ABSTRACT

The increase in global oil and gas demand is pushing operators to constantly increase production from existing assets. Within these facilities, the separation of oil, water, gas and added chemicals plays a key role in meeting export gas specifications.

The carryover of liquids from a gas - liquid separator vessel, however, can easily lead to compressor failure or increase the dewpoint outside of specifications. This can be the result of improper separator design (e.g. undersizing or incorrect type selection) or due to unforeseen operating conditions and limits. Conventional solutions to this problem are to modify the separator (new internals), or to install additional equipment (e.g. a downstream coalescer).

Furthermore, upstream pressure reducing valves inherently create flow shear, which reduces the mean droplet size. The throttling process (where fluids expand in the valve once sufficient pressure drop is created over the valve), can also lead to flashed liquid or a condensed gas, which must be separated afterwards, and an intensive mixing of the gas and liquid phases, thereby diminishes the efficiency of the separators downstream of the chokes.
EXECUTIVE SUMMARY

The effective separation of the phases in separators strongly depends on the sizes of the dispersed droplets. The bigger the droplets, the more efficient the separation. By improving the droplet size in the feed stream, the separator capacity can thus be increased without modifications to the separator, or the installation of additional equipment.

In this paper, we introduce the gains from employing a cyclonic valve, SwirlValve – an intelligent adaptation of existing and proven axial control valve technologies, achieved by applying tangential slots in the valve cage. We achieve this by demonstrating the working principles and benefits of the cyclonic valve both qualitatively - using fundamental physical arguments (particularly in regard to droplet size and droplet coalescence), and numerical simulations - and quantitatively, based on experimental test data.

The qualitative analysis shows that for gas-liquid applications, the separation performance gained by employing a cyclonic valve is primarily due to the strong centrifugal liquid slip and the emergence of a liquid wall film that acts as a droplet coalescing sink. For liquid-liquid applications, reduced droplet breakup also plays an important role and cyclonic valves can thus be regarded as low-shear coalescing valves in such applications.

The NAM test data shows tangible benefits and returns on investment, delivering in a gas-liquid application at the Opende Oost production facility, an improvement of export hydrocarbon dewpoint by 6 [°C] and increased flow rates of up to 20%. The results clearly showed that SwirlValve improved dewpointing in all cases and that the effect becomes more pronounced at higher flow rates. In a liquid-liquid SwirlValve test campaign in Brazil, test results showed that in most conditions SwirlValve significantly improved separation performance.

We thus conclude in this paper, that SwirlValve is playing a significant role in both gas-liquid and liquid-liquid applications in increasing plant flow and gas handling capacity; improving separation efficiencies; and enhancing hydrocarbon dew-pointing performance. The result is a commercially viable and more sustainable alternative to traditional JT valves and significant and almost immediate, returns on investment.
1. Introduction – The Emergence of Cyclonic Valve Technology

The increase in global oil and gas demand is pushing operators to constantly increase production from existing assets. Within these facilities, the separation of oil, water, gas and added chemicals plays a key role in meeting export gas specifications. The carryover of liquids from a liquid gas separator vessel, however, can easily lead to compressor failure or increase the dewpoint outside of specifications.

Insufficient separation efficiency, or the need to maximise throughput beyond the original design capacity can result in liquid carryover. This can be the result of improper separator design (e.g. undersizing or incorrect type selection) or due to unforeseen operating conditions and limits. Conventional solutions to this problem would be to modify the separator (new internals) or to install additional equipment (e.g. a downstream coalescer).

The separation of the phases in separators strongly depends on the sizes of the dispersed droplets: the bigger the droplets, the more efficient the separation. Upstream pressure reducing valves inherently create flow shear, which reduces the mean droplet size. The increased shear forces, cause a reduction in droplet size, leading to the need to install new separator vessels, higher costs, long overhaul times and diminishing returns.

An ideal solution to this problem would be to apply a control valve that minimizes shear and/or performs the pressure letdown more gradually. The maximum flow shear rate encountered by the droplets will then be reduced because the energy dissipation takes place over a larger volume. Cyclonic valves, which entered the market in 2008 and have been further developed by Twister, effectuate an increase in mean droplet size (compared to conventional valves) and consequently improve a downstream separator's ability to separate the phases (conceptually visualized in Figure 1). Alternatively, for a certain carryover specification, these valves can be used to optimize operational parameters, such as flow rate, pressure drop and glycol consumption.

Cyclonic valve technology can be beneficially employed in both gas-liquid and liquid-liquid separation applications. The valves can be configured (in terms of material selection, size and actuator method) in any manner a conventional valve could be; in Figure 2, a commercially delivered hand-controlled 2” SwirlValve from Twister is shown.

![Figure 1: SwirlValve effect on separated liquid fraction.](image1)

![Figure 2: Hand-controlled 2” SwirlValve.](image2)

The patented (Ref. 1) SwirlValve cyclonic valve concept is, in essence, the same as a conventional axial flow valve. The difference (schematically illustrated in Figure 3 and Figure 4) is in the cage design: for SwirlValve, the cage flow passages are angled with respect to the radial direction, resulting in a tangential (instead of radial) cage outflow.

![Figure 3: Conventional axial flow valve](image1)

![Figure 4: SwirlValve](image2)

With regard to the outlet droplet size distribution, this has the following main benefits:

- The energy dissipation (pressure letdown) not only takes place in the cage, but also in the downstream vortex. As such, the maximum flow shear rate the droplets are exposed to is lower and the droplet breakup in the cage is reduced. This benefit is more pronounced for liquid-liquid than for gas-liquid applications, which are related to the much higher viscosity of the continuous phase.
- The centrifugal force induced by the vortex causes the droplets to move towards and concentrate along the wall, which significantly increases the rate of droplet coalescence.

The results of these two key benefits are larger droplets and more efficient separation. We look at droplet breakup and droplet coalescence qualitatively in more detail below.

2.1 Droplet Breakup

The effect of increased dissipation volume is similar to that of multistage pressure reduction, which was shown by Van der Zande (Ref. 11) to result in a larger downstream mean droplet size. In his research, Van der Zande measured and compared the droplet size distributions resulting from the flow through an orifice, a single needle valve and two needle valves in series (for the same flow rate and pressure drop).

Fundamentally, the relation between shear rate and droplet size can be understood by looking at the local Weber number \( (We) \) - a dimensionless quantity describing the ratio of inertial flow forces over interfacial surface tension. Note that throughout this document, subscripts \( c \) and \( d \) denote continuous and dispersed phase properties respectively. When the Weber number exceeds a certain limit, the droplet surface tension will not be sufficient to balance the dynamic pressure exerted by the flow on the droplet surface; the droplet will then eventually break up into (several) smaller droplets.

The dynamic force can be present either due to local relative velocity (droplet slip), or due to turbulent eddies of a small scale (relative to the droplet size); these forms of droplet breakup are hereafter referred to as drag- and turbulent eddy-induced breakup. A general expression for this critical Weber number is:

\[
We_{cr} = \frac{\rho U^2 d_{max}}{\sigma} \quad \text{Eq. (1)}
\]
Where \( \rho_c \) is the continuous phase mass density (with dimensionality \([\text{kg/m}^3]\)), \( d_{\text{max}} \) the maximum stable droplet diameter (with dimensionality \([\text{m}]\)), \( \sigma \) the interfacial surface tension (with dimensionality \([\text{N/m}]\)) and \( U \) a characteristic velocity scale (dimensionality \([\text{m/s}]\)). Following Ref. 5, using expressions for the characteristic velocity scales for both types of breakup, Eq. (1) can be rewritten to obtain expressions for the maximum sustainable droplet size:

\[
 d_{\text{max}, \text{turb}-\text{drag}} = \left( \frac{\sigma \rho \epsilon}{k_1 \rho_c} \right)^{\frac{3}{5}} e^{-\frac{2}{5}} \left( \frac{\rho_c - \rho_d}{\rho_d} \right)^{\frac{2}{5}} e^{-\frac{2}{5}} \tag{Eq. 2}
\]

\[
 d_{\text{max}, \text{turb}-\text{eddy}} = \left( \frac{\sigma \rho \epsilon}{k_2 \rho_c} \right)^{\frac{3}{5}} e^{-\frac{2}{5}} \tag{Eq. 3}
\]

Where \( \rho_c, \rho_d \) are the phase mass densities (dimensionality \([\text{kg/m}^3]\)), \( \epsilon \) the turbulence dissipation rate (with dimensionality \([\text{m}^2/\text{s}^3]\)) and \( k_1, k_2 \) dimensionless constants with a value close to 1. The most limiting value obtained from the two expressions determines the maximum sustainable droplet size. Taking the ratio of the two expressions we find that:

\[
 \frac{d_{\text{max}, \text{turb}-\text{drag}}}{d_{\text{max}, \text{turb}-\text{eddy}}} = \left( \frac{\rho_c}{\rho_d} \right)^{\frac{2}{5}} \left( \frac{\rho_c - \rho_d}{\rho_d} \right)^{\frac{2}{5}} e^{-\frac{2}{5}} \tag{Eq. 4}
\]

Evaluating Eq. (4), it can be seen that for typical gas-liquid applications (\( \rho_c/\rho_d \ll 1 \)) turbulent drag-induced breakup will be limiting, while for oil-water applications (\( \rho_c/\rho_d \approx [0.5-1.0] \)) turbulent eddy-induced breakup will be limited. It should however be realized that the drag-induced breakup considered here is based on a turbulent velocity scale (evidenced by the presence of \( \epsilon \) in Eq. (2)).

For the flow through a cyclonic valve, the limiting drag velocity scale can also be the droplet slip velocity \( U_s \) (dimensionality \([\text{m/s}]\)), due to centrifugal acceleration, defined as:

\[
 U_s = |\vec{U}_d - \vec{U}_s| \tag{Eq. 5}
\]

Neglecting unsteady forces (e.g. Basset history force), the slip velocity can be calculated from the balance between droplet drag, the pressure gradient force and acceleration forces:

\[
 U_s = \sqrt{\frac{8 \rho_d - \rho_c}{6 \rho_c} \frac{a \cdot d}{C_d(Re_d)}} \tag{Eq. 6}
\]

Where \( a \) is the acceleration (dimensionality \([\text{m/s}^2]\)) and \( C_d \) is the dimensionless drag coefficient. The drag coefficient is a function of the droplet slip Reynolds number \( (Re_d) \) of which the definition is given in Eq. (7) below.

\[
 Re_d = \frac{\rho U_s d}{\mu_c} \tag{Eq. 7}
\]

Where \( \mu_c \) is the continuous phase dynamic viscosity (with dimensionality \([\text{Pa} \cdot \text{s}]\)). For Stokes flow \( (Re_d<1) \), the drag curve is well approximated by \( C_d=24/Re_d \), which results in an explicit relation for the droplet slip velocity. The same holds for the inertial range of the curve \( (750<Re_d<3.5 \cdot 10^5) \), which is well approximated by a constant value: \( C_d=0.45 \). In the intermediate range \( (Re_d<750) \), the drag curve is well approximated by the Schiller-Naumann relation presented in Eq. (8) (see Ref. 2). Using this relation, the droplet slip velocity equation (Eq. (6)) results in an implicit expression that needs to be solved iteratively.

\[
 C_d = \frac{24}{Re_d} \left(1 + 0.15 Re_d^{0.687} \right) \tag{Eq. 8}
\]
Inserting Eq. (6) into Eq. (1) and rewriting, we find an expression for the maximum sustainable droplet size under the influence of slip-induced breakup:

\[
d_{\text{max,slip}} = \sqrt[8]{\frac{6}{8} \frac{W_{ce}}{r_c} \frac{\sigma C_d \left( Re_{\text{max}} \right)}{\left( \rho_d - \rho_c \right) a}}
\]

Eq. (9)

For a centrifugal force field, the droplet acceleration can be calculated as:

\[
a_{\text{cent}} = \frac{U^2}{r}
\]

Eq. (10)

Where \( r \) is the radial coordinate (with dimensionality [m]). Combining Eq. (9) and Eq. (10) results in an expression for the maximum sustainable droplet size under the influence of centrifugally-induced breakup:

\[
d_{\text{max,cent}} = \sqrt[8]{\frac{6}{8} \frac{W_{ce}}{r_c} \frac{\sigma C_d \left( Re_{\text{max}} \right)}{\left( \rho_d - \rho_c \right) \frac{U^2}{r}}}
\]

Eq. (11)

Experiments have shown that for low viscosity fluids the critical Weber number is in the range of [5-25] and is dependent on the droplet Reynolds number. In Ref. 6 an empirical correlation for the critical Weber number, fitted to the experimental results of many authors, is provided:

\[
W_{ce} = \begin{cases} 
55 & , \text{for } Re_d < 200 \\
\frac{24}{Re_d} + \frac{20.1807}{Re_d^{0.615}} - \frac{16}{Re_d^{2/3}} & , \text{for } 200 < Re_d < 2000 \\
5.48 & , \text{for } Re_d > 2000
\end{cases}
\]

Eq. (12)

For typical gas-liquid cyclonic valve applications, the centrifugal acceleration (\( a \)) shortly after the cage exit will be between 10,000-50,000 times the gravitational acceleration (\( g \)). With such high inertial forces, Eq. (9) will equate to lower values than Eq. (2) and droplet breakup downstream of the valve will thus be driven by the droplet slip drag force. Though this slip-induced breakup that, at first glance, appears to have a deleterious effect on the droplet size distribution, it should be realized that this type of breakup only pertains to droplets that have not yet been captured by (or are re-entrained from) the wall-bound liquid film. It can be calculated that the slip velocity of the largest sustainable droplet size will be in the range of 10-50 [m/s²]; even with typical residence times in the order of just milliseconds, this is enough for the majority of the droplets to reach the wall film.

2.2 Droplet Coalescence

The downstream droplet size can also be increased through enhanced coalescence. The droplet collision rate for a monodispersed system can be written in general form (see Ref. 8), as:

\[
N = K \cdot n^2
\]

Eq. (13)

Where \( N \) is the collision rate (dimensionality [s⁻¹]), \( K \) the coalescence coefficient (dimensionality [cm³/s]) and \( n \) the droplet number density (dimensionality [cm⁻³]).

Due to the centrifugal forces induced by SwirlValve, the droplets become concentrated in a much smaller volume, i.e. the average droplet number density \( n \) increases. As Eq. (13) shows, the coalescence rate scales with \( n^2 \) and as such, cyclonic valves bring about a very direct and substantial improvement in the rate of coagulation. This direct effect will be stronger for gas-liquid than for liquid-
liquid applications, because both the centrifugal forces (velocities) and phase density difference are much higher.

Besides this direct effect via the centrifugal droplet densification, cyclonic valves also have an impact on the coalescence rate through the coalescence coefficient. The coalescence coefficient is in essence a measure for the relative velocities of the droplets and its form depends on the specific mechanism of coalescence. For the flow through a valve, the relevant forms are: Brownian, turbulent and inertial. Turbulent coalescence can be further subdivided based on the local scale of the turbulent structures. The subdivision can be made based on the ratio of the droplet diameter and the Kolmogorov length scale \( \lambda_0 \). The Kolmogorov length represents the smallest scale in the turbulence energy spectrum at which the dissipation of the turbulent energy takes places through viscous effects. An expression for the Kolmogorov length can be derived on dimensional grounds (e.g. see Ref. 10):

\[
\lambda_0 = \left( \frac{\nu^3}{\varepsilon} \right)^{1/4} \quad \text{Eq. (14)}
\]

With \( \nu \) the kinematic viscosity of the continuous phase (with dimensionality \([m^2/s]\)). The coalescence mechanisms and the regime in which they are dominant can now be described in more detail:

- **Brownian motion**
  - Collisions between droplets as a result of the random motion induced by collisions with gas molecules.
  - Dominant mechanism for very small droplets: \( d/\lambda_0 << 1 \).
  - Coalescence coefficient scales with the ratio of temperature \( T \) over dynamic viscosity \( \mu \) (Ref. 4).
- **Turbulence in the dissipation range**
  - Collisions between droplets as a result of turbulent motion of the gas in the small scale range of the turbulence energy cascade spectrum, where viscosity plays a role.
  - Dominant mechanism for small droplets: \( d/\lambda_0 < 1 \).
  - Coalescence coefficient scales with the square root of turbulence dissipation rate \( \varepsilon \) (Ref. 8).
- **Turbulence in the inertial range**
  - Collisions between droplets as a result of turbulent motion of the gas in the large scale range of the turbulence energy cascade spectrum, where viscosity does not play a role.
  - Dominant mechanism for larger droplets: \( d/\lambda_0 > 1 \).
  - Coalescence coefficient can be shown to scale with the cubic root of turbulence dissipation rate \( \varepsilon \) on dimensional grounds (Ref. 4).
- **Inertial motion**
  - Collisions between droplets due to differences in slip velocity with respect to the continuous phase, for example due to gravity or centrifugal forces.
  - All scales, as this type of collision occurs mainly between droplets of different size.
  - Slip velocity is strongly dependent on the droplet size, so collisions between droplets will occur more frequently when the size distribution is relatively wide.

Brownian motion can be seen as a temperature effect not strongly altered in cyclonic flow. The turbulent coalescence regimes are a function of the levels of turbulence (dissipation), which will be different for cyclonic valves. The effect is however expected to be limited as the scaling goes with the root and cubic root of the turbulence eddy dissipation rate, respectively for the dissipation and inertial range of the spectrum. Moreover, the difference will likely be more pronounced for liquid-liquid than for gas-liquid applications, because a larger part of the energy dissipation takes place in the downstream vortex.

Finally, it is obvious that cyclonic valves will promote collisions between droplets of different sizes, due to velocity differences induced by centrifugal force.

The use of Computational Fluid Dynamics (CFD), allows researchers and engineers to assess a concept’s validity at low expense (both in terms of time and cost). Moreover, it enables the exploration of features and details of the flow that would be extremely difficult, or even impossible, to measure experimentally.

In its most basic form, CFD can be used to properly design a valve for a specific application and to gain an understanding of the flow path and local velocities. As a case study, simulations were performed for a typical gas-condensate application with the following characteristics:

- Inlet pressure/temperature: 100 [bara] / 25 [°C]
- Outlet pressure: 65 [bara]
- Gas: Typical Groningen gas composition (81 [mol%] methane)
- Condensate density: 720 [kg/m³]
- Inlet droplet size distribution: Rosin-Rammler - power 1.5 and Harwell median size

The simulations included Lagrangian particle tracking and took into account the effects of breakup via the Enhanced Taylor Analogy Breakup (ETAB) model. In Lagrangian particle tracking, fluid parcels are released in the continuous phase flow and their trajectories calculated by solving the particle equations of motion. Note that the actual physics are much more complex, involving droplet nucleation, condensation and coalescence. Though possible in principle, for this numerical study, the focus is on the droplet breakdown process, which puts an upper limit on the droplet size.

The inlet droplet size distribution is based on the Rosin-Rammler distribution (with power 1.5), which is defined by Eq. (15):

\[
\phi_i = \left( \exp \left( - \frac{d_i}{d_{med}} \right) \right)^n \tag{Eq. (15)}
\]

Where \(\phi_i\) is the retained mass fraction of size class \(i\) (i.e. the cumulative mass fraction of all sizes above \(d_i\)) and \(n\) is the Rosin-Rammler power (or spread rate). The median droplet size \(d_{med}\) is determined here by means of the Harwell method (see Ref. 3):

\[
d_{med} = 1.42 \cdot 1.91 \cdot D_{in} \cdot \frac{Re_{in}^{0.1}}{We_{in}^{0.6}} \left( \frac{\rho_g}{\rho_l} \right)^{0.6} \tag{Eq. (16)}
\]

Where \(D_{in}\) is the diameter of the inlet pipe (with dimensionality [m]), \(Re_{in}\) and \(We_{in}\) are the inlet Reynolds and Weber numbers and \(\rho_g\) and \(\rho_l\) are the gas and liquid mass densities (dimensionality [kg/m³]). For the case study this leads to the inlet droplet size distribution presented in Figure 5.

![Figure 5: Simulation inlet droplet size distribution - retained mass fraction (cumulative fraction above a certain size). The square represents the median (Harwell) droplet size.](image-url)
Two valve geometries were simulated. The geometries were identical except for their cages, which were sized to yield a close match in valve flow capacity coefficient ($C_v$).

- **3" conventional axial valve with radial channels.**
  - Length of 493 [mm] and outer diameter of 144 [mm].
  - 3 rows of 40 tangentially arranged channels, with a 48.5 [mm$^2$] rectangular cross-section.
- **3" SwirlValve with tangential channels.**
  - Length of 493 [mm] and outer diameter of 144 [mm].
  - 2 rows of 12 tangentially arranged channels, with a 218 [mm$^2$] rectangular cross-section.

In Figure 6, the flow through the conventional valve and SwirlValve are visualized by means of the streamlines extracted from the simulation results. The figure clearly shows the difference in the flow fields, with SwirlValve achieving tangential velocities of more than 200 [m/s]. These high velocities translate (via Eq. (10)) to a strong centrifugal force, presented in Figure 7 as the centrifugal g-force (the ratio of centrifugal over gravitational acceleration).

![Figure 6: Computational Fluid Dynamics simulation results: streamlines coloured with local velocity, for a conventional valve (left figure) and SwirlValve (right figure).](image)

![Figure 7: SwirlValve Computational Fluid Dynamics simulation results for a typical gas-condensate application: cross-section isocontours of centrifugal g-force (ratio of centrifugal over gravitational acceleration).](image)

The various expressions developed to assess the maximum sustainable droplet size under various forms of breakup (Eq. (2), Eq. (3) and Eq. (11)) can be evaluated throughout the flow fields, with the results presented in Figure 8 and Figure 9. The figures show the following:

- Turbulent drag breakup is indeed dominant over turbulence eddy breakup for this application.
• Downstream of SwirlValve, breakup due to centrifugal slip is dominant over turbulent drag breakup, resulting in a maximum stable size of about 20 [µm]. At the cage exit, the criterions are equally limiting, both resulting in a maximum size of about 30 [µm].

Figure 8: Conventional valve Computational Fluid Dynamics simulation results for a typical gas-condensate application: cross-section isocontours of maximum sustainable droplet size under turbulent eddy breakup, based on Eq. (2), (left figure) and under turbulent drag breakup, based on Eq. (3), (right figure).

Figure 9: SwirlValve Computational Fluid Dynamics simulation results for a typical gas-condensate application: cross-section isocontours of maximum stable droplet size under turbulent drag breakup, based on Eq. (3), (left figure) and under centrifugal slip (drag) breakup, based on Eq. (11), (right figure).

Note that in evaluating Eq. (11) for Figure 9, it was assumed that $C_d=0.45$ (valid for $Re_d>750$) and $We_{crit}=5.48$ (valid for $Re_d>2000$). From Figure 10, in which the droplet Reynolds number of this maximum size (at terminal velocity) is presented, it can be seen that the $Re_d$ is indeed larger than 750 but not more than 2000 in the downstream flow field. From Eq. (12) it can be concluded that this leads to a conservative (lower) estimate of the maximum centrifugal droplet size, because the critical Weber number will actually be slightly higher in the sub-2000 Reynolds number range (i.e. droplets are able to withstand higher inertial forces).

Figure 10: SwirlValve Computational Fluid Dynamics simulation results for a typical gas-condensate application: cross-section isocontours of droplet Reynolds number for the maximum sustainable droplet size under centrifugal slip drag.
The above means that the swirling flow indeed limits the maximum droplet size downstream of SwirlValve. However, if the droplets of this maximum size are slipping fast enough to reach the wall film (or do so before they reach terminal velocity), this limit does not severely impede the droplet coalescence process. In order to assess whether this is the case, it is useful to investigate the particle tracks in more detail. In Figure 11 the averaged liquid volume fractions are presented on cross-section planes for both valves. The figure shows the strong centrifugal effect of SwirlValve, with most particles reaching the wall at a short distance behind the cage. This will lead to the formation of a wall film and, as argued in the previous section, the droplet densification will strongly promote coagulation. While most droplets reach the wall 20-50 [cm] downstream of SwirlValve, it's been estimated to require 20-50 [m] for a conventional valve.

![Figure 11: Particle tracking results (including break-up) for a typical gas-condensate application: averaged condensate volume for a conventional valve (left figure) and SwirlValve (right figure).](image)

In Figure 12 periodic sections of the actual particle trajectories are shown, coloured by the mean particle diameter. The right hand figure shows that downstream of SwirlValve’s cage, most particles migrate to the wall very quickly. Only the smallest droplets, with a diameter below 5 [µm], do not instantly reach the wall. As the critical centrifugally-induced breakup size was found to be roughly 20 [µm], this finding adds credence to the assertion that high centrifugal drag forces are not limiting the coalescing ability of SwirlValve.

This fact is visualized even more clearly in Figure 12, which shows the particle tracks coloured by their relative mean particle size. The relative size is defined here as the ratio of the mean particle diameter over the maximum sustainable droplet size (based on both turbulent drag and centrifugal slip). As such, this ratio gives an indication of the margin the droplets have with respect to becoming unstable. The figure reinforces the earlier findings, showing that droplets of critical size easily reach the wall film. It also shows that the assumed particle distribution and employed breakup model appear to be in agreement with the theoretically derived maximum droplet size expression.

![Figure 12: Particle tracking results (including break-up) for a typical gas-condensate application: particle tracks coloured with mean particle diameter for a conventional valve (left figure) and SwirlValve (right figure). Note that only periodic sections of the tracks are shown for visualization clarity.](image)
Figure 13: SwirlValve particle tracking results (including break-up) for a typical gas-condensate application: particle tracks coloured with relative droplet size ($d_{\text{mean}}/d_{\text{max}}$), where $d_{\text{max}}$ is based on the turbulent drag breakup criterion of Eq. (3) (left figure) and the centrifugal criterion of Eq. (11) (right figure). Note that only periodic sections of the tracks are shown for visualization clarity.

In Figure 14 the particle tracks are presented again, now coloured with the turbulence eddy dissipation rate. Though the distribution is slightly different, the figure shows a similar order of dissipation magnitude. Given the fact that the coalescence coefficient scales with the square or cubic root of this quantity, it can therefore be concluded that the coalescence rate is not significantly modified via the coefficient. The main beneficial effect will be due to the droplet densification and formation of the wall film.

Note that this only pertains to similar high pressure drop gas-liquid applications; for lower pressure drops and liquid-liquid applications, turbulence-enhanced coalescence might play a more prominent role. However, examining a previously generated result for SwirlValve employed in a water-glycol flow application (see Figure 15) shows that the turbulence damps out rather quickly in liquid-liquid flow. The improvement of coalescence rate through turbulence enhancement is therefore probably rather limited for most cases.

Finally, isocontours of stagnation (or total) pressure are shown in Figure 16. The figure clearly visualizes the fact that the total pressure loss occurs mainly in the cage (outlet) area for both the conventional valve and SwirlValve. Downstream of this area the total pressure distribution remains fairly constant.

Figure 14: Particle tracking results (including break-up) for a typical gas-condensate application: particle tracks coloured with turbulence eddy dissipation for a conventional valve (left figure) and SwirlValve (right figure). Note that only periodic sections of the tracks are shown for visualization clarity.
4. Field Trials – The Netherlands and Brazil

Now that the benefits of SwirlValve and its ability to increase droplet size and encourage droplet coalescence have been demonstrated both qualitatively and numerically, we examine real life trials in the field.

Two major test campaigns have been conducted for SwirlValve: the first in gas-liquid and the second for a liquid-liquid application. Both campaigns and the obtained results are discussed in the following sections.

4.1 Gas-liquid: JT-LTS

In collaboration with the Nederlandse Aardolie Maatschappij (NAM), a field trial was performed with SwirlValve employed as a Joule-Thomson (JT) valve in a Low Temperature Separator (LTS) train at the Opende Oost production facility (see Figure 17). The aim of the trials was to establish whether SwirlValve would be able to decrease LTS (type: SMSM) carryover and associated glycol losses. After new wells came on stream, plant capacity became the limiting factor because of increasing carryover from the LTS (and the associated impact on hydrocarbon dewpoint). With the conventional valve, the operational maximum was 600,000 [Nm³/day] while the operator was aiming to produce at 670,000 [Nm³/day].

The export gas quality was monitored during the test period using an online hydrocarbon dew-point (HCDP) analyzer and a mobile automatic condensate metering unit, both provided and operated by Gasunie Netherlands. The latter method measured the liquid dropout at -3 °C and 27 bar, which was reported as the Potential Hydrocarbon Liquid Content (PHLC) (with dimensions [mg liquid / Nm³ gas]).
The JT valve was operated with its traditional low-noise valve trim during the first half of the test plan. Subsequently, the valve was taken out and the labyrinth cage changed out for a SwirlValve cage. The valve was then operated for the remainder of the tests in the SwirlValve configuration.

Figure 17: NAM Opende Oost JT-LTS gas processing train.

Results from these tests have already been extensively described in Ref. 7. Here, only an overview of the tests and the main findings are presented. The trials can be subdivided into five test cases, for which the results are summarized below:

- **Low flow: 100,000 [Nm³/day]**
  - Small difference between results, with a small HCDP improvement using SwirlValve. Both valves operate well within the export specification.

- **Nominal flow: 600,000 [Nm³/day]**
  - Slightly improved dewpointing with SwirlValve.

- **High flow: 650,000 [Nm³/day]**
  - Much improved dewpointing with SwirlValve

- **Ultimate flow: >700,000 [Nm³/day]**
  - SwirlValve only case, showing that the flow could be increased up to 735,000 [Nm³/day] while still meeting export specifications.

- **Ultimate flow optimization: 710,000 [Nm³/day] at highest possible LTS temperature.**
  - SwirlValve only case, showing that (at very high flow rate) the LTS could be operated at 4-5 [°C] higher temperature.

The performance quantities of the comparative cases are presented in Figure 18, clearly showing that SwirlValve improved dewpointing in all cases and that the effect becomes more pronounced at higher flow rates.

The achieved increase of LTS operating temperature (within export specification), can be calculated to be equivalent to a 3-4 [bar] feed pressure reduction. This would translate to a deferral of planned compression installation and an associated significant saving in fuel costs could be realized.
4.2 Liquid-Liquid: Oil-Water Separation

A liquid-liquid SwirlValve test campaign was executed by Petrobras at their experimental center (NUEX) in Aracajú (Sergipe, Brazil). The results of this work were presented earlier in Ref. 9. For these tests, a flow loop (Process Flow Diagram shown in Figure 19) was set up in which a mixture of oil and salt water\(^1\) (fed from separate storage tanks TQ-02/TQ-03) was passed through either the globe valve or SwirlValve (installed in parallel, see Figure 20) after which it was directed to a settling tank, equipped with four sampling points at different elevations.

When the tank was filled to a certain predetermined volume, the flow was stopped and the mixture was allowed to settle for 10 minutes. Directly following this period, samples were swiftly taken at all sample points, which were subsequently photographed and taken for further laboratory analysis (measuring water and oil contents and assessing emulsion stability using a centrifuge). The samples obtained from the globe valve and SwirlValve runs (under identical conditions), could then be used to compare the performance, both visually and based on measured water content.

After finishing a specific test run, the sampling tanks were drained into a large storage vessel. From there, the mixture was then passed through a Compact Electrostatic Separator (CES) in which the oil and water phases were separated and subsequently returned to their storage tanks.

The valve inlet mixture conditions and flow rates could be varied over a wide range, using the individual feed pump rates and by choosing whether or not to pass the combined stream through a mixture device and/or a shear-inducing upstream valve. A test matrix was setup in which both flow rate and water content were varied in 3 steps: flow rates of 30 [m\(^3\)/hr], 50 [m\(^3\)/hr] and 60 [m\(^3\)/hr] and water cuts of 30 [vol%], 50 [vol%] and 70 [vol%]. Furthermore, each combination of flow rate and water cut was run in three different inlet flow scenarios (in- or excluding the mixing device and upstream shear valve): 1. No mixer, no shear valve (i.e. full bypass), 2. Mixer only 3. Both mixer and shear valve.

\(^1\) The water had a salinity of 135 [g/L] and the specific oil used was a blend of 55 [vol%] Bonsucesso oil and 45 [vol%] diesel, complemented with 60 [ppm] of a de-emulsifying agent to establish a stable and uniform oil mixture.
Figure 19: Process Flow Diagram of the SwirlValve test loop at the Petrobras NUX Aracajú facilities.

Figure 20: Aracajú test site: parallel valve installation.

Figure 21: Aracajú test site: settling/sampling tank with tag numbers identifying the sample points.
Examples of results are given in Figure 22 to Figure 25 below. In the captions of these figures, the flow rate is denoted as $Q$, the water cut as $WC$ and the pressure drop over the tested valve as $\Delta p_{\text{valve}}$.

The Petrobras researchers concluded the following (Ref. 9):

- For most conditions, SwirlValve significantly improved separation performance. For some conditions the difference was negligible, but in no cases, was the performance worse with SwirlValve.

- The degree of separation enhancement appeared to depend on:
  
  o Pressure drop: at higher pressure drop the swirling velocities are higher and the centrifugal effect is stronger. See for example the difference between Figure 22 and Figure 23.
  
  o Emulsion stability: results can change strikingly when the inlet flow regime is different (effectuated by the upstream mixer/shear valve). See for example the difference between Figure 24 and Figure 25 on the following page.

**Figure 22: Results for $Q = 45 \, [m^3/hr], \, WC = 70\%, \, \Delta p_{\text{valve}} = 9 \, [bar], \, \text{no mixer / no inlet shear}$. Tag numbers identify sample point locations (higher number being at lower elevation).**

**Figure 23: Results for $Q = 45 \, [m^3/hr], \, WC = 70\%, \, \Delta p_{\text{valve}} = 4 \, [bar], \, \text{no mixer / no inlet shear}$. Tag numbers identify sample point locations (higher number being at lower elevation).**
Figure 24: Results for $Q = 30 \text{ [m}^3\text{/hr]}$, $WC = 70\%$, $\Delta p_{valve} = 6 \text{ [bar]}$, mixer and 3 [bar] inlet shear. Tag numbers identify sample point locations (higher number being at lower elevation).

Figure 25: Results for $Q = 30 \text{ [m}^3\text{/hr]}$, $WC = 70\%$, $\Delta p_{valve} = 6 \text{ [bar]}$, no mixer / no inlet shear. Tag numbers identify sample point locations (higher number being at lower elevation).
5. Conclusions

The working principles of cyclonic valve technology were assessed by means of theoretical and numerical analysis. This analysis showed that two aspects of droplet dynamics play a key role:

- Reduced droplet breakup, because the energy dissipation takes places over a larger volume (both cage and vortex).
  - This effect is mostly relevant for liquid-liquid applications.
  - In gas-liquid applications, the high centrifugal droplet slip velocities become the dominant driver of droplet breakup. It was shown that droplets with sizes below this centrifugally induced upper limit still easily reach the wall where they'll merge with the liquid film. As such, the centrifugal drag does not truly impede the coalescence process downstream of a cyclonic valve.

- Increased coalescence rate, because of four effects:
  - Formation of a wall film acting as a droplet sink.
  - Centrifugal droplet densification. This effect is stronger in gas-liquid applications due to higher centrifugal forces and a larger phase density difference.
  - Slip velocity differences between droplets of different size (large droplets spin out faster). As above, this is more important in gas-liquid applications.
  - Increased levels of downstream turbulent velocity fluctuations. For the gas-liquid case, downstream dissipation rates were shown to be fairly similar for a conventional axial valve and SwirlValve. For liquid-liquid applications, it is more relevant because a much higher fraction of the total dissipated energy is consumed by the downstream vortex.

For gas-liquid applications, the separation performance gained by employing a cyclonic valve is primarily due to the strong centrifugal liquid slip and the emergence of a liquid wall film that acts as a droplet coalescing sink. For these cases, cyclonic valves can thus be thought of as coalescing valves. For liquid-liquid applications, reduced droplet breakup also plays an important role. Cyclonic valves can thus be regarded as low-shear coalescing valves in liquid-liquid applications.

Field trials for both a gas-condensate (JT-LTS) and an oil-water (phase separation) application, showed that separation performance is significantly improved by employing SwirlValve. For the gas-liquid application, a 6 °C improvement of the export hydrocarbon dewpoint was achieved at the intended flow rate. This performance gain was used to increase the flow rate even further (20%). Alternatively, this enables operations at a lower feed pressure (3-4 [bar]), to significantly decrease lifetime field operation costs.

The liquid-liquid tests revealed that SwirlValve significantly improved separation performance in almost all test conditions. Direct visual comparison of the samples taken from the settling vessel often showed a striking difference in contamination levels for the tested valves. The degree of performance improvement was dependent on factors such as flow rate, pressure drop and inlet flow regime but in none of these cases, was a deleterious effect observed.

SwirlValve is playing a significant role in increasing plant flow and gas handling capacity; improving separation efficiencies; and enhancing hydrocarbon dew-pointing. The significantly improved separation and flow performance over traditional solutions, for both gas-liquid and liquid-liquid applications, will also ultimately generate improved returns on investment.
Abbreviations

CES: Compact Electrostatic Separator
CFD: Computational Fluid Dynamics
DEG: Diethylene Glycol
HCDP: Hydrocarbon Dewpoint
JT: Joule-Thomson
LTS: Low Temperature Separator
NAM: Nederlandse Aardolie Maatschappij
PHLC: Potential Hydrocarbon Liquid Content
SMSM: Schoepentoeter Mistmat Swirideck Mistmat
WC: Water Cut

References Cited