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MASS TRANSFER IN A HORIZONTAL GAS – LIQUID FLOW

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A method and results are presented for local mass transfer from a wall to a two-phase gas-liquid flow; electrochemical techniques are used.

Much research is currently being done on two-phase flows, mainly to provide a detailed description of the turbulent-transport processes and to provide an adequate physical model, which can ensure the development of better methods of calculating two-phase systems. This research on purely hydrodynamic characteristics such as the velocity distribution or the local gas content should be accompanied by research on the turbulent heat and mass transport, in view of its considerable interest. An important aspect is the analogy between the transport of momentum, heat, and mass in a two-phase flow. The rather scanty data on heat transfer in two-phase flows [1-7] are largely incomparable (see, e.g., [7]).

A detailed study has been made [8, 9] of the frictional stress at the wall in a horizontal two-phase gas–liquid flow; we have examined the behavior of the mass-transfer coefficient under similar conditions at large values of the Schmidt number. An electrochemical method was used [10] with an apparatus described in [8]. The working section was a horizontal tube of internal diameter 19 mm and length 6 m. The reduced velocity of the liquid varied over the range 0.1-4 m/sec, while the same for the gas was 0-110 m/sec.

The mixture was produced in a T-shaped mixer, with the liquid supplied through an annular slot of variable width under various conditions. The following flow conditions were implemented: stratified, bolus, bubble, and dispersed-annular (Table 1).

The working section of Fig. 1a was used to measure the local mass-transfer coefficient; this was a nickel block of length 230 mm with a hole of diameter 19 mm separated into two electrically insulated sections. The smaller section of width 6 mm acted as the cathode, while the larger part, which occupied the rest of the perimeter of the tube, was the anode. The part of the cathode adjoining the anode was insulated with a film of foam plastic of thickness 30-50 μm. The cathode contained local mass-transfer transducers (nickel wire of diameter 2 mm and plates of size 0.2 × 2 mm). These transducers were cemented into holes

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TABLE 1. Mass-Transfer Coefficients for a Two-Phase Flow Measured in a Single Section at 30° Intervals: 1 – Top, 4 – Side, 7 – Lower Generator

N%	W_0'=0,25 m/sec	Nu	W_0'=0,5 m/sec	Nu	W_0'=2 m/sec	Nu	W_0'=3 m/sec	Nu
1	Bubble W_0'' = 0,07 m/sec	502	Bubble W_0'' = 0,49 m/sec	2080	Bubble W_0'' = 0,17 m/sec	2235	Bubble W_0'' = 0,35 m/sec	2814
2		575		1980		2130		2877
3		551		1800		2010		2793
4		427		1700		1805		2709
5		402		1570		1755		2625
6		381		1400		1650		2583
7		410		1430		1695		2541
1	Plug W_0'' = 0,85 m/sec	675	Plug W_0'' = 1,48 m/sec	2120	Plug W_0'' = 1,4 m/sec	3195	Plug W_0'' = 1,1 m/sec	3646
2		778		2040		3225		3646
3		918		2000		3100		3589
4		1026		2150		2895		3493
5		877		2150		2805		3376
6		764		1940		2745		3275
7		726		1880		2700		3230
1	Bolus W_0'' = 11,8 m/sec	1822	Bolus W_0'' = 12,4 m/sec	4670	Bolus W_0'' = 8,4 m/sec	3960	Bolus W_0'' = 3,12 m/sec	4368
2		1844		4800		3965		4347
3		1962		4780		4065		4305
4		2092		5100		4095		4158
5		2138		5340		4095		3996
6		1811		4940		3965		3927
7		1630		4870		3960		3906
1	Annular W_0'' = 76 m/sec	5470	Annular W_0'' = 66,3 m/sec	9720	Dispersed W_0'' = 21,8 m/sec	5730	Rod W_0'' = 16,3 m/sec	6447
2		5480		9760		5790		6447
3		5562		9960		5835		6447
4		5643		10200		5835		6449
5		5697		10600		5834		6489
6		5750		10800		5805		6552
7		5805		10800		5800		6510

in the cathode with epoxide resin, which provided insulation; the thickness of the layer around the transducer was 5-10 μm. The transducers of diameter 2 mm were set at 3 and 5 times the diameter from the start of the cathode, while the rectangular ones were set at 10 diameters. In addition, the measuring head contained transducers for the frictional stress at the wall. The block after assembly was ground to the required size with the required surface finish, and then it was installed at 200 diameters from the mixer.

The working liquid was a solution in distilled water 0.5 N in caustic soda, 0.01 N in potassium ferrocyanide, and 0.001 N in potassium ferricyanide. This relation between the ferricyanide and ferrocyanide was chosen so that the diffusion of the ions to the anode had virtually no effect on the current flowing in the electrochemical cell, because the area ratio for the cathode and anode was about 1:30. The physical parameters of the electrolyte were stabilized by keeping the temperatures of gas and liquid at the inlet constant at 25 ± 0.2°C. The value of the Schmidt number for the system was Sc = 1300.

Figure 1b shows the measuring system. A dc potential was applied to the cathode and local transducers, which was adjustable in the range 0-1 V, while the differences between the voltages were not more than 1 mV; the voltages were monitored with a multirange meter. The current flowing to the electrochemical cell was measured with a special electronic circuit, which stabilized the cathode potential in spite of large fluctuations in the current, which occur with this two-phase mixture. The current to a local transducer was amplified with an electrodiffusion device [11], and the mean output voltage from the latter was measured with an integrator consisting of a V2-23 voltmeter and an F-5080 frequency meter. The integration time was 100 sec. The local mass-transfer coefficient was measured with the rectangular transducer at a distance of 10 diameters from the start of the mass-transfer part (the diffusion boundary layer is well developed at 4-5 diameters from the start [12-14]).

The mass-transfer coefficient was calculated from the measured current to a local transducer via

$$K_0 = I/AFC, \tag{1}$$

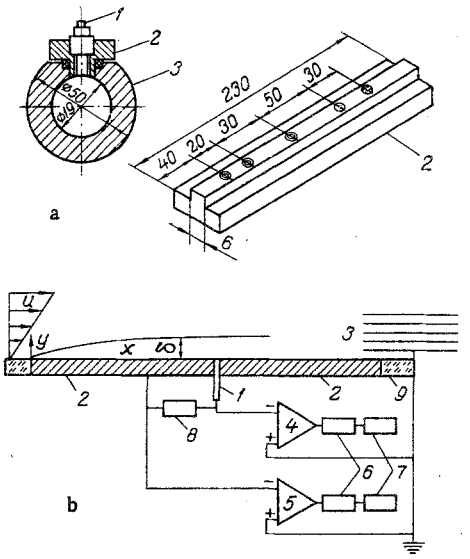

Fig. 1. a) Working unit: 1) local cathode; 2) cathode; 3) anode; b) measuring circuit: 1) local cathode, 2) cathode; 3) anode; 4) electrodiffusion converter; 5) current-measurement circuit; 6) integrating voltmeter; 7) frequency meter; 8) F-30 multirange meter; 9) insulating surface (lucite).

and the Nusselt number was given by

$$\mathrm{Nu}_0 = K_0 d/D. \tag{2}$$

The ferricyanide concentration C was determined by potentiometric titration; the operation of the transducers and measuring equipment was checked out with pure liquid. Figure 2 compares the measurements on the Nusselt number for a single-phase flow with a published formula [12],

$$\mathrm{Nu}_0 = 0.115\mathrm{Pr}^{0.25}\,\mathrm{Re}\sqrt{\lambda/8}, \tag{3}$$

as derived by the dissolution of benzoic acid and by the electrochemical method at large values of the Schmidt number ($4.8 \cdot 10^2 < \mathrm{Sc} < 1.2 \cdot 10^6$) for a single-phase flow. The deviations of Nu_0 from (3) were not more than 2%.

The gas–liquid flow in a horizontal tube is unsymmetrical on account of rise of the gas; therefore, there can be considerable variations in the local mass-transfer coefficient K_0 and Nusselt number Nu around the perimeter of the tube. The mass-transfer coefficient was measured at seven points on the half-perimeter of the tube at intervals of 30° by rotating the working channel around the axis. Table 1 gives the values of Nu measured for various flow conditions. There are substantial differences between the readings around the perimeter. The asymmetry in the flow decreases as the speed rises and vanishes in the dispersed-annular mode. The values for the mass-transfer coefficient given below were obtained by averaging the points in single section.

This method is better than the normal ring-cathode method because the resistance increases in a two-phase flow in the annular and dispersed-annular modes, and this can increase the error of measurement.

Figure 3 shows results for the mass-transfer coefficient in all conditions (the Nusselt number as a function of the ratio of the reduced velocities of the gas and liquid). The diffusion value of Nu for a two-phase flow is referred to the corresponding value for a one-phase flow Nu_0 as measured with a flow speed equal to the reduced speed of the liquid. It is clear that the observed points fit satisfactorily to two straight lines: for $W_0''/W_0' < 10$ (bubble and plug models) the result is

$$\mathrm{Nu}/\mathrm{Nu}_0 = (W_0''/W_0')^{0.63}, \tag{4}$$

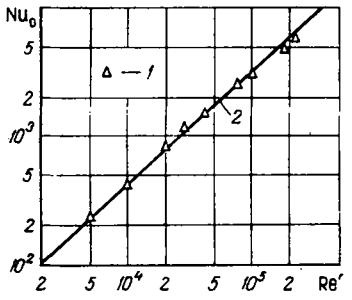

Fig. 2

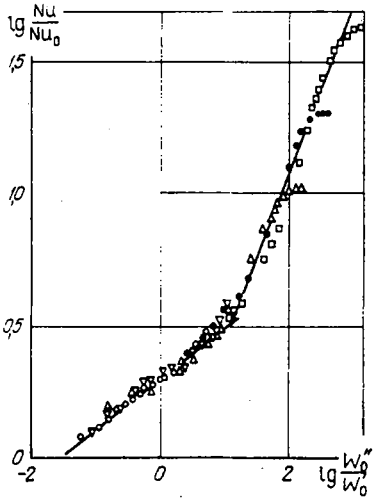

Fig. 3

Fig. 2. Calibration curve for mass-transfer transducer in a single-phase flow: 1) observed; 2) curve given in [12].

Fig. 3. Ratio of the Nusselt diffusion numbers for two-phase and single-phase flows as a function of the ratio of the reduced gas and liquid velocities: ●) Re = 5000; Δ) 10,000; ▽) 40,000; ○) 60,000; □) 2000.

and for $10 < W_0''/W_0'$ (bolus, annular, and dispersed-annular modes).

$$Nu/Nu_0 = (W_0''/W_0')^{2.42}. \quad (5)$$

At the high gas contents corresponding to the dispersed-annular mode, the points deviate substantially downward from the line, which has been reported previously [15] for this mode of flow.

The thickness of the diffusion layer is small at high Schmidt numbers, and all the changes in concentration occur very close to the wall; in a one-phase flow, there is a region near the wall that can be characterized by a single parameter, the dynamic velocity $v_* = \sqrt{\tau/\rho}$ [16, 17]. No evidence has been published on the distribution of the mean velocity or of the turbulent characteristics near the wall in a two-phase flow, so we do not know what parameters govern the conditions near the wall. It is not to be expected that the frictional stress τ at the wall or the dynamic velocity v_* are the only flow parameters under these conditions, although it may be that v_* is the basic parameter for a two-phase flow. It is therefore of interest to compare the measured values for the diffusion Nusselt number in a two-phase flow with the corresponding numbers for a one-phase flow in which the wall stress is the same. This may be called the equivalent Nusselt number Nu*, which is calculated as follows. From the measured frictional stress at the wall in the two-phase flow we determine the Reynolds number Re* for the one-phase turbulent flow with the same value of τ; Nu* was determined from this value of Re* via Fig. 2. The result for the ratio Nu/Nu_0 is shown in Fig. 4 as a function of W_0''/W_0'; for $W_0''/W_0' > 0.5$ m/sec (Re ≥ 10,000), all the results lie in the region $1 < Nu/Nu_0 < 1.5$, and here there is an analogy resembling the Reynolds analogy for the heat and mass transfer. We also show here Johnson's data for the heat transfer to a horizontal oil–air flow [2] (Prandtl number in the range 300-400, channel diameter 18.7 mm) and for water–air [1] (Prandtl number 3-5, channel diameter 22.2 mm).

The values of Nu, Nu*, Re*, Re for the heat carrier were determined from the results of [16, 17] as for the mass transfer.

Clearly, the heat-transfer results for the two-phase flow at high Prandtl numbers are in satisfactory agreement with the mass-transfer data; the results of [1] for the water–air system at $W_0''/W_0' < 5$ deviate upward from the general trend, evidently because the thickness of the thermal boundary layer at small Prandtl numbers is substantially greater than that at high Pr, which may disrupt the analogy in the transport processes. Also, there is an appreciable deviation of the data of [1, 2] for the dispersed-annular mode ($W_0''/W_0' > 20$),

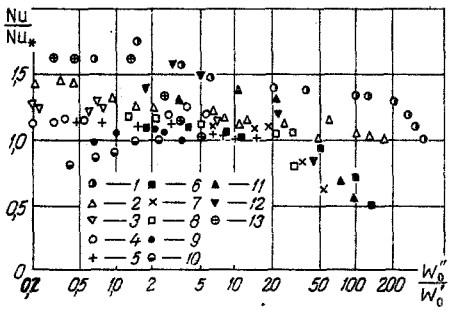

Fig. 4. Comparison of measurements of the heat and mass transfer in single-phase and two-phase flows for the same value of the dynamic velocity: mass transfer (our results): 1) Re = 5000; 2) 10,000; 3) 40,000; 4) 60,000; 5) 80,000; heat transfer (oil–air [2]): 6) Re = 260; 7) 520; 8) 780; 9) 1230; 10) 1840; (water–air [1]): 11) Re = 12,800; 12) 25,600; 13) 118,000.

which may be ascribed to the substantial loss of pressure in the acceleration. This means that the frictional stress calculated from the complete pressure difference is more than the actual value, which was not measured. As a result Re_* and Nu_* are too high, while Nu/Nu_0 is too low.

NOTATION

A	is the transducer area, m^2;
K_0	is the mass-transfer coefficient, m/sec;
I	is the transducer current, mA;
F	is the Faraday number, C/g-eq;
C	is the ferricyanide concentration, g-eq/m^3;
d	is the channel diameter, m;
D	is the diffusion coefficient, m^2/sec;
W'_0, W''_0	are the reduced liquid and gas velocities, m/sec;
u	is the transverse velocity component, m/sec;
v_*	is the dynamic velocity, m/sec;
τ	is the tangential wall stress, N/m^2;
ρ	is the density, kg/m^3;
y	is the transverse coordinate, m;
x	is the longitudinal coordinate, m;
δ	is the diffusion boundary-layer thickness, m;
Nu_0	is the diffusion Nusselt number;
Re'	is the Reynolds number;
Sc	is the Schmidt number;
Pr	is the Prandtl number.

Subscripts

0	one-phase;
*	equivalent.

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CHOICE OF PULSE-FLUIDIZATION CONDITIONS FOR MIXING GRANULATED MATERIALS

UDC 66.096.5

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A method is given for determining the best pulse-fluidization conditions for accelerating the mixing of components in a granular material.

These measurements on the mixing kinetics of granular materials in pulse fluidization provide evidence on the rate of the process in relation to the working conditions, i.e., the frequency f, mark-space ratio S, and gas flow rate. Visual observations show that rapid mixing corresponds to a definite state of the layer, which may be characterized as the active-piston state. This occurs if the maximum pressure under the bed during the pulse does not exceed $(1.5\text{-}2.5)Mg/S_e$ and causes bed expansion, with only a small amplitude of the oscillation of the upper boundary relative to the mean position and the absence of ejection above the bed. This is termed region A (Fig. 1) in S–f coordinates. In region B, there are large bubbles, whose escape into the space above the bed is accompanied by considerable ejection of material. In region C, the layer remains largely immobile and very little particle displacement occurs.

However, the mixing rate is not constant within region A for any combination of f and S; there is a definite pair of values for f and S (in our case f = 5 Hz and S = 0.5) that produces the most vigorous mixing. This has been observed elsewhere [1], where the effect was described in terms of a natural oscillation frequency of the bed.

The hydrodynamic features of the bed were examined by cinephotography in parallel with gas pressure measurement under the distribution grid. Figure 2 shows characteristic pressure oscillograms for f of 1-5 Hz

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