

Development of an Effective Diffusivity Model for Regular Packed Liquid Extraction Columns

Ahmad Rahbar Kelishami, Hossein Bahmanyar, Laleh Nazari, Mohammad Ali Moosavian

Engineering College, Chemical Engineering Faculty, University of Tehran, Iran

Abstract: The accurate specification of the mass transfer coefficient plays an important role in packed columns precise design. This research is based on the substitution of effective diffusivity for molecular diffusivity in calculation of mass transfer coefficient using an experimental setup which is a regular packed extraction column. A novel correlation is obtained for effective diffusivity as a function of Reynolds number and the height of the packings. Since the effect of height is considered in estimation of the mass transfer coefficient it has less average errors comparing other available theoretical equations and is in great compatibility with experimental values.

Key words: Packed column, Mass transfer coefficient, Regular packing, Effective diffusivity, Liquid - liquid extraction

INTRODUCTION

Liquid-liquid extraction is one of the most important separation processes. This process has a high ability in separation of temperature-sensitive mixtures, mixtures in which boiling point of components are very close, separation of components with low relative volatility or generally for mixtures whose separation with distillation process is impractical or very difficult and it is used widely in different chemical industries (Slater, M.J., 1994).

Different kinds of liquid - liquid contactors are being used in industries which are classified according to the mixing type of the two phases. The generally poor performance of spray columns can be improved by introducing column internals such as packing. It is possible to increase the mass transfer coefficients in these extractors using random and regular packings. Packing reduces axial mixing, increases drop coalescence and breakage rates resulting in increased mass transfer rates, and affects the mean residence time of the dispersed phase. It also reduces the area available for flow and so somewhat lowers the capacity of the column. Recently a number of structured packings have been developed for distillation and are being considered for use in liquid extraction.

Therefore understanding of relevant fluid dynamics and the mass transfer coefficients in these columns is of paramount importance for the precise design. Because of the presence of packing and its effect on producing internal circulation, the mass transfer coefficient will be increased. The transfer of a solute between a single drop and a continuous liquid phase has been widely studied, both theoretically and experimentally. The importance of this area is reflected in the vast number of works published over the years and is of interest in a variety of industrial processes, such as liquid extraction (Newman, A.B., 1931; Skelland, A.H.P., R.M. Wellek, 1964; Slater, M.J., 1995; Kumar, A., S. Hartland, 1999).

Since one phase is dispersed in the other phase the explanation of mass transfer behavior in droplets is necessary for designing the extraction process. The mass transfer rate is strongly affected by drop size and hydrodynamics of the two phases. Usually small drops are considered as rigid spheres whose internal circulations are negligible and the mass transfer inside the droplet is controlled by molecular diffusion. Furthermore the surface active agents affect the internal circulation and consequently the amount of mass transfer (Slater, M.J., 1995; Brodkorb, M.J., D. Bosse, 2003).

In the next section a set of theoretical equations for prediction of the mass transfer coefficient are reviewed briefly. Then a new relation for effective diffusivity is introduced. A similar job has been done on pulsed sieve plate extraction columns previously (Bahmanyar, H., L. Nazari, 2008). In this research we are trying to obtain an equivalent equation for regular packed extraction columns and moreover consider the effect of height in our relation which is a new and useful idea for more precise design.

Corresponding Author: Hossein Bahmanyar, Engineering College, Chemical Engineering Faculty, University of Tehran, Iran
Tel.: +98 2161112213.
Email address: hbahmany@ut.ac.ir (H. Bahmanyar)

Previous Works:

Many studies have been done on mass transfer from a single drop with vertical motion in a column and in a continuous phase with uniform concentration. Models describing the mass transfer into or out of drops rising or falling in a continuous medium may also be classified into categories of stagnant and circulating (with laminar and turbulent flow) patterns inside drops. Investigation of mass transfer coefficient for stagnant drops was first analyzed by Grober for the case of resistance in both phases. The consequence model is (Grober, H., 1925):

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

The next model is proposed by Newman (1931) for the particular case of no resistance in the continuous phase and unsteady- state mass transfer. This model can be used for drop Reynolds number values less than 10 (Newman, A.B., 1931):

$$K_d = -\left(\frac{d}{6t}\right) \operatorname{Ln} \left[\left(\frac{6}{\pi^2}\right) \sum_{n=1}^{\infty} \left(\frac{1}{n^2}\right) \exp \left(-\frac{4D_d \pi^2 n^2 t}{d^2} \right) \right] \tag{2}$$

Where:

$$\operatorname{Re} = \frac{\rho_c V d}{\mu_c} \tag{3}$$

By retaining only the first term of series, the mass transfer coefficient for long contact times may be derived in simple form from the above equation:

$$K_d = \frac{2\pi^2}{3} \times \frac{D_d}{d} \tag{4}$$

Kronig and Brink (1950) developed a model for mass transfer of drops with internal circulations for the case of negligible external resistance and the drop Reynolds number values between 10 and 200. The result for the mass transfer coefficient is (Kronig, R., J.C. Brink, 1950):

$$K_d = \left(\frac{d}{6t}\right) \left[\left(\frac{3}{8}\right) \sum_{n=1}^{\infty} B_n^2 \exp \left(\frac{-64\lambda_n D_d t}{d^2} \right) \right] \tag{5}$$

Their model was extended by Elzinga and Banhero (1959) for the case of a finite continuous-phase resistance. The resulting expression for the mass-transfer coefficient, K_{od} , is still given by equation (5) but the values of the coefficients B_n and λ_n are defined as functions of continuous phase resistance which is given

by $\frac{K_c d}{D_d}$. They carried out their model when the drop Reynolds numbers up to 80 (Elzinga, E.R., J.T.

Banhero, 1959).

Calderbank and Korchinski (1956) proposed an alternative approach involving the use of an enhanced

molecular diffusivity, $\mathfrak{R}D_d$, (also referred to as effective diffusivity) in the equation of Kronig and Brink with the dimensionless enhancement factor, $\mathfrak{R} = 2.25$ (Calderbank, P.H., I.J.O. Korchinski, 1956):

$$K_d = -\left(\frac{d}{6t}\right) \text{Ln} \left[1 - \left\{ 1 - \exp\left(-\frac{4\pi^2 \mathfrak{R} D_d t}{d^2}\right) \right\}^{\frac{1}{2}} \right] \quad (6)$$

Handlos and Baron (1957) proposed a model which describes the mass-transfer mechanism for drops that have strong turbulent internal circulations. They found out that for the high drop Reynolds number values drops have strongly turbulent or oscillating circulations. By considering resistance in both phases, the resulting expression for the mass transfer coefficient is given by (Handlos, A.E., T. Baron, 1975),

$$K_d = -\left(\frac{d}{6t}\right) \text{Ln} \left[2 \sum_{n=1}^{\infty} B_n^2 \exp\left(\frac{-\lambda_n V_i}{128 \left(1 + \frac{\mu_a}{\mu_c}\right) d}\right) \right] \quad (7)$$

The circulation patterns and mixing intensity in oscillating drops are not well understood. The values of the mass transfer coefficients are experimentally similar to those predicted by the simple model of Handlos and Baron. This might be due to the internal circulation of drops. Photographic study of oscillating drops by Rose and Kintner shows that the toroidal circulation patterns postulated by Handlos and Baron deviate from reality. When the drop oscillates, the surface area changes with time (Rose, P.M., R.C. Kintner, 1966).

Skelland and Wellek (1964) presented the following empirical equation in which Kronig and Brink's model have taken into account for binary liquid-liquid extraction systems with negligible resistance in continuous phase (Skelland, A.H.P., R.M. Wellek, 1964),

$$\frac{K_d d}{D_d} = 31.4 \left(\frac{4 D_d t}{d^2}\right)^{-0.338} \left(\frac{d V_i^2 \rho_c}{\gamma}\right)^{0.371} Sc_d^{-0.125} \quad (8)$$

Wellek and Skelland (1965), and Patel and Wellek (1967) have also extended the model of Handlos and Baron to include a finite resistance to mass transfer in the continuous phase (Wellek, R.M., A.H.P. Skelland, 1965).

Johnson and Hamielec (1960) suggested the following equation with a new formula for determining \mathfrak{R} (Johnson, A.I., A.E. Hamielec, 1960):

$$K_{od} = \left(-\frac{d}{6t}\right) \ln \left[6 \sum_{n=1}^{\infty} B_n \exp\left(\frac{-4 \lambda_n^2 \mathfrak{R} D_d t}{d^2}\right) \right] \quad (9)$$

Since in their experiments which was the transfer of ethyl acetate into (vigorously circulating) water drops a rapid approach to equilibrium was observed, they considered only the first term of the series in the above equation and determined \mathfrak{R} values:

$$\mathfrak{R} = \frac{d V_i}{2048 D_d (1 + \kappa)} \quad (10)$$

Boyadzhiev *et al.* (1969) presented the following equation for calculating \mathfrak{R} in equation (6) based on their own experimental data (Boyadzhiev, L., D. Elenkov, 1969):

$$\mathfrak{R} = 0.0125 \left[\frac{\text{Re}}{2(1+\kappa)} \right], \quad \text{for} \Rightarrow \frac{D_d t}{d^2} > 10^{-2} \quad (11)$$

Steiner (1986) also used equation (9) reduced to its first term of the summation series and evaluated \mathfrak{R} values on the basis of data from nine sources. The equation proposed was (Steiner, L., 1986):

$$\mathfrak{R} = 1 + 0.177 \text{Re}^{0.43} \text{Sc}_d^{0.23}, \quad \text{for} \Rightarrow \mathfrak{R} < 10 \quad (12)$$

Temos *et al.* (1993) presents the relation between eddy diffusivity and molecular diffusivity in a simple way:

$$\mathfrak{R} = 1 + 0.44 D_g / D_d \quad (13)$$

Where:

$$D_g = 3.29 \times 10^{-4} (\rho_d V_i d / \mu_d) \times \left[1 - \exp \left(-3.29 \times 10^{-4} (\rho_d V_i d / \mu_d) \right) \right] \times \left(\frac{\mu_d}{\rho_d} \right) \quad (14)$$

V_i for $\text{Re} \gg 1$ is given by:

$$V_i = \left\{ 1 - \left[\frac{2 + 3\kappa}{1 + (\mu_d \rho_d / \mu_c \rho_c)^{0.5}} \right] \frac{1.45}{\text{Re}^{0.5}} \right\} V_t \quad (15)$$

Slater (1995) used the Newman equation to calculate the mass transfer coefficient. He used the overall effective diffusivity D_{OE} ($D_{OE} = D_d + D_E$) instead of the molecular diffusivity. The obtained equation is (Slater, M.J., 1995):

$$K_d = - \left(\frac{d}{6t} \right) \text{Ln} \left[\left(\frac{6}{\pi^2} \right) \sum_{n=1}^{\infty} C_n \exp \left(- \frac{4\lambda_n^2 D_{OE} t}{d^2} \right) \right] \quad (16)$$

C_n and λ_n are functions of $\frac{K_c d}{D_{OE}}$ which are calculated by Elzinga and Banchemo (1959).

Kumar and Hartland (1999) proposed a new equation for calculating the overall dispersed phase mass transfer coefficients for circulating and oscillating drops in extraction columns. Reynolds number (Re) and Schmidt number (Sc_d) are defined on the basis of continuous phase and dispersed phase, respectively (Kumar, A., S. Hartland, 1999):

$$\frac{K_d d}{D_d} = 17.7 + \frac{3.19 \times 10^{-3} (\text{Re} \text{Sc}_d^{1/3})^{1.7}}{1 + 1.43 \times 10^{-2} (\text{Re} \text{Sc}_d^{1/3})^{0.7}} \times \left(\frac{\rho_d}{\rho_c} \right)^{2/3} \times \frac{1}{1 + \kappa^{2/3}} \quad (17)$$

$$\kappa = \frac{\mu_d}{\mu_c} \quad (18)$$

Experimental Apparatus:

Set-up Description:

Experimental set-up contains a regular packed column in pilot scale. This column is made of a Pyrex glass tube with inside diameter of 7.2 cm and height of 65 cm. At the lower end of column, there is a discharge

valve and glass entrance nozzle which can be used to connect to different nozzles for dispersed phase inlet. The dispersed phase enters through the bottom of column. The entrance nozzle of the dispersed phase has a valve which controls the speed of drops to the continuous phase. To collect the drops a reversed funnel with wide mouth that is connected to a capillary tube is used.

In this column, Rashig ring packing with the diameter of 1.6 cm and height of 2 cm were used. These packing were put on the top of each other in two rows. Each packing row was placed in a circular structure with a diameter equal to the internal diameter of extraction column. To hold the regular packing in a certain height, thin wires made of stainless steel were used. The drops in heights of 10, 15, 20 cm were collected and analyzed. Figure 1 shows Schematic diagram of the apparatus and Figure 2 shows schematic configuration of each bundle of packed bed.

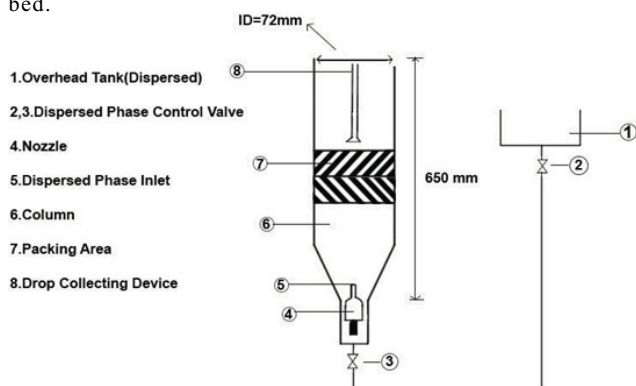


Fig. 1: Schematic diagram of the apparatus

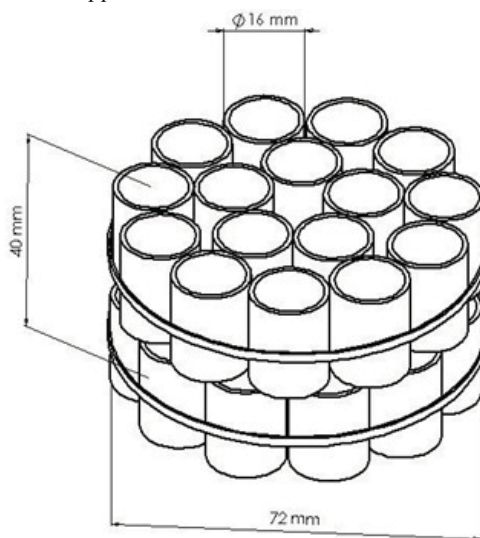


Fig 2: Schematic configuration of each bundle of packed bed

Chemical and Physical Properties of the Systems Used:

The chemical system were distilled water saturated with Toluene as continuous phase, and toluene saturated with distilled water also with a particular percentage of acetic acid as dispersed phase. The physical properties of the systems are tabulated in Table (1).

Table 1: Physical properties of the systems used

	Density (kg / m ³)	viscosity (cp)
Toluene saturated with Water	0.643	858
Water saturated with Toluene	0.931	996
Interfacial tension of the system without solute:	$\gamma = 22 \times 10^{-6}$	

Experimental Results:

In this section the mass transfer coefficients are measured experimentally. Considering the mass balance for a single drop, we get:

$$K_d = \left(-\frac{d}{6t} \right) \ln(1 - E) \tag{19}$$

Where:

$$E = \frac{C_0 - C}{C_0 - C^*} \tag{20}$$

c_0, c, c^* are solute concentration in primary drop (before contact), concentration in specific position and the concentration in equilibrium with continuous phase, respectively which is measured through collecting the drops and doing titration using normal NaoH.

In each experiment and for different heights of the column by measuring Acetic Acid concentration, mean diameter and terminal velocity of droplets and the contact time between two phases, the mass transfer coefficients are calculated considering equation (19). The experimental results for mass transfer coefficient are summarized in Table 2.

Table 2: Experimental values of the mass transfer coefficients (m/s) in different heights of the column

d (cm)	h = 10 cm	h = 15 cm	h = 20 cm
0.65	1.08×10^{-3}	0.94×10^{-3}	0.82×10^{-3}
0.87	1.19×10^{-3}	1.12×10^{-3}	1.09×10^{-3}
0.92	1.32×10^{-3}	1.19×10^{-3}	1.09×10^{-3}
1.02	1.35×10^{-3}	1.24×10^{-3}	1.14×10^{-3}
1.11	1.43×10^{-3}	1.25×10^{-3}	1.18×10^{-3}

Achieving Effective Diffusivity for the System in Different Operating Conditions:

The experimental values of the mass transfer coefficients obtained in the previous section are used in Newman equation to find the effective diffusivity which is replaced with molecular diffusivity. Besides, the Reynolds number is calculated in each experiment considering the physical properties of the continuous phase and terminal velocity of drops. The obtained values of the effective diffusivity versus Reynolds number are drawn in a logarithmic scale. Figures 3 to 5 show the results for three different heights of column. In each figure the best curve is drawn through the data.

The obtained equation for each curve is:

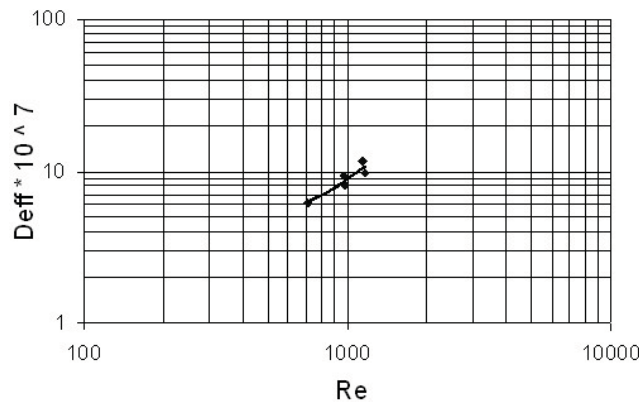


Fig. 3: D_{eff} v.s. Re for $h=10$ cm of the packed column.

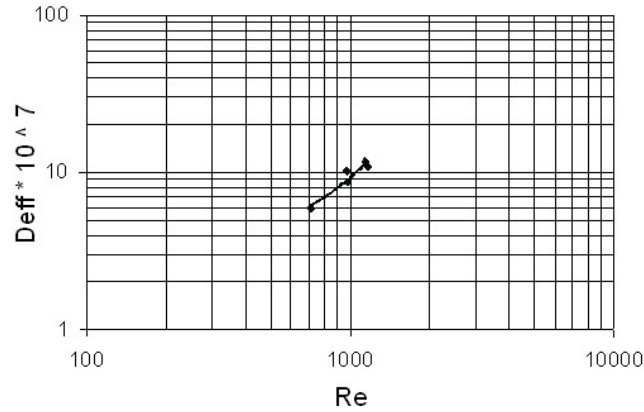


Fig. 4: D_{eff} v.s. Re for $h=15$ cm of the packed column.

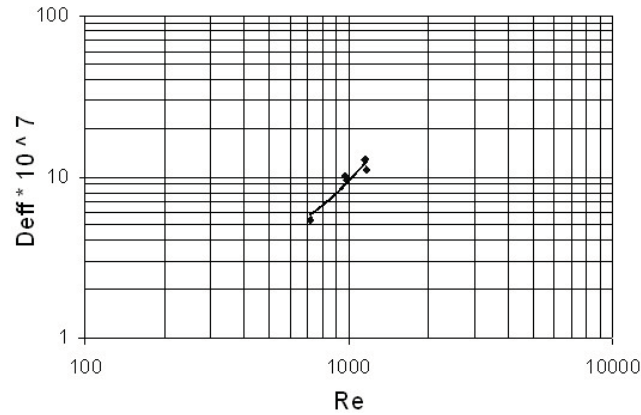


Fig. 5: D_{eff} v.s. Re for $h=20$ cm of the packed column.

$$D_{eff} = 2.6517 \exp(0.0012 Re), \quad h = 10cm \tag{21}$$

$$D_{eff} = 2.26 \exp(0.0014 Re), \quad h = 15cm \tag{22}$$

$$D_{eff} = 1.7492 \exp(0.0017 Re), \quad h = 20cm \tag{23}$$

Considering equations (21) to (23) it is obvious that the values of D_{eff} are increasing while Re is increasing. This happens due to more turbulent regimes in droplets. The slope of D_{eff} v.s. Re is also increased by increasing of the height.

In order to have a unique equation for different heights of the column, the effect of height in calculating the values of effective diffusivity is considered as a linear function of the constants of equations (21) to (23). The obtained equation is:

$$D_{eff} = (-9.025 \times h + 3.5741) \times 10^{-7} \exp [10^{-3} (5h + 0.7) Re] \tag{24}$$

The ratio of effective diffusivity obtained by above equation to actual diffusivity in different heights and Reynolds numbers is shown in table 3.

Table 3: Ratio of effective diffusivity to actual diffusivity in different heights and Reynolds numbers

Re	h (cm)	D_{eff} / D_{actual}
713	10	277
969	10	377
980	10	381
1147	10	466
1168	10	478
713	15	275
969	15	399
980	15	405
1147	15	516
1168	15	532
713	20	262
969	20	405
980	20	412
1147	20	548
1168	20	567

Investigating the Accuracy of the Obtained Equation for the Effective Diffusivity by Calculating the Mass Transfer Coefficients:

Considering the operating conditions and using the equations presented in section 2 and the obtained equation in the previous section mass transfer coefficients for the system are calculated for three different heights and for different droplet diameters and compared with the experimental values presented in Table 2. Since a large number of correlations for mass transfer coefficient are available and some of them were introduced in section 2, only the major correlations were selected for comparison with the experimental data. The results are given in Tables 4 to 6.

Table 4: Calculated values for mass transfer coefficients and the percentage of errors, h = 10 cm

d (cm)	Newman		Kronig and Brink		Handlos and Baron		The obtained equation	
	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %
0.65	5.2×10^{-5}	95.19	1.25×10^{-4}	88.43	1.22×10^{-3}	-12.96	1.098×10^{-3}	-1.67
0.87	6.8×10^{-5}	94.29	1.44×10^{-4}	87.89	1.68×10^{-3}	-41.18	1.236×10^{-3}	-3.87
0.92	7.58×10^{-5}	94.26	1.59×10^{-4}	87.95	1.95×10^{-3}	-47.73	1.246×10^{-3}	5.61
1.02	8.75×10^{-5}	93.52	1.82×10^{-4}	86.52	2.33×10^{-3}	-72.59	1.448×10^{-3}	-7.26
1.11	8.56×10^{-5}	94.01	1.77×10^{-4}	87.62	2.35×10^{-3}	-64.34	1.348×10^{-3}	5.73
d (cm)	Johnson and Hamielec		Steiner		Temos		Lochiel and Calderbank	
	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %
0.65	5.393×10^{-4}	50.06	1.540×10^{-4}	85.74	1.654×10^{-4}	84.69	5.039×10^{-4}	53.34
0.87	6.166×10^{-4}	48.18	1.643×10^{-4}	86.19	2.187×10^{-4}	81.62	5.822×10^{-4}	51.08
0.92	6.299×10^{-4}	52.28	1.724×10^{-4}	86.94	2.251×10^{-4}	82.95	5.949×10^{-4}	54.93
1.02	7.129×10^{-4}	47.19	1.889×10^{-4}	86.01	2.761×10^{-4}	79.55	6.768×10^{-4}	49.87
1.11	6.694×10^{-4}	53.19	1.806×10^{-4}	87.37	2.588×10^{-4}	81.90	6.353×10^{-4}	55.57

Table 5: Calculated values for mass transfer coefficients and the percentage of errors, h = 15 cm

d (cm)	Newman		Kronig and Brink		Handlos and Baron		The obtained equation	
	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %
0.65	4×10^{-5}	95.74	9.89×10^{-5}	89.48	8.7×10^{-4}	7.45	9.791×10^{-4}	-4.16
0.87	5.34×10^{-5}	95.23	1.16×10^{-4}	89.64	1.28×10^{-4}	-14.29	1.118×10^{-4}	0.18
0.92	5.19×10^{-5}	95.64	1.12×10^{-4}	90.59	1.27×10^{-4}	-6.72	1.060×10^{-4}	10.92
1.02	6.13×10^{-5}	95.06	1.29×10^{-4}	89.59	1.56×10^{-4}	-25.81	1.243×10^{-4}	-0.24
1.11	6.12×10^{-5}	95.10	1.29×10^{-4}	89.68	1.61×10^{-4}	-28.8	1.165×10^{-4}	6.80
d (cm)	Johnson and Hamielec		Steiner		Temos		Lochiel and Calderbank	
	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %
0.65	4.665×10^{-4}	50.37	1.299×10^{-4}	86.18	1.399×10^{-4}	85.12	4.351×10^{-4}	53.71
0.87	5.446×10^{-4}	51.38	1.416×10^{-4}	87.36	1.909×10^{-4}	82.96	5.156×10^{-4}	53.96
0.92	5.209×10^{-4}	56.23	1.359×10^{-4}	88.58	1.816×10^{-4}	84.74	4.913×10^{-4}	58.71
1.02	5.964×10^{-4}	51.90	1.495×10^{-4}	87.94	2.263×10^{-4}	81.75	5.655×10^{-4}	54.39
1.11	5.657×10^{-4}	54.74	1.444×10^{-4}	88.45	2.142×10^{-4}	82.86	5.363×10^{-4}	57.09

Table 6: Calculated values for mass transfer coefficients and the percentage of errors, $h = 20$ cm

d (cm)	Newman		Kronig and Brink		Handlos and Baron		The obtained equation	
	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %
0.65	3.29×10^{-5}	95.99	8.41×10^{-5}	89.74	6.8×10^{-4}	17.07	9.086×10^{-4}	-10.80
0.87	4.22×10^{-5}	96.13	9.48×10^{-5}	91.30	9.74×10^{-4}	10.64	1.020×10^{-4}	6.42
0.92	4.2×10^{-5}	96.15	9.3×10^{-5}	91.47	9.8×10^{-4}	10.09	9.708×10^{-4}	10.94
1.02	4.82×10^{-5}	95.77	1.04×10^{-5}	90.88	1.18×10^{-4}	-3.51	1.129×10^{-4}	0.96
1.11	4.69×10^{-5}	96.03	1.0×10^{-5}	91.53	1.19×10^{-4}	-0.85	1.045×10^{-4}	11.44
d (cm)	Johnson and Hamielec		Steiner		Temos		Lochiel and Calderbank	
	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %	Calculated (m/s)	Error %
0.65	4.213×10^{-4}	48.62	1.152×10^{-4}	85.95	1.242×10^{-4}	84.85	3.923×10^{-4}	52.16
0.87	4.874×10^{-4}	55.28	1.233×10^{-4}	88.69	1.677×10^{-4}	84.61	4.591×10^{-4}	57.88
0.92	4.668×10^{-4}	57.17	1.190×10^{-4}	89.08	1.604×10^{-4}	85.28	4.397×10^{-4}	59.66
1.02	5.296×10^{-4}	53.54	1.289×10^{-4}	88.69	1.982×10^{-4}	82.61	5.016×10^{-4}	56.00
1.11	4.959×10^{-4}	57.97	1.224×10^{-4}	89.63	1.850×10^{-4}	84.32	4.696×10^{-4}	60.20

Considering tables (4) to (6), we can see that the calculated values of the mass transfer coefficients by the obtained equation are much closer to the experimental results and are in great compatibility with them. This can be found out by comparing the percentage of errors of the equations as well. Among the theoretical equations Handlos and Baron equation also gives better results than the others and the calculated values are closer to the experimental ones. But the average error of the obtained equation is below 9% in each height. Thus the proposed relation can predict the behavior of the mass transfer coefficients versus variation of droplet diameter and height of the column, perfectly.

Conclusion:

According to the results, the following conclusions are obtained:

- Many correlations are published for prediction of mass transfer coefficients. Although many of them have a good theoretical basis but they do not practically have accurate results.
- Even though Handlos-Baron’s model gives better results among existing models for droplets with relatively large sizes, but we can obtain the mass transfer coefficient with much better precision by applying effective diffusivity in Newman equation.
- While height of packing increases solute concentration decreases and this results in less mass transfer rates.
- Presenting of effective diffusivity as a function of height of packing is a novel approach which can be a conceptual basis for prediction of mass transfer coefficient. This method can be used for more precise design of regular packed columns.

Appendix A. Nomenclature

B_n	Constants of Eqs. (5), (7) and (9)
c	Solute concentration in dispersed phase (Kg/m^3)
c_0	Initial concentration of solute in dispersed phase (Kg/m^3)
c^*	Equilibrium concentration of solute in dispersed phase (Kg/m^3)
C_n	Constant of Eq. (16)
d	Drop diameter (m)
D_d	Molecular diffusivity (m^2/s)
D_{eff}	Effective diffusivity (m^2/s)
D_E	Eddy diffusivity in Temos and Slater equation (m^2/s)

D_{OE}	Overall effective diffusivity (m^2/s)
E	Local efficiency ($= (c_0 - c) / (c_0 - c^*)$)
h	Height of packings (m)
K_c	Continuous phase mass transfer coefficient (m/s)
K_d	Dispersed phase mass transfer coefficient (m/s)
K_{od}	Overall dispersed phase mass transfer coefficient (m/s)
\mathfrak{R}	Enhancement factor for mass transfer
Re	Reynolds number ($= \rho_c V d / \mu_c$)
Sc	Schmidt number ($= \mu / \rho D_a$)
t	Contact time(s)
V	Drop velocity (m/s)
V_i	Terminal velocity of a single drop (m/s)

Greek Symbols:

β_n	Constant of equation (1)
γ	Interfacial tension (N/m)
κ	Ratio of dispersed phase viscosity to continuous phase viscosity ($= \mu_d / \mu_c$)
λ_n	Constants of Eqs. (1), (5), (7), (9) and (16)
μ	Viscosity (Pas)
ρ	Density (kg/m^3)

Subscripts:

c	Continuous phase
d	Dispersed phase

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