mean values of feed and outlet concentrations were adopted in estimating the distribution ratio using Eqs. (1)–(3). The constant values of the ratio enable the outlet concentrations from the 2nd stage to be calculated by the familiar KSB (Kremser, Souder and Brown) equation. The predicted values thus obtained are shown in parentheses in Table 2. For molybdenum, more than 97% of the feed was extracted into the organic phase. For vanadium, the extent of extraction is strongly dependent on the concentration of TBP used, as indicated by Eqs. (2) and (3). It was not easy to determine with accuracy concentrations less than 10^{-4} mol/l for both metals. The limit of detectable concentration was 10^{-5} mol/l. This limitation in the present experiments causes a significant difference between the observed and predicted values of the separation factors.

Conclusion

The results of extraction of two metals in both batch and flow operations agreed well with prediction using data for single-metal extraction. Scrubbing of loaded organic solution with hydrochloric acid removed 98% of vanadium and only 1–8% of molybdenum, thus increasing greatly the purity of the solution. Continuous flow extraction was successfully operated, using two-stage mixer-settler batteries, under the conditions of mean residence time of solutions of 4.2 min in the mixer and 6.7 min in the settler. Almost all the molybdenum feed was extracted into organic phase and the separation factor for the two metals increased to the order of 10^4.

Nomenclature

D	= distribution ratio of metal	[—]
K_{ex}	= extraction equilibrium constant	[—]
[ ]	= concentration of species in brackets	[mol$\cdot l^{-1}$]
β	= $([\overline{MoO_2}]/[MoO_2^{2+}])/([\overline{VO}]/[VO^{2+}])$, separation factor	[—]

⟨Subscripts⟩
Mo = molybdenum
V = vanadium

⟨Superscript⟩
— = organic-phase species or organic-phase concentration

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A SIMPLIFIED METHOD OF ESTIMATING MASS AND HEAT TRANSFER COEFFICIENTS FOR TURBULENT GAS STREAMS IN WETTED-WALL COLUMNS

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Introduction

Cylindrical wetted-wall columns are widely employed as equipment for mass or heat transfer between gas and liquid phases. For the rational design of industrial wetted-wall columns, a knowledge of the gas-phase mass or heat transfer coefficient for turbulent gas streams is required. In a previous paper,[8] a theoretical analysis of mass transfer in turbulent gas streams in the wetted-wall column was reported and the basic convective-diffusion equation was solved numerically. The solution for the average gas-phase Sherwood number Sh_G was found to be approximately expressed as a function of the gas-phase Reynolds number Re_{Gr} based on the gas velocity relative to the liquid surface and the gas-phase Schmidt number Sc_G, if the column height is sufficiently high and the Sc_G value is not very large. In another paper,[6] Hikita *et al.* have shown that the solution for the gas-phase Nusselt number Nu_G can be

obtained by replacing Sh_G and Sc_G in the numerical solution for Sh_G by Nu_G and the gas-phase Prandtl number Pr_G, respectively. Further, they found that the measured values of the average Sherwood and Nusselt numbers obtained for ammonia absorption and the adiabatic vaporization of water under rippling and non-rippling conditions were correlated well by the numerical solution for the average Sherwood and Nusselt numbers.[5,6,8]

The calculation procedure for the numerical solution is tedious and time-consuming, because of the numerical form of the solution for the average gas-phase Sherwood and Nusselt numbers and the integrative form of the solutions for the flow characteristics. In the present paper, therefore, a simplified method of estimating the gas-phase mass and heat transfer coefficients for turbulent gas streams in cylindrical wetted-wall columns is proposed and compared with experimental data reported in the literature.

1. Empirical Correlation of Gas-Phase Sherwood and Nusselt Numbers

For mass transfer into a turbulent gas stream from the surface of a falling laminar liquid film in the wetted-wall column, the theoretical values of the average Sherwood number in a large number of hypothetical systems covering a wide range of Re_{Gr} and Sc_G values were computed rigorously by a digital computer and all calculated values of Sh_G could be correlated satisfactorily by the following expressions:

$$Sh_G = 3.66 + \alpha(Re_{Gr} + 100 \log Sc_G - 1700)^m \quad (1)$$

$$\alpha = [0.0165 + 0.0565/(\sqrt{Sc_G} + 2)]Sc_G^{1/3} \quad (2)$$

and

$$m = 0.88 - 0.12/(Sc_G + 0.043)^{1/4} \quad (3)$$

For heat transfer, Sh_G and Sc_G in Eqs. (1) to (3) should be replaced by Nu_G and Pr_G, respectively.

To estimate the values of Sh_G and Nu_G, it is necessary to know the value of Re_{Gr}, that is, the thickness b and the surface velocity u_i of the falling laminar liquid film. These values can be estimated from the exact analytical expressions presented by Hikita et al.[7] The values of the gas-phase mass or heat transfer coefficient calculated from the expression for b and u_i and Eqs. (1) to (3) for Sh_G are in excellent agreement with the numerical solution, with an accuracy better than 1.2% for $4 \times 10^3 < Re_{Gr} < 10^5$ and $0.1 < Sc_G < 10$. However, use of the exact analytical expressions for b and u_i is still cumbersome because of the integrative form of the expressions. In the present study, therefore, further simplification was made and the following equations are used as the approximate expressions for b and u_i:

$$\frac{b}{d} = 0.909 Ga_L^{-1/3} Re_L^{1/3} \quad (4)$$

$$\frac{du_i \rho_L}{\mu_L} = 0.413 Ga_L^{1/3} Re_L^{2/3} \quad (5)$$

Equations (4) and (5) are identical to the well-known Nusselt equations.

As can be seen from the expressions described above, the gas-phase mass or heat transfer coefficient can be easily obtained by the following procedure:

(1) Calculate the values of u_i and $r_i (= d/2 - b)$ from Eqs. (4) and (5) to obtain the value of Re_{Gr}.

(2) Substitute the value of Re_{Gr} into Eq. (1) with Eqs. (2) and (3) to determine the value of the gas-phase Sherwood or Nusselt number.

(3) Estimate the gas-phase mass or heat transfer coefficient from the value of Sh_G or Nu_G and r_i.

The values of the gas-phase mass or heat transfer coefficient calculated by the method described above are in good agreement with the numerical solution, with an accuracy better than 8% for $10^5 < Ga_G < 10^7$, $10^{-5} < \rho_G/\rho_L < 10^{-2}$, $10^{-5} < \mu_G/\mu_L < 10^{-1}$, $4 \times 10^3 < Re_{Gr} < 10^5$, $Re_L < 1600$ and $0.1 < Sc_G < 10$.

2. Comparison of Empirical Correlation with Experimental Data

Figures 1, 2 and 3 show a comparison of the measured Sh_G values obtained by previous and present investigators[1-6,9-12] with the empirical correlation given by Eq. (1). The data shown in Fig. 1 are results for the absorption of ammonia from air streams into an aqueous sulfuric acid solution;[5] the data in Fig. 2 represent the results of the vaporization of water into air streams;[1-4,6,9-10,12] and the data in Fig. 3 show the results for the vaporization of various organic liquids into air streams.[9-11] In these figures all the data except those by McCarter and Stutzman[10] were obtained under the conditions of rippling liquid film. The solid lines represent the empirical correlation for Sh_G given by Eq. (1). The experimental results agree in general with the empirical correlation given by Eq. (1), although the data points shown in Fig. 2 indicate some scatter and the data of Jackson and Ceaglske[9] shown in Fig. 3 lie about 10 to 30% below the solid line representing Eq. (1). The deviations of the data points from the solid line may be due to the fact that the previous experiments were not carried out under conditions of true adiabatic vaporization.

Figure 4 compares the Nu_G values obtained by previous and present investigators[1,2,6,12] on the vaporization of water into air streams with the empirical correlation given by Eq. (1). As can be seen in this figure, the data of Barnet and Kobe[1] and Yoshida and Hyodo[12] lie above the solid line representing Eq.
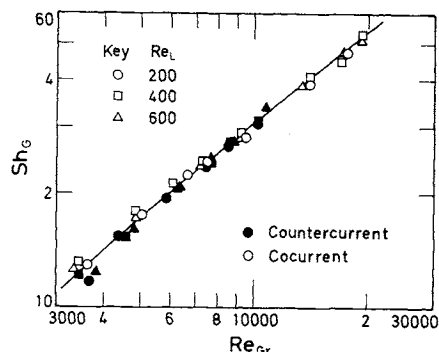
Fig. 1. Absorption of ammonia from air streams. (Countercurrent and cocurrent flows, $Sc_G = 0.66$)
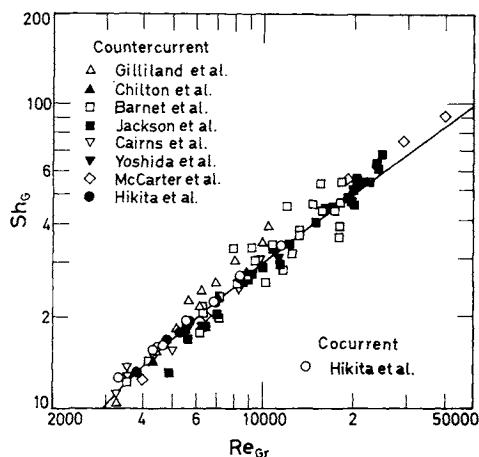

Fig. 2. Vaporization of water into air streams. (Countercurrent and cocurrent flows, $Sc_G = 0.60$)
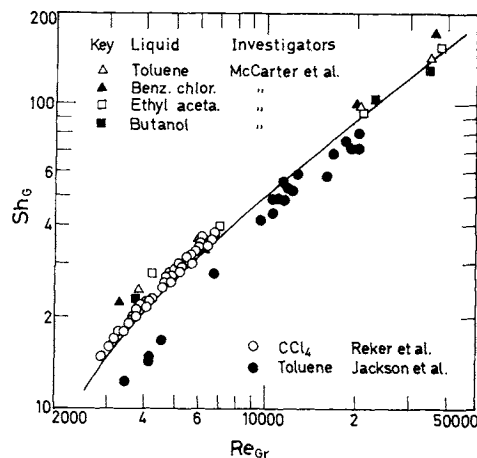

Fig. 3. Vaporization of organic liquids into air streams. (Countercurrent flow, $Sc_G = 1.82$–1.97)
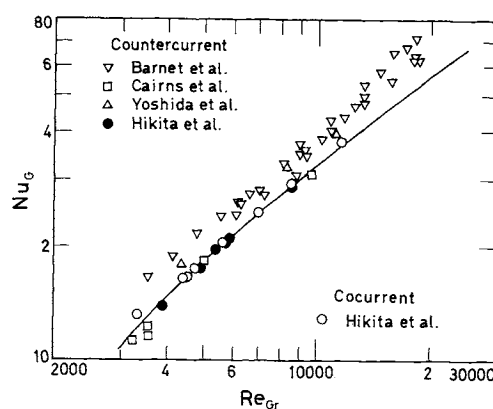

Fig. 4. Heat transfer on vaporization of water into air streams. (Countercurrent and cocurrent flows, $Pr_G = 0.72$)

(1), probably because of the overestimation of the heat transfer rate due to end effects in the lower calming sections of the wetted-wall column. The deviations of the data points, however, are about 20% and so are not very large. Therefore, it may be concluded that the experimental data for Nu_G can be correlated well by the empirical correlation given by Eq. (1).

Nomenclature

b	= liquid film thickness	[m]
c_G	= heat capacity of gas	[J/kg·K]
D_G	= gas-phase diffusivity of solute gas	[m²/s]
d	= column diameter	[m]
Ga_G	= gas-phase Galileo number, $\rho_G^2 d^3 g / \mu_G^2$	[—]
Ga_L	= liquid-phase Galileo number, $\rho_L^2 d^3 g / \mu_L^2$	[—]
g	= acceleration due to gravity	[m/s²]
h_G	= average gas-phase heat transfer coefficient	[W/m²·K]
k_G	= average gas-phase mass transfer coefficient	[m/s]
m	= constant defined by Eq. (3)	[—]
Nu_G	= average gas-phase Nusselt number, $h_G(2r_i)/\kappa_G$	[—]
Pr_G	= gas-phase Prandtl number, $c_G \mu_G / \kappa_G$	[—]
Re_{Gr}	= gas-phase Reynolds number based on gas velocity relative to liquid surface,	

	$2r_i \rho_G (u_m - u_i)/\mu_G$	[—]
Re_L	= liquid-phase Reynolds number, $4\Gamma/\mu_L$	[—]
r_i	= radius of gas passage, $d/2 - b$	[m]
Sc_G	= gas-phase Schmidt number, $\mu_G/\rho_G D_G$	[—]
Sh_G	= average gas-phase Sherwood number, $k_G(2r_i)/D_G$	[—]
u_i	= surface velocity of liquid film	[m/s]
u_m	= average velocity	[m/s]
α	= constant defined by Eq. (2)	[—]
Γ	= mass flow rate of liquid per unit perimeter	[kg/m·s]
κ_G	= thermal conductivity of gas	[W/m·K]
μ_G, μ_L	= viscosities of gas and liquid	[Pa·s]
ρ_G, ρ_L	= densities of gas and liquid	[kg/m³]

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CHARACTERISTIC FLOW BEHAVIOR OF HIGH SWIRLING JET IN A CIRCULAR VESSEL

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A swirling jet has been put to practical use to control hot air flow in spray driers, flames in combustors, etc. by using characteristic flow behavior such as reverse flow. If the intensity of swirl is weak, the flow behavior is relatively simple and its structure can be estimated by appropriate operational parameters, e.g. the Swirl number Sw.[1,5] However, the three-dimensional flow behavior becomes more complicated as the intensity of swirl increases and reverse flow appears. Hence, the investigation of a high swirling jet with reverse flow is considered necessary.

In this study, a high swirling turbulent jet in a circular vessel is taken up and the characteristic axial flow behavior is mainly investigated by measuring three-dimensional velocity components of water flow with an electrochemical technique.

1. Experimental Apparatus and Procedure

The experimental apparatus consists of two vertical acrylic-resin coaxial circular pipes with inside diameters $D_1 (= 100 \text{ mm})$ and $D_2 (= 290 \text{ mm})$. The test fluid in the smaller pipe spouts into the larger one as jet flow. In the smaller pipe the test fluid is supplied through two entrances. One is an axial entrance and the other is a tangential entrance. The flow rates are denoted by Q_z and Q_θ respectively and the total flow rate Q is represented by the sum of Q_z and Q_θ. The tangential entrance consists of two inlet pipes with an inside diameter $d (= 10 \text{ mm})$, which are attached tangentially to the smaller test pipe. An axial stream and a tangential stream join and spour into the larger

pipe 0.5 m downstream from the tangential entrance as a swirling jet.

Three-dimensional measurements of velocity were performed by using an electrochemical multi-electrode probe[3] in downstream cross sections 0.1 m and 0.3 m from the jet exit. In almost the same measuring range, some visualization experiments of flow were performed by using polystyrene particles with a diameter of about 1.2 mm.

In this study, two kinds of dimensionless operational parameters, i.e., the general Reynolds number Re and the swirl Reynolds number Re_θ, which were proposed in the previous paper,[4] are used. Re is based on the cross-sectional average velocity U_a in the small pipe as follows:

$$Re = D_1 U_a / v$$
$$= 2Q/(\pi v R_1) \tag{1}$$

Re_θ is based on the characteristic angular velocity ω_i at the jet exit. To determine the value of ω_i, the idea of the intensity of swirl Γ (the circulation) which was proposed in the previous paper[2] on swirling pipe flow is used. The value of Γ at the jet exit, i.e., Γ_i $(= 2\pi R_1^2 \omega_{i1} = 2\pi R_2^2 \omega_{i2})$, can be calculated from the initial condition at the tangential entrance by using the experimental equation[2] of the decay process of Γ in the small pipe as follows:

$$\Gamma_i / (2\pi R_1^2 \omega_0)$$
$$= 0.26 \exp\{-7100(0.5/R_1)(v/(R_1^2 \omega_0))\} \tag{2}$$

$$\omega_0 = 2Q_\theta / (\pi d^2 R_1) \tag{3}$$