CONVECTIVE HEAT TRANSFER IN A DYNAMIC TWO-PHASE BED

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An electrochemical method is used to investigate the average mass transfer coefficients and it is shown that the experimental data are correlated very satisfactorily by a relationship [4, 5] obtained on the assumption of no evaporation at the surface of the liquid film.

Heat-transfer processes in a dynamic two-phase bed are usually investigated on thermal models consisting of heat-transfer surface elements in the form of calorimeters heated by an electric current, a liquid heat transfer medium, or condensing vapor. The conduction of the experiments in this case involves several difficulties – averaging of the temperature over the heat transfer surface, the effect of heat loss through the ends of the calorimeter and the response time of the latter, the relative difficulty of constructing α calorimeters, and so on.

The most important feature, however, is that with this method of investigation it is impossible to distinguish the slowest stage in the over-all heat transfer process and, hence, to determine the role of evaporation. In view of this, some additional method of separating the convective and evaporative components of the heat transfer must be available. Integral investigations on thermal models are then used to determine the effect of evaporation and, ultimately, to determine the slowest stage of heat transfer.

The electrochemical method has definite advantages in this respect, since it allows the determination of the convective mass-transfer relationships with the hydrodynamic factors isolated from the over-all process. The electrochemical method was used in [1] to investigate convective mass transfer in the case of a flow of homogeneous liquid, and in [2] to investigate unsteady processes in a homogeneous liquid.

This method was first used and developed in 1967 by S. S. Kutateladze and his colleagues for the investigation of two-phase systems in the Laboratory of Physical Hydrodynamics, Institute of Thermophysics, Siberian Division of the Academy of Sciences of the USSR [3].

In the present work, proceeding from the premises and results of investigations conducted on thermal models [4, 5, 7], we used the electrochemical method to investigate convective mass transfer in the case of cylindrical surfaces and tube bundles in a dynamic bed. The average mass-transfer coefficient was measured on the experimental apparatus shown diagrammatically in Fig. 1a.

In the first series of experiments the experimental element 3, in the form of a clear plastic cylinder with $d = 4$ mm, was mounted in a 300 mm high column of square section ($100 \times 100\,\mathrm{mm}^2$) made of alkali-resistant clear plastic and fitted with a gas-distributing grid at the bottom. A nickel probe 4, 26.5 mm long ($335\,\mathrm{mm}^2$) was mounted flush with the cylinder surface. The experimental cylinder was worked on a lathe and then the probe surface was ground and polished. The cylinder 3 was positioned at a distance of 150 mm from the dielectric gas-distributing grid with a free cross section of 2.4%. A constant foam level (for a given gas flow rate) was maintained by replacement of the electrolyte entrained by the gas from the collecting tank 5 by means of pump 6.

Since the electrolyte was oxidized in the presence of oxygen and the decomposition potentials for oxygen and ferricyanide ions are the same, we used pure nitrogen as the light phase. The nitrogen was supplied from cylinder 8 through the rotameter 7 into chamber 9.

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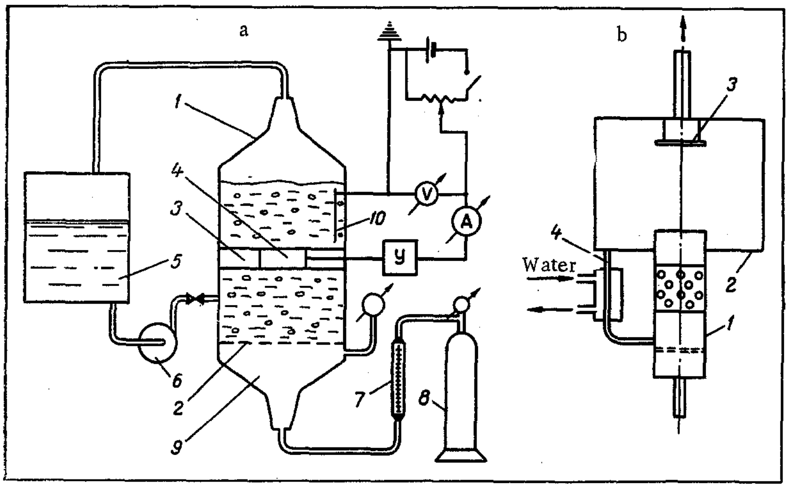

Fig. 1. Diagram of experimental apparatus (a) and apparatus for high velocities (b).

The electrochemical cell consisted of the probe-cathode 4 and an anode 10 mounted directly in the column. The experiments were carried out in the following sequence. With a prescribed nitrogen flow rate (i.e., for a given reduced velocity w_0'') we recorded the volt-ampere curve by means of a variable resistor and a 6 V battery and established maximum current operation. The signal from the probe was applied to a dc amplifier and then to a milliammeter.

For work at high velocities ($w_0'' > 0.2$ m/sec) some modifications were made to the apparatus (Fig. 1b) to prevent appreciable entrainment of the electrolyte. An expansion chamber 2 with a spray baffle 3 was attached to column 1; the entrained electrolyte was returned through a heat exchanger 4. It was confirmed during the conduction of the experiments that the electrolyte temperature t significantly affected its physical constants ν and D and, hence, the mass flow. With temperature fluctuations in the range 20 ± 5°C the mass flow varied by 2.5% per degree. The experiments were conducted at $t = 25$°C and the temperature was measured to an accuracy of 0.1°C by a thermocouple. The ferricyanide ion concentration in the solution was determined by volumetric chemical titration before and after each series of experiments.

To correlate the experimental data we used the formula obtained in [5] from experiments on thermal models in the form of cylindrical surfaces and also from the data of other authors on the assumption of no evaporation on the surface of the liquid film covering the heat-exchange element. The formula has the form

$$\overline{\mathrm{St}}^* = 0.15\,(\mathrm{Re}^*)^{-0.28}\,(\mathrm{Fr}'')^{-0.28}\,(\mathrm{Pr}')^{-0.56}. \tag{1}$$

The difference in the constant c and the indices of the powers of the groups in comparison with the formula given in [5] can be attributed to the fact that formula (1) was obtained from the correlation of experimental data for cylinders and spheres alone.

In the case of the electrochemical method the experimental mass transfer results were correlated by a similar relationship

$$\overline{\mathrm{St}}_D = f\,[(\mathrm{Re}^*\mathrm{Fr}'')^{\frac{1}{3}}\,\mathrm{Pr}_D^{\frac{2}{3}}], \tag{2}$$

where

$$\overline{\mathrm{St}}_D \equiv \overline{\beta}/w_0''; \quad \mathrm{Pr}_D \equiv \nu'/D.$$

Figure 2a compares the experimental results with relationship (1), which is shown by a continuous line. Figure 2 also shows the region of the experimental data obtained on thermal models.

Since a probe with $d = 4$ mm was used in the described experiments, and the NETI experiments showed that $\mathrm{St}^* \sim d^{-0.14}$, the effect of the probe diameter on mass transfer was examined. The experiments were conducted with single cylindrical probes of diameter $d = 2.5$, 8, 12, 16, and 20 mm; the results are shown in Fig. 2a, which reveals that all the experimental results are predicted fairly accurately by

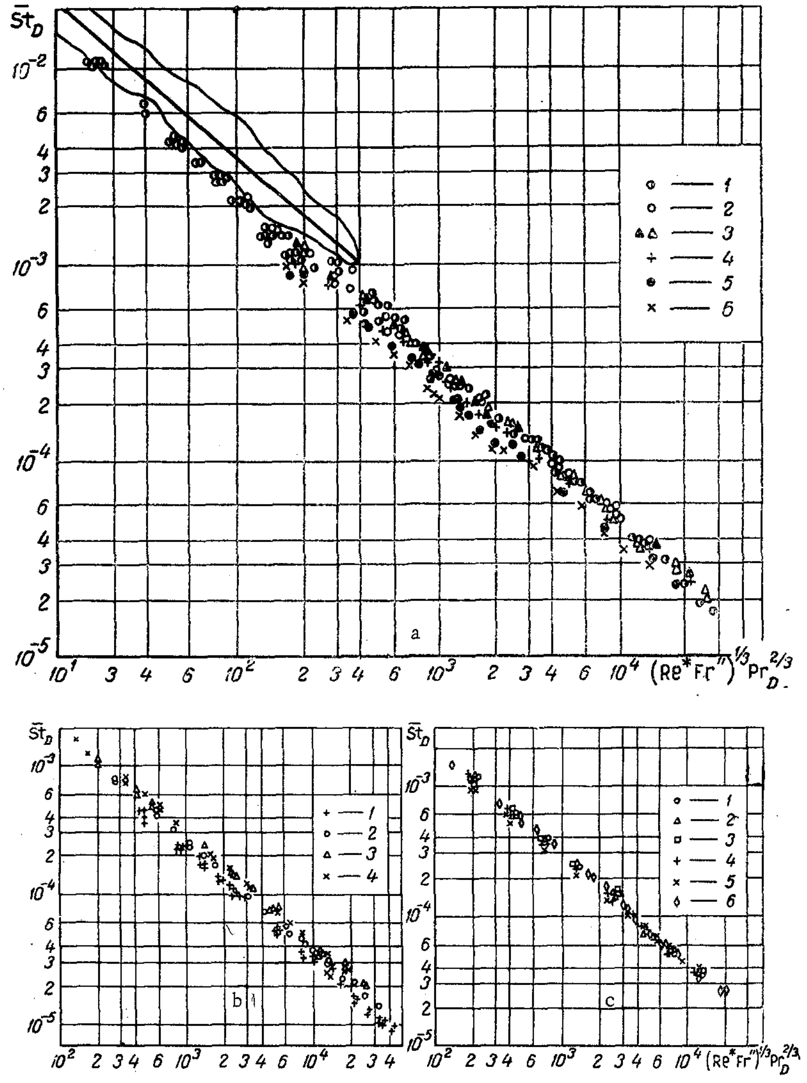

Fig. 2. Relationship $\bar{St}_D = f[(Re^*Fr'')^{1/3} \cdot Pr_D^{2/3}]$: a, 1) d = 4 mm; 2) 2.5; 3) 8; 4) 12; 5) 16; 6) 20; h_0 = 0.48 m. b, normality of NaOH (N = var) solution: 1) 6 N; 2) 4 N; 3) 2 N; 4) 0.5 N. c) 1) h_0 = 0.23 m; 2) 0.36; 3) 0.60; 4) 1.00; 5) 1.44; 6) 0.54.

relationship (2) up to d = 8-10 mm. A further increase in diameter leads to some stratification of the experimental data, and it can be assumed that $St_D \sim d^{-0.2}$. Thus, we have confirmed that the reduction of $\bar{\alpha}$ with increase in the diameter of the probe is due to the conditions of flow of the gas–liquid mixture. The velocity w_0'' in these experiments varied from $0.15 \cdot 10^{-2}$ to ~2.0 m/sec.

In the region in which we are interested St* differs from $\bar{St}_D$ by 10-15% on the average. This can be attributed to the effect of evaporation in the experiments on thermal models.

In the second series of experiments we investigated the mass transfer coefficient for individual tubes in a staggered tube bundle (seven horizontal rows and 14 vertical) with s/d = 3.5 (d = 4 mm) and found that calculations could be made from formula (2). The investigation was conducted on one tube bundle, since experiments on thermal models of bundles showed that when s/d > 3.5 the liquid film at the wall is responsible for most of the thermal resistance. It was necessary to confirm this result. The tube bundle was composed of glass cylinders and was mounted at a distance of 150 mm from the gas-distributing grid. The effect of the hydrodynamic conditions of flow of the gas–liquid mixture past the bundle was investigated by putting the experimental cylinder in different positions in the tube bundle. The experiments were conducted with the electrolyte at room temperature (t = 16-18°C). As a characteristic velocity we used the reduced velocity w_0'' in the narrow cross section of the bundle (up to ~0.2 m/sec). We found that for a single cylinder and individual tubes in a bundle with s/d = 3.5 the convective mass transfer varied in the same way, irrespective of the position of the experimental cylinder, i.e., the hydrodynamic conditions in the two cases were the same.

In the third series of experiments we used the same tube bundle with velocities in the narrow section of the bundle up to 0.2-3.35 m/sec. In these experiments the experimental cylinder was in a fixed position; the experimental results are shown in Fig. 2a.

We can conclude that heat transfer in a dynamic two-phase bed depends on the rate of heat transfer between the heat-transfer surface and the liquid film. The interphase surface in the gas–liquid bed is so extensive and evaporation is so rapid that it has practically no effect on the over-all rate of heat transfer.

In the main experiments in [4, 5, 7] $\mathrm{Pr}' = 4\text{-}60$, and to derive the correlation (1) we also used the experiments of other authors for $\mathrm{Pr}' = 10^4$. Since the diffusion criteria Pr_D in the case of the electrochemical method and the heat criteria Pr' in the main heat-transfer experiments differ considerably we investigated the effect of Pr_D on the mass-transfer rate. In these experiments we used a probe with $d = 8$ mm and an electrolyte in which Pr_D varied from $1.9\cdot10^3$ to $2.52\cdot10^4$.

This variation was achieved by using an electrolyte (NaOH solution) of several normalities (0.5, 2, 4, and 6 N) at various temperatures. The viscosity of the electrolyte was determined experimentally on a capillary viscosimeter kept at constant temperature by a TS-16 thermostat. We determined D from the formula used in [6], which has been tested in a wide range of temperatures and concentrations:

$$\frac{D\mu'}{T} = 2.5\cdot10^{-11}. \qquad (3)$$

The values of D agreed satisfactorily with the data of other authors who have determined D experimentally. Figure 2b indicates that for Pr_D up to $\sim5\cdot10^3$ the experimental results are correlated satisfactorily by relationship (2). At higher values of Pr_D the stratification of the experimental data is appreciable. Since all the mass-transfer experiments were carried out at relatively low Pr_D (values of a few thousand) we can assume that the mass-transfer processes by the electrochemical method and the heat-transfer experiments are practically similar.

An investigation of the effect of the height h_0 of the initial liquid layer on the mass-transfer rate for different velocities w_0'' is of definite interest, since the experiments were carried out with practically the same height h_0. The effect of h_0 is probably due to the onset at a particular velocity w_0'' of different hydrodynamic conditions due to the change in gas content of the two-phase bed. To ascertain the effect of h_0 we conducted experiments with a probe with $d = 8$ mm mounted at a distance of 0.215 m from the gas-distributing grid and with the height of the initial liquid layer varying from 0.23 to 1.44 m. As an electrolyte we used 0.5 N and 2 N NaOH solutions. An examination of Fig. 2c shows that h_0 has practically no effect on mass transfer.

Thus, we have established that the change in $\bar{\alpha}$ with change in h_0 in a dynamic two-phase bed must be due to a change in the interphase surface [7].

NOTATION

$\mathrm{Re}^* \equiv w_0'' d/\nu'$	is the Reynolds number referred to kinematic viscosity of liquid phase;
$\mathrm{Fr}'' \equiv w_0''^2/gd$	is the Froude number;
$\mathrm{Pr}' \equiv \nu'/a'$	is the Prandtl number;
$\overline{\mathrm{St}}^* \equiv \bar{\alpha}'/C_p'\rho' w_0''$	is the Stanton number referred to velocity of light phase;
$\mathrm{Pr}_D \equiv \nu'/D$	is the Prandtl diffusion number;
$\overline{\mathrm{St}}_D \equiv \bar{\beta}/w_0''$	is the Stanton diffusion number;
w_0''	is the reduced velocity of light phase, m/sec;
ν'	is the kinematic viscosity of phase, $\mathrm{m^2/sec}$;
μ'	is the dynamic viscosity of phase, $\mathrm{N\cdot sec/m^2}$;
D	is the molecular diffusivity, $\mathrm{m^2/sec}$;
d	is the probe diameter, m;
S	is the pitch between tubes of staggered bundle, m;
h_0	is the initial height of liquid layer, m;
N	is the normality of NaOH solution;
T	is the temperature, °K.

Indices

' and " denote the liquid and gas phases, respectively.

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