Breakup and coalescence regularity of non-dilute oil drops in a vane-type swirling flow field

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\section*{ABSTRACT}

In this work, in order to investigate the behavior of non-dilute drops, including breakup, coalescence and trajectory, in a 100-mm inner-diameter horizontal swirling flow field with low inlet mixture velocity, both an experimental study and numerical simulation were conducted. Inlet oil phase concentration was under 3.0\% volume fraction, with an inlet flow rate ranging from 12 m\(^3\)/h to 18 m\(^3\)/h. Malvern RTsize and Electrical Resistance Tomography were applied for measuring the drop size distribution and oil phase concentration, respectively. Correspondingly, numerical simulations applying a Renormalization-group \(k-\varepsilon\) turbulent model, coupled with a Discrete Phase Model simulating oil phase, were conducted as well. The results showed that small drops in the flow field tended to coalesce, while the behavior of large drops was determined by the inlet flow rate. A higher inlet flow rate led to a thinner oil core with constant inlet oil concentration. Moreover, the simulation results, which corresponded well with the experimental observations, presented oil drops distribution laws of breakup, coalescence and trajectory in a 100-mm inner diameter swirling flow field and established a prediction model in a similar flow field. Finally, regularity of swirling intense distribution and drop-turbulence interaction in a swirling flow field with a low inlet velocity was established. These results provide new information helpful for the design of vane-type separators.

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\section*{1. Introduction}

1.1. Background

As oil production continues, an increasing amount of water is present in the reservoir fluids. To extract crude oil from reservoir fluids with water holdup of over 95\%, the separation mechanism should be thoroughly studied, and a separation system should be meticulously refined. This is also true for cyclone separation, the most broadly used process for secondary treatment (Motin, 2015). Cyclone separation is proposed to process dispersed flow through a high swirling motion generated by a centrifuge, which makes separation possible due to the density difference. With the assistance of a static mixer and other pre-processing, reservoir fluids turn into a turbulent dispersed flow and entry cyclone separators. Considering the water holdup and various operating conditions, it is necessary for the cyclone separator to process dispersions with a non-dilute oil phase in low inlet flow conditions.

In previous literature, two types of cyclone separators have been applied industrially, namely, tangent inlet cyclonic separators and axial inlet cyclone separators, also called vane-type separators. The former has been studied and modified for decades to cater to various crude reservoir and strict environmental standards, while the latter has been only recently proposed and is initially designed for gas-solid separation in the field of air cleaning (Cai et al., 2014) or gas-liquid separation in the atomic field (Klujzso et al., 1999). Compared with a hydrocyclone, a vane-type separator has the advantages of simplicity in structure, high separation efficiency and low pressure drop in operation (Cai et al., 2014). When encountering the strict space demands for down-hole separation, the vane-type separator is a better choice. Compared with a hydrocyclone separator (Chakraborti and Miller, 1992; Khezzar, 2010;
Gao et al., 2013; Motin, 2015; Vakamalla and Mangadodd, 2016), some researchers (Klujszo et al., 1999; Rahimzadeh and Motin, 2010; Cai et al., 2014; Shi and Xu, 2015) have concentrated efforts on vane-type separators, mainly via vane-type designs and continuous phases with high inlet velocity to address issues in the breakup and coalescence behavior of drops, let alone drop breakup and coalescence in non-dilute dispersions under low inlet flow rates. As a consequence, it is necessary to provide insight into the exploration of the breakup and coalescence regularity of non-dilute drops in the vane-type swirling flow field with low inlet velocity, especially for guide vanes in 100-mm inner diameter pipes, whose size is nearly equal to those in real production.

1.2. Theoretical model

Dispersed multiphase flows are common in many engineering and environmental applications, and they are often turbulent. Examples of dispersed multiphase flows include particles suspended in gas or liquid, the dispersion of drops in a stream of gas, and bubbly flow. To investigate the breakup and coalescence regularity of dispersed oil-in-water flow, theoretical models of drop size and size distribution and of drop breakup and coalescence procedure microscopically are gradually developed. It is generally recognized that the drop size and their breakup are determined by the balance between the forces trying to break them up (high shear stress, turbulence in the continuous phase and rapid acceleration) and those that tend to stabilize them (namely, interfacial tension and dispersed phase viscosity) (Angeli, 2001; Brauner, 2003; Rahimzadeh and Motin, 2003). In regard to turbulent flow, Kolmogorov, based on the concept that kinetic energy in large structures does not dissipate into heat under the effect of molecular viscosity until transferred to smaller dissipative scales, proposed the Kolmogorov scale $l_k$.

$$l_k = \frac{v^{3/4}}{\varepsilon^{1/4}}$$  (1)

where $v_t$ is the kinematic viscosity of the continuous phase, and $\varepsilon$ is the energy dissipation per unit mass. Hinze (1955) proposed $W_{\text{Crit}}$ and suggested that in homogeneous and isotropic turbulent flow, drop break up occurs when the Weber number reaches $W_{\text{Crit}}$, namely:

$$W_{\text{Crit}} = W_{\varepsilon}$$  (2)

where $W_{\text{Crit}} = \tau d_{\text{max}}/\sigma$, and $W_{\varepsilon} = \rho_{\varepsilon} u'^2 D/\sigma$, where $\rho_{\varepsilon}$ is the density of the continuous phase.

The mean fluctuation velocity $u'$ for eddies of size $l_k$ is then related to the local energy dissipation rate per unit mass of the fluid $\varepsilon$:

$$u' = (\varepsilon l_k)^{1/3}$$  (3)

More generally, in the inertial field for isotropic turbulence in the drop length scale, $u'$ is given by:

$$u' = C_1 \varepsilon^{1/3} d_{\text{max}}^{1/3}$$  (4)

where $C_1$ is a constant of order unity (Walstra, 1993). The maximum diameter of drops, $d_{\text{max}}$ is:

$$d_{\text{max}} = C_2 \varepsilon^{-0.6} \delta_{0.6} \rho_{\varepsilon}^{-0.6}$$  (5)

Eq. (5) indicates that in a turbulent flow of diluted dispersed flows, the viscosity forces are negligible, and drops are often considered to break up in the inertial subrange of turbulence.

**Nomenclature**

- $a$ and $b$: Drop size distribution parameters
- $a_0$: Inlet hub length, m
- $C_d$: Drag coefficient
- $c$: Outlet hub length, m
- $D$: Pipe diameter, m
- $D_p$: Arithmetic mean diameter of two parcels, m
- $d_p$: The particle diameter, m
- $d_{\text{max}}$: Maximal drop diameter, m
- $d_{\min}$: Minimal drop diameter, m
- $d_{u_{10}}$: Diameter that corresponds to 10% by volume of drops with diameters less than $d_{u_{10}}$, m
- $d_{u_{90}}$: Diameter that corresponds to 90% by volume of drops with diameters less than $d_{u_{90}}$, m
- $d_{u_{50}}$: Diameter that corresponds to 50% by volume of drops with diameters less than $d_{u_{50}}$, m
- $d_{32}$: Sauter mean drop diameter, m
- $d_0$: Shell diameter, m
- $d_i$: Hub diameter, m
- $f$: The friction factor
- $f_h$: Diameter of Malvern sampling pipe, m
- $F$: Additional acceleration, N
- $G_k$: Generation of turbulence kinetic energy due to mean velocity gradients
- $G_{\varepsilon}$: The generation of turbulence kinetic energy due to buoyancy
- $l_k$: Kolmogorov scale, m
- $m_p$: Mass of the drop, kg
- $Q$: Inlet flow rate, m$^3$/h
- $R_e$: The relative Reynolds number
- $R$: Radius of the pipe, m
- $R_1$: Inner radius of cross section, m
- $R_2$: Outer radius of cross section, m
- $u_c$: Velocity of the continuous phase, m/s
- $\bar{u}$: Fluid phase velocity, m/s
- $\bar{u}_p$: The particle velocity, m/s
- $v_r$: Local radial velocity, m/s
- $v_x$: Local axial velocity, m/s
- $v_t$: Local tangential velocity, m/s
- $v^*$: Mixture viscosity, m$^2$/s
- $v_c$: Kinematic viscosity of the continuous phase, m$^2$/s
- $W_{\text{Crit}}$: Critical Weber number
- $\chi_0$: Displacement of the drop oscillator, m
- $\gamma_M$: The contribution of the fluctuating dilatation in compressible turbulence to the overall dissipation rate
- $\sigma$: The interfacial tension, N/m$^2$

**Greek letters**

- $\alpha$: Inlet oil phase concentration, %
- $\gamma$: Diffusion number
- $\varepsilon$: Energy dissipation per unit mass, m$^2$/s$^3$
- $\varepsilon^*$: Modified energy dissipation, m$^2$/s$^3$
- $\theta$: Swirling vane outlet angle, $^\circ$
- $\mu$: Molecular viscosity of the fluid, Pa s
- $\rho_c$: The density of the continuous phase, m$^3$/s
where the velocity of the continuous phase is \( u_c \), the density of the continuous phase is \( \rho_c \), the interfacial tension is \( \sigma \), and the pipe diameter is \( D \). The friction factor, \( f \), can be calculated for a smooth conduit from the Blasius equation:

\[
f = \frac{0.0791}{Re^{0.25}}
\]

Moreover, Kubie and Gardner (1977) developed the following relationship to predict \( d_{\text{max}} \) (Angeli and Hewitt, 2000):

\[
\left( \frac{d_{\text{max}} \rho c^2}{\sigma} \right)^{2/3} \left( \frac{f d_{\text{max}}}{4D} \right)^{2/3} = 0.369
\]

As for dense dispersions, a group of researchers have addressed drop breakup. Tsouris and Tavlarides (1994) suggest modification of energy dissipation considering the turbulence damping of dispersed drops. Depending on the mixture viscosity, \( \nu^* \), the modified energy dissipation can be shown as \( \epsilon^* = \epsilon_c (\nu^*)^{3/2} \). Brauner (2001) suggested that in dense dispersions where local coalescence is prominent, the maximal drop size, \( d_{\text{max}} \), can be evaluated based on a local energy balance. Considering quality problems for photographs, which will discount the sample size, an indirect way to obtain \( d_{\text{max}} \) with assistance of Sauter mean diameter is developed according to Eq. (10):

\[
d_{32} = \frac{d_{\text{max}}}{1 + a \exp \frac{\delta}{1 + 2\xi}}
\]

where \( a \) and \( \delta \) are distribution parameters, which can be determined by Eqs. (11) and (12).

\[
a = \frac{d_{\text{max}} - d_{\nu_0}}{d_{\nu_0}}
\]

\[
\delta = 0.394 \frac{\log_{10} \nu_{0}}{\log_{10} \nu_{c_0}}
\]

where \( \nu_0 = d_{\nu_0} / (d_{\text{max}} - d_{\nu_0}) \) and \( \nu_c = d_{\nu_c} / (d_{\text{max}} - d_{\nu_0}) \).

In the dynamic drop renewal process, the turbulent kinetic energy flux in the continuous phase should exceed the rate of surface energy generation required for the renewal of drops in the coalescing system. As research continues, more drop breakup models are proposed and employed in numerical simulations, including the Taylor analogy breakup (TAB) model, the wave breakup model, the KHRT breakup model, which combines the effects of Kelvin–Helmholtz waves driven by aerodynamic forces with Rayleigh–Taylor instabilities, and the discrete random event based stochastic secondary drop (SSD) model.

2. Experimental design

2.1. Measuring method and systematic error

In the experiment, measuring methods are first to be taken into consideration. According to the objective of the research work, the measurements should show the variation of drop diameter distribution before and after the swirling guide vane and phase fraction. Based on the literature (Nigmatulin et al., 2000; Balachandar and Eaton, 2010; Zhang and Xu, 2015; Ding et al., 2017), approaches able to gauge drop size distribution are standard laser-doppler anemometers (LDA), phase-doppler anemometers (PDA), photomicrographs, and Malvern RTSizer. Among them, the malvern RTSizer, developed on the theory of Fraunhofer diffraction for drop size measurement, has been proven to be suitable for oil–water dispersion measurement (Simmons et al., 2000).

Principle of malvern RTSizer is shown as Fig. 1. Drops passing through the beam scatter light. At small forward angles, the scattering is predominantly diffraction. The scattered light is detected by a set of concentric annular detectors placed at the local point of a Fourier transform lens. This means that each detector picks up light scattered at a specific angle and independent of the position of the drops. Thus, the drop size distribution would be directly obtained or through mathematical complications.
Jeelani et al. (2005) investigated size distribution and concentration of rapeseed oil drops on the kinetics of creaming, and interdrop and interfacial coalescence in surfactant-free dispersions with Malvern as measuring method. Brunazzi et al. (2003) presented new experimental data of three different axial separators and a new design model at atmospheric conditions with assistance of Malvern Particle Sizer. As with systematic error analysis of Malvern RTsizer, researchers have performed investigation for decades. Hirleman et al. (1984) studied response characteristics of Malvern 2200 theoretically and experimentally. He found that variations due to a combination of detector calibration errors and nonideal lens effects bought about 15% variations in the instrument response. Simmons et al. (2000) conducted comparison on measuring methods of Malvern and laser backscatter technique and found that the Malvern perform well under 3% volume fraction.

For the phase fraction, electrical resistance tomography (ERT) has been used to study the multiphase flow by measuring the voltages of the electrodes (Olerni et al., 2013; Singh et al., 2016; Faraj et al., 2015; Jia et al., 2017). As with in a liquid–liquid system, Jia et al. (2017) succeeded to apply ERT for an oil-in-water flow measurement and compared with visual observation, demonstrating that ERT could visualize the oil phase distribution and monitor the oil volume fraction up to 0.4. In this work, there were two circles of electrodes, named section I and section II, they were placed 11.28 and 13.48 downstream of the guide vanes, in relation to the developing period and the stable period of the swirling flow field, respectively. In each plane, sixteen stainless steel electrodes are mounted flush to the surface at equal intervals. The oil volume fraction was determined from Maxwell equations. The concentration profile obtained using ERT could be erroneous to a certain level due to the highly sensitivity of factors, such as the accuracy of the electrical measurement made at the system boundary and image reconstruction algorithm used. Reference measurement error of 1% could lead to conductivity error of up to 10% depending on the magnitude of the conductivity charge (Wang et al., 1999). Prior to collecting data we calibrated the ERT system and took the reference frame when the sensor was full of liquid only so that the reference measurement error could be controlled within 1%.

2.2 Materials and swirling guide vane configuration

In this work, tap water and industrial white oil PS-40, produced by the Yanshan Petrochemical Company in China, are applied to prepare oil–water dispersions. The white oil is a refined mineral oil consisting of saturated hydrocarbons and can be classified due to viscosity. At 20 °C, the density of white oil is 840 kg/m³ with a viscosity of 36 mPa s, and the oil–water surface-tension coefficient of 0.032 N m. All experiments are conducted at approximately 20 °C.

Since this work mainly concentrates on the drop behavior in a swirling flow field, guide vane design is not the primary target in the text. The swirling vane applied in the experiment loop is based on the work of Cai et al. (2014), and the key parameters are listed in Table 1. A total of six swirling vanes with a 0° inlet angle and 60° outlet angle are fixed on the hub equally at 60° intervals. The vane thickness is 5 mm, and detailed parameters for the swirling vane can be found in Table 1.

### Table 1 - Dimension for swirling guide vane.

| Structural parameters | Unit | Dimension |
|------------------------|------|-----------|
| Shell diameter(d_s)    | mm   | 100       |
| Shell length           | mm   | 121       |
| Hub diameter(d_h)      | mm   | 90        |
| Inlet hub length(d_i)  | mm   | 80.5      |
| Outlet hub length(d_o)| mm   | 67.5      |
| Swirling vane outlet angle(α) |  | 45        |
| Swirling vane inlet angle |  | 0         |
| Swirling vane axial rotation angle |  | 60        |

2.3 Experimental design and operating methods

Considering current laboratory conditions, the swirling vane, measurement equipment and observational components are assembled in turns, as shown in Fig. 2. In the test section, a Coriolis mass flowmeter (CMF for abbreviation) is applied for density and flow rate measurement of the mixture flow. Two pressure gauges are fixed next to the CMF and at the end of the test section to measure the pressure drop in the tube. Two pieces of sampling equipment are connected to the Malvern RTsizer and arranged before and after the swirling vane to test the drop size change. Elbows of sampling equipment can be moved to ensure that the maximum points along the radius in flow cross section are sampled. Two sampling locations are arranged at the section as shown in Fig. 3. Adjacent to the sampling equipment is the double pixels for EIT 3000, which measures the phase fraction of the swirling field. Moreover, a one-meter-long horizontal tube is set for visual observation and photographs. The entire test section is made up of plexiglass. The testing section tube has a 100-mm inner diameter, which is similar to the separator applied in the oil extraction.

The laboratory experiment loop is able to address oil–water dispersions. A dosing pump capable of controlling the oil flow rate with enough accuracy is applied to drive the white oil into the water. The white oil exiting the dosing pump (with

![Fig. 1 - Principle of Malvern RTsizer.](image-url)
For 3.1.1. Collecting Fig. through tence size angle drop of several volume drop are simulation L/min and of the maxima measured the case, far the inlet tube. The diameter diameters in each measurement gauge the local diameter. While, dv10, dv50, and dv90 are diameters that correspond to 10%, 50%, and 90% by volume fraction of drops with diameters less than dv10, dv50, and dv90, respectively. And, EIT data are included. The existence of random errors during metering, gauge reading and drop size distribution parameter is considered and repeated several times.

3. Numerical study

3.1. Numerical simulation model

3.1.1. Turbulent model

For simulation of continuous phase, though difference exists, RNG k-ε turbulence model is proper for cases involved. The RNG k-ε turbulent model, derived using renormalization group theory, including refinements on accuracy for rapidly strained flows, has an effect on swirl turbulence while enhancing the accuracy of swirling flows. What’s more, swirl or rotation is considered in the model which makes it better for simulation of swirling flow field. The model’s transport equations are as follows:

\[
\frac{\partial}{\partial t} (\rho k) + \frac{\partial}{\partial x_j} (\rho k u_j) = \frac{\partial}{\partial x_j} \left( \alpha_k \mu_{\text{eff}} \frac{\partial k}{\partial x_j} \right) + G_k + G_b - \nu \varepsilon - Y_M + S_k
\]

(13)

\[
\frac{\partial}{\partial t} (\rho \varepsilon) + \frac{\partial}{\partial x_j} (\rho \varepsilon u_j) = \frac{\partial}{\partial x_j} \left( \alpha_\varepsilon \mu_{\text{eff}} \frac{\partial \varepsilon}{\partial x_j} \right) + G_\varepsilon - G_\varepsilon - Y_M - \nu k + S_\varepsilon
\]

(14)

in the equations, \( G_k \) represents the generation of turbulence kinetic energy due to mean velocity gradients. \( G_b \) is the generation of turbulence kinetic energy due to buoyancy. \( Y_M \) represents the contribution of the fluctuating dilatation in compressible turbulence to the overall dissipation rate. \( Y_M \) can be neglected in the numerical simulation due to low velocity compared with local sound velocity. The quantities are the inverse effective Prandtl numbers for \( k \) and \( \varepsilon \), respectively. \( S_k \) and \( S_\varepsilon \) are user-defined source terms. \( E \) is nonzero due to existence of turbulence.

Researchers tend to apply RNG k-ε turbulent model for simulation of swirling flow field. Escue and Cui (2010) conducted thorough investigation on the differences of turbulence models in simulating the swirling pipe flows and found that RNG k-ε model was in well accordance with experimental results in conditions that swirling strength was not so strong. Besides, Javadi et al. (2016) gave an experimental and numerical investigation on the swirl unsteadiness generated in a swirl apparatus. In the research, RNG k-ε turbulent model was applied on the numerical simulation of swirling flow field which has similar configuration on vane topology investigated in this work and showed well accordance with experimental in mean velocity prediction. Jawarneh et al. (2008) conducted a numerical study of two-phase, strongly swirling flow in a cylindrical separator with double vertex generators to predict the separation efficiency of oil and sand RNG k-ε with turbulent model. In the research, simulation results had ideal agreement with experiment in terms of mean tangential velocity and the mean pressure profiles.

3.1.2. Discrete phase model

The discrete phase model (DFM) is able to simulate the discrete phase with attention paid to drop breakup and coalescence. The fluid phase is treated as a continuum by solving the Navier–Stokes equations, while the dispersed phase is solved.
by tracking a large number of particles, bubbles, or drops through the calculated flow field. The force balance equates the particle inertia with the forces acting on the particle and can be written (for the x direction in Cartesian coordinates) as:

\[ \frac{du_x}{dt} = F_D (\bar{u} - \bar{u}_p) + \frac{\partial (\rho_p - \rho)}{\rho_p} + \bar{F} \]  \hspace{1cm} (15)

where \( \bar{F} \) is an additional acceleration (force/unit particle mass) term. \( F_D (\bar{u} - \bar{u}_p) \) is the drag force per unit particle mass. \( F_D \) is related with relative Reynolds number \( Re_r \), which can be calculated in form of Eq. (19). In this work, \( Re_r \) is less than 1000, which means laminar assumption is proper and drag coefficient, \( C_d \) can be calculated as \( C_d = \frac{24}{Re_r} \) (Moris and Alexander, 1972). Hence, the \( F_D \) can be calculated as Eq. (17).

\[ Re_r = \frac{\rho_p b \bar{u}_p - \bar{u}}{\mu} \]  \hspace{1cm} (16)

\[ F_D = \frac{18 \mu C_D Re_r}{\rho_p b^2} \]  \hspace{1cm} (17)

where \( \bar{u} \) is the fluid phase velocity, \( \bar{u}_p \) is the particle velocity, \( \mu \) is the molecular viscosity of the fluid, \( \rho_p \) is the fluid density or the density of the particle, and \( b \) is the particle diameter.

The DPM model has been in use for discrete phase analyzing in the hydrocyclone flow field. Elsayed and Lacor (2011) investigated cyclone separator and the effect of the cone tip-diameter on the flow field with assistance of large eddy simulation (LES) and detected separation efficiency through DPM model. El-Batsh (2013) coupled RSM-DPM model to optimize dimensions of the exit pipe to improve cyclone performance in gas-liquid separation. Recently, Mokni et al. (2015) applied DPM model in order to simulate particle trajectories and predict the separation efficiency.

3.1.3. Drop breakup

In the present study, TAB model is applied to simulate drop break up. This model is proposed on the basis upon Taylor’s analogy between an oscillating and distorting drop and a spring mass system. The equation governing a damped, forced oscillator is:

\[ F - k x_0 - d_p \frac{dx_0}{dt} = m_p \frac{d^2 x_0}{dt^2} \]  \hspace{1cm} (18)

where \( F \) is the external force of the drop, \( x_0 \) is the displacement of the drop equator form its spherical position and \( m_p \) is the mass of the drop. The coefficients of the equation are taken from Taylor’s analogy. The drop is assumed to break up if the distortion grows to a critical ratio of the drop radius. This breakup requirement is given as \( x_0 > C_0 r \) where \( C_0 \) is a constant equal to 0.5. The size of child drops is determined by equating the energy of the parent drop to the energy of the child drops.

3.1.4. Drop coalescence

Drop collision may result in drop coalescence in the dispersion. O’Rourke (1981) proposed parcels, groups of particles, to enhance calculation efficiency. He made the assumption that two parcels might collide only if they were located in the same continuous-phase cell. If the continuous-phase cell size is small in size, the method is second-order accurate at estimating the chance of collisions. Probability of two parcels to collide is shown as Eq. (19):

\[ P_1 = \frac{\pi (r_1 + r_2)^3 v_{rel} \Delta t}{V} \]  \hspace{1cm} (19)

where \( v_{rel} \) is the relative velocity between two parcels, \( r_1 \) and \( r_2 \) is radius of the two parcels respectively, \( \Delta t \) is the time increment and \( V \) is the volume of the cell involving the two parcels.

Moreover, critical offset number is proposed to check whether the parcels coalescence or bound off. In case of a grazing collision, the new velocities are calculated on the basis of conservation of momentum and kinetic energy. O’Rourke (1981) derived the expression for new velocity on the assumption of kinetic energy of the drop is lost to viscous dissipation and angular momentum generation.

\[ v' = \frac{m_1 v_1 + m_2 v_2}{m_1 + m_2} \left( 1 - \frac{b - b_{crit}}{b_{crit}} \right) \]  \hspace{1cm} (20)

In the equation, \( m_1 \) and \( m_2 \) are mass of two parcels, \( v_1 \) and \( v_2 \) are velocity of the parcels before collision, respectively. And \( b \) is the offset number.

3.2. Geometry and boundary conditions

A numerical study is conducted on the basis of a real swirling vane applied in the experiment with the assistance of commercial simulation software ANSYS workbench and ANSYS multiphysics. The detailed configuration for the swirling vane can be found in Figs. 2 and 4. The vane thickness employed in the experiment is less than 5 mm, which can be neglected compared to the 100-mm inner diameter. To compare with the experimental results, the simulation model should contain all the listed test sections. As a result, the numerical domain is a 2.5-m-long solid cylinder, which has a hollow core corresponding to the hub of the swirling vane. Moreover, the flow domain is divided by the space curved face fitted to the vane shape.

In terms of boundary conditions, the work should be done in two parts including continuous phase of Navier–Stokes turbulence model and DPM model. (i) For continuous phase of Navier–Stokes and turbulence model, the inlet of the flow domain for all case is defined as velocity inlet, with fixed hydraulic diameter 0.1 m. The velocity ranges from 0.424 to 0.637 m/s, corresponding with inlet flow rate, respectively. As with outlet of the flow domain, pressure outlet is fixed for all cases with specified pressure corresponding with experimental gauge pressure. Other boundaries such as vanes, hub and pipe wall are defined as fixed wall boundary condition. (ii) For DPM model, drop injection type is defined as surface injected Rosin–Rammier drop size distribution with specified mass flow rate calculated from inlet flow rate and volume fraction. In the Rosin–Rammier distribution, key parameters \( d_{50} \) and \( n \) are defined on the basis of logarithmic linear regression on the inlet drop distribution obtained from Malvern. At the inlet and outlet of the flow domain, particle behavior is defined as escape, while on other faces, the discrete phase is set as trapped, namely, drops may be adhesioned to walls in cases of collisions with walls.

Fig. 5 present comparison between Malvern sampling and numerical discretize for drop size distribution at the inlet of flow domain. The DPM model predicts behavior of drop by solving parcel equation of force balance. Discrete drop is injected into the flow field with \( N \) groups classified by diameter in order to simulate drop size distribution gotten from experi-
mental observation. As the inlet area is consist of numbers of faces in the grid, each face are injected with \( N \) independent parcels as the basic calculation unit in the simulation. The section drop distribution parameter is obtained from these parcels. Considering the computational cost, \( N \) should not be larger than 15 in the simulation. As a consequence, discrepancy may come out between experimental observation and numerical discretize. Moreover, in the numerical simulation, drop break up and coalescence exist, deviation in drop size distribution sampling location between numerical simulation and experiment may bring slightly variation as well. Though deviations exist, drop distribution between numerical simulation and experiment for all cases are nearly the same in amplitude and peak value diameter.

3.3. Solution method

On the basis of experimental operating conditions, Reynolds number for the fluid domain ranges from 42,400 to 63,700, all of which locates in the turbulent region. When conducting a numerical simulation, RNG \( k-\varepsilon \) model is applied for turbulent model of continuous phase. Discrete Phase Mode (DPM) based on the Euler-Lagrange equation is used for drop simulation in the swirling flow field. On the basis of numerical model in Section 3.1, transient manner is essential when drop break up and coalescence are considered. Coupling between the continuous phase and discrete phase is necessary. Moreover, influence of turbulence on the drop behavior should be considered. Hence, stochastic collision and stochastic and discrete random walk model are selected. In terms of time step, Motin (2015) performed numerical oil separation simulation on hydrocyclone with tangential inlet. In his work, the time step was set as 20–50 ms. Similarly, considering grid quality, time step size is set as 50 ms in this work. In terms of global simulation time, the slowest inlet velocity is 0.424 m/s. With 2.5-m-length flow domain, 5.9 s is needed for the fluid flowing through flow domain averagely. Moreover, velocity in the pipe center is slower than the average velocity, let along oil core formation and stabilizing. As a consequence, 10 s is essential for the flow domain simulation.

Weighing both solution convergence and accuracy, the semi-implicit method for pressure-linked equations (SIMPLE) is applied for coupling between the pressure and velocity. Discrete formats for momentum, kinetic energy, dissipation and Reynolds stress equation are operated in a second-order upwind form. Time discretization is achieved by the first-order implicit scheme, while all gradients are discretized based on the least squared cell-based approach. And all the residual scales are set as \( 10^{-6} \).

3.4. Mesh quality and independence study

A grid is created with the assistance of commercial software ICEM CFD in the form of an O-grid structure hex cell. Fig. 6 illustrates the grid configuration in the vane part, and nearly all grids are hexahedral. Quality criteria for the grids are provided in the text as well.

A mesh independent study, whose reference parameters are facial average static pressure and velocity magnitude with 0.637 m/s inlet velocity, is conducted with the RNG \( k-\varepsilon \) model before the numerical study. Grid refinement is conducted axially, peripherally and radially, as is shown in Fig. 7. The figure is based on three levels of grid quality: a relatively coarse grid (a) with 230,796 cells, a relatively medium grid (b) with 548,424...
cells, and a relatively fine grid (c) with 829,692 cells. Fig. 7 depicts the static pressure and velocity magnitude change axially with different grid quality. It can be seen that deviations from grid (c) to grid (b) are much smaller than those of grid (b) to grid (a), both on perspective of static pressure and velocity magnitude. The average relative deviations for static pressure and velocity magnitude from grid (b) to grid (a) are 0.24% and 4.1% respectively, and 0.21% and 0.7% from grid (c) to grid (b). As a consequence, grid (c), with 829,000 cells, is fine enough to satisfy the grid independence requirements. A numerical study is conducted on the grid (c) scheme.

4. Results and discussion

4.1. Experimental observation

Experimental data are summed up into two parts: a data group characterizing drop size distribution measured by Malvern RTSizer, and a data group characterizing discrete phase volume fraction gauged by EIT 3000.

4.1.1. Error analysis for measurement system

The theoretical model is applied to prove the reliability of particle size distribution obtained by Malvern. According to research conducted by Simmons et al. (2000), linear fittings are conducted on \( d_{90} \), \( d_{\max} \), and \( d_{32} \) obtained from the experiment according to Eqs. (10)–(12), as shown in Fig. 8. A linear relationship does exist between \( d_{\max} \) and \( d_{32} \), with distribution parameters \( a = 1.075 \) and \( b = 0.73 \), and the corresponding values are 1.2 and 0.9, respectively for kerosene-water dispersion (Karabelas, 1978). Though discrepancies may exist due to the material attributes of white oil and kerosene and pipe diameter variation, the drop size distribution measured by Malvern is acceptable.

As is said above, in the experiment, measurement of drop size distribution was carried several times for each case. As a consequence, the maximum and minimum for drop distribution characteristic parameter of certain case exists, which contributes the error bar in Figs. 9–12. Moreover, the range of error bar is less than 15% of the value for certain case, which correspond well with Hirleman et al. (1984). For the measurement by ERT, the range of error bar is 10% of the measured pixel originated from possible error in the sensor, as shown in Fig. 23. Error analysis above proved measuring method is proper and the data matrix is reliable for investigation next.

4.1.2. Drop breakup and coalescence

On the basis of the data matrix obtained with assistance of Malvern, the oil drop behavior of breakup and coalescence can be detected. Two perspectives of analysis are performed with consideration of drop size distribution parameters \( d_{32} \), \( d_{10} \), \( d_{50} \) and \( d_{90} \).

When the flow rate is set at 18 m³/h, drop distribution characteristic parameters are carefully recorded, as shown in Figs. 9 and 10. It is obvious that \( d_{32} \) and \( d_{10} \) rise as dispersion flows through the vane zone, while \( d_{50} \) decreases for inlet oil concentrations higher than 0.1%. \( d_{90} \) shows little change in size before and after the vane zone. Trends for the drop size distribution characteristic parameters above are coinci-
dent for sampling location both adjacent to the center and away from the center. This phenomenon is theoretically reasonable. As for small drops characterized by $d_{v10}$, they tend to coalescence after the vane zone. For large drops characterized by $d_{v90}$, they tend to breakup due to shearing force of continuous phase. As a parameter characterizing median diameter, with small drops coalescence and large drops breaking up, $d_{v90}$ may not change drastically. As with $d_{32}$, a parameter characterizing turbulence dissipation is related to turbulent dissipation energy. According to Zhang and Xu (2015), augmentation of $d_{32}$ indicates a decline of turbulent dissipation energy. Lower turbulent dissipation means more ordered flow. From a change of $d_{32}$ when dispersion flows through the vane zone, it can be seen that turbulent dispersion changes to a more ordered rotational flow.
Another perspective links oil volume fraction with the deviation of continuous phase flow rate. Figs. 11 and 12 illustrate a drop behavior flowing through the swirling vane with an oil volume fraction of 0.5%. In the sampling location near the center, coalescence is prominent for small drops, while large size drop behavior turns from coalescence to breakup as the dispersion inlet flow rate increases from 12 m$^3$/h to 18 m$^3$/h. The median diameter changes little except for a small flow rate of 12 m$^3$/h. That is, for large flow rates, small drops easily coalesce, while large drops tend to breakup with totally average diameter remaining nearly unchanged. Comparatively, in a location away from the center, all drop distribution parameters are inclined to rise with low flow rate, especially for 12 m$^3$/h. This is probably because in relatively weak swirling fields, more dispersion phases are located away from the center; thus, chances for drop collision and coalescence are higher than high flow rates. For medium flow rates, drop distribution parameters do not change drastically. Though the increase of velocity might aggravate drop collision probability microscopically, the decline of dispersion phase concentration away from the pipe center likely counteracts the velocity’s effect. Until the flow rate rises to 18 m$^3$/h, the speed up effect wins over the phase dilute effect, and the drop coalescence becomes prominent again for small and medium size drops. However, large drops break up as the shearing effect is enhanced by the higher flow rate. In addition, the phenomenon of $d_{90}$ near the center after the vane zone is slightly larger than $d_{90}$ away from the center, in accordance with the theoretical analysis, as the shearing effect is more intense away from the pipe center, where the tangential velocity is faster.

On the basis of discussion above, influence of parameters on drop breakup and coalescence performance for dispersion flowing through vane zone could be compared. Comparison for parameters is inlet volume fraction and inlet flow rates. As shown in Figs. 9 and 10, though inlet volume fraction changes, character diameter almost keep constant before and after vane zone. On the other hand, as is shown in Figs. 11 and 12, performance of drop breakup and coalescence various obvi-
ously as the inlet volume flow rate varies from 12 to 18 m³/h. This is reasonable. For drop in relatively dilute dispersions, influence of drop collision originated from drop density is far less than turbulence of continuous phase which is resulted from inlet flow rates. So compared with inlet flow rate, inlet volume fraction is a secondary parameter that influences drop breakup and coalescence in relatively dilute dispersion.

In terms of drop distribution character parameters, trend is also clear from Figs. 9 and 10. Removing influential parameter inlet volume fraction as the inlet flow rate fixed at 18 m³/h, d_{50} and d_{10} are more susceptible of breakup and coalescence. Moreover, though inlet flow rate is considered in Figs. 11 and 12, d_{50} and d_{10} are more susceptible than other character parameters when drop breakup and coalescence happens. This means that small drops are easier to coalescence and the energy dissipation is lower in the swirling flow field. Large drops, on the other hand, are limited by the shearing effects of continuous phase, which obstructs them from changing so much in size though breakup and coalescence exists.

4.1.3. Dispersion phase concentration
The laws for drop migration in a swirling flow field are characterized through the EIT data matrix. Fig. 13 shows a comparison of oil phase concentration contours and photographs of the observation tube. In the cross-sectional oil phase concentration images, the blue region represents low oil phase concentration, while the red region represents high oil phase concentration, with transitions from green to yellow. For convenience of depicting oil distribution, the red region represents diverse peak values in different operating conditions. Obviously, the photographs illuminating the oil core width show an increase with inlet phase concentration rising in a fixed flow rate. Moreover, the core distribution in the contour obtained from EIT corresponds well with the photograph. Though low phase concentration and possible oil contamination may lower the sensitivity of EIT electrodes and cause discrepancies in some cases, the total trend collected from EIT is obviously credible.

As shown in Figs. 14 and 15, oil cores do exist in the center for the test section. Interestingly, when presenting images of different flow rates and volume fractions together, it can be seen that the oil core size narrows with the augmentation of the dispersion flow rate once the oil phase concentration is specified. In contrast, oil core size magnifies with increased oil volume fraction in the fixed dispersion flow rate, as the photographs indicate. Theoretically speaking, radial acceleration increases with an increase in the flow rate, lighter phase drops migrate faster to the center and converge in high flow rates, which results in a narrower and more stable core when the volume fraction is fixed. In contrast, as with a constant flow rate, assuming constant axial slip velocity between oil and water, higher volume fractions result in dense drops converging to the center, resulting in a wider core.

Further discussion is conducted on the horizontal centerline of the EIT sampling cross section, as illustrated in Figs. 16 and 17. Corresponding to the drop behavior, Fig. 16 depicts the volume fraction distribution under a constant flow rate of 12 m³/h and 18 m³/h with various oil volume fractions. A higher volume fraction induces a larger cross-sectional volume fraction magnitude and a wider core. In contrast with the volume fractions, the augmentation of the dispersion flow rate steepens the curve and increases the curve amplitude, which leads to narrow cores. This further proves the discussion above. Though discrepancies exist in the 3.0% phase concentration of the 18 m³/h flow rate, the total trend above is obvious. A high flow rate is helpful for discrete phase con-
vergence and core stabilization due to high slip velocity after dispersion flows through the swirling vane.

4.2. Numerical simulation

4.2.1. Simulation for continuous phase

In the transient simulation, the flow domain becomes stable at 10s. Fig. 18 illustrates the flow field distribution with an 18 m$^3$/h inlet flow rate at a time of 10s. The red color represents high values, while the blue color represents low values, with a transition from green to yellow. Physical values are high in the red region and low in the blue region. Cross sections A, B, C and D correspond to 0.11 m, 0.13 m, 0.32 m and 0.68 m from the entrance area of the blades. In cross section A, the outline for the guide vane rotates counter-clockwise from the view of the inlet-to-outlet direction. In each zone separated by adjacent blades, axial fluids are obstructed and guided to flow tangentially, especially for fluids near the right blades.
As a result, in the blade-separated zones, the right side has relatively high pressure and low velocity, while the left side has relatively low pressure and high velocity. A vortex starts to appear in each blade-separated zone. After fluids flow out of the swirling zone, as depicted in cross section B, six small vortexes begin to merge to a larger one. A vortex core with high tangential and axial velocity is still obvious at section B. When dispersion arrives at cross section C, large stable vortexes have formed, replacing the six small vortexes. In the core region, pressure and velocity is low. Additionally, axial velocity even indicates back flow. A longitudinal profile for pressure and stream line are better able to describe the phenomenon above. In Figs. 19 and 20, pressure drop have been observed radically for a long distance after vane zone. As with the stream line, rotational flow is more obvious. Under the pressure gradient and rotational flow, oil drops transfer to the center and are gathered in the core, as shown in Fig. 13.

Fig. 21 further depicts the pressure and velocity distribution in the Malvern RTsizer sampling area (i.e. 200 mm after the vane zone) with different inlet flow rates at 10 s, when the fluid domain is stable. For convenience of comparison of pressure distributions, outlet pressures for all cases are set as 0. In the figure, it is clear that as the inlet velocity is fixed to a certain value, take 16 m/s as an example, the pipe center suffers from low pressure and high velocity. When the position moves from the pipe center to the wall, velocities (velocity magnitude, tangential velocity and axial velocity) increase gradually to a maximum and decrease to zero at the wall. Unlike the velocities, the pressure gradually rises to a maximum at the pipe wall. Moreover, the pressure and velocity gradients become steeper as the inlet flow rate increases, while the curves for the axial velocity of different inlet flow rates remain nearly parallel. With a steeper gradient of pressure drop and tangential velocity in the high flow rate, the vortex becomes stronger. In addition, the position of maxima for all the variables is nearly fixed.

4.2.2. Swirling number and diffusion number

Swirling number is a dimensionless number characterizing swirl field intensity and has been applied in field of hydraulic turbomachines and turbulent jets (Wannassii and Monnoyer, 2016). Several forms of swirling number are in use by different groups of researchers (Ahmed et al., 2015). Considering literature till now, swirling number in this work is defined as ratio of the axial flux of tangential momentum to the axial flux of axial momentum:

\[
\Omega = \frac{1}{R_2^3} \int_{R_1}^{R_2} r^2 v_r v_r dr
\]

(21)

where \( R_2 \) is the outer radius of the cross section, \( R_1 \) is the inner radius of the cross section, \( v_r \) and \( v_x \) are tangential and axial velocity respectively.

Similarly, a dimensionless number characterizing diffusion of swirling field is defined as ratio of axial flux of radical momentum to the axial flux of axial momentum:

\[
\tau = \frac{\int_{R_1}^{R_2} r v_r v_r dr}{\int_{R_1}^{R_2} r^2 v_x^2 dr}
\]

(22)
where $v_r$ is the local radical velocity. In the equation, positive diffusion number means divergence while negative diffusion number means convergence radically. Diffusion number gives depiction of radical transfer in the flow domain. Through change in diffusion number, drop behavior can be clarified to some extent.

Fig. 22 gives a through depiction of swirling number and diffusion number in the flow domain where swirling number values are corresponding to left Y axis while diffusion number

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**Fig. 18** – Contour for flow field distribution. (For interpretation of the references to colour in the text, the reader is referred to the web version of this article.)

**Fig. 19** – Longitudinal pressure distribution.
values are corresponding to right Y axis. As with X axis, values are nondimensionalized by inner diameter D to character where zero means section right after the vane zone. It’s obvious from the figure that curves for different inlet flow rates are nearly overlapping to each other which means inlet flow rate has little effect on the swirling number and diffusion number under low inlet flow rate operating conditions. The swirling number continuous to raise for a short distance after flowing out of vane zone to a maximum of nearly 1.3 at axial location z/D = 3 then gradually decrease until a abruptly decrease at location of z/D = 23. Correspondingly, curves for diffusion number are negative with abrupt increase from z/D = 0 to z/D = 3 til 0.1 of local maximum, then decrease slowly until abruptly increase at location of z/D = 23. Considering two group of curves together, regularity for swirling flow field is illuminated after vane zone. Swirling flow firstly intensified in the location of z/D = 0 to z/D = 3 due to influences of swirling guide vanes, then keeps stable with little dissipation from z/D = 3 to z/D = 23 axially. At axial location of z/D = 23, swirling intensity precipitately decrease with low swirling number and high diffusion number which means swirling flow has already diverged. Through curves of swirling number and diffusion number, it can be assert that device for oil core collection should be placed between z/D = 5 and z/D = 23 axially for this type of swirling vane where oil core is compact and stable. Similar system can conduct analogous analysis to determine position of collection equipment.

4.2.3. Drop trajectory
The reliability for the discrete phase simulation is examined on the basis of Malvern and EIT 3000 data matrices. Inlet drop size distributions are the first to be examined. As shown in Fig. 5 above, the simulation drop size distributions are in accordance with the experiment in terms of amplitudes, band widths and peak value locations. Further detection is conducted on the distribution parameters of $d_{\text{max}}$, $d_{\text{min}}$, $d_{52}$ and $d_{30}$. They show good accordance between the numerical and
experimental distributions. Phase concentration on the EIT electrode section is another aspect to examine. Fig. 23 shows the numerical and experimental phase concentrations on the EIT electrode section under 18 m$^3$/h inlet flow rate with various inlet oil phase fractions. In comparison, the phase concentrations of the numerical results show narrower bands and higher amplitudes. These differences may be ascribed to the EIT sampling theory of conductivity detection. In low oil phase concentrations (under 3.0% by volume), operating conditions and sensitivity might be weakened as peak value deviation is determined (Li et al., 2009). However, this has little influence on the total research, as areas enclosed by the curve and diameter axis are nearly equal numerically and experimentally in each operating condition.

After demonstrating the correctness of the discrete phase simulation, drop trajectories are detected through the numerical study. Fig. 24 depicts the drop distributions at different times under an 18 m$^3$/h inlet flow rate with 3.0% oil phase fraction. In the figure, dots representing drops are colored by drop size. The color blue represents small drops, and the color red represents large drops, with other colors indicating the transition. In the dynamic developing procedure for oil phase distribution, small drops move faster in the queue and are located far away from the pipe center radially due to their smaller inertial forces and higher continuous phase axial velocity distribution away from the pipe center. This is in accordance with the experimental observations. Most medium and large size drops begin to transfer to the pipe center, leading to the formation of the oil core. The oil core processes slower than the periphery small drops as a result of their larger inertial forces and low continuous phase axial velocity distribution near the pipe center. Furthermore, the oil core gradually converges to a thinner line and becomes stable in the procession. A long axial trip is not required for the oil core to stabilize as a thinner line. Coalescence is prominent in the process after the developing region, where more large drops exit the core in replacement of medium size drops as a result of medium size drops colliding and coalescing in the pipe center.

4.2.4. Drop distribution in the flow field

The detailed drop distribution in the flow field is shown in Figs. 25 and 26. In the figures, the longitudinal section means the section passes through pipe center line from inlet to outlet direction. Fig. 25 shows the drop size distribution of the longitudinal section with both volume fractions and characteristic parameters. The distribution histograms contain information from the inlet to the outlet. Large drops and maximum drop size should be concerned with evaluating drop coalescence behavior after the vane zone. It can be seen that with a fixed inlet oil phase fraction of 3.0%, the flow field with 12 m$^3$/h inlet flow rate produces larger maximum drops, and more drops are larger than 12,000 µm. This is because the higher inlet flow rate produces a larger phase slip velocity and more intense shearing effect of the continuous phase, which reduces the corresponding maximal drop size. In addition, it should be noted that at the pipe center region, tangential and radial velocity is so low that it can be neglected. As a result, drop shearing is limited to the axial direction. Together with the extrusion of the continuous phase, possible large drops become long strips.

Fig. 26 further illustrates the drop distribution along the flow field with 1.0% inlet oil phase concentration under various inlet flow rates at 10.0 s. Distances along axial is nondimensionalized with pipe inner diameter $D$ for analysis of convenience. Parameters $d_{32}$, $d_{90}$, and $d_{\text{max}}$ increase while parameter $d_{\text{min}}$ decreases. Effects of instantaneous turbulence velocity fluctuations on drops enhances chances for both drop collision that is beneficial for drop coalescence and drop oscillation and distortion, which is beneficial for drop break up. This is proper for all sized drops in the swirling flow field. As a consequence of instantaneous turbulence and swirling flow field, part of small drops breakup which lowers $d_{\text{min}}$ in the flow domain and part of the large drops coalescence which raises $d_{\text{max}}$ in the flow field. Together with augment of parameter $d_{90}$ and $d_{32}$, it can be seen that coalescence is prominent in the swirling flow field. Moreover, trends for parameter change are not constant along the flow field. Drop parameter changes abruptly in short distance after entering into the vane zone,
then the change rate gradually slow down to nearly stable though small raise exists, as shown in Fig. 26. Another point attracts attention is the value of $d_{\text{max}}$, which slightly decreases with inlet flow rate increases. Phenomenon described above can be ascribed to shearing effect of continuous phase. High inlet flow rates result in more intense shearing effect that lowers maximum drop stable size in the dispersion. This is in consistent with both experimental results above and theoretical models. Fig. 27 compares numerical results and theoretical model predicting maximum drop size in turbulent dispersion develop by Kubie and Gardner (1977). Numerical maximal drop sizes under different inlet flow rates falls around curves obtained by Kubie and Gardner’s prediction model which validates numerical simulation results are reliable again.

In fact, drop-turbulence interaction can be characterized through relationship between $d_{32}$ and turbulent dissipation energy. Relationship between Sauter mean diameter $d_{32}$ and turbulent energy dissipation is investigated in the numerical simulation as well. On the basis of numerical simulation, Fig. 28 illuminates face average $d_{32}$ and face average turbu-
lent energy dissipation of different cross section in the flow domain under different inlet flow rates. It is clear that linear relationships exist in all the flow rates besides solid dots sampled from sections right after vane zone where the swirling flows are not fully developed. Moreover, the slopes for each flow rate are nearly equal to each other. This phenomenon is resemble to research work done by Zhang and Xu (2015) experimentally, where drop size in the turbulent emulsions decreases with increasing mean energy dissipation and have linear relationship in double logarithmic coordinate system as well. This further demonstrates validity of RNG $k$-$\varepsilon$ – DPM coupling numerical method in simulation of vane type separator in low inlet flow rates and low oil phase concentration.

Another perspective of discussion is conducted on the relationship between parameter $d_{32}$, $d_{\text{max}}$ and swirling number. Shown as Fig. 29, $d_{32}$ and $d_{\text{max}}$ decrease with augment of the swirling number. What’s more, linear relationship appears in the double logarithmic coordinate system. Swirling curves is monotonous axially after peak value near vane zone. In the monotonous section, the swirling number gradually decrease, while turbulent swirling flow turns to be more ordered which means less turbulent energy dissipation. Correspondingly, $d_{32}$ shows linear relationship with turbulent energy dissipation in the double logarithmic coordinate system as shown in Fig. 29.

As a result, nearly linear relationship comes out between swirling number and $d_{32}$ in the double logarithmic coordinate system. The same goes for $d_{\text{max}}$, smaller swirling number means less intense shearing effect and instantaneous turbulent dispersion, all of which are beneficial for maximum drop size increase. So it’s no wonder linear relationship with negative slope exists between swirling number and $d_{\text{max}}$ in the double logarithmic coordinate system.

5. Conclusions

Both experimental and numerical studies were conducted to investigate the drop behavior, including breakup, coalescence and trajectory, of a 100-mm inner diameter horizontal swirling flow field with low inlet velocity oil-in-water dispersions. The inlet oil phase concentration $\alpha$ was under 3.0% by volume, and the inlet flow rate ranged from 12 m$^3$/h to 18 m$^3$/h.

In the experimental study, Malvern KTsizer was applied to measure the drop size distribution both before and after the swirling vane, with sampling points near and away from the pipe center. At the same time, EIT 3000 was used for measuring the oil phase concentration. The experimental results showed that both small and large drops in the flow field tend to coalescence in low inlet flow rates. As the inlet flow rate increases, large drops tend to break up due to the shearing effect of the continuous phase. In most cases, the swirling vane enhances $d_{32}$ of the oil-water turbulence dispersion, which means less energy dissipation and more ordered dispersion. Moreover, the inlet flow rate and oil phase concentration have an influence on oil phase concentration after the swirling zone. A higher inlet flow rate, which creates a more intense swirling flow field, makes the fixed oil phase of the inlet oil fraction dispersion more concentrated in the center, resulting in a narrower oil core. Augmentation of the inlet oil concentration, in contrast, widens the oil core of the swirling flow field.

In the numerical study, the RNG $k$-$\varepsilon$ turbulent model, coupled with the DPM model, was applied for a simulation of the swirling flow field corresponding to the experimental operating conditions. The numerical simulations, conducted in
transient behavior in consideration of drop breakup and coalescence and stochastic tracking model considering effect of instantaneous turbulent velocity fluctuations on drop trajectories, corresponded well with the experimental observations and instrument measurements. Through swirling number and diffusion number, section for steady oil core was ascertained to be between 5–23 time inner diameter, and the section was suitable for arrangement of oil collection equipment. In addition, oil drop behavior of breakup, coalescence and transfer was revealed. Breakup and coalescence occurred simultaneously, resulting increase of $d_{\text{max}}$ and decrease of $d_{\text{min}}$. What's more, drop-turbulence interaction was investigated through detection of relationships between drop parameter $d_{\text{d2}}$, $d_{\text{max}}$, and parameter of continuous turbulent phase, namely turbulent dissipation energy and swirling number. Approximately linear relationship with negative slope existed between drop parameter and parameter of continuous turbulent phase under double-logarithmic coordinate system.

These conclusions above are beneficial for understanding of drop-turbulence interaction and site production of similar equipment. Besides, nondimensionalized analysis process enables research work to be expanded to similar swirling systems such as down-hole separation.

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