Conclusion

Column flotation has become the standard proven industrial flotation technique rather than an experimental method during the last decade. Nevertheless, its use in mineral-processing plants is mainly restricted at present to cleaning operations. The future of flotation equipment development lies in the combination of the advantages of impeller and column flotation and in the use of pneumatic machines in roughers.

As a greater share of flotation operations are used for unconventional areas such as environmental applications (water treatment, soil remediation, etc.) and ultrafine and colloid particle separation, special machines will be developed combining flotation attachment at intensive aeration and mixing conditions and three-phase separation in a quiescent environment. This leads to the concept of pre-aeration in a reactor (a unit for attachment of recovering phase on to gas bubbles) and de-aeration in a separator (a unit for separation of loaded bubbles from the bulk of three-phase suspension). Additional coarser bubbles can be added in a separator as a carrier to enhance the removal of loaded microbubbles by coalescence.

This concept and other types of new combined flotation machines will provide for more effective and efficient separation for a wide range of applications.

See also: **I / Flotation. II / Flotation:** Froth Processes and the Design of Column Flotation Cells.

Further Reading

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Column Flotation Cells

See **II / FLOTATION / Froth Processes and the Design of Column Flotation Cells**

Cyclones for Oil / Water Separations

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Synopsis

Though the solid–liquid hydrocyclone has been established for most of the 20th century, satisfactory liquid–liquid separation performance did not arrive until the 1980s. The offshore oil industry had a need for compact, robust and reliable equipment for removing finely divided contaminant oil from water. This need was satisfied by a significantly different type of hydrocyclone, which of course had no moving parts.

After explaining this need more fully and comparing it with solid–liquid cyclonic separation in mineral processing, the advantages that the hydrocyclone conferred over types of equipment installed earlier to meet the duty are given.

Separation performance assessment criteria are listed prior to discussing performance in terms of feed constitution, operator control and the energy required, i.e. the product of pressure drop and flowrate.

The environment for petroleum production sets some constraints for materials and this includes the problem of particulate erosion. Typical materials used are mentioned. Relative cost data for types of oil separation plant, both capital and recurrent, is outlined, though sources are sparse. Finally, some pointers to further development are described, as the oil industry looks to equipment installed on the sea bed or even at the bottom of the wellbore.

Introduction to Liquid–Liquid Hydrocyclones

This article covers the application of hydrocyclones to remove or concentrate dispersed oil from water. Two main classes of operation relating to their use with water-continuous liquid exist. Firstly as removers of oil contaminant from water (clean-up units) and secondly as a method of de-watering crude from wet oil fields (concentrator units). It excludes usage in relation to oil spills at sea, though feasibility studies have been technically successful. It also excludes applications where dispersed water (brine) is found in oil, though articles on this application – as yet only on the fringe of commercialization – may be found in the Hydrocyclone Conferences listed in the bibliography. Operation with oil-continuous liquid is difficult since interfacial effects are larger and break-up more likely as the brine droplets are less viscous than the continuous liquid.

Table 1 lists the key stages in arriving at the present near-universal usage of hydrocyclones for removal of oil contaminants from water in the oil industry offshore and more recently onshore. Concentrators, sometimes called 'dehydration hydrocyclones', are used in very wet oilfields as the initial stage to reduce the water (brine) content from, say, 95% to 50% or less. Under some conditions the overflow stream inverts to become oil continuous.

The two decades since de-oiling hydrocyclones were first seen to be capable of reaching legal standards of cleanliness offshore, e.g. $40\,\mathrm{mg\,kg^{-1}}$ maximum free oil on the UK Continental Shelf, are brief when compared with the much longer and widespread employment of solid–liquid hydrocyclones as in mineral processing or china clay production. Some salient comparisons between the two categories and the consequences of the clean-up or concentrator duty are set out in **Table 2**.

Flowfield and Geometry

The dispersion of fine oil droplets and their low differential density necessitate a high radial acceleration field. Since the oil is buoyant it will migrate towards the vortex core. To understand separation performance it is helpful to stress consequences of these two points. The oil is much more sensitive to the flow pattern than solid particles so this leads to the requirement for a low turbulence, reasonably linear vortex core and low peak shear flowfield to avoid droplet break-up leading to lower oil removal efficiencies.

The overflow stream should be a small fraction of the feed, since the oil content in clean-up applications is typically below 1%. An approximate volumetric balance gives the interrelationship between the parameters of a de-contaminating hydrocyclone. If the volumetric feed concentration of oil $C_f \sim 1\%$ and the overflow and underflow concentrations C_o, C_u are assumed to be 50% and 0 respectively, then the underflow rate Q_u will be 98% of the feed rate Q_f. For a wet oil concentrator however assuming $C_f \sim 10\%$ Q_u will be $\sim 80\%$ of Q_f. The overflow (reject) stream flow rate clearly is $Q_o = Q_f - Q_u$.

Colman and Thew at the University of Southampton in the late 1970s found that a cone angle as small as $1\text{–}1\frac{1}{2}^\circ$ (total angle) produced the necessary stable vortex, with a small relatively fast moving core moving to the overflow. This reverse axial flow penetrated a long distance downstream, so that in the early work a cylindrical tailpipe was put on the end of the cone. They also found that an enlarged entry section, as illustrated on **Figure 1**, reduced the pressure drop and reduced peak shear; the left-hand side of Figure 1 shows the approximate proportions of the

Table 1 Advances in the use of oil–water hydrocyclones

1950s and 1960s	Sporadic work on liquid–liquid hydrocyclones, especially in relation to use in the atomic energy industry
1965	Bradley produces his classic text on 'Hydrocyclones' clarifying the problems of liquid–liquid separation
1974	Kimber and Thew achieve 90% oil separation at Southampton University, UK
1978–80	The Southampton Group achieve $>99\%$ separation with crude oil (Colman and Thew, 1980 Hydrocyclone conference)
1983–84	First field trials offshore
1985	First large offshore installation ($15\,\mathrm{m^3\,min^{-1}}$)
Late 1980s	Installations worldwide
1990s	Virtually only method of oil–water separation. Number of manufacturers grows, prices fall. Higher mechanical packing density in pressure vessels
Early 1990s	Use in concentrator mode begins
Late 1990s	'Downhole' trials begin

Table 2 Comparisons between solid–liquid and oil–water hydrocyclones

Factor	*Solid–liquid*	*Oil–water*
Differential density	Often water–quartz, 1650 kg m^{-3}	50–300 kg m^{-3}
Split ratio (Q_U/Q_F)	Fixed; set by outlet orifice size	Controllable by external valves
Outlets	Usually one or both open to atmosphere	Closed system
Axial pressure gradient near the centre line	Usually very small (air core)	Substantial, overflow at lower pressure than the underflow
Pressure drop	1–5 bar	5–20 bar
Inlet pressure	1–5 barg	10–10^2 barg
Shear	May cause some size reduction due to particle-particle interaction	Likely to cause droplet break-up
Particle size	1 μm–10 mm	Usually droplets $\sim 1\,\mu$m–100 μm
Concentration (by volume)	Varies greatly; can be slurry at underflow	Typical oil contaminant concentration in feed to first stage $< 0.1\%$, oil contaminant concentration in feed to second clean-up stage 5–10%, well-head oil concentration in feed to wet oil concentrator 5–50%
Orientation	Fixed, 'g' important	No limitation, lateral acceleration up to ~ 0.1 'g' on floaters unimportant

Southampton bi-cone design, which had twin tangential inlets to produce a linear core. The right-hand side illustrates a typical later development with a single 180° wrap-round involute inlet and a curved wall. Note the absence of a projecting vortex finder, since there is no loss of oil in any short circuit flow in the end wall boundary layer.

Both the clean-up and concentrator units use the same wall profile but the latter has a larger overflow (or reject) port.

Later systematic work by Young (1994) came up with similar proportions.

Stronger swirl increases the acceleration field, but also raises the pressure drop and too much swirl seems to increase vortex instability and shear. Using a swirl number S ($S = \pi D_R X_i (2A_i)^{-1}$, where A_i is the total inlet area measured at a point where the flow is normal to the radius from the hydrocyclone axis to the centroid of the area, X_i, and D_R is a reference diameter of the hydrocyclone), the range of values for *de-oilers* is typically 7–11 or more usually 8–10. The reference value D_R for the bi-cone design is at the junction of the two cone angles (see Figure 1), and for a curved wall at, say, the point where the tangent is at about 10° to the cyclone axis.

The axial pressure gradient near the centre line is *not* zero: see section on pressure drop later.

De-oiling hydrocyclones have shown a range of sizes but there are no large units as for solid–liquid separators, since large droplets do not exist and thus no requirement for units with a relatively small acceleration field. Size is a compromise, taking into account the factors shown in **Table 3**. Early installations showed D_R values rising from about 30 mm to 70 mm, but more recently this has tended to drop back to about 15–30 mm.

Installation

Unlike solid–liquid units, clean-up or concentrator hydrocyclones have hitherto almost always been installed in pressure vessels, as shown schematically in **Figure 2**. This allows easy fabrication of individual units (sometimes called liners) from relatively thin walled material and reduces the number of connections to be made. Several assemblies complete with instrumentation and controls are built into a skid. Some larger units have serious vibration at very high flows so that a mid-length damper is built in.

The overflow or reject stream, being a small proportion of the feed, is readily incorporated into a manifold which may double up as a mounting plate.

Though series operation is possible and has been laboratory proven, in the field the use of a single stage with many units in parallel has hitherto been worldwide. Even though the turn-down ratio (comparison of maximum and minimum usable flowrates) for an individual unit is limited – see later discussion of the influence of feed flowrate on separation – the installation of units in banks which can be valved out in turn, allows overall turn-down ratio of about four for two banks, eight for three banks and so on.

Increased cleaned water throughput of an existing installation can be achieved by the addition of more hydrocyclone units.

Apart from modularity and improved separation, clean-up hydrocyclones have other advantages over the equipment formerly used such as induced gas

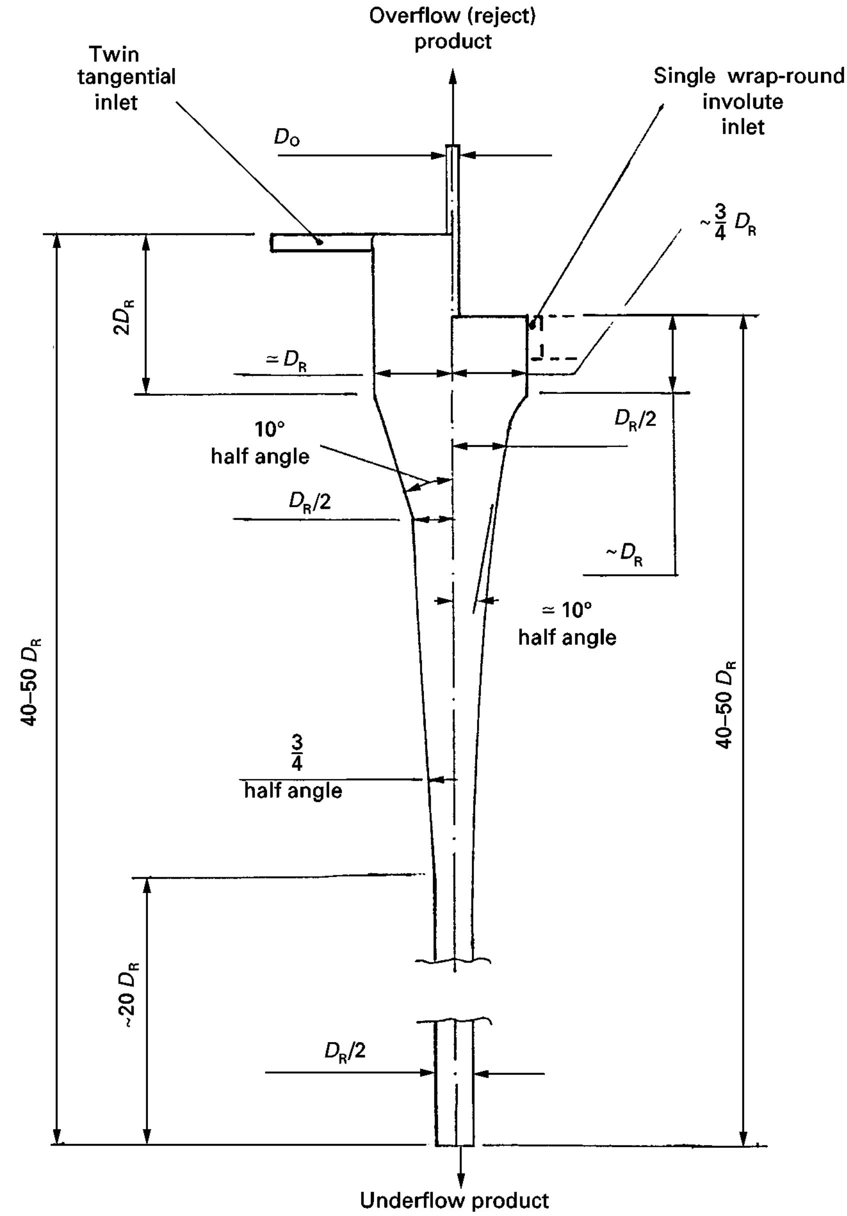

Figure 1 Hydrocyclone proportions. (left) Southampton University bi-cone design; (right) Typical later industrial development.

flotation (IGF) plant or gravity separators with plate packs. The installed size and weight of hydrocyclones has only been found to be about 10% of the older plant. Since hydrocyclones have negligible free surface effects they are insensitive to orientation and the motion of floating equipment. Unlike IGF they have no large requirement for chemicals and their cleaned underflow water does not require subsequent processing apart from occasional flocculant addition to reduce oil content below the legal limit.

As the unit residence time is only 1–2 s the hydrocyclones cannot cope with a slug of oil. Any zones of high shear upstream from these de-oiling hydrocyclones will reduce the size of oil droplets, thus worsening the separation efficiency. New installations can largely alleviate this problem by re-design to re-position flow-control valves to come after a hydrocyclone rather than upstream from it.

Performance

Though variables will often interact, for convenience their effects are discussed separately.

Separation Efficiencies

To achieve the purity required of the underflow stream from a clean-up unit, its output rate is usually

Table 3 Factors influencing the size of de-oiling hydrocyclones

Factor	*Influence*
Flowfield Reynolds Number, $u_t D_i / \nu$	In early development its perceived influence on separation suggested larger units. Now ignored
Pressure drop	Larger units have higher pressure drop, for the same peak radial acceleration field. Even though the reservoir pressure is often high enough to remove the need for pumps this factor tends to limit size
Separation of smaller drops*	As the hydrocyclone size reduces to a D_R of approximately 20 mm improved separation efficiency of droplets smaller than about 15 μm is achieved
Avoidance of excessive shear	For D_R values below about 10 mm experimental observation suggests that high shear causes droplet break-up and consequent loss of separation efficiency
Cost for a given flow rate	Favours few, larger units if cost for unit alone is considered
Packing density*	Units are normally installed in a pressure vessel. Higher overall throughputs can be achieved for a fixed volume pressure vessel fitted with many smaller rather than fewer larger units. This favours reduced cost for complete assembly.

*Generally dominant factors in current practice.

restricted to about 98% of the feed stream rate. This can result in the overflow reject stream having a significant water content; fortunately the oil droplets in this stream coalesce readily causing the oil and water phases to separate easily. The oil may then be pumped into an oil storage vessel or delivery line.

The inability of hydrocyclones to achieve perfect phase separation in a single step has previously been noted by Bradley (1965).

For an oil-water hydrocyclone operating in the concentrator mode, the yield of oil in the overflow product stream is the principal criterion.

Definitions for purity and yield follow below.

The performance of any separation unit is usually defined in terms of the fractional recovery η of the valuable product and its purity in the product stream ε. For the clean-up units the fractional recovery η_0 is however defined by the fraction of oil in the feed which appears in the reject (overflow) stream, and ε_u for this operation is defined in terms of the oil (waste) content of the product (underflow) stream:

$$\eta_0 = \frac{Q_o \times C_o}{Q_f \times C_f}$$

and

$$\varepsilon_u = 1 - \frac{C_u}{C_f}$$

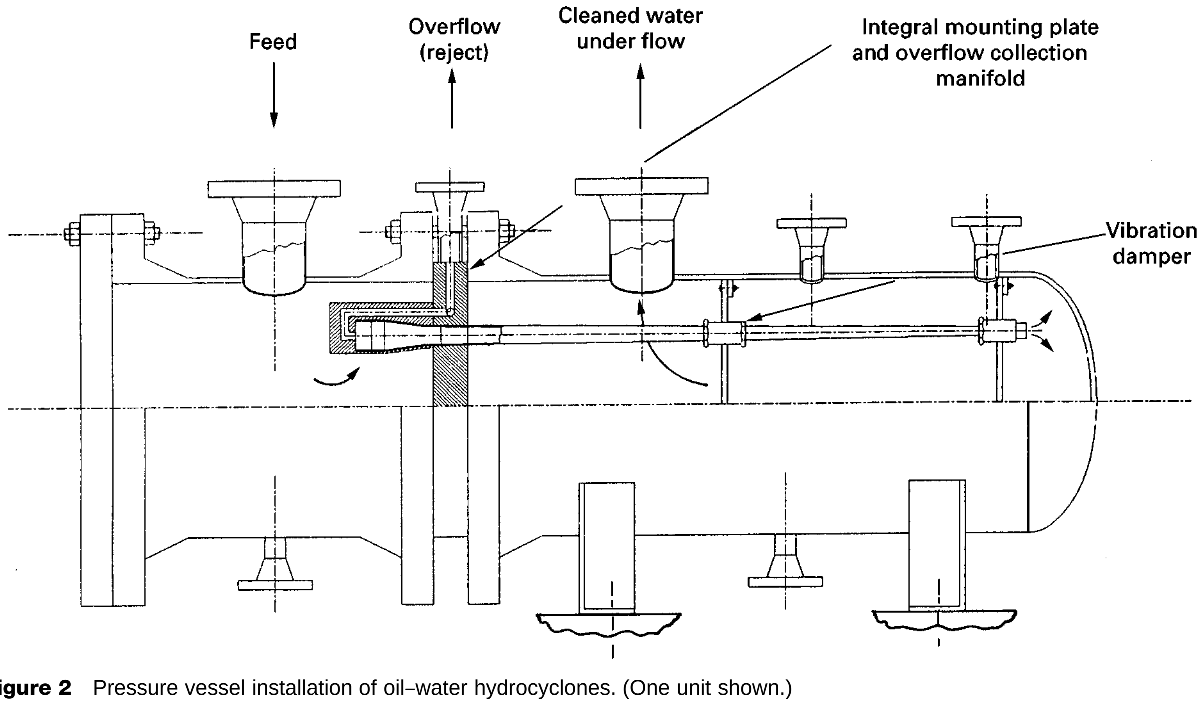

Figure 2 Pressure vessel installation of oil–water hydrocyclones. (One unit shown.)

if the split ratio R_f and its complement F are defined by $R_f = Q_u/Q_f$ and $F = Q_o/Q_f$.

Then a relationship between the fractional recovery parameter η_0 and the underflow purity ε_u follows:

$$\eta_0 = 1 - \frac{Q_u \times C_u}{Q_f \times C_f}$$

thus $\eta_0 = 1 - R_f \times (1 - \varepsilon_u)$.

One bonus for the de-oiling duty is that the produced water, i.e. brine, is often warm. Higher temperature is beneficial as the water viscosity is reduced and the differential density increased. Possible problems arising solely from the elevated temperature, with reduced interfacial tension and easier distortion of oil drops as they are less viscous, seem unimportant. Thus values of ε_u of 0.99 or even 0.999 have been obtained, even though the differential density may only be $100\,\mathrm{kg\,m^{-3}}$ with a mean drop size 20–30 μm.

Generalized prediction of ε_u and η_0 is very difficult. Apart from the practical difficulty of predicting the feed drop size distribution, droplet break-up is influenced by interfacial tension. In the complex liquid mixtures of crude oil and produced water, interfacial tension is influenced by surface-active agents whose presence and effects are difficult to determine.

Correlation of experimental results could ideally use the Stokes Number ($\mathrm{St} = 2Q_f\Delta\rho d^2/\pi 9 D_R^3\,\mu$, where $\Delta\rho$ = differential density, d = characteristic droplet diameter, μ = viscosity of the continuous fluid). Strictly, this only applies to dilute dispersions with Stokesian flow with a droplet at a Reynolds Number below unity. The major problem which tends to restrict usage to laboratory investigations is in the determination of d. This is usually taken as the d_{50} diameter of a droplet that has an equal chance of reporting to underflow or overflow, but in most applications it is impractical to measure it.

Influence of Feed Characteristics

Flow rate At low flow rates the tangential velocity is too low to generate an adequate inward radial acceleration. This is reflected in the field results of Meldrum *et al.* in 1987 which have been converted from tabular results to the plot in **Figure 3**. These show sharply decreasing values of ε_u for a 60 mm D_R hydrocyclone on the Hutton field and for a 35 mm D_R hydrocyclone on the Murchison field for feed flows below $\sim 100\,\mathrm{L\,min^{-1}}$ and $60\,\mathrm{L\,min^{-1}}$, respectively.

Figure 3 also shows rapid falls in ε_u values for the 35 mm Murchison unit for feed flows above $\sim 200\,\mathrm{L\,min^{-1}}$ and for the 60 mm Hutton unit above $\sim 425\,\mathrm{L\ min^{-1}}$, which reflects a split ratio falling below the critical value. This is discussed fully later but essentially as flow rate rises the outlet pressures fall, and as the overflow outlet is at the lower pressure, a stage is reached where the overflow stream is so diminished that oil can only leave via the underflow.

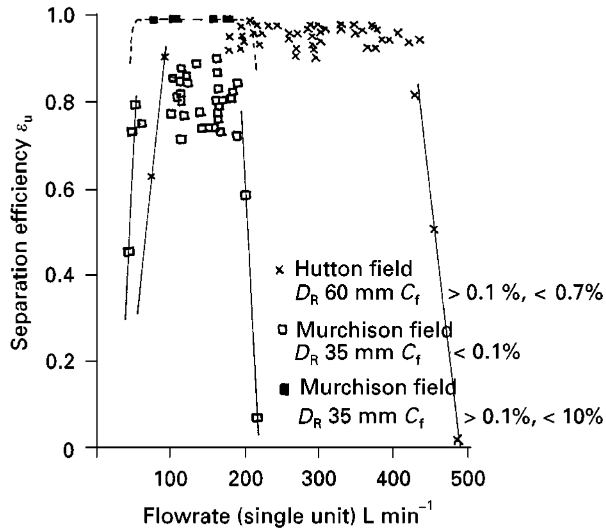

Figure 3 Field results: separation efficiency (ε_u) vs flow rate (Q_f). Calculated from Meldrum *et al.* data.

The turn-down ratios illustrated by Figure 3 are 3 for the Murchison unit operating on a feed concentration C_f < 0.1% and 4.5 for the Hutton unit, but other fields with a larger driving pressure available have achieved values up to about 7. Pumped installations or oilfields with lower reservoir pressures might fall to about 2, for a single unit.

A plateau region separation efficiency is usual. Higher flow rates raise the acceleration field but,residence time falls, turbulent re-mixing rises and if interfacial tension is low droplet break-up may become significant. Within the plateau region, rapid transients in flow rate should not reduce separation efficiency appreciably.

Oil–water ratio Over a wide range of oil concentrations clean-up and concentrator hydrocyclones are considered to act as a flow divider, i.e. ε_u remains constant and C_u will rise in proportion to C_f if droplet size remains constant, however field results show this to give a pessimistic view as droplet sizes rise with oil content. This means that C_u may even remain constant as C_f rises.

As C_f rises the split ratio may need to be raised also. For clean-up the value of F is often maintained at 2–3%, but a rational control strategy is discussed later. In concentrator operations when C_f rises to

Table 4 Transient oil separation performance

d_f μm	ε_u Steady-state	% drop (ε_u steady state − ε_u transient) 5 s injection	2 s injection
50	0.93	0	1
17	0.53	4	6

d_f = mean feed drop size, BP Forties oil in cold tap water, C_f steady state ~500 mg kg^{-1}.

5–10% the ratio F/C_f may be conservatively maintained at 2. This means the overflow stream is 50% water; for it to be, say, 70% oil (favourable result) F/C_f falls to about 1.4. A moderate transient increase in oil content, provided the F/C_f ratio remains satisfactory, has very little effect on ε_u. **Table 4** shows results for a short-lived oil pulse, obtained with on-line oil content measurements compared with steady-state results.

A larger increase in C_f, but still within the acceptable range, may show a rise in ε_u if droplet size has also risen. Any further rise in C_f will exceed the capacity of the overflow and ε_u will fall substantially when F/C_f falls below about 1.2.

Droplet size Like all separation devices based on differential density, in a hydrocyclone, reduction in droplet size will give poorer results. An acceptable minimum value for d_f under favourable conditions may be as low as 5–10 μm (elevated temperature, $\Delta\rho > 150$ kg m^{-2}, adequate interfacial tension), however the usual acceptable minimum of the feed droplets is 15–20 μm.

Droplets of 20 μm are less vulnerable to low values of interfacial tension in promoting break-up than larger ones. However, the presence of surfactants that drastically lower interfacial tension will almost certainly reduce the effectiveness of hydrocyclone separators, as shown by Colman, Thew and Corney at the First Hydrocyclone Conference (Cambridge, UK) in 1980.

In laboratory work, grade efficiency curves d_f vs ε_u have been obtained as for solid–liquid hydrocyclones, but because of the considerable difficulties in obtaining representative samples and in measurement such curves are seldom available in the field. When sampling, both isokinetic conditions and avoidance of droplet break-up are necessary and gas bubbles may complicate interpretation. A suitable technique was discussed by Colman, Thew and Corney.

Free gas Until fairly recently most clean-up hydrocyclones were installed downstream from three-phase separators. This means that the free gas content of the hydrocyclone feed was a fairly minor constituent. As a result any gas core was relatively small in diameter. Because of the significant axial pressure gradient at the centre line this gas leaves with the overflow (oil-rich) stream.

Provided the gas content of the feed is reasonably invariant with time, laboratory tests have demonstrated that oil separation is little affected up to a threshold of 20–30% by volume free gas. This figure relates to conditions at the hydrocyclone entry. Field experience has generally confirmed this satisfactory picture except when the gas flow exhibits significant slugging. An entering gas slug does not have the angular momentum to maintain rotation of liquid and with the breakdown of inward radial acceleration, separation performance falls sharply. Amongst the thousands of units in service such a loss of performance is uncommon.

One consequence of appreciable gas leaving via the overflow is that it reduces the area for liquid to an annulus in the overflow exit port. This changes the relationship between the pressure drop and flow rate for the overflow liquid thus adversely affecting the control.

With the pressure drop in the hydrocyclone, some evolution of dissolved gas would be expected. This does occur but is too slow to be appreciable within the hydrocyclone and is manifest downstream from it, being most noticeable in the overflow stream. The evolving gas has been used to achieve post-cyclone separation of some more very fine drops in a suitable vessel possibly because the evolving gas bubbles nucleate on the oil droplets. In terms of Henry's Law the mass of gas coming out of solution in the overflow, will be proportional to $Q_o \times H\ (p_f - p_o)$ where p_f is the upstream feed pressure, p_o is the downstream overflow pressure and H is Henry's constant. However, the volume evolved will also depend inversely on the absolute pressure and in any case Henry's Law gives a maximum value as it relates to equilibrium conditions.

Solid particles Crude oil–brine mixtures commonly contain small amounts of reservoir solids. This problem is growing as more fields have larger produced water contents and it is made worse by the trend to produce from unconsolidated formations. The amount may range from a few hundred mg L^{-1} to about 10 g L^{-1} in worst cases. The solids may be water-wetted or oil-wetted, both usually report to the underflow unless the oil-wetted solids are very fine, in which case they may tend to be neutrally buoyant. In practice the overflow stream very seldom contains solids. Overflow blockages are rare and when they do occur they are usually associated with debris left in the system at installation or after maintenance.

Erosion, if it occurs, is usually restricted to the inlet region where velocities are higher. The use of harder materials, for example Stellite, in the inlet region has allowed long periods of satisfactory operation. A number of installations have run continuously for five years or more.

A development of the last 2–3 years is the arrival of de-sanding hydrocyclones installed ahead of the de-oiling units. These have been used prior to the choke with a containment vessel to withstand the very high pressure. Both relatively large units in appropriate steels and smaller ceramic units are entering service.

In off-shore fields the water in the oil–water mixtures may be quite corrosive, particularly if it is sour (containing H_2S). This has necessitated the use of alloy steels in the fabrication of the clean-up hydrocyclones. However, particularly for low cost, low flow land-based installations, cheaper materials may be satisfactory. Polyurethane units, possibly in a carbon steel casing or even bare are on the threshold of commercial usage.

Pressure Drop (the Cost Implications for Separation)

As in solid–liquid hydrocyclones, pressure drop $\Delta p \sim Q_f^m$ where $2.1 < m < 2.2$. In dimensionless terms defining:

$$\text{Euler Number } N_{Eu} = \frac{\Delta p}{\frac{1}{2}\rho u_f^2}$$

$$\text{Reynolds Number } N_{Re} = \frac{u_f D_i}{v}$$

where $u_f = Q_f/A_i$ and the inlet port diameter $= D_i$ (for multiple or noncircular inlets D_i is the diameter of the circular port with equal A_i).

Leads to the dimensionless relationship

$$N_{Eu} \sim N_{Re}^n$$

where

$$0.1 < n < 0.2$$

The use of u_f in terms of Q_f and A_i will be reflected by changes in the swirl number S.

Values of ρ and v are based on arithmetic averages of the mixed liquids. Uncertainty in the value of the kinematic viscosity v is not serious due to the small value of n. Typical Euler Number values are in the range 10–20 and do not seem to be affected by variations in C_f.

The de-oiling hydrocyclones have a substantial pressure gradient along the hydrocyclone axis not present in solid–liquid hydrocyclones with an air core. Thus pressure drop between the feed and the overflow Δp_{fo} is greater than the pressure drop between the feed and the underflow Δp_{fu}, the relationship between the two being variable and set by external valves. The ratio between the two is important for control and optimization and this pressure drop ratio (PDR $= \Delta p_{fo}/\Delta p_{fu}$) varies with split ratio, R_f and also its complement F. Δp_{fo} has two principal components – one due to the radial pressure gradient and the other arising from the velocity through the overflow port. The radial pressure gradient and Δp_{fu} are both proportional to the u_f^2 and S, since $u_f = Q_f/A_i$. For a fixed geometry the PDR $= B_1 + B_2 \times F^2$ where B_1 and B_2 are constants, where $B_1 = f_1(\text{S})$ and $B_2 = f_2(\text{S}, A_0^{-1})$. The oil concentrated overflow rate is regulated by using a fixed PDR value as set-point.

Pumped installations In younger oilfields the driving pressure stems from reservoir pressure. In older fields where pumping is necessary, this represents a cost. To reduce droplet break-up a low shear positive displacement pump should be used, however centrifugal pumps have proved satisfactory provided their speed is not too high and the duty is not too far from their best efficiency point. Fields with electrical submersible pumps have utilized de-oiling hydrocyclones satisfactorily.

Operator Control

For water clean-up units the PDR set-point is usually in the range 2–3 but for concentrators, with their larger overflows, values below 1.5 may be set. A simplified control layout is shown on **Figure 4**. The effectiveness of the control is restricted to some extent by the relative sizes of the underflow and overflow exit ports.

The PDR is a weak function of the feed Reynolds Number ($N_{Re} = u_f D_i/v$), see **Figure 5**. This dependence is usually ignored.

Critical split ratio (F_{CRIT}) At low oil contents it is desirable to reduce F thereby reducing the amount of the overflow stream. But below a critical value F_{CRIT} the axial pressure gradient near the centre line is inadequate to sustain the reverse flow. Though oil still gathers in the vortex core it is unable to reach the overflow outlet, becomes remixed near the inlet and leaves via the underflow so ε_u falls sharply. The effect is illustrated on **Figure 6**. F_{CRIT} rises with an increase in the diameter D_o of the overflow port and is therefore higher for concentrator units. Though extremely small overflow outlets will permit very small F_{CRIT} values, they are impractical as the smallest upset is too much for them to handle.

Table 5 summarizes the factors to be considered in control of the split ratio via the easily measured

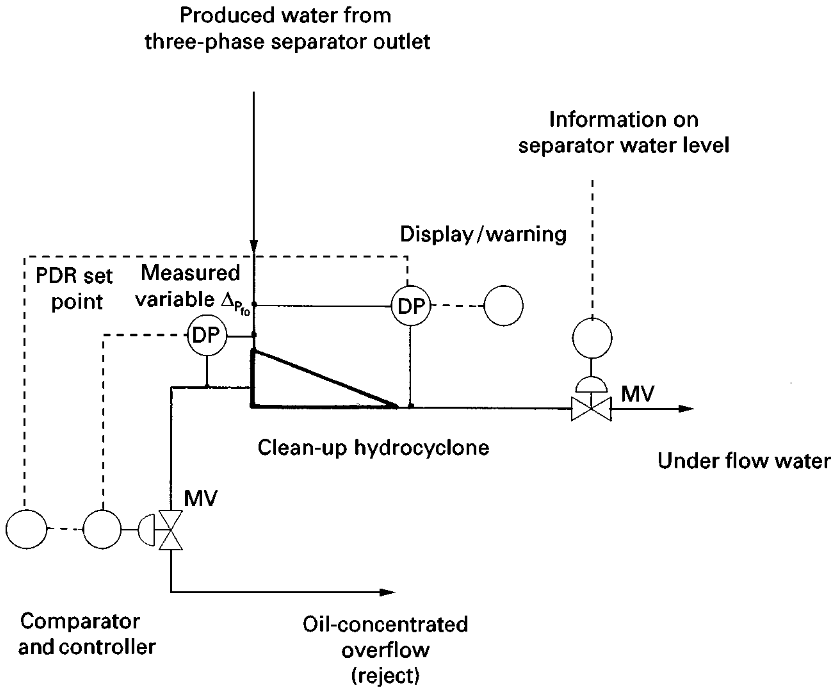

Figure 4 Simplified control layout.

variable, PDR. Hydrocyclone manufacturers will advise on the size of the overflow outlet. It may require enlargement during the life of an oilfield.

Cost Comparisons

Cost data are sparse but some information produced by BP about five years ago has been recast on **Table 6** in terms of ratios. The flow rate used for deriving the information was applicable to a field producing about 16 000 m^3 per day of oil.

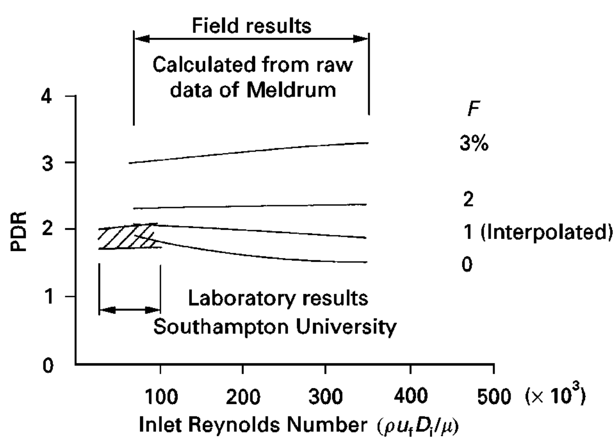

Figure 5 Pressure drop ratio (PDR) vs inlet Reynolds Number. Southampton University laboratory results and field results deduced from the data of Meldrum *et al.*

Table 6 shows that both the capital and running costs increase in the sequence hydrocyclone-IGF-filter/coalescer-centrifuge. Hydrocyclones also produce the most compact plant and are the least sensitive to orientation or lateral acceleration as encountered in installations on a floating base.

Future Developments

In 1979 an oil industry task force recommended plate pack gravity separators or IGF for produced water clean-up. Six years later the first large hydrocyclone installation – about 15 m^3 min^{-1} – was operating successfully in the North Sea, so the points below relate only to the immediate future, perhaps prior to 2003.

- Improving oil separation (in hydrocyclones) means removing smaller drops. There is probably limited scope for further optimizing geometry, though the claims for computational fluid mechanics (CFD) in rapid optimization are likely to become valid in a year or two. (Adequate representation of turbulence in confined swirling flow has proved difficult.)
- For new systems and even for some retro-fits, there is often room for worthwhile improvement by reducing shear upstream of the hydrocyclones or re-locating them. The resulting larger drops ease the separation task.

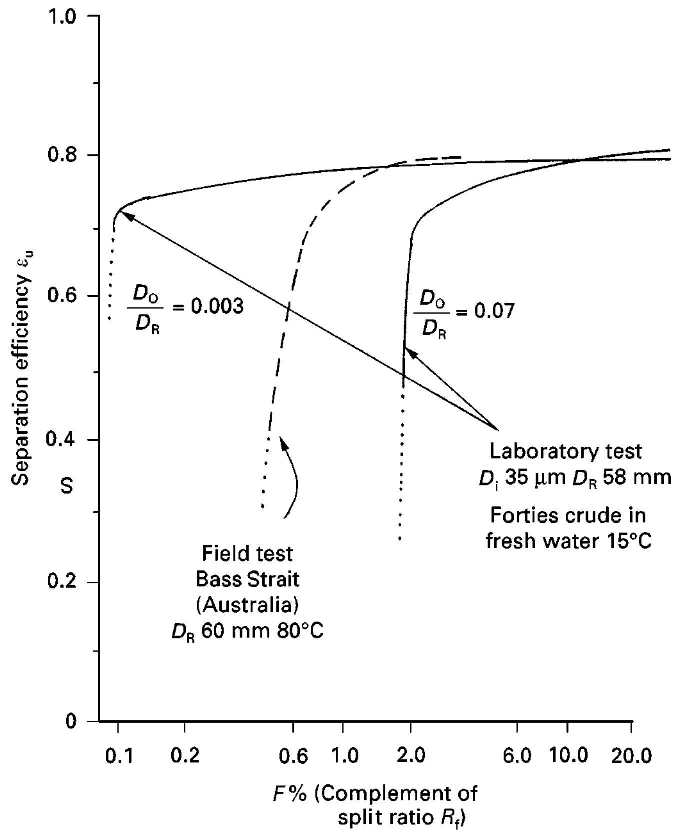

Figure 6 Separation efficiency (ε_u) vs complement split ratio (F). The collapse in separation consequent on too low an overflow is shown.

- If the split ratio needs frequent adjustment, a variable area overflow (two values would be adequate) simply controlled would be very useful. Laboratory tests have shown that this is readily feasible but moving parts are still not readily accepted, as they are associated with reduced reliability.
- As the cost of dealing with produced water mounts, economic/technical studies have strongly suggested hydrocyclone installation at the bottom of the wellbore, 'downhole'. If the geology is suitable the water could be re-injected in adjacent strata. Preliminary trials using electric submersible pumps (ESP) (some concepts suggest two) have shown and are showing considerable promise. Access after installation is costly so there are problems in control and reliability, though ESP speed variation is proven. Reduced standards of separation may still be acceptable. A halfway stage to use pre-choke hydrocyclones on the sea bed is also being investigated.

Table 5 Factors affecting PDR

Factor	*Correction to PDR*	*Comment*
$F < F_{CRIT}$	Increase*	Oil is lost in underflow, $\varepsilon_u \rightarrow 0$
$F/C_f < 2$	Increase*	Some oil lost to underflow (conservative criterion)
Excessive free gas in underflow	Increase*	Effective overflow area for liquids is reduced. Condition may be difficult to detect
F too high	Reduce	Δp_{fo} excessive
F too high	Reduce	Too much water in overflow stream

*Opening the valve in the overflow raises the flowrates hence increasing PDR.

Table 6 Relative cost for clean-up plant (16 000 $m^3 d^{-1}$)

Plant type	*Capital cost*	*Running cost*	*Oil separation performance*†
Hydrocyclones*	1.0	1.0	3
Induced gas flotation	1.1	1.1	4
Filter/coalescer	1.3	1.6	2
Centrifuge	4.6	4.0	1

*Capital cost is 6–7 × annual running cost based on data from several North Sea fields with installations ~10 years old.
† Ranking order for oil content in cleaned flow with identical oil content and mean drop size in the feed; 1 is best.

- De-sanding hydrocyclones are now being installed upstream from the choke in some fields. There could be an energy saving if de-sanding and de-oiling could be performed in a single unit, but simultaneous optimization of both functions is unlikely. A successful laboratory research project has been reported in France, but initial field trials in West Africa were disappointing.
- Heavy oils, i.e. those with a higher density and viscosity, appear unpromising for cyclonic separation processes. Nevertheless success has been reported for commercial de-oiling units used in the concentrator mode in trials in western Canada.
- Feasibility studies and preliminary field trials are in progress on integrated de-watering plus de-oiling cyclonic separation and/or de-sanding plus de-oiling. The attraction is an ultra-compact plant suited particularly to floating installations. Though de-watering units have not yet met with widespread success, the impending arrival of compact, robust electrocoalescers to raise water droplet size prior to separator entry, could transform the situation.
- With success in dealing with petroleum it is surprising that applications to edible oils, which are about 10 times more valuable, have yet to materialize. Laboratory trials have been very satisfactory. Not only is lost oil a revenue drain but it generates a potential environmental hazard.

Further Reading

Note: The five conferences on hydrocyclones all contain several papers on oil–water hydrocyclones

First Hydrocyclone Conference, Cambridge (UK) (1980) Priestley G and Stephens HS (eds). Cranfield: BHRA.

Second Hydrocyclone Conference, Bath (UK) (1984) Watts GA and Pickford R (eds). Cranfield: BHRA.

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Fourth Hydrocyclone Conference, Southampton (UK) (1992) Svarovsky L and Thew MT (eds). Dordrecht: Kluwer.

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Paige R and Ferguson M (1993) Water injection: practical experience and future potential (A BP Study). *Conference on Offshore Water and Environmental Management*. Business Seminars International, London.

Smyth IC and Fay B (1998) Further developments of the Hydrosep^{TM} System for downhole oil/water separation. *Conference on Downhole Production and Subsea Processing*, Aberdeen. Organised by BHR Group, London: MEP.

Svarovsky L (1984) *Hydrocyclones*. London: Holt, Rinehart and Winston.

Young GAB, Wakley WD *et al.* (1994) Oil-water separation using hydrocyclones: an experimental search for optimum dimensions. *Journal of Petroleum Science* 11: 37–50.

Dissolved Air

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Introduction

Dissolved air flotation (DAF) is a solid–liquid separation process for the removal of fine suspended material from an aqueous suspension. The basic principle underlying DAF is Henry's law, which gives the solubility of air in water. According to Henry's law, the solubility of air in water is directly proportional to its partial pressure. A supersaturated solution of water is produced using high pressure in a saturator. The bubbles are generated by the pressure release of this water stream. These bubbles attach to suspended material present in the aqueous stream, causing them to float to the surface, where they are collected as floc.

DAF can be carried out by vacuum or pressurized methods. In the vacuum flotation method the water to be treated is saturated with air at atmospheric pressure. The bubbles are produced by applying a vacuum to the flotation tank, releasing the air as fine bubbles. The vacuum flotation process has several disadvantages. These are (a) the amount of air available for flotation is limited by the vacuum achievable, (b) it is a batch process, and (c) it requires special equipment to produce and to maintain high vacuum. These disadvantages limit the application of vacuum flotation and it is only used in wastewater sludge thickening.