archives
of thermodynamics
Vol. **37**(2016), No. 2, 89–106
DOI: 10.1515/aoter-2016-0014

Prediction of flow boiling heat transfer coefficient for carbon dioxide in minichannels and conventional channels

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Abstract In the paper presented are the results of calculations using authors own model to predict heat transfer coefficient during flow boiling of carbon dioxide. The experimental data from various researches were collected. Calculations were conducted for a full range of quality variation and a wide range of mass velocity. The aim of the study was to test the sensitivity of the in-house model. The results show the importance of taking into account the surface tension as the parameter exhibiting its importance in case of the flow in minichannels as well as the influence of reduced pressure. The calculations were accomplished to test the sensitivity of the heat transfer model with respect to selection of the appropriate two-phase flow multiplier, which is one of the elements of the heat transfer model. For that purpose correlations due to Müller-Steinhagen and Heck as well as the one due to Friedel were considered. Obtained results show a good consistency with experimental results, however the selection of two-phase flow multiplier does not significantly influence the consistency of calculations.

Keywords: Heat transfer coefficient; Flow boiling; Carbon dioxide; Minichannels

Nomenclature

B	–	blowing parameter
Bo	–	boiling number, q/Gh_{LG}

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C	–	constant
Con	–	confinement number, $(\sigma/[d(\rho_L - \rho_G)]^{0.5}/d$
c_p	–	specific heat, J/kgK
d	–	tube diameter, m
f	–	friction factor
f_1, f_{1z}	–	function
Fr	–	Froude number, $G^2/(\rho_L^2 gd)$
G	–	mass velocity, kg/m^{2}s
h	–	heat transfer coefficient, W/m^{2}K
h_{LG}	–	latent heat of vaporization, J/kg
M	–	molecular weight, kg/kmol
Nu	–	Nusselt number, $\alpha d/\lambda$
P	–	empirical correction
p	–	pressure, Pa
Pr	–	Prandtl number, $\mu_L c_p/\lambda_L$
q	–	heat flux, W/m^2
R	–	two-phase multiplier
Re	–	Reynolds number, Gd/μ_L
s	–	slip ratio
T	–	temperature, K
We	–	Weber number, $G^2d/(\sigma\rho_L)$
x	–	quality

Greek symbols

α	–	heat transfer coefficient, W/m^{2}K
σ	–	surface tension
λ	–	thermal conductivity, W/mK
ρ	–	density, kg/m^3
μ	–	dynamic viscosity, Pa s
$\xi = \frac{f_r}{4}$	–	friction factor

Subscripts

0	–	reference case
cr	–	critical
exp	–	experimental
F	–	Friedel correlation
G	–	vapor
L	–	liquid
LO	–	total liquid flow rate
MS	–	Müller-Steinhagen and Heck correlation
Pb	–	pool boiling
r	–	reduced
sat	–	saturation
TPB	–	two-phase flow boiling
th	–	theoretical

1 Introduction

A widely used group of synthetic compounds in refrigeration technology is to be withdrawn from technical applications under the Montreal Protocol [51]. It is widely acknowledged that such compounds contribute to the reduction of ozone layer in the upper atmosphere. Natural refrigerants, such as hydrocarbons or carbon dioxide are likely to fully replace them in the very near future. That strikes interest to fully understand heat transfer performance of these fluids.

Carbon dioxide as compared to the contemporary used fluids is a relatively safe one. The fluid is nontoxic, nonflammable, nonexplosive and can be coupled with most metals and plastics [1]. At the moment its applications can be found mainly in small refrigeration, food industry, and air-conditioning units. Design of evaporators for use of the carbon dioxide requires the exact determination of heat transfer coefficient during flow boiling as well as flow resistance. Carbon dioxide, as compared to other fluids at the same saturation temperature is characterized by higher vapor density, lower surface tension and lower dynamic viscosity of vapor. Available in the literature empirical correlations give different results as compared to the results obtained experimentally. There is hardly any robust and recommended correlation for the purpose of calculation of carbon dioxide two-phase heat transfer, despite some devoted contributions [19,41]. The literature contains a number of reports on experimental research for this fluid. The research regards heat transfer in channels with conventional diameters and minichannels. For a more extensive literature survey of flow boiling in conventional size channels the reader is referred to a review by Thome [2] or in small diameter channels to Bergles *et al.* [3] or Kandlikar [4]. There are several approaches to distinguish between minichannels and conventional size channels. Kandlikar's [4] systematization of channel sizes with respect to the diameter reads:

- Conventional channels – hydraulic diameters greater than 3 mm.
- Minichannels – hydraulic diameters to range of 600 μm – 3 mm.
- Microchannels – hydraulic diameters to range of 50 μm – 600 μm.

In a general opinion the physical mechanism should be employed to distinguish the transition threshold between minichannels and conventional size channels, Thome [2]. A criterion based on the Laplace constant which allows to distinguish between conventional channels and minichannels was proposed by Kew and Cornwell [5]. This criterion is based on the so called confinement number Con, defined as

$$\text{Con} = \frac{1}{d}\sqrt{\frac{\sigma}{d(\rho_l - \rho_v)}} \ . \tag{1}$$

It has been postulated that when the confinement number Con is greater than 0.5 then the flow exhibits the properties of the flow in minichannels, in which the surface tension plays an important role.

Zhao and Bansal [6] conducted experimental study for flow boiling of carbon dioxide in tubes with internal diameter of 4.57 mm at very low saturation temperatures equal to $-30\,^{\circ}$C. The results of that experimental research were compared with the empirical correlations due to Cooper [7], Gungor and Winterton [8], Jung *et al.* [9], Kandlikar [10], Liu and Winterton [11], Kattan *et al.* [12] and Yoon *et al.* [13]. It was found that none of the mentioned above empirical methods were able to predict the boiling heat transfer coefficient of carbon dioxide to a satisfactory extent in relation to the experimental data.

Mastrullo *et al.* [14,15] compared the results of their experimental research with some of the established correlations for conventional channels. Their studies were carried out for flow boiling of carbon dioxide in the channels with internal diameter of 6 mm. Authors obtained 217 points for the mass velocity ranging from 200 to 349 kg/m^{2}s, heat flux ranging from 10 to 20.6 kW/m^2 and saturation temperature from -7.8 to 5.8 °C. These experimental data were compared with correlations due to Shah [16], Gungor and Winterton [8], Jung *et al.* [9], Steiner and Taborek [17], Panek [18], Yoon *et al.* [13] and Cheng *et al.* [19]. Presented results showed that the good agreement with experimental data was attained in the case of Jung *et al.* correlation. In that case the mean absolute deviation of data was 21.6% [14,20].

Experimental studies of flow boiling heat transfer of carbon dioxide, ammonia and propane in a single tube with internal diameters of 1.5 mm and 3 mm, were carried out by Pamitran *et al.* [21]. The mass velocity ranged from 50 to 600 kg/m^{2}s, heat flux from 5 to 70 kW/m^2 and saturation temperature from 0 °C to 10 °C. The results of experimental research carried out for carbon dioxide have been compared with six popular correlations due to Shah [16], Gungor and Winterton [8], Jung *et al.* [9], Wattelet *et al.* [22], Tran *et al.* [23] and Kandlikar and Steinke [24]. The results obtained from the comparison of experimental data and theoretical research shown that the good agreement was obtained with experimental data only in the case of Gungor and Winterton correlation. In the case of that correlation the smallest mean absolute deviation of 21.6% was obtained [20,21].

Comparison between experimental and theoretical studies has also been carried out by Docoulombier *et al.* [25]. Accomplished was research of flow boiling of carbon dioxide in tubes with the internal diameter of 0.529 mm. Study was conducted for three saturation temperatures, namely -10, -5, and 0 °C. Studied were also three levels of heat flux, namely 10, 20, and 30 kW/m^2. Mass velocity was varied from 200 to 1200 kg/m^{2}s. The results were compared with empirical correlations due to Chen [26], Shah [16], Gungor and Winterton [8], Jung *et al.* [8], Kandlikar [10], Liu and Winterton [11], Wattelet *et al.* [22], Satioh *et al.* [27], Cheng *et al.* [19], Wang *et al.* [28] and finally Hihara and Tanaka [29]. The results showed that the best agreement with experimental data was obtained for the case of Hihara and Tanaka correlation. In that case the mean absolute deviation of 17,9% was found [20,25].

In literature there are many empirical correlations for modeling of boiling heat transfer. Some of them have been mentioned above. However, in the case of a fluid such as carbon dioxide they did not prove a good consistency with experimental data. Several publications, which recently appeared, for example due to Ribatski [30], Tibirica and Ribatski [31], Sardeshpande and Ranade [32] or Alagesan [33] analyze the experimental data for validation of heat transfer coefficient predictions using the correlations available in literature. It was authors intention to show the performance of their approach in predicting flow boiling of carbon dioxide, a fluid which usually turns out to be a severe test for heat transfer and pressure drop predictions. In the paper the results of the collected from literature experimental evidence were compared with the predictions of the model [34–37]. Based on the evidence of comparisons with mentioned above experimental data a correction incorporating the effect of reduced pressure has been applied to the authors own model to provide feasibly the best consistently of the predictions with the experimental data.

In the paper considered are data due to Docoulombier *et al.* [25], Pamitran *et al.* [21], Mastrullo *et al.* [14,15], Yun *et al.* [38,39], Choi *et al.* [40], Yoon *et al.* [41], Oh *et al.* [42], Oh *et al.* [43,44], Dang *et al.* [45], Kim *et al.* [46], Wu *et al.* [47], Cho *et al.* [48] and Zhao *et al.* [6]. The range of parameters analyzed in the experimental research is shown in Tab. 1.

2 The model

The versatile semiempirical model for calculations of flow boiling and flow condensation due to J. Mikielewicz [34] and the final version due to D. Mikielewicz *et al.* [35–37] has been tested for a significant number of experimental data and has returned satisfactory results for the case of the flow boiling process for numerous fluids. The fundamental hypothesis of the model is the fact that heat transfer during flow boiling with bubble generation can be modeled as a sum of two contributions constituting the total energy dissipation in the flow, namely the energy dissipation due to shearing flow without the bubbles and dissipation resulting from the bubble generation. The final version of the model [37] reads:

$$\frac{\alpha_{TBP}}{\alpha_{LO}} = \sqrt{R_{MS}^n + \frac{C}{1+P}\left(\frac{\alpha_{Pb}}{\alpha_{LO}}\right)^2}\ . \tag{2}$$

In Eq. (2) α_{LO} is the heat transfer coefficient for the liquid only case. For turbulent flow it may be determined using for example the Dittus-Boelter equation, (for turbulent flow) or in case of laminar flow, Nu = 3.66. In the model given by Eq. (2) was introduced the empirical correction P and a modified two-phase multiplier due to Müller-Steinhagen and Heck [35], R_{MS}. The modified form of the two-phase multiplier is

$$R_{MS} = \left[1 + 2\left(\frac{1}{f_1} - 1\right) x\,\mathrm{Con}^m\right](1-x)^{\frac{1}{3}} + x^3\,\frac{1}{f_{1z}}\ . \tag{3}$$

It should be noted that the two-phase multiplier R_{MS} present in Eq. (2) is raised to the power n, where $n = 0.76$ for turbulent flows and $n = 2$ for laminar flows. Functions f_1 and f_{1z} in Eq. (3) are denoted as the ratio of the pressure drop in flow of liquid to flow of gas and heat transfer coefficient in vapour heat transfer coefficient for that of liquid, respectively. For the case of turbulent flow these functions can be determined from the following relations: $f_1 = (\rho_L/\rho_G)(\mu_L/\mu_G)^{0.25}$, $f_{1z} = (\mu_G/\mu_L)(\lambda_L/\lambda_G)^{1.5}(c_{pL}/c_{pG})$. In the case of laminar flow $f_1 = (\rho_L/\rho_G)(\mu_L/\mu_G)$ and $f_{1z} = (\lambda_G/\lambda_L)$. Furthermore, the exponent m in Eq. (3) is equal $m = 0$ for flow in conventional channels or$m = -1$ for flow in minichannels. The form of empirical correction P in Eq. (2), should be calculated as

$$P = 2.53 \times 10^{-3}\,\mathrm{Re}^{1.17}\,\mathrm{Bo}^{0.6}\,(R_{MS} - 1)^{-0.65}\ . \tag{4}$$

The pool-boiling heat transfer coefficient, α_{Pb}, present in Eq. (2) can be calculated using a generalized model due to Cooper [7]. This model describes the heat transfer coefficient in the fluid in terms of the reduced pressure, molecular weight and applied wall heat flux. The Cooper equation which describes the pool-boiling heat transfer coefficient has the form

$$\alpha_{Pb} = A\, p_r^{0.12} \,(-\log p_r)^{-0.55}\, M^{-0.5}\, q^{\frac{2}{3}} \ . \tag{5}$$

It was expected that the accuracy of model predictions could be improved by some modifications to the empirical correction P. The modified empirical correction P yields

$$P = 2.53 \times 10^{-3}\, \mathrm{Re}^{1.17}\, \mathrm{Bo}^{0.6}\, (R^*_{MS} - 1)^{-0.65} \left(\frac{p_{sat}}{p_{cr}}\right)^a \ . \tag{6}$$

The two-phase flow multipier R^*_{MS} in Eq. (6) is calculated using the original version of Müller-Steinhagen and Heck correlation [49]. Exponent a was adjusted to the available data bank for carbon dioxide. Furthermore, in calculations tested was the sensitivity of the developed model to the selection of the two-phase flow multiplier. For that purpose two models were introduced into Eq. (2), namely the Müller-Steinhagen and Heck correlation [35,49] and the Friedel correlation [50]. Additionally present in the calculation procedure is the so called blowing parameter B. That parameter is responsible for evaluation of the nonadiabatic effects present due to modification of shear stress on liquid vapour interface [36,37] and is defined by Eq. (8).

The modified two-phase multiplier inclusive of nonadiabatic effects, denoted as R_B, and relationship which describes the modifications has the following form [37]:

$$R_B = \begin{cases} R\left(1 - \frac{B}{2}\right) & \text{for } 0.1 < x \le 1 \ , \\ R\sqrt{1 + \left(\frac{8\alpha_{Pb}\, d}{\lambda_L\, \mathrm{Re}\, \mathrm{Pr}\, \xi_0\, R_{MS}}\right)^2} & \text{for } 0 \le x \le 0.1 \ . \end{cases} \tag{7}$$

In Eq. (7) the two-phase multiplier should be calculated using any formulation, however the modified Müller-Steinhagen and Heck correlation is recommended for use in the case of refrigerants. The blowing parameter which occurs in Eq. (7) is defined as [51]

$$B = \frac{2\, q\, \frac{\rho_L}{\rho_G}}{f\, G(s-1)\, h_{LG}} \ . \tag{8}$$

In Eq. (8) s is the slip ratio, which can be determined from Zivi relationship [37]

$$s = \sqrt[3]{\frac{\rho_L}{\rho_G}} \ . \tag{9}$$

As a result of application of correction (7), a modified heat transfer model is obtained, which was adopted for calculations in the present work

$$\frac{\alpha_{TBP}}{\alpha_{LO}} = \sqrt{R_B^n + \frac{C}{1 + 2.53 \times 10^{-3}\,\mathrm{Re}^{1.17}\,\mathrm{Bo}^{0.6}\,(R_{MS}^* - 1)^{-0.65}\,p_r^a}\left(\frac{\alpha_{Pb}}{\alpha_{LO}}\right)^2} \ . \tag{10}$$

As mentioned earier, in the study another two-phase flow multiplier was also considered, namely the Friedel correlation [50]. According to this method the two-phase multiplier R_F can be determined in terms of Weber and Froude numbers as follows:

$$R_F = E + \frac{3.24\,FH}{\mathrm{Fr}^{0.045}\,\mathrm{We}^{0.035}} \ . \tag{11}$$

The terms E, F and H are determined by the following equations:

$$E = (1-x)^2 + x^2\left(\frac{\rho_L f_G}{\rho_G f_L}\right) \ , \tag{12}$$

$$F = x^{0.78}\,(1-x)^{0.2224} \ , \tag{13}$$

$$H = \left(\frac{\rho_L}{\rho_G}\right)^{0.91}\left(\frac{\mu_G}{\mu_L}\right)^{0.19}\left(1 - \frac{\mu_G}{\mu_L}\right)^{0.7} \ . \tag{14}$$

The heat transfer model utilizing the Friedel two-phase flow multiplier therefore reads:

$$\frac{\alpha_{TBP}}{\alpha_{LO}} = \sqrt{R_F^n + \frac{C}{1 + 2.53 \times 10^{-3}\,\mathrm{Re}^{.1.17}\,\mathrm{Bo}^{0.6}\,(R_{MS}^* - 1)^{-0.65}\,p_r^a}\left(\frac{\alpha_{Pb}}{\alpha_{LO}}\right)^2} \ . \tag{15}$$

3 The results

In the following part, the basic model and its subsequent modifications, which have been selected for discussion, will be analyzed with respect to predictions of heat transfer coefficient. These models are denoted respectively as: model I – Eq. (2), model II – Eq. (10) and model III – Eq. (15). Moreover, exponent a, present in the modified two-phase flow multiplier in Eq. (6) was adjusted to the available data bank for flow boiling of carbon dioxide.

Using the Kew and Cornwell [5] criterion, the available data bank was divided into conventional size channels and minichannels. Amongst collected data the criterion of minichannels, i.e., Con > 0.5, is fulfilled only by the research due to Docoulombier *et al.* [25], Wu *et al.* [47] and Yun *et al.* [38,39] for the case of data corresponding to d = 0.98 mm. It can therefore be concluded that in case of carbon dioxide the transition from conventional size channels to minichannels takes place at a channel diameter smaller than 1.5 mm. The value of the confinement number Con for carbon dioxide together with values of reduced pressure and the range of variation of experimental parameters are presented in Tab. 1. Analysis of the parameters from Tab. 1 indicates the fact that the collected for scrutiny experimental research covers a full range of quality variation and a relatively wide range of mass velocity.

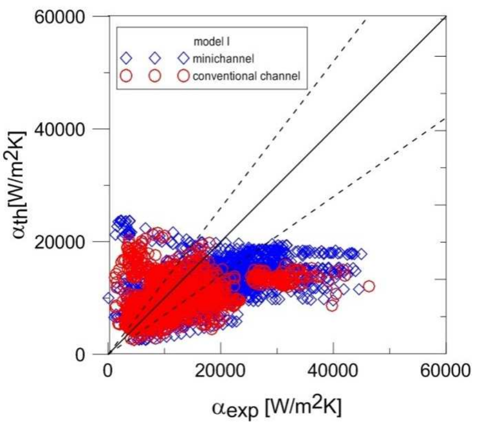

Figure 1: Comparison of test results, α_{exp} with predictions obtained using Eq. (2), α_{th}.

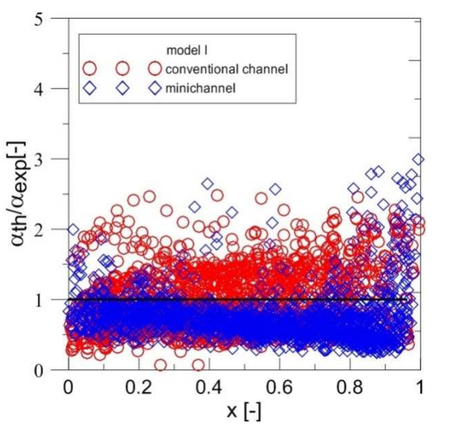

Figure 2: Comparison of the ratio of experimental values of α_{exp} to the ones obtained using Eq. (2), α_{th}, in function of quality.

Figures 1 to 6 show the results of calculations of heat transfer coefficient for carbon dioxide obtained using mentioned earlier methods based on Eqs. (2), (10), and (15). The version of the model applicable to minichan-

Table 1: The range of variation of experimental data for flow boiling of carbon dioxide and the confinement number Con, and reduce pressure.

The authors data	d [mm]	G [kg/m^{2}s]	q [kW/m^2]	T_{sat} [^{O}C]	Con	p_{sat}/p_{cr}
Docoulombier *et al*. [25]	0.529	600 – 1200	10 30	-10	**1.611**	0.359
				-5	**1.515**	0.413
				0	**1.412**	0.472
Pamitran *et al*. [21]	3	200 – 600 200 – 350	20 30 9 – 20	1	0.245	0.485
				2	0.241	0.498
				3	0.237	0.511
				10	0.207	0.610
	1.5			2	0.483	0.498
Mastrullo *et al*. [14,15]	6	170 – 340 1000, 1500	10 – 20 7 – 40	-7.8	0.138	0.382
				-3.2	0.130	0.433
				4.2	0.116	0.527
				5	0.115	0.538
				5.7	0.113	0.548
Yun *et al*. [38,39]	6	300 – 600	10 – 30 12 – 18 20 – 30	5	0.115	0.538
		318		10	0.104	0.610
	2	300		5	0.344	0.538
	0.98	400		10	**0.635**	0.610
Choi *et al*.[41]	1.5	400 – 900 200 – 500	20 – 40 10 – 30	10	0.415	0.610
Yoon *et al*. [41]	7.53	360, 720	4.5, 9, 18	0	0.099	0.472
				5	0.091	0.538
				10	0.083	0.610
Oh *et al*. [42]	3	212 – 424	15 – 40	10	0.207	0.610
				1	0.245	0.485
Oh *et al*. [43,44]	7.75	300 – 500 200 – 650	7.5, 14.9, 29.8 6 – 20	-5	0.104	0.413
				0	0.096	0.472
				5	0.089	0.538
				15	0.071	0.690
	4.57			5	0.145	0.538
				10	0.131	0.610
				15	0.115	0.690
				20	0.096	0.777
Dang *et al*. [45]	2	139 – 231	12.6 – 19.3	15	0.274	0.690
Kim *et al*. [46]	5	600 – 1200	10 30	-5	0.160	0.413
				0	0.149	0.472
				20	0.092	0.777
				5	0.136	0.538
Wu *et al*. [47]	1.42	200 – 600	20 30	0	**0.526**	0.472
				-10	**0.600**	0.359
				-20	**0.666**	0.267
				-30	**0.725**	0.194
				-35	**0.753**	0.163
				-40	**0.780**	0.136
Cho *et al*. [48]	5	200 – 350 170 – 340 1000, 1500	9 – 20 10 – 20 7 – 40	0	0.149	0.472
				5	0.138	0.538
				10	0.124	0.610
				20	0.092	0.777
	9.52			0	0.078	0.472
				5	0.072	0.538
				10	0.065	0.610
				20	0.048	0.777
Zhao *et al*. [6]	4.57	300 – 600	10 – 30	-30	0.223	0.194

nels was used if Con > 0.5 (importance of surface tension effects), and the version of the model applicable to conventional size channels was used for the case when Con < 0.5.

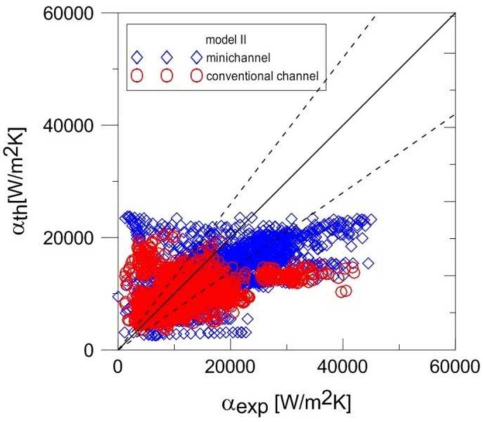

Figure 3: Comparison of test results, α_{exp} with predictions obtained using Eq. (10), α_{th}.

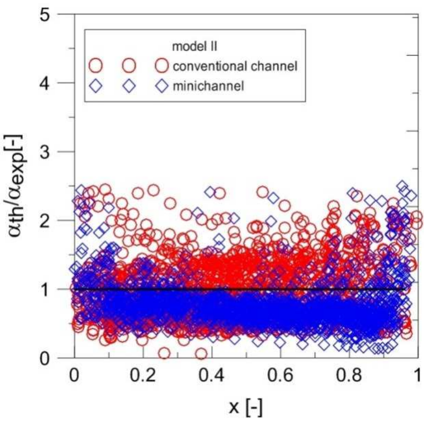

Figure 4: Comparison of the ratio of experimental values of α_{exp} to the ones obtained using Eq. (10), α_{th}, in function of quality.

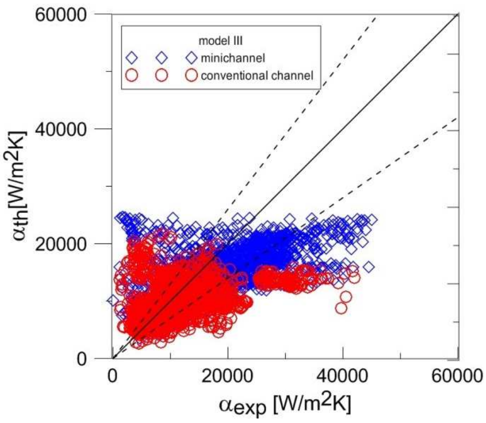

Figure 5: Comparison of test results, α_{exp} with predictions obtained using Eq. (15), α_{th}.

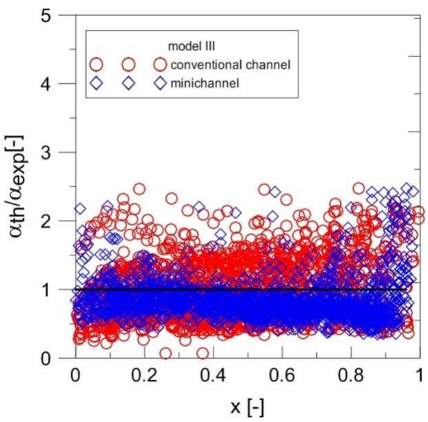

Figure 6: Comparison of the ratio of experimental values of α_{exp} to the ones obtained using Eq. (15), α_{th}, in function of quality.

The modification to the empirical correction described by Eq. (3) includes the effect of reduced pressure (p_{sat}/p_{cr}). The new version of the correction P is presented in Eq. (6), where the reduced pressure is raised to the power a. The value of the exponent a was adjusted using the regression analysis. Accomplished calculations indicate that best consistency is obtained if $-1.9 \leq a \leq -3$. For that reason a representative value of $a = -2$ has been selected. The results of calculations, which were obtained with the account of the reduced pressure are presented in Figs. 7 to 12, whereas the information about mean absolute deviation and correlation factors is shown in Tab. 2. Values of correlation factors are not very high, which indicates the dispersity of experimental data.

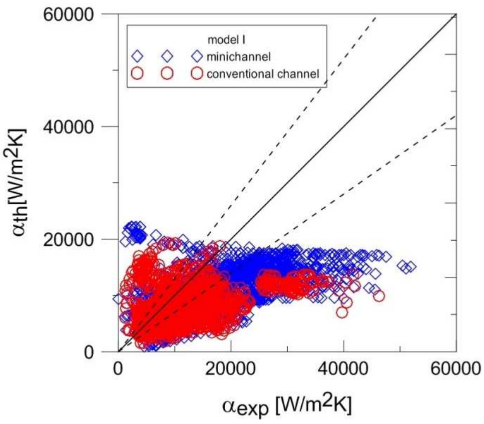

Figure 7: Comparison of test results, α_{exp} with predictions obtained using Eq. (15), α_{th}, and $a = -2$

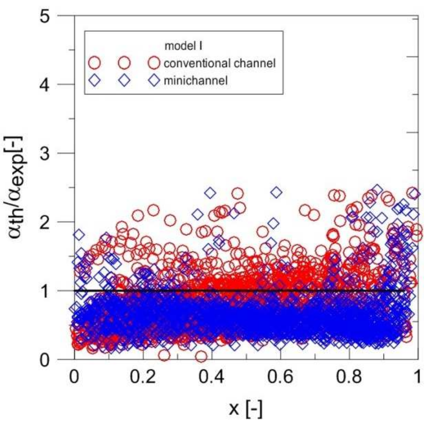

Figure 8: Comparison of the ratio of experimental values of α_{exp} to the ones obtained using Eq. (15), α_{th}, in function of quality; $a = -2$

Table 2: Values of exponent a correlation coefficient, R^2, and mean absolute deviation (MAD)

Figure	Model	Value of exponent a	R^2	MAD [%]
1	Model I	0	0.2325	60.99
2	Model II	0	0.3095	59.50
3	Model III	0	0.3227	59.98
4	Model I	-2	0.2729	61.09
5	Model II	-2	0.3277	59.29
6	Model III	-2	0.3381	58.79

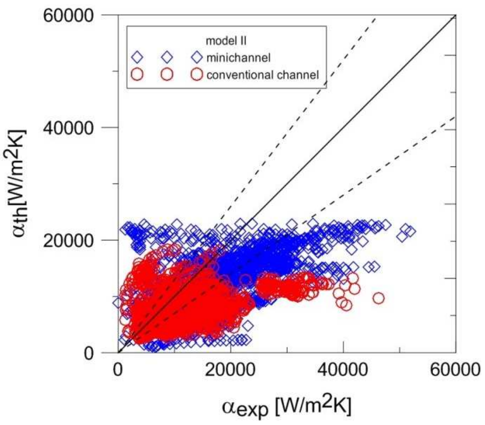

Figure 9: Comparison of test results, α_{exp} with predictions obtained using Eq. (15), α_{th} and $a = -2$.

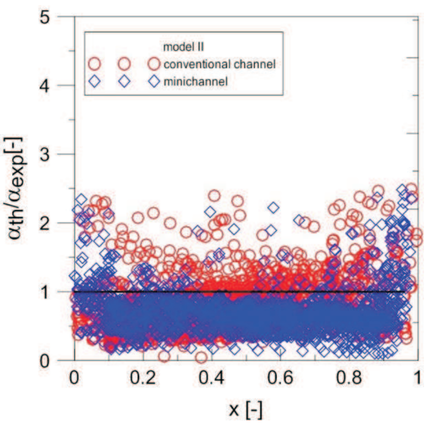

Figure 10: Comparison of the ratio of experimental values of α_{exp} to the ones obtained using Eq. (15), α_{th}, in function of quality; $a = -2$

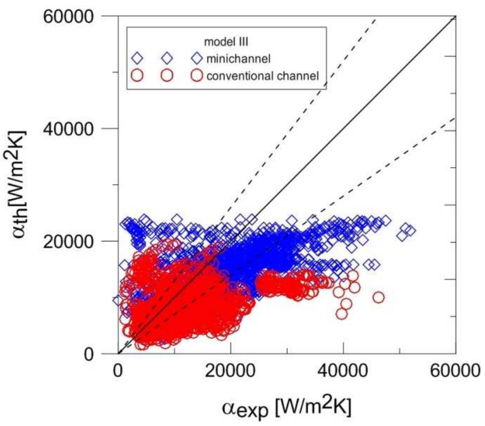

Figure 11: Comparison of test results, α_{exp} with predictions obtained using Eq. (15), α_{th}, and $a = -2$.

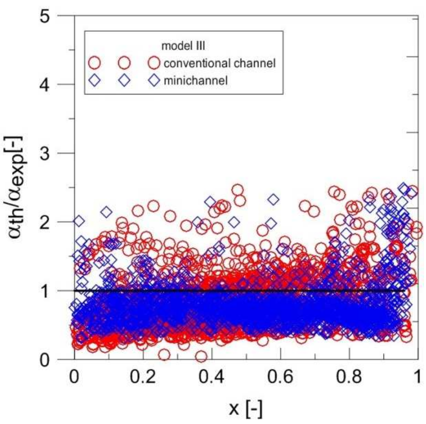

Figure 12: Comparison of the ratio of experimental values of α_{exp} to the ones obtained using Eq. (15), α_{th}, in function of quality; $a = -2$

Based on the analysis of presented comparisons it can be said that the greatest discrepancy between experimental and theoretical values is obtained in case of high values of quality. The discrepancy may be caused by the presence of dryout in experiments, which renders the reduction in measured heat transfer coefficient. It stems from Table 2 that in case of model II without the term considering the effect of reduced pressure obtained is the smallest mean absolute deviation (MAD) equal to 59.50%. At the same time in case of model III without the effect of reduced pressure obtained was the highest value of correlation coefficient R^2, which is equal in this case to 0.3227. In case, when the reduced pressure effect is considered, the best results were obtained for model III, where the Friedel correlation for prediction of the two-phase multiplier is used. In case of model III, where the reduced pressure effect is considered, obtained were the results where the mean absolute deviation is equal to 58.90% and the correlation coefficient is $R^2 = 0.3383$.

Based on the presented results of calculations, which were obtained using the versions of the heat transfer model described by Eqs. (2), (10) and (15), with and without the account of reduced pressure, $(p_{sat}/p_{cr})^a$, from Eq. (6), it can be concluded that the effect of appropriate selection of the two-phase multiplier also does not bear a significant influence on the results.

4 Conclusions

The paper presents the analysis of the results of calculations using a model developed earlier to study experimental data for flow boiling of carbon dioxide. The model was studied in several ways, i.e., it was used as the original one and also in a modified version where into the empirical correction P was included the reduced pressure effect (value of exponent a was modeled). The results show that the effect of reduced pressure does not significantly change the performance of the original model, however slightly improves the consistency of the results. The same conclusion can be drawn in case of selection of the model of two-phase flow multiplier, an inherent term in the heat transfer coefficient model. The results of calculations show better compliance with experimental data, in case of application of the Friedel correlation, however improvement over the Muller-Steinhagen and Heck formulation is not very significant. Potential improvements to the consistency of predictions will be expected if the effect of varying properties of carbon dioxide are introduced into the analysis.

Received 13 May 2016

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