Special automation and regulation strategies for enhancing sequencing batch reactor (SBR) performances

S. G. E. Rönner-Holm and N. C. Holm

ABSTRACT

Dynamic simulation analyses of five different sequencing batch reactor (SBR) wastewater treatment plants (WWTPs) were used in order to optimise developed regulation strategies and to develop new strategies. The results were applied directly to 15 full-scale SBR plants. To do this, the cycle strategies were extended through the use of appropriate aggregates, or were anchored in the programmable logic controller (PLC) and process control system (PCS) with the help of online sensors. This enabled all regulation strategies to be introduced and automated without problems.

Key words | cycle strategy, online sensors, process control, regulation strategy, SBR, simulation

INTRODUCTION

In the SBR process, the complete activated sludge and sedimentation processes take place successively in one reactor in continuous cycles (Irvine & Busch 1979). Various modified SBR processes with and without a buffer tank have been developed since then, and have been increasingly used in practice through the technological developments in automation engineering (Wilderer et al. 2001). Nowadays, the SBR process can be considered as a fully-fledged alternative to continuous pass-operated WWTPs. The plant control system can be regulated and monitored by the PLC and the PCS. This enables newly developed regulation strategies to be easily transferred to full-scale technology.

In the differential internal cycle strategy (DIC) SBR process, the complete activated sludge and sedimentation processes take place successively in one reactor in continuous cycles as in normal SBR plants (Holm et al. 2000). One or more upstream buffer tanks are crucial for DIC-SBR plants. The buffer tank is used simultaneously for influent storage and hydrolysis. Furthermore, it allows the compensation of daily wastewater flow variations. In the special DIC process, the buffer tank is also used to generate two different filling charges during one cycle. During the first filling phase, the reactor is only fed with the supernatant from the unstirred buffer tank, which consists of wastewater at a low C/N ratio. The first filling phase forms the largest amount proportion of the overall filling phase within a cycle. In this manner, particulate, slowly degradable substrate from the wastewater is concentrated and is then further hydrolysed and changed into easily degradable substrate due to the longer hydraulic retention time in the buffer tank (Holm et al. 2000). Due to the increased proportion of easily degradable substrate in the first filling phase, a complete and accelerated denitrification normally takes place at the beginning of the cycle without aeration. The anaerobic conditions which result from this positively support biological phosphorus (BioP) removal. In the subsequent aeration phase, the COD degradation and the largest proportion of the overall nitrification process takes place. Before the second filling phase of a particular cycle the wastewater in the buffer tank is mixed, yielding a feed with a C/N ratio and COD concentration approximately three times higher. A correspondingly larger portion of COD is now added at comparatively low volume due to the previous concentration phase, whereas the proportion of ammonium is not increased. An anoxic/anaerobic phase...
and a subsequent aeration phase performing the usual COD degradation, BioP removal and nitrification processes is then repeated. Due to the low volume of the second filling phase, there is a comparatively low NO3-N concentration at the end of the aeration. This method achieves nitrogen elimination rates of 94–98% (Holm et al. 2000). For the sedimentation, mixing is stopped and the clear supernatant is removed.

In the return sludge (RS) process, a part of the activated sludge is loaded from the reactors into the buffer tank, which is thereby additionally used as a pre-denitrification basin and for the BioP removal (Rönner-Holm & Holm 2008).

The dynamic simulation analysis of WWTPs is a useful tool for research into the activated sludge processes, for plant design, for analysis of control strategies and for optimising studies (Keller et al. 2002). A series of models such as ASM1, ASM2d, ASM3 and EAWAG-BioP for the activated sludge processes has been established over the last few decades (Henze et al. 2000; Rieger et al. 2001).

Five DIC-SBR plants were investigated using different models for the purposes of dynamic simulation (Rönner-Holm et al. 2006, 2009; Rönner-Holm & Holm 2008). First, the ASM1 model was used (Rönner-Holm et al. 2006). This model is most commonly used for full-scale WWTPs and simulates degradation of carbohydrate compounds, nitrification and denitrification processes. A protocol for the characterisation of the wastewater flows is available and the calibration procedure is well established (STOWA 2000; Hulsbeek et al. 2001; Roeleveld & van Loosdrecht 2002). Since, however, a pronounced BioP removal was observed at full-scale, all the following plants were investigated in ASM3 in combination with the EAWAG-BioP model (Rönner-Holm & Holm 2008; Rönner-Holm et al. 2009). This combined model additionally allowed simulation of the storage of carbohydrate compounds, nitrification and denitrification processes or the biological and chemical phosphorus removal can be analysed.

Table 1: Main characteristics of the investigated plants

<table>
<thead>
<tr>
<th>SBR plant</th>
<th>Size PE</th>
<th>SBR process</th>
<th>Number of SBR</th>
<th>Online sensors</th>
<th>Simulation model</th>
<th>Effluent values† (mg/l)</th>
<th>SVI‡</th>
</tr>
</thead>
<tbody>
<tr>
<td>Spenge</td>
<td>22,500</td>
<td>DIC/RS</td>
<td>3</td>
<td>P, NH, NO</td>
<td>ASM3 + EAWAG-BioP</td>
<td>21 1.3 1.2 65</td>
<td></td>
</tr>
<tr>
<td>Weisstal</td>
<td>9,200</td>
<td>DIC/RS</td>
<td>2</td>
<td>P, NH, NO</td>
<td>ASM3 + EAWAG-BioP</td>
<td>18 5.0 0.8 70</td>
<td></td>
</tr>
<tr>
<td>Heßheim</td>
<td>62,000</td>
<td>DIC</td>
<td>4</td>
<td>P, NH, NO</td>
<td></td>
<td>30 2.0 1.2 55</td>
<td></td>
</tr>
<tr>
<td>Büddendenst 4</td>
<td>4,400</td>
<td>DIC</td>
<td>2</td>
<td>–</td>
<td></td>
<td>35 3.0 1.5 80</td>
<td></td>
</tr>
<tr>
<td>Deuz</td>
<td>12,800</td>
<td>DIC/RS</td>
<td>3</td>
<td>P, NH, NO</td>
<td>ASM1/ASM3 + EAWAG-BioP</td>
<td>18 2.0 1.0 80</td>
<td></td>
</tr>
<tr>
<td>Bad Zwischenahn</td>
<td>31,000/41,000</td>
<td>DIC</td>
<td>3</td>
<td>P, NH, NO</td>
<td>ASM1</td>
<td>50 3.0 0.7 70</td>
<td></td>
</tr>
<tr>
<td>Polenz</td>
<td>20,000</td>
<td>DIC</td>
<td>3</td>
<td>P</td>
<td></td>
<td>29 3.0 1.7 80</td>
<td></td>
</tr>
<tr>
<td>Prossen</td>
<td>10,500</td>
<td>DIC</td>
<td>2</td>
<td>P</td>
<td></td>
<td>37 2.5 0.5 70</td>
<td></td>
</tr>
<tr>
<td>Hettstedt</td>
<td>30,000</td>
<td>DIC</td>
<td>3</td>
<td>P, NH, NO</td>
<td>ASM1</td>
<td>27 3.0 1.5 65</td>
<td></td>
</tr>
<tr>
<td>Nammen</td>
<td>8,500</td>
<td>DIC</td>
<td>2</td>
<td>P</td>
<td></td>
<td>31 1.4 0.5 60</td>
<td></td>
</tr>
<tr>
<td>Zierenberg</td>
<td>5,250</td>
<td>DIC</td>
<td>2</td>
<td>P</td>
<td></td>
<td>28 0.6 1.4 70</td>
<td></td>
</tr>
<tr>
<td>Bruchmühlen</td>
<td>11,250</td>
<td>DIC</td>
<td>3</td>
<td>P, NH, NO</td>
<td></td>
<td>26 0.5 1.2 80</td>
<td></td>
</tr>
<tr>
<td>Huntlsonen</td>
<td>10,000</td>
<td>DIC/RS</td>
<td>2</td>
<td>P, NH, NO</td>
<td></td>
<td>29 2.0 1.5 60</td>
<td></td>
</tr>
<tr>
<td>Radeburg</td>
<td>20,000</td>
<td>DIC/RS</td>
<td>3</td>
<td>P, NH, NO</td>
<td></td>
<td>32 3.0 1.4 60</td>
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<tr>
<td>Gemünden</td>
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<td>P, NH, NO</td>
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<td>26 0.9 0.9 70</td>
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<tr>
<td>Herbertshausen</td>
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<td>DIC/RS</td>
<td>2</td>
<td>P, NH, NO</td>
<td></td>
<td>25 2.0 1.2 70</td>
<td></td>
</tr>
</tbody>
</table>

| n = PO4-P, NH = NH4-N, NO = NO3-N. |
| Effluent values = mean effluent values, mean BOD values <5 mg/l. |
| SVI = sludge volume index. |
compounds and the biological and chemical phosphorus removal. Furthermore, this model enabled estimation of operating costs consisting of plant energy consumption, the precipitant consumption and excess sludge production.

Established regulation strategies already in use at full-scale were optimised, and new strategies were developed by dynamic simulation. The most promising strategies were introduced at various plants in full-scale tests, and the effects were monitored and documented over longer periods. Introduction of online sensors helped to introduce the developed optimisation strategies in 16 full-scale SBR plants. For automation, regulation strategies were implemented in the PLC and visualised in the PCS. The way in which developed strategies were implemented is shown for various examples in this paper based on the simulation results.

**METHODS**

**Investigated plants and processes**

All the investigated plants are listed in Table 1, and their equipment and characteristics after implementation of the strategies are shown. All listed WWTPs located in Germany treat mainly municipal wastewater, and most plants have been designed for the extended aeration SBR process (25 days sludge age) with the exception of the Bad Zwischenahn and Weißtal WWTPs (approx. 15 days).

**Dynamic simulation analysis**

The plants which were investigated by dynamic simulation and the used model are indicated in Table 1. The dynamic simulation studies were performed according to the HSG and STOWA guidelines (STOWA 2000; Langergraber et al. 2004). Further descriptions of the measuring campaigns, characterisation of the influent, the calibration procedure and the performance of studies are given in Röörer-Holm et al. (2006, 2009) and Röörer-Holm & Holm (2008). For analysis of the comparability of the different models used, the model of Deuz WWTP was calibrated in ASM1 and ASM3 + EAWAG-BioP. For monitoring the introduced optimisation strategies, 24 h hydraulic equalised influent and effluent samples were collected and analysed every week for Hettstedt and Spenge WWTP over a couple of years. For all other plants, the introduction was monitored over a couple of months.

**Cross-plant automation systems, special control strategies and online sensors**

All plants investigated are equipped with Siemens SPS S7 PLCs for cross-plant automation and control. The PC-based supervisory control and data acquisition (SCADA) system functioning as the PCS is based upon Wonderware or Siemens WinCC Software. Here, all online sensor data is processed into curves and is used as variables in formulas for operation, control and default strategies. All plants are equipped with level and flow online sensors by Endress + Hauser and O2 sensors by Endress + Hauser, Hach Lange or WTW, some have MLSS sensors by Endress + Hauser, Hach Lange or WTW. Online sensors such as NH4-N, NO3-N and PO4-P by Endress + Hauser, Hach Lange or WTW are installed as shown in Table 1.

![Figure 1](https://iwaponline.com/wst/article-pdf/60/5/1161/448438/1161.pdf)
RESULTS AND DISCUSSION

Comparison calibration results in ASM1 and ASM3 + EAWAG-BioP

A comparison of the calibration results from the Deuz plant model in ASM1 and ASM3 + EAWAG-BioP showed no significant differences in the degradation of carbohydrate compounds, nitrification and denitrification processes, which means that the results of the models used are comparable in this respect (Figure 1).

RS-SBR process

The RS-SBR process was investigated through the use of dynamic simulation at the Weiβtal and Spenge WWTPs. In the Weiβtal model, this led to the reduction of NO$_3$-N concentrations in the reactor, and therefore to lower effluent values (Figure 2). In addition, the process had a beneficial effect on the BioP removal, and also resulted in the P$_{\text{total}}$ effluent values being reduced. These results were confirmed by simulation analyses at Spenge WWTP. The operating mode was closely investigated and optimised in further simulations (Figure 3). It was shown that removal of the RS at the end of the sedimentation phase was particularly appropriate. Due to the sludge sedimentation characteristics, higher RS quantities were able to be drawn off in the short-term using this method than in the mixing phase. This additionally improved the cleansing performance in the buffer tank at lower operating costs (Figure 3). However, it was also shown that drawing off the RS too heavily tended to be counter-productive (Figure 3) and that it needed to be limited by a previously specified MLSS content in the reactor. This value varied between plants and over the summer/winter seasons. For example, a value was chosen for the Spenge WWTP which was 0.5 g MLSS/l below the operating MLSS of 3–3.5 g/l. This ensured that sufficient activated sludge was still available in the reactor, and that the cleansing performance remained stable and unaffected.

The simulation results were transferred into the PLC and PCS and an additional slide valve was connected. The surplus activated sludge (SAS) pumps were also used to draw the RS off. The new valve, however, ensured that the RS was only fed into the buffer tank and not into the SAS holding tank. This unit was integrated at cycle strategy level (Figure 4). The duration can be selected individually in 5-minute stages. This method was used to automate the RS draw-off. The existing MLSS sensors were used for
premature termination of the draw-off if falling below a limit value made this necessary. Using this automation control system, the RS process was able to be introduced in the investigated plants. The simulation results were also confirmed on plants where the RS process was introduced (Table 1).

The effect of introduction of the RS process in full-scale on the effluent values of Spenge WWTP is shown in Figure 5. After implementation in September 2005 except for three months the monthly mean $N_{\text{total}}$ and $P_{\text{total}}$ values of the following year decreased which confirms the simulation results. It has to be considered that the influent quantity and influent load in 2006 was even higher than in 2005 (Table 2).

### Optimisation of wastewater flow equalisation

As a main control strategy, a type curve for the buffer tank (type level hydrograph) is used to regulate filling the SBR tanks so that daily variations of wastewater flows are partially compensated. Since the flow rates differ considerably between the WWTPs, only a standard type level hydrograph (based on experience with different plants) is provided during start-up. This means further adjustments can be performed after monitoring the WWTP-specific daily flow rates in order to calculate an optimised hydraulically compensated type level hydrograph. In a next step, a load-dependent type level hydrograph can be calculated on the basis of the measuring campaign data and subsequent dynamic simulation adjustments. The different type level hydrographs and their influence on the effluent values are shown in Figure 6.

The hydraulic type level hydrograph rapidly led to better effluent values, but the lowest effluent values were predicted with the $\text{NH}_4$-N load-dependent type level hydrograph optimised by simulation (Figure 6a). The operation costs including aeration energy and precipitant amount remained almost unaffected. These results were confirmed by all the other investigated plants by dynamic simulation analysis (Table 1), but the calculated type level hydrographs differed depending on the influent characteristic of each plant (Figure 6b).

The type level hydrograph was broken down into 48 values for transfer into the PCS (Figure 7). A higher resolution does not make sense. Figure 7 highlights the value which is currently active. Formula (1) indicates how much is supplied to a single cycle depending on the current height level in the buffer tank (measured with a fill level measurement) and the current active type level.
Reactor filling height = \( \frac{1}{4} \) (level buffer tank - type level hydrograph) \times plant specific value \hspace{1cm} (1)

The influence of the type level hydrograph during dry weather conditions is shown in Figure 8. When the filling is calculated using a constant type level hydrograph the ammonium peaks in the reactor do vary strongly in their heights, and sometimes one reactor is loaded more than the others, which indicates different loading situations (Figure 8a). In contrast, the peaks are more equalized after introduction of the NH\(_4\)-N load-dependent type level hydrograph, which indicates similar loading situations resulting in an lower maximum effluent value (Figure 8b). The slight differences in the filling times are due to previous cycle length changes caused by rain events as explained below but have no effect on the wastewater flow equalisation in the buffer tank. The simulation results were confirmed by all the other plants investigated (Table 1) after introduction of a NH\(_4\)-N load-dependent type level hydrograph subsequently monitored by NH\(_4\)-H online sensors.

### Table 2

<table>
<thead>
<tr>
<th>WWTP Properties</th>
<th>People equivalent</th>
<th>Influent quantity (m(^3)/a)</th>
<th>Rain quantity (mm/m(^2))</th>
<th>Influent BOD (kg/a)</th>
<th>Elimination rates [%]</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>N(_{\text{total}})</td>
</tr>
<tr>
<td>2005(^5)</td>
<td>18,517</td>
<td>2,394,008</td>
<td>712</td>
<td>390,223</td>
<td>92</td>
</tr>
<tr>
<td>2006(^6)</td>
<td>18,547</td>
<td>2,434,017</td>
<td>756</td>
<td>574,428</td>
<td>94</td>
</tr>
<tr>
<td>2007(^7)</td>
<td>18,308</td>
<td>2,948,291</td>
<td>987</td>
<td>513,003</td>
<td>93</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th></th>
<th>Total electricity demand (kWh/kg BOD(_5))</th>
<th>Total electricity demand (kWh/PE(^8))</th>
<th>Sludge quantity (kg/kg BOD(_5))</th>
<th>Precipitant quantity (l/kg BOD(_5))</th>
<th>Lime quantity (kg/kg BOD(_5))</th>
<th>Waste water tax (€/kg BOD(_5))</th>
<th>Operation costs(^9) (€/kg BOD(_5))</th>
</tr>
</thead>
<tbody>
<tr>
<td>2005(^5)</td>
<td>1.464</td>
<td>32.21</td>
<td>3.3</td>
<td>0.112</td>
<td>0.237</td>
<td>0.129</td>
<td>0.49</td>
</tr>
<tr>
<td>2006(^6)</td>
<td>0.885</td>
<td>19.47</td>
<td>3.2</td>
<td>0.074</td>
<td>0.299</td>
<td>0.057</td>
<td>0.35</td>
</tr>
<tr>
<td>2007(^7)</td>
<td>1.122</td>
<td>26.84</td>
<td>3.8</td>
<td>0.087</td>
<td>0.268</td>
<td>0.060</td>
<td>0.41</td>
</tr>
</tbody>
</table>

\(^5\)DIC/RS-SBR plant not optimised.  
\(^6\)DIC/RS-SBR plant partial optimised.  
\(^7\)0.16/kWh + (0.055c/kg sludge) + (0.213c/l precipitant) + (0.04505c/kg lime).  
\(^8\)People equivalent.  

\(^9\)Operation costs\(^9\) = (0.1€/kWh) + (0.055€/kg sludge) + (0.213€/l precipitant) + (0.04505€/kg lime).
Optimisation of the DIC-SBR process

For the DIC-SBR process, the buffer tank is used to generate two different filling charges for the SBR tank during one cycle. In general, the first filling feed is calculated with 80% of the total cycle loading amount, whereas the second filling comprises 20%. In dynamic simulation studies with different wastewater flow examples at the investigated plants (indicated in Table 1), results showed a wastewater-dependent optimum of the filling ratio at 90/10, 80/20, 70/30 or 60/40 with regard to the effluent values. The results
of the simulation investigations with the specific wastewater at the Hettstedt WWTP with different filling ratios are shown in Figure 9. The lowest ratio of 60/40 led to the lowest maximum effluent values with regard to $N_{\text{total}}$ and $\text{NO}_3$; in contrast, the $\text{NH}_4$-concentration rose (Figure 9).

It must be emphasised that this finding is probably only valid for plants with a DIC operation mode where an unusually high COD concentration is generated in a second filling charge. Simulation analysis of all the plants investigated so far in the normal SBR operation mode using only one filling per cycle with mixed buffer tank influent always revealed higher effluent values. This confirms the improved denitrification rate of the DIC operation mode in the second filling as outlined in lab-scale analysis by Holm et al. (2000) and indicates the advantageous influence of a concentrated second filling event at such a low amount.

Although 6-hour cycles often generate slightly lower effluent values than 8-hour cycles in the case of Hettstedt WWTP (Figure 9), it has to be considered that the aeration energy consumption was always higher in shorter cycles. Therefore operation costs in 6-hour cycles increase.

However, the effluent values and aeration energy consumption were lower in 8-hour cycles at Deuz WWTP. This confirms the results of Spenge WWTP where the 8-hour cycle had the best cleansing performance and the lowest running costs in comparison to shorter cycles (Rönder-Holm et al. 2009). The optimal cycle length might depend on the influent characterisation of each plant with regard to the effluent values. Further investigations are necessary.

In the PLC and PCS, the implementation of different filling charges for the DIC-SBR operation mode were realised using automatically guided pumps or valves in the buffer tank at the cycle strategy level, and these were activated either for the first or second filling (Figure 10). For the first filling of the reactors, the buffer tank top valve was opened while the buffer was not being mixed. In contrast, the buffer content was mixed during the second filling and both valves left open. After the filling quantity has been calculated for one cycle as in Formula (1), the quantity is divided in accordance with the previously stipulated but freely selectable filling ratios for the first and second filling.

![Figure 9](image_url) Influence of different filling ratios in the Hettstedt WWTP model on $N_{\text{total, max}}$ (a), $\text{NH}_4_{\text{max}}$ and $\text{NO}_3_{\text{max}}$ effluent values (b).

![Figure 10](image_url) Automation of the DIC-SBR process on the basis of cycle strategy in 8-hour cycle regulated in 5 min stages (dark grey = activated, light grey = not activated).
In August 2006 the optimal filling ratio 90/10 for the 8-hour cycle was introduced in full-scale at Spenge WWTP which resulted in lower monthly mean $N_{\text{total}}$ and $P_{\text{total}}$ effluent values the following months except for November 2006 (Figure 5). In contrast to the simulation results, which indicated an increase of the $P_{\text{total,max}}$ effluent values, the $P_{\text{total}}$ effluent values decreased at full-scale. We suggest that the phosphorus release or the lysis of polyphosphate in the Spenge WWTP model might be too weakly adjusted.

Optimisation of aeration using online sensors

The 8 and 6-hour cycles are operated in the DIC-SBR mode with two aeration phases, whereas the 4 and 3-hour cycles are operated with one filling and aeration phase only. Optimisation studies for the aeration strategies were performed for all plants which were analysed by dynamic simulation.

The following strategies for the Hettstedt WWTP were investigated in the following order (Figure 11):

1. (1), shortening of the aeration time in the cycle strategy;
2. (2), (1) and limitation of the first aeration phase depending on the $NH_4-N$ concentration (stop aeration 10 minutes after 0.3 $NH_4-N$ g N/m$^3$ is reached);
3. (3), (1) and limitation of the aeration depending on the $NH_4-N$ concentration (stop aeration immediately after 0.5 $NH_4-N$ g N/m$^3$ is reached);
4. (4), (3) and earlier 1st aeration phase and 2nd filling phase, later 2nd aeration phase;
5. (5), (4) and intermittent aeration between 0.1 and 1.0 g DO/m$^3$ instead of 1.0 to 2.0 g DO/m$^3$.

In all cases, the $N_{\text{total,max}}$ effluent values decreased in stages (Figure 11a). With the exception of strategy (4), the aeration time behaved similarly (Figure 11b). Nevertheless, since strategy (4) led to more complete denitrification rates, this strategy was also considered for the implementation in the WWTP. The positive effect of strategy (5) was not suggested for implementation since, under this condition, high NO$_2$ concentrations might arise. The two-step metabolic cycle of the nitrification is not included in ASM series. Furthermore, the COD$_{\text{total,max}}$ effluent values were slightly higher through the use of strategy (5). The aeration minimum of 0.5 g DO/m$^3$ was therefore used for safety reasons.

Simulation results from other plants confirmed the above-mentioned results, but slight differences occurred regarding the length of aeration, the $NH_4-N$ limitation concentration and the minimum and maximum DO concentration for the intermittent blower performance. We suggest that these values have to be adjusted depending on the influent characteristics and blower capacity for each plant. This can be done using $O_2$ and $NH_4-N$ online sensors. For example, the DO increase during aeration should not be too low and the resulting $NH_4-N$ concentration may not be higher than 2 mg/l for the first aeration and 1 mg/l for the second aeration phase, otherwise the minimum and maximum DO concentration for the intermittent blower performance must be increased. In contrast, the $NH_4-N$ limitation concentration might be decreased. Furthermore, simulation results confirmed that continuous aeration instead of intermittent aeration regulation is even more energy-consuming and increases $N_{\text{total}}$ effluent values.

![Figure 11](https://iwaponline.com/wst/article-pdf/60/5/1161/448438/1161.pdf)
The number of intervals of intermittent aeration during one phase does vary depending on the plant. In addition, the blower performance can be reduced once the oxygen uptake rate of the activated sludge decreases, which again depends on the plant characteristics.

The length, start and end of an aeration phase for automation in PLC and PCS is regulated variably on the level of cycle strategy (Figure 12). In addition, online sensors for O2, NH4-N, and also recently for NO3-N, are installed for further regulation purposes. The O2 online sensor is used for intermittent aeration strategy, which allows stopping of aeration once a maximum value of 1.5 to 2 g DO/m3 is reached and starting of aeration once the DO concentration drops below a minimal value of 0.5 to 1 g DO/m3 (Figure 13). Furthermore, aeration is stopped when this strategy reaches a certain number of intervals. The aeration finishes when the NH4-N measured by online sensors is below 2 to 0.5 mg/l, and the NO3-N measured by online sensors also increases above a previous fixed amount which again depends on the influent characteristic of each plant.

After introduction of the above-mentioned aeration strategies at Hettstedt WWTP at the end of 2003, the energy consumption and the N\textsubscript{total} mean effluent value in the following operation year decreased by approx. 12%.

**Strategy for optimised cycle length change**

The SBR WWTPs listed in Table 1 treat dry weather inflow in 8 or 6-hour cycles depending on the hydraulic situation. However, combined sewer flow during storm events is treated by increasing the decantation quantity and/or operating in shorter cycles such as 6, 4 and 3-hour cycles. Suitable activators for the switch to shorter cycles in addition to the inflow rate are the rates of the rain quantity in addition to the level and the velocity of the level increase in the buffer tank (Wiese et al. 2006; Rönner-Holm et al. 2009). These results were confirmed by simulation analyses of Weißtal WWTP. In addition the conductivity online sensor in the influent showed a characteristic behaviour (Figure 14) at full-scale during all investigated rain events. Although the conductivity has a daily floatation behaviour during dry weather conditions due to daily loading variations, the value dropped dramatically under the lowest value (Figure 14a) at the beginning of a rain event. Nonetheless, this behaviour was not particularly clear for Spenge WWTP and thus further investigations of other plants are necessary to confirm this result (Figure 14b).

**Figure 12** Automation of the aeration strategy on the basis of the cycle strategy in the 4-hour cycle regulated in 5 min steps (dark grey = activated, light grey = not activated).

**Figure 13** Automation of the aeration strategy in the 8-hour cycle (1st aeration left, 2nd aeration right).

**Figure 14** Behaviour of conductivity online sensor during rain events in full-scale Weißtal (a) and Spenge (b) WWTP; DW = dry weather conditions, SW = storm weather conditions.
Dynamic simulation showed that an optimum adjustment of suitable activators leads to lower effluent values and running costs (Rönner-Holm et al. 2009). Furthermore it is possible to raise the purification performance and reduce operation costs if the strategy switches automatically back into longer cycles in good time after the end of storm water conditions.

The automated control strategy of the switch to the 4-hour cycle is shown in Figure 15. If the level of the buffer tank (Hi) or the influent amount (Zi) exceeds a certain freely selectable value, the operation switches to a shorter cycle. This strategy is adapted to each cycle length of 8, 6 or 4 hours in operation. The switch back to longer cycles can be regulated in the same manner.

However, the switch to a different cycle length may only occur at an optimal point in time during one cycle which has been previously set in the automated strategy for each cycle length. For example, a switch is never allowed into the decantation phase of one reactor if the sedimentation phase in the previous cycle length was not long enough. It is therefore necessary to have as many suitable activators as possible to avoid hydraulic overloading of the buffer tank.

**Full-scale optimisation potential example**

Many of the optimised strategies determined by means of dynamic simulation have been implemented at full-scale as indicated in Table 1. As in the model, the optimisation measures led to increased, stabilised elimination rates in Spenge WWTP effluent in 2006/2007 compared with 2005 without optimisation strategies (Table 2). Although the new plant was extended by 1,000 PE in stages, which resulted in 2–23% higher influent amounts and 10–50% higher loads of N_{total}, NH$_4$-N, P_{total}, BOD and COD, higher elimination rates were still achieved (Table 2). The operational costs of the optimised DIC/RS-SBR plant in 2006/2007 turned out to be lower then that of the non-optimised DIC/RS-SBR plant in 2005. Moreover there was an annual saving of operational costs at the SBR plant with regard to the mean annual BOD load including the sewage tax of 22% equivalent to approx. 49,000 € (without tax 18,000 €) at the optimised DIC/RS-SBR plant in 2006/2007 compared with 2005. It has to be considered that the influent quantity in 2007 was the highest due to higher rain quantities, which caused more electricity demand for pumping, aeration and mixing. In comparison to 2005, the mean electricity demand of the whole plant in 2006/2007 decreased to approx. 69% for the purification of one kg BOD, the precipitant quantity decreased to approx. 72%, whereas the surplus sludge quantity rose to 107%.

**CONCLUSIONS**

Implementation of new and subsequent optimisation of automation and control strategies is only possible within the range of flexibility specified by the plant design. For example, the anaerobic, anoxic and aerobic volumes cannot be changed in many continuous pass plants, one basin can only be used alternatively for anoxic or aerobic operation mode in some cases.

In SBR plants, especially those with an upstream buffer tank, the operational flexibility is considerably higher. The optimal treatment strategy can be dynamically adjusted to the actual inflow conditions by automatic and cycle phase adjustments controlled by online sensors. No fixed volume constraints exist because of the dynamic phase length adjustments of the anaerobic/anoxic/oxic duration times. This enhanced flexibility increases the potential of dynamic simulation studies, making the ASM models valuable tools for comparing different treatment strategies before implementing at full-scale plants.

This paper shows clearly that previously known regulation strategies can be optimised and new strategies can be developed for SBR plants with the help of dynamic simulation. In addition, these strategies can be simply transferred to the PLC and PCS. In this context, online sensors such as MLSS, O$_2$, NH$_4$-N, NO$_3$-N, PO$_4$-P are a valuable aid. The introduction of such optimised and new
strategies can result in considerable savings in operating costs, especially for power costs, and persistently increased the cleansing performance of the sewage works. Both aspects are valuable contributions to environmental protection, especially with respect to greenhouse gas emissions and pollution in addition to the eutrophication of waterways.

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