Optimisation of micro-processes for shear-assisted solid–liquid separation in a rotatory batch flow vortex reactor

B. Oyegbile, P. Ay and S. Narra

ABSTRACT

This paper reports the study of micro-processes in a novel pre-treatment technique using a pellet forming batch flow vortex reactor of cylindrical shape that consists of axially revolving rotor plates between fixed stator plates. The suspension was first mixed with high molecular weight synthetic polymers by stirring for approximately 50 seconds and then agitated for 20 minutes. The process was optimised for a number of operating conditions including polymer type and dosing regimen, rotation speed, wall-plate gap distance, residence time and suspension filling method. The results of the investigation show that optimising a number of process variables that influence floc formation and growth, along with specific apparatus construction and geometry, help to maintain the suspension in a metastable state that is crucial for the formation of pellet-like compact agglomerates with better dewaterability and uniform aggregate size. A maximum dry solids content of 28.3% after gravity dewatering through a 0.5 mm sieve was recorded during the investigation, with a maximum particle removal efficiency of 97.5%.

Key words | floc stability, hydrodynamics, pelleting flocculation, sludge, turbulence

NOMENCLATURE

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
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<tbody>
<tr>
<td>N</td>
<td>Particle number concentration per unit volume (m⁻³)</td>
</tr>
<tr>
<td>F₀</td>
<td>Orthokinetic collision efficiency (-)</td>
</tr>
<tr>
<td>Fₚ</td>
<td>Perikinetic collision efficiency (-)</td>
</tr>
<tr>
<td>r</td>
<td>Particle radius (m)</td>
</tr>
<tr>
<td>d, dₚ</td>
<td>Floc or particle diameter (m)</td>
</tr>
<tr>
<td>β</td>
<td>Rate constant (-)</td>
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<tr>
<td>η, μ</td>
<td>Dynamic viscosity (Ns⁻¹m⁻²)</td>
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<td>ν</td>
<td>Kinematic viscosity (m²s⁻¹)</td>
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<tr>
<td>T</td>
<td>Absolute temperature (K)</td>
</tr>
<tr>
<td>ε</td>
<td>Mean dissipation rate of kinetic energy (Nms⁻¹kg⁻¹ or m²s⁻³)</td>
</tr>
<tr>
<td>G, G̅</td>
<td>Mean velocity gradient or effective shear rate (s⁻¹)</td>
</tr>
<tr>
<td>U</td>
<td>Pair bonding energy (f)</td>
</tr>
<tr>
<td>D₆₅₅ₐₓ</td>
<td>Maximum floc diameter (m)</td>
</tr>
<tr>
<td>p</td>
<td>Flocs' porosity (-)</td>
</tr>
<tr>
<td>K</td>
<td>Bond destruction coefficient (-)</td>
</tr>
<tr>
<td>V</td>
<td>Stirring vessel volume or fluid volume (m³)</td>
</tr>
<tr>
<td>F₉</td>
<td>Turbulent hydrodynamic force (N)</td>
</tr>
<tr>
<td>P</td>
<td>Power input (W)</td>
</tr>
<tr>
<td>Tq</td>
<td>Shaft torque (Nm or kgm⁻²s⁻²)</td>
</tr>
<tr>
<td>N, n</td>
<td>Agitation speed (s⁻¹, or m⁻¹)</td>
</tr>
<tr>
<td>Np</td>
<td>Dimensionless power number (-)</td>
</tr>
<tr>
<td>λ</td>
<td>Kolmogorov micro scale of turbulence (m)</td>
</tr>
<tr>
<td>D</td>
<td>Agitator or stirrer diameter (m)</td>
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<tr>
<td>Dₑ</td>
<td>Dimensionless Reynolds number (-)</td>
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<td>ω</td>
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<tr>
<td>ρ₁</td>
<td>Fluid density (kgm⁻³)</td>
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<tr>
<td>t</td>
<td>Time (s)</td>
</tr>
<tr>
<td>Rx</td>
<td>Removal efficiency (%)</td>
</tr>
<tr>
<td>Vₜ₁ᵖ</td>
<td>Tip or tangential velocity (ms⁻¹)</td>
</tr>
<tr>
<td>σ</td>
<td>Global hydrodynamic stress (Nm⁻²)</td>
</tr>
<tr>
<td>F₉</td>
<td>Aggregate binding or cohesive force (N)</td>
</tr>
<tr>
<td>τ</td>
<td>Aggregate binding or cohesive strength (Nm⁻²)</td>
</tr>
</tbody>
</table>

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INTRODUCTION

Effective management of concentrated slurries from minerals processing, oil and gas production, water and wastewater treatment, and other process industries is very important from an environmental and economic perspective. This constitutes one of the biggest challenges facing the operators in these sectors, especially in their downstream operations. Typically, a large proportion of sludge management costs are related to downstream processes such as solid–liquid separation and dewatering prior to eventual disposal or utilisation. Improvement in sludge processing can serve as a tool for improving environmental quality, and can contribute to natural resource conservation through water re-use, nutrient recycling and biomass utilisation.

The aggregation of fine particles in flowing suspension is mainly caused by the effects of Brownian motion (perikinetic), turbulence-induced shear or velocity gradient (orthokinetic) and differential settling (Tsai et al. 1987; Lick et al. 1992, 1993; Atkinson et al. 2005; Van Leussen 2011). Of these mechanisms, turbulence is the most complex and least well understood (Gregory 2013; Argyropoulos & Markatos 2015). Turbulence-induced orthokinetic flocculation is typically observed in many engineered and natural flow conditions such as pipe flow, open channel flow, stirred tanks, and bioreactors. However, progress in this field of research relies on the development of a new generation of shear reactors based on optimised physicochemical and process engineering conditions that are tailored to the specific processing equipment. In order to make a significant breakthrough in this field, more work is needed to improve our understanding of the complex fluid–particle interactions in these reactors so as to take full advantage of their potential for improved solid–liquid separation. Pelleting flocculation, an offshoot of classical orthokinetic flocculation, is an effective structure formation process based on the ‘meta-stable state’ concept, and has been recognised by many researchers as one of the most promising solid–liquid separation techniques (Higashitani & Kubota 1987; Tambo 1990; Amirtharajah & Tambo 1991; Hemme et al. 1995; Walaszek & Ay 2005).

The principal factors governing the pelleting flocculation process have been identified as suspended solids concentration, the stirrer-vessel system (stirrer configuration and geometry), stirrer rotation speed, and mixing time (Yusa 1987).

Process optimisation is therefore a crucial step in the design, development and testing of new solid-liquid separation equipment (Wakeman et al. 1989). It involves the systematic investigation and control of a number of process variables that influence the slurry pre-treatment and agglomeration processes. In conducting such optimisation studies, the different scales, which range from bench to pilot and full-scale, must be considered. An apparatus constructed on a small scale offers the flexibility required to evaluate or simulate a process at a reasonable cost to the extent that is desired (Yusa 1987; Hess et al. 1997).

Aggregation process in sheared suspensions

A shear reactor or flocculator is typically used in laboratory flocculation experiments in which the shear rate must be precisely determined and strictly controlled. In most of these devices, the flow is largely turbulent (Kratzer et al. 1995; Farrow & Swift 1996; Svarovsky 2000; Owen et al. 2008; Sievers et al. 2008; Concha 2014). The mechanism of particle aggregation under laminar and turbulent flow conditions has been the subject of many scientific publications. The rate at which this occurs depends to a great extent on the motion of the suspended particles relative to the motion of the fluid in the developed turbulent flow field (Shamlou & Hooker-Titchener 1993).

The formation of aggregates in any turbulence or shear-induced flocculation essentially consists of destabilisation; collision and adhesion; floc growth and deformation phases (Hogg 2000). Owing to the dynamic nature of several micro-processes that play key roles in the formation of aggregates, the knowledge of polymer–particle interactions (addition profiles and mixing conditions) is indispensable (Farinato et al. 1993). Ives (1984) and more recently Bergenstahl (1995) expressed the flocculation rate in terms of the change in the initial particle concentration for both Couette and shear-induced turbulent flow conditions, while Attia (1992) provides a theoretical estimate of the floc growth rate (Equations (1) and (2)).

\[
\frac{-dN}{dt_{(\text{Couette})}} = \frac{2}{3} f_0 G d^2 N^2 + \frac{4}{3} f_p \frac{kTN^2}{\mu} \tag{1}
\]
\[
\frac{dN}{dt_{\text{Turb.}}} = -6\pi\beta\sqrt{\frac{\text{Intensity}}{\eta}} r^2 N^2
\] (2)

The growth and stability of flocs under laminar and turbulent flow conditions is strongly influenced by the hydrodynamic conditions in the flocculation unit. Previous studies by Coufort et al. (2005), Bouyer et al. (2005a) and Boyle et al. (2005) in their study of orthokinetic flocculation in a Taylor-Couette reactor, 1 litre-jar, and cone-and-plate device (CPD), respectively, pointed out that the hydrodynamic condition is a function of both the geometry of the flocculation reactor as well as the stirrer speed and its configuration.

Increasing power input into the reactor speeds up the flocculation rate, with the floc growth limited only by the increasing hydrodynamic viscous stress that tends to disrupt the flocs (Rulyov 2010). Dobias & Von Rybinski (1999), on the basis of similar empirical studies by Mühle (1993), Mühle & Domasch (1990), as well as Jarvis et al. (2005) and Rulyov (2010), provided a theoretical estimate of the maximum floc size for laminar and turbulent flow conditions (Equations (3) and (4)), although the expression for laminar flow condition is more accurate owing to the complex nature of the turbulent flow.

\[
D_{F, \text{max}} = \frac{6U(1 - p)^{\frac{1}{2}}}{\pi^2 KGd_p^2}
\] (3)

\[
d_{F, \text{max}} = \frac{(F_B/d_p)}{\rho Gv}
\] (4)

The size of aggregates varies from molecular dimensions to a range that is visible to the unaided eye, with the smaller sizes being associated with the primary particles of diameter \(d_p\), while the largest size \(d_{F, \text{max}}\) is determined by the balance of floc growth and rupture within the fluid (Spicer & Pratsinis 1996; Bache 2004; Bridgeman et al. 2009; Partheniades 2009). Therefore, in the case of floc growth beyond the initial swift phase, further attachment of particles or microflocs to the larger agglomerates is prevented, as they are more susceptible to breakage or the existing agglomerates are destroyed by the increasing hydrodynamic force \(F_{H}\) (Gregory 2006; Van Leussen 2011; He et al. 2015). Hogg (2005) suggested that floc growth and breakage occur simultaneously: growing flocs are subject to breakage while fragments of broken flocs undergo reagglomeration until a levelling off of the floc sizes at steady state occurs (Patrick & Pratsinis 1996). In the viscous subrange, Lu et al. (1998) expressed this phenomenon in terms of the kinetic equation of flocculation. Their model is presented in Equation (5). The maximum limit of floc growth is attained when the hydrodynamic force is equal to the floc cohesive or binding force \(F_B\).

\[
\frac{dN}{dt} = -\frac{2}{\sqrt{\pi}} \frac{d_p^{\frac{3}{2}}}{cGv} \frac{N^2}{\sigma_s} + \beta_1 \frac{r_s}{d_p} \left( \frac{d_p}{d_f} \right)^{\frac{1}{2}} d_B N_B
\] (5)

where \(A\) and \(B\) denote smaller and larger particles, respectively.

Several theoretical and empirical models have been proposed to account for the hydrodynamic stress exerted on particle agglomerates and their cohesive strength (Kobayashi et al. 1999; Boyle et al. 2005; Yuan & Farnood 2010). Bouyer et al. (2005b) and Thomas et al. (2013) in their study of hydrodynamics in a jar test and swirl flow reactor, respectively, as well as in other similar contributions elsewhere (Mhaisalkar et al. 1986; Edzwald et al. 1997; Lu et al. 1998; Bridgeman et al. 2009), indicated a number of expressions that are widely used to characterise the hydrodynamics and turbulent flow phenomena typically encountered in most shear reactors (Equations (6)–(14)).

The turbulence parameters of interest in any agitated system (with respect to the mixer capacity and performance) are the Kolmogorov microscale \(\lambda\) and the turbulence kinetic energy dissipation rate \(\epsilon\) (Wu & Patterson 1989). The Kolmogorov length scale is an indicator of the rate of micromixing and mode of aggregate deformation under turbulent flow conditions. The energy dissipation rate is the rate of dissipation of kinetic energy from larger to smaller eddies due to the shearing action of the fluid motion (Thoenes 1998). In mixing experiments performed under laboratory conditions (\(D < 1\) m), the micro-scale of turbulence predominates (Zlokarnik 2008). Under such conditions, particle collisions are promoted by an eddy size similar to those of the colliding particles (Thomas et al. 1999).

\[
P = T_{\phi\theta} = \frac{2\pi n T_s}{60}
\] (6)

\(\Phi\theta\) is the angular velocity of the impeller and \(n\) is the rotation speed of the impeller.
The mechanism of floc break up under turbulent flow conditions has also been the subject of numerous publications. Based on the empirical models of floc formation and destruction presented by Mühle (1993), Dobias & Von Rybinski (1999), and Peng & Williams (1993), Thomas et al. (1999) identified floc splitting or fracture, and surface erosion as the main mechanisms of floc deformation. In a similar study by Boyle et al. (2005), erosion and bulk rupture were reported as the principal mechanisms of agglomerate cohesive failure. A comprehensive review of floc stability in turbulent flow under the influence of hydrodynamic stress is, however, beyond the scope of the present communication, as the topic has received considerable attention in a number of similar contributions (Parker et al. 1972; Leentvaar & Rebhun 1983; Baldyga & Bourne 1984; François 1987; Mühle & Domasch 1991; Partheniades 1993; Serra et al. 2008).

In the past, shear flocculation studies were mostly confined to the use of existing apparatuses such as Taylor-Couette devices, stirred tank reactors or pipe flocculators with well-defined geometry and stirrer configurations. There have been few studies on unconventional stirrer-vessel systems (geometry and stirrer types), partly due to the extreme complexities of the flow structure in such devices. The aim of this study is to fill this gap in knowledge by investigating the performance of a new slurry pre-treatment technique using a shear flocculator with an alternative geometry and stirrer configuration under different operating conditions. All experiments were performed in a single-cell rotating plate batch flow vortex reactor system with kaolin slurry as the substrate. The assessment of the aggregation process was based on indirect performance indicators such as turbidity, dry solids content, residual charge and particle removal efficiency $R_e$.

**MATERIALS AND METHODS**

**Batch flow vortex reactor**

The batch flow flocculator, fabricated from Plexiglass, consists of a single-cell reactor with rotor-stator configuration, and was used to investigate a patented sludge pre-treatment technique (Oyegbile et al. 2014). The outer wall consists of a semi-circular cross-section that acts as the stator, and axially revolving plates that act as the rotor, which induce flow in an anticlockwise direction. The angular rotation is achieved by means of a motor attached to the outer plate, and includes a provision for making speed adjustments. The torque and rotation speed (rpm) measurements were taken directly from the built-in digital display attached to the electric motor. The suspension is loaded to fill one-half of the reactor volume when operated in batch and quasi-continuous mode, such that the pelletisation takes place in this area only. The revolving plates are perforated at the centre to the outer edge in order to facilitate cleaning of the device and the removal of the supernatant and pellet flocs. The dosing and mixing of the suspension as well as subsequent pelletisation take place in this reactor. This patented device was used to simulate the rotor-stator principle by using the wall of the reactor as the stator and the revolving plate as the rotor (Figure 1; Table 1).
Agglomerate preparation and characterisation

A number of synthetic polymers of high molecular weights and low charge densities: Superfloc® C-492 and N-300 (Kemira Oyj, Finland) were selected based on preliminary screening and physicochemical evaluation and used as bridging materials in this investigation. The working slurry, a poly-dispersed kaolin suspension with characteristics described in detail elsewhere (Oyegbile et al. 2016), was pre-treated in the shear device by adding 8 ml each of dilute solution of polyelectrolyte with different concentrations (0.2, 0.3, 0.4, 0.5 and 0.6 g/l), after which the suspension was continuously stirred for a sufficient time to allow for the pellet floc formation. Subsequently, the flocs were separated from the supernatant by gravity drainage through a 0.5 mm diameter sieve, and then dried in a thermostatically controlled oven with forced air ventilation at 105°C for 24 hours to a constant mass (EN 15934 2012). The solids content of the pellet flocs and the particle removal efficiency $R_x$, which is a measure of the efficiency of particle removal from the slurry to the solids fraction, were then determined according to Equation (15).

$$R(x) = \frac{\text{Dry mass of pellet flocs filtered through 0.5 mm sieve}}{\text{Dry mass of kaolin particles in suspension}} \times 100\%$$  

(15)

Characterisation of the reactor supernatant

The working slurry, a polydispersed kaolin suspension, was first pre-conditioned in the reactor by the addition of suitable flocculants as bridging materials. A sequential dual-addition of a cationic high molecular weight with low charge activity and a high molecular weight non-ionic polymer at a mixing speed of 155 rpm, 175 rpm, and 195 rpm for about 45 s was followed by a further mixing time of 5, 10, 15 and 20 minutes at slower speed of 145 rpm, 165 rpm, and 185 rpm, respectively.

Thereafter, samples of the supernatant were withdrawn after 2 minutes of settling for turbidity and residual charge measurements using PCD 03 (BTG Instruments GmbH, Herrsching, Germany) and Turbiquant® 3,000IR (Merck Millipore GmbH, Hessen, Germany), respectively. The experiments were performed with varying total polymer doses of 3, 4 and 5 kg/t total solids (TS). This dual-additions scheme was found to provide better aggregates in terms of structural attributes such as shape, size distribution, compactness, and shear resistance.

RESULTS AND DISCUSSION

Optimisation of separation process variables

In optimising the processing technique, the main aim is to realise the rolling and collisional effects necessary for the pelleting flocculation process by generating a Taylor-like vortex within the rotor-stator plates. The agglomeration

Table 1 | Technical data and operating conditions

| Reactor dimension | 180 × 180 × 60 mm |
| Construction material | Plexiglass |
| Operational mode | Discontinuous–batch, continuous |
| Agitation speed | 145–185 rpm |
| Polymer type | HMW cationic & non-ionic polymers |
| Polymer dosage | 3–5 kg/t TS |
| Wall-plate gap | 3–6 mm |
| Residence time | 5–20 minutes |
| Slurry concentration | 1–4% |
| Separation method | Gravity dewatering (φ < 0.5 mm) |

process was randomly optimised for a number of process variables and reference points using factorial experimental design to select the optimisation points (Table 2). Several batch flocculation and gravity dewatering experiments were performed over a range of different process conditions to determine a number of optimum reference points in which the process conditions (physicochemical and hydrodynamics) will result in the formation of pellet-like aggregates. A critical mixing speed of 165 rpm was observed, above which there was a significant rupture of the pellet flocs. This trend was seen to intensify with increased mixing speed.

### Hydrodynamic characteristics and flow pattern

In both laminar and turbulent flow conditions, the effectiveness of the flocculation process and the stability of the floc is controlled by two principal forces: hydrodynamic and cohesive or binding force (Gregory 2006; Carissimi & Rubio 2015). The former is determined by the turbulence conditions, whereas the latter depends on both the physicochemical conditions and structure of the flocs. The hydrodynamic parameters for the reactor are given in Table 3. Power input into the reactor is determined from torque, measured directly from the shaft of the motor attached to the shearing device that provides the axial rotation.

The stream pattern of the turbulent flow, which can be characterised by the velocity and eddy scale (Boller & Blaser 1998), and its effect on the pre-treatment process was observed both visually and with the aid of a digital camera. The observations showed that the path of the agglomerates’ motion aligned closely with that of the flow stream within the developed Taylor-like vortex flow field. The turbulent flow in the reactor results in the formation of a spiral forced vortex flow on either side of the reactor, providing both the necessary rotational and collisional effect as well as enhanced mass and energy transport, which are crucial for the formation of pellet flocs as depicted in Figure 2 (Visscher et al. 2013).

### Effects of micro-processes on the process efficiency

The effect of the micro-processes on the flocculation performance was investigated by assessing the clarity of the reactor supernatant and the particle removal efficiency \( R \), after gravity dewatering at intervals of 5, 10, 15 and 20 minutes. The results of the observed turbidity values and the particle removal rate at time intervals of 5, 10, 15 and 20

### Table 2 | Process optimisation parameters and reference points

<table>
<thead>
<tr>
<th>Operational parameters</th>
<th>Reference points</th>
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<tbody>
<tr>
<td>Polymer type</td>
<td>Cationic, Anionic, Cationic-Non-ionic, Anionic-Non-ionic</td>
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<td>Polymer dose (kg/t TS)</td>
<td>2, 3, 4</td>
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<tr>
<td>Dosing sequence</td>
<td>Once, Twice, Thrice</td>
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<tr>
<td>Slurry concentration (% wt.)</td>
<td>1, 2, 3, 4</td>
</tr>
<tr>
<td>Filling mode</td>
<td>Batch, Quasi-continuous, Fully continuous</td>
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<tr>
<td>Wall-plate gap (mm)</td>
<td>3, 4, 5</td>
</tr>
<tr>
<td>Mixing speed (min⁻¹)</td>
<td>125, 145, 165, 185</td>
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<tr>
<td>Residence time (min)</td>
<td>5, 10, 15, 20</td>
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</table>

### Table 3 | Hydrodynamic characteristic values as a function of the agitation speed

<table>
<thead>
<tr>
<th>Hydrodynamic parameters</th>
<th>Mixing speed (rpm)</th>
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<tr>
<td></td>
<td>145</td>
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<tr>
<td>( T_r ) (Nm)</td>
<td>0.229</td>
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<tr>
<td>( P ) (W)</td>
<td>3.48</td>
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<tr>
<td>( R_f ) (s⁻¹)</td>
<td>48,077</td>
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<tr>
<td>( N_p ) (s⁻¹)</td>
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<td>( G ) (s⁻¹)</td>
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<tr>
<td>( e ) (m² s⁻³)</td>
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<td>( \lambda ) (μm)</td>
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</tr>
<tr>
<td>( V_w ) (ms⁻¹)</td>
<td>1.0254</td>
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<tr>
<td>( \sigma ) (Nm⁻²)</td>
<td>5.7615</td>
</tr>
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</table>
Figure 2 | Relative motion of aggregates and fluid in the developed turbulent vortex flow field.

Figure 3 | Relationship between residence time and process performance at an operating speed of 145 rpm and polymer dose of 3 kg/g TS.

Figure 4 | Relationship between residence time and process performance at an operating speed of 145 rpm and polymer dose of 4 kg/g TS.
The results of the residual turbidity show a gradual decrease with residence time and agitation speed, whereas the minimum residual turbidity increases with polymer dosage. The lowest turbidity of 7 NTU and 25.2 NTU were obtained with a polymer dose of 3 kg/t TS at agitation speeds of 145 and 165 rpm. This optimum polymer dose shows a good correlation with the physicochemical optimisation conducted prior to this study (Oyegbile et al. 2016).

The observed trend in the turbidity measurements with respect to residence time can be attributed to an increased floc attachment and growth that allows micro-particles to be incorporated onto the macro-flocs, which in turn results in lower turbidity. With respect to the agitation speed, the observed reduction in turbidity can be attributed to the increased flocculation kinetics. It has been reported that increases in agitation speed also increase the shear rate, which promotes particle–particle contacts and floc growth up to the steady state condition, and results in improved particle aggregation and lower turbidity (McConnachie 1991; Patrick & Pratsinis 1996; Rulyov 2010). However, in terms of the polymer dosage, a higher polymer dose appears to correlate well with an increase in turbidity of the supernatant for all the polymers. The increase in turbidity at a higher dosage can be attributed to unadsorbed polymer molecules that remain in solution due to increasing polymer adsorption beyond the optimum. This results in lower particle aggregation due to the weakening of the polymer bond (Besra et al. 2002; Betatache et al. 2014).
The observed values of the removal efficiency $R_x$ generally show a gradual reduction with residence time and operating speed except in Figure 3, Figure 4, and Figure 8. In these cases, the removal efficiency shows a gradual increase at a residence time of 20 minutes. This observation can be attributed to the fact that in these cases the compact agglomerate undergoes a gradual deformation with time as the floc size increases at a constant shear rate as shown by the increase in the proportion of fines ($<0.5\, \text{mm}$) in the reactor system (Wu & Patterson 1989; Yeung & Pelton 1996; He et al. 2015). While this trend does not seem to correlate with the trend in the turbidity measurements, the observed variation might be attributed to the fact that the eroded micro-particles were not effectively captured by the sieve, giving a lower than expected removal efficiency.

The residual charge in the supernatant remains negative irrespective of the operating speed and polymer dose. This negative charge is due to the partial theoretical charge neutralisation of approximately 35.6%, 52.5%, and 69.4% for polymer doses of 3 kg/t, 4 kg/t, and 5 kg/t TS, respectively. This confirms the theory that an optimum flocculation performance does not entail a complete charge reversal even in the case of dual polymer additions (Lee & Liu 2000).

**Effects of micro-processes on the solids fraction**

The effect of the micro-processes on the solid–liquid separation efficiency by simple gravity dewatering was investigated by determining the solids content of the pellet flocs. The results of the agglomerates’ solids content at time intervals of 5, 10, 15 and 20 minutes are presented in Figures 9 and 10.

The dry solids content of the agglomerates does not appear to show a wide variation with respect to the investigated process conditions once the pellet flocs are fully formed after 5 minutes. (The highest solids content of 28.3% was recorded after 15 minutes in all experimental conditions.) However, the highest solids content of 27.7% and 28.3% were obtained at a polymer dose of 3 kg/t TS and agitation speeds of 145 rpm and 165 rpm, respectively. This seems to confirm a polymer dose of 3 kg/t TS as the optimum (Lee & Liu 2000).

In contrast, there seems to be no correlation between solids content and removal efficiency $R_x$; in fact, contrary to what might be expected at a lower solids content, there is a higher removal efficiency. One possible explanation for this might be that the pellet flocs with higher removal efficiency are more porous and less dense than those with lower removal efficiency. This would explain the reduced solids content after thermal treatment to remove the pore water. In addition, it has been reported that an increase in pore space or porosity leads to a decrease in agglomerate density (Walaszek & Ay 2005).

The observed process efficiency for this pre-treatment technique appears to compare favourably with conventional solid–liquid separation techniques (Mahmoud et al. 2013; Carissimi & Rubio 2015). In addition, the optimum polymer
dosage values of 3 kg/t TS obtained in this investigation appear to be much lower than those reported elsewhere, although a different substrate was employed for their study (Sievers et al. 2008).

CONCLUSIONS

This paper presents an evaluation of the effect of microhydrodynamics and physicochemical micro-processes on the aggregation of a model suspension. Results of the investigation showed that the effectiveness of the aggregation process is highly dependent on an optimum combination of several process variables such as the polymer type, concentration and dosing regimen, agitator rotation speed, wall-plate gap distance, residence time, suspension loading method, reactor geometry and stirrer type. Floc splitting, fracture or bulk rupture was identified as the main floc deformation mechanism as shown by lower particle removal efficiency $R_x$ with respect to residence time. The highest particle removal efficiency of 97.5% was attained at mixing speeds of both 145 and 165 rpm. These data confirm that optimised micro-processes during the aggregation of slurry are a crucial step in the formation of compact pellet agglomerates. It is anticipated that future studies will address the process performance and flow characterisation for a continuous reactor system.

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