Coalescence of the Oil-in-Water Emulsions in Fibrous and Granular Beds

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Petroleum Systems Engineering

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By

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ABSTRACT

In the daily oil production operations, the oil-in-water (O/W) emulsions are often encountered in the produced liquids. It has been a challenging technical problem to effectively separate oil droplets from the O/W emulsions for many years. In this thesis, a mechanical coalescing method was proposed and tested to separate oil droplets from O/W emulsions. The coalescence of the O/W emulsions in a fibrous or granular bed was thoroughly studied. More specifically, the pressure drop across the coalescing bed was successfully predicted by using the Carman-Kozeny equation for single-phase flow. A geometric method and an experimental method were used to calculate the Kozeny constant for the fibrous bed and granular bed, respectively. For two-phase flow tests, pressure drop is slightly increased as compared to that for the single-phase flow. In addition, the effects of flow rate, inlet oil concentration, length of the coalescing bed and fiber size on the coalescence performance were examined. The Sherony and Kintner model was modified and the accumulative drop size distribution curves at the inlet and the outlet were compared to evaluate the coalescence performance, respectively. In the parametric studies, first, it has been found that the coalescence efficiency was not a monotonous function of flow rate in the granular bed. The coalescence efficiency was high at the superficial velocity of 1.0-1.5 mm/s and its maximum was 10.42%. Secondly, the higher inlet oil concentration is, the lower coalescence efficiency is acquired. Thirdly, it has been concluded that, in this study, the coalescence performance in the fibrous and granular beds is insensitive to the length of the coalescing bed in terms of the accumulative drop size distribution curves. Finally, the coalescence performance was better for

smaller fiber size. In summary, this thesis study achieves a better understanding of the O/W emulsion flow in the coalescing beds and helps to design a more effective mechanical coalescing bed for separation of the O/W emulsions.

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CHAPTER 1 INTRODUCTION

Effective separation of oil from the produced fluids has long been a challenging task in petroleum industry. Recent studies have shown that a substantial portion of the operating cost associated with daily oil production is actually spent on oil-water separation [Croasdale *et al.*, 1999]. In fact, there exist several oil-water separation methods, which are proved to be effective in some applications. However, most of them have rather limited applications in separating oil from the produced fluids due to their large capital and/or operating costs, low separation efficiency and/or long residence time. Therefore, some improved or even novel separation techniques are still required.

In general, oil is produced as a mixture of oil and water. This oil-water mixture can be classified into the oil-in-water (O/W) emulsions or the water-in-oil (W/O) emulsions. The O/W emulsions are often encountered in the oil production operations. It should be noted that usually the O/W emulsions are highly stable. This is because some water-soluble surface-active materials and fine mineral particles are often attached onto the oil droplets so as to make the emulsions difficult to demulsify. These O/W emulsions in the produced fluids can be further categorized as the secondary emulsion, which has a droplet size in the range of 1-20 μm . Thus, it takes much longer time to separate these emulsions by using conventional gravity separators.

At present, produced oil and water are usually pumped to a primary gravity separator for initial separation, which is commonly termed as free water knockout

(FWK). After passing through the FWK unit, the W/O emulsions are generally treated with heaters to further remove water and minerals. The separated water from the FWK unit is transported to a skim tank, i.e., so-called API separator, where sufficient long residence time is needed for the O/W emulsion droplets to rise up. Then they can be skimmed off and returned to the oil stream. The nature gravity separation process may be assisted by adding some demulsifiers. In practice, the API separators are effective in removing a considerable amount of oil. However, separation of produced liquids in such a system is incomplete. For the oil droplets in the range of 1-10 μm , it is extremely difficult, if not impossible, to separate them by merely using conventional gravity methods.

The mechanical coalescer, a passive separation apparatus, has been used for some time as an inexpensive means of separating the secondary emulsion. For example, coalescing bed has been used in pollution control, such as wastewater treatment [Redmon, 1963]. Inside the coalescing bed, the drop size is enlarged due to the coalescence process. The enlarged emulsions droplets are then delivered to a separator, where they can be removed. Although the coalescing bed does retain some of the emulsion droplets at the beginning, it is rather different from a conventional filter. In fact, the coalescing bed operates under a steady state, in which the volume concentration of the dispersed phase at the inlet equals to that of the dispersed phase at the outlet. In this thesis, a mechanical coalescing method was proposed and tested to separate the O/W emulsions. Also, the coalescence of the O/W emulsions in a fibrous or granular bed was studied.

1.1 Existing Separation Methods

1.1.1 Gravity separation method

Gravity separation method is one of the oldest separation techniques [Gu, 2001]. The application of this method has been discussed elsewhere [Rao and Patil, 1998; Holland-Batt, 1998; Kenawy *et al.*, 1997; Padilla *et al.*, 1996; Ruiz and Padilla, 1996]. The gravity separation method represents all the relevant phase separation techniques, which separate substances with different densities by gravity. This method finds wide applications in petroleum industry for the separation of produced petroleum liquids. The gravity separation method involves no moving parts and is cost-effective. Up to now, this method is still the most popular separation technique in the oil fields. However, the gravity separation facilities usually occupy an extremely large space and may pose environmental hazards. Also this method suffers from long residence time. Therefore, the gravity separation method may be improved or combined with other separation methods.

1.1.2 Hydrocyclone separation

In the past, the hydrocyclone separation method has been among the most widely used separation techniques particularly for mineral processing and liquid-solid separation. This method has been increasingly used in the surface treatment of produced petroleum fluids since 1980 [Colman and Thew, 1980a; 1980b; Medlum, 1987]. Sometimes the pre-separating hydrocyclone is used to separate the bulk water, largely depending on the oil concentration in the produced liquids. And then a de-oiling hydrocyclone or a de-watering hydrocyclone is used for further separation.

Hydrocyclones have no moving parts and separate substances with different densities by the centrifugal force. The centrifugal field generated in the conical portion makes the lighter phase to move towards the center and reach the overflow outlet. Meanwhile, the heavier phase moves towards the wall and reaches the underflow outlet. It is the back pressure formed at the underflow outlet that pushes the lighter phase towards the overflow outlet. In such a way, phase separation occurs.

It has been well recognized that the hydrocyclone separation method has some distinct advantages, such as high efficiency, continuous operation and high volume handling capacity. However, it is difficult to use this method to separate stable O/W emulsions with small oil droplets (less than 20 μm). Moreover, the density difference between brine phase and medium or heavy oil is small. In these two cases, the hydrocyclone separation method becomes ineffective.

1.1.3 Thermal and chemical methods

The thermal method is an old separation technique for surface treatment of produced liquids. In the literature, Jean *et al.* [1999] reported separation of oil from oily sludge by freezing and thawing. Other researchers used microwave or laser-induced heating to separate the W/O emulsions [Lai and Fang, 1995; Delville *et al.*, 1993]. Although the thermal method greatly accelerates the separation, the high heating cost and low separation efficiency are always the major concerns. Also, the greenhouse gases generated during the thermal separations present a serious environmental concern.

In chemical method, separation is made possible by adding appropriate surfactants as demulsifiers into the O/W emulsions to induce phase separation

[Hartland, 1990]. The dominant separation mechanism is to significantly reduce the interfacial tension between the oil phase and the aqueous phase so that coalescence can readily occur. Nevertheless, this method is not cost-effective and suffers from chemical pollutions. In general, these two methods are not recommended for large-scale applications. In oil fields, as the auxiliary methods, they are often used to assist the gravity separation.

1.1.4 Mechanical coalescing method

The mechanical coalescer, which is a mechanical passive separation apparatus, has been known and studied for a long time. In the mechanical coalescing method [Sokolovic *et al.*, 1992; Akagi *et al.*, 1990; Rulev and Sedlukho, 1990], the sizes of the dispersed drops can be enlarged due to their deposition onto the coalescers such as fibrous beds and granular beds. The feed emulsions are brought into contact with solid materials, which interact with either one or both of the dispersed and continuous phases, strongly depending on the surface properties of the solid, such as wettability. The incoming droplets onto the solid surface will coalesce with other deposited droplets. The mechanical coalescing method has unique advantages. It is reliable, cost-effective and has much less environmental impact. In the past, this method is extensively applied in wastewater treatment and some filtration processes. However, there are still few studies with respect to understanding the coalescence mechanisms and determining the most suitable flow conditions for separation of the O/W emulsions. In this regard, the full potential of the mechanical coalescing method is yet to be developed.

1.1.5 Other separation methods

Other separation methods are available for particular applications. In the past, electrically enhanced phase separation method [Taylor, 1996, Gupta *et al.*, 1993] was often used to separate the W/O emulsions since the continuous oil phase is a good electrical insulator. The air flotation separation method [Zheng and Zhao, 1993] was frequently applied to separate the emulsions with rather small dispersed droplets. As a relatively novel separation technique, the membrane separation method [Fernandez *et al.*, 2001, Kong and Li, 1999] has recently been utilized to separate the dispersed phase from either the O/W emulsions or the W/O emulsions.

1.2 Mechanical Coalescers

Mechanical coalescers can be sub-categorized as quiescent settler, plate coalescers, porous beds, granular beds, and fibrous beds. Among these coalescers, the porous beds, granular beds and fibrous beds are most frequently used in industry.

1.2.1 Porous beds

Bradley [1985] discussed porous coalescers under the name of “fixed media”. Porous beds rely on the passage of emulsions through a porous medium with an appropriate structure. In general, laboratory studies and field tests are often conducted because there are few commonly established criteria for design of such coalescers. Usually, the medium material is selected in terms of its preferential affinity for one of the two phases in the emulsions, preferably the dispersed phase. Polyurethane foam is a typical packing material for the porous beds. When waxy crude oil or crude oil containing entrained solid fines are treated, expectedly, frequent cleaning or

regeneration of the medium becomes necessary. Some incompressible media have also been considered. The fretted discs (aluminum oxide and porous ceramic) and high-density polyethylene are used for coalescing Freon emulsified in water [Johnson, 1980]. Polyethylene is oleophilic and has excellent coalescence performance, while aluminum oxide and porous ceramic (both hydrophilic) have fairly good coalescence performance.

1.2.2 Granular beds

Among the coalescing media considered here, granular beds are the least consolidated. Such coalescers consist of granular particles. The packing materials are selected so as to have preferential affinity for either the dispersed phase or the continuous phase. A granular bed coalescer is designed solely by using either laboratory experimental data or field experience, in conjunction with numerical simulation of the coalescence performance.

Madia *et al.* [1976] examined the effects of wettability on the coalescence performance. By using a chromatographic technique to determine the wettability of the materials, they tested anthracite coal (30-40 mesh, oleophilic), Ottawa sand (30-40 mesh, hydrophilic), polypropylene (20-40 mesh, oleophilic), and XAD-2 (highly cross-linked styrene-divinylbenzene co-polymer; 20-50 mesh, oleophilic) in the coalescing dilute O/W emulsions. They concluded that the coalescence performance of oleophilic packing is superior to that of hydrophilic packing for the separation of the O/W emulsions.

Spielman and Su [1977] tested dilute emulsions of fairly viscous (50-500 cP) silicone liquids, which has a specific gravity close to that of water, through beds of

glass beads. These glass beads have the mean diameter of 0.26, 0.36, and 0.50 mm, respectively. The experimental data were also compared with the predictions of a numerical model. However, the coalescence efficiency was not discussed in this study.

Moses and NG [1985] built a two-dimensional glass flow cell to observe the O/W emulsion flow through a granular coalescer. It was inferred that a graded medium or a coalescer with mixed wettability might enhance the coalescence performance. Specifically, the wettability at the outlet of the coalescer should be nonwetting to the dispersed phase.

Since 1987, the Saskatchewan Research Council (SRC) in Regina has been endeavoring to find a suitable material that can effectively coalesce water from the produced petroleum liquids [Viraraghavan *et al.*, 1988; 1993; Meszaros, 1989; Hack, 1989]. Their results showed that granular media could remove up to 97.9% of water from heated W/O emulsions at 90 to 95 °C.

1.2.3 Fibrous beds

Fibrous beds are among the oldest and most widely used coalescers for coalescing emulsions. Such beds mainly include unconsolidated beds, knitted mats, cartridges and sock filters of woven materials. Fibrous media, because of their large specific surface area and porosity, are among the best mechanical treating materials in terms of droplet coalescing ability. Design of fibrous beds is based on the premises similar to those for the granular beds: experimental trial-and-error, possibly in combination with the existing numerical coalescence models. No general criteria are agreed upon.

Burtis and kirbride [1946] published one of the first known studies of a fibrous coalescing bed. They examined the efficiency of fiberglass beds (hydrophilic, fiber diameter: 7 μm , bed thickness: 0.32 m, bed density: 208 kg/m^3) in handling the W/O emulsions of brine and crude oil. Their throughputs ranged from 0.001 to 0.005 $\text{m}^3/\text{m}^2/\text{s}$. They found that a bed thickness of 0.01 m was adequate to accomplish the desired separation. Hazlett [1969] examined fiberglass (hydrophilic) and resin-bonded fiberglass (mixed wettability; fiber diameter: 0.4-11 μm) in separating water/JP-5 emulsions. It has been concluded that surfactants have no effects on coalescence behavior within the fibrous beds in this work.

Bitten and Fochtman [1971] studied the effects of liquid velocity, bed density, bed thickness, and emulsion composition on separation of JP-5 jet fuel from tap water. They also used fiberglass bed (hydrophilic, fiber diameter: 1-2 μm , bed thickness: 0.01-0.02 m, bed density: 96-225 kg/m^3). A layer of oleophilic material (knit cotton cloth) was placed at the downstream of the fiberglass bed to simulate the structure found in commercial coalescing units. The dispersed phase saturation dropped sharply from a maximum at the entrance of the bed to a much lower value at the exit. This indicates that the thick bed is not needed as long as bed thickness is beyond a certain minimum value.

Based on the impaction contact mechanism, Sherony and Kintner [1971a; 1971b] developed a model to determine the overall coalescence efficiency. Sherony *et al.* [1978] also presented an extensive review of the theory of fibrous beds, including a summary of the literature and patents. There had been 1,109 American patents alone up to 1967 directly regarding the application of specific or general coalescing media,

including polyoxide-treated amine-modified thermoplastic phenol-aldehyde resins, glass-wool, permeable resins, sands and others, for separating a variety of emulsions.

Abdel-ghani and Davies [1985] established a model for non-woven fiber materials used in coalescers. In this model, a Monte Carlo technique was used to compute droplet penetration into a fiber mat. This model can be employed to determine the capture efficiency, which is dependent on the fiber diameter, free surface area and droplet diameter.

1.3 Coalescence Process

In the coalescing bed, the coalescence may occur at the liquid-liquid interface and liquid-solid interface. The coalescence at the liquid-liquid interface is usually initiated by a collision between two droplets, which impacts sufficient forces to rupture the interfacial film. Once the droplets are in physical contact, the coalescence process is completed due to the surface forces involved [Adamson and Gast, 1997].

Coalescence at the liquid-solid interface is of the major interest in studying flow behavior of the O/W emulsions through porous media. Basically, the coalescence of two or more droplets to form a large single droplet consists of bringing the droplets into close contact so that hydrodynamic force and van der Waals attractive forces can promote the rupture of the thin water film. In the coalescing bed, porous media have the ability to attract many droplets, thereby bringing droplets sufficiently close to each other. After the water film is ruptured, the surface forces control adhesion of the droplet on the solid surface. In particular, a wetting material can provide a large adhesion force [Davies and Jeffreys, 1969].

The coalescence process in a fibrous bed can be divided into three major steps: approach of a droplet onto a fiber, attachment of a droplet onto a fiber or to a droplet already attached onto a fiber, and release of an enlarged droplet from a fiber surface [Hazlett, 1969]. These three steps are discussed in the following sections.

Approach

An oil droplet must be sufficiently close to the fiber for the subsequent coalescence to occur. Approach of a dispersed phase onto a fiber can occur due to direct interception, diffusion, inertia impaction, electrostatic attraction and repulsion, and sedimentation.

Attachment

Once the dispersed droplet has reached the fiber, it attaches onto the fiber for further coalescence. Resistance to the attachment is the continuous film between the droplet and the fiber or the attached droplet. Therefore, successful droplet attachment depends upon the thinning of the continuous film to the point of rupture.

Release

The larger size of the droplets released from the solid surface, the more successful the coalescing beds for separation of oil from the O/W emulsions. In general, the adhesion force between a droplet and the fiber in conjunction with wettability determines whether the droplet can remain on the fiber long enough to coalesce with other droplets. Larger adhesion and wetting forces allow the droplets to grow larger before they are released.

Coalescence can be achieved only if droplets can approach and adhere to a packing surface and reside there long enough for coalescence with an adjacent or

approaching droplet to take place. The retention of droplets on a packing surface occurs when the adhesion force is larger than the local shear force on the droplets. Hydrodynamic effects are important in controlling the rate of arrival of droplets on a surface. There exists a maximum velocity above which the retention time of droplets on the surface is too short for coalescence to occur. This maximum velocity is called the critical velocity, above which the emulsions droplets pass through the coalescing beds without apparent change of their sizes.

1.4 Coalescence Models

Several mathematical models have been developed to predict the coalescence efficiency, filter coefficient, oil saturation and pressure drop.

Hazlett Model

Hazlett [1969] presented a coalescence model for a fibrous coalescer. In this model, direct interception is assumed to be the dominant mechanism of approach. An equation developed by Langmuir [1942] is used to describe approach of the droplets:

$$E_s = \frac{1}{4 - \ln Re} \left[2(1 + R) \ln(1 + R) - (1 + R) + \frac{1}{1 + R} \right], \quad (1.1)$$

where E_s is the collection efficiency by a single isolated fiber from a fluid stream with the width equal to the fiber diameter; R is the ratio of the droplet diameter to the fiber diameter; and Re is the Reynolds number. Re is defined as $u\rho L/\mu$, where u is the flow velocity; ρ is the fluid density; L is characteristic length; and μ is the fluid viscosity. The extension of the collection efficiency for a fiber to that for a fibrous

bed can be made by considering the surface coverage. This equation shows that the efficiency increases with the Re , i.e., flow velocity.

Spielman and Goren Model

Based on air and water filtration theory and two-phase flow theory, a mathematical model was developed for coalescence of liquid-liquid dispersions through porous media [Spielman and Goren, 1971]. It has been assumed that all the captured oil droplets coalesce readily in the porous media and reach the capillary equilibrium with the aqueous phase. In this model, the filter coefficient, capillary pressure, and relative permeabilities of both liquid phases are treated as known functions of the oil saturation and other relevant parameters.

Spielman and Su Model

Experimental studies of oil-water coalescence by using spherical glass beads as a medium were conducted by Spielman and Su [1977]. Oil drop size distributions and pressure drop are measured as functions of position along the bed length by varying flow rate, bead diameter, and oil viscosity. Three distinguishable regimes of the coalescing phase are proposed to exist in the porous medium: oil droplets suspended in a capillary-conducted aqueous phase; capillary-conducted oil forming well connected channels; held-up oil as discrete coalescing globules. The globules facilitate the local transport of oil from the suspended droplets to the capillary-conducted oil regime.

Sherony and Kintner Model

Another coalescence model was presented to determine the overall coalescence efficiency [Sherony and Kintner, 1971a; 1971b]. In this model, it is assumed that the

impaction approach mechanism is dominant, i.e., droplets moving with the continuous phase collide with other droplets already attached onto the medium. The steady-state equation used for relating some parameters and the overall coalescence efficiency is as follows.

$$Y = \exp \left\{ - \frac{3}{4} \frac{S}{(1-S)} \frac{(1-\phi) \left(1 + \frac{d_i}{d_f} \right)}{d_f} \eta_c L \right\}, \quad (1.2)$$

where Y is the ratio of outlet to inlet drop number densities; ϕ is the porosity; S is the degree of saturation; d_i is the mean inlet droplet diameter; d_f is the characteristic size of the fiber; η_c is the overall coalescence efficiency; and L is the bed length.

1.5 Flow Behavior in Coalescing Beds

1.5.1 Single-phase flow

The pressure drop is one of important factors that should be considered in design of the coalescing beds. The flow of a single phase through a fibrous bed is usually described by the Darcy law:

$$\frac{\Delta P}{L} = - \frac{U \mu}{B_0}, \quad (1.3)$$

where B_0 is the permeability; U is the superficial velocity; μ is the fluid viscosity; and L is the length of the bed.

According to Carman-Kozeny theory [Carman, 1956], the permeability applied to a fibrous bed is equal to:

$$B_0 = \frac{d_f^2 \phi^3}{16k(1-\phi)^2}, \quad (1.4)$$

where k is the Kozeny constant. Substituting Equation (1.4) into Equation (1.3) yields the pressure drop, ΔP , across the bed of length L :

$$\Delta P = \frac{16LU\mu k(1-\phi)^2}{d_f^2 \phi^3}, \quad (1.5)$$

where ΔP is the pressure drop across the bed for single-phase flow. For single-phase flow, Equation (1.5) shows a linear relationship between the pressure drop and the velocity under the laminar flow condition.

In addition, a number of investigators have attempted to modify Carman-Kozeny equation to describe the flow behavior of different types of fibers. A pore model was developed for single-phase flow through a packing bed of fibers by Kyan *et al.* [1970], where a three-term polynomial extension of the Carman-Kozeny equation is provided as follows.

$$\frac{\Delta P}{L} = k_1 \frac{\mu U}{d_f^2 Re^4 (1-\phi)} + k_2 \left(\frac{\mu}{d_f U \rho} \right) \frac{\rho U^2}{d_f Re^6 (1-\phi)^2} + k_3 \left(\frac{\mu}{d_f U \rho} \right)^2 \frac{\rho^2 U^4}{d_f Re^{10} (1-\phi)^{5.5}}, \quad (1.6)$$

where k_1 , k_2 , and k_3 are coefficients and ρ is the density of fluid. This equation gives a better estimation of ΔP across beds with long fiber than that determined from one- or two-term equation.

1.5.2 Two-phase flow

As shown in Figure 1.1, the pressure drops of two-phase flow are always higher than that of the single-phase flow if compared at the same Reynolds number defined

for porous media. These discrepancies are due to the fact that the true porosity for two-phase flow, ϕ_t , is smaller than the original porosity ϕ for single-phase flow. The degree of saturation S is related to ϕ_t by

$$S = 1 - \frac{\phi_t}{\phi}. \quad (1.7)$$

The degree of saturation S represents the amount of the dispersed phase already attached onto the solid. If S is equal to zero, no dispersed phase retains on the solid surface in the coalescing beds.

Equation (1.7) used for determining the degree of saturation is based on an assumption that the dispersed phase, which is already attached, uniformly distributes on the solid surface along the coalescing bed. Owing to the lack of an appropriate relation between S and ϕ_t , this assumption is made for simplicity.

1.6 Research Objectives

The main objective of this thesis study is to examine the feasibility and potential of the mechanical coalescing method in separating the O/W emulsions. More specifically, the research purpose of this thesis is to identify the effects of some important parameters on coalescence performance. In addition, this study aims at achieving a better understanding of the coalescence of the O/W emulsions in fibrous beds and granular beds, respectively. This thesis research has the following specific objectives:

- (1) To apply a geometric method to calculate the Kozeny constant and predict the pressure drop across the coalescing bed;

- (2) To modify the Sherony and Kintner model for evaluating the overall coalescence efficiency;
- (3) To study the effects of flow rate, inlet oil concentration, porous medium characteristics such as bed length and fiber size on the coalescence performance;
and
- (4) To identify the dominant coalescence mechanisms occurring in the coalescing beds.

Figure 1.1 Friction factor f_k versus Reynolds number defined for porous media

$$N_{Re}/(1-\phi) \text{ [Sherony } et al., 1978], \text{ where } N_{Re} = \frac{d_f u \rho}{\mu}, f_k = \frac{k(1-\phi)}{N_{Re}}.$$

CHAPTER 2 EXPERIMENTAL

In this thesis research, a coalescing bed has been designed and constructed. Meanwhile, a complete flow system is established to conduct the dynamic tests. This experimental study examines the effects of flow rate, inlet oil concentration, bed length and different packing materials on coalescence performance. Both fibrous beds and granular beds are tested. In addition, measurements of several parameters are undertaken in the experimental study.

2.1 Apparatus

A block diagram of the experimental set-up for the dynamic tests is shown in Figure 2.1. The entire experimental setup consists of the following parts: homomixer, two feed tanks, a magnetic stirrer, two gear pumps, a coalescing bed and a differential pressure transducer as shown in Figure 2.2.

A homomixer (Model: 1L, Greerco, Chemineer, Inc., USA) is shown in Figure 2.3 and utilized to make a stable emulsion. The maximum operating speed of this homomixer can reach 7,600 rpm. In addition, this homomixer can be set at any speed in the range of 0-7,600 rpm with a stepless transformer (Model: 3PN1010B, Energy Products Co., USA). Two feed tanks are used to store the bulk water and the concentrated emulsion. The bulk water feed tank is a plastic container, which has a capacity of approximately 22.3 liters. The concentrated O/W emulsion is stored in a glass beaker and stirred on-line with a magnetic stirrer (Model: SP 46925, Cimarec Co., USA). Two gear pumps (Cole-Parmer Instrument Co., Canada), shown in Figure

2.4, are used in this study. One pump (EW-75211-10) is used to deliver the bulk water so that it will mix with the concentrated emulsion before the mixture enters the coalescing bed, which has a flow rate range of 13.1-939.6 ml/min. The other gear pump (EW-07002-32) is applied to transport the concentrated emulsion from the emulsion tank into the main flow pipe and its flow rate can be adjusted from 2.6 to 85 ml/min.

A stainless cylindrical coalescing bed (See Figure 2.5) used in this study is 70 cm long with the inner diameter (ID) of 73 mm. The bed is flanged with a 15.2×15.2 cm² square plate on each end. The inlet and outlet connections are made possible through these two end plates. The outer diameter (OD) of all the tubing used is 1/4 inch. Four ports are made available along the coalescing bed at 0, 20, 40, and 70 cm from the entrance for monitoring the pressure drop across the bed and/or taking the emulsion samples. Emulsion samples are collected through four ports, where a ball valve is installed. A differential pressure transducer (Model: P55D 1-N-1-58-W-4-A, Validyne Company, USA) is used to measure the pressure drop across the coalescing bed. The accuracy of the pressure transducer is 0.25% of its full scale, which is adequate for this experimental study.

2.2 Materials

2.2.1 Fibrous and granular materials

Three types of coalescing media are used in this study, which include polypropylene fiber and nylon fiber for fibrous beds and polypropylene particles for granular beds. The polypropylene fiber (See Figure 2.6) and nylon fiber are purchased

from National Filter Media Corporation, USA. The polypropylene particles (See Figure 2.7) are obtained from ExxonMobil Chemical Co., USA. The characteristic sizes of these packing materials are listed in Table 2.1. The height and the width of fibers are measured by using digital vernier caliper and their characteristic sizes (e.g., the hydraulic diameters) are determined. For the granular polypropylene particles, a sieve analysis is undertaken to determine the mean diameter by following procedures suggested in ASTM [1963]. The detailed procedures for the sieve analysis can be found in 2.4.2.

2.2.2 Oil and aqueous phases

Instow crude oil, mineral oil and tap water are used in the experiments. The physical properties of the oil and water samples are given in Table 2.2. The Instow crude oil is obtained from Instow oil field. The mineral oil is purchased from Fisher Scientific (Canada). In the experiment, an emulsifier, OP-70 (Huntsman Petrochemical Corporation, Canada) is also used to make a stable O/W emulsion. This emulsifier is a non-ionic surfactant with a full chemical name of Octylphenoxy polyethoxyethanol.

2.3 Experimental Procedure

The dynamic tests require determination of the porosity of the coalescing bed, preparation of the emulsion, and measurement of flow rate, pressure drop and droplet size. In the coalescing bed, the fiber is packed in a layered packing format, while the granular material is randomly packed. The porosity of the coalescing bed is measured with the vacuum method. After being evacuated, the coalescing bed is saturated with

tap water. Then the porosity can be determined by measuring the amount of water needed to fill the coalescing bed.

How to make a stable emulsion is essential to the dynamic tests. First, a concentrated emulsion is made by using a Greerco homomixer. In this step, the surfactant is dissolved and the oil is dispersed in the tap water with known concentrations. Then the homomixer is used to mix the oil and water at a preset speed of 2,400 rpm for five minutes. Such generated O/W emulsion is then further maintained in a magnetic stirrer until a stable emulsion is formed. It is found that, by trial and error, this method gives a mean droplet diameter in the range of 5-9 μm . Figure 2.8 shows an image of the O/W emulsion made by using this method. A stable O/W emulsion can be made in this way.

After all the tubing is connected and a stable emulsion is prepared, the dynamic test can be carried out according to the following steps. At the beginning, single-phase (water) flow test is conducted. The bulk water is pumped through the coalescing bed. The equilibrium pressure drops are recorded at different flow rates. The steady-state pressure drop is assumedly reached if voltage reading has the smallest fluctuations. This voltage reading is converted to the corresponding pressure drop by using the pre-determined calibration curve for this differential pressure transducer as shown in Figure 2.9. Then the two-phase flow test starts. In this step, the concentrated emulsion is delivered into the main flow pipe. For each run, the flow rate is preset for each pump and then the inlet oil concentration is known. The flow rate is determined by measuring the accumulative liquid volume. Mainly depending on the flow rate, the length of the bed and the inlet oil concentration, it takes about

10-20 minutes to reach the steady-state condition. This steady-state condition is achieved once the equilibrium pressure drop is established. Subsequently, the equilibrium pressure drop is recorded and emulsion sample is taken from each port under the two-phase flow conditions. During the sampling process, the sampling speed is too low to disturb the emulsion flow in the coalescing bed. The sample is then analyzed by using the drop size measurement system available in the SRC. The detailed procedure for determining the drop size distribution is described in 2.4.3.

2.4 Experimental Measurements

2.4.1 Pressure drop

In this study, the pressure drop across the coalescing bed is measured by using a differential pressure transducer. It should be noted that the displayed reading of this pressure transducer is in voltage. Then a calibration curve is needed to convert voltage to pressure in Pascal. As shown in Figure 2.9, three pressures and their corresponding voltages are used to plot this calibration curve. The linear regression is employed to obtain the correlation equation, which can be used for determining the pressure drop from a measured voltage.

2.4.2 Characteristic size of granular particles

For polypropylene particles, a sieve analysis is carried out. The detailed procedure for the sieve analysis is as follows. A total of 393.88 g of polypropylene particles is used to perform the sieve analysis. By using a shaker, these samples are graded in several groups with different sieves and the weight of each group is measured and given in Table 2.3. Accordingly, weight percentage of each group can be calculated

so that the accumulative mass distribution versus the particle diameter can be plotted. As shown in Figure 2.10, d_{50} represents the mean particle diameter at which the accumulative percentage is equal to 50%. In this study, d_{50} is chosen as the characteristic size of the granular particles and found to be 684 μm .

2.4.3 Droplet size

After the samples are taken from sampling ports, the emulsion droplet size is measured to evaluate the coalescence performance. In this experimental study, a drop size measurement system (See Figure 2.11) available in the SRC is employed to measure the drop size distribution of the O/W emulsion samples. With a microscope (Nikon, Japan) and a digital camera (Nikon, Japan), the emulsion samples can be observed and the droplet images can be taken. Two stages are involved in this measurement. First, a glass slide is placed under the microscope. Then an emulsion drop is introduced on the glass slide and a cover plate is placed on the top of the drop. In this process, much attention should be paid to minimize the possible effect of squeezing the emulsion drop on the oil droplet size determination. A photograph is taken after the image of the emulsion drop is well focused. For each sample, usually 5-6 drop images are acquired. Secondly, the drop images are transferred to a personal computer, where they are analyzed by using the SimplePCI software. An interface of the SimplePCI software for the drop size analysis is shown in Figure 2.12. The output of this software includes each droplet diameter, droplet volume, droplet surface area and overall droplet mean diameter.

Table 2.1 Characteristic sizes of the coalescing media.

Coalescing media	Fiber		Granular
	Polypropylene	Nylon	Polypropylene
Characteristic size (μm)	170	80	684

Table 2.2 Physical properties of oil and water samples (at 25 °C).

Properties	Density (g/cm^3)	Viscosity (cP)
Instow crude oil	0.905	45.6
Mineral oil	0.850	33.5
Tap water	0.998	1.0

Note: The density and viscosity data for the two oil samples are obtained from Saskatchewan Research Council (SRC) and Fisher Scientific (Canada), respectively.

Table 2.3 Sieve analysis of polypropylene particles.

Particle size (μm)	Weight (g)	Accumulative Percentage (%)
0-250	7.20	1.828
250-500	41.81	12.443
500-600	35.45	21.443
600-850	291.03	95.331
850-1000	17.12	99.678
>1000	1.27	100

Figure 2.1 Schematic diagram of the experimental setup used to study coalescence performance.

Figure 2.2 Experimental setup used in coalescing test.

Figure 2.3 Greerco homomixer (Model: 1L, Chemineer, Inc., USA) and stepless transformer (Model: 3PN1010B, Energy Products Co., USA).

Figure 2.4 Two gear pumps (Cole-Parmer Instrument Co., Canada).

Figure 2.5 A stainless cylindrical coalescing bed.

Figure 2.6 Polypropylene fiber used as packing material for the fibrous beds.

Figure 2.7 Polypropylene particles used as packing material for the granular beds.

Figure 2.1 Image of a prepared O/W emulsion.

Figure 2.9 Calibration curve of pressure versus output voltage for the differential pressure transducer.

Figure 2.10 Accumulative mass distribution versus the particle diameter for polypropylene particles.

Figure 2.11 Drop size measurement system at SRC.

Figure 2.12 Interface of the SimplePCI software for the drop size analysis.

CHAPTER 3 EMULSION FLOW IN COALESCING BEDS

Flow of the O/W emulsions in porous media is a very complex process due to the natures of the emulsions and porous media. The flow behavior may be rather different in various situations. In general, it can be classified as either a straining-controlled process or an interception-controlled process. As shown in Figure 3.1, if the emulsion droplets are equal to or larger than the size of the flow channels, the flow is straining-controlled. On the other hand, if the emulsion droplets are much smaller than the pore size, the interception mechanism becomes dominant. In this study, the size of the emulsion droplets is sufficiently smaller than the size of the flow channels. This interception-controlled process is accordingly studied and the Sherony and Kintner model is modified to determine the overall coalescence efficiency.

3.1 Coalescence in Fibrous and Granular Beds

In the coalescing beds, coalescence may occur among flowing droplets in the stream. These flowing droplets may also attach onto the solid surface. In addition, these flowing droplets may coalesce with those droplets already attached onto fibers or trapped at the pore throat of the randomly packed granular beds. A variety of interactions may lead to the coalescences among these droplets. These interactions can be described by using some basic theories, which have been developed for the coagulation of colloidal particles. Some major interaction mechanisms are summarized as follows.

3.1.1 Interception

In the interception mechanism, it is assumed that a droplet having finite size follows exactly the fluid streamline around the solid. When the droplet intercepts the fiber or particles, capture takes place. Figure 3.2 shows capture of a droplet by a spherical solid collector. In this figure, any droplets within the limiting streamline will intercept the solid collector.

3.1.2 Inertia impaction

When the dispersed droplets do not follow the streamlines due to their inertia, their capture by a collector is referred to as inertial deposition. For droplets of larger than 10 μm , inertial deposition becomes important. When there exists a large density difference between the dispersed phase and the continuous phase, inertial impaction can easily occur. However, large droplets and large density difference are not frequently encountered in the emulsion flow. Thus, the inertial impaction has no important contribution to the coalescence of the emulsions.

3.1.3 Gradient collision

The first theory of collision in a shear flow field was proposed by Smoluchowski [1916]:

$$K_G = \frac{4}{3} d_m^3 \Gamma, \quad (3.1)$$

where K_G is the gradient collision coefficient; Γ is the shear rate; and d_m is the mean droplet diameter. In the fibrous or granular beds, Γ can be as high as 6,000/sec. This mechanism causes the collisions among the droplets in the stream.

3.1.4 Brownian diffusion

A droplet moving with a fluid tends to deviate from the streamline partially due to the random transverse Brownian diffusion. This random Brownian motion enhances a droplet capture on a solid surface in the fibrous or granular beds. The diffusion coefficient of the emulsion droplets can be expressed as follows [Masliyah, 1994]:

$$D_{\infty} = \frac{kT}{6\pi\mu_c d}, \quad (3.2)$$

where D_{∞} is Brownian diffusion coefficient; k is Boltzmann constant; T is absolute temperature; μ_c is viscosity of the continuous phase; and d is the droplet diameter.

The significance of diffusion mechanism was examined in a fibrous coalescer to remove water from jet fuel [Hazlett, 1969]. It was concluded that diffusion at a flow rate of 0.9 cm/s is important only when both the droplets and the fibers are less than 1 μm in diameter.

3.2 Coalescence Models for Fibrous and Granular Beds

In the fibrous or granular beds, the coalescence process changes the drop size distribution. Although a coalescing bed retains some dispersed phase at the beginning, it differs from a filter because a coalescing bed is a steady-state device where the mass concentration of the dispersed phase at the inlet remains the same as that at the outlet. Then a question that naturally arises is how to evaluate the coalescence performance of a coalescing bed. In this study, the Sherony and Kintner model [Sherony and Kintner, 1971a] is modified to determine the overall coalescence efficiency of the fibrous and granular beds, respectively.

In the Sherony and Kintner model, there are two important assumptions. First, the droplets held by the fiber or granular particles are referred to as “holdup” and their size distribution is represented by $g(m, x, t)$, where m is mass; x is the distance from the entrance; and t is time. This holdup is considered to have a steady-state distribution, i.e., the bed is operated in a steady-state fashion. In the second assumption, the emulsion is considered not to form a continuum in the fibrous beds or granular beds, i.e., the surface coverage is rather low. A qualitative description of this model is shown in Figure 3.3.

3.2.1 Conservation law of mass

The conservation of the dispersed phase for the emulsion flow system can be expressed as below [Sherony and Kintner, 1971a]:

$$\frac{\partial f}{\partial t} + \frac{\partial(uf)}{\partial x} = h(m, x, t), \quad (3.3)$$

where f represents the drop size distribution in the stream; u is interstitial speed of the fluid; and $h(m, x, t)$ accounts for the rate of the coalescence between a droplet in the stream and a droplet on the solid surface and between two droplets in the stream. These two kinds of coalescence are represented by $h_1(m, x, t)$ and $h_2(m, x, t)$, respectively.

3.2.2 Two-droplet collisions

Two-droplet collision (a droplet in the stream and a droplet on the solid surface)

A collision of a droplet of mass m_1 from f (drop size distribution in the stream) with a droplet of mass m_2 from g (drop size distribution in the holdup) may result in

a new droplet of mass $m = m_1 + m_2$. It is assumed that the resultant droplet of mass m stays in the stream and becomes a member of f . It is further assumed that the droplet of mass m_2 from g on the solid surface is replaced by a similar droplet of mass m_2 from f in the stream. In this way, the total mass in the stream f and in the holdup g remains constant, respectively. In this Sherony and Kintner model, the number of collisions of the droplets with mass m_1 from f and the droplets with mass m_2 from g resulting the coalescence per unit volume and per unit time is proportional to the product of their number densities such that [Sherony and Kintner, 1971a]:

$$h_1(m, x, t) \propto f(m_1, x, t)g(m_2, x, t). \quad (3.4)$$

If two droplets collide, one droplet of mass m_1 from f and the other droplet of m_2 from g , and form a droplet of mass $m = m_1 + m_2$, the rate of formation of droplet of mass m can be written as:

$$k_1 \left\{ \sum_{m_2=0}^m f(m_1, x, t)g(m_2, x, t) \right\}, \quad (3.5)$$

where k_1 is the accommodation coefficient associated with $h_1(m, x, t)$. As $m = m_1 + m_2$, this equation becomes:

$$k_1 \left\{ \sum_{m_2=0}^m f(m - m_2, x, t)g(m_2, x, t) \right\}. \quad (3.6)$$

Equation (3.6) can be expressed in an integral format:

$$k_1 \left\{ \int_0^m g(m', x, t)f(m - m', x, t)dm' \right\}. \quad (3.7)$$

Since there exist droplets of mass m in f and g , respectively, these droplets of mass m also undergo collisions with the droplets in f and g whose mass are not equal

to m . Each of collision causes a droplet of mass m to disappear. Therefore, the rate of loss of droplet of mass m is equal to:

$$k_1 \left\{ g(m, x, t) \int_0^\infty f(m', x, t) dm' + f(m, x, t) \int_0^\infty g(m', x, t) dm' \right\}. \quad (3.8)$$

Based on Equation (3.7) and Equation (3.8), the net rate of formation of droplet of mass m can be expressed as:

$$h_1(m, x, t) = k_1 \left\{ \int_0^\infty g(m', x, t) f(m - m', x, t) dm' \right\} - k_1 \left\{ g(m, x, t) \int_0^\infty f(m', x, t) dm' + f(m, x, t) \int_0^\infty g(m', x, t) dm' \right\}, \quad (3.9)$$

The function $h_1(m, x, t)$ expresses the rate of change of a droplet of mass m in the stream f due to the interactions between droplets in f and those in g . The accommodation coefficient k_1 represents a sum of two mechanisms of collision, one due to interception between droplets in f and those in g and the other due to Brownian diffusion.

Two-droplet collision in the stream

The same method can be applied when two droplets in the stream collide each other. Under this circumstance, the rate of change of a droplet of mass m in the stream can be expressed as follows [Sherony and Kintner, 1971a]:

$$h_2(m, x, t) = k_2 \left\{ \int_0^\infty \frac{1}{2} f(m', x, t) f(m - m', x, t) dm' - f(m, x, t) \int_0^\infty f(m', x, t) dm' \right\}, \quad (3.10)$$

where k_2 is the accommodation coefficient associated with $h_2(m, x, t)$.

The following moments can be defined:

$$\int_0^{\infty} m^n f(m, x, t) dm = \alpha_n, \quad (3.11a)$$

$$\int_0^{\infty} m^n g(m, x, t) dm = \beta_n, \quad (3.11b)$$

When n is equal to zero, the 0th-order moments α_0 and β_0 represent the number densities of droplets in the stream f and on the solid surface g , respectively.

Substituting Equations (3.9) and (3.10) into Equation (3.3), multiplying it by m^n and integrating the resultant equation over m from 0 to ∞ , one can obtain the following general equation,

$$\begin{aligned} \frac{\partial \alpha_n}{\partial t} + \frac{\partial(u\alpha_n)}{\partial x} = & k_1 \left\{ \sum_{k=0}^n \frac{n!}{k!(n-k)!} \alpha_{n-k} \beta_k - \alpha_0 \beta_n - \beta_0 \alpha_n \right\} \\ & + k_2 \left\{ \frac{1}{2} \sum_{k=0}^n \frac{n!}{k!(n-k)!} \alpha_k \alpha_{n-k} - \alpha_0 \alpha_n \right\} \end{aligned} \quad (3.12)$$

In particular, when n is equal to zero, Equation (3.12) becomes,

$$\frac{\partial \alpha_0}{\partial t} + \frac{\partial(u\alpha_0)}{\partial x} = -k_1 \alpha_0 \beta_0 - k_2 \frac{\alpha_0^2}{2}, \quad (3.13)$$

where $k_1 \alpha_0 \beta_0$ represents the number of collisions between droplets in the stream f and on the solid surface g resulting in coalescence per unit volume and per unit time; and $k_2 \frac{\alpha_0^2}{2}$ represents the number of collisions between droplets in the stream f resulting in coalescence per unit volume and per unit time.

3.2.3 Number of collisions between droplets in the stream and those on the solid surface

In Equation (3.13), $k_1\alpha_0\beta_0$ represents the number of collisions between droplets in the stream f and on the solid surface g resulting in coalescence per unit volume and per unit time. Therefore,

$$k_1\alpha_0\beta_0 = \frac{\text{total number of collisions}}{\text{volume} \cdot \text{time}} \times \eta_1\eta_2, \quad (3.14)$$

where η_1 is the contact efficiency and η_2 is the coalescence efficiency. The product of contact efficiency and coalescence efficiency is the overall coalescence efficiency η_c . The total number of collisions among droplets in Equation (3.14) can be further divided into:

$$\frac{\text{total number of collisions}}{\text{volume} \cdot \text{time}} = \frac{\text{number of collisions between droplets \& solid surface}}{\text{volume} \cdot \text{time}} \cdot \frac{\text{solid surface area covered by holdup}}{\text{the total solid surface area}}. \quad (3.15)$$

The second term in Equation (3.15) on the right hand side represents the so-called surface coverage. The first term in Equation (3.15) on the right hand side can be obtained according to the theory of Calcus for the mean free path of gas [Loeb, 1961]. In this theory, the number of collisions of droplets with a solid surface per unit time for per droplet can be expressed as:

$$n_1 = \frac{UA_f}{\pi\phi V} \text{ (Fibrous beds),} \quad (3.16a)$$

$$n_1 = \frac{UA_g}{\pi\phi V} \text{ (Granular beds),} \quad (3.16b)$$

where U is the superficial velocity; A_f and A_g are the total surface area of fiber and granular particles, respectively; ϕ is porosity; and V is the total free volume available for the emulsion flow. Hereafter, all the equations for fibrous and granular beds are represented by Equation (a) and Equation (b), respectively.

The volume of fiber or granular particles is $(1-\phi)V_B$, where V_B is the bulk volume of the fibrous or granular beds. The volume of holdup can be expressed as $V_H = S\phi V_B$, where S is the degree of saturation; and ϕV_B is often referred to as the pore volume. Therefore the total free volume is:

$$V = \phi V_B (1 - S). \quad (3.17)$$

The respective total surface area of fibrous and granular beds available for collision is equal to:

$$A_f = \frac{4V_B(1-\phi)}{d_f}, \quad (3.18a)$$

$$A_g = \frac{6V_B(1-\phi)}{d_g}, \quad (3.18b)$$

where d_f and d_g are the characteristic sizes of the fiber and the granular particle, respectively.

The hydraulic radius for a fibrous or granular bed can be defined as:

$$m_H = \frac{2V}{A_f} = \frac{\phi(1-S)d_f}{2(1-\phi)}. \quad (3.19a)$$

$$m_H = \frac{2V}{A_g} = \frac{\phi(1-S)d_g}{3(1-\phi)}. \quad (3.19b)$$

From Equations (3.16a) and (3.16b), the number of collisions in V volume per unit time can be obtained as:

$$N_0 = m_0 n_1 = \frac{m_0 U A_f}{\pi \phi V}, \quad (3.20a)$$

$$N_0 = m_0 n_1 = \frac{m_0 U A_g}{\pi \phi V}, \quad (3.20b)$$

where m_0 is the total number of droplets in the volume V , m_0/V is the number of droplets per volume or number density, which is also referred to as α_0 , N_0/A_f or N_0/A_g is the number of collisions per unit surface area and per unit time for fibrous or granular bed, which is expressed by n_0 ,

$$n_0 = \frac{N_0}{A_f} = \frac{\alpha_0 U}{\pi \phi}. \quad (3.21a)$$

$$n_0 = \frac{N_0}{A_g} = \frac{\alpha_0 U}{\pi \phi}. \quad (3.21b)$$

Thus, the number of collisions between droplets and the fiber or granular particles per unit volume and per unit time is:

$$n_2 = \frac{n_0 A_f}{V} = \frac{4(1-\phi)\alpha_0 U}{\pi d_f \phi^2 (1-S)}. \quad (3.22a)$$

$$n_2 = \frac{n_0 A_g}{V} = \frac{6(1-\phi)\alpha_0 U}{\pi d_g \phi^2 (1-S)}. \quad (3.22b)$$

The above analysis for the number of collisions between droplets and the fiber or granular particles per unit volume and per unit time is based on an assumption that the size of the colliding droplets is much smaller than the size of the solid interceptor. In the fibrous beds, sometimes the size of drops in the stream is equal to or even

larger than that of the fiber. In the granular beds, the same situation may occur. In this case, the effect of large droplets on n_2 should be considered and thus a factor of $(1+d_i/d_f)$ or $(1+d_i/d_g)$ needs to be incorporated into Equations (3.22a) and (3.22b), respectively.

$$n'_2 = \frac{4(1-\phi)\alpha_0 U}{\pi d_f \phi^2 (1-S)} \left(1 + \frac{d_i}{d_f} \right), \quad (3.23a)$$

$$n'_2 = \frac{6(1-\phi)\alpha_0 U}{\pi d_g \phi^2 (1-S)} \left(1 + \frac{d_i}{d_g} \right), \quad (3.23b)$$

where d_i is the mean diameter of the droplets at the inlet.

Calculation of the surface coverage

The solid surface area A_D covered by the deposited droplets per unit volume is equal to:

$$\frac{A_D}{V} = \frac{\pi \beta_0 d_{2,0}^2}{4}, \quad (3.24)$$

where $d_{2,0}$ is the area-mean droplet diameter. In general, $d_{i,j}$ is defined as follows.

$$d_{i,j} = \left[\frac{\int_0^\infty D^i f(D) \cdot d(D)}{\int_0^\infty D^j f(D) \cdot d(D)} \right]^{\frac{1}{(i-j)}} \quad (i > j). \quad (3.25)$$

In Equation (3.11b), when n is equal to one, β_1 becomes to the mass concentration. According to this definition,

$$\beta_1 = \int_0^\infty m g(m) dm = \bar{m} \beta_0, \quad (3.26)$$

$\bar{m}$ can be calculated as $\bar{m} = \rho_0 \frac{\pi}{6} d_{3,0}^3$, where $d_{3,0}$ is the volume-mean droplet

diameter. In the fibrous or granular bed, β_1 can also be determined in another way,

$$\beta_1 = \frac{V_B \phi S \rho_0}{V_B} = \phi S \rho_0, \quad (3.27)$$

Thus,

$$\beta_0 = \frac{\phi S}{\frac{\pi}{6} d_{3,0}^3}. \quad (3.28)$$

Combining Equations (3.17), (3.18a), (3.18b) (3.24), and (3.28) gives:

$$\frac{A_D}{A_f} = \frac{3\phi^2 S d_f (1-S)}{8(1-\phi) d_{3,2}}. \quad (3.29a)$$

$$\frac{A_D}{A_g} = \frac{\phi^2 S d_g (1-S)}{4(1-\phi) d_{3,2}}. \quad (3.29b)$$

where $d_{3,2}$ is the Sauter mean diameter. Rigg and Churchill [1964] showed that, in the packing beds, the Sauter mean diameter $d_{3,2}$ for the deposited droplets is dependent upon the geometry of the bed.

$$d_{3,2} = m_H = \frac{\phi(1-S)d_f}{2(1-\phi)}. \quad (3.30a)$$

$$d_{3,2} = m_H = \frac{\phi(1-S)d_g}{3(1-\phi)}. \quad (3.30b)$$

Therefore, Equations (3.29a) and (3.29b) can be simplified using the above two equations.

$$\frac{A_D}{A_f} = \frac{3\phi S}{4}. \quad (3.31a)$$

$$\frac{A_D}{A_g} = \frac{3\phi S}{4}. \quad (3.31b)$$

3.2.4 Number of collisions between droplets in the stream

In this case, Equation (3.1) is used to determine the number of collisions between droplets in the stream. In the fibrous beds, Γ in Equation (3.1) becomes [Sherony *et al.*, 1978]:

$$\Gamma = \frac{4U(1-\phi)}{d_f \phi^2}. \quad (3.32a)$$

$$\Gamma = \frac{6U(1-\phi)}{d_g \phi^2}. \quad (3.32b)$$

Based on the above definitions, the gradient collision coefficient K_G in Equation (3.1) can be expressed as:

$$K_G = \frac{16Ud_m^3(1-\phi)}{3d_f \phi^2}. \quad (3.33a)$$

$$K_G = \frac{8Ud_m^3(1-\phi)}{d_g \phi^2}. \quad (3.33b)$$

Physically, for a given emulsion, the gradient collision coefficient K_G represents the ability for collisions to occur, while the accommodation coefficient k_2 associated with $h_2(m, x, t)$ represents the ability for coalescence to occur. Thus, the accommodation coefficient k_2 is equal to the product of K_G and overall coalescence efficiency, i.e., $k_2 = K_G \eta_1 \eta_2$.

3.2.5 Overall coalescence efficiency

In the preceding sections, the number of collisions has been determined for the two cases, i.e., collisions between droplets in the stream and those on the solid surface and collisions between droplets in the stream. In this section, a general equation is obtained to describe the overall coalescence efficiency.

Combining Equations (3.14), (3.15), (3.23a) or (3.23b), and (3.31a) or (3.31b) gives,

$$k_1 \alpha_0 \beta_0 = \frac{3U(1-\phi)S}{\pi d_f \phi (1-S)} \left(1 + \frac{d_i}{d_f} \right) \eta_c \alpha_0. \quad (3.34a)$$

$$k_1 \alpha_0 \beta_0 = \frac{9U(1-\phi)S}{2\pi d_g \phi (1-S)} \left(1 + \frac{d_i}{d_g} \right) \eta_c \alpha_0. \quad (3.34b)$$

Similarly, $\frac{k_2 \alpha_0^2}{2}$ can be expressed as:

$$\frac{k_2 \alpha_0^2}{2} = \frac{8Ud_m^3(1-\phi)}{3d_f \phi^2} \eta_c \alpha_0^2. \quad (3.35a)$$

$$\frac{k_2 \alpha_0^2}{2} = \frac{4Ud_m^3(1-\phi)}{d_g \phi^2} \eta_c \alpha_0^2. \quad (3.35b)$$

In Equation (3.13), the interstitial velocity u can be approximated as U/ϕ . Substituting Equations (3.34a) or (3.34b), and (3.35a) or (3.35b) into (3.13) and considering only the steady-state case yields:

$$\frac{d\alpha_0}{dx} = -\frac{3(1-\phi)S}{\pi d_f (1-S)} \left(1 + \frac{d_i}{d_f} \right) \eta_c \alpha_0 - \frac{8d_m^3(1-\phi)}{3d_f \phi} \eta_c \alpha_0^2. \quad (3.36a)$$

$$\frac{d\alpha_0}{dx} = -\frac{9(1-\phi)S}{2\pi d_g (1-S)} \left(1 + \frac{d_i}{d_g} \right) \eta_c \alpha_0 - \frac{4d_m^3(1-\phi)}{d_g \phi} \eta_c \alpha_0^2. \quad (3.36b)$$

In the above equations, physically, $\frac{1}{6}\pi d_m^3 \alpha_0$ is the volume concentration of the dispersed phase and can be expressed by C_o .

If S and η_c are considered as constant inside the coalescing beds, then Equations (3.36a) and (3.36b) can be integrated from $x=0$ to $x=L$ to obtain:

$$Y = \frac{\alpha_0(L)}{\alpha_0(0)} = \exp\left(-\left[\frac{3(1-\phi)S}{\pi d_f(1-S)}\left(1+\frac{d_i}{d_f}\right) + \frac{16C_o(1-\phi)}{\pi d_f\phi}\right]\eta_c L\right), \quad (3.37a)$$

$$Y = \frac{\alpha_0(L)}{\alpha_0(0)} = \exp\left(-\left[\frac{9(1-\phi)S}{2\pi d_g(1-S)}\left(1+\frac{d_i}{d_g}\right) + \frac{24C_o(1-\phi)}{\pi d_g\phi}\right]\eta_c L\right), \quad (3.37b)$$

where Y is the ratio of the outlet to the inlet drop number densities and L is the length of the coalescing bed.

η_c is accordingly obtained as follows.

$$\eta_c = -\frac{\ln Y}{\left[\frac{3(1-\phi)S}{\pi d_f(1-S)}\left(1+\frac{d_i}{d_f}\right) + \frac{16C_o(1-\phi)}{\pi d_f\phi}\right]L}. \quad (3.38a)$$

$$\eta_c = -\frac{\ln Y}{\left[\frac{9(1-\phi)S}{2\pi d_g(1-S)}\left(1+\frac{d_i}{d_g}\right) + \frac{24C_o(1-\phi)}{\pi d_g\phi}\right]L}. \quad (3.38b)$$

In Equations (3.38a) and (3.38b), Y and the degree of saturation S can be measured experimentally. Then the overall coalescence efficiency η_c can be determined. In the above equations, the first term in the denominator on the right hand side represents the contribution of collisions between the oil droplets in the stream and those deposited on the solid surface. The second term denotes the contribution of collisions among the oil droplets in the stream.

3.3 Summary

In this chapter, coalescence of the O/W emulsions in coalescing beds is analyzed. First, several major interaction mechanisms causing the coalescence in the coalescing beds are described. These interactions include interception, inertial impaction, gradient collision, and Brownian diffusion. Secondly, the Sherony and Kintner model is modified to evaluate the coalescence performance. Both the collision between two droplets in the stream and the collisions between a droplet in the stream and a droplet on the solid surface may result in coalescence in the coalescing bed. In the modified model, these two kinds of collisions are considered. Finally, an equation is derived to determine the overall coalescence efficiency for a fibrous or granular coalescing bed, respectively. In this equation, the degree of saturation, drop number densities at the inlet and outlet, and porosity can be measured experimentally. Therefore, the overall coalescence efficiency can be calculated.

Figure 3.1 Schematic of the O/W emulsion flow through porous media.

Figure 3.2 Capture of droplets due to interception by a spherical solid collector.

Figure 3.3 The Sherony and Kintner coalescence model.

CHAPTER 4 RESULTS AND DISCUSSION

4.1 Pressure Drop across Coalescing Beds

4.1.1 Carman-Kozeny equation

Many models have been developed to describe single- or two-phase flow through porous media. These models are based on two fundamentally different cases: fluid flow inside conduits and fluid flow around solid objects. For fluid flow inside a porous medium with low and intermediate porosities, the conduit flow case is more appropriate, whereas the second case is more applicable inside a porous medium with extremely high porosities. In this thesis study, the porosity is in the intermediate range of 0.34 to 0.57. Therefore, the conduit flow case is chosen to describe the flow in the fibrous and granular beds.

It is well known that Carman-Kozeny theory is often used to describe the flow through porous media [Carman, 1956]. In Carman-Kozeny theory, the porous medium is assumed to be equivalent to a conduit, whose cross-section has an extremely complicated shape but on the average, a constant surface area. The conduit diameter is defined as:

$$D_H = \frac{4 \times \text{void volume of a porous medium}}{\text{surface area of channels in a porous medium}} = \frac{4V_B\phi}{A} = \frac{4\phi}{A_0(1-\phi)}, \quad (4.1)$$

where A_0 is the specific surface area based on the solid volume of porous medium;

V_B is the bulk volume; and A is the total surface area of the porous medium.

$$A_0 = \frac{\text{total surface area}}{\text{solid volume}} = \frac{A}{V_B(1-\phi)}. \quad (4.2)$$

The Hagen-Poiseuille equation is used to determine the average interstitial or seepage velocity u in the flow channels [Dullien, 1992]:

$$u = -\frac{\Delta P}{L_e} \frac{D_H^2}{16k_0\mu}, \quad (4.3)$$

where k_0 is the shape factor; L_e is the effective length; and μ is the fluid viscosity.

The superficial velocity U used in Darcy law is defined as:

$$U = -\frac{k_{ck}}{\mu} \frac{\Delta P}{L}, \quad (4.4)$$

where k_{ck} is the permeability and L is the length of the porous medium.

In the porous medium, “Dupuit-Forchheimer assumption” is often used to define an average interstitial velocity. According to this assumption,

$$u = (U/\phi)(L_e/L). \quad (4.5)$$

Multiplication of U/ϕ by L_e/L is due to the fact that a fluid, flowing with velocity u , actually covers an effective length L_e rather than L . Combining Equations (4.3), (4.4) and (4.5) gives:

$$k_{ck} = \frac{\phi D_H^2}{16k_0(L_e/L)^2}. \quad (4.6)$$

This is the general form for calculating the permeability in all the conduit flow models, differing only in the method of determining the conduit diameter and in the value used for $k_0(L_e/L)^2$.

Substituting Equation (4.1) into (4.6) gives:

$$k_{ck} = \frac{\phi^3}{k_0(L_e/L)^2(1-\phi)^2 A_0^2}, \quad (4.7)$$

where L_e/L is usually called tortuosity T and the combined factor $k = k_0(L_e/L)^2 = k_0T^2$ is the so-called Kozeny constant.

Thus, the general Carman-Kozeny equation is obtained to predict the pressure drop across the coalescing beds if Equation (4.7) is substituted in Equation (4.4) and rearranged:

$$\Delta P = \frac{UL\mu k(1-\phi)^2 A_0^2}{\phi^3}. \quad (4.8)$$

4.1.2 Prediction of pressure drop in a fibrous bed

Kozeny constant

In order to calculate the Kozeny constant, the shape factor k_0 and the tortuosity T need to be determined. In this study, these two quantities are determined based on the geometric consideration.

In the dynamic test, the fiber is packed in the coalescing bed in a layered packing format. As shown in Figure 4.1, the cross-section of the fiber is rectangular. In this study, for the polypropylene fiber, the height a and the width b are 0.47 mm and 0.19 mm, respectively.

The shape factor for a single particle is defined as the ratio of the surface area of the particle to the surface area of a sphere with the same volume [Scheidegger, 1960]. According to this definition, the shape factor k_0 for the polypropylene fiber is equal to:

$$k_0 = \frac{2(ab + al + bl)}{4\pi\left(\frac{3abl}{4\pi}\right)^{2/3}}. \quad (4.9)$$

Substituting the values for a and b into this equation gives:

$$k_0 = \frac{0.1666 + 1.32l}{0.9224(l)^{2/3}}$$

It should be noted that shape factor varies with the length. In this study, the fiber length is ranged from 3 to 73 mm. Applying the geometric mean to the fibers with different lengths gives:

$$k_0 = 4.8826$$

The tortuosity T is the ratio of actual length of flow path to straight length of flow path, depending on the pore shape of the medium.

The geometry of the polypropylene fiber and the flow path are shown in Figure 4.2. In this study, the distance c between two fibers is found to be 0.26 mm, which is used in the following calculations.

Two assumptions are made regarding the flow path in order to calculate the tortuosity. As can be seen in Figure 4.2, one can assume that the flow follows either a triangle path or a semi-circle path. In these two cases,

$$T_1 = \frac{L_e}{L} = \frac{0.2685 \times 2 + 0.19}{0.26 + 0.19} = 1.616 \quad (\text{Triangle path})$$

$$T_2 = \frac{L_e}{L} = \frac{\pi}{4} \times \frac{0.26 + 0.19 + 0.47}{0.26 + 0.19} = 1.604 \quad (\text{Semi-circle path})$$

Once the shape factor and tortuosity are determined, the Kozeny constant can be calculated,

$$k_1 = k_0 T_1^2 = 12.75 \quad (\text{Triangle path})$$

$$k_2 = k_0 T_2^2 = 12.56 \quad (\text{Semi-circle path})$$

Here, triangle path is used to determine the Kozeny constant for simplicity.

Dynamic experiments for single-phase flow

In the dynamic experiments, first the single-phase flow (water flow) test is performed. As shown in Figure 4.3, the pressures drop versus flow rate is plotted according to Equation (4.8). In this equation, the porosity is measured prior to the dynamic experiment. The specific surface area A_0 for the fiber can be calculated by using Equation (4.2).

By applying the linear regression to the experimental data, the Kozeny constant is determined to be $k = 12.81$ with a correlation coefficient (R^2) of 0.9993. Physically, the Kozeny constant is largely affected by the specific surface area and porosity. The specific surface area is a function of the fiber shape and size. Therefore, the Kozeny constant varies with the fiber shape and size. The Kozeny constant is determined to be 8-10 in the fibrous bed, where the fiber size is around $10\ \mu\text{m}$ [Sherony and Kintner, 1971b]. However, in this study, the fiber has a characteristic size of $170\ \mu\text{m}$. This is why the Kozeny constant obtained in this experiment is relatively larger than the Sherony and Kintner's results.

As previously discussed, the Kozeny constant is determined to be $k = 12.75$ with the assumption of the triangle flow path. A comparison between the predicted pressure drop using the calculated Kozeny constant and the experimental results is plotted in Figure 4.4. There exists an excellent agreement between the experimental and predicted results in this figure.

4.1.3 Prediction of pressure drop in a granular bed

Similarly, the pressure drop across the granular beds can be predicted by using the Carman-Kozeny equation. However, it is difficult to determine the shape factor k_0

for the granular bed geometrically. In terms of its definition, the shape factor varies with the particle shape. Meanwhile, the particles used in this study consist of different sizes. In this regard, an experimental method needs to be adopted to determine k_0 for the granular bed.

For a porous medium with a non-straight path of pores, the tortuosity can be estimated as $1/\phi$ [Johnston, 1983], where ϕ is the porosity of the medium. Once the porosity is known, the tortuosity can be determined. Since the Kozeny constant can be obtained experimentally, then the shape factor can be found and further used to predict the pressure drop in the granular beds with different porosities.

As mentioned in Chapter 2, the mean particle diameter for the granular bed used in this study is 0.684 mm. Figure 4.5 shows the pressure drop versus the flow rate for the single-phase flow in the granular bed with porosity $\phi=0.367$. The tortuosity is accordingly estimated as $T=1/\phi=2.725$. By applying the linear regression, the Kozeny constant is found to be $k=10.83$. Therefore, the shape factor can be obtained as follows.

$$k_0 = \frac{k}{T^2} = 1.459$$

In the literature, the shape factor is reported in the range of 1.5-3.0, depending on the geometric shape of the porous medium [Sakthivadivel *et al.*, 1972]. Thus, the shape factor of the granular particles obtained in this study is reasonable.

In addition, this shape factor is utilized to predict the pressure drop across the granular beds with porosity $\phi=0.384$. The results of the measured and the predicted pressure drops across the granular beds are given in Figure 4.6. It can be seen from

this figure that a good match is achieved between the measured and the predicted pressure drops. Furthermore, by applying the linear regression, the Kozeny constant in this case is found to be $k=9.668$. Accordingly, the shape factor can be calculated,

$$k_0 = \frac{k}{T^2} = 1.426.$$

It should be noted that the shape factors obtained for the above two tests are rather close. This fact indicates that the proposed experimental method to determine the shape factor is applicable.

4.1.4 Dynamic tests of two-phase flow

Friction factor

Reynolds number is usually defined as follows [Dullien, 1992].

$$Re_p = D_p U \rho / \mu, \quad (4.10)$$

where D_p is the characteristic particle or fiber diameter; and U is the superficial velocity.

Meanwhile, the resistance of porous media to flow can be expressed by using the friction factor f_p , which is defined as:

$$f_p = D_p \Delta P / (\rho U^2 L). \quad (4.11)$$

It is well known that the transition from the linear relationship between Re_p and f_p to the nonlinear one is gradual. Usually, a critical Reynolds number is used to separate these two cases. For most of the data analyzed by Macdonald *et al.* [1979], the deviation from linearity becomes noticeable starting at $Re_p/(1-\phi) \approx 10$. In this

study, $Re_p/(1-\phi)$ is in the range of 0.3-3.0 and the experimental results show that there exists a good linearity between Re_p and f_p .

Figures 4.7, 4.8 and 4.9 show the plots of f_p versus $Re_p/(1-\phi)$ for single- and two-phase flows at 2,000, 5,000, and 10,000 ppm inlet oil concentrations, respectively. It is found that the friction factors for the two-phase flow at three different inlet oil concentrations are only slightly increased in comparison with those for the single-phase flow. No obvious permeability reduction occurs in the two-phase flow. In fact, two mechanisms may lead to the permeability reduction when the O/W emulsions flow through the coalescing beds [Soo, 1983]. One is interception, which causes oil droplets to attach onto the solid and thus reduce the porosity. The porosity reduction eventually increases the pressure drop across the coalescing beds in the two-phase flow case. In general, permeability reduction caused by the retained droplets is not significant. The other mechanism is straining capture due to some large oil droplets entrapped in some small pores. Physically, the second mechanism usually results in a larger pressure drop. In this case, the pores are either completely blocked or greatly restricted by the strained droplets. In this study, the experimental results for the two-phase flow show that the pressure drop is slightly increased in the test. Therefore, it can be concluded that the O/W emulsion flow in the coalescing beds is mainly interception controlled.

True porosity

As mentioned above, some oil droplets attach onto the solid and cause the porosity reduction. Then a question that naturally arises is to determine the true

porosity for two-phase flow. In the literature, the true porosity can be determined by the following method [Sherony *et al.*, 1978].

The Carman-Kozeny equations for single-phase and two-phase flows are expressed as follows.

$$\Delta P_1 = \frac{UL\mu k_1 (1-\phi)^2 A_0^2}{\phi^3}, \quad (4.12a)$$

$$\Delta P_2 = \frac{UL\mu k_2 (1-\phi_t)^2 A_0^2}{(\phi_t)^3}, \quad (4.12b)$$

where k_1 and k_2 are the Kozeny constant for single-phase and two-phase flows, respectively; ΔP_1 and ΔP_2 are the pressure drops across the coalescing bed for single-phase and two-phase flows; and ϕ_t is the true porosity.

Combining Equations (4.12a) and (4.12b) gives,

$$\frac{\Delta P_1}{\Delta P_2} \frac{k_2}{k_1} \frac{\phi^3}{(1-\phi)^2} = \frac{(\phi_t)^3}{(1-\phi_t)^2}. \quad (4.13)$$

In this equation, k_1 , ΔP_1 , and ΔP_2 can be obtained from the experiment. Then the two unknown variables are Kozeny constant for two-phase flow k_2 and true porosity ϕ_t . In order to solve this equation, a relation between these two variables is needed. In the literature, a curve of Kozeny constant versus porosity is available [Sherony *et al.*, 1978]. With this curve, these two variables can be determined by solving Equation (4.13). However, it should be noted that this curve is only applicable for the fibrous beds with high porosity. Therefore, a new method is required to determine the true porosity for the coalescing beds with intermediate porosity.

As previously mentioned in 4.1.3, for a granular medium with non-straight paths of pores, the tortuosity T can be estimated as $1/\phi$ [Johnston, 1983]. This relation provides a new way to determine the true porosity for the two-phase flow. Since the shape factor k_0 is a function of the particle shape only, it is the same in the single-phase flow and the two-phase flow. Hence, the Kozeny constant k_2 ($k = k_0 T^2$) depends on the tortuosity only (i.e., true porosity). Thus, the Kozeny constant k_2 and the true porosity ϕ_t can be easily determined from Equation (4.13).

4.2 Parametric Studies

4.2.1 Effect of flow rate

A series of tests are performed to determine the effect of flow rate on the coalescence performance. The detailed results are listed in Tables 4.1, 4.2, and 4.3. In these tables, the true porosity can be determined by using the method discussed in 4.1.4. The degree of saturation is calculated according to Equation (1.7). As the mean droplet diameters and the number densities at the inlet and the outlet can be measured experimentally, the overall coalescence efficiency can be determined by using the Equations (3.38a) and (3.38b) derived in Chapter 3.

$$\eta_c = - \frac{\ln Y}{\left[\frac{3(1-\phi)S}{\pi d_f(1-S)} \left(1 + \frac{d_i}{d_f} \right) + \frac{16C_o(1-\phi)}{\pi d_f \phi} \right] L} \quad (\text{Fibrous bed}).$$

$$\eta_c = - \frac{\ln Y}{\left[\frac{9(1-\phi)S}{2\pi d_g(1-S)} \left(1 + \frac{d_i}{d_g} \right) + \frac{24C_o(1-\phi)}{\pi d_g \phi} \right] L} \quad (\text{Granular bed}).$$

As shown in Figure 4.10, the overall coalescence efficiency versus the superficial velocity is plotted at three different inlet oil concentrations. It can be found from this figure that the overall coalescence efficiency is not a monotonous function of the superficial velocity at lower inlet oil concentrations (2,000 and 5,000 ppm). There is a velocity range in which high coalescence performance can be achieved. In the literature, Sherony *et al.* [1978] found that coalescence efficiency increases with flow rate in their theoretical analysis. They assumed that the direct interception is the only mechanism in the capture of oil droplets by solids. The aerosol filtration theory is then applied to determine the interception efficiency. It should be noted that such determined interception efficiency may not fully represent the overall coalescence efficiency. In fact, the overall coalescence efficiency may be affected by at least three processes including approach, attachment and release as described in Chapter 1. Thus, the overall coalescence performance cannot be accurately evaluated using the theoretical analysis. In this study, it has also been found that the overall coalescence efficiency decreases as the superficial velocity increases at high inlet oil concentration (10,000 ppm). This finding is in agreement with published data [Viraraghavan *et al.*, 1997], where there was a decrease in overall coalescence efficiency with an increase in superficial velocity. In the previous study [Viraraghavan *et al.*, 1997], the emulsions used were water-in-oil emulsions and the dispersed phase (water) concentration was around 10,000 ppm. Thus, at higher inlet oil concentrations, lower superficial velocity is needed to achieve the better coalescence efficiency.

In Figure 4.10, it can be seen that the overall coalescence efficiency is higher in the range of superficial velocity of 1.0-1.5 mm/s for the emulsions tested in this study. The corresponding Reynolds number is equal to 0.74-1.11.

4.2.2 Effect of inlet oil concentration

In this study, three inlet oil concentrations are chosen to examine their effects on the coalescence performance in the granular beds as shown in Figure 4.10. It is found that, in general, the overall coalescence efficiency decreases with an increase in the inlet oil concentration. The maximum overall coalescence efficiency is 10.42%, 9.37%, and 4.48% for 2,000, 5,000, 10,000 ppm inlet oil concentrations, respectively. As previously discussed, there are two types of collisions in the coalescing beds. The first type of collisions occurs between an oil droplet in the stream and an oil droplet on the solid surface. The second type of collisions happens between two oil droplets in the stream. Usually, the number density of oil droplets increases as the inlet oil concentration increases. This results in more collisions among the oil droplets in the stream. It is expected that increasing the number of collisions can improve the overall coalescence efficiency. As shown in Figure 4.10, on the contrary, the overall coalescence efficiency decreases with an increase of the inlet oil concentration. Thus, it can be inferred from this fact that the second type of collisions between two oil droplets in the stream may play an unimportant role in the coalescing beds studied here.

The specific surface area of the granular material in this study is $A_0=8.77\times10^3$ m²/m³. This value is relatively small as compared with those for typical fibrous coalescers, whose specific surface areas can be as large as 2.0×10^5 m²/m³. It is

speculated that inadequate specific surface area is also the partial reason why the overall coalescence efficiency is low in this study.

4.2.3 Effect of bed length

In this test, three different bed lengths are chosen to study the effect of bed length on the coalescence performance in the granular beds. Here, the accumulative drop size frequency is used to assess the coalescence performance. Figures 4.11 and 4.12 show the accumulative drop size frequency versus drop size at the inlet and three outlets at two different flow rates. Here, the accumulative drop size frequency is defined as:

$$f_d = \frac{\text{Number of droplets with diameters less than } d}{\text{Total number of droplets}}. \quad (4.14)$$

Physically, the accumulative drop size frequency can reflect the effectiveness of the fibrous or granular beds in coalescing emulsion droplets. In Figures 4.11 and 4.12, the outlet accumulative drop size distributions are plotted at bed lengths of 20, 40, 70 cm from the inlet. It can be seen that the accumulative drop size distributions at the three outlets are rather different from that at the inlet. It can also be observed that the accumulative drop size frequency at each outlet is smaller than that at the inlet. This fact indicates that the overall drop size becomes larger after the emulsions pass through the coalescing bed. In these two figures, the maximum drop sizes are around 20 μm at the inlet, while the largest droplets can be as large as 38 μm at the outlets.

As shown in Figures 4.11 and 4.12, it can be found that the accumulative drop size distributions at the three outlets are rather similar. This fact implies that the length of the coalescing bed plays an unimportant role in the coalescence performance. There may exist a minimum length. As long as the length of the

coalescing beds is larger than this minimum length, there is no need to further increase the bed length.

4.2.4 Effect of fiber size

In general, large specific surface area can improve the coalescence performance. In the fibrous beds, the smaller the fiber size, the larger the specific surface area for the bed with a given volume. In order to examine the effect of the fiber size on the coalescence performance, two kinds of fibers are tested in this study, both of which are made from oil-wet materials. One is the polypropylene fiber with a characteristic fiber size of 170 μm and the other is the nylon fiber with a characteristic fiber size of 80 μm .

Figures 4.13, 4.14, 4.15, and 4.16 show the accumulative drop size frequency versus drop size for the polypropylene fiber and the nylon fiber at two superficial velocities ($U=1.446$ and 2.286 mm/s). In each figure, the accumulative drop size frequencies at the inlet and at the outlet $L=20$ cm are compared. The smaller the accumulative drop size frequency compared at the same drop size, the better the coalescence performance. Here, the accumulative drop size frequency at the mean inlet drop size is chosen to assess the coalescence performance at the outlet. It can be seen from Figures 4.13 and 4.14 that the accumulative drop size frequency decreases about 15% for the polypropylene fiber at the mean inlet droplet diameter of 5.53 μm , whereas the accumulative drop size frequency decreases about 37% for the nylon fiber at the mean inlet droplet diameter of 7.66 μm . These two tests are conducted at the superficial velocity of 1.446 mm/s. This trend holds at the superficial velocity of 2.286 mm/s, as shown in Figures 4.15 and 4.16, where the accumulative drop size

frequency decreases roughly 10% for the polypropylene fiber at the mean inlet droplet diameter of 4.09 μm and 40% for the nylon fiber at the mean inlet droplet diameter of 7.55 μm , respectively. Thus, it is concluded that smaller fibers have better coalescence performance than larger fibers if other parameters are kept invariant.

This conclusion is further supported by the previous theoretical study [Spielman and Goren, 1971]. In the past, many researchers have studied the coalescence due to interception, diffusion and inertial impaction in the fibrous beds. In general, the diffusion becomes relatively unimportant as long as the drop size is greater than 1 μm . Also the inertial impaction can be ignored when the drop size is less than 10 μm . Thus, the interception plays a major role in the coalescence performance.

In the literature, the interception efficiency has been theoretically determined for the emulsion flow around a fiber. An expression for the coalescence performance in the case of a cylindrical collector is given by Spielman and Goren [1971]:

$$\eta = 2A_{cyl} \left(\frac{d_m}{d_f} \right)^2, \quad (4.15)$$

where η is the interception efficiency; d_m is the mean droplet diameter; and d_f is the fiber size. The term A_{cyl} represents the hydrodynamic flow parameter for the fibrous bed and is given by:

$$A_{cyl} = - \left[\ln \phi + \frac{(1 - \phi^2)}{(1 + \phi^2)} \right], \quad (4.16)$$

where ϕ is the porosity of the fibrous beds.

Equation (4.15) clearly shows that decreasing the fiber size can greatly increase the interception efficiency and thus improve the overall coalescence performance.

The experimental results obtained in this study are consistent with the theoretical predictions.

4.3 Summary

In this chapter, different methods for predicting the pressure drop in the fibrous and granular beds are presented. The predicted pressure drops across both the fibrous beds and the granular beds for single-phase flow are rather consistent with the experimental results. It is found that the Kozeny constant can be geometrically determined for the fibrous beds. Nevertheless, the shape factor of the granular beds needs to be determined experimentally. These findings will be helpful in designing the coalescing beds.

The effects of four important parameters on the coalescence performance are examined. These parameters include flow rate, inlet oil concentration, bed length, and fiber size. In this study, the coalescence performance is evaluated by using either the overall coalescence efficiency or the accumulative drop size frequency. It is found that there is a flow rate range in which better coalescence performance can be achieved. Meanwhile, it is also found that the coalescing beds tested in this study may be effective at low inlet oil concentrations. In addition, it can be further concluded that increasing the length of the coalescing beds does not improve the coalescence performance as long as the bed length is beyond a minimum value. Also, it has been proved that the coalescence performance is better if smaller fibers are used.

Table 4.1 Experimental results of single- and two-phase flows at L=20 cm, 2,000 ppm inlet oil concentration (granular bed).

	Flow rate (ml/min)			
	206.21	308.92	468.59	542.68
Superficial velocity U (mm/s)	0.829	1.242	1.883	2.181
ΔP for single-phase flow (Pa)	799.49	1259.90	1985.20	2298.40
ΔP for two-phase flow (Pa)	799.49	1284.60	2082.30	2587.70
True porosity for two-phase flow*	0.384	0.381	0.375	0.362
Saturation S	0	0.0078	0.0234	0.0573
Mean inlet droplet diameter (μm)	8.89	7.81	8.48	7.47
Mean outlet droplet diameter (μm)	8.90	8.84	9.53	9.26
Overall Coalescence efficiency η_c	0	0.1042	0.0414	0.0323

* The original porosity: 0.384

Table 4.2 Experimental results of single- and two-phase flows at L=20 cm, 5,000 ppm inlet oil concentration (granular bed).

	Flow rate (ml/min)			
	180.43	267.11	410.00	523.70
Superficial velocity U (mm/s)	0.725	1.074	1.648	2.105
ΔP for single-phase flow (Pa)	995.00	1499.00	2313.97	2966.80
ΔP for two-phase flow (Pa)	1118.80	1624.25	2861.50	3411.70
True porosity for two-phase flow*	0.322	0.328	0.307	0.318
Saturation S	0.0557	0.0381	0.0997	0.0674
Mean inlet droplet diameter (μm)	7.99	9.19	6.77	7.43
Mean outlet droplet diameter (μm)	10.75	13.70	11.62	8.70
Overall coalescence efficiency η_c	0.0463	0.0937	0.0440	0.0202

* The original porosity: 0.341

Table 4.3 Experimental results of single- and two-phase flows at $L=20$ cm, 10,000 ppm inlet oil concentration (granular bed).

	Flow rate (ml/min)			
	177.37	276.36	403.08	494.93
Superficial velocity U (mm/s)	0.713	1.110	1.620	1.989
ΔP for single-phase flow (Pa)	708.26	1090.97	1607.53	1976.26
ΔP for two-phase flow (Pa)	766.54	1171.14	1895.09	2381.18
True porosity for two-phase flow*	0.370	0.371	0.354	0.350
Saturation S	0.0365	0.0339	0.0781	0.0886
Mean inlet droplet diameter (μm)	7.22	7.87	6.47	7.49
Mean outlet droplet diameter (μm)	9.48	9.79	9.29	8.78
Overall coalescence efficiency η_c	0.0448	0.0365	0.0314	0.0122

* The original porosity: 0.384

Figure 4.1 Geometric shape of the polypropylene fiber.

Figure 4.2 Schematic diagram of flow path.

Figure 4.3 Pressure drop versus flow rate for single-phase flow in the fibrous bed.

Figure 4.4 Comparison of the measured and predicted results for a fibrous bed.

Figure 4.5 Pressure drop versus flow rate for the single-phase flow in the granular bed ($\phi=0.367$).

Figure 4.6 Comparison of the measured and predicted results for a granular bed.

Figure 4.7 Friction factor f_p versus $Re_p/(1-\phi)$ for single- and two-phase flows at inlet oil concentration of 2,000 ppm in the granular bed.

Figure 4.8 Friction factor f_p versus $Re_p/(1-\phi)$ for single- and two-phase flows at inlet oil concentration of 5,000 ppm in the granular bed.

Figure 4.9 Friction factor f_p versus $Re_p/(1-\phi)$ for single- and two-phase flows at inlet oil concentration of 10,000 ppm in the granular bed.

Figure 4.10 Overall coalescence efficiency η_c versus superficial velocity at inlet oil concentrations of 2,000, 5,000, and 10,000 ppm in the granular bed.

Figure 4.11 Accumulative drop size frequency versus drop size at superficial velocity of 1.110 mm/s and inlet oil concentration of 10,000 ppm in the granular bed (mean inlet droplet diameter: 7.87 μm).

Figure 4.12 Accumulative drop size frequency versus drop size at superficial velocity of 1.620 mm/s and inlet oil concentration of 10,000 ppm in the granular bed (mean inlet droplet diameter: 6.47 μm).

Figure 4.13 Accumulative drop size frequency versus drop size at superficial velocity of 1.446 mm/s and inlet oil concentration of 5,000 ppm in the polypropylene fiber bed (mean inlet droplet diameter: 5.53 μm).

Figure 4.14 Accumulative drop size frequency versus drop size at superficial velocity of 1.446 mm/s and inlet oil concentration of 5,000 ppm in the nylon fiber bed (mean inlet droplet diameter: 7.66 μm).

Figure 4.15 Accumulative drop size frequency versus drop size at superficial velocity of 2.286 mm/s and inlet oil concentration of 5,000 ppm in the polypropylene fiber bed (mean inlet droplet diameter: 4.09 μm).

Figure 4.16 Accumulative drop size frequency versus drop size at superficial velocity of 2.286 mm/s and inlet oil concentration of 5,000 ppm in the nylon fiber bed (mean inlet droplet diameter: 7.55 μm).

CHAPTER 5 CONCLUSIONS AND RECOMMENDATIONS

In this thesis, coalescence of O/W emulsions through a fibrous bed or granular bed is studied. The Sherony and Kintner model has been improved to evaluate the coalescence performance in a coalescing bed. The Carman-Kozeny equation has been successfully applied to predict the pressure drop across the coalescing beds. A geometric method and an experimental method are used to calculate the Kozeny constant for the fibrous bed and granular bed, respectively. In addition, an experimental setup is established to conduct the laboratory study. Two fibrous materials and one granular material are tested in the experiment. Furthermore, parametric studies are undertaken. These parameters include flow rate, inlet oil concentration, bed length, and fiber size. The coalescence performance is assessed in terms of the overall coalescence efficiency obtained from the improved Sherony and Kintner model or the accumulative drop size frequency. Then the detailed effects of these parameters on the coalescence performance are analyzed. The following major conclusions can be made based on this thesis study:

1. An improved Sherony and Kintner model for determining the overall coalescence efficiency in the coalescing bed is presented. In this model, two types of collisions are considered, collisions between the oil droplets in the stream and those onto the solid surface and collisions between the oil droplets in the stream.
2. The Carman-Kozeny equation can be used to accurately predict the pressure drop for the fibrous bed and granular bed.

3. It is found that for a fibrous bed, a geometric method can be used to accurately calculate the Kozeny constant. For a granular bed, an experimental method can be used to determine the Kozeny constant.
4. The pressure drop is only slightly increased for the O/W emulsion flow in comparison with that for the single-phase flow. It is inferred that interception is the dominant mechanism in the coalescence bed studied in this thesis.
5. The overall coalescence efficiency for the O/W emulsion flow in the coalescing bed is not a monotonous function of flow rate. In this study, it is found that better coalescence performance occurs in the intermediate range of superficial velocity.
6. Higher inlet oil concentration leads to lower overall coalescence efficiency. This fact implies that the collisions between the droplets in the stream may play an unimportant role in determining the coalescence performance.
7. In terms of the accumulative drop size frequency, it is found that there exists a minimum bed length above which the coalescence performance remains almost the same.
8. Smaller fibers result in better coalescence performance. This may be attributed to the larger specific surface area in the coalescing bed made of smaller fibers.

Based on this thesis study, the following recommendations for future research are made:

1. It is to be ascertained whether interception is the dominant coalescing mechanism for different packing materials and emulsions.
2. A comprehensive mathematical model is desired to simulate the emulsion flow in coalescing beds. In particular, some important factors, such as interfacial tension,

electrical double-layer force, van der Waals force and wettability, should be incorporated into the model.

3. Visual study may be applied to observe the saturation distribution along the bed length.
4. The detailed interactions between emulsion droplets and porous media need to be thoroughly studied.

NOMENCLATURE

A_{cyl} = hydrodynamic flow parameter for a fibrous bed.

A_0 = specific surface area based on the solid volume of porous medium, m^{-1} .

A_D = solid surface area covered by the deposited droplets, m^2 .

A_f = total surface area of fiber, m^2 .

A_g = total surface area of granular particles, m^2 .

C_0 = volume concentration of the dispersed phase, v/v %.

$d_{2,0}$ = area-mean droplet diameter of distribution g , μm .

$d_{3,0}$ = volume-mean droplet diameter of distribution g , μm .

$d_{3,2}$ = Sauter mean diameter, μm .

d_f = characteristic size of the fiber, μm .

d_g = characteristic size of the granular particle, μm .

d_i = mean inlet droplet diameter, μm .

$d_{i,j}$ = mean droplet diameter of distribution g as defined in Equation (3.25), μm .

d_m = mean droplet diameter, μm .

D_∞ = Brownian diffusion coefficient, m^2/s .

D_H = hydraulic diameter of the coalescing bed, μm .

D_p = characteristic diameter of particle or fiber, μm .

f_d = accumulative drop size frequency at a drop size of d .

f_p =friction factor.

$f(m, x, t)$ =drop size distributions in the stream, m^{-3} .

$g(m, x, t)$ =drop size distributions on the solid surface, m^{-3} .

$h(m, x, t)$ =total rate of coalescence among the droplets inside the coalescing bed, $m^3 s^{-1}$.

$h_1(m, x, t)$ =rate of coalescence between a droplet in the stream and a droplet on the solid surface, $m^{-3} s^{-1}$.

$h_2(m, x, t)$ =rate of coalescence between two droplets in the stream, $m^{-3} s^{-1}$.

k =Kozeny constant.

k_1 =accommodation coefficient associated with $h_1(m, x, t)$ as defined in Equation (3.5), m^3/s .

k_2 =accommodation coefficient associated with $h_2(m, x, t)$ as defined in Equation (3.10), m^3/s .

k_1 =Kozeny constant for single-phase flow as defined in Equation (4.12a).

k_2 =Kozeny constant for two-phase flow as defined in Equation (4.12b).

k_{ck} =permeability, m^2 .

k_0 =shape factor.

K_G =gradient collision coefficient, m^3/s .

L =length of the coalescing bed, m.

L_e =effective length of coalescing bed, m.

m =mass of a droplet, kg.

m_0 =total number of droplets in the coalescing bed.

m_H =hydraulic radius for fibrous or granular bed, m.

N_0 =number of collisions between a droplet in the stream and a droplet on the solid surface in V volume per unit time, s^{-1} .

n_0 =number of collisions of droplets with a solid surface per unit surface area and per unit time for fibrous or granular bed, $m^{-2}s^{-1}$.

n_1 =number of collisions of droplets with a solid surface per unit time for per droplet, s^{-1} .

n_2 =number of collisions between droplets and the fiber or granular particles per unit volume and per unit time, $m^{-3}s^{-1}$.

Re_p =Reynolds number.

S =the degree of saturation.

T =tortuosity.

u =interstitial speed of the fluid, m/s.

U =superficial velocity, m/s.

V =total free volume available for the emulsion flow, m^3 .

V_B =bulk volume of fibrous or granular bed, m^3 .

V_H =volume of holdup, m^3 .

Y = ratio of the outlet to inlet droplet number densities.

α_0 =drop number densities in the stream f , m^{-3} .

β_0 =drop number densities in the holdup g , m^{-3} .

β_1 =droplet mass concentration per unit volume in the holdup g , kgm^{-3} .

Γ =shear rate, s^{-1} .

η_1 = contact efficiency.

η_2 = coalescence efficiency.

η_c = overall coalescence efficiency.

ΔP_1 = pressure drop across the coalescing bed for single-phase flow, Pa.

ΔP_2 = pressure drop across the coalescing bed for two-phase flow, Pa.

μ_c = viscosity of the continuous phase, mPa·s.

ρ_0 = density of the dispersed phase, kgm⁻³.

ϕ = porosity.

ϕ_t = true porosity for two-phase flow.

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Appendix 1.1 Drop image at the inlet (polypropylene fiber at superficial velocity U of 2.286 mm/s and inlet oil concentration of 5,000 ppm).

Appendix 1.2 Drop image at the outlet $L=20$ cm (polypropylene fiber at superficial velocity U of 2.286 mm/s and inlet oil concentration of 5,000 ppm).

Appendix 1.3 Drop image at the inlet (polypropylene granular particle at superficial velocity U of 1.242 mm/s and inlet oil concentration of 2,000 ppm).

Appendix 1.4 Drop image at the outlet $L=20$ cm (polypropylene granular particle at superficial velocity U of 1.242 mm/s and inlet oil concentration of 2,000 ppm).

Appendix 1.5 Drop image at the inlet (polypropylene granular particle at superficial velocity U of 1.110 mm/s and inlet oil concentration of 5,000 ppm).

Appendix 1.6 Drop image at the outlet at $L=20$ cm (polypropylene granular particle at superficial velocity U of 1.110 mm/s and inlet oil concentration of 5,000 ppm).

Appendix 1.7 Drop image at the inlet (polypropylene granular particle at superficial velocity U of 1.074 mm/s and inlet oil concentration of 10,000 ppm).

Appendix 1.8 Drop image at the outlet at $L=20$ cm (polypropylene granular particle at superficial velocity U of 1.074 mm/s and inlet oil concentration of 10,000 ppm).

Appendix 2.1 Flow rates and pressure drops for single-phase flow (polypropylene fiber bed at $L=20$ cm and $\phi=0.5611$)

Actual flow rate (ml/min)	Average flow rate (ml/min)	Average U (mm/s)	ΔP (Pa)	Average ΔP (Pa)	Re
771.28	776.23	3.120	1770.17	1770.17	1.879
777.94			1770.17		
779.95			1770.17		
775.75			1770.17		
690.05	694.67	2.792	1454.74	1453.42	1.681
691.38			1454.74		
699.70			1452.11		
697.52			1452.11		
586.46	587.78	2.362	1210.28	1210.28	1.424
586.57			1210.28		
586.79			1210.28		
591.31			1210.28		
509.83	509.29	2.047	1028.91	1028.91	1.234
507.12			1028.91		
510.72			1028.91		
509.49			1028.91		
409.15	410.18	1.648	802.85	2007.40	0.994
409.28			802.85		
411.18			800.22		
411.11			800.22		
318.27	318.16	1.279	592.56	592.56	0.771
319.13			592.56		
316.21			592.56		
319.02			592.56		
249.36	248.27	0.998	424.33	408.56	0.601
246.29			411.19		
248.69			403.30		
248.71			395.42		
192.39	192.07	0.772	248.22	242.30	0.465
192.71			242.96		
192.08			240.33		
191.09			237.70		
143.24	143.86	0.578	114.16	109.56	0.348
144.04			108.90		
145.23			108.90		
142.93			106.27		

Appendix 2.2 Flow rates and pressure drops for single-phase flow (polypropylene fiber bed at $L=40$ cm and $\phi=0.5611$)

Actual flow rate (ml/min)	Average flow rate (ml/min)	Average U (mm/s)	ΔP (Pa)	Average ΔP (Pa)	Re
772.23	777.89	3.127	3639.09	3648.43	1.884
777.34			3639.09		
779.83			3639.09		
782.15			3636.46		
657.62	661.92	2.792	3060.80	3060.80	1.601
659.75			3060.80		
662.83			3060.80		
664.38			3060.80		
561.44	562.80	2.262	2600.80	2600.80	1.363
561.89			2600.80		
563.67			2600.80		
564.20			2600.80		
469.11	469.76	1.888	2172.34	2172.34	1.138
470.79			2172.34		
469.50			2172.34		
469.65			2172.34		
386.74	385.81	1.551	1778.05	1777.40	0.935
385.87			1778.05		
387.19			1778.05		
383.45			1775.43		
304.19	304.39	1.223	1389.02	1389.02	0.737
304.68			1389.02		
303.87			1389.02		
304.02			1389.02		
206.09	205.74	0.827	931.65	408.56	0.498
205.81			931.65		
205.37			931.65		
205.71			931.65		
167.99	167.40	0.673	745.02	744.36	0.406
167.68			745.02		
166.93			742.39		
167.02			745.02		
124.99	125.23	0.503	547.88	547.22	0.303
124.69			545.25		
125.35			547.88		
124.87			547.88		

Appendix 2.3 Flow rates and pressure drops for single-phase flow (polypropylene fiber bed at L=70 cm and $\phi=0.5611$)

Actual flow rate (ml/min)	Average flow rate (ml/min)	Average U (mm/s)	ΔP (Pa)	Average ΔP (Pa)	Re
765.60	769.14	3.091	5607.90	5509.33	1.863
767.76			5486.99		
770.33			5463.33		
772.88			5479.10		
669.09	668.52	2.687	4616.93	4616.93	1.619
669.26			4616.93		
668.55			4616.93		
667.18			4616.93		
559.47	562.40	2.260	3757.38	3754.75	1.362
561.57			3757.38		
563.32			3754.75		
565.23			3749.49		
455.31	455.27	1.830	3000.36	2998.37	1.103
456.70			2997.72		
456.22			2997.72		
452.86			2997.72		
370.67	370.92	1.491	2422.06	2420.74	0.898
370.32			2419.43		
371.34			2422.06		
370.79			2419.43		
310.85	310.48	1.248	2004.11	2007.40	0.752
310.37			2006.74		
310.89			2009.37		
309.83			2009.37		
243.09	244.02	0.981	1551.99	1553.97	0.591
243.43			1554.63		
244.16			1554.63		
245.41			1554.63		
165.16	164.50	0.661	1010.51	1009.85	0.398
164.61			1010.51		
163.71			1010.51		
164.50			1007.88		
124.14	124.18	0.499	697.71	695.73	0.301
124.80			695.08		
123.95			695.08		
123.82			695.08		

Appendix 2.4 Test results for polypropylene fiber (at inlet oil concentration of 5,000 ppm and $\phi=0.4788$)

Flow rate (ml/min)	Superficial velocity U (mm/s)	Mean inlet droplet diameter (μm)	Mean droplet diameter at 20 cm (μm)	Mean droplet diameter at 40 cm (μm)	Mean droplet diameter at 70 cm (μm)
254.73	1.018	5.01	6.59	6.34	6.08
367.80	1.446	5.53	6.33	7.16	7.67
469.47	1.877	5.41	6.21	6.80	7.50
581.50	2.286	4.09	7.28	6.60	6.13

Appendix 2.5 Test results for nylon fiber (at inlet oil concentration of 5,000 ppm and $\phi=0.4500$)

Flow rate (ml/min)	Superficial velocity U (mm/s)	Mean inlet droplet diameter (μm)	Mean droplet diameter at 20 cm (μm)	Mean droplet diameter at 40 cm (μm)	Mean droplet diameter at 70 cm (μm)
367.82	1.446	7.66	9.40	9.82	10.30
475.48	1.869	7.80	9.24	9.16	9.55
581.72	2.286	7.55	9.12	9.97	9.64
694.82	2.731	8.48	10.32	10.02	9.99
813.70	3.198	6.74	8.08	8.28	8.40

Appendix 2.6 Test results for polypropylene granular particles (at inlet oil concentration of 2,000 ppm and L=20 cm)

Flow rate (ml/min)	Superficial velocity U (mm/s)	ΔP_1 for single-phase flow (Pa)	ΔP_2 for two-phase flow (Pa)	Mean inlet droplet diameter (μm)	Mean outlet droplet diameter (μm)
206.21	0.829	799.49	799.49	8.89	8.90
308.92	1.242	1259.90	1284.60	7.81	8.84
468.59	1.883	1985.20	2082.30	8.48	9.53
542.68	2.181	2298.40	2587.70	7.47	9.26

Appendix 2.7 Test results for polypropylene granular particles (at inlet oil concentration of 5,000 ppm and L=20 cm)

Flow rate (ml/min)	Superficial velocity U (mm/s)	ΔP_1 for single-phase flow (Pa)	ΔP_2 for two-phase flow (Pa)	Mean inlet droplet diameter (μm)	Mean outlet droplet diameter (μm)
180.43	0.725	995.00	1118.80	7.99	10.75
267.11	1.074	1499.00	1624.25	9.19	13.70
410.00	1.648	2313.97	2861.50	6.77	11.62
523.70	2.105	2966.80	3411.70	7.43	8.72

Appendix 2.8 Test results for polypropylene granular particles (at inlet oil concentration of 10,000 ppm and L=20 cm)

Flow rate (ml/min)	Superficial velocity U (mm/s)	ΔP_1 for single-phase flow (Pa)	ΔP_2 for two-phase flow (Pa)	Mean inlet droplet diameter (μm)	Mean outlet droplet diameter (μm)
177.37	0.713	708.26	766.54	7.22	9.48
276.36	1.110	1090.97	1171.14	7.87	9.79
403.08	1.620	1607.53	1895.09	6.47	9.29
494.93	1.989	1976.26	2381.18	7.49	8.78