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Direct contact evaporation of nearly saturated R 114 in water

G. P. CELATA, M. CUMO and F. D'ANNIBALE
ENEA, Energy Department, Via Anguillarese, 301 I-00060, Rome, Italy

and

F. GUGLIERMETTI and G. INGUÌ
Energy Department, Rome University "La Sapienza", I-00100, Rome, Italy

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Abstract—This paper reports the results of an experimental investigation on direct contact boiling of immiscible liquids. The continuous phase, water, is under stagnant conditions, while the dispersed one, Freon 114, is injected in the test section with different velocity and thermodynamic conditions through a nozzle. The injection system has been designed to vary the quantity of injected refrigerant and/or the liquid injection velocity. The test section is a Plexiglas vertical cylinder 72 mm i.d., 2.0 m long. Experimental data are obtained from high-speed movies of the continuous phase level during and after the Freon 114 injection, as well as from the movies of the rising boiling dispersed phase (injected under nearly saturation conditions). Vaporization rate has been characterized as a function of thermal hydraulic conditions (i.e. water temperature, system pressure and Freon mass flow rate). Direct contact boiling efficiency was derived by the evaluation of the fraction of Freon that did not undergo the boiling process during the transit in the test section.

1. INTRODUCTION

Direct contact heat transfer between two immiscible liquids has the advantage of avoiding metallic heat transfer surfaces that may be subjected to fouling and corrosion. Direct contact heat exchangers find applications in water desalination, geothermal heat recovery, ocean thermal energy conversion, thermal energy storage systems, etc. In a direct contact heat exchanger, the heat transfer process taking place can be sensible heat transfer or latent heat transfer. Sensible heat transfer requires large volumes of the fluids, which may be avoided if we utilize latent heat transfer due to the evaporation of a dispersed phase (volatile liquid) in a continuous liquid phase medium.

Experimental research on direct contact boiling of immiscible liquids, one of which is more volatile, has essentially been devoted so far to the study of the boiling of droplets of a volatile liquid (dispersed phase) injected in a continuous phase (a high-boiling liquid). This latter is obviously characterized by a temperature higher than the saturation value of the low-boiling liquid, corresponding to the system pressure.

Among the first systematic studies conducted on the heat transfer of volatile liquid droplets rising in a continuous immiscible liquid phase, Sideman and Taitel [1] carried out experiments on the boiling of pentane and butane in distilled water and sea water at atmospheric pressure, using a high-speed camera. However, the picture did not allow an accurate evaluation of the time of complete droplet evaporation and of the actual heat transfer surface. It is, in fact, difficult to get from the pictures the evolution of a three-dimensional (3D) system.

To overcome the above difficulties, Prakash and Pinder [2] measured the total evaporation time of drops of furan, isopentane and cyclopentane in water, using a dilatometric technique. Prakash and Pinder also evaluated the heat transfer coefficient using a cine-photographic technique [3].

Shimaoka and Mori [4] performed experiments with *n*-pentane drops 2.0–6.5 mm in diameter evaporating in water under pressures of 0.11–0.49 MPa. They complemented the results obtained by Shimizu and Mori [5] with smaller drops (1.4–1.7 mm in diameter) of *n*-pentane and R 113 under pressures of up to 0.48 MPa. It was pointed out that the instantaneous heat transfer coefficient for each drop, evaporating and thereby taking the form of a liquid–vapour two-phase bubble, can be defined in four different ways, depending on how one takes into account possible evaporation of the surrounding water into the bubble. The effect of the pressure on the evaporation length is also studied.

Tadrist *et al.* [6] studied the vaporization by direct contact of R 113 and *n*-pentane dispersed into a column of water flowing countercurrently. The vaporization of a single droplet in a stagnant liquid medium, and the evaporation of a multidroplet flowing system

NOMENCLATURE

A	cross flow area [m^2]
a	jet factor
b	drops falling down factor
C	parameter defined in equation (10)
C_p	specific heat [$J\ kg^{-1}\ K^{-1}$]
D	diameter [m]
$Eö$	Eötvös number, $(\rho_c - \rho_{d,v}) g D_f^2/\sigma$
g	gravitational acceleration [$m\ s^{-2}$]
h	local heat transfer coefficient [$W\ m^{-3}\ K^{-1}$]
h_{lg}	latent heat [$J\ kg^{-1}$]
H_v	evaporation length [m]
h_v	volumetric heat transfer coefficient [$W\ m^{-3}\ K^{-1}$]
Ja	Jakob number, $\rho_c C_p \Delta T/\rho_{d,v} h_{lg}$
K	thermal conductivity [$W\ m^{-1}\ K^{-1}$]
Nu	Nusselt number, hD/K
p	pressure [Pa]
p^*	saturation pressure at T^* [Pa]
Pe	Peclet number, $uD\rho C_p/K$
T	temperature [°C]
t	time [s]
T^*	temperature at which, in the bubble, $p_c^* + p_d^* = p_\infty$ [°C]
t_i	injection time of the dispersed phase [s]
u	velocity [$m\ s^{-1}$]
V	volume [m^3]
v	specific volume [$m^3\ kg^{-1}$]
v^*	specific volume at T^* [$m^3\ kg^{-1}$]
W	thermal power [W].

Greek symbols

ΔV	volumetric variation of the continuous medium [m^3]
ΔT	temperature difference between continuous and dispersed phase [K]
ΔT^*	$T_\infty - T^*$ [K]
Γ	mass flow rate [$kg\ s^{-1}$]
ρ	density [$kg\ m^{-3}$]
σ	surface tension [$N\ m^{-1}$]
τ	time of total evaporation [s].

Subscripts

c	continuous phase
cr	pertains to the critical condition
D	pertains to the drop
d	dispersed phase
i	injection
in	inlet conditions
j	pertains to the jet
l	liquid
n	pertains to the nozzle
o	relative to the initial time
s	superheating
sat	saturated conditions
v	vapour
w	water
∞	pertains to continuous phase undisturbed.

were studied. They outlined the importance of the determination of the effective exchange surface for the bubble droplet during its rise in an immiscible liquid in determining a formula for its evaluation.

Seetharamu and Battya [7] studied the direct contact evaporation of R 113 and *n*-pentane in a stagnant column of distilled water. The effects of operational parameters such as the column height, the phase temperature difference, the dispersed phase flow rate, and the diameter and number of orifices in the distributor on the volumetric heat transfer coefficient were investigated. A positive influence of dispersed phase flow rate and column height was observed, while the effect of the initial drop diameter, at high rates of coalescence, was found to be almost negligible.

Fuyita *et al.* [8] performed experiments where droplets of R 113 were countercurrently injected through a nozzle into hot water flowing upwards in a vertical channel, analyzing the vapour flow structure and boiling flow patterns along the channel. Boiling R 113 in the region just behind the nozzle exit takes the form of one of three flow patterns, unstable, stable or column-like, and the heat transfer rate is improved by this order of change in the boiling regimes. Boiling regimes depend on thermal hydraulic and geometric conditions.

The present work deals with an experimental study of direct contact evaporation of a liquid jet of R 114 injected at nearly saturation conditions in a hot column of stagnant demineralized water. The dispersed phase is injected from the bottom of the vertical test section, where, coming into contact with the liquid having a temperature higher than the R 114 saturation temperature at the system pressure, it tends to evaporate, rising with the non-vaporized fraction. As the water temperature approaches the Freon saturation temperature, i.e. the water superheating with respect to the refrigerant decreases, vaporization of droplets from the jet break-up takes place only partially. Below a given water superheating drop boiling does not occur. In the above conditions, non-vaporized liquid drops tend to fall down on the bottom of the test section (R 114 density is higher than that of water). Here, if the surface has sufficient roughness to provide the nucleation sites, missing in the drops rising in the water column, the evaporation of collected drops takes place. Because of the lack of nucleation sites in the direct contact evaporation of drops in a con-

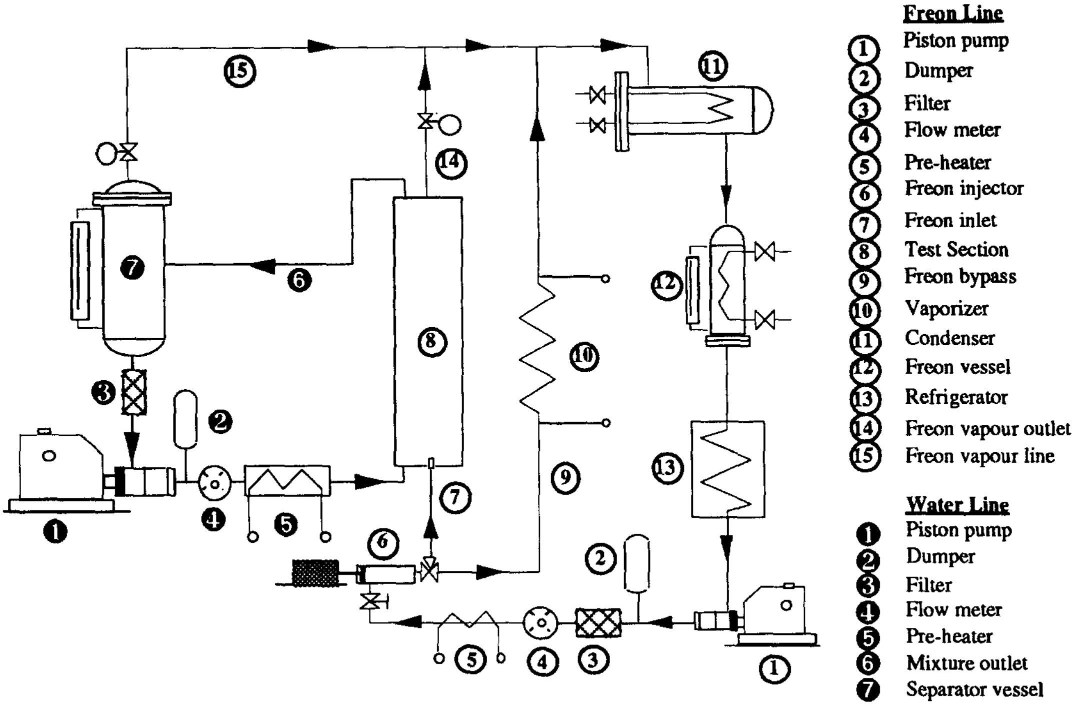

Fig. 1. Schematic of the experimental facility.

tinuous medium, it may happen that the drops of the volatile liquid also remain in a metastable state for a long time. Experimental results will show the influence of the water superheating on the vaporization rate, providing the threshold below which drops falling down are no longer negligible. The latter information is useful for design purposes, as, in continuously operating equipment, drops falling down can cause a piling up of refrigerant in a metastable state in the lower part of it. In addition to the off-normal operating conditions, this condition could present risks of overpressure if a trigger event caused the flashing of the metastable volatile liquid.

2. THE EXPERIMENTAL APPARATUS

The experimental facility employed in the present research is sketched in Fig. 1. The loop consists of two lines : the demineralized water line (continuous phase) and the R 114 line (dispersed phase). Water enters by a three-head piston pump, is filtered, electrically heated, delivered to the test section, and finally to the water–Freon separator. The separator is a heated-wall vessel where the two-phase mixture (water and R 114) is tangentially injected to make phase separation easier because of the centrifugal force and R 114 evaporation on the separator wall. R 114 vapour comes out from the upper part of the separator through a labyrinth towards the refrigerant tank, through a condenser, while water is collected in the water tank. The refrigerant, before the three-head piston pump, is cooled to avoid pump cavitation and, through the filter and an electric heater, flows through the injection system. The latter is a piston moved by a dc motor, capable of single injection of given volumes of R 114. The velocity of the piston and the volume of the liquid to be injected may be changed independently. The top venting line from the water–R 114 separator is connected to the condenser to ensure, under operating conditions, a uniform pressure in the whole loop and to prevent a possible overpressure in the separator. The experimental loop is made up of stainless steel AISI 316, while the two available test sections are made of Plexiglas to allow visualization of the phenomenon.

The first test section, used only to obtain some information for the development of the present study, is a square duct channel made up of Plexiglas 20 × 20 cm, 60 cm long. In this test channel, experimental runs are carried under steady-state conditions, injecting R 114 through a nozzle and the water through a porous medium, in co-current upward flow. The level is maintained at the pre-fixed value by a weir tube, through which the R 114–water mixture flows to the separator. Jet evaporation is analyzed using a high-speed shutter video camera and digital image processing to obtain a statistical distribution of the vapour volume and of bubble velocity as a function of the quota.

The test section employed for the experiments reported in the present paper consists of a vertical 2 m long Plexiglas channel, 72 mm i.d. and 4 mm wall thickness. The inlet nozzle is made of stainless steel AISI 316, with 2.0 mm i.d. (D_n). The outer part of the nozzle is insulated with Teflon, to limit heat transfer from the water side. The inlet temperature, $T_{d,in}$, is measured using a K-type thermocouple placed at the entrance of the nozzle. The temperature is measured after about 0.3 s from the beginning of the injection. The single experimental test is performed by injecting a fixed volume of R 114 (in saturated or nearly saturated condition) in a given time, obtained through the variation of the direct current motor voltage. The position of the water column level during the injection of the refrigerant and the time necessary for the volatile liquid to leave the top of the water column is recorded using a high-speed video camera system (KODAK Ektapro 1000). The variation of the level as a function of time is due to the vaporization of the injected volatile liquid (provided one accounts for the volume of the volatile liquid), and is the parameter observed in the experiments as a function of a thermal hydraulic parameter: R 114 inlet temperature, water temperature, R 114 injection velocity, R 114 injection volume.

It is necessary therefore, starting from the recorded water column level, as a function of time to obtain the corresponding values of the parameters of interest for the analysis of the thermal performances of the system, such as the jet evaporation length (or the total evaporation time) and the volumetric heat transfer coefficient. The first parameter is defined as the distance from the injection point at which the volatile liquid is completely evaporated. The second parameter is defined as the ratio between the exchanged thermal power and the volume involved in the process.

The above parameters cannot be directly derived from experimental data, as the total evaporation function, $\Delta V(t)$, is given by the sum of contributions due to the evaporation of a series of R 114 volume elements, injected in a time interval not completely negligible with respect to the total vaporization time. A possible procedure is to define a model, depending on the above parameters, for the theoretical evaluation of the level as a function of time. By equating the experimental values of $\Delta V(t)$ to model predictions it is possible to achieve an estimation of the jet evaporation length and the volumetric heat transfer coefficient. The experimental uncertainty in $\Delta V(t)$ measurements may be evaluated in $\pm 1.5\%$. The experimental runs were carried out as a combination of the following values:

R 114 inlet temperature, $T_{d,in}$ (nearly saturated)	17, 20, 23, 25, 30°C
Water superheating, $T_w - T_{d,sat} = \Delta T_S$	10, 13, 15, 17, 20, 25 K
R 114 injection velocity, u_n	from 0.56 to 1.8 m s^{-1}
Injected volume of R 114, $V_{od,l}$	from 1.5 to 3 cm^3
System pressure, $p_\infty = p_{sat}(T_{d,in})$	from 0.16 to 0.25 MPa

3. THE THEORETICAL MODEL AND DATA ANALYSIS

The variation of volume, $\Delta V(t)$, due to boiling of the volatile liquid, R 114, may be considered as given by the sum of contributions due to the vapour generated at time t from the vaporization of all the infinitesimal volumes of liquid Freon injected in the time increments dt^* between 0 and t. Assuming a constant R 114 flow rate during the injection time, $0 \leqslant t \leqslant t_i$, it is necessary to find a function that provides for each infinitesimal liquid volume, $dV_{od,l} = V_{od,l}\,dt^*/t_i$, the time-dependent expression of the vapour generated. Such a function will have then to be integrated between 0 and t_i to obtain $\Delta V(t)$.

From a preliminary study of the jet evolution, performed with the square duct test section, a continuous injection of R 114 and a digital image processing of vapour bubbles, the jet was observed to break into bubbles only in the proximity of the injection point, as shown in Fig. 2.

A typical jet evolution in the low part of the round duct test section is shown in Fig. 3. The four pictures are taken 1/7 s after each other. Jet evolution in the round duct test section confirms the assumption of the early fragmentation into drops of the liquid jet. It is therefore thought necessary to schematize the jet evolution in the water column as a series of drops rising independently of each other. To fully describe the liquid jet evaporation it is sufficient, as a starting point, to have a model that provides the description of the evaporation of single drops. The study of the jet boiling may therefore be turned to the evaluation of the difference between the actual behaviour (the experimental one) and the ideal behaviour of a series of isolated single drops. On the other hand it will be necessary to account for all possible phenomena characterizing the actual behaviour with respect to the ideal one. As for each drop it is possible to calculate the total evaporation time and the volumetric heat transfer coefficient, such parameters will also be available for the actual situation, once they have been corrected by taking into account the above differences.

A correlation for the description of single liquid drops evaporation in an immiscible liquid is given by Shimizu and Mori [5]:

$$Nu = 0.121 Pe^{0.5} \tag{1}$$

recommended for R 113–water systems. To calculate the bubble growth it is also necessary to know the average initial diameter of the drops generated by the liquid jet break-up, and the bubble rising velocity. The calculation of the average initial diameter, D_D, may be obtained from a procedure given by Horvath *et al.* [9]:

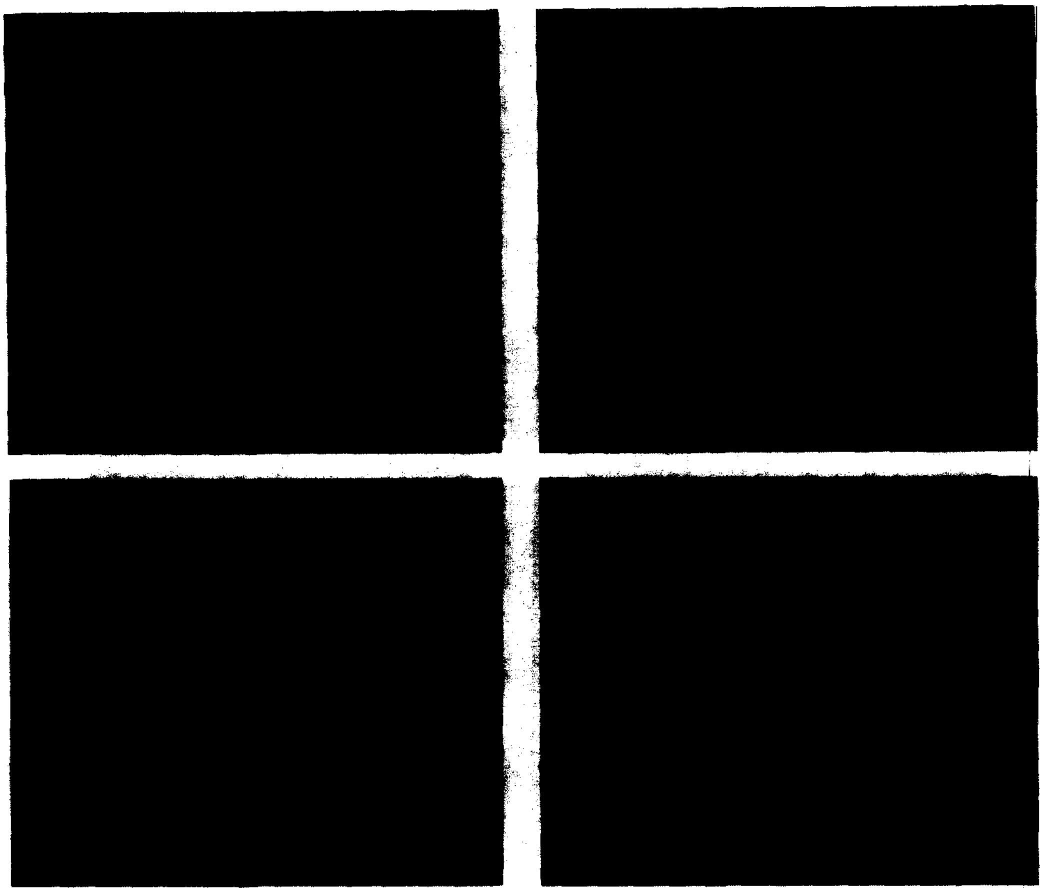

Fig. 2. Jet fragmentation at the nozzle exit (square duct test section).

$$D_D = D_{j,cr}\left[2.06\frac{u_{j,cr}}{u_n} - 1.47\ln\left(\frac{u_{j,cr}}{u_n}\right)\right] \tag{2}$$

where $u_{j,cr}$ is the critical jetting velocity as given by Steiner and Hartland [10]:

$$u_{j,cr} = 2.69\left(\frac{D_{j,cr}}{D_n}\right)^2\left[\frac{\sigma_{d,l}}{D_{j,cr}(0.514\rho_{d,l}+0.472\rho_c)}\right]^{0.5} \tag{3}$$

and $D_{j,cr}$ is also given by Skelland and Johnson [11]:

$$D_{j,cr} = \frac{D_n}{0.485E\ddot{o}_n+1} \quad \text{for } E\ddot{o}_n < 0.615 \tag{4}$$

$$D_{j,cr} = \frac{D_n}{1.51E\ddot{o}_n^{0.5}+0.12} \quad \text{for } E\ddot{o}_n \geqslant 0.615. \tag{5}$$

In the above equations (3)–(5) $\sigma_{d,l}$ is the interfacial tension between R 114 and water. As no data for R 114–water interfacial tension have been reported in published material, the surface tension of the dispersed phase has been used.

The bubble rising velocity is calculated from

$$u = (0.507gD)^{0.5} \tag{6}$$

proposed by Shimaoka and Mori [4]. The above correlation shows a fairly good agreement with rising velocities observed in present experiments.

Time-dependent expression of bubble volume may be obtained by equation (1) written as

$$h = 0.121\left(\frac{u\rho_c C_{p,c} K_c}{D}\right)^{0.5} \tag{7}$$

considering that

$$h = \frac{W}{\pi D^2 \Delta T_s^*} \tag{8}$$

h being the local heat transfer coefficient. From an energy balance written for the bubble, accounting for mass conservation and integrating, after also considering the steam inside the bubble and using equation (6) for the bubble rising velocity, we get

$$V(t) = (Ct + V_{oD}^{5/12})^{12/5} \tag{9}$$

where

$$C = \left(\frac{\pi}{6}\right)^{5/12}\frac{0.255}{h_{lg}}(\rho_c C_{p,c} K_c)^{0.5} g^{0.25} \Delta T_s^*(v_{d,v}^* - v_{d,l}). \tag{10}$$

The above equation (9) represents the time-depen-

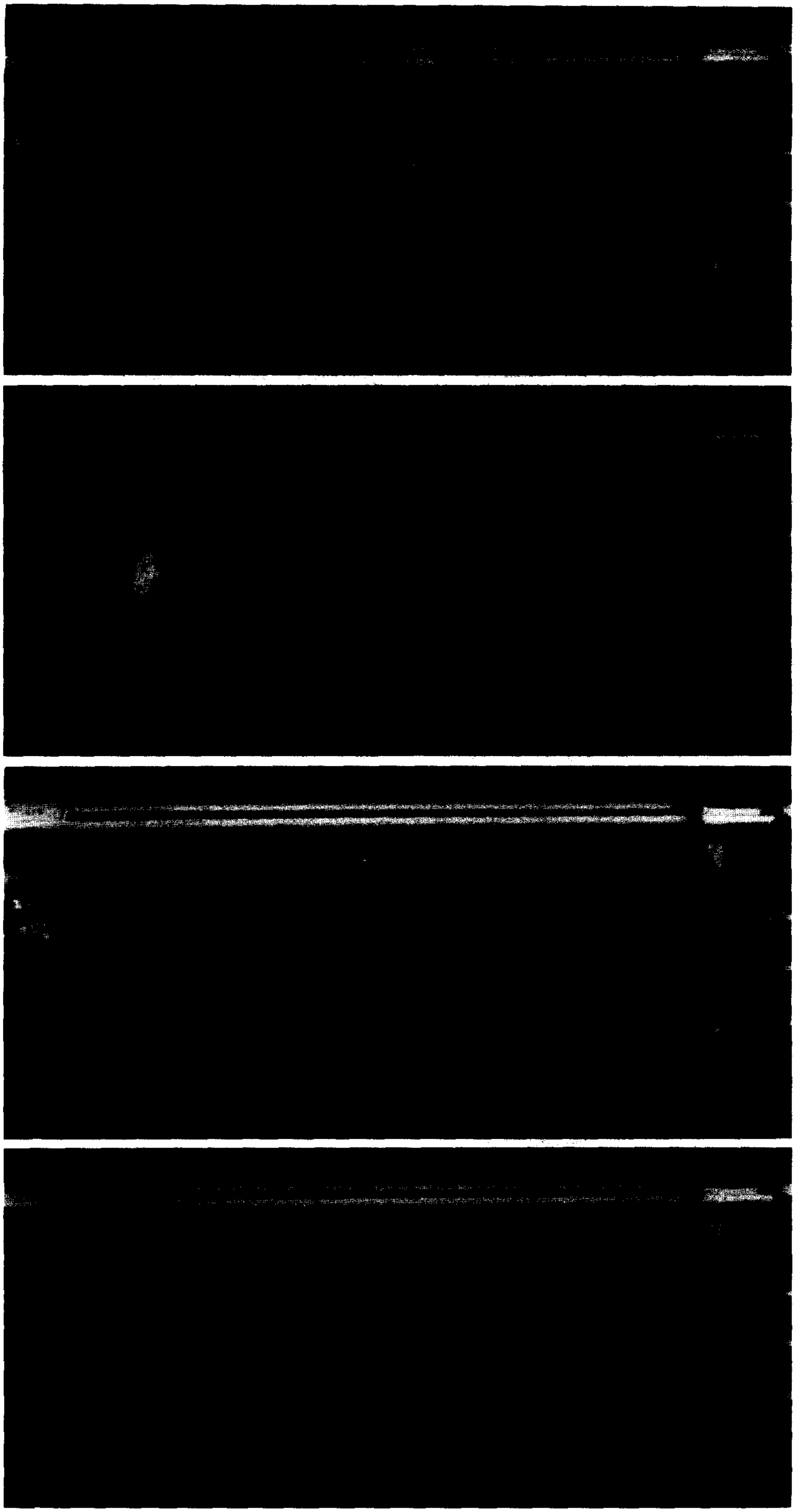

Fig. 3. Typical pictures of the experiment taken 1/7 s after each other.

dent expression of single drop evaporation. The total evaporation time of the drop, τ, is then given by

$$\tau = \frac{V_{\mathrm{oD}}^{5/12}}{C}\left[\left(\frac{v_{\mathrm{d,v}}^{*}}{v_{\mathrm{d,l}}}\right)^{5/12} - 1\right] \tag{11}$$

where $V_{\mathrm{oD}}\, v_{\mathrm{d,v}}^{*}/v_{\mathrm{d,l}}$ is the final volume occupied by the single drop of liquid R 114, V_{oD}.

Of course, in the actual situation, drops coming from jet fragmentation do not show the same thermal and dynamic behaviour as single drops injected separately. Typically, drop coalescence and drop falling down phenomena occur. These phenomena may lead to a reduction of drop evaporation, as calculated for isolated and single drops. In fact, coalescence of drops reduces the heat transfer surface, for a given volume, and drops falling down, not vaporizing, do not contribute to the expected volumetric variation due to the complete vaporization of the injected liquid.

Besides, it is necessary to account for the local cooling of the continuous medium, due to the passage of an evaporating drop, when describing the thermal behaviour of the following drops. Wherever, during the falling down, the drop impacts on the following evaporating drop, this may cause an increase of the liquid–liquid heat transfer surface and, consequently, an increase of the evaporation that, partially, counterbalances the effect of the drops falling down. Therefore, to take into account the above phenomena, generally tending to change the total evaporation, we introduce the jet factor, a, and the drop falling down factor, b. The jet factor is practically a correction for the heat transfer with respect to the single drop, while the drop falling down factor depends on the volume of the non-vaporized Freon (falling down in drops) and represents the fraction of the injected volume that evaporates. The above factors, a and b, are functions of the thermal hydraulic conditions, and of the water column height, wherever this is less than the evaporation length.

Equation (9), for the case of a drop coming from the jet fragmentation (actual case) becomes

$$V(t) = (aCt + V_{\mathrm{oD}}^{5/12})^{12/5} \tag{12}$$

where $a = 1$ for a single evaporating drop, and $a \neq 1$ for liquid jets, unless we have triggering effects. Similarly, in the case of liquid jets, eq. (11) for calculation of the total evaporation time gives

$$\tau_{\mathrm{j}} = \frac{V_{\mathrm{oD}}^{5/12}}{aC}\left[\left(\frac{v_{\mathrm{d,v}}^{*}}{v_{\mathrm{d,l}}}\right)^{5/12} - 1\right]. \tag{13}$$

In the present schematization, the volume of the injected Freon may be assumed as subdivided into a number of drops given by $V_{\mathrm{od,l}}/V_{\mathrm{o}}$.

Again for the local heat transfer coefficient and the thermal power exchanged, in the case of a jet we have, using equations (6) and (7),

$$h_{\mathrm{j}} = 0.102a(\rho_{\mathrm{c}} C_{\mathrm{p,c}} K_{\mathrm{c}})^{0.5} g^{1/4} D^{-1/4} \tag{14}$$

$$\bar{W}_{\mathrm{j}} = b h_{\mathrm{lg}} \Gamma. \tag{15}$$

The volumetric heat transfer coefficient, h_{v}, is defined as

$$h_{\mathrm{v}} = \frac{\bar{W}_{\mathrm{j}}}{\Delta T_{\mathrm{s}} H_{\mathrm{v}} A} \tag{16}$$

where H_{v} can be calculated by integrating equation (6) up to τ, making use of equation (12):

$$H_{\mathrm{v}} = \int_0^{\tau} u(t)\,\mathrm{d}t = \frac{5}{7}\sqrt{0.507gC^{2/5}\tau^{7/5}} \tag{17}$$

where τ is obtained from equation (13).

Once the volume of vapour generated as a function of time from a single drop is known, it is necessary to calculate the volume of the total vapour generated as a function of time, $\Delta V(t)$, considering the jet as consisting of liquid drops, continuously injected, that evaporate according to equation (12). The calculation of $\Delta V(t)$ is obtained from the integration up to time t of contributions to vapour generation due to the liquid volumes injected at times t^* between 0 and t_{i}. Such contributions will have a trend that follows equation (12) up to time $t^* + \tau_{\mathrm{j}}$ (substituting t with $t - t^*$), and then is kept constant, all the liquid being evaporated. The above integration is achieved by distinguishing four different times, $t = 0$; $t = t_{\mathrm{i}}$; $t = \tau_{\mathrm{j}}$; $t = \tau_{\mathrm{j}} + t_{\mathrm{i}}$, and two cases: $t_{\mathrm{i}} \leqslant \tau_{\mathrm{j}}$ and $t_{\mathrm{i}} > \tau_{\mathrm{j}}$. We obtain eight $\Delta V(t)$ functions depending on C, as defined in equation (10), D_{D} as calculated using equation (2), t_{i} and the two factors a and b described above. The above functions are used to get the parameters of interest to describe the jet direct contact evaporation, starting from the experimental time-dependent value of the total vapour volume. The latter parameter, *per se*, does not allow any direct analysis and, above all, does not provide useful information for the steady-state direct contact boiling.

The model above described, once initial conditions of the test are given, is a function of factors a and b only. Therefore a best-fit procedure for the experimental values of $\Delta V(t)$ allows the determination of a and b values. From these, using equation (13) it is possible to calculate the jet total evaporation time, while using equation (16) it is possible to evaluate the volumetric heat transfer coefficient.

Uncertainty of calculated values of a and b parameters may be evaluated on the basis of dummy variations of the most critical parameters in the best-fit procedure, in particular the initial injection time. Using such a procedure we obtained an uncertainty of $\pm 15\%$ on the jet factor a, and of $\pm 4\%$ on the drop falling down factor b. To the above uncertainties must be added the uncertainty related to the inlet subcooling of the dispersed phase. The latter is not present in data analysis, as inlet conditions are supposed to be saturated. However, it may affect the experimental behaviour, and may be estimated as ± 1 K.

Typical examples of the obtained values of factors a and b as well as τ_{j} are plotted in Figs. 4 and 5. Figure 4 shows the jet factor a (top figure), the drop falling

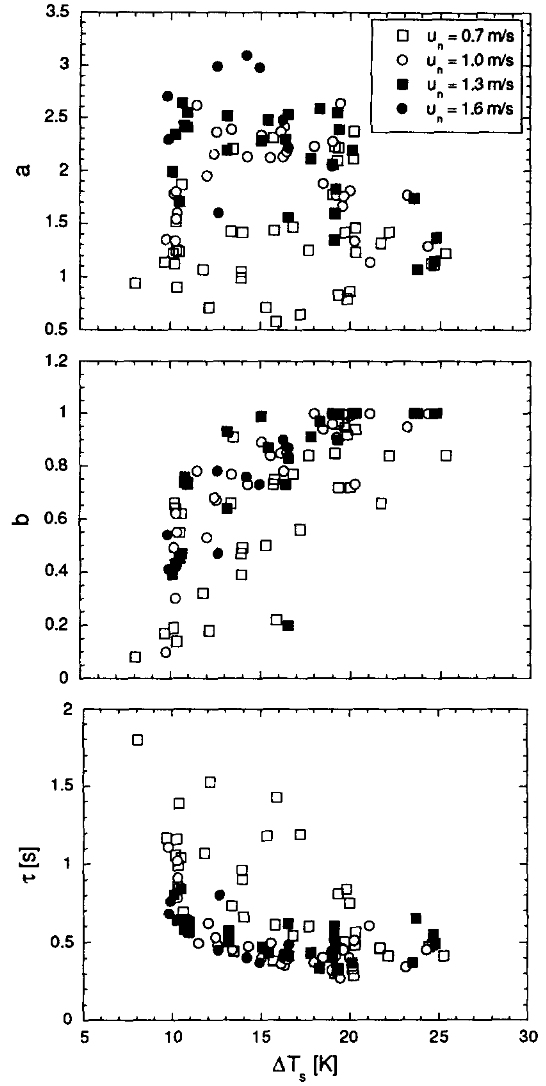

Fig. 4. Jet factor, a, drop falling down factor, b, and total evaporation time, τ_j, as functions of water superheating, ΔT_s.

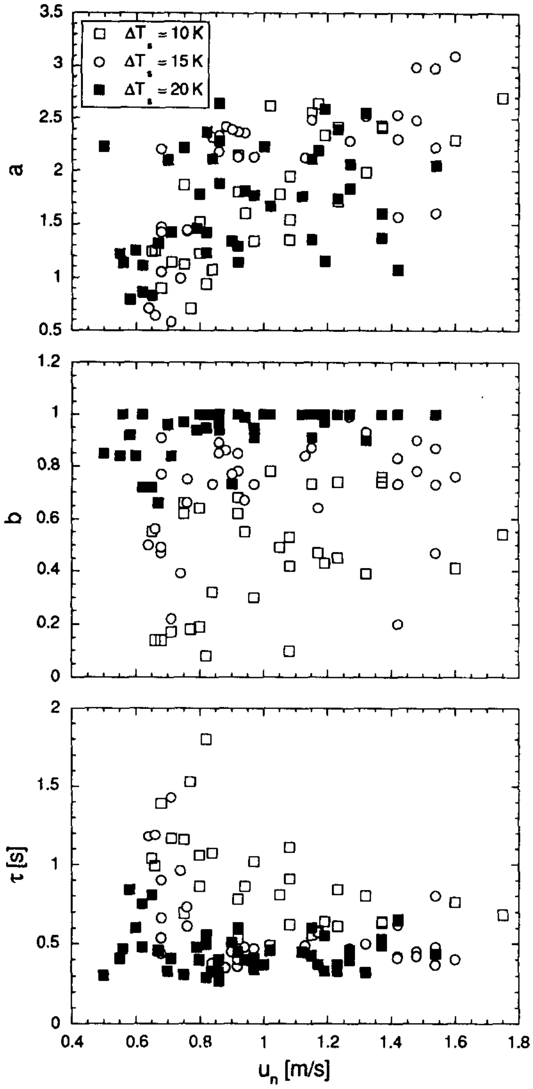

Fig. 5. Jet factor, a, drop falling down factor, b, and total evaporation time, τ_j, as functions of jet velocity at the nozzle, u_n.

down factor b (centre figure) and the total evaporation time τ_j (bottom figure) vs water superheating, ΔT_s, for high injection velocity. Figure 5 shows the same variables as functions of the injection velocity, u_n.

The jet factor a turns out to be independent of ΔT_s, while increasing as u_n increases, starting from values close to unity (single drop behaviour) for low velocities. As injection velocity increases, the heat transfer enhancement due to turbulence induced by the injection overcomes the effect of the local cooling of the water column due to the drop evaporation on the following drops.

The drop falling down factor, b, tends to unity (all the injected liquid evaporates and drop falling down vanishes) for ΔT_s greater than 20 K. For water superheating between 12 and 20 K it ranges between 0.4 and 0.8, while for $\Delta T_s < 12$ K factor b drops to a very low value. That means that for low values of water superheating ($<$ 10–12 K) most of the injected liquid falls down to the bottom plate of the test section without contributing to the evaporation process, at least from the viewpoint of direct contact boiling. It is expected that, on increasing the water superheating, i.e. the boiling driving force, the lack of nucleation sites becomes less and less important. As the absence of nucleation sites is the main reason for the drop falling down, the latter effect is also expected to vanish as water superheating increases.

From Fig. 5, we may notice that the drop falling down factor, b, also increases as injection velocity increases, as the higher initial velocity causes an enhancement of the evaporation (as already observed for factor a). Therefore, we have a greater upwards entrainment of the non-evaporated dispersed phase that, linked to the higher turbulence, i.e. higher probability of boiling occurrence, leads to a reduction of the falling down.

Here also the dependence of b on water super-

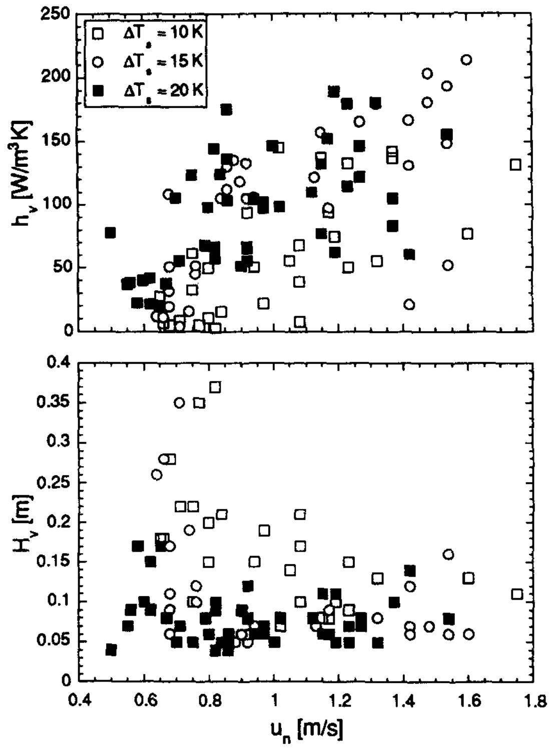

Fig. 6. Volumetric heat transfer coefficient, h_v, and evaporation length, H_v, trends as functions of injection velocity, u_n.

heating is clearly evident, together with the wide scattering of data points for the lowest value of water superheating.

The behaviour of the total evaporation time, is, of course, linked to the observed trends of factors a and b. For $\Delta T_s > 12$ K, the total evaporation time τ_j is more constant and tends to 0.3 s. As the water superheating decreases below 12 K, τ_j correspondingly increases. This could be due to the global reduction of the heat transfer as observed in the behaviour of factor a. Volumetric heat transfer coefficient, h_v, and evaporation length, H_v, trends as functions of injection velocity, u_n, are shown in Fig. 6. The former shows a trend similar to a, as also expected, considering that its variation with characteristic parameters of the single drop is very small with respect to that due to the enhancement of the jet factor. Evaporation length, instead, is related to the evaporation time, τ.

Data reduction can be obtained by using the following empirical correlations for factors a and b:

$$a = 1.94u_n \tag{18}$$

$$b = 1 - 3.28/\exp(0.156\Delta T_s + 0.29u_n). \tag{19}$$

A comparison of a, b and h_v with predictions obtained using equations (18) and (19) is plotted in Fig. 7. The error in h_v is greater, as h_v is a parameter derived from a and b. The standard deviation of ratios between calculated and experimental value is as follows:

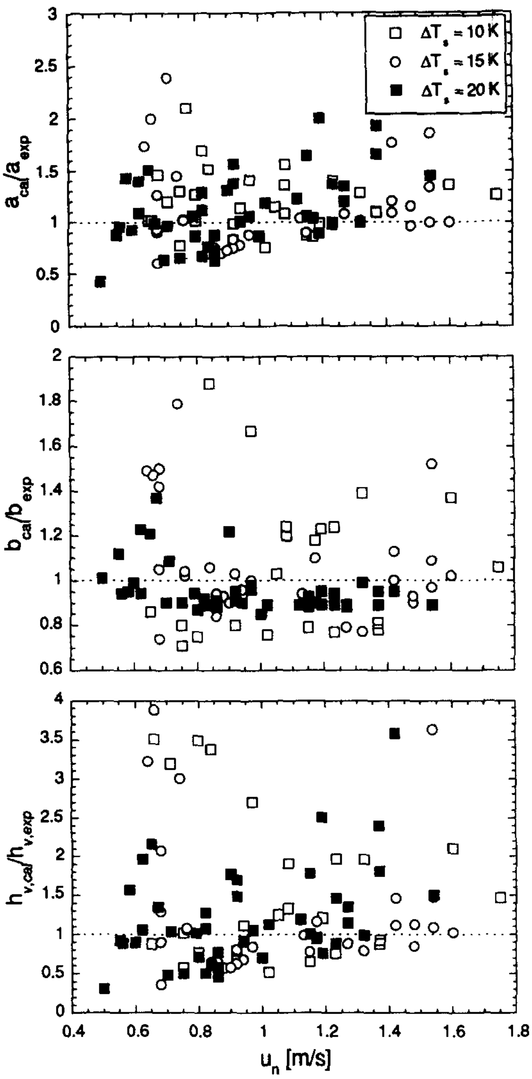

Fig. 7. Comparison of a, b and h_v with predictions obtained using equations (18) and (19).

Factor	a	b	τ [s]	h_v [W m^{-3} K^{-1}]	H_v [m]
σ	0.38	0.73	0.30	1.9	0.43

4. CONCLUDING REMARKS

Experimental research on direct contact evaporation of liquid jets of R 114 (volatile liquid) flowing upwards in a stagnant water column has been carried out with the particular aim of ascertaining the effect of water superheating (difference between water temperature and R 114 saturation temperature at system pressure) on the heat transfer process.

The experimental data are obtained by observing the level variation of the water column after the injection of a known volume of volatile liquid. The level variation, due to the evaporation of the injected liquid, has been correlated to the volatile liquid evaporation in the column using a model which, starting from the ideal situation of an isolated single drop evaporating as it rises in the water column, accounts for the actual phenomena such as the local cooling of the continuous medium, due to the passage of evaporating drops, and the drops falling down.

The evaporation of a liquid jet flowing upwards in a water column has been described using an isolated single drop model, providing correction factors determined from the ideal behaviour. In particular, the drops falling down are only important for water superheating less than 18–20 K. The local cooling of the water column due to the drop evaporation on the following drops is practically negligible when $\Delta T_s > 12$–15 K as turbulence induced by injection plays a more relevant role in the heat transfer process.

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