



## Technical Note

# A note on the influence of droplet interchange on evaporation and condensation of multicomponent mixtures in annular flow

J.R. Barbosa Jr.<sup>1</sup>, G.F. Hewitt<sup>\*</sup>, S.M. Richardson*Department of Chemical Engineering and Chemical Technology, Imperial College of Science, Technology and Medicine, Prince Consort Road, London SW7 2BY, UK*

Received 16 July 2001; received in revised form 30 October 2002

## 1. Introduction

Droplet entrainment and deposition are vigorous mass exchange phenomena which disrupt the hydrodynamic equilibrium of annular flow. Drops are created by the shearing action of the gas flow at the crest of large amplitude waves which travel on a liquid film attached to the wall. Any given droplet travel for a while in the gas core and eventually redeposit back on the film. Amongst many other effects, these phenomena are known to exert a high influence on important flow parameters like pressure drop.

The objective of the study, of which the work described in this paper was part, was to investigate the role of droplet interchange on the mechanisms by which heat is transferred to and from the two-phase mixture in convective boiling and condensation. The significance of droplet interchange seems to have been overlooked in previous multicomponent phase change studies and the correct prediction of this mechanism may hold the key to understanding the behaviour of the heat transfer coefficient at high qualities.

The heat transfer coefficient for mixture systems is defined by the following equation:

$$\alpha = \frac{\dot{q}_w}{T_w - T_E}, \quad (1)$$

where  $T_E$  is, for mixtures undergoing condensation, the dew point temperature. For boiling or evaporation,  $T_E$  is the bubble point temperature.

<sup>\*</sup> Corresponding author. Tel.: +0171-594-5562; fax: +0171-594-5564.

E-mail address: [g.hewitt@ic.ac.uk](mailto:g.hewitt@ic.ac.uk) (G.F. Hewitt).

<sup>1</sup> Present address: Departamento de Engenharia Mecânica, Universidade Federal de Santa Catarina, UFSC, 88040-900 Florianópolis, SC, Brazil.

In design calculations of phase change equipment, the two-phase flow pattern is usually ignored (Fig. 1a and b). A flash calculation is used to determine the amount of liquid (vapour) which has evaporated (condensed) for a given wall heat flux and to determine the bubble (dew) point temperature of the mixture. The preferential evaporation of the more volatile component gives rise to axial gradients of saturation temperature and of mean concentration in both phases.

In a real situation, however, where the annular flow pattern is the dominant configuration, not all of the liquid is present as a film coating the inner wall of the pipe. Rather, the core is a fairly homogeneous mixture of vapour and entrained droplets. Due to the distribution of the liquid phase within the pipe, the temperature driving force across the *liquid film* is much higher than that at the vicinity of a droplet. Therefore, in real annular flow, with a significant fraction of the liquid entrained as droplets, not all of the liquid plays a significant role in phase change. In mixture systems, this hydrodynamic effect changes the relationship between quality and interfacial saturation temperature, i.e., it can no longer be assumed that when  $X = 0$ ,  $T_{\text{sat}} = T_{\text{bub}}$  and when  $X = 1$ ,  $T_{\text{sat}} = T_{\text{dew}}$ .

If droplet evaporation (or condensation) is negligible compared with that of the film, then it is not illogical to assume that the droplets will retain a concentration equal to that of the liquid film at the point at which they were created. As an axial gradient of concentration is established in the liquid film, droplets generated at distinct co-ordinates will have distinct concentrations. When calculating the rate at which droplets are deposited on the film at a certain distance, one must take into account the difference in concentration between groups of droplets and the difference in concentration between the droplets and the liquid film.

### Nomenclature

$\alpha$	heat transfer coefficient ( $\text{W m}^{-2} \text{K}^{-1}$ )
$\dot{m}_T$	total mass flux ( $\text{kg m}^{-2} \text{s}^{-1}$ )
$p$	pressure (Pa)
$\dot{q}_w$	wall heat flux ( $\text{W m}^{-2}$ )
$T_{\text{bub}}$	bubble point temperature (K)
$T_{\text{dew}}$	dew point temperature (K)

$T_E$	equilibrium temperature (K)
$T_I$	interfacial temperature (K)
$T_{\text{sat}}$	saturation temperature (K)
$T_w$	wall temperature (K)
$X$	quality

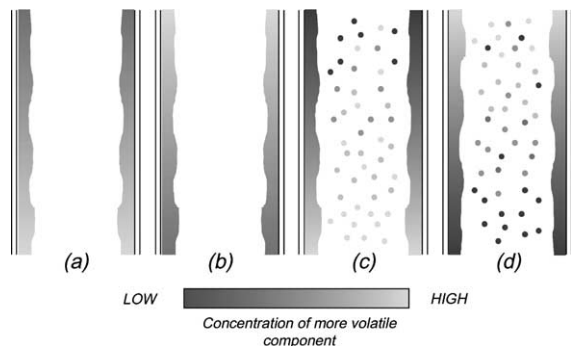


Fig. 1. An illustration of the problem of phase change in annular flow: (a) assumed in design (evaporation); (b) assumed in design (condensation); (c) evaporation and (d) condensation.

The implications of the mechanism described above for evaporation and condensation are as follows—see Fig. 1c and d. The film will have a different concentration of the less volatile component than the droplets compared to the ideal situation where entrainment and deposition do not occur. The concentration of the less volatile material in the film will be higher than the ideal in evaporation and lower in condensation. This effect will give, at a certain distance along the pipe, a local saturation temperature at the film interface,  $T_I$ , which is different to that which would be expected (i.e.,  $T_E$ ) in the absence of entrainment. Thus,  $T_I > T_E$  for evaporation and  $T_I < T_E$  for condensation. If the liquid physical properties remain the same, then the temperature driving force across the liquid film must remain unchanged to satisfy conservation of energy and this can only be achieved if there is a change in the wall temperature between the idealised (equilibrium) and real cases equal to  $T_I - T_E$  for evaporation and  $T_E - T_I$  for condensation.<sup>2</sup>

<sup>2</sup> This assertion is based on the assumption that the *liquid film* heat transfer coefficient is primarily a function of local turbulence and of physical properties. In theory, thermal energy transport by counter mass diffusion would also be of some influence. However, Shock [6] showed that for any mixture, irrespective of the width of its boiling range, the effect of diffusional mass transfer on the *liquid film* heat transfer coefficient is small enough to be ignored.

In summary, droplet interchange processes lead to a change in  $\alpha$  for multicomponent forced convective phase change heat transfer due to a definition of the heat transfer coefficient based on the temperature difference  $T_w - T_E$ ; when defined in this way, the coefficient for evaporation apparently decreases and that for condensation apparently increases compared with the ‘ideal’ value.

As far as mass transfer is concerned, gradients of concentration building up adjacent to both sides of the interface would tend to offer a further influence. Even though Shock [6] revealed that these effects are negligible on the liquid film side, they may be of some significance in the vapour phase. Shock analysed the evaporation of a water–ethanol mixture, with a boiling range of the order of 20 K. He could not evaluate precisely the significance of mixture effects as the differences in wall temperature due to such effects may be small compared with those due to limitations of his model for determination of the temperature drop across the film. Some of the systems studied here have boiling ranges at atmospheric pressure nearly 2.5 times greater than those analysed by Shock and mass transfer effects are likely to be much more significant.

In practice, the effect of droplet interchange on the condensation heat transfer may be more difficult to identify than in evaporation for two reasons. Firstly, typical boiling ranges of mixtures used in condensing systems (and therefore experiments) are much lower than some observed in boiling experiments (R-407C at 1.5 MPa has a boiling range of the order of 5 K). This makes the saturation temperature imbalance between the liquid film and droplets much more subtle. Secondly, in evaporation, diffusive mass transfer resistance and droplet interchange work in the same way, i.e., towards a decrease in the heat transfer coefficient.<sup>3</sup> In condensation, however, depending on the mixture used, the heat transfer coefficient deterioration due to mass transfer resistance may be sufficient to cancel out or overcome any increase associated with droplet interchange.

<sup>3</sup> The sole influence of droplet interchange effects on heat transfer was quantified in previous work [7] by assuming no mass transfer resistance in both phases.

## 2. Modelling and results

The physical model and its main hypotheses will not be repeated here. The reader is referred to previous work [1–3] for a complete description of the methodology. Here, it is worth mentioning that in the condensation case, only the macroconvective term of the correlation for the liquid film heat transfer coefficient is used (i.e., there is, of course, no nucleate boiling). Also, 0% and 10% were chosen as initial conditions for the initial entrained liquid fraction by mass at the onset of annular flow for condensation and evaporation, respectively. Finally, for ternary mixtures, the diffusive mass fluxes were calculated using an effective diffusivity method [8].

Comparisons are made with the experimental data obtained by Kandlbinder [4] for boiling and Cavallini et al. [5] for condensation. A detailed description of the experiments by Kandlbinder using hydrocarbon mixtures is given elsewhere [3].

Cavallini et al. conducted condensation experiments using R-407C (a non-azeotropic ternary mixture of R-32, R-125 and R134a, respectively 23%, 25% and 52% by weight) at pressures of the order of 17.0 bar in a 1 m long, horizontal stainless steel double-pipe test section. The inner diameter of the inner pipe was 8.0 mm and condensation of the test fluid was achieved by passing cold water through the outer annular gap. In the present study, care was taken to ensure that the conditions analysed fell in a region of a condensation flow pattern map [9] in which gravity effects (those resulting on a variation of the liquid film thickness around the periphery of the pipe) are expected to be less significant. Refrigerant and wall temperatures at the inlet and outlet of the test section were measured using thermocouples inserted at both the centre-line of the inner tube and at the tube wall. The average condensation heat transfer coefficient is given by the ratio between the wall heat flux (obtained from the axial temperature gradient on the coolant side) and the logarithmic mean temperature difference between the two-phase mixture dew point temperature and the pipe wall.

Figs. 2 and 3 illustrate typical variations of liquid phase (film and droplets) mean concentration for boiling and condensation, respectively. The effect of hydrodynamic non-equilibrium can be observed in both cases. In the boiling example, as the quality increases, the droplets become richer in the more volatile component than the liquid film. In condensation, the film gets richer in the more volatile components as the quality increases.

As mentioned before, the relative change in concentration for the R-407C case is much less pronounced. However, one can perform a hypothetical numerical experiment for condensation of the *n*-pentane/*iso*-octane mixture whose behaviour in evaporation is shown in Fig. 2. It can be observed (Fig. 4) that the difference in concentration between the liquid film and the droplets is

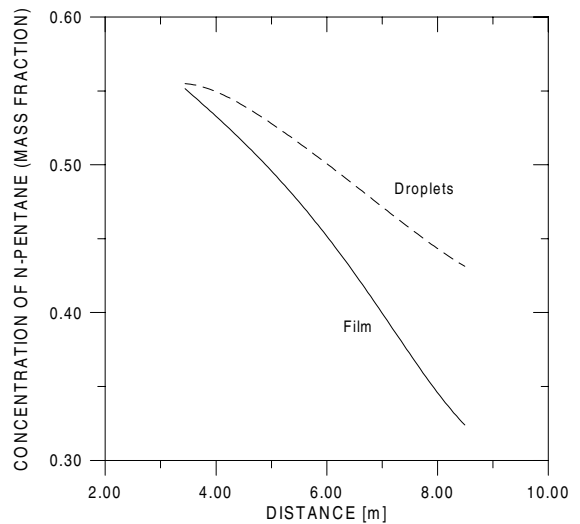


Fig. 2. Liquid film and droplet concentration in boiling. Mixture 0.7/0.3 (*n*-pentane/*iso*-octane),  $p = 0.23$  MPa,  $\dot{q}_w = 49.8$  kW m<sup>-2</sup>,  $\dot{m}_T = 292.8$  kg m<sup>-2</sup> s<sup>-1</sup>.

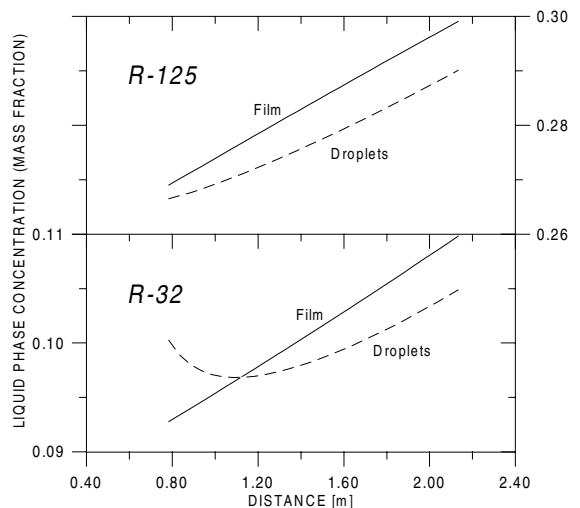


Fig. 3. Liquid film concentration in condensation. R-407C,  $p = 1.7$  MPa,  $\dot{q}_w = -33.7$  kW m<sup>-2</sup>,  $\dot{m}_T = 402$  kg m<sup>-2</sup> s<sup>-1</sup>.

now considerable, but not quite as high as in the boiling case under the same conditions. This is probably due to different directions of the mass transfer resistance on the vapour side (see above).

A typical set of local heat transfer coefficient predictions for the boiling data is presented in Fig. 5. Predictions using other methods, such as those by Kandlikar [10], Palen [11] and Sardesai et al. [12], are also shown. Of these, only the equilibrium resistance method by Sardesai et al. picks up the decreasing heat transfer

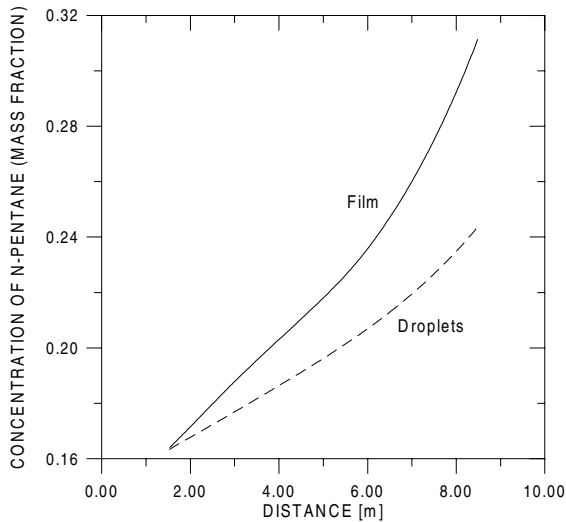


Fig. 4. Liquid film and droplet concentration in condensation (hypothetical case). Mixture 0.7/0.3 (*n*-pentane/*iso*-octane),  $p = 0.23$  MPa,  $\dot{q}_w = -49.8$  kW m<sup>-2</sup>,  $\dot{m}_T = 292.8$  kg m<sup>-2</sup> s<sup>-1</sup>.

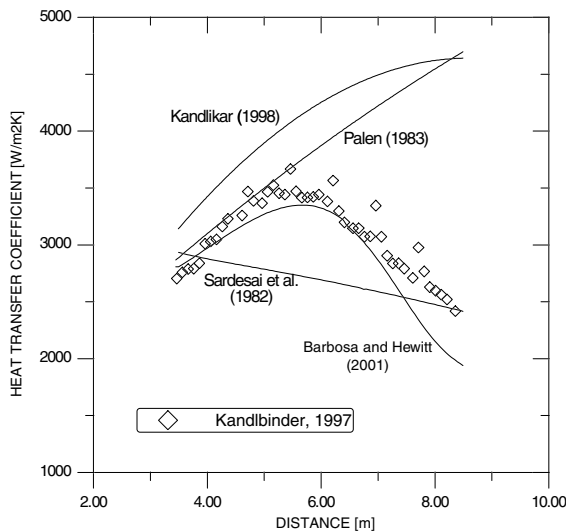


Fig. 5. Heat transfer coefficient predictions in boiling. Mixture 0.7/0.3 (*n*-pentane/*iso*-octane),  $p = 0.23$  MPa,  $\dot{q}_w = 49.8$  kW m<sup>-2</sup>,  $\dot{m}_T = 292.8$  kg m<sup>-2</sup> s<sup>-1</sup>.

coefficient trend. However, it underpredicts the data quite considerably.

Average heat transfer coefficients for condensation are shown in Fig. 6. These were calculated for three different total mass fluxes, namely 200, 400 and 750 kg/m<sup>2</sup> s. The agreement between experimental and predicted data is satisfactory, presenting deviations always lower than  $\pm 20\%$ .

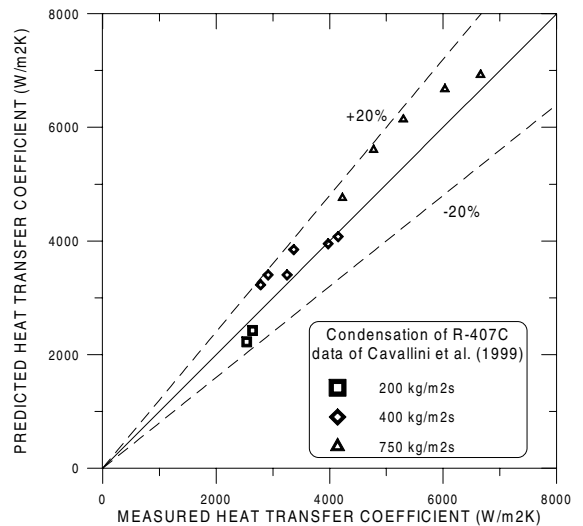


Fig. 6. Prediction of condensation heat transfer coefficients for R-407C.

### 3. Conclusions

This note presented an analysis of the influence of droplet interchange on phase change heat transfer to binary and ternary mixtures in annular flow. The main conclusions arising from this study are as follows:

1. the departure from hydrodynamic equilibrium caused by droplet interchange (entrainment and deposition) in multicomponent systems leads to an imbalance between the mean concentration of the entrained droplets and the liquid film. The analysis suggests that the heat transfer coefficient decreases with increasing quality for boiling/evaporating flows, whereas it increases with increasing quality for condensation;
2. the calculation methodology presented was successful at predicting both boiling and condensation heat transfer data;
3. a more detailed set of experiments, aimed at investigating the influence of droplet interchange on condensation heat transfer of multicomponent mixtures, is required in order to consolidate the analysis.

### Acknowledgements

The authors thank Dr. Luisa Rossetto (Università di Padova, Italy) for kindly providing the experimental data on condensation. JRB thanks the Brazilian National Research Council (CNPq—Conselho Nacional de Desenvolvimento Científico e Tecnológico) for the award of a scholarship (grant no. 200085/97-2).

## References

- [1] J.R. Barbosa Jr., G.F. Hewitt, Forced convective boiling of binary mixtures in annular flow. Part I. Liquid phase mass transport, *Int. J. Heat Mass Transfer* 44 (2001) 1465–1474.
- [2] J.R. Barbosa Jr., G.F. Hewitt, Forced convective boiling of binary mixtures in annular flow. Part II. Heat and Mass Transfer, *Int. J. Heat Mass Transfer* 44 (2001) 1475–1484.
- [3] J.R. Barbosa Jr., T. Kandlbinder, G.F. Hewitt, Forced convective boiling of ternary mixtures at high qualities, *Int. J. Heat Mass Transfer* 45 (2002) 2655–2665.
- [4] T. Kandlbinder, Experimental investigation of forced convective boiling of hydrocarbons and hydrocarbon mixtures, Ph.D. thesis, University of London, Imperial College, UK, 1997.
- [5] A. Cavallini, D. del Col, L. Doretti, G.A. Longo, L. Rossetto, Condensation of R-22 and R-407C inside a horizontal tube, 20th International Congress of Refrigeration, IIR/IIF, Sydney, Australia, 1999.
- [6] R.A.W. Shock, Evaporation of binary mixtures in upward annular flow, *Int. J. Multiphase Flow* 2 (1976) 411–433.
- [7] J.R. Barbosa Jr., G.F. Hewitt, Forced convective boiling of hydrocarbon mixtures, *Int. J. Heat Technology—Calore e Tecnologia* 17 (2) (1999) 37–47.
- [8] R. Taylor, R. Krishna, *Multicomponent Mass Transfer*, Wiley, New York, 1993.
- [9] T.N. Tandon, H.K. Varma, C.P. Gupta, A new flow regime map for condensation inside horizontal tubes, *J. Heat Transfer* 107 (1982) 763–768.
- [10] S.G. Kandlikar, Boiling heat transfer with binary mixtures: Part II—Flow boiling in plain tubes, *J. Heat Transfer* 120 (1998) 388–394.
- [11] J.W. Palen, Shell-and-tube reboilers: thermal design, in: G.F. Hewitt (Ed.), *Handbook of Heat Exchanger Design*, Begell House, New York, 1992, pp. 3.6.2-1–3.6.2-12.
- [12] R.G. Sardesai, R.A.W. Shock, D. Butterworth, Heat and mass transfer in multicomponent condensation and boiling, *Heat Transfer Eng.* 3 (3–4) (1982) 104–114.