Hydrodynamics and Mass Transfer Characteristics of a Scheibel Extractor

Part I: Drop Size Distribution, Holdup, and Flooding

The hydrodrynamic behavior of a 1.0×10^{-1} m dia. Scheibel column packed with a material preferentially wetted by the dispersed phase has been investigated with and without mass transfer. Drop size and size distribution and dispersed phase holdup were measured as a function of flow rates, agitator speed, packing pad height, and mass transfer direction in a 2^4 factorial design. Flooding rates were measured and found to be dependent on the impeller position.

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SCOPE

Previous studies of drop size and size distribution, dispersed phase holdup, and flooding rates in a Scheibel extraction column have been limited to columns operated with packing material that was not wetted by the dispersed phase. For this type of operation the prediction of column performance is not totally reliable at the present time, and the mechanism describing the drop behavior inside the packing has not been established. Honekamp and Burkhart (1962) concluded, from data obtained in a 7.5×10^{-2} m diam. standard York-Scheibel column, that drop size and holdup at constant stirrer speed closely paralleled that of a packed column. Later, Jeffreys et al. (1971), using the same type of stainless steel knitmesh packing, found that the drop behavior was rather more complex, and only in those cases where drops were relatively smaller than the interstices of the packing could an analogous behavior to the packed column be found. This behavior was confirmed quantitatively by Wilkinson et al. (1975).

There is some evidence to indicate that a Scheibel column has a much greater capacity when operated with a packing material preferentially wetted by the dispersed phase. In addition, under some operating conditions, the overall mass transfer rate has been reported to be higher than those obtained in columns where the continuous phase wets the packing. Hence it appeared to be worthwhile to investigate the hydrodynamics and mass transfer performance of a Scheibel column operated under conditions of total coalesence in the packing zones.

The effect of drop size, size distribution, and dispersed phase holdup on the mass transfer performance of an extraction column are the most important hydrodynamic characteristics, since, from these, the interfacial area can be calculated. A knowledge of the interfacial area coupled with a suitable model will permit calculation of the overall mass transfer coefficient. This paper presents a study of drop size, size distribution, dispersed phase holdup, and flooding rates for a wide range of operating conditions.

CONCLUSIONS AND SIGNIFICANCE

Droplet size was determined from photographs of the dispersion taken through the column wall in different mixing compartments for the system toluene-acetone-water. Sauter mean drop sizes under nonmass transfer and different mass transfer direction were correlated using different regression models. Deviations from Kolmogoroff's law were large in the case of nonmass transfer and when the mass transfer direction was dispersed to continuous phase. This was found to be due to nonisotropic turbulent conditions existing in the dispersion. Hence care must be taken when applying Kolmogoroff's law in extraction equipment.

The holdup results were subjected to an analysis of variance when only agitator speed, times, mass transfer direction interaction. This indicates that the positive effect of agitator speed on holdup is greater when the mass transfer direction is continuous to dispersed phase with the system studied.

Attempts to correlate the flooding point results in a similar manner to that of Crawford and Wilke (1851) for the flooding of packed columns were unsuccessful. Finally, the capacity of the column was increased significantly when the dispersed phase wet the packing and the agitator was located close to the exit surface of the backing. In this study the optimum location was D/5 but this is believed to vary with column diameter.

INTRODUCTION

The Scheibel column consists of an alternate series of mixing sections and packed sections with a centrally located shaft upon which are mounted open turbine agitators. The packing or calming sections usually consist of a woven wire mesh that acts as an entrainment separator and coalescer for the two liquid phases. It has been identified as a low throughput high efficiency rotary column.

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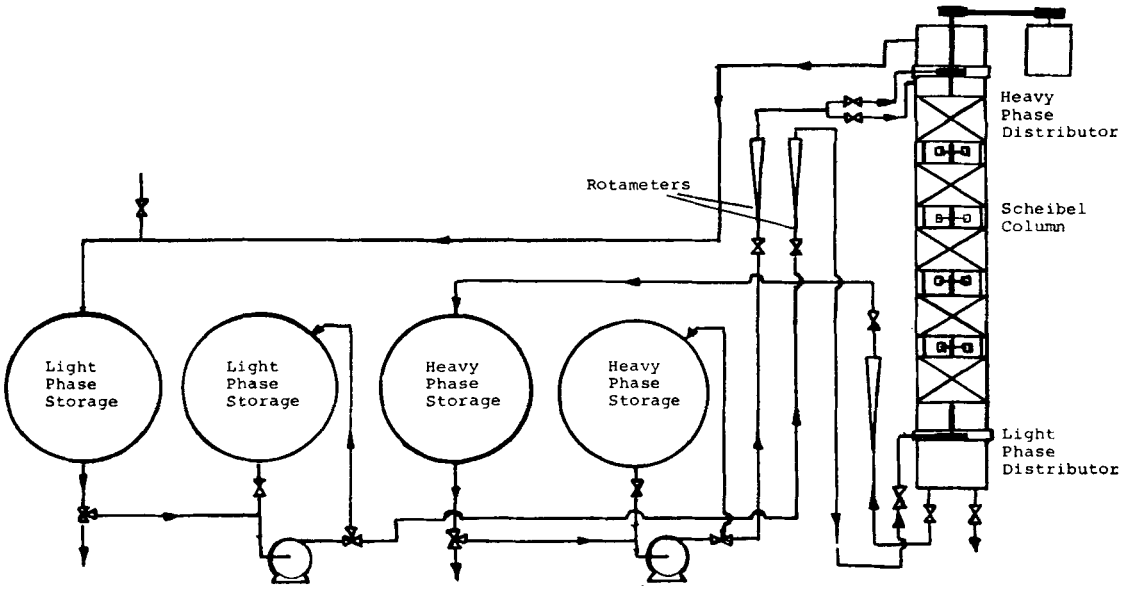

Figure 1. Flow diagram of experimental equipment.

The limiting capacity has been the major cause of its relatively low industrial application, even though the possible mass transfer efficiency is the highest of all agitated extraction columns.

Early research by Karr and Scheibel (1954) and Honekamp and Burkhart (1962) was concerned with the operation of the column with packing material that was not wetted by the dispersed phase, and a small number of correlations have been developed for this type of operation. However, Piper (1969) has shown that greater column capacity can be achieved when the column is operated with the packing material preferentially wetted by the dispersed phase. Also he stated that better mass transfer performance was achievd when a total coalescence packing pad was used. Unfortunately, insufficient experimental data were presented, and consequently a detailed analysis was not possible.

None of the above investigations has produced correlations for drop size and holdup that may give firm guidelines in the design of this equipment. At present, the drop size correlations developed from data obtained in baffled or unbaffled agitated tanks (Coulaloglou and Tavlarides, 1976) have been used to estimate drop sizes. Similar correlations (Strand et al., 1962; Misek, 1964; Arnold. 1974; Mumford and Al-Hemeri, 1974) have been obtained for the rotating disc contactor (RDC) and Oldshue-Rushton columns.

Dispersed phase holdup correlations have only been published for the RDC and Oldshue-Rushton extractors, and the analysis of holdup data has been considered in one of two ways. In the first, the holdup is considered to be a function of the characteristic velocity V_N, assuming that the theoretical approach (Gayler and Pratt, 1951) developed for packed columns is applicable to agitated columns (Laddha et al., 1978; Kung and Bechman, 1961). In the second, the holdup is correlated as a function of the physical properties of the system, column geometry, and the power input to the rotors by a standard dimensional analysis (Katsakin et al., 1962; Murakami et al., 1978).

Although no specific correlations have been reported for the Scheibel column, the hydrodynamic similarity with the packed column extractor suggests that the holdup may be correlated with other parameters by the Pratt characteristic velocity approach. This should be possible for Scheibel columns packed with material wetted by the continuous phase, operating under conditions that produce drops of size less than the interstices of the packing. In all other cases, including the Scheibel column packed with material preferentially wetted by the dispersed phase, no correlations have been reported.

EXPERIMENT

A flow diagram of the equipment is shown in Figure 1. Two columns with the same number of stages and consisting of a packed section and mixing section but with different packing thickness were used in this study. The first consisted of a 0.101 m dia. × 0.70 m long QVF glass pipe section divided into four Scheibel stages, each 0.11 m high containing a baffled mixing section and packing pads 0.06 m high and two nonagitated end sections. The second column was a 0.101 m dia. × 0.50 m long QVF glass pipe section with packing pads of 0.03 m in length. The nonagitated end sections of both columns were similar in order to ensure similar fluid dynamic and end effects. This was believed to be important in the experimental design because it was essential to avoid exaggerating the error variance and mislead the statistical analysis.

Eight sampling points were provided in column 1 and six in column 2. They were located at the middle of each mixing compartment and at the middle of the top and bottom distributor plates. These last two sampling points were installed to measure the phase concentration jump at the feed inlets. In addition, a sample point was provided for each end section in column 1. Each point comprised a 13 mm dia. drilled hole through which the sampling head probes could be set in position.

The column internals were fabricated entirely from 18/8 stainless steel. The rotor shaft was fabricated from 9.6 mm stainless steel rod and was supported by bearings at three points throughout the column length. There was no support bearing within the effective column length. Each four-bladed standard disc turbine of 33 mm dia. was fabricated from 2 mm thick stainless steel elements and secured to the shaft by grub screws. The packing pads were supported on a stainless mesh that was welded to four vertical baffles each 8 mm wide and 2 mm thick located 90° apart and machined to obtain a close fit at the column walls. The original Scheibel column contained an unbaffled mixing chamber, but in this work the baffles were necessary because at a low rpm a good dispersion was required for an accurate measurement of holdup by sampling.

A 1.5 cm dia. hole was drilled into the wall of both columns at the second chamber from the column bottom to accommodate the on-off brass valve for holdup measurements.

The dispersed light phase was introduced into the column through a stainless steel distributor located in the bottom of the column and discharged via a pipe located above the coalescence interface, as shown in Figure 1. A similar distributor plate was located at the top of the Scheibel column for the introduction of the heavy phase. Each distributor consisted of 46, 2.0 mm dia. holes arranged on 6.0 mm triangular pitch in a 50 mm diameter plate.

The column end plates were fabricated from 8.0 mm thick 18/8 stainless steel. The bottom plate supported a QVF pipe section located below the lower distributor plate that contained the necessary lines for draining the column and for withdrawal of the heavy phase. The top end plate housed a Teflon bearing and a stuffing box filled with Teflon granules that formed the seal.

The agitator shaft was driven by 0.25 HP DC motor controlled by a single phase 1/4 kW thyristor with an effective speed range of 0–1200 rpm.

In the Scheibel extractor studies reported in the literature, stainless steel knitted wire mesh packings were extensively used. Plastic wire mesh wettable by an organic dispersed phase has only been briefly studied by Piper (1969). In all cases the only variable used as criteria for the packing

TABLE 1. HOLDUP DATA: DISPLACEMENT VS. SAMPLING TECHNIQUE

Velocity, m/s			Holdup Fraction	
$V_D \times 10^3$	$V_c \times 10^3$	rpm	ϕ_D Displacement	ϕ Sampling
1.42	0.98	400	0.0415	0.0432
1.42	1.74	400	0.0454	0.0469
1.42	0.98	500	0.0643	0.0660
2.08	0.98	500	0.0843	0.0946
1.42	0.98	700	0.1438	0.1467
2.08	0.98	700	0.2000	0.2060

selection was the percentage voidage. A lower limit of 97% has been empirically stated, and no further information concerning the number of interlocking loops and wire diameter has been given. It has been shown (Bonnet, 1982) that pad voidage alone is not a sufficient criterion for the selection of the proper mesh and that the specification wire diameter and the number of interlocking loops in both the horizontal and vertical directions are required. Since the main objective of this investigation was to study a Scheibel column under conditions where the packing was wetted by the dispersed phase, a composite knitted mesh was selected. This packing consists essentially of two dissimilar materials, i.e., polypropylene wire of 0.25 mm dia. and stainless steel wire of 0.15 mm dia. knitted together. The dimension of the mesh chosen consisted of a structure of asymmetrical interlocking loops with 1.5–2.0 stitches/cm (Jeffreys and Davies, 1971). Thus, total wettability of the pad by the dispersed phase, with good column capacity performance, is obtained. A total of five pads of 0.03 m or 0.06 m thickness were inserted in each column with an average percentage voidage of 96%.

The system toluene-acetone (solute)-water was selected for this study. Toluene was the dispersed phase throughout the investigation. In the nonmass transfer runs the experiments were conducted with mutually saturated toluene and water phases and the turbine impellers were positioned at about (D_I/5) from the upper surface of the pads.

Holdup of the dispersed phase was measured by removing a two-phase sample from the second mixing compartment via the on-off valve. Due to the large bore of the valve, 100 ml of the dispersion could be taken out in about 2 s. In order to assess the accuracy of this technique, a Scheibel column was fitted with packing pads 15 mm thick and operated under similar conditions and the results of the sampling technique were compared with the overall holdup obtained by stopping the agitator and isolating the column (displacement technique). For this column the dispersed phase entrainment (static holdup) inside the wetted pads can be considered negligible in comparison with the overall holdup. The results of applying both techniques are presented in Table 1 and they clearly indicate that the sampling method gives reliable data. Replicated experiments using the sampling technique gave the percentage holdup values with a standard deivation of about 0.72 for all operating conditions.

The photographic technique was chosen to measure drop size and drop size distribution. An Asahi Pentax Spotmatic still camera with a Macro Takumar 100 mm lens, together with appropriate back lighting, were employed to photograph the dispersion. For each experiment two photographs were taken of each mixing chamber after steady state had been attained. Distortion of the droplet photographs was corrected by comparing the horizontal and vertical diameters of the drop with the shaft diameter and impeller blade width, and enlargement of the negatives to give an overall magnification of three to four times the true drop size gave sufficient magnification and contrast for counting the drops on a Zeiss TG3 Particle Counter.

Flooding rates under nonmass transfer conditions were determined at various agitator speeds for both Scheibel columns, and the flooding point was taken to be the condition at which the dispersed phase accumulated as a coalesced layer below a packing pad. If these conditions persisted, this layer would occupy the whole stage and eventually spread to the whole column inverting the phases. The flooding points were determined by setting the continuous phase flow rate and the rotor speed at predetermined values, then increasing the dispersed phase flow rate by small increments until flooding occurred. After each increase in flow, the column was allowed at least 15 min to come to steady conditions.

The experimental program followed the 2^4 factorial design shown in Table 2. This program was planned primarily to interpret the experimental concentration profiles for the models presented in Part II of this investigation. The experimentally determined drop sizes and holdups are tabulated in Part II.

RESULTS

Column Flow Behavior

The flow regimes obtained for different impeller positions and speeds are illustrated in Figure 2. The toluene drops were found to wet the mesh fiber and to coalesce fully inside the packing, leaving the pad surface in the form of jets as shown. The position of the impeller at low agitator speed was found to have an important effect on the hydrodynamics of the system, as is seen in Figure 2, (a-2) and (b-2). The resulting differences between the two modes of operation are summarized below.

1. Impellers positioned as shown in (a-2) produced, as observed in the majority of runs at low agitator speeds, larger drops when compared with the ones obtained under the same operating conditions but with the impellers close to the pads as in (b-2). This situation arises from the fact that some of the organic phase exit points at the packing surface are outside the area swept by the impeller. In this case the organic jets pass through zones of relatively low energy, compared with the high energy zone immediately below the impeller, with the result that breakage of the streams is incomplete and large globules form. It was noticed that this situation was more pronounced when the exit point was very close to the column wall, since a bypass flow was then produced.

2. The location of the impellers at a distance of about D_I/5 from the upper surface of the pad, as illustrated in Figure 2, (b-2), eliminates or substantially decreases the randomness in the discharge mechanism of the organic phase. The toroidal flow pattern generated by the impeller goes deep into the pad, inducing the coalesced organic phase to move toward the high *voidage center* zone in the neighborhood of the shaft. This phase then leaves the pad as a single stream surrounding the shaft below the disc of the impeller. This ensures that all the fluid elements of the coalesced organic phase emerging from the packing are subjected to the same

TABLE 2. TWENTY-FOUR EXPERIMENTAL DESIGN

Impeller Speed, N, rpm (A)	Pad Height, H_p, m (B)	Dispersed Phase Superficial Velocity, V_D (D)			
		1.28×10^{-3} m/s Mass Transfer Direction (C)		1.94×10^{-3} m/s Mass Transfer Direction (D)	
		D → C	C → D	D → C	C → D
400	0.03	(1) (Run 9)	C (Run 13)	6 (run 11)	CD (Run 15)
	0.06	B (Run 1)	BC (Run 5)	BD (Run 3)	BCD (Run 7)
600	0.03	A (Run 10)	AC (Run 14)	AD (Run 12)	ACD (Run 16)
	0.06	AB (Run 2)	ABC (Run 6)	ABD (Run 4)	ABCD (Run 8)

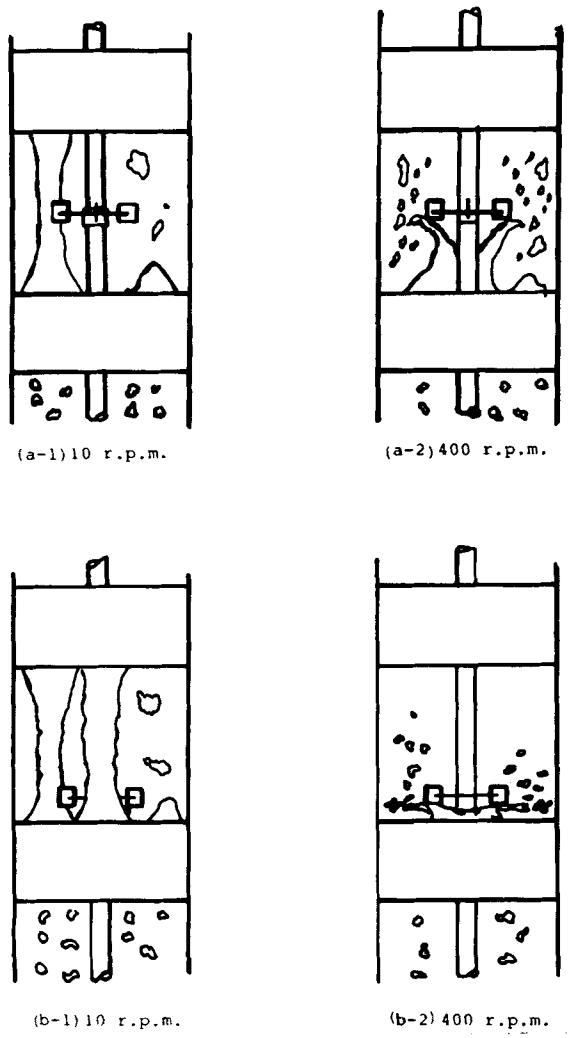

Figure 2. Effect of impeller position on flow regimes. No mass transfer.

high turbulence intensity, with much better control over the desired drop size and size distribution. Furthermore as discussed below, the position of the impellers close to the pads maximized the columns limiting capacity.

Flooding

The flooding characteristics of the first and second columns were investigated under nonmass transfer conditions initially. The influence of the agitator location on the flooding behavior was tested by running each of the columns with the impellers positioned at the middle of the mixing chambers and at $D_I/5$ from the upper surface of the pads. Typical flooding results are presented in Figures 3 and 4. These figures show that by positioning the impellers close to the upper surface of the pads, the column capacity increases with the impeller speed. A possible explanation for this behavior is that the coalesced organic phase inside the pads is collected and drawn out of the packing toward the shaft by the pumping action of the impeller. When the impellers are positioned for normal operation, as shown in Figure 2, (a-2), the results presented in Figure 3 illustrate that the relationship flooding rates vs. rpm follows the expected behavior; that is, increasing agitator speed decreases column capacity. In this case the pumping effect tends to be less intense because the organic phase is not drawn into the vicinity of the shaft and discharge is from many points with the result that the holdup of dispersed phase in the packing impedes the continuous phase flow and thereby reduces flooding rates.

Figure 4 presents the replicated flooding points for the second column for impellers positioned close to pads. These were obtained from the completed factorial experimental program. The results

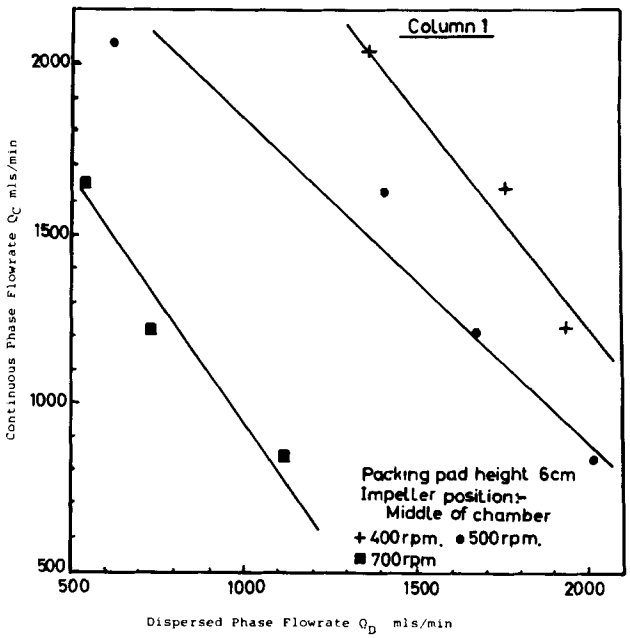

Figure 3. Relation between flow rates and flooding when impeller is in middle of mixing chamber.

confirm the unusual behavior reported above, but the poor reproducibility suggests that another variable has a strong influence upon the flooding mechanism. It is believed that this variable is random in nature and is closely related to the voidage distribution inside the pads.

Drop Size and Size Distribution

The Sauter mean drop diameter data from the nonmass transfer runs and the mass transfer experiments were correlated by regression (Bonnet, 1982). The results obtained from the nonmass transfer runs confirmed that drop size was independent of compartment number. This would be expected since similar coalescence and redispersion conditions exist in each stage. The preferred correlation was found to be

$$d_{32} = 5.719 \times 10^{-5} \cdot 10^{(1+1.397\phi_D)\bar{\epsilon}^{-0.708}} \qquad (1)$$

with a multiple correlation coefficient of 0.999 and a residual error of 0.11×10^{-4}. Holdup was found to be a significant variable.

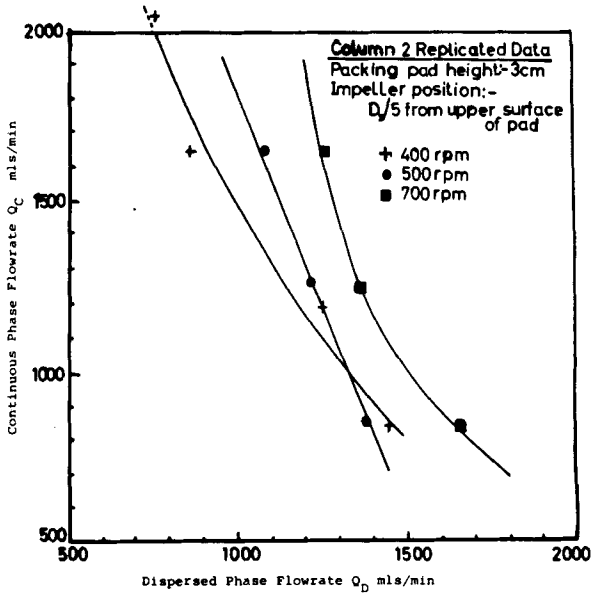

Figure 4. Replicated experimental results of flooding data for column 2.

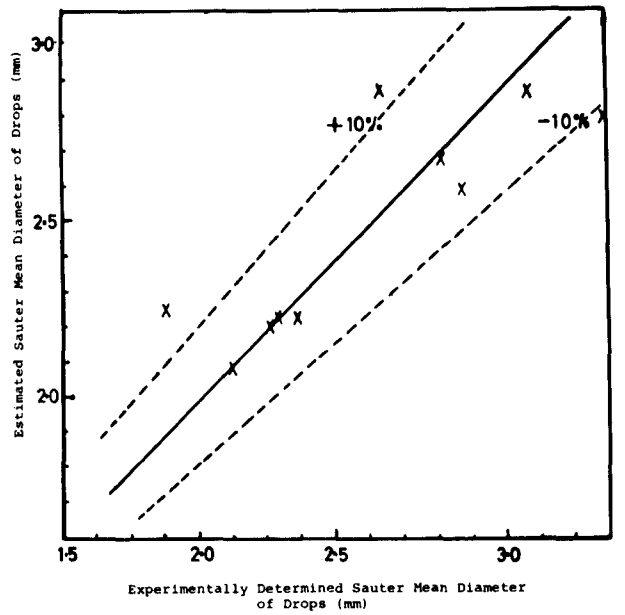

Figure 5. Comparison of calculated and experimental drop sizes (mass transfer direction vs. dispersed to continuous phase).

For the drop diameter data resulting from the experimental design of Table 2 and preliminary exploratory experiments, the best correlations found were for

1. Mass transfer direction; dispersed to continuous phase (Figure 5)

$$d_{32} = 1.240 \times 10^{-4} \cdot 10^{(1+3.417\theta_D)\bar{\epsilon}^{-0.277}}$$

residual error = 3.67×10^{-4} (2)

2. Mass transfer direction; continuous to dispersed phase (Figure 6)

$$\frac{d_{32}}{D_I} = (1.763 \times 16.117\ \phi_D)\,(We)^{-0.907}$$

residual error = 2.26×10^{-4} (3)

In these correlations the influence of the superficial velocity V_D on d_{32} is implicit, being included in the holdup factor. The influence of the agitator speed can be isolated to give

$d_{32} \alpha N^{-2.1}$	No mass transfer
$d_{32} \alpha N^{-0.8}$	Dispersed to continuous phase
$d_{32} \alpha N^{-1.8}$	Continuous to dispersed phase

Kolmogoroff's law is not strictly followed ($d_{32} \alpha N^{-1.2}$). Nevertheless, the exponents of the correlations for continuous to dispersed phase mass transfer direction were found to be in the range -1.2 to -1.8, close to -1.2 required by that law. The values of d_{32} were in the range 2.2×10^{-3} to 3.8×10^{-3} m for dipsersed to continuous mass transfer direction, compared with 1.3×10^{-3} to 2.40×10^{-3} m for nonmass transfer. For transfer in the continuous to dispersed phase direction, the drop size range was 0.8×10^{-3} to 2.2×10^{-3} m. The integral scale of turbulence Le, calculated according to the approximation given by Cutter (1966) for the impeller discharge stream of an agitated baffled vessel ($L_e = 0.08\ D_I$), is 2.7×10^{-3} m, and the Kolmogoroff length scale η was in the range 0.05×10^{-3} to 0.03×10^{-3} m. Therefore drop sizes for the conditions no mass transfer and dispersed to continuous phase mass transfer direction were most likely similar in size to the large and medium-sized eddies, for which isotropic turbulence does not exist and consequently Kolmogoroff's law cannot be applied.

Less deviation from Kolmogoroff's law is expected when the mass transfer is in the direction continuous to dispersed phase because the sizes of the drops were close to the upper limit of the domain of isotropic turbulence.

In terms of a characteristic particle size X, the log-normal distribution is given by

$$y = \frac{d\theta}{d\ln X} = \frac{1}{\ln\sigma_g\sqrt{2\pi}} \exp\left(-\frac{\ln X - \ln X_g)^2}{2(\ln\sigma_g)^2}\right) \qquad (4)$$

For $X = d$, Eq. 4 reduces to the simple log-normal distribution function, and for $X = d/(d_m - d)$, Eq. 4 corresponds to the upper limit laws of Mugele and Evans (1951), with d_m representing the maximum stable drop size. Combining the third and second moments of the distribution for Eq. 4, the following expressions for the X_{32} can be obtained in terms of the geometric mean by number, $X_{g,N}$:

$$X_{32} = X_{g,N} \exp\,[2.5\,(\ln\sigma_g)^2] \qquad (5)$$

or

$$X_{32} = X_{g,V} \exp\,[-0.5\,(\ln\sigma_g)^2] \qquad (6)$$

in terms of the geometric mean by volume, $X_{g,V}$ (Allen, 1981).

A transformation of Eq. 4 into a standard normal distribution enables the log-normal standard deviation expression to be

$$\ln\,\sigma_g = \ln\,(X_{90}/X_g)/1.2816 \qquad (7)$$

Figures 7 and 8 give typical cummulative volume percentage undersize data plotted on log-probability paper for two experiments for the simple and upper limit log-normal distributions. Each figure shows the experimental and the calculated Sauter mean drop diameters d_{32} for both distributions. A reasonably good fit to a straight line was obtained in all cases, although relatively large variation about the line is observed in Figure 8. This is most likely due to the small number of drops measured (100 to 150 drops per run).

The graphical method of testing has its shortcomings; it takes no account of the extent of the scatter of the points about the straight line, and the position and direction of the line itself can only be regarded as approximate. Therefore, the chi-square (χ^2) test was applied and the 5% level of probability was adopted. If the tabulated χ^2 (χ^2 tab) for the 5% level and for a certain number of degrees of freedom is exceeded by the χ^2 calculated (χ^2 calc) from the set of observed results, the differences between the observed and the theoretical distributions are regarded as signifying that the hypothesis of normality is not justified. The results of applying the χ^2 test to the data obtained in the above experiments are summarized in Table 3.

The χ^2_{calc} values for the single log-normal distribution are in the

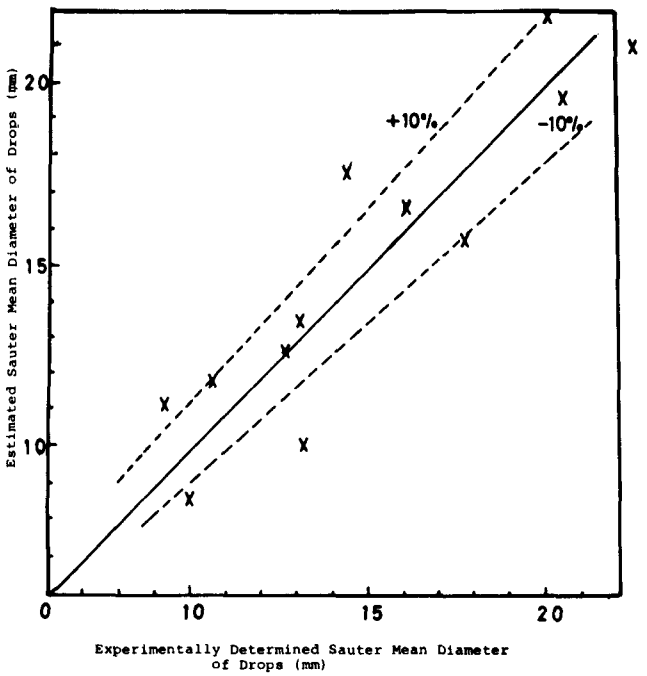

Figure 6. Comparison of calculated and experimental drop sizes (mass transfer direction vs. continuous to dispersed phase).

TABLE 3. χ^2 TEST FOR NORMALITY

Run	d_{32} Exp, mm	Distribution Log-Normal	d_{32} Calc., mm	Deg. of Freedom	χ^2 tab	χ^2 calc
1	1.89	Simple	1.86	6	12.59	13.01
		Upper Limit	1.90	5	11.07	15.41
6	0.82	Simple	0.828	4	9.5	11.01
		Upper Limit	0.833	4	9.5	15.92

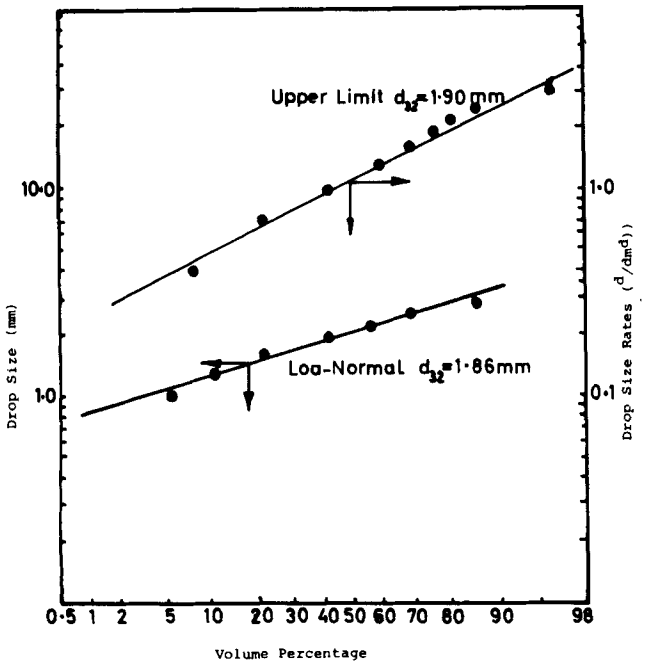

Figure 7. Drop size distribution; mean drop size (experimental); d_{32} = 1.89 mm.

borderline of significance with the largest deviation occurring in the latter experiment, due to the larger scatter of the data shown in Figure 8. For the upper limit distribution χ^2_{calc} largely exceeded χ^2_{tab} in all cases, and therefore this function cannot represent the observed distribution. Therefore, it is concluded that the simple log-normal probability density function is an acceptable representation of the measured drop size distributions.

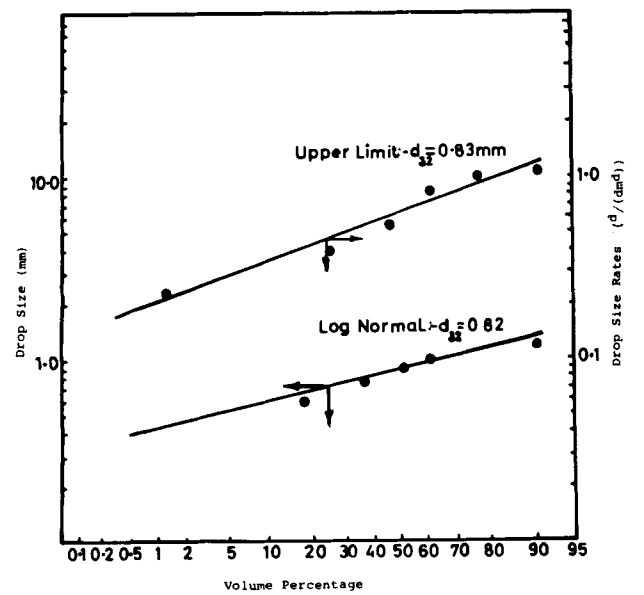

Figure 8. Drop size distribution; mean drop size (experimental); d_{32} = 0.82 mm.

TABLE 4. ANALYSIS OF VARIANCE OF HOLDUP DATA

Source of Variation	Degrees of Freedom	Mean Squares	F
A	1	47.680	95.58*
B	1	0.366	0.733
C	1	129.846	250.30*
D	1	33.698	67.54*
AB	1	0.308	0.617
AC	1	25.452	51.02*
AD	1	0.297	0.59
BC	1	1.311	2.62
BD	1	1.550	3.10
CD	1	1.311	2.62
ABC	1	0.632	
ABD	1	0.648	
ACD Error	1 = 5	0.011	$= \frac{2.494}{5} = 0.498$
BCD	1	0.648	
ABCD	1	0.555	

$F(1,5) = 6.61$ at 5% level.
$F(1,5) = 16.30$ at 1% level.
* Highly significant.

Dispersed Phase Holdup

Holdup data from the factorial experiment shown in Table 2 were subjected to an analysis of variance in which the experimental error variance was estimated from the three- and four-factor interactions. The validity of this step was tested by Bartlett's criterion (Davies, 1978). Table 4 shows that all factors except pad height are significant with positive effects on the response; that is, holdup increases when the level of the factor increases. The relatively large interactions for agitator speed times mass transfer direction indicate that the positive effect of the agitator speed on holdup is greater for the continuous to dispersed phase mass transfer direction. This strongly suggests that when holdup data are correlated, a factor must be included to represent the effect of the directions of transfer or correlations for each direction be obtained.

No attempt was made to obtain a correlation, mainly because insufficient data, especially at different values of the dispersed and continuous phase velocities, were available. The available data for both mass transfer directions were plotted with the coordinates

$$\left[V_D + \frac{\phi_D}{(1-\theta)} \cdot V_c\right] \text{ vs. } \left[\phi(1-\phi_D)\right]$$

at constant agitator speed. In all cases the plots failed to give satisfactory straight lines, which indicates that if the Pratt correlation is used as a basis, the characteristic volocity V_N must include a factor quantifying the effect of the flow rates.

NOTATION

d_{32} = Sauter mean drop diameter, m
D_I = impeller diameter, m:
H_P = packing pad thickness, m
H_M = Height of a mixing compartment in a Scheibel stage, m
L_e = integral scale of turbulence, m
N = impeller speed, rpm
n_d = number of data points
n_p = number of parameters to be estimated

Q_C = volumetric flow rate of continuous phase, ml/min
Q_D = volumetric flow rate of dispersed phase, ml/min
$(Re)_s$ = tank or impeller Reynolds number = $D_I^2 \cdot N \cdot P_c / \mu_c$
V_D = dispersed phase superficial velocity based on extractor cross section, m/s
V_C = continuous phase superficial velocity based on extractor cross section, m/s
V_N = characteristic velocity, m/s
v = volume fraction of droplets having diameter d
We = vessel Weber number = $D_I^3 \cdot N^2 \rho_c / G$
X = characteristic particle size, m
X_g = geometric mean particle size, m
$X_{g \cdot N}$ = geometric mean particle size by number, m
$X_{g \cdot V}$ = geometric mean particle size by volume, m
X_{90} = particle size at $v = 0.90$
X_{32} = volume surface mean particle size, m
X_{50} = 50% median particle size, m

Greek Letters

$\bar{\epsilon}$ = average energy dispersion per unit mass, $17.6 N^3 D_I^5 \rho_c / \pi D_c^2 H_M \rho$
η = Kolmogorof's length scale, m
θ = general term for frequency
ϕ_D = dispersed phase volumetric fractional holdup
μ_c = continuous phase dynamic viscosity, N/s·m^2
ρ_c^1 = density of continuous phase, kg/m^3
$\bar{\rho}$ = mean density of dispersion, mg/m^3
σ = interfacial tension, N/m
σ_g = geometric standard deviation
χ = chi-square probability test

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