

VERIFICATION OF MULTICOMPONENT MASS TRANSFER MODELS FOR CONDENSATION INSIDE A VERTICAL TUBE

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Abstract—This paper discusses and provides experimental evidence of the applicability of the following theories to multicomponent condensation.

- (1) The Krishna–Standart method (1976).
- (2) Toor's linearised theory (1964).
- (3) An effective diffusivity model based on Colburn & Hougen (1934).

It is found that all three models adequately quantify the vapour-side diffusional process for the experimental systems and conditions investigated but that multicomponent diffusional effects can be important. Criteria are discussed for the determination of the significance of diffusional interactions. The experimental data indicate that the analysis of Ackermann (1937) gives improved prediction of the sensible heat transfer from the bulk vapour phase and that the liquid side mass transfer resistance is not important under conditions studied. Computer simulation of the condenser predicts the condenser performance within an accuracy of ± 10 per cent.

INTRODUCTION

Physical models of the process of multicomponent condensation can be classified into two categories:

(1) Models such as those of Silver (1947) and Bell & Ghaly (1973) which assume that the multicomponent condensation process follows the condensation curve, an approach which is extensively used in industrial practice to design multicomponent condensers. The reasons for this widespread use are

- (i) Simplicity.
- (ii) Rapidity in computation.
- (iii) No requirement to know intermediate vapour compositions or to obtain diffusivity data.

(2) Models which are more physically realistic. Such models (film, penetration or boundary layer) lead to design procedures involving calculation of the local heat and mass transfer rates and their integration over the entire condenser area. The film theory of mass transfer forms the basis of the present study.

Multicomponent mass transfer has been described by an empirical generalisation of Fick's Law and by the Maxwell–Stefan equations. For isothermal, isobaric conditions of ideal vapours the MS equations are particularly useful. The generalisation of Fick's Law has been applied to film models in the linearised theory of Toor (1964) and Stewart & Prober (1964) while more recently an exact solution of the Maxwell–Stefan equations has been proposed by Krishna & Standart (1976). The linearised theory and the exact matrix method are commonly referred to as “interactive” models of mass transfer because they allow for “diffusional interactions”, where the diffusional behaviour of a constituent depends on all independent concentration gradients. A simpler approach to the application of film models in multicomponent mass transfer is by definition of effective diffusion coefficients for constituents with respect to the mixture as a whole, i.e. by assuming independent diffusion of the individual species. The mass transfer calculations are then effectively based on the approach of Colburn & Hougen (1934). This type of multicomponent mass transfer model is referred to as a “non-interactive” model because the interaction phenomena described by Toor (1957) cannot be accounted for.

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In this paper, a comparative study is made of "interactive" and "non interactive" models with experimental data for multicomponent condensation inside a vertical tube. The simultaneous sensible heat transfer rate is calculated using the analysis of Ackermann (1937). The experimental performance of the condenser is compared with that predicted to assess. (1) the relative validity of the three multicomponent diffusion models, (2) the significance of diffusional effects, and (3) the validity of the Ackermann analysis.

Criteria suggested by Webb & Panchal (1981) to determine the significance of diffusional interactions are assessed for the experimental conditions prevailing in the work.

MULTICOMPONENT MASS TRANSFER MODELS

The exact matrix and the linearised theories of multicomponent condensation make differing assumptions but the equations which arise for predicting diffusive fluxes are identical in form. Given the physical situation of a film model in which all concentration changes take place across a single film, the diffusive flux, J_{iG} , of the i th component in an n component mixture is given by Krishna & Standart (1976) as [1], but the result of Toor (1964) may be expressed in identical form.

$$(J_G) = [K][\Phi]\{\exp[\Phi] - {}^rI\}^{-1} (y_G - y_I). \quad [1]$$

In the above equation diffusive fluxes, and mole fraction driving forces, $(y_G - y_I)$, are column matrices of dimension $(n - 1)$: $[K]$ and $[\Phi]$ are squared matrices and rI is the diagonal matrix, all three of dimension $(n - 1) \times (n - 1)$. Subscripts G and I refer to quantities evaluated in the bulk gas phase, and at the condensate surface respectively.

In a binary mixture there is only one independent mole fraction driving force, $(y_{iG} - y_{iI})$ between bulk and interface and J_{iG} is readily obtained from Fick's Law as,

$$J_{iG} = k_{in} \frac{\phi_i}{\exp(\phi_i) - 1} \{y_{iG} - y_{iI}\} \quad i = 1, n - 1 \quad [2]$$

$$\phi_i = N_i/k_{in}. \quad [3]$$

The appropriate rate factor, ϕ_i , for binary mass transfer is the ratio of the total flux, N_i , to the mass transfer coefficient, k_{in} based on the binary diffusivity, D_{in} . (For a binary mixture $(n = 2)$, i takes the value 1.) Equation [2] may be written for a multicomponent mixture by defining effective mass transfer coefficients, (k_{in}^e) . In matrix form [2] becomes,

$$(J_G) = [k_n^e] {}^r \frac{\phi}{\exp \phi - 1} (y_G - y_I) \quad [4]$$

where $[\]$ indicate diagonal matrices of dimension $(n - 1)$. Equation [4] is a multicomponent generalisation of the approach of Colburn & Hougen (1934).

From the comparison of [1] and [4] it is clear that $[K]$ may be interpreted as a matrix of mass transfer coefficients and $[\Phi] \{\exp[\Phi] - [I]\}^{-1}$ is the multicomponent analogue of the high flux correction of [4]. It can be shown that [4] is the limiting form of [1] in the case of equal diffusivities of all binary pairs of constituents in the mixture, i.e. $\{k_{in}^e = k\}$ and in the limit of dilute mixtures of all condensable components in a non-condensable gas, $\{with k_{in}^e = k_{in}\}$. For ternary condensation in the presence of a non-condensable gas, Webb & Panchal (1981) have shown how to predict whether [4] is an adequate representation of the full equation [1].

The total flux of condensation, N_i , of a component is made up of a diffusive and convective contribution,

$$N_i = J_{iG} + y_{iG}N_i = J_{iI} + y_{iI}N_i. \quad [5]$$

The condensation fluxes, N_i are constant across a planar film but the diffusive fluxes, J_i , vary.

Subscript G indicates that a diffusive flux, J_{iG} , has been evaluated at the bulk vapour end of the diffusion film but it may also be evaluated at the interface, subscript I . Taylor & Webb (1980) have shown that there is a better choice of equations [5] to evaluate (N) using the method of Krishna & Standart (1976).

In using [1] to predict the diffusive fluxes by the "interactive" methods, it is convenient to define a matrix $[M]$ which is related to a column matrix (m) by

$$M_{ii} = \frac{m_i}{k_{in}} + \sum_{j=1, j \neq i}^n \frac{m_j}{k_{ij}}$$

$$M_{ij} = -m_i \left\{ \frac{1}{k_{ij}} - \frac{1}{k_{in}} \right\} \quad i = 1, \dots, n-1$$

$$j = 1, \dots, n-1. \quad [6]$$

The following table then shows how $[K]$ and $[\Phi]$ must be evaluated, by the methods of Toor (1964) and Krishna & Standart (1976) respectively. The binary mass transfer coefficients, k_{ij} , are

Table 1.

Method	Matrix $[M]$	Column matrix (m)
Toor (1964)	$[K]^{-1}$	$(m) = (y_G + y_I)/2.0$
	$[\Phi]$	$(m) = N_t (y_G + y_I)/2.0$
Krishna and	$[K]^{-1}$	$(m) = (y_G)$
Standart (1976)	$[\Phi]$	$(m) = (N)$

calculated from the corresponding heat transfer coefficients, h_G , by the Chilton-Colburn j_H analogy (1934) written in the form,

$$k_{ij} = \frac{h_G}{C_p \bar{M}_G} \left(\frac{D_{ij}}{\alpha} \right)^{2/3} \quad [7]$$

where \bar{M}_G , C_p and α are the molecular weight, molar specific heat capacity and thermal diffusivity respectively. In the present study the J_H factor was taken as

$$j_H = 0.039 \text{Re}^{-0.24}$$

where the coefficients were obtained as the best fit to extensive data obtained with binary mixtures of the constituents used in this study of multicomponent mass transfer.

The closure of the equations for condensation fluxes, N_i , is achieved by the continuity of energy flux, E , across the interface.

$$q_l = q_t + \sum_{i=1}^n N_i \Delta H_{li} = E \quad [8]$$

where subscript l refers to the condensate and the ΔH_{li} are the partial molar enthalpy differences at interfacial conditions,

$$q_l = h_G \epsilon \exp(\epsilon) (T_G - T_I) \{ \exp(\epsilon) - 1 \}$$

$$\epsilon = \sum_{i=1}^n N_i C_{pGi} / h_G > 0. \quad [9]$$

In [8] the dependence on ϵ (itself dependent on the molar specific heats, C_{pGi}) was introduced by Ackermann (1937) to account for sensible heat transfer brought about by the condensing species.

The solution of the above set of equations requires the knowledge of interfacial temperature, T_I , and interfacial vapour compositions, y_{iI} . The method of solution of the equations will also depend on the modelling of liquid side mass transfer processes.

THE INTERFACIAL CONDITIONS

The interfacial temperature, T_I , is not known *a priori* and has to be calculated iteratively by considering the continuity of energy at the interface. Hence T_I is evaluated such that the following equation is satisfied.

$$h_0(T_I - T_c) = q_i \quad [10]$$

where q_i is given by [7] and h_0 is the overall heat transfer coefficient of the condensate, tube wall and coolant. The condensate heat transfer coefficient may be calculated using the analysis of Nusselt (1916).

The interfacial vapour composition, y_{iI} , is obtained by assuming equilibrium at the vapour-liquid interface. Thus, the interfacial vapour composition is given in terms of the liquid composition, x_{iI} ,

$$y_{iI} = \frac{\gamma_i x_{iI} p_i^0}{p_i} \quad [11]$$

where p_i^0 is the vapour pressure of component i at interfacial temperature T_I , p_i the pressure and γ_i the activity coefficient of component i in the liquid.

THE LIQUID-SIDE MASS TRANSFER MODEL

The limiting values of the mass transfer resistance of the condensate can be calculated from simple models of the degree of mixing. These models allow the determination of the interfacial liquid composition, x_{iI} .

(i) If the condensate is assumed to be unmixed,

$$x_{iI} = N_i / N_t \quad [12]$$

(ii) If the condensate is assumed to be totally mixed,

$$x_{iI} = x_{ii} \quad [13]$$

At the vapour inlet, x_{iI} is necessarily given by [12] because there is no liquid present and the first drop formed will have a composition given by the ratio of fluxes, N_i / N_t . Downstream the interfacial liquid composition will lie somewhere between that given by [12] and [13]. Thus these models are the two limiting cases corresponding to infinite and zero mass transfer resistance respectively of the condensate.

In [12], N_i and N_t are not known *a priori* but are an outcome of the calculations. Hence an iterative procedure will be necessary to calculate x_{iI} using [12]. Since x_{ii} is known *a priori*, no iteration is involved if the condensate is assumed to be completely mixed. The effect of liquid mixing in condenser modelling is a feature that has not previously been investigated experimentally.

DOWNSTREAM DEVELOPMENT OF CONDITIONS

The equations for downstream development of conditions involve differential mass and energy balances to allow prediction of the variation of temperatures and flow rates along the

condenser length. The equations have been presented by Price & Bell (1974) and were given in their following form by Krishna & Panchal (1977).

A differential molar balance gives the rate of change of molar gas flow rate, G_G , with area, A , for each component i ,

$$\frac{dG_{iG}}{dA} = -N_i \quad i = 1, n. \tag{14}$$

For a non-condensing gas (say component n), N_n will be zero and G_{nG} constant.

The variation of the bulk vapour temperature, T_G , is described by a differential energy balance,

$$G_G C_{pG} \frac{dT_G}{dA} = -h_G \epsilon (T_G - T_I) / \{\exp(\epsilon) - 1\} \tag{15}$$

The variation of coolant temperature, T_c , is given by

$$G_C C_{pc} \frac{dT_c}{dA} = \pm q_i \tag{16}$$

where q_i is given by [7]. The sign of the r.h.s. of [16] depends on whether coolant and vapour flows are concurrent or countercurrent, and subscript c refers to coolant.

The condensate flow rate, G_i , and composition, x_{i1} , at any downstream location can be obtained by a material balance between the entrance of the condenser and that location.

The above differential equations of downstream development were solved using a fourth order Runge-Kutta method, with experimentally measured conditions providing the boundary conditions of the initial value problem.

EXPERIMENTAL PROCEDURE

A schematic diagram of the experimental apparatus used in this study to investigate multicomponent condensation of the vapours of miscible liquids in the presence of a non-condensing gas is given in figure 1.

The test condenser was a 1-m long, 0.023 m i.d. vertical tube with a surrounding water jacket, condensation taking place inside the tube. The tube wall was made of 3-mm thick copper with thermocouples embedded to measure the wall temperature profile along the length of the condenser. The vapours and the coolant were admitted at the top of the tube.

The non-condensing gas was circulated around a closed loop by a blower. The gas was heated before it was mixed with vapours generated in two boilers operating under identical conditions. The temperature of the vapour-gas mixture entering the condenser was adjusted to a value slightly above saturation by controlling the power input to the gas-heater.

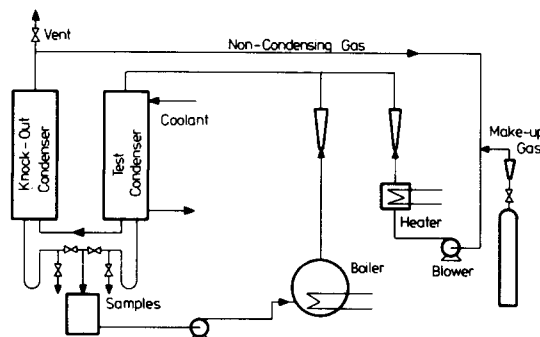


Figure 1. Experimental apparatus.

A fraction of the condensable vapours was condensed in the test condenser while the remainder (usually more than half) was condensed in a knock-out condenser. The condensates from both condensers were returned to the condensate reservoir via U-tube seals. The liquid from the reservoir was pumped back to the boilers at a constant rate by two peristaltic pumps. The pumping rate was adjusted so that the levels of the liquid in the boilers remained constant. The non-condensing gas leaving the knock-out condenser was recycled by the blower, so that the whole system worked as a closed loop.

At steady state, the gas-vapour mixture flowrate, its composition and temperature were measured both at the inlet and outlet of the condenser. Also measured were the condensate flowrate and composition at the outlet of the condenser, the coolant flowrate and temperatures at the inlet and outlet of the condenser and the wall temperature profile along the length of the condenser. These data are presented in table 2.

EXPERIMENTAL RESULTS AND DISCUSSION

Experiments were carried out with the Reynolds number at the condenser inlet lying in the range 6000–20,000 and with inlet compositions of condensable components ranging from 0.1 to 0.6 by mole fraction. A total of 15 runs were carried out widely distributed over the above range of feed conditions, for the following two mixtures

- (i) Water (1), isopropyl alcohol (2), nitrogen (3)
- (ii) Water (1), isopropyl alcohol (2), refrigerant-12 (3).

The vapour phase mass transfer behaviour of these two systems is characterised by the relative binary diffusivities of the pairs of components.

	D_{13}/D_{23}	D_{12}/D_{23}
System (i)	2.25	1.30
System (ii)	2.50	3.30

It has been pointed out by Webb & Panchal (1981) that the limiting values of the fractional error, ϵ'_+ and ϵ'_- , incurred by using the effective diffusivity (E) method rather than the Krishna–Standart (KS) method can be estimated as follows, the equation holding at any downstream location

$$\epsilon'_+, \epsilon'_- = (-1)^{i+1} \left\{ \frac{D_{13}}{D_{23}} - 1 \right\} y_{iG} / \left\{ y_{1G} + \frac{D_{13}}{D_{23}} y_{2G} + \frac{D_{12}}{D_{23}} y_{3G} \right\} \quad i = 1, 2 \quad [17]$$

where

$$\epsilon' = \frac{(N_t)_{KS} - (N_t)_E}{(N_t)_E} \quad [18]$$

and subscripts KS or E indicate the method used to estimate N_t .

Clearly the absolute value of ϵ' becomes large if $D_{13}/D_{23} \gg 1.0$ and the above systems have been chosen as having the possibility of large discrepancies between the interactive and non interactive models.

The experimentally measured inlet conditions were used to predict the performance of the condenser which involved the integration of [14]–[16] over the area. The predictions of condenser performance are used to assess the mass transfer models themselves, the two models of liquid mixing and the accuracy of prediction of rates of sensible heat transfer. These topics are considered in turn.

(i) Comparison of mass transfer models

A comparison of the experimental condensation rates with those predicted theoretically

Table 2. Experimental results
 CONDENSATION OF A MIXTURE OF ISOPROPYL ALCOHOL, WATER VAPOURS AND NITROGEN

RUN NO.	VAPOUR FLOW RATES AT INLET			RE. NO. AT INLET	VAP. INLET TEMP. K	VAP. EXIT TEMP. K	COOL. FLOW RATE kg/s $\times 10^2$	COOL. INLET TEMP. K	COOL. EXIT TEMP. K	CONDEN. SATION RATE kmol/s $\times 10^5$	CON-DEN SATION RATE COMP x_1	WALL TEMPERATURE AT LOCATION ζ						HEAT LOAD		
	I.P.A. kmol/s $\times 10^5$	WATER kmol/s $\times 10^5$	N ₂ kmol/s $\times 10^5$									0.025 K	0.167 K	0.333 K	0.5 K	0.667 K	0.833 K	1.0 K	GAS W $\times 10^4$	COOL ANT W $\times 10^4$
1	1.14	4.31	4.70	9790	378.1	350.7	3.90	301.9	310.1	2.87	.084	330.7	325.8	324.8	323.9	323.0	322.3	323.1	.137	.134
2	.62	4.19	3.43	7370	361.3	351.1	3.78	304.8	312.6	2.81	.041	330.4	327.9	326.4	325.0	324.1	323.3	324.0	.133	.124
3	.74	3.20	4.12	7420	380.0	340.8	3.94	306.7	312.8	2.10	.060	325.5	325.0	324.3	323.3	322.5	321.9	322.4	.102	.100
4	.71	3.23	3.16	6670	380.9	350.8	3.90	309.4	315.7	2.28	.068	328.9	328.0	327.1	325.9	325.2	324.4	324.9	.100	.103
5	.61	3.92	3.24	7100	376.5	347.2	3.94	299.8	307.8	2.96	.058	324.6	324.4	322.7	321.1	319.9	319.1	319.5	.139	.132
6	.74	3.91	4.55	8380	375.3	345.5	3.82	298.6	306.3	2.71	.058	321.1	321.9	320.7	319.7	318.5	318.0	318.4	.130	.124
7	.79	3.95	6.72	9440	373.7	344.4	3.86	298.6	306.1	2.53	.052	319.8	320.7	320.1	319.1	318.5	317.8	318.3	.123	.121
8	.99	3.87	5.64	9810	375.1	346.0	3.94	301.5	308.5	2.50	.070	320.7	322.3	321.9	321.0	320.3	319.6	320.1	.122	.114

TABLE 2 EXPERIMENTAL RESULTS (contd)
 CONDENSATION OF A MIXTURE OF ISOPROPYL ALCOHOL, WATER VAPOURS AND REFRIGERANT-12

RUN NO.	VAPOUR FLOW RATES AT INLET		RE. NO. AT INLET	VAP. INLET TEMP. K	VAP. EXIT TEMP. K	COOL. FLOW RATE kg/s $\times 10^2$	COOL. INLET TEMP. K	COOL. EXIT TEMP. K	CONDEN. RATE kmol/s $\times 10^5$	CONDEN. SAT. COMP $\times 10^5$	WALL TEMPERATURE AT LOCATION ζ						HEAT LOAD			
	I.P.A. kmol/s $\times 10^5$	WATER kmol/s $\times 10^5$									0.075 K	0.167 K	0.333 K	0.5 K	0.667 K	0.833 K	1.0 K	GAS W $\times 10^4$	COOL W $\times 10^4$	
9	0.75	3.16	2.12	14990	360.7	342.9	3.65	305.1	312.2	2.50	.084	325.2	326.0	324.8	323.5	322.5	321.7	322.2	.115	.109
10	0.67	3.18	1.53	11870	366.5	346.3	3.86	308.8	316.2	2.71	.086	328.7	329.3	327.9	326.7	325.2	324.4	325.1	.124	.120
11	0.73	3.18	2.62	17110	363.7	341.4	3.66	299.9	307.1	2.52	.086	320.7	321.7	320.8	319.4	318.3	317.6	317.8	.119	.110
12	1.13	2.51	2.59	17640	366.4	343.6	3.94	301.5	307.8	2.20	.207	318.3	320.6	320.3	319.1	318.2	317.6	317.7	.106	.104
13	1.01	2.68	1.94	14050	373.0	347.3	3.78	303.0	309.9	2.41	.167	322.5	323.5	322.3	321.4	320.3	319.6	319.8	.115	.110
14	0.90	2.60	1.31	10820	374.1	348.5	3.78	303.5	310.7	2.56	.178	324.7	325.0	323.6	322.0	320.7	319.7	319.9	.121	.114
15	0.94	2.67	3.34	20330	367.5	344.4	3.75	302.7	309.2	2.07	.124	320.1	321.7	321.6	320.4	319.8	319.2	319.5	.101	.102

provides the best assessment of the different vapour side mass transfer models. The liquid is assumed to be unmixed in all these comparisons which are summarised in figures 2 and 3. The comparison is made separately for each condensing vapour, water in figure 2 and isopropyl alcohol in figure 3, as the percentage deviation between the predicted rate of condensation, $(\bar{N}_i)_p$, and the experimental value, $(\bar{N}_i)_e$, with respect to $(\bar{N}_i)_e$ where the bar indicates values over the condenser as a whole. In assessing these results it must be remembered that the condensation rates of water range from 5 to 10 times greater than the condensation rates of the isopropyl alcohol. Thus conclusions reached on the basis of condensation rates of water apply also to total condensation rates. The diagrams show the predictions of all three models of the mass transfer process for both nitrogen and the refrigerant as the non condensable gas.

The models are compared first. It is apparent that the "interactive methods", the Krishna-Standart method and Toor's linearised theory, are in very close agreement with one another

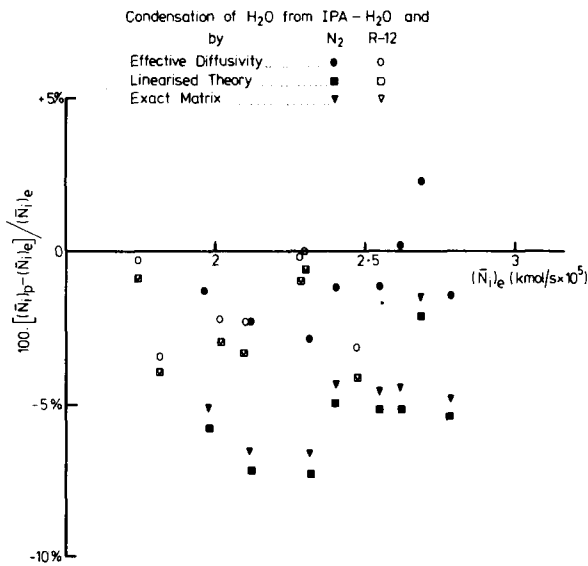


Figure 2. Comparison of methods.

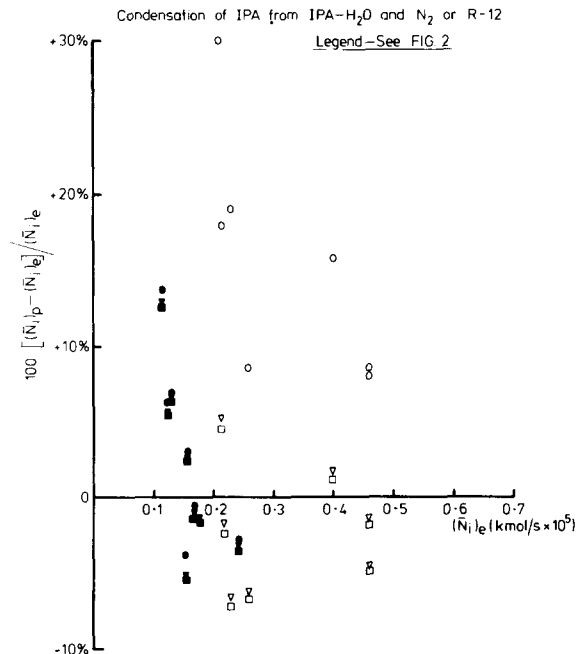


Figure 3. Comparison of methods.

always showing less than 2 per cent discrepancy in their predictions of the individuals rates of condensation of both water and isopropyl alcohol. The "non interactive" effective diffusivity method is in good agreement with the above two models in its predictions of condensation rates of water and hence total condensation rates (< 5 per cent discrepancy). However significant discrepancies (up to 25 per cent) are found between the predictions of the condensation rates of isopropyl alcohol by the "interactive" and "non interactive" methods, figure 3, when the refrigerant is the non-condensing gas.

The models are assessed relative to the experimental data on the basis of the standard deviations, \bar{S} , defined and reported in table 3.

Table 3.

$\bar{S} = 100 \sqrt{\frac{1}{(IK-1)} \sum_{k=1}^K \sum_{i=1}^I [(\bar{N}_i)_p - (\bar{N}_i)_e]^2}$	No of runs K = 15		
	Water & Ipa (I=2)	Water (I=1)	Total (i=2) (I=1)
Krishna and Standart method (1976)	4.9	4.2	4.1
Toor's Linearised Theory (1964)	5.0	4.7	4.4
Effective diffusivity method	9.1	2.0	1.7

The first column of table 3 assigns equal importance to each condensing component so that the values of \bar{S} presented are a very sensitive measure of the performance of the mass transfer model. The "interactive" models (which are equivalent) are somewhat better than the effective diffusivity method. This is occasioned entirely by the failure of the effective diffusivity method to predict accurately the condensation rates of isopropyl alcohol from freon (see figure 3). The second column of table 3 is for the condensation of water alone. In the case the effective diffusivity method is marginally better than the "interactive methods" though this is not regarded as significant. Finally the third column shows the comparison based on total fluxes ($\bar{N}_i = \sum \bar{N}_i$), which because of the much lower condensation rate of the alcohol is similar to column two based on the condensation of water alone.

There is no clear advantage in using any of the mass transfer models to predict total condensation rates in a partial condenser for the systems studied. This is significant in that these systems have been chosen as having the potential for large diffusional effects according to [17]. On the other hand if individual condensation rates are important there appears to be a significant advantage in using a model which accounts for diffusional interactions.

Attention is now turned to the relevance of [17] in assessing whether there will be significant differences between the predictions of an "interactive" and "non-interactive" model. The deviations, ϵ' , between the Krishna-Standart and effective diffusivity models defined by [18] were calculated at the vapour inlet for all 15 runs of the present study. The individual values of ϵ' varied from -0.037 to -0.079 with a mean of -0.063. The largest value of -0.079 was found for run 4 of the isopropyl alcohol-stream-nitrogen system. For this run the inlet vapour composition by mole fraction was,

$$y_{iG} = (0.100, 0.455, 0.455).$$

These values may be substituted directly into [17] with the stated ratios of binary diffusivities to give the upper and lower limits of the values of ϵ' (i.e. ϵ'_+ and ϵ'_- respectively). Thus

$$\epsilon'_+ = 0.07; \epsilon'_- = -0.33.$$

Clearly [17] is conservative in this case in quantifying the magnitude of diffusional interactions

and this is true for all the experimental runs of this study. If the values of ϵ' estimated by [17] are within acceptable limits then the effective diffusivity may be confidently used to estimate the total rate of condensation.

Equation [17] provides a local estimate of the limits of ϵ' and it is important to ask how its value will vary downstream of the vapour inlet. The dominant effect on ϵ' in [17] is due to y_{iG} in the numerator and hence in condensation from a non condensing gas there is a strong tendency for ϵ' to decrease downstream as the condensation proceeds. However the mole fraction of a minor condensing component may increase downstream and in addition the magnitude of the denominator may decrease. Both these effects occur in the isopropyl alcohol-water-nitrogen system. Using the Krishna–Standart predictions, the outlet vapour composition for run (4) is

$$y_{iG} = (0.114, 0.250, 0.636)$$

and hence

$$\epsilon'_+ = 0.10 \quad \epsilon'_- = -0.17.$$

Thus from [17] it is apparent that in some situations there may be an increase in the discrepancy between the predictions of “interactive” and “non interactive” models through a condenser. However the dominant effect of y_{iG} in the numerator must eventually assert itself and the discrepancy between models must reduce as total condensation is approached.

The discrepancy between the “interactive” and “non-interactive” models is apparent in figure 2 though of course the figure presents average discrepancies over the condenser as a whole (i.e. $\bar{\epsilon}'$) not local values of ϵ' at a particular position. For the fifteen runs of the study the predicted $\bar{\epsilon}'$ vary from -0.022 to -0.042 with a mean of -0.032 . These are smaller than the predicted inlet values (which showed a range from -0.037 to -0.079 with a mean of -0.063). Thus for the experimental conditions of the present study there is a smaller discrepancy in predictions of rates of total condensation between “interactive” and “non-interactive” models for the condenser as a whole than for local conditions at the inlet. This is consistent with the general behaviour predicted by [17]. The implication of these arguments is that if the effective diffusivity and Krishna–Standart methods show a reasonably small deviation, ϵ' , based on specified inlet conditions, their predictions for overall design will be in acceptable agreement.

(ii) *Liquid side models*

Two models of liquid side mass transfer were introduced, described by [12] and [13]. In the “unmixed” model the rate of mass transfer in the condensate film is assumed to be infinitely slow. New condensate is added to the surface and does not mix with the condensate produced earlier. In the “mixed” model the rate of mass transfer in the condensate film is assumed to be infinitely fast and the condensate composition is uniform. Actual condensate film behaviour will be somewhere between these limits and hence predictions based on these models should bracket observed behaviour.

The predictions of the two models are presented as figure 4. Clearly there is not a significant difference between the two models, both of which predict the individual condensation rate of both species within 15 per cent with a maximum discrepancy between models of about 10 per cent for the rate of condensation of the minor condensing species.

It is apparent that the “mixed” model predicts a higher condensation rate of iso-propyl alcohol and this is consistent with expected physical behaviour. The surface composition of ipa will be lower in the “mixed” model because initially the less volatile constituent, water, will dominate the condensation process producing an upstream condensate richer in water. Hence at downstream locations, as the relative condensation rate of the alcohol increases, the “mixed” model which disperses the fresh condensate through the entire film shows a lower surface concentration than the “unmixed” model where fresh condensate gathers at the surface.

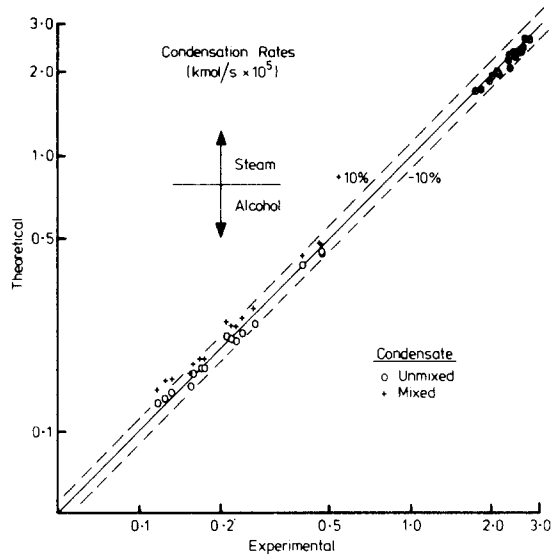


Figure 4. Comparison of liquid models.

Thus for the conditions investigated in this study liquid condensate mass transfer resistance was not important.

(iii) Sensible heat transfer

The sensible heat transferred from the gas phase during condensation is reflected in the change of temperature of the gas-vapour mixture. The modelling of the process of sensible heat transfer may therefore be evaluated by comparing the experimental gas temperature drops with those predicted theoretically by [15]. It will be seen that the Ackermann correction factor, $\epsilon/(\exp(\epsilon) - 1)$ has been included to allow for sensible heat transfer arising from the simultaneous mass transfer process. The importance of this factor is assessed by predicting the temperature drop in the gas phase in the limiting case of $\epsilon = 0$ when the r.h.s. of [15] simplifies to $-h_G(T_G - T_I)$. The heat transfer coefficient, h_G , has been calculated by the j_H correlation of Chilton-Colburn applicable for a single phase flow and no allowance has been made for any enhancement of heat transfer in the gas phase due to the presence of the liquid condensate.

The comparison is shown in figure 5. Equation [15] allows the prediction of gas temperature drops within ± 5 per cent for all the experimental data of this study. Omission of the Ackermann correction factor has an adverse effect on the accuracy of the predictions. In this case the temperature drop of the gas is consistently overestimated by between 5 and 20 per cent. The excellent results obtained with the use of [15] seem to indicate that enhancement of the gas film heat transfer coefficient by effects arising from the presence of a liquid condensate are not important.

CONCLUSIONS

A comparative study of multicomponent diffusion models has shown that, for the experimental conditions and systems investigated, both "interactive" and "non-interactive" models adequately predict the total rate of condensation (within 5 per cent). However if individual rates are to be calculated evidence is given showing that the "interactive" models are to be preferred. The two "interactive" models considered (Toor 1964, Krishna & Standart 1976) are equivalent for the conditions of this study.

A criterion suggested for determining *a priori* the significance of multicomponent diffusional effects is assessed on the basis of the experimental data and is shown to be conservative.

The modelling of the liquid side mass transfer process is found to be unimportant under the experimental conditions investigated.

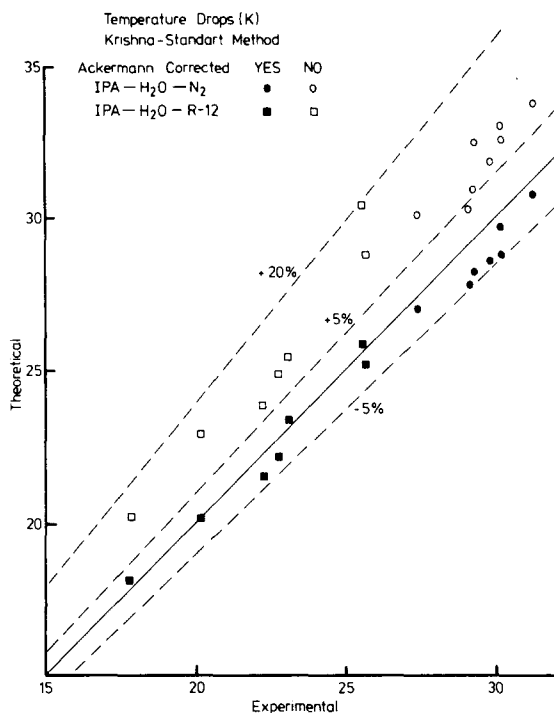


Figure 5. Temperature drops.

The Ackermann correction factor for sensible heat transfer greatly improves the accuracy of predictions of temperature drop through the condenser.

A computer simulation of the behaviour of the test condenser based on the Krishna & Standart model and including the Ackermann correction factor has allowed the major quantities of interest (rates of condensation of individual species, heat load and temperature drop) to be predicted within ± 10 per cent.

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