

## Three-phase Wellstream Gravity Separation

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### I. INTRODUCTION

Production of petroleum is a stepwise process. The reservoir fluid is taken through a wellhead and transported to a production site by using the reservoir pressure, submersible pumps and/or gas/water injection. The production site takes the pressure stepwise down to shipping conditions (often atmospheric pressure) and removes water, gas, and solids from the oil.

The first process equipment the incoming reservoir fluid enters after the initial pressure reduction is the primary separator. This unit removes most of the water from the oil, which continues to a secondary and possibly ternary separator (which is often equipped with electrodes to enhance coalescence). Often the primary separator removes all the water, and the other stages are used for pressure reduction and gas removal only. The final goal is to match the specifications of the refineries and/or transport companies.

The water separated from the oil is postprocessed for dumping or reinjection by flotation-based or centrifugal equipment, according to the applicable specification. The gas is similarly dried and compressed for shipping or reinjection.

All water/oil separation processes utilize the immiscibility and density difference between the two phases (the electrostatic unit uses the difference in polarity as well).

The primary gravity separator is an important factor, especially offshore, in making the process cost effective. At an offshore platform where volume is an expensive resource, it is important to design the separator as small and light as possible. This is particularly important with regard to gravity separators since they are usually larger than other equipment and have a potential for size reduction.

In the petroleum production on the Norwegian continental shelf new trends will emerge during the next 3—5 years. First, the amount of water produced from offshore platform separators will increase mainly as a result of aging fields where water breakthrough has taken place, giving a concomitant production of injection water together with the oil. This high production rate of water will place high demands on separator efficiency and treatment of wastewater. In addition to this, many new fields to be explored in the future will be complicated to develop, as the crude oil produced will contain large amounts of heavy components such as asphaltenes and resins. These components strongly increase the capability of the crude oil to bind water, which increases the necessary retention time in the separator. These new types of crude oils will also most likely lead to an increase in the use of production chemicals in the separator and in the transport process.

The effects of these two trends have to be implemented into the design tools used for optimizing gravity separators.

Tools in use today do not have a coalescence model for the dispersion entering the separator, but use only modified versions of Stokes' law when describing the settling/creaming of droplets and hence the separation. The influence of higher watercut and more stabilizing components increases the need for a coalescence model.

## A. Basic Principles of Gravity Separation

The two major phenomena recognized in phase separation are drop break-up and coalescence. Drop breakup is the process where one phase in an immiscible (multiphase) system forms an unstable, heterogeneous state of two or more distinct phases (drops) dispersed in a continuous phase. Coalescence is the reverse process where the system returns to the state of lowest total energy, i.e., separate homogeneous phases with a minimized common interface.

Of these two phenomena, break-up is by far the best understood. This is because turbulent forces, usually correlated with turbulent dissipation, often dominate break-up, and can be analyzed in terms of equilibrium states by considering the energy transport. Coalescence is often dominated by kinetics, and depends heavily on the chemical composition of the system.

The aim of a separator vessel is to give the coalescence process the necessary time and create conditions for satisfactory phase separation. The classical approach is to use an overall residence time criterion, which allows the drops to: (1) reach a bulk interface by sedimentation; and (2) coalesce with this bulk interface, forming a single homophase. One seeks to minimize any events contributing to drop break-up.

## B. Classical Design Philosophy-Sizing the Vessel

Correct modeling of the coalescence process for use in separator design is very difficult. The engineering solution has been to circumvent the problem by focusing on settling as described by Stokes' law, assuming that coalescence is sufficient. A standard developed throughout the years is to design the separator to handle a cut size of 200  $\mu\text{m}$  in both the water and oil phases, assuming that there exists a sharp interface controlled by the interface control system. Thus, more residence time is allowed for viscous oils. In addition, the different engineering companies may have proprietary safety factors/cut-size relations based on various parameters available at the time of design.

Often the design based on separation (or settling) characteristics is overridden by the necessity of certain fill-up times between the different alarm settings. These are based on the operation of the outlet valves versus shutdown criteria for the process. Standard shutdown criteria range from 30 s to 1 min between the different alarm levels. Following the applicable API specification (1), the vessel size is selected and possibly increased until the 200- $\mu\text{m}$  cut-size criterion is met. Proprietary factors and relations may modify the design slightly.

The separation involves settling and coalescence mechanisms. The settling velocity is a function of droplet size. The local velocity in the vessel and the settling regime (usually Stokian or Newtonian) determines the exact relationship between these two mechanisms (2). The coalescence of droplets within the dispersion and the dispersed continuous phase boundary is a complex function of droplet diameter as the gravity and surface forces that control the coalescence are both related to the droplet diameter, or more precisely the local curvature at the interface. It is known that when the droplet size reduces, so does the separation rate. When the droplets reach a size of about 30–60  $\mu\text{m}$  the separation is found to be settling controlled (2). In these cases the settling time is usually longer than the residence time of the bulk phases. Systems of this type are classified as secondary dispersions. Although they are thermodynamically unstable they lead to poor outlet quality of the continuous phase. Knowledge of the characteristics of the dispersion (droplet size distribution) is obviously of great importance in designing separator equipment.

There are several obvious flaws with this technique. First, there is no reason to believe that Stokes' law will be valid for the problems it is applied to, especially in the continuous phase (normally the oil phase) where the high droplet concentration may lead to significant droplet/droplet interaction. Second, the 200- $\mu\text{m}$  cutsize seems rather arbitrarily chosen (possibly from API specifications for refinery separators, following Ref. 3), and as far as the authors have been able to find out, no evidence exists that this necessarily resembles any actual droplet entering the separator. Third, the coalescence process is disregarded and the separator is assumed to be settling controlled. This is in reality a limiting case for real fluid systems, and by no means generally applicable. This also leads to the assumption of a sharp interface. This is generally incorrect, as the coalescence process normally creates a band of noncoalesced drops residing at the interface. This region has special characteristics, and will hereafter be referred to as the dispersion band.

These problems with classical design are the core of this chapter, and will be discussed below.

## II. FLUID PROPERTIES AND STABILIZING MECHANISMS

This section describes the various chemical properties of the fluids entering the separator, and how they interact with the coalescence rate. The intention is not to describe a coalescence model for design use, but merely to give an overview of the different stabilizing mechanisms that must be taken into account when trying to model coalescence.

The overall coalescence rate of a dispersion/emulsion in a separator is the most important design criterion. Unfortunately, this rate is a product of several complex mechanisms like binary coalescence, interfacial coalescence, and settling/creaming. Each of these mechanisms is further related to other even more complex processes/factors like hydrodynamic micro- and macro-motion, droplet size distribution, and interfacial components. In order to understand the overall coalescence rate one must also understand the interactions between these mechanisms. This makes it difficult to separate the overall rate into a sum of distinct rates, and is probably the reason why there exists no generalized coalescence model for concentrated dispersions with a sound theoretical foundation.

Coalescence is the process where two or more droplets combine and form a larger droplet. This is necessary to form a clear liquid layer from an initially dispersed phase. Droplets can coalesce owing to binary or interfacial coalescence.

Binary coalescence is coalescence between droplets that are settling/creaming or packed in the dispersion band, while interfacial coalescence is the coalescence of a droplet with its own phase (a droplet with infinite dimensions). In both cases a liquid film of continuous phase separates the dispersed droplets and this film has to be drained and broken in order to complete the coalescence process. Hartland describes this draining process in detail (4).

The following will mainly focus on the effect and mechanisms that are associated with the hindering of film drainage by the interfacial components.

When describing stabilizing mechanisms, one can distinguish between three types:

1. Steric stabilization;
2. Electrostatic stabilization;
3. Mechanical stabilization.

Steric stabilization occurs when the interfacial components have a long chained part of the molecule that stretches into the continuous phase. The term steric stabilization can be associated with several different mechanisms, but they are all related to an increase in the system free energy. The penetration mechanism is known to be the most important and can be described as a local increase in concentration of polymer segments in the film separating the droplets. If the continuous phase is a good solvent for the polymer segments, the local increase in concentration when the drops move closer will be thermodynamically unfavorable ( $\Delta G > 0$ ). The chemical potential of the solvent in the area between the droplets will decrease. This creates an osmotic pressure,  $\pi$ , that will oppose the increase in concentration by making the continuous phase flow into to the film between the droplets. The mechanisms is depicted in Fig. 1.

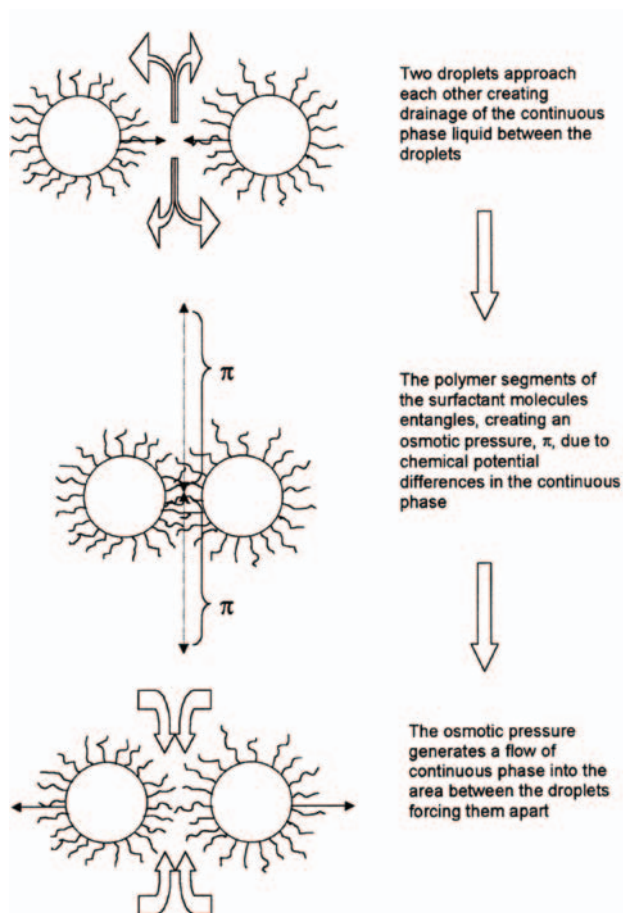


Figure 1 Sketch of the steric stabilization of drops.

Electrostatic stabilization occurs when the interfacial components are charged and the electric double layer between two or more droplets overlap. The resulting repulsive force counteracts further drainage of the film. Authors have, however, disregarded this repulsion as a significant stabilizing factor in describing water-in-crude oil stability (5, 6).

Mechanical stabilization is a process where interfacial components act as particles, creating a mechanically stable film on the surface of the droplets. This film encapsulates the droplets and, due to its immobility and low solubility in both water and oil, creates a very stable emulsion. Asphaltenes, resins, wax particles, minerals, and clay are compounds believed to enhance the formation of mechanically stable films.

All these mechanisms can be present when an emulsion is formed (although some are more predominant than others). This makes it very difficult to model emulsion stability as a function of fluid properties and interfacial components.

A more quantitative theory based on interfacial gradients can also be used to describe how film drainage is retarded, thereby stabilizing the dispersion. When interface-active components adsorb on a water/oil interface the interfacial tension will decrease monotonically as a function of the surface concentration. As two droplets approach each other the resulting film drainage will carry interfacial components away in the drainage direction, creating a concentration gradient. This gradient results in an inward interfacial-tension gradient, creating a positive inward force counteracting the film drainage. This theory assumes mobile W/O interfaces. Figure 2 shows a simplified scenario of this theory.

Accepting these stabilizing mechanisms may give qualitative ideas on how to explain the known separation of a given system, but cannot be used to predict the separation in advance. Even if the composition of the probable stabilizing components (asphaltenes, resins, wax, minerals, and clay) is known, it is very difficult to predict the stability of the resulting emulsion. One of the reasons is that the history of the fluid greatly influences the stability of the emulsion. Factors such as temperature-pressure variation and well combinations affect the solubility and thereby the size distribution of the stabilizing particles, which is thought to be an important stabilizing factor. Aging of an emulsion is also known to greatly affect the emulsion character. It is currently impossible to control all these factors, and difficult to gain detailed information about them.

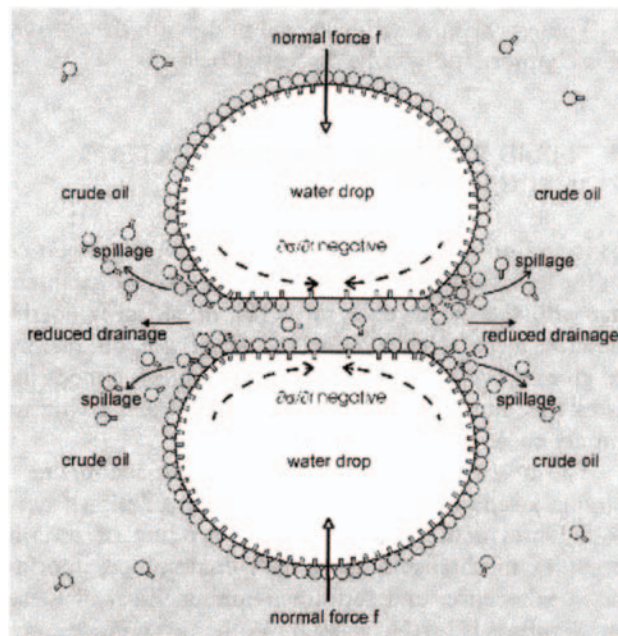


Figure 2 Sketch of the marangoni effect. (From Ref. 22.)

The petroleum industry generally “solves” the emulsion problem by adding demulsifiers in an ad hoc manner, often based on simple bottle tests. There are many problems associated with this solution. First, the chemical composition of a given well changes with time and can in a worst-case scenario result in a composition totally incompatible with the given demulsifier. Second, little is known about the exact interaction between demulsifiers and other chemical additives (e.g., corrosion inhibitors and “flow enhancers”). One may, therefore, create a new problem by solving another.

Another well-known term used in the petroleum industry is “watercut curves”. For a given crude oil they give the fraction of separated water as a function of the watercut at a given retention time (normally 4 min). A typical watercut curve is shown in Fig. 3. As can be seen from the figure, a higher watercut gives better separation. The lowest watercut that is associated with complete separation is called the critical watercut. A qualitative explanation of this could be that at this watercut, the total area of the dispersed droplets is larger than the maximum area that the interfacial components can cover for proper stabilization of the droplets. Operating a separator at this watercut solves the problem of a stable emulsion, but is not always cost effective as the water production increases at the expense of oil production.

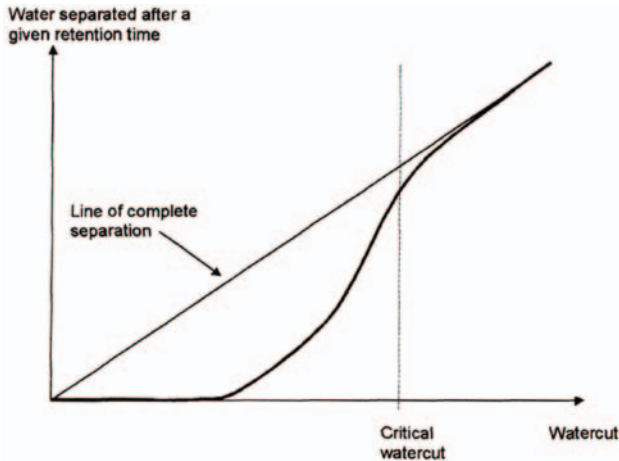


Figure 3 Sketch of a watercut curve.

### III. DROP SIZES IN TURBULENT REGIMES

It is widely recognized that the size of the drops entering the separator is a parameter of great importance, since it will affect both the settling and coalescence mechanics inside the vessel. The drop size at a given point in the process is dependent on the turbulent fluctuations, the history of the fluids up to that point, and the physical properties of the mixture. Traditionally, the existence of a specific drop size equilibrium in any turbulent field is assumed. This has been investigated experimentally by several authors, particularly for stirred systems (7). Some authors have also looked more specifically into tube flow (8).

Turbulent modeling is a large field of research in itself. Various turbulence models have been developed with the evolution of computers, as this is a field of high computing intensity. The bulk of experimental work has, as mentioned above, been performed on simple geometries, and the isotropic turbulence model (see below) has been used with spatially averaged values to approximate break-up effects in these experiments. For the more complex geometries typically found in an oil-producing process, such as manifolds, various valves, bends, and inlets, the relevance of these relations is questionable. Additionally, when following a multiphase flow from the well to the separator, equilibrium cannot always be assumed at the various regions observed. Depending on the stability of the emulsion(s) formed, the approach toward equilibrium will depend on the residence time in the various regions. For highly stable systems, the drop size used for design basis should probably be in the regime of highest turbulent intensity, corresponding to the smallest drops. This also leads to the

statement that, in order to make the separator inlet conditions more favorable, chemical destabilization should be performed as early as possible in the process. Turbulence theory is traditionally limited to single-phase flow, and the extension to multiphase behavior should be carried out with care. Kolev (9) gives a comprehensive review of various approaches and models. This chapter is, however, limited to the traditional approach of modeling the multiphase mixtures with single-phase turbulence relations.

Drop break-up is usually associated with turbulence and is most prominent in sections with high turbulent shear. Kolmogoroff [from Davies (7)] showed that the maximum diameter a drop can have in a local isotropic turbulent field is given by

$$d_{\max} = C \left[ \frac{\sigma}{\rho_c} \right]^{3/5} \epsilon^{-2/5} \quad (1)$$

Here,  $C$  is a constant near unity,  $\sigma$  is the interface tension,  $\rho_c$  is the continuous phase density, and  $\epsilon$  is the turbulent energy dissipation (energy input per mass unit). This equation has been verified experimentally for dilute, stirred systems of various types.

Davies (10) concluded that it is the turbulent fluctuation velocity  $\epsilon$  that is responsible for drop breakup. The dissipation term  $\epsilon$  is given by the following relation:

$$\epsilon = -\frac{1}{2} \left[ \frac{d}{dt} \right] [(\bar{v}'_x)^2 + (\bar{v}'_y)^2 + (\bar{v}'_z)^2] \quad (2)$$

The assumption of isotropic turbulence arises when one assumes that the various mean turbulent velocity fluctuations are equal, and thereby obtains a mean dissipation for a given mass (knowing the energy input):

$$\epsilon = -\frac{3}{2} \frac{d(\bar{v}')^2}{dt} \quad (3)$$

If the turbulent velocity fluctuations are known from simulations or measurements, the maximum drop sizes can also be calculated more accurately for anisotropic turbulence.

Davies (10) proposed a viscosity correction as shown in Eq. (4):

$$d_{\max} = C \left[ \frac{4\sigma + \mu_d \bar{v}'}{4\rho_c} \right]^{3/5} \epsilon^{-2/5} \quad (4)$$

Here,  $\mu_d$  is the dispersed phase viscosity.

For dilute pipe flow, Eq. (1) is expected to hold for the flow near the centre of the pipe where the turbulence can be

regarded as isotropic. However, the shear in turbulent pipe flow is mainly located near the walls where isotropy is a poorer assumption. Karabelas (8) showed that for a dilute pipe-flow system the maximum diameter surviving in turbulent pipe flow is given by

$$d_{\max} = 4.0D \left[ \frac{D\rho_c \mu^2}{\sigma} \right]^{-3/5} = 4.0 D We^{-3/5} \quad (5)$$

It should be noted that later work (11) has shown that this (and similar) equations may not be entirely correct as the experiments may not have been performed under steady-state conditions. This, however, falls beyond the scope of this chapter.

Polderman et al. (12) suggested a water-cut dependency on droplet size distribution at the Draugen field:

$$d_{\max} \propto \Delta P^{-3/5} (1 + 20\phi_0^2) \quad (6)$$

The pressure term represents the turbulent energy input across a valve. An interesting feature of this equation is that it shows a dependence of dispersed phase fraction. This equation is of a more empirical nature than the others, and the extra term could include both binary coalescence in the downstream region of the valve and the possibility that a larger dispersed mass would absorb and dissipate energy internally at larger eddy sizes. It addresses, however, the nondilute situation usually encountered in crude oil/water separation processes.

For the problem at hand, three regions are regarded as important for drop break-up: the choke valves, the tube from the choke to the inlet, and the inlet. For noncentrifugal inlets, the inlet momentum from the tube can be regarded as the energy input, and the length scale of the inlet can be used for scaling the dissipative volume. For cyclonic inlets, the dominating velocity is usually the tangential one, and the region of large dissipation is the liquid outlet region. Hence, inlet cyclone-related drop break-up is related to the cyclone liquid-outlet momentum and the liquid-outlet length scale. Depending on the stability of the system and the residence time in the tube, the engineer will have to estimate the size of the drop entering the separator, based on these regimes.

### A. Turbulence-induced Coalescence

Meijs and Mitchell (13) examined the possibilities of inducing coalescence by gentle turbulent mixing in tube flow, and found that droplets with an initial  $d_{\max} = 20 \mu\text{m}$  were coalesced to  $d_{\max} = 250 \mu\text{m}$ , by applying a turbulent mix-

ing intensity  $\epsilon$  of  $2 \text{ m}^2/\text{s}^3$ . This effect decreased linearly with  $\log \epsilon$  until  $d_{\max} = 39 \mu\text{m}$  for  $\epsilon$  of  $10^3 \text{ m}^2/\text{s}^3$ , with the same initial drop size distribution. They also indicated that a time span of the order of  $10 + \text{min}$  was needed to reach these new equilibria, depending on the amount of dispersed phase and energy input. The most efficient energy input level for the tests was found to be  $0.59 \text{ m}^2/\text{s}^3$ , and the efficiency decreased with decreasing dispersed fraction. The energy input level was estimated by the relation:

$$\epsilon = \frac{\mu^3 f}{D} \quad (7)$$

where  $u$  is the mean velocity in the tube,  $D$  is the tube diameter, and  $f$  is the friction factor defined as  $16/\text{Re}$ . They also found that Eq. (1) overpredicted the drop sizes by a factor of 1.5–3 for the system investigated.

## IV. GRAVITY SEPARATOR MECHANISMS

This section covers the status of mechanistic understanding of the processes in a gravity separator. Recently, several new philosophies (12, 15, 20) have emerged which study the mechanisms inside a separator, using different approaches.

### A. Settling Laws

The classical approach to settling is Stokes' law, a balance between gravity forces and drag on a solid, spherical particle in infinite dilution and for creeping flow where the Reynolds number is  $\ll 1$  (experiments have shown that the equation has validity for systems approaching  $\text{Re}=1$ ). The particle's terminal vertical velocity is given by

$$v_{st} = \frac{\Delta \rho g d^2}{18 \mu_c} \quad (8)$$

Stokes' law is an analytic solution of the Navier-Stokes equation for the simplified flow case with solid particles and creeping flow. If the particles are fluid and in the absence of surface-active components, internal circulation inside the particle will reduce the drag. (Note that this is not necessarily valid for small fluid particles, but these are irrelevant in gravity separation.) The viscosity correction term for this case is given in Eq. (9). From this equation it can be seen that, for large viscosity differences between the dispersed and continuous phases, the settling will approach the Stokes velocity or 3/2 Stokes velocity (the two limiting

cases), depending on what is dispersed. Viscous liquid drops in a gas will approach Stokes' law (negligible circulation), while gas bubbles in viscous liquids will approach 3/2 Stokes' law (high degree of circulation inside the bubbles).

$$v_{St, visc} = \frac{\Delta\rho g d^2}{18\mu_c} \frac{\mu_c + \mu_d}{2/3\mu_c + \mu_d} \quad (9)$$

Kumar and Hartland (14) reviewed 14 different data sources published, with 998 results for 29 different liquid/liquid systems and correlated Eq. (10) by nonlinear fitting. This equation is different from Eqs (8) and (9) in that it includes the initial amount of dispersed phase  $\Phi_0$ . This feature is called hindered settling, as it does not use infinite dilution as an initial assumption.

$$v_{KH} = \frac{12\mu_c}{0.53\rho_c d} \left[ -1 + \sqrt{1 + \frac{0.53\rho_c \Delta\rho g d^3 (1 - \phi_0)}{108\mu_c^2 (1 + 4.56\phi_0^{0.73})}} \right] \quad (10)$$

By calculating the mean residence time between two heights inside the separator, the smallest particle diameter that will traverse the vertical distance can be calculated. This is known as cut size. It is customary in gravity-separator design to calculate the cut size for an oil drop between the bottom of the vessel (BV) and the various oil/water interfaces (NIL, LIL, HIL). Likewise, the cut size for a water drop between the various liquid interfaces (NOL, LOL, HOL) and the interface is calculated. The vertical velocity criterion thus becomes:

$$v_c = \frac{h_1 - h_2}{\tau\eta} \quad (11)$$

where  $h_1$  and  $h_2$  are the chosen levels,  $\tau$  is the residence time of the fluid between these levels, and  $\eta$  is an efficiency depending on the flow regime;  $\eta = 1$  corresponds to a plug-flow approximation. The cut sizes for the various settling equations by this relation become:

$$\begin{aligned} d_c^{St} &= \sqrt{\frac{18v_c\mu_c}{\Delta\rho g}} \\ d_c^{St, visc} &= \sqrt{\frac{18v_c\mu_c(\mu_d + 2/3\mu_c)}{\Delta\rho g(\mu_d + \mu_c)}} \\ d_c^{KH} &= \frac{\frac{0.53\rho_c v_c^2}{12\mu_c} + \sqrt{\frac{0.53^2\rho_c^2 v_c^4}{12^2\mu_c^2} + \frac{\Delta\rho g(1-\phi_0)v_c}{\mu_c(1+4.56\phi_0^{0.73})}}}{\frac{\Delta\rho g(1-\varepsilon_0)}{4\mu_c(1+4.56\phi_0^{0.73})}} \quad (12) \end{aligned}$$

### 1. Plug Velocities and Retention Times

The retention-time calculations are based on the volume between the perforated plate and the weir plate. The cross-sectional area available to the phases (oil, water (or liquid), and gas) is calculated and the mean plug velocity and retention times are given. The cross-sectional area below a level  $h$  in a horizontal cylinder is generally given by the geometric relation:

$$A_c = r^2 \arccos\left(\frac{r-h}{r}\right) - r(r-h)\sin\left[\arccos\left(\frac{r-h}{r}\right)\right] \quad (13)$$

The plug velocity of a phase is then found by dividing a phase flow by its respective cross-sectional area, and the retention time is given by the ratio between the effective separator length and this plug velocity.

Also, a dispersion retention time can be defined, as in Eq. (14), by calculating the available dispersion volume between the O/W interface setting for the dispersed [water] phase:

$$\tau_d = F \frac{A_{NIL}(h_w - NIL)}{Q_w^{out}} \quad (14)$$

here,  $F$  is a currently unknown function depending on flow conditions [e.g., the concentration gradient within the dispersion layer, suggested as linear by Polderman et al. (12)];  $F > 1$ . Setting  $F = 1$  will give the volume available above the interface setting, thus being incorrect as the dispersion layer will extend below this level. The available region below NIL will depend on the suction from the water outlet, i.e., water outlet geometry and velocity, and weir height. If the concentration gradient through the dispersion layer is known, the correct volume occupied by the dispersion layer can be calculated. Note that this approach also puts stringent demands on knowing the absolute value of the interface setting which, depending on the control method, is often slightly inaccurate.

The splitting of the dispersed region into a (dense) dispersion layer and a (dilute) settling region has been developed for batch tests, where the only transport is parallel to gravity. These types of tests are common in laboratories for characterizing the separability of an oil/water system, are the origin of the "critical water cut" concept as depicted in Fig. 3, and are discussed briefly in Sec. II. In continuous systems there will also be a transport normal to gravity, and this will possibly create a concentration gradient within the dense region, as suggested by Polderman, et al. (12).

## B. Load Factors

The load factors attempt to compare different separator performances. Their origin is uncertain (possibly API). Three different load factors have been observed: the liquid (LLF), oil (OLF), and water (WLF) load factors. These are described in Eq. (15). Load factors have a unit of  $(\text{m}^2\text{s}^2)^{-1}$ .

$$\begin{aligned} LLF &= \frac{Q_1 \mu_c}{A_{\text{NIL}} \Delta \rho} 1000 \\ OLF &= \frac{Q_o \mu_w}{A_{\text{NIL}} \Delta \rho} 1000 \\ WLF &= \frac{Q_w \mu_o}{A_{\text{NIL}} \Delta \rho} 1000 \end{aligned} \quad (15)$$

The physical interpretation of the load factors is that a liquid flow should be transported through the interface, and this transport is augmented by the density difference and decreased by the continuous phase viscosity. The liquid load factor accounts for transport of both phases through the interface, while the other two only transport the applicable dispersed phase. Hence, the liquid load factor is in line with the new separator design philosophy proposed by Polderman et al. (12, 15), while the oil and water load factors are in line with the dispersion layer theory developed by Jeelani and Hartland (16-20) (for the corresponding dispersed phase).

## C. Binary Coalescence and Settling Hindered Systems

In systems with rapid coalescence the limiting parameter for separation will be the settling velocity (e.g., the droplet size distribution entering the separator). Hafskjold et al. (21) modeled the outlet oil quality from a continuous model separator by using batch data. The basis for their model was, in addition to Stokes' law [Eq. (8)] and the plug-flow approximation [Eq. (11)], the water concentration in the oil outlet as a function of the concentration gradient over the oil phase:

$$\phi_{\text{oil outlet}} = \frac{1}{h_{\text{oil}} - h_{\text{water}}} \int_{h_{\text{water}}}^{h_{\text{oil}}} \phi(z) dz \quad (16)$$

here,  $\Phi$  is the fraction of dispersed phase (e.g., water as all systems investigated were oil continuous),  $z$  is the height above the water level, and  $\Phi(z)$  is the local water concentration profile immediately upstream of the weir. This profile was modeled from batch tests as a Padé approximant

[Eq.(17)], with  $\Phi_0$  as the separator inlet concentration and  $\Phi_{\text{residual}}$  as the concentration still in the oil phase after a long separation time;  $t$  refers to a time in the batch tests corresponding to a retention time  $\tau$  in Eq. (11).

$$\frac{\phi_0 - \phi}{\phi - \phi_{\text{residual}}} = \left[ \frac{z}{h_{\text{oil}} - h_{\text{water}}} - z \right]^{a_1} \left[ \frac{t}{a_3} \right]^{a_2} \quad (17)$$

Here,  $a_1$  and  $a_2$  are characteristic parameters for the drop size distribution, and  $a_3$  characteristic separation time. This model was tested on an offshore test separator (light crude), with good results.

## D. Dispersion Layer Theory

For the past 15 years, Hartland and coworkers have developed a theory referred to here as the dispersion layer theory (20). The theory has been developed for batch tests, and has the following assumptions:

1. The incoming fluid has a defined, pseudohomogeneous continuity. Thus, the incoming mixture consists of one defined continuous phase and one defined disperse phase. The total volume of drops entering the separator is equal to the incoming dispersed phase flow.
2. All of the dispersed phase (in drop form) has to be transported through the continuous phase layer and the interface in order to achieve separation. This is divided into steps:
  - a. Transport through the continuous layer to the dispersion layer by settling.
  - b. Transport through the dispersion layer by stackwise removal of the interfacial drop layer by coalescence. The dispersion layer is considered to be a packed layer with a fixed drop concentration.
  - c. Transport through the interface by coalescence.
3. The settling through the continuous layer is hindered, and described by Eq (10).
4. The coalescence rate for a drop of diameter  $d_0$  at the interface is either correlated from experiments or calculated by theory.

Following these assumptions, there will exist a point in time where the last drop settles at the top of the dispersion layer, the inflection point  $t_i$ . A mass (or rather a volume) balance is calculated from this point, describing the heights of the region boundaries:

$$\frac{dh_c}{dt} = \Psi = \Psi_i \left[ \frac{\Delta h}{\Delta h_i} \right]^p \quad (18)$$



Here,  $h_c$  is the height of the coalescing interface [the interface between the two different liquid continuous regimes corresponding to  $h_{water}$  in Eq. (16)],  $\Delta h$  is the thickness of the dispersion layer, and  $\Delta h_i$  is the thickness at the inflection point;  $p$  is the interfacial coalescence index, a parameter describing the degree of packing within the dispersion layer, ranging from 0 to 1;  $\Psi_i$  (m/s) is the coalescence rate at the inflection point, the velocity with which the coalescing interface moves, given by

$$\Psi_i = \frac{2 \phi_p d_0}{3 \tau_0} \left[ \frac{\Delta h_i}{d_0} \right]^p \tag{19}$$

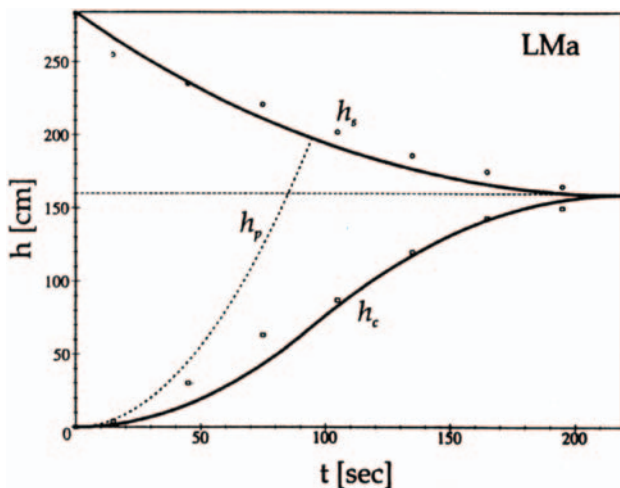
Here,  $d_0$  is the initial mean droplet size,  $\Phi_p$  the packed dispersed-phase hold-up, and  $\tau_0$  the coalescence time for drops of size  $d_0$ .

A modeling of a light/medium crude oil in batch settling is shown in Fig. 4 (from Ref. 22).

Panousopoulos (22) has studied several real and model systems, using this model, and reported values of  $\epsilon_p$  ranging from 0.65 to 0.875 and  $p$  ranging from 0.23 to 0.68. The current lack of a theory predicting these variations is considered to be the major weakness of this model.

### E. Design Philosophy for Dewatering Vessels (Developed by Shell)

In recent years, Shell has published a new design philosophy (12-15) based on extensive laboratory tests and field trials. The basis of this philosophy is close to that of Jeelani



**Figure 4** Theoretical and experimental variation of the sedimenting and coalescing interface with time for a crude oil system—batch test (From Ref. 22.)

and Hartland (16-20) and looks at the transport of the dispersed phase through an interface. This philosophy has certain prerequisites, being:

1. The dispersed phase has to be appropriately destabilized (in order to make viscosity the only stabilizing factor)
2. The model includes the vertical transport of both continuous and dispersed phases.

The theory is developed for vertical separators, with the oil flowing upward, opposing the settling velocity of the dispersed water. This may explain the inclusion of oil flow into the relations.

The main feature of the theory is the prediction of the dispersion layer thickness  $h_d$  as a function of liquid flow rate and interface area:

$$\frac{Q_l}{A_{\text{interface}}} = \frac{h_d}{a + bh_d} \tag{20}$$

$$h_d = \frac{a Q_l}{A_{\text{interface}} + b Q_l}$$

here,  $a$  and  $b$  are constants depending on feed properties and operating conditions. These are determined by batch tests in the laboratory or in the field, where the decay of the dispersion band with time  $dh_d/dt$  is as follows:

$$\frac{dh_d}{dt} \Big|_{h_d} \equiv \frac{Q_l}{A_{\text{interface}}} \Big|_{h_d} \tag{21}$$

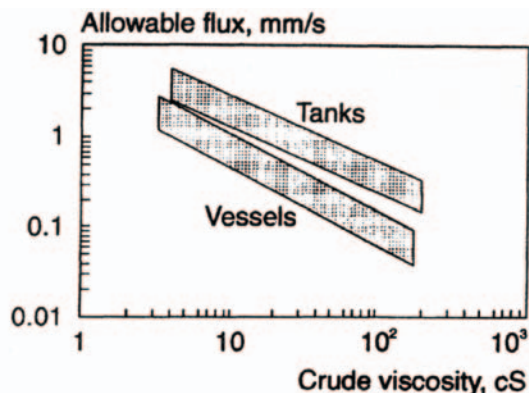
Combining Eqs (20) and (21) yields the relationship(22), giving  $a$  and  $b$  from batch tests for a given system:

$$\frac{h_d}{\left[ \frac{dh_d}{dt} \right]_{h_d}} = a + bh_d \tag{22}$$

Combined with experimental and field data, Polderman et al. (15) have deduced generalized design windows for destabilized crude oils, as shown in Fig. 5. For comparison, the applicable API specification (1) suggests a design basis as shown in Table 1.

### F. Brief Discussion of Models Presented in Sections IV.C, IV.D, and IV.E

The main differences between the different models presented are as to where the coalescence processes take place, and whether coalescence or settling is the limiting factor. Hafskjold et al. (21) discusses a system where rapid coalescence takes place, and attributes the separation characteris-



**Figure 5** Generalized design window for primary separators. (From Ref. 15, © Society of Petroleum Engineers.)

tics to binary coalescence (increasing the settling rate) and flow conditions. Jeelani and Hartland (20) neglect binary coalescence and focus on systems where coalescence is slow and takes place only at the interface. They distinguish between a packed layer formed by the settled drops where coalescence takes place and a dilute layer where settling takes place. Polderman et al. (15) have a combined view, suggesting that binary coalescence is important in the inlet tube upstream of the separator and at the inlet (or that the actual break-up is reduced by increasing the dispersed phase), and that interfacial coalescence is the main parameter within the separator, together with flow conditions. Our view is that all the mentioned interpretations are valuable, and that the different mechanisms will appear in varying degrees for different systems and conditions.

However, one important shortcoming is that the quality of the dispersed-phase effluent is not accounted for by any of the models. In real operation this parameter may also be limiting, e.g., in oil continuous systems the water quality is often the limiting factor. The instrumentation used in

batch tests may be inadequate for measuring dispersed-phase effluent quality, as the concentration levels are much lower in this phase. However, the various interpretations of continuous-phase behavior (in particular the degree of binary and bulk phase coalescence) would have an effect on the predicted dispersed-phase effluent quality.

## V. INTERNALS

Several internal components for improving the performance of gravity separators are available on the market. These address different problems that may occur within the process, and are here divided into foam handling, mist handling, flow distributing, and settling enhancing devices. Because of the costs associated with maintenance, the normal trend in design is to keep the number of internals at a minimum, and to maximize the simplicity of the process. It is therefore important to choose the correct configuration based on the actual problem at hand.

### A. Devices Used

#### 1. Foam Handling Devices

Mechanical foam handling is normally done at the inlet section, and seeks to separate the gas phase from the liquid phase by utilizing the inlet momentum. Two inlet types, traditionally regarded as efficient for foam handling, are the baffle type and the cyclonic inlets. Both types rely on a smooth reduction of the inlet momentum, to reduce the mixing energy at the inlet. The design of the baffles aims at using the difference in momentum between the gas and liquid phases by directing them separately into the separator and thereby avoiding foam generation. Cyclonic inlets force the incoming mixture to rotate, thus creating centrifugal forces enhancing the separation of the lighter gas phase from the liquids. The separated phases exit through separate outlets into the separator.

The obvious advantage of the cyclone type is that it physically separates the phases before the completion of momentum reduction; baffle-type inlets rely only on the capability to reduce the momentum smoothly. As such, it is difficult to design a baffle-type inlet that will perform satisfactorily for high loadings, especially if the mixture is highly susceptible to foaming. Cyclonic inlets may have a higher associated pressure drop than baffle-type inlets and/or occupy a larger volume. Note that these effects are strongly dependent on design.

**Table 1** Design Criteria for Liquid-Liquid Separation

Oil gravities ( $^{\circ}$ API)	Typical residence time, equal in both liquid phases (min)
Above 35	3–5
Below 35, $T > 100^{\circ}\text{F}$	5–10
Below 35, $80^{\circ}\text{F} < T < 100^{\circ}\text{F}$	10–20
Below 35, $60^{\circ}\text{F} < T < 80^{\circ}\text{F}$	20–30

Source: Ref. 1.

## 2. Mist Handling Devices

Mist handling devices seek to remove liquid drops from the gas phase to meet the specifications of downstream equipment. Demisters are typically based on impingement and centrifugal separation mechanisms.

Mesh pads and filters have the highest efficiency with respect to the drop sizes they are able to remove. They work on the impingement principle, guiding the gas through channels formed by the media, and making the liquid drops coalesce at solid surfaces. The pressure drop and efficiency are functions of the density of the mesh, and are often high. Unfortunately, mesh pads and filters have a low turndown ratio and are highly susceptible to fouling by liquid overload and clogging. The main drawback is, however, the onset of flooding which will occur at relatively low gas velocities (depending on system pressure). Flooding is characterized by a re-entrainment of the liquid resulting from high gas velocities through the mesh pad. This makes the mesh pads and filters relative large units compared to vane packs and axial flow cyclones since the gas velocity must be kept low.

Vane packs also operate on the impingement principle, but the channels are much wider, providing a pressure drop lower than that of mesh pads. These are, therefore, less susceptible to foaming, but do not have the same efficiency versus drop size (commercial claim: 10+  $\mu\text{m}$ , depending on design). They are also limited with respect to liquid loading. When it comes to flooding, the vane packs are less sensitive than the mesh pads. The liquid drainage is arranged as slots to guard the liquid drain from the gas flow.

Axial cyclones are flow-through devices that set the entering fluids in rotation and separate the liquid from the gas by centrifugal forces. The liquid forms a film on the wall, and is drained through slots and flows back to the bulk liquid phase through a downcomer. These units have a turndown ratio and efficiency versus drop size (commercial claim: 4+  $\mu\text{m}$ , depending on design) higher than those of vane packs. The pressure drop is the dimensional criterion for axial-flow cyclones. Axial-flow cyclones are characterized by a relatively high-flow throughput, which makes them the most compact alternative.

Finally, reversible-flow cyclones can be used for gas cleansing. These have the highest turndown ratio and good efficiency versus drop size (commercial claim: 3+  $\mu\text{m}$ , depending on design). They do, however, have a pressure drop higher than that of vane packs and axial flow cyclones.

Generally, separation efficiency is a function of pressure drop — increasing the available pressure drop improves separation (until flooding occurs for the vane packs and the mesh pads). Pressure drop is often critical for demisting devices. The liquid removed from the gas will be at a lower pressure than the gravity separator as the pressure drop is the driving force of the separation. To reintroduce the liquid into the gravity separator, the static height between the demister and the liquid surface in a downcomer pipe is utilized. If the pressure drop across the device becomes larger than the liquid static height in this pipe, there will be no liquid transport in the pipe and the liquid will follow the gas.

## 3. Flow Distributing Devices

Separation is normally a function of time, as either the settling or the coalescence process is limiting. It is, therefore, imperative that the flow through a separator is controlled in some manner. The design equations presented in [Sec. IV](#) are all based on an even distribution of flow over the cross-section of the separator. A skewed inlet distribution will lead to a distribution in residence times and impaired separation.

The conventional method to avoid uneven inlet velocity distributions is to divide the separator into compartments with baffle plates that provide a low pressure drop in the flow direction. The effect of this will be briefly discussed in [Sec. V.B](#).

## 4. Settling Enhancing Devices

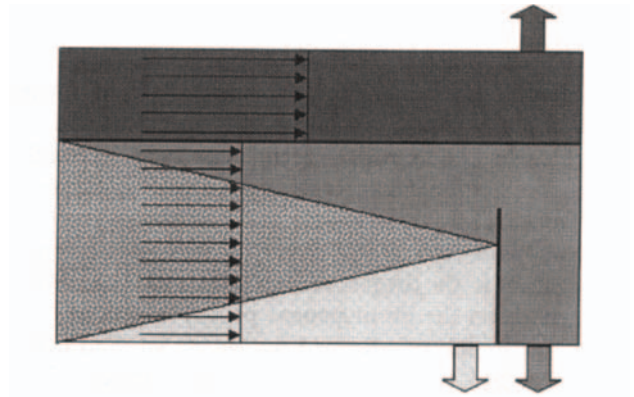
Settling enhancing devices seek to improve the settling process by providing channels with reduced height. This has two effects: reducing the diameter available for the flow reduces the Reynolds number and the turbulence, and shortening the vertical space reduces the residence time required for a drop to reach its bulk phase. The bulk phase in this respect is the film that is formed on the channel surface. The channels are usually formed by several plates, which are inclined to allow the liquid film to drain vertically. Meon and Blass (23) studied the performance of inclined plates, and found that performance variation was a function of plate inclination, drop size, and flow regime. Similar structures are also used for foam suppression if the inlet does not perform satisfactorily.

## B. Modeled Flow Patterns in Gravity Separators - Impact of Various Models and Internals

The flow pattern inside a gravity separator is complex owing to the simultaneous transport of multiple, heterogeneous bulk phases, mass transfer between these phases, and the impact of various internals. This section seeks to give a qualitative description of understanding flow patterns, in the light of the different effects of these results, by case descriptions.

The simplest mechanical design basis possible is a separator with a homogeneous inflow in which gas and liquid are separated, a weir plate to provide suitable safety criteria for the oil phase, and three outlets for the respective (clean) water, oil, and gas phases. The traditional simplified view is to assume plug flow in the liquid and gas phase, and slip between the liquid and gas, thus neglecting inlet effects and the possible slip between oil and water. Furthermore, Stokes' law is used for mass transport between the bulk phases (assuming rapid coalescence). This view is depicted in Fig. 6.

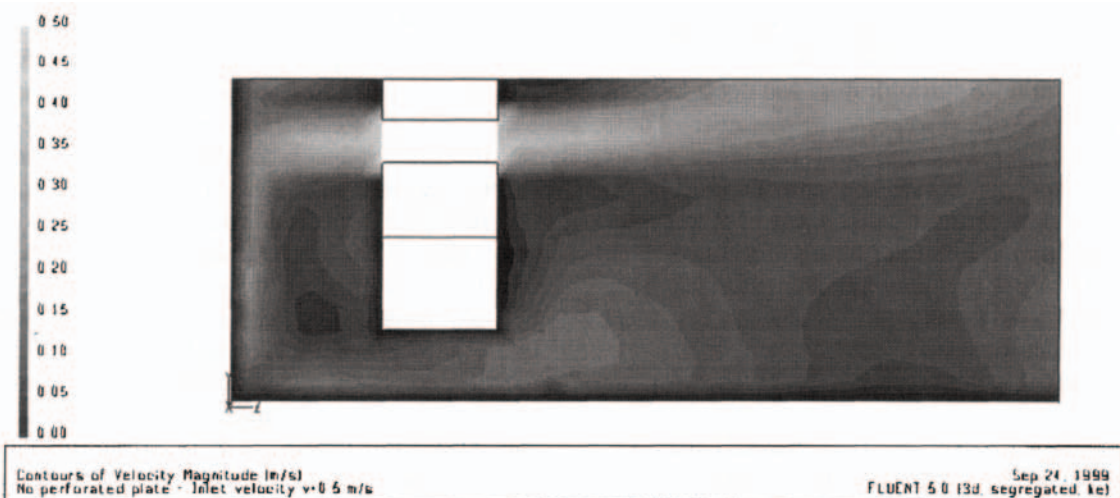
This situation is idealized and outlet driven, in that plug flow is initiated at the inlet in some manner. This is certainly not the case for any inlet reported. As inlets have to handle an incoming liquid momentum from the inlet nozzle, the incoming fluid will enter the vessel in a defined region. Depending on the length and diameter of the vessel, the flow pattern will be partly driven by the inlet and outlet boundaries of the flow, as imposed by the inlet and outlet geometries/velocities. The best tool to look at these effects



**Figure 6** Sketch of velocity profiles according to a plug-flow assumption, with cut-size calculation from Stokes' law.

is computational fluid dynamics (CFD) codes. Figures 7 and 8 shows the velocity profile and velocity vectors of a CFD single-phase, three-dimensional simulation of a separator with a cyclonic inlet (thus only the liquid phase is shown, as the gas phase is assumed to be separated by the inlet). Even though the velocities are greatly reduced by the removal of the gas, the velocity profile downstream of the inlet is far from the plug-flow assumption

In CFD codes, coalescence models have not yet been implemented and multiphase solutions should be used with great care. As the hydrodynamics often are controlled by gravity and coalescence as well as by momentum, the absence of coalescence models will affect CFD results, and consequently a complete quantitative evaluation of two-



**Figure 7** Velocity profile in a simulated gravity separator inlet liquid section; inlet velocity 0.5 m/s.

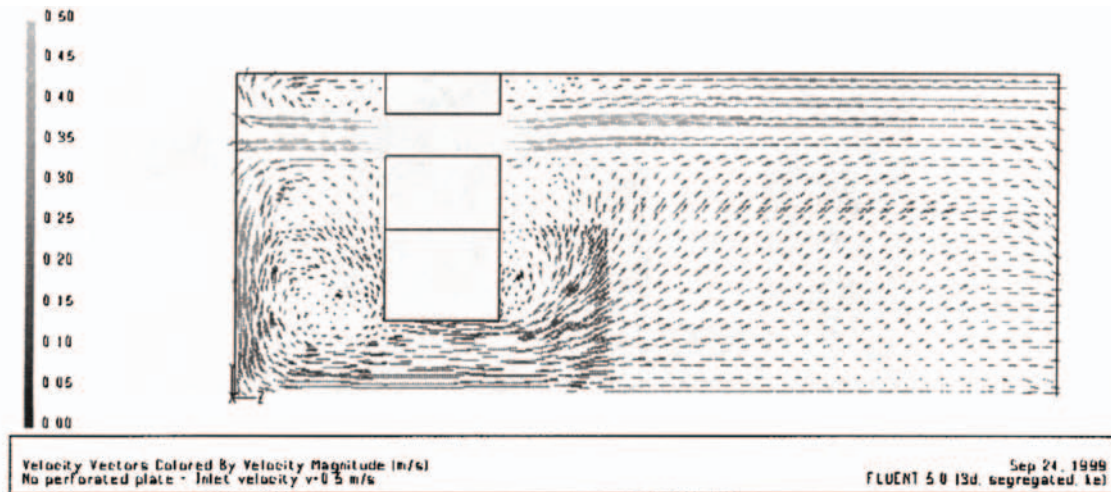


Figure 8 Cross-sectional views of the velocity profile for the simulation in Fig. 7.

phase flow through a gravity separator is still not within reach. However, sections of the separator are expected to be controlled by momentum (typically inlet and outlet regions), and CFD modeling of these parts can be expected to yield reasonable results.

The implementation of a flow distributor, such as a perforated plate, will greatly enhance the downstream conditions with regard to flow distribution. Figure 9 shows the same case as Figures 7 and 8, with a porous region resembling two baffle plates, each with 20% open area. The velocity distribution downstream of this region is rather unaffected by the inlet region, as can be seen by the figure.

This is still the case when doubling the inlet velocity, as shown in Figs 10 and 11.

Moving toward the impact of the models mentioned in Secs IV.D and IV.E, these will also have significance on the modeling of velocity and phase distribution. Assuming a case where settling is adequately completed within the residence time in the oil phase, a packed layer may form near the oil/water interface. This layer will have a dispersed-phase fraction of approximately 0.7 according to Jeelani and Hartland (20).

Assuming further that the outlet-driven flow will not give local velocities large enough to draw from this layer,

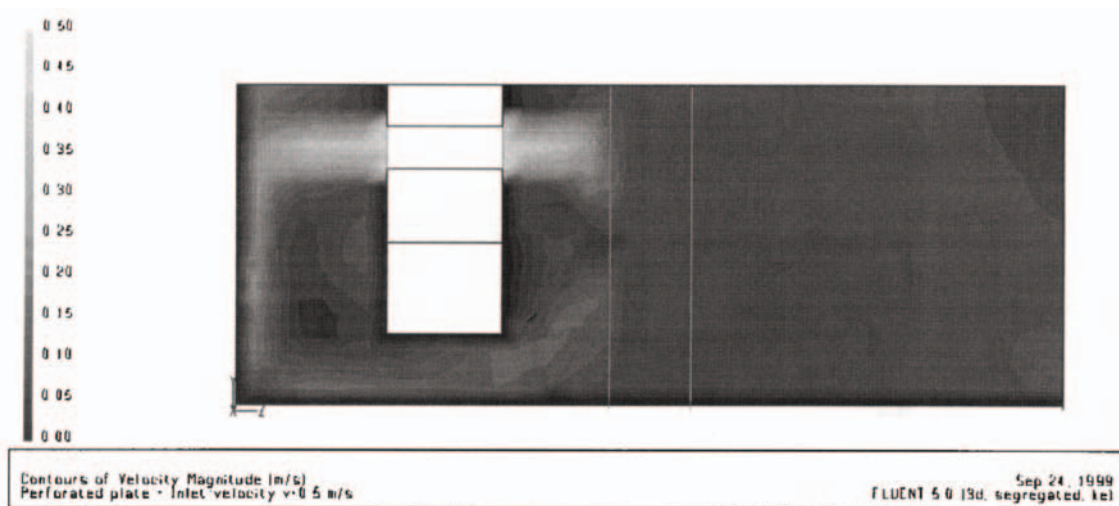


Figure 9 Simulation of the case in Fig. 7, with porous section.

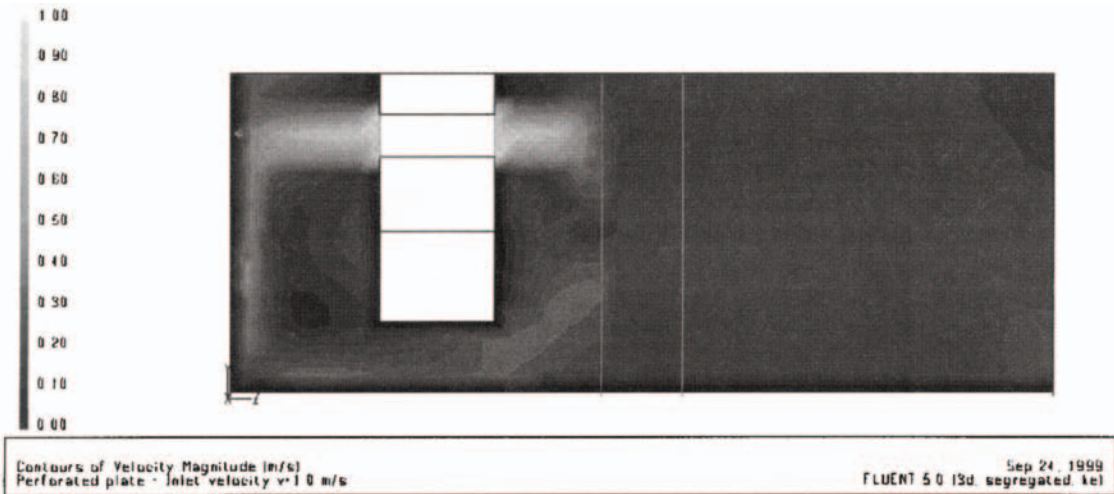


Figure 10 Simulation of the case in Fig. 7, with added porous section; inlet velocity is doubled (1 m/s).

the source term of this layer will be the incoming dispersed phase, and the removal term will be coalescence. As the velocities found by the plug-flow assumption are very low, the dynamic head of this layer will be several orders of magnitude lower than the (vertical) force induced by gravity. The packed layer will therefore remain between two fixed horizontal planes as determined by the interface control setting.

This suggests that the dispersed layer may be treated as if it were stagnant. The momentum transfer between the “pure” bulk oil and water phase will be greatly reduced, implying that they will have independent flow behavior and may be treated separately. Also, the residence time within

the packed layer will be independent of its horizontal velocity (which is assumed to be zero) and will vary only with the coalescence rate (which can be interpreted as a vertical velocity through the applicable interface). As the dispersed layer grows, it will eventually reach down to the region of suction from the water outlet or flow over the weir, and a step change in the respective outlet quality will occur.

## VI. DOWNSTREAM PROCESSING

The outlet specifications for a primary gravity separator are normally given by the downstream processing equipment,

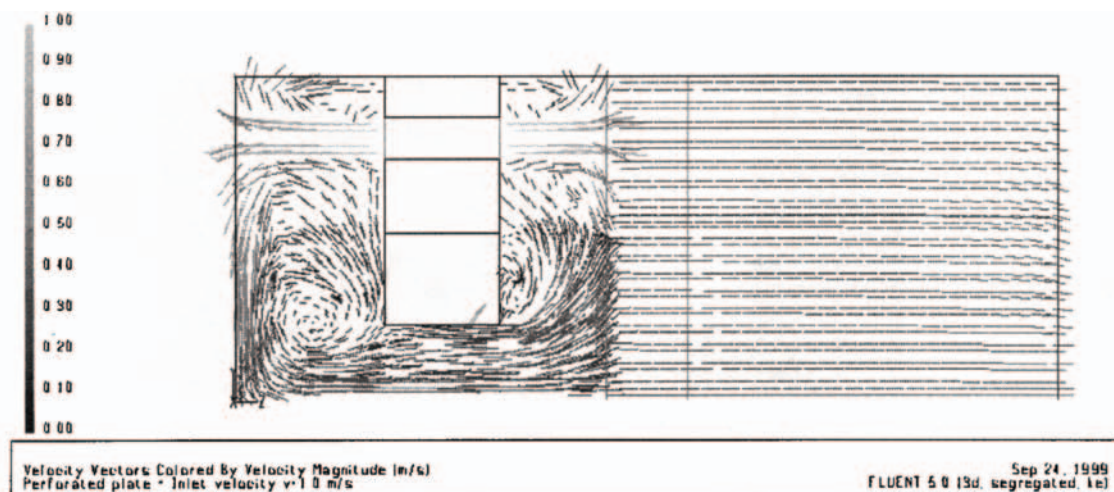


Figure 11 Cross-sectional views of the velocity profile for the simulation in Fig. 10.

and a brief review of these conditions is in order.

## A. Water Downstream Processing

The standard downstream processing equipment on the produced water side is hydrocyclones, also known as deoiler cyclones. Standard specifications for this type of equipment are often given as parts per million levels, and normal commercial specifications are inlet qualities no higher than 1000 ppm. This should give an outlet quality no higher than 40 ppm. It is appropriate to mention that these qualities are with respect to saturated hydrocarbons. Aromatic and polar groups are not included. As such, crude oils high in aromatic and polar components will give outlet values lower than those of paraffinic crudes, and this is also reflected in government regulations in different oil-producing regions—the requirements for dumping water quality are often more restrictive in regions with traditionally aromatic crude oils.

Deoiler hydrocyclones separate oil from water by inducing a strong centrifugal field, of the order of 500-1000 *g*. As such, their performance is not necessarily a function of inlet concentration, but rather of drop size distribution and continuous phase viscosity. A high concentration will increase the coalescence rate (by increased collision frequency) and therefore give a slight improvement in performance. It is, however, imperative that the droplets are protected from excessive shear, particularly as the concentration of dispersed phase is low (1000 ppm) and binary coalescence will not usually prevail. For example, deoiler efficiency will suffer greatly if centrifugal pumps are used for upstream pressure boosting.

The deoiler hydrocyclone is considered to be at a mature technology level, and little improvement is expected in mechanical design. Current research on improving the performance tends to focus on upstream chemical additives, such as flocculants and low-viscosity solvents.

## B. Oil Downstream Processing

Downstream processing of the oil is more complicated than for produced water, as cost-effective pressure reduction versus compression work has to be considered alongside oil/water separation. The objective of the downstream oil processing is to provide the necessary quality required for shipping/transporting, usually known as “stock tank oil” or “dead oil.” Normal requirements are less than 0.5 Wt %

BS&W (basic sediment and water) as defined by a standard method, such as that in Ref. 24.

Downstream oil processing normally consists of subsequent gravity separator units, possibly enhancing the last step with electrostatic equipment. The water separation efficiency of these steps will normally be much lower than for the primary separator, and higher residence times are often required.

## C. Gas Processing

The gas is processed for either reinjection or sale. In both cases, the gas phase has to be dried and compressed to the relevant pressure. As mentioned above, this is an important feature of the process as compressors are among the most expensive and mechanically complex equipment at a production facility. The downstream processing of the oil is therefore often associated with minimizing the compressor work, and the number of stages and pressures at each stage is determined by the gas/liquid equilibrium. Compressors are highly sensitive to liquid following the gas, and there is often a sharp focus on removing the liquid from the gas prior to it entering the compressors.

## VII. EMERGING TECHNOLOGY

There is a strong drive in offshore processing to reduce cost and weight. This is done by exchanging large and bulky components on a platform with more compact units, or by moving them closer to the well (subsea) or indeed into the well itself (downhole). These possibilities will be briefly discussed below.

### A. Compact Separation Units

Compact separation equipment seeks to reduce weight and size, demanding less of the costly support structure. This is done primarily by reducing the retention time needed for separation, by increasing the separating forces. Typical examples are cyclones and centrifuges, creating a centrifugal force field several times the magnitude of gravity by inducing rotational flow and radial mass transfer. Also, the utilization of turbulent coalescence and electric fields has been implemented in a compact electrostatic coalescer unit. Recently, equipment of this type has become commercially

available, customized for the different tasks needed. Typical compact components are:

1. Gas-liquid (liquid dominated) separation done by reverse-flow cyclonic devices operating at 5—20 g. These are already extensively used as inlets in ordinary gravity separators, to eliminate foaming.
2. Gas-liquid (gas dominated) separation done by axial- and reverse-flow cyclonic devices operating at 20-200 g. These are also used for polishing at gravity separator gas outlets.
3. Sand-liquid separation done by reverse-flow cyclonic devices operating at 10-500 g. These have been commercially available in the mining industry for a long time, and are extensively used for sand cleaning in the offshore industry
4. Oil/water (high oil content, water continuous) separation done by cyclonic devices operating at 50—200g.
5. Oil/water (ppm oil content, “deoilers”) separation done by cyclonic devices operating at 500—1000 g.
6. Oil/water (up to 10-20% water) separation done by compact coalescers (actually droplet growth promoters), reducing the size of downstream separators.
7. Oil/water (in principle any range) separation, done by centrifuges.

Items 1, 2, and 5 are units with a large number of applications within the oil-producing industry, while the others are at pilot unit level and will probably be installed commercially in the near future.

Cyclonic and centrifugal devices (all except item 6) work under the principle of setting a multiphase flow into rotational movement, thereby forcing the heavy phase radially outward and the light phase inward. For cyclones, the difference between reverse- and axial-flow mode is the mechanism of splitting the separated phases. The axial flow drains the heavy phase at the wall, while the reverse-flow types apply a back pressure forcing the light phase to exit countercurrently within the low-pressure core of the swirling flow. Reverse-flow types have a larger turndown, but also a larger pressure drop versus efficiency than that of axial-flow types.

All of these units have a large potential as individual components, but the major impact on cost/weight reduction would appear when using them throughout a truly compact process. The main concern of such a compact process is the lack of an accumulator volume, needed for safety during shutdown. The lack of an accumulator volume puts higher demand on the control system and makes the equipment more sensitive to variations in flow. Thus, the compactness

cannot be exploited fully unless these restrictions can be modified. In addition, the lack of compact equipment managing high-pressure, oil continuous oil/water separation is also a drawback. The two available compact units for this separation regime are the compact coalescer and the centrifuge - the compact coalescer has an upward limit in water content while the centrifuge will be sensitive to operating pressure and gas content of the oil. Several attempts at producing (static) cyclonic equipment for oil continuous flows have been reported, but so far it has not been possible to prove the general handling of oil continuous flows.

## B. Subsea and Downhole Separation

Another trend is to move processing equipment to the seabed, known as subsea processing. This requires reliability with regard to process control and stability/durability. In particular, sand handling is often a concern. The first subsea separation unit is due for installation this summer, at Norsk Hydro's Troll field (25).

Downhole separation is perhaps the hottest subject in separation today. Experiments suggest that the multiphase fluids are easier to separate at conditions normally present in the well without free gas, at high pressures and temperatures (26). Prototype tests with two-stage cyclonic devices and hydraulic pumps for offshore wells up to 20,000 barrels/day have recently been performed, with promising results (27). Also, a concept involving gravity mechanisms in horizontally drilled wells is being patented (28) and is due for onshore prototype testing in the near future.

The purpose of performing separation downhole is to increase the production rate to the platform, by removing the main part of the water for reinjection. The lowering of the liquid volume will enhance the effect of gas lift, by lowering the slope of dynamic pressure versus gas flow rate, and the reduced produced volume will also reduce the strain on the existing process equipment. Cyclonic separation will, as mentioned earlier, often be constrained by the demand of a water continuous flow.

## ACKNOWLEDGMENTS

Richard Arntzen would like to thank the Norwegian Research Council (NFR) and Kvaerner Process Systems (KPS) for a PhD grant. Per Arild Andresen would like to acknowledge the technology program Flucha financed by NFR and the oil industry. KPS is also thanked



for the kind permission to publish the material underlying the chapter.

## NOMENCLATURE

Symbols, with units where applicable, are shown in Table 2.

**Table 2** Nomenclature

Symbol	Unit	Definition
<i>A</i>	m <sup>2</sup>	Area
<i>C</i>		Constant
<i>d</i>	m	Droplet diameter
<i>D</i>	m	Pipe diameter
<i>f</i>		Friction factor
<i>g</i>	m/s <sup>2</sup>	Acceleration by gravity
<i>G</i>	J/mol	Gibbs free energy
<i>h</i>	m	Level height
<i>P</i>	Pa	Pressure
<i>Q</i>	m <sup>3</sup> /h	Volumetric flow rate
<i>r</i>	m	Radius
<i>t</i>	s	Time
<i>T</i>	°F	Temperature
<i>u</i>	m/s	Mean velocity
<i>v</i>	m/s	Terminal (vertical) velocity
<i>We</i>		Weber number
<i>ε</i>	m <sup>2</sup> /s <sup>3</sup>	Energy dissipation
<i>μ</i>	kg/(ms)	Kinematic viscosity
<i><math>\tilde{v}</math></i>	m/s	Turbulent fluctuation velocity
<i>ρ</i>	kg/m <sup>3</sup>	Density
<i>σ</i>	N/m	Interface tension
<i>τ</i>	s	Residence time
<i>φ</i>		Fraction of internal phase
<i>Ψ</i>	m/s	Coalescence rate

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Fremgangsmåte og anordning for separering av et fluid omfattende flere fluidkomponenter, fortrinnsvis separering av et brønnfluid i forbindelse med et rør for produksjon av hydrokarboner/vann. Berg, Norway: André, 1998, p. 25.