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A generalized correlation for two-phase forced flow heat transfer—second assessment

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Abstract—A generalized correlation for two-phase forced flow heat transfer (nucleate boiling and vaporization) is suggested which is valid for vertical and horizontal channels with a fully wetted perimeter. It represents, with a mean absolute deviation equal to 14.4%, experimental data for 21 different liquids (water, organic liquids, freons and cryogens) in the following ranges of the main parameters: pressure, 0.61–196 bar; heat flux density, $10–8\times10^6$ W m^{-2}; mass flow rate, 5.6–6240 kg m^{-2} s^{-1}; channel diameter, 0.47–74.7 mm. It has been found that both the nucleate boiling and the vaporization heat transfer intensity depend on the thermal conductivity of the channel wall material. It is demonstrated that nucleate boiling heat transfer is described by a single equation in dimensionless variables containing an individual constant for each of the four groups of fluids (water, organic liquids, freons and cryogens). Vaporization heat transfer is described by a single equation with a constant universal for all the fluids.

INTRODUCTION

DURING the last decade, a number of papers [1–4] were published by the present author, where, on the basis of extensive information accumulated by that time on boiling and vaporization heat transfer in two-phase flow of various fluids, an attempt was made to suggest a universal predictive correlation. In a recent paper [5], a correlation for forced convection vaporization was obtained, which is valid for various fluids (water, freons, cryoagents) in a wide range of the main parameters specifying the process, as well as the relationship for nucleate boiling. But since the latter relation had been initially developed exclusively for cryogenic fluids, it was not subjected to intensive verification by experimental data for other substances. The aim of the present paper is to refine this relationship and to extend it to non-cryogenic fluids. This reassessment is based on extensive experimental information accumulated over the last four decades and, which is of no less importance, on the up-to-date data for thermal properties the acquisition and processing of which are being constantly carried out at the author's laboratory. Simultaneously the correlation for forced convection vaporization [5] was also subjected to verification against an expanded body of data.

PREVIOUS STUDIES

In a recent paper [5], the following correlation has been suggested for two-phase flow heat transfer:

$$Nu_{TP} = \begin{cases} Nu_b & \text{with} \quad N_{CB} < 1.6\times10^4 \\ Nu_c & \text{with} \quad N_{CB} > 1.6\times10^4 \end{cases} \qquad (1)$$

where the nucleate boiling heat transfer is found from the formula

$$Nu_b = 7.4\times10^{-3}\, Pe_*^{0.6}\, K_p^{0.5}\, Pr_l^{-1/3}\, K_\lambda^{0.15} \qquad (2)$$

and the forced convection vaporization heat transfer is given by the relation

$$Nu_c = 0.087 Re_m^{0.6}\, Pr_l^{1/6} (\rho_v/\rho_l)^{0.2} K_\lambda^{0.09}. \qquad (3)$$

Equation (3) was verified by experimental data on evaporation of nine different liquids (water, freons, cryogens) in wide ranges of the main governing parameters: pressure (0.61–30.4 bar), mass flow rate (50–2690 kg m^{-2} s^{-1}), quality (0.017–1.00), channel diameter (1.63–41.3 mm). Altogether, 553 experimental points were used for comparison, with the mean absolute deviation from the calculation amounting to 12.9% against 24.1 and 34.2% for the familiar relationships of Chen and Shah, respectively.

Equation (2) was derived on the basis of the data for nucleate boiling of only cryogenic liquids (helium, hydrogen, neon, nitrogen and argon). Originally, 440 experimental points were used for processing, of which 417 points (i.e. almost 95%) deviated from those calculated by no more than 35%; the mean absolute deviation did not exceed 17%. Equation (2) was compared with selected data on boiling water and freons [8–12] showing a satisfactory agreement. However, on the whole, the question about the possibility of applying equation (2) to nucleate boiling heat transfer calculation for non-cryogenic liquids remains unsolved, because up to now this equation has not been thoroughly checked on a great number of experimental data.

EXPERIMENTAL DATA USED TO DEVELOP AND TEST A NEW CORRELATION

Table 1 contains the principal information concerning experimental investigations that provided a

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NOMENCLATURE

a	thermal diffusivity
A	Philippov's number, 100π at $\tau = 0.625$
b	Laplace constant, $\sqrt{(\sigma/g(\rho_l-\rho_v))}$
C_p	specific heat per unit mass at constant pressure
d	channel diameter
g	body force acceleration
G	total mass flow rate
K_λ	relative thermal conductivity, λ_w/λ_l
K_p	dimensionless parameter, pb/σ
l	channel length
N_{CB}	convective boiling number, $(rG/q)[1+x(\rho_l/\rho_v-1)](\rho_v/\rho_l)^{1/3}$
Nu	Nusselt number, $\alpha b/\lambda_l$
p	pressure
Pe_*	modified Peclet number, $qb/r\rho_v a_l$
Pr	Prandtl number
q	heat flux density
r	latent heat of vaporization
R_a	mean microroughness
Re_m	Reynolds number of the mixture, $W_m b/\nu_l$
T	temperature
ΔT	temperature difference, $T_w - T_s$
w_m	two-phase mixture velocity, $(G/\rho_l)[1+x(\rho_l/\rho_v-1)]$
x	quality.

Greek symbols

α	heat transfer coefficient
θ	contact wetting angle
λ	thermal conductivity
μ	dynamic viscosity
ν	kinematic viscosity
π	reduced pressure, P/P_{cr}
ρ	density
σ	surface tension
τ	reduced temperature, T/T_{cr}.

Subscripts

b	boiling
c	two-phase forced convection (vaporization)
calc	calculated
cr	critical
e	equivalent
exp	experimental
l	liquid at saturation
m	mixture
s	saturation
TP	two-phase
v	vapour at saturation
w	wall.

basis for developing a new correlation. Use was made of nucleate boiling data for 21 various fluids (water, organic fluids, freons, cryogens) in vertical and horizontal pipes and annular channels (without stratified flow regimes). It should be pointed out that the initial file of experimental points (about 3000 points) was thoroughly analysed and after preliminary calculations part of these were withdrawn from it on the following grounds.

(1) A considerable part of the experiments, especially for freons and cryogens, was conducted in horizontal channels in which there is a possibility for the existence of regimes with periodically or continuously unwetted upper generatrices of the tube (plug, slug, stratified and wavy regimes). This may result in a sharp asymmetry of circumferentially local heat transfer coefficients and in the decrease of the section-averaged α's. The latter regimes are not considered in the present work. Flow patterns were identified with the aid of a map given in ref. [42] and based on an extensive experimental material for non-adiabatic two-phase flows. As one should expect, this new map predicts a marked expansion of the boundaries of stratified regimes in comparison with the maps based on adiabatic flow observations [91, 92]. It should be emphasized that in rather a broad range of regime parameters the new map is in good agreement with Schicht's map modified in ref. [5] (see Fig. 1 in that work).

(2) The accuracy of measurements of the local T in two-phase channel flow strongly depends on the type of experimental set-up, sensors and instruments employed and lies within the range from 0.01 K for precision measurements on helium [15] to 1.5 K [28]. Bearing in mind the fact that the desired accuracy of the predictive correlation is at the level of, at least, $\pm 35\%$, the experimental data, for which the relative error in the measurements of the local ΔT, and, hence, also of α, exceed 30% were discarded as unreliable.

(3) A characteristic feature of nucleate boiling is a strong dependence of the heat transfer coefficient on the heat flux density described by the power dependence $\alpha \sim q^n$, where n lies in the range 0.4–0.7, in accordance with the data for channels. A diminution in this dependence for high heat fluxes recorded, for instance, in refs. [30, 43] seems to be attributed to the heat transfer crisis developing in downstream channel sections. The development or even the approach of the crisis radically changes the hydrodynamic and thermal situation in the channel, resulting, in particular, in the appearance of a large axial heat flow, which eventually leads to the apparent weakening of the $\alpha(q)$ dependence. For this reason, the points for which $n < 0.4$ in the region of high fluxes were ignored.

(4) As is known, in the region of low heat fluxes

or high G's and x's a transition to forced convection vaporization occurs, which is accompanied by the weakening of the function $\alpha(q)$ until complete disappearance of this dependence.

Vaporization is a specific type of heat transfer in which the relationship between the heat transfer coefficient and regime parameters essentially differs from that for nucleate boiling. In order to separate the nucleate boiling region from that of two-phase convection (vaporization), use was made of the convective boiling number

$$N_{CB} = (rG/q)[1+x(\rho_l/\rho_v-1)](\rho_v/\rho_l)^{1/3}$$

obtained in ref. [5]. The condition $N_{CB} < 1.6\times10^4$ identifies the nucleate boiling region, while the condition $N_{CB} > 1.6\times10^4$ identifies the vaporization region.

Taking into consideration the restrictions considered, the total number of 2500 experimental points were processed for the region of nucleate boiling. All the experimental data used were directly obtained from the experiments for which an exhaustive primary information was available and therefore no prior data smoothing was made. Thus, the correlation given below indirectly takes into account rather a large scatter (up to $\pm15\%$) characteristic for the two-phase flow heat transfer data.

DEVELOPMENT OF CORRELATION FOR NUCLEATE BOILING WITH FORCED CONVECTION

The correlation is constructed in the system of dimensionless numbers which was originally suggested in ref. [2], i.e. in the form: $Nu_b = Nu_b(Pe_*, K_p, Pr, K_\lambda)$, with the solution being sought in a conventional form of the product of power functions

$$Nu_b = \text{const.}\, Pe_*^{n_1} K_p^{n_2} Pr_l^{n_3} K_\lambda^{n_4}. \quad (4)$$

In the first place, attention was focused on finding the dependence of the heat transfer coefficient on the thermal properties of the channel wall material given by the number K_λ. The question about the effect of the wall material remains unsolved up to now; moreover, the correlations suggested in refs. [1–5] are the only ones that predict the existence of such an influence. It is natural that the discovery of this effect is more probable, the broader the range of K_λ that varied in the experiment. Of the four groups of liquids listed in Table 1, cryogenic liquids are most preferable for research, because K_λ varied in a very broad range—from 14 to 8.2×10^5—and the tubes used in experiments were made of five different materials: silver, copper, inconel, monel and stainless steel. The treatment of experimental data from Table 1 is shown in Fig. 1, from which it follows that the effect of the thermal properties of the wall can be described by a simple power function to which there corresponds an equation of the form

$$Nu_b = 9.7\times10^{-3} Pe_*^{0.6} K_p^{0.5} Pr_l^{-0.33} K_\lambda^{0.12}. \quad (5)$$

Expression (5) differs from equation (2) by predicting a somewhat weaker dependence of the heat transfer coefficient on the wall material. It is necessary to emphasize that expression (5) takes into account additional experimental information accumulated over the last few years [21, 22, 24, 35, 39–42] and, besides, rests on the constantly updated bank of the thermal properties of cryogenic fluids. References [93–95] are an example of recent contributions to this bank. It should also be noted that the existence of the effect of the wall material thermal properties was experimentally demonstrated in ref. [96]. Measurements carried out in the latter work displayed an increase in the heat transfer coefficient for nitrogen nucleate boiling in a copper tube by approximately 50% compared with a stainless steel tube in the same set-up and at the same values of all regime parameters (heat flux, pressure, mass flow rate and quality). This effect corresponds exactly to the value of $n_4 = 0.12$.

Having thus established the degree of the influence of the wall thermal conductivity on heat transfer rate, one can turn to further improvement of the accuracy of equation (4) representing the functional relationship of the heat transfer coefficient with heat flux, pressure, and the thermal properties of the fluid and of the wall. As earlier [1–5] it will be assumed that $n_1 = 0.6$. This assumption reflects the fact that in channel nucleate boiling there is a weaker function

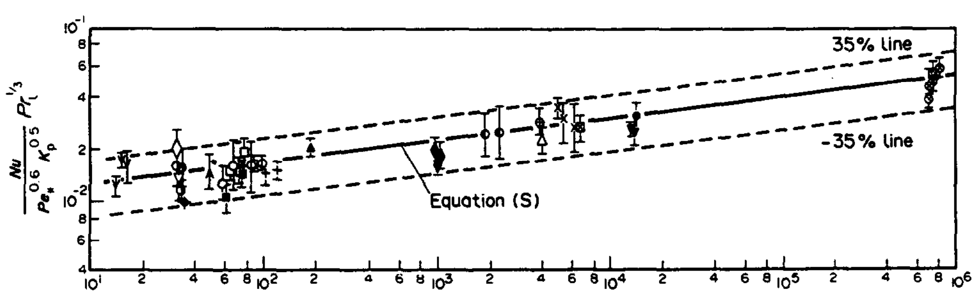

FIG. 1. Nucleate boiling heat transfer rate vs the thermal conductivity of tube wall material for cryogens in a forced flow. For symbols, see Table 1.

Table 1. Experimental investigations analysed to develop and test the proposed correlation

Ordinal No.	Authors	Fluid	Test section: dimensions (mm); orientation	Material	Pressure (bar)	Mass flow rate (kg m^{-2} s^{-1})	Heat flux density (kW m^{-2})	Number of experimental points: Vaporization $N_{CB} > 1.6\times10^4$	Number of experimental points: Nucleate boiling $N_{CB} < 1.6\times10^4$	Symbols
1	2	3	4	5	6	7	8	9	10	11
				Cryogenic fluids						
1	La Harpe *et al.* [13]	Helium	Coiled tube: $l = 4885$, $d = 3.0$; vertical	Monel	1.2	25–120	0.021–0.56	0	13	*
2	Johannes [14]	Helium	Tube: $l = 296$, $d = 2.12$; vertical	Monel	1.11	62–230	0.145–3.2	0	18	∧
3	Hildebrandt [15]	Helium	Tube: $l = 20$, $d = 1.0$; vertical	Silver	0.2–0.9	5.6–76.3	0.018–1.77	0	29	⊗
4	Ogata and Sato [16]	Helium	Tube: $l = 85$, $d = 1.09$; vertical	Stainless steel	1.1–1.9	79–163	0.040–4.0	0	35	◑
5	Keilin *et al.* [17]	Helium	Tube: $l = 100$, $d = 2.0$; vertical	Copper	1.11–1.52	18–96	0.145–3.6	0	25	◆
6	Giarratano *et al.* [18]	Helium	Tube: $l = 100$, $d = 2.13$; vertical	Stainless steel	1.11	67–315	0.080–1.8	0	12	▽
7	Antipov and co-workers [19, 20]	Helium	Tube: $l = 130$, $d = 0.67$; vertical	Stainless steel	1.0	14–64	0.010–1.13	0	26	◇
8	Romanov *et al.* [21]	Helium	Tube: $l = 100$, $d = 0.47$; vertical	Copper	1.02	34	0.060–2.15	0	7	⊕
9	Anisimov [22]	Helium	Tube: $l = 130$, $d = 1.8$; vertical	Stainless steel	1.0	22–135	0.035–4.0	0	53	
10	Solodovnikov and co-workers [23, 24]	Helium	Tube: $l = 180$, $d = 1.63$; vertical	Stainless steel	1.0–1.82	65–110	0.030–1.82	8	35	∨
11	Core *et al.* [25]	Hydrogen	Tube: $l = 40$, $d = 4.25$; vertical	Stainless steel	6.45	520	32.8–131	0	2	♦
12	Walters [26]	Hydrogen	Tube: $l = 152$, $d = 6.35$; horizontal	Copper	1.62–2.23	470–950	10.4–129	0	14	▼

13	Mohr and Runge [27]	Neon	Tube: $l = 100$, $d = 4.0$; horizontal	Copper	1.5	77–131	0.4–49	42	27	●
14	Lewis *et al.* [28]	Nitrogen	Tube: $l = 144$, $d = 14.1$; vertical	Stainless steel	3.46–3.8	20.5–76.6	30.1–75.7	0	8	□
15	Troshin and Episthov [29]	Nitrogen	Tube: $l = 900$, $d = 3.87$; horizontal	Stainless steel	0.9–1.0	92.4–199	4.74–34.5	19	10	○
16	Klein [30]	Nitrogen	Tube: $l = 1000$, $d = 12$; horizontal	Copper	3.0	206–257	1.0–31.0	28	5	△
17	Steiner and co-workers [31, 32]	Nitrogen	Tube: $l = 525$, $d = 14$; horizontal	Copper	5.2–20.4	117–562	0.30–37.8	18	102	×
18	Papell and Hendricks [33]	Nitrogen	Tube: $l = 250$, $d = 1.98$; vertical	Inconel	21.7	2210	212	0	5	▲
19	Dolgoy *et al.* [34]	Nitrogen	Tube: $l = 200$, $d = 1.6$; horizontal	Stainless steel	1.46–3.0	61–203	1.76–45.4	0	15	■
20	Deev *et al.* [35]	Nitrogen	Tube: $l = 212$, $d = 2.6$; vertical	Stainless steel	3.4–15.0	490–2130	7.2–145	0	23	+
21	Anisimov [22]	Nitrogen	Tube: $l = 130$, $d = 1.8$; vertical	Stainless steel	1.0	67–300	1.5–70	0	47	
22	Klimenko and Sudarchikov [36–38]	Nitrogen	Tube: $l = 1850$, $d = 10$;	Stainless steel	1.8–9.7	130–790	2.3–46	7	193	⦶
23	Grigoriev [41]	Nitrogen	Tube: $l = 1630$; $d = 10$; vertical	Copper	1.75–4.6	242–517	3.2–46	57	104	◐
24	Klimenko *et al.* [39, 40]	Nitrogen	Tubes: $l = 500$, $d = 4$; 9; 15.8 and 18.9; vertical	Stainless steel	1.5–8.0	60–800	0.65–65	43	102	◫
25	Fyodorov [42]	Nitrogen	Tube: $l = 600$, $d = 14.1$; horizontal	Stainless steel	2.0–9.0	160–700	2.8–52	11	102	
26	Müller-Steinhagen *et al.* [43, 44]	Argon	Tube: $l = 175$, $d = 14$; horizontal	Copper	1.67–19.6	120–447	0.22–16	4	11	⊠
	Total cryogenic fluids							237	1023	

Table 1 (*continued*)

Ordinal No.	Authors	Fluid	Test section: dimensions (mm); orientation	Material	Pressure (bar)	Mass flow rate (kg m^{-2} s^{-1})	Heat flux density (kW m^{-2})	Number of experimental points: Vaporization $N_{CB} > 1.6 \times 10^4$	Nucleate boiling $N_{CB} < 1.6 \times 10^4$	Symbols
1	2	3	4	5	6	7	8	9	10	11
				Refrigerants						
27	Bryan and Quaint [45]	Freon 11	Tube: $l = 3048$, $d = 8$; horizontal	Copper	1.2–1.6	137–587	3.17–18.1	17	0	
28	Chawla [46]	Freon 11	Tube: $l = 119$, $d = 6$; horizontal	Copper	0.61	78.7–251	4.65–93	43	24	
29	Kihara *et al.* [12]	Freon 11	Annular channel: $l = 686$, $d_e = 27.6$; vertical	Inconel	1.37	335–411	4.32–49.6	0	8	
30	Bogdanov [48]	Freon 12	Tube: $l = 1500$, $d = 12$; horizontal	Copper	4.23	273–663	3.49–23.3	0	16	
31	Uchida and Yamaguchi [49]	Freon 12	Tube: $l = 1000$, $d = 6.4$; horizontal	Stainless steel	3.92	259–518	2.44–27.9	65	7	
32	Bandel and Schlünder [51]	Freon 12	Tube: l not specified, $d = 14$; horizontal	Copper	3.13	239–441	5.0–71	17	10	
33	Chaddock and Noerager [50]	Freon 12	Tube: $l = 1930$, $d = 11.7$; horizontal	Stainless steel	4.4	350–587	6.84–35.0	15	13	
34	Vaihinger [52]	Freon 12	Tube: $l = 1600$, $d = 20.5$; vertical	Stainless steel	11.8–33.4	860–5600	9.0–87	0	21	
35	Mayinger and Ahrens [53]	Freon 12	Tube: $l = 5000$, $d = 14$; vertical	Stainless steel	9.0–22	300–1200	10–67	0	29	
36	Malyshev *et al.* [10]	Freon 12	Tube: $l = 90$, $d = 10$; horizontal	Stainless steel	1.63	224	2.0–25	14	3	
37	Kirin *et al.* [54]	Freon 12	Tube: $l = 400$, $d = 8$; vertical	Copper	5.67	50–240	2.0	11	2	
38	Pujol and Stenning [55]	Freon 113	Tube: $l = 1295$, $d = 15.8$; vertical	Stainless steel	2.0	586	16.9–56	16	16	

39	Lazarek and Black [56]	Freon 113	Tube: $l = 126$, $d = 3.15$; vertical	Stainless steel	1.54	502	64–178	0	14
40	Gouse and Coumou [57]	Freon 113	Tube: $l = 3810$; $d = 10.9$; horizontal	Glass	1–1.5	519–700	12.9–22.1	3	7
41	Fagerholm *et al.* [58, 59]	Freon 114	Tubes: $l = 9000$ and 2800, $d = 8$; vertical and horizontal	Stainless steel	3.1–5.3	625–2120	15.4–84.1	5	27
42	Altman *et al.* [60]	Freon 22	Tube: $l = 1219$, $d = 8.7$; horizontal	Copper	5.7–10.5	146–573	9.41–35.9	3	3
43	Bogdanov [61]	Freon 22	Tube: $l = 1500$, $d = 12$; horizontal	Copper	6.8	376–626	5.8–23	0	10
44	Kirin *et al.* [62]	Freon 22	Tube: $l = 100$, $d = 10$; vertical	Stainless steel	2.45–9.1	50–250	5–20	34	40
45	Bartau [63]	Ammonia	Tube: $l = 474$, $d = 30.3$; vertical	Nickel	34.3–108	1000	52–1240	0	18
	Total refrigerants							243	268
				Organic fluids					
46	Blinov *et al.* [64]	Propane	Tube: $l = 990$, $d = 8$; vertical	Stainless steel	3.0–11.5	300	4.9–135	0	16
47	Blinov *et al.* [65]	Butane	Tube: $l = 990$, $d = 7.75$; vertical	Stainless steel	1.45	118–708	10.1–79.5	0	24
48	Guerrieri and Talty [66]	Pentane	Tube: $l = 1981$, $d = 25.4$; vertical	Brass	1.0	165–409	17.7–38.2	0	4
49	Guerrieri and Talty [66]	Heptane	Tube: $l = 1981$, $d = 25.4$; vertical	Brass	1.0	184–448	18.5–30	0	4
50	Yusuphova *et al.* [67]	Isooctane	Tube: $l = 700$, $d = 7.2$; vertical	Stainless steel	2.5–20.5	684	55–490	0	23
51	Guerrieri and Talty [66]	Benzene	Tube: $l = 1981$, $d = 25.4$; vertical	Brass	1.0	247–602	27.1–42	0	4
52	Guerrieri and Talty [66]	Cyclohexane	Tube: $l = 1981$, $d = 25.4$; vertical	Brass	1.0	392–484	7.89–30.9	0	21

Table 1 (*continued*)

Ordinal No.	Authors	Fluid	Test section: dimensions (mm); orientation	Material	Pressure (bar)	Mass flow rate (kg m^{-2} s^{-1})	Heat flux density (kW m^{-2})	Number of experimental points: Vaporization $N_{CB} > 1.6 \times 10^4$	Nucleate boiling $N_{CB} < 1.6 \times 10^4$	Symbols
1	2	3	4	5	6	7	8	9	10	11
53	Guerrieri and Talty [66]	Methanol	Tube: $l = 1981$, $d = 25.4$; vertical	Brass	1.0	225–571	18.7–54.9	0	4	
54	Lukomsky and Madorskaya [47]	Ethanol	Tube: $l = 210$, $d = 30$; vertical	Chromo-molybdenum steel	1.57–30.2	112–474	31.6–256	0	15	
	Total organic fluids							0	115	
55	McAdams *et al.* [68]	Water	Annular channel: $l = 95.3$, $d_e = 47.8$; vertical	Stainless steel	2.07–6.21	1100–3440	237–726	0	8	
56	Sterman [69]	Water	Annular channel: $l = 90$, $d_e = 3.8$; vertical	Stainless steel	6.86	4520–6020	814	0	64	
57	Dengler and Addoms [70]	Water	Tube: $l = 6096$, $d = 25.4$; vertical	Copper	2.2	724	99.9–463	5	0	
58	Aladiyev *et al.* [71]	Water	Tubes: $l = 62.5$ and 135, $d = 8.2$; vertical	Stainless steel	5.88–138	312–6240	523–4650	0	56	
59	Tarasova *et al.* [72, 73]	Water	Tube: $l = 350$, $d = 8$ Annular channel: $l = 376$, $d_e = 4.61$ and 18.3; vertical	Nickel	128–196	1250–3640	502–1230	0	57	
60	Averin and Kruzhilin [74]	Water	Annular channel: $l = 60$, $d_e = 74.7$; vertical	Stainless steel	0.98–8.83	792–5270	815–3610	0	396	
61	Isbin *et al.* [75]	Water	Tube: $l = 235$, $d = 9.55$; horizontal	Brass	1.5–5.7	248	80.8–142	26	0	
62	Bennet *et al.* [76]	Water	Annular channel: $l = 737$, $d_e = 447$ and 6.12; vertical	Stainless steel	1.5–1.8	152–290	300–400	16	1	
63	Wright [77]	Water	Tubes: $l = 1430$ and 1728, $d = 12$ and 18.3; vertical	Stainless steel	1.41–4.6	667–3054	117–278	52	24	

64	Alexeev *et al.* [78]	Water	Tube: $l = 700$, $d = 9$; vertical	Nickel	29.4–147	250–2000	98.8–692	0	54
65	Morozov [79]	Water	Tube: $l = 250$, $d = 32$; vertical	Stainless steel	30.4–40	116–515	112–515	0	100
66	Collier *et al.* [80]	Water	Annular channel: $l = 1219$, $d_e = 6.12$; vertical	Stainless steel	1.6	136–258	199–401	10	7
67	Rounthwaite [81]	Water	Tube: $l = 6248$; $d = 41.3$; horizontal	Mild steel	14.8	235	63.1	14	0
68	Bertoletti *et al.* [8]	Water	Tube: $l = 1399$, $d = 9.8$; vertical	Stainless steel	70	1100–3800	200–1980	0	39
69	Thom *et al.* [82]	Water	Tube: $l = 127$, $d = 12.7$ Annular channel: $l = 305$, $d_e = 11.6$; vertical	Stainless steel	52.8–138	1040–3800	284–1670	0	23
70	Müller [83]	Water	Tube: $l = 240$, $d = 13.1$; vertical	Stainless steel	1.52–181	609–5700	175–8000	0	168
71	Hodgson [84]	Water	Tube: $l = 152$, $d = 11.7$; vertical	Stainless steel	13.8–34.5	1780	1913	0	21
72	Andreyevsky *et al.* [85, 86]	Water	Tubes: $l = 800$, $d = 8.12$ and 18; vertical	Stainless steel	4.1–30.4	183–2690	352–1490	75	58
73	Herkenrath and Mörk-Mörkenstein [87]	Water	Tube: $l = 5100$, $d = 10$; vertical	Stainless steel	205	1500	600	0	3
74	Cheng and co-workers [9, 88]	Water	Tubes: $l = 50.8$ and 108, $d = 12.7$; vertical	Copper Inconel	1.01	136	470–1400	0	8
75	Kenning and co-workers [89, 90]	Water	Tube: $l = 500$, $d = 9.6$; vertical	Copper–nickel alloy	1.0–3.9	105–304	50–400	37	7
	Total water							235	1094
	Grand total							715	2500
	Total number of experimental points							3215	

$\alpha(q)$ than in pool boiling. An assumption that the expression $\alpha \sim \mu_l^{-0.33}$ [97] describes the influence of viscosity on the heat transfer coefficient yields $n_3 = -0.33$. The exponent at dimensionless pressure K_p was found on the basis of all the data given in Table 1. It appeared to be the same for all the groups of fluids and equal to $n_2 = 0.54$. Hence, the final correlation for calculating the heat transfer for the forced flow nucleate boiling takes the form

$$Nu_b = C\,Pe_*^{0.6}\,K_p^{0.54}\,Pr_l^{-0.33}\,K_\lambda^{0.12}. \qquad (6)$$

The constant C in equation (6) turns out to be different for different groups of fluids and can be found from Table 2 given below.

The fact that dimensionless equation (6) for the heat transfer coefficient contains an individual constant for each group of fluids may seem at first glance illogical. However, it is explained by a number of objective reasons, the most important of which are the following.

(1) The system of dimensionless parameters (4) is not complete because it lacks variables characterizing the channel inner wall roughness (R_a/b) [2], wettability (θ), individual properties of substance (A or some other parameter of thermodynamic similarity).

(2) The influence of the above listed parameters is rather weak compared with the influence of, say, q or p, and which is of no less importance, is rather uncertain. Moreover, there has always been and will continue to exist the shortage of information about the quantities R_a and θ.

(3) Nobody has ever experimentally investigated the dependence of the heat transfer coefficient on viscosity in nucleate boiling under the conditions of forced flow. That is why the exponent of a power $n_3 = -0.33$ is likely to be the least substantiated one of all those entering into expression (6). It is of interest that a certain correlation between the value of the constant in equation (6) and Prandtl numbers for separate groups of fluids (see Table 2) is observed. However, it is hardly probable that at the moment there is sufficient information for drawing a responsible conclusion.

Taking into account the above, the appearance of an individual constant in equation (6) seems to be quite natural now.

Finally, it should be noted that the attempts to correlate the data on pool boiling, that made claims to a sufficiently high accuracy, have always led to the appearance of individual constants or even equations not only for separate groups of fluids, but also for individual combinations of liquid-heating surface [98–100]. Equation (6) predicts a markedly weaker dependence of heat transfer coefficient on heat flux and pressure in nucleate boiling under the conditions of forced flow compared with boiling in a pool. It is most clearly seen if written side by side with one of the most well-known equations for boiling in a pool, i.e. Kutateladze's equation [101]

$$Nu_b = 1.5\times10^{-3}\,Pe_*^{2/3}\,K_p^{2/3}(1-\rho_v/\rho_l)^{2/3}. \qquad (7)$$

The reasons for such a discrepancy were discussed in detail earlier [4].

Completing the consideration of the correlation obtained, it is worth adding that, to simplify the calculations of nucleate boiling heat transfer rate for water, one can use the following equation:

$$\alpha = \frac{9.6p^{0.14}}{1-4\times10^{-3}p}\,q^{0.6} \qquad (8)$$

where α, q and p are measured in W m^{-2} K^{-1}, W m^{-2} and bar, respectively. The coefficient 9.6 corresponds to boiling in a channel with metallic walls, with their thermal conductivity being $\lambda_w = 14$–20 W m^{-1} K^{-1} (stainless steel, inconel, titanium and zirconium alloys). Equation (8) approximates correlation (6) in the range of $p = 1$–200 bar within 11%. According to equation (6), for water boiling in a nickel tube, the coefficient in equation (8) increases up to 11.2 and in a copper unoxidized tube increases up to 13.6.

In Fig. 2 calculations by equation (6) for water are compared with calculations by Rassokhin *et al.*'s equation [102]

$$\alpha = \begin{cases} 3.29q^{2/3}p^{1/4} & \text{for } 1 < p < 80 \text{ bar} \\ 0.0288q^{2/3}p^{4/3} & \text{for } 80 < p < 200 \text{ bar} \end{cases} \qquad (9)$$

which was recommended [103] as the best equation for representing the experimental data. As is seen from Fig. 2, a good agreement of calculations made by equation (6) and by equations (9) is observed: the deviation never exceeds 40% and in the majority of cases it lies within 20%. This fact is evidence of the correctness of the analysis undertaken above.

Table 2

Fluids	C	Pr
Freons	7.6×10^{-3}	3.8 ± 2.5
Organic fluids	6.8×10^{-3}	3.5 ± 1.3
Cryogenic fluids	6.1×10^{-3}	1.6 ± 1.0
Water	4.9×10^{-3}	1.2 ± 0.4

CALCULATION RECOMMENDATIONS AND COMPARISON WITH EXPERIMENTAL DATA

To calculate heat transfer to a two-phase flow in vertical and horizontal channels in the case when the perimeter is fully wetted by a liquid, the correlation given below can be used:

$$Nu_{TP} = \max\{Nu_b;\,Nu_c\} \qquad (10)$$

where Nu_b and Nu_c are found from equations (6) and (3), respectively.

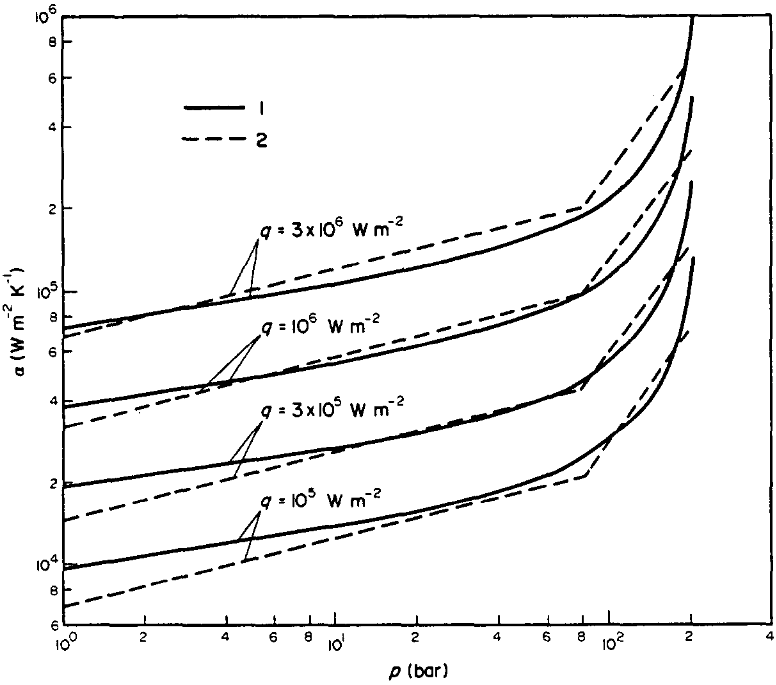

FIG. 2. Comparison of the water nucleate boiling heat transfer rate under conditions of forced convection calculated by different correlations: 1, equation (6) (tube material, stainless steel); 2, equation (9).

The position of the boundary between the regions of the predominant influence of nucleate boiling and two-phase forced convection is defined by the condition $N_{CB} = 1.6\times10^4$ with an accuracy within $\pm25\%$. The latter means that in the case when N_{CB} numbers are beyond the range of 1.2×10^4–2.0×10^4, the need for choosing the maximum heat transfer coefficient out of the two coefficients is avoided, and the calculation can be made directly either by expression (6) (with $N_{CB} < 1.2\times10^4$), or by expression (3) (with $N_{CB} > 2.0\times10^4$). For water nucleate boiling equation (6) can be replaced by dimensional formula (8) which approximates it. Equations (3) and (6) show that in two-phase flow the heat transfer is independent of the channel diameter. This statement is valid if the condition $(d/b) > 1.5$ is satisfied [1].

Tables 3–5 list the results of comparison of all the experimental data from Table 1 with correlation (10). Altogether 3215 experimental points on two-phase flow heat transfer were analysed (2500 for the nucleate boiling region and 715 for the vaporization region). Out of this number, only 96 points, i.e. 3%, deviate from predictive correlation (10) by more than 35% (82 points for nucleate boiling and only 14 points for vaporization). The average absolute deviation given by the relation

$$\bar{D} = \frac{1}{N}\sum_{i=1}^{N}\left(\frac{|\alpha_{exp}-\alpha_{calc}|}{\alpha_{calc}}\right)_i \tag{11}$$

Table 3. Comparison of experimental data with prediction. Nucleate boiling

Fluid	Parameter: Number of fluids investigated	Total number of experimental points	Number of points outside the $\pm35\%$ range (percentage of the total number)	Number of points outside the 73% to -42% range (percentage of the total number)	Mean absolute deviation, $\bar{D}$ (%)	Maximum deviation $\frac{\alpha_{exp}-\alpha_{calc}}{\alpha_{calc}}$ (%)
Cryogens	5	1023	2.1	0.1	13.6	68 and −43
Freons	6	268	6.0	1.1	19.3	51 and −56
Organic fluids	9	115	1.7	0.0	13.5	48 and −35
Water	1	1094	3.9	0.7	14.3	64 and −65
All fluids	21	2500	3.3	0.5	14.5	68 and −65

Table 4. Comparison of experimental data with prediction. Forced convection vaporization

Fluid	Parameter: Number of fluids investigated	Total number of experimental points	Number of points outside the $\pm 35\%$ range (percentage of the total number)	Number of points outside the 73% to -42% range (percentage of the total number)	Mean absolute deviation, $\bar{D}$ (%)	Maximum deviation $\frac{\alpha_{exp}-\alpha_{calc}}{\alpha_{calc}}$ (%)
Cryogens	4	237	0.8	0.0	13.8	39 and -35
Freons	5	243	2.9	0.0	14.5	59 and -34
Organic fluids	0	0	0.0	0.0	0.0	0
Water	1	235	2.1	0.0	13.3	46 and -25
All fluids	10	715	2.0	0.0	13.9	59 and -35

Table 5. Comparison of experimental data with prediction. Two-phase flow heat transfer

Fluid	Parameter: Number of fluids investigated	Total number of experimental points	Number of points outside the $\pm 35\%$ range (percentage of the total number)	Number of points outside the 73% to -42% range (percentage of the total number)	Mean absolute deviation, $\bar{D}$ (%)	Maximum deviation $\frac{\alpha_{exp}-\alpha_{calc}}{\alpha_{calc}}$ (%)
Cryogens	5	1260	1.8	0.1	13.6	68 and -43
Freons	6	511	4.1	0.6	17.0	59 and -56
Organic fluids	9	115	1.7	0.0	13.5	48 and -35
Water	1	1329	3.6	0.6	14.1	64 and -65
All fluids	21	3215	3.0	0.4	14.4	68 and -65

in which N is the number of experimental points, amounted to 14.4% which is much better than for any other correlations (see for instance, Table 2 from ref. [5]).

Tables 3–5 include, besides three rather often used deviation parameters, one more parameter, i.e. a number of points beyond the interval 73% to -42%. Its introduction is based on the following arguments. It is postulated that in the calculation of such a complex process as the two-phase flow heat transfer, the scale of 'quality' of the predictive correlation might look as follows:

$\bar{D} < 15\%$—a very good one (Chen's relationship, for instance, agrees with the data from refs. [66, 70, 76, 104, 105], on the basis of which it was derived);

$\bar{D} = 15$–20%—a good one (Bjorge *et al.*'s relationship [106] compared with the data on water from eight references);

$\bar{D} = 20$–25%—acceptable (Chen's relationship compared with available data on forced convection vaporization, see ref. [5]);

$\bar{D} > 25\%$—unacceptable (any correlation, known to the present author, compared with the data from Table 1).

The mean deviation of 25% implies that approximately 95% of the points from the described bulk of the experimental data lie within the range of $\pm 50\%$ of the calculated one, i.e. the ratio of the maximum and minimum heat transfer coefficients lying in this interval constitutes $\alpha_{max}/\alpha_{min} = 1.5/0.5 = 3.0$. When dealing with rather large deviations it would be more correct, in the author's opinion, to operate not with percentage of deviation, but rather with the value showing by how many times the experiment deviates from the calculation. Otherwise, the deviation equal, say, to 80% can be considered to be 'worse' than that equal to -50%, in spite of the fact that in the second case the disagreement is two-fold, while in the first case it is only 1.8-fold. Thus, to large deviations there should correspond an 'asymmetric' error band. If one assumes that $\alpha_{max}/\alpha_{min} = 3.0(\bar{D} \approx 25\%)$, then in the range 73% to -42% one will find data deviating from the calculation by not more than $\sqrt{3}$ times. Let us assume that the data beyond this range are not represented by that correlation. For the correlation considered the number of such data is very small. They include:

(a) data on water boiling at $p = 205$ bar [87], as well as some points from refs. [68, 88];

(b) data on ammonia boiling at $p = 108$ bar [63].

It is not difficult to conclude that basically the proposed correlation overpredicts the data which are obtained at high reduced pressure ($p/p_{cr} > 0.92$). It should be noted that the error of the experiments under such pressures is rather high ($\Delta T < 1$ K) and the information about the thermal properties of the fluid is rather inaccurate. Additional and precise experiments are required to specify with certainty the character of the function $\alpha(p)$ in the vicinity of the critical point. At the Moscow Power Engineering Institute a special programme is being carried out aimed at the research of forced flow nucleate boiling under high reduced pressures.

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UNE FORMULE GENERALE POUR LE TRANSFERT THERMIQUE EN CONVECTION FORCEE DIPHASIQUE—SECONDE DETERMINATION

Résumé—Une formule générale pour la convection forcée thermique diphasique (ébullition nucléée et vaporisation) est suggérée pour des canaux verticaux et horizontaux avec un périmètre complètement mouillé. Elle représente les données expérimentales, avec une déviation moyenne égale à 14,4% pour 21 liquides différents (eau, liquides organiques, Fréons et cryogènes) dans les domaines suivants des paramètres principaux: pression 0,61–196 bar, densité de flux thermique $10–8 \times 10^6$ W m^{-2}, débit masse 5,6–6240 kg m^{-2} s^{-1}, diamètre du canal 0,47–74,7 mm. Le flux de chaleur transféré par l'ébullition nucléée comme par vaporisation dépend de la conductivité thermique de la paroi du canal. On montre que le transfert de chaleur par ébullition nucléée est décrit par une seule équation en variables dimensionnelles contenant une constante individuelle pour chacun des quatre groupes de fluides (eau, liquides organiques, Fréons et cryogènes). Le transfert de chaleur par vaporisation est décrit par une seule équation avec une constante universelle pour tous les fluides.

EINE VERALLGEMEINERTE KORRELATION FÜR DEN WÄRMEÜBERGANG BEI ERZWUNGENER ZWEIPHASENSTRÖMUNG—ZWEITE FORMULIERUNG

Zusammenfassung—Es wird eine verallgemeinerte Korrelation für den Wärmeübergang bei erzwungener Zweiphasenströmung (Blasensieden und Verdampfung) vorgeschlagen, die in vertikalen und horizontalen Kanälen bei vollständig benetztem Umfang gültig ist. Sie gibt die Meßergebnisse für 21 verschiedene Flüssigkeiten (Wasser, organische Flüssigkeiten, Kältemittel und kryogene Flüssigkeiten) mit einer mittleren absoluten Abweichung von 14,4% und folgendem Bereich der Hauptparameter wieder: Druck, 0,61–196 bar; Wärmestromdichte, $10–8 \times 10^6$ W m^{-2}; Massenstromdichte 5,6–6240 kg m^{-2} s^{-1}; Kanaldurchmesser, 0,47–74,7 mm. Es zeigt sich, daß der Wärmeübergang sowohl beim Blasensieden als auch bei der Verdampfung von der Wärmeleitfähigkeit des Wandmaterials abhängt. Der Wärmeübergang beim Blasensieden wird mit einer einzigen Gleichung mit einer individuellen Konstanten für jede der vier Flüssigkeitsgruppen dargestellt. Der Wärmeübergang bei der Verdampfung läßt sich mit einer einzigen Gleichung mit einer universellen Konstanten für alle Fluide darstellen.

ОБОБЩЕННОЕ СООТНОШЕНИЕ ДЛЯ РАСЧЕТА ТЕПЛООТДАЧИ К ДВУХФАЗНОМУ ПОТОКУ ПРИ ВЫНУЖДЕННОЙ КОНВЕКЦИИ—ПОВТОРНАЯ ОЦЕНКА

Аннотация—Получено обобщенное соотношение для расчета теплоотдачи при вынужденном движении двухфазного потока (пузырьковом кипении и испарении), справедливое для вертикальных и горизонтальных каналов с полностью смоченным периметром. Соотношение со средним абсолютным отклонением в 14,4% описывает экспериментальные данные для двадцати одной различной жидкости (вода, органические жидкости, фреоны, криогены) в следующем диапазоне основных параметров: давление 0,61–196 бар, плотность теплового потока $10–8 \times 10^6$ Вт $м^{-2}$, массовая скорость 5,6–6240 кг $м^{-2}$ $с^{-1}$, диаметр канала 0,47–74,7 мм. Установлено, что интенсивность теплоотдачи как при пузырьковом кипении, так и при испарении зависит от теплопроводности материала стенки канала. Показано, что теплоотдача при пузырьковом кипении описывается единым уравнением в безразмерных переменных, содержащим индивидуальную для каждой из четырех групп жидкостей (вода, органические жидкости, фреоны, криогены) константу. Теплоотдача при испарении описывается единым уравнением с константой, универсальной для всех жидкостей.