GENERALIZATION OF EXPERIMENTAL DATA ON HEAT-TRANSFER CRISIS IN PIPES

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The article presents generalized experimental data on heat-transfer crisis upon boiling in pipes with various types of intensifiers and without them, obtained by superposition of convective heat removal and mass transfer of liquid to the heating surface. For finned and corrugated pipes relationships are recommended that are suitable for use in calculations.

At present, much attention is given to investigations of two-phase flows; in consequence, the concepts concerning the processes occurring in such systems have been fairly exhaustively formulated [1, 2]. However, the interrelation between the processes determining the observed regularities of heat transfer to boiling two-phase flows is so complicated that it is often considered most expedient under the given conditions to use strictly empirically obtained data in the form of "skeleton" tables, as is done for heat-transfer crisis in pipes [3, 4]. Such a notion is probably most objective, but it is not exhaustive and convenient. At the present development level of the notions on heat transfer and crisis in boiling, empirical correlations are established for practical needs between the parameters determining the course of the process [5, 6, 8]. Physically substantiated models submitted by various authors yielded only partial successes in limited regions where the mechanisms of exchange processes are most obvious (e.g., in disperse annular or disperse flow regimes) [1, 4, 9, etc.]. This does not mean that attempts to find general models are bound to fail. It seems only that the physical foundations determining heat transfer and crisis in boiling processes have to be sought in a somewhat different direction.

It is suggested to use a common approach that is correct for any flow regime and structure of two-phase flow, based on the fact that heat removal in boiling is due to two physically mutually independent processes: convective heat removal and mass transfer of the liquid on the heating surface where this liquid evaporates. When a supercooled liquid boils, the heat accumulated by the vapor is transferred to the liquid in its subsequent condensation in the flux.

The principle of superposition that is used was previously used by several authors for representing the regularities of heat transfer in boiling [2, 9] as well as for representing heat-transfer crisis [7, 9, 18, etc.]. However, these attempts were confined to partial problems. Thus, the difference of the suggested approach is that the determining processes (convective heat removal and mass transfer in flux) are physically substantiated and that the approach is extended to the entire range of parameters, regardless of the regime and structure of the two-phase flow. Different regimes have their own characteristic dependences.

The following principal assumptions are adopted: 1) the processes of convective heat removal and mass transfer of liquid on the heating surface proceed independently in the flux; 2) the regularities of heat transfer in turbulent flow of a single-phase heat carrier are applicable to the determination of the heating of the liquid in the process of boiling.

It is well known that the processes of heat and mass transfer are analogs and are described by the Newton—Richmann equation

$$q = \alpha \Delta t, \tag{1}$$

$$m = k \Delta c. \tag{2}$$

The heat and mass transfer coefficients contained in this equation are determined basically by the hydrodynamics of the flux and by the properties of the liquid [11, 12].

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In accordance with such an approach, we may write, proceeding from the heat balance on the heating surface in the section under examination, that the total magnitude of the heat flux removed in boiling is equal to

$$q = \alpha (t_{\text{hs}} - \bar{t}_l) + k\Delta c (t_{\text{hs}} - t_s) c_p' + k\Delta c \Delta i_{\text{sub}} + k\Delta c r. \tag{3}$$

For further examination it is convenient to divide the first term in Eq. (3) into two:

$$q = \alpha (t_{\text{hs}} - t_s) + \alpha \Delta t_{\text{sub}} + k\Delta c (t_{\text{hs}} - t_s) c_p' + k\Delta c \Delta i_{\text{sub}} + k\Delta c r. \tag{4}$$

The buoyant forces of temperature and concentration of the liquid in the examined section of the flux, contained in these relations, are taken analogously:

$$\Delta t = t_{\text{hs}} - \bar{t}_l , \tag{5}$$

$$\Delta c = \bar{c} - c_{\text{hs}}. \tag{6}$$

Since the concentrations of liquid cannot practically be measured, it is impossible to determine the mass-transfer coefficients from the known heat transfer in boiling. But such a determination can be made by proceeding from the experiments on heat transfer crisis. A sign of the onset of heat-transfer crisis is the drying of the heating surface, i.e., attainment of the condition $c_{\text{hs}} = 0$. The causes disturbing the inflow of liquid into the crisis zone are not being examined here. Hence it follows that in heat-transfer crisis

$$\Delta c_{\text{cr}} = \bar{c}. \tag{7}$$

Consequently, we obtain from (4) for heat-transfer crisis that

$$q_{\text{cr}} = \alpha (t_{\text{hs}}^{\text{cr}} - t_s) + \alpha \Delta t_{\text{sub}} + k\bar{c} (t_{\text{hs}}^{\text{cr}} - t_s) c_p' + k\bar{c} \Delta i_{\text{sub}} + k\bar{c} r. \tag{8}$$

In accordance with the adopted assumptions, the heat-transfer coefficient is determined here by Mikheev's formula [11]:

$$\text{Nu} = 0.021\, \text{Re}^{0.8} \text{Pr}^{0.43} (\text{Pr}_l / \text{Pr}_{\text{hs}})^{0.25}. \tag{9}$$

Such an assumption does not contradict the nature of things, because in boiling the vapor bubbles turbulize the liquid intensely. On the other hand, by this equation we determine the heat-transfer coefficient to the liquid, even if it exists under somewhat specific conditions of saturation but receives heat as a result of the same physical processes as a single-phase liquid. An error therefore need not introduce fundamental changes into the obtained results, because the bulk of the heat evaporating the liquid, which as a result of mass transfer is conveyed to the heating surface, is taken separately into account. The suggested approach is correct in principle for determining the heat flux in the boiling of any liquid under any conditions. As the first verification of the possibility of using it, the most thoroughly investigated field was chosen: the determination of the generalizing dependences for the process of mass transfer in heat-transfer crisis in uniformly heated round pipes.

The authors of [19, 20] used a similar approach to the notion of heat-transfer crisis as a process that is due to the mass transfer of liquid to the heating surface of a pipe. The difference of the suggested approach is: Firstly, the approach to heat removal in boiling, as determined by mass transfer, is extended to all flow regimes, not only to vapor drop flows; secondly, the heat for heating the liquid to the temperature of the heating surface and for its evaporation is taken into account; and thirdly, the convective heat removal by the moving flux is taken into account.

The thermophysical properties contained in formulas (8), (9) were determined within the framework of the model of homogeneous flow using the theoretical magnitude of the true volumetric quantity of vapor. To simplify the calculation procedure, a method suggested by Miropol'skii, with Molochnikov's correction [10], was used. From the magnitude of the true volumetric quantity of vapor thus found, the mean concentration of the liquid phase over the section was also determined:

$$\bar{c} = (1 - \varphi) \rho'. \tag{10}$$

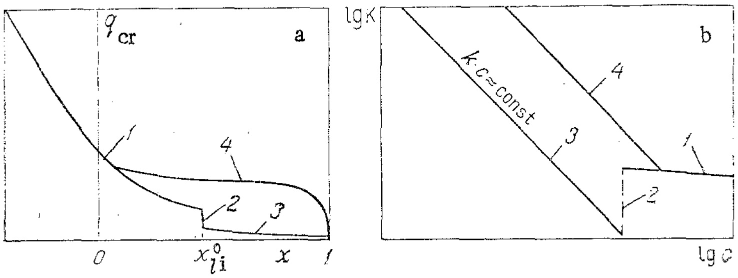

Fig. 1. Typical dependences of heat flux (a) and of the mass-transfer coefficient (b) in crisis of the first kind (1), second kind (2), deposition-controlled crisis (3), and in crisis in pipes with intensifiers (4).

Unfortunately, Miropol'skii recommended his method for use in the ranges of pressures 0.5-9.8 MPa, of mass velocities 100-3600 $kg/m^2 \cdot sec$, and of heat fluxes 0.2-2 MW/m^2, which obviously do not coincide with the range of change of heat fluxes for pipes upon heat-transfer crisis. Thus, according to the tables of coordinated values [3], critical heat fluxes change with pressures below 9.8 MPa from 1.6 to 12.6 MW/m^2. This level of heat fluxes is characteristic of quantities of vapor below the limit value x^0_{li}. With quantities of vapor above x^0_{li}, i.e., in deposition-controlled crisis [1, 15], the level of the heat fluxes makes it possible to use Miropol'skii's method, all the more so since this range of quantity of vapor is calculated by practically any method by dependences for adiabatic flow. In this case it is therefore probably justified to extrapolate the method of calculation to higher pressures than 9.8 MPa.

In continuing the analysis of the processes of heat and mass transfer, it should be expected that the regularities of change of the mass-transfer coefficient in dependence on the hydrodynamics of the flow and the thermophysical properties of the liquid will be generalized by the Sherwood numbers (Sh) — an analog of the Nusselt number, the Reynolds number, and the Schmidt number (Sc) — an analog of the Prandtl number. The coefficient of molecular diffusion contained in the Sherwood and Schmidt numbers was calculated by the Stokes—Einstein equation

$$\frac{D\mu}{kT_s} = \frac{1}{6\pi r}, \tag{11}$$

where k is the Boltzmann constant and r is the radius of the molecule of liquid. The calculation is carried out for the saturation temperature, because near the heating surface this temperature of the liquid, in particular, is characteristic.

To find generalized equations, we processed tabulated data for calculating heat-transfer crisis in the boiling of water in uniformly heated pipes [3], and experimental data of the author obtained in pipes with internal helical finning [13], internal transverse corrugation [14], and in the measurement of mass-transfer coefficients in smooth pipes [15]. These data relate to different parts of the dependence of the critical heat flux (Fig. 1a). According to the terminologies used for this purpose in [1, 4, 15], data on the "crisis of the first kind" and on "deposition-controlled crisis" were used. The boundary between these two zones is the limit quantity of vapor x^0_{li}." In Fig. 1b the change of the mass-transfer coefficients for the mentioned regions is shown schematically.

As was mentioned before, for the range of quantities of vapor below x^0_{li} a method of calculating the true quantity of vapor was used which was, strictly speaking, extrapolated to a range of heat fluxes not recommended by the author. This extrapolation was carried out on the basis of the following considerations: firstly, the method is simple, and the necessary calculations are correspondingly abridged; secondly, the nature of the dependence of the true volumetric quantity of vapor on the balance quantity of vapor for any fluxes is the same, only the level is different. It was therefore expected that the quantitative nature of the dependences as a result of such an extrapolation would be fairly fully revealed. However, errors in determining the mass-transfer coefficient naturally increase in an uncontrolled manner. When Miropol'skii's method is used for determining the true quantity of vapor in heat-transfer crisis in the range of crisis of the first kind, only the qualitative nature of the dependence of the mass-transfer coefficient on the regime parameters is revealed. As regards deposition-controlled crisis, there we have a quantitative evaluation of the accuracy of the suggested method of calculation.

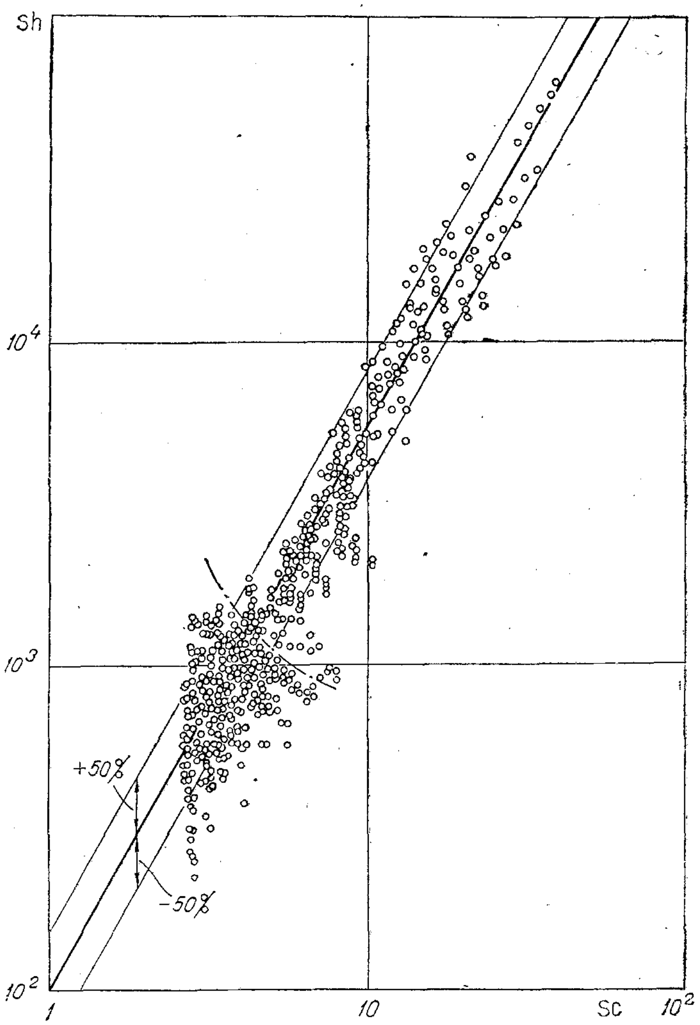

Fig. 2. Results of comparing the calculation by the tabulated data of [3] with Eq. (12); the dot-dash line shows the limit of applicability of the method of determining the true quantity of vapor from the pressure that was used.

The results of the calculation showed that for quantities of vapor below x°_{li} the course of the dependence $q_{cr}(P, \rho W, x)$ is determined principally by the change of true quantity of vapor, whereas the mass-transfer coefficients increase only slightly with increasing balance quantity of vapor, decreasing asymptotically from the level of 0.010-0.015 m/sec at 2.95 MPa to ∿0.005 m/sec at 17.6 MPa. A dependence of the mass-transfer coefficients on the mass velocity was not found.

Analysis of the parametric dependences made it possible to generalize the results of calculating the mass-transfer coefficients from the tabulated data of [3] by the criterial equation

$$\mathrm{Sh} = 100\,\mathrm{Sc}^{1.75}. \tag{12}$$

A comparison of the results of the calculation with Eq. (12) is presented in Fig. 2.

It is significant that the generalizing formula does not contain a Reynolds number; this indicates that the process of mass transfer in boiling is decisively affected by phase transformations on the heating surface, and not by the flow velocity in the channel.

With quantities of vapor equal to x°_{li}, the mass-transfer coefficient decreases abruptly by approximately one order of magnitude, attaining 0.001 m/sec or less. In the range of quantities of vapor below x°_{li}, the nature of the dependence of the mass-transfer coefficient on the quantity of vapor (and consequently, on the concentration of the liquid phase) is differ-

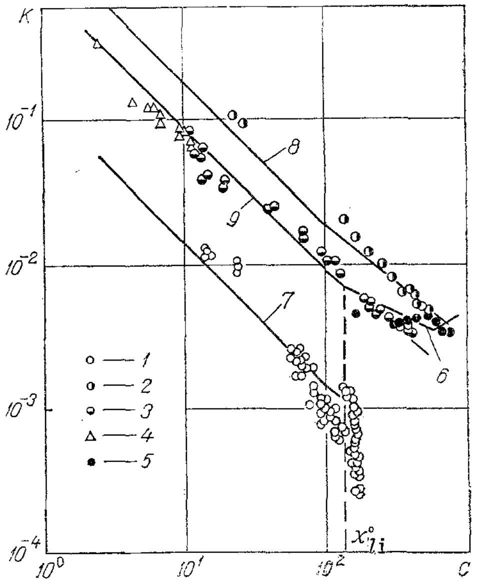

Fig. 3. Comparison of the experimental data for deposition-controlled crisis (1) with the dependences obtained for pipes with corrugation 1 mm deep and pitch 100 mm (2), pipes with fins 1.2 mm high, pitch 200 mm (3), and pitch 250 mm (4); 5) calculation with tabulated data of [3]; 6) by Eq. (12); 7) (13); 8) (14), (15); 9) (14), (16). K, m/sec; $\bar{c}$, kg/m^3.

ent. Whereas for $x < x^{\circ}_{li}$ $k \approx \text{const}$, for $x > x^{\circ}_{li}$ $k\bar{c} \approx \text{const}$ (Figs. 1b and 3), while the mass-transfer coefficient remains practically unchanged when the pressure increases from 4 to 20 MPa. The mass velocity in this range affects the heat flux via the true quantity of vapor and the mass-transfer coefficient. Therefore, the generalizing criterial formula expresses such a correlation:

$$\mathrm{Sh}_0 = \frac{0.5(\mathrm{Re}'')^{1.73}}{1 - \mathrm{Re}/\mathrm{Re}''}\,\mathrm{Sc}^{0.5}\cdot 10^{-8}. \tag{13}$$

This formula generalizes the experimental data used in [15]. They were obtained with a pipe with 11 mm inner diameter, with pressures of 4 to 20 MPa and mass velocities of 500-2800 kg/m^2·sec.

Unlike Eq. (12) for crisis of the first kind, the regularities of deposition-controlled crisis are generalized with the use of the Reynolds number. This is logical because the hydrodynamics of a flow with a quantity of vapor exceeding the limit is entirely different. In this range of quantities of vapor, part of the liquid is dispersed in the vapor phase, and consequently its transfer to the heated surfaces of the channel is determined primarily by the level of turbulization of the flow, which is measured by the Reynolds number.

On the boundary of the zones under examination, i.e., with the limit quantity of vapor, the flow is rearranged, primarily by redistribution of the flow rates in the core of the flow and in the film next to the wall. As a result, the mass-transfer characteristics and their level change radically.

In channels with heat-removal intensifiers it is to be expected that the nature of the dependences is analogous to the dependence in the case of deposition-controlled crisis because, firstly, intensifiers, no matter of what type, lead to heat-transfer crisis at high quantities of vapor, and secondly, the intensifiers themselves lead to redistribution of liquid over the channel section that characterizes flows with quantities of vapor exceeding the limit value. A comparison of the experimental data with the results of calculation of the mass-transfer coefficients for pipes with internal transverse corrugation [14] and with the results obtained later for different corrugation parameters with the participation of the present author, and also data for pipes with internal helical finning [13], showed that the nature of the dependences is perfectly identical (Fig. 3). Therefore, the form of the generalizing dependence was chosen as follows:

$$Sh = Y\,Sh_0, \tag{14}$$

where Sh_0 is determined by formula (13).

In processing the available experimental data we made the previously agreed assumption: the true volumetric quantity of vapor was determined according to the dependence of the adiabatic flow, and the intensifiers apparently bring the state of the flow closer to the equilibrium state. Convective heat removal was calculated by using the recommendations for pipes with internal helical finning [16] and for corrugated pipes [17].

The correction factor has to take into account the geometric characteristics of the intensifiers and the nature of the effect of the mass velocity on the increase in critical thickness. Thus, for corrugated pipes it has the form

$$Y = 1 + 2.3\left[\left(\frac{h}{S}\cdot 10^3\right)^{0.35} + 8\cdot 10^{-3}\left(\frac{h}{S}\,\frac{d_e}{d}\right)^{1.0}\left(1 + \frac{10^{13}}{(\mathrm{Re}'')^{2.32}}\right)\right]. \tag{15}$$

Its range of application is: pressure 9.8-17.6 MPa, diameter 10-12 mm, Reynolds number for saturated vapor $(0.33-4)\cdot 10^6$. These ranges were verified by the experimental data. The obtained rms deviation is 18.3%.

The correction factor for finned pipes has an analogous form:

$$Y = 1 + 2.3\left[\left(\frac{h}{S}\cdot 10^3\right)^{0.1} + 6\cdot 10^{-3}\left(\frac{h}{S}\,\frac{d_e}{d}\right)^{3.4}\left(1 + \frac{10^{13}}{(\mathrm{Re}'')^{2.32}}\right)\right]. \tag{16}$$

The parameters of the dependence were selected on the basis of an array of experimental data obtained with the participation of the present author and which find partial expression in [13]. An rms deviation of 16.5% was obtained. The range of application is the same as for the expressions of [15].

Thus, the work shows that the use of the principle of superposition for representing a heat flux in mass-transfer crisis, as in boiling generally, opens up good prospects. Such an approach makes it possible to use fairly simple criterial equations for convective heat removal and mass transfer of liquid to the heating surface.

NOTATION

c, concentration of the liquid phase, kg/m³; $\Delta c = \bar{c} - c_{hs}$, concentrational buoyant force, kg/m³; c_p, heat capacity at constant pressure, kJ/kg·°C; d, characteristic channel diameter, m; D, coefficient of molecular diffusion, m²/sec, calculated by (11); q, heat flux, mW/m², W/m²; h, height of fins or depth of corrugation, mm; i, enthalpy, kJ/kg; $\Delta i = i - i'$; K, mass transfer coefficient, m/sec; m, mass flow of liquid transferred on the heating surface, kg/m²·sec; r, latent heat of vaporization, kJ/kg; S, pitch of the fin helix or of the corrugation, mm; t, temperature, °C; T, temperature, °K; $\Delta t = t_{hs} - \bar{t}_l$, temperature gradient, °C; α, heat-transfer coefficient, W/(m²·sec); μ, dynamic viscosity, Pa·sec; λ, thermal conductivity, W/m·°C; ρ, density, kg/m³; ρW, mass velocity of flow, kg/m²·sec; φ, true volumetric quantity of vapor; $Nu = \alpha d/\lambda$, Nusselt number; $Re = \rho W d/\mu$, Reynolds number; $Pr = \mu/\rho\lambda$, Prandtl number; $Sc = \mu/\rho D$, Schmidt number; $Sh = Kd/D$, Sherwood number. Subscripts: hs, *l*, sub, parameters of the heating surface, of the liquid, of subcooling to the saturation temperature, respectively; per, superheating above the saturation temperature; s, saturation parameter; ′, ″, parameters of water and vapor, respectively, on the saturation line; —, sign of averaged parameter (over the channel section).

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