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Modelling and Control of Activated Sludge Processes with Nitrogen Removal

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Abstract

Stricter legislations on nitrogen removal together with increasing wastewater loads have successively increased the need for optimization and control of activated sludge processes. The aim of this thesis is primarily to propose and illustrate methods for modelling and control of activated sludge processes, and thus possibly improve process efficiency. The first part of this thesis describes a JAVA based simulator of the activated sludge process. Its features are discussed, and some control strategies are illustrated. The second part of the thesis presents a simple model based feedforward controller for controlling the nitrate level by using an external carbon source. Several simulation examples are used to illustrate the control law. In the third part, a model based strategy for control of the ammonium level by manipulating the aeration volume in the activated sludge process is presented. The strategy is based on exact linearization combined with some logical programming rules in order to deal with discrete volume changes. The control law is also evaluated in simulations.

List of publications

Chapter 2 is based on:

Samuelsson, P., M. Ekman and B. Carlsson (2001). A JAVA based simulator of activated sludge processes. *Journal of Mathematics and Computers in Simulation*, **56**(4–5), 333–346.

Ekman, M., P. Samuelsson and B. Carlsson (2001). Control strategies in a JAVA based activated sludge process simulator. In: *Proceedings of the 1st IWA Conference on Instrumentation, Control and Automation*, Malmö, Sweden, June 3–7, pp 539–546.

Carlsson, B. and P. Samuelsson (1999). A JAVA based simulator for activated sludge processes. In: *Proceedings of the Third International Symposium on Mathematical Modelling and Simulation in Agricultural and Bio-Industries*, Uppsala, Sweden, June 7–9, pp 127–132.

Chapter 3 is based on:

Samuelsson, P. and B. Carlsson (2001b). Feedforward control of the external carbon flow rate in an activated sludge process. *Water Science and Technology* **43**(1), 115–122. Also presented in *Proceedings of the 1st World Water Congress of the International Water Association*, Vol. 3, Paris, July 3–7 2000, pp 76–83. A shorter version is also available in: *Preprints of Reglermöte 2000*, Uppsala, Sweden, June 7–8 2000, pp 101–106.

Samuelsson, P. and B. Carlsson (2000). Feedforward control of the external carbon flow rate in an activated sludge process. Technical Report 2000-005. Systems and Control Group, Uppsala University. Uppsala, Sweden.

Chapter 4 is partly published as:

Samuelsson, P. and B. Carlsson (2001a). Control of the aeration volume in an activated sludge process for nitrogen removal. *Water Science and Technology* (accepted for publication). Also presented in: *Proceedings of the 1st IWA Conference on Instrumentation, Control and Automation*, Malmö, Sweden, June 3–7 2001, pp 43–50.

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Contents

Chapter 1

Introduction

The theme of this licentiate thesis is modelling, simulation and control of the activated sludge process with nitrogen removal in wastewater treatment plants. Three main contributions are included.

- Firstly, a JAVA based simulator of the activated sludge process is presented. Its functionality is described, and some features are discussed.
- Secondly, the problem of adding an external carbon source in order to automatically control the denitrification process is considered. A simple feedforward control method, which performs well in simulations, is suggested.
- Finally, a model based method for controlling the aeration volume in an activated sludge process is derived and evaluated in simulations. The method is a combined feedforward-feedback strategy.

In this chapter, a short introduction to wastewater treatment in general, and modelling, simulation and control of the activated sludge process in particular is given. The aim is to give the reader a short overview of the topic before the more specific approaches are considered.

1.1 A Short Background to Swedish Wastewater Treatment

Due to urbanization, wastewater treatment is becoming more and more important in the society of today. In the past, people lived in relatively small population units, often on the countryside, and the disposal of garbage was not a major problem. However, with the expanding cities and towns, disposal and accumulation of human waste products rapidly became a hygienic problem. A first step towards solving the problem was the introduction of the water closet.

Unfortunately, this just moved the problem to the surrounding environment, since the pollutant load on specific recipients such as lakes and rivers then increased.

Only after severe negative impact on the recipients was detected, the first steps towards modern wastewater treatment were taken in Sweden. First, in the 1930's, the visual problems in the recipient caused by the wastewater were partly handled by the introduction of sedimentation tanks. As the problem with oxygen depletion in the recipients was discovered, biological treatment was introduced in the 1950's. In the 1970's, the biological treatment was combined with chemical treatment in order to also deal with over fertilization of lakes and rivers due to phosphorus compounds.

In the 1990's the next step towards a sustainable development and resource efficient nitrogen removal was taken. This was imposed by the EU Urban Water Directive (91/271/EC) increasing the demands on nitrogen removal in the member countries. In Sweden it was decided to decrease the nitrogen discharges with approximately 50 percent for larger wastewater treatment plants at the coast from the Norwegian border to Stockholm. The legislations became even stricter in 1999, and the maximum yearly nitrogen average for all plants with more than 100000 person equivalents is now 10 mg/l. This is one motivation for the work in this thesis, which mainly concentrates on biological nitrogen removal. The development of new legislations for Swedish wastewater treatment plants during the last decade is further described in Winnfors (1998). The Swedish legislations are collected in SNFS (1994:7). More historical material on Swedish wastewater treatment can be found in Finnson (1994), Lundgren (1994) and Isgård (1998).

Today, treating wastewater is generally a complex, multi step industrial process, see Metcalf and Eddy Inc. (1991) for a detailed discussion of the subject. Figure 1.1 shows a typical setup for a modern wastewater treatment plant (WWTP). The first step in a WWTP is generally some kind of mechanical treatment, where large objects and heavy material are removed. The mechanical treatment may consist of, like in Figure 1.1, a grid, a sand filter and a primary sedimentation unit. To remove soluble organic matter and possibly also nitrogen from the wastewater, biological treatment is often the second step. Depending on the setup of the plant, different processes can be conducted by microorganisms. These processes are used to convert the many different compounds present in the influent water to less harmful substances. It is also common to treat the wastewater chemically. Phosphorus can be removed from the water by chemical precipitation. This may be performed in a third step, as depicted in Figure 1.1. It is also possible to add chemicals before or simultaneously with the biological treatment. Another possibility is to remove the phosphorus biologically, see for instance Isaacs and Henze (1995). As seen from Figure 1.1, sludge is removed from all processes. A step where the removed sludge can be treated is also necessary. To reduce pathogenic hazards, and to avoid odours, the sludge must be stabilized. Anaerobic digesters are often used for this purpose. The sludge is also dewatered in order to make it easier to handle.

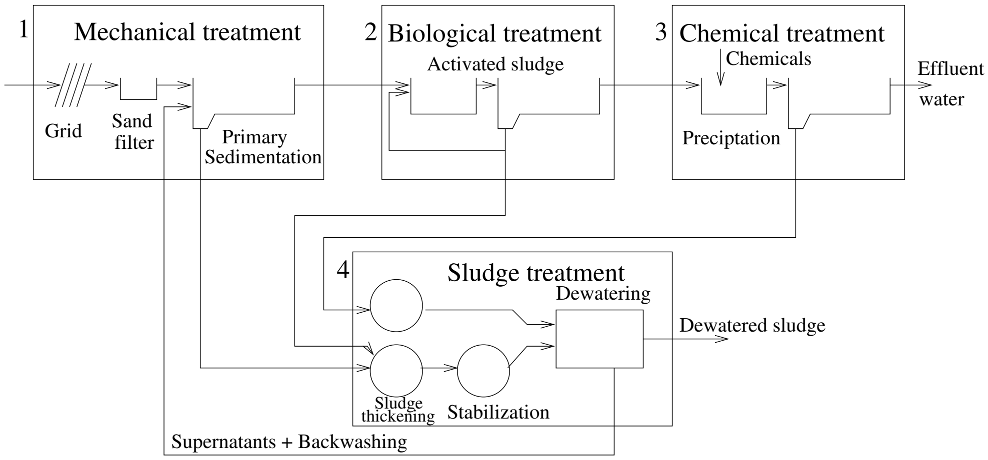

Figure 1.1: Layout of a typical wastewater treatment plant (Kommunförbundet (1988)).

The dewatered sludge may then be used as fertilizer, fuel or soil conditioner. Unfortunately, the levels of heavy metals and other toxic components in the sludge are often higher than the limiting levels imposed by the Swedish Environmental Protection Agency. Therefore it can often not be used for agricultural purposes in Sweden. This problem is however not further discussed in the thesis. More information on sludge management in general can be found in Metcalf and Eddy Inc. (1991). For an overview of the current sludge debate in Sweden, see Hultman *et al.* (2000).

1.2 The Activated Sludge Process

A common concept for biological wastewater treatment is the activated sludge process (ASP). The ASP is a biological process where activated sludge, *i.e.* microorganisms, are used to oxidize and mineralize organic matter. A schematic layout of the simplest form of an ASP is shown in Figure 1.2.

In the aerated compartment different biological reactions take place. The microorganisms in the compartment are kept suspended in the water by means of aeration. A sedimentation unit, where particulate matter is allowed to settle, is an important part of the ASP. The settled material is rich in biomass, and

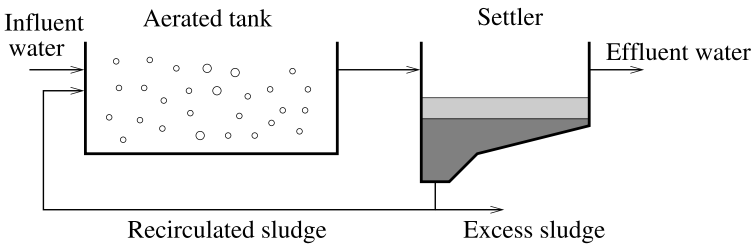

Figure 1.2: A basic activated sludge process consisting of an aerated compartment and a settler.

to maintain the biomass population in the system, part of the settled sludge is recirculated back into the aerated compartment. The other part is removed from the process as excess sludge.

The reactions occuring in the process have different time scales. The organic matter that enters the process via the influent water consists of both readily biodegradable compounds, that can be used directly for biomass growth, as well as more complex compounds. These more complex molecules are slowly degraded into molecules that are more accessible to the biomass by the hydrolysis process. The decay of biomass releases complex organic compounds, as well as inert material.

The basic ASP described above can be extended to also include biological nitrogen removal. In the case when both ammonium and nitrate is to be removed, a two step configuration with both aerated and anoxic (absence of dissolved oxygen) compartments is often used, see Figure 1.3.

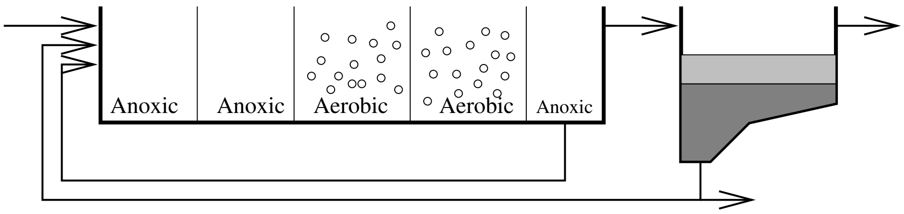

Figure 1.3: An activated sludge process configured for nitrogen removal (predenitrification). The last anoxic compartment serves the purpose to achieve a low oxygen concentration in the internally recirculated water.

Most of the nitrogen in the influent water is in the form of ammonium (NH_4^+). In the aerated compartments, the ammonium may be converted into nitrate (NO_3^-). This process is called nitrification. For the nitrification process to work well, several environmental factors need to be fulfilled including a sufficiently high concentration of dissolved oxygen (DO) and a sufficiently long sludge retention time (the average time a particle stays in the system). A simplified picture of the nitrification is that two different groups of autotrophic bacteria oxidize ammonium to nitrate with nitrite as an intermediate product:

$$NH_4^+ + 1.5O_2 \Rightarrow NO_2^- + H_2O + 2H^+$$
$$NO_2^- + 0.5O_2^+ \Rightarrow NO_3^-$$

In the anoxic compartments, where no dissolved oxygen should occur, nitrate may be converted to gaseous nitrogen (N_2) by a process called denitrification. In the conversion, heterotrophic microorganisms use the nitrate produced by the nitrification process to oxidize organic and inorganic compounds. The reduction of nitrate leads to an emission of gaseous nitrogen to the atmosphere. This can be described by the simplified formula:

$$2NO_3^- + 2H^+ \Rightarrow N_2(g) + H_2O + 2.5O_2$$

Since the growth rate of heterotrophic microorganisms is higher than for the autotrophic microorganisms, the denitrification process is not fully as dependent on a long sludge retention time as the nitrification process is. Instead it relies on sufficient amounts of readily biodegradable substrate and low DO levels in order to convert nitrate into gaseous nitrogen.

In reality, both the nitrification and the denitrification processes are far more involved than described above, see for instance Hallin (1998) for a more detailed treatment of the subject.

There are many different ways to configure the ASP for nitrogen removal. The two most common concepts are to use either a post-denitrifying or a pre-denitrifying plant structure. In a post-denitrifying plant, the aerated compartments are placed before the anoxic. Since the readily biodegradable substrate in the influent water here will be oxidized in the aerated compartments, carbon addition from an external carbon source is often necessary in the anoxic part. In a pre-denitrifying plant, the aerated compartments are placed after the anoxic. The advantage with this structure is that available readily biodegradable substrate in the influent water can be used for denitrification. A drawback is, however, that water rich in nitrate has to be recirculated from the aerated compartments to the anoxic where denitrification can take place. For a deeper discussion about the mechanisms behind the biological processes in ASP's, see also Henze *et al.* (1995) and Metcalf and Eddy Inc. (1991).

1.3 Modelling and Control of Wastewater Treatment Plants

1.3.1 Mathematical Models

There are a number of different models for the ASP. The most commonly used today is the IAWQ activated sludge model No. 1 (ASM1), see Henze *et al.* (1987). This model describes removal of organic matter, nitrification and denitrification. A more recent model, that also deals with biological phosphorus removal is ASM2, see Gujer *et al.* (1995). A model constructed to deal with some defects in the ASM1 is ASM3, see Gujer *et al.* (1998).

In order to provide some basic understanding of bioreactor modelling, a simple example will here be examined. Consider a completely mixed bioreactor with one substrate, S, and one biomass, X, according to Figure 1.4.

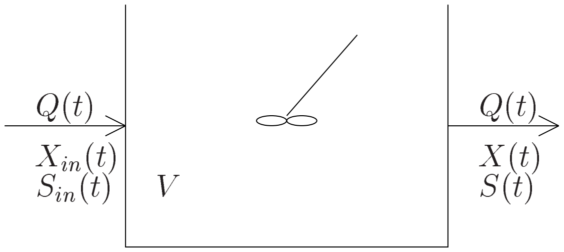

Figure 1.4: A completely mixed bioreactor. The reactor has volume V and the influent flow rate Q is equal to the effluent flow rate. The concentrations of substrate and biomass are denoted S and X, respectively.

For the concentration of one general process component, Z, a simple mass balance yields the following dynamical model, (Bastin and Dochain (1990)):

$$\frac{dZ(t)}{dt} = R_Z(t) + \frac{Q(t)}{V}(Z_{in}(t) - Z(t)) \tag{1.1}$$

where Q is the flow rate, V is the reactor volume, Z_{in} is the concentration of the process component in the influent water and R_Z is the reaction rate describing biological and chemical processes.

The reaction rate, $R_Z(t)$, can be modelled in different ways. A common choice is to use Monod functions to model the specific growth rates (the rate of increase in cell concentration per unit cell concentration). The Monod function describes how the specific growth rate depends on the concentration of the different substrates. For the simple one substrate process in Figure 1.4 the specific growth rate according to the Monod function is

$$\mu(S) = \mu_{max} \frac{S(t)}{K_S + S(t)} \tag{1.2}$$

where μ_{max} is the maximum specific growth rate and K_S is a half saturation constant. It is easily seen from (1.2) that the specific growth rate is zero if the substrate concentration is zero and that it is approximately μ_{max} for large substrate concentrations.

When the specific growth rate is given by (1.2), the total reaction rate for the biomass component X is given by:

$$R_X(t) = \mu(S)X(t) \tag{1.3}$$

Using the reaction rate in (1.3) together with a yield factor describing the substrate consumption by the biomass, and the basic structure (1.1) will now result in the differential equations:

$$\frac{dX(t)}{dt} = \mu(S)X(t) + \frac{Q(t)}{V}(X_{in}(t) - X(t)) \tag{1.4}$$

$$\frac{dS(t)}{dt} = -\frac{1}{Y}\mu(S)X(t) + \frac{Q(t)}{V}(S_{in}(t) - S(t)) \tag{1.5}$$

where Y is the yield factor.

As mentioned, there are many other possible ways to model the specific growth rates. The Haldane law, for instance, takes substrate inhibition into account. An overview of different specific growth rate models is given in Bastin and Dochain (1990).

In the ASM1, the number of components is larger and the reaction rates are far more complicated than in the simple example given in (1.4)-(1.5). The principle is, however, the same, and the specific growth rates are modelled as coupled Monod functions. ASM1 also contains other processes than pure biomass growth, such as microbial decay, hydrolysis and ammonification. A description of the ASM1 model is given in Appendix A.

Due to its complexity, ASM1 is hard to calibrate and to use directly for control purposes. Therefore, many reduced models and methods for model simplification have been suggested. One of the methods used for reducing nonlinear models is the singular perturbation technique. This method applied on bioreactor models is discussed by Bastin and Dochain (1990) and Weijers (2000). Another method is the direct use of simplifying assumptions or process knowledge that reduces the model. In Jeppsson (1996) and Vanrolleghem (1994) such simplified model structures are proposed. Grey-box modelling of Monod kinetics is dealt with in Carstensen *et al.* (1995). In Lukasse (1999) identification schemes for simplified models are discussed. A review of respirometric methods for model calibration is given in Vanrolleghem *et al.* (1999), and also in Petersen *et al.* (2000). Besides the methods based on physical modelling, it is also possible to model the ASP using black-box methods. Lindberg (1998) presents a multivariable model describing nitrification and denitrification. The major drawback with these methods is that the model parameters will generally not have a physical interpretation. The black-box identification problem in general is described in Söderström and Stoica (1989).

1.3.2 Simulators

With the development of faster computers and more accurate models, the possibility to efficiently simulate activated sludge processes has increased in the last years. A simulation is an imitation of the behaviour of the real system. If a sufficiently good model is used and a careful calibration of that model has been done, it might from simulations be possible to predict the behaviour of a real plant. It should however be emphasized that any conclusions about the real process based on simulation results should be drawn with caution.

Besides forecasting, a simulator can be used for other purposes, for example:

- Education and training.
- To develop process understanding.
- Optimization.
- Design and evaluation of different automatic control strategies.

Simulators can be constructed in many different ways, using different models. The result of the simulation reflects the choice of model and the choice of parameters in that model. Often simulators are equipped with a graphical user interface (GUI). In a GUI the simulated model is graphically represented, commonly together with data presentation. Usage of GUI's has become more common in the past years, and the GUI's have become more technically advanced. Of course, from the user's point of view, the layout and construction of the GUI is of great importance. A well constructed and pedagogical GUI could help to increase process understanding, while a badly constructed could be of more harm than benefit.

Today there are many simulators available for simulating the ASP, both commercial and non commercial ones. A few examples of commercial simulators are

EFOR (Aug 13, 2001), GPS (Aug 13, 2001), SIMBA (Aug 13, 2001) and WEST (Aug 13, 2001). Simulation and different simulation softwares for WWTP's are more thoroughly treated in Olsson and Newell (1999).

1.3.3 Control of Wastewater Treatment Plants

Few sciences have so many application areas as automatic control. Automatic control is used in many industrial processes, among which wastewater treatment is one. To control a process, it must be possible to affect it with some control signal. In general, the aim of the control is to keep some state of the process at a desired value, despite disturbances. A very basic concept is to use feedback from measurements or estimates of the state for the control. The feedback is used to calculate a suitable control signal. One may also use feedforward information from different process disturbances in the control law, which may possibly improve the disturbance attenuation. Nowadays, most controllers are implemented in computers, and advanced algorithms may thus be used for computing the control signal. General sources for information on the topic of automatic control theory are, among many others, Åström and Wittenmark (1989), Åström and Wittenmark (1990), Skogestad and Postlethwaite (1996), and Goodwin *et al.* (2001).

Traditionally there has been little interest in using advanced control strategies in wastewater treatment. There are several possible explanations for this, see Olsson (1993). For instance, since treating wastewater has been considered to be a non productive process, it could be hard to prove and point out that long term profits could be gained by introducing an advanced control system. The introduction of the control system, in its turn, is often associated with short term investment costs that are readily visible. Another part of the problem has been the equipment at the plants. Unreliable sensors and bad actuators have in many cases made attempts towards advanced control meaningless. A historical perspective on automatic control in Swedish WWTP's can be found in Olsson *et al.* (1998).

However, development towards tighter effluent demands and better equipment has contributed to an increasing interest for applying advanced control strategies for WWTP's. In Jeppsson *et al.* (2001), it is stated that the nations having the highest levels of instrumentation and automatic control also have been the most successful fulfilling the stricter effluent demands imposed by the European Community. Bastin and Dochain (1990), Lindberg (1997), Lukasse (1999), Olsson and Newell (1999) and Weijers (2000) propose several model based automatic control strategies. Ayesa *et al.* (1998) considers selection of operational strategies based on optimization. See also Rivas *et al.* (2001) and Galarza *et al.* (2001*b*) for a case study. Supervisory plant control and monitoring is considered in Rosen and Yuan (2001). This is also the theme in Roda *et al.* (2001).

Some general incentives for using advanced control strategies in WWTP's are (see Olsson and Newell (1999)):

- To make it possible to meet stricter effluent demands.
- To achieve cost efficient solutions, *i.e.* to optimize the operation in order to minimize operating costs.
- To dampen the impact of process disturbances, since the load to a WWTP generally shows large variations.
- Today, many WWTP's have very complex processes for which manual control is difficult.

A first step in the design of an automatic control strategy is to determine which signals to control, which signals to measure or estimate, and which control signals to use. In Olsson and Jeppson (1994), cause-effect relationships for the ASP are established. In a WWTP, there are many variables that can be used for affecting the nitrogen removal. Only a few of the possible control strategies will be outlined here, and the discussion will be concentrated on variables affecting the nitrogen removal.

- Control of the DO concentration. As mentioned earlier, in order for nitrification bacteria to convert ammonium to nitrate, oxygen is needed. DO control has been practiced for many years, and several variants exists. See for example Olsson *et al.* (1985) for cascaded DO control and Lindberg and Carlsson (1996*b*) for nonlinear control of the DO. An approach using auto tuning is found in Carlsson *et al.* (1994).
- Instead of using a fixed DO set-point, the DO set-point can be manipulated in order to control the ammonium concentration, see among others Nielsen and Lyngaard-Jensen (1993), Lindberg and Carlsson (1996*b*), Lindberg (1998), Olsson and Newell (1999), Waltin (2000), Suescun *et al.* (2001), Serralta *et al.* (2001), Galarza *et al.* (2001*a*) and Ingildsen *et al.* (2001).
- The aeration volume can be used to control the nitrification. A strategy for this is presented in Chapter 4, see also Samuelsson and Carlsson (2001*a*), Hoen *et al.* (1995), Brouwer *et al.* (1998) and Krause *et al.* (2001).
- When there is a lack of readily biodegradable substrate in the influent water, or when a post-denitrifying plant structure is used, addition of an external carbon source is commonly used to control the denitrification. In Chapter 3 and in Samuelsson and Carlsson (2001*b*) such a control strategy

is presented. Examples of carbon dosage control can also be found in Hellström and Bosander (1990), Isaacs *et al.* (1993), Hallin *et al.* (1996), Yuan *et al.* (1997), Lindberg and Carlsson (1996*a*), Barros and Carlsson (1997), de Arruda and Barros (2001) and Milocco and Carlsson (2001).

- Control of hydraulic variables
  - Control of the internal recirculation flow rate. In Rehnström and Carlsson (2001) and Ekman *et al.* (2001) a control strategy that combines addition of an external carbon source with control of the internal recirculation flow rate is designed and evaluated. The study in Rehnström and Carlsson (2001) is performed on the COST benchmark, see Alex *et al.* (1999). Control of the internal recirculation flow rate is also explored in Suescun *et al.* (2001), Vrecko *et al.* (2001) and Yuan *et al.* (2001*a*).
  - The influent flow rate. By using storage volumes, it is possible to equalize the variations in influent flow rate. Thus an important part of the process disturbances are reduced. See for instance Nyberg *et al.* (1996) for an example of this strategy.
  - The sludge recycling flow rate. Simulations on the COST benchmark suggest that control of the sludge blanket height could improve the nitrate removal, see Yuan *et al.* (2001*b*).
  - The excess sludge flow rate can be used to control the sludge age. See Vaccari and Christodoulatos (1989) for a discussion.
- It is also common to control the addition of precipitation chemicals used for phosphorus removal, see Hellström *et al.* (1984) or Devisscher *et al.* (2001) for a more recent contribution. Control of biological phosphorus removal can be found in Munk-Nielsen *et al.* (2001).

With the development of modern control technologies, it has also become possible to use multivariable approaches and control a number of outputs using a number of control signals. In Lindberg (1997) the DO set-point, the external carbon flow rate and the internal recirculation flow rate are used for simultaneous control of the nitrification and denitrification. García-Sanz and Ostolaza (1998) use multivariable QFT-techniques to control the ASP, and the internal recirculation flow rate and DO set-point are here used to control the nitrate and ammonium levels. Another multivariable approach is found in Weijers *et al.* (1997).

1.4 Outline of the Thesis

This section presents a brief summary of the three remaining chapters in the thesis.

In Chapter 2, a JAVA based simulator of the activated sludge process is presented. The incitement for the development of the simulator was to construct a user friendly simulation platform. A guideline has been that both plant personnel and academic undergraduate students should be able to use the simulator to increase their process knowledge. The simulator is also used as a tool to illustrate the importance of automatic control strategies. The implemented simulation model is the ASM1 with ten completely mixed compartments. Also a settler is implemented, using an one-dimensional layer model. The GUI is another important part of the simulator. From the GUI, it is possible to change different parts of the simulation setup on-line. Simulation results are also presented on line in the GUI. From practical experiences of laboratory work, the simulator is believed to be user friendly, and therefore most useful for educational purposes. Since the simulator can be loaded via Internet it is also suitable for distance education. The possibility to run the simulator via Internet using a standard browser makes the simulator very accessible.

A feedforward controller for the external carbon flow rate is presented in Chapter 3. The controller is based on steady state analysis of a simplified ASM1 model. The main control objective is to keep a low constant nitrate concentration in the last anoxic compartment despite variations in influent load. Although the control law is simple, load variations are quickly attenuated compared to when ordinary PID feedback control is used. In the paper it is suggested how to choose a feasible nitrate set-point. A strategy for estimating the heterotrophic yield is also developed.

Chapter 4 presents a model based feedforward strategy for controlling the aeration volume in an ASP. The controlled output is here the ammonium concentration in the last aerated compartment. The control objective is to reject process disturbances quickly, without needing excessively high DO levels. Exact linearization is used in the controller derivation. It is discussed how to implement the control strategy in practice, and a suggestion of how to discretize the control law with respect to volume is made. The resulting control law combines volume control with supervisory control of the DO set-points. Simulation results seem promising.

Chapter 2

A JAVA Based Simulator of Activated Sludge Processes

Modelling and simulation studies are becoming more and more common in the wastewater treatment area. Typical application areas include process design, controller design, optimization, forecasting, and research. Several simulators and models of the activated sludge process have been developed during the last decade, see (Olsson and Newell 1999). It is, however, fair to conclude that most simulators have not been used very much by plant operators. There are several reasons for this, but one major cause is that the simulators so far have not been very user friendly. We will here present a simulator for the activated sludge process, which is believed to be more user friendly than most existing simulators. It also has some other advantages to be discussed below.

1. The simulator is written in the programming language JAVA and can hence be run via Internet using an ordinary browser such as Netscape Communicator or Microsoft Explorer.

2. The simulator is easy to understand and run. For example, to change a flow rate, the user needs just to click on the corresponding pump symbol, and enter the desired flow rate in a dialog box.

3. A novel graphical user interface (GUI) is used to present process information in a comprehensive way.

4. Special concern has been devoted to implement automatic control strategies including DO control, supervisory DO control, external carbon flow rate control and control of the internal recirculation flow rate.

Items 1-3 are believed to make the simulator ideal for educational and training purposes. Item 4 may be a tool to promote the benefits of using automatic control in WWTP's. With the increased use of the Internet, the simulator may be used also for distance education.

The chapter is organized as follows: In Section 2.1 the underlying model and the implementation of the JASS simulator are discussed. The implemented control strategies are shortly presented in Section 2.2. In Section 2.3 experiences of JASS as a tool for teaching and training are reviewed.

2.1 The Model and the Implementation

The underlying process model used in the simulator is the IAWQ Activated Sludge Model no 1 (ASM1). In Appendix A the ASM1 model is outlined, see also Henze *et al.* (1987) and Jeppsson (1996) for a general model description. Three different processes are described by the model, removal of organic matter, nitrification and denitrification. The model contains 13 different components, the behaviour of each component is described by a nonlinear differential equation. ASM1 is probably the most used advanced model for the activated sludge process. When simulating the activated sludge process in a WWTP it is also necessary to model a settler since biomass must be settled and recirculated back into the process. The settler unit is modeled as a traditional one-dimensional layer model. The settler is an important part of the activated sludge process. It is used for two purposes: clarification and thickening. In a settler, particulate matter in the water from the bioreactor sink to the bottom of the settler (thickening), and clear water is produced (clarification) and removed in the top of the settler. The sludge is recirculated to the bioreactor to maintain a desired solids level. There exist many different settler models in the literature. Grijspeerdt *et al.* (1995) performed a comparative study where it was found that the model by Takács *et al.* (1991) gave the most reliable results. This model is used in the simulator. Some information about the used settler model is given in Appendix A, see also Takács *et al.* (1991) for details.

In the default version of the simulator, an activated sludge process with post-denitrification is implemented. The implemented process consists of a basin divided into 10 completely mixed compartments, each modeled by ASM1, and a settler. This structure is depicted in Figure 2.1. The number of compartments is fixed. However, the settings are possible to change in order to simulate *e.g.* a pre-denitrifying plant. The settler is modeled with 10 vertical layers. All differential equations are solved with a fourth order Runge-Kutta method with time step 0.01 h. Since the simulated activated sludge process is divided into compartments, the use of an object oriented language, such as JAVA is natural. Each compartment is implemented as an instance of a class, which makes the structure of the program simple. The settler is an instance of another class. Another advantage with JAVA is its support for multithreading and graphics.

Two threads are implemented in the program, one for handling events and one for running the simulation.

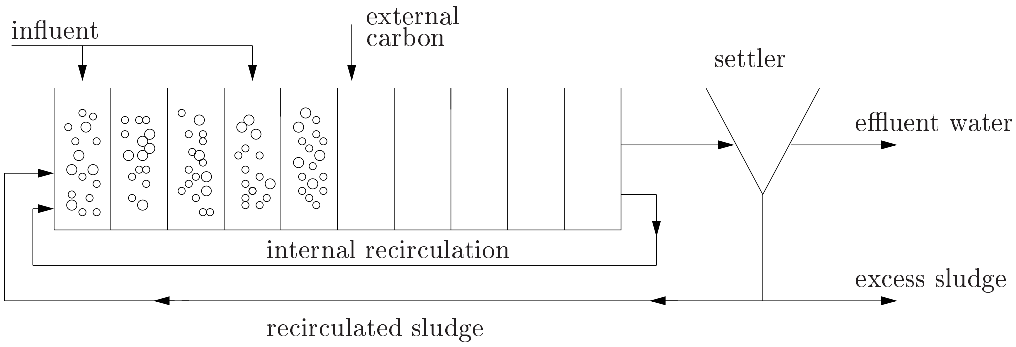

Figure 2.1: Schematic layout of the activated sludge process in the simulator.

The simulator is operated from a GUI where also the status of the plant is shown in real time, see Figure 2.2. In the lower part of the GUI, the layout of the simulated plant is shown. The aim of this picture is to illustrate some fundamental settings of the model, for instance aerated compartments and active pumps. The picture also contains clickable fields. When such a text field is activated, new windows, either used for presenting data or changing process parameters, appear. To increase clarity, the fields are located as near the corresponding part of the process picture as possible. For instance, text fields containing information connected to the settler are displayed close to the settler.

To improve the presentation of simulated process parameters, three bar diagrams are used. Each bar diagram displays the concentration of a selectable component in influent water, all compartments, the recirculated sludge and the effluent water. Numerical values of the concentrations are also presented. In the GUI an ordinary graph is also used. The concentrations of two components in a selectable compartment can be plotted versus time. Most of the functions in the simulator are very easy to learn, see further Appendix A for a short user manual. For more detailed information on the program implementation see Samuelsson (1998).

2.2 Control Strategies

In the current version of the simulator, four control strategies are implemented:

- DO control. The air flow rate is controlled so that the DO concentration is kept close to some set-point. A standard PID controller which can be

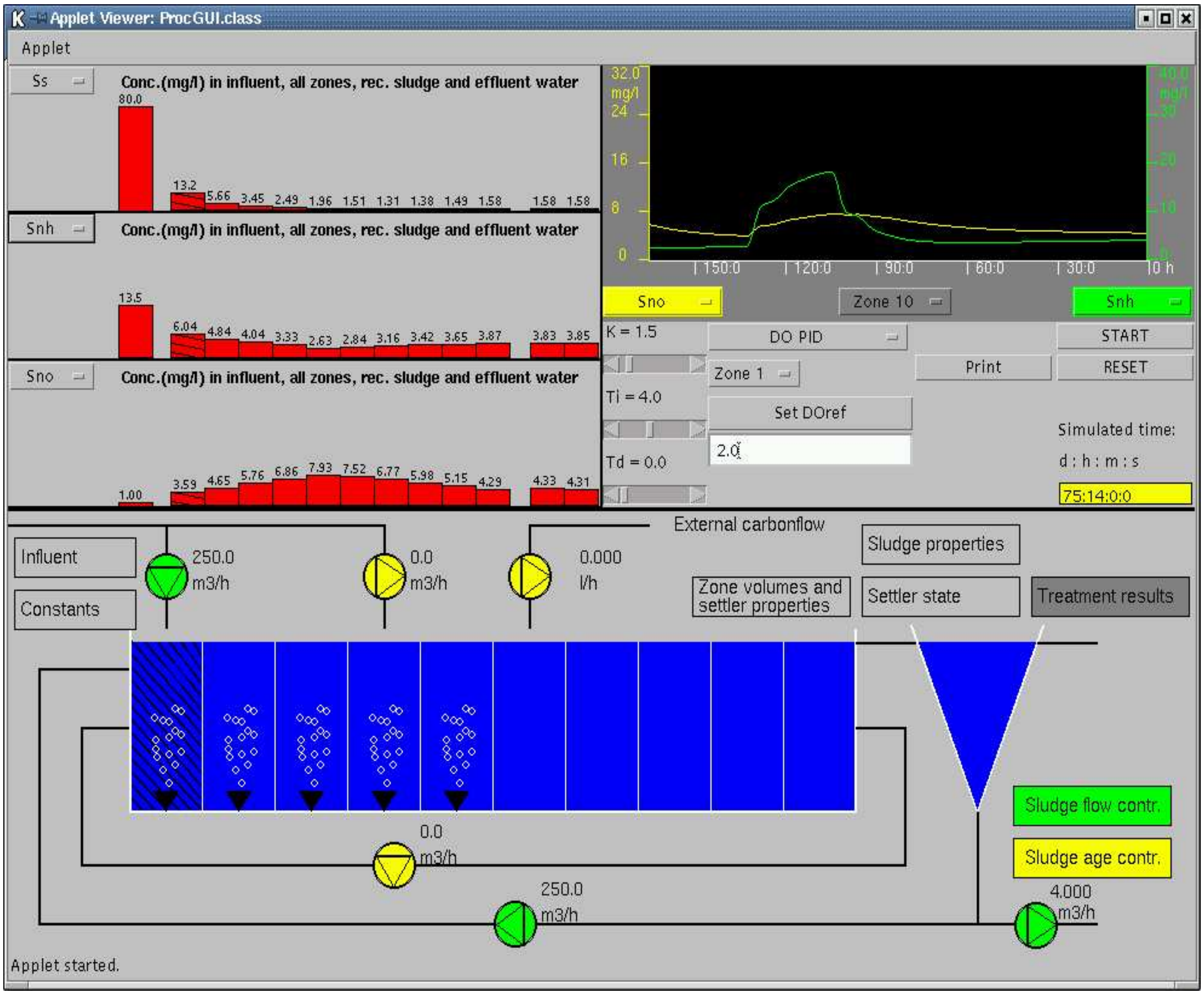

Figure 2.2: The graphical user interface in the simulator.

tuned from the GUI is used. This DO control strategy is standard, and not further discussed here.

- Supervisory DO control. Given a desired ammonium level and a measurement of the ammonium concentration in the last aerated compartment, a supervisory controller calculates the set-point for the DO controller above. The controller is a standard PID and can be tuned from the GUI. The control strategy works for both pre- and post-denitrifying configurations. In the GUI it is possible to choose between standard DO control and the supervisory DO control strategy. This controller is further discussed below.

- External carbon flow rate control. The flow rate of external carbon is used to control the nitrate level in the last anoxic compartment. Here, a feedforward strategy based on a simple mass balance is applied together with a feedback controller. Only the nitrate set-point can be changed from the GUI. This controller is discussed in detail and illustrated with simulations in Chapter 3.

- Internal recirculation flow rate control. The nitrate concentration in the last aerobic compartment is controlled by manipulating the internal recirculation flow rate. The controller is derived using simple mass balances, a general description of the control law can be found in Rehnström and Carlsson (2001) and Ekman *et al.* (2001). The controller is also further described below.

2.2.1 Supervisory Control of the DO Set-Points

The nitrification bacteria in the aerobic compartments require a sufficient high concentration of DO in order to obtain a satisfactory nitrification. On the other hand an excessively high DO level will lead to an unnecessary high energy consumption. In the case when a pre-denitrifying structure is used on the plant, a too high DO level in the internally recirculated water may also inhibit the denitrification, see Olsson and Jeppson (1994).

Presently, the most common DO control strategy is to use a fixed DO set-point. The goal of the DO control is to keep a constant DO level by varying the airflow rate. An interesting alternative to a constant DO set-point is to control the DO set-point from on-line measurements of the ammonium concentration, see Lindberg and Carlsson (1996*b*) and Lindberg (1997), letting the ammonium concentration in the last aerated compartment determine a time-varying DO set-point. This offers the possibility to design a supervisory DO controller, also known as cascade control, see Figure 2.3. The supervisory DO controller aims to maintain an ammonium concentration set-point in the last aerated compartment by varying the DO set-point. If the ammonium concentration is chosen properly, the controller may give a lower average DO set-point, compared to when a constant DO set-point is used, which saves energy. Other advantages may be better control of effluent ammonium, lower nitrate concentration in the effluent because of improved denitrification (due to a lower average DO) and a decreased need of external carbon, Lindberg (1997).

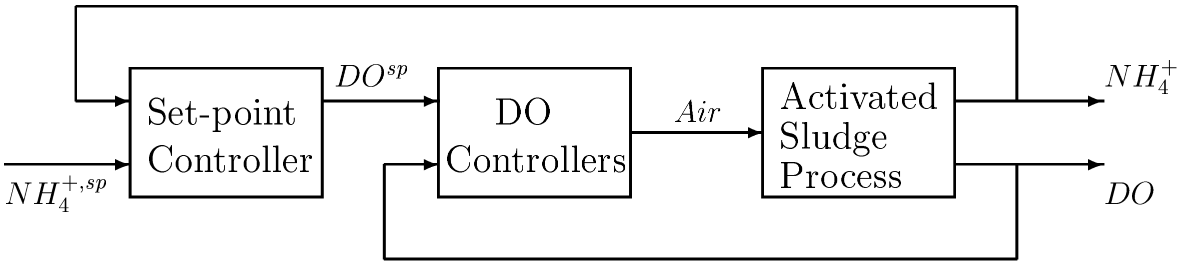

Figure 2.3: A block diagram for the DO set-point controller. The time-varying DO set-point is determined by the ammonium concentration in the last aerobic compartment. The superscript *sp* denotes the set-point. The same DO set-point is fed to all DO controllers.

In the simulator, a supervisory DO controller has been successfully implemented. The PID-parameters have been manually tuned and can be changed from the GUI. The ammonium concentration is measured in the last aerated compartment and the controller works for post-denitrification as well as for pre-denitrification. For information on how to run the supervisory DO controller, see Appendix A. Some simulation results can be found in Samuelsson *et al.* (2001).

2.2.2 Control of the Internal Recirculation Flow Rate

In a pre-denitrifying system the denitrification process takes part in the first anoxic compartments and the nitrification in the last aerobic compartments. The nitrogen rich water in the last aerobic compartment is consequently necessarily recirculated back to the first anoxic compartment. A reasonable control strategy for controlling the nitrate concentration in the last aerobic compartment is to adjust the internal recirculation flow rate so that the nitrate level is kept close to a given set-point. The performance of the internal recirculation flow rate controller is dependent on the denitrification in the anoxic compartments and thus dependent on sufficient amounts of readily biodegradable substrate. It is therefore necessary to complete the control strategy with addition of an external carbon source. The impact of the internal recirculation flow rate controller is more significant if the nitrate concentration in the last anoxic compartment is kept at a low value. The internal recirculation flow rate controller is schematically illustrated in Figure 2.4. For the full mathematical description and simulation results, see Ekman *et al.* (2001).

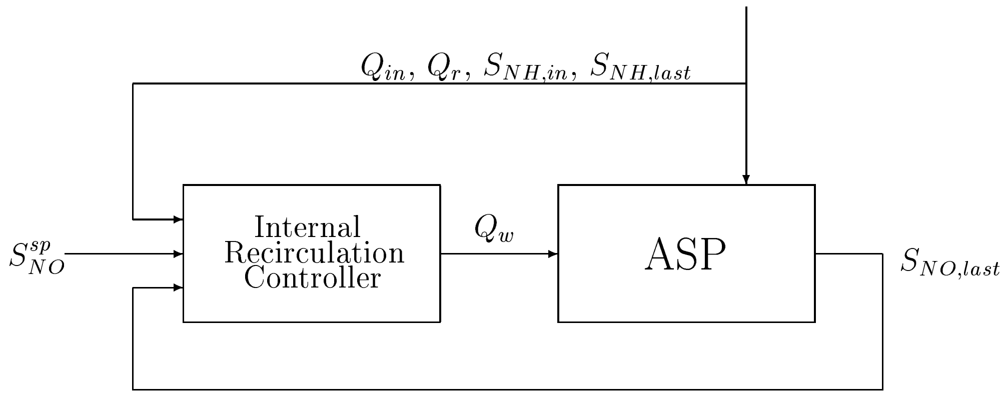

Figure 2.4: A block diagram over the internal recirculation flow rate controller. The controller uses feedforward from influent flow rate Q_{in}, sludge recirculation flow rate Q_r, influent ammonium $S_{NH,in}$ and ammonium in last aerated compartment $S_{NH,last}$. Feedback from the nitrate concentration in the last aerated compartment ($S_{NO,last}$) is also used together with a nitrate set-point (S_{NO}^{sp}). The internal recirculation flow rate is denoted Q_w.

2.3 JASS as a Teaching and Training Tool

Since 1998 we have used the JASS simulator as an integrated teaching tool for undergraduate students and WWTP personnel. Currently, around 150 students have been using the simulator in our course "Wastewater treatment" given in the Aquatic and Environmental MSc program at Uppsala University. Further, around 80 participants from Swedish WWTP's have been using the simulator in a 3 day course entitled "Automatic Control in WWTP's". This course has currently been given four times. The simulator has also been used by some other universities, and special editions of the simulator have been developed for three WWTP's. The simulator is run via Internet with a standard browser. This feature is especially appealing for distance courses, which is an interesting topic for further consideration. It can be concluded that the simulator has been very appreciated by students and plant personnel. Below we give some experiences from the courses.

The simulator is mainly used for illustrating various aspects of nitrogen removal in the ASP. Typical issues that are studied are advantages and disadvantages with pre- and post-denitrification processes. For example, the need for an external carbon source in the latter case. Other aspects which are illustrated is the risk for wash-out if a too low sludge age is chosen, and the risk for sludge overflow in the settler if a too high sludge age is selected. Another use of the simulator is to illustrate and motivate various automatic control strategies.

A problem when the simulator is used for students is to find suitable pen and paper problems for the students to do as preparation exercise and during the simulation session. It is important to have instructions which include some theoretical exercises and verifications since this reduces the risk that the students just push buttons and passively write down the simulation results without any reflections. Also it is important to emphasis that the simulator uses a very idealistic model of an ASP. For example, the well known problem with malfunctioning sensors are not encountered in the simulator. Further, the difficult problem to calibrate a model like the ASM1 to a specific plant must not be underestimated.

When using the simulator as a training and teaching tool for personnel from WWTP's it is important to recognize that their background and skills differs a lot. Also, the experience of using computers may be limited for some participants. In the course, morning sessions are devoted to theory, including automatic control and microbiology. In the afternoon sessions, the simulator is used to further illustrate various aspects of the activated sludge process. In order for the simulation session to be appreciated by all types of plant personnel it is important to have quite simple (but meaningful) laboratory instructions. It can also be good to have non compulsory exercises where more advanced material is covered. We usually start the next session after a simulation session by going through the lab exercises, which also serves as a repetition. We have found that the user interface used in the JASS simulator has been easy to use and interpret by plant personnel. A big advantage is that users, almost immediately can start to simulate and interpret the results. On the other hand, many operators want to start to simulate their "own" plant. A drawback with the JASS simulator is that it is not possible to make big changes in the basic layout without programming.

2.4 Conclusions

A JAVA based activated sludge process simulator has been presented. Special emphasis has been devoted to create a user friendly platform. Some control strategies have been implemented. The simulator can be run with traditional DO level control as well as with supervisory DO control. A simple but effective control strategy for the external carbon flow rate is used in the simulator. Also, an internal recirculation flow rate controller has been implemented. The simulator has been used for academic education and also in courses for personnel from wastewater treatment plants. The simulator is then used as an integrated teaching tool to illustrate different process alternatives in nitrogen removal using the activated sludge process.

Chapter 3

Feedforward Control of the External Carbon Flow Rate in an Activated Sludge Process

The interest to use the ASP for biological nitrogen removal has increased drastically during the last 5-10 years due to stricter effluent standards. To remove the nitrogen from the wastewater in an ASP, two biological processes, nitrification and denitrification, are needed. In aerobic compartments, ammonium may be converted into nitrate (nitrification) and in anoxic compartments nitrate may be converted into gaseous nitrogen (denitrification). The denitrification process is dependent on sufficient amounts of readily metabolized carbon. In a post-denitrification system, an external carbon source is normally needed and in a pre-denitrification system, an external carbon source may be needed if the influent water has a too low carbon/nitrogen ratio.

It is possible to use several different external carbon sources. Common choices are for instance acetate, methanol, ethanol, industrial waste products and primary sludge. Different carbon sources have different characteristics and may give different denitrification rates, see Isaacs and Henze (1995) and Hallin *et al.* (1996). When external carbon is first added to the process, it may take several weeks for the microorganisms to adapt to it. The use of a specific carbon source benefits certain microorganisms which become very dependent of the presence of that particular carbon source. For the very sensitive microorganisms to be sustained, once in use, some external carbon sources should not be turned off for any longer periods, see Hallin *et al.* (1996).

An important problem when an external carbon source is needed, is how much carbon should be added. Too little external carbon gives incomplete denitrification, while too much carbon increases the sludge production, may give carbon spill in effluent, and gives an unnecessary increase of the operating costs, see for instance Lindberg (1997). Note also that a constant flow rate of external carbon may not be feasible since the load to a typical plant may vary a factor of six or more, see Van Impe *et al.* (1992).

Several strategies for controlling an external carbon source have recently been suggested, a natural strategy is to control the flow rate of external carbon so that the concentration of nitrate in the last anoxic compartment is kept at a (low) pre-specified level. In Lindberg and Carlsson (1996*a*), an adaptive controller was suggested and evaluated, see also Olsson and Newell (1999). Barros and Carlsson (1997) considered an iterative pole placement design. Another approach based on exact linearization is developed in Milocco and Carlsson (2001). An interesting contribution related to the work in this thesis is given in Yuan *et al.* (1997). That paper considers a model based on the IAWQ Activated sludge Model No. 1 (ASM1), using different specific growth rates for external carbon and readily biodegradable substrate. The paper of Yuan *et al.* (1997) concentrates on a pre-denitrifying system, and the focus is on optimization of the nitrate set-point in the anoxic compartment given average influent parameters and effluent demands for that particular configuration. The work presented here concentrates on controller derivation for the ordinary ASM1 (with some simplifications), and also the case with several anoxic compartments is considered.

In this chapter a model based feedforward control strategy is developed using a steady state analysis of a simplified ASM1. This strategy is combined with a standard feedback PI controller in order to compensate for unmeasurable disturbances and model simplifications. Since major disturbances are effectively attenuated by the feedforward part, the feedback controller can be slowly tuned. The feedforward part is very simple and requires little a priori knowledge in contrast to more complex control strategies. However, the simple strategy attenuates process disturbances such as changes in the influent flow rate effectively.

The organization of the chapter is as follows: Section 3.1 presents a reduced model for describing the denitrification process in an anoxic compartment. Section 3.2 considers derivation of the controller for one completely mixed reactor. In section 3.3, a general expression for the controller when the reactor consists of several anoxic compartments is derived. A method for estimating the heterotrophic yield is proposed in Section 3.4. Some results for simulations on the ASM1 model are shown in Section 3.5. Finally, results and conclusions are presented in Section 3.6.

3.1 A Reduced Order Model for an Anoxic Compartment

The IAWQ Activated Sludge Model No. 1 is a widely spread model for simulating the activated sludge process. However, the ASM1 model is quite complex and may therefore be hard to use for controller design. In this section, some model simplifications which give a more feasible model for controller design are suggested. In order to describe the denitrification process in an anoxic compartment, at least three components are needed; Heterotrophic biomass (denitrifying bacteria) $X_{B,H}$ [mg COD/l], readily biodegradable substrate S_S [mg COD/l] and the soluble nitrate nitrogen S_{NO} [mg/l]. See Figure 3.1 for a schematic picture of a completely mixed anoxic compartment.

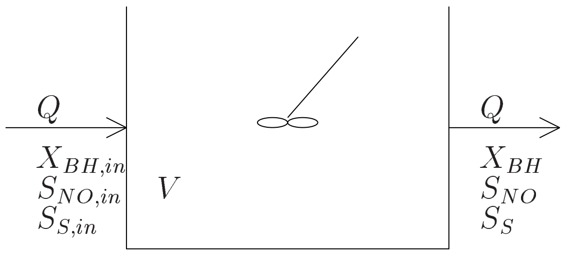

Figure 3.1: A completely mixed anoxic compartment in an activated sludge process. The compartment has volume V, and the influent flow rate Q is equal to the effluent flow rate.

By applying a simple mass balance using the notation in Figure 3.1 the following model structure is obtained (see also Bastin and Dochain (1990))

$$\frac{dX_{B,H}}{dt} = R_{X_{B,H}} + D(X_{B,H,in} - X_{B,H}) \tag{3.1}$$

$$\frac{dS_S}{dt} = R_{S_S} + D(S_{S,in} - S_S) \tag{3.2}$$

$$\frac{dS_{NO}}{dt} = R_{S_{NO}} + D(S_{NO,in} - S_{NO}) \tag{3.3}$$

Here $R_{X_{B,H}}$ denotes the reaction (growth) rate for the component $X_{B,H}$ etc. and $D = \frac{Q}{V}$ is the dilution rate.

In the common case when the influent flow consists of several partial flows, $Q_1,\ \ Q_2,\ \ldots\ Q_n$, the influent concentration of a general component Y_{in} is obtained as a weighted mean (assuming a perfect mix) as:

$$Y_{in} = \frac{1}{Q}\sum_{j=1}^{n} Q_j Y_{in,j}$$

where $Q = \sum_{j=1}^{n} Q_j$, Q_j denotes the j:th inflow to the compartment, and $Y_{in,j}$ the influent concentration corresponding to the flow Q_j.

In the ASM1 model, the reaction rates in (3.1), (3.2) and (3.3) are described as:

$$\begin{aligned}
R_{X_{B,H}} =& \hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{S_O}{K_{O,H} + S_O})X_{B,H} \\
&+ \hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{K_{O,H}}{K_{O,H} + S_O})(\frac{S_{NO}}{K_{NO} + S_{NO}})\eta_g X_{B,H} - b_H X_{B,H} \qquad (3.4)\\
R_{S_S} =& -\frac{1}{Y_H}\hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{S_O}{K_{O,H} + S_O})X_{B,H} \\
&- \frac{1}{Y_H}\hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{K_{O,H}}{K_{O,H} + S_O})(\frac{S_{NO}}{K_{NO} + S_{NO}})\eta_g X_{B,H} \qquad (3.5)\\
&+ k_h \frac{\frac{X_S}{X_{B,H}}}{K_X + \frac{X_S}{X_{B,H}}} \left((\frac{S_O}{K_{O,H} + S_O}) + \eta_h (\frac{K_{O,H}}{K_{O,H} + S_O})(\frac{S_{NO}}{K_{NO} + S_{NO}}) \right) X_{B,H} \\
R_{S_{NO}} =& -\frac{1 - Y_H}{2.86 Y_H}\hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{K_{O,H}}{K_{O,H} + S_O})(\frac{S_{NO}}{K_{NO} + S_{NO}})\eta_g X_{B,H} \qquad (3.6)\\
&+ \frac{1}{Y_A}\hat{\mu}_A (\frac{S_{NH}}{K_{NH} + S_{NH}})(\frac{S_O}{K_{O,A} + S_O})X_{B,A}
\end{aligned}$$

where S_{NH} is soluble ammonium nitrogen, $X_{B,A}$ is the autotrophic biomass, X_S is the slowly biodegradable substrate and S_O is the dissolved oxygen concentration. See Henze *et al.* (1987) for a detailed discussion of the involved parameters in the reaction rates.

In order to reduce the model complexity, three assumptions are made:

A.1 It is assumed that $S_O \approx 0$ and hence $S_O << K_{O,H}$.

A.2 The hydrolysis of entrapped organics is considered a first order reaction with specific reaction rate k.

Assumption A.1 is natural since the compartment is assumed to be anoxic and yields that:

$$\frac{S_O}{K_{O,H} + S_O} \approx 0 \qquad (3.7)$$

$$\frac{K_{O,H}}{K_{O,H} + S_O} \approx 1 \qquad (3.8)$$

Assumption A.2 is made for convenience and may be a realistic choice given the poor knowledge of the hydrolysis process, see Henze *et al.* (1987). Using the assumptions A.1 and A.2 in (3.4)-(3.6) gives the simplified reaction rates:

$$
\begin{aligned}
R_{X_{B,H}} =& \hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{S_{NO}}{K_{NO} + S_{NO}})\eta_g X_{B,H} - b_H X_{B,H} && (3.9)\\
R_{S_S} =& -\frac{1}{Y_H}\hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{S_{NO}}{K_{NO} + S_{NO}})\eta_g X_{B,H} + k X_{B,H} && (3.10)\\
R_{S_{NO}} =& -\frac{1 - Y_H}{2.86 Y_H}\hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{S_{NO}}{K_{NO} + S_{NO}})\eta_g X_{B,H} && (3.11)
\end{aligned}
$$

Here, the parameter $\hat{\mu}_H$ is the maximum heterotrophic growth rate, K_S and K_{NO} are the half saturation constants for S_S and S_{NO} respectively. The factor η_g is a correction factor for anoxic growth of heterotrophic biomass.

Inserting (3.9)-(3.11) in (3.1)-(3.3) gives the model

$$
\begin{aligned}
\frac{dX_{B,H}}{dt} &= \mu(S_S, S_{NO}) X_{B,H} - b_H X_{B,H} + D(X_{B,H,in} - X_{B,H}) && (3.12)\\
\frac{dS_S}{dt} &= -\frac{1}{Y_H}\mu(S_S, S_{NO}) X_{B,H} + k X_{B,H} + D(S_{S,in} - S_S) && (3.13)\\
\frac{dS_{NO}}{dt} &= -\frac{\beta}{Y_H}\mu(S_S, S_{NO}) X_{B,H} + D(S_{NO,in} - S_{NO}) && (3.14)
\end{aligned}
$$

where the factor $\beta = (1 - Y_H)/2.86$ and

$$
\mu(S_S, S_{NO}) = \hat{\mu}_H (\frac{S_S}{K_S + S_S})(\frac{S_{NO}}{K_{NO} + S_{NO}})\eta_g \qquad (3.15)
$$

In order to improve the denitrification, an external carbon source will be used. It is assumed that the microorganisms have adapted to the carbon source so that the following assumption holds:

A.3 The external carbon source is assumed to be readily biodegradable.

For notational simplicity, we now define

$$
u = Q_{car} COD_{car} \qquad (3.16)
$$

where Q_{car} is the flow rate of the external carbon source, and COD_{car} is its COD content. Here, the external carbon source is assumed to have the COD of ethanol, 1200000 mg (COD)/l. With an external carbon source added, using assumption A.3, (3.13) is modified to:

$$
\frac{dS_S}{dt} = -\frac{1}{Y_H}\mu(S_S, S_{NO}) X_{B,H} + k X_{B,H} + D(S_{S,in} - S_S) + \frac{1}{V}u \qquad (3.17)
$$

3.2 Derivation of a Feedforward Controller for a Completely Mixed Compartment

In this section, a feedforward control strategy is presented using a steady state analysis of the model (3.12), (3.14) and (3.17). A related approach can be found in Yuan *et al.* (1997). The objective is to control the flow rate of the external carbon source so that the nitrate concentration stays at a given set-point. A key result is given in the following lemma.

Lemma 1 *Consider the model described by (3.12), (3.14) and (3.17) and assume that Q, $X_{B,H,in}$, $S_{NO,in}$, and $S_{S,in}$ are constant and that the desired value of effluent nitrate (the set-point) S_{NO}^{ref} is chosen so that*

$$S_{NO}^{ref} > -Z + \sqrt{Z^2 + X} \qquad (3.18)$$

$$S_{NO}^{ref} < S_{NO,in} \qquad (3.19)$$

where

$$Z = \frac{\hat{\mu}_H \eta_g S_{NO,in} + \hat{\mu}_H \eta_g \frac{\beta}{Y_H} X_{B,H,in} + (D + b_H)(K_{NO} - S_{NO,in})}{2(D + b_H - \hat{\mu}_H \eta_g)}$$

$$X = \frac{D + b_H}{D + b_H - \hat{\mu}_H \eta_g} K_{NO} S_{NO,in}$$

The following flow rate of external carbon then asymptotically gives $\bar{S}_{NO} = S_{NO}^{ref}$.

$$\bar{u} = Q[\frac{\mu(\bar{S}_S, \bar{S}_{NO}) - kY_H}{\beta\mu(\bar{S}_S, \bar{S}_{NO})}(S_{NO,in} - S_{NO}^{ref}) - (S_{S,in} - \bar{S}_S)] \qquad (3.20)$$

where $^-$ is used to denote steady state (asymptotic) values and

$$\bar{S}_S = \frac{K_S(D + b_H)(S_{NO,in} - S_{NO}^{ref})}{\hat{\mu}_H \frac{S_{NO}^{ref}}{K_{NO} + S_{NO}^{ref}} \eta_g (S_{NO,in} + \frac{\beta}{Y_H} X_{B,H,in} - S_{NO}^{ref}) - (D + b_H)(S_{NO,in} - S_{NO}^{ref})} \qquad (3.21)$$

Proof See Appendix B. ■

The conditions (3.18) and (3.19) assure that $\bar{S}_S$ is positive and finite. Equation (3.21) is relevant for two reasons. Firstly, $\bar{S}_S$ is needed to calculate the desired external carbon flow rate of (3.20), secondly it may indicate in what range S_{NO}^{ref} may be selected.

Figure 3.2 shows $\bar{S}_S$ as a function of S_{NO}^{ref} for the case $K_S = 20$ mg/l, $D = 1.6667$ h^{-1}, $b_H = 0.026$ h^{-1}, $Y_H = 0.57$, $S_{NO,in} = 20$ mg/l, $X_{B,H,in} = 1000$ mg/l, $\hat{\mu}_H = 0.25$ h^{-1} and $\eta_g = 1$. For this case, the bound (3.18) becomes approximately 0.43 mg/l. The plot indicates in what range S_{NO}^{ref} may be selected. A set-point less than say 2 mg/l, corresponds to an undesirably high substrate concentration in the effluent water.

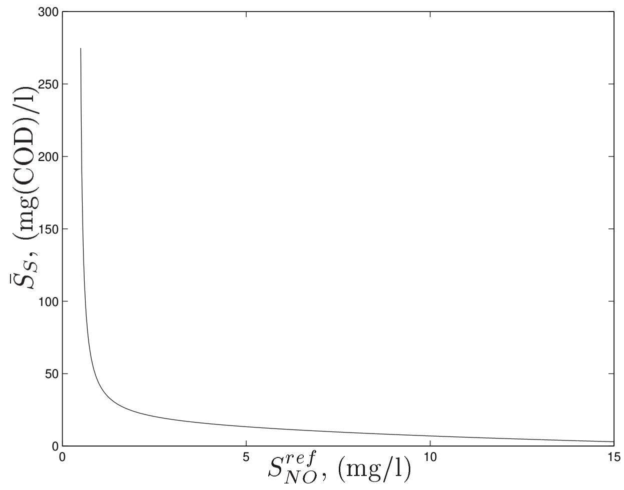

Figure 3.2: Steady state values of S_S as a function of S_{NO}^{ref}

One problem when calculating the flow rate (3.20) is that several process parameters need to be known including the Monod function $\mu(\bar{S}_S, \bar{S}_{NO})$ and the hydrolysis parameter k. A suggestion is therefore to ignore the hydrolysis by setting $k = 0$. Then also $\mu(\bar{S}_S, \bar{S}_{NO})$ disappears from the expression (3.20). Secondly, $\bar{S}_S$ is needed in the calculation. As seen from (3.21), several process parameters are needed for calculating $\bar{S}_S$. However, in practice S_{NO}^{ref} should be chosen so that $\bar{S}_S$ is small. A reasonable approximation may therefore be to set $\bar{S}_S = 0$. By using $k = 0$ and $\bar{S}_S = 0$ in (3.20) the following simple feedforward controller is proposed

$$u(t) = Q(t)[\frac{1}{\beta}(S_{NO,in}(t) - S_{NO}^{ref}(t)) - S_{S,in}(t)] \tag{3.22}$$

where we have included the time argument t in order to indicate that in practice sensor values are used in the control law. This feedforward controller requires measurements of $S_{NO,in}$ and $S_{S,in}$ and knowledge of $\beta = (1 - Y_H)/2.86$. Since several simplifications are made in the derivation of the feedforward controller, a feedback strategy may also be needed. A standard PI controller which gives the following feedforward-feedback strategy is suggested:

$$
\begin{aligned}
u(t) =& K(S_{NO}(t) - S_{NO}^{ref}(t)) + K_I \int_{t_0}^{t} (S_{NO}(\tau) - S_{NO}^{ref}(\tau))d\tau \\
&+ Q(t)[\frac{1}{\beta}(S_{NO,in}(t) - S_{NO}^{ref}(t)) - S_{S,in}(t)]
\end{aligned} \tag{3.23}
$$

The control strategy is also depicted in Figure 3.3.

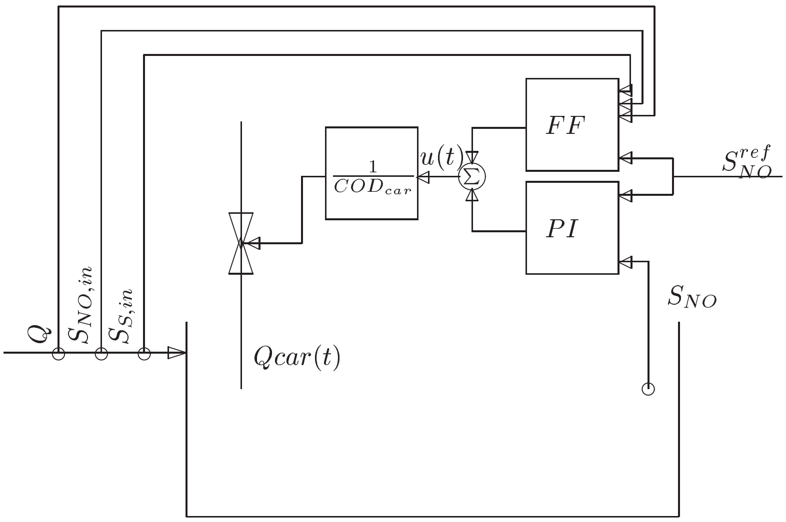

Figure 3.3: The schematic structure of the control law in (3.23).

3.3 Extension to Several Anoxic Compartments

Often, the ASP is configurated with several anoxic compartments. It is therefore of interest to extend the result in Lemma 1 to the multi compartment case. It is here assumed that an external carbon source is added in the first compartment. The nitrate concentration in the last anoxic compartment is to be controlled. Figure 3.4 shows the layout of the process.

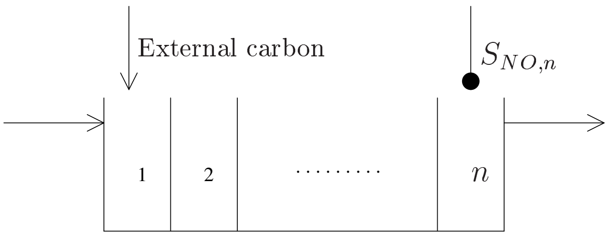

Figure 3.4: Dosage of an external carbon source in a multi compartment anoxic reactor.

The generalization of the control law for the case of n completely mixed compartments is given in the following lemma:

Lemma 2 *Consider n completely mixed anoxic compartments. The concentration of a general component Y in the compartment j, $j = 1, 2...n$, is denoted Y_j. The first compartment is described by (3.12), (3.14) and (3.17) and the following compartments are described by (3.12), (3.13) and (3.14). Assume that Q, $X_{B,H,in}$, $S_{NO,in}$, and $S_{S,in}$ are constant and that the desired value of effluent nitrate (the set-point) S_{NO}^{ref} is chosen so that S_S is finite in the last anoxic compartment. Then the flow rate of external carbon that asymptotically gives $\bar{S}_{NO,n} = S_{NO,n}^{ref}$ is given by:*

$$\begin{aligned} \bar{u} =& Q[\frac{1}{\beta}(S_{NO,in} - S_{NO,n}^{ref}) - (S_{S,in} - \bar{S}_{S,n}) \\ & - \frac{kY_H}{\beta} \sum_{i=1}^{n} \frac{(\bar{S}_{NO,i-1} - \bar{S}_{NO,i})}{\mu(\bar{S}_{S,i}, \bar{S}_{NO,i})}] \end{aligned} \tag{3.24}$$

where $\bar{S}_{NO,0}$ and $\bar{S}_{NO,n}$ correspond to $S_{NO,in}$ and S_{NO}^{ref}, respectively.

Proof See Appendix B. ■

Note that in Lemma 2 it is assumed that the condition (3.18) is fulfilled also in the case of n compartments. It is, however, difficult to give an analytical expression for the readily biodegradable substrate, $\bar{S}_{S,n}$, in the multi compartment case. The expression in (3.24) is quite complex. By setting k=0 (no hydrolysis) and $\bar{S}_{S,n}$=0 (neglecting the substrate in the last compartment) it is obtained:

$$u = Q[\frac{1}{\beta}(S_{NO,in} - S_{NO}^{ref}) - S_{S,in}] \tag{3.25}$$

which is basically the same expression as in the single compartment case, c.f (3.22).

In order to obtain a control law, we parallel the discussion in Section 3.2 and include a time argument in (3.25), c.f (3.22). The controller is also completed with a PI part. This gives:

$$\begin{aligned} u(t) =& K(S_{NO,n}(t) - S_{NO}^{ref}(t)) + K_I \int_{t_0}^{t} (S_{NO,n}(\tau) - S_{NO}^{ref}(\tau))d\tau \\ & + Q(t)[\frac{1}{\beta}(S_{NO,in}(t) - S_{NO}^{ref}(t)) - S_{S,in}(t)] \end{aligned} \tag{3.26}$$

Thus, basically the same controller is obtained in the multi compartment case as in the single compartment case. This controller requires measurements of $S_{NO,in}$,

$S_{NO,n}$ and $S_{S,in}$ and knowledge of $\beta = (1 - Y_H)/2.86$. $S_{NO,in}$ can be measured directly using an on-line sensor. Measurements of $S_{S,in}$ could be obtained from respirometry, see for instance Petersen (2000). Another possibility is to estimate $S_{S,in}$ by approximating it as a fraction of the soluble chemical oxygen demand (COD_S), see Gernaey *et al.* (2001). The factor β could be determined from default values of the heterotrophic yield, Y_H, or from an estimation strategy, see further Section 3.4.

3.4 Estimating the Heterotrophic Yield, Y_H

In order to realize the control law, (3.23) or (3.26), the parameter $\beta = (1 - Y_H)/2.86$ must be known. Here, an estimation procedure for the heterotrophic yield in an anoxic compartment is proposed. The estimator also quantifies the hydrolysis.

Let us define

$$\psi(t) = -\frac{1 - Y_H}{2.86 Y_H}\hat{\mu}_H\left(\frac{S_S}{K_S + S_S}\right)\left(\frac{S_{NO}}{K_{NO} + S_{NO}}\right)\eta_g X_{B,H} \tag{3.27}$$

and further

$$\theta_1 = \frac{2.86}{1 - Y_H} \tag{3.28}$$

Assume that the term in (3.5) describing the hydrolysis could be described as a slowly time varying function denoted $\theta_2(t)$. Following this notation, (3.2) and (3.3) in an anoxic compartment can be written as:

$$\frac{dS_S(t)}{dt} = \theta_1\psi(t) + \theta_2(t) + D(S_{S,in}(t) - S_S(t)) \tag{3.29}$$

$$\frac{dS_{NO}(t)}{dt} = \psi(t) + D(S_{NO,in}(t) - S_{NO}(t)) \tag{3.30}$$

If an external carbon source is added in the compartment in question a term $\frac{1}{V}u(t)$ adds to the right hand side of (3.29) and should be taken into account in the following derivation of the estimator.

Assume further that for an arbitrary anoxic compartment, the nitrate level, $S_{NO}(t)$, the nitrate level in the influent, $S_{NO,in}(t)$, the level of readily biodegradable substrate, $S_S(t)$, and the level of readily biodegradable substrate, $S_{S,in}(t)$ can be properly measured or estimated and that these are sampled or estimated with sampling interval h. Sampling the system and approximating the derivatives in (3.29) and (3.30) with the Euler backward method yields:

$$\frac{S_S(t) - S_S(t-h)}{h} \approx \theta_1 \psi(t) + \theta_2(t) + D(S_{S,in}(t) - S_S(t)) \tag{3.31}$$

$$\frac{S_{NO}(t) - S_{NO}(t-h)}{h} \approx \psi(t) + D(S_{NO,in}(t) - S_{NO}(t)) \tag{3.32}$$

It is now possible to obtain an estimate of $\psi(t)$ directly from (3.32) as:

$$\hat{\psi}(t) \approx \frac{S_{NO}(t) - S_{NO}(t-h)}{h} - D(S_{NO,in}(t) - S_{NO}(t)) \tag{3.33}$$

This approximation of $\psi(t)$ will then after low pass filtration be used directly in the further estimations. To reformulate (3.31) in a more convenient way, set $y(t) = \frac{S_S(t)-S_S(t-h)}{h} - D(S_{S,in}(t) - S_S(t))$. By using (3.33) and (3.31) the $y(t)$ may be approximated as:

$$y(t) = \phi(t)^T \theta(t) \tag{3.34}$$

where $\phi(t) = \begin{bmatrix} \hat{\psi}(t) & 1 \end{bmatrix}^T$ is known and $\theta(t) = \begin{bmatrix} \theta_1 & \theta_2(t) \end{bmatrix}^T$ is the parameter vector to be estimated. From the estimate of the parameter θ_1, the heterotrophic yield can be calculated as $Y_H = (\theta_1 - 2.86)/\theta_1$.

Since the estimation problem includes tracking of time varying parameters, some tracking method has to be used. One possibility is to use a Kalman filter, see Söderström and Stoica (1989).

In this chapter, the parameter vector θ has been estimated on-line via a Kalman filter. To improve the tracking properties, the parameters θ is then parameterised as a random walk process. The full parameterisation of the system then becomes:

$$\theta(t+1) = \theta(t) + v(t) \tag{3.35}$$

$$y(t) = \phi(t)^T \theta(t) + e(t) \tag{3.36}$$

Here, $v(t)$ is a (white) process noise with covariance matrix R_1 and $e(t)$ is a (white) measurement noise with variance R_2. These variances are often seen merely as design variables in the Kalman filter, since the uncertainties of the true system is not known. An increasing R_1 will give a faster tracking, while an increasing R_2 will give the filter more low pass character. With notation according to the above, the Kalman estimator is given by:

$$
\begin{aligned}
\varepsilon(t) &= y(t) - \phi(t)^T\hat{\theta}(t-1) && (3.37)\\
P(t) &= P(t-1) - P(t-1)\phi(t)^T\phi(t)P(t-1)/(R_2 + \phi(t)^T P(t-1)\phi(t)) \\
&\quad + R_1 && (3.38)\\
K(t) &= P(t)\phi(t) && (3.39)\\
\hat{\theta}(t) &= \hat{\theta}(t-1) + K(t)\varepsilon(t) && (3.40)
\end{aligned}
$$

where $\varepsilon(t)$ is the estimation error, $\hat{\theta}(t)$ is the estimate of $\theta(t)$, $P(t)$ is the covariance matrix of the estimate and $K(t)$ is the Kalman gain.

A problem when dealing with parameter identification is that the system must be properly excited in order to obtain sufficiently good estimates. All estimation procedures are dependent of an input signal that excites the output sufficiently for the parameters to be estimated. In particular, when on-line identification methods are used, periods of low excitation may cause estimator windup. This causes the covariance matrix, $P(t)$, to grow fast while the estimation error, $\varepsilon(t)$ is close to zero. When the system is excited again, the estimated parameters will change abruptly. Different methods to avoid estimator windup exist, the main idea is to stop updating the parameters when the absolute value of the estimation error is small and to limit the covariance matrix $P(t)$. More detailed information on estimator windup can be found in Åström and Wittenmark (1989).

Due to that approximations of derivatives are used in the parameter $\psi(t)$ and $y(t)$, the estimator described by (3.37)-(3.40) is very sensitive to high frequency measurement noise. This problem can be partly solved by low-pass filtration of the signals. A way to do this is to use a Butterworth filter with a properly chosen cut-off frequency. Since the estimation is supposed to be done off-line, the parameters may then be filtered forward and reverse in order to avoid phase shift.

The estimation of the heterotrophic yield and hydrolysis is further illustrated by simulations in Section 3.5.

3.5 Simulations

The control strategy (3.26) was evaluated in a simulation study where the full ASM1 was used to model the process. Hence no model simplification was used when simulating the process. A post-denitrifying ASM1 model with five aerobic and five anoxic compartments was implemented, see Figure 3.5. The parameters in the model were the default values reported in Henze *et al.* (1987). In

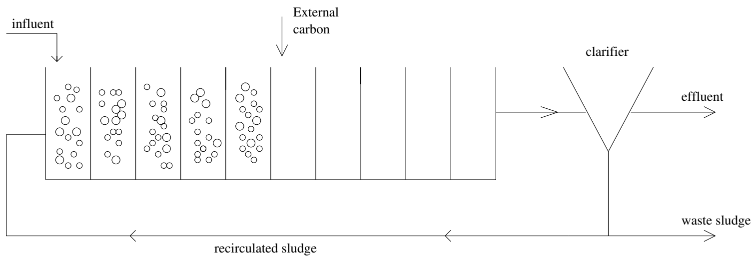

Figure 3.5: Layout of the ASP used in the simulation study.

Appendix B some additional data for the simulation model are given.

3.5.1 Scenario I

The influent disturbances shown in Figure 3.6 were used. After 20 hours, $S^*_{S,in}$ is changed from 70 mg/l to 10 mg/l, after 60 hours the influent flow rate is changed from 3000 m^3/h to 3500 m^3/h, and after 100 hours, $S^*_{NO,in}$ is changed from 1.0 to 7.0 mg/l. The superscript * denotes that these are influent flow concentrations to the plant, and therefore differs from the influent to the controlled anoxic compartments. Finally, after 120 hours, the set-point S^{ref}_{NO} is changed from 2.0 to 1.0 mg/l.

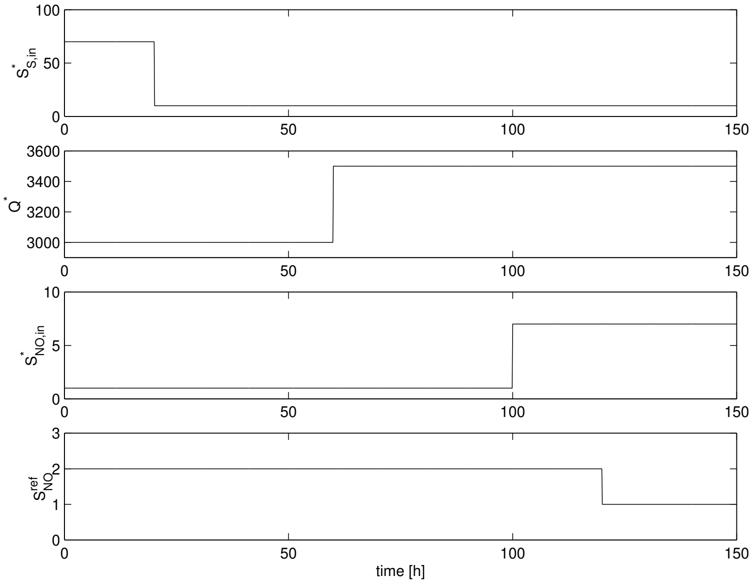

Figure 3.6: Influent variation used in the simulation.

Figure 3.7 shows the uncontrolled process (a constant external carbon flow rate is applied), the process controlled with a PI controller and the process

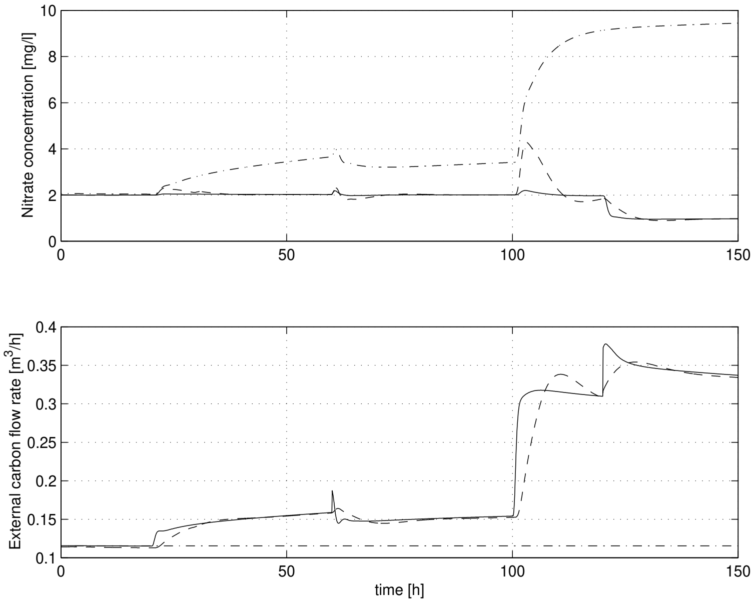

Figure 3.7: Upper plot: The solid line shows the nitrate concentration in the last anoxic compartment when the feedforward-feedback controller is used, dashed line shows the nitrate concentration when the PI controller is used, dashed-dotted line shows the nitrate concentration for a constant flow rate of external carbon. Lower plot: The external carbon flow rates are shown for the three controllers respectively.

controlled with the feedforward-feedback controller (3.26). The response of the uncontrolled process shows the necessity of automatic control. By comparing the PI and feedforward-feedback controller it is seen that the feedforward-feedback controller gives a faster disturbance rejection than the PI controller for all three disturbances. The feedforward-feedback controller also has a faster set-point tracking. It can be argued that a faster PI controller will have a better performance than the PI controller in the figure. However, the same PI parameters have been used both in the feedforward-feedback controller and in the pure PI controller, and this makes the comparison relevant. Note also that a too fast PI controller increases the risk for oscillations and instabilities.

3.5.2 Scenario II

In order to show the controller performance for a more realistic time variation of the disturbances, the influent flow rate and the influent concentrations in Figure 3.8 were used. Other influent concentrations were the same as in Appendix B.

Figure 3.9 shows the simulations when only the PI controller part of (3.26) was used to control the process. Figure 3.10 shows the simulations of the system when the feedforward-feedback control law of (3.26) has been used. The design parameters K and K_I were in both cases chosen as in Scenario I, and the nitrate set-point was 2 mg/l. The same tendency as in Scenario I is seen also in Scenario II. The disturbance rejection is faster for the feedforward feedback controller than for the PI controller. In particular, the response to changes in influent concentrations is fast. The peaks in nitrate concentration in Figure 3.10 (upper plot) are due to the peaks in the flow rate, Q.

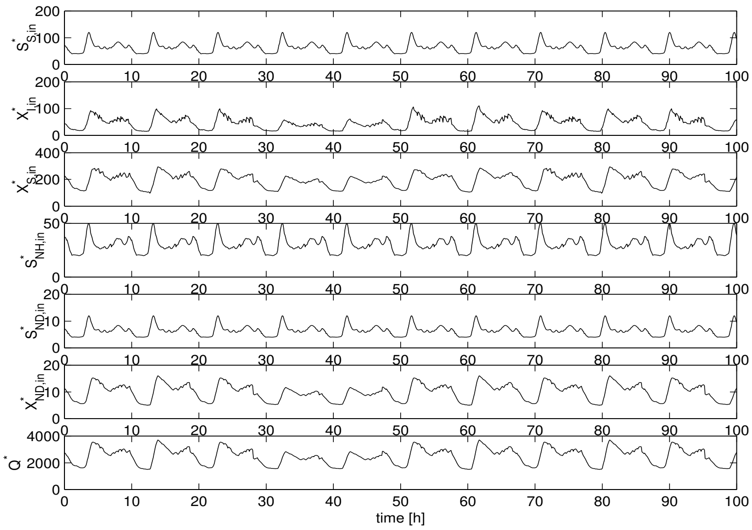

Figure 3.8: Time variations in influent water. Concentrations are given in mg/l

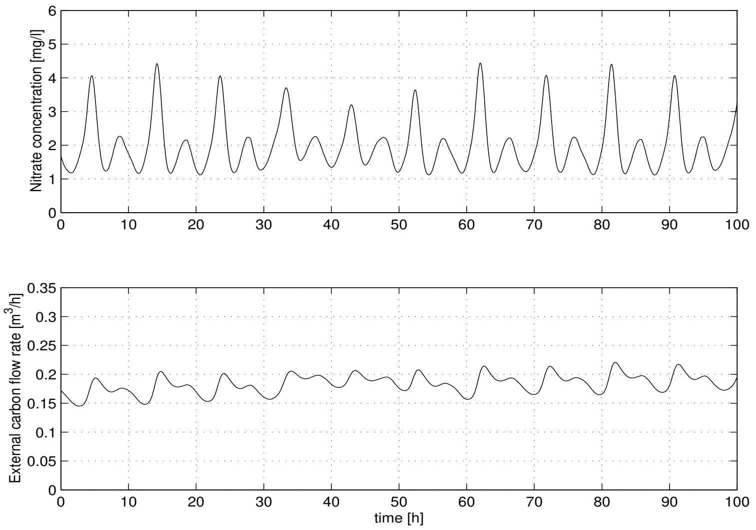

Figure 3.9: Upper plot: Nitrate concentration for simulations performed with the PI controller, influent water according to Figure 3.8. Lower plot: The corresponding external carbon flow rate is shown.

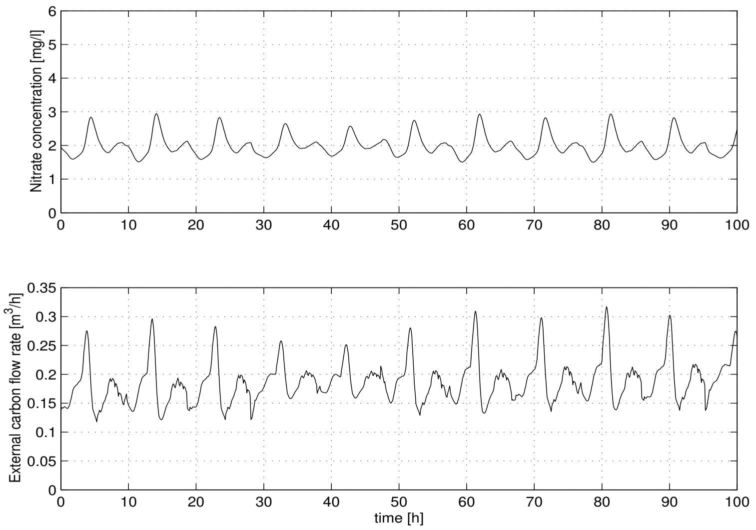

Figure 3.10: Upper plot: Nitrate concentrations for simulations with the feedforward-feedback controller, influent water according to Figure 3.8. Lower plot: The corresponding external carbon flow rate is shown.

3.5.3 Estimation of Y_H

To illustrate the estimation strategy for the heterotrophic yield Y_H outlined in Section 3.4 simulations were performed. The setup of the simulator was the same as depicted in Figure 3.5, and the detailed setup is given in Appendix B. The estimation was performed in the second anoxic compartment. The design variables in the Kalman filter were chosen as $R_1 = \begin{bmatrix} 0.1 & 0 \\ 0 & 2 \end{bmatrix}$ and R_2=0.5. The initial value of the covariance matrix was $P(0)$=0.00001I, and the initial parameter estimate was $\hat{\theta}(0) = \begin{bmatrix} 0.1 & 0 \end{bmatrix}^T$. Figure 3.11 shows the external carbon flow rate used to excite the system.

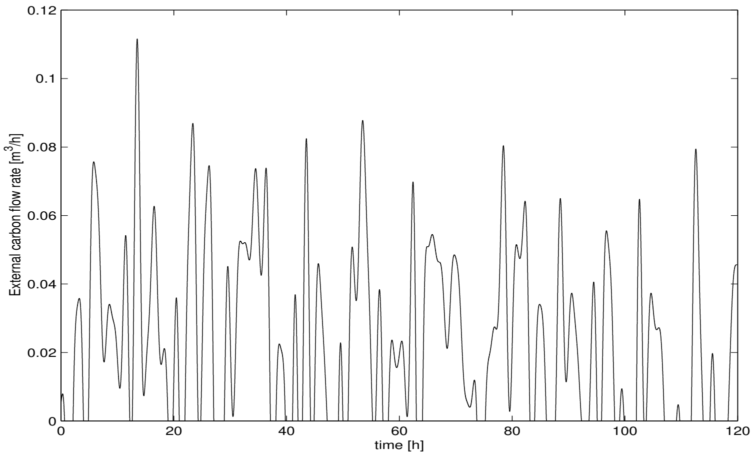

Figure 3.11: The input signal (external carbon flow rate) used to excite the system is shown.

In Figure 3.12 the results of the estimator in the nominal case is shown, *i.e.* when no measurement noise was added to the nitrate concentration. From the plot it is seen that the estimate of θ_1 converges towards an unbiased value. Figure 3.13 shows the results of the estimator in the case when a measurement noise with a standard deviation of 0.1 was added to the nitrate concentration. A small bias in the estimated parameters is here detectable. A bias analysis is currently conducted but not presented here, see Samuelsson *et al.* (2002)

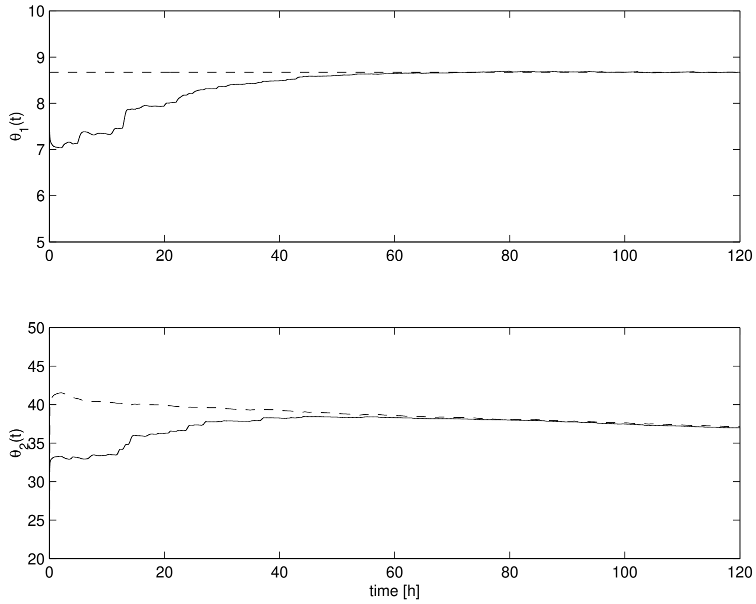

Figure 3.12: Simulations for the nominal case. Upper plot: Solid line shows the value of $\hat{\theta}_1(t)$ estimated with the Kalman filter in Section 3.4. Dashed line shows the true value. Lower plot: Solid line shows the corresponding estimated value of the hydrolysis, $\hat{\theta}_2(t)$. Dashed line shows the true value of the hydrolysis.

In the case when a measurement noise was added to the nitrate concentration, S_{NO}, the parameters $\psi(t)$ and $y(t)$ used in the estimation were low-pass filtered in the manner described in Section 3.4. The order of the Butterworth filter was 5, and the cut-off frequency 20 rad/h. These parameters have to be chosen carefully, to eliminate as much of the noise as possible without loosing too much information in the excited frequencies. The input signal and filter used above have been chosen in an *ad hoc* manner. Ideally the input signal should contain a lot of energy for frequencies smaller than the cut-off frequency of the filter, and no energy at all above it. However, in this application it is important for the input signal to be persistently exciting to avoid bias, meaning that its frequency content must be sufficiently high. The low-pass filter and the impact of the measurement noise also induces bias in the estimation. The choice of filter and input signal is therefore a very delicate tradeoff.

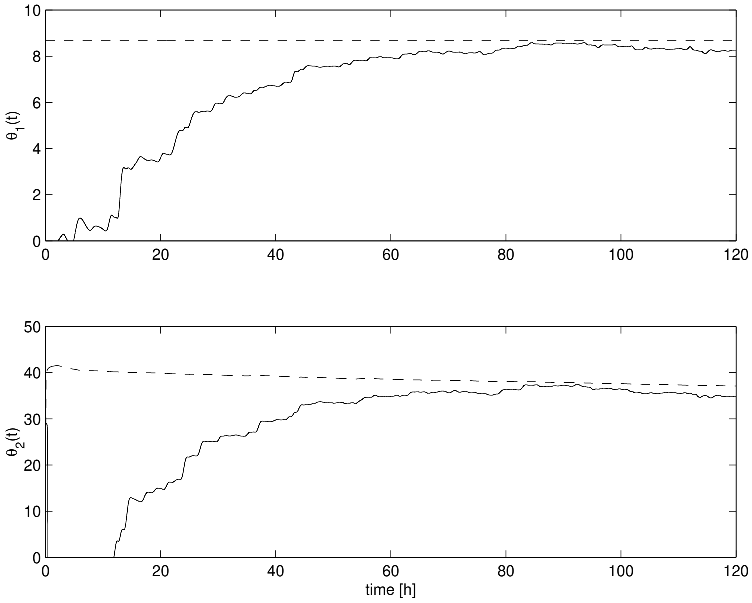

Figure 3.13: Simulations performed with a measurement noise on S_{NO} with standard deviation 0.1. Upper plot: Solid line shows the value of $\hat{\theta}_1(t)$ estimated with the Kalman filter in Section 3.4. Dashed line shows the true value. Lower plot: Solid line shows the corresponding estimated value of the hydrolysis, $\hat{\theta}_2(t)$. Dashed line shows the true value of the hydrolysis.

3.6 Conclusions

A feedforward strategy for controlling the nitrate concentration in an activated sludge process has been developed. The main objective was to reject disturbances quickly. The controller was derived from a simplified version of the ASM1 model. In a simulation study, the controller performed very well and the nitrate level was kept close to its set-point despite major process disturbances. Topics for further research could include evaluation on a full scale plant and to modify the control law to improve the performance for time varying influent flow rates. The strategy for estimating the heterotrophic yield could also be further improved.

Chapter 4

Control of the Aeration Volume in an Activated Sludge Process for Nitrogen Removal

To remove the nitrogen from the wastewater in an ASP, the biological processes, nitrification and denitrification, are needed. In aerobic compartments, ammonium may be converted into nitrate (nitrification) and in anoxic compartments nitrate may be converted into gaseous nitrogen (denitrification). For the nitrification to work well, several environmental factors must be fulfilled, including a sufficiently high concentration of DO together with a sufficiently large aeration volume, see Henze *et al.* (1995).

A common way of controlling the aeration in an ASP is to keep the DO concentration at some pre-specified level. One way to do this is in a single loop arrangement, where the DO controller directly changes the air flow valve to adjust the DO level to the desired value (set-point). In order to decrease the impact of nonlinearities in the air flow valve, cascaded DO control can be used, see Olsson *et al.* (1985). A master control loop calculates the desired air flow rate for adjusting the DO to a given set-point. In an inner loop, a slave controller is then used to adjust the valve opening to get the desired air flow rate. The cascaded control method requires measurements of DO levels and air flow rates, while the direct method requires measurements of DO levels only. Many different strategies for the DO control have been suggested. An early interesting contribution is Flanagan *et al.* (1977) where a PI control strategy using feedforward from respiration rate and flow rate is used. In Carlsson *et al.* (1994) an auto tuning approach for control of the DO is described. An approach based on exact linearization of the nonlinear oxygen transfer function is made in Lindberg and Carlsson (1996*b*). A comprehensive overview of different approaches to DO control is given in Olsson and Newell (1999).

One major drawback with the DO control methods above when a nitrifying aerobic process is used, is that neither feedback from effluent ammonium nor feedforward from influent ammonium concentrations are used. Since the load to a wastewater treatment plant may vary a factor six or more, see Van Impe *et al.* (1992), this may cause large variations of ammonium in the effluent water. Even with a constant load it is hard to manually choose the set-point of the DO controller so that it corresponds to a desired level of ammonium in the effluent.

The problem described above can partly be solved by using supervisory control (cascade control) of the set-point of the DO controller. This is done by implementing a control law that adjusts the set-point of the DO controller so that the ammonium concentration in the effluent is kept at some pre specified level. Such an approach can be found in, for instance, Lindberg and Carlsson (1996*b*).

However, the supervisory control structure may cause other problems. If the aerated volume is not sufficiently large, the air flow rate may saturate and the DO may not reach the desired level. When a pre- denitrifying structure is here used for plant operation and the DO level in the internally recirculated water is too high, this may inhibit the denitrification, Olsson and Jeppson (1994). On the other hand, a too low DO level may decrease the sludge quality and induce formation of nitrous oxide (N_2O). Also, mixers are generally needed when the DO level becomes too low. It is possible to use multivariable controllers for controlling the nitrification and denitrification simultaneously. Such an approach using quantitative feedback theory can be found in García-Sanz and Ostolaza (1998). Another multivariable approach using model predictive control is found in Weijers *et al.* (1997).

Since there are some drawbacks with the control methods described above, attempts to use other control variables have recently been made. For sequential batch reactors, an approach is to control the length of the aerated phase, see Nielsen and Lyngaard-Jensen (1993). For continuous reactors Brouwer *et al.* (1998) suggest a model based feedforward control strategy of the aeration volume. A simple ASP model is used for calculating an aeration volume that in steady state gives the desired ammonium level. This strategy was also evaluated in a pilot plant. Rule based strategies for manipulating the aerobic volume can also be found in Hoen *et al.* (1995) and Krause *et al.* (2001).

In this chapter, a method for controlling the aerobic volume of a wastewater treatment plant is presented. The method relies on exact linearization together with estimation of the reaction rate of ammonium. A related approach is the one suggested in Brouwer *et al.* (1998). The main difference is that here a model based strategy, using both traditional feedback and feedforward information, for controlling the aeration volume is suggested. Also, another strategy for on-line estimation of the reaction rate of ammonium is used here, using no assumptions regarding the model structure of the reaction rate. In the derivation of the controller it is assumed that the volume can change continuously over time. In practice, wastewater treatment plants are often divided into compartments that can be either aerated or anoxic, so a continuously changing volume is not realisable. Therefore a suggestion on how to discretize the control law with respect to volume is made. It is believed that a combination of DO control and volume control could be an interesting alternative to pure DO control. It should be emphasized that the resulting controller in the discrete volume case does not linearize the system exactly, but should be looked upon as an *ad hoc* model based control method.

The organization of this chapter is the following: The model to be used for controller design is presented in Section 4.1. In Section 4.2 the aeration volume control law is designed, and aspects of practical implementation are discussed in 4.3. Some simulation results are presented in Section 4.4. Finally, the results are discussed and conclusions are drawn in Section 4.5.

4.1 The Model

In order to design a model based control strategy for how to select the aeration volume, it is of interest to model the dynamical behaviour of the ammonium concentration, $S_{NH}(t)$ [mg/l]. By applying a simple mass balance using the notation in Figure 4.1 the following model structure is obtained (see also Bastin and Dochain (1990)):

$$\frac{dS_{NH}(t)}{dt} = R_{S_{NH}}(t) + \frac{Q(t)}{V(t)}(S_{NH,in}(t) - S_{NH}(t)) \tag{4.1}$$

where $R_{S_{NH}}(t)$ is the reaction rate of ammonium, $S_{NH,in}(t)$ is the concentration of influent ammonium, $Q(t)$ is the flow rate and $V(t)$ is the volume.

In the general case when the influent flow consists of several partial flows, $Q_1(t)$, $Q_2(t)$, ... $Q_n(t)$, the influent concentration of $S_{NH,in}(t)$ can be obtained as a weighted mean (assuming a perfect mix):

$$S_{NH,in}(t) = \frac{1}{Q(t)} \sum_{j=1}^{n} Q_j(t) S_{NH,in,j}(t)$$

The total flow rate $Q(t)$ is here given by $Q(t) = \sum_{j=1}^{n} Q_j(t)$, where $Q_j(t)$ denotes the j:th inflow to the compartment. The influent concentration corresponding

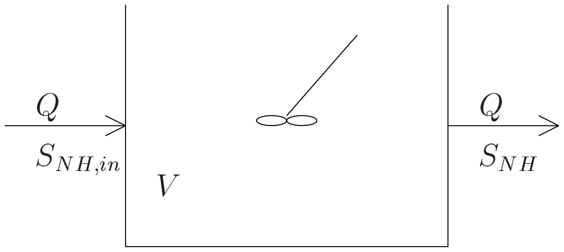

Figure 4.1: A completely mixed aerobic compartment in an activated sludge process. The compartment has volume V, the influent flow rate Q is equal to the effluent flow rate.

to the flow $Q_j(t)$ is denoted $S_{NH,in,j}(t)$.

In the ASM1 model, the reaction rate, $R_{S_{NH}}(t)$, in (4.1) is a highly nonlinear term, described by (see Henze *et al.* (1987)):

$$\begin{aligned}R_{S_{NH}} =&(-i_{XB} - \frac{1}{Y_A})\hat{\mu}_A(\frac{S_{NH}}{K_{NH}+S_{NH}})(\frac{S_O}{K_{O,A}+S_O})X_{B,A} \\ &- i_{XB}\hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\eta_g X_{B,H} \\ &- i_{XB}\hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{S_O}{K_{O,H}+S_O})X_{B,H} + k_a S_{ND} X_{B,H} \end{aligned} \tag{4.2}$$

where S_{ND} is the concentration of soluble biodegradable organic nitrogen, $X_{B,A}$ is the autotrophic biomass, $X_{B,H}$ is the heterotrophic biomass, S_S is the readily biodegradable substrate and S_{NO} is the nitrate nitrogen. The time index has here been omitted for convenience. See Henze *et al.* (1987) for a detailed discussion of the involved parameters in the reaction rate.

4.2 Derivation of the Controller

Since major process disturbances such as time varying flow rates and influent concentrations affect the process, a control strategy that combines feedforward knowledge and traditional feedback could be of use. This could be achieved by exact linearization via feedback, see for instance Khalil (1996) for a general description, and Bastin and Dochain (1990) for applications in control of bioreactors. In the controller derivation here, the output is chosen as the ammonium concentration in a totally mixed aerated compartment, and the control signal is chosen as the volume $V(t)$. It is seen directly from (4.1) that a linearizing control law is given by:

$$V(t) = \frac{Q(t)(S_{NH,in}(t) - S_{NH}(t))}{u_{lin}(t) - R_{S_{NH}}(t)} \tag{4.3}$$

yielding

$$\frac{dS_{NH}}{dt} = u_{lin}(t) \tag{4.4}$$

Theoretically, the problem is thus reduced to control an integrator with the control signal $u_{lin}(t)$.

There are many possible ways to choose the remaining control signal $u_{lin}(t)$ in (4.4). Here, a simple PI controller is chosen:

$$u_{lin}(t) = Ke(t) + K_I \int_{t_0}^{t} e(\tau)d\tau \tag{4.5}$$

where

$$e(t) = S_{NH}^{sp}(t) - S_{NH}(t) \tag{4.6}$$

Here, $S_{NH}^{sp}(t)$ is the ammonium set-point.

4.3 Practical Implementation

When implementing the control law in practice, some modifications of (4.3) have to be done:

1. The reaction rate $R_{S_{NH}}(t)$ first has to be estimated.
2. In a typical ASP, a fixed number of compartments can be aerated or non aerated. Hence, only discrete volume changes can be utilised, i.e the realisable volume V belongs to a set $V = [V_1, \ V_2, \ ... \ V_n]$, where V_1 corresponds to the minimum aeration volume and V_n to the maximum aeration volume.
3. The control law has to be discretized with respect to time, and anti windup should be implemented. This is standard procedure and is not discussed further.

4.3.1 Estimation of the Reaction Rate

A simple approach to estimate $R_{S_{NH}}(t)$ is to use (4.1) directly. From (4.1) it is clear that:

$$R_{S_{NH}}(t) = \frac{dS_{NH}(t)}{dt} - \frac{Q(t)}{V(t)}(S_{NH,in}(t) - S_{NH}(t)) \tag{4.7}$$

If the backward difference method is used to approximate the derivative in (4.7), an estimate of $R_{S_{NH}}(t)$ can be written as:

$$\hat{R}_{S_{NH}}(t) = \frac{S_{NH}(t) - S_{NH}(t-h)}{h} - \frac{Q(t)}{V(t)}(S_{NH,in}(t) - S_{NH}(t)) \tag{4.8}$$

Here h denotes the sampling interval, the index t denotes the current sample and the index $t-h$ denotes the previous sample. Since the estimate of the derivative in (4.8) may be affected by measurement noise and other fast transients, $\hat{R}_{S_{NH}}(t)$ should be low pass filtered before being used in the controller. A suitable choice may be a first order low pass filter with the transfer function:

$$H(q) = \frac{\alpha q}{q + \alpha - 1} \tag{4.9}$$

where q is the forward shift operator.

4.3.2 Discrete changes of the aerobic volume

As mentioned, the nature of the possible volume changes is discrete, since the plant is divided into compartments. Thus, to implement the control law (4.3), it has to be combined with some form of logical programming that takes this into account. The suggestion is therefore to control the ammonium in the last aerated compartment, by using (4.3) to calculate the desired aeration volume. The nearest larger volume V_j in the possible set $V = [V_1, \;\; V_2, \;\; ... \; V_n]$ should then be aerated. The rationale for this is to keep a low DO level, by using a larger volume. In case the total aeration volume becomes too large and the non aerated volume too small, it is possible to instead select the nearest smaller volume in the set.

To achieve the ammonium set-point, and to avoid fluctuations in the ammonium concentration, the strategy above needs to be combined with a slow supervisory DO controller that adjusts the set-point of the DO controllers in the aerated compartments. One possibility is to use the same DO set-point in all aerated compartments. A simple choice that is used in this study, is to use a PI controller for calculating the DO set-point:

$$DO^{sp}(t) = -Ke(t) - K_I \int_{t_0}^{t} e(\tau)d\tau \tag{4.10}$$

where $e(t)$ is the control error of ammonium in the last aerated compartment, see (4.6). To prevent the DO concentration to take too high or too low values, $DO^{sp}(t)$ should be given both an upper and a lower bound.

It should be mentioned that the control method works better the smaller the discrete volumes are. If sufficiently small volumes can be aerated, it might be possible to exclude the DO control part. This could be the case when techniques such as “finger-clipping” are used, see Waltin (2000).

4.3.3 Summary of the Control Strategy

Below, a short summary of the control strategy is given:

1. Estimate $R_{S_{NH}}(t)$ according to (4.8).

2. Calculate the desired aeration volume (4.3), using the estimated reaction rate (4.8).

3. Aerate the nearest larger volume, V_j, in the realisable set $V = [V_1,\ \ V_2,\ ...\ V_n]$.

4. Use (4.10) to adjust the DO set-point.

When using discrete volumes, as above, the control method should be looked upon as an *ad hoc* method rather than an exact linearization strategy.

4.4 A Simulation Study

4.4.1 Simulator Setup

The suggested control law was evaluated in a simulation study. A post denitrifying ASM1 model with ten compartments was implemented and used in the study. Each compartment had a volume of 235 m^3. Matlab was used to implement the model. The model parameters were in both cases the default values reported in Henze *et al.* (1987). Some additional data regarding default flow rates and influent concentrations can be found in Appendix B. The layout of the simulated plant is shown in Figure 4.2.

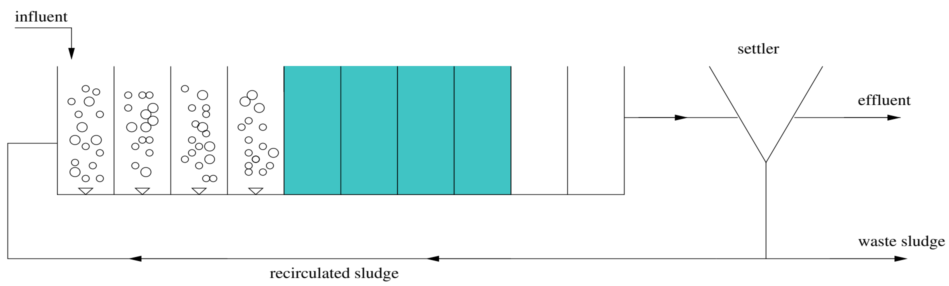

Figure 4.2: The layout of the simulated ASP. The first four compartments are always aerated. The shaded compartments are allowed to be aerated when needed.

4.4.2 Simulations

Influent load variations according to Figure 4.3 was used. The superscript * in the figure denotes that these are influent flow rate and concentration, not to be mixed up with the total flow rate to the controlled aerobic compartment. After 15 hours the ammonium concentration in the influent water is increased from 24 to 32 mg/l. After 65 hours the influent flow rate is changed from 250 to 300 m^3/h.

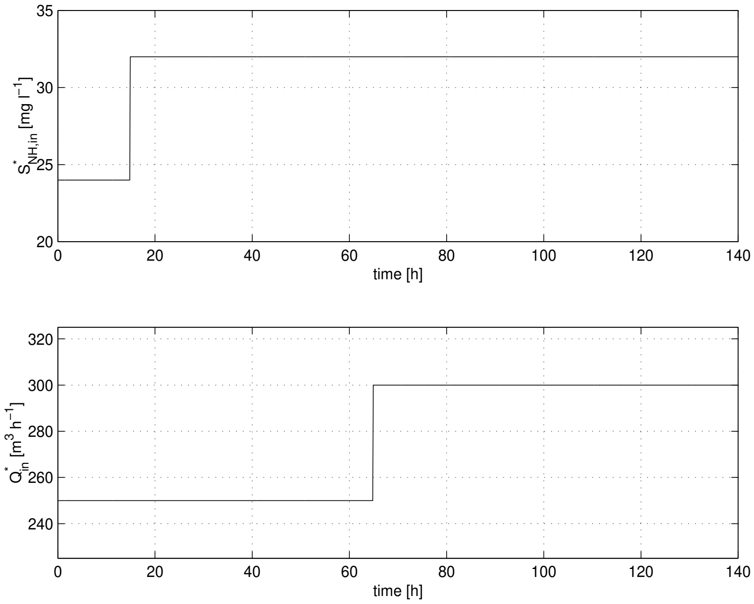

Figure 4.3: Disturbances in the influent water. Upper plot: Influent ammonium. Lower plot: Influent flow rate.

This influent scenario was used first when simulating ordinary supervisory DO control with a PI controller, see (4.10). The sampling interval was 0.001h. The parameters of the supervisory PI controller were chosen as K=0.0012 and K_I=0.08. The set-point of ammonium, S_{NH}^{sp}, was chosen to be 2.5 mg/l. The constant total volume of the aerobic compartments was 1175 m^3. In Figure 4.4 the simulation results for this experiment is shown. It is clearly seen that the disturbance rejection, especially for the second disturbance is poor, and that the DO concentration takes very high values. This is caused by a too small aeration volume. The suggested volume controller was then simulated in combination with the supervisory DO controller, (4.10). In the simulations, the volume of the aerated compartment was allowed to take the discrete values V = [940, 1170 ... 1880]. The maximum allowed value of $DO^{sp}(t)$ was 5 mg/l and the minimum value was 1.5 mg/l. The influent characteristics and the set-point of ammonium in the last aerated compartment were the same as in the

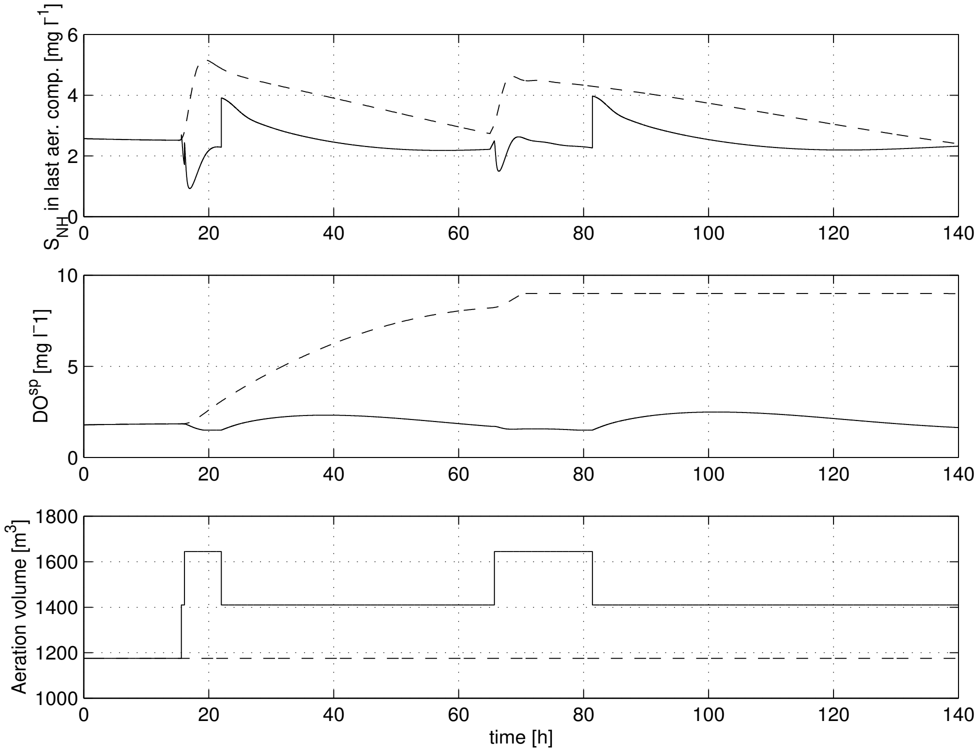

Figure 4.4: Comparison between supervisory DO control with fixed volume (dashed lines) and aeration volume control (solid lines). Upper plot: Ammonium concentration in the last aerobic compartment. Intermediate plot: Set-point of the DO controllers. Lower plot: Aeration volume.

simulation where only the supervisory DO controller was used. The PI parameters of the supervisory DO controller were the same, as well as the sampling rate. The volume control loop was sampled with the interval 0.1 h, and the PI parameters in (4.3) were chosen as K=0.00001 and K_I=0.0000001. The estimation of the reaction rate $\hat{R}_{S_{NH}}(t)$ in (4.8) was performed over only the first four compartments to avoid fast transients, since these compartments were assumed to always be aerobic. The parameter in the low pass filter, (4.9), was chosen as α=0.07. Compared to when only the supervisory DO controller is used, it is seen that the disturbance rejection is significantly faster and that the required DO level for this control strategy is much lower, see Figure 4.4.

In Figure 4.5 the behaviour of the control strategy when $S^*_{NH,in}$ varies periodically with a 7.5 hour periode is shown. The ammonium set-point and all PI parameters were the same as in the previous example. The DO levels are still kept at intermediate values, and the disturbance attenuation is fast. The volume controller and the DO set-point controller cooperates in order to achieve the ammonium set-point.

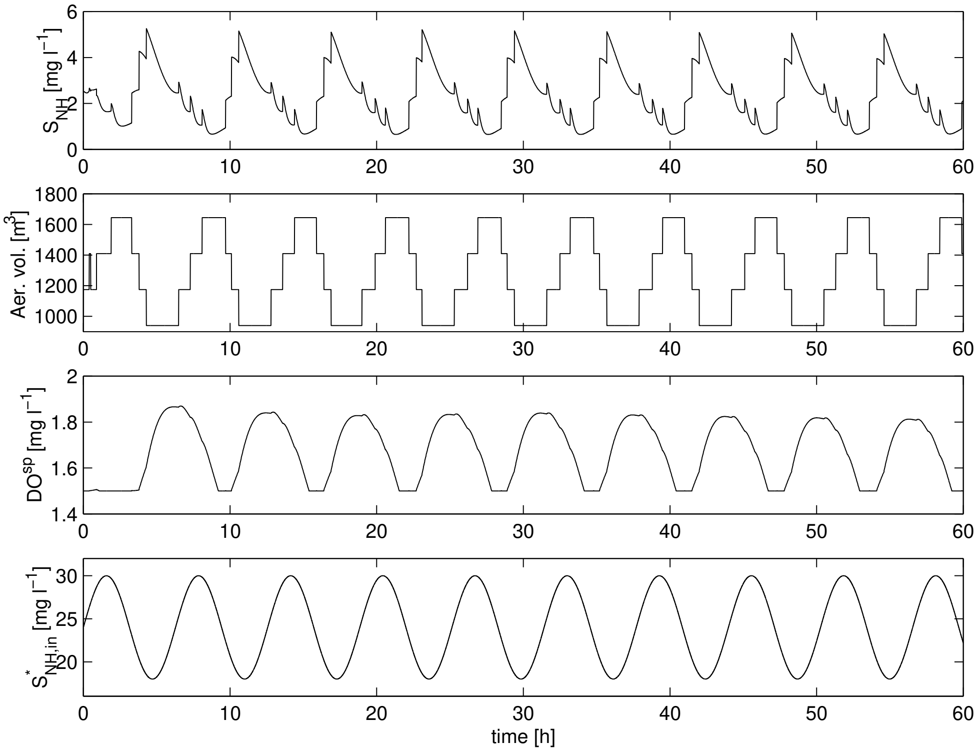

Figure 4.5: Illustration of the volume controller for a periodical variation in influent water. Upper plot: Ammonium concentration in the last aerobic compartment. Second plot: Aeration volume. Third plot: DO set-point. Lower plot: Influent ammonium concentration.

4.5 Conclusions

In this chapter, a method based on linearization via feedback, for controlling the aeration volume in an activated sludge process has been presented. The main objective was to reject changes in influent load variations quickly. Simulations suggest that the aeration volume could be an efficient control variable, and that feedforward methods could give a good result for controlling the aeration volume. It is also suggested how to implement the control law in practice, i.e when the possible volume changes are discrete. It should be mentioned that the tuning of the aeration controller must be done with care to avoid oscillations. More advanced logics for handling the discrete volume changes could also be incorporated. For example, the impact on the denitrification in the remaining anoxic compartments could be considered. An interesting step would be to evaluate the suggested strategy in a pilot plant.

Appendix A

Appendix to Chapter 2

A.1 The Simulation Model

To evaluate different controllers and control strategies it is important to use a model which realistically simulates a true plant. In the simulator, the IAWQ Activated Sludge Model No. 1 (ASM1) by (Henze *et al.* 1987) has been used to model each compartment of the implemented process. The settler is modeled with the clarification-thickening model by (Takács *et al.* 1991).

For a completely mixed compartment, the following model structure holds for a general component Z, see (Bastin and Dochain 1990):

$$\frac{dZ}{dt} = R_Z(t) + D(t)(Z_{in}(t) - Z(t))$$

where $Z_{in}(t)$ is the influent concentration of the actual component, $Z(t)$ is the effluent concentration which is equal to the concentration in the compartment under the assumption of complete mixing, $D(t)$ is the dilution rate (flow/volume), and $R_Z(t)$ denotes the reaction (growth) rate for the component.

The different components of the ASM1 model are given in Table A.1.

$S_{NH}(t)$	soluble ammonium nitrogen
$S_{NO}(t)$	soluble nitrate nitrogen
$S_{ND}(t)$	soluble biodegradable organic nitrogen
$S_O(t)$	dissolved oxygen
$S_S(t)$	soluble readily biodegradable substrate
$S_I(t)$	soluble inert organic matter
$S_{ALK}(t)$	alkalinity
$X_{B,A}(t)$	active autotrophic biomass
$X_{B,H}(t)$	active heterotrophic biomass
$X_{ND}(t)$	particulate biodegradable organic nitrogen
$X_S(t)$	slowly biodegradable substrate
$X_I(t)$	particulate inert organic matter and products

Table A.1: Components included in the model.

According to ASM1 the reaction rates for the components are:

$$
\begin{aligned}
R_{S_S} &= -\frac{1}{Y_H}\hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{S_O}{K_{O,H}+S_O})X_{B,H} \\
&- \frac{1}{Y_H}\hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\eta_g X_{B,H} \\
&+ k_h\frac{\frac{X_S}{X_{B,H}}}{K_X+\frac{X_S}{X_{B,H}}}\left((\frac{S_O}{K_{O,H}+S_O})+\eta_h(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\right)X_{B,H} \\
R_{X_I} &= f_P(b_H X_{B,H}+b_A X_{B,A}) \\
R_{X_S} &= (1-f_P)(b_H X_{B,H}+b_A X_{B,A}) \\
&- k_h\frac{\frac{X_S}{X_{B,H}}}{K_X+\frac{X_S}{X_{B,H}}}\left((\frac{S_O}{K_{O,H}+S_O})+\eta_h(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\right)X_{B,H} \\
R_{X_{B,H}} &= \hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{S_O}{K_{O,H}+S_O})X_{B,H} \\
&+ \hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\eta_g X_{B,H} - b_H X_{B,H} \\
R_{X_{B,A}} &= \hat{\mu}_A(\frac{S_{NH}}{K_{NH}+S_{NH}})(\frac{S_O}{K_{O,A}+S_O})X_{B,A} - b_A X_{B,A} \\
R_{S_O} &= -\frac{1-Y_H}{Y_H}\hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{S_O}{K_{O,H}+S_O})X_{B,H} \\
&- \frac{4.57-Y_A}{Y_A}\hat{\mu}_A(\frac{S_{NH}}{K_{NH}+S_{NH}})(\frac{S_O}{K_{O,A}+S_O})X_{B,A} + K_L a(S_{O,sat}-S_O)
\end{aligned}
$$

$$
\begin{aligned}
R_{S_{NO}} &= -\frac{1-Y_H}{2.86Y_H}\hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\eta_g X_{B,H} \\
&+ \frac{1}{Y_A}\hat{\mu}_A(\frac{S_{NH}}{K_{NH}+S_{NH}})(\frac{S_O}{K_{O,A}+S_O})X_{B,A} \\
R_{S_{NH}} &= -i_{XB}\hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{S_O}{K_{O,H}+S_O})X_{B,H} \\
&- i_{XB}\hat{\mu}_H(\frac{S_S}{K_S+S_S})(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\eta_g X_{B,H} \\
&- (i_{XB}+\frac{1}{Y_A})\hat{\mu}_A(\frac{S_{NH}}{K_{NH}+S_{NH}})(\frac{S_O}{K_{O,A}+S_O})X_{B,A} + k_a S_{ND} X_{B,H} \\
R_{S_{ND}} &= -k_a S_{ND} X_{B,H} \\
&+ \frac{X_{ND}}{X_S}k_h\frac{\frac{X_S}{X_{B,H}}}{K_X+\frac{X_S}{X_{B,H}}}\left((\frac{S_O}{K_{O,H}+S_O})+\eta_h(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\right)X_{B,H} \\
R_{X_{ND}} &= (i_{XB}-f_P i_{XP})(b_H X_{B,H}+b_A X_{B,A}) \\
&- \frac{X_{ND}}{X_S}k_h\frac{\frac{X_S}{X_{B,H}}}{K_X+\frac{X_S}{X_{B,H}}}\left((\frac{S_O}{K_{O,H}+S_O})+\eta_h(\frac{K_{O,H}}{K_{O,H}+S_O})(\frac{S_{NO}}{K_{NO}+S_{NO}})\right)X_{B,H}
\end{aligned}
$$

The reaction rate of the alkalinity, S_{ALK}, has been omitted since it does not affect other components and the reaction rate of S_I since it is inert and does not react. For an explanation of the different process parameters and their default values, see (Henze *et al.* 1987).

The settler is modeled with ten horizontal layers where each layer is assumed to be completely mixed. The model is based on the solids mass balance around each layer. The solids flux is denoted J and is assumed to depend on settling velocity v_s and concentration of a particulate components X according to $J = v_s(X)X$. The concentration of the particulate components in a specific layer is then found from

$$\frac{dX}{dt} = \frac{1}{h}\Delta J \tag{A.1}$$

where ΔJ is the difference in flux over the layer, see (Jeppsson 1996) or (Takács *et al.* 1991) for a further description of the numerical implementation of (A.1).

A.2 Short Users Manual

A short overview on how to run the simulator is given below:

1. Start an Internet browser, i.e. Netscape Communicator.

2. Load the following URL:
   http://www.syscon.uu.se/JASS/
   where a demo version of the simulator is available.

3. Click on the *START* button in the GUI.

Plotting and presenting data

4. To choose a certain component to plot in the bar diagrams, use the menu bar to the left of a bar diagram. For instance, to plot the concentration of dissolved oxygen, choose S_O. The oxygen concentrations in influent water, all compartments, return sludge and effluent water are then displayed.

5. In the scrolling graph it is possible to plot the concentrations of two components at the same time. Three menu bars are located under the graph. In the middle menu bar it can be chosen whether to plot concentrations from influent water, any of the compartments, return sludge or the effluent water. Two components may then be selected from the menu bars to the left and to the right.

6. The simulated time is presented in two ways. First, it can be seen approximately from the timescale of the graph, secondly it is presented in a separate text field to the right in the GUI.

7. When the field *Sludge properties* is clicked, a window presenting the actual sludge age, the excess sludge flow and the sludge concentration of the plant appears.

8. The state of the settler is shown when the field *Settler state* is clicked. Here the particulate concentration in the effluent is presented. Also the particulate concentration profile in the settler is shown, see Figure A.1. When a sludge overflow occurs, a warning message appears in the window.

9. When the field *Treatment results* is clicked, a window illustrating the concentrations of COD, BOD, total N, ammonium (*Snh*), nitrate (*Sno*) and Total Suspended Solids (TSS) in influent and effluent of the plant appears. The reduction percentage is also shown in the window, see further Appendix 3.

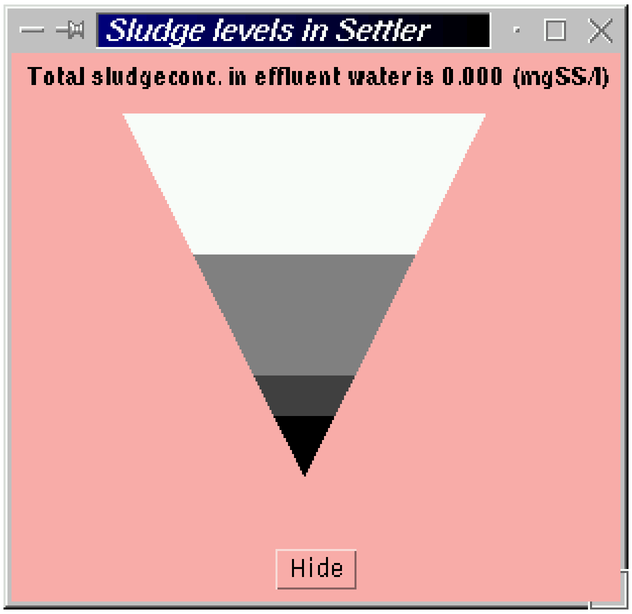

Figure A.1: The particulate concentration profile of the settler. The darkness of the colour is proportional to the sludge concentration.

Changing the process

10. To change a flow rate, click on the corresponding pump symbol in the process image and enter a new value in the appearing window. An active pump will be colored green, and an inactive yellow. Also external carbon (ethanol) can be added by clicking on *Carb. flow zone 1* .

11. To change ASM1 process parameters, click on the field *Constants* (upper left corner) in the process image. A new window where the new parameters may be entered is shown. Here, it is also possible to choose the simulated temperature of the water. To make the text fields writeable, they must be clicked with the cursor. When a new value has been chosen, push the *ok* button in the window. The new values are then used in the process, and the window closes. If the button *Cancel* is chosen, no changes in the process are done and the window disappears. The window is depicted in Figure A.2.

12. To change concentrations of the components in the influent water, click in the field *Influent*. A window similar to the one described above appears, and new values can be entered in the same way.

13. By clicking in the field *Zone volumes and settler properties*, a window appears where the volumes of all compartments can be changed. Here, it

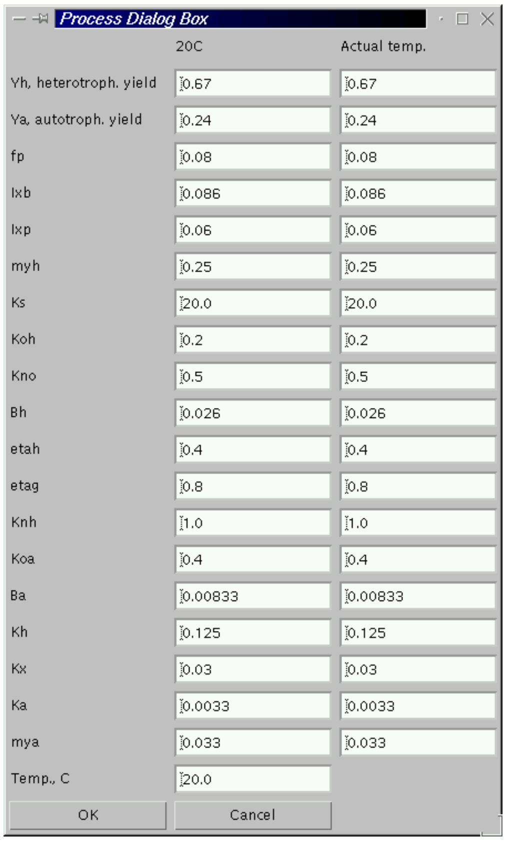

Figure A.2: The window for changing the ASM1 process parameters.

is also possible to change the settler parameters such as settler area, settler height and settling velocity.

Controlling the process

14. During simulation, it is possible to choose between running the simulator with a fixed excess sludge flow or with a fixed sludge age. In the default setup, the simulator is running with a fixed sludge flow. By clicking in the field *Sludge age contr.* and enter a desired fixed sludge age in the appearing window, the simulator will run with a fixed sludge age instead. Clicking on the field *Sludge flow contr.* and enter a desired flow value will change the mode back, so that the simulations again run with a fixed excess sludge flow.

15. To change the reference values of dissolved oxygen, go to the panel under the scrolling graph (located under the menu bar *DO PID*). First select in which compartment (zone 1 - zone 10) to change the reference value. Next, enter a DO reference value in the text field just below the button

Set DOref and then push the *Set DOref* button. Note that the air bubbles in the compartment will disappear when $DOref = 0$ is chosen.

16. When the option *DO set point controller* is chosen from the menu bar *DO PID* under the graph, it is possible to run a supervisory control of the DO controllers. The user may then enter a set-point for the ammonium concentration in the last aerated compartment. When the button *Set Snhref* is selected, the supervisory control starts. Once the option *DO PID* is chosen in the menu bar again, the supervisory control will be stopped.

17. When the option *C controller* is chosen from the menu bar *DO PID* under the graph, it is possible to automatically control the flow rate of external carbon source. The controller parameters are fixed, but the user may enter the set-point of nitrate in the last anoxic compartment by entering a value in the textfield and then push the *Ok* button.

Reseting the simulator

18. In case the simulator needs to be reseted, press the button *RESET* in the GUI followed by *shift+reload* in the Communicator view menu (Netscape).

Appendix B

Appendix to Chapter 3

B.1 Proof of Lemma 1

Assume that $\bar{S}_S$ is finite (this means that (3.18) is fulfilled). From (3.17) and (3.14) it is seen that, at steady state:

$$\frac{1}{V}\bar{u} = \frac{1}{Y_H}\mu(\bar{S}_S, \bar{S}_{NO})\bar{X}_{B,H} - k\bar{X}_{B,H} - D(S_{S,in} - \bar{S}_S) \tag{B.1}$$

$$\bar{X}_{B,H} = \frac{DY_H}{\beta\mu(\bar{S}_S, \bar{S}_{NO})}(S_{NO,in} - \bar{S}_{NO}) \tag{B.2}$$

To see that (3.20) gives the desired set-point for S_{NO}, insert (3.20) and (B.2) in (B.1). This gives immediately:

$$\bar{S}_{NO} = S_{NO}^{ref}$$

which proves the first part of Lemma 1.

The steady state value of S_S can be obtained as follows. Equations (3.12) and (B.2) give

$$\bar{X}_{B,H} = \frac{D}{D + b_H - \mu(\bar{S}_S, \bar{S}_{NO})}X_{BH,in} \tag{B.3}$$

$$\bar{S}_{NO} = S_{NO,in} - \frac{\beta}{Y_H D}\mu(\bar{S}_S, \bar{S}_{NO})\bar{X}_{B,H} \tag{B.4}$$

Inserting the expressions (3.15) and (B.3) in (B.4) and solve for $\bar{S}_S$ now yields :

$$\bar{S}_s = \frac{K_S(D+b_H)(S_{NO,in}-S_{NO}^{ref})}{\hat{\mu}_H \frac{S_{NO}^{ref}}{K_{NO}+S_{NO}^{ref}}\eta_g(S_{NO,in}+\frac{\beta}{Y_H}X_{B,H,in}-S_{NO}^{ref})-(D+b_H)(S_{NO,in}-S_{NO}^{ref})} \tag{B.5}$$

which proves the second part of the lemma.

In order for (B.5) (and (3.21)) to be finite, its denominator must not be equal to zero. According to (3.18), the same denominator must be positive for $\bar{S}_s$ to be positive. This gives after straightforward manipulations the expression (3.18). The condition (3.18) guarantees positive solutions to exist for (B.5).

B.2 Proof of Lemma 2

Assume that the external carbon source is added only to the first compartment in the line. This gives the model:

$$\frac{dX_{B,H,1}}{dt} = \mu(S_{S,1}, S_{NO,1})X_{B,H,1} - b_H X_{B,H,1} + D(X_{B,H,in} - X_{B,H,1}) \tag{B.6}$$
$$\frac{dS_{S,1}}{dt} = -\frac{1}{Y_H}\mu(S_{S,1}, S_{NO,1})X_{B,H,1} + kX_{B,H,1} + D(S_{S,in} - S_{S,1}) + \frac{1}{V}u \tag{B.7}$$
$$\frac{dS_{NO,1}}{dt} = -\frac{\beta}{Y_H}\mu(S_{S,1}, S_{NO,1})X_{B,H,1} + D(S_{NO,in} - S_{NO,1}) \tag{B.8}$$
$$\frac{dX_{B,H,2}}{dt} = \mu(S_{S,2}, S_{NO,2})X_{B,H,2} - b_H X_{B,H,2} + D(X_{B,H,1} - X_{B,H,2}) \tag{B.9}$$
$$\frac{dS_{S,2}}{dt} = -\frac{1}{Y_H}\mu(S_{S,2}, S_{NO,2})X_{B,H,2} + kX_{B,H,2} + D(S_{S,1} - S_{S,2}) \tag{B.10}$$
$$\frac{dS_{NO,2}}{dt} = -\frac{\beta}{Y_H}\mu(S_{S,2}, S_{NO,2})X_{B,H,2} + D(S_{NO,1} - S_{NO,2}) \tag{B.11}$$
$$\vdots$$
$$\vdots$$
$$\frac{dX_{B,H,n}}{dt} = \mu(S_{S,n}, S_{NO,n})X_{B,H,n} - b_H X_{B,H,n} + D(X_{B,H,n-1} - X_{B,H,n}) \tag{B.12}$$
$$\frac{dS_{S,n}}{dt} = -\frac{1}{Y_H}\mu(S_{S,n}, S_{NO,n})X_{B,H,n} + kX_{B,H,n} + D(S_{S,n-1} - S_{S,n}) \tag{B.13}$$
$$\frac{dS_{NO,n}}{dt} = -\frac{\beta}{Y_H}\mu(S_{S,n}, S_{NO,n})X_{B,H,n} + D(S_{NO,n-1} - S_{NO,n}) \tag{B.14}$$

For the last compartment in the line, compartment n, it will during steady state hold that, according to (B.13) and (B.14):

$$0 = \frac{1}{Y_H}\mu(\bar{S}_{S,n}, \bar{S}_{NO,n})\bar{X}_{B,H,n} - k\bar{X}_{B,H,n} - D(\bar{S}_{S,n-1} - \bar{S}_{S,n}) \tag{B.15}$$
$$\bar{X}_{B,H,n} = \frac{DY_H}{\beta\mu(\bar{S}_{S,n}, \bar{S}_{NO,n})}(\bar{S}_{NO,n-1} - \bar{S}_{NO,n}) \tag{B.16}$$

Combining (B.15) and (B.16) gives:

$$\begin{aligned} 0 =& \frac{D}{\beta}(\bar{S}_{NO,n-1} - \bar{S}_{NO,n}) - k\frac{DY_H}{\beta\mu(\bar{S}_{S,n}, \bar{S}_{NO,n})}(\bar{S}_{NO,n-1} - \bar{S}_{NO,n}) \\ &- D(\bar{S}_{S,n-1} - \bar{S}_{S,n}) \end{aligned} \tag{B.17}$$

The exactly similar procedure for compartment $n-1$ gives the expression:

$$\bar{S}_{S,n-1} = -\frac{1}{\beta}(\bar{S}_{NO,n-2} - \bar{S}_{NO,n-1}) + \frac{kY_H}{\beta\mu(\bar{S}_{S,n-1}, \bar{S}_{NO,n-1})}(\bar{S}_{NO,n-2} - \bar{S}_{NO,n-1}) + \bar{S}_{S,n-2} \tag{B.18}$$

Inserting (B.18) in (B.17) now yields:

$$0 = \frac{D}{\beta}(\bar{S}_{NO,n-2} - \bar{S}_{NO,n}) - D(\bar{S}_{S,n-2} - \bar{S}_{S,n}) - \sum_{i=n-1}^{n} k\frac{DY_H}{\beta\mu(\bar{S}_{S,i}, \bar{S}_{NO,i})}(\bar{S}_{NO,i-1} - \bar{S}_{NO,i}) \tag{B.19}$$

Clearly, continuing this procedure replacing $S_{S,n-2}$ by the expression from compartment $n-2$ and so on down to the expression for compartment 2 finally yields:

$$0 = \frac{D}{\beta}(\bar{S}_{NO,1} - \bar{S}_{NO,n}) - D(\bar{S}_{S,1} - \bar{S}_{S,n}) - \sum_{i=2}^{n} k\frac{DY_H}{\beta\mu(\bar{S}_{S,i}, \bar{S}_{NO,i})}(\bar{S}_{NO,i-1} - \bar{S}_{NO,i}) \tag{B.20}$$

Finally, inserting the expression for $\bar{S}_{S,1}$ obtained from (B.7) and (B.8) in (B.20) and solve for the control signal u gives the general expression for the carbon flow rate,

$$\bar{u} = Q[\frac{1}{\beta}(S_{NO,in} - \bar{S}_{NO,n}) - (S_{S,in} - \bar{S}_{S,n}) - \frac{kY_H}{\beta}\sum_{i=1}^{n}\frac{(\bar{S}_{NO,i-1} - \bar{S}_{NO,i})}{\mu(\bar{S}_{S,i}, \bar{S}_{NO,i})}] \tag{B.21}$$

which proofs Lemma 2

B.3 Default Simulator Setup

Comp no.	Volume of cell (m^3)	Operation
1	896	Aerobic
2	896	Aerobic
3	1792	Aerobic
4	896	Aerobic
5	896	Aerobic
6	1878	Anoxic
7	936	Anoxic
8	936	Anoxic
9	936	Anoxic
10	3136	Anoxic

Table B.1: The simulator simulates 10 compartments, compartments 1 to 5 are aerobic and compartments 6-10 are anoxic.

Flow	Rate [m^3/h]
Influent	3000
Excess sludge	20
Return sludge	3000
Internal recirculation	0

Table B.2: Nominal values of the flow rates in the the simulation model.

Standard abbrev.	Influent conc. (mg/l)
$X_{B,H}$	0
$X_{B,A}$	0
X_S	82
$X_I and X_P$	43
$X_{N,D}$	5.8
$S_{N,H}$	23.5
$S_{N,D}$	2.1
$S_{N,O}$	1
S_I	37
S_O	0.1
S_S	70

Table B.3: Influent wastewater composition.

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