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Nucleate pool boiling on horizontal tubes: a convection-based correlation

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Abstract—Pool boiling on horizontal tubes leads to a bubbly flow layer around the lower periphery of the tube in which bubbles slide around the surface. Heat transfer is by enhanced liquid convection at the surface around the bubble and by rapid evaporation of a thin layer under the sliding bubble. Both these mechanisms are a function of the vapour flow rate and a correlation of the following form is developed: $Nu = AF(p)\,Re_b^{0.67}Pr^{0.4}$. A is a function of the critical pressure alone, $F(p)$ is a function of the reduced pressure alone and Re_b is based on the mean vapour mass flux from the surface and the tube diameter. The correlation applies to water, refrigerants and organics boiling on tubes of 8–50 mm in diameter. Bearing in mind the wide scatter found in pool boiling experiments the correlation fits well with the available world data and is particularly suitable for use in equipment design calculations.

1. INTRODUCTION

ALTHOUGH studies of pool boiling are prolific in heat transfer literature, there is one important aspect which has attracted limited attention. This is the geometry of the boiling situation. Pool boiling correlations generally contain no physical dimension and are not categorised according to the physical arrangement. This is curious because pool boiling is largely convection driven with a small latent heat transport at the surface and the equivalent single-phase convection correlations are always associated with a particular geometry. Furthermore, experimental evidence supports a strong dependence on geometry. At low heat flux, boiling is influenced by the natural convection flow as shown in Fig. 1 from our earlier work [1] while at higher heat flux it is influenced by the bubble stream at the surface as shown later.

There is however a fundamental difference between the turbulent convection of the liquid at the surface in pool boiling and in single-phase forced convection. In boiling the convection is driven by vapour bubbles at the surface creating liquid movement while in forced convection the liquid movement is caused by external flow. Thus in boiling the mass flux influencing the convection is given by:

$$G_g = \frac{\dot{m}_g}{A}$$

where $\dot{m}_g$ is the mean vapour production rate. Under saturated pool boiling conditions this is directly related to the heat flux and the *boiling Reynolds number* becomes:

$$Re_b = \frac{G_g D}{\mu_f} = \frac{qD}{\mu_f h_{fg}}. \tag{1}$$

The case for using this form of the boiling Reynolds number is by no means new (see ref. [2] for example) but it does not appear to have been logically extended to various geometries.

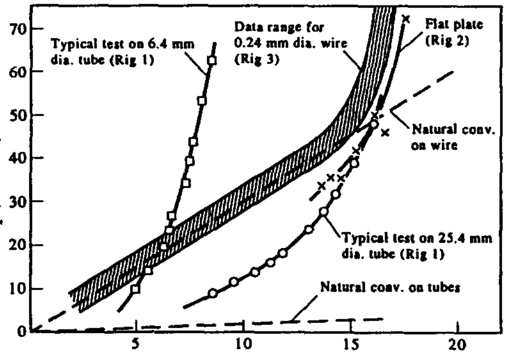


FIG. 1. Pool boiling of R113 at 1 atmosphere on similar surface finishes in three different geometries (from ref. [1]).

This work extends that of our earlier studies [1,3] and is aimed at the development of a simple design correlation for pool boiling on tubes under conditions of fully developed nucleate boiling on normal engineering surfaces. The correlation is presently restricted to normal process fluids (water, organics and refrigerants) and thus excludes liquid metals and cryogenic fluids.

2. MECHANISM OF POOL BOILING ON HORIZONTAL TUBES

Convective nucleate boiling on a tube differs considerably from that on a flat plate. In the latter case bubbles are formed at scattered nucleation sites and depart taking part of the superheated boundary layer with them and causing an in-rush of fresh liquid to the surface. The convective effects are often characterised

NOMENCLATURE

A	heat transfer area [m^2]
D	tube diameter [m]
$F(p)$	pressure function, equation (3)
G_g	mass flux of vapour [kg m^{-2} s^{-1}]
h_{fg}	latent enthalpy [kJ kg^{-1}]
k_f	thermal conductivity of fluid [kW m^{-1} K^{-1}]
$\dot{m}_g$	mass flow rate of vapour [kg m^{-1} s^{-1}]
Nu	Nusselt number, $\alpha D k_f^{-1}$
Pr	Prandtl number
p	pressure [bar]
p_c	critical pressure [bar]
p_r	pressure ratio, p/p_c
q	heat flux [kW m^{-2}]
Re_b	boiling Reynolds number, $qD/\mu_f h_{fg}$.

Greek symbols

α	heat transfer coefficient [kW m^{-2} K^{-1}]
μ_f	dynamic viscosity [kg m^{-1} s^{-1}].

using the bubble diameter at departure, although this is a gross over-simplication [4].

In the case of a horizontal tube observation shows that nucleation occurs primarily at the under-side of the tube and that the bubbles then slide *parallel with the surface* to a point near the top before they depart. Thus a bubble-layer is formed around the tube as shown in Fig. 2 and analysis [3] of a radial control volume shows a voidage varying from virtually zero at the base to around 0.5 at the sides with a corresponding rapid increase in velocity from the base to the sides. It is also shown [3] that there is a considerable peripheral variation of heat transfer coefficient on the tube. (This is largely smoothed out in most experiments by the peripheral conduction in the tube.) Typical test results for a 27 mm diameter tube in R113 at 1 atmosphere are shown in Fig. 3. Here U is the vertical liquid velocity and pool boiling is therefore at $U = 0$. The value of α increases from the base to the sides in direct opposition to that for single phase flow where it is known [5] to drop. This increase is caused by the increase in bubble layer velocity from the base and by the additional latent heat transport from the thin layers formed under the bubbles as they slide around the surface.

This latter mechanism becomes more important as the liquid velocity increases and accounts for the bulge which develops at higher velocities at the sides in Fig. 3. In tube bundles, where velocities are higher still, sliding bubble evaporation becomes predominant as shown in ref. [6].

The existence of this bubble layer raises a question mark over the validity of considering this type of boiling as *pool boiling*. Here we take the definition of pool boiling as that occurring when there is no externally imposed liquid velocity in the general potential field around the tube. It is accepted that, to an observer on the tube surface, the tangential bubbly flow will differ little between pool boiling and boiling with an imposed liquid flow.

In summary, pool boiling on a tube induces a bubbly layer around the tube in which the local heat transfer from the surface is largely due to liquid convection and evaporation under sliding bubbles. The driving force for both these mechanisms is the bubble production rate and hence the mean vapour mass flux at the surface. This concept is used as the basis for development of a working correlation suitable for application to thermal design.

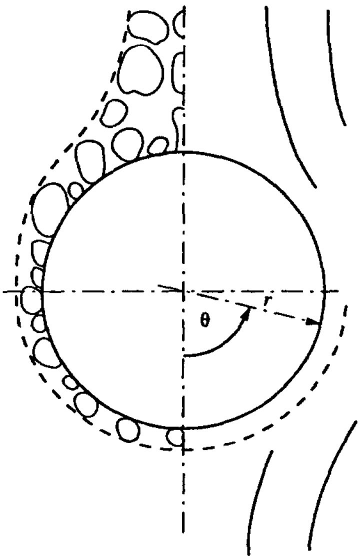


FIG. 2. The bubble-layer model.

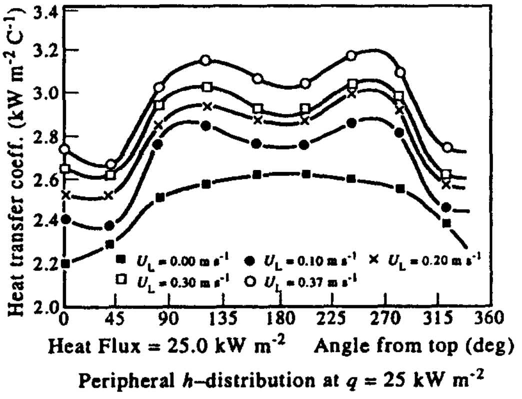


FIG. 3. Peripheral variation of heat-transfer coefficient with various approach velocities for boiling of R113 at 1 atmosphere at $q = 25$ kW m^{-2}.

Table 1. Data for pool boiling on tubes

Fluid	p_c (bar)	D (mm)	p (bar)	q (kW m^{-2})	α (kW m^{-2}K^{-1})	Reference
Water	221.2	19.1	1.013	22.8–97.3	3.5–7.0	T1
Water	221.2	19.1	1.013	56.8–199	4.9–12.5	T2
Water	221.2	13.7, 19.1	0.29–1.013	54.8–951.0	9.4–47.3	T3
Water	221.2	19.1	37.6, 106.4	2.2–61.8	3.2–44.2	T4
Water	221.2	25.4	1.013	34.1–242.9	3.1–18.2	T5
Water	221.2	12.0	0.24–2	22.3–232.6	3–22.9	T6
R113	34.15	19.1	1.013	2.7–100.0	0.5–4.6	T7
R113	34.15	25.4	1.013	10.0–60.0	0.9–2.6	T8
R113	34.15	8.0	1.013	20.0–100.0	1.6–4.2	T9
R113	34.15	12.1–50.8	1.013	9.7–47.8	1.0–7.0	T10
R113	34.15	19.1	1.013	10.0–50.0	1.0–3.5	T11
R113	34.15	19.1	1.013	10.0–200.0	2.1–9.1	T12
R113	34.15	12.7	1.013	11.4–74.8	1.1–4.3	T13
R113	34.15	25.4	1.013	5.1–68.0	0.6–2.7	T14
R113	34.15	25.4	1, 2, 5, 10	0.6–53.8	0.4–9.6	T15
R113	34.15	14.0	0.5, 1.0	30.0–146.3	2.0–9.1	T16
R113	34.15	12.7	1.7, 2.7	10.0–50.0	1.7–3.8	T17
R113	34.15	25.4	1.013	1.0–25.6	0.3–2.2	T18
R113	34.15	19.1	1.013	3.4–155.7	0.5–6.4	T19
R113	34.15	15.8	1.013	2.8–116.8	0.4–6.1	T20
R113	34.15	31.7, 44.5	1.013	72.7–165.2	4.5–6.6	T21
R113	34.15	15.9, 28.6	1.01, 2.03	0.8–83.9	0.3–5.1	T22
R11	44.15	15.9	1.01, 2.03	3.1–73.1	0.5–3.8	T22
R11	44.15	10.0	1.013	0.4–10.0	0.16–1.0	T23
R11	44.15	15.9	1.013	1.0–70.0	0.2–3.0	T24
R11	44.15	15.9, 22.2	1.01, 2.03	21.0–84.6	1.6–3.9	T25
R11	44.15	22.2	1.013	5.6–55.2	0.5–1.8	T26
R11	44.15	22.4	1.013	5.6–22.3	0.9–2.5	T27
R12	41.25	30.0	1, 2, 3 atm	5.4–50.8	0.5–3.1	T28
R12	41.25	15.9	1.013	3.1–68.2	0.7–4.0	T29
R12	41.25	19.1	3.8	2.7–15.1	0.8–2.6	T30
R12	41.25	8.0	5.9–20.5	0.2–39.9	0.2–21.0	T31
R12	41.25	14.0	7, 14, 30	26.3–138.7	8.0–60.3	T16
R114	32.57	14.0	3, 6, 12, 20	45.6–150.0	6.9–45.4	T16
R114	32.57	8.0	3.7	2.1–90.0	0.6–12.9	T32
R115	31.5	8.0	10.2	1.7–40.9	1.4–15.3	T33
R115	31.5	8.0	7.9–20.2	0.1–68.7	0.3–39.5	T31
R22	49.8	8.0	5, 10, 25	1.5–50.0	0.5–40.0	T34
Nonane	22.8	19.1	1.013	16.0–56.0	1.8–3.7	T1
Pentane	33.73	19.1	1.013	10.0–50.0	1.1–2.6	T35
Propane	42.47	8.0	2.3–17.1	1.5–50.0	0.4–32.6	T34
Hexane	30.3	31.8, 50.8	1.013	20.1–92.5	1.2–4.3	T36
Hexane	30.3	8.0	0.3–8.2	0.1–100.0	0.5–21.9	T37
Ethane	48.7	8.0	13.3–31.7	0.1–100.0	0.4–67.8	T37
Benzene	49.0	19.1	1.013	37.5–122.8	1.9–4.4	T2
Benzene	49.0	9.5	0.9–20.4	8.2–148.6	0.7–12.2	T38
Methanol	80.9	31.7, 50.8	1.013	26.7–203.3	1.5–5.0	T36
Methanol	80.9	19.1	1.013	49.2–181.2	1.9–4.9	T2
Ethanol	61.4	15.0	0.1–1.0	40.0–227.8	1.7–13.5	T39
Ethanol	61.4	19.1	1.013	38.9–186.6	2.6–8.4	T2
Isobutanol	43.0	19.1	0.3, 1.3	51.7–318.6	2.4–10.3	T3
p-Xylene	35.1	12.7	1.013	20.0–80.3	1.1–2.8	T40

3. THE POOL BOILING *Nu–Re* RELATIONSHIP

Given the essentially convective nature of the boiling on the tube due to the vapour production as discussed earlier, it is reasonable to attempt a correlation of the type commonly used for turbulent convection thus

$$Nu = C\,Re_b^m\,Pr^n. \tag{2}$$

In fully developed nucleate boiling m is usually taken as 2/3 or 0.7 and n for heating is 0.4. The adequacy of this correlation is examined for the available experimental data listed in Table 1. This table includes data in the open literature found by the authors for pool boiling on tubes under the following restrictions.

The *fluid* is water, a refrigerant or an organic. Liquid metals and cryogenic fluids were excluded owing to the strong influence of surface effects and the wider experimental errors due to the low temperature differences. Mixtures were excluded although data for pure components were extracted from studies on the effects of mixtures. The *pressure* range was taken as about 0.001–0.8 p_c, this being the range over which data were readily available.

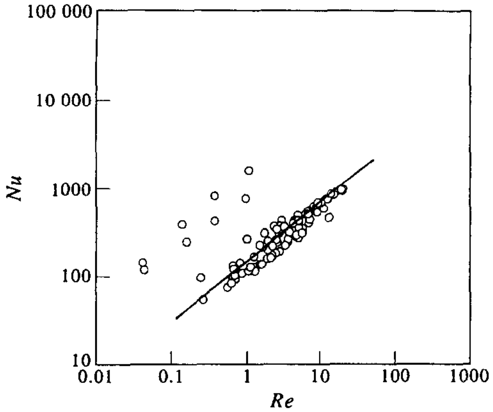


FIG. 4. The Nu–Re_b plot for data on water.

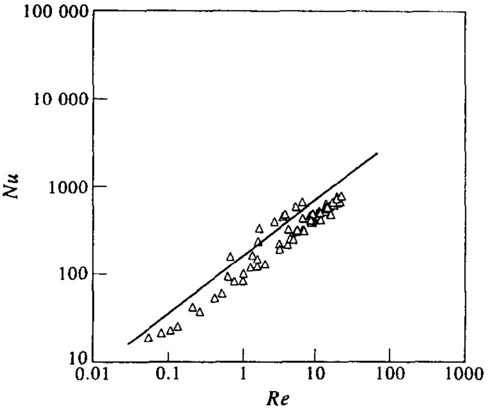


FIG. 6. The Nu–Re_b plot for data on R11.

The *heat flux range* covers fully developed nucleate boiling, say 10–80% of the critical heat flux. This excludes low heat fluxes where the availability of nucleation sites is predominant and high heat fluxes where liquid starvation may occur.

The *tube diameter* range was taken as 8–50 mm to cover normal industrial usage. The lower limit is dictated by the influence of departure bubble size on the bubble layer. There is a sharp rise of α at small diameters [1,7] although thin wires appear to have lower values. The upper limit reflects the limited data available above a diameter of 25 mm. At very large diameters it would appear [8] that the influence of diameter on α is dependent on the heat flux. The *tube surface* is generally as machined or drawn.

Some attempt to correctly apportion the weighting given to the data has been made by using about 6 to 10 of the q–ΔT values for each set of tests reported. The mean experimental curve has been used where possible to give some elementary smoothing of the data.

Only three fluids, water, R113 and R11 had data for a sufficiently wide range of parameters to construct reliable Nu–Re plots, and these are shown in Figs. 4–6. The degree of correlation between the data points along a slope of 0.67 (as indicated by the line in each figure) is outstanding in view of the well-known scatter of boiling results. For the record the remaining refrigerants and the organics are plotted in a similar way in Figs. 7 and 8. Although the range of parameters is less for these cases and the variations of Pr will alter the relative positions the same prevailing slope is evident. The implied variation with diameter

$$\alpha = f(D^{-0.3})$$

has been shown [1, 7] to be reasonable over the diameter range in question.

It is concluded that the Nu–Re representation is acceptable and the influence of the pressure and the fluid is now considered.

4. PRESSURE AND WORKING FLUID EFFECTS

For turbulent convection in single-phase fluid equation (2) is pressure independent. This is unlikely to be the situation for our case owing to the reduction of initial bubble size with pressure and the increase of vapour flow rate at the same heat flux among other effects. A number of papers contain data over a wide

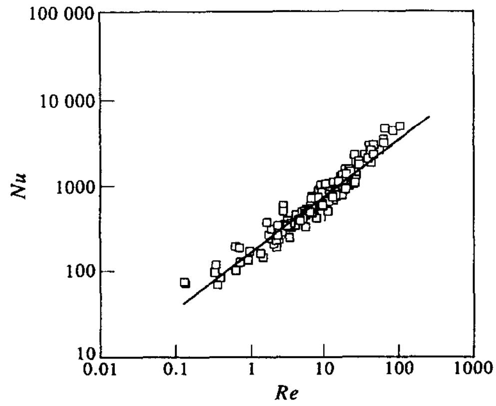


FIG. 5. The Nu–Re_b plot for data on R113.

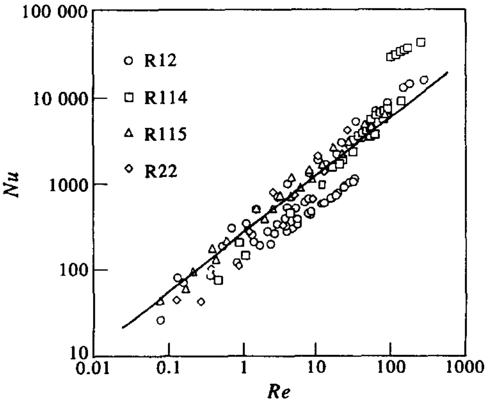


FIG. 7. The Nu–Re_b plot for data on refrigerants.

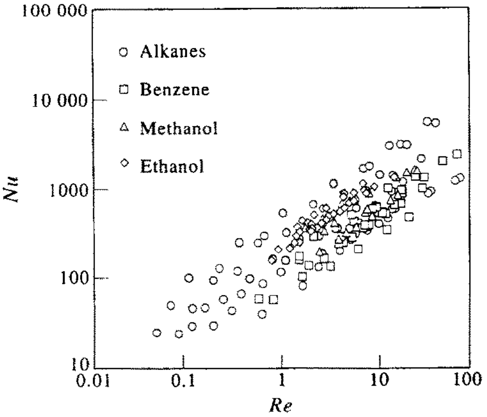


FIG. 8. The Nu–Re_b plot for data on organics.

pressure range and, while there are considerable differences in detail between these data for the same fluid, there is a general rise of Nu with Pr similar to that reported for pool boiling in other geometries.

With this in mind the pressure term in the Mostinski [9] equation based on corresponding states which has been widely used for 30 years was found to adequately describe the pressure dependence

$$F(p) = 1.8p_r^{0.17} + 4p_r^{1.2} + 10p_r^{10} \tag{3}$$

where p_r is p/p_c.

The Nu–Re relationships in Figs. 4–8 indicate a surprising insensitivity to the working fluid. This is itself a further indication of the appropriateness of this form of correlation. Cooper [10] discusses the influence of properties on (flat-plate) nucleate boiling and finally uses the molecular weight to indentify the fluid while accepting that other properties could be used. (He also mentions that copper tubes appear to have α-values 1.8 times higher than flat plates.) In the Mostinski relationship the fluid is identified by its critical pressure if an experimental constant of proportionality is not available and it seemed appropriate to follow this practice here.

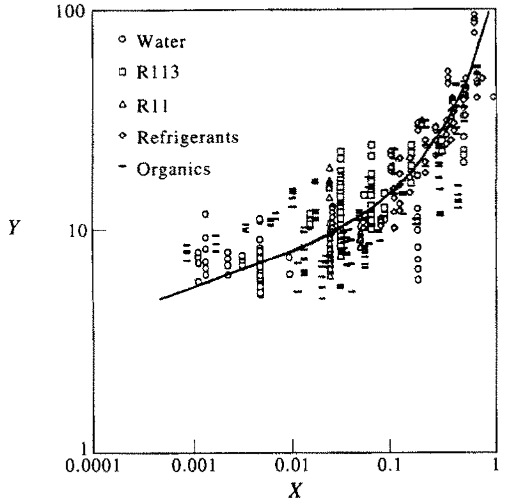
There were insufficient fluids with a wide range of critical pressures to make a precise analysis, especially in view of the low sensitivity to α as mentioned earlier. However, the best fit yielded a square-root correlation. The universal correlation is thus of the form:

$$\frac{Nu}{Re_b^{0.67}\,Pr^{0.4}} = AF(p), \tag{4}$$

where

$$A = C_{p_c}^{0.5}.$$

Figure 9 shows the data fit to the correlation from equations (3) and (4) as a function of the pressure. The symbols for water, R113 and R11 are the same as those used in Figs. 4–6 for ease of identification. Bearing in mind the wide range of fluids, diameters, pressures and heat fluxes and the inherent scatter in boiling experiments the data fit is not unreasonable.



$$Y = \frac{Nu}{Re_b^{0.67}Pr^{0.4}p_c^{0.5}}\ ;\ X = C\left(\frac{p}{p_c}\right)$$

FIG. 9. General correlation for all the fluids.

The curve shows the best fit value of $C = 9.7$ yielding the *final correlation* as

$$Nu = AF(p)Re_b^{0.67}\,Pr^{0.4}, \tag{5}$$

where $A = 9.7p_c^{0.5}$ with p_c, in bar (10^5 N m^{-2}) and $F(p)$ is given by equation (3). (The lines drawn in Figs. 4–6 actually show this correlation for water, R113 and R11 at 1 atmosphere. The line in Fig. 7 indicates the slope only.)

5. DISCUSSION AND CONCLUSION

There are dangers in recommending a general correlation, even when its area of application is clearly defined. These dangers are evident on a closer inspection of Fig. 9. For example the column of data for R113 at $X = 0.03$ (~ 1 atm) indicates an experimental scatter from $Y = 8$–30 for a single fluid and pressure. This scatter of data is of similar size to that found between different fluids at a common pressure, perhaps indicating that inherent randomness in pool boiling precludes greater accuracy. Furthermore, the general trend may mask differences within particular fluids. The data for water, for example, appears to exhibit a fairly constant mean value of Y at low X before rising at the higher pressure (but this may arise through over-emphasis on four low points at $X = 0.2$). Finally, seemingly minor effects can have a significant influence on boiling with sliding bubbles. Additives to water may alter the surface tension and hence the generated bubble size and the design of the boiling chamber may lead to an imposed liquid velocity on the tube (see Fig. 3).

In spite of these dangers the designer of equipment needs a core correlation which will give him the best mean value from the available data. Equation (5) is presented as this core correlation for pool boiling on horizontal tubes within the limitations given in Section 3.

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