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Analytic modelling of a falling film absorber and experimental determination of transfer coefficients

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ABSTRACT

Accurate interpretation of the experimental data on falling film flows is a critical part of the investigations in the field of absorption energy system research. However, there is no theoretically proven way to determine experimental heat and mass transfer coefficients for non-isothermal absorption falling film flows. In this article, firstly, it is shown how the governing equations of a falling film absorber can be reduced to two ordinary differential equations and analytic expressions can be obtained for the temperature and concentration profiles along the absorber. Secondly, a new method is proposed to determine heat and mass transfer coefficients from experimental data and its application is demonstrated by reprocessing the experimental data from two experimental studies reported in the literature. The results show that some of the experimental data were misinterpreted by conventional methods and the errors were negligible only when heat and mass fluxes were small, which agrees with the fact that the obtained analytic solutions approach the conventional logarithmic heat and mass transfer equations in such conditions.

1. Introduction

In contrast with single-phase heat transfer or isothermal absorption, significant changes both in composition and temperature accompany the sorption processes in absorption refrigeration systems. The complexity of physical mechanisms in the non-isothermal sorption processes has been forcing researchers to resort mostly on experiments. However, there has been a great confusion in the interpretation of experimental data regarding the true driving potentials in the heat and mass transfer processes. Various definitions of heat and mass transfer coefficients can be found in the literature and to make matters worse, most of them lack theoretical basis. The invalidity of conventional methods has been recently investigated by Islam et al. [1–3] and Fujita and Hihara [4] and some alternative methods have been proposed. However, these alternatives have some weaknesses, which will be pointed out at the end of this section, and the problem remains largely unsolved.

First of all, various definitions of heat and mass transfer coefficients found in the literature are briefly discussed in the following (see Fig. 1 for notations).

Wall heat flux is commonly defined by

$$\dot{q}^w = k\left(\frac{\partial T}{\partial y}\right)_{y=0} \equiv \alpha'(T - T^w) \tag{1}$$

where T^w is the wall temperature, T is the solution temperature and α' is a local heat transfer coefficient. In the literature, different solution temperatures have been chosen for T including bulk solution temperature T^b, equilibrium solution temperature T^s and interface temperature T^i.

The heat rejection to the cooling water can be written as

$$\dot{q}^w_{avg} = U\Delta T_{avg} \tag{2}$$

where U is an overall heat transfer coefficient and ΔT_{avg} is an average temperature difference most commonly defined as

$$\Delta T_{avg} \equiv \frac{(T-t)_{top} - (T-t)_{bot}}{\ln[(T-t)_{top}/(T-t)_{bot}]} \tag{3}$$

Many studies [5–8] used T^b for T in Eqs. (1)–(3). On the other hand, [9–11] used equilibrium bulk solution temperatures calculated from bulk concentration and pressure, i.e. $T^s(x^b,p)$, and [12] used T^i calculated from T^b and x^b using the model of Yüksel and Schlünder [13]. Takamatsu et al. [14] measured local wall temperatures and calculated average heat transfer coefficients with the arithmetic averages of T^b and T^w.

Rather unusually, Miller and Keyhani [15] used $\Delta T_{avg} = T^i_{top} - t_{bot}$, which is the maximum temperature difference in the

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Nomenclature

A	area, m^2
a	constant in Eq. (A.4)
B	constant in Eq. (18)
b, c	constant in Eq. (B.4)
C	constant in Eq. (17)
Cp	heat capacity, kJ kg^{-1} K^{-1}
c_{1-3}	constants in Eq. (13)
D	mass diffusivity, m^2 s^{-1}
F	constant in Eq. (B.6)
g	gravity constant, m s^{-2}
h	specific enthalpy, kJ kg^{-1}
h^{fg}	latent heat, kJ kg^{-1}
$\bar{h}$	partial specific enthalpy, kJ kg^{-1} of a species
k	thermal conductivity, kW m^{-1} K^{-1}
L	absorber length, m
Le	Lewis number, $(k/\rho Cp)/D$
Nu	Nusselt number, $\alpha^b/k \times (\nu^2/g)^{1/3}$
n	exponent of Lewis number in Eq. (B.10)
$\dot{n}$	mass flux, kg m^{-2} s^{-1}
p	pressure, kPa
$\dot{q}$	heat flux, kW m^{-2}
Re_f	film Reynolds number, $4\Gamma_s/\mu$
s	distance along vapour-liquid interface, m
Sh	Sherwood number, $\beta/D \times (\nu^2/g)^{1/3}$
St_h	Stanton number for interface heat transfer, $\alpha^i/\dot{n}Cp_w$
St_m	Stanton number for mass transfer, $\rho\beta/\dot{n}$
T	temperature of solution or vapour, K
t	temperature of cooling water, K
U	average overall heat transfer coefficient, kW m^{-2} K^{-1}
u	velocity in z direction, m s^{-1}
$\overline{U}$	dimensionless heat transfer coefficient, $UL/\Gamma_s Cp_s$
$v_{1,2}$	eigenvector components
x	mass fraction of absorbent in solution
y	distance perpendicular to flow direction, m
z	distance in flow direction, m

Greek symbols

α'	local heat transfer coefficient, kW m^{-2} K^{-1}
α	average heat transfer coefficient, kW m^{-2} K^{-1}
β'	local mass transfer coefficient, m s^{-1}
β	average mass transfer coefficient, m s^{-1}
$\bar{\beta}$	dimensionless mass transfer coefficient, $\rho\beta L/\Gamma_s$
Γ	mass flow per unit perimeter, kg m^{-1} s^{-1}
Δh	heat of absorption ($\Delta h = ah^{fg}$), kJ kg^{-1}
ΔT	temperature difference, K
$\varDelta_T$	thermal boundary layer thickness at the interface, m
Δx	concentration difference or driving potential for mass transfer
$\varDelta_x$	concentration boundary layer thickness at the interface, m
δ	film thickness, m
ζ	dimensionless distance in flow direction, z/L
$\lambda_{1,2}$	eigenvalues
μ	dynamic viscosity, Pa s
ν	kinematic viscosity, m^2 s^{-1}
ρ	density, kg m^{-3}
$\varPhi_h$	correction factor for interface heat transfer in Eq. (B.8)
φ	dimensionless thermal mass flux, $\dot{n}Cp_w/\alpha^i$
$\varPhi_m$	correction factor for mass transfer in Eq. (B.2)
ϕ	dimensionless mass flux or driving potential for mass transfer, $\dot{n}/\rho\beta(= x^b - x^i)$
ω	dimensionless wall heat flux, $\dot{q}^w Cp_s/U\Delta h[= (T^b - t)Cp_s/\Delta h]$

Superscripts

b	bulk solution
i	vapour-liquid interface
l	liquid
s	saturated or equilibrium
v	vapour
w	wall
$*$	dew point

Subscripts

○	reference condition
avg	average
bot	absorber bottom
h	heat transfer
m	mass transfer
s	solution
sub	subcooling
top	absorber top
w	cooling water

system. [16–17] adopted a similar approach but used $\Delta T_{avg} = T^s(x^b_{top}, p) - t_{bot}$.

Similar to Eq. (1), mass flux at vapour-liquid interface can be written as

$$\dot{n} = \rho D\left(\frac{\partial x}{\partial y}\right)_{y=\delta} \equiv \rho\beta'\Delta x \tag{4}$$

where β' is a local mass transfer coefficient and Δx is the driving potential for mass transfer.

The average mass flux at vapour–liquid interface is expressed as

$$\dot{n}_{avg} \equiv \rho\beta\Delta x_{avg} \tag{5}$$

where β is an average mass transfer coefficient and Δx_{avg} is an average concentration difference most commonly defined as

$$\Delta x_{avg} \equiv \frac{\Delta x_{top} - \Delta x_{bot}}{\ln(\Delta x_{top}/\Delta x_{bot})} \tag{6}$$

For Δx in Eq. (6), $\Delta x = x^b - x^i$ was used in the isothermal absorption studies including [18,19] and also in many non-isothermal absorption studies including [5,8,11,12,17].

On the other hand, Kim et al. [20] used a rather unusual average driving force called "the logarithmic mean concentration difference of pressure difference" to take account of the non-condensable gas in vapour phase. [14] used $\Delta x_{avg} = (x^b - x^i)/x^i$ determined by the arithmetic averages of T^b and x^b. [15] used the difference between the bulk concentrations at the inlet and outlet, i.e. $\Delta x_{avg} = x^b_{top} - x^b_{bot}$, to avoid using the interface concentrations which they did not measure.

Except for the studies such as Yüksel and Schlünder [13] where interface concentration x^i has been determined from measured interface temperature, x^i has to be calculated somehow from the condition of bulk solution. In the isothermal absorption studies [18,19], it was calculated with Henry's law. And among the non-isothermal absorption studies above, [12] used the model of Yüksel and Schlünder [13], Kim and Infante Ferreira [8] used the method described in Appendix B, Hihara and Saito [5] and Yoon et al [11] assumed $x^i = x^s(T^b, p)$ and Bourouis et al. [17] did not clearly mention how they determined it.

As previously mentioned, a few recent studies have addressed this problem, namely Islam et al. [1–3] and Fujita and Hihara [4].