Experimental Investigations in Liquid–Liquid Dispersion System: Effects of Dispersed Phase Viscosity and Impeller Speed

Mohd Izzudin Izzat Zainal Abidin, Abdul Aziz Abdul Raman,* and Mohamad Iskandr Mohamad Nor

Department of Chemical Engineering, Faculty of Engineering, University of Malaya, Kuala Lumpur 50603, Malaysia

ABSTRACT: Experimental investigation was conducted on a liquid–liquid dispersion in a stirred vessel in which the effects of dispersed phase viscosity were studied. Different grades of silicone oils were used to create oil-in-water dispersion by using Rushton turbine as an impeller, and drop sizes were measured by laser diffraction technique. Dispersion with higher uniformity of drop sizes was produced at low viscosity and high impeller speed. The dispersed phase viscosity influenced the equilibrium Sauter mean diameter, \(d_{32}\), by contributing to drops stabilization. The decrement of \(d_{32}\) with an increase in impeller speed is larger for high dispersed phase viscosities. It shows the influence of number and size of drop fragments formed after drop breakup on the mean drop size. Correlations relating \(d_{32}\) and dispersed phase viscosity were proposed with an accuracy of more than 90% between the predicted and experimental values.

1.0. INTRODUCTION

A liquid–liquid dispersion operation is required in industrial processes such as extraction, suspension polymerization, and multiphase reactions.\(^1\) The drop size distribution of a dispersion is related to the interfacial area which controls the amount of mass transferred between the phases.\(^2\) Therefore, accurate prediction of interfacial area in dispersions is important in determining the mass transfer and reaction rates.\(^3,4\)

Drop breakup and coalescence occur simultaneously during a liquid–liquid dispersion process, and the dynamic equilibrium between both processes determines the final drop size distribution of the dispersion. A drop in moving fluid experiences turbulent pressure fluctuations, and viscous stress which tend to cause drop breakup. The drop then resists the deformation by its surface force and internal viscous force if its viscosity is high.\(^5\) Drop coalescence occurs when two drops collide and combine into bigger drops, involving drainage and rupture of the intervening liquid film between the drops which is governed by the physical properties of the drops.\(^6\) The mean drop size and distribution are influenced by several factors, such as operating parameters of stirring process and properties of liquids which include viscosity and interfacial tension.\(^7\) An accurate prediction equation could be developed by understanding how mean drop size and distribution are by different factors to allow better control of drop size and uniformity of the distribution.\(^8\)

2.0. BACKGROUND

Available works in liquid–liquid dispersion have mainly focused on investigating effects of various influencing factors such as dispersed phase fraction, viscosity, and impeller design on the mean drop size and distribution.\(^3,9–11\) The main physical properties of liquids which influence mean drop size are interfacial tension and viscosity. Earlier works on liquid–liquid dispersion however have often been conducted for dilute and low dispersed phase viscosity\(^2,12,13\) in which the surface force dominates drops stabilization and the internal viscous force can be neglected. Usually, the outcome of these works led to the development of models or correlations which relate the Sauter mean diameter \(d_{32}\) to different influencing factors.

Most of the previous works involving prediction of drop size have applied the concept of turbulent energy cascade to estimate the maximum stable diameter \(d_{\text{max}}\) referring to the Hinze and Kolmogrov theory.\(^14\) On the basis of the theory, drops in the inertial region of turbulence have the maximum diameter \(d_{\text{max}}\) which is related to a dimensionless Weber number \(W\) and mean energy dissipation rate \(\varepsilon_T\) as shown in eq 1;\(^15\)

\[
\frac{d_{\text{max}}}{D} \propto \varepsilon_T^{-0.4} \propto W^{-0.6}
\]

(1)

\[
W = \frac{\rho N^2 D^3}{\sigma}
\]

(2)

The mean energy dissipation rate, \(\varepsilon_T\) at constant power number can be determined from power consumption \(P\) as shown in eq 3\(^16\)

\[
\varepsilon_T = \frac{P}{\rho V_T}
\]

(3)

The \(\varepsilon_T\) or mean flow which is characterized by the impeller tip speed \(ND^2\) is usually used to develop scale-up correlations for liquid–liquid systems, but the use of each of the parameters alone is not sufficient.\(^16\) Zhou and Kresta\(^16\) suggested that the interaction between energy dissipation rate and mean flow had to be considered, in which both \(\varepsilon_T\) and \(ND^2\) were included for a better correlation. They also claimed that the maximum energy dissipation rate \(\varepsilon_{\text{max}}\) should be used instead of \(\varepsilon_T\) as \(\varepsilon_{\text{max}}\) was more suitable to describe the breakup across different impellers. However, there are arguments whether one can easily and accurately measures \(\varepsilon_{\text{max}}\). Therefore, scale-up correlations were

Received: January 22, 2014
Revised: March 16, 2014
Accepted: March 19, 2014
Published: March 19, 2014
developed in this work by relating $d_{32}$ to $\varepsilon_s$ and the mean flow, ND$^2$.

Sauter mean diameter ($d_{32} = \sum d_i^3 / \sum d_i^2$) is preferred in characterizing dispersion as it relates interfacial area of dispersed phase to its volume.\(^7\) A relation between $d_{32}$ and $W_e$ is obtained based on the fact that $d_{32} \propto W_e^{0.6}$ as proven by several researchers:\(^7\,18\)

$$\frac{d_{32}}{D} \propto W_e^{-0.6} \quad \text{or} \quad \frac{d_{32}}{D} = C_4 W_e^{-0.6} \quad (4)$$

Equation 4 is a well-known relation and has been widely used to correlate $d_{32}$ for a dilute system with low dispersed phase viscosity. The constant $C_4$ in eq 4 needs to be evaluated experimentally where the published values in the literature are in the range of 0.05–0.081\(^9\) at low dispersed-phase viscosity. Although several researchers\(^20\,21\) have successfully applied eq 4 to correlate their data, some have discovered that the exponent of $W_e$ in eq 4 should be modified at different operating parameters where the applications of $d_{32} \propto W_e^{-0.6}$ have been argued at high dispersed phase fractions.\(^21\,22\) At high phase ratio, the value of the $W_e$ exponent has been found to be lower than $-0.6$.\(^23\) This is because the application of eq 4 is limited to a dilute and low viscosity system, based on the assumption that $d_{32}$ depends only on $d_{\text{max}}$ linearly. Without this assumption, Pacek et al.\(^21\) suggested that a general equation should be used where the exponent of $W_e$ should be determined experimentally;

$$\frac{d_{32}}{D} = C_4 (1 + C_5 \phi) W_e^{-\alpha} \quad (5)$$

Equation 5 is applicable when the effects of viscous force are negligible, especially for an inviscid liquid.\(^3\) However, the viscous force has significant effects on the cohesive stress at high viscosity, which stabilizes the drops and should be considered in the relation. The effects of viscosity can be included in the prediction model by introducing a dimensionless viscosity number $\nu_l$ which represents the ratio of viscous to surface forces:\(^5\,25\)

$$\nu_l = \left( \frac{\mu}{\rho_s} \right)^{0.5} \frac{\mu_c \text{ND}}{\sigma} \quad (6)$$

The viscosity number is added into the $d_{32}$ relation to account for drop stabilization by both surface and viscous forces as shown in the relation\(^5\)

$$\frac{d_{32}}{D} = C_4 W_e^{-0.6} \left[ 1 + C_5 \nu_l \left( \frac{d_{32}}{D} \right)^{1/3} \right]^{0.6} \quad (7)$$

When the viscosity term in eq 7 is too small and approaching zero, the equation reverts to eq 4. The relation is limited to dilute dispersions ($\phi < 0.01$)\(^26\) but could be modified to account for the effects of dispersed phase fraction.

The effects of dispersed phase viscosity are important as it is one of the main factors that influence the final drop size besides interfacial tension. However, most works on liquid–liquid dispersion have focused on low viscosity dispersed phase which has a narrow range of applications, and studies at high dispersed phase are limited. Viscous and nonviscous drops undergo breakup in different ways where the number of fragments formed after breakup is different.\(^27\) Therefore, $d_{32}$ is not only influenced by the rate of drop breakup but also the number and drop size of drop fragments produced after breakup.\(^28\) Relevant studies on effects of viscosity have been conducted in a surfactant-free system by Calabrese et al.\(^29\) and Podgorska\(^1\) and in a surfactant-stabilized system which focused only on drop breakup.\(^30\,31\) The objectives of this research work are to investigate the effects of viscosity on drop breakup and mean drop size apart from obtaining the correlations relating $d_{32}$ to the influencing parameters. The studies were conducted at different power inputs to gain better understanding on the effects of viscosity on drop breakup.

Various grades of silicone oils with different viscosities (moderate to high) were used to study the effects of viscosity on mean drop size and distribution. Silicone oil was used as it exhibits nearly constant interfacial tension at different viscosities with its density not too much different from the density of the continuous phase used in this work. Since the works focused on drop breakup only, surfactant was added to the system to prevent coalescence.\(^32\) Different techniques are available for drop size measurement which can be categorized based on their measurement principles.\(^33\) Laser diffraction is one of the laser system techniques that yields fast and accurate measurements. It is based on the measurement and interpretation of angular distribution of light diffracted by the drops, referring to Fraunhofer diffraction theory. This technique was used in this work because of its versatility, broad dynamic range, and high reproducibility.\(^34\)

### 3.0. EXPERIMENTAL PROCEDURES

Experiments were conducted in a clear cylindrical flat bottom stirred tank with an internal diameter of $T = 0.20$ m, equipped with four identical baffles with width equal to $T/10$ each. The tank was filled with distilled water as continuous phase to a height of $H = T$. A standard six-blade stainless steel Rushton turbine with diameter $D = 0.08$ m was used with bottom clearance equal to $T/3$. Lighting was provided by four fluorescent lights at the corners of the vessel to allow observation of the dispersion process. Various grades of silicone oils (20, 350, and 500 mPas) supplied by Sigma Aldrich were used as dispersed phase. Silicone oils have a high refractive index which makes them suitable for measurement by laser diffraction and image analysis techniques. The interfacial tensions of the oils were measured using the du Nouy ring method. The properties of the oils are tabulated in Table 1.

<table>
<thead>
<tr>
<th>Table 1. Properties of the Silicone Oils</th>
</tr>
</thead>
<tbody>
<tr>
<td>properties (at 25 °C)</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>silicone oil 20</td>
</tr>
<tr>
<td>silicone oil 350</td>
</tr>
<tr>
<td>silicone oil 500</td>
</tr>
<tr>
<td>viscosity, $\mu_c$ (mPas)</td>
</tr>
<tr>
<td>20</td>
</tr>
<tr>
<td>350</td>
</tr>
<tr>
<td>500</td>
</tr>
<tr>
<td>interfacial tension, $\sigma$ (kg/m²)</td>
</tr>
<tr>
<td>0.017</td>
</tr>
<tr>
<td>0.0182</td>
</tr>
<tr>
<td>0.0189</td>
</tr>
<tr>
<td>density, $\rho$ (kg/m³)</td>
</tr>
<tr>
<td>1001</td>
</tr>
<tr>
<td>970</td>
</tr>
<tr>
<td>refractive index</td>
</tr>
<tr>
<td>1.39</td>
</tr>
<tr>
<td>1.403</td>
</tr>
<tr>
<td>1.403</td>
</tr>
</tbody>
</table>

Approximately 0.3% w/w of sodium dodecyl sulfate (SDS) was diluted by distilled water before being added to the oil. The temperature in the vessel was monitored during the mixing process to make sure that there was no significant temperature increase. The required amount of silicone oil (equivalent to $\phi = 0.01$) was added into the distilled water near the impeller region by using a syringe. The impeller speeds were varied from 300 rpm to 500 rpm, which are equivalent to Reynold’s numbers in the range of 33 000 to 55 000. The Reynold’s number is calculated as $Re = \rho \text{ND}^2 / \mu_c$. The minimum impeller...
speed was chosen to be above the minimum speed required for a complete dispersion, and the maximum speed was chosen below the speed where air entrainment starts to occur in the vessel. The power number determined for the Rushton turbine was 5.8 in the turbulence region. The details of the experiments are shown in Table 2.

Table 2. Details of the Experiment

<table>
<thead>
<tr>
<th>experimental conditions</th>
<th>symbol</th>
<th>value</th>
</tr>
</thead>
<tbody>
<tr>
<td>mixing time, min</td>
<td>θ</td>
<td>60</td>
</tr>
<tr>
<td>impeller diameter, m</td>
<td>D</td>
<td>0.08</td>
</tr>
<tr>
<td>tank diameter, m</td>
<td>T</td>
<td>0.2</td>
</tr>
<tr>
<td>tank height, m</td>
<td>H</td>
<td>0.4</td>
</tr>
<tr>
<td>impeller clearance, m</td>
<td>C</td>
<td>0.067</td>
</tr>
<tr>
<td>baffle width, m</td>
<td>B</td>
<td>0.02</td>
</tr>
<tr>
<td>liquid height, m</td>
<td>H_l</td>
<td>0.2</td>
</tr>
<tr>
<td>dispersed phase fraction</td>
<td>ϕ</td>
<td>0.01</td>
</tr>
<tr>
<td>SDS concentration, wt %</td>
<td></td>
<td>0.3%</td>
</tr>
</tbody>
</table>

3.1. Drop Size Measurements by Laser Diffraction. Samples of the dispersion were collected at a fixed position near the impeller region where the rate of drop breakup is high. The samples were taken to a Malvern Mastersizer 2000 for drop size measurement process. The laser diffraction equipment is well-known and widely used in this area by several researchers. 20,24,30,35,36 It is capable of measuring drops in the range of 0.02–2000 μm with an accuracy of ±1%. The equipment requires the information on the refractive index of the dispersed phase (see Table 1). Drop size distribution curves and Sauter mean diameter $d_{32}$ for all dispersions were produced and determined from the measurement.

![Figure 1. Schematic Diagram of the Stirred Vessel.](image-url)

4.0. RESULTS AND DISCUSSION

4.1. Drop Size Distributions at Different Impeller Speeds. The drop size distribution curves at different impeller speed for each dispersed phase viscosities were plotted as shown in Figure 2a,b,c. The curves show the range of drop sizes in the dispersion. In Figure 2 the curves shifted to the right where the size of the drops was larger as the dispersed phase viscosity increased. In most cases, the distribution curves for 20 mPas were much taller (see Figure 2) compared to 350 and 500 mPas which show that the drop sizes for low viscosity are more uniform and the drop size distribution also broadens as $μ_d$ increases. Besides, at constant impeller speed, the size of the smallest drops decreased while the volume percentage increased as $μ_d$ decreased. For example, at 400 rpm, the $d_{50}$ (50% of drops were below this diameter) for 20 mPas, 350 mPas, and 500 mPas was 82.93, 126.32, and 273.31 μm, respectively. Therefore, it shows that low viscosity produces dispersion with smaller drops compared to high viscosity, which leads to higher surface area. Similar trends were reported by Podgorska1 and El-Hamouz et al.35 Since the interfacial tension differences between the three oils were minimal, the influence on drop-size distribution by reducing the rate of drop breakup comes from $μ_d$.

As the impeller speed increased, the curves shifted to the left which was toward the lower drop sizes range. This is because smaller and finer drops were produced and the size frequency increased at high impeller speed. It occurs as a result of increased in drop breakup rate when the power input to the system is increased. For example, at $μ_d = 20$ mPas, $d_{50}$ decreased as the impeller speed increased where $d_{50}$ at the impeller speed of 300, 400, and 500 rpm was 290.65 μm, 126.32 and 117.18 μm respectively. The observations were expected as increase in the power input to the impeller leads to increase in shear and energy dissipation rate. In this condition, turbulent pressure fluctuation which is the main external force that causes drop breakup increases.

4.2. Effect of Dispersed Phase Viscosity on Mean Drop Size. The $d_{32}$ for every $μ_d$ are tabulated in Table 3 at different impeller speeds. As seen in Table 3, larger $d_{32}$ was produced as $μ_d$ increased from 20 to 500 mPas at the same impeller speed, leading to smaller surface area in the dispersion. The observation can be explained by how the dispersed phase viscosity affects the rate of drop breakup where drops are stabilized not only by the surface force but also the internal viscous force at high viscosity. The viscous force of the drops provide additional cohesive force, and thus more energy is required to overcome the cohesive force, reducing drop breakup.37 Drops with higher viscosity also have lower drop breakup probability, where the breakup events area in the stirred vessel is smaller, compared to drops with low viscosity due to the stability of the oil.4

4.3. Sauter Mean Diameter as a Function of We. A dimensionless size ratio ($d_{32}/D$) was plotted against the dimensionless Weber number, $We$ for every dispersed-phase viscosities as shown in Figure 3. The data fitted well into eq 4 with slightly smaller $We$ exponents. The constants $C_1$ and $We$ exponents as in eq 4 were obtained for the three $μ_d$ and are shown in Table 4.

The exponents of $We$ obtained from the data were between −0.65 to −0.68 for low to high $μ_d$ which were close to the theoretical value of −0.6 as in eq 4. As seen in Table 4, the value of $C_1$ for $μ_d = 20$ mPas was in the range 0.05–0.081 that was reported in the literature for low viscosity as discussed earlier. It shows that the system satisfies the theoretical conditions which are based on a breakup dominant process. The constant $C_1$ increased to 0.174 and 0.364 as $μ_d$ increased, values which were out of the range as the reported values were only for a low viscosity system. The increment of $C_1$ was due to the increase in $d_{32}$ at higher $μ_d$. It can be concluded from the result that eq 4 could be applied in a high viscosity system with slight modifications of the values of the $We$ exponent.
4.4 Effect of Impeller Speed on \( d_{32} \) at Increasing Dispersed Phase Viscosity. The \( d_{32} \) data for each \( \mu_d \) was plotted against impeller rotational speed \( N \) in Figure 4. It is observed that \( d_{32} \) decreased as \( N \) increased for every \( \mu_d \). An increase in \( N \) gives higher energy dissipation into the system where the energy is used to overcome the cohesive forces which act to keep the drops to its original shape. Thus, the rates of drop breakup increase as \( N \) increases and smaller \( d_{32} \) is produced. The behavior can also be explained based on circulation time, \( \theta_c \) where it is related to \( N \) by \( \theta_c \propto 1/ND^2 \). Therefore, the circulation time in the vessel becomes shorter as \( N \) increases, resulting in more circulations or flows in the vessel. Therefore, the drops take less time to travel to the impeller region, which is the most important zone that controls drop breakup and where the rate of drop breakup is high.

In Figure 4, a straight line was used to fit each of the data sets in order to show the slope which represents the change in \( d_{32} \) as a function of impeller speed. Interestingly, different slopes were obtained for different dispersed phase viscosities, which were \(-0.149\), \(-0.455\), and \(-0.881\), where the largest slope was achieved by \( 500 \) mPas followed by \( 350 \) mPas. It means that high viscosity drops undergo larger change in \( d_{32} \) as the impeller speed increases, although \( d_{32} \) for high viscosity is...
bigger compared to that for low viscosity. On the basis of the findings, the changes in drop breakup rates are varied at different viscosities.

Different values of the slopes can be explained by the influence of drop diameter on drop breakup rate. The breakup of two drop diameters was investigated by Hermann et al. and Maß et al. by using computational fluid dynamics (CFD). It was observed that a smaller drop is more stable than a larger drop. The region of drop breakup is larger for a bigger drop diameter leading to an increase in drop breakup probabilities. Since low viscosity produced smaller \( d_{32} \) in this study, the drops were more stable and the change in \( d_{32} \) was smaller as the impeller speed increased, as shown in Figure 3. Besides, Andersson\(^\text{27} \) pointed out that a viscous drop in turbulent flow undergoes deformation by forming a thin liquid thread before it breaks into many smaller fragments, while a nonviscous drop breaks into three fragments only. Thus, a high-viscosity drop produces more daughter droplets, and therefore larger \( d_{32} \) changes were observed as \( N \) increased. Similar trends of the slopes were observed by Sechremeli et al.\(^\text{41} \) but at different dispersed phase fractions. The slope of \( d_{32} \) versus \( N \) for \( \phi = 0.1 \) which produced bigger \( d_{32} \) was larger than the slope for \( \phi = 0.1 \) with smaller \( d_{32} \).

### 4.5. Sauter Mean Diameter as a Function of Energy Dissipation Rate and Mean Flow

Both mean energy dissipation rate \( \varepsilon_T \) and mean flow \( ND^2 \) were used to develop correlations for scale-up, and the graph of \( d_{32} \) vs \( \varepsilon_T (ND^2) \) is shown in Figure 5. Zhou and Kresta\(^\text{16} \) reported that \( \varepsilon_T \) could also give reasonably good correlation when it was used with mean flow, \( ND \), or \( ND^2 \) just like in the case of \( \varepsilon_{\text{max}} \). The correlations for each \( \mu_d \) were obtained with a \( R^2 \) value of 0.87 for 20 mPa·s and above 0.9 for 350 and 500 mPa·s as shown below:

\[
d_{32} = 34.31(\varepsilon_T ND^2)^{-0.2} \quad \text{for 20 mPa·s} \\
d_{32} = 27.48(\varepsilon_T ND^2)^{-0.4} \quad \text{for 350 mPa·s} \\
d_{32} = 28.51(\varepsilon_T ND^2)^{-0.52} \quad \text{for 500 mPa·s}
\]

It is seen from eq 9 to eq 11 that the power of \( \varepsilon_T (ND^2) \) decreases from 20 mPa·s to 500 mPa·s resulting in a bigger difference between \( d_{32} \) value when the impeller speed is increased. It shows how much \( d_{32} \) at different \( \mu_d \) is affected by the power input to the system. Therefore the change in \( d_{32} \) is bigger as \( N \) is increased at high viscosity, as explained in the previous section. A model was developed to represent the three scale-up correlations in a single equation by adding viscosity number \( V_i \):

\[
d_{32} \left( \frac{D}{L} \right) = 0.000014[1 + 1.428(\varepsilon_T ND^2)^{-0.7}]V_i^{0.27}
\]

The predicted \( d_{32} \) data obtained from eq 12 was plotted against the experimental data as shown in Figure 6. It can be observed that reasonably good fits were achieved for \( \mu_d = 20 \) and 350 mPa·s and there was little scattering for 500 mPa·s.

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**Table 4. Constants and \( R^2 \) Values of Data Fitting into the Generalized Form of \( d_{32}/D = C_1 We^n \)**

<table>
<thead>
<tr>
<th>viscosity, ( \mu_d ) (mPas)</th>
<th>( C_1 )</th>
<th>( N )</th>
<th>( R^2 )</th>
</tr>
</thead>
<tbody>
<tr>
<td>20</td>
<td>0.06</td>
<td>-0.65</td>
<td>0.900</td>
</tr>
<tr>
<td>350</td>
<td>0.174</td>
<td>-0.65</td>
<td>0.951</td>
</tr>
<tr>
<td>500</td>
<td>0.364</td>
<td>-0.68</td>
<td>0.950</td>
</tr>
</tbody>
</table>

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**Figure 3.** \( d_{32}/D \) vs \( We \).

**Figure 4.** Graph \( d_{32} \) vs impeller speed.

**Figure 5.** Graph \( d_{32} \) vs \( \varepsilon_T (ND^2) \).
4.6. Mean Drop Size Correlations. The correlation of $d_{32}$ was developed as a function of We and viscosity ratio, $\mu_d/\mu_c$ to represent the effects of dispersed phase viscosity. The exponent of We and other constants were determined by linear regression. The data are best correlated by the following equation

$$\frac{d_{32}}{D} = 0.285 \left[ 1 + 0.0035 \left( \frac{\mu_d}{\mu_c} \right)^{0.9} \right] We^{-0.83}$$

(13)

In eq 13, the value of the We exponent is $-0.83$, which is slightly smaller but quite close to the value in the theoretical relation ($-0.6$). An almost similar value of We exponent ($-0.85$) was obtained by Singh et al. in their model which related $d_{32}$ to dispersed phase fraction, $\phi$. Figure 7 shows that eq 13 gives good prediction of $d_{32}$ at low $\mu_d$ where almost all the points are in the range of less than 10% deviation from the experimental $d_{32}$ values. The overall percentage of similarity between experimental and predicted $d_{32}$ by eq 13 is 92.25%.

![Figure 6. $d_{32}/D$ experimental vs $d_{32}/D$ predicted.](image)

Although the $d_{32}$ model (eq 13) is able to give reasonably good estimation of $d_{32}$ at different impeller speed and dispersed phase viscosity, the deviations between the predicted and experimental values at high impeller speed are quite significant (see Figure 7). Thus, another correlation relating $d_{32}$ to $\mu_d$ and impeller speed has been developed. In this model, viscosity number $Vi$ is used instead of viscosity ratio to represent the effects of dispersed phase viscosity. Since there are small differences in the interfacial tension values at different $\mu_d$, the use of viscosity number may be more suitable as it represents the ratio of viscous to surface forces. The proposed model is

$$\frac{d_{32}}{D} = 1.588(1 + 0.0227 Vi^{1.4})We^{-1.05}$$

(14)

The comparison between the predicted and experimental $d_{32}$ is shown in Figure 8. As seen in Figure 8, the predicted $d_{32}$ values by eq 14 are in good agreement with the experimental $d_{32}$ values. The points for $\mu_d = 500$ mPas are nearer to the 10% line, which means that the deviations between the predicted experimental $d_{32}$ at high $\mu_d$ are reduced. Therefore, the model (eq 14) gives better $d_{32}$ estimation compared to eq 13 at high $\mu_d$. It also shows that the use of viscosity number $Vi$ as a parameter to represent the effects of dispersed phase viscosity on mean drop size is more suitable at high $\mu_d$ compared to viscosity ratio only.

In eq 14, the value of the We exponent ($-1.05$) is smaller than the value in the theoretical relation ($-0.6$). The value obtained in the model deviates from the theoretical model because the model developed in this work covers high viscosity dispersed phase while the theoretical relation only applies at low viscosity. Although eq 13 gives better accuracy at low $\mu_d$, a higher percentage of similarity (95.5%) between the predicted and experimental $d_{32}$ was achieved by eq 14 for the ranges of $\mu_d$ investigated in this work. Both correlations developed in this work show that the predicted $d_{32}$ correlated well with the experimental $d_{32}$ with slight scatters for 500 mPas points. At high $\mu_d$ it can be observed that both correlations under-predicted $d_{32}$ values where the deviations between the predicted and experimental $d_{32}$ values were quite significant. Similar observations were reported at this $\mu_d$ value by Wang et al. and
Calabrese et al., where their models showed scattering of \(d_{32}\) values at \(\mu_2 \geq 500\) mPas. The authors explained that this behavior might be caused by different nature of the silicone oil where the oil tends to change its nature from Newtonian to viscoelastic at high viscosity. Therefore, the use of a single model to represent the \(d_{32}\) data from low to high \(\mu_2\) might not be suitable because of different mechanisms in the model.

### 5.0. CONCLUSIONS

The drop size distribution for dispersion of moderate to high dispersed phase viscosity was presented at different impeller speeds. Smaller drop sizes with higher uniformity were produced at low dispersed phase viscosity and high impeller speed. Since the silicone oils used in the study exhibited almost constant interfacial tension, the change in \(d_{32}\) at different dispersed phase viscosity showed the contribution of viscous force on the stabilization of drops against breakup.

The data showed that the theoretical value of the \(W_e\) exponent \((-0.6)\) was slightly modified at a higher dispersed phase fraction as the theoretical value is only valid for dilute and inviscid systems.

Although decrease in \(d_{32}\) as the impeller speed increased was observed for all \(\mu_2\) the percentage of \(d_{32}\) decrement was different. It shows that the drop breakup mechanism is varied at different viscosities, proving that the mean drop size is also influenced by the size and number of daughter droplets formed after a drop breakup.

Scale-up models based on \(\varepsilon\) and mean flows were developed at different \(\mu_2\) which fit reasonably well into individual models. Therefore, it is agreed that both parameters should be used together in a \(d_{32}\) scale-up model. Two \(d_{32}\) correlations were proposed on the basis of the framework of the classical relation with the addition of \(\mu_2/\mu_1\) and \(V_i\) as parameters to represent the effects of dispersed phase viscosity. The predicted \(d_{32}\) values by both correlations correlated well with the experimental \(d_{32}\) values with slight scatters at high \(\mu_2\). The accuracy of both correlations were higher than 90%.

### NOMENCLATURE

- \(d_{32}\) = Sauter mean diameter (\(\mu m\))
- \(d_{\text{max}}\) = maximum stable diameter (\(\mu m\))
- \(D\) = impeller diameter (m)
- \(N\) = impeller rotational speed, (1/s)
- \(P\) = power consumption (kgm\(^2\)/s\(^3\))
- \(V_i\) = volume of liquid in the tank (m\(^3\))
- \(VI\) = viscosity number, \((\rho_1/\rho_2)^{3/2}(\mu_2N D/\sigma)\) (dimensionless)
- \(We\) = Weber number, \(\rho_2N^2D^3/\sigma\) (dimensionless)
- \(\sigma\) = interfacial tension (kg/s\(^2\))
- \(\rho_1\) = continuous phase density (kg/m\(^3\))
- \(\rho_d\) = dispersed phase density (kg/m\(^3\))
- \(\mu_1\) = continuous phase viscosity (mPas)
- \(\mu_d\) = dispersed phase viscosity (mPas)
- \(\phi\) = dispersed phase fraction (dimensionless)
- \(\varepsilon\) = mean energy dissipation rate per unit mass (m\(^2\)/s\(^3\))

### REFERENCES


