



HEAT TRANSFER AT EVAPORATIVE CONCENTRATION OF DOWN FLOWING IN VERTICAL PIPES SOLUTIONS IN ANNULAR REGIMES

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Abstract: In many branches of food processing, particularly in sugar production, the solution concentration takes place within industrial multi effect evaporator; here, the solution moves down as a liquid film forming annular descending flow in vertical heated pipes. It should be noticed that within the evaporation station the liquid which is being evaporated significantly changes its physical properties due to the fact that the concentration of the solution increases by the range of 14...65% of dry matter. The operational process parameters (pressure, temperature, etc.) also change to a noticeable degree. These factors determine a wide range of liquid films flow patterns, namely film thickness, surface wavy structures, velocity distribution across the films, interphase shear stress and, as a result of these, the making of a certain heat transfer mechanism including bubble boiling or its suppression. Unfortunately, the reliable methods of prediction and calculation of heat transfer coefficients are unavailable, which seriously hampers the designing and calculation of efficient evaporators and other heat transfer equipment. The work presents the results of modeling heat transfer to saturated liquid films of solutions flowing down inside the vertical pipes as in the regime of evaporation from the interphase surface and in that of surface boiling. The correlations given allow calculating the heat transfer coefficients for liquid films being concentrated in the vertical pipes of industrial evaporators

Keywords: heat flux, heat transfer coefficient (HTC), depression, interphase surface, nucleate boiling

1. Introduction

An extensive analysis of the mechanisms of heat transfer to the saturated turbulent and laminar liquid films of sugar solutions in the regime of evaporation from the developed wavy structures is given in [1,2]. The analyzed pattern of film movement is considered to be dominant in the long vertical pipes of industrial evaporators [3]. The data presented in [1,2] and respective correlations for calculation heat transfer coefficients (HTC) for laminar and turbulent films adequately

reflect the mechanisms of heat transfer in long vertical pipes of industrial evaporators within the regime of evaporation from film free surface. At the same time in [4-7] it has been shown that the regime of heat transfer which has been identified as that of evaporation from the interphase surface can be realized only to the certain limiting value of heated wall superheat related to the local saturation temperature, above which the mechanism of heat transfer undergoes a drastic change as being affected by pulses resulted by appearing and move of steam bubbles.

Nomenclature

δ – film thickness;
 Γ_v – volumetric liquid flux;
 a – temperature conductivity;
 ν – cinematic viscosity coefficient;
 t_{sat} – saturation temperature;
 t_w – wall temperature;
 t_i – interphase temperature
 λ – heat conduction of liquid;
 q – heat flow;
 ρ – liquid density;
 ρ_2 – steam density;
 σ – surface tension;
 C – mass concentration
 \bar{u} – average liquid film velocity
 u_2 – steam velocity;
 g – acceleration of gravity;
 $Pe = \frac{4\Gamma_v}{a} = \frac{4\bar{u}\delta}{a}$ – the Peclet number;
 $Re = \frac{4\Gamma_v}{\nu}$ – the Reynolds number.

In [6], the intensifying effect of the nucleate boiling upon the heat transfer at a free down flowing water films on the outside of a cylinder and a constant along the channel heat flux has been determined as a multiplier:

$$\left[1 + 0.05 \left(\frac{q_w - q_{incip}}{q_{incip}} \right)^{1.36} \right] \quad (1)$$

to the basic equation for HTC determination at the regime of evaporation from the film free surface. In Eq.(1) the subscripts denote: *w*-related to wall, *incip*-at bubble boiling inception.

The minimum value of the heat flux at which the bubble boiling starts effect the heat transfer mechanism has also been found. According to [6], the heat flux at which the bubble boiling starts in some inception points is recommended to

determine as $q_{incip} = 14.2 \frac{kW}{m^2}$. At the same time, it has been noted that the onset of boiling is influenced by liquid convection, therefore a much more adequate factor which reflects the beginning of boiling as well as its effect upon the intensity of heat transfer would be a complex $\frac{q}{r\bar{u}\rho}$ presented in [4] in which the onset of boiling is identified as:

$$\frac{q}{r\bar{u}\rho} \geq 1.5 \cdot 10^{-4} \left(\frac{\bar{u}\delta}{a} \right)^{-0.24} Pr^{-0.3}. \quad (2)$$

It should be marked that the complex in Eq. (2) $\frac{q}{r\bar{u}\rho}$ does not account for the

effect of pressure or vacuum upon the critical heat flux at which the surface nucleate boiling starts. On top of this, neither q_{incip} in Eq. (1), nor $\frac{q}{r\bar{u}\rho}$ account

for the local roughness of heat transfer surface, which presents a decisive factor in the determination of a critical value of the wall superheat leading to the onset of bubbles generation. It is necessary to stress specifically, that the majority of experimental works deal with the heat transfer processes in the liquid films heated to the saturation temperatures in experimental stands which model the conditions that occur in the industrial evaporators. In these conditions of simultaneous effect of mutually regulating factors, it is practically impossible to separate and distinguish effect of any individual factor.

Therefore the recommended expressions which are used for calculations of heat transfer intensity in the saturated liquid films have a limited applicability due to the narrow range of the regime parameters, in

which they are valid, as well as due to their applicability only to some geometrical configurations. Due to the above mentioned, they hardly may be applied to determine the conditions of the bubble boiling onset in the pipes of real industrial evaporators.

2. Materials and methods

A direct experimentation of heat transfer in down flowing liquid films of sugar solutions heated to the saturation temperatures has been carried out in the experimental unit with the independent formation of phases' mass flow rates and heat flux. The main core of the experimental unit is formed by a stainless still pipe with the inside diameter of 20 mm and 1.8 m long. The experimental tube was separated into the initial 1.5 m stabilization section and 0.3 m measurement section. The down flow of water (sugar solutions) film has been formed by means of overflowing over the tube's upper rim. In the event of steam-liquid flow modelling, dry saturated steam has been supplied in co-current regime. The liquid falling film has been heated by dry saturated steam which was supplied into outside heating sections attached to the experimental tube. The heating chambers were designed in a way as to provide an individual heating of the stabilization section and the experimental one. The said sections were hooked up to the individual vacuum-condensation sections which allowed for the keeping of different pressures in each chamber. Such arrangements allowed also maintaining vacuum down to 0.85 bars and thus, vary the temperature head between the heating steam temperature and evaporation temperature. Special probes for taking samples of liquid to determine its concentration and measurements of temperatures were positioned directly after the measurement section. A detailed

description of the experimental unit is given in [11].

The experimental unit closely reflects the actual operational conditions of sugar industrial film evaporators. The installation consist of heat transfer pipe from stainless steel, 9 m long with internal diameter of 30 mm partitioned in 20 section each of 40 mm long. Each section was equipped with the special chamber which allowed collecting condensate which was generated on the respected section. The experimental pipe and condensate chambers were positioned into the heating jacket. Then this condensate was transferred to the adiabatic measurement glasses which allowed measurements of the condensate collected from each test section and thus to determine the heat flux acting on the respective section. The heating of the experimental pipe has been done by means of supplying dry saturated steam into the pipe's heating jacket. Between each of 20 test section a special probe for taking samples of liquid and static pressure were installed. The wall temperature of each section and the temperature at the access of the pipe were taken by means of copper-constantan thermo-couple (type T). A detailed description of the experimental unit is given in [4].

3. Results and discussion

A generally accepted parameter for the identification of the onset of nucleate boiling in films has been suggested by Chun and Seban [5]. The value of the minimal superheat of a rough wall over which a liquid film flows down looks a reliable factor which may be used for identification of the onset of steam bubbles appearing. The value can be calculated based upon the combining the Laplace and Clapeyron-Clausius expressions, which correlate the value of a wall critical superheat Δt_{min} with a certain radius of

indentations and micro cracks R_c at a certain pressure (or vacuum):

$$\Delta t_{min} = \frac{2\sigma T_{sat}}{r\rho_2 R_c}, \quad (3)$$

where T_{sat} - absolute saturation temperature. In the event of solution boiling, the wall superheat will be bigger by the value of the physical-chemical depression Δ_{fc} , and thus Eq.(3) is to be rearranged as:

$$\Delta t_{min} = \frac{2\sigma T_{sat}}{r\rho_2 R_c} + \Delta_{fc}. \quad (4)$$

It is clear that the intensification of heat transfer at the surface boiling is caused by a gradual increase in the new centers of steam bubbles generation along with the growth of the temperature superheat, as it is evidently follows from (3,4). Equally this parameter adequately reflects the effect of pressure upon the inception of bubble appearing on the heated surface. Therefore the parameter which may be used to characterize the heat transfer intensification might be developed by use of the temperature head $\Delta t \geq \Delta t_{min}$ which later will be presented as a multiplier to the basic equation for calculation of HTC in liquid films at evaporating from the interphase surface as:

$$K_{boil} = 1 + c \left(\frac{\Delta t - \Delta t_{min}}{\Delta t_{min}} \right)^n, \quad (5)$$

where c and n - empirical coefficients derived from the experimental data. The basic equation for calculation the intensity of heat transfer at the regime of evaporation from the free interphase of liquid films of water and sugar solutions derived in [2] is extremely complex, especially in case of co-current steam flow, therefore its simplified version for
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engineering calculations is given as Eq. (6):

$$\frac{\alpha}{\lambda} \left(\frac{v^2}{g} \right)^{1/3} = 1.12 \text{Re}^{-1/3} (0.85 + 0.01 \text{Pe}^{0.2} + 4.5 \cdot 10^{-4} \text{Pe}^{0.86} \text{Pr}^{-0.2}) \times \sqrt{1 + \left[7.5 \cdot 10^{-6} \text{Re}_2 \left(\frac{\rho}{\rho_2} \right)^{0.2} \right]^2} K_{boil} K_L, \quad (6)$$

where K_{boil} - correction factor accounting for the nucleate boiling, K_L - correction factor accounting for the channel length and diameter.

A comparison of experimental and calculated according to Eq.(6) data on HTC to water and sugar solution free flowing films in 20 mm diameter pipes at the regime of evaporation from the free interphase surface is given in figure 1 and in case of co-current steam flow – on figures 2, 3 and 4.

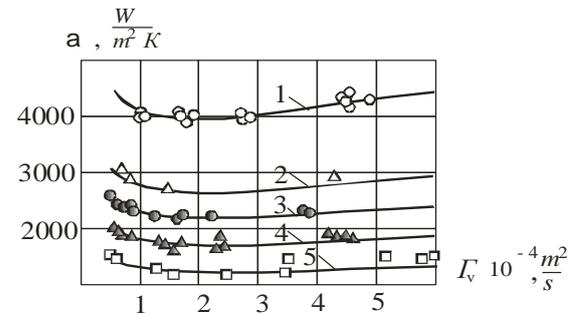


Fig.1. Comparison of data calculated as per (6) and experimental data [11]
1 – water, $t = 100^\circ\text{C}$; 2, 3, 4, 5 – sugar solution;
2 –DM = 30%; 3 – 40; 4 – 50; 5 – 60

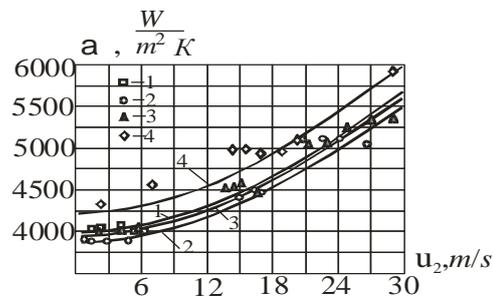
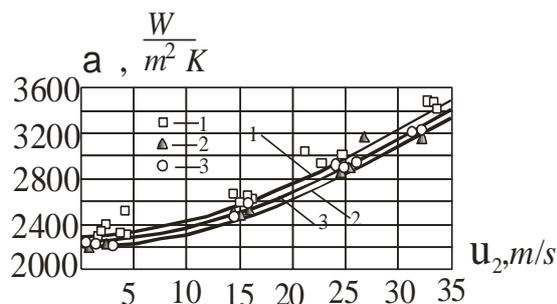
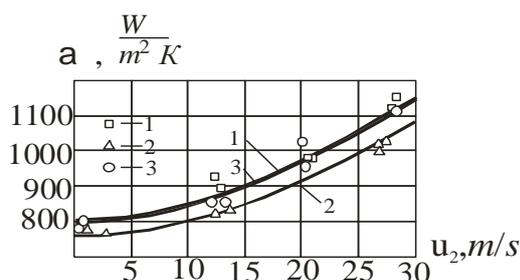


Fig.2. Correlation $\alpha = f(u_2)$. Comparison of data calculated by Eq.(6) with experimental data [11] water, $t = 100^\circ\text{C}$.
– $\Gamma_v = 0.1 \cdot 10^{-3}$ 2 – 0.2; 3 – 0.3; 4 – 0.5



**Fig.3. Correlation $\alpha = f(u_2)$. Comparison of data calculated by Eq.(6) with experimental data [11] Sugar solution, DM = 40%, $t = 100^\circ\text{C}$;
1 – $\Gamma_v = 0.1 \cdot 10^{-3}$ 2 – 0.2; 3 – 0.35**



**Fig.4. Correlation $\alpha = f(u_2)$. Comparison of data calculated by Eq.(6) with experimental data [11] Sugar solution DM = 70 %, $t = 100^\circ\text{C}$,
1 – $\Gamma_v = 0.1 \cdot 10^{-3}$ 2 – 0.3; 3 – 0.55**

Analysis of the obtained results shows that a close concurrence may be achieved by substituting into (3) the values of micro cracks dimensions of $R_c = 0.5 \cdot 10^{-5}$ m, which corresponds to the values of new heat transfer pipes' roughness. Respective values of $c = 0.4$ and $n = 1.2$ found from experimental data are to be used in Eq. (5). An additional factor affecting the accuracy and singularity of the experimental data on heat transfer to the solutions processing would be a unevenness of the concentration distribution across the film. As a result of the gradual evaporation of solvent (water in case of sugar solutions or fruit juices) from the liquid film, there appears a thin layer adjacent to the interphase in which the concentration of solution is somewhat bigger than that in

the bulk of the film. If only the molecular diffusion had caused the equalizing of the concentration field, the process would have taken an extremely long time. In fact, the surface of the film is covered by a complex wavy structure, which even at low Reynolds numbers exerts turbulence penetrating the bulk of the film, which in turn results in an uncertain form of the concentration distribution across the film.

The temperature of liquid film of solution being concentrated at a regime of evaporation from the free surface t_i should be equal to the saturation temperature of the solvent at a given pressure and physical-chemical depression at a concentration on the interphase surface C_i , thus $t_i = t_{sat} + \Delta_{fc}(C_i)$. Keeping in mind stated above concerning the uncertainties in the concentration field across the film, it would be grounded enough to suggest that the waves equalize the concentrations in the bulk of the film, so that the mean film temperature will be $t_i \cong t_{sat} + \Delta_{fc}(\bar{C})$. Using this value as a reference temperature and determining HTC as $\alpha = q / (t_w - t_i)$, one may obtain:

$$\alpha = \frac{q}{t_w - (t_{sat} + \Delta_{fc}(\bar{C}))}, \quad (7)$$

where \bar{C} - mean across the film solution concentration.

Since the film temperature profile is nonlinear (usually in the laminar stationary film it is parabolic), the mean film temperature $t_{av.m}$ in general may be derived as Eq.(8):

$$t_{av.m} = \int_0^{\delta} t(y) \frac{u(y)}{\bar{u}} dy \approx \frac{t_w + t_i}{2}, \quad (8)$$

where $t(y), u(y)$ - temperature and velocity profiles, respectively.

At the same time, the results of direct measurements have proven that the experimental mean mass film temperature of solution $t_{av.m}^{exp}$, which had been measured in the adiabatic calorimeter immediately at the exit of the experimental stand correlates with Eq.(8) only at conditions of free down flowing at the absence of interphase shear stress. The schematic of measurements is given in figure 5.

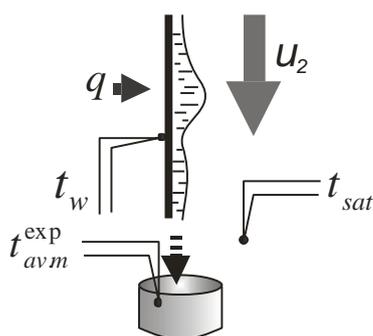


Fig. 5. Schematics of calorimetric temperature measurements in flowing down liquid films

In case of a co-current flow of steam above the film, the experimental mean mass solution temperature $t_{av.m}^{exp}$ turned out lower than that determined by Eq.(8) $t_{av.m} \geq t_{av.m}^{exp}$. The deviation observed was the bigger the bigger was the steam flow velocity and liquid mass flow rate.

Since the experimental temperature $t_{av.m}^{exp}$ in the contrast to t_i adequately reflects any change in the regime parameters and being the value which is directly measured in the experiments, then the HTC to solutions are to be correctly defined as Eq.(9):

$$\alpha_m = \frac{q}{t_w - t_{av.m}^{exp}}, \quad (9)$$

which will adequately reflect the effect of phases mass flow rates change. This immediately yields $\alpha \geq \alpha_m$. Since HTC determined as per Eq.(6) led to the generalized correlation Eq.(9), the calculation of heat flux (or as a result evaporation capacity) should take into consideration the value of the mean concentration of the solution in the liquid film and thus appearing effect of physical-chemical temperature depression. It is necessary to stress specifically that since the correlation presents data of the discrete-local character, which means that they were obtained within comparatively short sections of the channel, in which it was acceptable to assume a constancy of all regime parameters, phase flow rates and their properties. This determines a successive-iterative method of channel evaporative capacity calculations. A channel being under the calculation is to be split into a number of sections, and the calculation process starts from the upper inlet of the channel. Insofar as it can be seen from Eqs.(6) and (11), the local values of heat flux do not affect HTC directly, there is no need to initial guess of q to start iterative calculations to reach a convergence of guessed and obtained q values. Rather, it would be necessary to first calculate the local value of heat flux on the first section as a first approach, then to calculate the amount of steam generated in this section, and after this – determine all parameters dependent on the steam velocity. The next iteration would be calculation of a new value of HTC by Eq.(6) and (11) with the respective new values of heat flux and obtained in the previous calculation q and eventually - steam velocity, phase flow rates, concentrations, etc. The values of the local heat flux are to be determined with taking account of local value of physical-chemical temperature depression as well as its suppression by the concurrent steam flow

$$q = \alpha_m (t_w - t_{sat} - \Delta_{fc}(\bar{C}) + \delta\Delta t), \quad (10)$$

where $\delta\Delta t = (t_w - t_{av.m}^{exp}) - (t_w - t_i)$ - a function which accounts for the suppression effect of interphase shear stress.

The suppression effect may be explained by the fact that the steam being generated from the film has the temperature of the saturated solvent (water), whereas the upper layer of the film has local temperature sufficiently higher due to the concentration growth of the soluble substance. Being in the contact with the steam flow at a lower temperature, the surface of the film cools down; this leads to the decrease in physical-chemical temperature depression.

Analysis of the experimental data on the heat transfer to liquid films of sugar solutions with the concentrations up to 70% DM (Dry Matter) in the regimes of evaporation from the interphase surface as well as at bubble boiling with the concurrent steam flow allowed deriving a temperature function as Eq.(11):

$$\delta\Delta t = [1 - \exp(-2.3 \cdot 10^{-2} \sqrt{We}^3 \sqrt{Pe})] \Delta_{fs}. \quad (11)$$

The comparison of HTC calculations by Eq.(6) with the experimental data on the heat transfer to the free down flowing liquid films and films flowing co-currently with the steam at a velocity of 25 m/s of water and sugar solutions with the concentration up to 75% within the regimes of evaporation from the free surface. Unfortunately, it is impossible to separate the effect of certain factors on the intensity of heat transfer within the system under consideration. It is necessary to mark specifically that within the experimental stands, which simulate the conditions of actual film evaporators, all working and regime parameters manifest themselves in the unseparable mutually regulating unity.

surface and nucleate boiling are shown in figures 6 and 7. The transition to the regime of nucleate boiling is identified by the beginning of upward line $\alpha = f(q)$ on the graphs. It is clearly seen from the data given above that the transition to regime of nucleate boiling with the steam flow growth shifts towards to greater heat fluxes, which satisfactorily is identified by Eqs. (4) and (5).

For rough pipes ($R_c \geq 0.5 \cdot 10^{-5} m$), for example, after cleaning with the hard metal brushing or superimposing an artificial roughness according to Eq. (3) the transition to the nucleate boiling regime accompanied with the respective heat transfer intensification at lower values of heat flux.

Apart from the mentioned above factors, the heat transfer intensity is greatly affected by geometry of the channel, since as the height as well as the diameter of the channel determines the character of wavy structure on the film surface. Namely, in the 9 m boiling tubes of industrial evaporators along with the distance from the film distributor down to 3...4 m takes place an increase in heat transfer intensity compared to that in the short pipes at the same regime parameters and mass flow rates. On top of this, it has been found in [8,9] that an increase in pipe diameter entails the increase in length and amplitude of big waves that significantly intensifies the mixing in films.

For example, at a certain set of regime parameters and initial solution flow rate, down the Therefore, modelling of heat transfer to free falling films in such conditions becomes possible only within the upper sections of experimental pipe. Here the velocity of cocurrent steam is negligibly low but, at the same time, the wavy structure will be undeveloped. The

data for films with the developed waves might be taken at the middle or exit sections of pipes, but here the film flow

will be affected by the strong shear stress from the steam flow.

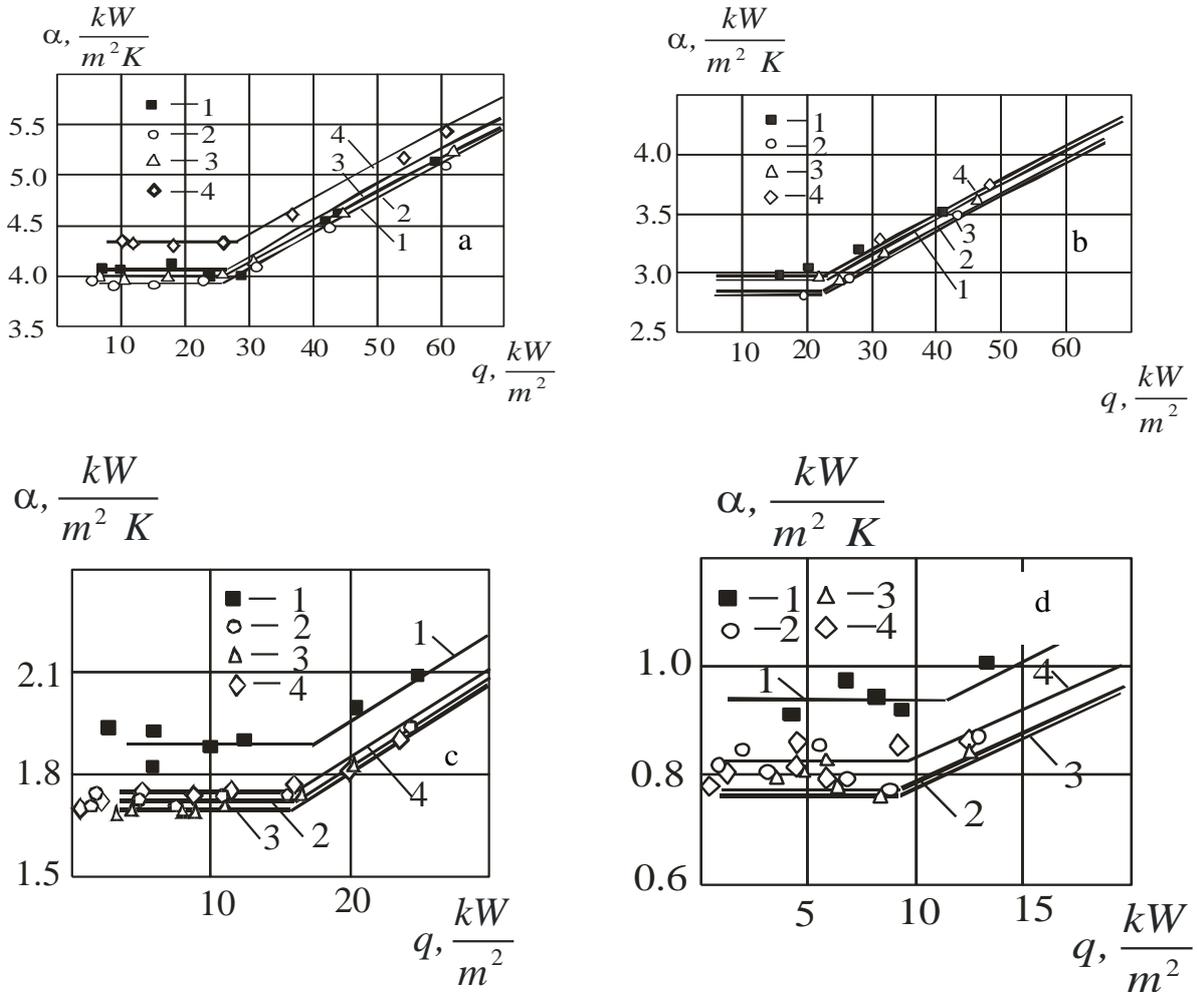


Fig. 6. Correlation $\alpha = f(q)$ for free falling films of water and sugar solutions,

($t = 100 \text{ } ^\circ\text{C}$, $R_c = 0.5 \cdot 10^{-5} \text{ m}$). Data[11].

a. water, $1 - \Gamma_v = 1 \cdot 10^{-4} \text{ m}^2/\text{s}$; $2 - 2 \cdot 10^{-4}$; $3 - 3 \cdot 10^{-4}$; $4 - 5.5 \cdot 10^{-4}$;

b. sugar solution, DM = 30 %, $1 - \Gamma_v = 1 \cdot 10^{-4} \text{ m}^2/\text{s}$; $2 - 1.6 \cdot 10^{-4}$; $3 - 2.2 \cdot 10^{-4}$; $4 - 4.5 \cdot 10^{-4}$;

c. sugar solution, DM = 50 %, $1 - \Gamma_v = 0.7 \cdot 10^{-4} \text{ m}^2/\text{s}$; $2 - 1.5 \cdot 10^{-4}$; $3 - 2.2 \cdot 10^{-4}$; $4 - 4 \cdot 10^{-4}$;

d. sugar solution, DM = 70 %, $1 - \Gamma_v = 0.5 \cdot 10^{-4} \text{ m}^2/\text{s}$; $2 - 2 \cdot 10^{-4}$; $3 - 3 \cdot 10^{-4}$; $4 - 5.5 \cdot 10^{-4}$.

Lines – calculated to Eq.(6) at the same mass flow rates.

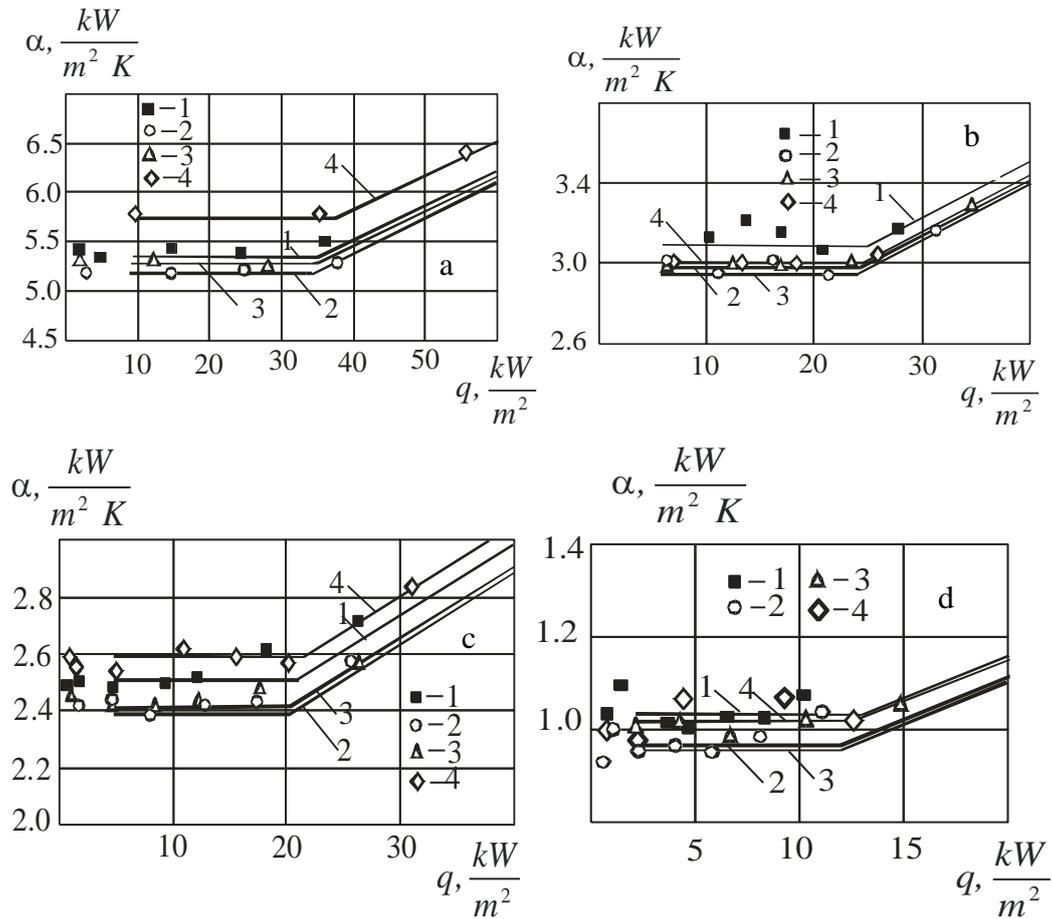


Fig.7. Correlation $\alpha = f(q)$ at steam velocity $u_2 = 25 \frac{m}{s}$, ($t = 100 \text{ } ^\circ C$, $R_c = 0.5 \cdot 10^{-5} m$).

Data [11]

- a. water, 1 – $\Gamma_v = 1 \cdot 10^{-4} m^2/s$; 2 – $2 \cdot 10^{-4}$; 3 – $3 \cdot 10^{-4}$; 4 – $5.5 \cdot 10^{-4}$;
- b. sugar solution, DM = 40 %, d = $1 \cdot 10^{-4} m^2/s$; 2 – $1.5 \cdot 10^{-4}$; 3 – $2 \cdot 10^{-4}$; 4 – $3.5 \cdot 10^{-4}$;
- c. sugar solution, DM = 50 % $\Gamma_v = 1 \cdot 10^{-4} m^2/s$; 2 – $2 \cdot 10^{-4}$; 3 – $3 \cdot 10^{-4}$; 4 – $6 \cdot 10^{-4}$;
- d. sugar solution, DM = 70 %, 1 – $\Gamma_v = 1 \cdot 10^{-4} m^2/s$; 2 – $2 \cdot 10^{-4}$; 3 – $3 \cdot 10^{-4}$; 4 – $5.5 \cdot 10^{-4}$

Lines – calculated to Eq. (6) at the same mass flow rates.

Therefore, the comparison and generalization of the experimental data obtained in long evaporating channels at the local level (within a comparatively short sections to suggest the constancy of parameters) appear grounded only in the event of equality all respective regime parameters, flow rates and physical properties of phases. In such case the data can be compared and generalized with

those obtained in the stands allowing individual formation of pphase flows and thus providing for the comparable working conditions. Data in figure 8 illustrate the given above. It shows the fields of scatter for data obtained in the experiments in short and long pipes of different diameters in the comparable conditions and their comparison with the results of calculation by Eq.(6).

The correction factor accounting for the effect of pipe length and diameter on the heat transfer was found by examining data obtained in pipes of different geometries but exclusively in the comparable conditions as Eq.(12):

$$K_L = \left\{ 1 + 0,06 \left(\frac{v}{v_o + v} \right) \left[1 - \exp(-0,05 L^3) \right] \right\} \times \left(\frac{d}{d_o} \right)^{0,35 - 0,06 \left(\frac{d}{d_o} \right)}, \quad (12)$$

where $d_o = 0.02 \text{ m}$, $v_o = 0.3 \cdot 10^{-6} \frac{\text{m}^2}{\text{s}}$,
valid for $L > 1.0 \text{ m}$.

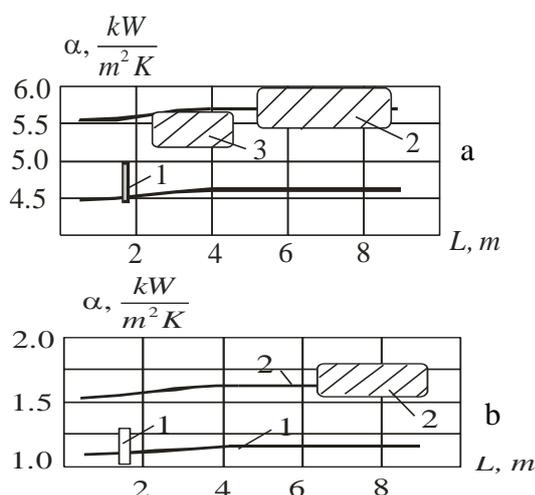


Fig. 8. Correlation $\alpha = f(L)$.

a. water, $u_2 = 15 \frac{\text{m}}{\text{s}}$; b. sugar solution,

DM = 70 %, $u_2 = 30 \frac{\text{m}}{\text{s}}$.

1 – data scatter field, $d = 20 \text{ mm}$, $L = 1.8 \text{ m}$ [11];
2 – data scatter field [4], $d = 30 \text{ mm}$, $L = 9 \text{ m}$;
3 – data scatter field [10] $d = 32 \text{ mm}$, $L = 4.9 \text{ m}$.
Lines – calculated by Eq.(6), 1 – $d = 20 \text{ mm}$, 2 – 30 mm .

The comparison of calculated by Eq. (6) values of heat flux along the height of a 9 m pipe with the experimental data [4] at various temperature differences is given in figure 9. As it is clearly seen from figure 9 the proposed correlations and methodology allows quire closely predict the evaporative capacity of boiling channels of industrial film evaporators with pipes up to 9 m height within a wide range of working parameters. Even at a big temperature

difference (wall – saturation) of 5°C , figure 9(c) when the local values of heat flux reach 35 kW/m^2 at the lower sections and, thus, one may expect rather significant values of the cocurrent steam flow velocity with affecting film surfacial wave structure, its supressing effects, named above, which manifest itself inseparably with the local heat flux – the recommended method gives remarkably good coincidece of measured and predicted data.

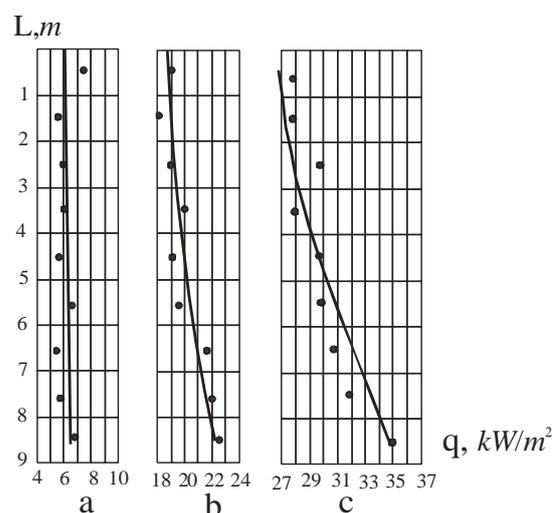


Fig.9. Heat flux distribution at a height of the evaporative pipe

Lines – calculation by (6) (6); Dots- experimental data [4].

a – $\Delta t = 2.2^\circ\text{C}$, $t_w - t_{\text{sat}} = 1.2$; b – 7.5 , $t_w - t_{\text{sat}} = 3.7$; c – 11 , $t_w - t_{\text{sat}} = 5.1$;

$\Gamma_v = 0.3 \cdot 10^{-3} \text{ m}^2/\text{s}$, water, $t_{\text{sat}} = 100^\circ\text{C}$.

4. Conclusion

Equations (3) and (4) which take into account the value of local wall superheat and heating surface roughness are recommended to be used as a factor determining the onset of nucleate boiling in films.

The temperature of liquid solution film at a regime of evaporation from free surface with the co-current steam flow is to be determined with taking into account the value of physical-chemical temperature depression Eq. (11).

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The Eq. (6) along with the methodology of its application may be used for the calculations of evaporative capacities of boiling channels of industrial film evaporators of sugar industry. The methodology is valid in application to water and sugar solutions with the concentrations up to 70% DM, at regimes of evaporation from free surface and nucleate boiling, within the range of vacuum up to 0.85 bar, heat fluxes up to 60 kW/m², liquid mass flow rate density 0.05...600 m²/s, co-current steam flow velocities up to 40 m/s, for evaporative pipes 20...32 mm diameter.