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HEAT EXCHANGE AND HYDRODYNAMICS DURING BOILING ON A HORIZONTAL TUBE BUNDLE IN A FLUIDIZED BED OF SOLID PARTICLES

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The results are given on the experimental study of heat transfer and hydrodynamics with boiling on a horizontal tube bundle located in a dispersed bed of solid particles and of this process hydrodynamics simulation by gas bubbling within the fluid filtration rate variation via the bed $(0\text{-}3.5)v_{l\,i.f}$. An analysis and correlation of experimental results are presented.

Arranging for boiling to occur under conditions of thermal fluidization [1] makes it possible to prevent the deposition of scale on the heating surfaces of tube bundles as a result of the collision of solid particles with these surfaces [2-4]. The cause of particle motion in this case is filtration through a dispersed layer of vapor generated on the tube surfaces. Since the amount of filtering vapor increases over the height of a bundle, scale removal will be adversely affected on the lower tubes compared to the overlying tubes [4].

One method of equalizing tube operating conditions, together with that proposed in [5], is arranging for boiling of a saturated liquid on the tube bundle under conditions of simultaneous thermal and hydrodynamic fluidization of a disperse bed of particles. The process of boiling on a horizontal tube located in a water-fluidized particle bed was studied in [6, 7]. High degrees of subheating in tests in [6, 7] caused condensation of the vapor phase in the boundary layer of the heating surface. The vapor phase generated here had no effect on hydrodynamic processes in the core of fluidized bed, as is typical for the case of boiling of a saturated liquid [3, 4]. The relations obtained for the conditions in [6, 7] cannot be used to calculate heat exchange at $T_l = T_{sat}$.

Presented below is an analysis and generalization of empirical data on heat exchange and the hydrodynamics of water boiling under the conditions $T_l = T_{sat}$ on a horizontal bundle of tubes located in a disperse bed of spherical particles. The tests were conducted in the range p = 0.05-1 bar and $\bar{q} = 10^4\text{-}1.5 \cdot 10^5$ W/m². The hydrodynamics of the process were also modeled using gas bubbling. The bundles consisted of a staggered arrangement of stainless steel tubes with ∅ 18/2 . The bundles were heated by condensing vapor. A bundle with a heating surface of 0.192 m² consisted of four vertical and five horizontal rows with relative spacings $S_{hor} = 3.4$ and $S_v = 2.2$. The 0.35-m-high disperse bed was made of particles of glass ($\rho_p = 2.5 \cdot 10^3$ kg/m³) and agalite ($\rho_p = 2.4 \cdot 10^3$ kg/m³) with mean diameters of 1.5, 2.75, and $4.75 \cdot 10^{-3}$ m. The circulating liquid, under conditions of simultaneous thermal and hydrodynamic fluidization, was fed into the bottom half of the bed. The filtration rate in the tests was variable: from 0 to 0.12 m/sec for the liquid, and from 0 to 0.4 m/sec for the vapor at the border of the bed.

The hydrodynamic model was a vertical cross section of a vessel with a tube bundle measuring 1.2 × 0.4 × 0.05 m. The vessel contained a transparent front glass window. The tube bundle was made up of tubes with ∅ 18/0.5 . Holes 0.7 mm in diameter were drilled in the sides of the tubes, 25 holes to each cm². The gas phase in the model was air, while the working liquid was water. The ranges of v_l and v_g in the bubbling tests were the same as in the boiling tests.

Preliminary tests showed that, as in [2-4], the relations $\alpha_i = f(q_i)$ under the conditions studied are the same for all tube bundles if the temperature head, in the determination of α_i, is reckoned from the temperature of the liquid at the level of the i-th horizontal row $\Delta T_{wa\,i} = T_{wa\,i} - T_{l\,i}$, where $T_{l\,i} = T_{sat}(p) + \Delta T_{h.d\,i}$. It was shown in [3] that the value of $\Delta T_{h.d}$ can be

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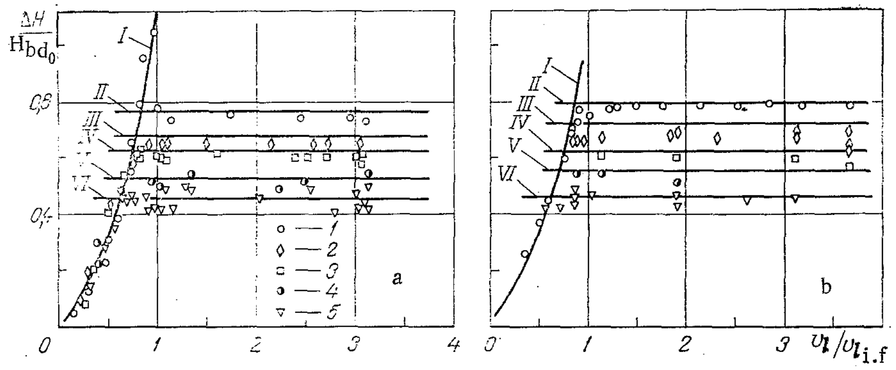

Fig. 1. Effect of velocities of liquid and gas (vapor) on the drag of a disperse particle bed during generation of the gas (vapor) phase on a tube bundle (I — calculation from (3); II — calculation from (2) with $\bar{\omega} = 0$; III-VI — calculation from (2), (6)): a) gas bubbling; $m_b = 0.9$; ΔH for bed depth $\bar{h}$ = 0.85; 1, 2, 3, 4, 5) v_g = 0; 0.06; 0.137; 0.216; 0.343 m/sec; b) boiling; $m_b = 0.9$; ΔH for $\bar{h}$ = 0.8: 1, 2, 3, 4, 5) v_g = 0; 0.052; 0.144; 0.198; 0.316 m/sec.

calculated for thermal fluidization conditions ($v_l = 0$) as

$$T_{h.d} = T_{sat}(p + \Delta p_{h.d}) - T_{sat}(p), \tag{1}$$

where

$$\Delta p_{h.d} = (1 - \bar{\omega})[m_{lp}\rho_l + (1 - m_{lp})\rho_p]gH, \tag{2}$$

and an analytic expression can be found for $\bar{\omega}(\bar{h})$ on the basis of the assumption that m_{lp} is independent of the velocity of the vapor ($m_{lp} \simeq m_{lp_0}$). This assumption is in turn a reflection of the data on m_{lp} in [8].

The possibility of calculating $\Delta T_{h.d}$ from (1) and (2) for the investigated conditions ($v_l > 0$) is not immediately apparent, but the only obstacle is the lack of information on $\bar{\omega}(\bar{h})$. Thus, calculation of $\Delta T_{h.d}$ from (1), (2) at $v_l > 0$ for boiling without a disperse bed of particles ($m_{lp} = 1$) yielded values considerably different from experimental values of $\Delta T_{h.d}$ for the conditions under discussion. Since the motion of the liquid in the apparatus in this case was closer to a flow with ideal displacement, it may be proposed that the above-noted difference is due mainly to incorrect (for the present conditions) use of Eq. (1), i.e., it can be suggested that T_l at any point of the system takes the equilibrium value T_{sat} corresponding to the pressure at the given point.

The use of the relations (2) and $\bar{\omega}(\bar{h})$, based on assumption of constant m_{lp}, for the case of forced motion of the liquid is also not obvious. It was noted in [9] that m_{lp} increases with v_l under conditions of three-component fluidization.

We checked the applicability of relations of type (2) for the conditions $v_l > 0$ on a hydrodynamic boiling model with a tube bundle. Such modeling can be backed up by the results obtained in [10] in a successful modeling of liquid boiling without a disperse particle bed by gas bubbling. However, the insufficient study yet devoted in the hydrodynamics of gas—liquid—particle systems makes it impossible to resolve the question of hydrodynamic similitude of systems with boiling and bubbling on the basis of the methods of similitude theory for the case under discussion. The possibility of ignoring — by analogy with [11, 12] — the effect of the physical properties of the gas phase on the hydrodynamics of processes in the core of the flow allowed us to reduce the requirements for similitude in the present case to the requirement of approximate identity of the investigated and modeled systems: geometric identity of the conditions under which the process is organized, identity of the working liquids and particles, and equality of the values of v_l and v_g in both cases.

The results of calibration tests on the model with $v_g = 0$ agreed well with well-known data in the literature [13]: the pressure drop in the stationary bed agreed with that calculated from Krgun's relation (curve I in Fig. 1a; the values of Δp in Fig. 1 are expressed in m H_2O — ΔH)

$$\frac{\Delta p}{H} = 150\,\frac{(1 - m_{lp})^2}{m_{lp}^3}\,\frac{\mu_l v_l}{d_p^2} + 1.75\,\frac{(1 - m_{lp})}{m_{lp}^3}\,\frac{\rho_l v_l^2}{d_p}, \tag{3}$$

initial fluidization velocity $v_{l_{i.f}}$ agrees with that calculated from Todes' relation (the pressure peak in Fig. 1a corresponds to the transition to the fluidized state at $v_l/v_{l_{i.f}} \approx 1$)

$$\mathrm{Re}_{i.f} = \frac{\mathrm{Ar_p}}{1400 + 5.22\sqrt{\mathrm{Ar_p}}}, \tag{4}$$

and the pressure drop in the fluidized bed agrees with that calculated from (2) at $\omega = 0$ (curve II in Fig. 1a).

Visual observation and photographing of the process at $v_g > 0$ showed that each value of v_g corresponds to a characteristic value $v_g \equiv v_{g_{i.f}}$. If this value is exceeded, the bed particles in the vicinity of gas jets generated by the tubes of the bundle are brought into motion, similar to the spouting that takes place in a two-component gas-particle system. In the present case, there is ascending motion of the disperse three-component system in the cores of the jets and descending motion of the dense, liquid-inundated bed of particles outside the jets. These conditions result in turbulent pulsation of the three-component system in the vicinity of the jets. The scale of these pulsations increases with an increase in v_g and v_l up to values commensurate with the spacing of the tubes in the bundle. The value of $v_{l_{i.f}}$ decreases with an increase in v_g, with the degree of effect of v_g on $v_{l_{i.f}}$ decreasing as well. The empirical dependence of $v_{l_{i.f}}$ on v_g obtained on the basis of visual observations of the moment of transition of the bed to the moving state has the form

$$\frac{v_{l\mathrm{i.m}}}{v_{l\mathrm{i.f}}} = \exp(-3.0 v_g). \tag{5}$$

At liquid velocities $v_l \geqslant v_{l_{i.f}}$, regardless of v_g, the disperse bed is fluidized, without distinct ascending and descending zones. The scale of the turbulent pulsations here is considerably smaller than in the case $v_l < v_{l_{i.f}}$. An increase in v_g in this region also leads to an increase in the scale of the pulsations, while an increase in v_l has the opposite effect — stabilizing the motion of the components of the disperse system.

According to Fig. 1a, an increase in v_g up to the moment of the transition to the moving state does not, under the conditions of our tests, cause an increase in the drag of the bed (points 2-5 on curve 1). The value of $\Delta p(\Delta H)$ is nearly independent on v_g. This evidently has to do with the fact that the gas phase in this case occupies only a small part of the horizontal section of the bed, filtering among individual drops, and the laws governing filtration of the liquid through the rest of the cross section are the same as for the case $v_g = 0$.

The transition of the bed to the spouting condition is accompanied by a decrease in the slope of the curve $\Delta p(\Delta H) = f(v_l/v_{l_{i.f}})$ with an increase in v_l to the value $v_{l_{i.f}}$. When $v_l \geq v_{l_{i.f}}$, the drag of the bed, as in the case of a two-component fluidized system, is independent of v_l. An increase in v_g leads to a decrease in $\Delta p(\Delta H)$ (points 2-5 in Fig. 1a). Thus, the two qualitatively different regimes of motion of the disperse bed seen here correspond to qualitatively different laws of the effect of v_g and v_l on Δp. As was shown earlier, the effect of v_g on heat exchange is qualitatively different in these regimes, and we may evidently also expect different laws of scale removal as well. The different scale-removal laws will be related to the velocity of the particles and the frequency of their collision with the heating surface. These differences in hydrodynamic and thermal laws make it necessary to differentiate the above regimes in the case being considered and to critically review several data sources in the literature on three-component fluidization. For example, the theoretical relation constructed in [9] for $v_{l_{i.f}}$ for conditions similar to those investigated here — feeding of the gas and liquid through the bottom half of the bed — used a value of v_l for invitation of fluidization which corresponded to the intersection of the curves $\Delta p = f(v_l,\ v_g = \mathrm{const})$ for stationary and fluidized beds. The dependence of $v_{l_{i.f}}$ on v_g presented in [9] is qualitatively and quantitatively close to relations which can be obtained from the data in Fig. 1a by a method similar to that used in [9]. However, the resulting values of $v_l = v'_{l_{i.f}}$ will be quite different from both $v_{l_{i.m}}$ and $v_{l_{i.f}}$. Thus, for $v_g = 0.35$ m/sec, $v'_{l_{i.f}} = 1.9 v_{l_{i.m}} = 0.65 v_{l_{i.f}}$.

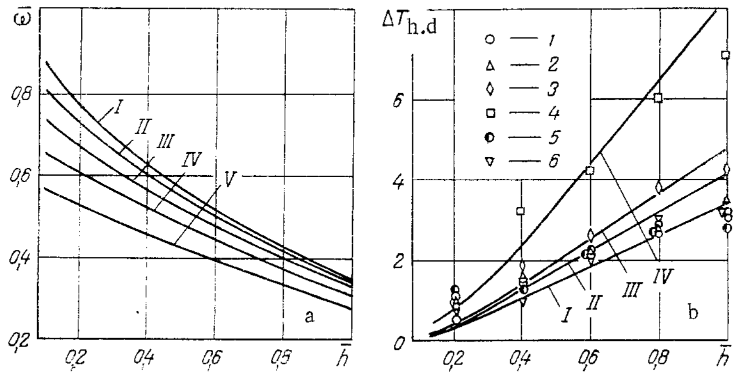

Fig. 2. Effect of relative depth of bed on mean volumetric vapor content and $\Delta T_{h.d}$: a) calculation with (6); I, II, III, IV, V) m_b = 1.0; 0.9; 0.8; 0.7; 0.6; b) $\Delta T_{h.d}$ at $\bar{q} = 5.1 \cdot 10^4$ W/m²; I, II, III, IV) calculation with (1), (2), (6); 1, 2, 3, 4) $v_l = 0$; 5, 6) 0.035-0.10 m/sec; 1, 5) p = 0.2 bar; 2, 6) 0.15 bar; 3) 0.125 bar; 4) p = 0.05 bar.

Figure 1a compares values of $\Delta p(\Delta H)$ calculated from (2) and values of $\bar\omega(\bar h)$ at $m_{lp} \simeq m_{lp_0}$ with test data on Δp for $v_l \geqslant v_{l_{i.f}}$ (curves II-VI). The value of $\bar\omega(\bar h)$ for the calculation with Eq. (2) was determined from the analytical relation

$$\bar\omega(\bar h) = \frac{1}{\bar h}\left\{\mathrm{Fr}\frac{\sqrt{\frac{1}{m_b^2}+\frac{4\bar h}{\mathrm{Fr}}m_b^2}-\frac{1}{m_b}}{2m_b^2} - \mathrm{Fr}^2\left(\frac{H}{L}+1\right)\frac{\left(\frac{2\bar h}{\mathrm{Fr}}m_b^2-\frac{1}{m_b^2}\right)\sqrt{\frac{1}{m_b^2}+\frac{4\bar h}{\mathrm{Fr}}m_b^2}+\frac{1}{m_b^3}}{12m_b^4} + \mathrm{Fr}^3\frac{H}{L}\times\right.$$
$$\left.\times\frac{\left(\frac{6}{\mathrm{Fr}^2}m_b^4\bar h^2-\frac{2}{\mathrm{Fr}}\bar h+\frac{1}{m_b^4}\right)\sqrt{\frac{1}{m_b^2}+\frac{4\bar h}{\mathrm{Fr}}m_b^2}-\frac{1}{m_b^5}}{60m_b^6}\right\}, \tag{6}$$

which is similar to the relation in [3] and which, is contrast to the one in [3], allows for the effect on $\bar\omega$ of the effective porosity of the tube bundle in the disperse bed (Fig. 2a). The value of H/L here was taken as 0, since the identical gas flow rate was maintained on all of the tubes of the bundle in the bubble tests. The good agreement between the experimental and theoretical values in Fig. 1a shows the applicability of Eqs. (2) and (6) and the validity of the assumption $m_{lp} \simeq m_{lp_0}$ under the conditions investigated. The possibility of ignoring the effect of the change in $m_{lp} = f(v_l, v_g)$ on Δp evidently owes to the phenomenon of contraction of the bed [8], which compensates for the effect of m_{lp} increasing with v_g.

The data shown in Fig. 1b on the drag of the bed in boiling on a tube bundle is in good qualitative agreement with the test data obtained on the hydrodynamic model. Comparison of these sets of data with the data calculated with Eqs. (2)-(6) (curves I-VI in Fig. 1b) shows good quantitative agreement and confirms the correctness of modeling the hydrodynamics of the boiling process in the present case by gas bubbling. According to Fig. 1, under the conditions investigated, the drag of the bed $\Delta p_s(v_l < v_{l_{i.m}})$ can be calculated from Eq. (3); the drag of the fluidized bed $\Delta p_{fb}(v_l \geqslant v_{l_{i.f}})$ can be calculated from Eqs. (2) and (6); drag in the intermediate regime $(v_{l_{i.m}} \leqslant v_l < v_{l_{i.f}})$ can be calculated from (3) up to $\Delta p < \Delta p_{fb}$ and from (2) and (6) at $\Delta p > \Delta p_{fb}$; the value of $v_{l_{i.f}}$ can be calculated from (4).

The increase in pressure in the disperse bed $p + \Delta p(H)$ relative to the pressure above the evaporation surface causes an increase in the temperature of the liquid in the bed during boiling compared to $T_{sat}(p)$ by the amount $\Delta T_{h.d}$. The characteristic dependence of $\Delta T_{h.d}$ on bed depth and pressure is shown in Fig. 2b. Values of $\Delta T_{h.d}$ obtained in similar [3] tests with thermal fluidization of a disperse bed (points 1-4, Fig. 2b, $v_l = 0$) agree with the values calculated from (1) and (2), as in [3]. It follows from Fig. 2b that the character of the effect on p and $\bar h$ on $\Delta T_{h.d}$ with simultaneous thermal and hydrodynamic fluidization (points 5-6) is the same as under the conditions $v_l = 0$, and the values of $\Delta T_{h.d}$ in this case are independent of v_g. Thus, for both these conditions and the conditions in [3], it is possible to use Eq. (1) in calculating $\Delta T_{h.d}$, with calculation of $\Delta p_{h.d}$ from (2) and (6). The fact that such is possible of $v_l > 0$ is evidently due to the fact that the presence of a disperse bed of particles in the boiling liquid makes the properties of this system more closely resemble the properties of a system with ideal mixing (according to [14], the coefficient of longitudinal mixing in a water-fluidized bed with values of d_p, ρ_p, and v_l similar to those investigated here reached 0.1 m^2/sec). The good agreement between the calculation (curves I-IV, Fig. 2b) and experiment confirms the correctness of using (1) and (2) for the investigated conditions, as opposed to the conditions of boiling at $v_l > 0$ without a disperse bed of particles.

The results of calibration tests of heat exchange on a bundle of steel Kh18N10 tubes without a disperse bed of particles agree with data in [4] for a single heating tube and agree qualitatively with the data in [4] for a bundle of copper tubes, proving that the number of horizontal rows of tubes i has no effect on the function $\alpha_i = f(q_i)$ for the investigated values $S > 2$.

Data from tests on heat exchange on a bundle of stainless steel tubes under conditions of thermal fluidization ($v_l = 0$) of a bed of particles with $d_p = 1.5$, 2.75, and $4.75 \cdot 10^{-3}$ m confirmed the laws obtained earlier in [3, 4] on a copper tube bundle with particles $d_p = 2.75 \cdot 10^{-3}$ for describing the effect of q_i and p on heat exchange — the independence of α_i on q_i and p in the investigated range of p and at $q_i \geqslant q^*$, where q^* is the value of q at which thermal fluidization begins. In accordance with Fig. 3, during thermal fluidization on a tube bundle when $q_i \geqslant q^*$, as under conditions of boiling when $q \geqslant q^*$ on a single tube [15], α_i is independent of d_p

Comparison of the results obtained in [3] for a copper tube bundle with the data in Fig. 3 for steel Kh18N10 tubes shows qualitatively the same character of the effect of the tube material on heat exchange as is seen in the case of boiling on a single tube [15] — α_i increases with an increase in the thermal conductivity of the material of the tube wall. Test data on heat exchange under the investigated conditions (steel Kh18N10) is generalized, with a spread of ±25% by the relation

$$q_i = 8.9 \cdot 10^3 (T_{wa_i} - T_{sat} - \Delta T_{h.d_i}), \tag{7}$$

which is valid for $p = 0.05$–1.0 bar; $q_i = 1.9 \cdot 10^4$–$1.5 \cdot 10^5$ W/m^2; $i = 1$–5; $d_p = 1.5$–$4.75 \cdot 10^{-3}$ m. The value of $\Delta T_{h.d_i}$ in (7) is calculated from (1), (2), and (6).

When $q_i < q^*$, the heat-exchange data, as the results in [3] for these conditions, are generalized by the relation in [15] for boiling on a single tube at $q < q^*$. This relation has the form

$$\alpha = f(\tilde{p}, \tilde{d}_p) \left(\frac{\lambda_l^2}{v_l \sigma T_{sat}} \right)^{1/3} q^{2/3}. \tag{8}$$

The dependence of α on d_p in (8), as in [14], is nonmonotonic in character, with a maximum in the region $\tilde{d}_p \approx 1$. The nature of the effect of d_p on α seen under conditions of thermal fluidization on a tube bundle, agreeing well with the data in [4, 15] for a single tube within a broad range of d_p, additionally confirms the analysis of heat-exchange laws under conditions of thermal fluidization presented in [3] on the basis of comparison of data in [3] for a single value of d_p with data and model representations in [4, 15].

Heat-exchange tests involving boiling with forced motion of a liquid through a disperse particle bed were conducted in the velocity range $0 < v_l < 3.5 v_{l_{i.f}}$. The lower boundary of vapor velocity in the tests $v_{y_{min}} = f(q_i, p)$ was limited by the value of $q_{i_{min}}$, at which boiling began on a tube bundle. All of the values of $v_{l_{i.m}}$ in our tests, corresponding, according to (5), to the values of $v_{y_{min}}$, were less than the values of v_l at which the tests were conducted. Thus, it may be suggested that a disperse bed in boiling tests is either in a spouting condition ($v_l < v_{l_{i.f}}$) or a state of fluidization ($v_l \geq v_{l_{i.f}}$) when $v_l > 0$. The heat-exchange tests showed that these hydrodynamic regimes are characterized by different effects of v_l and $v_g(q)$ on heat exchange. Given low $q_i < q^*(v_l)$, the relation $\alpha_i = f(q_i)$, as in (8), has the form $\alpha_i \sim q_i^{2/3}$ as v_l increases to $v_{l_{i.f}}$. Here, an increase in v_l is accom-

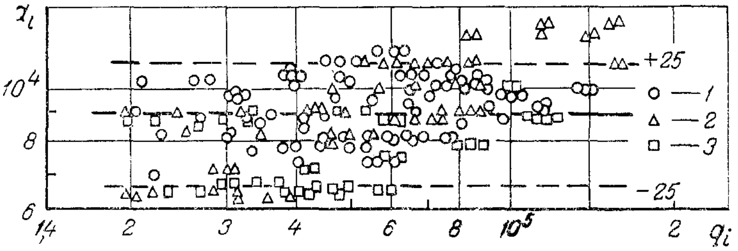

Fig. 3. Heat exchange in boiling on a horizontal tube bundle under conditions of thermal fluidization of a disperse bed of particles, i = 1-5: 1, 2, 3) d_p = 2.75; 4.75; $1.5 \cdot 10^{-3}$ m; glass, agalite. The lines represent calculations with (7); the dashed line corresponds to a spread of 25%. q_i, W/m²; α_i, W/m²·deg.

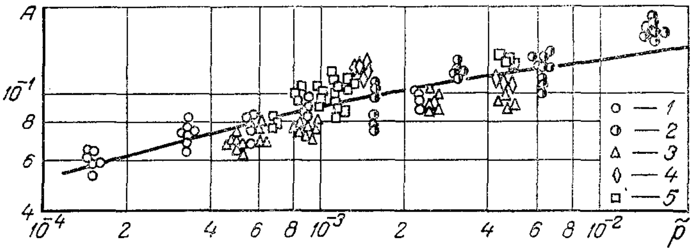

Fig. 4. Generalization of test data on heat exchange in boiling $A=\alpha\left(\frac{v_l\,\sigma T_{sat}}{\lambda^2}\right)^{1/3} q^{-2/3}$; T_{wa} = const: 1, 2, 3) boiling without a disperse bed of particles; 4, 5) boiling in a disperse particle bed at $v_l \geqslant v_{l_{i.f}}$; 1) single tube, water; 2) single tube, ethanol; 3) tube bundle, water; 4) tube bundle, water, v_l = 0.109 m/sec; 5) tube bundle, water, v_l = 0.065 m/sec. The solid line denotes calculation with (10); the material of the heating surface is steel Kh18N10.

panied by an exponential decrease in the quantity $\alpha_i/q_i^{2/3}$. For example, for p = 1 bar and $d_p = 2.75 \cdot 10^{-3}$ m, this relation has the form

$$\frac{\alpha_i/q_i^{2/3}}{(\alpha_i/q_i^{2/3})_0} = 1 + 1.5\exp\left(-2.25\frac{v_l}{v_{l_{i.f}}}\right). \tag{9}$$

When $q_i > q^*(v_l)$, the heat-exchange laws agree with the character of heat exchange under conditions of thermal fluidization (7). The value of q^*, increasing with v_l at $v_l > 0$, can be determined from simultaneous solution of equations of types (9) and (7). The character of the effect of $v_l < v_{l_{i.f}}$ on heat exchange is qualitatively similar to the character of the effect of $\tilde{d}_p$ on heat exchange under the conditions v_l = 0 and can be explained, within the framework of the model representations in [3, 15] as the result of the same factor: the heat-transfer rate at $q_i < q^*$ is determined by the effect of the thickness $\bar{\delta}$ of the nonsteady liquid boundary film formed by periodically merging vapor bubbles as they move along the heating surface. The value of $\bar{\delta}$ is determined by the corrected velocity of the vapor and the effective size of the pore space in the boundary region d. An increase in v_l increases the mobility of the disperse bed and also increases $\bar{\delta}(d)$, due to the greater displacement of the dense bed from the heating surface by the vapor bubbles. As a final result, α decreases. When $q_i > q^*$, a steady vapor–liquid layer is created at the surface. The laws of heat exchange of this layer are the same as the laws under conditions of thermal fluidization [3].

The heat-transfer rate during boiling after the transition of the disperse bed to the fluidized state ($v_l \geqslant v_{l_{i.f}}$) is similitudinous with respect to v_l and corresponds to the rate in boiling on a single tube and on a tube bundle without a disperse bed of particles. Since the values of v_l in the velocity range investigated ($v_l \leqslant 0.12$ m/sec) were an order less than the values of v_l at which v_l is seen to affect α in boiling without a disperse particle bed

[16], v_l can affect heat exchange only by changing the geometric structure of the bed (the value of d) or the number of active centers of vapor formation and the convective component of heat transfer at sites of particle impact with the heating surface. The characteristic size of the pore space d in our tests exceeded $\sqrt{\sigma/g\rho_l}$. In accordance with the model representations in [4, 15], d has no effect on heat exchange at such values. The independence of α from v_l allows us to conclude that the effect of direct collision of the particles with the heating surface is also insignificant from the point of view of heat exchange under the conditions investigated here.

We have the following relation for heat exchange during boiling. The relation was obtained on the basis of the structural relations [15, 16] and least-squares analysis of test data for boiling on the surface of a bundle of steel Kh18N10 tubes in a fluidized bed at $v_l \geqslant v_{l_{i.f}}$, data for a bundle without a disperse particle bed, and data from [4] for a single tube under free conditions in the ranges $\tilde{p} = 1.6 \cdot 10^{-4}-1.6 \cdot 10^{-2}$, $v_l/v_{l_{i.f}} = 1-3.5$, and i = 1-5

$$\alpha_i = \left(0{,}13 - \frac{1}{10^4\tilde{p}+12{,}3}\right)\left(\frac{\lambda_l^2}{\nu_l\, \sigma T_{\text{sat}}}\right)^{1/3} q_i^{2/3}. \quad (10)$$

To use (10) for boiling on a tube bundle, it is necessary to know the distribution of q_i and $\Delta T_{h.d_i}(\omega_i)$ over the horizontal rows. The inverse relationship between $T_{h.d_i}$ makes it impossible to calculate α_i directly from (10), (1), (2), and (6). Heat exchange can be calculated from (10), (1), (2), and (6) under these conditions by the method of successive approximations, similar to the procedure described in [3].

NOTATION

v, velocity; T, temperature; p, pressure; q, heat flux; S, spacing of tube bundle; g, acceleration due to gravity; H, bed height; ΔH, drag in m H_2O; d_p, particle diameter; d, linear dimension; m, porosity; L, characteristic depth of bed; $\bar{h} \equiv H-x/H$, where x is the height coordinate; $Re \equiv vd/\nu$; $Ar_p \equiv (\rho_p-\rho_l)gd_p/\rho_l\nu^2$; $Fr \equiv v_{g_0}^2/gH$, $v_{g_0} \equiv v_g(H)$ at $\bar{q} \equiv q(\Delta T_{h.d} = 0)$; $\tilde{p} \equiv p/p_{cr}$; $\tilde{d}_p \equiv d_p\sqrt{\sigma/g\rho_l}$; ρ, density; α, heat-transfer coefficient; ω, volumetric vapor content; μ, absolute viscosity; λ, thermal conductivity; ν, kinematic viscosity; σ, surface tension; δ, thickness. Indices: *l*, liquid; i.*f*, initiation of fluidization; sat, saturation; g, gas, vapor; hor, horizontal; v, vertical; p, solid particle; i, number of row in vertical direction; wa, heating wall; h.d, hydrodynamic depression; ($\bar{\ }$), mean; *l*p, liquid–particle system; 0, initial conditions, free conditions; b, tube bundle; i.m, initiation of motion; s, stationary; *f*b, fluidized bed; cr, critical.

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UNIDIMENSIONAL MODEL FOR PIPE FLOW OF A GAS MIXTURE ALLOWING FOR CONDENSATION

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Condensation kinetics is described with allowance for the dependence of the temperature of the phase transformation on the pressure and concentration of the condensing gas at the phase boundary.

We will examine the turbulent axisymmetric flow of a binary gas mixture in a pipe. Let the flow conditions be such that one of the components of the mixture condenses on the pipe surface, forming a film of liquid condensate of a thickness which increases with time.

We will describe the flow of the condensing gas with a two-layer model of turbulent flow — a laminar sublayer close to the surface of the condensate layer and a turbulent core in the remaining part of the flow. Intensive turbulent transport of the substance in the flow core makes the temperature, pressure, density, velocity, and concentration of the condensing gas practically constant across the pipe. Thus, we will use cross-sectional-mean values of these parameters to describe transport processes in the flow core, with the parameters changing only with time and station (from cross section to cross section). In the laminar sublayer between the flow core and the surface of the condensate film, the parameter values change from the values in the core to the values on the surface.

The liquid in the film will be assumed stationary in order to simplify the construction of a one-dimensional model of the process, here retaining all of the essential features of the flow.

The unidimensional equations for the mean (across the gas flow) quantities have the form

$$\frac{\partial \rho S_w}{\partial t} + \frac{\partial \rho u S_w}{\partial x} = \dot{M} S_w,$$

$$\dot{M} = -\frac{\Pi_w}{S_w}\left[\rho v_w - \rho u_w \frac{\partial R_w}{\partial x} - \rho \frac{\partial R_w}{\partial t}\right],$$

$$\frac{\partial \rho u S_w}{\partial t} + \frac{\partial \rho u^2 S_w}{\partial x} = -S_w \frac{\partial p}{\partial x} + \tau_w \Pi_w + u_w \dot{M} S_w - \rho S_w g \sin\alpha^*,$$

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