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Substantiation and the range of application of a new method for heat transfer prediction in condensing inside plain tubes

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Since the first work of Tepe and Mueller [1] and until now, there have been published hundreds of studies with the results of heat transfer investigations in vertical and horizontal tubes with vapour condensing of various liquids. Dozens of methods and formulae based both on the results of the theoretical research and on the experimental data have been proposed.

The existence of more than 50% discrepancy in different experimental data and various empirical and theoretical relationships is shown. Thus, the absence of both substantiation of different methods and explanations of methods disagreement both between themselves and with different experiments is noted. Also, there are often no remarks concerning boundaries for the use of proposed relationships.

There is proposed a simple semi-empirical correlation for heat transfer prediction in condensing inside the plain tubes at the annular and intermediate flow of the phases. This correlation is based on the nature of film condensation process and on the specific features of the results of theoretical solutions. The range of application of complexes, which determine the heat transfer process, is also substantiated.

Good convergence of the new method with the experimental data on condensation of steam, carbon dioxide, hydrocarbon refrigerants and other various fluids inside horizontal and vertical tubes is shown.

Keywords: film condensation, heat transfer, plain tube

INTRODUCTION

The interest to investigate the hydrodynamics and heat transfer in condensing inside tubes is always relevant due to the demand for improvement of various heat exchangers design, particularly in the evaporative systems of thermal desalinating plants, air conditioning systems, safety systems of reactors, heaters of power plants and condensers of cooling equipment. It is very important to know the exact values of condensation heat transfer coefficients when their value is close to heat transfer from the cooling side.

Since the publication of the first work of Tepe and Mueller [1] and till now, there have been published hundreds of researches on condensation inside plain tubes and channels. Reviews of some of these researches were carried out by Rifert et al. [2, 6], Garcia-Valladares [3], Kandlikar et al. [4], Dalkilic and Wongwises [5]. Also, there are good reviews in numerous dissertations, starting from the first one of Royal [7] to one of the recent of Macdonald [8]. However, in most reviews, the list of ten–twenty works with a brief description of the objects and findings of the cited authors, investigations are given without their critical analysis. In addition, there is no discussion on the limitations of performed investigations, the shortcomings of the proposed calculation methods, as well as there is no substantiation for new experiments. In most researches, the authors compare their own results with those of various theoretical and empirical calculation methods without any substantiation of their choice. In most cases, new correlations for heat transfer prediction are also proposed without any substantiation. Rifert et al. [6] noted a different degree of convergence of the same correlations for heat transfer calculation by different authors. In addition, it is shown that the accuracy of heat transfer prediction in theoretical solutions depends on the knowledge of shear stress or friction coefficient for which there have been no correct equations up to now.

A new correlation for heat transfer prediction in condensing inside tubes is proposed in this paper. This correlation is obtained by improving the existing semi-empirical relationships using the results of theoretical solutions for the laminar and turbulent flow of a condensate film under the influence of interfacial shear stress τ_w.

ANALYSIS OF THEORETICAL SOLUTIONS

The first theoretical solution for heat transfer prediction during film condensation under the influence of vapour velocity is made by Nusselt [9]. In the dimensionless form, Nusselt's solution takes the form:

$$Nu_f = 0.5(C_f Fr_l / Re_f)^{0.5}, \tag{1}$$

where $Nu_f = \alpha/\lambda_l(\nu_l^2/g)^{2/3}$ is film Nusselt number.

Equation (1) is used in many works for heat transfer coefficients definition in vapour condensing inside tubes. The main problem in determining Nu_f by (1) is the correct estimation of the friction coefficient C_f.

There are several solutions for heat transfer prediction under laminar and turbulent flow of a condensate film in vertical and horizontal tubes at annular phase flow. Particularly, Bae et al. [10] and Traviss et al. [11] presented the results of their solutions in a dimensionless form:

$$Nu_f = f(\beta, Re_l, Pr_l), \tag{2}$$

where $\beta = 0.5\ C_f Fr_l$ is parameter related to shear stress in the interphase.

Correlation (2) is plotted in [10] for Pr_l numbers from 1 to 5. As well as in the case of using (1), it is necessary to know the exact value of the friction coefficient C_f in a two-phase flow for determining Nu_f by (2). At low values of Re_l (less than 200) and equal values of parameter β, prediction of Nu_f numbers by (1) and (2) gives the same results.

Cavallini et al. [12] measured local heat transfer coefficients and pressure drop for condensation of freons R410A, R32, R134a, R125, R236ea. The authors reached general agreement between their experimental data and Kosky and Staub [13] theoretical solution. It should be noted that the results of the calculations by solution [13] are also dependent on the values of Re_l and C_f.

Rifert et al. [14] measured local heat transfer coefficients for condensation of steam and R22 inside a horizontal tube by the "thickness wall" method. The authors showed a good convergence of their own experimental data with the calculation by (2), on condition that the friction coefficient C_f is determined by the method they proposed. This method allows estimating the influence of the two-phase flow (parameter Φ_v^2) as well as vapour mass suction on interphase (parameter Φ_q) on the friction coefficient C_f by:

$$C_f = C_{fo}\Phi_v^2\Phi_q, \tag{3}$$

where $C_{fo} = 0.079/Re_v^{0.25}$ at $Re_v < 10^5$ or $C_{fo} = 0.046/Re_v^{0.2}$ at $Re_v \geq 10^5$; Re_v is vapour Reynolds number.

Parameter Φ_v^2 is determined by Miyara's [15] equations:

$$\Phi_v^2 = 1 + CX_{tt}^n + X_{tt}^2, \tag{4}$$

$$C = 21[1 - \exp(1 - 0.28\,Bo^{0.5})]\,[1 - 0.9\exp(-0.02Fr_l^{1.5})], \tag{5}$$

$$n = 1 - 0.7\exp(-0.08Fr_l). \tag{6}$$

Parameter Φ_q is calculated from [16] by:

$$\Phi_q = 1 + 17.5\,Re_v^{0.25}\,j, \tag{7}$$

where the suction parameter j can be obtained:

$$j = q/(rGx). \tag{8}$$

Comparison of heat transfer coefficients predicted by the Rifert et al. [14] method with experimental data in condensing propylene, propane, dimethyl ether, isobutene [17] and FC-72 [18] is shown in Figs. 1–2. Good agreement of the experiments with the calculations (divergence within 25%) proves the correctness of the Rifert et al. [14] method. The practical application of the Rifert et al. [14] method for calculating heat transfer under various condensation parameters is complicated by the necessity of the solution of the dependence (2) given in an implicit form. Solving the equation (2) is possible in two ways. Graphically, using the interpolation of graphs from [10] or numerically. However, the numerical solution (2), presented in [10, 11], has a very cumbersome and uncomfortable look for permanent use.

ANALYSIS OF SEMI-EMPIRICAL RELATIONSHIPS

In our previous work [6] we performed a detailed analysis of existing semi-empirical design methods. In this paper we discus three most well-known semi-empirical correlations. One of them belongs to Ananiev, Boyko and Kruzhilin [19]. The authors think that heat transfer in mix condensing is completely analogous to convective heat transfer in turbulent liquid flow in a tube:

$$Nu_d = c\,Re_{lo}^{0.8}\,Pr_l^{0.43}[1 + x(\rho_l/\rho_v - 1)]^{0.5}, \tag{9}$$

where Re_{lo} is only liquid Reynolds number.

In equation (9) two-phase flow was taken into account by introduction of the following complex $[1 + x(\rho_l/\rho_v - 1)]^{0.5}$.

As shown in [6], equation 9 is one of the most successful correlations for generalization of experimental data on vapour condensation inside horizontal and vertical tubes in a wide range of G, x and refrigerant physical properties. However, equation (9)does not work in the regime of phase flow, which is close to the stratified one.

Shah [20] suggested a simple method for heat transfer prediction, in which the correlation for one-phase convection in liquid turbulent flow in

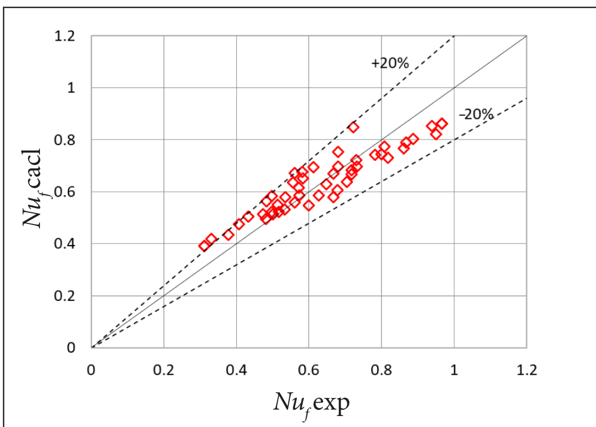

Fig. 1. Comparison of the Park at al. [17] experimental data with the Rifert et al. [14] method

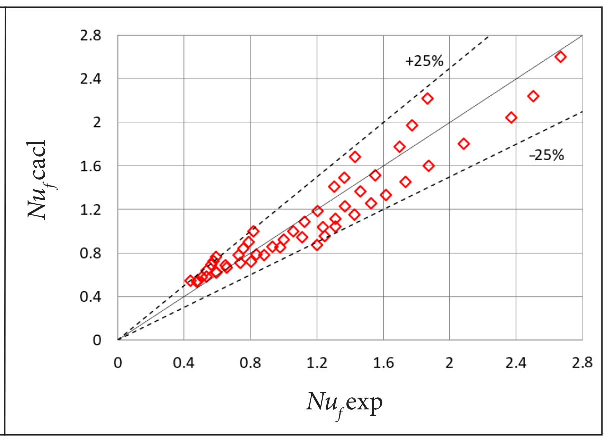

Fig. 2. Comparison of the Lee et al. [18] experimental data with the Rifert et al. [14] method

tubes is corrected by the function $\Psi = (1 - x)^{0.8} + 3.8x^{0.76}(1 - x)^{0.04}/p_r^{0.38}$, including vapour content x and reduced pressure p_r:

$$\alpha = 0.023\, Re_{lo}^{0.8}\, Pr_l^{0.43}(\lambda_l/d)\Psi. \tag{10}$$

In [21], Shah specified the correlation (10) inserting the new complex $(\mu_l/14\mu_v)^n$:

$$\alpha_1 = 0.023\, Re_l^{o0.8}\, Pr_l^{0.43}\,(\mu_l/14\mu_v)^n\Psi, \tag{11}$$

where $n = 0.0058 + 0.557\, p_r$.

Shah [21] proposed to apply equation (11) for heat transfer prediction if

$$J_g \geq 0.98(Z + 0.263)^{-0.62}, \tag{12}$$

where $Z = (1/x - 1)^{0.8}\, Pr_l^{0.4}$; J_g is dimensionless vapour velocity.

It should be noted that Borishanskij et al. [22] used reduced pressure p_r for heat transfer prediction in condensing earlier than Shah [20]. This fact was mentioned in [23].

In Table 1 it is shown the comparison of the additional complexes from (9) and for condensation of steam [24], methane [25], isobutene [17], freon R22 [14] and freon R245fa [26]. As seen from Table 1, the values of the additional complexes are practically the same (divergence within 25%). However, Ananiev's et al. correlation (9) is preferred to be used, because in contrast to Shah's formula (11), it has theoretical substantiation and clearer range of application.

The third known semi-empirical equation for heat transfer prediction in condensing inside the horizontal tube was devised by Thome et al. [27]:

$$Nu_d = 0.0039 f_i\, Re_\delta^{0.7}\, Pr_l^{0.5}, \tag{13}$$

where

$$Nu_d = \frac{\alpha\delta}{\lambda_l},\ Re_\delta = \frac{4G(1-x)\delta}{(1-\varepsilon)\mu_l},$$
$$f_i = 1 + (w_v / w_l)^{0.5}\,[(\rho_l - \rho_v)g\,\delta^2/\sigma]^{0.25}.$$

Table 1. **Comparison of the additional complexes from (9) and (11)**

No.	Work Fluid	$\frac{t_s}{p_r}$	$\frac{G}{x}$	$\frac{Re_{lo}}{Pr_l}$	$(\mu_l/14\mu_v)^n\Psi)$	$[1 + x(\rho_v - 1)]^{0.5}$
1	[24] steam	$\frac{189.1}{0.0557}$	$\frac{213}{0.84}$	$\frac{19395}{0.95}$	9.36	10.85
			$\frac{405}{0.55}$	$\frac{48322}{0.95}$	7.42	8.8
2	[25] methane	$\frac{-107.3}{0.4348}$	$\frac{200}{0.87}$	$\frac{19342}{1.9}$	3.61	2.95
			$\frac{200}{0.54}$		2.96	2.4
3	[14] freon R-22	$\frac{43.5}{0.3352}$	$\frac{226}{0.88}$	$\frac{28907}{2.4}$	4.65	3.68
		$\frac{45.5}{0.3515}$	$\frac{28}{0.55}$	$\frac{3669}{2.4}$	3.7	2.89
4	[17] isobutane	$\frac{40}{0.1459}$	$\frac{200}{0.82}$	$\frac{13725}{3.9}$	6.68	5.66
			$\frac{100}{0.5}$	$\frac{6863}{3.9}$	5.18	4.46
5	[26] freon R-245fa	$\frac{50}{0.0940}$	$\frac{300}{0.95}$	$\frac{7750}{5.2}$	8.37	7.92
			$\frac{500}{0.5}$	$\frac{12917}{5.2}$	6.11	5.79

The authors did not support the introduction of correct function f_i with any experimental data. This method is often used by researchers to compare with different experiments. It should be mentioned that formula (13) is used for the annular, intermediate and stratified phase flow. As shown in [28], there is a good convergence (about 20%) between the experiments on condensation of carbon dioxide inside the horizontal tube of d = 3.42 mm and l = 3.5 m at G = 200–800 kg/(m^2·s) with the calculations by equation (13). However, it was also shown in another article [29] that there is the discrepancy in 172% between the experimental data on condensation of carbon dioxide with similar specification and the data calculated by (13).

SUBSTANTIATION AND THE RANGE OF APPLICATION OF A NEW SEMI-EMPIRICAL CORRELATION

The analysis of the scientific papers devoted to comparing the experimental data on condensation inside tubes grounded on various methods of calculation shows great differences in the results obtained by different authors. In this case, the differences are observed when using the same methods (dependencies) for calculating heat transfer for both identical and different liquids. The discrepancy in the identical calculation methods for the same substances by different authors reaches up to 100% or more.

When using theoretical solutions, the main reason for such significant discrepancies is the use of different methods for calculating the friction coefficient C_f on the boundary of the phases. At the same time, it is also important to know the limits of applying this or that dependency for its comparison with theoretical solutions. Many empirical correlations are obtained in the narrow ranges of G, x, physical properties and therefore they correspond only to specific experimental methods.

The theory of film condensation when influenced by the vapour velocity [10] shows that the Nu_f number is the function of three parameters β, Re_l, Pr_l (see eq. (2)). As can be seen from [10], depending on the value of each of these parameters, the theory predicts a different degree of their effect on heat transfer. For instance, in the region of predominantly laminar or laminar-wave film flow, the effect of Pr_l on Nu_f number decreases with decreasing Re_l up to the complete absence.

The accuracy of calculating Nu_f number in the region of β influence depends on the accuracy of the determination of friction coefficient C_f. Up till present, there has been no valid methodology for calculating C_f. For example, Macdonald [8] considers more than 20 formulae for calculating C_f. None of them has sufficient (within ±25%) accuracy when compared to the experiments.

The analysis of the theory [10] shows that for β > 10 (under the influence of vapour velocity on heat transfer) and Pr_l = 1–3 under the changes of Re_l from $8 \cdot 10^2$ to $2 \cdot 10^4$, the impact of Re_l on Nu_f can be neglected. Figure 3 shows the influence of

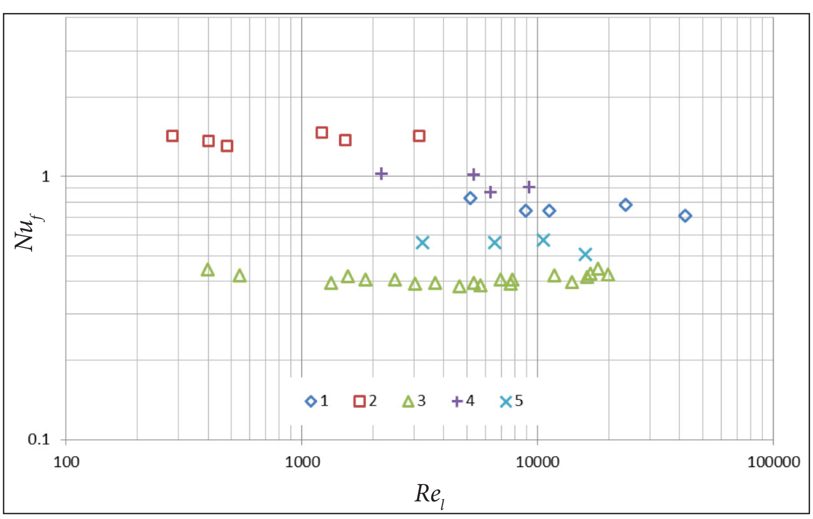

Fig. 3. Influence of Re_l on Nu_f at constant values of Pr_l and β: 1 – data from Boyko et al. [24] for steam at Pr_l = 0.95, β = 150; 2 – data from Lee et al. [18] for fluid FC-72 at Pr_l = 8.2, β = 40; 3 – data from Yu et al. [30], Rifert et al. [14], Park et al. [17], Kwon et al. [31] for freon R22 and from Kim et al. [28] for carbon dioxide at Pr_l = 2.4, β = 10; 4 – data from Kim et al. [28] for carbon dioxide at Pr_l = 2.4, β = 100; 5 – data from Boyko et al. [24] for steam at Pr_l = 0.95, β = 100

Re_l on Nu_f at constant values of Pr_l and β, based on the experimental data of different authors. As Fig. 3 displays, only at high vapour velocities (β more than 100) there is an insignificant decrease of Nu_f with an increase of Re_l. The boundaries for Re_l variations from $8 \cdot 10^2$ to $2 \cdot 10^4$ are characteristic of all freons (R22, R134a, R125, R32, R410A), hydrocarbons (propane, isobutane, propylene, dimethyl ether) and carbon dioxide at G from 200 to 800 kg/m²s and at β more than 10. These particularities of theoretical calculations [10] became the basis to generalize the data on the local heat transfer coefficients of many works on condensation inside both vertical and horizontal tubes by the formula $Nu_f = f(Fr_l, Pr_l)$.

The absence of the friction coefficient in this dependence (it is used in the theoretical solutions [10, 11] to determine the parameter β) can be explained as follows. The number Fr_l increases along with increasing w_v^2, and the value C_f in this case (if the influence of q/rG is ignored) proportionally decreases $w_v^{2-0.25}$ depending on the value Re_v. For this reason, the influence of vapour velocity on Fr_l is much greater than the influence of vapour velocity on the friction coefficient.

We summarized the following data: Park et al. [17] data on condensation of isobutene, dimethyl ether, propane, propylene; Kim et al. [28] data on condensation of carbon dioxide; Cavallini et al. [12] data on condensation of freons R134a, R125, R32, R410A, R236ea; data on steam condensation inside horizontal [24] and vertical [32] tubes; Ghim and Lee [26] data on condensation of organic fluids R245fa, NOVEC649 and HFE-7000. Of all the experiments, the data were processed at such values of mass velocities G and vapour contents x when, according to the recommendation of [27], there was observed an annular or intermediate regime. This usually occurs when β > 5 and when $Fr_l > 500$. In Fig. 4 the data of authors mentioned above are generalized by such correlation:

$$Nu_f = 0.0144 Fr_l^{0.36} Pr_l^{0.43}. \quad (14)$$

As seen in Fig. 5, equation generalizes all experimental data with the error less than 25%.

In order to confirm the accuracy of the new method, its verification is performed with experimental data from the works of the following scientists: concerning the condensation of freon R-134a in a vertical tube – Dalkilic et al. [33]; fluid FC-72 – Lee et al. [18]; propane – Macdonald [8]; methane – Zhuang et al. [25] and carbon dioxide – Peihua Li et al. [34]. The results are shown in Fig. 6, which make it evident that the developed method of calculation with an accuracy of ±30% generalizes all experimental data in the annular and intermediate modes of phase flow.

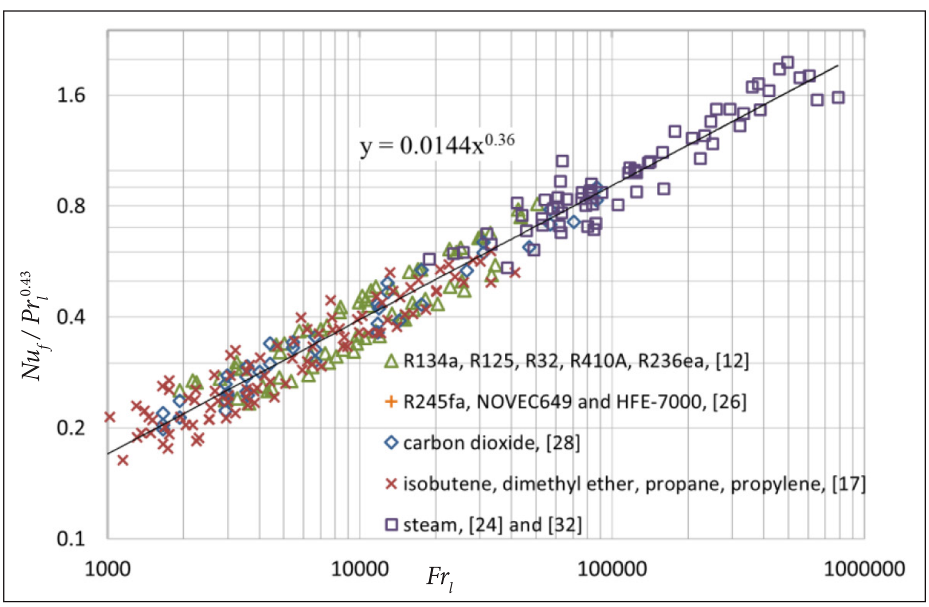

Fig. 4. Approximation of the experimental data of different authors in the form $Nu_f = f(Fr_l, Pr_l)$

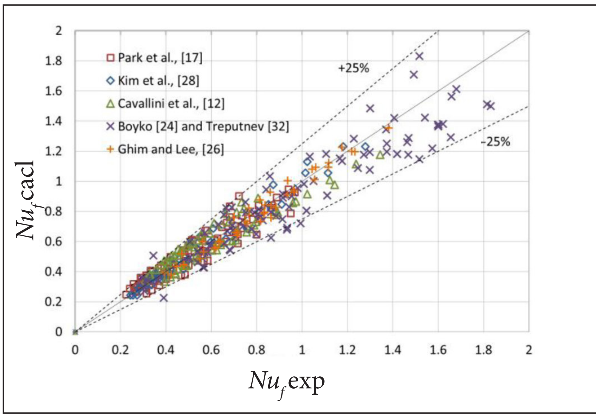

Fig. 5. Calculated vs. experimental heat transfer coefficients: predictions by the proposed method

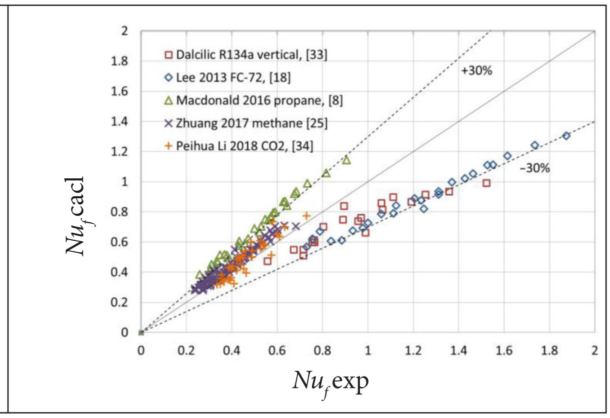

Fig. 6. Application of the new method to various experimental data

CONCLUSIONS

The outcomes of our theoretical investigation are the following:

1. There was obtained a new correlation for heat transfer prediction, which is based on the theoretical model of turbulent condensation and can be expressed in the form offered: $Nu_f = 0.0144 Fr_l^{0.36} Pr_l^{0.43}$. This equation generalizes a large number of experimental data on condensation of different refrigerants inside horizontal and vertical tubes.

2. The formula suggested the calculation does not need any correct estimation of phase flow regimes, condensate accumulation and a friction coefficient.

3. The two restrictions should be considered when using the suggested correlation: firstly, Fr_l must be more than 500; secondly, β must be more than 5.

NOMENCLATURE

Bo – Bond number (= $gd^2(\rho_l - \rho_v) / \sigma$)
C_f – friction coefficient
d – inner diameter of tube, m
Fr_l – liquid Froude number (= $\frac{\tilde{n}_v(\tilde{n}_l - \tilde{n}_v)w_v^2}{\tilde{n}_l^2(\nu_l g)^{2/3}}$)
G – mass velocity, kg/(m²s)
g – gravitational acceleration, m/s²
J_g – dimensionless vapour velocity (= $xG/[gd\rho_v(\rho_l - \rho_v)]^{0.5}$)
l – length of the tube, m
Nu – Nusselt number
p_s – saturated pressure, Pa
p_{cr} – critical pressure, Pa
Pr – Prandtl number
p_r – reduced pressure (= ρ_s/ρ_{cr}), Pa
q – heat flux, W/m²
r – heat of vaporization, J/kg
Re_f – film Reynolds number ($\mu = ql/(r\mu_l)$)
Re_l – liquid Reynolds number (= $G(1 - x)d/\mu_l$)
Re_{lo} – only liquid Reynolds number (= Gd/μ_l)
Re_v – vapour Reynolds number (= Gxd/μ_v)
t_s – saturated temperature, °C
w – velocity, m/s
x – mass vapor quality

Greek Symbols

α – heat transfer coefficient, W/(m²K)
β – parameter related to shear stress in the interphase
δ – thickness of the condensate film, m
λ – thermal conductivity, W/(mK)
μ – dynamic viscosity, Pa·s
ν – kinematic viscosity, m²/s
ρ – density, kg/m³
σ – surface tension, N/m
τ_w – shear stress, Pa
Φ_v^2 – parameter that takes into account the influence of two-phase flow on shear stress
Φ_q – parameter that takes into account surface suction at the interphase

Subscripts

d – dimensionless
f – film
l – liquid
v – vapour
exp – experimental
calc – calculated

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NAUJOJO ŠILUMOS MAINŲ PROGNOZAVIMO METODO PAGRINDIMAS IR NAUDOJIMO SRITIS VYKSTANT KONDENSACIJAI VAMZDŽIUOSE

Santrauka

Nuo pirmojo J. Tepe ir A. Muellerio [1] darbo paskelbti šimtai tyrimų, kuriuose buvo pateiktos šilumos perdavimo vertikaliuose ir horizontaliuose vamzdeliuose su įvairių skysčių garų kondensacija tyrimų išvados. Pasiūlyta daugybė metodų ir formulių, pagrįstų tiek teorinių tyrimų rezultatais, tiek eksperimentiniais duomenimis.

Parodyta, kad eksperimentinių duomenų ir įvairių empirinių bei teorinių skaičiavimų neatitiktys siekia per 50 %. Trūksta skirtingų metodų pagrįstumo ir metodų neatitikimų tiek tarpusavyje, tiek ir su skirtingais eksperimentais paaiškinimų. Be to, dažnai nepateikiama pastabų dėl pasiūlytų sąsajų naudojimo ribų.

Pasiūlyta paprasta pusiau empirinė koreliacija šilumos perdavimui ir kondensacijai paprastų vamzdžių viduje žiediniame ir tarpiniame fazių srautuose prognozuoti. Ši koreliacija pagrįsta plėvelės kondensacija ir teorinių sprendimų rezultatų specifika. Taip pat pagrįsta šilumos perdavimo taikymo sritis.

Parodyta gera naujojo metodo konvergencija su garų, anglies dvideginio, angliavandenilių aušalų ir kitų įvairių skysčių, esančių horizontaliuose ir vertikaliuose vamzdeliuose, kondensacijos eksperimentiniais duomenimis.

Raktažodžiai: plėvelinė kondensacija, šilumos perdavimas, paprastas vamzdis