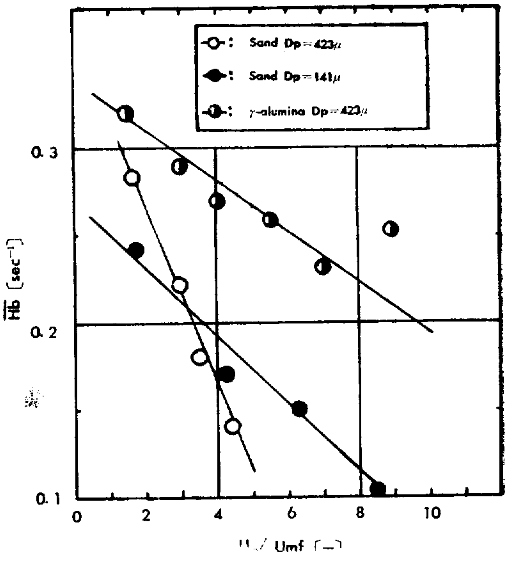


Fig. 7. $\overline{H_b}$ vs U_0/U_{mf}

Table 2. Measurement value of U_b, $\overline{D_b}$, and $\overline{H_b}/C_{pg}\ \rho_g U_b$ for the 30—60 mesh sand.

U_0/U_{mf}	[—]	1.05	1.8	2.9	3.5	4.6
$\overline{D_b}$	[cm]	0.25	1.72	3.40	4.35	6.30
U_b	[cm/sec]	42	48	60	65	74
$\overline{H_b}/C_{pg}\rho_g U_b$	[cm^{-1}]	6.2×10^{-2}	5.5	4.7	3.9	3.2

Table 3. Measurement value of U_b, $\overline{D_b}$, and $\overline{H_b}/C_{pg}\ \rho_g U_b$ for the 80—150 mesh sand

U_0/U_{mf}	[—]	1.9	4.3	6.4	8.6
$\overline{D_b}$	[cm]	0.73	1.99	3.02	4.13
U_b	[cm/sec]	50	70	85	95
$\overline{H_b}/C_{pg}\ \rho_g U_p$	[cm⁻¹]	6.0×10^{-2}	3.4	2.5	1.6

Table 4. Measurement value of U_b, $\overline{D_b}$ and $\overline{H_b}/C_{pg}\ \rho_g U_b$ for the 30—60 mesh Neobead

U_0/U_{mf}	[—]	1.7	2.9	4.3	5.7	7.2	8.8
$\overline{D_b}$	[cm]	1.05	1.85	3.0	4.2	5.5	7.0
U_b	[cm/sec]	47	59	70	78	81	98
$\overline{H_b}/C_{pg}\ \rho_g U_b$	[cm⁻¹]	9.2×10^{-2}	6.9	5.6	5.1	4.2	3.8

$\ln(\Delta T/\Delta T_0)$ are decreased rapidly with the increase of U_0/U_{mf}, the slope of $\ln(\Delta T/\Delta T_0)$ curve are increasing with the decrease of particle density as shown in Figs. 4, 5 and 6. For high superfical gas velocity, bubble rise upper direction of the bed with coalescence, where the center of the cylindrical fluidized bed is lower fluid friction portion than other portion of the bed, perhaps solid particles are sticked to the wall of the bed, so that the bed is separated perfectly with two region; bubble phase and emulsion phase. From this reason, we can find out that the $\ln(\Delta T/\Delta T_0)$ are decreased with the encrease of U_0/U_{mf}.

$\overline{H_b}$, heat transfer coefficients per unit bubble volume are plotted against $(U_{mf}/\overline{D_b})$ as shown in Fig. 8.

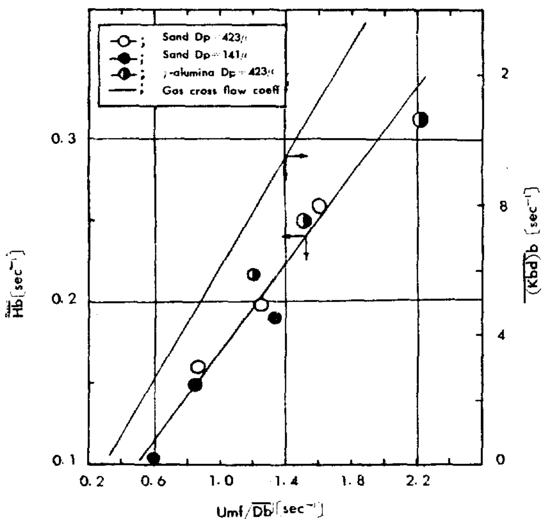


Fig. 8. Heat transfer coefficients $\overline{H_b}$ and gas cross flow coefficients $(\overline{K_{bd}})_b$ vs $U_{mf}/\overline{D_b}$

It is found that there is a linearlity between $\overline{H_b}$ and $U_{mf}/\overline{D_b}$; $\overline{H_b}\cong 0.14\,U_{mf}/\overline{D_b}$ and $\overline{H_b}$ value is about one-tenth of gas cross flow coefficients $(\overline{K_{bd}})_b$[6]

Conclusions

1. Observed correlation between heat transfer coefficient $\overline{H_b}$ and $U_{mf}/\overline{D_b}$ is as follows
   $\overline{H_b}\cong 0.14\,U_{mf}/\overline{D_b}$
2. Heat transfer coefficient $\overline{H_b}$ is about one-tenth of the gas cross flow coefficient $(\overline{K_{bd}})_b$ of same unit: [sec], and volumatric mixing rate of gas in the bed is estimated to be about several thousand times of that of solid, for heat capacity of sand and that of solid is about several times.
3. Factors of increase in the heat transfer coefficient is as follows,
   a. decrease in particle size
   b. decrease in particle density.

Nomenclature

A_t	cross sectional area of bed	[cm²]
$\overline{D_b}$	average bubble diameter	[cm]
$\overline{D_p}$	average particle diameter	[cm]
$\overline{H_b}$	heat transfer coefficient per unit volume of bubble	[sec⁻¹]
h_p	heat transfer coefficient of single particle	[kcal/cm⁻²hr°C]
K_g	thermal conductivity of gas	[kcal/cm hr°C]
$(\overline{K_{bd}})_b$	gas cross flow coefficient	[sec⁻¹]
L	height of measuring position	[cm]
L_0	static bed height	[cm]
n	bubble frequency	[sec⁻¹]
S_b	surface area of bubble	[cm²]
T_g	temperature of gas	[°C]
T_s	temperature of particles	[°C]
U_0	superficial gas velocity	[cm/sec]
U_b	average bubble rise velocity	[cm/sec]
U_{mf}	minimum fluidization velocity	[cm/sec]
V_b	volume of bubble	[cm³]
γ_b	particle density ratio	[g/cm³bubble/g/cm³ bed]
α	volume ratio of wake and bubble	[—]
δ	fraction of bed cross sectional area of bubble	[—]
ε_{mf}	void fraction of incipient fluidized bed	[—]
ρ_g	density of gas	[g/cm³]
ρ_p	particle density	[g/cm³]

Literature cited

1) Davidson, J. F., Paul, R. C., Smith, M. J. S., Duxbury, H. A.; *Trans, Instn. Chem, Engrs. London*, **37**, 323 (1959)

2) Rowe, P. N., Partridge, B. A., Lyall, E.; *Chem. Sci.*, **19**, 973 (1964)

3) Kunu, D. ;*Chem: Eng. (Japan)*, **21**, 342 (1957)

4) Namkoong, S., Moon, S. K.; *International Chemical Eng.* **6**, 4, 668 (1966)

5) Yoshidome; *Bull. TOKYO In. Tech. No.* 52

6) Namkoong, S., Yoo, K. O.: *J. of K. I. Chem. Eng.* **2**, 2, 82(1964)