



Separation Science and Engineering

Prediction of mass transfer coefficients in an asymmetric rotating disk contactor using effective diffusivity

Meisam Torab-Mostaedi, Mehdi Asadollahzadeh ^{*}, Jaber Safdari

Nuclear Fuel Cycle Research School, Nuclear Science and Technology Research Institute, P.O. Box: 11365-8486, Tehran, Iran

ARTICLE INFO

Article history:

Received 31 May 2016

Received in revised form 18 August 2016

Accepted 23 August 2016

Available online 31 August 2016

Keywords:

Asymmetric rotating disk contactor

Mass transfer coefficient

Enhancement factor

Interfacial area

Dispersed phase holdup

ABSTRACT

Mass transfer characteristics have been investigated in a 113 mm diameter asymmetric rotating disk contactor of the pilot plant scale for two different liquid–liquid systems. The effects of operating parameters including rotor speed and continuous and dispersed phase velocities on the volumetric overall mass transfer coefficients are investigated. The results show that the mass transfer performance is strongly dependent on agitation rate and interfacial tension, but only slightly dependent on phase flow rates. In this study, effective diffusivity is used instead of molecular diffusivity in the Gröber equation for estimation of dispersed phase overall mass transfer coefficient. The enhancement factor is determined experimentally and there from an empirical expression is derived for prediction of the enhancement factor as a function of Reynolds number. The predicted results compared to the experimental data show that the proposed correlation can efficiently predict the overall mass transfer coefficients in asymmetric rotating disk contactors.

© 2016 The Chemical Industry and Engineering Society of China, and Chemical Industry Press. All rights reserved.

1. Introduction

Liquid–liquid extraction columns are widely used on commercial scale in the hydrometallurgical, pharmaceutical, chemical and nuclear industries. Of the available counter-current extraction columns, the rotating disk contactor (RDC) is widely used in industrial solvent extraction processes due to its high throughput, low investment, flexible operation, and easy maintenance. Due to limitation of its structure, the mass transfer performance of commercial RDC columns is low because of severe back-mixing [1–6]. Therefore, modifications of the RDC have been attempted in order to reduce the extent of back-mixing which causes significant reduction in mass transfer efficiency. For this reason, the RDC has appeared with modified structure such as the RDC with perforated disks and the asymmetric rotating disk contactors. It goes without saying that, modified RDC columns are finding wide applications in different industries [7–10]. Asymmetric rotating disk contactors (ARDCs) are widely used in pharmaceutical, petrochemical, and chemical industries because of dual advantages of a high mass transfer rates and reduced back-mixing in both phases [11]. In this extractor, a shaft carrying the disks is mounted off-centre, and two sets of staggered stator plates are provided, connected by a vertical segmental baffle. By this means, the mixing chambers are enclosed, and are connected to one another via openings each side of the vertical baffle leading to chambers which partial coalescence occurs [11]. Since the coalescence zone in ARDC is isolated from the mixing zone, the coalescence of the dispersed droplets and the transport of the separated phase occur

more efficiently in this contactor as compared to those in the RDC. This feature not only reduces the back-mixing between the stages, but also increases the power consumption in the mixing zone and high efficiency per stage through higher values of the dispersed phase holdup and interfacial area [11].

Although the ARDC column has been used effectively for a number of separation processes, there are limited data in the literature on the performance of this type of extraction column [11]. On this basis, pilot plant experiments on hydrodynamic and mass transfer characteristics of ARDC column is necessary for the purpose of establishing the optimal design procedure for this column.

The present research work has examined the influence of operating variables including the rotor speed as well as the dispersed and continuous phase velocities on the mass transfer performance in a pilot scale ARDC column. An empirical correlation for prediction of enhancement factor is developed and used for the estimation of overall mass transfer coefficient.

2. Previous Work

The reliability of the design of liquid–liquid extraction columns depends upon the theories or correlations used for the calculation of overall mass transfer coefficients. In the literature, a large number of such theoretical and empirical equations are reported for the dispersed and continuous phase, and each has a particular range of application. For the dispersed phase, theories rather than correlations have been mostly used for the estimation of the mass transfer coefficient [12]. These theories have usually been considered in the light of three mechanisms of mass transfer inside drops.

^{*} Corresponding author.

E-mail address: masadollahzadeh@aeoi.org.ir (M. Asadollahzadeh).

The Gröber equation (Eq. (1)) [13] concerns rigid drops which do not have any internal circulation and where mass transfer is more controlled by a transient molecular diffusion:

$$K_{od} = -\frac{d}{6t} \ln \left[6 \sum_{n=1}^{\infty} B_n \exp \left(-\frac{4\lambda_n^2 D_d t}{d^2} \right) \right] \quad (1)$$

The Kronig–Brink model (Eq. (2)) [14] assumes a laminar diffusion with an inner circulation inside the drop, induced by its relative motion, with respect to the continuous phase.

$$K_{od} = -\frac{d}{6t} \ln \left[\frac{3}{8} \sum_{n=1}^{\infty} B_n^2 \exp \left(-\frac{64\lambda_n D_d t}{d^2} \right) \right] \quad (2)$$

The Handlos–Baron model (Eq. (3)) [15] deals with the case of drops with internal turbulent circulation and where mass transfer is controlled by turbulent diffusion.

$$K_{od} = -\frac{d}{6t} \ln \left[6 \sum_{n=1}^{\infty} B_n^2 \exp \left(-\frac{\lambda_n V_t t}{128d(1+\kappa)} \right) \right] \quad (3)$$

Although mass transfer in to or out of drops have been investigated for many years, it is still not fully understood, since it depends on several factors. These factors include the fact that the dispersed phase mass transfer coefficient depends upon the nature, size and behavior of the drop. The presently available equations for calculation of the dispersed phase mass transfer coefficient are not usually valid over a range of drop sizes and behaviors in a typical extraction column. An attractive method uses an enhanced molecular diffusivity, RD_d (also referred to as effective diffusivity, D_{eff}) in the equation for diffusion in rigid spheres (Eq. (1)). The enhancement factor contains the effects of all known and unknown parameters that influence the mass transfer coefficient. The resulting equation for the dispersed phase mass transfer coefficient, given by Johnson and Hamielec [16], is the following equation:

$$K_{od} = -\frac{d}{6t} \ln \left[6 \sum_{n=1}^{\infty} B_n \exp \left(-\frac{4\lambda_n^2 RD_d t}{d^2} \right) \right] \quad (4)$$

Several empirical correlations for estimation of the enhancement factor in extraction columns are described in the literature. These correlations are listed in Table 1.

The idea of using enhancement factor provides a very appropriate practical method for estimation of mass transfer coefficient for a variety of drop sizes in distribution, with different residence time in an extraction column.

3. Experimental

The pilot plant comprises a 36 compartments asymmetric rotating disk contactor of 113 mm diameter. The main section of the ARDC column consists of a 1430 mm long outer Pyrex glass shell and stainless steel internals. The disks are mounted onto a shaft and driven by an electric motor via a variable gearbox. The flow rates of both phases are

controlled via rotameters. The interface is maintained at the required level by using an optical sensor, which has been previously described [22,23]. Centrifugal pumps (Penax model) were used to circulate both liquid phases through the column. A scheme of the ARDC pilot scale unit used in the present study is shown in Fig. 1. The column dimensions are listed in Table 2.

Two liquid–liquid test systems recommended by the European Federation of Chemical Engineering (E.F.C.E.), namely toluene–acetone–water and *n*-butyl acetate–acetone–water are used in the experiments. The physical properties of the liquid–liquid systems are given in Table 3. The equilibrium data were obtained from Mišek *et al.* [24].

Before starting each run, the aqueous and organic phases were first mutually saturated, after which acetone was added to the dispersed (organic) phase to give a concentration of about 3.5 wt% acetone. The samples of each phase were taken at their inlets to the column and used for determination of the initial solute concentration. The rotor speed and the continuous phase flow rate were set at the desired valves and the dispersed phase gradually admitted into the column up to the desired volumetric flow rate. The interface location was then maintained at the desired height, and the system was allowed to reach steady-state. For all experiments the steady-state condition could be achieved after three times the residence times. At the end of each run, samples of the aqueous and organic phases were taken at their respective outlet. The solute concentrations were then determined by UV–visible spectroscopy. All experiments were performed far from flooding conditions. For each liquid–liquid system the operating variables were systematically varied to determine their influence on the volumetric overall mass transfer coefficient.

Drop size was determined by the photography method by means of a Nikon D3100 digital camera. Drop dimensions were then determined using Digimizer software. For elliptical drops both the vertical and horizontal axes were measured. In all cases, the stators thickness served as the reference for the drop size measurements. At least 400 drops were analyzed for each experiment to guarantee the significance of the determined Sauter mean drop diameter. The Sauter mean drop diameter was then calculated by the following equation:

$$d_{32} = \frac{\sum_{i=1}^N n_i d_i^3}{\sum_{i=1}^N n_i d_i^2} \quad (12)$$

where n_i denotes the number of drops of diameter d_i .

The dispersed phase holdup was measured by the shut down (displacement) method where the continuous and dispersed inlet and outlet valves were shut simultaneously and the dispersion height between the initial and final interface was measured.

4. Modeling

In the design of extraction columns, an important step is to determine the required column height. Several models have been developed in recent years, since the effect of axial mixing on the performance of liquid–liquid extraction columns was recognized. The backflow and the

Table 1
Previous correlations for estimation the enhancement factor in extraction columns

Investigator	Correlation	Column type
Steiner (1986) [17]	$R = 5.56 \times 10^{-5} \left(\frac{2Re}{1+\kappa} \right)^{1.42} \left(\frac{g\Delta\rho d_d^2}{\sigma} \right)^{0.12} \times Sc_d^{0.67} (1-x_d)$ (5)	Spray column
Bahmanyar <i>et al.</i> (2008) [18]	$D_{eff} = RD_d = 4.5151 \times 10^{-9} \exp(0.0067Re)$ (6)	Pulsed sieve plate column
Amanabadi <i>et al.</i> (2009) [19]	$D_{eff} = RD_d = 0.4755 \times 10^{-9} Re^{0.65}$ (7)	Rotating disk contactor (RDC)
Torab-Mostaedi <i>et al.</i> (2009) [20]	$R = 0.51 + 213.35 Re^{-0.257} (1-x_d) E\bar{\sigma}^{1.42}$ (8)	Hanson mixer-settler extraction column
Torab-Mostaedi and Safdari (2009) [21]	$R = -2.57 + 1326.07 Re^{0.50} Sc_c^{-0.94} (1+\kappa)^{-0.80}$ (9)	Pulsed packed column
Torab-Mostaedi <i>et al.</i> (2012) [22]	$R = -11.86 + 63.38 Re^{-0.237} (1-x_d) E\bar{\sigma}^{0.30}$ (10)	Pulsed disk and doughnut column
Hemmati <i>et al.</i> (2014) [23]	$R = 0.243 + 0.785 Re^{0.85} (1-x_d)^{7.07}$ (11)	Kühni column

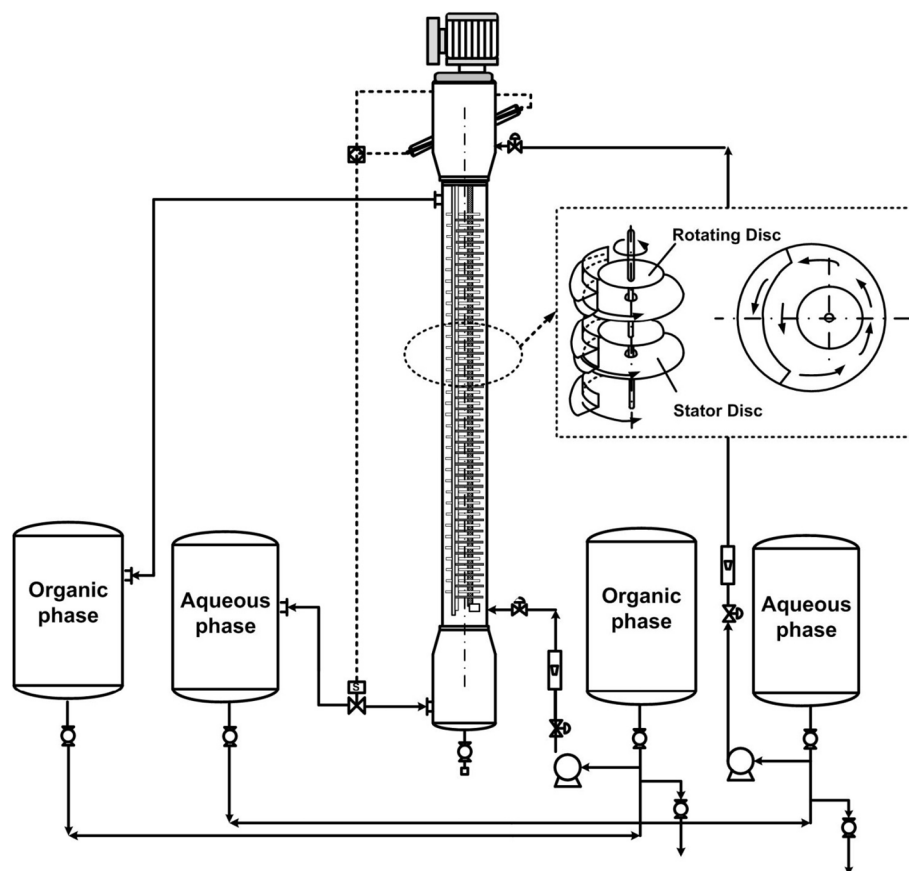


Fig. 1. Schematic flow diagram of the ARDC pilot plant.

Table 2
Dimensions of the pilot plant ARDC

Column diameter /m	Rotor diameter /m	Column working height /m	Compartment height /m	No. of Compartments
0.113	0.042	1.43	0.033	36

Table 3
Physical properties of liquid-liquid systems investigated at 20°C [19]

	Toluene-acetone-water	n-Butyl acetate-acetone-water
$\rho_c/\text{kg}\cdot\text{m}^{-3}$	994.4–995.7	994.3–995.8
$\rho_d/\text{kg}\cdot\text{m}^{-3}$	864.4–865.2	879.6–881.4
$\eta_c/\text{mPa}\cdot\text{s}$	1.059–1.075	1.075–1.088
$\eta_d/\text{mPa}\cdot\text{s}$	0.574–0.584	0.723–0.738
$\sigma/\text{mN}\cdot\text{m}^{-1}$	27.5–30.1	12.4–13.2
$D_c/\text{m}^2\cdot\text{s}^{-1}$	$(1.09\text{--}1.14)\times 10^{-9}$	$(1.01\text{--}1.06)\times 10^{-9}$
$D_d/\text{m}^2\cdot\text{s}^{-1}$	$(2.7\text{--}2.8)\times 10^{-9}$	$(2.16\text{--}2.18)\times 10^{-9}$

axial diffusion models are the most important practical approaches for considering the effect of axial mixing on the mass transfer performance of the column. In the present study, the mass transfer data are interpreted in terms of this model as described by Pratt and Stevens [25]. Based upon the axial diffusion model and mass balance in the column, over the differential elements of the column with a total effective height H , the equation set for the steady state process is established as follows, under the constant superficial velocities V_c and V_d at any given rotor speed:

$$\frac{\partial}{\partial Z} \left(x - \frac{1}{P_c} \frac{\partial x}{\partial Z} \right) + N_{oc}(x - x^*) = 0 \quad (13)$$

$$\frac{\partial}{\partial Z} \left(y + \frac{1}{P_d} \frac{\partial y}{\partial Z} \right) - N_{oc} \frac{V_c}{V_d} (x - x^*) = 0 \quad (14)$$

where $P_c = HV_c/E_c$, $P_d = HV_d/E_d$, $Z = h/H$. Note that in Eqs. (15) and (16) $N_{oc} = \frac{K_{oc}aH}{V_c}$ represents $(NTU)_{oc} = H/(\text{HTU})_{oc}$.

In this study, the dispersed phase axial dispersion is assumed to be negligible, with the continuous phase axial mixing coefficient calculated by the following equation proposed by Kumar and Hartland [26]:

$$\frac{E_c}{V_c h_c} = 0.42 + 0.29 \left(\frac{V_d}{V_c} \right) + \left[1.05 \times 10^{-2} \left(\frac{ND_R}{V_c} \right) + \frac{13.38}{3.18 + \left(\frac{ND_R}{V_c} \right)} \right] \left(\frac{V_c D_R \rho_c}{\eta_c} \right)^{-0.08} \left(\frac{D_c}{D_R} \right)^{0.16} \left(\frac{D_c}{h_c} \right)^{0.1} e \quad (15)$$

The three boundary conditions are as follows:

At the top of the column ($Z = 0$):

$$-\frac{dx}{dZ} = Pe_x(x^* - x) \quad (16)$$

At the bottom of the column ($Z = 1$):

$$\frac{dx}{dZ} = 0 \quad (17)$$

$$y = y^* \quad (18)$$

By using the axial dispersion coefficient of the continuous phase, the measured continuous and dispersed phase concentrations and the boundary conditions together with equilibrium data, the continuous phase volumetric overall mass transfer coefficients ($K_{oc}a$) are calculated from Eqs. (13) and (14).

The dispersed phase volumetric overall mass transfer coefficients are then calculated as follows:

$$\frac{1}{mK_{od}} = \frac{1}{K_{oc}} \quad (19)$$

5. Results and Discussion

Fig. 2 illustrates the effect of the rotor speed on the volumetric overall mass transfer coefficient for both studied liquid–liquid systems. This figure shows that the mass transfer performance of the column is markedly dependent on the rotor speed. As expected, the increase of agitation rate results in smaller drops. Moreover, a higher dispersed phase holdup is obtained upon increasing rotor speed. The value of the interfacial area increases with both effects. However, the overall mass transfer coefficient decreases with an increase in the rotor speed. The reduction of internal circulation and turbulence in drops leads to decrease mass transfer coefficients. The results show that the effect of interfacial area becomes more predominant than the overall mass transfer coefficient effect and consequently, the mass transfer performance will increase. At high values of the rotor speed, however, the overall mass transfer coefficient starts to fall significantly with the formation of the rigid droplets, whereby the molecular diffusion controls the drop's mass transfer rate. For *n*-butyl acetate–acetone–water system, the effect of interfacial area may be compensated by the effect of overall mass transfer coefficient at high agitation rate and consequently, the column performance is not affected by the rotor speed in this range of operating conditions. Fig. 2 also shows that the mass transfer performance is significantly influenced by the interfacial tension of the system. As seen in this figure, the column performance improves within the system of lower interfacial tension due to the formation of smaller drops and higher interfacial area. It is also found that the effect of the rotor speed on the column performance of the toluene acetone–water system (high interfacial tension) is greater than that of butyl acetate–acetone–water (medium interfacial tension), because the breakup of the drops into smaller ones is limited within the latter system due to its lower interfacial tension.

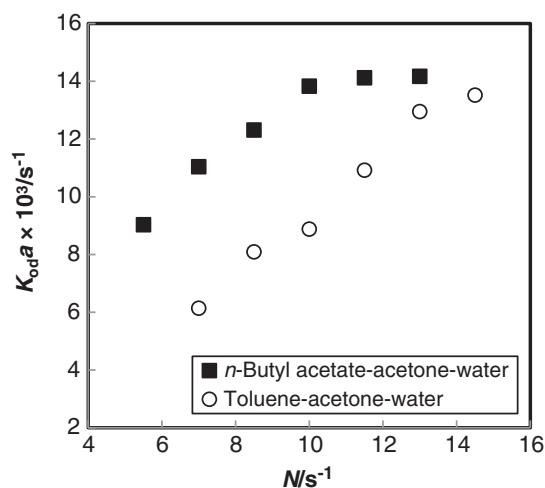


Fig. 2. Variation in the volumetric overall mass transfer coefficient with rotor speed ($V_c = 1.33 \times 10^{-3} \text{ m} \cdot \text{s}^{-1}$).

As shown in Fig. 3, the effect of the dispersed phase velocity clearly shows that the higher the dispersed phase flow rate, the better the mass transfer of the column becomes. Increasing the dispersed phase velocity tends to increase the mean drop size. A higher dispersed phase velocity results in not only a larger drop formation but also the higher coalescence frequencies. As expected, an increase in the dispersed phase flow rate leads to the increased holdup because a greater volume of the organic phase is fed to the column. In the present work, it is observed that the effect of the dispersed phase holdup on the interfacial area is greater than that of the drop size and, consequently, the interfacial area increases with an increase in the dispersed phase velocity. Furthermore, the dispersed phase mass transfer coefficient increases with an increase in the drop size due to the internal circulation inside the drop. The column performance increases with the increase in both mass transfer coefficient and interfacial area. A comparison of Figs. 2 and 3 shows that the effect of dispersed phase velocity on the mass transfer performance is less pronounced than rotor speed and interfacial tension.

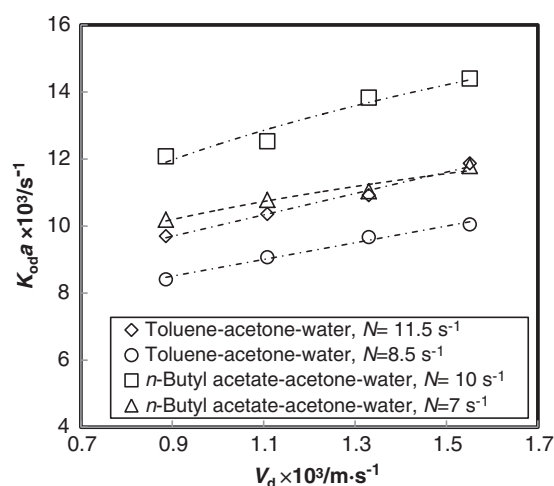


Fig. 3. Variation in the volumetric overall mass transfer coefficient with dispersed phase velocity ($V_c = 1.33 \times 10^{-3} \text{ m} \cdot \text{s}^{-1}$).

The effect of continuous phase velocity on the volumetric overall mass transfer coefficient is illustrated in Fig. 4. As seen in this figure, in the case of continuous phase velocity, no significant change in mass transfer performance is verified for the operating conditions

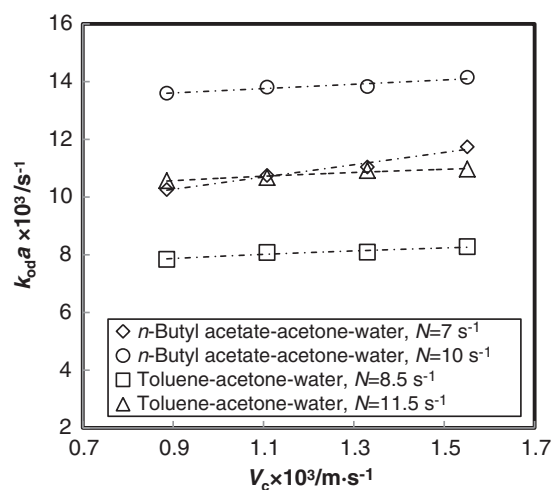


Fig. 4. Variation in the volumetric overall mass transfer coefficient with continuous phase velocity ($V_d = 1.33 \times 10^{-3} \text{ m} \cdot \text{s}^{-1}$).

investigated in this research work. The dispersed phase holdup increases with an increase in the continuous phase velocity due to the increment of the drag force between the dispersed drops and the continuous phase. Moreover, the experiments show that the mean drop size slightly increases with an increase in the continuous phase flow rate. As effects of the continuous phase velocity on the volumetric overall mass transfer coefficient, the contribution of holdup is positive and that of the mean drop size is negative because the interfacial area is directly proportional to the holdup and inversely proportional to the drop size. The results indicate that the former may be compensated by the latter and the volumetric coefficient is not affected by the continuous phase velocity within both systems.

Predictive correlation for enhancement factor

The overall dispersed phase mass transfer coefficient is determined by dividing the volumetric coefficient by the interfacial area $a (= 6x_d/d_{32})$. The experimental values of the overall mass transfer coefficient and the interfacial area are given in Table 4.

One of the main objectives of this investigation is to develop a correlation that is capable of predicting the mass transfer coefficient in ARDC columns. The experimental results for the dispersed phase overall mass transfer coefficient are compared with those of the theoretical models and the models suggested for other types of extraction columns. The values of the average relative deviation (ARD) of the calculated values of the overall mass transfer coefficient obtained by applying the previous correlations to the experimental results are summarized in Table 5. As

Table 4

Experimental values of dispersed phase overall mass transfer coefficient, interfacial area, and enhancement factor

Run No.	$N/r \cdot \text{min}^{-1}$	Q_d/h^{-1}	Q_c/h^{-1}	$a/m^2 \cdot m^{-3}$	$K_{od}/m \cdot s^{-1}$	R
Toluene-acetone-water system						
1	420	48	48	174.54	3.52×10^{-5}	5.38
2	510	48	48	249.95	3.25×10^{-5}	4.29
3	600	48	48	307.86	2.88×10^{-5}	3.46
4	690	48	48	435.34	2.51×10^{-5}	2.57
5	780	48	48	635.07	2.04×10^{-5}	1.77
6	870	48	48	819.67	1.65×10^{-5}	1.28
7	510	32	48	231.58	3.13×10^{-5}	4.08
8	510	40	48	247.60	3.17×10^{-5}	4.21
9	510	56	48	309.13	3.27×10^{-5}	4.77
10	690	32	48	415.50	2.33×10^{-5}	2.29
11	690	40	48	423.61	2.44×10^{-5}	2.46
12	690	56	48	467.54	2.54×10^{-5}	2.74
13	510	48	32	243.01	3.225×10^{-5}	4.06
14	510	48	40	249.95	3.23×10^{-5}	4.15
15	510	48	56	252.02	3.30×10^{-5}	4.45
16	690	48	32	430.54	2.46×10^{-5}	2.41
17	690	48	40	427.05	2.50×10^{-5}	2.53
18	690	48	56	431.39	2.54×10^{-5}	2.71
n-Butyl acetate-acetone-water system						
1	330	48	48	298.41	3.02×10^{-5}	3.34
2	420	48	48	410.27	2.69×10^{-5}	2.77
3	510	48	48	519.53	2.37×10^{-5}	2.32
4	600	48	48	673.37	2.05×10^{-5}	1.85
5	690	48	48	848.21	1.665×10^{-5}	1.48
6	780	48	48	1050.30	1.35×10^{-5}	1.07
7	420	32	48	391.80	2.60×10^{-5}	2.61
8	420	40	48	406.38	2.65×10^{-5}	2.67
9	420	56	48	424.57	2.77×10^{-5}	2.89
10	600	32	48	646.60	1.87×10^{-5}	1.68
11	600	40	48	645.19	1.94×10^{-5}	1.77
12	600	56	48	686.87	2.09×10^{-5}	1.92
13	420	48	32	398.37	2.63×10^{-5}	2.65
14	420	48	40	404.45	2.66×10^{-5}	2.70
15	420	48	56	422.19	2.78×10^{-5}	2.88
16	600	48	32	674.26	2.02×10^{-5}	1.77
17	600	48	40	679.89	2.03×10^{-5}	1.80
18	600	48	56	688.67	2.05×10^{-5}	1.86

Table 5

The ARD values in the predicted values of K_{od} obtained by previous equations with respect to the experimental data

Equation No.	ARD value/%
(1)	56.35
(2)	27.26
(3)	328.47
(4)	48.27
(5)	86.29
(6)	30.95
(7)	47.04
(8)	147.23
(9)	25.16
(10)	57.52
(11)	64.47

can be seen in this table, none of the previous correlations give reasonable estimates of the overall mass transfer coefficients in ARDC column.

Therefore, the experimental values of overall mass transfer coefficient are used in Eq. (4) to define the enhancement factor. This equation is reduced to its first term in determining the R values. The experimental values of R are also given in Table 4.

After calculating the experimental values of the enhancement factor for the investigated operating conditions, Eq. (20) is derived in terms of the Reynolds number by using the least squares method.

$$R = -0.79 + 0.48Re^{0.67} \quad (20)$$

where:

$$Re = \frac{d_{32}V_s\rho_c}{\eta_c} \quad (21)$$

in which V_s is the slip velocity between the two phases through the column. The slip velocity between the phases is obtained as follows:

$$V_s = \frac{V_d}{x_d} + \frac{V_c}{1-x_d} \quad (22)$$

The predicted values of R are used in Eq. (4) to calculate K_{od} values. A comparison between the predicted and experimental values of K_{od} is depicted in Fig. 5 where a good agreement is observed. The proposed method predicts the experimental values of K_{od} with an average relative deviation of 7.13%.

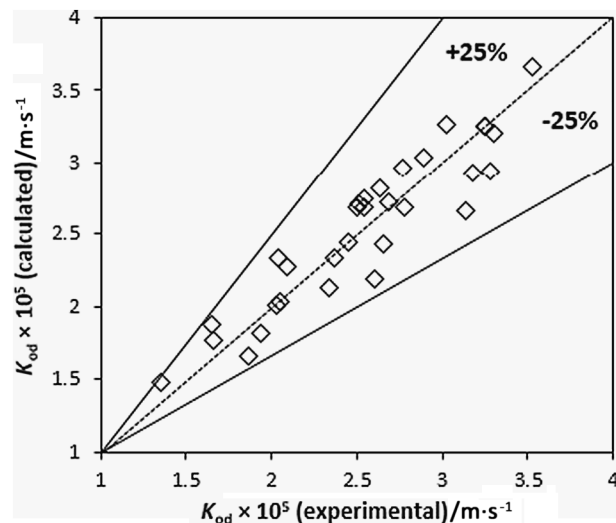


Fig. 5. Parity plot of experimental values of dispersed phase overall mass transfer coefficient against those predicted by Eqs. (4) and (20).

6. Conclusions

This paper presents an experimental study on the mass transfer performance of an asymmetric rotating disk contactor. The experimental findings show that the mass transfer performance is strongly dependent on the rotor speed and interfacial tension. Improved column performance is observed within the system of lower interfacial tension. The results show that the continuous phase velocity has little influence the value of $K_{od}a$, while $K_{od}a$ increases with an increase in the dispersed phase velocity. The results also show that the correlations developed in the other types of extractors cannot be used to predict the mass transfer performance of the ARDC column. An empirical expression for the enhancement factor as a function of Reynolds number is also proposed. The proposed correlation which predicts the enhancement factor in the ARDC column can be applied to estimate the column height in different separation processes. The present study has provided valuable information on the mass transfer characteristics of ARDC column about which there are currently limited pieces of information.

Nomenclature

a	interfacial area, $\text{m}^2 \cdot \text{m}^{-3}$
B_n	n th coefficient in Eqs. (1)–(4)
D	molecular diffusivity, $\text{m}^2 \cdot \text{s}^{-1}$
D_c	column diameter, m
D_{eff}	effective diffusivity, $\text{m}^2 \cdot \text{s}^{-1}$
D_R	impeller diameter, m
d_{32}	Sauter mean drop diameter, m
E	axial mixing coefficient, $\text{m}^2 \cdot \text{s}^{-1}$
$Eö$	Eötvös number ($=g\Delta\rho d_{32}^3/\sigma$)
e	fractional free cross-sectional area
g	acceleration due to gravity, $\text{m}^2 \cdot \text{s}^{-1}$
H	effective height of the column, m
h_c	compartment height, m
K	overall mass transfer coefficient, $\text{m} \cdot \text{s}^{-1}$
m	distribution ratio
N	rotor speed, s^{-1}
N_{ox}	number of 'true' transfer unit ($=K_{oc}aH/V_c$)
P	Péclet number ($=HV/E$)
Pe_c	continuous-phase Péclet number ($=d_{32}V_s/D_c$)
Q	flow rate of the continuous or dispersed phase, $\text{m}^3 \cdot \text{s}^{-1}$
R	enhancement factor for mass transfer
Re	Reynolds number ($=d_{32}V_s\rho_c/\eta_c$)
Sc	Schmidt number ($=\eta/\rho D$)
t	time, s
V	superficial velocity, $\text{m} \cdot \text{s}^{-1}$
$\bar{V}_c$	true velocity for continuous phase ($=V_c/(1-x_d)$), $\text{m} \cdot \text{s}^{-1}$
V_s	slip velocity, $\text{m} \cdot \text{s}^{-1}$
x	mass fraction of acetone in continuous phase
x_d	dispersed phase holdup
x^*	equilibrium mass fraction of acetone in continuous phase corresponding to dispersed phase
y	mass fraction of acetone in dispersed phase
$\Delta\rho$	density difference between phases, $\text{kg} \cdot \text{m}^{-3}$
λ_n	n th coefficient if Eqs. (1)–(4)
ρ	density, $\text{kg} \cdot \text{m}^{-3}$
κ	viscosity ratio (η_d/η_c)
η	viscosity, $\text{Pa} \cdot \text{s}$
σ	interfacial tension, $\text{N} \cdot \text{m}^{-1}$

Subscripts

c	continuous phase
d	dispersed phase
o	overall value

x	x-phase (continuous phase in present case)
y	y-phase (dispersed phase in present case)

Superscripts

*	equilibrium value
°	inlet to column

References

- [1] M. Asadollahzadeh, Sh. Shahhosseini, M. Torab-Mostaedi, A. Ghaemi, Mass transfer performance in an Oldshue–Rushton column extractor, *Chem. Eng. Res. Des.* 100 (2015) 104–112.
- [2] M. Asadollahzadeh, Sh. Shahhosseini, M. Torab-Mostaedi, A. Ghaemi, The effects of operating parameters on stage efficiency in an Oldshue–Rushton column, *Chem. Ind. Chem. Eng. Q.* 22 (1) (2016) 75–83.
- [3] M. Asadollahzadeh, M. Torab-Mostaedi, Sh. Shahhosseini, A. Ghaemi, Holdup, characteristic velocity and slip velocity between two phases in a multi-impeller column for high/medium/low interfacial tension systems, *Chem. Eng. Process.* 100 (2016) 65–78.
- [4] M. Asadollahzadeh, M. Torab-Mostaedi, Sh. Shahhosseini, A. Ghaemi, Experimental investigation of dispersed phase holdup and flooding characteristics in a multistage column extractor, *Chem. Eng. Res. Des.* 105 (2016) 177–187.
- [5] A. Kumar, S. Hartland, Prediction of dispersed phase hold-up in rotating disc extractors, *Chem. Eng. Commun.* 106 (1987) 56–87.
- [6] É. Moreira, L.M. Pimenta, L.L. Caneiro, P.C.L. Faria, M.B. Mansur, C.P. Pibeiro, Hydrodynamic behavior of a rotating disc contactor under low agitation conditions, *Chem. Eng. Commun.* 192 (2005) 1017–1035.
- [7] Y.D. Wang, W.Y. Fei, J.H. Sun, Y.K. Wan, Hydrodynamics and mass transfer performance of a modified rotating disc contactor (MRDC), *Chem. Eng. Res. Des.* 80 (2002) 392–400.
- [8] M. Asadollahzadeh, A. Ghaemi, M. Torab-Mostaedi, Sh. Shahhosseini, Experimental mass transfer coefficients in a pilot plant multistage column extractor, *Chin. J. Chem. Eng.* 24 (8) (2016) 989–999.
- [9] Y. Kawase, Dispersed-phase holdup and mass transfer in a rotating disc contactor with perforated skirts, *J. Chem. Technol. Biotechnol.* 48 (1990) 247–260.
- [10] S. Soltanali, Y. Ziaie-Shirkolaei, G. Amoabediny, H. Rashedi, A. Sheikhi, P. Chamanrokh, Hydrodynamics and mass transfer of rotating sieved disc contactors used for reversed micellar extraction of protein, *Chem. Eng. Sci.* 64 (2009) 2301–2306.
- [11] B.D. Kadam, J.B. Joshi, R.N. Patil, Hydrodynamic and mass transfer characteristics of asymmetric rotating disc extractors, *Chem. Eng. Res. Des.* 87 (2009) 756–769.
- [12] N. Outili, A.–H. Meniai, G. Gneist, H.–J. Bart, Model assessment for the prediction of mass transfer coefficients in liquid–liquid extraction columns, *Chem. Eng. Technol.* 30 (2007) 758–763.
- [13] H. Gröber, Die Erwärmung und abkühlung einfacher geometrischer körper, *Z. Var. Dtsch. Ing.* 69 (1925) 705–711.
- [14] R. Kronig, J.C. Brink, On the theory of extraction from falling drops, *Appl. Sci. Res. A2* (1950) 142–154.
- [15] A.E. Handlos, T. Baron, Mass and heat transfer from drops in liquid–liquid extraction, *AIChE J.* 3 (1957) 127–136.
- [16] A.I. Johnson, A.E. Hamielec, Mass transfer inside drops, *AIChE J.* 6 (1960) 145–149.
- [17] L. Steiner, Mass transfer rates from single drops and drop swarms, *Chem. Eng. Sci.* 41 (1986) 1979–1986.
- [18] H. Bahmanyar, L. Nazari, A. Sadr, Prediction of effective diffusivity and using of it in designing pulsed sieve extraction columns, *Chem. Eng. Process.* 47 (2008) 57–65.
- [19] M. Amanabadi, H. Bahmanyar, Z. Zarkeshan, M.A. Moosavian, Prediction of effective diffusion coefficient in rotating disc columns and application in design, *Chin. J. Chem. Eng.* 17 (2009) 366–372.
- [20] M. Torab-Mostaedi, J. Safdari, M. Ghannadi-Maragheh, M.A. Moosavian, Prediction of overall mass transfer coefficient in a Hanson mixer-settler using effective diffusivity, *J. Chem. Eng. Jpn.* 42 (2009) 78–85.
- [21] M. Torab-Mostaedi, J. Safdari, Prediction of mass transfer coefficients in a pulsed packed extraction column using effective diffusivity, *Braz. J. Chem. Eng.* 26 (2009) 685–694.
- [22] M. Torab-Mostaedi, A. Ghaemi, M. Asadollahzadeh, Prediction of mass transfer coefficients in a pulsed disc and doughnut extraction column, *Can. J. Chem. Eng.* 90 (2012) 1569–1577.
- [23] A. Hemmati, M. Torab-Mostaedi, M. Asadollahzadeh, Mass transfer coefficients in a Kühni extraction column, *Chem. Eng. Res. Des.* 93 (2015) 747–754.
- [24] T. Mišek, R. Berger, J. Schröter, Standard Test Systems for Liquid Extraction Studies, *EFCE Publ. Ser.* 46 (1985) 1.
- [25] H.R.C. Pratt, G.W. Stevens, Axial dispersion, in: J.D. Thornton (Ed.), *Science and Practice in Liquid–Liquid Extraction*, Oxford University Press, UK, 1992, pp. 461–492.
- [26] A. Kumar, S. Hartland, Prediction of axial mixing coefficients in rotating disc and asymmetric rotating disc extraction columns, *Can. J. Chem. Eng.* 70 (1992) 77–87.