

# DISPERSED PHASE HOLDUP IN A PULSED DISC AND DOUGHNUT EXTRACTION COLUMN

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(Submitted: September 26, 2010 ; Revised: March 16, 2011 ; Accepted: March 22, 2011)

**Abstract** - Dispersed phase holdup has been measured in a 76 mm diameter pulsed disc and doughnut extraction column for three different liquid-liquid systems. The effects of operational variables such as pulsation intensity and dispersed and continuous phase flow rates on holdup have been investigated and found to be significant. The existence of three different operational regimes, namely mixer-settler, dispersion, and emulsion regimes, was observed when the energy input was changed. The results indicated that the characteristic velocity approach is applicable to this type of extraction column for analysis of holdup in the transition and emulsion regions. Empirical correlations are derived for prediction of the dispersed phase holdup in terms of the physical properties of the system and the operating variables in different operating regimes. Good agreement between prediction and experiments was found for all operating conditions investigated.

**Keywords:** Dispersed phase holdup; Pulsed disc and doughnut column; Slip velocity; Characteristic velocity.

## INTRODUCTION

Various types of solvent extraction contactors have been used for a range of applications in the hydrometallurgical, pharmaceutical, petrochemical, and nuclear industries for many years. In recent years, a relatively new type of solvent extraction contactor, called the pulsed disc and doughnut extraction column, has been developed and extensively applied to reprocess spent nuclear fuel in France and Japan (Angelov and Journe, 1990; Angelov et al., 1998). On an industrial scale, the pulsed disc and doughnut column has also been installed at the Western Mining Corporation (WMC), Olympic Dam, South Australia, for extraction of uranium. In comparison to mixer-settlers, which are frequently used in the mining industry and nuclear fuel reprocessing, this extractor is attractive from both safety and economic standpoints, in particular its simplicity of design, smaller space consumption, reduction of organic loss and no internal moving parts (Movsowitz et al., 1997).

Introduction of a solvent into an extractor produces droplets that undergo repeated coalescence and breakage, leading to an equilibrium drop size distribution. The resulting fractional volumetric holdup is defined as the volume fraction of the active section of the column that is occupied by the dispersed phase:

$$x_d = \frac{v_d}{v_e} \quad (1)$$

where  $v_d$  represents the volume of the dispersed phase and  $v_e$  the total volume of the two phases for the effective length of the column (Oliveira et al., 2008; Stella et al., 2008).

The dispersed phase holdup is a key variable in the design of solvent extraction columns as it is related to the interfacial area for mass transfer. Prediction of dispersed phase holdup is also important in relation to the flood point of a pulsed disc and doughnut column. Upon increasing flow rates, a second interface forms at some points in the

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column and the column is thus characterized as flooded and any further increase in the dispersed phase flow is rejected from the column resulting in inefficient operating conditions. The dispersed phase holdup will tend to increase unstably as the flood point is approached. The dispersed phase holdup is also necessary to calculate the slip velocity between the phases. The slip velocity is the single most important parameter for the control of mass transfer.

At present, little information has appeared in the literature regarding the performance of pulsed disc and doughnut extraction column. Many investigations have concentrated on the development of numerical solutions to describe the operational characteristics of the column (Aoun Nabli et al., 1998; Aoun Nabli and Guiraud, 1997; Bardin-Monnier and Guiraud, 2003); however, no firm design and scale-up criteria have yet been proposed.

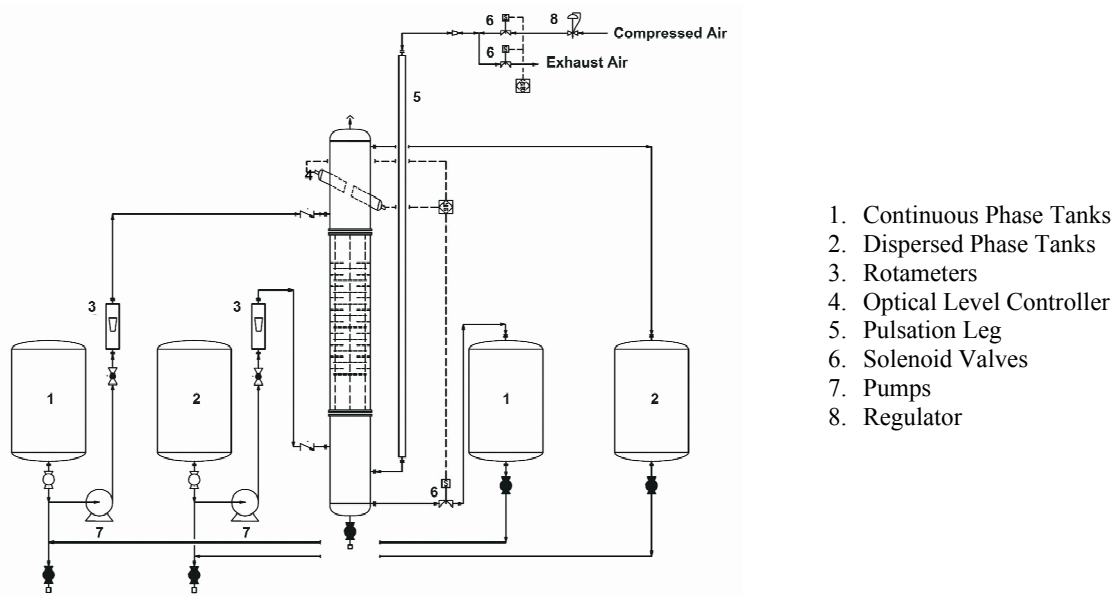
The present paper describes an investigation of dispersed phase holdup of a pulsed disc and doughnut column as a function of pulsation intensity and phase flow rates with three different systems. Holdup data are correlated in terms of phase flow rates by the characteristic velocity method. Empirical correlations are developed for dispersed phase holdup in terms of physical properties and operating variables for different operating regimes.

## EXPERIMENTAL

A schematic diagram of the experimental apparatus is shown in Figure 1. The main column section consisted of a 74 cm long glass tube of 76 mm internal diameter, enclosing a stack of 30 pairs of disc and

doughnut, made of 2 mm stainless steel sheet. The discs and doughnuts were arranged alternately and spaced 10 mm apart, resulting in a 20 mm compartment height; they were held in place by means of three 3.2 mm o.d. SS tie rods with SS spacer sleeves. The discs were 67 mm in diameter and the doughnut apertures were 36 mm, giving an open free area of 23.5%. A settler of 112 mm diameter at each end of the column permitted the liquids to coalesce and decant separately. The column was pulsed by blowing air at the required amplitude and frequency into the pulse leg. The air pressure was controlled by a regulator to provide pulses of the required amplitude in the column while the frequency of the pulses was controlled using two solenoid valves. The inlet and outlet of the column were connected to four tanks, each of 80 l capacity. The flow rates of the two phases were indicated by two rotameters. The liquid-liquid interface was maintained approximately 250 mm above the top compartment. This was achieved by using an optical sensor. A solenoid valve (a normally closed type) was provided at the outlet stream of the heavy phase. This valve received electronic signals from the sensor. When the interface location was going to change, the optical sensor sent a signal to the solenoid valve and the aqueous phase was allowed to leave the column by opening the diaphragm of the solenoid valve. The organic phase was allowed to leave the column via overflow.

The studied liquid-liquid systems were kerosene-water, toluene-water, and n-butyl acetate-water. Distilled water was used as the continuous phase and technical grade solvents were used as the dispersed phase. The physical properties of the system used in these experiments are listed in Table 1. The physical properties were measured under equilibrium conditions.



**Figure 1:** Schematic diagram of the pulsed disc and doughnut column

**Table 1: Physical properties of liquid systems at 20°C (Míšek et al., 1985; Jahya et al., 1999)**

Physical property	Kerosene-water	Toluene-water	n-Butyl acetate- water
$\rho_c$ (kg / m <sup>3</sup> )	998	998.2	997.6
$\rho_d$ (kg / m <sup>3</sup> )	804	865.2	880.9
$\eta_c$ (mPa.s)	1.0	0.963	1.0274
$\eta_d$ (mPa.s)	1.66	0.584	0.734
$\sigma$ (mN / m)	46.5	36.0	14.1

When starting a run, the solvent and water phases were first mutually saturated by repeated circulation through the column. After filling the column with the continuous phase, the dispersed phase was introduced. The amplitude and frequency of pulsation were next adjusted to the desired values. The column was operated for a period before steady state was reached. After this period, the holdup was measured. The interface position was then maintained at the desired height and the system was allowed to reach the steady state. The period of running was doubled and holdup was measured again. If there is no change then it can be assumed that it is at the steady state. In these experiments, 3-4 changes of the column volume were needed to reach steady state conditions. All experiments were carried out far from flooding conditions.

Holdup measurements were obtained using the shut down (displacement) method. In order to use this method, at the end of a run, the inlet and outlet valves were shut simultaneously and the dispersed phase was allowed to disengage to the interface at the top of the column. A period of 10-15 min was allowed for the dispersed phase to settle. The change in interface height between operation and after settling was measured and then converted into the corresponding volume to determine holdup. In the settler of an extraction column, there are no internals and agitation and the droplets are very large compared to the active part of the column. Consequently, the value of the holdup is very low in the settler compared to the active part of the column. So, the method has enough accuracy for measuring of holdup in the column.

Visual observation of the column in operation with these three liquid systems indicated that three regimes, mixer-settler, dispersion (transition), and emulsion, occurred depending upon pulsation intensity. A stepwise movement of the droplets within the column indicated the occurrence of the mixer-settler regime. The existence of the dispersion regime was revealed by a non-uniform drop size distribution and no coalescence of the dispersed phase droplets. The emulsion regime followed the dispersion regime and was characterized by a uniform distribution of dispersed phase drops in the continuous phase in each compartment of the column.

## RESULTS AND DISCUSSION

The effect of pulsation intensity on dispersed phase holdup is shown in Figure 2. In the mixer-settler regime at low pulsation intensity, the dispersed phase holdup is high due to the presence of thick layers of the dispersed phase under the discs and doughnuts. As the pulsation intensity increases, large drops formed due to pulsation do not have enough time to collect under the internals. These large drops have a low residence time. Therefore, the holdup decreases with an increase in pulsation intensity and reaches a minimum value. This minimum corresponds to the transition from the mixer-settler to the dispersion regime. As shown in Fig. 2, dispersed phase holdup increases with an increase in pulsation intensity in both the dispersion and emulsion regimes. The inertial and shear forces on droplets increase with an increase in pulsation intensity. These forces enhance drop breakage. It appears that the number of drops in the column increases due to the decrease of the relative velocity between the two phases and, consequently, the value of dispersed phase holdup will increase in these regimes.

The rising velocity of the dispersed phase droplets in the aqueous phase depends on drop diameter. The interfacial tension is the physical property that has greatest effect on drop size. Therefore, the dispersed phase holdup is expected to vary with interfacial tension. As shown in Fig. 2, holdup decreases with an increase in interfacial tension. It is known that the size of the drops increases with increasing interfacial tension. Therefore, the residence time of the drops decreases and, consequently, dispersed phase holdup will increase.

The effect of dispersed phase velocity on holdup is given in Figure 3. As can be seen in this figure, the dispersed phase holdup increases with an increase in dispersed phase velocity in all operating regimes. The number of dispersed drops increases with an increase in dispersed phase velocity and, consequently, holdup of the dispersed phase will increase according to Equation (1).

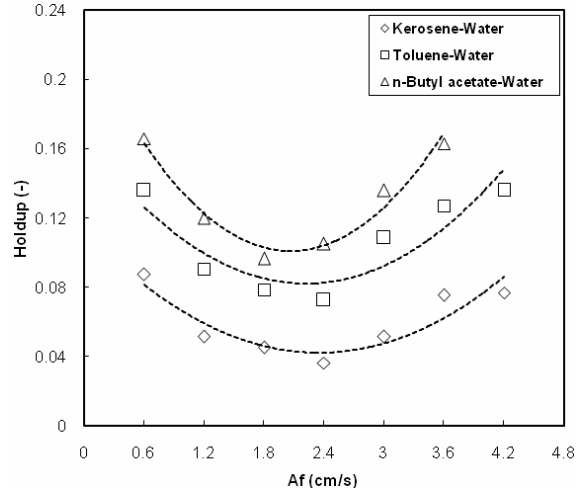


Figure 2: Effect of pulsation intensity on dispersed phase holdup ( $V_c=V_d=1.132$  mm/s)

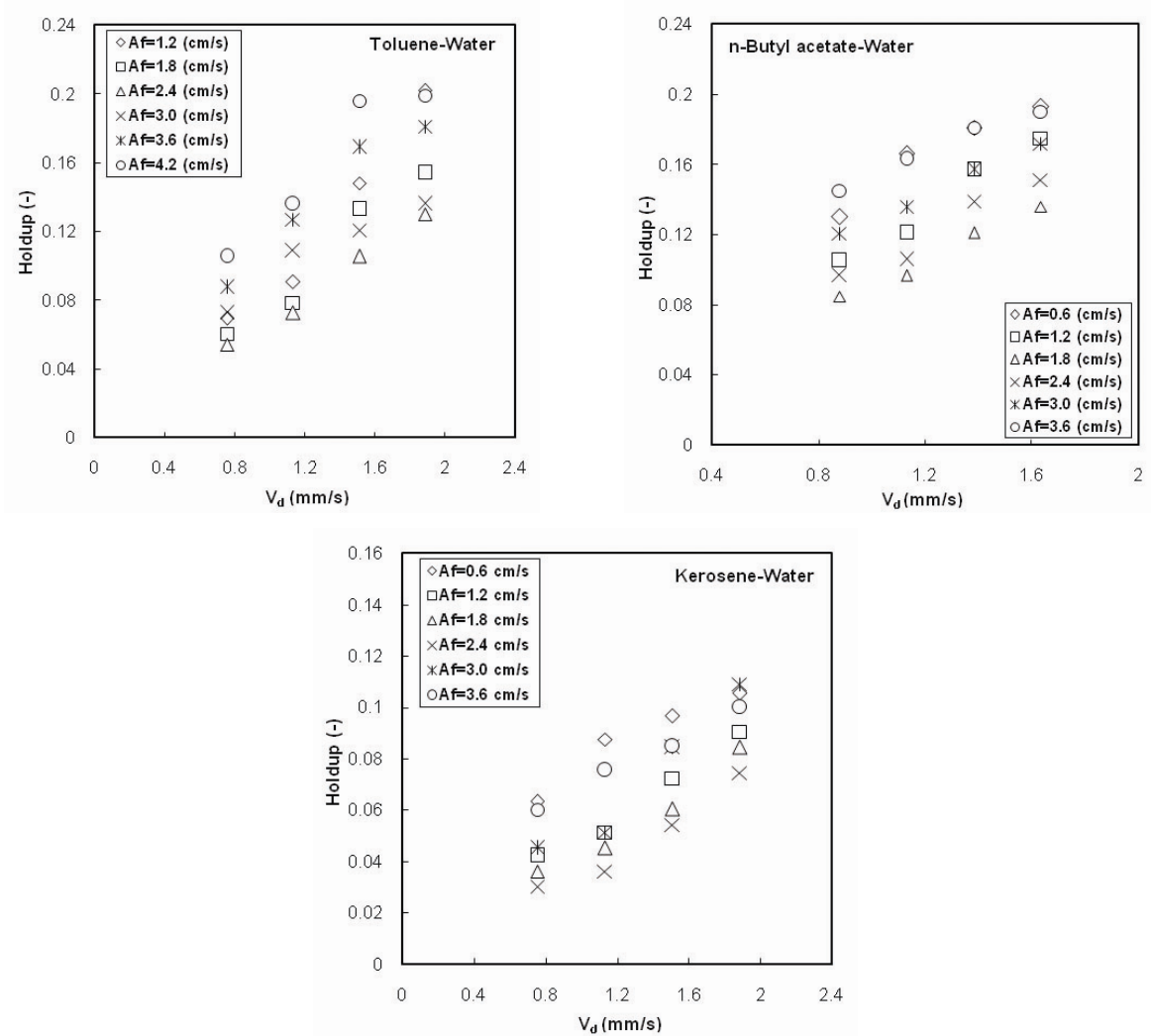


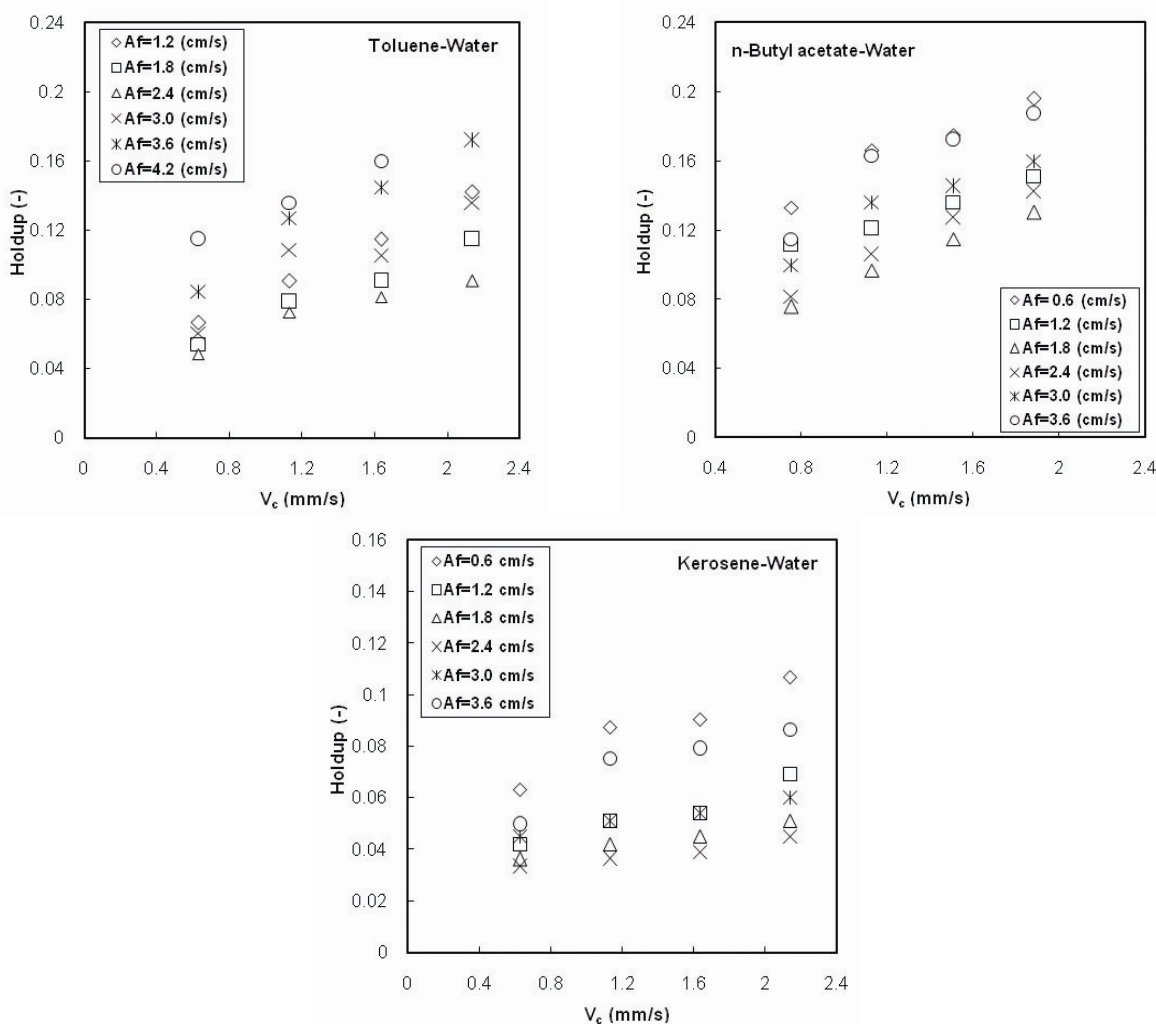
Figure 3: Effect of dispersed phase velocity on dispersed phase holdup at constant  $V_c$  ( $V_c=1.132$  mm/s)

Figure 4 illustrates the effect of continuous phase velocity on dispersed phase holdup. By increasing the continuous phase velocity, the drag force between the dispersed drops and the continuous phase increases, so the drop movement will be limited and the residence time will increase. Consequently, the value of the holdup increases with increasing continuous phase velocity. Comparing Figures 3 and 4, it can be seen that the effect of dispersed phase velocity on holdup is larger than that of continuous phase velocity. The operating variables and experimental results are listed in the Appendix.

The concept of characteristic velocity (Gayler et al, 1953),  $V_k$ , is very useful for relating the dispersed phase holdup and the phase flow rates:

$$V_s = \frac{V_d}{x_d} + \frac{V_c}{1-x_d} = V_k(1-x_d) \quad (2)$$

where  $V_s$  is the slip (or relative) velocity between the continuous and dispersed phase through the column. The applicability of the above equation to the present column depends on the linearity of the characteristic velocity plot. Typical graphs of this plot are shown in Figure 5 using Equation (2). The results confirm that the method is applicable to the present column in the transition and emulsion regions. Even if there is a little discrepancy in the graphs, this approach can be used for the column like similar results observed by other investigators (Jahya et al., 2000). However, a marked discrepancy was observed in the values of  $V_k$  obtained in the mixer-settler regime. The resulting values of  $V_k$  are listed in Table 2. As shown in this Table, the characteristic velocity decreases with an increase in pulsation intensity. Table (2) also shows that the characteristic velocity increases with an increase in interfacial tension.



**Figure 4:** Effect of continuous phase velocity on dispersed phase holdup at constant  $V_d$  ( $V_d=1.132$  mm/s)

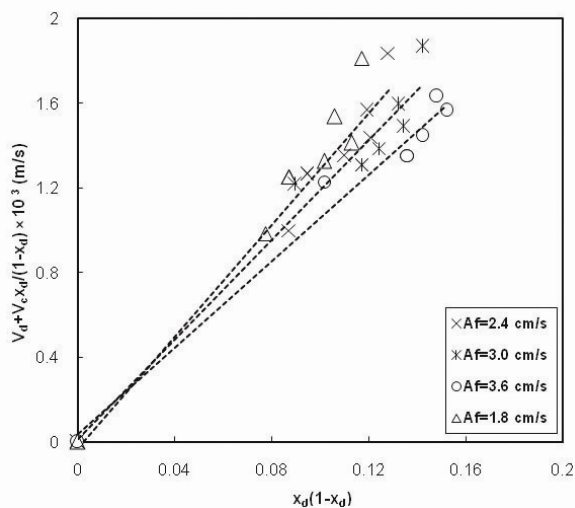


Figure 5: Characteristic velocity plots for the n-butyl acetate-water system

Table 2: The values of characteristic velocity for the transition and emulsion regions

Af (cm/s)	V <sub>k</sub> (m/s)	V <sub>k</sub> (m/s)	V <sub>k</sub> (m/s)
	n-Butyl acetate-Water	Toluene-Water	Kerosene-Water
1.8	0.0145	-	-
2.4	0.0136	0.0183	0.0299
3.0	0.0117	0.0152	0.0237
3.6	0.0106	0.0123	0.0189
4.2	-	0.0112	0.0179

Jahya et al. (2009) proposed the following correlation for prediction of the characteristic velocity in terms of the physical properties of the systems and the column variables in the dispersion and emulsion regimes:

$$V_k = 6.889 \times 10^{-4} \left( \frac{d}{d_a} \right)^{1.59} \left( \frac{A^2 \rho_c g}{\sigma} \right)^{0.13} \left( \frac{\sigma}{\mu_c} \right) \left( \frac{\rho_c \sigma^4}{\psi \mu_c^5} \right)^{0.32} \left( \frac{d_a \rho_c \sigma}{\mu_c^2} \right)^{0.47} \left( \frac{\mu_c^4 g}{\Delta \rho \sigma^3} \right)^{0.81} \left( \frac{\Delta \rho}{\rho_c} \right)^{0.98} \left( \frac{\mu_d}{\mu_c} \right)^{0.12} \quad (3)$$

where

$$\psi = \frac{\pi^2 (1 - \varepsilon^2)}{2 \varepsilon^2 C_o^2 h_c} (Af)^3.$$

Equation (3) was used for estimation of characteristic velocity obtained in the present work. The comparison of the experimental results with those predicted by Eq. (3) is illustrated in Fig. 6. As can be

seen in this figure, the experimental results are in very good agreement with the calculated values obtained by using Eq. (3) for both the toluene-water and n-butyl acetate-water systems. However, a relatively marked deviation is observed for the characteristic velocity data of the kerosene-water system.

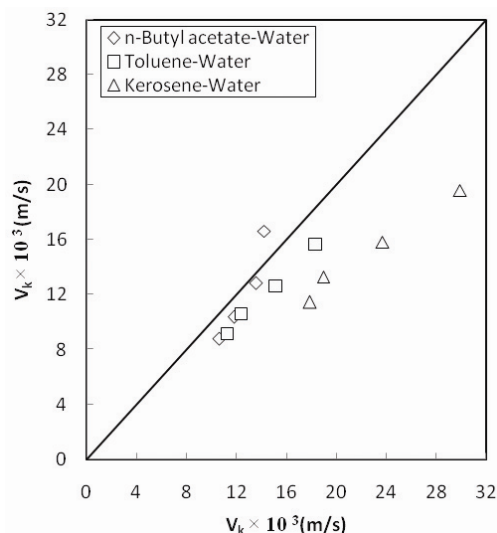


Figure 6: Comparison of experimental results with the values calculated using Eq. (3)

Empirical correlations are also derived for dispersed phase holdup for different operating regimes in the present work. The values of holdup,  $x_d$ , are correlated with physical properties of the systems and the operating variables as follows:

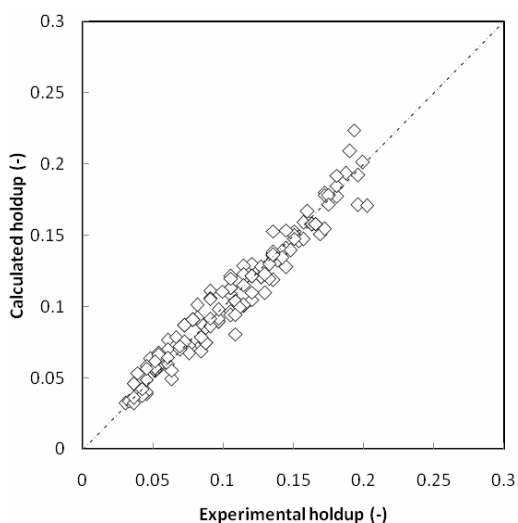
$$x_d = 2.57 \left( \frac{Af^4 \rho_c}{g\sigma} \right)^{-0.095} \left( \frac{V_d^4 \rho_c}{g\sigma} \right)^{0.35} \left( \frac{\mu_d^4 g}{\rho_c \sigma^3} \right)^{-0.06} \left( 1 + \frac{V_d}{V_c} \right)^{-0.88} \left( \frac{\Delta\rho}{\rho_c} \right)^{-0.91} \quad (4)$$

(mixer-settler)

$$x_d = 12.31 \left( \frac{Af^4 \rho_c}{g\sigma} \right)^{0.20} \left( \frac{V_d^4 \rho_c}{g\sigma} \right)^{0.32} \left( \frac{\mu_d^4 g}{\rho_c \sigma^3} \right)^{-0.08} \left( 1 + \frac{V_d}{V_c} \right)^{-0.92} \left( \frac{\Delta\rho}{\rho_c} \right)^{-0.60} \quad (5)$$

(transition and emulsion)

The experimental values of the dispersed phase holdup are compared with those predicted by the proposed correlations (Eqs. (4) and (5)) in Figure 7. This figure indicates that the suggested correlations are in good agreement with experimental data. Equations (4) and (5) reproduce the experimental results with average relative deviations (ARD) of 8.81% and 9.50%, respectively.



**Figure 7:** Comparison of experimental data for dispersed phase holdup with calculated values

In order to test the validity and applicability of the suggested correlations, the data of two investigators (Hocq, 1994; Milot, 1990) were selected for use in this work. The comparison of the experimental results with calculated values is given in Table 3. The results show that there is reasonable agreement between the calculated values and experimental results obtained by Milot (1990). However, a large deviation is observed for Hocq's (1994) experimental data. In those experiments, the organic phase (light phase) was used as the continuous phase and the aqueous phase (heavy phase) was the dispersed phase. This might be the main reason for this marked deviation between experimental data and calculated values.

**Table 3: Comparison of experimental and calculated values of dispersed phase holdup (H1-H4= experimental data of Hocq (1994), M1-M3= experimental data of Milot (1990))**

Case	$x_d$ (exp.)	$x_d$ (cal.)
H1	0.0525	0.0182
H2	0.0625	0.0224
H3	0.090	0.0437
H4	0.0933	0.0563
M1	0.213	0.172
M2	0.268	0.223
M3	0.296	0.212

## CONCLUSIONS

This paper presents an experimental study of the dispersed phase holdup of a pulsed disc and doughnut extraction column. The results show that the holdup is drastically affected by the pulsation intensity and interfacial tension. Significant, but weaker, are the effects of the continuous and dispersed phase velocities. Three operational regions, described as the mixer-settler, transition, and emulsion regimes, are observed. These are primarily controlled by the pulsation intensity. The dispersed phase holdup data are well correlated in terms of phase flow rates by the characteristic velocity in the transition and emulsion regimes. Additionally, empirical correlations are proposed to predict the dispersed phase holdup as a function of the physical properties of the systems and the operating variables for different operating regimes. Since there is little experimental data on this type of extraction column, the present work is of use to those desiring to use this type of contactor.

## ACKNOWLEDGMENTS

The authors are grateful to Prof. G. W. Stevens (Department of Chemical and Biochemical Engineering, The University of Melbourne) for providing some important references and for his useful information regarding the concept of characteristic velocity in extraction columns.

## NOMENCLATURE

A	amplitude of pulsation	m
C <sub>o</sub>	Orifice coefficient	(-)
d	column diameter	m
d <sub>a</sub>	doughnut aperture diameter	m
f	frequency of pulsation	s <sup>-1</sup>
g	acceleration due to gravity	m/s <sup>2</sup>
h <sub>c</sub>	compartment height	m
v	volume	m <sup>3</sup>
V	superficial velocity	m/s
V <sub>s</sub>	slip velocity	m/s
V <sub>k</sub>	characteristic velocity	m/s
x <sub>d</sub>	dispersed phase holdup	(-)

### Greek Symbols

σ	interfacial tension	N/m
Δρ	density difference between phases	kg/m <sup>3</sup>
μ	viscosity	Pa.s
ρ	density	kg/m <sup>3</sup>
ε	fractional free area	(-)
Ψ	power dissipated per unit mass	m <sup>2</sup> /s <sup>3</sup>

### Subscripts

c	continuous phase
d	dispersed phase

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## APPENDIX

The experimental values obtained in the present work are listed in Table A-1 to A-3.

**Table A-1: Experimental data for the toluene-water system**

No.	A (cm)	f (1/s)	Q <sub>c</sub> (l/h)	Q <sub>d</sub> (l/h)	x <sub>d</sub>	Operating regime
1	1.2	0.5	18	18	0.135	mixer-settler
2	1.2	1	18	12	0.0694	mixer-settler
3	1.2	1	18	18	0.0905	mixer-settler
4	1.2	1	18	24	0.147	mixer-settler
5	1.2	1	18	30	0.202	mixer-settler
6	1.2	1	10	18	0.0664	mixer-settler
7	1.2	1	26	18	0.115	mixer-settler
8	1.2	1	34	18	0.142	mixer-settler
9	1.2	1.5	18	12	0.603	mixer-settler
10	1.2	1.5	18	18	0.0785	mixer-settler
11	1.2	1.5	18	24	0.132	mixer-settler
12	1.2	1.5	18	30	0.153	mixer-settler
13	1.2	1.5	10	18	0.0543	mixer-settler
14	1.2	1.5	26	18	0.0906	mixer-settler
15	1.2	1.5	34	18	0.114	mixer-settler
16	1.2	2	18	12	0.054	transition
17	1.2	2	18	18	0.0724	transition
18	1.2	2	18	24	0.0815	transition
19	1.2	2	18	30	0.0905	transition
20	1.2	2	10	18	0.0483	transition
21	1.2	2	26	18	0.0815	transition
22	1.2	2	34	18	0.0906	transition
23	1.2	2.5	18	12	0.0725	transition
24	1.2	2.5	18	18	0.108	transition
25	1.2	2.5	18	24	0.121	transition
26	1.2	2.5	18	30	0.136	transition
27	1.2	2.5	10	18	0.0603	transition
28	1.2	2.5	26	18	0.1056	transition
29	1.2	2.5	34	18	0.136	transition
30	1.2	3	18	12	0.0875	emulsion
31	1.2	3	18	18	0.126	emulsion
32	1.2	3	18	24	0.169	emulsion
33	1.2	3	18	30	0.181	emulsion
34	1.2	3	10	18	0.0845	emulsion
35	1.2	3	26	18	0.145	emulsion
36	1.2	3	34	18	0.172	emulsion
37	1.2	3.5	18	12	0.105	emulsion
38	1.2	3.5	18	18	0.136	emulsion
39	1.2	3.5	18	24	0.192	emulsion
40	1.2	3.5	18	30	0.199	emulsion
41	1.2	3.5	10	18	0.114	emulsion
42	1.2	3.5	26	18	0.16	emulsion
43	1.2	3.5	34	18	0.151	emulsion

**Table A-2: Experimental data for the n-butyl acetate-water system**

No.	A (cm)	f (1/s)	Q <sub>c</sub> (l/h)	Q <sub>d</sub> (l/h)	x <sub>d</sub>	Operating regime
1	1.2	0.5	18	14	0.13	mixer-settler
2	1.2	0.5	18	18	0.166	mixer-settler
3	1.2	0.5	18	22	0.181	mixer-settler
4	1.2	0.5	18	26	0.193	mixer-settler
5	1.2	0.5	12	18	0.132	mixer-settler
6	1.2	0.5	24	18	0.175	mixer-settler
7	1.2	0.5	30	18	0.196	mixer-settler
8	1.2	1	18	14	0.105	mixer-settler
9	1.2	1	18	18	0.120	mixer-settler
10	1.2	1	18	22	0.157	mixer-settler
11	1.2	1	18	26	0.175	mixer-settler
12	1.2	1	12	18	0.111	mixer-settler
13	1.2	1	24	18	0.136	mixer-settler
14	1.2	1	30	18	0.151	mixer-settler
15	1.2	1.5	18	14	0.0845	transition
16	1.2	1.5	18	18	0.0966	transition
17	1.2	1.5	18	22	0.121	transition
18	1.2	1.5	18	26	0.135	transition
19	1.2	1.5	12	18	0.0754	transition
20	1.2	1.5	24	18	0.114	transition
21	1.2	1.5	30	18	0.129	transition
22	1.2	2	18	14	0.0966	transition
23	1.2	2	18	18	0.105	transition
24	1.2	2	18	22	0.139	transition
25	1.2	2	18	26	0.151	transition
26	1.2	2	12	18	0.0815	transition
27	1.2	2	24	18	0.127	transition
28	1.2	2	30	18	0.142	transition
29	1.2	2.5	18	14	0.12	transition
30	1.2	2.5	18	18	0.136	emulsion
31	1.2	2.5	18	22	0.157	emulsion
32	1.2	2.5	18	26	0.172	emulsion
33	1.2	2.5	12	18	0.0996	emulsion
34	1.2	2.5	24	18	0.145	emulsion
35	1.2	2.5	30	18	0.16	emulsion
36	1.2	3	18	14	0.145	emulsion
37	1.2	3	18	18	0.163	emulsion
38	1.2	3	18	22	0.181	emulsion
39	1.2	3	18	26	0.19	emulsion
40	1.2	3	12	18	0.114	emulsion
41	1.2	3	24	18	0.172	emulsion
42	1.2	3	30	18	0.187	emulsion

**Table A-3: Experimental data for the kerosene-water system**

No.	A (cm)	f (1/s)	Q <sub>c</sub> (l/h)	Q <sub>d</sub> (l/h)	x <sub>d</sub>	Operating regime
1	1.2	0.5	18	12	0.0634	mixer-settler
2	1.2	0.5	18	18	0.0875	mixer-settler
3	1.2	0.5	18	24	0.0966	mixer-settler
4	1.2	0.5	18	30	0.105	mixer-settler
5	1.2	0.5	10	18	0.063	mixer-settler
6	1.2	0.5	26	18	0.0906	mixer-settler
7	1.2	0.5	34	18	0.108	mixer-settler
8	1.2	1	18	12	0.0422	mixer-settler
9	1.2	1	18	18	0.0513	mixer-settler
10	1.2	1	18	24	0.0724	mixer-settler
11	1.2	1	18	30	0.0905	mixer-settler
12	1.2	1	10	18	0.0423	mixer-settler
13	1.2	1	26	18	0.0543	mixer-settler
14	1.2	1	34	18	0.0694	mixer-settler

Continuation Table A-3

No.	A (cm)	f (1/s)	Q <sub>c</sub> (l/h)	Q <sub>d</sub> (l/h)	x <sub>d</sub>	Operating regime
15	1.2	1.5	18	12	0.0362	mixer-settler
16	1.2	1.5	18	18	0.0422	mixer-settler
17	1.2	1.5	18	24	0.0603	mixer-settler
18	1.2	1.5	18	30	0.0845	mixer-settler
19	1.2	1.5	10	18	0.0362	mixer-settler
20	1.2	1.5	26	18	0.0452	mixer-settler
21	1.2	1.5	34	18	0.0513	mixer-settler
22	1.2	2	18	12	0.0302	transition
23	1.2	2	18	18	0.0362	transition
24	1.2	2	18	24	0.0543	transition
25	1.2	2	18	30	0.0754	transition
26	1.2	2	10	18	0.0332	transition
27	1.2	2	26	18	0.0392	transition
28	1.2	2	34	18	0.0452	transition
29	1.2	2.5	18	12	0.0452	transition
30	1.2	2.5	18	18	0.0513	transition
31	1.2	2.5	18	24	0.0845	transition
32	1.2	2.5	18	30	0.108	transition
33	1.2	2.5	10	18	0.0452	transition
34	1.2	2.5	26	18	0.0543	transition
35	1.2	2.5	34	18	0.0603	transition
36	1.2	3	18	12	0.0603	emulsion
37	1.2	3	18	18	0.0754	emulsion
38	1.2	3	18	24	0.085	emulsion
39	1.2	3	18	30	0.10	emulsion
40	1.2	3	10	18	0.0497	emulsion
41	1.2	3	26	18	0.079	emulsion
42	1.2	3	34	18	0.0867	emulsion
43	1.2	3.5	18	12	0.0539	emulsion
44	1.2	3.5	18	18	0.0766	emulsion
45	1.2	3.5	18	24	0.0961	emulsion
46	1.2	3.5	18	30	0.113	emulsion
47	1.2	3.5	10	18	0.0561	emulsion
48	1.2	3.5	26	18	0.0892	emulsion
49	1.2	3.5	34	18	0.0979	emulsion