OPTIMUM DESIGN OF YOUR CENTER WELL: USE OF A CFD MODEL TO UNDERSTAND THE BALANCE BETWEEN FLOCCULATION AND IMPROVED HYDRODYNAMICS

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ABSTRACT

The use of center wells (CWs) in circular secondary settling tanks (SST) has become a common practice. According to Parker et al. (1996) the main purpose of the CW is to promote the aggregation of remaining dispersed particles into settleable flocs, i.e., to promote the flocculation process inside the settling tank. Parker et al. (1996) and Parker and Stenquist (1986) showed strong evidence that, under similar operating conditions, circular tanks equipped with CWs can achieve lower effluent suspended solids and tend to perform better than tanks without a CW. A large study presented by Wahlberg et al. (1994) concluded that a residence time of about 20 minutes in a flocculation zone inside a SST can lead to more than 90% of the achievable flocculation. This 20 minute residence time, in conjunction with the average dry weather flow and a 50% return activated sludge, has been use as rule of thumb to determine the volume of the center well (WEF, 2005).

Merrill et al. (1992) used a two-dimensional (2-D) hydrodynamic clarifier model to optimize the geometry of the CW in a circular SST. The numerical model did not include flocculation in the CW; however, Merrill and co-workers found that the optimum placement coincided with the optimum diameter recommended for flocculation. These results open the discussion about the main role of the CW. Is it to promote flocculation or to improve the tank hydrodynamics? This paper uses a computational fluid dynamics model to help answer this question. Furthermore, this numerical model is used to better understand the effects that the geometry of the CW has in the hydrodynamics of the SST, and how this geometry affects the flocculation process.

Results demonstrate that the CW promotes the aggregation of unflocculated particles and this effect has a major effect on the clarifier performance. However, an even greater benefit of the CW is the improvement of the tank hydrodynamics. The CW helps to control the re-entrainment of clarified fluid with the influent flow, reducing the strength of the upflow current close to the launder. Sensitivity analyses demonstrate that the optimum dimension of the CW is about 20 to 35% of the total clarifier diameter. The incoming MLSS does not seem to have a major effect on the optimum dimension, while it appears that the optimum dimension tends to increase as the surface overflow rate increases.

KEYWORDS

Secondary Settling Tank, flocculation, modeling, computational fluid dynamics.
INTRODUCTION

The effects of flocculation in clarifier performance and the causes that promote flocculation in settling tanks have been largely acknowledged. Camp (1945) recognized that flocculation in settling tanks is due to two causes:

1. Differences in the settling velocities of the particles whereby faster settling particles overtake those which settle more slowly and coalesce with them; and
2. Velocity gradients in the liquid, which cause particles in a region of higher velocity to overtake those in adjacent stream paths moving at slower velocity.

The differential settling and the flocculation due to velocity gradients are referred to as orthokinetic flocculation. The orthokinetic flocculation is also known as macroflocculation since it mostly affects the aggregation of particles greater than about 1 μm (Tchobanoglous et al. 2003). The term microflocculation is used to refer to the aggregation of smaller particles (from about 0.001 to 1 μm) that are brought together through the random thermal “Brownian” motion of the particles; this is commonly referred to as perikinetic flocculation.

Parker et al. (1970, 1971, 1972) demonstrated the utility of a flocculation zone previous to the final settling zone. They showed that often, the highly turbulent condition in the aeration chamber is so intense that it favors floc breakup over aggregation, resulting in a high level of dispersed solids (similar results were found by Starkey and Karr, 1984; Das et al., 1993; Wahlberg et al., 1994). They recommended additional flow conditioning, through the incorporation of a mildly stirred flocculation step between the aeration basin and the clarifier to promote the incorporation of dispersed particles into the floc. This practice became popular to improve the final effluent of attached growth systems, e.g. trickling filter-solids contact processes (TF/SC). The idea has been supported by other researchers (Das et al., 1993; Wahlberg et al., 1994; La Motta et al., 2003; Jimenez et al. 2003). Wahlberg et al. presented an evaluation of 21 full-scale wastewater treatment plants (WWTP) that showed that the provision of additional flocculation would reduce the supernatant SS (SSS).

The inclusion of flocculation zones inside the SSTs is a relatively new practice. Knop (1966) reported, on the basis of pilot plant studies, that the placement of a flocculator in the inlet zone of a rectangular SST improved effluent turbidity. Lately, a full-scale plant was constructed including two sets of paddle flocculators. The induced flocculation in the clarifier improved the effluent SS by about 5 mg/L. Flocculation wells (FWs) were initially operated with mixers to impart G values of 20 – 70 s⁻¹, levels that had been found optimal in bench scale research (Parker et al., 1970, 1971). Parallel operation of SSTs showed that turning the mixer off had no effect on effluent quality (Ekama et al., 1997). This can be explained by the favorable G values present in the flocculation well produced by conversion of kinetic energy into turbulent kinetic energy.

The use of center wells (CWs) in circular secondary settling tanks (SST) has become a common practice. According to Parker et al. (1996) the main purpose of the CW is to promote the aggregation of remaining dispersed particle into settleable flocs, i.e., to promote the flocculation process inside the settling tanks. Parker et al. (1996) and Parker and Stenquist (1986) showed strong evidence that, under similar operating conditions, circular tanks equipped with CWs can
achieve lower effluent suspended solids (ESS) and tend to perform better than tanks without a CW. A large study presented by Wahlberg et al. (1994) concluded that a residence time of about 20 minutes in a flocculation zone inside a SST can lead to more than 90% of the achievable flocculation. This 20 minute residence time, in conjunction with the average dry weather flow (ADWF) and a 50% return activated sludge, has been use as rule of thumb to determine the volume of the center well (WEF, 2005).

Merrill et al. (1992) used a two-dimensional (2-D) hydrodynamic clarifier model to optimize the geometry of the CW in a circular SST. The numerical model did not include flocculation in the CW; however, Merrill et al. found that the optimum placement coincided with the optimum diameter recommended for flocculation. These results open the discussion about the main role of the CW. Is it to promote flocculation or to improve the hydrodynamics of the tank? In this paper a computational fluid dynamics (CFD) model is used to help answer this question. Furthermore, this numerical model is used to better understand the effects that the geometry of the CW has in the hydrodynamics of the SST, and how this geometry affects the flocculation process. In general the objectives of the paper are: (1) present the use of a CFD model to estimate the achievable flocculation in a SST; (2) determine the optimum placement of the CW using a 2-D hydrodynamic model that includes the flocculation process.

**Flocculation Models**

The net flocculation in a turbulent environment depends upon the balance of the opposing processes of aggregation and floc breakup (Parker et al., 1972; Oles, 1992; Spicer and Pratsinis, 1996; Serra and Casamitjana, 1998). At steady state, a successful flocculation model should include the balance between particle growth and breakage. Spicer and Pratsinis (1996) described the dynamics of floc formation; they indicated that several phases of floc growth occur during flocculation. Initially floc growth is dominant, particles combine in the presence of polymers (coagulation), and their size increases rapidly. As the floc grows larger, porous and open structures are formed that are more susceptible to fragmentation by fluid shear. Parker et al. (1972), Galil et al. (1991) and Biggs and Lant (2000) found that the floc size for activated sludge decreased with increasing shear; similar results have been found in the flocculation of inorganic systems (Spicer and Pratsinis, 1996; Serra and Casamitjana, 1998).

Parker et al. (1971) identified two breaking mechanisms: 1) floc breakup as a result of erosion caused by surface shearing forces exceeding the shear strength of the bonds joining the primary particles to the floc thus releasing primary particles in the suspension; and 2) floc breakup as a result of filament fracture that occurs when excessive tensile stresses are applied on the floc (which produces fragmentation of the floc instead of primary particles).

Based on a detailed theoretical analysis Parker et al. (1970, 1971) developed a differential equation describing the overall kinetics of flocculation in turbulent mixing:

\[
\frac{dn}{dt} = K_B \cdot X \cdot G^m - K_A \cdot X \cdot n \cdot G
\]
where $X$ is the MLSS concentration (g/L), $G$ the root-mean-square velocity gradient ($s^{-1}$), $K_A$ a floc aggregation coefficient (L/g), $K_B$ a floc breakup rate coefficient (number $s^{m-1}/g$), $m$ the floc breakup rate exponent (dimensionless), and $n$ is the primary particle number concentration (numbers/L). Parker’s experimental primary particle concentration was based on the weight concentration of SS in the supernatant after 30 minutes of settling. They performed a series of flocculation tests in a continuous-flow reactor (CFSTR) to support their development. A mass balance of primary particles for a CFSTR without recycle yields at steady state the following equation:

$$\frac{n_0}{n_t} = \frac{\left[1 + K_A \cdot X \cdot G \cdot t\right]}{\left[1 + K_B \cdot D \cdot G^m \cdot t\right]}$$

Other researchers have supported the development presented by Parker and his co-workers. Wahlberg et al. (1994) presented an integrated form of Equation 1 for the calculation of flocculation in a batch flocculator. Wahlberg et al. assumed $K_A$ and $K_B$ as true constants and used a value of $m$ equal to 2 [\(m = 2\)] was selected based on analysis presented by Parker et al. (1971). This number indicates that floc breakup occurs by erosion of primary particles from floc surfaces due to eddies in the viscous dissipation range:

$$n_t = \frac{K_B \cdot G}{K_A} + \left(n_o - \frac{K_B \cdot G}{K_A}\right) \cdot e^{-K_A \cdot X \cdot G \cdot t}$$

Setting $G$ as a constant and for a given MLSS concentration, Equation 3 can be expressed in the form:

$$n_t = \alpha + \beta e^{\lambda t}$$

in which $\alpha = K_B \cdot G/K_A$ is the equilibrium primary particle number concentration (number/L), $\beta = n_o - \alpha$, is the difference between the initial and equilibrium primary particle number concentration, and $\lambda = K_A \cdot X \cdot G$ is the overall primary particle removal rate ($s^{-1}$).

Wahlberg et al. (1994) tested Equations 3 and 4 with activated sludges obtained at 21 full-scale facilities with different aeration methods. They measured the primary particle concentration as the turbidity of the supernatant after 30 minutes of settling; turbidity was later correlated to SS mass concentration. They observed that the flocculation data was well described by the curves, concluding the applicability of the theoretical development of Parker et al. (1970, 1971) to the description of batch flocculation data. Their study indicates that 99% flocculation was achieved within 10 minutes under batch conditions for most sludge. They concluded that a similar performance improvement could be obtained in the field using a completely-mixed flocculation zone with a residence time of at least 20 minutes. The authors reported values of $K_A$, $K_B$, $n_o$, $\alpha$, $\beta$ and $\lambda$ for 30 activated sludge samples.
Jimenez (2002) and La Motta et al. (2003) used equations similar to Equation 2 and 4 to evaluate the removal of suspended solids (SS) and Particulate Chemical Oxygen Demand (PCOD) in continuous flow and batch flocculators. For a batch reactor, operated a constant $G$, they presented:

$$C = a + (C_0 - a) \cdot e^{-k_F t \cdot X} \quad \text{5}$$

where $C$ (mg/L) is the concentration of unflocculated particles remaining in the supernatant at reaction time $t$ (min) after 30 minutes settling, $a$ is the residual concentration of particles (mg/L), $k_F$ is the reaction rate coefficient, $C_0$ is the initial concentration of influent particles (mg/L), and $X$ is the MLSS concentration (mg/L). Batch experiments and curve fitting procedures can be used to find the flocculation constants $K_A$ and $K_B$ using Equations 4 or 5.

As indicated before, the previous models describe the kinetics of flocculation in a turbulent mixed environment, and therefore they do not include the possible effect of differential settling flocculation. Griborio (2004) presented the development of a differential equation predicting the rate of change of primary particles into flocs due to differential settling. This Equation is presented in the next section.

**METHODS**

The 2-D hydrodynamic model used in this research was developed at the University of New Orleans (UNO) with the financial support of the U.S. Environmental Protection Agency (EPA). Details about the model development, physics and features can be found in Griborio (2004), Griborio and McCorquodale (2005) and McCorquodale et al. (2005). Among other features, this model includes flocculation due to velocity gradient, differential settling and filtration in the sludge blanket. The next section provides a brief description about the development of the flocculation sub-model.

**Shear Induced Flocculation**

The differential equation (Eq. 1) presented by Parker et al. (1970, 1971) is used to model the floc aggregation and break up in a turbulent environment. The studies of Wahlberg et al. (1994), Jimenez (2002), and La Motta et al. (2003) support the use of this model. Beyond the fact that this model has been used widely; it was selected on the basis of its simplicity and kinetic constant documentation. For implementation purposes, the parameter $n$ in Eq. 1 was taken as the concentration of unflocculated-primary particles. Griborio (2004) described a simple procedure to obtain the concentration of unflocculated-primary particles and flocculated flocs-particles for a MLSS sample.

The average $G$ values in the center well are calculated based on the total energy dissipation. Eq. 6 predicts the average $G$ values in the center well based on the inlet kinetic energy (Camp and Stein, 1943):
\[ G = \frac{P}{V\mu} \approx \frac{u_o}{\sqrt{2i\nu}} \]

Since the \( G \) values in the inlet jet can be considerably higher than the average value predicted by Eq. 6, the \( G \) values in the inlet jet are adjusted depending on the inlet height \( (B) \) and the horizontal distance from the inlet \( (X) \) using an equation for a plane jet proposed by Larsen (1977):

\[ G = 1.53 \nu^{-1/2} B^{3/4} u_o^{3/2} X^{-5/4} \]

Velocity gradient values outside of the center well or outside of the inlet jet influences are probably too small to induce shear flocculation. However, these values are incorporated in the model using a general definition of \( G \):

\[ G = \sqrt{\left(\frac{\partial u}{\partial y}\right)^2 + \left(\frac{\partial v}{\partial r}\right)^2} \]

In Equations 6, 7 and 8; \( P \) is the total energy loss or the power imparted to the water; \( V \) is the center well volume; \( \mu \) is the absolute viscosity of the fluid; \( \nu \) is the kinematic viscosity; \( \bar{t} \) is the hydraulic retention time in the center well; \( u_o \) is the inlet velocity; and \( u \) and \( v \) are the velocities in the \( y \)-vertical and \( r \)-radial directions, respectively.

**Differential Settling Flocculation**

The differential settling flocculation is due to the differences in the settling velocities of the particles whereby large particles overtake smaller particles during the settling process. The particles coalesce increasing the mass and settling at a faster rate. McCorquodale et al. (2004) developed a differential equation predicting the rate of change of primary particles into flocs due to differential settling:

\[ \frac{dC_1}{dt} = -\frac{9}{4} k_{ds} \frac{C_1 C_2}{\rho_1 \rho_2} \left(1 + \frac{d_1}{d_2}\right)^2 \frac{C_1}{d_1} \left(V_{S2} - V_{S1}\right) \]

In Equation 9, \( C_1 \) and \( C_2 \) are the concentrations of unflocculated–primary and flocculated–flocs particles, respectively; \( d_1 \) and \( d_2 \) are the cross sectional diameters of unflocculated and flocculated particles, respectively; \( \rho_1 \) and \( \rho_2 \) are the densities of the primary and flocculated particles, respectively; \( t \) is time; \( k_{ds} \) is a kinetic constant between 1 and 2 accounting for the increase in the rate of collision due to the turbulence in the flow; and \( V_{S1} \) and \( V_{S2} \) are the settling velocities of the primary particles and flocs, respectively.

The cross sectional diameter is a physical characteristic of the floc that requires sophisticated techniques for its determination. These techniques are not available in most wastewater treatment

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installations. To avoid this limitation, the determination of the cross sectional diameter is based on the equation proposed by Li and Ganczarczyk (1987) that relates the settling velocity of the individual activated sludge floc to the cross sectional diameter:

\[ V_{si}(mm/s) = 0.35 + 1.77d_{pi}(mm) \]

where \( V_{si} \) is the settling velocity of the \( i \) floc class, and \( d_{pi} \) is the cross sectional diameter of the \( i \) floc class. Table 1 presents an estimation of the cross sectional diameter for different types of flocs based on their settling velocities.

**Table 1 - Settling Velocities and Cross Sectional Diameters for Different Particles**

<table>
<thead>
<tr>
<th>Type of Particle</th>
<th>Settling Model</th>
<th>Settling Velocity (m/h)</th>
<th>Cross Sectional Diameter (microns)*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Small Floc</td>
<td>Primary Particles</td>
<td>&lt; 1.50</td>
<td>&lt; 37</td>
</tr>
<tr>
<td>Medium Floc</td>
<td>Flocculated Particles</td>
<td>1.5 ≤ Vs &lt; 6</td>
<td>37 ≤ d &lt; 740</td>
</tr>
<tr>
<td>Large Floc</td>
<td>Flocculated Particles</td>
<td>≥ 6</td>
<td>≥ 740</td>
</tr>
</tbody>
</table>

* Cross sectional diameter estimation based on Equation 10

**Calibration and Validation**

This model was calibrated and validated with field data collected at the Marrero Wastewater Treatment Plant (WWTP) located in Marrero, Louisiana. The operating condition of the SST and the flocculation constants of the mixed liquor suspended solids (MLSS) were measured during the calibration and validation of the model. Details about the calibration and validation of the model are presented in Griborio (2004), Griborio and McCorquodale (2005) and McCorquodale et al. (2005). Table 2 shows the geometry of the SST, and the loading and performance data measured during the calibration.
### Table 2 - Marrero WWTP SST – Geometry, Loading and Performance Data

<table>
<thead>
<tr>
<th>Geometry</th>
<th>Value</th>
<th>Loading</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Radius of the clarifier</td>
<td>17.6 m</td>
<td>SOR</td>
<td>1.0 m/h</td>
</tr>
<tr>
<td>Radius of the inlet pipe</td>
<td>0.5 m</td>
<td>RAS Ratio</td>
<td>0.5</td>
</tr>
<tr>
<td>Depth of outer wall</td>
<td>4.3 m</td>
<td>Qeffluent</td>
<td>972 m³/h</td>
</tr>
<tr>
<td>Bottom Slope</td>
<td>8.33%</td>
<td>Qras</td>
<td>486 m³/h</td>
</tr>
<tr>
<td>Center Well radius</td>
<td>4.5 m</td>
<td>MLSS</td>
<td>2800 mg/L</td>
</tr>
<tr>
<td>Center Well Depth</td>
<td>2.6 m</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Outboard launder</td>
<td>----</td>
<td>ESS</td>
<td>15 mg/L</td>
</tr>
<tr>
<td>Radial length of hopper</td>
<td>2.0 m</td>
<td>RAS SS</td>
<td>8400 mg/L</td>
</tr>
</tbody>
</table>

### Calibration of the Flocculation Sub-model.

A simple batch flocculation test was conducted to determine the flocculation constants used in the sheared induced flocculation model. The flocculation test was performed on-site at the Marrero WWTP. A six-paddle stirrer (Phipps and Bird Stirrer) was used to flocculate the activated sludge samples. Flocculation was induced mechanically by stirring at a rotational velocity the samples inducing a $G$ value of about 40 s⁻¹. The stirrers were equipped with rectangular flat-blades of 2.54 x 7.62 cm (1.0 x 3.0 in) that rotate in a horizontal plane about the centerline of their long axis. The bottoms of the paddles were situated approximately 5 cm (1.97 in) above the bottom of the jar during the tests. Square, 15 x 15 cm (5.91 x 5.91 in), glass jars were used for the flocculation tests, to prevent vortexing.

The activated sludge samples were taken from the contact chamber of the Marrero WWTP (MLSS sample). The average concentration of the MLSS samples was 2800 mg/L. Two liters of MLSS were poured in each jar, and a flocculation time was assigned to each jar. The selected flocculation times were 0, 2.5, 5, 10 and 20 minutes. After the prescribed flocculation time had elapsed, the jar was removed carefully from the stirrer. After 30 minutes of settling, approximately 250 mL of supernatant were withdrawn very carefully by a siphon mechanism avoiding suction of floating solids. Each supernatant sample was analyzed for TSS. The $C_O$ concentration was measured from a sample taken from the inlet zone of the Marrero contact chamber (SS concentration at time equal zero, see Equation 5); this value could also be determined experimentally by mixing the influent to the solids contact chamber and recycle sludge in proportion to the influent flow rate and the recycle flow rate, respectively, and by measuring the SS concentration of the supernatant of the mixture after 30 minutes of settling.

The flocculation characteristics were defined by the parameters $a$ and $k_F$. Estimates of the parameters were made by fitting Equation 5 to the experimental data. Figure 1 shows the removal of the supernatant suspended solids with the flocculation time. This figure shows a good correlation between the model proposed by Equation 5 and the measured field data. The flocculation constants $a$ and $k_F X$ were obtained by a non-linear regression analysis. The values obtained were $a = 4.3$ mg/L, and $k_F X = 0.497$ min⁻¹. A value of $k_F = 0.178$ L/g SS min is obtained by dividing $k_F X$ by the MLSS concentration. The values of the flocculation kinetic
constants obtained with the batch test are similar to values reported in the literature by Parker et al. (1970), Wahlberg et al. (1994), Jimenez (2002) and La Motta et al. (2003).

Figure 1 - Supernatant SS versus Flocculation Time in a Batch Test

With the values of \(a\), \(k_F\), \(C_O\), \(G\) and \(X\), the values of the kinetic constants \(K_A\) and \(K_B\) were calculated by comparing Equations 3, 4 and 5. Table 3 summarizes the values of the flocculation kinetic constants found during the calibration period.

Table 3 - Kinetic Constants for the Flocculation Sub-Model

<table>
<thead>
<tr>
<th>Batch Test Information</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>(C_0)</td>
<td>65 mg/L</td>
</tr>
<tr>
<td>MLSS ((X))</td>
<td>2800 mg/L</td>
</tr>
<tr>
<td>Velocity Gradient ((G))</td>
<td>40 s(^{-1})</td>
</tr>
<tr>
<td>Flocculation Kinetic Constants for Equation 5</td>
<td>Value</td>
</tr>
<tr>
<td>(a)</td>
<td>4.3 mg/L</td>
</tr>
<tr>
<td>(k_F)</td>
<td>0.178 L/g SS min</td>
</tr>
<tr>
<td>Flocculation Kinetic Constants for Equations 1 and 2</td>
<td>Value</td>
</tr>
<tr>
<td>(K_A)</td>
<td>7.4 \times 10^{-5} L/g SS</td>
</tr>
<tr>
<td>(K_B)</td>
<td>8.00 \times 10^{-9} s</td>
</tr>
</tbody>
</table>
Model Simulation

To clarify the effect of the center well on hydrodynamics and flocculation, four different conditions were simulated using the information presented in Tables 2 and 3. The four different simulated scenarios were:

1. SST with a center well and flocculation sub-model on.
2. SST with a center well and flocculation sub-model off.
3. SST without center well and flocculation sub-model on.
4. SST without center well and flocculation sub-model off.

The simulations were run until steady conditions were reached, or until failure of the clarifier. Failure was identified as a rise of the sludge blanket to the water surface or as an ESS value higher than 30 mg/L. Steady state conditions were assumed to be reached when the return activated sludge (RAS) suspended solids (SS) concentration was within ± 2% of the equilibrium value obtained with a mass balance of suspended solids around the secondary clarifier; provided that the effluent suspended solids (ESS) of the simulation did not change more of than 5% in the last 30 minutes of simulation. For the four scenarios, steady conditions were reached before six hours of simulation time. The SST’s performance (ESS and RAS SS) were obtained for each scenario after 360 minutes of simulation.

The effect of the position of the center well on the clarifier performance was also evaluated by simulating the loading conditions presented in Table 2 with different center well radii. The ESS was obtained for different CW radii and different solids loading rates (SLR). The different SLRs were simulated by keeping the surface overflow rate (SOR=1 m/h) and the recirculation ratio constant (0.5) and changing the MLSS (e.g., 1.8, 2.8, and 4.2 kg/m^3). The depth of the CW was kept constant at 2.6 m.

Finally, the effects of the CW geometry in the SST performance were also evaluated for different SORs. In this case, the recirculation ratio (0.5) and the MLSS (2.8 kg/m^3) were kept constant during the simulations, thus changing the SLR with the change on the SOR. The SORs evaluated were 0.75, 1.0, 1.5 and 2.0 m/h and the baffle positions were defined during the runs to find the optimum radius for each SOR.

RESULTS

Effect of Center Well on Flocculation and Hydrodynamics

Figure 2 shows the evolution of the ESS with the simulation time for the four scenarios described in the previous section. Table 4 presents the values of the ESS and the RAS SS concentrations after 360 minutes of simulation time. The SS contours and velocity vector fields for the two scenarios with the flocculation sub-model on are presented in Figure 3.
Table 4 - ESS and RAS SS for Study Cases in Center Well’s Effects

<table>
<thead>
<tr>
<th>Suspended Solids Concentration (mg/L)</th>
<th>Center Well Flocculation On</th>
<th>Center Well Flocculation Off</th>
<th>No Center Well Flocculation On</th>
<th>No Center Well Flocculation Off</th>
</tr>
</thead>
<tbody>
<tr>
<td>ESS</td>
<td>15.3</td>
<td>24.6</td>
<td>108.6</td>
<td>126.2</td>
</tr>
<tr>
<td>RAS SS</td>
<td>8364</td>
<td>8335</td>
<td>8086</td>
<td>8044</td>
</tr>
</tbody>
</table>
Figure 2 and Table 4 show that the clarifier performs much better with the CW that without it. Also the ESS improves when the flocculation sub-model is on. The clarifier fails for the case with no CW, regardless of the status of the flocculation sub-model. On the other hand, the clarifier achieves ESS below 30 mg/L for the case with the CW, without the flocculation sub-model. However, as previously mentioned, it predicts better performance when the flocculation sub-model is activated. The RAS SS was not affected by the status of the flocculation sub-model.

The simulations of the four different scenarios demonstrate that flocculation plays a major role in the ESS on secondary clarifiers. Results show that CWs improve the hydrodynamics in the SST and also promote aggregation of dispersed particles. However, the improved hydrodynamics is the major benefit of the CW by controlling of the re-entrainment of clarified fluid into the influent flow. The center well promotes flocculation by allowing enough contact time for the mixture in a zone of relatively high velocity gradients.

Figure 3 shows that the density current and the upflow velocities at the outlet zone are stronger for the tank without the center well. The center well decreases the strength of the density current...
by controlling the entrainment in the inlet zone; however, entrainment still occurs under the
center well, suggesting that the dimensions of the center well could be improved.

**Optimum Dimension of the Center Well – Effect of SLR.**

The effect of the position of the center well on the clarifier performance was evaluated by
simulating the loading conditions presented in Table 2 with different center well radii. The
results are presented in Figure 4 which shows the ESS for different CW radii and different SLRs.

**Figure 4 - Effect of SLR on Optimum Position of the Center Well (Baffle Depth = 2.6 m,
SOR = 1 m/h, SLR= Variable; RAS = 50%)**

![Figure 4](image)

Figure 4 shows that for the typical operating conditions of the Marrero WWTP (SOR= 1 m/h,
MLSS = 2.8 kg/m³, Recirculation Ratio = 0.5, SLR = 4.20 kg/m²/h), the optimum radius for the
CW is about 28% of the total clarifier radius (for the 2.6 m baffle depth). The 28% of the total
clarifier radius yields a 5 m baffle radius and allows a hydraulic residence time for flocculation
of about 18 minutes in the center well. It seems that smaller center wells do not provide enough
contact time for flocculation but may slightly decrease the strength of the density current. On the
other hand, large center wells do not provide good control of the re-entrainment of the fluid from
the sedimentation zone, resulting in a strengthened density current. Figure 4 indicates that an
optimum placement of the center well becomes more distinct as the SLR increases. For the
range of simulated SLRs, between 2.7 and 6.3 kg/m²/h, the optimum placement of the CW
ranges between 20 and 32 percent of the clarifier radius. This range is almost in the same design
range proposed by Ekama et al. (1997) who recommend a center well diameter extending from
20% to 35% of the tank diameter.
Optimum Dimension of the Center Well – Effect of SOR

To define the effect of the SOR on the optimum dimension of the CW, different baffle positions were evaluated under different SORs. The recirculation ratio (0.5) and the MLSS (2.8 kg/m³) were kept constant during the simulations, thus changing the SLR with the change on the SOR. The SORs evaluated were 0.75, 1.0, 1.5 and 2.0 m/h and the baffle positions were defined during the runs to find the optimum radius for each SOR. The ESS was affected by the baffle position, while it did not seem to have any effect on the RAS SS. The results for the ESS concentrations are shown in Figure 5.

Figure 5 - Effect of SOR on Optimum Position of the Center Well (Baffle Depth = 2.6m, MLSS = 2.8 Kg/m³, SLR= Variable, RAS= 50%)

Figure 5 shows that the optimum size of the center well decreases as the SOR decreases: (1) for SOR= 0.75 m/h, the optimum radius is about a 20% of the total clarifier radius; (2) for the SORs equal to 1 and 1.5 m/h, the optimum radius were about a 28% of the total radius; and (3) for the SOR = 2.0 m/h, the optimum radius increased to 37%. This tendency can be observed in Figure 6. For SORs between 0.75 and 2.0 m/h the optimum dimension of the center well radius changes in a range from 20 to 37%.
Figure 6 - Optimum Center Well Radius versus SOR (Baffle Depth = 2.6m, MLSS = 2.8 Kg/m$^3$, SLR= Variable, RAS = 50%)

Figure 6 shows that the optimum size of the center well increases as the SOR increases. Again, the reason for this behavior is the control of the re-entrainment and the flocculation process in the center well. As discussed before, large center wells do not provide a good control of the re-entrainment for low and medium SORs. In the presence of high SORs, strong influent momentum dominates the flow in the center well, restricting the re-entrainment of the fluid from the sedimentation zone. Obviously, the same effect will occur with smaller baffles and high SOR, but in the case of the larger baffle there is additional contact time for the flocculation process at adequate G values (produced by the conversion of the high inlet kinetic energy). Figure 7 shows how the re-entrainment of the clarified fluid in the center well significantly decreases as the SOR increases for the large baffle.

Figure 7 - Flow Pattern and SS Contours for a Large Center Well under Different SOR Loadings (CW Radius = 6.5 m, CW Depth = 2.6 m, MLSS = 2.8 kg/m$^3$, RAS = 50%)
CONCLUSIONS

- The center well promotes flocculation, but its major benefit is the improvement of the tank hydrodynamics. Model sensitivity studies indicate that the dominant role of the center well is the control of the re-entrainment of clarified fluid with the influent flow thus inducing a stronger upflow at the launder.

- The optimum dimension of the center well diameter is 20% to 35% of the total clarifier diameter. The incoming MLSS does not seem to have a major effect on the optimum dimension, while it appears that the optimum dimension tends to increase as the SOR increases. The increase in the underflow rate might also affect the optimum dimension of the center well in the same way the SOR does.

- The performance of SST depends on several interrelated processes and therefore its behavior can not be explained with a single variable. Clarifiers should not be designed on the basis of a single parameter, e.g. the SOR, which alone can not provide the center well dimensions for secondary clarifiers.

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