Aeration control – a review

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ABSTRACT

This review covers automatic control of continuous aeration systems in municipal wastewater treatment plants. The review focuses on published research in the 21st century and describes research into various methods to decide and control the dissolved oxygen (DO) concentration and to control the aerobic volume with special focus on plants with nitrogen removal. Important aspects of control system implementation and success are discussed, together with a critical review of published research on the topic. With respect to DO control and determination, the strategies used for control span from modifications and developments of conventional control methods which have been explored since the 1970s, to advanced control such as model-based predictive and optimal controllers. The review is supplemented with a summary of comparisons between control strategies evaluated in full-scale, pilot-scale and in simulations.

Key words | activated sludge, aeration, control strategies, dissolved oxygen control, review, wastewater treatment

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INTRODUCTION

Making dissolved oxygen (DO) transfer from gas phase to liquid phase is an energy intensive process in the wastewater treatment plant (WWTP), as well as crucial for the biological process to operate satisfactorily. Oxygen serves as an electron acceptor when organic carbon and nitrogen in the form of ammonium are oxidised. Blowers (not considering influent pumping) are the largest single user of energy at treatment plants today, motivating appropriate aeration control. Aeration energy is commonly responsible for around half of the plant power usage (WEF 2009) but numbers up to 75% have been reported (Rosso *et al*. 2008). Control of aeration systems becomes even more important when treatment plants face more stringent discharge limits and when energy efficiency is high up on the agenda.

This review paper considers aeration control of municipal nitrogen removal activated sludge systems with an emphasis on continuous, diffused aeration. Alternating or intermittent aeration systems, sequencing batch reactors and industrial applications are only mentioned briefly when applicable. Blower control and pressure control are only mentioned in the introductory section.

One of the first attempts to measure DO continuously was made at the Water Research Centre in Stevenage, UK, in 1954, using a semi-continuous colorimeter in conjunction with the Winkler method. By the early 1970s the use of on-line DO sensors was well established in many WWTPs, making DO control possible. This review does not cover the whole history of aeration control, but has an emphasis on published research during the 21st century but does not claim to be exhaustive. More information on early developments within aeration control can be found in Olsson (2012).

Earlier published material on the topic includes the annual literature reviews published by Water Environment Research (e.g. Sweeney & Kabouris 2011), and text books like Olsson & Newell (1999) and Olsson *et al*. (2005) where different aspects of ICA (instrumentation, control and automation) within the wastewater and water industries are presented. Weijers (2000) has documented a detailed list of control laws for wastewater treatment control up to then, including aeration control. Another overview of different control systems is found in Vanrolleghem (2001). Jeppsson *et al*. (2002) provide an overview of ICA from a European perspective and conclude that PI (proportional–integral) control or variations thereof were the most common strategies in full-scale at the turn of the last century.

In the paper we start by presenting terminology, and mention important elements of DO control, including hardware requirements and process dynamics. The most common controller structures within aeration control are categorised, followed by a presentation of research on

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different control algorithms. The descriptions of the control algorithms are divided into two sections representing the two ways in which nitrification and/or BOD (biological oxygen demand) removal capacity can be influenced: control of aeration intensity through DO control and control of the aerobic volume. The review is ended by a critical discussion.

TERMINOLOGY

The heart of automatic control is the feedback loop. A picture of a simple feedback loop is found in Figure 1.

There are many ways to name the variables in the closed control loop in Figure 1. Examples are listed in Table 1, with the terminology used in this paper in bold.

In most control loops the set-point is constant and the controller tries to minimise the control error caused by disturbances. Sometimes the set-point is changed either manually or by another controller. The latter appears in cascade control. Examples of other control structures will follow.

AERATION CONTROL: THE TASK

Aeration is important for providing sufficient DO for aerobic organisms performing BOD removal and nitrification in activated sludge plants, as well as keeping the biomass in suspension. The nitrification capacity can be varied in relation to DO control in two ways: by adjusting the aeration intensity or by adjusting the aerated volume. Apart from DO concentration, several other factors have been reported to affect nitrification rates including inorganic substrates, solids retention time (SRT), temperature, pH and toxic inhibition. Other control handles which also will affect the nitrogen removal, and hence might have an impact on the DO control loops, are return and waste activated sludge flows and nitrate recycle or external carbon dosage for plants with denitrification.

The nitrifier growth rate depends on the DO concentration and is commonly described by Monod kinetics (Monod 1942). The growth rate function will increase significantly with the DO at low DO concentrations but the dependence of DO becomes limited at high DO concentrations when approaching the maximum growth rate. Already in 1965, scientists at the Stevenage site in the UK reported that DO concentrations had very limited effects on nitrifier growth rates above 2.0 mg/l, but there is a wide range of reported effects of DO on maximum nitrifier growth rates (Stenstrom & Poduska 1980). A half-saturation concentration of 0.5–2.0 mgDO/l is reported (Henze *et al.* 2000).

The DO concentration should not be viewed on its own without considering temperature and aerobic SRT. At lower SRT and temperature, higher DO concentrations might be required to balance a loss in nitrification rate. For processes with denitrification, elevated levels of DO can hamper denitrification performance if DO-rich water is recirculated to the anoxic zones.

Low DO concentrations have been associated with high emissions of nitrous oxide (N_2O) (Kampschreur *et al.* 2009). Some groups of filamentous microorganisms can compete with floc-forming organisms during low DO concentrations (<1.5 mg/l), which could affect sludge settleability (Martins *et al.* 2004). This may set a lower bound on the acceptable DO level in an aeration basin.

DRIVERS FOR PROCESS CONTROL

For a control system to be successful it is important to consider what incentives can motivate an organisation or individual to support good performance. Important insights into the challenge of creating a successful control system are presented in Rieger & Olsson (2012).

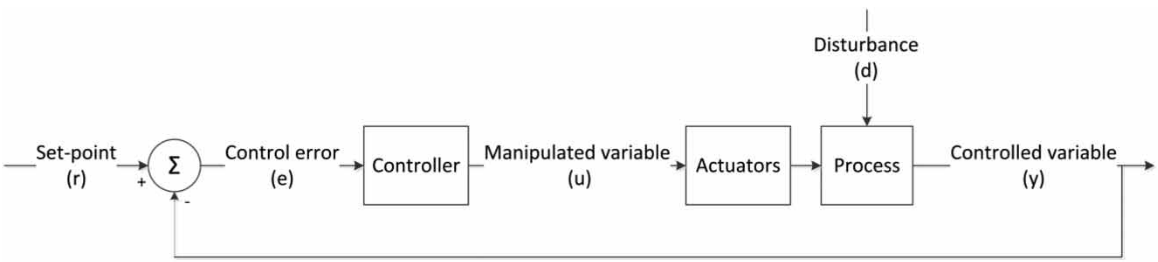

Figure 1 | A simple feedback system. The sensor is included in the process box.

Table 1 | List of variable names in a closed loop system. The terminology in this paper in bold

Variable	Control science name	Other names
Controller set-point	**Set-point** (r)	Reference value
Control error	**Control error** (e)	Deviation
Process input (output from controller)	Input signal (u)	**Manipulated variable**, Control signal
Process output	Output signal (y)	**Controlled variable**, Process value (PV), Measured value (MV)

With automatic controllers the process supervision can be tighter, enabling operation closer to any constraint such as effluent criteria. Of course, cost is an important incentive at the management level, and cost-benefit analysis is an important tool to be able to include all aspects of the control system.

IMPORTANT ELEMENTS IN DISSOLVED OXYGEN CONTROL

Air supply system

The blowers are the first stage in a diffused air system. Adequate blower system design is required for efficient control of the DO concentration. During the design process it is important to consider that the air flow demand varies over the day, week and year as well as along the aeration tank. The flexibility of a blower system is crucial for the performance of the aeration system since plants need to handle a large variation in load.

Historically, inlet vanes or outlet dampers have been used to meet a varying demand but not in an energy efficient manner (WPCF & ASCE 1988). Today, blowers supplied with variable frequency drives (VFD) allow turndown of the aeration capacity. Centrifugal or positive displacement blowers are the two main types of blower (Keskar 2006). Centrifugal blowers – such as turbo blowers – can be controlled at a fixed air flow rate set-point by varying the blower capacity, while positive displacement blowers provide a constant flow independent of the system pressure. Positive displacement compressors with VFD have a nominal efficiency of 50–60%, while centrifugal blowers have a higher efficiency (65–85%) (Keskar 2006). The positive displacement blowers are more common for small installations.

The air passes through a valve before it is diffused into the aeration basin. Butterfly valves, damper valves, globe valves, plug valves, etc. have different mechanical design. The flow dynamics of the valve describes the flow rate as a function of the valve position. The flow characteristics for a fixed pressure drop over the valve can either be linear (flow is proportional to valve lift), equal percentage (flow is proportional to the first derivative of the flow with respect to the valve lift) or quick opening (a small change in valve lift produces a large change in flow) (Seborg *et al.* 2010). An example of a damper valve with quick opening dynamics is illustrated in Figure 2. The valve has an actuator which is commonly pneumatic or electric, setting the valve opening depending on the control signal to the actuator (Keskar 2006).

When the aeration control loops call for a certain air flow rate the actuator changes the valve position in the air

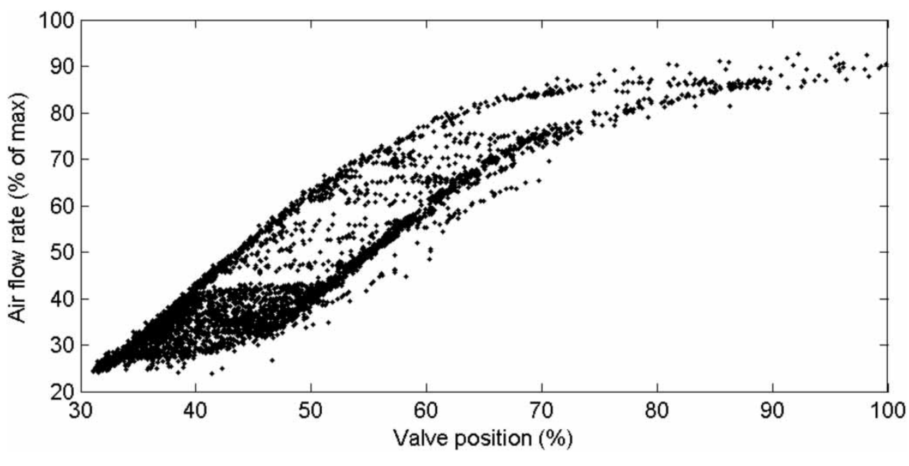

Figure 2 | Example of non-linear valve characteristics with hysteresis (damper valve).

grid, which will cause a change in header pressure. Given constant pressure control the change in header pressure will be compensated for by changing the blower capacity or by using inlet/outlet throttling. Apart from blower capacity control, blower control also includes start–stop procedures for multiple blowers as well as safety procedures during start and stop and surge control which prevents instability at low flows by maintaining a minimum flow (Keskar 2006).

One option to minimise the pressure loss over the air supply valves is to use the most-open-valve principle (MOV) (Alex *et al.* 2002). MOV will vary the pressure in the air headers until the most open valve in the system is nearly completely open. Another alternative is to relate the pressure set-point to the total air requirement. There is also the option of using direct flow control where pressure control is omitted and the blower capacity is adjusted to meet a total air flow requirement. MOV and direct flow control are examples of power-minimising control strategies, not aeration control strategies.

There are several types of diffuser used in activated sludge basins. One way to categorise them is (1) porous or fine porous diffusers, (2) non-porous equipment and (3) other devices, including jet aerators (Metcalf & Eddy Inc. 2003). Another way to categorise diffusers is by bubble size: coarse bubbles and fine bubbles. It is recognised that coarse bubble aeration has a lower oxygen transfer efficiency than fine bubble aeration (Groves *et al.* 1992). Other factors that have been reported to affect transfer efficiency are diffuser layout, diffuser density, diffuser age, and SRT (Groves *et al.* 1992; Rosso *et al.* 2008).

The oxygen requirement along a plug-flow aeration tank will decrease as the concentrations of organic material and nitrogen decrease. To avoid unnecessary aeration the aeration intensity along the tank should be decreased to balance the requirement. During aeration system design, this is often handled by means of tapered aeration. Tapered aeration decreases the diffuser density along the tank, but does not satisfactorily compensate for load variations. A more flexible way is to divide the tank into zones and control the air flow rate to the individual zones, to be able to compensate for spatial and temporal load variations. An example of the ammonium and DO profiles in a plant with tapered aeration but without individual zone control is illustrated in Figure 3. When ammonium is removed the DO concentration rises to high levels.

Sensors

When measuring different properties in the activated sludge process there are a range of methods to use. Vanrolleghem & Lee (2003) present state-of-the-art on-line measuring equipment in WWTPs. The authors list general parameters (e.g. suspended solids and temperature) and unit process specific sensors (e.g. DO and BOD measurements for the activated sludge process).

There are two main ways to measure DO: by electrochemical cells (galvanic or polarographic) or by using optical sensors with luminescent techniques (Keskar 2006). Galvanic cells are the dominant electrochemical technology today. With luminescent techniques less maintenance is required compared to membrane sensors since there is no need for membrane cleaning and maintenance. However, the sensor cap needs regular replacement.

In the nutrient removal process there is an option to measure ammonium and nitrate with automated wet chemistry techniques, with *in situ* ion-selective electrodes (ISE) or with titrimetric sensors (Vanrolleghem & Lee 2003).

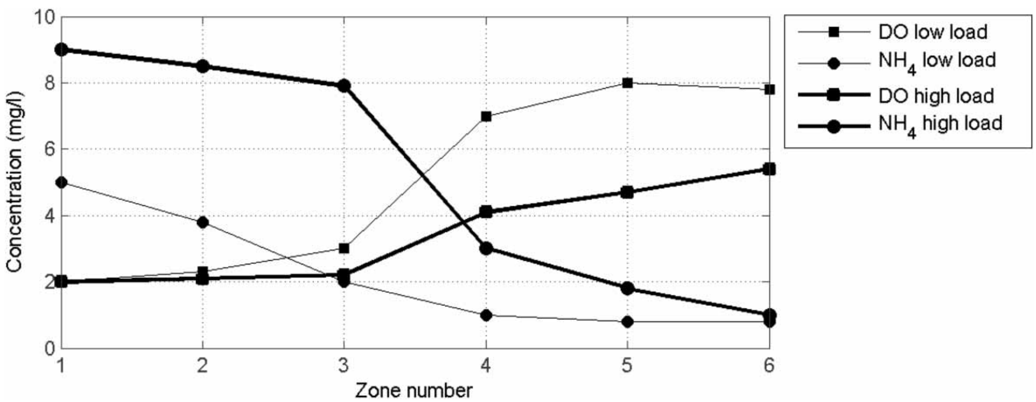

Figure 3 | DO and ammonium profiles along an aeration tank with nitrification and no individual zone control, using one valve to the whole tank, which is adjusted to control the DO in the first zone to 2 mg/l.

The ISE sensor has a faster response time than the other two methods since no sample pretreatment such as filtering is required.

A sensor has certain properties, depending on the measurement technique and device, such as measurement range, response time and accuracy. Different response times for sensors were determined in Rieger *et al.* (2003). Sensors requiring filtration were estimated to have a total response time of 10–30 minutes depending on the speed of filtration, while the ISE sensors and optical sensors without filtration were modelled with a 1-minute response time.

Measurements are often prone to be noisy, which can hinder the performance of a controller. Noisy signals should therefore be filtered. Fast sampling may allow for averaging or more sophisticated filtering such as exponential filters. Many sensors have implemented an internal filter.

Compared to using a sensor purely for monitoring, a sensor in a control loop can seriously hamper the control performance if the signal is faulty. Rosen *et al.* (2008) discuss different types of sensor faults in their attempt to model sensor and actuator behaviour. The list of sensor faults include normal noise according to specifications, excessive drift, shift (off-set), fixed value, complete failure (no signal), wrong gain and erroneous calibration. Thomann *et al.* (2002) present a monitoring concept for on-line sensors having drift, a shift or outlier problems. A detailed method on how to quantify sensor uncertainty is found in Rieger *et al.* (2005).

Sensor maintenance is an important factor to achieve good performance, and the cost for maintenance work should be included in a cost-benefit analysis when a new control strategy is considered. *In situ* ISE sensors have been reported to require around 2 hours of maintenance per week and sensor (Kaelin *et al.* 2008), to mainly take care of dirt around the electrode, particularly chemical film formation on the membrane. The sensor location will impact the need for maintenance. Influent waste streams constitute a more hostile environment for *in situ* measurements than secondary settler effluents.

The sensor location will also impact what information is available for the controller. There are often large time delays in treatment plants. Time delays are – in a feedback system – not easily managed by a controller. If the ammonium is to be measured for aeration control the sensor can be placed in the aeration basin *in situ*. In a plug-flow system and especially at larger plants, a sensor placed in the last aerated zone will be delayed in relation to the concentration in the first aerated zone. An option would be to place the sensor in the middle of the aeration tank, providing feedback with respect to the first aerated zones and feedforward action with respect to the last aerated zone.

In theory, plants with a common sludge return could be expected to need less instrumentation than plants with separate sludge return for individual treatment trains. In reality there is often an individual behaviour in treatment trains with a common sludge return due to influent variation, status of equipment, etc.

Since hardware sensors or measurements can be difficult or expensive to handle, there has been a development towards using software sensors (soft sensors) where models are used together with simple measurements in order to calculate a variable that may be more complex to measure directly. The soft sensor has to be calibrated and validated based on measurement data, typically from a dedicated measurement campaign. During operation the soft sensor relies on information from other hardware sensors. The soft sensor can be used as a ‘shadowing’ sensor to be able to provide information about estimated sensor faults (Lumley 2002). As an ordinary sensor, the model in the soft sensors needs to be calibrated at regular intervals to maintain its prediction capability.

PROCESS DYNAMICS

The key manipulated variable to the aeration process is either the valve position – for diffused bottom aeration – or the power input – for surface aeration. Several steps are taking place before the actual nitrogen concentration is influenced (Figure 4).

If the valve is non-linear, the system is non-linear in each of the steps in Figure 4. The non-linearities are smooth and monotonically increasing. This makes them readily manageable in control.

The origins of the non-linearities are as follows:

- Non-linear valve characteristics, as described above.
- Decreased oxygen transfer efficiency at higher air flow rates due to aggregation of bubbles, decreasing the total transfer area towards the water phase as well as increasing bubble rising times.
- Saturation of the DO concentration.
- Growth rate dynamics of nitrifiers. At lower DO concentration the relationship is approximately linear, while at higher DO concentrations an increase in DO has a limited effect on the growth rate.

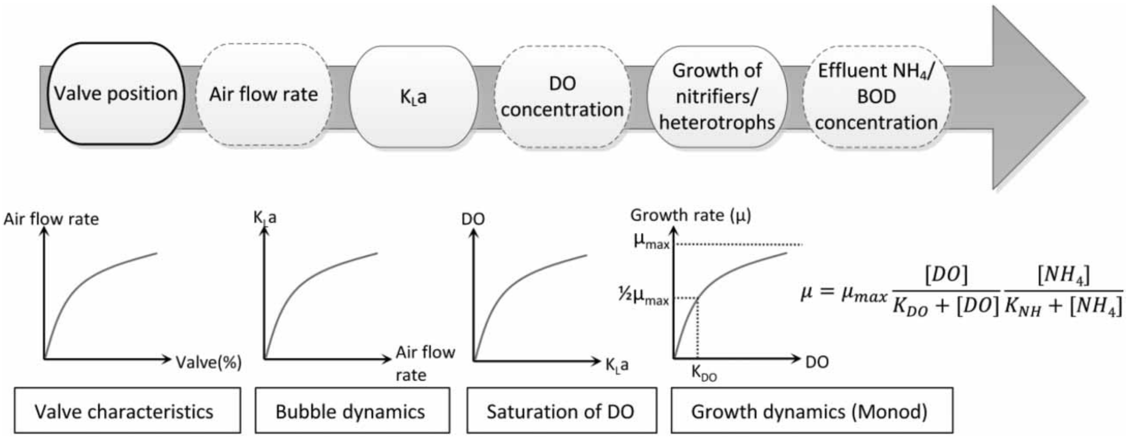

Figure 4 | The multistep process from valve opening to effluent ammonium concentration in a nitrification basin with diffused bottom aeration for the case of pressure controlled blowers. The manipulated variable (bold) and variables available for on-line measurements with standard sensors (dashed) are marked in the figure. The steps in the process are considered to be non-linear as schematically depicted in the figure with brief explanations. The equation describes the Monod functions for DO and ammonium. K_La is the oxygen transfer rate. K_{NH} and K_{DO} are half-saturation constants.

The response time in each of the steps in Figure 4 increases along the arrow from a few seconds (change in valve position) to hours (change in effluent concentrations). The aeration system has been modelled to have a response time of approximately 4–5 minutes (including control loops but excluding DO control, rise time of bubbles and delays in air supply system) (Rieger *et al.* 2006). The response time of the aeration system including DO dynamics is of the order of 30 minutes.

MODELLING AS AN EVALUATION TOOL

Many of the studies performed on aeration control and on other unit processes in municipal WWTPs are investigated through modelling and simulation of the processes. There are several benefits of modelling of WWTP processes. Jeppsson (1996) mentions plant design, testing hypotheses in research, development and testing of control strategies, forecasting, analysis of total plant performance, and education as general purposes for using models in the field of wastewater treatment.

A commonly used tool for evaluating control strategies in a model is the IWA/COST Simulation Benchmark (BSM1) (Copp 2001), developed to provide a unified framework for control strategy evaluation. BSM1 is an implementation of the Activated Sludge Model No. 1 (ASM1) (Henze *et al.* 1987). In research on control and modelling of WWTPs, ASM1 or simplifications of ASM1 are widely used for evaluation purposes, but also implemented in model-based controllers. There are further developments of the ASM1 model into the ASM2d and ASM3 models (Henze *et al.* 2000), and the BSM concept is continuously developed. One important extension of BSM1 is BSM2 (Nopens *et al.* 2010), covering not only the activated sludge process but also primary treatment and sludge handling (Jeppsson *et al.* 2006). Jeppsson & Vanrolleghem (in press) present publications related to the BSM. The BSM models are available in several simulation platforms including BioWin™, EFOR™, GPS-X™, Matlab/Simulink™, Simba®, STOAT™ and WEST® (Copp 2001).

To evaluate plant and controller performance, performance indices have been used for evaluation. Copp *et al.* (2002) presented performance indices for the evaluation of control strategies based on simulations in the BSM. The indices include the effluent quality index (EQI), which integrates the total amount of pollutants for the process with different weights depending on their severity, as well as cost indices for aeration energy, pumping energy, etc.

CONTROL STRUCTURES

Control structure design is about how to set up the control system, namely which variables to control, which variables to manipulate and how to combine these two sets of variables to create control loops. Depending on the process at hand different types of controller structure can be considered in a process control scheme.

There is no unique way to categorise the control structures for aeration control. In this paper we have chosen the following four categories:

(A) DO cascade control
(B) Ammonium-based supervisory control

(B1) Feedback control
(B2) Feedforward–feedback control
(C) Advanced single input single output (SISO) and multiple input multiple output (MIMO) controllers
(D) Control of the aerobic volume.

The motivation behind this categorisation is that each of the levels requires a different level of complexity (programming, sensors) in the control system. The block diagrams for each of the strategies are presented in Figure 5. The controllers in A and B are SISO controllers. Advanced controllers have often a MIMO structure, but they can also be SISO. To be precise, advanced SISO and MIMO controllers (structure C) are not necessarily part of a unique control structure. An advanced controller can be a part of control structure A or B. But in C, we consider control strategies that are typically not included in a basic course in automatic control.

In a mutivariable system each manipulated variable affects several controlled variables causing loop interactions. In order to decide if the system should be controlled by conventional SISO controllers (decentralised control) or if a MIMO controller should be used, loop interaction analysis is needed. Examples of interaction analysis with application to activated sludge systems are given in Machado *et al*. (2009), Ingildsen (2002) and Samuelsson *et al*. (2005). There are several other control loops besides those found in aeration control at a treatment plant. To manage these loops with a unified approach is often referred to as plant-wide control.

DO cascade control

There are several levels of control in an aeration control structure. The innermost controllers are generally simple feedback controllers, often set up in cascade. Air flow rate control and DO control benefit from cascade control, since it is a non-linear process with increasing response times and an intermediary measured variable. The intermediary variable (air flow rate) is measured and controlled in an inner (slave) loop, and the outer (master) loop controls the controlled variable (DO). The benefit of a cascade control system is that non-linear dynamics of the elements in the slave control loop can be compensated for by the slave controller, meaning the master controller can 'see' a more ideal behaviour, which simplifies controller tuning.

Ammonium-based supervisory control

The DO set-point in the DO cascade controller is decided by the operator. To improve the control performance an externally calculated set-point can be used. The DO set-point can be calculated based on the measured ammonium concentrations in the outlet of the activated sludge process or from an *in situ* sensor (structure B1). This is nothing else than a triple cascade controller. Another way to calculate a supervisory set-point is by feedforward control for improved disturbance rejection (structure B2, including feedback ammonium control). Commonly, the key disturbances are the influent ammonium concentration and influent flow rate to the plant. Feedforward control has the ability to react faster to a disturbance, since it will predict the impact of the disturbance before it affects the process by using a feedforward model. The accuracy of the prediction will depend on the model quality. A perfect prediction never occurs, so feedback control should be added to feedforward control in order to make the final correction based on the true measurement.

Advanced SISO and MIMO control

With advanced control we here refer to different model-based and optimal controllers. Model-based controllers include a large group of control algorithms which all make use of a process model in the control law. The model can be either black-box or be based on 'physical' process equations. Often the model can be used to find a controller output that is optimal in some sense. Optimal control, as defined here, assumes a cost function to be mathematically minimised, and attempts to find the best solution to the minimisation problem given constraints on the system.

Control of aerobic volume

Additional to adjusting the aeration intensity, parts of the aerobic volume can be switched on and off. The control is often feedback or feedforward, as can be seen in Figure 5, and the output of the controller is a decision on whether or not a zone should be aerated. Therefore the controller commonly needs a defined rule on whether or not aeration should operate.

CONTROL ALGORITHMS

The task of a controller is to keep the process value at the set-point. The most widely used control algorithm to

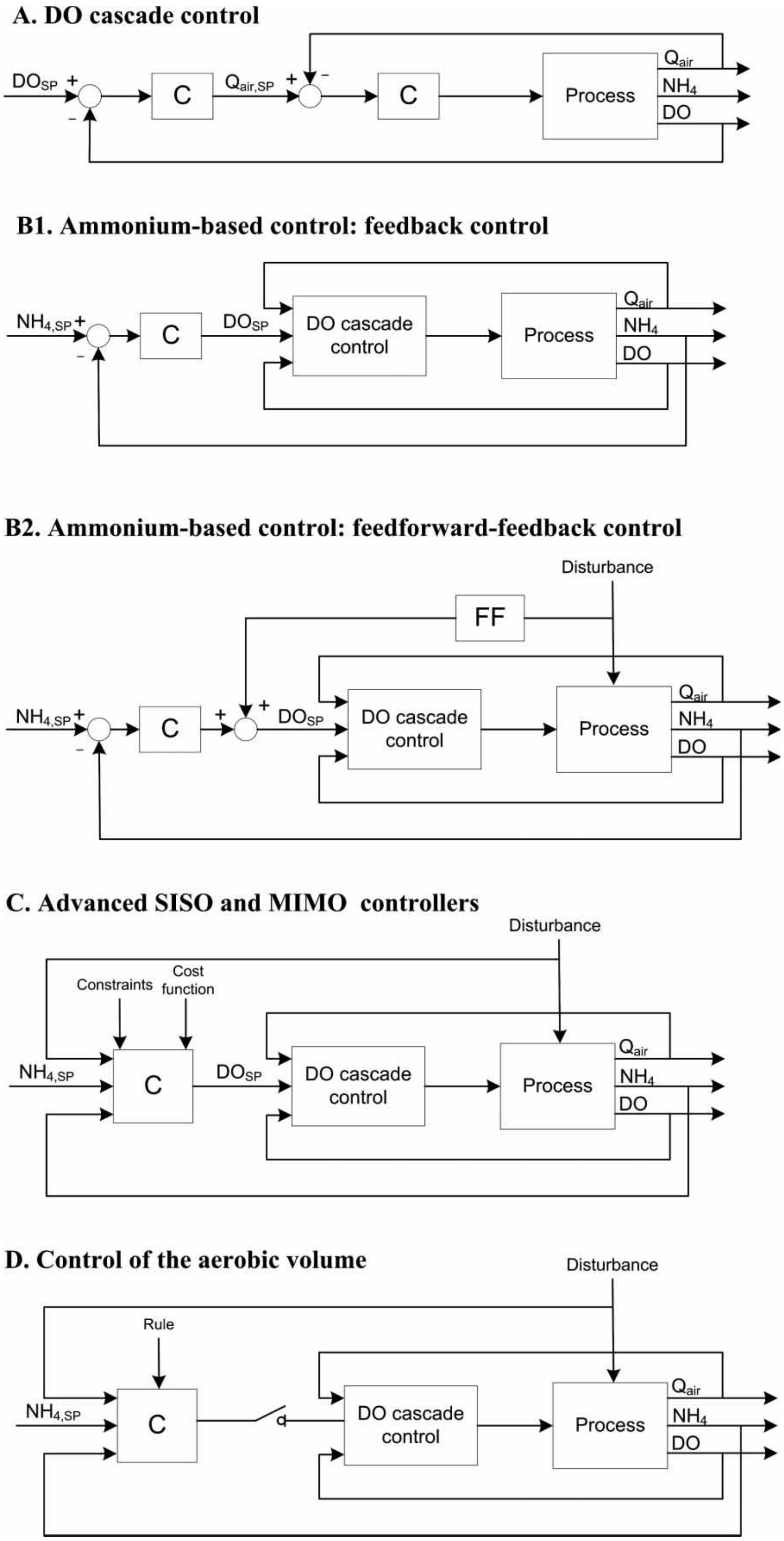

Figure 5 | Categorisation of controller structures for aeration control. SP = set-point, C = controller, FF = feedforward controller. For control structure C, the example is a model-based controller with constraints and cost-function controlling the DO set-point. Control structure A is included in structure B–D.

achieve this in process control is the PID (proportional–integral–derivative) controller (Åström & Hägglund 1995). The PID controller consists of three parts, as presented in Figure 6. The controller can either be used with all its parts or with only the P, PI or PD terms. The proportional part reacts to the present control errors, the integral part sums up previous control errors and the derivative part predicts future control errors by using the derivative of the control error. The integral part provides what is referred to as integral action. Integral action leads to an elimination of steady-state offset.

A large group of controllers can be joined under the name *rule-based control*. The most simple form of a rule-based controller include *if*…*then* rules to determine, for instance, set-points of DO based on a feedforward or feedback signal. Control of the aerated volume is commonly based on rule-based control rules.

Rule-based control in the form of *fuzzy logic control* (FLC) was traditionally applied to alternating systems and batch reactors; see for instance Traoré *et al*. (2005) and Fiter *et al*. (2005). The trend has been to expand the application towards continuous operation. Historically, fuzzy logic is an extension of Boolean logic where not only 0 and 1 are considered as alternatives but also the continuous interval in between. Membership functions are used to 'fuzzify' the controller and apply rules. At the end the fuzzy controller has to be 'defuzzified' and the end product is a nonlinear controller. FLC is appreciated for its transparency and the possibility to include process knowledge (such as operator experience) in the controller.

There are many types of model-based controllers used in advanced control strategies. Both feedback model-based control, such as *linear quadratic control* (LQC), and predictive control, such as *model predictive control* (MPC), minimise a cost function. MPC has become popular within many industries for its ability to handle constraints and to include multiple variables. MPC has been a research topic for WWTPs since the mid 1990s. There are many developments of the classical MPC method, such as robust MPC, adaptive MPC and non-linear MPC, see Weijers (2000).

$$u(t) = K\left(\underbrace{e(t)}_{\substack{\text{Proportional}\\ \text{(P)}}} + \underbrace{\frac{1}{T_i}\int e(t)dt}_{\substack{\text{Integration}\\ \text{(I)}}} + \underbrace{T_d\frac{de(t)}{dt}}_{\substack{\text{Derivation}\\ \text{(D)}}}\right)$$

Figure 6 | The three parts of a PID controller.

CONTROL OF AERATION INTENSITY

DO cascade control

DO control has been common practice in process control for many decades, and DO control was first implemented more than 40 years ago. Ingildsen (2002) reproduce a table originally published in Andersson (1979) with results on energy savings at seven Danish WWTPs in the 1970s due to implementation of DO control. The total power savings range from 2.5 to 60% with an average of 27%, emphasising that the gain achieved from implementation of a new control strategy very much depends on the situation before the upgrade.

An example of a full-scale evaluation of individual zone control of DO at the Käppala WWTP (Stockholm, Sweden) is reported in Thunberg *et al*. (2009). The goal was to distribute the air according to the oxygen demand along the length of the bioreactor, and to avoid high air flow rates at the beginning of the basin. The original control strategy is based on a linear air flow distribution and makes use of two DO sensors: one at the beginning and the other at the end of the four aerated zones. The first sensor decides the total air flow to the reactor and the second sensor the slope of the step-like air flow profile. With individual control of each zone a saving of 26% of air flow rate was achieved over a 1-year period.

Air flow distribution was also investigated by Sahlmann *et al*. (2004) where four different DO set-point combinations were compared for three aerobic zones in an A^2O (anaerobic–anoxic–oxic) process. The zone distribution of air flow and standardised oxygen transfer efficiency (αSOTE, measured by off-gas method) were analysed for different loads. An air flow rate saving of 15% was achieved by using a DO profile of 1.2/1.2/1.5 mgDO/l compared to 2/2/2 mgDO/l. There is no information on variations in nitrogen removal performance, other than that the effluent concentrations met the discharge criteria.

Classical PID control has been investigated and developed further. As an example, Tzoneva (2007) evaluated two standard PID tuning methods (Ziegler–Nichols and a relay tuning method) using the BSM1. The paper presents a method on how to perform real-time tuning for the purpose of adaptive control. Adaptive control uses controllers which have rules for updating controller parameters with the purpose to adjust to changes in process dynamics or disturbances.

Gerkšič *et al.* (2006) evaluate gain scheduling of DO PI control in BSM1 and in a pilot-plant MBBR (moving bed biofilm reactor) based on a model-based estimation of the respiration rate. Gain scheduling changes the controller parameters depending on the value of a scheduling variable, in this case the estimated respiration rate. The goal was to compensate for process non-linearities. The pilot-plant results show a slight improvement in DO control performance.

Another example of adjusting controller parameters is presented by Han *et al.* (2008) who simulate a fuzzy DO controller. The simulation study employs a piecewise linearised relationship between the air flow rate and the DO. The PI parameter values are tuned for each linear section. A blending of the output of the PI controllers is performed based on Gaussian membership functions. The result is a controller acting similarly to gain scheduling which changes the controller parameters based on the DO set-point.

Many DO cascade controllers have been investigated in full-scale and in simulation studies during the 21st century. Most often they are used as a reference strategy when evaluating more advanced controllers as can be seen in the following sections. The case studies presented in this section are compared in Figure 7.

Ammonium-based supervisory control – simulation studies

In Krause *et al.* (2002) the determination of the DO set-point using feedforward–feedback control is combined with rule-based control. The feedforward controller compares the nitrification load to the nitrification capacity and use rules to switch aerated compartments on and off and to step the DO set-point up and down. The set-points are compared with the set-points of a feedback controller which bases the set-point on measured ammonium in the outlet of the aeration tank. The feedforward controller excels at reducing ammonium peaks due to an early increase of aeration during peak load. This is important, particularly in Germany since the German effluent standards never allow the plant to exceed the effluent limits in grab samples.

Rosen (2001) describes challenges involved in monitoring and control of wastewater treatment operation and outlines the possibilities for multivariate monitoring and control. With respect to control of aeration systems, the thesis covers set-point adjustments based on clustering to make the process return to its preferred process state and a multivariate feedback controller that calculates appropriate set-points for lower level controllers. The techniques are applied to DO set-point control by simulations of a reduced order ASM1.

Serralta *et al.* (2002) present simulations where DO in the last aerobic tank and nitrate are controlled with supervisory and fuzzy control in a model of a Bardenpho process using ASM2. Pressure control in air mains, air flow control, DO control, ammonium control and nitrate control by control of internal recycling all had fuzzy controllers.

A rule-based feedforward methodology was developed by Shen *et al.* (2010) for an A^2O process, based on the cumulative frequency distribution of the influent ammonium and the C/N (carbon/nitrogen) ratio. Optimal set-points for the feedforward rules were created based on steady-state simulations and validated in dynamic simulations. An 8.5% reduction in air flow rate was found for similar total nitrogen concentrations in the effluent.

In Murphy *et al.* (2009), DO set-points are determined with a feedforward rule-based approach depending on the influent ammonium load. The DO set-points (between 1 and 2.5 mg/l) were calculated for the four aerated zones at the Mangere WWTP in New Zealand using steady-state Monte-Carlo simulations searching for the lowest DO profile for a chosen ammonium load which satisfies the DO limitation and the discharge criteria. Other data such as temperature are not included in the calculations.

Using a hierarchical control structure, Machado *et al.* (2009) compute the most economical set-point for a number of decentralised controllers, controlling for instance the ammonium in the effluent of an A^2O process. The topmost controller in the hierarchy was a so-called cost controller, constituting three PI controllers (representing ammonium, nitrate and phosphorous set-point manipulation) with the total operating cost as the controlled variable. The PI controllers were designed based on first-order-plus-deadtime (FODT) models created from step-response tests. The cost controller is not really control structure B (supervisory DO control); it is rather a supervisory ammonium controller.

The energy consumption for different controller settings in an ammonium feedback controller was investigated in Åmand & Carlsson (2012a). Simulations were performed in a BSM1 model with one aerobic compartment and with only daily influent variations. Different ammonium controllers were compared to an optimised K_La vector, where all simulations reach the same daily ammonium concentration in the effluent. The optimal solution is 1–4% more efficient in terms of K_La than constant DO control depending on the variation in load, and close to optimal performance

Authors		Process	Structure		Control goal	NH_4		NH_4 SP	NH_4 max		NO_3		TN		DO		Energy		Temp °C	Eval. time	NO_3 ctrl	Comment/ effluent limit
			Exp	Ref		Exp	Ref	Exp	Exp	Ref	Exp	Ref	Exp	Ref	Exp	Ref	Exp (%)	Ref (unit)				
Husmann *et al.* (1998)	f	Step-feed	B1+D	A	Low NH_4 Energy reduction	0-2	0-6	2.0	10	14	1-12	6-20			0.5-2.0	2.5	-16	1 Y data (AE)	16	1 W	N	NH_4 <10 TN<18
Husmann *et al.* (1998)	f	Step-feed	B1+D	A	Maintain nitrification	0-6	2-18		6	18	4-12	4-14			0.5-2.0	2.5	-16	1 Y data (AE)	7	1 W	N	NH_4 <10 TN<18
Steffens and Lant (1999)	s	Pre-DN	C	B2	Operating costs License violation Deferred cap. exp.								3-15	4-22	0-4.5	0.5-5	-33	453 (AUD/d)		4.5 D	Y	TN<10 (GS) +1015 % carbon
Galarza *et al.* (2001)	s	RDN/DRDN	B1	A	Reduce costs Denitrification perf.	0.5-2	0.5-2		2.5	4					1-2.5	2.5	-16	(AFR)		8 M	Y	T>13: NH_4 +NO_3 <13
Suescun *et al.* (2001)	s	Pre-DN	B1+D	A	NH_4 +NO_3 ctrl Energy reduction	0-2	0-4	2.0	10	14	0-12	0-12			1.0-2.25	2.0	-11	(AFR)		1 Y	Y	T<13: NH_4 <3, NO_3 <15, T>13: NH_4 <2, NO_3 <10
Serralta *et al.* (2002)	s	Bardenpho	B1	A	TN below limit Minimise NO_3	0.5-1.5			1.5		<0.2	<0.4			1.3-1.6	2.0	-10	(estimate, AE)		5 D	Y	
Krause *et al.* (2002)	s	Pre-DN	B2+D	B1+D	Effluent permit Energy reduction Reduced C source	0.8	2		6	7.5			6-16	6-16	0.5-2.0		+6	(AFR)	11-16	17 D	N	T>12: NH_4 <5 (GS)
Ingildsen *et al.* (2002)	f	Pre-DN	B2*	A	Effluent permit Energy reduction Simplest controller	0.5	0.4		2.5	3.5	5.6	5.6			0.25-2.5		-5 to -15	(estimate, AE)		35 D	N	
Ingildsen (2002)	f	Pre-DN	B2	A	See above	1.6	1.1	2.0	6	6	4.1	5.6			0.25-2.5	2/1.8	-11.7	31 968 (AFR)		24 D	Y	
Ingildsen (2002)	f	Pre-DN	B1	A	See above	1.9	2.4	1.5	5						0.25-4		-13.5	37 296 (AFR)		17 D		
Ingildsen (2002)	f	Pre-DN	B1	A	See above	3.3		3.0							0.25-3		-27.5	42912 (AFR)		41 D		
Meyer & Pöpel (2003)	s	Pre-DN	B2+D	A			0.5		1.7	3.1	5.7	9.1	6.6	9.6		2.0	-25	323 (AFR)		3 D	N	
Meyer & Pöpel (2003)	s	Pre-DN	B2+D	A+D			0.5		1.7	3.9	5.7	7.7	6.6	8.5		2.0	-12	282 (AFR)		3 D	N	
Vrečko *et al.* (2003)	s	BSM1	B2	A		0.5-3.2	0.5-4	1.0	3.2	4			12-22	12-22	0.2-5.0					1 W	N (Y)	
Yong *et al.* (2005)	s	BSM1	B1	A		2.4	2.8	2.0	7.9	8.4			17.7	18.5	3.2	3.5	-9.6	15569 (AP)		1 W	N	
Yong *et al.* (2005)	s	BSM1	B2	A		2.1	2.8		6.9	8.4			17.3	18.5	3.0	3.5	-9.3	15569 (AP)		1 W	N	
Yong *et al.* (2005)	p	Pre-DN	B2	A		4.2	4.9	4.0	9.2	21	As ref.				0-6	2.5	Up to -9.6	(AE)	20	60 D	N	
Liu *et al.* (2005)	f	MLE	B2	A									4.9	7.7		2.75/ 2/0.5	-19	(AE)		11 W	Y	
Vrečko *et al.* (2006)	p	Pre-DN (MBBR)	B1	A		1.1	3.5	1.0	3.7	7.8					5.5-8.5	6.8-7.8	-23	1098 (AFR/kgNH_4)	15	5 D per strat.	N	
Vrečko *et al.* (2006)	p	Pre-DN (MBBR)	B2	A		1.0	3.5	1.0	2.2	7.8					5-8	6.8-7.8	-45	1098 (AFR/kgNH_4)	15	5 D per strat.	N	
Ekman *et al.* (2006)	p	Pre-DN	A+D	A		4.5	6.7				2.2	6.7	6.7	13			-37	348 (AFR)	22/19	5 W	N	

Figure 7 | Comparison between control strategies. *Type of study*: f = full-scale, p = pilot-scale, s = simulation study. *Type of process*: ASP = activated sludge process, Pre-DN = predenitrification, A^2O = anaerobic–anoxic–oxic, MBBR = Moving bed biofilm reactor, MLE = Modified Ludzack–Ettinger process, UCT = University of Cape Town process. *Type of control strategy:* A. DO cascade control, B1. Ammonium-based feedback control, B2. Ammonium-based feedforward–feedback, C. Advanced SISO and MIMO controllers, D. Control of the aerobic volume. *Other abbreviations:* Exp = experiment, Ref = reference case, TN = total nitrogen, SP = set-point, AFR = air flow rate (Nm^3/d), AP = aeration power (kWh/d), AE = aeration energy (no unit), GS = grab sample. *The feedforward controller is not supported by feedback control. (*Continued*)

Authors		Process	Structure		Control goal	NH4		NH4 SP	NH4 max		NO3		TN		DO		Energy		Temp °C	Eval. time	NO3 ctrl	Comment/ effluent limit
			Exp	Ref		Exp	Ref	Exp	Exp	Ref	Exp	Ref	Exp	Ref	Exp	Ref	Exp (%)	Ref (unit)				
Baroni *et al.* (2006)	f	Pre-DN	B1	A	Process eff.+stab. Energy reduction	5.4	3.0		11.6	10.1					0.3-1.7	2.2	-4	16506 (AP)		1 Y	N	
Ayesa *et al.* (2006)	f	DRN/DRDN	B1	A	Process stability Effluent quality Cost			1.0							1.5-4	1.5	-15 to -20	(AE)		1 Y	Y	
Stare *et al.* (2007)	s	BSM1	C	B1	Operating costs Treatment perf.	1.5-3	1-6	2.0	3.0	6.0			8-16	8-18	0.2-2.5	0.2-3	-0.2	538 (€/d)		1 W	Y	Effluent fines + 2.7 % carbon
Stare *et al.* (2007)	s	BSM1	B2	A	Operating costs Treatment perf.	1.5-4	0.5-8	2.0	4.0	8.0			8-18	11-16	0.2-3	0.5	+5.7	512 (€/d)		1 W	Y	Effluent fines -53 % carbon
Zhang *et al.* (2008)	s	Reduced BSM1	B2	A	Effluent standards Minimal cost	3.15	3.23		7.5	8	10.8	11.3			0.5-3		-4.7	47794 (AP)		2 W	N	
Walz *et al.* (2009)	f	MLE	B2	A	Energy reduction	As ref.					As ref.				2/1.3	2.5/2	-15	(AP)		48 D	N	
Thunberg *et al.* (2009)	f	UCT	B1	Other	Energy reduction Maintain treatment	0.2	0.2	0.2			3.5	3.6			1.8-1.2/1	1.8/1	-18	(AFR)		8 D	N	
Thunberg *et al.* (2009)	f	UCT	A	Other	Energy reduction Maintain treatment	0.2	0.4				8.1	6.4					-26	280 800 (AFR)		1 Y	N	
Thornton *et al.* (2010)	f	Bardenpho	B2	A		0-1	<0.5	1.0	1.5	0.5	6-10	7-12			0.75-1.4	2/2.1/ 1.6/0.5	-20	(AFR)		1 W	Y	-50 % carbon
Shen *et al.* (2010)	s	A^2O	B2*+D	A	Effluent criteria Energy reduction	~4	~4				~12	~14			0-1/2- 3/2-3/1	1/3/3/1	-8.5	228 392 (AFR)	20		N	Improved P-removal
Kandare and Reviriego (2011)	f	ASP	C	A	DO control Pressure control										0.5-3	0.5-6	-27	235 (kW)		1 D	N	Includes pressure optimisation
O'Brien *et al.* (2011)	f	ASP	C	On/ off	Energy reduction DO control										1-1.5	2-3	-20	(AP)		2 W	N	
Vrečko *et al.* (2011)	p	Pre-DN (MBBR)	C	B2		1.1	1.0	1.0	2.8	2.2					6-8		+19	602 AFR/kgNH_4		3 D	N	
Rieger *et al.* (2012 b)	s	Pre-DN	B1	A	Improved TN rem. Energy reduction Reduced C source			1.0					~5	~5	0-2.5	2.5/2/1	-5	(AE)	12	1 W	Y	TN< 8, TP<2 -5 to -10 % carbon
Rieger *et al.* (2012 b)	s	Pre-DN	B1	A	Improved TN rem. Energy reduction Reduced C source			1.0					~5	~5	0-2.5	2.5/2/1	-20	(AE)	20	1 W	Y	TN< 8, TP<2 -20 to -40 % carbon
Rieger *et al.* (2012 b)	s	Pre-DN	B1	A	Improved TN rem. Energy reduction Reduced C source			1.0					~5	~5	0-2.5	2.5/2/1	-30	(AE)	30	1 W	Y	TN< 8, TP<2 -20 to -40 % carbon
Rieger *et al.* (2012 c)	s	Pre-DN	B1	A	Increase N rem. Energy reduction								+32% rem			2.0	-25	($/Y)	15		N	Effluent tax, incl. supernatant ctrl -50 % carbon
Rieger *et al.* (2012 c)	s	Pre-DN	B2	A	Energy reduction	0-1	1-2		2.5	3.5			+48% rem		0-2	0.5-4	-30	($/Y)			N	
Rieger *et al.* (2012 c)	f	A^2O	B1	A	Red. op. costs Higher N removal Higher P removal	1.0		2.0	4.0				+40% rem			2.0	-17	($/Y)		4 W (B1)	N	Effluent tax

Figure 7 | continued.

can be reached with supervisory ammonium control. Further developments of this approach is presented in Åmand & Carlsson (2012b).

In Rieger *et al.* (2012a), the authors review important aspects of ammonium control in general and ammonium feedforward in particular, including disadvantages of feedforward controllers and selection criteria. The authors discuss the limitations to ammonium removal created by the mass of nitrifiers in the system. The limitation of the nitrifier mass cannot be compensated for by an increased aeration above DO concentrations of about 2 mg/l, since the average mass is based on the average ammonium load removed. To mitigate the effect of high nitrogen load when DO is high, the options remaining are increased aerobic volume (swing zones) or looking at load buffering.

Ammonium feedforward is not always beneficial, as exemplified by Rieger *et al.* (2012b). Using simulations for control system design of the Nansemond treatment plant, the authors test different ammonium feedback and feedforward controllers, including continuous and discrete (on–off) determination of the DO set-point. See Figure 7 for savings at different water temperatures for total nitrogen concentrations just below 5 mg/l. Continuous ammonium feedback was considered the preferable option compared to discrete feedback mainly due to less wear and tear on the equipment. Feedforward control (comparing incoming ammonium load to nitrification capacity) was evaluated compared to feedback ammonium control for dry weather flow and for an ammonium peak. The impact of the feedforward controller was limited since it lost its control authority quickly at DO concentrations above 2 mg/l.

Ammonium-based supervisory control – full-scale and pilot-scale case studies

Rule-based feedback control of outlet ammonium was evaluated in full-scale by Husmann *et al.* (1998) during warm and cold temperatures. DO concentrations were changed in steps and a facultative aerated zone was controlled. The controller managed to reduce effluent ammonium and nitrate concentrations with an aeration energy reduction of 16%, and the controller maintained the ammonium concentration below the effluent permit.

The control of the DO set-point was also reported in a study by Suescun *et al.* (2001). The DO set-point was adjusted every 4 hours to compensate for the deviation in actual effluent ammonium compared to the ammonium set-point. The DO was controlled with conventional feedback. There are facultative zones which can be aerobic/anoxic depending on operational aspects. The simulations demonstrated an air flow rate reduction of 11%. The controller was eventually combined with a similar control loop for suspended solids and verified in full-scale at the Galindo-Bilbao WWTP (Galarza *et al.* 2001; Ayesa *et al.* 2006).

Meyer & Pöpel (2003) performed simulations and pilot-plant testing of a predenitrification system controlling the DO set-point and the ratio of aerobic and anoxic zones using a fuzzy controller. The system combined feedforward of influent ammonium with feedback of the outlet ammonium concentration and nitrate as well as the outlet ammonium time variation. Compared to a fixed set-point of DO with relay control alternating between 0 (meaning prolonged denitrification) and 2 mg/l with constant zone division, the fuzzy controller air flow was decreased by 24%.

A similar approach to control was made by Yong *et al.* (2006), where, apart from the DO set-points, the external carbon dosage was controlled. Inlet and outlet ammonium concentrations were used as inputs to the controller, which was tested in simulations and in pilot-scale testing. Pilot-plant testing showed an increased removal of ammonium of 16% and an air flow rate decrease of 10%. No graphs were shown of pilot-plant performance.

In full-scale experiments, Ingildsen (2002) concluded that *in situ* nutrient measurements in combination with simple control strategies can improve the plant performance significantly. An important advantage of the *in situ* instruments was the fact that different sensor locations could readily be tested and compared. Different controllers based on feedback of ammonium and feedforward of ammonium load were tested, see Figure 7. The best performance was achieved by a feedback ammonium controller.

In Vrečko *et al.* (2006), the air flow per kg of ammonium removed was reduced by 45% with a combined feedforward and ammonium cascade control method, compared to PI control with a constant DO set-point. Using only ammonium feedback control reduced the air flow per kg of ammonium removed by 23%. The results came from a pilot-plant MBBR experiment. The oxygen set-point in the reference case was higher than the average DO concentration during the evaluation of the ammonium-based controllers; nevertheless the effluent ammonium concentration was substantially higher during constant DO control (see Figure 7).

Baroni *et al.* (2006) present a full-scale implementation of a fuzzy logic system in a predenitrification system. The DO set-point and the air supply were controlled through fuzzy logic. The installations were running for approximately a year and produced long-term and short-term process stability as well as energy savings.

Two examples of full-scale implementations of the Bioprocess Intelligent Optimisation System (BIOS) were presented in Liu *et al.* (2005) and Walz *et al.* (2009). The BIOS uses feedforward control to update the DO set-point and internal recycle flow in a predenitrification plant based on on-line influent measurements and process data. In Liu *et al.* (2005), energy savings of around 19% were achieved with improved nitrogen removal, while Walz *et al.* (2009) demonstrated a 15% energy reduction with maintained nitrogen removal. It is not specified in Liu *et al.* (2005) whether the total nitrogen reduction from implementing BIOS originated from improved aeration control or from improved control of the internal recycle.

Yoo & Kim (2009) performed full-scale testing where different autotuning methods for PID DO control were evaluated in an industrial WWTP. Together with estimation of the K_La and respiration rate (R) proposed by Lindberg (1997) and a DO set-point decision law based on the estimated R, the study demonstrated more stable treatment results for COD (chemical oxygen demand) and an energy saving potential of 5%. The control structure is not identical to that depicted in Figure 5 (B2), but it is a supervisory DO controller without ammonium measurement.

Slow ammonium feedback control together with feedback of DO in the last aerated zone to quickly counteract oxygen peaks was evaluated for 1 week in full-scale in Thunberg *et al.* (2009). Compared to the air flow distribution method with two DO sensors described before, the strategy saved 18% of air flow rate for similar treatment performance. The larger part of the energy saving can be explained with the introduction of individual zone control of DO.

Thornton *et al.* (2010) investigated a feedforward controller in a full-scale plant in the south of the UK. The feedforward model uses information about ammonium in the first anoxic zone, flow into the first aerobic zone, water temperature, settled sewage suspended solids and COD concentration and the ammonium set-point. The model is based on the ASM1 model and the controller caused the DO in zone 1 to vary between the minimum allowed level of 0.75 mg/l up to about 1.2–1.4 mg/l. A reduction in air flow rate of about 20% along with an increase in effluent ammonium was achieved compared to fixed DO set-points of 0.5–2.1 mg/l depending on zone.

Extensive evaluations of different rule-based feedforward and feedback controllers are presented in Rieger *et al.* (2012c), simulating and performing full-scale testing at three Swiss WWTPs. The authors use rules to determine switching points for the DO set-points. Energy savings and improved total nitrogen removal are achieved through reducing aeration, thereby improving denitrification and at the same time allowing for operation closer to the ammonium set-point. The new controllers are compared to the full-scale base cases with constant DO control, sometimes limited by, for example, blower constraints. Energy savings in full-scale operation amount to up to 20%, compared to fixed DO set-point control, see Figure 7.

Advanced control – simulations

Steffens & Lant (1999) describe simulation results based on different model-based controllers in comparison with classical PI control with fixed or variable DO set-points using ammonium feedback control. The DO set-point, internal recycle flow rate, return activated sludge flow rate and external carbon dosing are used as control handles. The model-based multivariable controllers included were LQC, DMC (dynamic matrix control) and non-linear predictive control (NPC). When simulating a real influent, the NPC controller displayed tighter process control with respect to a never-to-exceed total nitrogen license limit compared to the other controllers, and improvements were experienced using DMC and LQC compared to feedback control. All model-based controllers showed an increase in total costs of 16–25% compared to feedback ammonium and nitrate control. Aeration costs and license costs were reduced while carbon costs significantly increased (factor 10). It is unknown how the ammonium and nitrate concentrations were affected since only total nitrogen was reported. The model-based controller with the lowest aeration cost (LQC) is presented in Figure 7. The model in the NPC is ideal, since it is the same model as the simulation model.

Later on, Shen *et al.* (2009) compared DMC, quadratic DMC (QDMC) and various nonlinear MPC algorithms. According to their simulations, a nonlinear MPC improves performance (measured as EQI) but at the cost of increased energy consumption. QDMC is not reported to outperform DMC. Also, feedforward for disturbance rejection was investigated in combination with DMC. Ammonium feedforward brings better results than only measuring flow rate and a combination of the two signals is even better. Performance is improved through feedforward but with increased energy consumption. A similar study published by Shen *et al.* (2008) the year before compared QDMC, QDMC with feed-forward and nonlinear MPC. The authors conclude that non-linear MPC handles disturbances best and with acceptable energy consumption.

Weijers (2000) develops a methodology for an improved control system design approach. The author searches to construct a method for control goal formulation and argues in support of a mathematical approach to set up the design problem. The case studies of the thesis where the proposed control system design approach is applied are exclusively examples of model-based control. Both a predenitrification system and a carousel system are evaluated by use of simulations with the ASM1.

Several studies report on the oxygen concentration tracking. Brdys & Konarzcak (2001) investigated a nonlinear SISO MPC based on the oxygen dynamics. The method was improved through increased computational efficiency by a fuzzy predictive controller in Brdys & Diaz-Maiquez (2002). A nonlinear MPC and an adaptive model reference controller (DMRAC) were compared in Chotkowski *et al.* (2005). In Piotrowski *et al.* (2008), the nonlinear MPC is supplemented by a model of the blower system. The DO controllers above fit into the hierarchical control structure set up by Brdys *et al.* (2008), which is divided into the supervisory control layer, optimising control layer and follow-up control layer. The structure involves integrated control of a treatment plant and sewer system, and the optimising control layer involves slow, medium and fast time scales. The optimising control layer contains a MIMO robust MPC and other advanced methods. In this layer the DO trajectory is produced to the lower level DO controller in the follow-up control layer. In Brdys *et al.* (2008), the low-level DO controller in the follow-up layer consists of a simple proportional controller; however, the authors argue that a much better solution would be to apply an MPC for this purpose. The simulations in the report are based on real data from the Katurzy UCT (University of Cape Town) treatment system in Poland.

The set-point of DO was determined with MPC in Sanchez & Katebi (2003). Sub-space identification is used to create models for DO. The authors compare three different MPC controllers with a single PI controller with constant set-point. The evaluation is performed mainly through comparing system overshoot and settling time, criteria which are not as important in DO control as they are in servo systems.

The thesis by Holenda (2007) investigates methods to optimise the aeration length in an alternating system, but also develops an MPC controller for the purpose of DO set-point tracking. The MPC is based on a linearised version of the ASM1.

Stare *et al.* (2007) compare several control strategies in BSM1: constant manipulated variables, DO control, nitrate and ammonium feedback control (with and without feedforward control) and MPC. The lowest aeration cost is required by the ammonium and nitrate feedback controllers. When operational costs (including effluent fines) are included the MPC and feedforward–feedback controller performs the best, with slightly better performance by the MPC. The MPC model is the benchmark model with perfect measurement, meaning the MPC model is ideal since it incorporates perfect process knowledge, which is never achievable in a real process. The authors conclude that improvements by process control are more related to control structure than to choice of control algorithm.

In Ekman (2008), a bilinear discrete time model is estimated using a recursive prediction error method. Data from a simulated activated sludge process with post-denitrification were used. A bilinear MPC algorithm was derived and applied to the simulation model. The results reveal that, even though the identified bilinear model describes the dynamics of the activated sludge process better than linear models, bilinear MPC only gives moderate improvements of the control performance compared to linear MPC laws.

Zarrad *et al.* (2004) and Vilanova *et al.* (2009) compare decentralised PI controllers to multivariable model-based controllers in the BSM1. In the first paper, a nitrate recycle PI loop and an air flow rate PI loop are compared to two model-based controllers (LQC, disturbance accommodation controller (DAC)). The PI controllers demonstrated better results (measured as EQI) than the model-based controllers with respect to EQI and aeration energy. This was motivated by the fact that the dynamics of the processes at hand are very different and can therefore be controlled with decentralised controllers. Vilanova *et al.* (2009) compare the performance of a multiloop PI controller to a multivariable controller in a single aerated reactor. DO and substrate concentration are considered. The results when analysing step responses are comparable for the two controllers.

DO control by MPC with a process model incorporating classical DO dynamics is verified by Holenda *et al.* (2008). The effect of sampling time is investigated, indicating improved controller performance with decreased sampling time but this does not impact plant performance or operating costs. The MPC controller compared to standard PI control shows marginal improvements on EQI and a small increase in aeration energy when comparing simulation results in BSM1. The controller was also tested on an alternating process. DO control with MPC is also found in Ostace *et al.* (2011), who also consider ammonium MPC, which performs better than only DO control. Simulations are performed in BSM1, where only the last aerobic compartment is controlled. Effluent nitrogen is increased with ammonium MPC while aeration energy is decreased by 15%.

Model-based set-point optimisation has been investigated by Guerrero *et al.* (2011). The optimisation searched for the set-points of, for example, ammonium using a pattern search method. The optimisation minimised total operational costs, including effluent fines, pumping energy, etc. Due to this, all the control strategies did not treat the same amount of ammonium (effluent ammonium varied from 0.6 to 7 mg/l), making comparisons difficult if only looking at aeration control performance. Several control strategies are evaluated, including using constant DO of 4 mg/l and optimal set-points of ammonium and nitrate, which are fixed or updated in different intervals. The best strategies with regard to the operating costs were using separate but fixed nutrient level set-points for weekdays and weekends respectively.

Genetic algorithms (GAs) have been used to calculate model-based controllers. GAs have search strategies inspired by the process of natural evolution. Yamanaka *et al.* (2006) evaluate a cost-minimisation control scheme using BSM1. The DO set-points are calculated using GAs and a simplified process model. Multiobjective GAs were used by Beraud *et al.* (2009) to find the best set-point in three consecutive aerobic zones and obtain energy reductions of 10–20% compared to the original benchmark performance.

Advanced control – full-scale and pilot-scale case studies

The STAR control system (Superior Tuning and Reporting) (Thornberg *et al.* 1992; Önnerth *et al.* 1996) is an early example of an advanced controller in full-scale systems. Ingildsen (2002), chapter 3, presents a summary of benefits from implementations of the STAR control system in Denmark and Sweden. The summary was originally presented at a seminar in 2001 arranged by the Society of Civil Engineers in Denmark. The benefit-cost ratio ranged from approximately four to up to around 10. The savings were largely due to avoided capital costs from plant extensions.

STAR is a supervisory model-based control system using on-line measurements and process data. It was primarily developed for intermittent aeration of the Biodenitro process, and is put on top of the plant's SCADA (Supervisory Control and Data Acquisition) system. Today, STAR consists of several modules for treatment plant control. Full-scale examples of STAR are described by Nielsen & Önnerth (1995) and Thornberg *et al.* (1998). STAR was published before the main period of interest for this review and to the authors' knowledge there are published results on only alternating processes. STAR is nevertheless mentioned since it is an early example of an advanced controller.

Stare *et al.* (2006) developed reduced order non-linear models based on mass balances, and subsequently performed simulations of an MPC controller for an MBBR. The results demonstrated tighter process control with a non-linear model than with a linear model. For this reason, a non-linear MPC was evaluated in pilot-scale in Vrečko *et al.* (2011), where the MPC controller was compared with the feedforward and feedback controllers previously published in Vrečko *et al.* (2006). The MPC controller performed better than conventional feedback control, but compared with the present MPC model the feedforward controller used 16% less air flow per kilogram of ammonium removed.

Kandare & Nevado Reviriego (2011) used adaptive predictive expert control to keep the DO concentration at a specified level and to achieve an air pressure level that minimises power consumption. The controller has been evaluated in a pilot plant, showing smoother DO control compared to PID DO control with fixed parameters. The reference PID controller oscillates and does not appear to be performing as a well-tuned state-of-the-art PID controller. The study also looked at pressure set-point optimisation.

O'Brien *et al.* (2011) present full-scale results from a MPC for aeration control. The previous on/off control (0.5/1.5 mgDO/l) of the surface aerators in an activated sludge process for BOD removal was improved by using a MPC based on a black-box model. The model uses a feedforward term from measured incoming BOD. The BOD was measured with a spectral analyser ('spectrolyser'). The MPC has aerator powers as manipulated variables and DO in the two lanes as controlled variables. The constraints in the MPC are minimum power levels and minimum required DO concentrations for carbonaceous removal. Compared to the reference on-off controller the MPC saves 20% of the energy by keeping the DO concentration closer to the set-point of 1 mg/l. The on-off controller operates in the DO range of 1–2 mg/l.

Two of the studies mentioned above (O'Brien *et al.* 2011; Kandare & Nevado Reviriego 2011) do not involve ammonium control, but the controller structure is closer to control structure A (DO control). The model-based controllers are categorised as advanced control since the controllers incorporated an advanced control strategy.

CONTROL OF THE AEROBIC VOLUME

Volume control – simulation studies

Suescun *et al.* (2001) incorporated a facultative zone in the control, which could be made aerobic if the DO set-point was at its maximum level. The facultative control strategy was needed to meet both ammonium and nitrate effluent requirements.

Samuelsson & Carlsson (2002) controlled the ammonium concentration in the last compartment in a plant model through a model-based feedforward strategy which changed the aerobic volume. The approach was based on on-line estimation of the reaction rate of ammonium and combined feedforward with feedback control.

As mentioned, Krause *et al.* (2002) simulated the control of the DO set-point and aerated volume in simulations. Through a feedforward model several switching points were determined for the three aerated compartments in a predenitrification plant. The improved handling of ammonium peaks was due to the control of aerated volume together with an early increase in aeration intensity created by the feedforward controller.

Meyer & Pöpel (2003) used fuzzy logic to determine the DO set-point and the fraction of aerobic volume. Compared to using on-off control based on effluent ammonium to decide the aerated volume, the fuzzy controller could reduce the nitrogen peaks.

Volume control – full-scale and pilot-scale studies

Brouwer *et al.* (1998) is an early example where a feedforward model-based approach was used to determine the aerobic volume needed for complete nitrification. A simple process model together with estimation of biokinetic parameters through batch respirometric measurements in one of the plant compartments decided the necessary aerobic volume. Evaluations were performed in a pilot plant.

Baeza *et al.* (2002) varied the total aerobic volume in a pilot plant A^2O process fed with synthetic wastewater, leading to a 10% increase in nitrogen removal compared to operation with a fixed volume. Estimation of COD was performed by turning off the aeration for short periods, hence enabling OUR (oxygen uptake rate) calculation. The COD estimations served as inputs to a neural network model which determined the total volume to be aerated. From the description in the paper, it is not obvious if the fixed volume is smaller or equivalent in size to the average volume in the case with a varied volume.

Svardal *et al.* (2003) used the measured air flow rate and DO concentration to decide how to adjust the aerobic volume to the ammonium load. The method maximises the anoxic volume given the goal of complete nitrification. The OUR is in this study a good indicator of nitrification requirement at low ammonium concentrations, and is approximately proportional to the air flow rate. The method is based on increasing the aerobic volume when the total air flow passes certain thresholds. The paper presents full-scale results from the Linz-Asten WWTP in Austria, which uses an oxidation ditch for nitrification and denitrification.

Ekman *et al.* (2006) developed a method for aeration volume control that only requires measurements of the DO concentration. The method makes use of supervisory control where two out of three zones can be either aerobic or anoxic depending on the DO concentration in all three zones, creating a disturbance rejection effect. The strategy is evaluated in the BSM1 and in the large pilot-plant facility Hammarby Sjöstadsverk in Stockholm, Sweden, treating municipal wastewater. The pilot-plant evaluation suggests that volume control can give lower nitrogen concentrations in the effluent with less energy consumption compared to constant DO control.

COMPARISON BETWEEN CASE STUDIES

Figure 7 aims to guide the reader through the research within aeration control of activated sludge processes. It summarises comparative studies between control strategies and covers full-scale, pilot-scale and simulation results. The figure includes research that compares one or several strategies to some reference case. Some papers are listed more than once to simplify reading, since they cover several comparisons.

In order to judge the efficiency of a control strategy or control structure in a case study the reader would need some quantitative indicator, such as energy consumption, in combination with a measure of the treatment performance. Other important data include whether or not nitrate is controlled (through carbon addition or internal recycle control), changes in carbon addition, the water temperature, the duration of the evaluation and the type and level of effluent permit. As can be seen in Figure 7, not all data are available from all the studies. Even if the publications were carefully read there is no guarantee that all available data were found. If available, the control goals are listed in the order mentioned in the paper. If internal recycle is used, improved nitrogen removal could be a result of nitrate control rather than aeration control. This is not always clear from the presentations in the papers in Figure 7.

If effluent concentrations are listed as intervals in Figure 7, the numbers have been read from a graph, otherwise averages are listed. Maximum concentrations were often estimated from graphs and are not necessarily representative for the whole evaluation period.

Energy consumption is reported in various ways in the research papers. In Figure 7, the reference controller's energy consumption is given if reported by the paper, together with the type of energy measure given (air flow rate, aeration power or simply aeration energy without any specific unit). Air flow rate is recalculated to Nm^3/d and aeration power to kWh/d if reported in other units. The energy consumption of the investigated control strategy is reported as saving in percent compared to the reference case.

CRITICAL REVIEW OF AERATION CONTROL SYSTEMS

Full-scale considerations

It is challenging to draw conclusions about the superiority of one control algorithm over another in full-scale operation, given the many uncertainties in operation with regards to load, equipment and other local conditions, such as plant capacity.

It is important to recognise limited control authority due to constraints of the system such as blowers (minimum/maximum air flow), sensor signal quality or sensor location. Another reason for limited control authority is that plants have a limited mass of nitrifiers, which limits the possibility to treat peak loads. Also when operating at a long sludge age, the nitrification capacity will depend on the average load of ammonium that is treated, as discussed in Rieger *et al.* (2012a).

Controller performance has to be related to the legislative framework which the plant has to comply with, which in turn will determine the control goal for the process. Internationally, there are different levels of total nitrogen or ammonium limits. There are also different averaging periods when assessing effluent limit compliance. If the plant nitrifies more than is required by the effluent permit there is additional room for improvement by reducing aeration. Grab sampling and effluent fines on ammonium will require limitations of ammonium peaks and it might be worth an increased energy consumption to satisfy these demands. Similarly, if denitrification is limited it may be worthwhile reducing aeration and thereby increasing the total nitrogen removal while perhaps emitting more ammonium. Whether this is a good thing or not depends on plant-specific factors and effluent limits and thereby on how the control goal is defined. This is the reason why not all control structures will be useful at all plants. There is not a single evaluation criterion in place, but the evaluation is a plant-specific multi-criteria task.

Another aspect of judging the performance of a controller is the choice of the reference controller. Nitrification can only be improved if the reference case is limited in some way. Similarly, the energy saving with a new controller can be exaggerated if compared with a poor-performing reference case.

The time scale in a comparative study is important, especially in full-scale. Most full-scale studies are, due to the reality, not very long. By chance, it is possible to reach results that are not representative for the plant or for the control structure under study. In Figure 7, five of the 14 full-scale comparisons are made during longer periods than approximately 1 month, and three of these look at periods of about 1 year. It is not obvious from the papers in Figure 7 that the investigated full-scale controllers are still operating at the plants.

Modelling

Simulation is a valuable tool for evaluation and development of controllers as discussed in the introduction. Modelling can provide next to unlimited flexibility and opportunities in control and process choices. However, there is a gap between real world limitations, and the opportunities that a simulation model can offer. Simulations are a step ahead with regards to development and testing of new controllers, but it is important that the gap is not allowed to become too large.

The limitations and quality of the models used are critical in modelling and simulation, in the design of model-based control algorithms as well as in control system evaluation. It is said that 'All models are wrong, but some are useful' (the statistician George E.P. Box (Box 1979)). The models used for wastewater treatment modelling are constantly evolving (Jeppsson *et al.* 2011). It is encouraging that more efforts are being put into modelling e.g. sensor failure (Rosen *et al.* 2008), trying to add more realism to models used for controller comparison and development.

The BSM1 model and the BSM2 model differ in the design of the plant in relation to the load. Essentially, BSM1 is highly loaded. Ammonium controllers in BSM1 which are evaluated without adjusting the load or the volume of the modelled reactor have little control authority, rendering evaluations of aeration control strategies difficult. Several early investigations in the BSM1 model investigating feedforward–feedback control tolerates high DO concentrations above 2–3 mg/l (Vrečko *et al.* 2003; Yong *et al.* 2005). Despite this they cannot decrease the ammonium peaks to a large extent.

Ammonium-based control

Ammonium-based control in the form of feedforward and feedback control can add two advantages to the process: the potential to limit aeration during periods with low effluent ammonium (and possibly limit complete nitrification) and the possibility to increase aeration intensity to limit ammonium peaks during peak load. There are several examples from the last decade on well-performing supervisory controllers.

Feedforward control should be used with care, since it adds more sensors and extra complexity to the control system. The controller also has to be provided with a feedforward model. Using feedforward control can be motivated by discharge criteria where it is never allowed to be above the limit. Adding feedforward control can use more energy than using only feedback control (Krause *et al.* 2002; Stare *et al.* 2007), which could be justified by the effluent criteria. Using feedforward without a feedback loop is not recommended since feedback contributes to a more robust performance in the light of feedforward model uncertainty and can compensate for non-modelled disturbances.

A thorough discussion on feedforward ammonium-based control is given in Rieger *et al.* (2012a). Two of the points made are that (1) feedforward should only be used when there is a benefit of reacting fast to a disturbance (such as 'never-to-exceed' effluent limits) compared to pure feedback control and (2) feedforward control can easily lose control authority if nitrification capacity is limited.

Volume control

Volume control can offer three benefits. Firstly, volume control can add control authority at high loads to be able to decrease ammonium peaks. Unlike ammonium feedback control, volume control can rapidly increase nitrification capacity at times when feedforward control is limited by the lack of nitrifier mass and the DO concentrations should not be further increased. Secondly, volume control can be a tool to save energy when a plant is low loaded and parts of the volumes are not required for nitrification. Consequently the additional aeration merely contributes to endogenous respiration. Finally, if denitrification is limited then volume control can provide extra anoxic volume. Volume control can therefore be a means to balance total nitrogen removal. The efficiency of volume control is much improved if walls are limiting the zones that are switched on and off.

The possibility to control the aerobic volume demonstrates the important coupling between design and operation. Many plants are not designed to use available volumes in the best possible way. For example, the volumes for denitrification and nitrification are not always matched. Volume control can be used to better utilise the plant capacity for both organic removal, increased energy efficiency by good use of the denitrification volume, and for nitrification.

Advanced control

The wastewater treatment process is often referred to as being complex, non-linear, with a range of time constants and never being in steady-state. In the literature, it is possible to find arguments supporting decentralised SISO controllers, as well as MIMO controllers. Some argue that given the complex nature of the system, simple control cannot guarantee good performance under a full range of conditions (Brdys *et al.* 2008). Simple SISO controllers are not sufficiently robust, and control of an activated sludge system must be considered a multivariable control problem and should thereby best be handled with model-based control methods, such as MPC (Steffens & Lant 1999; Weijers 2000). On the other hand, other researchers argue that conventional well-tuned control algorithms are sufficient to achieve acceptable system performance under most conditions (Ingildsen 2002; Stare *et al.* 2007). It is possible to find sub-processes that can be controlled with linear controllers that demonstrate little coupling with other processes (Vrečko *et al.* 2002). That is particularly true for intermediate process variables – such as the DO concentration.

The DO concentration is under normal conditions well controlled with a properly implemented PI controller, as is demonstrated in several of the full-scale studies in Figure 7. Early attempts to use a self-tuning controller of higher order in a full-scale process resulted in a controller converging towards a PI-controller performance (Olsson *et al.* 1985). Despite this, attempts are still made to control the DO concentration with advanced control algorithms.

The benefit of an optimal controller such as MPC is the possibility to include constraints in combination with handling a MIMO system. If DO control is the goal, constraints in air flow rate/aerator power and limitations of DO concentrations are readily manageable with a simple controller. Hence, the benefit from MIMO and optimal controllers is not on the lower level aeration controllers, but in higher levels of control. As with feedforward control, having a never-to-exceed limit on ammonium could motivate advanced control since a predictive capacity together with handling of constraints could benefit treatment performance. Looking at Figure 7, there are so far no examples of advanced full-scale controllers outperforming supervisory controllers in continuous aeration where ammonium control is considered. Since the control strategy should be kept as simple as possible, conventional feedforward and feedback controllers should be used whenever this is adequate.

Advanced controllers could provide set-points of ammonium and other effluent concentrations, while lower level controllers can be allowed to be feedforward and feedback controllers. This concept would be similar to a plant-wide control approach, which is not within the scope of this review.

IMPLEMENTING ADEQUATE AERATION CONTROL FOR FULL-SCALE OPERATION

This review has provided a walk-through in recent developments within control of aeration in WWTPs, including important elements in an aeration control system. The process to implement a new control strategy is systematic, and involves defining the control goal and looking into process and system limitations and possibilities as well as evaluation criteria. To summarise, important aspects for aeration control success in full-scale operation are listed below. These should be considered irrespective of the control structure chosen.

Important conditions of aeration control system success are as follows:

- *Properly designed aeration system:* Proper design of blowers, piping, valves and diffusers is crucial for good aeration performance. Limitations in maximum and minimum air flow capacity are often limiting the controller performance. Without a blower with possibilities for turndown during low load there is no help to be given from a well-tuned and well-working controller further up in the aeration system. Process simulators can be a help in aeration system design.
- *Individual line and zone control:* This is the next step in adding flexibility. If the reactor is a plug-flow system, the varying oxygen demand along the reactor is best managed by separate DO control of zones along the basin, using separate actuators in each zone. Individual treatment lines could be controlled separately to handle individual behaviour.
- *Adequate instrumentation and adequate maintenance*: The main cause for controller failure is instrumentation. Sensors and actuators must always be maintained adequately, and before choosing the final control strategy this must be taken into account in the cost-benefit analysis.
- *Sensor location:* This is part of the control structure and should be carefully examined.
- *Wise controller implementation:* Basic requirements and safety nets such as maximum and minimum limits for all controller output signals, anti-windup (Åström & Hägglund 1995) of controllers with integration, a sufficient range in controller parameters and fall-back strategies are sometimes overlooked. The sampling times used in the control systems have to be carefully considered. They should primarily be selected based on response time of the controlled variable. Appropriate filtering of signals should be performed taking the sampling time into account.
- *Adjustable controller implementation:* It is important that the end user can get access rights to make necessary adjustments in the control systems. Depending on the possibilities and needs for the plant, the system vendor or the plant operating team can assume different levels of responsibility for the future developments of the system.
- *Always consider plant-specific aspects:* Given the size, load and location of a certain plant, additional aspects may be considered in control structure design and implementation. Such aspects may include effluent taxes, the level and averaging period of the effluent

criteria, the variability of the load over the day, energy cost structure, peak demand charges, design characteristics and how close the plant is to its design load.

- *Plan for control system maintenance:* To achieve successful implementation of a new control strategy as well as long-term stable operation, operators need to get sufficient training to understand and trust the basics of the controllers. Today there are several vendors offering PID controllers with autotuning capabilities (Åström & Hägglund 2006), but PID controllers can also be tuned with simple tuning rules. Process simulators can be helpful in operator training.

CONCLUSIONS

More than 10 years into the 21st century, the state-of-the-art within aeration control is changing. Ammonium feedback control is evolving towards state-of-the art in real application and there are several examples of energy savings and improved nitrogen removal from full-scale case studies. More advanced algorithms including model-based control have their success stories, but so far the main efforts have been focused on method development using simulation models. This review has not found any advanced controllers for continuous ammonium control, tested in full-scale or pilot-scale studies, which outperform conventional feedforward–feedback controllers. Generally, there is an increase of publications from full-scale case studies during the last 10–15 years, even though there is a lack of long-term studies. This development will hopefully continue to better establish the benefits of improved process control. The paper has also emphasised the importance of coupling plant design and operation, as well as the importance of taking operational flexibility into consideration during the whole plant design phase.

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