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IMPROVED SEMI-EMPIRICAL METHOD FOR DETERMINATION OF HEAT TRANSFER COEFFICIENT IN FLOW BOILING IN CONVENTIONAL AND SMALL DIAMETER TUBES

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ABSTRACT

In the paper presented have been considerations regarding a model of boiling heat transfer in conventional and small diameter channels. Modified has been a model postulated earlier by Mikielewicz (1973) to incorporate the surface tension effects into the two-phase flow multiplier as well as the empirical correction influencing the nucleate boiling heat transfer term. The changes to the model under scrutiny were both of qualitative and quantitative character as the entirely new data bank of experimental data points has been collected to devise a new form of a semi-empirical model applicable both to conventional and small diameter channels. The fluids examined were: R141b, R134a, R113, R12 and R11 and the total number of considered experimental points, corresponding to small diameters, amounted to over 260. The range of considered tube diameters varied from 1 to 3mm. The use of the model is very simple on account of its analytical form, which is its fundamental advantage. It is of general character, and its coefficients do not depend on the type of fluid. Comparisons with experimental data show good agreement.

Keywords: Flow boiling, heat transfer, small diameter channels

NOMENCLATURE

C_p - specific heat at constant pressure, J kg^{-1} K^{-1}
d - channel diameter, m
g - gravity, m s^{-2}
G - mass flowrate, kg m^{-2} s^{-1}
h_{fg} - latent heat of evaporation, J kg^{-1} K^{-1}
l - bubble characteristic length, $l = \sqrt{\sigma / \left(\rho_L - \rho_G \right) g}$, m
p - pressure, Pa
P - correction in equations (1) and (5)
q - heat flux density, W m^{-2}
R - two-phase flow multiplier
t - temperature, ^{o}C
x - quality

Non-dimensional numbers

$Bo = q / \left(G \cdot h_{fg} \right)$ - Boiling number
$\mathrm{Re} = \left(G \cdot d \right) / \mu_L$ - Reynolds number
$\mathrm{Pr} = \left(\mu_L \cdot C_L \right) / \lambda_L$ - Prandtl number

Greek letters

α - heat transfer coefficient, W m^{-2} K^{-1}
λ - thermal conductivity, W m^{-1} K^{-1}
σ - surface tension, N m^{-1}

Subscripts:

G - saturated vapour
GO - gas only
L - saturated liquid
LO - liquid only
PB - pool boiling
TP - two-phase flow
TPB - two-phase boiling

1. INTRODUCTION

Boiling heat transfer, as one of the most efficient techniques for removing high heat fluxes, has been studied and applied for a very long time. Nowadays, rapid development of practical engineering applications for micro-devices, micro-systems, advanced material designs and manufacturing of compact heat exchangers, high capacity micro heat pipes for spacecraft thermal control, electronic chip increases the demand for better understanding of small and micro-scale transport phenomena. The last decade of the twentieth century witnessed a rapid progress in the research of micro- and nano-scale transport phenomena, which borne important applications in modern technologies such as micro electronics etc.

There is a considerable number of empirical correlations that can be employed in calculation of the heat transfer coefficient for flow boiling in channels. Excellent review on that topic has been provided by Thome et al. (1996). A paramount difficulty in development of correlations for flow boiling is represented by the fact that the considered phenomenon occurs in practice at small mass flowrates and small to moderate heat fluxes (refrigeration, air-conditioning) or, alternatively, at large mass flowrates and moderate to high heat fluxes (power engineering). Capturing all these trends with one only correlation is a very difficult task. Heat transfer issues in small diameter channels received already attention in numerous papers, just to mention a few by Bergles et al. (2003), Kandlikar (2001) and Hausner and Herwig (2003). Hausner and Herwig (2003) wonder whether there are any other additional effects influencing heat transfer, which are more pronounced in small diameter channels, such as for example axial heat conduction (small Peclet numbers), conjugate heat transfer in walls (relatively thick walls), temperature dependent properties (large axial temperature gradients), pressure dependent properties (large axial pressure gradients) or the wall roughness. The limit between conventional channels and small channels was suggested by Kandlikar (2001). He proposed that channel diameters smaller than 3mm should be regarded as minichannels and also devised a limit for microchannels to be smaller than 600μm. Hausner and Herwig (2003) concluded that the phenomena in diameters greater than 600μm are just the scaled down effects from conventional channels, which is in some contradiction to other studies.

The empirical correlations suggested up to date are based on reduction of a restricted number of authors own experimental data or form a generalisation of a greater number of experimental data from various authors. In the latter case correlations are usually of a worsen accuracy in predicting the heat transfer coefficient or the pressure drop due to the fact that each experiment bears its own measurement error, however such correlations are of a more general character. All these empirical correlations, however, are the fits to experimental data, which is restricts their generality. The most popular approach to model flow boiling is to present the resulting heat transfer coefficient in terms of nucleate boiling heat transfer coefficient and convective boiling heat transfer coefficient:

$$\alpha_{TPB} = S\,\alpha_{PB} + F\,\alpha_{cb} \tag{1}$$

where α_{PB} – pool boiling heat transfer coefficient, α_{cb} – convective heat transfer coefficient, which can be evaluated using for example the Dittus-Boelter type of correlation. Such approach, equation (1) has originally been postulated by Chen (1966). Function S is the so called suppression factor which accounts for the fact that together with increase of vapour flow rate the effect related to forced convection increases, which on the other hand impairs the contribution from nucleate boiling, as the thermal layer is reduced. The parameter F accounts for increase of heat transfer with the increase of vapour quality. That parameter always assumes values greater than unity, as flow velocities in two-phase flow are always greater than in the case of single phase flow. Subsequently, functions F and S have been assigned mathematical formulas and are therefore more comfortable to be used.

Kattan et al. (1998) concluded that only the models based on distinguishing between flow regimes should be seriously considered for a general use. Wojtan et al. (2004) postulated models for stratified wavy and mist flow regimes. These models are very promising, however, fall to the class of the regime dependent models. They require knowledge of the particular flow regime, as well as the proportion of liquid and gaseous phase in the flow. For that reason such model is difficult to be applied in engineering practice, as it requires prior knowledge of the flow conditions. The model considered in the present paper possesses a feature enabling distinguishing the flow pattern. That is coded in the form of a two-phase flow multiplier, which renders it a general model. Such two-phase flow multiplier can be structure dependent, and in such case more accurate results can be obtained, but also a general two-phase flow multiplier, valid in the entire range of quality variation, can be used, which generalizes the method. The modern model should balance contributions to heat transfer coefficient coming from the bubble nucleation and convective heat transfer. The former case is adjusted to experimental data by the empirical correction, whereas in the case of convective heat transfer the two-phase flow multiplier is responsible for the mechanism of heat exchange (Mikielewicz (1973)). In authors opinion the correlation due to one of the co-authors, Mikielewicz (1973), recalled and further modified in the present paper, satisfies all these requirements.

Another issue regarding existing models of flow boiling is their application to small diameter channels as one must borne in mind that most of such correlations present in the literature has been developed for conventional channels, i.e. channels with diameter greater than 3mm. Therefore a great challenge for such correlations is to predict heat transfer data in the case of smaller diameters, where there are no new effects, but the ones dominant in the case of conventional channels become less pronounced (for example nucleate boiling) and other start to be more important (surface tension effects). Such attempt has been done by the authors of the present model. The model has been tested first on the available database for refrigerants R11, R12 and R22 in the case of conventional diameters of channels and subsequently the model was tested on some data available from literature for small diameter channels for R11, Bao et al. (2000), R12, Tran et al. (1996) and R141b, Lin et al. (2001). Considered range of parameters for small diameter channels has been put together in Table 1.

Author	**Fluid**	**p**	**G**	**q**	**D**
		bar	kg m^{-2} s^{-1}	kW m^{-2}	m
Lin et al. (2001)	R141b	1.35 ÷2.2	510	18 ÷ 72	0.0011
Tran et al. (1996)	R134a R12 R113	1.38 ÷ 8.56	30 ÷ 500	2.2 ÷ 90	0.00246 0.00292
Bao et al. (2000)	R11 R123	1 ÷ 5	50÷1800	5÷200	0.0019

Table 1. Experimental data used in the development of model/

2. THE MODEL FOR CONVENTIONAL CHANNELS

The model was developed on the basis of consideration of energy dissipation in flow boiling (Mikielewicz (1973)). That provided the ground for the underlying hypothesis of the method which enabled finding of a right set of terms describing the flow boiling heat transfer. As a result a model has been developed which was based on the knowledge of heat transfer coefficients for simpler cases, namely of pool-boiling and convective single-phase flow, as well as on the knowledge of hydrodynamical resistance coefficients for two-phase vapour-liquid adiabatic flow conditions, in other words the two-phase flow multiplier. Relations describing mentioned above modes of heat transfer, namely pool boiling heat transfer coefficient, liquid-only heat transfer coefficient as well as two-phase flow multiplier are usually empirical and that was the principal reason why the model was named a semi-empirical one. Mikielewicz (1973) made use of a generalization of investigations by various contemporary authors, as well as his own for R21, to devise a semi-empirical model for determination of the heat transfer coefficient for boiling flow. The accuracy of correlation was improved some time later, Mikielewicz et al. (1992), on the basis of the available data base for R12, R11 and R22, which amounted to over 2500 experimental points. That correlation proved to be very efficient in the case of calculation of heat transfer coefficient in flow boiling for conventional tube diameters, i.e. diameters greater than 3mm. In the discussed modification the model was fitted with an empirical correction modifying the pool boiling heat transfer coefficient. That was primarily due to the fact that the temperature gradient in pool boiling is different then in flow boiling. The correlation proved a very satisfactory consistency as almost 50% of experimental data considered felt within the error band of ±20% and almost 70% of data was within the error band of ±30%. The detailed description of the model and experimental results were presented in Mikielewicz et al. (1992).

3. THE IMPROVED MODEL FOR CONVENTIONAL CHANNELS

The model under scrutiny here is a modification of a model developed earlier by J. Mikielewicz (1973), which in its final form read:

$$\frac{\alpha_{TPB}}{\alpha_L} = \sqrt{R^{0.8} + \left(\frac{\alpha_{PB}}{\alpha_L}\right)^2} \tag{2}$$

In (2) R denotes the two-phase flow multiplier and originally the correlation due to Lottes and Flinn has been suggested, however any correlation describing the two-phase flow multiplier could be used, either developed for a particular flow structure or a one valid for the entire range of quality variation. In (2) α_L is a liquid only heat transfer coefficient and α_{PB} is a pool boiling heat transfer coefficient. In the present study that is evaluated from the widely known formula due to Cooper (1984).

In later studies Ould Didi et al. (2002) found that correlation due to Muller-Steinhagen and Heck (1986) performs best in the case of refrigerants and in the present study the original form of the Muller-Steinhagen and Heck correlation with a small adjustment discussed later is used. The correlation is valid in the entire range of quality.

Consideration of superheating or subcooling is usually another challenging issue to be modeled in correlations describing flow boiling. In refrigeration problems the evaporator is supplied with liquid (or even more often subcooled liquid) and hence the starting point for modelling of flow boiling problem is the heat transfer coefficient not corresponding to liquid, but some other value more appropriate for that region. In the case of empirical correlations it is common to assume the start of boiling at quality x=0. That is however, what is usually regarded as an equilibrium mass quality (in the saturated flow boiling). Beginning of flow boiling starts much earlier, where first alienated bubbles are growing on the wall and immediately condensing, then followed by a more pronounced generation of bubbles but still in the region where the core of the flow is below the saturation temperature, and finally the state is reached where the core attains the saturation temperature. That is presented in fig. 1, where the first described situation corresponds to the so called “wall voidage” region in the subcooled boiling regime. With increasing heating there starts to be more bubbles generated, but still in the subcooled boiling regime. Heat transfer in that region, which can be calculated in a very detailed way, but empirical or theoretical correlation rather do not consider such conditions. At small vapour production there is a lack of reliable experimental data hence as a starting point the heat transfer coefficient corresponding to the single phase liquid has been assumed, even though it may sound as a crude assumption. The similar situation holds for the quality corresponding to saturated vapour. The cases mentioned above are the limiting modelling cases, i.e. to be attained in the limit. In (3) there also appears the reference heat transfer coefficient $\alpha_{REF} = k\,\alpha_L$, which can account for the effect of subcooling at the channel inlet. The parameter k is responsible for subcooling. In presented calculations it has been decided that the correction accounting for subcooling k is equal 1 and will be given additional attention in further studies.

An improved version of the postulated method incorporates an empirical correction, P, to the bubble generation term to improve the model performance. The model now reads:

$$\frac{\alpha_{TPB}}{\alpha_{REF}} = \sqrt{R_{M-S}^{0.76} + \frac{1}{1+P}\left(\frac{\alpha_{PB}}{\alpha_{REF}}\right)^{2}} \tag{3}$$

Relation (3), in case when bubble generation is not present, yields $\alpha_{TPB} \approx R_{M\text{-}S}^{0.4}\alpha_{REF}$, whereas in the case when bubble generation is dominant then the second term in (3) plays more important role. In an improved model the correlation describing two-phase flow multiplier due to Muller-Steinhagen and Heck (1986) is used, $R_{M\text{-}S}$, however in a slightly modified form. Modification was introduced to assure consistency of the relation (3) at quality equal zero and one, i.e. limiting cases. A modified version of $R_{M\text{-}S}$ reads:

$$R_{M-S} = \left[1+2\left(\frac{1}{f_1}-1\right)x\right]\cdot(1-x)^{1/3} + x^3\frac{1}{f_{1z}} \tag{4}$$

where function $f_1 = \left(\frac{\mu_L}{\mu_G}\right)^{0,25}\cdot\left(\frac{\rho_L}{\rho_G}\right)^{-1}$ and function f_{1z}, $f_{1z} = \frac{\mu_G}{\mu_L}\cdot\frac{C_L}{C_G}\cdot\left(\frac{\lambda_L}{\lambda_G}\right)^{1,5}$. Function f_1 is formed from a ratio of pressure drop of liquid only flow to pressure drop of gas only flow (as in the original form of the Muller-Steinhagen and Heck correlation), whereas f_{1z} is derived from a ratio of heat transfer coefficient of liquid only flow to heat transfer coefficient in gas only flow. The correction P was sought in the following general form:

$$P = a\left(R_{M-S}-1\right)^{b}\,\mathrm{Re}_L^{c}\,Bo^{d} \tag{5}$$

The exponents a,b,c,d have been adjusted using the available data base for R11, R12 and R22 (2500 experimental points). The bank of data has been taken from the Table 2.

Author of data	Refrigerant	T_{SAT}	x	D	G	q
		^{o}C		M	kg/m^2s	kW/m^2
Bandel (1973)	R12	0	0.1÷0.9	0.014	92÷735	1.1÷71
Iwicki, Steiner (1979)	R12	0,-10,-20	0.1÷0.8	0.014	51÷241	1.5÷81
Bandel (1973)	R11	10	0.1÷0.8	0.014	113÷725	1.0÷70
Chawla (1967)	R11	0,10,20	0.1÷0.98	0.006 0.014 0.025	14÷216	1.2÷16
Bandel (1973)	R22	-20,-10,0	0.1÷0.8	0.014	90÷700	0.7÷70
Blaszewski (1977)	R22	0,-7	0.1÷0.85	0.014	260÷450	0.6÷67

Table 2. Experimental data for conventional tubes used in the study.

Actual form of a method for heat transfer coefficient calculation in flow boiling now reads:

$$\frac{\alpha_{TPB}}{\alpha_{REF}} = \sqrt{R_{M-S}^{0.76} + \frac{1}{1+2.53\times10^{-3}\,\mathrm{Re}^{1.17}\,Bo^{0.6}\left(R_{M-S}-1\right)^{-0.65}}\left(\frac{\alpha_{PB}}{\alpha_{REF}}\right)^{2}} \qquad (6)$$

The results of calculations for conventional channels obtained using relation (6) have been presented in fig. 2 and 3. As can be seen quite satisfactory consistency is obtained for the set of experimental data of R11, R12 and R22. Over 63% of experimental data is described by the correlation within ±30% error margin. Authors are aware that the number of experimental data considered is not very extensive and hence authors are open for all exchange of information with that respect. The postulated method in limiting cases reduces to the Chen's correlation (1). In case where bubble nucleation term is dominant (influence of convective flow is negligible) then suppression factor S in (1) is corresponding to the square root of the 1/(1+P), whereas in case where bubble nucleation is negligible then the parameter F in (1) is equivalent to $R_{M-S}^{0.4}$ in (6). As an example comparisons of the distributions of correction to the nucleate boiling term has been presented in Fig. 4, whereas the comparison with the parameter F has been presented in Fig. 5, respectively for the case of R134a. A good consistency can be found.

4. APPLICATION OF THE MODEL TO SMALL DIAMETER CHANNELS

In the case of a two-phase flow in small diameter channels the complexity of such flow is strongly influenced by the dynamics of the growing and flowing bubbles confined in a narrow space with a flowing liquid phase. Therefore additional factors must be considered, including the effects of the interface between vapour and liquid phases, and the wetted surface between liquid and the channel wall. An appropriate correlation should account for the effect of tube dimension as well as fluid surface tension. Considered correlation already contains the terms, which model theses influences, but their impact to the results is still too small. Therefore it has been decided to modify the two-phase flow multiplier to introduce the effect of surface tension. In the literature can be found a so called confinement number, introduced by Cornwell and Kew (1996), which is defined in the following manner:

$$Con = \frac{\left[\dfrac{\sigma}{g(\rho_L - \rho_G)}\right]^{0.5}}{D} \tag{7}$$

It ought to be stressed that such a term has already been used by the present authors during the development of the considered correlation, Mikielewicz et al. (1992). The confinement number, Con, is simply the ratio of the characteristic dimension during boiling to the channel diameter. In our nomenclature it bore the form of a simplex l/D. Such contribution has been applied to the correction P in Mikielewicz et al. (1992), which modifies only the bubble nucleation. In many cases of flow boiling in small diameter channels the nucleate boiling is not present at all and the heat transfer is mainly of the convective character. Following the idea of Tran et al. (2000) the confinement number will be applied to modify the two-phase flow resistance coefficient. In the paper by Tran et al. (2000) the correlation due to Chisholm has been modified to reflect the two-phase flow data in small diameter channels. Presented below is a final form of correlation derived by Tran et. al (2000):

$$R = 1 + \left(C \frac{\left(\frac{dp}{dz}\right)_{GO}}{\left(\frac{dp}{dz}\right)_{LO}} - 1 \right) \left[Con\, x^{0.875} (1-x)^{0.875} + x^{1.75} \right] \quad (8)$$

where C=4.3 and x is equilibrium mass quality. The above correlation is claimed to reduce data within 30% for all data considered by the authors and 93.8% of data falls in the error band of 20%. Three refrigerants have been considered, namely R134a, R12 and R113 at six different pressures ranging from 138 to 856 kPa and two sizes of round tubes of 2.46mm and 2.92mm inside diameters. However, careful inspection of (8) shows that it does not obey the limiting behaviour in the gas limit, i.e. the pressure drop for gas is not attained. Therefore it was decided to modify hitherto used correlation due to Muller-Steinhagen and Heck to incorporate the effect of surface tension. The modified correlation yields:

$$R_{M-S} = \left[1 + 2\left(\frac{1}{f_1} - 1 \right) x\, Con^m \right] \cdot (1-x)^{1/3} + x^3 \frac{1}{f_{1z}} \quad (9)$$

Preliminary tests show that a good agreement with the correlation proposed by Tran et al. is obtained when m=-1. Results of comparisons are presented in fig. 6. A very good consistency with Tran et al. correlation can be observed. Recapitulating, in calculations ought to be used a form of a correlation (4) with correction P (equation (7)) and R_{M-S} as in (9).

4. RESULTS OF CALCULATIONS

In order to perform comparisons some selected data from literature has been collected and the presented earlier correlation run to predict such cases. In figures 7-8 presented are the results of calculations carried out for the available data set of small diameter channels. The consistency is very satisfactory. Subsequently some runs were carried out for the flow development with quality for refrigerant R134a using the experimental data due to Yan and Lin (1998). The results show also a good quantitative character, see Fig. 9 and 10. It can be seen that the influence of change of heat flux still requires modification in the correction term P. The higher the heat flux the deterioration of heat transfer coefficient is much more pronounced. This means that the correlations developed earlier for larger diameters cannot be blindly used in predictions of data for smaller diameters, even in the light of the fact that the phenomena should only be scaled down in the considered cases. As we can observe the modifications performed in the course of present work significantly improve the correlation performance, as at least the qualitative trends are depicted properly.

6. CONCLUSIONS

In the paper presented have been considerations regarding two-phase flow boiling modeling in small diameter channels.

The presented correlation is of general character, and its coefficients do not depend on the type of fluid. The use of correlation is very simple on account of its analytical form, which is its fundamental advantage. Some heat transfer data exhibit deterioration with the flow quality in some cases, which is also present in the considered here experimental data. Such a case cannot be described by the majority of correlations known from literature on account, of their structure. The correlation under investigation possesses; such a capability which raises its quality. Such capability is coded within the relationship of the flow resistance coefficient and the correction P. The mutual relationship between these functions makes it possible to obtain a rising or falling dependence of heat transfer coefficient, during the boiling flow upon flow quality x. The accuracy of the proposed correlation, in the common range of parameters, is comparable with the best correlations known today.

Modelling of subcooled boiling requires further scrutiny. The correction P also requires further examination in order to incorporate the effects of surface tension explicitly. The model will not be completer until also the superheated region (post dryout) attracts sufficient attention. In such case it is also required to correct the heat transfer coefficient, as in the flow there are present droplets, which significantly enhance heat transfer.

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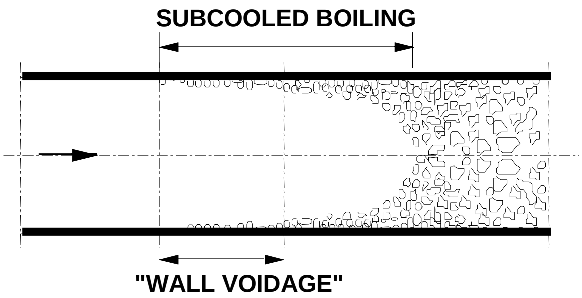

Fig. 1. Incipience of flow boiling.

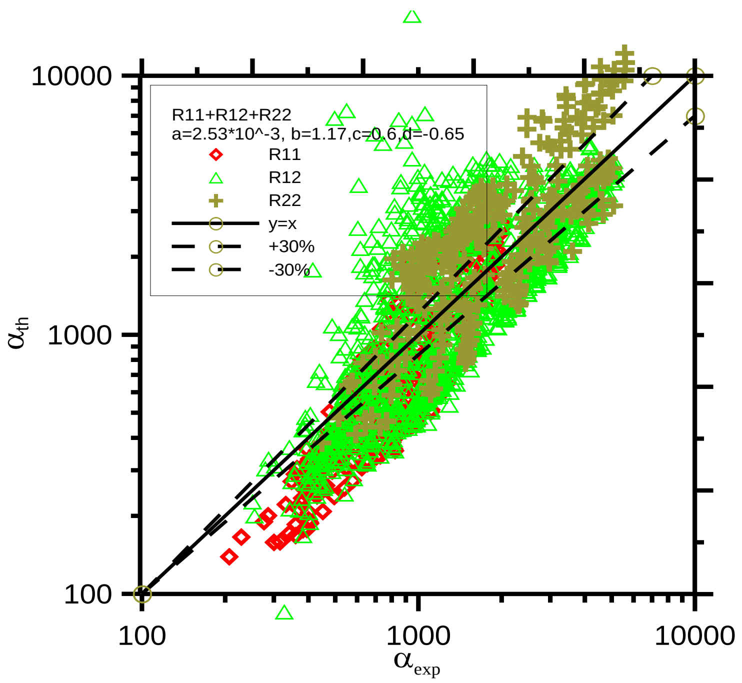

Fig. 2. Correlated database for conventional tubes using (7).

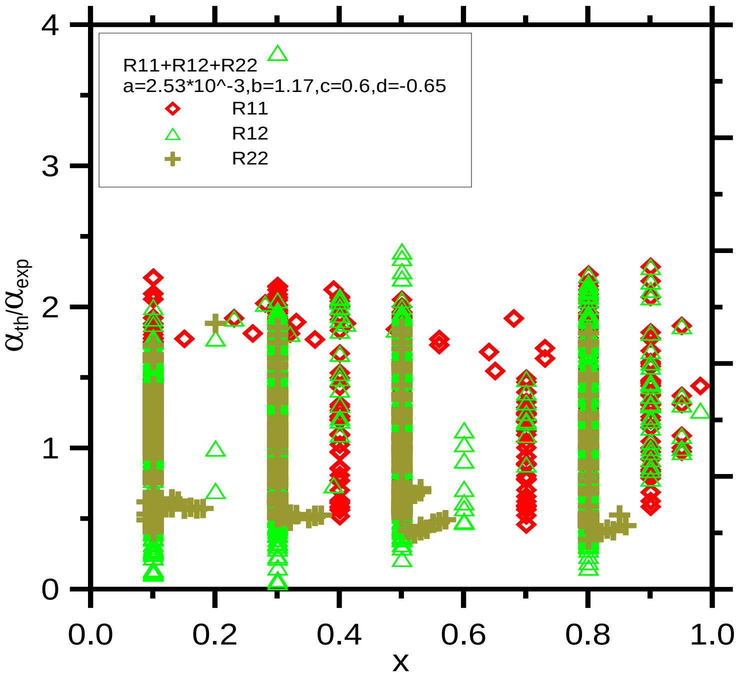

Fig. 3. Distribution of considered data bank described by (7) against quality.

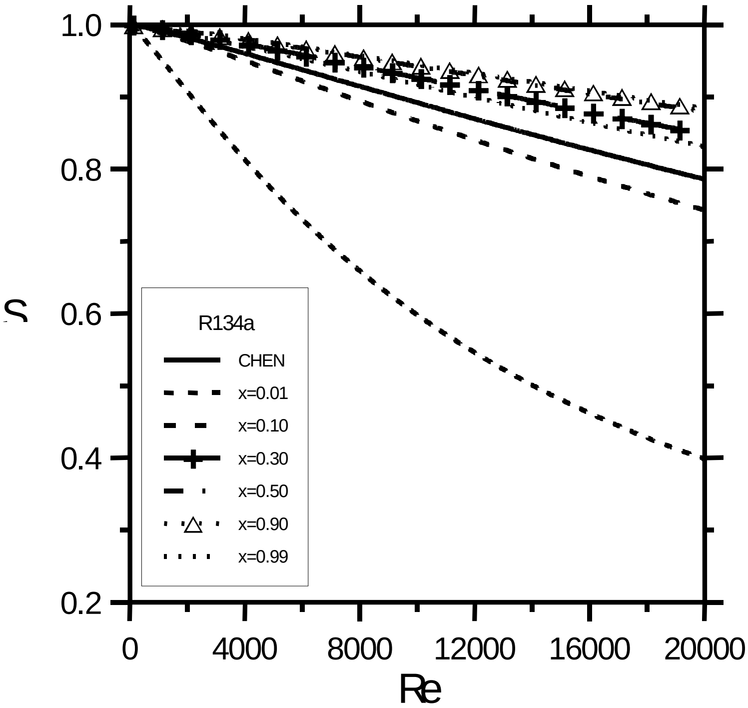

Fig. 4. Nucleate boiling correction from (6) in function of quality. Comparison with S parameter from (1) for the case of R134a.

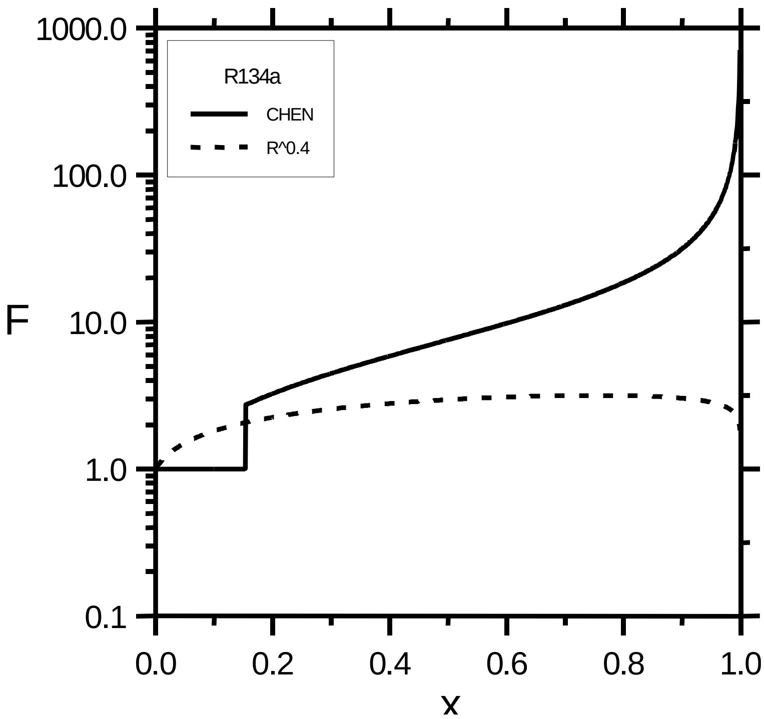

Fig. 5. Enhancement term F from (1) in comparison against the corresponding term $R_{M\text{-}S}^{0.4}$ in (6).

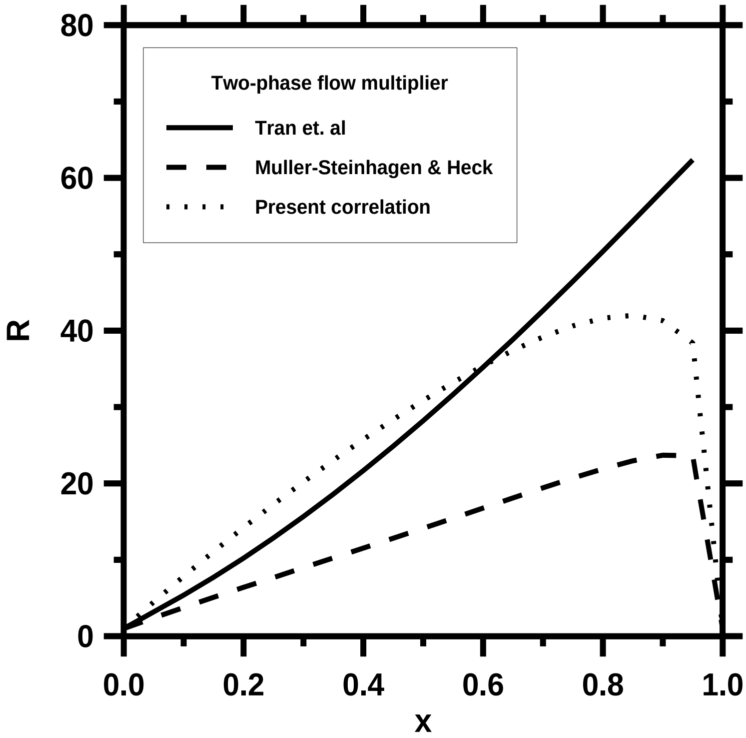

Fig. 6. Comparison of two-phase flow multipliers.

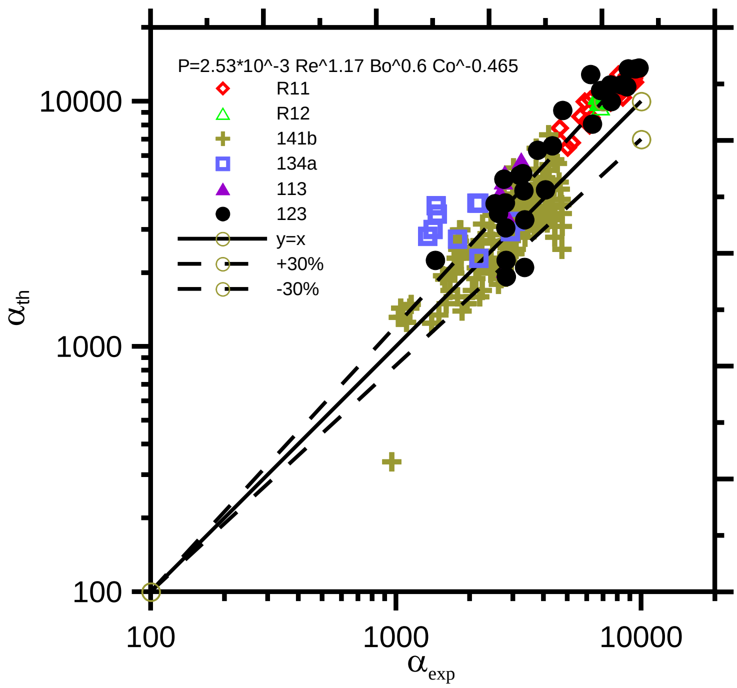

Fig. 7. Correlated database for small diameter tubes using (7) with application of (10).

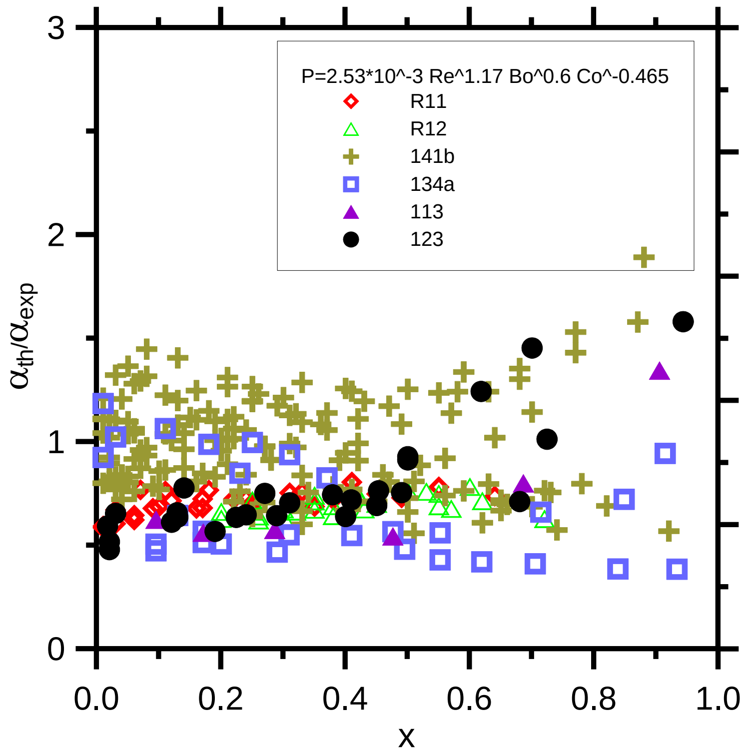

Fig. 8. Distribution of considered data bank described by (7) with (10) with quality for small diameter channels.

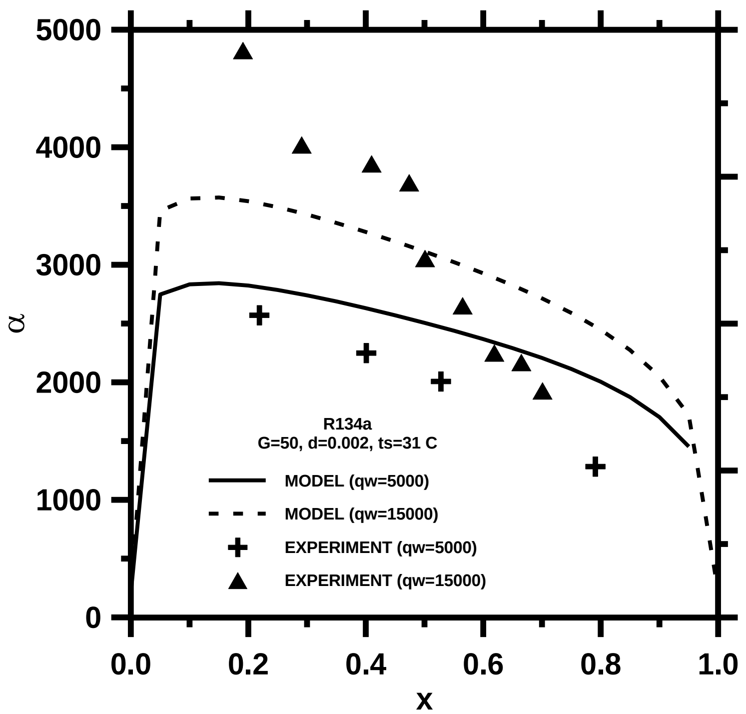

Fig. 9. Dependence of heat transfer coefficient with quality for r134a. Data due to Yan and Lin (1998). G=50 kg/m^{2}s, D=2mm.

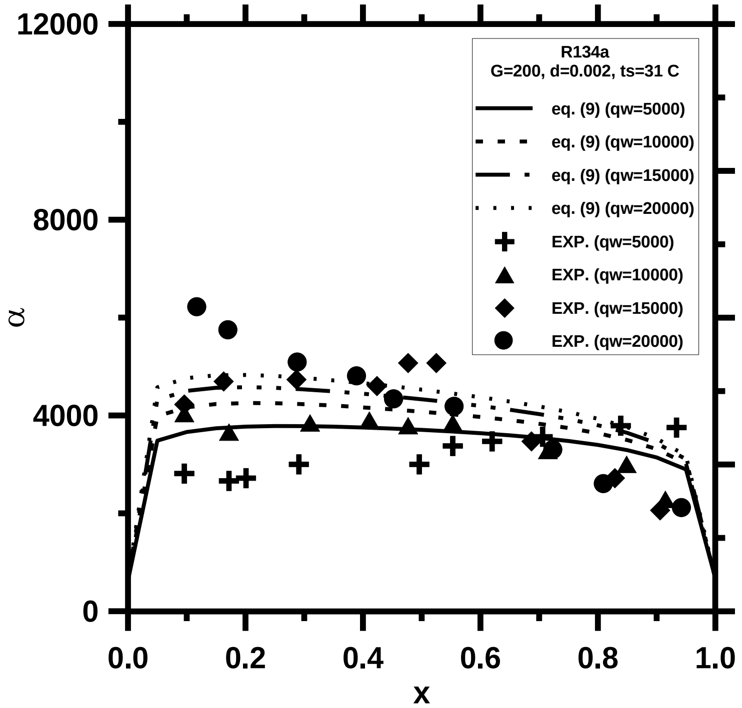

Fig. 10. Dependence of heat transfer coefficient with quality for r134a. Data due to Yan and Lin (1998). G=200 kg/m^{2}s, D=2mm.