

DIRECT-CONTACT MASS AND HEAT TRANSFER BETWEEN VAPOR AND LIQUID WITH CHANGE OF PHASE*

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Experimental study was carried out on direct-contact mass and heat transfer between vapor and liquid with change of phase, that is, simultaneous condensation of organic vapor and evaporation of water or evaporation of water alone in a bubble column equipped with gas dispersion nozzles. The systems used in this experiment were carbon tetrachloride-water and trichloroethylene-water, which were immiscible in the liquid phase.

Volumetric coefficients of vapor phase mass transfer were measured and compared with previous work.

From the experimental results, condensation and evaporation rates were calculated under the given operating conditions and the effects of these specified operating conditions on these rates were made clear.

Introduction

This investigation was undertaken to obtain the basic mechanism of direct-contact mass and heat transfer processes between vapor and liquid with change of phase. Sidemann and his co-workers^{9~11)} and Prakash and Pinder^{7,8)} carried out a series of experimental studies of direct-contact heat transfer between vapor and liquid with change of phase, related to water desalination processes and three-phase heat exchangers. In these studies, pentane, isopentane, furan and cyclopentane were used as dispersed phase, and evaporation of water or mass transfer in the vapor phase was neglected because the vapor pressure of water at the experimental temperature was very small. However, if the organic vapor, of relatively high boiling point, is contacted with hot water, vapor pressure of the water is comparable with that of organic solvent, and vapor-phase mass transfer should not be ignored. This complicated process is generally described by the mass and heat transfer in vapor and liquid phase. To simplify the problem, systems of two immiscible liquids were selected in this work so that mass transfer did not occur in the liquid phase.

On the typical phase diagram of the system as described above, shown in **Fig. 1**, $E(y_E, t_E)$ is the unique condition in which mixed vapor and two liquid phases can co-exist. Assuming equilibrium at the vapor-liquid interface, and condensation and evaporation occurring simultaneously (case A), one obtains the conditions of interface (y_i, t_i) as

$$y_i = y_E \quad (1)$$

$$t_i = t_E \quad (2)$$

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In this case, calculation of condensation rate $-dN^o$ and evaporation rate dN^w , equivalent with estimation of total mass transfer rate dN and z , defined by Eqs. (3) and (4), respectively, is the problem to be solved.

$$dN = dN^o + dN^w \quad (3)$$

$$z = dN^o/dN \quad (4)$$

On the other hand, when water temperature is much higher than t_E , organic vapor can not condense. Evaporation of water alone results (case B), and total mass transfer rate dN , or evaporation rate dN^w , and interfacial conditions (y_i, t_i) located on curve AE in **Fig. 1**, are to be estimated as general evaporation processes,

$$dN = dN^w \quad (3')$$

$$z = 0 \quad (4')$$

In this work, organic vapor was contacted with hot water in a bubble column and volumetric coefficients of vapor-phase mass transfer were obtained from the measured condensation and evaporation rates. The effects of the operating conditions on these rates were made clear.

Apparatus and Procedure

The apparatus used in this investigation is shown in **Fig. 2**: it is essentially a bubble column with gas dispersion nozzles, in which pre-heated water was contacted with vapor of an organic solvent immiscible in water. The inlet vapor partially condensed, while a little of the inlet water evaporated simultaneously. The inlet water was pre-heated to desired temperature, to prevent total condensation of vapor, by an electric immersion heater (PH). After removal of dissolved air (AS), water was introduced to the top of the aerated water pool through the inlet. Contacting organic

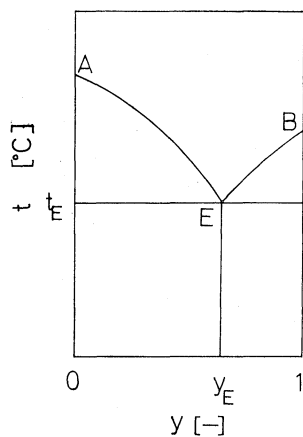
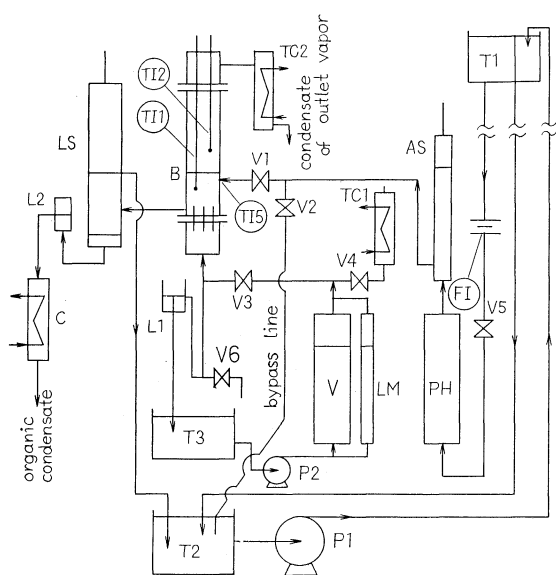


Fig. 1 Phase diagram



- B: Bubble Column
 PH: Pre-heater
 V: Vaporizer
 TC1, TC2: Total Condenser
 C: Cooler
 L1, L2: Level Controller
 LM: Level Meter
 LS: Level Controller and Settler
 AS: Air Separator
 T1-T3: Tank
 P1, P2: Pump
 FI: Flow Indicator
 TI1-TI2: Temperature Indicator
 V1-V6: Valve

Fig. 2 Experimental apparatus

vapor, water dispersed the organic condensate and this dispersion was withdrawn from the bottom of the water pool and settled gravitationally to two phases in the level controller (LS): water phase was overflowed from the top of LS to the storage tank (T2) and recirculated by pump (P1) through the constant level tank (T1), while organic liquid was underflowed from the bottom of LS and then overflowed at the second level controller (L2) into a mess cylinder through the cooler (C).

Organic solvent was vaporized in the vaporizer (V) and was supplied to the water pool through nozzles, after draining (L1) of organic condensate caused by

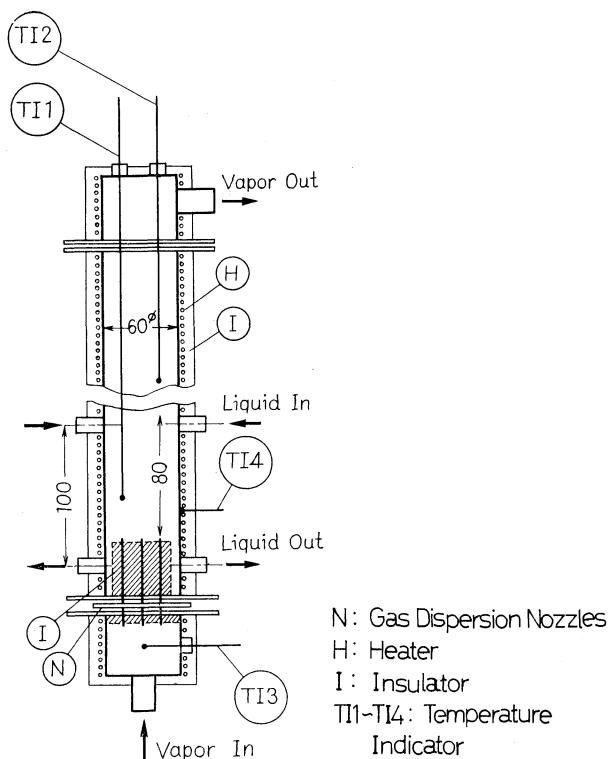


Fig. 3-a Details of the bubble column

$$D_0 = 0.5, 1.0, 1.5 \text{ mm.}$$

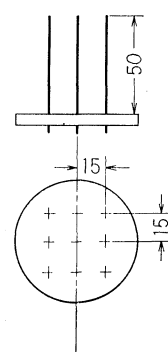


Fig. 3-b Gas dispersion nozzle

heat loss in the pipeline. To keep a certain constant vaporization rate, the liquid level in V, indicated by the level meter (LM), was controlled at constant height by adjusting feed rate of constant feed pump (P2). Exit mixed vapor of organic solvent and water was withdrawn from the top of the column and condensed in the total condenser (TC2).

Details of the bubble column are shown in Fig. 3-a. The column was made of brass tube of 60mm inside diameter and the outside nichrome wire heater was sandwiched with a layer of asbestos as thermal insulator. The temperature of the outside wall surface indicated by TI4 was adjusted to the liquid temperature in the column for prevention of heat loss or heat gain, by regulating the heater. Temperatures of liquid, vapor and outside wall surface were measured by

copper-constantan thermo-couples (diameter of wire = 0.3mm).

A typical example of the axial distribution of liquid temperature in the column measured by sliding T11 is shown in Fig. 4. The liquid in the column was found to be completely mixed, and regarding the reading of T11, located slightly above the liquid exit, as exit liquid temperature did not cause serious error.

To make the organic vapor disperse into the water layer directly (not into the condensate layer), the upper part of the nozzles was adjusted to a level slightly higher than the liquid exit. Nozzle diameters of 0.5, 1.0 and 1.5 mm I.D. were used, and nine nozzles were arranged at the corners of squares of 15 mm spacing, as shown in Fig. 3-b.

Experimental procedure was as follows. In the beginning, valve V3 was closed and valve V4 was opened. Then water was circulated through the system and heated to the desired temperature by PH. Organic vapor from V was recycled through the total condenser (TC1). When the temperature of circulating water was raised to the desired value, and vaporization rate became stable, then the circulating water was bypassed through valve V2 and water in the column was drained from valve V6. Organic vapor was introduced into the empty column through valve V3 after closing valve V4. Water was supplied slowly to the column, so as not to flow into the nozzles. Water and vapor were circulated until thermal and hydraulic steady state were obtained. Inlet and outlet temperature of liquid and vapor, flow rate of water, condensation rates at the cooler(C) and at the total condenser (TC2), were all measured under steady conditions.

The systems used in this investigation were carbon tetrachloride-water and trichloroethylene-water.

Experimental conditions were $L_T^w = 8 \sim 25$ kg/hr, $V_T^v = 1 \sim 6$ kg/hr, $T_i = 53 \sim 83^\circ\text{C}$ and $H_c = 8 \sim 10$ cm.

Experimental Results

Calculation of volumetric coefficient of vapor-phase mass transfer

Based on a section of water pool of differential height (dH), the following equations were obtained from material and enthalpy balances.

$$dV = dN \quad (5)$$

$$d(V \cdot y) = z \cdot dN \quad (6)$$

$$-dQ = V\{y \cdot di^o + (1-y) \cdot di^w\} + dN\{z \cdot i^o + (1-z)i^w\} \quad (7)$$

In Eq.(7), the first term of the right side is the enthalpy change due to change of vapor temperature, and the second term is the latent heat transferred by vapor-phase mass transfer. These are equated with the heat transfer rate in the liquid phase, $-dQ$.

Colburn and Drew's counter diffusion equation⁴⁾ is applicable to vapor-phase mass transfer,

$$dN = k_G a \cdot S \cdot dH \cdot \ln \frac{z-y}{z-y_i} \quad (8)$$

Introducing the heat transfer coefficients,

$$-dQ = h_L a \cdot S \cdot dH (T - t_i)$$

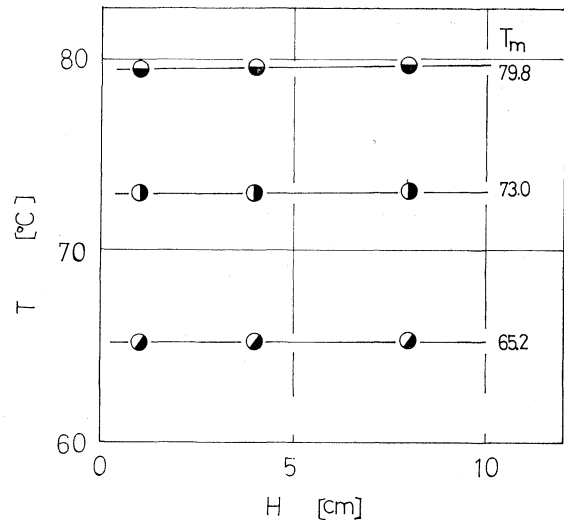


Fig. 4 Axial distribution of liquid temperature in the column

$$= h_L a \cdot S \cdot dH \cdot \Delta t_L \quad (9)$$

and

$$V\{y \cdot di^o + (1-y)di^w\} = h_G a \cdot S \cdot dH (t_i - t) = h_G a \cdot S \cdot dH \cdot \Delta t_G \quad (10)$$

Substituting Eqs.(8), (9) and (10) into Eq.(7) and dividing by $a \cdot S \cdot dH$,

$$h_L \Delta t_L = h_G \Delta t_G + k_G \ln \frac{z-y}{z-y_i} \{z \cdot i^o + (1-z)i^w\} \quad (11)$$

one can estimate z by this equation, if the transfer coefficients are given. Eq.(11) is rearranged,

$$z = \frac{y - \alpha y_i}{1 - \alpha} \quad (12)$$

where

$$\alpha = \exp \frac{h_L \Delta t_L - h_G \Delta t_G}{k_G \{z \cdot i^o + (1-z)i^w\}} \quad (13)$$

From Eqs.(5), (6) and (12),

$$\frac{dV}{V} = \frac{1-\alpha}{\alpha} \cdot \frac{dy}{y-y_i} \quad (14)$$

If the mean molar enthalpy of transferred substance, $z \cdot i^o + (1-z)i^w$, and the transfer coefficients are assumed to be constant in the column, Eq.(14) is integrated,

$$\ln \frac{V_I}{V_o} = \frac{1-\alpha}{\alpha} \ln \frac{y_I - y_i}{y_o - y_i} \quad (15)$$

When the first term of right side in Eq.(7) is negligible in comparison with the second term,

$$\alpha = \exp \frac{N}{k_G a S H} \quad (16)$$

From Eqs.(15) and (16),

$$k_G a = \frac{V_I - V_o}{S H \cdot \ln \left\{ 1 + \frac{\ln V_I / V_o}{\ln \frac{y_I - y_i}{y_o - y_i}} \right\}} \quad (17)$$

Eq.(17) was used for the calculation of $k_G a$ from the experimental data, in which the sensible heat terms were less than 15% of total heat transfer rates.

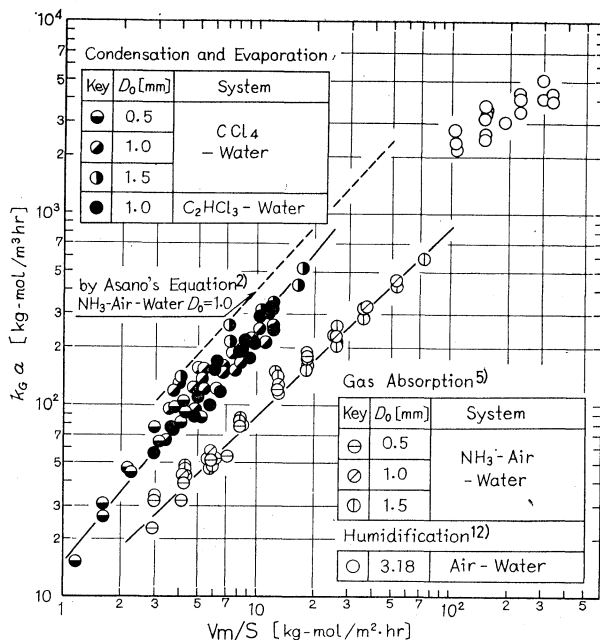


Fig. 5 Volumetric coefficients of vapor phase mass transfer

Results of calculation

When condensation and evaporation occurred simultaneously, $k_G a$ was calculated directly by Eq.(17) from experimental data, as y_i was given in Eq.(1). In the case of evaporation of water alone, interfacial temperature t_i was estimated from measured Q and T_o by Eq.(18) and y_i was calculated from this t_i , by using vapor pressure-temperature relationship of water.

$$Q = h_L a \cdot SH (t_i - T_o) \quad (18)$$

where liquid-phase heat transfer coefficient h_L was estimated from Calderbank's equation, Eq.(19)^{3)*}, and the contacting area was estimated from Abdel-Aal's equation, Eq.(20)¹⁾, which was obtained from the experiments by a light transmittance method in the bubble column with perforated plate (aerated liquid height was 15~90 cm).

$$h_L Pr^{1/2} = 0.42 (\Delta \rho \mu_L g / \rho_L^2)^{1/3} \quad (19)$$

$$aS/D_0 = 1.897 (U_0^2/gD_0)^{0.894} (D_0 U_0^2 \rho_L / \sigma g_C)^{1.408} (D_0 U_0 \rho_L / \mu_L)^{-0.488} N_N^{3.991} (D_0/\sqrt{q})^{0.963} \quad (20)$$

where the last factor of Eq.(20) is dimensional, $D_0 = ft$, $q = ft^3/sec$.

The experimental results are shown in Fig. 5 with the results of previous work in gas-phase mass transfer. The abscissa is mean molar mass velocity V_m/S . There is no remarkable difference between the two systems used in these experiments. Data of Furukawa⁵⁾ were obtained by the absorption of ammonia from air in water in a bubble column with perforated plate. The aerated liquid height was 25~30 cm and the number of perforated holes was 37. The ef-

* For the case A, as $t_i = t_E$, $h_L a$ can be calculated from Eq.(18). But in this experiment, Δt_L was so small (less than 1°C) that $h_L a$ was not calculated accurately, and was not correlated.

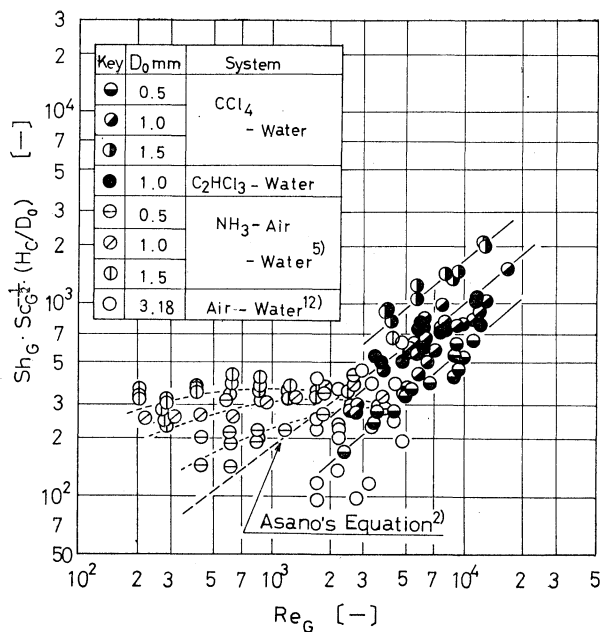


Fig. 6 Vapor phase mass transfer coefficients

fect of nozzle diameter D_0 on $k_G a$ was not appreciable in the authors' and Furukawa's results. Data of West¹²⁾ were obtained from humidification of air in a perforated-plate column. The clear liquid height was 1.5~5 cm and the number of holes was 83. The broken line was calculated from Asano's equation, Eq.(21)²⁾, obtained by the absorption of ammonia from air in water in a perforated-plate column, assuming the same dimensions as the column used in this experiment.

$$k_G D_0 / D_{AB} = 1.0 \cdot Sc^{0.5} (U_0 D_0 \rho_G / \mu_G)^{0.75} (H_C / D_0) \quad (21)$$

By the contacting area equation, Eq.(20), k_G was calculated from $k_G a$ and results are shown in Fig. 6 for comparison with Asano's equation. In the calculation in this work, the diffusivity was estimated by Fujita's equation, and other physical properties were estimated for mean concentration in the column. All the results are of the same order of magnitude with Asano's equation, but the results of this work and of Furukawa show an effect of nozzle diameter on k_G different from that in Asano's correlation.

Discussion

As described above, in the process of direct-contact mass and heat transfer between vapor and liquid with change of phase, dN^o and dN^w , or equivalently dN and z , are the terms to be estimated in case A, and dN and interfacial conditions (y_i, t_i) are unknown in case B. In any case, unknown quantities are estimated from the enthalpy balance for the vapor-liquid interface, Eq.(11), if the volumetric transfer coefficients are given; as in the condensation of mixed vapor on the solid surface, which was dealt with in the previous work⁶⁾.

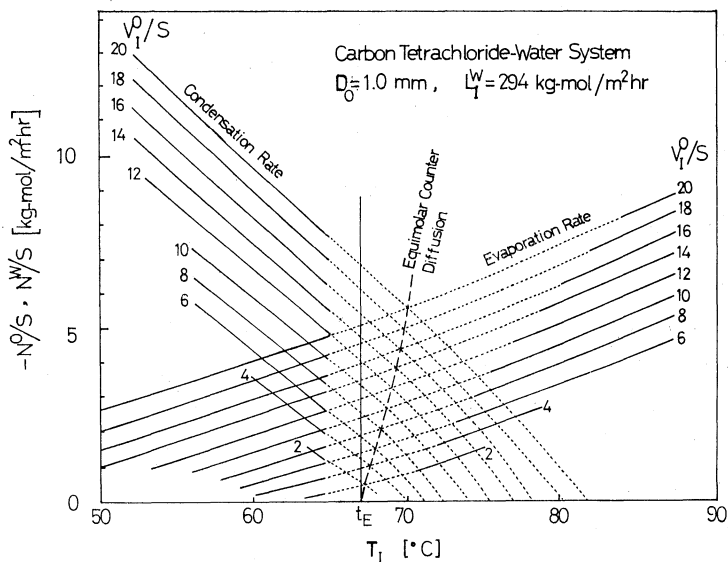
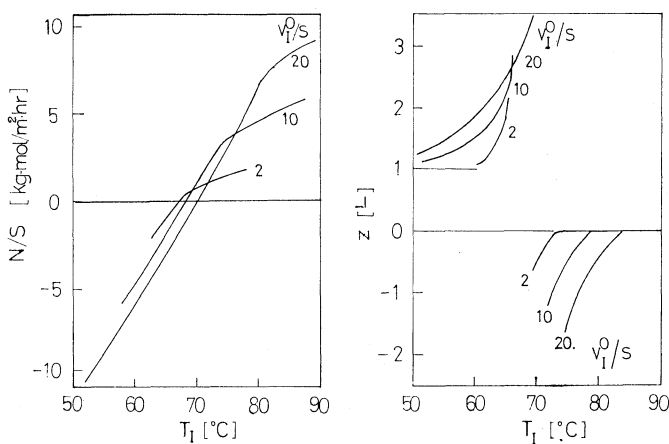


Fig. 7 Effects of T_I and V_I^0 on N^0 and N^W



Carbon Tetrachloride-Water System $D_0=1.0$ mm, $L_1^W=294$ kg-mol/m²·hr.

Fig. 8 Effects of T_I and V_I^0 on N_I and z

The enthalpy balance about the entire column is

$$V_I^0 C_G^0 (t_o - t_I) + N^0 i^0 + N^W i^W = (N^0 C_L^0 + N^W C_L^W) T_o - L_I^W C_L^W (T_o - T_I) \quad (22)$$

Condensation and evaporation rates, N^0 and N^W , for case A, and evaporation rate N^W and interfacial conditions (y_i, t_i), for case B, are calculated by Eqs. (22), (17) and (18), at any given operating conditions of flow rate of inlet water L_I^W , temperature of inlet water T_I , and flow rate of inlet vapor V_I^0 , if $k_G a$ and $h_L a$ are given.* An example of this calculation is shown in Fig. 7. In this calculation, $k_G a$ was obtained from the experimental results shown in Fig. 5 and $h_L a$ was estimated from Calderbank's and Abdel-Aal's equations. Fig. 7 shows the effect of T_I and V_I^0 on the condensation and evaporation rates. The solid line shows the calculated value, and the dotted line is drawn by interpolation of the calculated value. As the temperature of inlet water rises, condensation rate decreases and evaporation rate increases. Also, the

* The procedure of this calculation is show in Fig. 9.

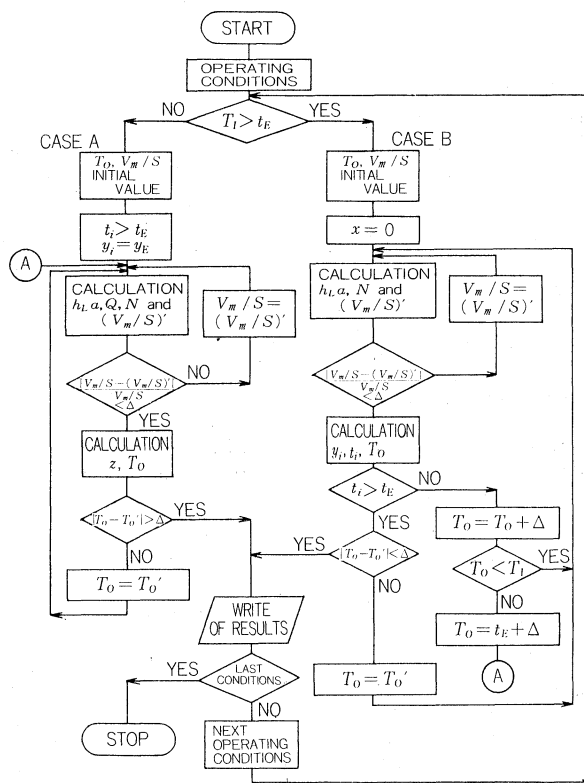


Fig. 9 Calculation procedure

larger the vapor flow rate, the higher are the condensation and evaporation rates. When T_I is higher than t_E' , an intercept of the curve for condensation rate with abscissa, organic vapor cannot condense at all. But it is interesting that if T_I is higher than t_E and lower than t_E' , condensation occurs to some extent, because the liquid temperature is reduced by evaporation of water. The intercept of the curves for the same fixed vapor flow rate, V_I^0/S , shows equimolar counter diffusion. The broken line passing through these intercepts gives the effect of V_I^0 and T_I on the rate of

equimolar counter diffusion. When the vapor flow rate, condensation and evaporation rates become quite small, this broken line approaches the eutectic point ($t_E, 0$). On the other hand, this line shows a tendency to deviate from the vertical line, $t = t_E$, as the vapor flow rate increases. From Fig. 7, total mass transfer rate N and z are easily obtained, and are shown in Fig. 8.

Conclusion

Experimental study was carried out to find the basic mechanism of direct-contact mass and heat transfer between vapor and liquid with change of phase in a bubble column. The system of water and an organic solvent immiscible in water was selected in order to simplify the problem by eliminating mass transfer in the liquid phase.

Volumetric coefficients of vapor-phase mass transfer under the conditions of simultaneous transfer of two components and transfer of single component were calculated from measured condensation and evaporation rates by Colburn and Drew's equation. These coefficients are of the same order of magnitude and similar dependency on vapor flow rate, compared with the results obtained previously for single-component transfer in such processes as absorption and humidification.

New calculation procedure of condensation and evaporation rates was presented and some results of calculation were exemplified. The effects of operating conditions, such as temperature of inlet water and flow rate of inlet vapor, on these rates were made clear.

Acknowledgement

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Nomenclature

a	= contacting area per unit volume of aerated water pool	$[m^2/m^3]$
C	= specific heat	$[kcal/kg\cdot mol\cdot ^\circ C]$
D_{AB}	= molar diffusivity	$[kg\cdot mol/m\cdot hr]$
D_0	= inside diameter of nozzles	$[m]$
g	= acceleration due to gravity	$[m/hr^2]$
g_c	= conversion factor	$[kg\cdot m/Kg\cdot hr^2]$
H	= height of aerated water pool	$[m]$
H_C	= clear liquid height	$[m]$
h_G	= vapor-phase heat transfer coefficient	$[kcal/m^2\cdot hr\cdot ^\circ C]$

h_L	= liquid-phase heat transfer coefficient	$[kcal/m^2\cdot hr\cdot ^\circ C]$
i	= molar enthalpy of vapor based on the liquid at interfacial temperature	$[kcal/kg\cdot mol]$
k_G	= gas-phase mass transfer coefficient	$[kg\cdot mol/m^2\cdot hr]$
L	= liquid flow rate	$[kg\cdot mol/hr]$
N	= total mass transfer rate	$[kg\cdot mol/hr]$
N^O	= condensation rate	$[kg\cdot mol/hr]$
N^W	= evaporation rate	$[kg\cdot mol/hr]$
N_N	= number of nozzles	$[-]$
Q	= heat transfer rate in liquid phase	$[kcal/hr]$
Re_G	= Reynolds number of gas, $U_0 D_0 \rho_G / \mu_G$	$[-]$
S	= cross-sectional area of the column	$[m^2]$
Sc	= Schmidt number	$[-]$
Sh_G	= Sherwood number of gas phase, $k_G D_0 / D_{AB}$	$[-]$
T	= temperature of liquid	$[^\circ C]$
t	= temperature of gas	$[^\circ C]$
U_0	= gas velocity through nozzles	$[m/hr]$
V	= vapor flow rate	$[kg\cdot mol/hr]$
y	= mole fraction of organic solvent in the vapor	$[-]$
z	= N^O/N	$[-]$
μ	= viscosity	$[kg/m\cdot hr]$
ρ	= density	$[kg/m^3]$
σ	= surface tension	$[Kg/m]$
<Subscript>		
E	= eutectic conditions	
G	= gas	
I	= inlet	
i	= vapor-liquid interface	
L	= liquid	
m	= mean value	
O	= outlet	
<Superscript>		
O	= organic solvent	
W	= water	

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