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Use of axial dispersion model for determination of Sherwood number and mass transfer coefficients in a perforated rotating disc contactor



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ABSTRACT

The mass transfer process in a perforated rotating disk contactor (PRDC) using a toluene–acetone–water system was investigated. The volumetric overall mass transfer coefficients are calculated in a PRDC column. Both mass transfer directions are considered in experiments. The influences of operating variables containing agitation rate, dispersed and continuous phase flow rates and mass transfer in the extraction column are studied. According to obtained results, mass transfer is significantly dependent on agitation rate, while the dispersed and continuous phase flow rates have a minor effect on mass transfer in the extraction column. Furthermore, a novel empirical correlation is developed for prediction of overall continuous phase Sherwood number based on dispersed phase holdup, Reynolds number and mass transfer direction. There has been great agreement between experimental data and predicted values using a proposed correlation for all operating conditions.

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1. Introduction

Mass transfer performance of a system is an inseparable part of many chemical engineering operations in two-phase systems. Solvent extraction (SE) is one of these chemical operations which applied in an enormous number of industrial systems. The solvent extraction process can be conducted in a number of different extraction columns. These extraction columns are divided into non-agitated and agitated extraction columns [1,2]. The rotating disk contactor (RDC) is one of the most important agitated extraction columns. This type of extraction column has been widely used in chemical industries and petroleum refining due to its high throughput, low investment, low energy consumption, and easy operation and maintenance [3]. However, this extraction column has some limitations due to its structure. There have been some drawbacks including severe axial back mixing and inappropriate drop distribution which cause a reduction in the mass transfer efficiency of commercial RDCs, particularly for low interfacial tension systems. Both the continuous and dispersed phases as they pass through the extraction column can affect axial mixing [4,5]. For these reasons, modified RDCs are developed to overcome the mentioned issues. Perforated rotating disk contactor is one of these modified extraction columns that leads to the enhancement in the mass transfer

performance of a system. PRDC extraction column is considered as a subset of the RDC extraction column with improved structure, for instance, the RDC with sieved disks [6,7]. Better drop dispersion, lower bond drops, more contact time and at the last count, the higher mass transfer are the advantages of PRDC extraction column in comparison to RDC extraction column [8]. Furthermore, low operating expenses, easy manufacturing in the industrial scale compared to other agitated columns, and easy scale up to commercial scale are the other plus points of PRDC extraction columns. Sieved disks were apparently more appropriate to liquid–liquid systems with low interfacial tension [9]. Based on the obtained experimental results [9], the height of the transfer unit for the sieved disk extraction columns is much lower than that of the rotating disk extraction columns. PRDCs are also used by Fan *et al.* [10] for phenol removal using liquid membrane separation. According to their conclusion, the mass transfer performance of sieved disk extraction columns is much better than that of the RDC [10]. Wang *et al.* [5] investigated how the perforations on disks cut the droplets and disperse them in the continuous phase. The droplet size in the PRDC extraction column can be controlled using holes on sieved disks and is rather uniform. Thus, it is expected that the extraction column with sieved disks has suitable mass transfer performance and controllability compared to the rotating disk extraction column [5].

The design and scale up of an extractor need the calculation of two independent parameters including the cross-sectional area and the height of the column. The first one is required to accommodate the desired flows without flooding and the second one should be determined

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to reach the appropriate performance of extraction columns. Ideal plug flow conditions were considered in continuous and dispersed phases for the calculation of the height needed for a separation system in early design and scale-up procedures. However, it is now accepted that axial mixing in the extraction column for one or two phases can lead to the enhancement of the required height for reaching desired separation [11]. Using a diffusion or backflow model for design and scale-up is now developed for determination of the height of the extraction column. In this method one parameter is considered for all deviations from the ideal plug flow conditions [12]. As a result, widespread availability of reliable correlations for calculation of mass transfer coefficients is required to develop an extraction column design. On the other hand, determining mass transfer coefficient in an extractor is difficult due to significant uncertainty. Although the PRDC was applied appropriately for a number of separation processes, limited experimental data are available from the literature on the mass transfer performance of PRDC extraction columns.

In the current study, the volumetric overall mass transfer coefficients of a pilot scale perforated rotating disk contactor have been determined. A thorough research was carried out for calculation of operating variable effects which include disk rotational speed, flow rates of continuous and dispersed phase on mass transfer performance of the extraction column for a water–acetone–toluene liquid–liquid system. Additionally, experimental results are compared with Kumar and Hartland [13] correlations and new correlations are developed so that the overall mass transfer coefficient can be precisely foreseen.

2. Experimental

2.1. PRDC pilot scale unit

A pilot scale unit of a perforated rotating disk contactor was constructed. Internal diameter of the perforated rotating disk contactor is 113 mm and it has 43 stages. The accessory equipment contains storage tanks, pumps and rotameters. The rotor shaft carries sieved disks with a diameter of 0.07 m and drilled with 30 holes of 0.015 cm diameter. All internal parts of the column and the blade impellers were made of AISI 316 stainless steel. The blade impellers are mounted on a shaft, which is driven by an electric motor via a variable speed gear box. An optical sensor was used to interface the location of two phases at the top of the extraction column which was automatically controlled. Centrifugal pumps (Penax model) were used to circulate both liquid phases through the column. A schematic of the PRDC pilot-scale unit used in this study is depicted in Fig. 1. The geometrical data of the investigated extraction column are given in Table 1.

2.2. Liquid–liquid system

All experiments were conducted at ambient temperature, using a liquid–liquid system according to the recommendations of the European Federation of Chemical Engineering [14]. The system was toluene–acetone–water. Distilled water was used as continuous phase in all experiments. The physical properties of this system are given in Table 2. Liquid–liquid equilibrium data of the system was taken from

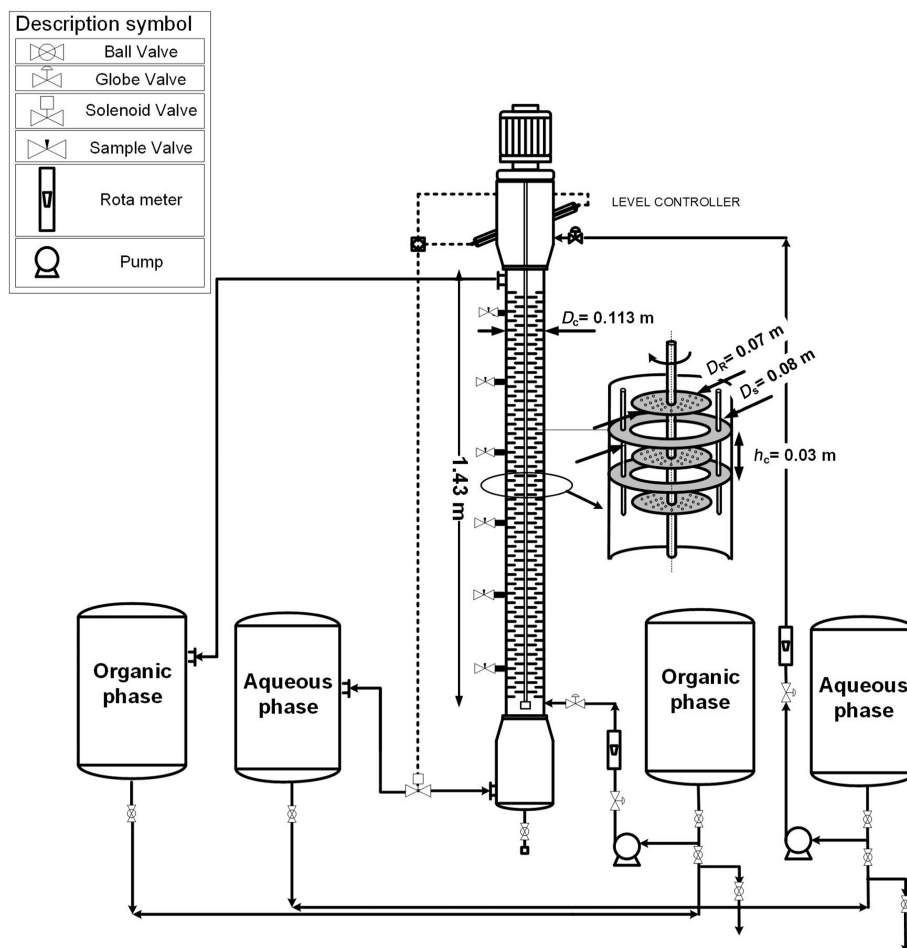


Fig. 1. A schematic diagram of the perforated rotating disk contactor (PRDC).

Table 1

The main dimensions of the PRDC column

Item	Symbol	Column dimensions/m
Column internal diameter	D_c	0.113
Stator ring diameter	D_s	0.08
Disk diameter	D_R	0.07
Compartment height	h_c	0.03
Column active height	Z	1.43
Settler diameter	–	0.5

Table 2

Physical properties of the systems studied at 20 °C [14]

Physical property	Toluene–acetone–water
$\rho_c/\text{kg} \cdot \text{m}^{-3}$	994.4–995.7
$\rho_d/\text{kg} \cdot \text{m}^{-3}$	864.4–865.2
$\eta_c/\text{mPa} \cdot \text{s}^{-1}$	1.059–1.075
$\eta_d/\text{mPa} \cdot \text{s}^{-1}$	0.574–0.584
$\sigma/\text{mN} \cdot \text{m}^{-1}$	27.5–30.1
$D_c/\text{m}^2 \cdot \text{s}^{-1}$	$(1.09\text{--}1.14) \times 10^{-9}$
$D_d/\text{m}^2 \cdot \text{s}^{-1}$	$(2.7\text{--}2.8) \times 10^{-9}$

Mišek *et al.*'s [14] paper. All experiments were performed with acetone concentrations of approximately 3.5 wt%.

2.3. Experimental procedures

Before running each experiment, the continuous and dispersed phases were mutually saturated, after which acetone was added to the toluene. The acetone content is approximately 3.5 wt% in organic phase. The extraction column was filled with the continuous phase and the agitation rate is then adjusted to the exact amount. Next, the organic phase as dispersed phase was entered into the extraction column. Optical sensor was applied to maintain the interface position at the desired height. The system was permitted to achieve steady state. Samples of aqueous and organic phases at 7 points along the extraction column were collected for determination of acetone concentration. Solute concentrations were then determined by UV–visible spectroscopy. The photography approach using a Nikon D3100 digital camera is used for determination of drop size. Several photographs of the extraction column contents were taken. The photos were analyzed by Digimizer Image Analysis software. The drops were compared with the size of the disks or stator thickness served as the reference for the drop size measurements. Most of the droplets are spherical based on experimental observations. For elliptical drops both the vertical and horizontal axes were measured and the following equation was used to calculate the mean drop size of droplets.

$$d_e = (d_1^2 d_2)^{1/3} \quad (1)$$

The Sauter mean diameter is then determined as follows:

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

where n_i is the number of mean diameter drop.

The hold-up is defined as the ratio of the dispersed phase to the total volume of the extraction region. At the end of each experiment, displacement approach was applied for determination of hold-up of dispersed phase in the extractor [15]. The holdup of dispersed phase was determined using change of interfacial height after coalescence of dispersed phase droplets.

3. Modeling

According to the axial diffusion model and material balance in the extraction column, over the differential elements of the extraction column with a total effective height, H , the equation set for the steady state process is established using following equation, under the constant superficial velocities V_c and V_d at any given agitation rate [16]:

$$\frac{\partial}{\partial Z} \left(x - \frac{1}{Pe_c} \frac{\partial x}{\partial Z} \right) + N_{oc}(x - x^*) = 0 \quad (3)$$

$$\frac{\partial}{\partial Z} \left(y + \frac{1}{Pe_d} \frac{\partial y}{\partial Z} \right) - N_{oc} \frac{V_c}{V_d} (x - x^*) = 0 \quad (4)$$

It should be noted that in Eqs. (3) and (4) $N_{oc} = K_{oc} a H / V_c$ represents $(NTU)_{oc} = H / (HTU)_{oc}$. Pe_c and Pe_d are the Peclet number and may be represented as Eqs. (5) and (6):

$$Pe_c = \frac{H V_c}{E_c} \quad (5)$$

$$Pe_d = \frac{H V_d}{E_d} \quad (6)$$

The axial mixing in the dispersed phase is generally less pronounced than that in the continuous phase. In the present work, dispersed phase axial dispersion is assumed to be negligible. The following boundary conditions can be used for the extraction column:

$$\frac{dx}{dZ} = 0, \quad y = y^0 \text{ (at the bottom of the extractor), } Z = 1 \quad (7)$$

$$-\frac{dx}{dZ} = Pe_c(x^0 - x), \text{ (at the top of the extractor), } Z = 0 \quad (8)$$

For calculation of continuous phase volumetric overall mass transfer coefficient from Eqs. (3) and (4), the axial dispersion coefficient, the measured concentrations of continuous and dispersed phase and boundary conditions along with the equilibrium data are required.

The specific interfacial area is obtained by the following equation:

$$a = \frac{6x_d}{d_{32}} \quad (9)$$

4. Results and Discussion

4.1. Effect of agitation rate

The influence of agitation rate on the volumetric overall mass transfer coefficient, mean drop size and holdup is shown in Fig. 2(a, b and c). It can be seen that higher agitation rate results in the enhancement of holdup but the reduction of mean drop size. Consequently, decreasing the mean drop size and increasing holdup would increase the interfacial area for mass transfer. However, a decrease in drop size causes a reduction of the overall mass transfer coefficient because of decreasing the internal circulations inside the droplets. Since the latter effect is not dominant, the overall result is the enhancement of the overall mass transfer coefficient via increasing the agitation rate.

Fig. 2 also reveals that the extraction column performance changes with the mass transfer direction. When the mass transfer carries out from the continuous (c) to the dispersed phase (d), acetone concentration at the top of the drop is lower than the bottom of the drop. This interfacial tension driving force results in the interface moving in opposition to the direction of the inner circulation generated in the droplet. So, interface deformation can happen when the direction of mass transfer is $c \rightarrow d$. Drop deformation such as the break-up process leads

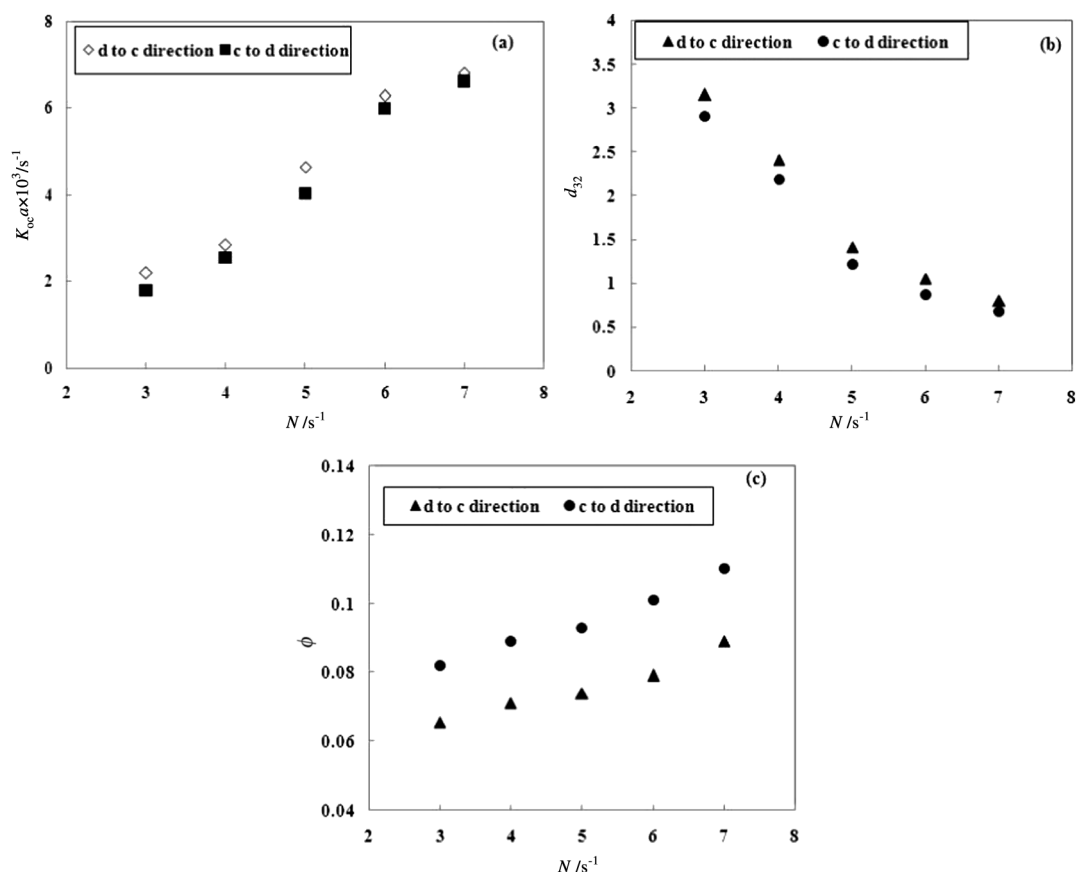


Fig. 2. (a) The volumetric overall mass transfer coefficient as a function of agitation rate and mass transfer direction, (b) the Sauter mean drop diameter as a function of agitation rate and mass transfer direction, (c) the dispersed phase holdup as a function of agitation rate and mass transfer direction.

to the enhancement of mass transfer. For $d \rightarrow c$ mass transfer, the interfacial tension driving force is opposite. The interface motion and the inner circulation are in the same direction. Consequently, the droplets can be more stable than in the $c \rightarrow d$ direction. Therefore, drops tend to be smaller in $c \rightarrow d$ mass transfer than the opposite direction (see Fig. 2(b)). In addition, the dispersed phase holdup is larger for $c \rightarrow d$ acetone transfer because of the smaller droplet formation (see Fig. 2(c)). Finally, in the case of the $c \rightarrow d$ direction, the interfacial area is higher than that of in the $d \rightarrow c$ because of the lower drop size and higher holdup.

The higher mass transfer coefficient in the $d \rightarrow c$ direction than $c \rightarrow d$ direction for acetone transfer can be explained via the interfacial turbulence induction of local differences in the acetone concentration. The interfacial tension driving force occurs due to interfacial movement and internal circulation generated in the drop. The latter mentioned phenomena are in the same direction in $c \rightarrow d$ acetone transfer, while in $d \rightarrow c$ acetone transfer, they are in the opposing directions. For the mentioned reason, internal agitation within the drop would carry out in $d \rightarrow c$ transfer systems which results in higher mass transfer coefficient than that realized for $c \rightarrow d$ transfer systems. It is seen that the influence of interfacial area on the mass transfer performance of the extraction column is lower than that of the overall mass transfer coefficient. Consequently, mass transfer in the direction of dispersed phase to continuous leads to appropriate column performance [17,18].

4.2. Effect of continuous phase velocity

The influence of continuous phase velocity on volumetric overall mass transfer coefficient, mean drop size and holdup is shown in Fig. 3(a, b and c). According to Fig. 3, generally, an increase in the velocity of the continuous phase has a minor effect on the volumetric overall

mass transfer coefficient, mean drop size and holdup. From Fig. 3(b), it can be seen that the Sauter mean drop diameter changes slightly with the increase of the continuous phase velocity. Also, based on section (c) in Fig. 3, there has been an increase in the dispersed phase hold-up with an increase in continuous phase velocity. This observation could be attributed the increment of drag force between the continuous phase and the drops. The interfacial area increases with the enhancement of dispersed phase holdup, when the mean drop size is almost constant. It can be also seen that the overall mass transfer coefficient increases a little with an increase in the continuous phase velocity. Thus, the extraction column mass transfer performance will improve slightly due to slight variation in the hold-up and mean drop size with variation of velocity of the continuous phase.

4.3. Effect of dispersed phase velocity

Fig. 4(a, b and c) shows the influence of dispersed phase velocity on volumetric overall mass transfer coefficient, mean drop size and holdup respectively. As it can be seen from Fig. 4(b) and (c), the Sauter mean drop diameter and dispersed phase holdup increase with the enhancement of dispersed phase velocity in the $c \rightarrow d$ and $d \rightarrow c$ mass transfer directions, although, the influence of the dispersed phase holdup on the interfacial area is dominant than that of the mean drop size. Thus, there is an increase in the interfacial area with the increase of dispersed phase velocity. As it is also seen, the overall mass transfer coefficient increases with the enhancement of the dispersed phase velocity for the $d \rightarrow c$ and $c \rightarrow d$ directions. For both mass transfer directions, the mass transfer performance in the extraction column would enhance with an increase in both overall mass transfer coefficient and the interfacial area.

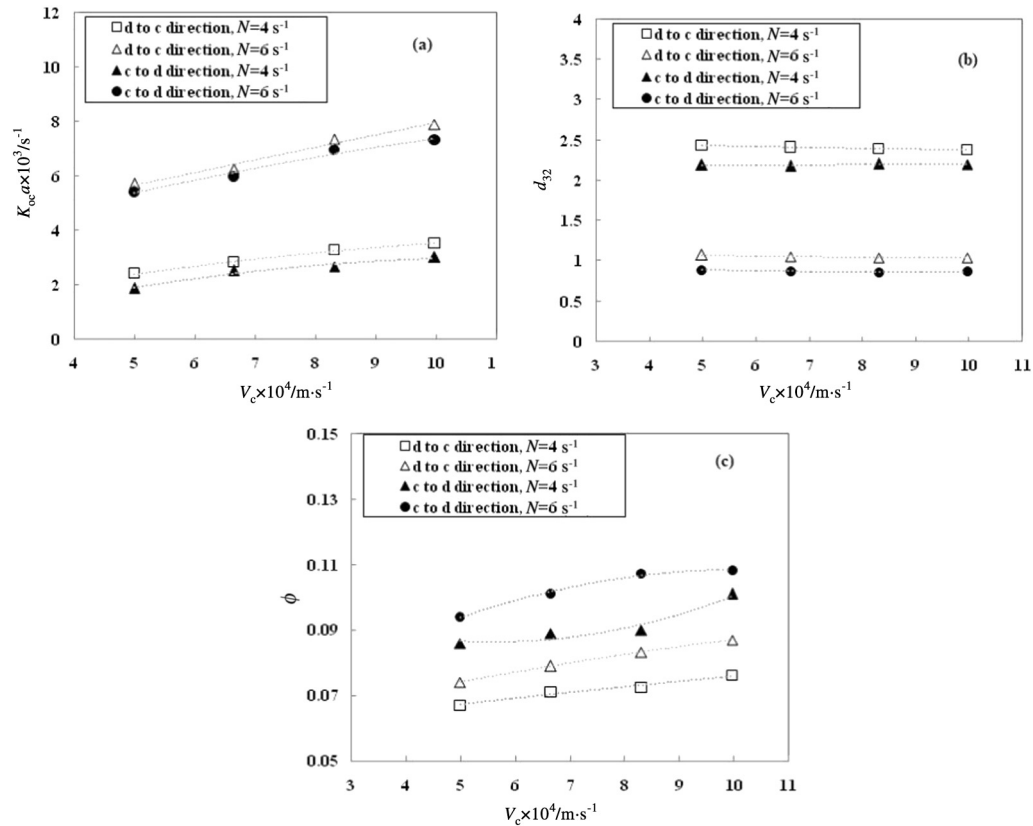


Fig. 3. (a) The volumetric overall mass transfer coefficient as a function of the continuous phase velocity, (b) the Sauter mean drop diameter as a function of the continuous phase velocity and (c) the dispersed phase holdup as a function of the continuous phase velocity.

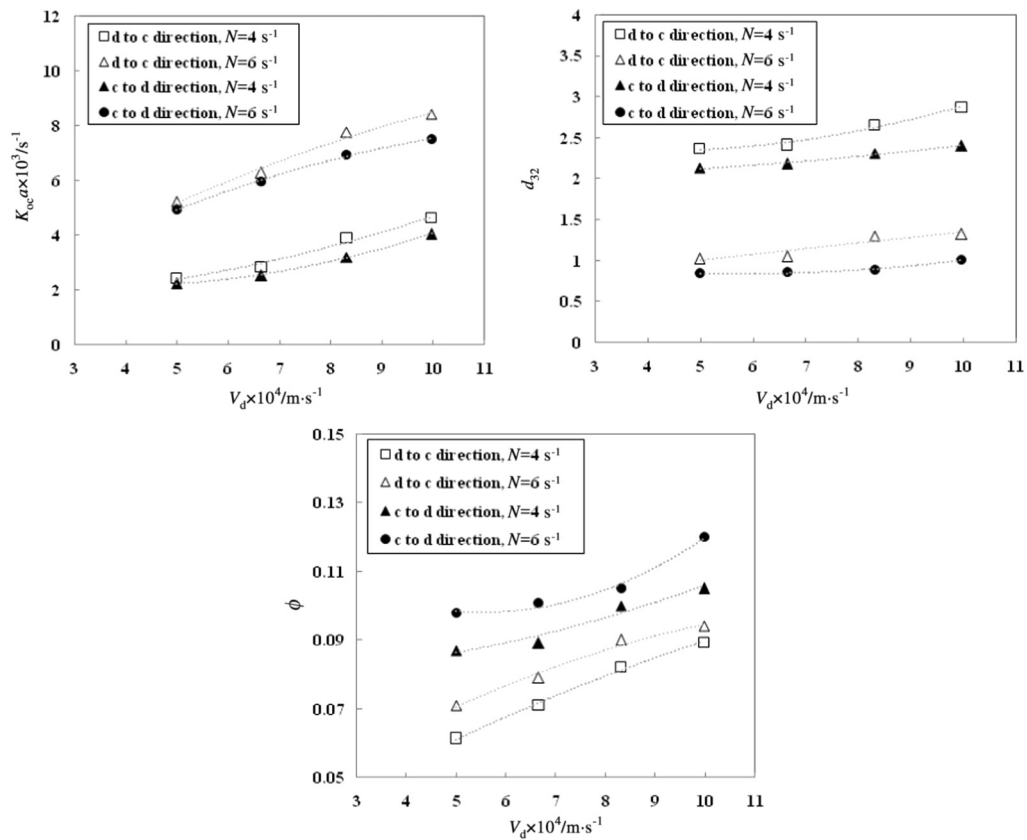


Fig. 4. (a) The volumetric overall mass transfer coefficient as a function of the dispersed phase velocity, (b) the Sauter mean drop diameter as a function of the dispersed phase velocity and (c) the dispersed phase holdup as a function of the dispersed phase velocity.

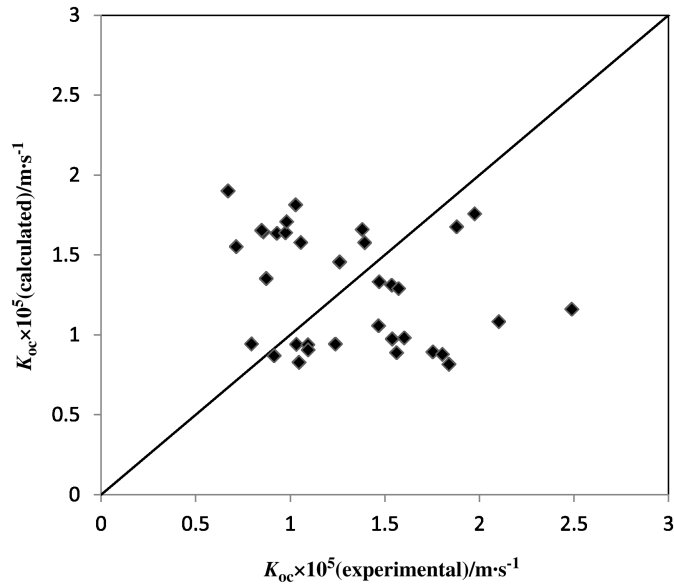


Fig. 5. Comparison of the experimental overall mass transfer coefficient data with the correlation by Kumar and Hartland [19] for water–acetone–toluene.

4.4. Predictive correlation of the overall mass transfer coefficient

Development of new correlation for overall mass transfer coefficient in a perforated rotating disk column is one of the aims of the current study. Kumar and Hartland [19] proposed semi-empirical predictive correlations for the individual mass transfer coefficients of the dispersed and continuous phases in different kinds of extraction columns including

spray, Kühni, rotating disk, EC and PSE columns. The developed correlations for the RDC extraction column may be written as follows [19]:

$$\frac{Sh_c/(1-x_d)-Sh_{c,rigid}}{Sh_{c,\infty}-Sh_c/(1-x_d)} = 5.26 \times 10^{-2} Re^{1/3} + 6.59 \times 10^{-2} Re^{0.25} \times Sc_c^{1/3} \left(\frac{V_s \eta_c}{\sigma} \right)^{1/3} \frac{1}{1+k^{1.1}} \quad (10)$$

$$Sh_d = 17.7 + \frac{3.19 \times 10^{-3} (Re Sc_d^{1/3})^{1.7}}{1 + 1.43 \times 10^{-2} (Re Sc_d^{1/3})^{0.7}} \left(\frac{\rho_d}{\rho_c} \right)^{2/3} \times \frac{1}{1+k^{2/3}} \quad (11)$$

$$Sh_{c,rigid} = 2.43 + 0.775 Re^{1/2} Sc_c^{1/3} + 0.0103 Re Sc_c^{1/3} \quad (12)$$

$$Sh_{c,\infty} = 50 + \frac{2}{\sqrt{\pi}} (Pe_c)^{1/2} \quad (13)$$

$$\frac{1}{K_{oc}} = \frac{1}{k_c} + \frac{1}{mk_d} \quad (14)$$

The overall mass transfer coefficient is determined by Eq. (14). The experimental continuous phase overall mass transfer coefficient for the water–acetone–toluene system is compared with the obtained result from the developed correlation by [19]. Fig. 5 compares the experimental values with determined results using proposed correlations. The average absolute value of the relative error is calculated as 44.43%.

Table 3 is provided for comparison of experimental results of overall mass transfer coefficient with the results of previous studies. Average Absolute Relative Error (AARE) is used to determine the accuracy of the current work. According to Table 3, it can be resulted that none of the previous correlations are suitable for determination of mass transfer

Table 3

The AARE values in the predicted values of K_{oc} obtained by the previous correlation to the experimental data

Researcher	Equation	Average absolute relative error		Ref
		d → c	c → d	
Lochiel and Calderbank (1964)	$Sh_c = 0.7 Re^{1/2} Sc_c^{1/3}$ high Reynolds numbers	73.5%	64.7%	[21]
Brauer et al. (1971)	$Sh_c = 2 + \left[\frac{0.66}{1+Sc_c} + \frac{Sc_c}{2.4+Sc_c} \left(\frac{0.79}{Sc_c^{1/6}} \right) \right] \frac{Pe_c^{1.7}}{1+Pe_c^{1.2}} 1 < Re < 100$	67.3%	53.7%	[22]
Work of several authors	$Sh_c = 2 + C_1 Re^{1/2} Sc_c^{1/3}$	75.4%	69.32%	[22]
Clift et al. (1978)	$Sh_c = 1 + 0.724 Re^{0.48} Sc_c^{1/3}$	72.4%	66.7%	[22]
Weber et al. (1975)	$Sh_c = 2 \left(\frac{Pe_c}{\pi} \right)^{1/2} \left[1 - \frac{1}{Re^2} (2.89 + 2.15 \kappa^{0.64}) \right]^{1/2} \kappa \leq 2$ and $\rho_d/\rho_c \leq 4$	44.3%	33.7%	[23]
Clift et al. (1978)	$Sh_c = 2 \left(\frac{Pe_c}{\pi} \right)^{1/2} \times \left\{ 1 - \frac{\frac{2+3\kappa}{3(1+3\kappa)}}{\left[1 + \left(\frac{(2+3\kappa) Re^{1/2}}{(1+\kappa)(8.67+6.45 \kappa^{0.64})} \right)^{n_1} \right]^{1/n_1}} \right\}^{1/2} \kappa < 2$ and $n_1 = \frac{4}{3} + 3\kappa$	57.5%	40.2%	[22]
Steiner et al. (1986)	$\frac{Sh_c - Sh_{c,rigid}}{Sh_{c,circ} - Sh_{c,rigid}} = 1 - \exp \left[-4.18 \times 10^{-3} Pe_c^{0.42} \right]$ $10 < Re < 1200$ $190 < Sc_c < 241000$ $Sh_{c,rigid} = 2.43 + 0.775 Re^{1/2} Sc_c^{1/3} + 0.0103 Re Sc_c^{1/3}$	49.5%	41.4%	[24]
Garner et al. (1959)	$Sh_c = -126 + 1.8 Re^{0.5} Sc_c^{0.42}$ $8 < Re < 800$	125.03%	119.3%	[22]
Lochiel and Calderbank (1964)	$Sh_c = 2 \left(\frac{Pe_c}{\pi} \right) \left[1 - \frac{2+3\kappa}{1 + (\kappa \rho_d/\rho_c)^{0.5} Re^{0.25}} \frac{1.45}{Re^{0.25}} \right]^{1/2}$ $20 < Re < 100$	78.9%	69.1%	[21]
Brauer et al. (1971)	$Sh_c = 2 + 0.0511 Re^{0.724} Sc_c^{0.70}$ $4 < Re < 1000$, $130 < Sc_c < 23600$	96.8%	87.2%	[22]
Garner and Tayeban (1960)	$Sh_c = 50 + 8.5 \times 10^{-3} Re Sc_c^{0.7}$	66.4%	58.3%	[22]
Hughmark (1967)	$Sh_c = 2 + 0.084 \left[Re^{0.484} Sc_c^{0.339} \left(\frac{d_g^{1/3}}{D_c^{1/3}} \right)^{0.072} \right]^{1.5}$	38%	29.7%	[25]
Clift et al. (1978)	$Sh_c = \frac{1.2}{\sqrt{D_c}} \left[\frac{48 \sigma d}{\pi^2 (2\rho_c + 3\rho_d)} \right]^{1/4}$	58.2%	39.3%	[22]
Torab-Mostaedi et al. (2012)	$R = c_1 + c_2 Re^{c_3} (1-\phi) E \phi^{c_4}$ for d → c transfer : $c_1 = -11.86$, $c_2 = 63.38$, $c_3 = -0.237$, $c_4 = 0.30$ for c → d transfer : $c_1 = -13.33$, $c_2 = 36.69$, $c_3 = -0.135$, $c_4 = 0.16$ $E \phi = \frac{g \Delta \rho d_{32}}{\sigma} Re = \frac{d_{32} V_s \rho_c}{\mu_c}$	56.3%	42.9%	[26]
Torab-Mostaedi et al. (2011)	$Sh_{oc} = -121.56 + 103.62 Re^{0.16} (1-\phi) (d \rightarrow c)$ $11.73 < Re < 69.43$ $Sh_{oc} = -119.5 + 113.3 Re^{0.12} (1-\phi) (c \rightarrow d)$ $9.45 < Re < 57.08$	53.2%	35.6%	[27]
Torab-Mostaedi et al. (2010)	$Sh_{oc} = 12.34 + 0.116 Re^{1.389}$ $Re > 10$ $Sh_{oc} = 2.586 + 0.000217 Re^{4.86}$ $Re < 10$	46.4%	23.3%	[28]

coefficient in the perforated rotating disk contactor. Also, the comparison between experimental data and the investigated data by Korchinsky [20] in the same experimental conditions was shown in Table 4. It is observed that the performance of the PRDC column improved better than the RDC column.

Table 4

The comparison between the performance of PRDC and RDC columns

Experimental conditions			RDC (Korchinsky [20])		PRDC (Present work)	
$Q_c/L \cdot \text{min}^{-1}$	$Q_d/L \cdot \text{min}^{-1}$	$N/r \cdot \text{min}^{-1}$	$K_c \times 10^3/\text{cm} \cdot \text{s}^{-1}$	$E_c/\text{cm}^2 \cdot \text{s}^{-1}$	$K_c \times 10^3/\text{cm} \cdot \text{s}^{-1}$	$E_c/\text{cm}^2 \cdot \text{s}^{-1}$
24	30	240	1.08	1.17	2.01	0.95
24	36	300	1.07	1.82	2.17	1.37
18	24	360	1.17	1.53	1.38	1.45
24	36	240	1.22	1.14	2.48	0.96
24	30	300	1.07	2.00	1.93	1.25
18	24	360	1.29	1.73	1.31	1.41

Consequently, based on the Sherwood number, Reynolds number and dispersed phase holdup in the differing directions of mass transfer novel empirical correlations are developed in order to predict overall mass transfer coefficients. The following correlations are calculated for the water–acetone–toluene system in the PRDC column:

$$Sh_{oc} = -22.05 + 13.49Re^{0.498}(1-x_d) \quad (d \rightarrow c) \quad 6.17 \leq Re \leq 31.97 \quad (15)$$

$$Sh_{oc} = -12.20 + 8.56Re^{0.514}(1-x_d) \quad (c \rightarrow d) \quad 4.99 \leq Re \leq 23.96 \quad (16)$$

After calculation of the Sherwood number using Eqs. (15) and (16), the overall mass transfer coefficient of continuous phase can be determined via the following equation:

$$Sh_{oc} = \frac{K_{oc}d_{32}}{D_c} \quad (17)$$

Also, the Reynolds number in Eqs. (15) and (16) is calculated by Eq. (18). In which, V_s is the slip velocity between the continuous and dispersed phases through the extraction column. The slip velocity between the continuous and dispersed phases is obtained by Eq. (19).

$$Re = \frac{d_{32}V_s\rho_c}{\eta_c} \quad (18)$$

$$V_s = \frac{V_d}{x_d} + \frac{V_c}{1-x_d} \quad (19)$$

The comparison of experimental data with those calculated by Eqs. (15) and (16) is shown in Fig. 6. The average absolute value of the relative error for prediction of the overall mass transfer coefficient using the proposed correlations for the water–acetone–toluene system is 12.53%.

Finally, mass transfer data for the water–acetone–toluene system in this study is compared with mass transfer data for other types of extraction columns which was obtained by Kumar and Hartland [19]. Fig. 7 shows the relation between $Re \cdot Sc_c$ and the Sherwood number in different extraction columns. It is observed that the mass transfer behavior of the PRDC extraction column is the same as the RDC extraction column.

4.5. Axial mixing coefficients for continuous and dispersed phases

The effect of rotor speed on the mass transfer and axial mixing coefficients for the continuous and dispersed phase was observed in Table 5. It is shown that the values of axial mixing coefficients for the dispersed

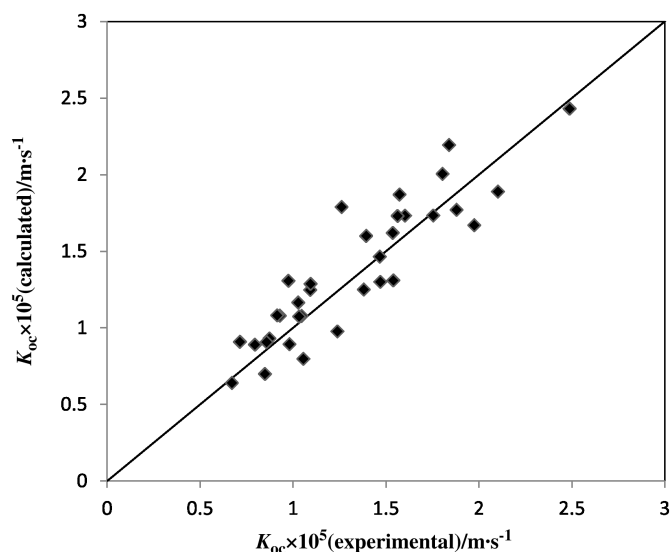


Fig. 6. Comparison of the experimental values of overall mass transfer coefficient with the correlations developed in the current study.

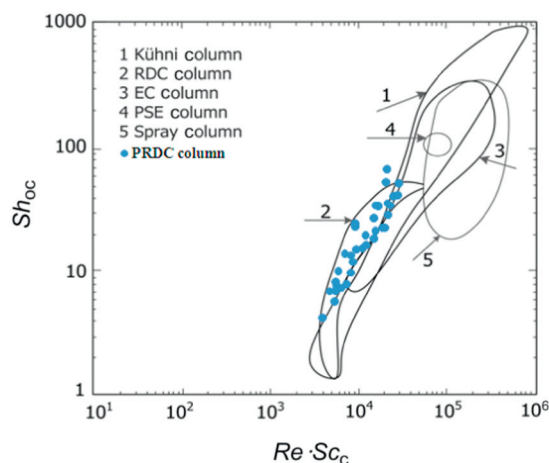


Fig. 7. The relationship between $Re \cdot Sc_c$ and Sh_{oc} in different extraction columns [19].

phase were lower than the continuous phase and the neglect from these values in modeling equations is correct.

4.6. Errors of d_{32} and acetone concentration

The experimental errors such as errors in measuring droplet size from pictures and errors in acetone concentration from each experiment may occur in the experimental work. Therefore, measurements for determination of the Sauter mean drop diameter were made in triplicate to verify experimental reproducibility. The Sauter mean drop diameter for each run was determined by the obtained average results from taking photos. The average of the Sauter mean drop diameter with a standard deviation for one of the experiments is shown in Fig. 8. The acetone concentrations were obtained after steady conditions with the five measurements from each valve in 10 min intervals to verify experimental reproducibility. The average of acetone concentrations with a standard deviation for one of the experiments is shown in Fig. 9.

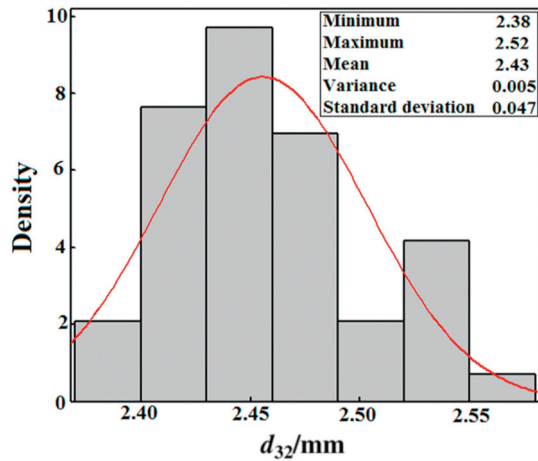
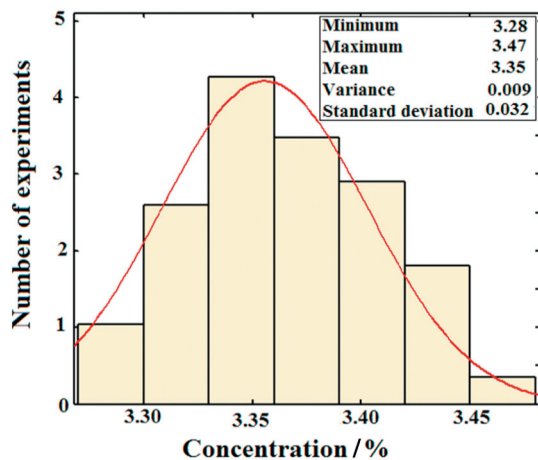
5. Conclusions

In the current study, the mass transfer performance of the perforated rotating disk contactor for the water–acetone–toluene system is

Table 5

Axial mixing coefficients for continuous and dispersed phases

$N/r \cdot \text{min}^{-1}$	$Q_c/L \cdot \text{min}^{-1}$	$Q_d/L \cdot \text{min}^{-1}$	$K_{od}a \times 10^5$	E_c	E_d	E_c/E_d
180	24	24	322.113	7.9050×10^{-5}	1.3150×10^{-7}	6.01×10^2
240	24	24	416.2933	9.4150×10^{-5}	6.8500×10^{-7}	1.37×10^2
300	24	24	680.6408	1.2850×10^{-4}	7.9000×10^{-6}	16.3
360	24	24	923.2175	1.4840×10^{-4}	2.2000×10^{-6}	67.5
420	24	24	1003.149	1.5450×10^{-4}	1.1500×10^{-6}	1.34×10^2

**Fig. 8.** Average Sauter mean drop diameter and standard deviation for toluene–water.**Fig. 9.** Average acetone concentration and standard deviation for toluene–water.

investigated. The effect of operating parameters including dispersed and continuous phase velocities and agitation rate is considered for investigation of mass transfer in the PRDC extraction column. The results revealed that the column performance increased with an increase in agitation rate, whereas it changes slightly with the continuous phase velocity. It is also indicated that the mass transfer performance increases with the dispersed phase velocity for $c \rightarrow d$ and $d \rightarrow c$ transfer. The results indicated that the volumetric overall mass transfer coefficient is higher for acetone transfer in the direction of continuous to dispersed phase because of the smaller drop size and higher hold-up for mass transfer. Furthermore, continuous phase overall mass transfer coefficients are predicted using new correlations which are developed in this study. The proposed correlations can be applied to determine the extraction column height for various extraction processes.

Nomenclature

a	interfacial area, $\text{m}^2 \cdot \text{m}^{-3}$
D	molecular diffusivity, $\text{m}^2 \cdot \text{s}^{-1}$
D_c	column diameter, m
D_i	rotor diameter, m
d_{32}	Sauter mean drop diameter, m
E	axial mixing coefficient, $\text{m}^2 \cdot \text{s}^{-1}$
g	acceleration due to gravity, $\text{m} \cdot \text{s}^{-2}$
H	effective height of the column, m
K	overall mass transfer coefficient, $\text{m} \cdot \text{s}^{-1}$
N	rotor speed, s^{-1}
Pe	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, $\text{m}^3 \cdot \text{s}^{-1}$
Re	Reynolds number ($= d_{32}V_s\rho_c / \mu_c$)
Sc	Schmidt number ($= \mu / (\rho D)$)
Sh	Sherwood number ($= kd_{32} / D$)
Sh_{oc}	overall continuous-phase Sherwood number ($= K_{oc}d_{32} / D$)
t	time, s
V	superficial velocity, $\text{m} \cdot \text{s}^{-1}$
V_s	slip velocity, $\text{m} \cdot \text{s}^{-1}$
V_t	terminal velocity, $\text{m} \cdot \text{s}^{-1}$
x	mass fraction of acetone in continuous phase
x^*	equilibrium mass fraction
y	mass fraction of acetone in dispersed phase
$\Delta\rho$	density difference between phases, $\text{kg} \cdot \text{m}^{-3}$
κ	viscosity ratio (μ_d/μ_c)
μ	viscosity, $\text{Pa} \cdot \text{s}$
ρ	density, $\text{kg} \cdot \text{m}^{-3}$
σ	interfacial tension, $\text{N} \cdot \text{m}^{-1}$
φ	dispersed phase holdup
ψ	power dissipated per unit mass, $\text{m}^2 \cdot \text{s}^{-3}$

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
o	inlet to column

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