

# Investigation of Mass Transfer Coefficient under Jetting Conditions in a Liquid-Liquid Extraction System

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**ABSTRACT:** In this research mass transfer coefficient under jetting regime in different directions (from dispersed to continuous and continuous to dispersed phase) has been studied using an experimental setup. *n*-Butanol-succinic acid-water with low interfacial tension has been selected as experimental chemical system. The effects of various parameters such as jet velocity, nozzle diameter and the height of the continuous phase above the nozzle, on mass transfer coefficient have been investigated. A correlation has also been derived in order to predict the mass transfer coefficient as a function of physical properties of both phases and aforementioned parameters. Based on the experimental results, mass transfer coefficient increases with an increase in the nozzle diameter and jet velocity, while increasing the height of the continuous phase above the nozzle decreases the mass transfer coefficient. These results may reveal the importance of mass transfer during the jet formation and breakage.

**KEY WORDS:** Liquid-liquid extraction, Mass transfer coefficient, Jet velocity, Dispersed phase, Continuous phase, Nozzle diameter.

## INTRODUCTION

Liquid-liquid extraction as well as several types of equipment such as mixer-settlers, packed columns, agitated towers, spray columns and perforated plate

columns employ diffusion process. In these devices, mass transfer is facilitated by dispersing one liquid phase into the other.

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When a liquid is injected into another liquid, several breakup modes are observed. For low injection velocities droplets are formed periodically at nozzle tip and no jet is observed. This breakup mode is called dripping. As the injection velocity is increased to the jetting velocity,  $U_{oj}$ , jetting mode appears. The length of the jet increases with the injection velocity up to  $U_{max}$ , the jet breaks up into the drops. For velocities greater than  $U_{max}$ , the jet breaks up and the length of the jet decreases. At still higher injection velocities, the jet breaks up by atomization, where many non-uniform droplets are formed near the nozzle tip. So we can see two regimes, drop and jet, in extraction.

In general, studies about drop(s) and jet(s) are divided into two parts, hydrodynamics and mass transfer. Therefore the design of liquid-liquid extraction equipment requires an understanding of hydrodynamics and mass transfer.

The hydrodynamics aspects of jet behavior in liquid-liquid systems have been studied by many researchers [1-17].

The first theoretical study of the breakup of a liquid jet in another immiscible liquid was done by *Tomotika* [1], who extended the linear stability analysis of *Rayleigh* [2]. *Tomotika's* equation has often been used to estimate the jet length and the droplet size for real jets in the axisymmetric jetting mode. *Meister & Scheele* [3-5] using the *Tomotika* stability equation, developed an expression for the drop size which was an improvement over the equation used by *Christiansen* [6]. They tried to develop an understanding of the jet and drop formation based on experiments obtained using fifteen liquid-liquid systems.

*Kitamura et al.* [7] experimentally varied the motion of the continuous phase to be either faster, same as, or slower than the jet down to the case of a stagnant continuous phase. They found that the jet shortened as the absolute value of the continuous phase velocity related to the jet increased from zero. They found that *Tomotika's* theory well predicts the size of the droplets when the surrounding fluid moves with the same speed as the jet.

*Bright* [8] attempted to perform a linear viscous stability analysis to calculate the disturbance wavelength assuming a constant, but unequal, velocity in each liquid phase and to predict drop volumes (after jet breakup) using *Tyler* [9] analysis.

*Richards et al.* [10-12] carried out direct numerical simulations of the formation of an axisymmetric jet and its breakup into droplets using the Volume of Fluid (VOF) method.

*Homma et al.* [13] investigated numerically the formation of a liquid jet and its breakup into droplets. *Richard and Scheele* [14] experimentally developed a flash photolysis dye technique to measure the velocity profiles in liquid-liquid jets and compared their experimental profiles with the models of *Yu & Scheele* [15] and *Gospodinov et al.* [16]. Although they found reasonable agreement with jet radius and the velocity profiles with both models at the highest Reynolds number examined, the agreement deteriorated at lower Reynolds number as the interface contraction increased.

*Garner et al.* [17] derived the steady state velocity distribution of liquid jet by neglecting the inertial terms in the momentum equation.

These studies have contributed significantly to our understanding of the formation of a jet and the parameters that related to the jet dynamics.

*Meister & Scheele* [18] measured jet length during the transfer of acetone between benzene jets and water. They found that mass transfer in both directions stabilized the jet relative to the non-mass transfer case, and that transfer into the jet produced significantly longer jets than transfer out.

*Sawistowski* [19], in contrast, noted that jet breakup in liquid-liquid systems is accelerated by mass transfer into a jet and decelerated by transfer out.

*Burkholde & Berg* [20, 21] studied the effect of mass transfer on laminar liquid jet breakup for liquid jets in gases and liquid jets in liquids. For the latter case, they predicted mass transfer of a surface tension lowering solute either into or out of the jet may be either stabilizing or destabilizing, depending on physical properties of the system and mass transfer rate.

*Skelland & Huang* [22] obtained a correlation of jet length for mass transfer out of the jet.

*Coyle et al.* [23] performed a linear stability analysis for liquid-liquid viscous jet assuming dilute solute mass transfer, with undisturbed solute concentration linear within jet and uniform outside, while the undisturbed velocity profile corresponded to plug flow in the jet. The difficulty with this solution is that the base state plug flow profile is inconsistent with the actual profile

in viscous liquid-liquid jets and the local nature for the analysis.

Kimura & Miyauchi [24] measured the approximate interfacial velocity by photographic technique in order to measure the rate of mass transfer in liquid-liquid laminar jet with diffusion equation.

Skelland & Huang [22] attempted to apply a Graetz-type analysis, regarding the jet as a straight cylinder. They have used penetration theory for mass transfer coefficient in jet (dispersed phase controlled), plug velocity distribution, with low rates of mass transfer were assumed and the results compared rather successfully with experimental data in liquid-liquid extraction systems.

As can be seen, the studies of mass transfer in liquid-liquid systems associated with jet phenomena are very scarce and normally based on simplifying assumptions. Moreover, most researches investigated the effect of mass transfer on jet length and did not consider the mass transfer coefficient. This is due to experimental difficulties and mathematical complexities. Therefore investigation of mass transfer under jetting conditions in liquid-liquid systems and measurements of mass transfer coefficient seems necessary.

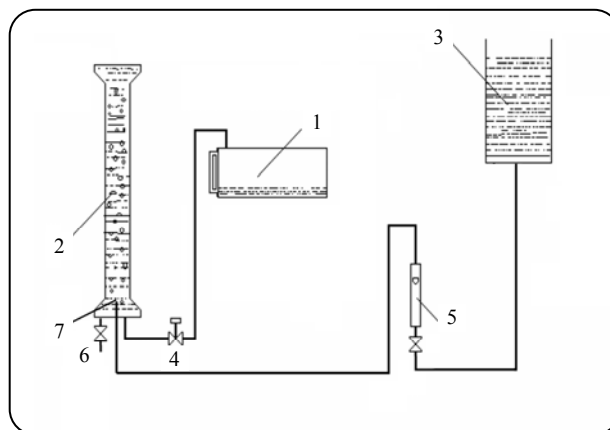
In this research mass transfer coefficient during jet formation and breakage in turbulent condition (the value of Re number in turbulent condition for jets is  $Re \geq 10-30$  [25, 26]) and low interfacial tension system was measured. It should be stressed that most of previous studies in liquid-liquid extraction are conducted in high interfacial tension systems. The mass transfer coefficients were measured at different heights of continuous phase which addresses the question whether mass transfer during jet formation is important or not and what portion of total mass transfer in an equipment occurs at this stage. The mass transfer coefficients have been measured in both directions and the direction in which the major effect on mass transfer rate happens determined.

## EXPERIMENTAL SECTION

### Experimental setup

The experimental setup (Fig.1) consisted of the following parts:

- 1- Water feed tank
- 2- A cylindrical vessel (10 cm diameter and 50 cm height) fitted with stainless steel base which contains continuous phase and is used as the extraction column.



**Fig. 1: Experimental setup:** (1) water feed tank; (2) cylindrical vessel; (3) scaled cylindrical vessel; (4) needle valve; (5) rotameter; (6) drain valve; (7) nozzle.

3- A scaled cylindrical vessel equipped with a control valve with a 0 to 10 mm opening to adjust the dispersed phase flowrate (and subsequently tuning Reynolds numbers). This vessel serves as the reservoir for dispersed phase.

4- Valve

5- Rotameter

6- Drain valve

7- Nozzles with dimensions of 1, 3, 5 mm inner diameters.

Both vessels and all nozzles were made of Pyrex.

### Chemicals

The chemical system was chosen according to the standard test system recommended by EFCE [27], n-butanol-succinic acid-water, as a typical example of liquid-liquid extraction processes, with low interfacial tension and partial mutual miscibility. Succinic acid was of analytical grade, but n-butanol was of industrial grade and distilled water was used as a continuous phase. At the beginning of the experiments the organic and aqueous phases were mutually saturated by another to prevent multi component diffusion in the phases and hence, succinic acid was the only diffusing species in the chemical system.

### Analytical procedure

The experiments were conducted in both directions and at 20°C. The concentration of solute was 20 g/L. When mass transfer direction was from continuous to dispersed dissolved in water and when the mass transfer

Table 1: Physical properties of the system at 20°C.

Physical property	d → c	c → d
$\rho_c$ (kg/m <sup>3</sup> )	987.4	993.5
$\rho_d$ (kg/m <sup>3</sup> )	855.5	847.3
$\mu_c$ (kg/m.s)	0.00134	0.00139
$\mu_d$ (kg/m.s)	0.0037	0.0032
$\sigma$ (dyn/cm)	2.12	2.01

$$D_c = 0.52 \times 10^{-9} \text{ (m}^2/\text{s)}$$

$$D_d = 0.22 \times 10^{-9} \text{ (m}^2/\text{s)}$$

The range of operating conditions is as follows: Temperature 20 °C. Nozzle diameter,  $d_N = 1, 3, 5$  mm, Reynold number for  $d_N = 1$  mm,  $Re\ 1 = 426.790$ ,  $Re\ 2 = 495.995$ ,  $Re\ 3 = 582.362$ . Reynold number for  $d_N = 3$  mm,  $Re\ 1 = 306.681$ ,  $Re\ 2 = 364.353$ ,  $Re\ 3 = 435.268$ . Reynold number for  $d_N = 5$  mm,  $Re\ 1 = 179.828$ ,  $Re\ 2 = 240.061$ ,  $Re\ 3 = 300.022$ . Height of continuous phase,  $H = 3.5, 6.5, 12.5$  cm.

direction was from dispersed to continuous phase the solute was dissolved in n-butanol.

Run times varied with nozzle diameter, dispersed phase flowrate and the height of continuous phase above the nozzles. The volumetric flowrate of the dispersed phase was measured by timing the change in level of scaled cylindrical vessel. The samples were collected and analyzed by titration with 0.1 M sodium hydroxide in presence of phenolphthalein indicator.

Each experiment was repeated five times. Thus each data point shown on the figures represents the mean value of at least five measurements of the outlet concentration of organic phase. The continuous phase was replaced in each experiment to avoid any contamination and to minimize the experimental error.

Physical properties of the system are given in Table 1.

Properties were measured using a pycnometer, Oswald viscometer and tensiometer. Molecular diffusion coefficients were reported by *Misek* [28]. The equilibrium distribution coefficient for succinic acid between water and butanol at 20°C is given by [29]:

$$C_d = 1.086 C_c - 0.849 \times 10^{-3} C_c^2 - 0.162 \times 10^{-4} C_c^3 \quad (1)$$

Where  $C_d$  and  $C_c$  are concentrations of succinic acid in butanol and water, respectively.

#### MASS TRANSFER COEFFICIENT CALCULATION

The extraction efficiency could be expressed as follows in both directions:

$$E(\%) = 100 \frac{C_{d,out} - C_{d,in}}{C_d^* - C_{d,in}} \quad c \rightarrow d \quad (2)$$

Where  $C_{d,in}$  and  $C_{d,out}$  are concentrations of succinic acid in the inlet and outlet of organic solutions respectively and  $C_d^*$  is the equilibrium concentration of the solute in aqueous phase.

$$E(\%) = 100 \frac{C_{d,in} - C_{d,out}}{C_{d,in} - C_d^*} \quad d \rightarrow c \quad (3)$$

For mass transfer direction from dispersed to continuous phase,  $C_d^*$  is equal to zero because the concentration of solute in aqueous phase is zero while for mass transfer from continuous to dispersed phase  $C_d^*$  is calculated from Eq. (1). Because the interfacial area between phases is not known the overall volumetric mass transfer coefficient,  $K_L a$ , was determined from the experimental results. The overall volumetric mass transfer coefficient is defined by the following equation:

$$R = Q_d (C_{d,in} - C_{d,out}) = K_L a V_c \Delta C_m \quad (4)$$

Where  $R$ ,  $Q_d$ ,  $V_c$ ,  $\Delta C_m$  are the extraction rate, volumetric flowrate of organic phase, contacting device volume and appropriate mean concentration driving force respectively. Since the state of mixing of the two phases was not known, the logarithmic mean concentration driving force,  $\Delta C_{ln}$ , was used to calculate the overall volumetric mass transfer coefficient,  $K_L a$ . The logarithmic mean concentration driving force,  $\Delta C_{ln}$ , is given by:

$$\Delta C_{ln} = \frac{(C_{d,out}^* - C_{d,out}) - (C_{d,in} - C_{d,in}^*)}{\ln[(C_{d,out}^* - C_{d,out}) / (C_{d,in} - C_{d,in}^*)]} \quad (5)$$

$U_{oj}$ ,  $U_{max}$ ,  $U$  and  $A_Z$  are defined as follows [27, 22]:

$$U_{oj} = \left[ \left( \frac{0.8 \mu_d g d_N}{\sigma} \right)^2 + \frac{3 \sigma / (\rho_d d_N)}{1 + d_N (\Delta \rho g / 2 \sigma)^{1/2}} \right]^{1/2} \quad (6)$$

$$\frac{0.8 \mu_d g d_N}{\sigma} \quad (7)$$

$$U_{max} = 2.94 \left( \frac{\sigma}{\rho_d d_N} \right)^{1/2} \quad (7)$$

$$U = \frac{Q}{A_Z} \quad (8)$$

$$A_Z = \pi \left( \frac{d_N + d_{jc}}{2} \right) L_j \quad (9)$$

## RESULTS AND DISCUSSION

### Effect of the height of continuous phase

In order to investigate the mass transfer rate as well as mass transfer coefficient during jet formation, the extraction column was filled with different heights of continuous phase.

In the lowest height of continuous phase only jet can be seen while in the other heights of continuous phase, jet and drops resulting from jet breakup are visible. Figs. 2 and 3 (8 and 9) demonstrate the dependency of the overall volumetric mass transfer coefficient,  $K_{La}$ , versus the height of continuous phase,  $H$ , at different values of Reynolds number (nozzle diameters). Reduction of volumetric mass transfer coefficient is observed when the height of continuous phase is increased. By increasing the height of continuous phase, the static pressure on the jet is increased which in turn results in decreasing the degree of turbulence of the dispersed phase. As we know when the turbulence decreases, the viscous forces dominate inertial forces and the motion of eddies and as a consequence diffusivity of eddies decreases resulting in less mixing. On the other hand by increasing the height of continuous phase, the contacting device volume is increased and for a constant extraction rate, the overall volumetric mass transfer coefficient will decrease (see Eq. (4)).

Figs. 4 and 5 (6 and 7) show an increase in the extraction efficiency,  $E$ , when the height of continuous phase,  $H$ , increases. When the height of continuous phase increases, the jet length decreases and interfacial area between the jet and the continuous phase decreases (see Eq. (9)) and the volumetric flowrate of dispersed phase will decrease. Therefore the continuous and dispersed phases have enough time for transferring the solute to each other. Since the extraction efficiency is related to the amount of solute which has been extracted by the opposite phase, the contacting time for transferring the solute between phases is increased and therefore the extraction efficiency increases.

### Effect of nozzle diameter

Figs. 10 and 11 (12 and 13) demonstrate the dependency of the extraction efficiency on the nozzles diameter. These figures show that an increase in nozzle diameter exhibits a minimum in extraction efficiency. With increasing the nozzle diameter from 1mm to 3mm,

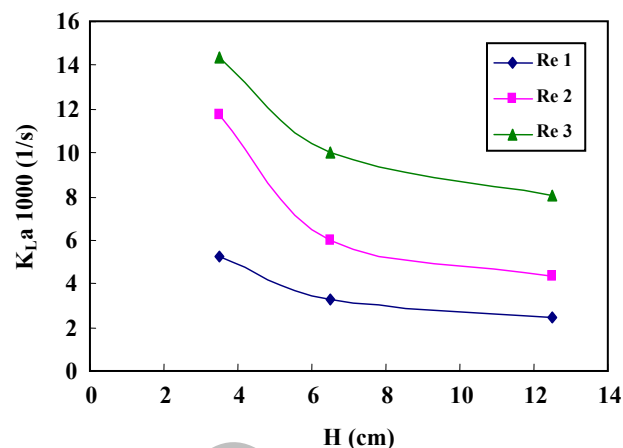


Fig. 2: Effect of continuous phase height on the overall volumetric mass transfer coefficient.  $d_N=3$  mm ( $d \rightarrow c$ ).

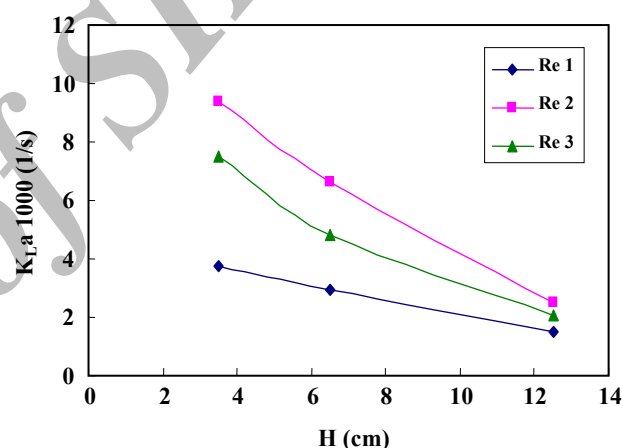


Fig. 3: Effect of continuous phase height on the overall volumetric mass transfer coefficient.  $d_N=3$  mm ( $c \rightarrow d$ ).

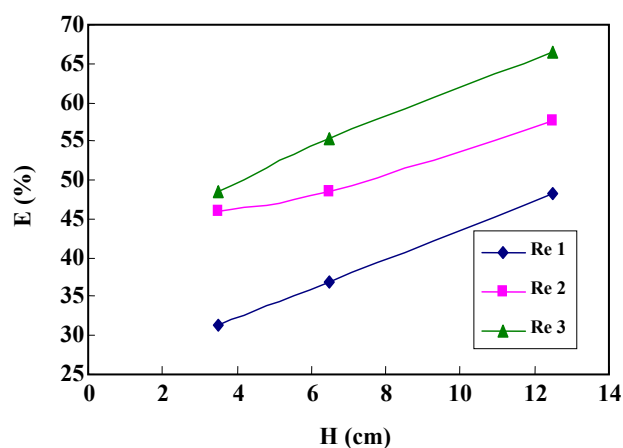


Fig. 4: Effect of continuous phase height on the extraction efficiency.  $d_N=3$  mm ( $d \rightarrow c$ ).

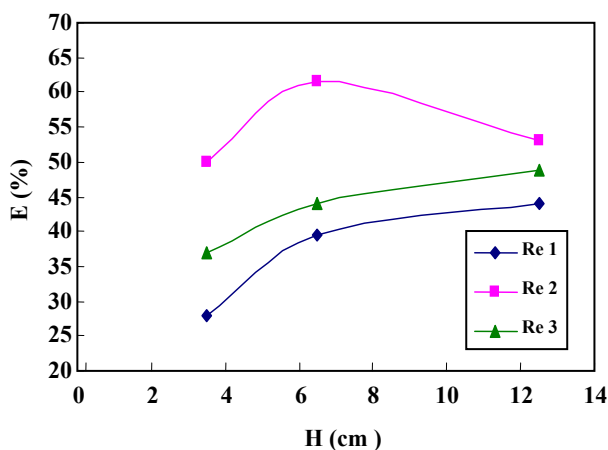


Fig. 5: Effect of continuous phase height on the extraction efficiency.  $d_N = 3$  mm ( $c \rightarrow d$ ).

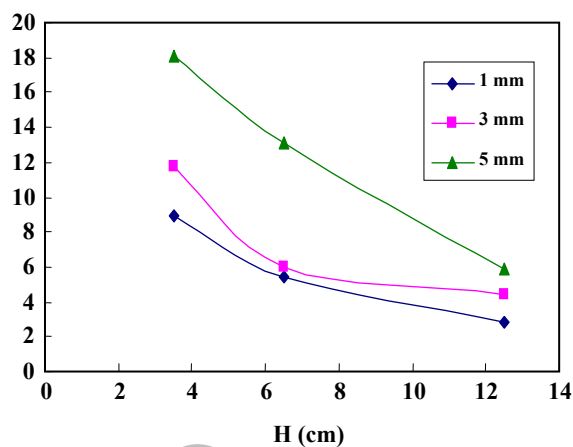


Fig. 8: Effect of continuous phase height on the overall volumetric mass transfer coefficient. Re 2 ( $d \rightarrow c$ ).

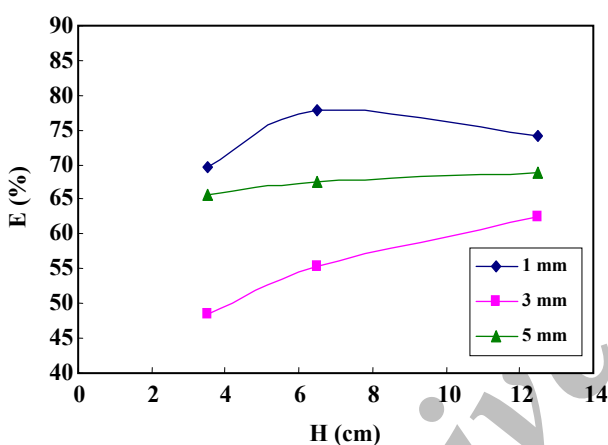


Fig. 6: Effect of continuous phase height on the extraction efficiency. Re 3 ( $d \rightarrow c$ ).

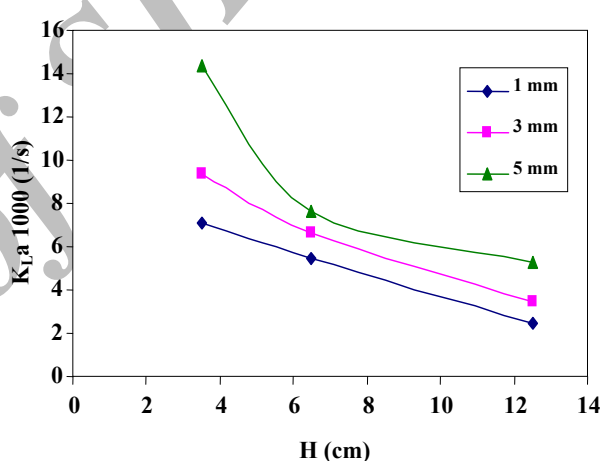


Fig. 9: Effect of continuous phase height on the overall volumetric mass transfer coefficient. Re 2 ( $c \rightarrow d$ ).

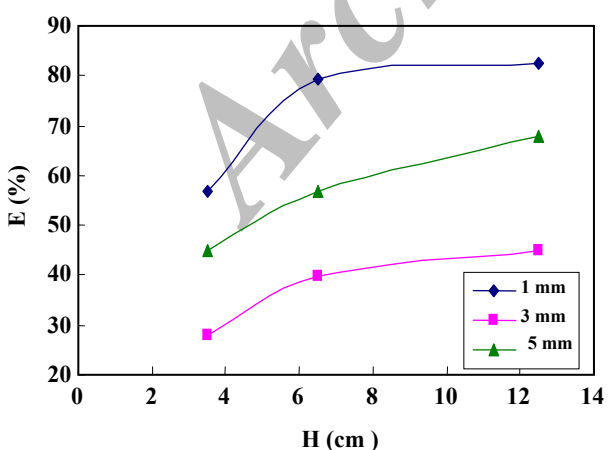


Fig. 7: Effect of continuous phase height on the extraction efficiency. Re 1 ( $c \rightarrow d$ ).

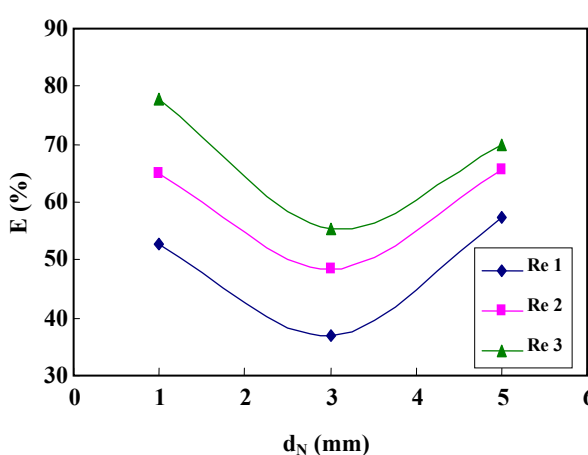


Fig. 10: Effect of nozzle diameter on the extraction efficiency.  $H = 6.5$  cm.

$(C_{d,in}-C_{d,out})$  decreases due to the decline in velocity and turbulence. However, by increasing the nozzle diameter up to 5mm,  $(C_{d,in} - C_{d,out})$  increases. This behavior can be explained as the result of increased contact time between jet and continuous phase which enhances the mass transfer rate.

Figs. 14 and 15 (16 and 17) show that by increasing the nozzle diameter, the volumetric flowrate of dispersed phase increases and at constant height of continuous phase and constant contacting device volume, the volumetric mass transfer coefficient increases (see Eq. (4)). On the other hand with increasing the nozzle diameter at constant height of continuous phase, the jet length decreases and as a consequence the individual mass transfer coefficient increases and this results in higher volumetric mass transfer coefficient.

#### Effect of jet velocity

Figs. 18 and 19 demonstrate the dependency of the overall volumetric mass transfer coefficient,  $K_L a$ , on the jet velocity,  $U$ , (or  $Re$  number), at different height of continuous phase,  $H$ . As it can be seen,  $K_L a$  increases by increasing the jet velocity.

It is obvious that at constant nozzle diameter and constant height of continuous phase, with increasing the jet velocity, the volumetric flowrate of dispersed phase and Reynolds number increases. This directly influences the overall volumetric mass transfer coefficient. Furthermore, this behavior is the consequence of an increase in the mixing and turbulence of the system which leads to an increase in the surface renewal mechanism, the eddy diffusivity and interfacial mass transfer.

#### Effect of mass transfer direction

Many researchers have obtained correlations of mass transfer coefficients for dispersed and continuous phases using some assumptions amongst which *Withman's* two-film theory and penetration theory can be mentioned.

The individual mass transfer coefficients in each phase can be estimated by the following relations [22, 27]:

$$k_d = (k_p^*)_{av} = 2 \left( \frac{D_d}{\pi t_e} \right)^{1/2} \quad t_e = \frac{L_j}{U} \quad (10)$$

$$Sh_c = \left[ \frac{2}{\pi^{0.5}} \right] Re^{0.5} Sc_c^{0.5} \quad k_c = \frac{Sh_c D_c}{d_N} \quad (11)$$

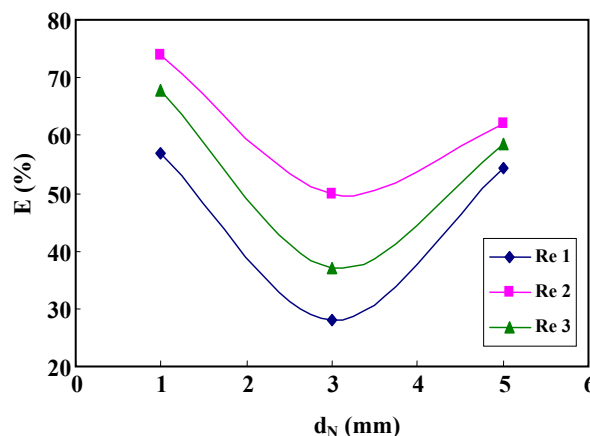


Fig. 11: Effect of nozzle diameter on the extraction efficiency.  $H=3.5\text{cm}$ . ( $c \rightarrow d$ ).

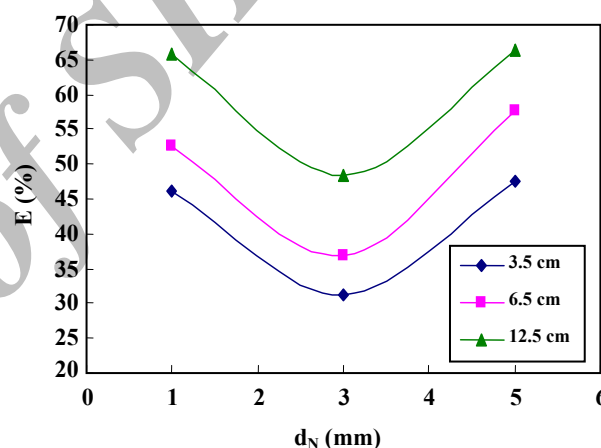


Fig. 12: Effect of nozzle diameter on the extraction efficiency.  $Re\ 1$  ( $d \rightarrow c$ ).

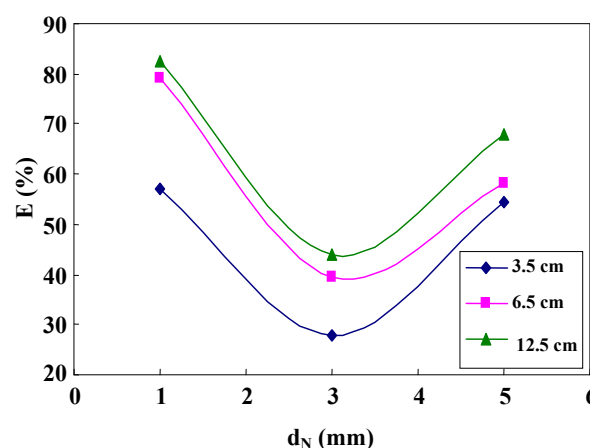


Fig. 13: Effect of nozzle diameter on the extraction efficiency.  $Re\ 1$  ( $c \rightarrow d$ ).

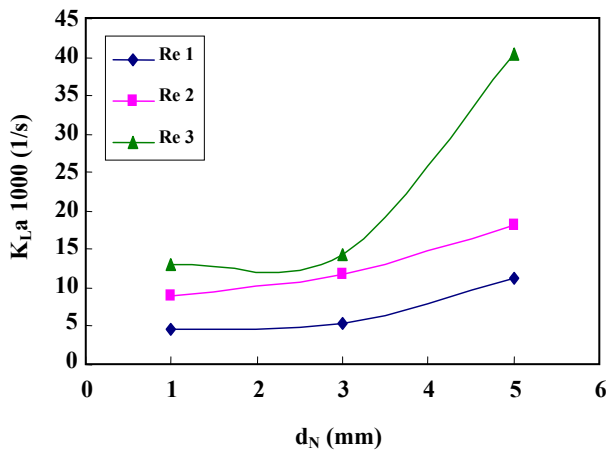


Fig. 14: Effect of nozzle diameter on the overall volumetric mass transfer coefficient.  $H = 3.5$  cm ( $d \rightarrow c$ ).

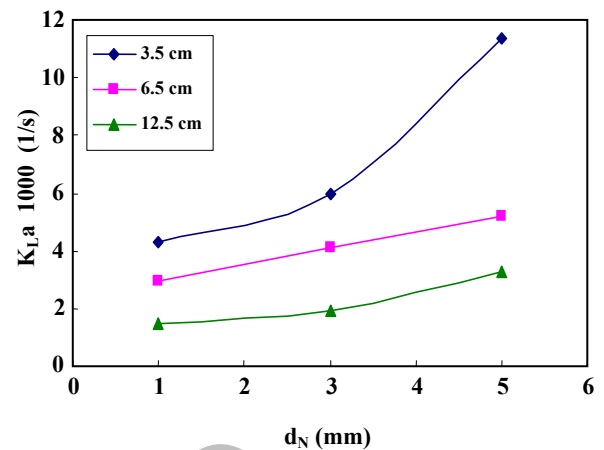


Fig. 17: Effect of nozzle diameter on the overall volumetric mass transfer coefficient. Re 1 ( $c \rightarrow d$ ).

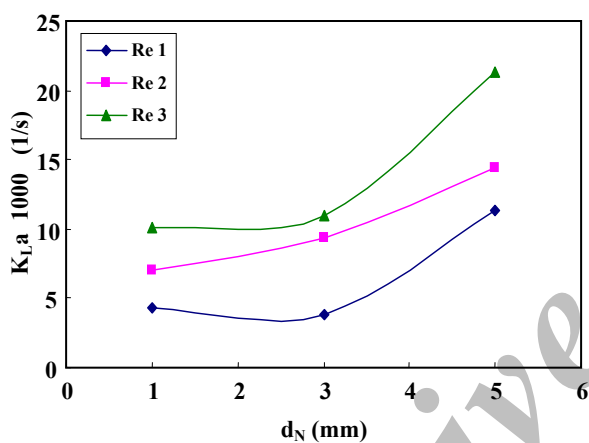


Fig. 15: Effect of nozzle diameter on the overall volumetric mass transfer coefficient.  $H = 3.5$  cm ( $c \rightarrow d$ ).

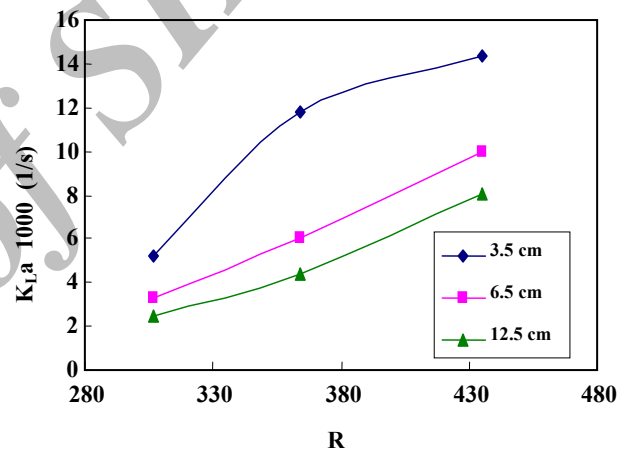


Fig. 18: Effect of jet velocity on the overall volumetric mass transfer coefficient.  $d_N = 3$  mm ( $d \rightarrow c$ ).

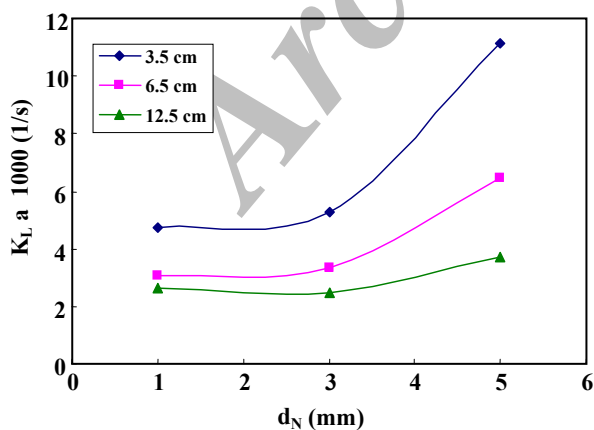


Fig. 16: Effect of nozzle diameter on the overall volumetric mass transfer coefficient. Re 1 ( $d \rightarrow c$ ).

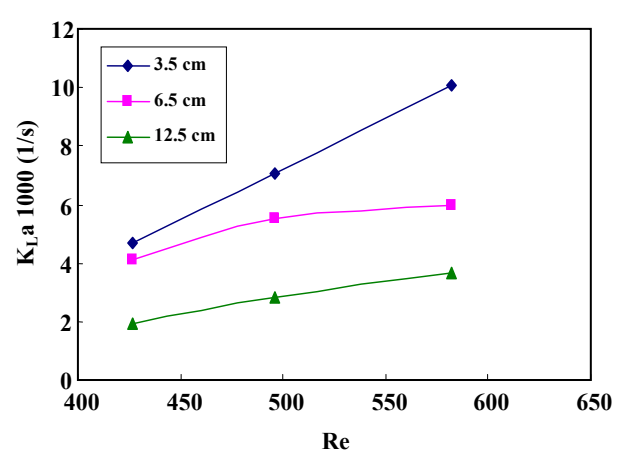


Fig. 19: Effect of jet velocity on the overall volumetric mass transfer coefficient.  $d_N = 1$  mm ( $c \rightarrow d$ ).



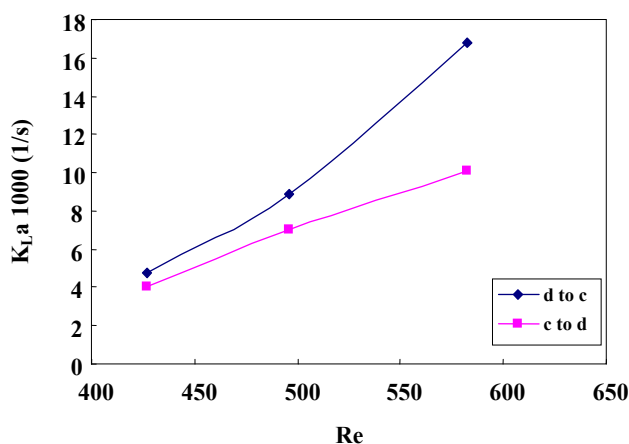


Fig. 20: Effect of mass transfer direction on the overall volumetric mass transfer coefficient.  $d_N = 1\text{ mm}$  and  $H = 3.5\text{ cm}$ .

The overall mass transfer coefficient is:

$$\frac{1}{K_{Ld}} = \frac{1}{k_d} + \frac{1}{m_{cd}k_c} \quad m_{cd} = \frac{\Delta C_c}{\Delta C_d} \quad (12)$$

When the mass transfer direction is from continuous to dispersed phase we have longer jets and more stability and it means that jet breaks up later than the opposite direction. Therefore, longer jets result in smaller individual mass transfer coefficient ( $k_d$ ) as well as smaller overall mass transfer coefficient.

On the other hand, for the mass transfer direction from dispersed to continuous, the drops formed in jet breakup have larger diameters and as the drop diameter becomes bigger the inner circulation increases and the resistance for mass transfer decreases resulting in mass transfer coefficient increase. These results are shown in Figs. 20 and 21.

### Correlation of data

In view of the complexity of mass transfer in actual equipment, fundamental equations for mass transfer in actual equipment are rarely available, and empirical methods, guided by dimensional analysis and by semi-theoretical analogies, are relied upon to give workable equations [39]. Thus dimensional analysis is used to plan the experiments and to interpret the results in the form of dimensionless groups which can be employed to obtain a relationship between the measured quantities. It is assumed that the following independent variables should play significant role in this process:

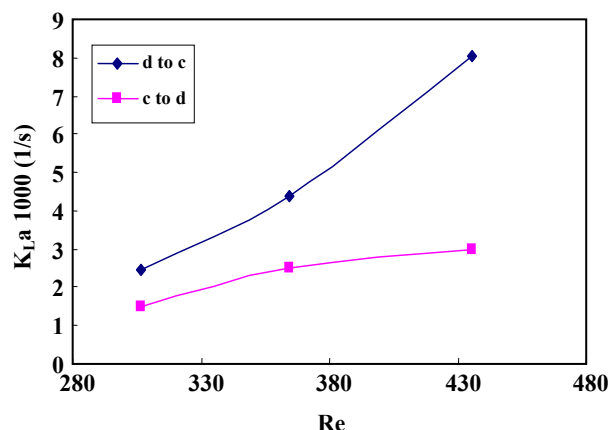


Fig. 21: Effect of mass transfer direction on the overall volumetric mass transfer coefficient.  $d_N = 3\text{ mm}$  and  $H = 12.5\text{ cm}$ .

$$K_{La} = f(\rho_c, \rho_d, \mu_c, \mu_d, \sigma, D_c, D_d, d_N, U, H, L_j) \quad (13)$$

Eq. (13) can be transformed into following groups by using the Buckingham Pi theorem:

$$\frac{K_{La} d_N}{U} = N(We^{-1})^A (Re_d^{-1})^B (Pe_d^{-1})^C (Pe_c^{-1})^D \times \left(\frac{\mu_c}{\rho_d U d_N}\right)^E \left(\frac{\rho_c}{\rho_d}\right)^F \left(\frac{L_j}{d_N}\right)^G \left(\frac{H}{d_N}\right)^M \quad (14)$$

The constant  $N$  and the exponents  $A, B, C, D, E, F, G$  and  $M$  are obtained by the least squared method. The final relationship which fits the experimental data satisfactorily is:

$$K_{La} = 7.306 We^{3.65} Re^{-9.22} Pe_d^{0.407} Pe_c^{3.74} L_j^{0.136} H^{-0.594} U^{2.313} d_N^{1.3} \quad (15)$$

### CONCLUSIONS

With increasing the height of continuous phase at constant nozzle diameter, the overall volumetric mass transfer coefficient decreases while the extraction efficiency increases. An increase in the nozzle diameter at constant continuous phase height, leads to an increase in volumetric mass transfer coefficient.

An increase in jet velocity at constant nozzle diameter (constant height of continuous phase) leads to an increase in the jet momentum, and hence increases the mixing and turbulence. This in turn increases the extraction efficiency and the volumetric mass transfer coefficient.

The importance of mass transfer direction is also investigated. It is observed that when the mass transfer direction is from dispersed to continuous phase, higher volumetric mass transfer coefficient is detected.

The major differences between this research and previous studies can be summarized as:

- 1- Considering turbulent flow (jet)
- 2 The effect of continuous phase height to be able to compare jet and droplet mass transfer coefficients
- 3- The influence of direction of solute transfer on mass transfer coefficient which has not been previously studied in a low interfacial tension system (n-butanol, succinic acid, water).

Clearly further studies with different immiscible systems are needed to enable generalization of the results. The correlation obtained in this study is valid for Re numbers between 100 and 1000, and more experimentation is needed to expand the results beyond this range.

Investigating the effect of adding surfactants to the system under study is also recommended.

#### Nomenclatures

$A_z$	Interfacial area between the jet and the continuous phase ( $\text{cm}^2$ )
$C$	Concentration of solute ( $\text{kg m}^{-3}$ )
$d_{jc}$	Jet diameter at breakup (mm)
$d_N$	Nozzle diameter (mm)
$D$	Molecular diffusion coefficient ( $\text{m}^2 \text{s}^{-1}$ )
$E$	Extraction efficiency
$H$	Height of continuous phase (cm)
$k$	Individual mass transfer coefficient, ( $\text{cm s}^{-1}$ )
$(k_p^*)_{av}$	Mass transfer coefficient for use with arithmetic mean driving force ( $\text{cm s}^{-1}$ )
$K_{La}$	Overall volumetric mass transfer coefficient ( $\text{s}^{-1}$ )
$L_j$	Length of the jet (cm)
$m_{cd}$	Equilibrium distribution coefficient
$Pe$	Peclet number ( $Pe = Re.Sc$ )
$Q$	Volumetric flow rate ( $\text{cm}^3 \text{s}^{-1}$ )
$R$	Extraction rate ( $\text{g s}^{-1}$ )
$Re$	Jet Reynolds number ( $Re = \rho_d U_d d_N / \mu_d$ )
$Sc$	Schmidt number, $Sc = \mu / \rho D$
$t_e$	Time of exposure of mass transfer (s)
$U$	Jet velocity ( $\text{m s}^{-1}$ )
$U_{oj}$	Jetting velocity ( $\text{m s}^{-1}$ )
$U_{max}$	Injection velocity where maximum breakup length is observed ( $\text{m s}^{-1}$ )

$V_c$	Contacting device volume ( $\text{cm}^3$ )
$We$	Weber number ( $U^2 d_N \rho_d / \sigma$ )

#### Greek symbols

$\Delta$	Difference operator
$\mu$	Viscosity ( $\text{kg m}^{-1} \text{s}^{-1}$ )
$\rho$	Density ( $\text{kg m}^{-3}$ )
$\sigma$	Interfacial tension ( $\text{N m}^{-1}$ )

#### superscripts and subscripts

$c$	Continuous phase
$d$	Dispersed phase
$in$	For inlet
$m$	Mean
$o$	For outlet
$org$	Organic phase
$*$	Equilibrium value

#### Abbreviation

EFCE	European Federation of Chemical Engineering
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