CONVECTIVE HEAT AND MASS TRANSFER IN THE NONADIABATIC EVAPORATION OF VOLATILE LIQUIDS FROM A POROUS PLATE

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This paper describes the apparatus, the experimental procedure, and the calculation of the characteristics of convective heat and mass transfer in the nonadiabatic evaporation of volatile liquids into a boundary layer from a flat porous plate situated longitudinally in a turbulent flow of air.

The change from liquid to gas is a feature of several widely employed processes and apparatuses in the chemical industry. An analysis of the local heat (enthalpy) and mass flows in the general hydrodynamics field is very complicated even in the simplest case of adiabatic evaporation of water [1]. In the scientific literature at present there are several controversial, and sometimes quite contradictory, views on the effect of mass transfer on heat transfer [2, 3]. There is no efficient and reliable method of determining the convective heat transfer coefficient when mass transfer is involved.

Experimental investigations [4, 5] have been based on the assumption of the adiabatic nature of the process, i.e., it is assumed that all the heat supplied from outside by convection is used up on the phase change and is returned to the boundary layer with the enthalpy of the vapor. It is well known, however, how difficult it is in experiments to secure stable and reproducible evaporation conditions. These difficulties are increased in the case of volatile liquids with low heats of transformation. Here we find that the flow rate of evaporated liquid depends very greatly on various kinds of nonconvective effects, and on the internal structure and physicochemical properties of the porous permeable material. In our experiments on the evaporation of acetone on permeable metal plates (Cr23Ni18 steel with a voidage of 40–43%) we found an appreciable difference in the temperature distribution according to different measurements of the plate. The inhomogeneity of the temperature field can be attributed to the strong effect of the structure of porous material on the internal flow of liquid and vapor.

For an incoming flux with the same parameters the equilibrium temperature of adiabatic evaporation of volatile liquids is lower than that of water. Hence, the temperature of the wall and the medium close to it can be lower than the "dew point" for water vapor, which is almost always present in the air. Then in the boundary-layer region adjacent to the phase interface bulk and surface condensation begins and is accompanied by additional heat release. In our experiments (carried out in summer, when the absolute moisture content of the air is very high), we visually observed abundant precipitation of water on the surface of the plate in the air flow. In some regions there were films of hoar frost or ice. Precipitation of water on the free surface of volatile liquids took place in the experiments in [6]. Hence, new components due to the heat of condensation, ice formation, and mixing of the liquid appear in the energy balance of the wall. This inevitably leads to a redistribution of the enthalpy and mass flows crossing the washed solid surface.

The curve of saturated vapor pressure of volatile liquids is very steep. Hence, even small deviations from equilibrium temperature have a pronounced effect on the partial pressure and concentration in the evaporation zone. In such conditions the use of the relationships for adiabatic processes leads to appreciable errors in the generalization of the results of investigations of external convective heat and mass transfer.

Local heat and mass transfer in the nonadiabatic evaporation of volatile liquids in a forced flow of heated gas was experimentally investigated in the following way. Liquids (ethyl alcohol, benzene, acetone), which had been thoroughly purified by double distillation and deaerated, were delivered from inside to the external surface of a porous plate. The design of the experimental model ensured that evaporation took place directly from the external surface of the specimen (without extension of the evaporation boundary into the porous body). The most suitable homoporous material was found to be a layer of slag-cotton spherules 0.3–0.5 mm in diameter [7]. Such a layer has a very homogeneous structure and can easily be reproduced. Clogging of the pores with liquid impurities is easily removed by mixing, washing, or replacement of the used batches. The spherules were contained in a rectangular quartz-glass holder divided into five compartments. The surface of the layer was leveled with a special template and its horizontality was checked with a spirit level. We thus obtained a five-sectioned porous plate 70 mm wide, 12 mm deep, with section lengths $l_1 = 25$ mm, $l_2 = 34$ mm, $l_3 = 39$ mm, $l_4 = 49$ mm, $l_5 = 54$ mm in the direction of the flow. The investigated liquid was fed through tubing into individual connections in the bottom of each section. Since each vessel was mounted on the platform of a VTK-500 dial scale the flow of evaporated liquid was continuously recorded. A special drainage system ensured the absence of air bubbles in the porous material.

Evaporation from the outer surface was achieved by varying the hydrostatic head of the liquid. The material of the model did not allow us to conduct an experiment even with a minimum extension of the

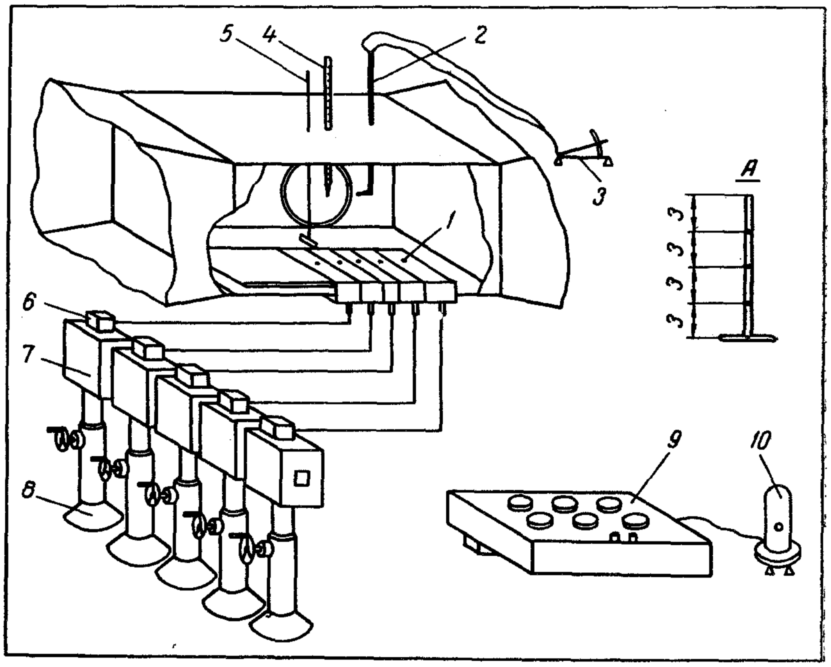

Fig. 1. Diagram of working section of wind tunnel: 1) experimental plate; 2) Pitot-Prandtl tube; 3) MMN-0.5 micromanometer; 4) thermometer; 5) microthermocouple; 6) reservoir; 7) VTK-500 balance; 8) coordinator; 9) R-306 potentiometer; 10) galvanometer; A) arrangement of thermocouples for measuring wall temperature in each section of model.

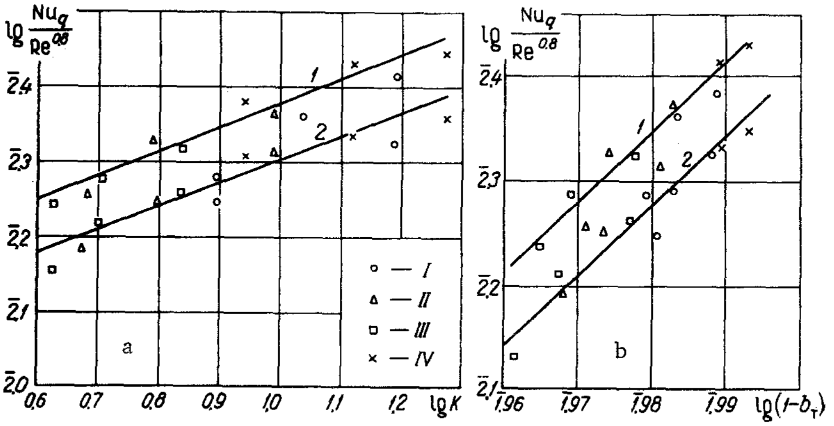

Fig. 2. Mean and local heat transfer as a function of a) $K = r/c_L\Delta T$; b) parameter $(1 - b_T)$: 1) mean; 2) local; I) alcohol; II) benzene; III) acetone; IV) water [5].

evaporation zone. When the top of the layer dried out the surface of the specimen rapidly disintegrated and the spherules were carried off in the flow. When the head was increased, the streams of filtering liquid washed away the surface of the model and made it uneven. We carried out special experiments to determine the range of optimum hydrostatic heads for which a change of level in the vessel had no effect on the flow rate of the evaporated liquid. The indicator of the balance made it possible to fix precisely and maintain the required head. Through windows in the side walls of the working section we could visually observe the state of the surface.

Nonconvective flows of heat to the model walls were reduced by putting the working body in a water jacket. A turbulent boundary layer was produced at the very start of the active mass-transfer surface by the attachment of a rough projection (l = 0.5 m) to the front edge of the porous plate.

The temperature at 25 points on the porous plate and the temperature of the side walls of the model and the working section of the wind tunnel were recorded by copper-constantan thermocouples. Their thermal emf was measured by an R-306 low-resistance potentiometer. The hot junctions of the thermocouples were mounted on a wooden post (0.5 × 2 × × 12 mm), fitted to a special base, in each section (Fig. 1). A copper-constantan microthermocouple 0.05 mm in diameter, butt-welded by the capacitor discharge method, without the head, was used to monitor the surface temperature on the flow side.

During the experiments we measured the velocity of the incoming flow, the distribution of static pressure along the model, and the temperature of the thermometers wetted by the investigated liquid and by water. For different regimes we recorded the temperature and velocity distributions in the boundary layer.

The investigations were carried out in a closed-circuit continuous-action wind tunnel with a closed working section [5]. A constant concentration of the volatile substance in the circuit was secured by providing for partial air exchange with the atmosphere. The flow velocity was regulated by a system of gates and slide valves. The air was heated to the required temperature by an electric heater, two sections of which were connected to the automatic control system, which maintained the prescribed temperature to within ±0.2° K.

The experiments with each liquid involved three hydrodynamic and three thermal regimes: Re_∞ varied from $2.83 \cdot 10^5$–$4.23 \cdot 10^5$ and the temperature T_∞ of the incoming flow from 323–353° K.

The convective heat transfer coefficient α_u was calculated from the heat-balance equation

$$\alpha_u = \frac{\dot{m}''r \pm c_1 \dot{m}''(T_1 - T_w) - q_r \pm \Sigma q_\tau}{T_\infty - T_w}, \tag{1}$$

where $\dot{m}''r$ is the amount of heat spent on evaporation of a specific flow of liquid; $c_1\dot{m}''(T_1 - T_w)$ is the amount of heat spent on heating the liquid from the temperature at the entrance to the evaporation temperature; q_r is the radiant heat flux; Σq_T is the sum of the heat losses and leakage of heat through the wall of the model.

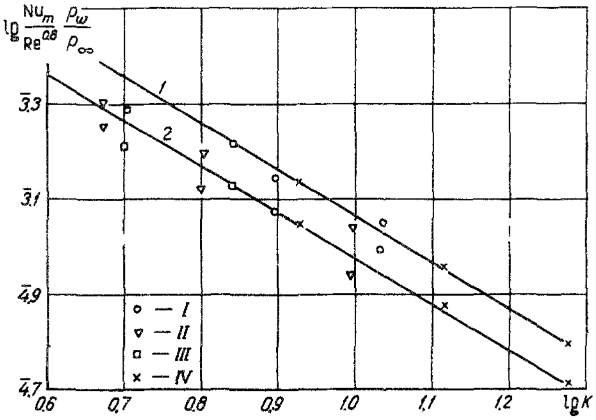

Fig. 3. Mean and local mass transfer as a function of $K = r/c_L\Delta T$: 1) mean; 2) local; I) alcohol; II) benzene; III) acetone; IV) water [10].

The radiant component was reduced by water cooling of the walls of the working section of the wind tunnel. The temperature of the section walls even in the hottest conditions did not exceed the plate temperature by more than 4–6° K, and the value of q_r was not more than 1% of the total heat flux.

A calculation of Σq_T showed that the low thermal conductivity of the layer material and holder of the model and the use of the water jacket reduced the nonconvective components to a minimum: 0.8–1.5% of the total heat flux.

The temperature of the evaporation surface for all the analyzed regimes was above the dew point of water vapor in the air flow and, hence, in the calculation of α_u thermal effects due to condensation of water vapor were ignored. The local values of the specific mass flow and the mass transfer coefficient was determined from the relationships

$$\dot{m}'' = \frac{m_x}{F\tau}, \tag{2}$$

$$\alpha_m = \frac{\dot{m}_x''}{z_{1w} - z_\infty}. \tag{3}$$

The concentration of vapor on the surface was deter-from the equation of state on the assumption that the evaporating liquid is in thermodynamic equilibrium with its vapor. The vapor concentration in the incoming flow was neglected in view of its smallness. In the calculation of the values of the similarity criteria the temperature T_∞ was taken as a characteristic temperature, and the linear dimension was measured from the start of the working section to the end of the averaging zone [8].

Treatment of the experimental data showed that for all thermal and hydrodynamic regimes there was a steady reduction of the heat and mass transfer coefficients along the plate, with the most rapid reduction occurring in the front part of the porous plate. In the same temperature conditions all the liquids showed stratification of the experimental Re numbers, Local and Mean Heat and Mass Transfer Coefficients in Relation to Temperature of Incoming Flow

T, °K	Alcohol		Benzene		Acetone	
	heat transfer	mass transfer	heat transfer	mass transfer	heat transfer	mass transfer
323.15	$\mathrm{Nu}_{qx} = 0.0212\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.0119\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{qx} = 0.0207\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.0097\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{qx} = 0.0182\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.011\,\mathrm{Re}_x^{0.8}$
	$\mathrm{Nu}_q = 0.0245\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0137\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_q = 0.0236\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0113\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_q = 0.0212\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0118\,\mathrm{Re}_\infty^{0.8}$
338.15	$\mathrm{Nu}_{qx} = 0.0196\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.0142\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{qx} = 0.0177\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.0141\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{qx} = 0.0164\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.0121\,\mathrm{Re}_x^{0.8}$
	$\mathrm{Nu}_q = 0.0229\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0164\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_q = 0.0215\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0162\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_q = 0.0193\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0140\,\mathrm{Re}_\infty^{0.8}$
353.15	$\mathrm{Nu}_{qx} = 0.0181\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.0173\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{qx} = 0.0157\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.0149\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{qx} = 0.014\,\mathrm{Re}_x^{0.8}$	$\mathrm{Nu}_{mx} = 0.01\,\mathrm{Re}_x^{0.8}$
	$\mathrm{Nu}_q = 0.0190\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0197\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_q = 0.0184\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0169\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_q = 0.0162\,\mathrm{Re}_\infty^{0.8}$	$\mathrm{Nu}_m = 0.0116\,\mathrm{Re}_\infty^{0.8}$

and in the case of the same Re numbers they showed stratification as regards temperature of the air flow. The table gives the approximating formulas. The index of the power of the Re number and the coefficients in the formulas were obtained by applying the method of least squares to the results.

The table shows that an increase in the rate of transverse mass flow has different effects on the heat and mass transfer. With increase in the transverse flow the mass transfer is increased, whereas the heat transfer is reduced. The stratification of the experimental data depends not only on the regime parameters but also on the physical properties of the evaporating liquid. Hence, in the subsequent treatment we used the criterion $K = r/(c_l \Delta T)$, which takes into account the phase change, the physical properties of the transported substance, and the temperature conditions.

The experimental heat transfer data lie on the approximating straight line with a spread of not more than ±4% (Fig. 2a) and are given by the following expressions:

$$\mathrm{Nu}_{qx} = 0.01\,\mathrm{Re}^{0.8} K^{0.3},$$
$$\overline{\mathrm{Nu}}_q = 0.0118\,\mathrm{Re}^{0.8} K^{0.3}. \qquad (4)$$

Figure 2a also gives the results of an investigation of the local heat transfer on evaporation of water from the surface of a capillary-porous plate [5].

A fact which merits attention is that the obtained relationships (4) can be used for calculations of the evaporation of liquids other than those used. However, (4) cannot be extrapolated for the calculation of pure heat transfer. This fault can be overcome by introducing a relative mass transfer parameter which takes into account the effect of the transverse flow of substance in the turbulent boundary layer (Fig. 2b):

$$b_{\mathrm{T}} = \frac{c_{p_n} \rho_w u_w}{c_{p_\infty} \rho_\infty u_\infty \mathrm{St}_0}.$$

Then

$$\mathrm{Nu}_{qx} = 0.0297\mathrm{Re}^{0.8}(1 - b_{\mathrm{T}})^9,$$
$$\overline{\mathrm{Nu}}_q = 0.0331\mathrm{Re}^{0.8}(1 - b_{\mathrm{T}})^9. \qquad (5)$$

When $b_{\mathrm{T}} = 0$ Eqs. (5) are very similar to the known expressions for heat transfer in the absence of evaporation [9].

The investigation and calculation of the mass transfer process showed that the experimental data are given by the approximating relationships (Fig. 3)

$$\mathrm{Nu}_{mx} = 4.07 \cdot 10^{-3}\mathrm{Re}^{0.8} K^{-0.6} \left(\frac{p_\infty}{p_{1w}}\right),$$

$$\overline{\mathrm{Nu}}_m = 4.75 \cdot 10^{-3}\mathrm{Re}^{0.8} K^{-0.6} \left(\frac{p_\infty}{p_{1w}}\right),$$

where p_∞/p_{1w} is a simplex expressing the ratio of the total pressure in the flow to the partial vapor pressure on the phase interface. Figure 3 also shows results of Fedorov's mass transfer investigations [10], which we treated by the method described. It is obvious that relationships (6) can be recommended for calculation of the mass transfer of different liquids.

NOTATION

α_u is the convective heat transfer coefficient; m" is the specific transverse flow of substance; c is the specific heat; r is the heat of phase transition; T is the temperature; z is the concentration of substance; α_m is the convective mass transfer coefficient; Re is the Reynolds number; Nu_q is heat transfer Nusselt number; Nu_m is the mass transfer Nusselt number; F is the area; τ is time. Subscripts: ∞ stands for incoming flow; w stands for wall; x stands for local values.

REFERENCES

1. B. M. Smol'skii, Z. P. Shul'man, and B. I. Fedorov, collection: Heat and Mass Transfer Between Bodies and Surroundings [in Russian], Minsk, 1965.
2. A. V. Luikov, IFZh, no. 11, 1962.
3. P. N. Romanenko, V. N. Kharchenko, and Yu. P. Semenov, IFZh [Journal of Engineering Physics], no. 6, 1965.
4. V. P. Isachenko and V. V. Vzorov, Teploenergetika, 9, no. 1, 1961.
5. B. I. Fedorov, IFZh, no. 1, 1964.
6. Yoshida Tetsuo and Tsutomu Hyodo, Chem. Ind., no. 7, Japan, 1958.
7. A. B. Verzhinskaya and R. Sh. Vainberg, IFZh [Journal of Engineering Physics], 12, no. 1, 1967.
8. A. B. Ambrazyavichus and A. A. Zhukauskas, Trudy AN LitSSR, seriya B3 (19), 1959.
9. E. Eckert and R. Drake, Heat and Mass Transfer [Russian translation], Gosenergoizdat, 1961.
10. B. I. Fedorov, Author's Abstract of Candidate's Dissertation, Minsk, 1964.

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