SEQUENCING BATCH REACTOR FOR ORGANICS

AND NITROGEN REMOVAL: MODELING AND

DESIGN OPTIMIZATION

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CHAPTER I

INTRODUCTION

1.1   Background

A sequencing batch reactor (SBR) process is a fill-and-draw activated-sludge system for

wastewater treatment. In this system, wastewater is added to a single "batch" reactor,

treated to remove undesirable components, and then discharged (U.S. EPA, 1999). SBRs

are essentially the batch reactors which have been widely used in the chemical and

pharmaceutical industries for a long time. The term sequencing batch reactor stems from

the sequence of steps the reactor goes through as it receives wastewater, treats it, and

discharges it, since all steps are accomplished in a single tank. This process is identical

in concept to a continuous flow activated sludge system, but the SBR is a self contained

system performing equalization, aeration and clarification in a single reactor. Although

the activated sludge process for wastewater treatment was first developed as a batch

system, the configuration was quickly changed to continuous flow. This was due to the

high demands on operator time, lack of specialized technological equipment and some

operational problems of batch systems like clogging of aeration diffusers. During the

past decades, the development of new hardware such as motorized and pneumatically

actuated valves, electronic and mechanical timers, level sensors, jet aerators and

microprocessors led to the revival of batch treatment technology (Katsogiannis et al., 1999).

In recent years, SBRs have gained great interest for wastewater treatment, because of their simple configurations (all necessary processes are taking place time-sequenced in a single basin). SBRs could achieve nutrient removal using an alternate mode of anoxic and aerobic periods, so nitrification and denitrification are achieved in the mentioned periods, while the separation of treated wastewater and biomass is accomplished by ceasing aeration and/or mixing at the end of process cycle. Due to its operational flexibility, it is quite simple to increase the efficiency in treating wastewater by changing the duration of each phase rather than adding or removing tanks in continuous flow systems (Mahvi et al., 2005).

1.2   Research Needs

Despite the growing popularity, a widely accepted approach to process analysis and modeling, a unified design basis, and even common terminology are still lacking for SBR systems. This situation is now regarded as the major obstacle hindering broader practical application of the SBR (Artran et al., 2001). The design engineers are reluctant to use mathematical models in practice because some mathematical models have complex expressions and have parameters which are either not normally available or difficult to estimate. Instead they prefer to use the conventional design methods based on simple expressions developed from simple models for steady state conditions, for example the model developed by Lawrence and McCarty (1970) for activated sludge processes. This

trend in design practice is still continuing although SBRs have been widely used in municipal wastewater treatment for the last two decades.

In United States, most SBR installations are used for smaller wastewater treatment systems of less than two million gallons per day (U.S. EPA, 1999). In recent years many larger installations are showing interest in the SBR system as it provides more operational flexibility to meet stringent effluent limits. However, capital and operating costs are the primary limitations to SBR process. Capital cost increases with reactor volume, with the size of single unit limited by practical construction and mixing limitations. Multiple parallel reactors will greatly increase capital costs, eventually offsetting savings in sludge separation and handling equipment. Generally, the transition from batch to continuous treatment is at 3,800 m^3/d (1 MGD) (Celenza, 2000). Also from this author's experience, the SBR system is more economical over conventional treatment process at smaller flows but for larger installation it does not provide much cost benefit due to its larger reactor volume, higher level of sophistication in instrumentation and increased solids production rate (Velmurugan and Veenstra, 2008). The following are the main reasons attributed to increased cost of SBRs for larger installations:

1. Successful SBR process design is more complex than conventional activated sludge process. The number of reactors, reactor volume and reactor proportions, which are used to receive incoming sewage and retain effluent and settled sludge, will each have an effect on the total plant volume and the capital cost of each particular plant (Boon, 2003). Process designs for SBR are mostly performed by

SBR equipment manufacturers using the conventional design approach based on mean cell residence time (MCRT), food to microorganism ratio (F/M ratio), mixed liquor suspended solids concentration (MLSS), and hydraulic retention time (HRT) (Geselbracht, 2007). The sizing of reactors and equipment based on this approach does not always result in a cost effective system as the designers usually use conservative design criteria. As SBRs are proven to perform well under varying influent flow and shock loadings (Aqua-Aerobic Systems, 1986; Nakhla et al., 1997), conservative design approach is not required to provide a safety margin against uncertainties in influent flow conditions.

2. Model-based design approach for SBR is not as widely practiced as continuous activated sludge process due to the complexity of the unsteady state nature of the process and sequencing of different environmental conditions such as anoxic, aerobic and anaerobic involved in SBR process.

3. The application of cost optimization techniques to the design of SBR has not been studied as extensively as in a continuous activated sludge system. Most of the optimization studies done in SBR were limited to operational policies such as reducing total cycle time or aeration time or improving efficiency. Even these optimization studies ignore the solids processing cost in the objective function. Solids processing costs are one of the major operational costs and should be given the same consideration as aeration energy cost.

1.3   Research Goal and Objectives

The overall research goal of this study, therefore, is to develop an optimization model for the design of SBR with a new process model that describes the process adequately with as few parameters as possible to predict the performance well under all operating conditions. This study will serve two purposes: 1) promote the use of model-based design methods among the practicing engineers, and 2) make the SBR system more attractive to larger installations if cost reduction can be achieved. In order to accomplish the overall research goal, the objectives of this work are formulated as follows:

1. Develop a new model specifically for the SBR for removal of biodegradable organics and nitrogen to promote the use of a model-based design approach. The proposed new model would require fewer parameters unlike advanced models but at the same time, describe all the processes involved without over-simplification.

2. Calibrate and validate the developed model with the data obtained from the operation of an existing full-scale 5.3 MGD (836 m^3/h) SBR system treating municipal wastewater. A sampling program and test parameters will be carefully planned and designed to meet the modeling requirements. An optimization technique called Simulated Annealing will be used to determine the process sensitive model parameters in model calibration.

3. Develop a model-based design methodology for the design of a SBR using the calibrated process model and apply the developed methodology to the full-scale SBR system to compare with the existing conventional design.

4. Extend the model-based design methodology into an optimization model to automate the design of a SBR. The optimization model will include both the capital cost and operational costs in the objective function.

5. Apply the optimization model to the existing full-scale 5.3 MGD (836 m^3/h) SBR system to produce the optimal design, and compare the cost savings with the existing design.

6. Perform sensitivity analysis to study the impact of possible variations in unit cost of sludge processing on the design parameters.

7. Develop an optimal strategy for operation of the existing SBR system for energy savings without making any major modifications to the current operating practice to provide some practical benefits to the existing system.

1.4   Originality of Research

The originality of this work lies in the holistic approach of producing optimal design information (volume of reactor, blower size, aeration time, length of time cycle) for a given set of constraints (process, influent and effluent constraints) taking into account both capital construction costs and operational costs. Although similar work has been

done in the continuous activated sludge process, cost optimization applied to design of sequencing batch reactors for municipal wastewater treatment has not been presented until now. The other salient features of this research are as follows:

1. A new model with fewer parameters has been developed for predicting the performance of the SBR employing kinetic expressions used in the continuous activated sludge process.

2. A new model-based design methodology has been proposed for the design of the SBR for removal of biodegradable organics and nitrogen substrates.

3. A calibration methodology using an optimization technique has also been presented for determination of unknown kinetic and stoichiometric parameters in the process model.

CHAPTER II

REVIEW OF LITERATURE

2.1   SBR Process Description

SBRs are a variation of the activated sludge process. They differ from continuous

activated sludge plants because they combine all of the treatment steps and processes into

a single basin, or tank, whereas continuous activated sludge plants rely on multiple

basins. According to U.S. EPA (1999), an SBR is no more than an activated sludge plant

that operates in time rather than space. The operation of a SBR is based on a fill-and-

draw principle, which consists of five steps - fill, react, settle, decant, and idle. These

steps as explained as follows can be altered for different operational applications (New

England Interstate Water Control Commission, 2005):

> *Fill*: During the fill phase, the basin receives influent wastewater. The influent
>
> brings food to the microbes in the activated sludge, creating an environment for
>
> biochemical reactions to take place. Mixing and aeration can be varied during the
>
> fill phase to create the following three different scenarios:
>
>> Static Fill - Under a static-fill scenario, there is no mixing or aeration
>>
>> while the influent wastewater is entering the tank. Static fill is used during

the initial start-up phase of plants that do not need to nitrify or denitrify, and during low flow periods to save power. Because the mixers and aerators remain off, this scenario has an energy-savings component. Mixed Fill - Under a mixed-fill scenario, mechanical mixers are active, but the aerators remain off. The mixing action produces a uniform blend of influent wastewater and biomass. Because there is no aeration, an anoxic condition is present, which promotes denitrification. Anaerobic conditions can also be achieved during the mixed-fill phase. Under anaerobic conditions the biomass undergoes a release of phosphorus. This release is reabsorbed by the biomass once aerobic conditions are reestablished.

Aerated Fill - Under an aerated fill scenario, both the aerators and the mechanical mixing units are activated. The contents of the basin are aerated to convert the anoxic or anaerobic zone over to an aerobic zone. No adjustments to the aerated fill cycle are needed to reduce organics and achieve nitrification. However, to achieve denitrification, it is necessary to switch the oxygen off to promote anoxic conditions for denitrification. By switching the oxygen on and off during this phase with the blowers, oxic and anoxic conditions are created, allowing for nitrification and denitrification.

React: During this phase, no wastewater enters the basin and the mechanical mixing and aeration units are on. Because there are no additional volumetric and

organic loadings, the rate of organic removal increases dramatically. Most of the carbonaceous BOD removal and further nitrification occurs in the react phase. The phosphorus released during mixed fill, plus some additional phosphorus, is taken up during the react phase.

Settle: During this phase, activated sludge is allowed to settle under quiescent conditions. No flow enters the basin and no aeration and mixing takes place. The activated sludge tends to settle as a flocculent mass, forming a distinctive interface with the clear supernatant.

Decant: During this phase, a decanter is used to remove the clear supernatant effluent. Once the settle phase is complete, a signal is sent to the decanter to initiate the opening of an effluent-discharge valve, and clear supernatant is discharged out as effluent. It is optimal that the decanted volume is the same as the volume that enters the basin during the fill phase, assuming the waste sludge volume withdrawn from the idle period is negligible compared to the influent volume entering the basin during the fill phase.

Idle: This step occurs between decant and fill phase of the next cycle. Idle time varies, based on the influent flow rate and the operating strategy. During this phase, excess sludge (concentrated solids) produced during the cycle is pumped out from the bottom of the SBR basin for further processing and disposal.
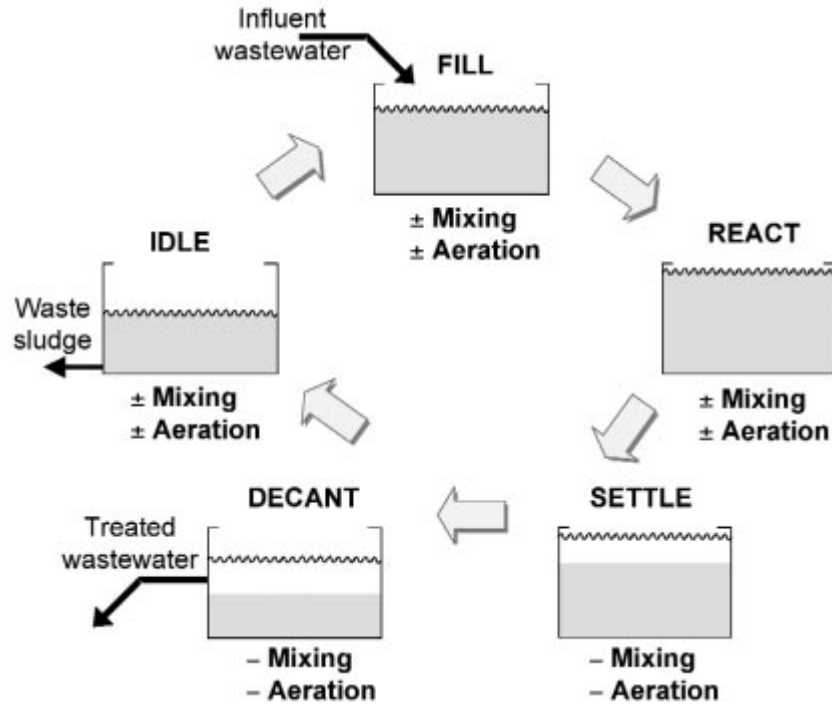
Figure 2-1: Schematic operation of SBR for one cycle (TREEO Center, 2000).

The treatment steps described above as shown in Figure 2-1 are repeated for every cycle when a fresh batch of wastewater is received. The typical duration of one cycle is 4 to 6 hours for typical municipal wastewater depending on the desired final effluent quality. One of the major benefits of an SBR is its operational flexibility i.e. different environments such as aerobic or anoxic or anaerobic conditions of different time duration can be incorporated in the batch cycle to accomplish desired removal of organics, nitrogen and/or phosphorus in a single basin. Other major advantages of the SBR process are as follows (Aqua-Aerobic Systems, 1986):

- Improved effluent quality over conventional activated sludge process,
- Elimination of separate clarifiers and sludge return pumps,

- Increased settling area,

- A perfectly quiescent settling environment,

- Demand-controlled energy consumption,

- Elimination of short-circuiting,

- A special ability to handle extremely high organic and hydraulic shock loads, and

- Capability to equalize flows and loads.

The establishment of stringent effluent limitations requiring nitrogen and/or phosphorus removal and the operational flexibilities that the SBR offers have led to its popularity in municipal wastewater treatment in recent years.

2.2   Review of Existing SBR Models

Mechanistic-based mathematical modeling of the wastewater treatment process provides insight into the understanding of the system and serves as a tool for designers to evaluate a wide range of system variables to optimize design to meet the system objectives at the lowest cost. Modeling also helps operators of the treatment facility to adjust or manipulate system control variables to realize the desired performance under a given scenario. Although a SBR is a variation of the activated-sludge process, unfortunately, modeling of the SBR process has not been studied as extensively as that of continuous flow conventional activated sludge systems. A continuous flow system operates under steady state conditions and hence changes in the substrate and biomass can be neglected. This simplifies the process and benefits the model building as it reduces the computational complexity by eliminating differential equations. In contrast, substrate

removal in the SBR system is carried out under unsteady state conditions in which substrate concentration and biomass concentration change significantly. This makes model building a complex situation and provides greater challenges to the researchers to mathematically model the system (Sun, 1996). However, the advent of computers and the availability of mathematical software and programs have made the task of solving complex differential equations much easier. Despite the modeling challenges, many researchers have developed mathematical models to describe the biological process involved in the SBR. These models range from simple to advanced, based on the components considered and hence the biochemical operations incorporated. Simple models have fewer parameters and are easier to apply. Sometimes, simple models may be too simplistic and may not describe the process precisely. Meanwhile advanced models are more complex and generally require software to solve the differential equations for dynamic simulations.

In their landmark paper, Lawrence and McCarty (1970) developed a mathematical model based on the fundamental principles of mass balance and biological kinetics. This has been the most commonly used model for activated sludge processes in wastewater treatment for the past many decades. Most of the earlier mathematical models developed specifically for an SBR (Irvine and Ritcher, 1978; Orhon et al., 1986; Droste, 1990) were based on the Lawrence and McCarty (1970) model. Sun (1996) in his study reported that the above models can either fit the fill period but fail for the period after the fill or vice versa. Theoretical results from all these models are the same, i.e., the predicted substrate concentration curve falls to zero after the fill stage as shown in Figure 2-2. The reasons

for the failure of these models are (1) they ignore the refractory organic materials in the influent and byproducts of substrate metabolism and sludge decay during the process, and (2) they classify all substrate as soluble and uniformly degradable (Sun, 1996). The above deficiencies were overcome in the intermediate models developed by
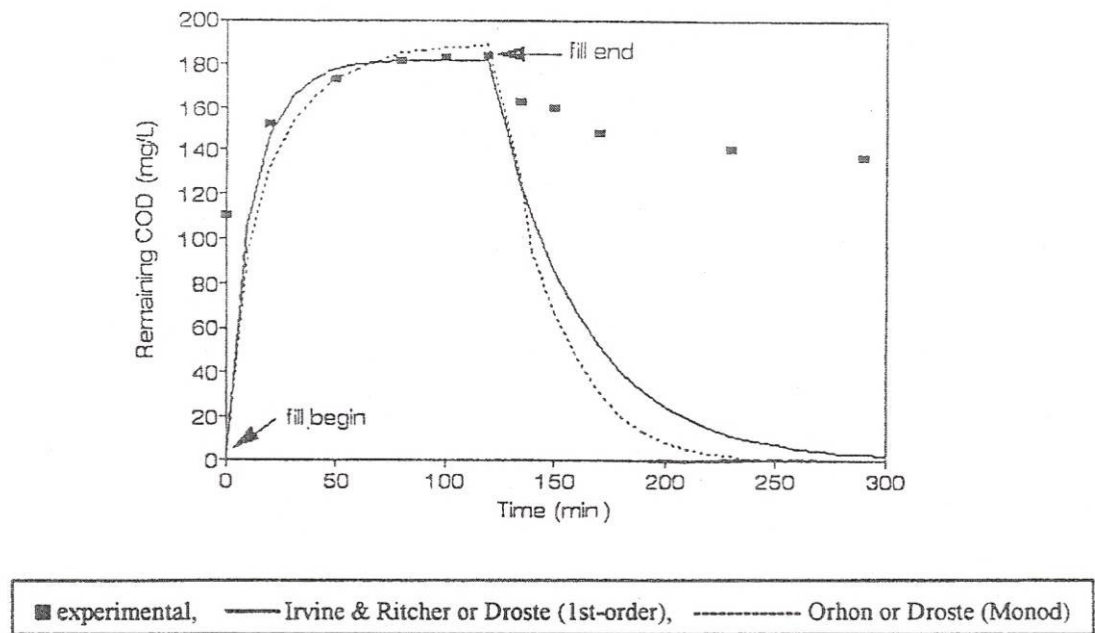
Figure 2-2: Comparison of results from different mathematical models (Sun, 1996).

Ibrahim and Abasaeed (1995) and Sun (1996) as these researchers classified carbonaceous substrates into three components such as soluble, difficult to degrade, and soluble inert. The limitations of both these models are that they were developed for completely aerobic systems and do not include the nitrification and denitrification processes.

In 1986, the International Association on Water Pollution Research and Control (IAWPRC) task group on mathematical modeling for design and operation of biological wastewater treatment systems advanced a mathematical model which allows prediction of organic matter degradation, nitrification, and denitrification in suspended sludge systems (Henze et al., 1987). This model commonly known as Activated Sludge Model No.1 (ASM1) was a major step forward in modeling activated sludge systems and is still considered as the "state-of-the-art" model. ASM1 model is expressed as a matrix system comprising 13 components, 8 process rates and 20 parameters and has been accepted by a wide range of scientists and engineers (Henze et al., 1987). Oles and Wilderer (1991) have applied this model successfully for process simulation of the SBR system. ASM1 does not include biological phosphorus removal. Model ASM1 has been updated to a higher version, ASM2, to include biological phosphorus removal (Henze et al., 1995). The ASM2 represents the state-of-the-art in the modeling of activated sludge processes with carbonaceous removal, nitrification and denitrification, and biological phosphorus removal. ASM2 model is described by 18 components (10 soluble and 8 particulate) and 17 biochemical reactions to portray the behavior of heterotrophs, autotrophs and phosphorus accumulating organisms (PAO) under aerobic, anoxic and anaerobic conditions. It allows simulation of nitrification, denitrification and biological phosphorus removal. Furumai et al. (1999) have successfully used ASM2 to address the long-term dynamic behavior of nutrients in a sequencing batch reactor (SBR) activated sludge process. These advanced models, ASM1 and ASM2, have been used with or without modification by several other researchers for process simulation and optimization (Katsogiannis et al., 1999; Coelho et al., 2000; Chang et al., 2000; Kim et al., 2001; Artan

et al., 2002; Hu et al., 2005). Another updated version of ASM1 called ASM3 has been introduced to correct a number of known defects in the original model (Gujer et al., 1998).

A common trait among the versions of these models is that each is high-dimensional and possesses a large number of kinetic and stoichiometric parameters. For example, ASM3 consists of 12 process rate equations involving seven dissolved and six particulate components, 21 kinetic parameters, and 13 stoichiometric and composition parameters. Though this level of model complexity is necessary to describe and relate dynamics over a wide range of operating conditions, it can present a significant computational encumbrance for performing simulations and analysis (Anderson et al., 2000). Some substrate components and model parameters are difficult to estimate (e.g., readily biodegradable substrate, slowly biodegradable substrate, and inert substrate), partly due to the limitation of available measurement techniques. Also, some processes listed in ASM2 (e.g., fermentation and hydrolysis) are hard to quantify, and for them deriving a rate equation is difficult, thus rendering calibration of the model more difficult (Zhao et al., 1997). Shahriari et al. (2006) evaluated different models ranging from simple to intermediate, and the International Water Association's complex activated sludge models (ASMs) to compare their ability to describe biomass growth and substrate removal in an activated sludge system. They reported that the intermediate model(s) is the practical choice for modeling considering the effort to determine parameter values, although the ASM models are better for research purposes because they provide more insight into the system components. The analysis of advanced models as discussed above clearly

indicates that the use of advanced models for an SBR is not a practical choice for design engineers considering the level of computational complexity and kinetic and stoichiometric parameters involved.

2.3   Review of SBR Design Methodologies

Design of a SBR system basically involves proper selection of reactor size, aeration system, cycle time duration, and sludge wasting rate for the given influent conditions to meet the desired effluent quality. Despite the growing popularity, a widely accepted approach to process analysis and modeling, a unified design basis, and even common terminology are still lacking for SBR systems. This situation is now regarded as the major obstacle hindering broader practical application of the SBR (Artran et al. 2001). The environmental engineers/design engineers are reluctant to use mathematical models in practice because some mathematical models have complex expressions and have parameters which are either not normally available or difficult to estimate. Instead they prefer to use the conventional design methods based on simple expressions developed from simple models for steady state conditions, for example the model developed by Lawrence and McCarty (1970) for activated sludge process. This trend in design practice is still continuing although SBRs has been widely used in municipal wastewater treatment for the last two decades. In United States, the process designs for SBRs are mostly performed by SBR equipment manufacturers. The process design calculations performed by many of the vendors are based on conceptual, time average models of the activated sludge process and do not take advantage of modern modeling tools. The

calculations performed by the manufacturer for the design engineers during the design process consist of the following steps (Geselbracht, 2007):

- Determine the average BOD_5 loading to the SBR system,

- Select a design F:M ratio,

- Determine the mass (lbs MLSS) required for the selected F:M ratio and average loading,

- Select a maximum MLSS concentration (average conditions) at the low water level (LWL) in the SBR (typically a value of 4500 mg/L is used),

- Calculate the total volume of the SBR at the LWL based on the maximum MLSS concentration and the solids inventory required under average loading conditions,

- Select the number of SBR reactors,

- Calculate the volume at the LWL in each SBR,

- Select the number of cycles per day,

- Calculate the maximum volume per decant based on the maximum daily flow and the number of cycles per day,

- Calculate the total volume per reactor by adding the volume at the LWL to the volume per decant,

- Calculate the total daily oxygen requirement based on average daily flow, BOD_5, and TKN concentrations using 1.25 lbs O_2/lb BOD_5 and 4.6 lbs O_2/lb TKN, adding in any denitrification credit (if appropriate),

- Divide the total daily oxygen requirement (under average loading) by the actual aeration time (only the feed-react and mix-react phases of the SBR cycle) to get an hourly oxygen supply rate required (field conditions), and

- Adjust the field oxygen transfer rate based on aeration system parameters to obtain the standard oxygen transfer rate required of the system.

The above design methodology is a conventional approach that has been in use for steady state systems for many decades, and it does not use a model-based design approach. The SBR process design procedures presented by Irvine and Ketchum (1989), Ketchum (1997) and International Water Association (IWA, 2001) scientific and technical report also follow a similar approach for municipal wastewater treatment. U.S. EPA (1999) suggested the engineers consult SBR manufacturers for recommended designs. However, as a guideline U.S. EPA (1999) has provided design values for key process parameters such as food to microorganism ratio (F/M ratio), mixed liquor suspended solids concentration (MLSS), hydraulic retention time (HRT), and the treatment cycle duration. The major disadvantage of this conventional method is that the designers usually prefer to use conservative design values for the above key design parameters which may result in an oversized system and increased costs. The conventional approach obviously does not take advantage of any mathematical modeling to produce a cost effective system.

Artran et al. (2001) developed a systematic rational approach for dimensioning of SBRs based on the principles of process stoichiometry. However, they considered the SBR as a steady state continuous flow system and derived the effective sludge age for the autotrophs and heterotrophs based on the time sequences of various cycle phases. They used this effective sludge age as a key parameter to estimate the size of the SBR to meet the effluent requirements. This method is a better approach than the conventional

approach as in this case the effective sludge age was not selected arbitrarily, but based on the process performance and effluent requirements. However it ignores the true condition of the SBR which is unsteady state in nature.

Most recent work reported in the literature on the design of SBRs is the design procedure developed by Abu-Ghunmi and Jamrah (2006) for treatment of textile wastewater in an SBR. This method uses the simple mass balance concept and experimental results to determine the reactor basin volume and cycle time. The experiments were carried out to monitor changes in substrate removal and MLSS concentration in the reactor for an extended period of time ranging up to 50 h. From the experimental observations, the reaction time and biomass concentrations were selected for the required removal efficiency to determine the reactor volume. It is a simple approach for a specific type of waste but requires experimental studies. It is clear from the above review of existing design methodologies that SBR designs lack model-based approaches primarily due to model complexity inherent in describing an unsteady-state process.

2.4   Review of Previous Work on SBR Optimization

Despite the multiple benefits and operational flexibility SBRs offer, in the United States most SBR installations are used for smaller wastewater treatment systems of less than two million gallons per day (US EPA, 1999). According to International Water Association scientific and technical report (IWA, 2001), there were about 1016 SBRs in operation for domestic wastewater treatment in North America. Out of 1016 SBRs, about 80 percent are small systems, sized for flow rates less than 4000 m^3/d (about 1.0 MGD).

According to Irvine et al. (1987), SBRs are perfectly suited for small wastewater flows (<10 MGD), while they perform satisfactorily even in large applications. From this author's experience, the SBR system is more economical than a conventional treatment process at smaller flows, but for larger installations it does not provide much cost benefit due to its larger reactor volume, higher level of sophistication in instrumentation and increased solids production rate. The possible reason for increased costs of SBR systems for larger installations is the use of conventional design methods for sizing of reactors and equipment, and use of conservative design values for key design parameters. Therefore, optimization in design methodology for SBRs is very important to make it more attractive to larger installations. Cost optimization in the design of SBRs was not studied as extensively as that of the continuous flow activated sludge process (Middleton and Lawrence, 1974; Grady, 1977; Craig et al., 1978; Tyteca and Smeers, 1981; Tang et al., 1987; Rivas et al., 2001; Doby et al., 2002; Espírito-Santo et al., 2005; Safaa et al., 2005). It is evident from the review of design methodologies that there has been little progress even now in switching from conventional design methods to model-based design approaches.

Most of the studies found in the literature on optimization of SBRs were aimed at the optimization of the operation of existing treatment units to determine optimal operational policies (Demuynck et al., 1994; Moreno, 1997; Katsogiannis et al., 1999; Coelho et al., 2000; Hvala et al., 2001; Artan et al., 2002). Demuynck et al. (1994), for instance, studied the optimization of an SBR for nitrogen removal. The authors used, for this purpose, the nitrification-denitrification biological excess phosphorus removal

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(NDBEPR) model of Wentzel et al. (1992) to optimize SBR time scheduling. The authors concluded that a sequence of short aerobic/anoxic phases performs better than the usual sequence of one aerobic phase followed by one anoxic phase. Moreno (1997) demonstrated how the optimization of cycle time can be used to increase the treatment capacity of the SBR for industrial wastewater. The author in this study used a simplified single substrate model based on Monod's kinetics. Katsogiannis et al. (1999) used an on-line adaptive optimization algorithm for identification of the batch cycle duration to minimize the cost of nitrification. On-line optimization models are black-box models based on input and output data and ignore physical, chemical or biological process knowledge. Coelho et al. (2000) developed an optimization algorithm to minimize the total batch time to maximize the reactor productivity. The decision variables in this optimization model were feed profile, fill time, and aeration time, and the model used in this study was ASM1. Work carried out by Hvala et al. (2001) and Artran et al. (2002) focused on the optimization of an SBR to determine optimal filling strategies and time sequences (such as aerobic, anoxic and anaerobic sequences). Recent work by Alhumaizi and Ajbar (2006) developed a design equation for cyclically operated biological reactors based on an unstructured first-order kinetic model that was used for optimization of volumetric efficiency and minimization of end-of-cycle pollutant concentration. However, the model used in this study is a first-order kinetic model for single substrate.

The review of existing SBR optimization models reveals that most of the work was focused on process optimization to determine the optimal operational policies to reduce batch time and improve treatment efficiency. Although these models will reduce the

operational costs to a certain extent due to minimal batch time, excess sludge wasting, a key decision variable having major cost implication, has been ignored. The impact of wasting excess sludge and its processing cost on the operational policy of the activated sludge process and the importance of including it in the optimization model was demonstrated in the work of Velmurugan and Veenstra (2007). Process optimization models are beneficial to only existing SBR systems, and any model ranging from simple to advanced can be used because the system configuration (such as reactor volume, equipment sizing) is already known. Design optimization models differ from process optimization models in the following aspects:

- Reactor volume and equipment sizes are to be determined and not known a priori. Analytical solution for determination of reactor volume will be a complex procedure for multiple substrates.

- The optimization model shall include both the capital costs (construction) and operational costs (operation).

- System shall perform and produce the desired effluent quality under varying conditions of influent flow and influent characteristics.

CHAPTER III

PROCESS MODEL DEVELOPMENT

3.1   Introduction

The activated sludge process is one of the most commonly used biological processes for

removing soluble and colloidal organic and nitrogenous constituents present in

wastewater. The activated sludge process is very flexible and can be adapted to almost

any type of biological waste treatment problem. SBRs are a variation of the activated-

sludge process. According to the U.S. EPA (1999), an SBR is no more than an activated-

sludge plant that operates in time rather than space. SBR basically operates on a fill-and-

draw basis. The reactor is filled during a discrete period of time and then operated as a

batch reactor. After desired treatment, the mixed liquor is allowed to settle and the

clarified supernatant is then drawn from the tank and discharged as a treated effluent

(Irvine and Ketchum, 1989). The excess biomass produced in the tank is wasted as

sludge for further treatment in the solids processing facility.

The biological process involved in the SBR can be modeled using activated sludge

process models by incorporating appropriate biochemical operations for different

environmental conditions such as anoxic, aerobic and anaerobic involved in the SBR

process. Several researchers have developed a number of models for studying and

understanding the activated sludge process. As discussed in Chapter II, many researchers have adopted the activated sludge process models for modeling of the SBR process. The objective of this study is to develop a new mathematical model specifically for SBRs for removal of multiple biodegradable organic substrates, ammonia nitrogen and nitrate nitrogen that would require fewer parameters than advanced models but at the same time describe all the processes involved without over-simplification.

3.2   Modeling Assumptions

The following assumptions were made in the model development to simplify the physical process to meet the given requirements:

1. This mathematical model has been developed only for removal of organic and nitrogen substrates for which SBRs are typically designed and biological phosphorus removal has not been considered. Therefore, the substrates that are considered are particulate and soluble organics, ammonia nitrogen and nitrate nitrogen.

2. The environmental conditions required for removal of organic and nitrogen substrates are aerobic and anoxic conditions. Therefore, the different phases of the reaction that are considered for the removal of these substrates are anoxic fill, aerobic fill, and aerobic react.

3. The reactor is completely mixed during all phases of reaction except settle, decant, and idle.

4. Typically no removal of substrates takes place during settle, decant and idle periods and hence these time periods are ignored in the model.

5. The kinetic expressions for removal rates for oxidation of dissolved organic matter (v_S), hydrolysis of particulate organic matter (v_P), nitrification of ammonia nitrogen (v_N), and denitrification of nitrate nitrogen (v_D) are adopted from Esposito et al. (2003) as follows:

$$v_S = v_{S,\max}\left(\frac{S}{K_S + S}\right) \tag{3.1}$$

$$v_P = v_{P,\max}\left(\frac{(S_P / X_H)}{(K_P + (S_P / X_H))}\right) \tag{3.2}$$

$$v_N = v_{N,\max}\left(\frac{S_{NH}}{K_N + S_{NH}}\right)\left(\frac{O_2}{K_{O2} + O_2}\right) \tag{3.3}$$

$$v_D = v_{D,\max}\left(\frac{S_{NO}}{K_D + S_{NO}}\right)\left(\frac{S}{K_{S,D} + S}\right) \tag{3.4}$$

Where

$v_{S,\max}$ = Maximum removal rate of dissolved organics (mg BOD$_5$/mg VSS-h),

S = Soluble BOD$_5$ (mg/L),

K_S = Half saturation coefficient for dissolved organics (mg BOD$_5$/L),

$v_{P,\max}$ = Max. hydrolysis rate of particulate organics (mg BOD_5/mg VSS-h),

S_P = Particulate BOD_5 (mg/L),

X_H = Active heterotrophic biomass (mg/L),

K_P = Half saturation coefficient for particulate organics (mg BOD_5/mg VSS),

$v_{N,\max}$ = Maximum removal rate of ammonia nitrogen (mg N-NH_4/mg VSS-h),

S_{NH} = Ammonia nitrogen (mg/L),

K_N = Half saturation coefficient for ammonia nitrogen (mg N-NH_4/L),

O_2 = Dissolved oxygen concentration (mg/L),

K_{O2} = Half saturation coefficient for dissolved oxygen (mg/L),

$v_{D,\max}$ = Maximum removal rate of nitrate nitrogen (mg N-NO_3/mg VSS-h),

S_{NO} = Nitrate nitrogen (mg/L),

K_D = Half saturation coefficient for nitrate nitrogen (mg N-NO_3/L), and

$K_{S,D}$ = Biodegradable organics half saturation coefficient for denitrification

process (mg BOD_5/L).

The above kinetic expressions have been used for nitrification and denitrification processes in continuous flow systems.

6. The kinetic parameters for maximum removal rates for oxidation of dissolved organic matter ($v_{S,\max}$), hydrolysis of particulate organic matter ($v_{P,\max}$), nitrification of ammonia nitrogen ($v_{N,\max}$), and denitrification of nitrate nitrogen

$(v_{D,\max})$ do not vary from one time phase of reaction to other i.e. anoxic fill or

aerated fill or react.

3.3   Mathematical Formulation

Based on the above assumptions and removal rate expressions, mass balance equations

have been developed as follows for all substrate concentrations and microbial cells

(autotrophic and heterotrophic) for each phase of reaction such as anoxic fill, aerobic fill,

and aerobic react as follows:

For anoxic filling: During anoxic filling, reactor basin receives wastewater and the

contents of the reactor are mixed but the aerator remains turned off. The volume of the

reactor varies based on the flow rate. Since aeration devices are turned off, the anoxic

condition prevails in the reactor. During this phase, removal of soluble organic substrate,

hydrolysis of particulate organic substrates and removal of nitrate nitrogen occurs. These

removal processes are carried out by the heterotrophic bacteria using nitrate nitrogen as

electron acceptor and their concentration increases during the phase. The following mass

balance equations provide the concentration of various substrates and heterotrophic

biomass at any time during the anoxic phase:

Time interval: $0 < t < t_{anf}$

$$V = V_0 + \int_0^{t_{anf}} q\, dt \tag{3.5}$$

$$\frac{dS}{dt} = \frac{q}{V}\left(S_{in} - S\right) - \omega v_D X_{H,D} + v_p X_{H,D} \tag{3.6}$$

$$\frac{dS_P}{dt} = \frac{q}{V}\left(S_{P,in} - S_P\right) - v_p X_{H,D} \qquad (3.7)$$

$$\frac{dS_{NH}}{dt} = \frac{q}{V}\left(S_{NH,in} - S_{NH}\right) - \varepsilon\omega v_D X_{H,D} \qquad (3.8)$$

$$\frac{dS_{NO}}{dt} = \frac{q}{V}\left(S_{NO,in} - S_{NO}\right) - v_D X_{H,D} \qquad (3.9)$$

$$\frac{dX_{H,D}}{dt} = \frac{q}{V}\left(X_{H,D,in} - X_{H,D}\right) + v_D Y_D X_{H,D} \qquad (3.10)$$

Where

t = Time (h),

t_{anf} = Time anoxic fill ends (anoxic fill time) (h),

V = Volume at any time, t (m^3),

V_O = Initial volume (m^3),

q = Influent flow rate (m^3/h),

ω = Unit mass of biodegradable organics removed per unit mass of nitrate

nitrogen reduced (mg BOD$_5$/mg N-NO$_3$),

ε = Nitrogen consumption for synthesis requirement (mg N-NH$_4$/mg BOD$_5$),

Y_D = Yield of heterotrophic biomass for denitrification (mg VSS/ mg N-NO$_3$),

and Subscripts - *in* and *D* refer to influent and denitrification, respectively.

For aerobic filling: During aerobic filling, reactor basin continues to receive wastewater

but the mixers and aeration devices are on. Since aeration devices are turned on, oxygen

is provided in the basin and the anoxic condition changes to aerobic condition. Both

heterotrophic and autotrophic growth of bacteria occurs in the reactor basin. The removal

of soluble organic substrates, hydrolysis of particulate organic substrates and removal of ammonia are carried out by the heterotrophic and autotrophic bacteria using oxygen as electron acceptor. As a result, the removal of nitrate nitrogen stops during this phase. The concentration of various substrates, heterotrophic biomass and autotrophic biomass at any time during the aerobic phase is expressed as follows:

Time interval: $t_{anf} < t < t_{aef}$

$$V = V_0 + qt_{anf} + \int_{t_{anf}}^{t_{aef}} q\, dt \qquad (3.11)$$

$$\frac{dS}{dt} = \frac{q}{V}(S_{in} - S) - v_S X_{H,N} + v_p X_{H,N} \qquad (3.12)$$

$$\frac{dS_P}{dt} = \frac{q}{V}(S_{P,in} - S_P) - v_p X_{H,N} \qquad (3.13)$$

$$\frac{dS_{NH}}{dt} = \frac{q}{V}(S_{NH,in} - S_{NH}) - v_N X_{Aut,N} - \varepsilon \omega v_S X_{H,N} \qquad (3.14)$$

$$\frac{dS_{NO}}{dt} = \frac{q}{V}(S_{NO,in} - S_{NO}) + v_N X_{Aut,N} \qquad (3.15)$$

$$\frac{dX_{H,N}}{dt} = \frac{q}{V}(X_{H,N,in} - X_{H,N}) + v_S Y_H X_{H,N} \qquad (3.16)$$

$$\frac{dX_{Aut,N}}{dt} = \frac{q}{V}(X_{Aut,N,in} - X_{Aut,N}) + v_N Y_A X_{Aut,N} \qquad (3.17)$$

Where

t_{aef} = Time aerobic fill ends (anoxic fill + aerated fill) (h),

X_{Aut} = Active autotrophic biomass (mg/L),

Y_A = Yield of autotrophic biomass (mg VSS/ mg N-NH₄), and

Subscript - N refers to nitrification.

For react phase: During react phase, wastewater flows to the reactor basin are stopped and the wastewater volume in the tank remains constant. The environmental conditions that prevail in this phase are same as the aerobic filling phase except the wastewater flows are shut off.   The removal mechanisms that take place during this phase and the concentration of heterotrophic biomass and autotrophic biomass at any time during the react phase are expressed as follows:

Time interval: $t_{aef} < t < t_r$

$$V = V_0 + qt_{anf} + q\left(t_{aef} - t_{anf}\right) \qquad\qquad (3.18)$$

$$\frac{dS}{dt} = -v_S X_{H,N} + v_p X_{H,N} \qquad\qquad (3.19)$$

$$\frac{dS_P}{dt} = -v_p X_{H,N} \qquad\qquad (3.20)$$

$$\frac{dS_{NH}}{dt} = -v_N X_{Aut,N} - \varepsilon v_S X_{H,N} \qquad\qquad (3.21)$$

$$\frac{dS_{NO}}{dt} = v_N X_{Aut,N} \qquad\qquad (3.22)$$

$$\frac{dX_{H,N}}{dt} = v_S Y_H X_{H,N} \qquad\qquad (3.23)$$

$$\frac{dX_{Aut,N}}{dt} = v_N Y_A X_{Aut,N} \qquad\qquad (3.24)$$

Where

t_r = Time react ends (anoxic fill + aerated fill + react) (h).

3.4 Model Parameters

The process model equations (3.5) through (3.24) have 15 kinetic and stoichiometric parameters. Some parameters are process dependent and vary depending on the environmental conditions specific to the wastewater treatment plant. Some of the parameters are mostly typical for the given type of waste at a given temperature. In general, the kinetic parameters are temperature dependent and must be adjusted for temperature variation. Some of the temperature dependent kinetic and stochiometric parameters such as maximum removal rates for oxidation of dissolved organic matter ($v_{S,\max}$), hydrolysis of particulate organic matter ($v_{P,\max}$), nitrification of ammonia nitrogen ($v_{N,\max}$), denitrification of nitrate nitrogen ($v_{D,\max}$) and half saturation coefficient for particulate organics (K_P) are adjusted for temperature variation from their values at 20 $^{\circ}$C using the following expressions:

$$v_{S,\max} = v_{S,\max,20\,^{\circ}C}\, 1.02^{(T-20)} \tag{3.25}$$

$$v_{P,\max} = v_{P,\max,20\,^{\circ}C}\, 1.04^{(T-20)} \tag{3.26}$$

$$K_P = K_{P,20\,^{\circ}C}\, 0.898^{(T-20)} \tag{3.27}$$

$$v_{N,\max} = v_{N,\max,20\,^{\circ}C}\, 1.12^{(T-20)} \tag{3.28}$$

$$v_{D,\max} = v_{D,\max,20\,^{\circ}C}\, 1.12^{(T-20)} \tag{3.29}$$

Where

T = Temperature in degrees Celsius ($^{\circ}$C).

3.5   Solution Technique

Once the model parameters are defined, the set of non-linear first-order differential

equations (3.5) through (3.24) formulated for the process model can be solved

numerically for determination of variables such as volume, different substrate

concentrations and microbial concentrations at any time with known initial conditions.

The Euler's method as described in Sewell (1988) has been employed in this study to

determine substrates and microbial cell concentrations for given influent conditions and

time periods of different reaction phases such as anoxic fill, aerated fill, and react phase.

There are several numerical methods available for solving non-linear first-order

differential equations. However, the Euler's method described in Sewell (1988) has been

used in this work because this method is computationally faster than other methods. The

optimization model which is developed in Chapter VI requires the results of the process

model several times during its iterations for finding the optimal design decision variables.

Therefore, a computationally faster solution technique to the process model is very

important. Euler's method for solving first-order initial value ordinary differential

equation is simple and is expressed as follows:

If the first-order initial value ordinary differential equation to be solved is of the form:

$$\frac{dx}{dt} = f(t,x) \qquad\qquad x(0) = x_0 \qquad\qquad\qquad (3.30)$$

Then, the solution takes the following form:

$$x(t+h) = x(t) + h\,f(t,x) \qquad\qquad\qquad (3.31)$$

Where t is the independent variable and h is the time increment. The solution to the problem starts with the given initial value and moved forward one (time increment) step at a time. In this problem, the time step of 0.0003 hr was used for h. Based on the above solution technique, a computer program has been developed in MATLAB for solving the process model equations. The MATLAB codes with explanations are listed in Appendix A.

CHAPTER IV

EXPERIMENTAL VALIDATION OF THE MODEL

4.1 Introduction

The process model developed in Chapter III is to be calibrated and validated prior to its

use in the design optimization model to obtain meaningful results. The objective of this

Chapter is to determine the unknown model parameters through model calibration and

test the validity of the model. The data for the model calibration were obtained from the

operation of a full-scale sequencing batch reactor system located at the City of

Tahlequah, Oklahoma by a carefully designed sampling program. A calibration

methodology using an optimization technique called Simulated Annealing has been

employed in this study to determine the unknown model parameters. The calibrated

model has been verified using the separate set of sampling data from the same SBR

system at the City of Tahlequah to evaluate model performance and its applicability. The

following sections of this chapter describe the operation of the full-scale SBR system,

sampling program, data collection, model calibration, model verification and discussion

of results on model performance and its applicability.

4.2 City of Tahlequah SBR System

The SBR treatment system at the City of Tahlequah, Oklahoma was designed to treat

836 m^3/h (5.3 MGD) of domestic sewage. This is one of the earliest and largest SBR

wastewater treatment systems in the State of Oklahoma, and built in 1992. The liquid

treatment portion of this treatment facility includes three grit removal units, four SBR

basins, an online chemical feeding system, two filters, two UV disinfection units and a

flow measurement structure. The solids treatment portion of this treatment facility

includes a gravity belt thickener, two aerobic digesters and a belt filter press. The

process flow diagram for the liquid treatment process which is relevant to this study is

shown in Figure 4-1. The raw wastewater entering the treatment facility is screened

initially before it passes through the grit removal units where heavy inorganic particles

are removed. The de-gritted wastewater is pumped into the SBR basins for removal of

biological oxygen demand and ammonia nitrogen. Each SBR basin is equipped with a

single center-mounted floating mixer for mixing reactor contents and eight retrievable

bubble diffusers for providing oxygen to the SBR basin. The SBR basins are the heart of

the treatment process which is a focus of this study.
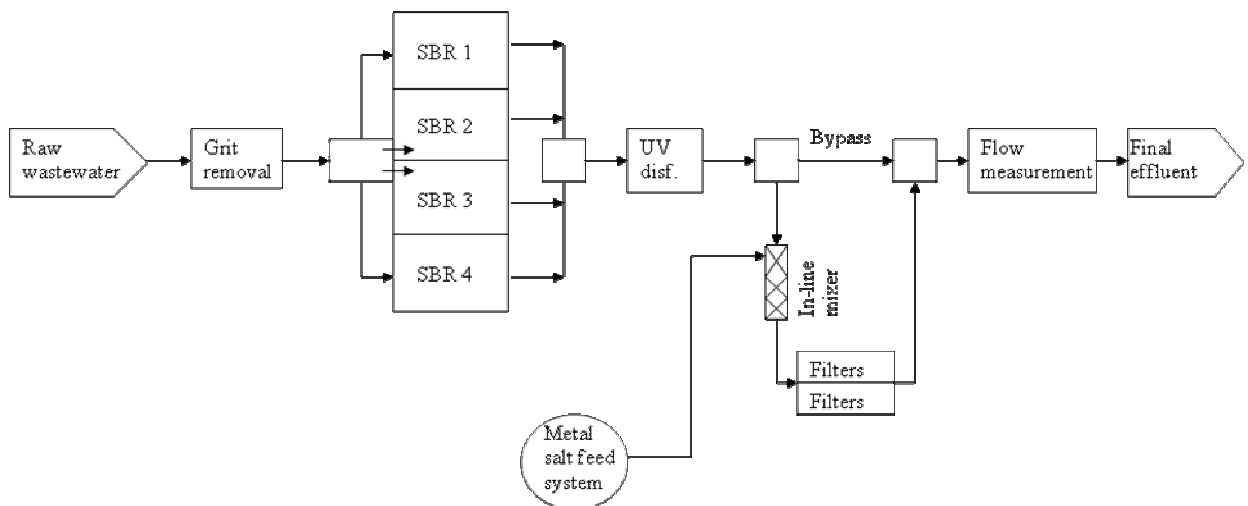
Figure 4-1: Process flow diagram for the City of Tahlequah SBR system.

The basis of design and operation of the SBR as designed and found in the Operation and Maintenance manual of the plant is shown as follows:

Number of Basins	4
Dimensions	
Length, ft	116.0
Width, ft	116.0
Sidewater Depth (Minimum), ft	9.0
Sidewater Depth (Maximum), ft	15.7
Aeration Equipment	
Type	Fine bubble
Diffuser Assemblies/Basin	8
Disc Diffusers/Assembly	40
Mixing Equipment	
Type	Downdraft
Number/Basin	1
Horsepower	50
Decanter Assemblies	
Number/Basin	2
Decant Rate, MGD	6
Blowers	
Number	2 Working
	1 Standby

Maximum capacity at 14.7 psia and 68 $^{\circ}$F, scfm   2500

Discharge pressure at 700 ft above MSL and 100 $^{\circ}$F, psig   7.9

Horsepower   125

Operation parameters

Retention Time @ Design Flow, hrs   14

F/M Ratio, lb BOD_5/lb MLSS   0.05 – 0.10

MLSS, mg/L   4000

Time cycle

Phase	Phase Duration	Cumulative Cycle Time
Static Fill	0 minutes	0 minutes
Mix Fill	45 minutes	45 minutes
React Fill	35 minutes	80 minutes
React	115 minutes	195 minutes
Settle	45 minutes	240 minutes
Decant/Idle/Waste	120 minutes	360 minutes

The decant (treated effluent) from the SBR basin is passed thorough a physico-chemical treatment system consisting of online alum addition, in-line mixing, and filtration units for removal of phosphorus. The filtration units are of dual media type filters which consist of 12 inch thickness of support gravel, 12 inch thickness of sand and 18 inch thickness of anthracite. The filter sand is placed on top of the gravel and the anthracite is placed on top of the sand. The treated final effluent is disinfected in UV disinfection units for removal of pathogens before it is discharged to Tahlequah Creek.

4.3 Sampling Program

The discharge of treated effluent from this treatment plant is regulated by an Oklahoma Discharge Elimination System (OPDES) permit. Table 4-1 below shows the OPDES effluent discharge limitations for various parameters. The effluent limits shown in Table 4-1 are the average for a month for composite samples at the plant outfall. Hence, the normal operation of the treatment plant for regulatory compliance requires only the analysis of the composite effluent for 5-day biochemical oxygen demand (BOD_5), ammonia nitrogen (NH_4-N), and total phosphorus at the plant outfall, and these data would not be adequate for calibration and verification of the model. Therefore, an

Table 4-1: Effluent limits for existing SBR system at the City of Tahlequah

Parameter	Effluent limits
5-Day biological oxygen demand	< 7.6 mg/L
Suspended solids	< 11.0 mg/L
Ammonia nitrogen	< 1.5 mg/L
Phosphorus	< 1.0 mg/L
Dissolved oxygen	> 6.8 mg/L
pH	6.5 – 9.0

intensive sampling program was conducted to collect and analyze samples for the purpose of model calibration and verification. One of the four SBR basins identified as SBR2 was selected for sampling, and samples from this basin were collected for a total of six cycles at 25 minute intervals for each cycle duration of 3 hours (excludes settle,

decant and idle periods). The first sampling episode consisting of three cycles was collected on May 14 – 15, 2008 and second sampling episode of three cycles was collected on June 18 – 19, 2008. In order to determine the initial conditions of the SBR, a grab sample was collected from basin SBR2 for each cycle before filling began, and the sample was analyzed for soluble chemical oxygen demand (COD) or soluble biochemical oxygen demand (BOD_5), ammonia nitrogen (NH_4-N), nitrate nitrogen (NO_3-N), and mixed liquor volatile suspended solids (MLVSS). After the initial sample, grab samples were collected from SBR2 at 25 minute time intervals until the end of reaction phase, and the collected samples were analyzed for soluble COD or soluble BOD_5, NH_4-N and NO_3-N. Temperature, pH, dissolved oxygen (DO) and MLVSS were measured for three samples, one each during the initial, middle, and end of the cycle (anoxic fill + react phase). For each cycle, at the end of react phase (before settle begins) a grab sample was collected for measuring sludge volume index (SVI). During filling time, a composite sample was collected at the outlet of the grit chamber to characterize the influent wastewater for total COD or total BOD_5, soluble COD or soluble BOD_5, and total kjeldahl nitrogen (TKN). The basic operational parameters and influent characteristics for the first and second sampling episodes are summarized in Tables 4-2 and 4-3, respectively. The summary of all data collected for all six cycles of sampling episodes 1 and 2 and the raw analysis data from the laboratory are provided in Appendix B. For the second sampling episode, soluble COD and total COD were measured instead of soluble BOD_5 and total BOD_5 for better accuracy. From analysis of previous samples collected during the sampling episode 1, it was found that the ratio of soluble COD to soluble BOD_5 and total COD to total BOD_5 were 2.25 and 1.30, respectively. COD parameter

Table 4-2: Reactor data, influent characteristics, initial conditions and operational parameters for sampling episode 1

			Cycle 1 May 14 8:50 am	Cycle 2 May 14 2:50 pm	Cycle 3 May 15 8:50 am
Reactor data					
Initial volume	V_O	m^3	3768	3768	3764
Total volume after fill	V_T	m^3	4687	4610	4847
Average influent characteristics					
Influent flow	q	m^3/h	613	561	722
Total BOD$_5$		mg/L	175	173	112
Soluble BOD$_5$	S_{in}	mg/L	38.1	57.5	23.2
Particulate BOD$_5$	$S_{P,in}$	mg/L	136.9	115.5	88.2
TKN	$S_{NH,in}$	mg/L	26.2	24.5	13.6
pH			6.9	6.8	6.9
Temperature		^{o}C	19.5	19.6	19.6
SBR initial conditions					
Soluble BOD$_5$	S	mg/L	14.3	15.50	5.80
NH$_4$-N	S_{NH}	mg/L	1.84	4.78	2.72
NO$_3$-N	S_{NO}	mg/L	0.11	0.10	0.13
Active autotrophic biomass	X_{Aut}	mg/L	314	280	306
Active heterotrophic biomass	X_H	mg/L	1256	1114	1224
Operational parameters					
Anoxic fill	t_{anf}	h	1.50	1.50	1.50
Anoxic fill + Aerated fill	t_{aef}	h	1.50	1.50	1.50
Anoxic fill + React	t_r	h	3.00	3.00	3.00
DO concentration	O_2	mg/L	0.87	0.73	0.50

Table 4-3: Reactor data, influent characteristics, initial conditions and operational parameters for sampling episode 2

			Cycle 1 June 18 8:50 am	Cycle 2 June 18 2:50 pm	Cycle 3 June 19 8:50 am
Reactor data					
Initial volume	V_O	m³	3791	3780	3784
Total volume after fill	V_T	m³	4934	4896	4827
Average influent characteristics					
Influent flow	q	m³/h	762	744	695
Total BOD$_5$		mg/L	164	115	91.5
Soluble BOD$_5$	S_{in}	mg/L	16.7	27.6	17.8
Particulate BOD$_5$	$S_{P,in}$	mg/L	147.3	87.4	73.7
TKN	$S_{NH,in}$	mg/L	15.0*	13.1	16.9
pH			7.0	6.9	6.9
Temperature		°C	21.6	22.5	22.4
SBR initial conditions					
Soluble BOD$_5$	S	mg/L	6.2	7.6	6.8
NH$_4$-N	S_{NH}	mg/L	1.03**	1.79	0.95
NO$_3$-N	S_{NO}	mg/L	0.93	0.87	0.77
Active autotrophic biomass	X_{Aut}	mg/L	180	184	192
Active heterotrophic biomass	X_H	mg/L	726	737	769
Operational parameters					
Anoxic fill	t_{anf}	h	1.50	1.50	1.50
Anoxic fill+Aerated fill	t_{aef}	h	1.50	1.50	1.50
Anoxic fill+Aerated fill+React	t_r	h	3.00	3.00	3.00
DO concentration	O_2	mg/L	1.60	2.00	1.60

* Data not available. Assumed to be average influent TKN of cycles 2 and 3.

** Data not available. Initial NH$_4$-N was assumed to be 80% value of NH$_4$-N data collected after 25 minutes into the cycle. This assumption was based on the trend that was observed for cycles 2 and 3.

were not normally measured at this plant. Therefore, there were no historical data at the plant to confirm the long term correlation for ratio of COD to BOD_5. However, ratios for soluble COD to soluble BOD_5 and total COD to total BOD_5 determined as above from the previous samples fall in the typical range of 1.25 to 2.50 reported in the literature for municipal wastewater (Metcalf and Eddy, 2003). Therefore, the measured soluble and total COD were converted to soluble BOD_5 and total BOD_5 before their use in the model. The sample data from the first two cycles of each episode were used for model calibration, and the third cycle of each episode was used for model validation.

4.4   Calibration Technique

The process model developed in this study has 15 kinetic and stoichiometric parameters, which shall be defined to study the effectiveness or usefulness of the model in predicting the process performance of the plant. For some parameters, default values recommended in the literature can be used as they are sufficiently accurate. However, some parameters are very sensitive to the process and can have a wide range of values depending on the process operating conditions. The preliminary process simulation has revealed that the process model is sensitive to five kinetic and stoichiometric parameters: maximum removal rate of dissolved organics ($v_{S,max}$), half saturation coefficient for dissolved organics (K_S), maximum hydrolysis rate of particulate organics ($v_{P,max}$), maximum removal rate of ammonia nitrogen ($v_{N,max}$) and maximum removal rate of nitrate nitrogen ($v_{D,max}$). These five parameters which showed marked influence on model response were determined by model calibration.

Model calibration is the process in which unknown parameters are adjusted to minimize the difference between the predicted and observed results. Model calibration can be done manually by adjusting the parameter values. This can be most tedious and difficult particularly for a large number of parameters. Therefore, an optimization algorithm has been employed in this study to determine these parameters. The optimization algorithm will determine the optimum values for the selected parameters by minimizing the objective function (f), which is the weighted relative least-square error between the predicted and observed results as shown in the following expression.

$$f = \sum_{j=1}^{k} \sum_{i=1}^{l} \left(w_S \left(\frac{S^* - S}{S^*} \right)_i^2 + w_{NH} \left(\frac{S_{NH}^* - S_{NH}}{S_{NH}^*} \right)_i^2 + w_{NO} \left(\frac{S_{NO}^* - S_{NO}}{S_{NO}^*} \right)_i^2 \right) \tag{4.1}$$

Where i is the ith observation, l is the number of observations, j is the jth cycle, k is the number of cycles from which the data are used for the calibration, w_S is the weight factor for BOD_5 or COD data, w_{NH} is the weight factor for NH_4-N data, w_{NO} is the weight factor for NO_3-N data and * is the predicted value from the model.

Using the relative least-square error as an objective function in place of the more commonly used least-square error is very appropriate for activated sludge processes where the magnitudes of observations differ significantly (Yuan et al., 1993). For example, the range for soluble BOD_5 observed in the SBR would be different from the range observed for ammonia nitrogen or nitrate nitrogen. Using the weight factor is important where the collected data set has different ranges of measurement errors. The optimization algorithm used in this study is Simulated Annealing (SA), a meta-heuristics method. It is a random search technique which exploits an analogy between the way

metal cools and freezes into a minimum energy crystalline structure (the annealing process) and searches for a minimum in a more general system. The algorithm employs a random search technique which not only accepts changes that decrease the objective function but also some changes that increase it with a specified probability. This feature allows the optimization to move towards the global minimum and avoids getting trapped in local minima. This optimization method is easy to implement in computers as it does not use the first derivative information of the objective function. More details on the development of the Simulated Annealing (SA) algorithm and its implementation can be found elsewhere (Bohachevsky et al., 1986; Ingber, 1993; Brooks and Morgan, 1995; Rao, 1996; Edgar et al., 2001). A flow chart describing the calibration methodology and a step by step procedures for using this optimization technique for determining the selected kinetic and stochiometric parameters is shown in Appendix C. A computer program has been written in MATLAB for the implementation of this optimization algorithm to estimate the optimum (calibrated) values for the selected kinetic and stochiometric parameters. The MATLAB codes with explanations are listed in Appendix C.

4.5   Model Calibration and Verification

4.5.1   Sampling Episode 1 (May 14 – 15, 2008)

The data collected from SBR2 on May 14, 2008 for two cycles, which began at 8:50 am and 2:50 pm, respectively, were used for model calibration. Out of 15 kinetic and stoichiometric parameters that appear in the process model, 10 parameters were assigned typical values recommended in the literature as shown in Table 4-4. The remaining 5

parameters: $v_{S,\max}$, K_S, $v_{P,\max}$, $v_{N,\max}$ and $v_{D,\max}$ (denoted as parameter vectors in the

calibration algorithm) which are sensitive to the process were determined through the

model calibration. The weight factors of 1, 2 and 0 were assigned to BOD$_5$ data, NH$_4$-N

Table 4-4: Typical and calibrated kinetic and stoichiometric parameters for sampling episode 1

Parameter	Unit	Value*
Typical values used in the model (Esposito et al., 2003)		
Y_H	mg VSS/mg BOD$_5$	0.55
K_P	mg BOD$_5$/mg VSS	0.095
K_N	mg N-NH$_4$/L	1
K_{O2}	mg/L	1
Y_A	mg VSS/ mg N-NH$_4$	0.15
ε	mg N-NH$_4$/mg BOD$_5$	0.05
K_D	mg N-NO$_3$/L	0.1
$K_{S,D}$	mg BOD$_5$/L	0.1
Y_D	mg VSS/ mg N-NO$_3$	0.512
ω	mg BOD$_5$/mg N-NO$_3$	4
Estimated values from calibration		
$v_{S,\max}$	mg BOD$_5$/mg VSS-h	0.5716
K_S	mg BOD$_5$/L	606.90
$v_{P,\max}$	mg BOD$_5$/mg VSS-h	0.0182
$v_{N,\max}$	mg N-NH$_4$/mg VSS-h	0.0054
$v_{D,\max}$	mg N-NO$_3$/mg VSS-h	0.0708

* Values are for 20° C. The measured temperature of the wastewater was close to 20°C (See Table 4-2), and hence they were used as such without applying temperature correction factors.

data, and NO$_3$-N data, respectively. The different weight factors were assigned to offset

the likely measurement error in the data set. The weight factor of zero was assigned to

NO$_3$-N data for two reasons: 1) the minimum and maximum values of NO$_3$-N in the data

set used for calibration were 0.05 mg/L and 0.11 mg/L, respectively. This is very close to

zero, would produce a significantly high value for the objective function, and would

ignore the importance of BOD$_5$ and NH$_4$-N data, and 2) the treatment plant has no

effluent limitation for NO$_3$-N. Therefore, it is more appropriate to use the weight factor

of zero for NO$_3$-N data. During model calibration, the calibration algorithm was run

several times with randomly generated initial guesses for the parameter vector. The

calibration algorithm has converged at different objective function values, indicating that

there are several local minima. The parameter vector that corresponds to the lowest

objective function value was chosen to give the calibrated values which are shown in

Table 4-4. The corresponding model fitting with the measured data for cycle 1 and cycle

2 are shown in Figures 4-2 and 4-3, respectively.

The model fitting was satisfactory for nitrate nitrogen and ammonia nitrogen. The

minimum and maximum deviations for NH$_4$-N from the measured values for both the

cycles were 0.04 mg/L and 1.46 mg/L, respectively, and for NO$_3$-N, the minimum and

maximum deviations were 0.03 mg/L and 0.93 mg/L, respectively. The minimum and

maximum deviations for soluble BOD$_5$ for the first cycle were 0.42 mg/L and 6.29 mg/L,

respectively and for second cycle, the minimum and maximum deviations were 2.70 and

16.87 mg/L, respectively. The high deviation noticed in soluble BOD$_5$ between model

fitting and measured values could be attributed to many factors varying from

Figure 4-2: Model calibration: measured data and model fitting for cycle 1 (Sampling Episode 1).
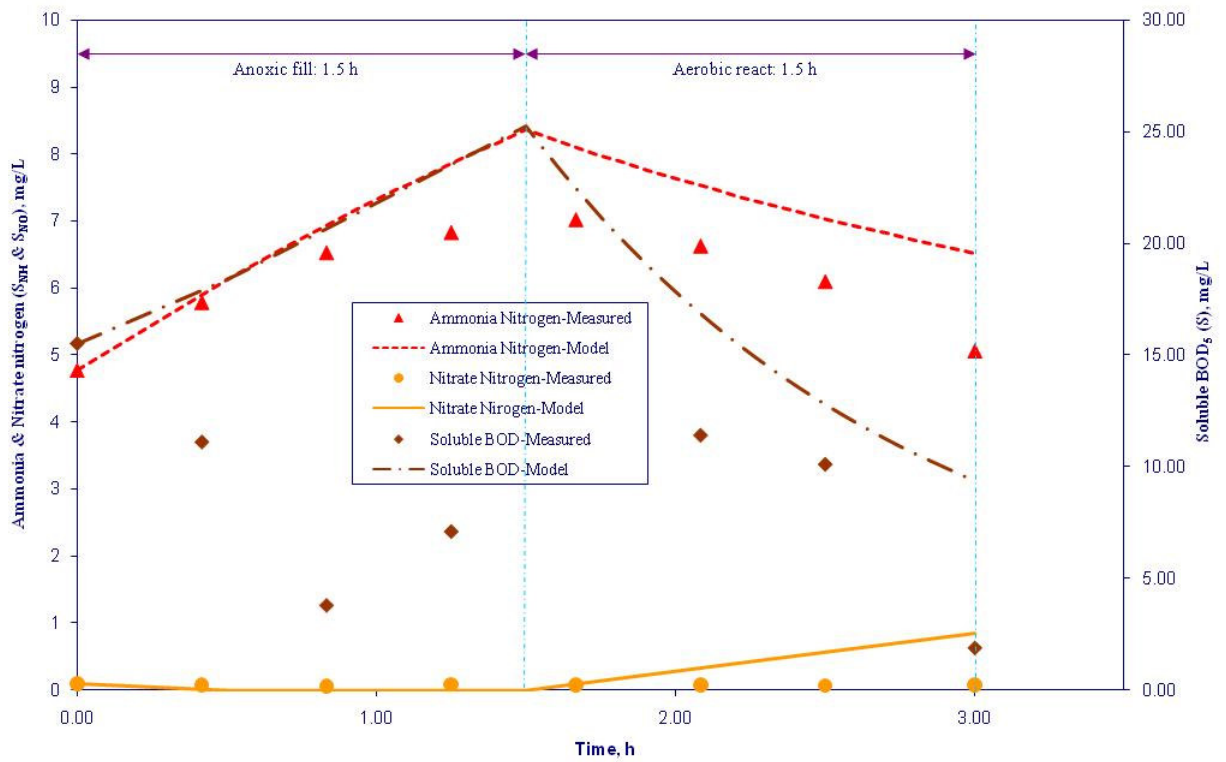
Figure 4-3: Model calibration: measured data and model fitting for cycle 2 (Sampling Episode 1).

measurement errors to the plant operating conditions. It is difficult to measure soluble BOD_5 precisely particularly in a low concentration range (reasons for inaccuracy are explained at the end of this section), and measured values for some control duplicate samples varied by about 2.5 mg/L. Further, the operating conditions of the SBR during the sample collection were not normal and the dissolved oxygen content was lower than the design condition. This was evident from the measured data which showed high concentrations of NH_4-N and low concentrations of NO_3-N. Although the deviations for soluble BOD_5 were high for samples during the middle intervals, the deviation for end of cycle concentration was low, and the model-predicted values were higher than the measured values. For example, the end of cycle BOD_5 concentrations for the first and second cycles were 1.3 mg/L and 7.5 mg/L, respectively higher than the measured values. This conservative model prediction is good for model-based design.

With the calibrated model parameters, the model verification was performed on the data collected from SBR2 on cycle 3 that began at 8:50 am on May 15, 2008. Figure 4-4 depicts the comparison of measured and model-predicted concentration profiles for soluble BOD_5, NH_4-N, and NO_3-N. The model prediction matches fairly well with the measured data for NH_4-N and NO_3-N. The model-predicted end of cycle concentrations for NH_4-N and NO_3-N were 1.93 mg/L and 0.62 mg/L, respectively higher than the measured values. The model prediction for NH_4-N and NO_3-N were conservative, which is good for design. However, the model predicted end of cycle concentration for BOD_5 was 3.67 mg/L lower than the measured value. The model prediction for BOD_5 was aggressive and may not be desirable for design purposes. This is not a surprise
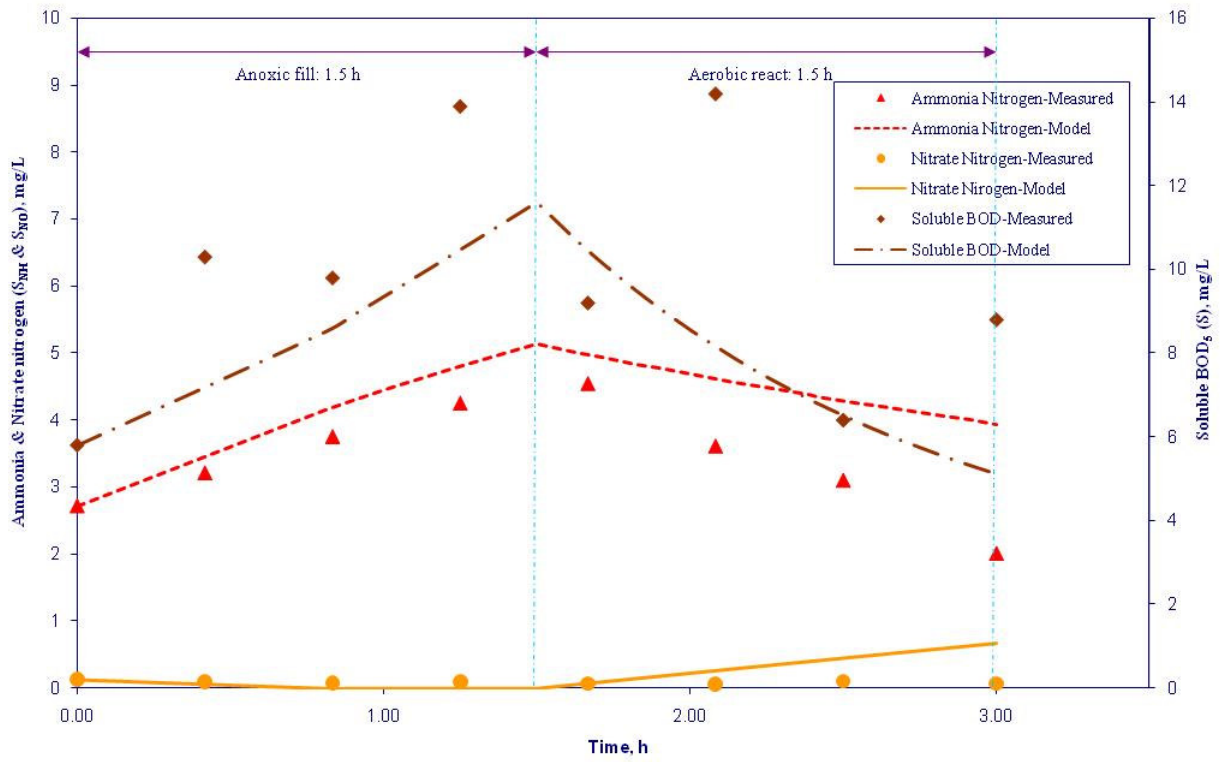
Figure 4-4: Model verification: measured data and model prediction for cycle 3 (Sampling Episode 1).

considering the fact that the deviation for BOD_5 during model fitting was much more pronounced than NH_4-N and NO_3-N, and the possible reasons for such a deviation was explained earlier.

In order to improve the model prediction for BOD_5, it is important to measure this parameter with more accuracy. However, the BOD test is highly sensitive particularly at low concentration range, and the test results are subject to 1) the variability of the seed used, 2) adherence to test procedures, 3) the effects of waste impurities inhibiting biodegradability, 4) introduction of contaminants in the test procedure, 5) seed acclimation procedures, 6) dilution of the test sample to the test limits, and any other intrinsic test limitations (Celenza, 2000). Therefore, it was decided to measure the Chemical Oxygen Demand (COD) as a parameter in place of BOD_5 for the next sampling episode 2. From previous sample analysis, it was found that the ratio of soluble COD to soluble BOD_5 and total COD to total BOD_5 were 2.25 and 1.30, respectively. These ratios fall in the typical range of 1.25 to 2.50 reported in the literature for municipal wastewater (Metcalf and Eddy, 2003). Therefore, the measured soluble and total COD could be converted to soluble BOD_5 and total BOD_5 as required.

4.5.2 Sampling Episode 2 (June 18 – 19, 2008)

In sampling episode 2, samples were collected from SBR2 on June 18 and June 19, 2008 for three cycles. The samples were analyzed for all parameters as described in Section 4.3 except for BOD_5, but in its place COD was measured. The measured soluble and total COD results were converted to soluble BOD_5 and total BOD_5 before their use in the

model. The sample data from the first two cycles, which began at 8:50 am and 2:50 pm, respectively, on June 18, 2008 were used for model calibration, and the third cycle which began at 8:50 am on June 19, 2008 was used for model validation. As described in Section 4.5.1, out of 15 kinetic and stoichiometric parameters that appear in the process model, 10 parameters were assigned typical values recommended in the literature as shown in Table 4-4 and the remaining 5 parameters: $v_{S,\max}$, K_S, $v_{P,\max}$, $v_{N,\max}$ and

$v_{D,\max}$ (denoted as parameter vectors in the calibration algorithm) which are sensitive to the process were determined through the model calibration. The weight factors of 1, 2 and 2 were assigned to BOD_5 data, NH_4-N data and NO_3-N data, respectively. The weight factors of 2 were assigned to NH_4-N data and NO_3-N data because they are likely to be more precise than BOD_5 data. In this sampling episode, the environmental conditions such as dissolved oxygen concentration and sludge volume index in basin SBR2 were different and more favorable than during sampling episode 1. The average dissolved oxygen concentration and sludge volume index for sampling episode 2 were 1.7 mg/L and 128 mL/g, respectively, which were near the design conditions. As a result, all the measured data including NO_3-N were assigned weight factors during this model calibration unlike the sampling episode 1 in which NO_3-N data were ignored. During model calibration, the calibration algorithm was run several times with randomly generated initial guesses for the parameter vector. The calibration algorithm converged at different objective function values, indicating that there are several local minima. The parameter vectors that correspond to the lowest objective function value were chosen as the calibrated values, and they are shown in Table 4-5 along with literature values reported Esposito et al., 2003 and Metcalf and Eddy, 2003. The calibrated values for

Table 4-5: Calibrated kinetic and stoichiometric parameters for sampling episode 2

Parameter	Unit	Calibrated value*	Literature values	
			Ref [1]	Ref [2]
$v_{S,max}$	mg BOD_5/mg VSS-h	0.7416	0.2083	0.1042 – 1.0416
K_S	mg BOD_5/L	598.32	80	25 - 100
$v_{P,max}$	mg BOD_5/mg VSS-h	0.0158	0.1167	0.075 – 0.125**
$v_{N,max}$	mg N-NH_4/mg VSS-h	0.0183	0.0783	0.0417 – 0.4167
$v_{D,max}$	mg N-NO_3/mg VSS-h	0.0004	0.0029	0.0139 – 0.0938

*   Values are for 20° C. The measured temperature of the wastewater was close to 20^oC (See Table 4-3), and hence they were used as such without applying temperature correction factors.

**   Orhon et al. (1999). Values not reported in Metcalf and Eddy, 2003

Ref[1] Typical values as reported in Esposito et al., 2003.

Ref[2] Necessary unit conversions were made to the values reported in Metcalf and Eddy, 2003 to match with units of model parameters.

$v_{S,max}$ and $v_{N,max}$ are within the range reported in the literature. The calibrated value of

$v_{D,max}$ is much lower than the values reported in the literature. It is reasonable as the

denitrification rate is generally low, and varies significantly depending on the

environmental conditions. The calibrated value of K_S is about 8 times higher than the

typical value and the calibrated value of $v_{P,max}$ is about 8 times lower than the typical

value. They are outside the range of values reported in the literature. Therefore, the

process model was forced to run with typical values for K_S (80 mg BOD_5/L) and

$v_{P,max}$ (0.1167 mg BOD_5/mg VSS-h) while retaining other three model parameters to its

calibrated values. But this only increased the relative mean square error from 15.3 (for all

five parameters with the calibrated parameters) to 88.2, and most of the error contribution

was from BOD_5 part of the model component. Therefore, it can be reasonably assumed

that the calibrated values are the ones that best describe the experimental observations for this plant. The corresponding model fitting with the measured data for cycles 1 and 2 are shown in Figures 4-5 and 4-6, respectively. The model fitting was excellent for nitrate nitrogen and ammonia nitrogen. The minimum and maximum deviations for NH_4-N from the measured values for both cycles were 0.07 mg/L and 1.15 mg/L, respectively, and for NO_3-N, the minimum and maximum deviations were 0.01 mg/L and 1.80 mg/L, respectively. The average deviation for NH_4-N for both cycles combined was 0.58 mg/L, and for NO_3-N, the average deviation for both cycles combined was 0.42 mg/L. The model fitting was generally satisfactory for BOD_5. The deviations for soluble BOD_5 for the first cycle ranged between 0.1 mg/L and 1.0 mg/L except for one experimental value in which the deviation was 4.7 mg/L. The deviations for soluble BOD_5 for the second cycle ranged between 0.3 mg/L and 1.6 mg/L except for one experimental value in which the deviation was 5.4 mg/L. The experimental data and the model fitting follow the same trend except at the end of the cycle where the elevated level of soluble BOD_5 resulted in the higher deviation mentioned above. It could be attributed to solubilization of particulate organic matter due to hydrolysis. It is worth investigating to confirm the possibility of solubilization by analyzing mores samples for few more cycles. However, these two experimental values which produced highest deviations could be considered as suspected outliers due to the following reasons: 1) the values of these two suspected outliers were 11.38 mg/L (cycle 1) and 8.09 mg/L (cycle 2), and they were collected at the end of react phase. If these were true values, then this trend would continue for every cycle and would potentially create effluent violation (effluent limit for BOD_5 is 7.6 mg/L). But no effluent violation for BOD_5 has been reported in the plant, 2) on contrast
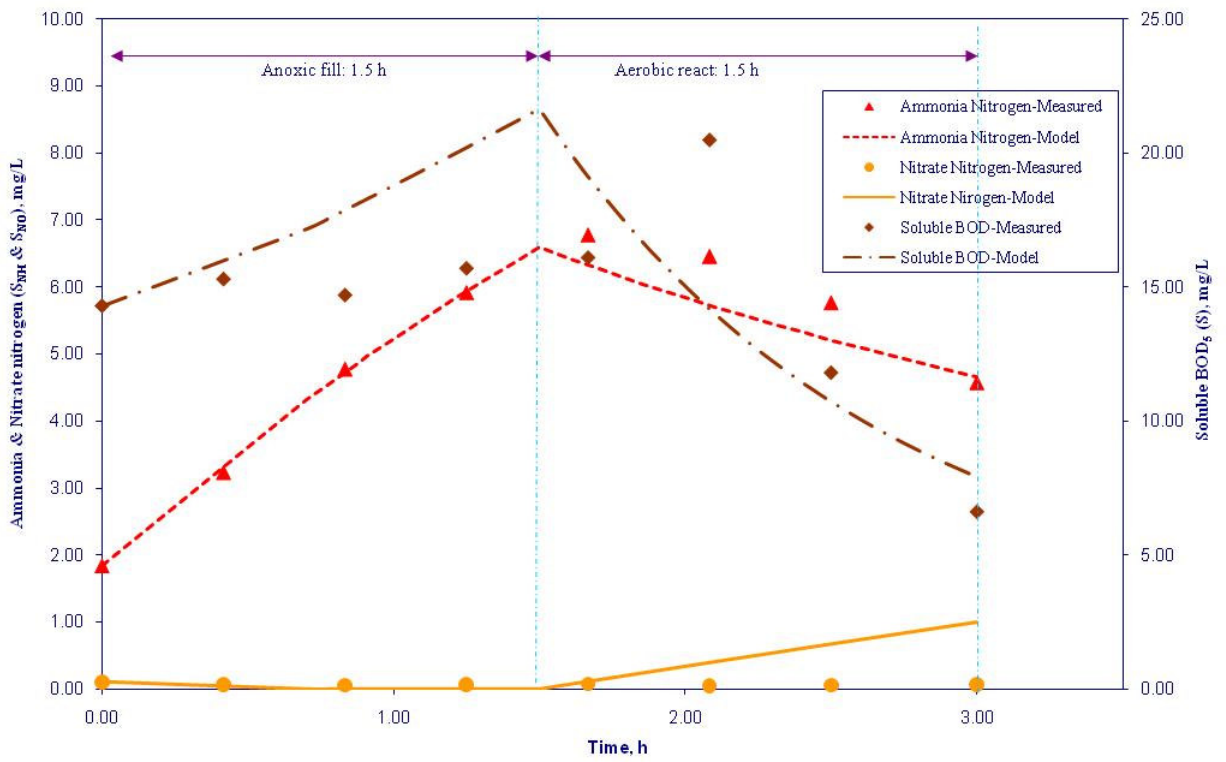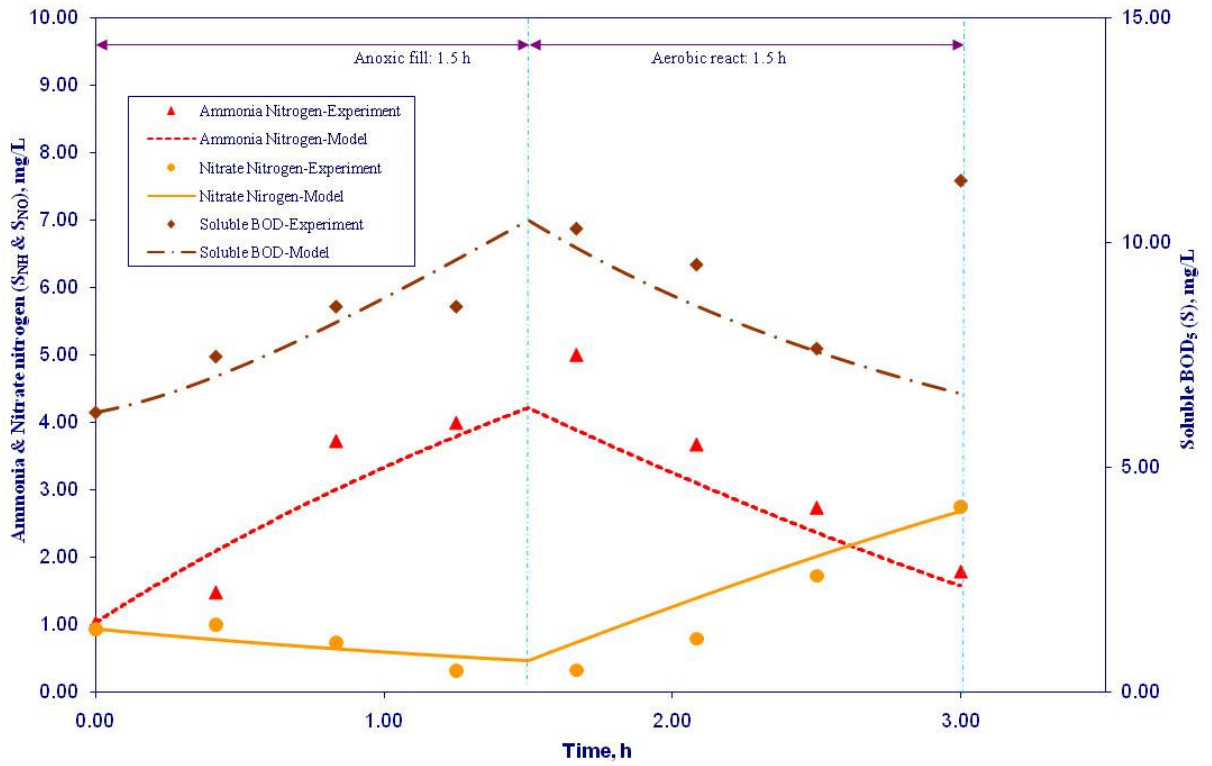
Figure 4-5: Model calibration: measured data and model fitting for cycle 1 (Sampling episode 2).
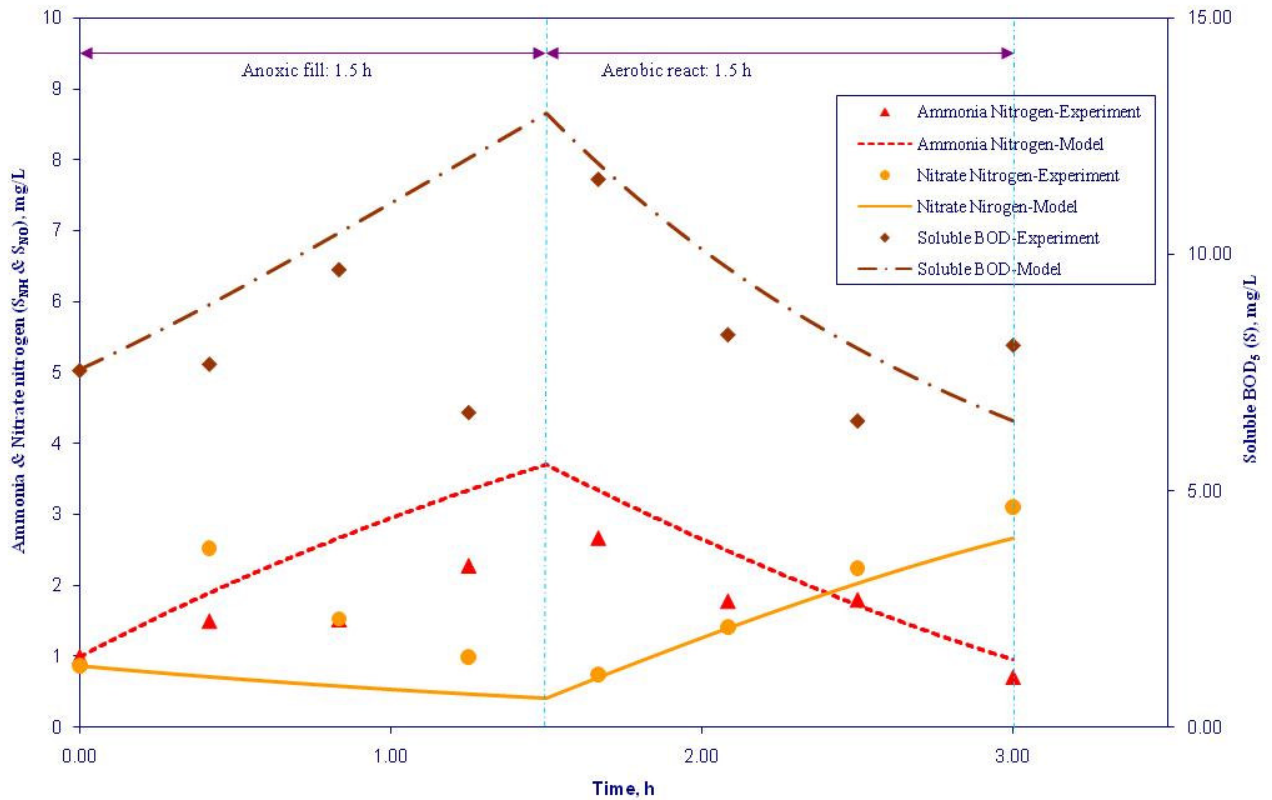
Figure 4-6: Model calibration: measured data and model fitting for cycle 2 (Sampling episode 2).

the end of cycle concentration for BOD_5 for cycles 1 and 2 in sampling episode 1 were

6.60 mg/L and 1.90 mg/L, respectively. These values were lower than the initial BOD_5

concentration as well as the model prediction, 3) the calibrated value for the kinetic

parameter, $\nu_{P,\max}$, which accounts for hydrolysis of particulate organic matter was 0.0158

mg BOD_5/mg VSS-h with relative mean square error of 15.3. The process model was

forced to run for different values of $\nu_{P,\max}$ ranging from 0.0005 to 5.0 BOD_5/mg VSS-h to

capture the suspected outliers. But the results were not satisfactory, and these values of

$v_{P,\max}$ only increased the relative mean square error ranging from 25.8 to 576.3, 4) the

concentration (experimental results) profile of soluble COD from SBR operations

reported in the literature (Carucci et al., 1995; Ibrahim and Abasaeed, 1995; Novak et al.,

1995; Chang et al., 2000; Boaventura et al., 2001; Mazouni et al., 2004) has not shown

any anomaly at the end of the cycle, 5) the elevated level BOD_5 measured at the end of

cycle could also be attributed to sampling or measurement errors. The sampling error is

most likely to occur in full-scale treatment plants where the basins are very large, and 6)

it is also difficult to measure soluble COD precisely, particularly at low concentration

range. Due to above reasons, those two experimental BOD_5 values could be considered

as the suspected outliers, and ignored. In that case, the average deviation for BOD_5 for

both cycles combined was 0.8 mg/L, which is reasonably good. Therefore, the estimated

calibrated parameters can be assumed to be acceptable.

With the calibrated model parameters, model verification was performed on the data

collected from SBR2 during cycle 3 that began at 8:50 am on June 19, 2008. Figure 4-7

depicts the comparison of measured and model-predicted concentration profiles for

soluble BOD_5, NH_4-N, and NO_3-N. The model prediction matches well with the

measured data for NH_4-N and NO_3-N. The model-predicted concentrations for NH_4-N

were within 0.03 mg/L and 0.61 mg/L from experimental values, and the average

deviation was 0.29 mg/L. The average deviation for NO_3-N from the experimental value

was 0.23 mg/L. The model-predicted concentrations for BOD_5 were generally between

0.5 mg/L and 1.8 mg/L except for two samples which are suspected outliers (reasons

explained above). Model-predicted results were generally satisfactory and matched well

with the measured data for most samples.
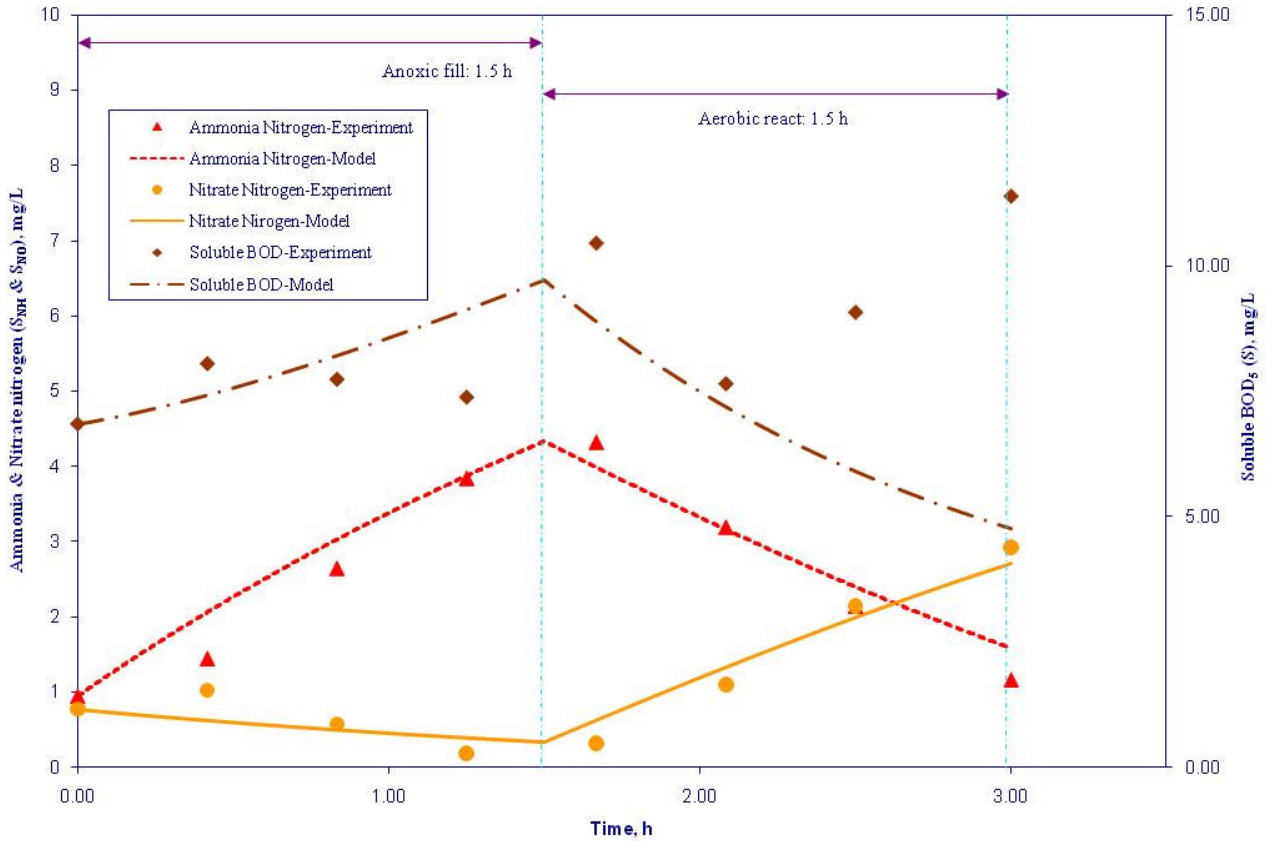
Figure 4-7: Model verification: measured data and model prediction for cycle 3 (Sampling episode 2).

4.6   Conclusions

In general, the model predictions described reasonably well the trend of biodegradable organics removal, nitrification during aeration, and denitrification during the anoxic period. Model-predicted results were generally satisfactory and matched well with the measured data for most samples except for a few outliers for sampling episode 2. Therefore, the model developed in this study could be used successfully to determine the system responses to changes in design variables such as reactor volume, duration of time cycles, and influent conditions. The model predictions for sampling episode 2 were more accurate than sampling episode 1 because the likely measurement errors in BOD_5 in sampling episode 2 were reduced by measuring COD instead of BOD_5.   This indicates the importance of careful planning and selection of sampling parameters for calibration. Fewer measurement errors in sampling will lead to improved model calibration and prediction. Since the model calibration and verification from sampling episode 2 was more accurate than for sampling episode 1, the calibrated model parameters from sampling episode 2 were used for the purposes of model-based design and design optimization in the following chapters. If the model developed in this chapter is to be used for the design or process simulations of wastewater treatment systems with different influent characteristics, it is recommended to perform a pilot study to determine unknown kinetic and stochiometric parameters through the model calibration.

CHAPTER V

MODEL-BASED DESIGN APPROACH

5.1   Introduction

Design of SBR systems involves determining design factors such as reactor size, cycle

time duration, oxygen demand and sludge wasting rate for the given influent conditions

to meet the desired effluent quality. In a conventional design method, these factors are

determined based on the parameters such as mean cell residence time (MCRT), hydraulic

retention time (HRT), food to microorganism ratio (F/M ratio) and mixed liquor

suspended solids concentration (MLSS) using the design procedure outlined in Section

2.4. The conventional design method is a conservative approach and would not always

result in a cost-effective design because it does not use mathematical modeling for

determination of effluent quality. Therefore, the objective of this chapter is to propose a

model-based design methodology for SBR design. The proposed model-based design

methodology has been, then, applied to the full-scale SBR, and results from the model-

based design were compared with the existing conventional design. The proposed

model-based design methodology is intended to serve two purposes: 1) to encourage

practicing engineers to use a model-based design approach, and 2) to demonstrate the

cost economics of using the model-based SBR design over the conventional design

method.

5.2   Model-Based Design Algorithm

The design factors or variables that will be determined in SBR design will include the reactor size; total cycle time including time duration of various phases such anoxic fill, aerated fill, and react; oxygen demand; and sludge wasting rate for the system. Prior to performing design calculations, the number of SBR reactors and number of cycles each SBR reactor will be operated in a day will be selected. Once the number of reactors and cycles/day are selected, the next step is to determine the initial volume (V_0) or low water level (LWL) of the reactor and total reactor volume (V_T) or high water level (HWL) of the reactor. The initial volume (V_0) of the reactor selected for the design shall be adequate to prevent solids carryover in the effluent during decanting. The initial volume that will prevent solids carryover can be expressed in terms of sludge volume index (SVI), desired safety factor (SF), MCRT (θ_c), growth yield coefficient (Y), influent flow rate (q) and influent and effluent BOD$_5$:

$$V_0 \geq SF \bullet SVI \bullet q(S_{in} - S\ )Y \bullet \theta_c \bullet 10^{-6} \qquad\qquad (5.1)$$

The total reactor volume (V_T) shall be able to handle the average and peak flow and meet the desired effluent quality within the given cycle duration. The effective cycle duration is the fill and react time which includes anoxic fill, aerated fill and aerated react time. The effluent substrate concentration at the end of the effective cycle duration shall be within the desired effluent quality, and the process model is required for determination of effluent substrate concentration. The aeration system sizing and sludge wasting rate are

dependent on MCRT and the amount of BOD_5 and nitrogen removed in the SBR basin. The design conditions described above are mathematically expressed and implemented in the proposed model-based design algorithm which consists of the following steps:

Step 1:   Start with known design data which include influent flow, specified influent characteristics, effluent limitations, design and operational criteria that need to be complied.

Step 2:   Assign influent flow as $q = q_{avg}$ (average influent flow rate (m^3/h)), and perform the following steps.

Step 3:   Calculate the initial volume V_0 using equation (5.1).

Step 4:   Assign time cycle at which various phases of reaction such as anoxic fill, aerated fill, and react phase end: t_{anf}, t_{aef}, t_r

Step 5:   Calculate the effective cycle time duration, called batch time t_b (anoxic fill + aerated fill + react) as follows.

$$t_b = t_{anf} + (t_{aef} - t_{anf}) + (t_r - t_{aef}) \qquad (5.2)$$

Step 6:   Check whether the batch time t_b is less than maximum allowable time,

$$t_{max} = T_C - t_{settle} - t_{decant} - t_{waste} - t_{idle,min} \qquad (5.3)$$

Where

t_{max} = Maximum allowable batch time (h),

T_C = Total cycle time (h),

t_{settle} = Settle time duration (h),

t_{decant} = Decant time duration (h),

t_{waste} = Sludge wasting time duration (h), and

$t_{idle,min}$ = Minimum idle time duration (h).

If $t_b < t_{max}$, then go to Step 7. Otherwise, go to Step 4.

Step 7:   Calculate the total final volume, V_T (m^3) using the following equation,

$$V_T = q\,t_{anf} + q\,(t_{aef} - t_{anf})$$ (5.4)

Step 8:   Calculate the end of cycle effluent substrate concentrations, S, S_{NH} and

S_{NO} using the process model.

Step 9:   Check whether the end of cycle effluent substrate concentrations are less

than the desired effluent quality as follows:

$$S \leq S_{LIMIT}$$ (5.5)

$$S_{NH} \leq S_{NH,LIMIT}$$ (5.6)

$$S_{NO} \leq S_{NO,LIMIT}$$ (5.7)

Where

S_{LIMIT} = Effluent limit for BOD$_5$ (mg/L),

$S_{NH,LIMIT}$ = Effluent limit for ammonia nitrogen (mg/L), and

$S_{NO,LIMIT}$ = Effluent limit for nitrate nitrogen (mg/L),

If equations (5.5), (5.6) and (5.7) are satisfied, then go to Step 10.

Otherwise, go to Step 4.

Step 10:   Assign influent flow as $q = q_{peak}$. Reassign t_{anf} and t_r for peak flow such

that t_b for peak flow is equal to or less than t_b for average flow in Step 5

and calculate t_{aef} as follows:

$$t_{f,peak} = (V_T - V_o)/q_{peak} \qquad (5.8)$$

$$t_{aef} = t_{f,peak} \qquad (5.9)$$

Where

q_{peak} = Peak influent flow rate (m^3/h), and

$t_{f,peak}$ = Fill time during peak flow (h).

Step 11:   Calculate the end of cycle effluent substrate concentrations, S, S_{NH}, and

S_{NO} for peak flow using the time cycles as shown in Step 10 and check

whether the equations (5.5) – (5.7) are satisfied. If equations (5.5) – (5.7)

are satisfied, then go to Step 12. Otherwise, go to Step 4.

Step 12:   Calculate the fill time, $t_{f,min}$ during the minimum flow q_{min} as follows:

$$t_{f,min} = (V_T - V_o)/q_{min} \qquad (5.10)$$

Where

q_{min} = Minimum influent flow rate (m^3/h), and

$t_{f,min}$ = Fill time during minimum flow (h).

If $t_{f,min} < t_b$, then the design ends with Step 13. The time cycle assigned

in Step 4, V_0 calculated in Step 3, and V_T calculated in Step 7 are the final

design results. Otherwise, go to Step 4.

Step 13:   Calculate the oxygen requirement and sludge to be wasted from the basin

using the following equations:

$$RO = (24)q_{avg}(S_{IN} - S)(\frac{1}{f_c}) - 1.42W_T + (4.57)(24)q_{avg}(S_{NH,IN} - S_{NH})$$

$$(5.11)$$

$$W_T = \frac{m\, V_T\, X}{\theta_c} \qquad\qquad (5.12)$$

Where

RO = Total actual oxygen requirement for all reactors (kg/d),

f_c = conversion factor for converting to ultimate BOD,

W_T = Amount of sludge wasted from all reactors (kg/d), and

m = Number of reactors.

The above expressions (5.11) and (5.12) are the total estimation for all reactors on a daily basis, and a straightforward adjustment may be made on a cycle basis per reactor based on number of cycles and number of reactors.

The above design algorithm can be successfully used for the design of SBR systems once the design information and model parameters are available.

5.3   Case Study

The developed model-based design algorithm has been applied to the same full-scale SBR system from which the process model has been calibrated and verified as described in Chapter IV. This existing SBR system constructed in 1992 was designed originally by conventional design methods. The design decision variables estimated from the model-based design were compared with the existing design. The design output results from the model-based design are shown in Table 5-1 along with the existing design for comparison.

Table 5-1: Comparison of results from model-based design with existing plant design data

	Existing Plant	Model-based design
Influent characteristics		
Influent flow, average (q_{avg}), m³/h	836	836
Influent flow, peak (q_{peak}), m³/h	1,672	1,672
Influent flow, minimum (q_{min}), m³/h	473	473
Total BOD5, mg/L	120	120
Soluble BOD5 (S_{in}), mg/L	30	30
Particulate BOD5 ($S_{P,in}$), mg/L	90	90
NH4-N ($S_{NH,in}$), mg/L	17	17
Design parameters		
MLVSS (X), mg/L	3,200	3,200
MCRT, (θ_c), d	25	25
F/M ratio	0.05 - 0.10	0.05 – 0.10
SVI, mL/g	150	150
Design output results		
Initial reactor volume (V_0), m³	3,430	2,900
Total reactor volume (V_T), m³	5,980	4,012
During average flow		
Anoxic fill (t_{anf}), h	0.75	0.90
Anoxic fill + Aerated fill (t_{aef}), h	1.33	1.33
Anoxic fill + Aerated fill + React (t_r), h	3.25	2.33
Effluent soluble BOD5(S), mg/L	< 7.6	1.9
Effluent NH4-N (S_{NH}), mg/L	< 1.5	< 0.10
Effluent NO3-N (S_{NO}), mg/L	------	4.93
During peak flow		
Anoxic fill (t_{anf}), h	------	0.66
Anoxic fill + Aerated fill (t_{aef}), h	------	0.66
Anoxic fill + Aerated fill + React (t_r), h	------	2.33
Effluent soluble BOD5(S), mg/L	------	1.6
Effluent NH4-N (S_{NH}), mg/L	------	< 0.10
Effluent NO3-N (S_{NO}), mg/L	------	4.87

Reactor volume shown in Table 5-1 is for one reactor.

The model-based design uses the same design criteria that were used in the original design for meaningful comparison. The results shown in Table 5-1 are for one reactor. The concentration profiles of soluble BOD_5 and NH_4-N for model-based design for average design flow conditions are presented in Figure 5-1. The concentration profile of NO_3-N is not shown as this plant has no effluent limits for NO_3-N or total nitrogen.
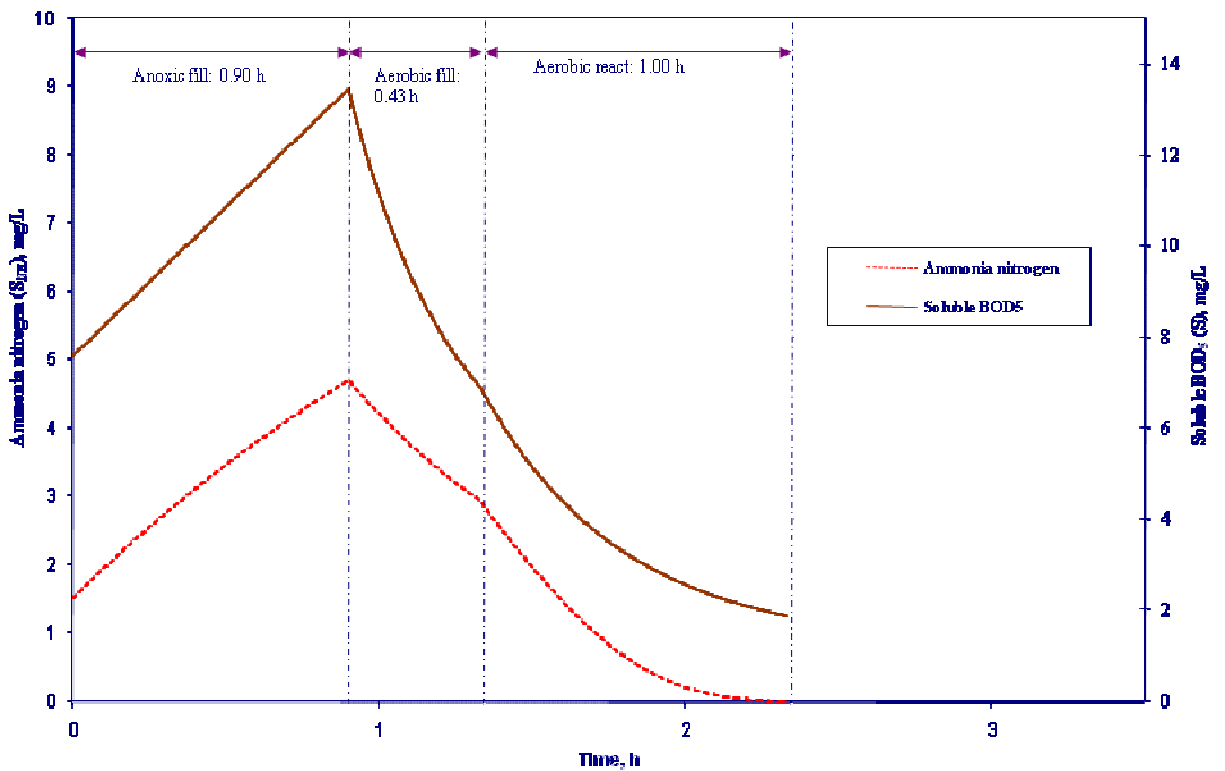
Figure 5-1: Model-based design: concentration profile of soluble BOD_5 and NH_4-N at average design flow rate.

Comparison of the values in Table 5-1 indicates that the reactor volume determined from the model-based design is about 33 percent less than the existing plant reactor volume. This would have meant considerable savings in the capital construction cost if the model-based design method had been used for the design. In conventional designs, the total reactor volume would usually be estimated based on the average flow rate. But review of the existing plant design information revealed that the total reactor volume was estimated based on the peak flow using the same fill time that was used for average flow. It may be noticed in the model-based design algorithm that it is not necessary to use the peak flow to estimate the total volume of the reactor if the fill time during the minimum flow ($t_{f,\min}$) is less than the batch time (t_b) and the effluent substrate concentrations during peak flow conditions meet the effluent limitations. The reduction in total reactor volume would have been about 11 percent for the model-based design, even if the existing plant reactor volume had been estimated based on average flow rate. Table 5-1 also reveals that the effective cycle duration and total aeration time from the model-based design is 0.92 hours and 1.07 hours, respectively less than the existing design. Aeration to the existing SBR basin is provided by the 74.5 kW (100 HP) blower. The above reduction in aeration time would correspond to energy savings of $5,820/year-basin at a unit energy cost of 5 cents/kW-h. This would result in a total energy savings of $23,280 per year for the 836 m^3/h (5.3 MGD) SBR system. Since the model-based design in this study involves using the same design criteria as was used in the original design, MCRT and MLVSS concentrations were not changed. As a result, the sludge production and aeration rate would remain the same. However, if design optimization is the goal, then it

is worthwhile to include design parameters such as MCRT and MLVSS also as design decision variables.

5.4   Conclusions

In this chapter, a model-based design algorithm has been proposed for the design of SBR. The proposed model-based design algorithm has been applied to a full-scale SBR system. Results from the proposed model-based design algorithm were compared with the existing design of a full-scale SBR system. Model-based design alone, without any design optimization method, produced a reactor volumetric reduction of about 33 percent and a total energy savings of $23,280 per year for the 836 m^3/h (5.3 MGD) SBR system. These benefits in cost savings would be very attractive and would encourage design engineers to move from conventional design methods towards model-based design methods. As the results from this study were encouraging, an optimization design methodology to determine all relevant design parameters to minimize the capital and operational costs has been developed in the following Chapter VI.

CHAPTER VI

DESIGN OPTIMIZATION MODEL

6.1   Introduction

The objective of this chapter is to develop cost optimization model to determine the

optimal design parameters such as the reactor volume, batch time, mean cell residence

time (MCRT), mixed liquor suspended solids concentration (MLSS), and waste sludge

quantity to keep the total capital and operation cost of SBR system to a minimum while

satisfying the effluent requirements and operational process stability criteria. The

developed optimization model was applied to the same full-scale SBR system that was

modeled and calibrated in Chapter IV to produce the optimal design, and the cost savings

were compared with the existing design. A sensitivity analysis has also been performed

to study the impact of the variation in unit sludge processing costs on optimum design

parameters. This optimization model is, in fact, the automation of the model-based design

approach presented in Chapter V with construction and operational cost as an objective

function. Some of the salient features of this optimization model are as follows:

1. The model includes both the capital construction and operational cost in the

    objective function. The model also includes the operational cost associated with

solids processing which is normally overlooked. The solids processing cost is

very important as it will have a competing effect on the aeration energy cost.

2. The model incorporates equipment system data such as diffuser and blower

    characteristics directly into the objective function. In most optimization models,

    these data are usually lumped as one parameter by unit energy cost factor for

    simplification. The use of equipment system data directly into the model allows

    for more precise energy computation as it accounts for variations in air

    temperature and dissolved oxygen concentration in the aeration basin.

In an earlier phase of this research, an optimization model was developed for the design

of SBR for removal of a single pollutant, biodegradable organics. The developed model

was applied to an existing wastewater treatment plant with a capacity of 315 m^3/h (2

MGD) to produce an optimum design. The findings of this study were published as a

paper titled "An Optimization Model for Design and Operation of Sequencing Batch

Reactor" in *4th Conference on Sequencing Batch Reactor Technology* proceedings held in

Rome, Italy on April 7-10, 2008. A copy of the conference proceedings paper is included

in Appendix D. The limitations of this model were: (a) the model as presented

considered only a single biodegradable organic substrate and did not include an anoxic

phase in the cycle, (b) the model was not calibrated and verified with experimental data

from actual plant operation, and (c) the model did not include equipment system data

such as diffuser and blower characteristics directly into the objective function. These

limitations were overcome in the multi-substrate optimization model developed in this chapter.

6.2   Optimization Model Development

6.2.1   Objective Function Formulation

The objective function of the optimization model in this problem is the annualized cost for constructing the SBR system and the costs associated with its operation. The construction cost includes the cost of building the reactor (C_R), and the cost for installing blowers (C_B), and diffusers (C_D). The operation costs include the energy cost for aeration (C_{AE}) and the solids processing cost (C_{SP}). The resulting objective function to be minimized is expressed as follows:

$$Minimize \quad TC = C_R + C_B + C_D + C_{AE} + C_{SP} \tag{6.1}$$

The cost components listed in equation (6.1) are directly related to the volume of reactor, the amount of oxygen provided to the reactor and the amount of excess sludge wasted from the SBR system. The construction cost for building the reactor is a product of annualized cost factor (CF), unit cost for building the reactor (c_v, in \$/m^3), and the total volume of the reactor (V_T, in m^3).

$$C_R = CF \; V_T \; c_v \tag{6.2}$$

The oxygen required for the process is provided into the SBR basin through a diffused air system by blowers. The installation cost of the blower (C_B) is dependent on the blower capacity (E_B) which is a function of standard oxygen requirement (SOR). Standard oxygen requirement in turn is dependent on influent and effluent conditions, the amount of sludge wasted from the reactor (W_T), the blower parameters and the diffuser parameters. The following expressions (Metcalf and Eddy, 2003; Sanitaire CD Catalog) are used to determine the blower capacity (E_B) and blower installation cost.

$$C_B = CF \ E_B \ c_{bl} \tag{6.3}$$

$$AOR = (24) q_{avg} \left(\frac{1}{m}\right)(S_{IN} - S)\left(\frac{1}{f_c}\right) - 1.42 W_T + (4.57)(24) q_{avg} \left(\frac{1}{m}\right)(S_{NH,IN} - S_{NH}) \tag{6.4}$$

$$\frac{AOR}{SOR} = \frac{\alpha \vartheta^{(T-20)} \left[\beta C_{sat\,20} \left(\frac{P_{site}}{P_{std}}\right)\left(\frac{C_{sutfT}}{9.07}\right) - DO \right]}{C_{sat\,20}} \tag{6.5}$$

$$w_{air} = \frac{SOR}{(86400)(0.232)SOTE} \tag{6.6}$$

$$E_B = \frac{w_{air} RT}{29.7 n(eff)} \left[\left(\frac{P_{del}}{P_{site}}\right)^n - 1 \right] \tag{6.7}$$

Where

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E_B = blower capacity in kW,

c_{bl} = blower installation cost in \$/kW,

AOR = actual oxygen transfer under field condition,

SOR = oxygen transfer at standard condition,

α = ratio of oxygen transfer efficiency (OTE) in wastewater to OTE in tap water,

ϑ = operating water temperature factor,

T = operating wastewater temperature (degree Celsius),

β = ratio of field to standard condition oxygen saturation ratio,

$C_{sat\,20}$ = oxygen saturation concentration at standard condition (mg/L),

P_{site} = site ambient pressure (atm),

P_{std} = ambient pressure at standard condition (1 atm),

C_{sutfT} = oxygen saturation conc. at operating temperature and altitude (mg/L),

DO = operating dissolved oxygen concentration (mg/L),

w_{air} = weight of air flow (kg/s),

E_B = blower capacity, function of aeration system parameters (kW),

$SOTE$ = Standard oxygen transfer efficiency in clean water,

R = gas constant (8.314 kJ/k mole $^{\circ}$K),

T_{air} = absolute inlet air temperature ($^{\circ}$K),

$P_{del.}$ = absolute outlet delivery pressure (atm),

n = constant for air (0.283), and

$eff.$ = blower efficiency in fraction.

The diffuser installation cost (C_D) is a product of annualized cost factor (CF), unit cost for diffuser installation (c_{df}, in $ per kg O_2/d), and the standard oxygen requirement (SOR, in kg O_2/d).

$$C_D = CF \ SOR \ c_{df} \tag{6.8}$$

The energy cost for providing aeration to the SBR basin is dependent on the duration of time in which blower is turned on to supply process oxygen to the SBR basin, and the annual energy cost is calculated using the following expression.

$$C_{AE} = (365) \, n \, E_B \, ((t_{aef} - t_{anf}) + (t_r - t_{aef})) \, c_e \tag{6.9}$$

Where

n = Number cycles/d,

E_B = Blower capacity (kW),

t_{aef} = Time aerobic fill ends (anoxic fill + aerated fill) (h),

t_{anf} = Time anoxic fill ends (anoxic fill time) (h),

t_r = Time react ends (anoxic fill + aerated fill + react) (h), and

c_e = Unit energy cost factor ($/kW-h).

The solids processing cost (C_{SP}) for excess sludge produced in the process is a product of unit sludge processing cost (c_s, in $/kg) and the amount of sludge wasted from the system (W_T, in kg/d). The excess sludge produced in the reactor must be wasted

periodically to maintain a desired mean cell residence time (θ_c, in d) and mixed liquor

volatile suspended solids concentration (X , in mg/L).

$$W_T = \frac{m\, V_T\, X}{\theta_c} \qquad\qquad (6.10)$$

$$C_{SP} = (365)(\frac{1}{m})W_T\, c_s \qquad\qquad (6.11)$$

Where

m = Number of reactors,

V_T = Total volume of the reactor (m^3),

X = Mixed liquor volatile suspended solids concentration (mg/L),

θ_c = Mean cell residence time (d), and

c_s = Unit sludge processing cost ($/kg).

The objective function stated in equation (6.1) is evaluated using the listed equations

(6.2) through (6.11). In the objective function above, the design decision variables are

the volume of the reactor (V_T), time duration of the phases in which anoxic fill, aerated

fill and react phase end (t_{anf} , t_{aef} , t_r) , the mean cell residence time (θ_c), and the

mixed liquor volatile suspended solids concentration (X) which is the sum of

autotrophic and heterotrophic biomass. The optimum values will have to be determined

for these design decision variables to keep the objective function to a minimum under the specified constraints on the system.

6.2.2 Constraints

The design decision variables shall meet several constraints to ensure the feasibility of process operation. These constraints arise from the process model equations, desired range for operational parameters, influent conditions, and effluent limitations. The process model equations (3.5) through (3.24) listed in Chapter III determine the concentrations of various substrates, the heterotrophic and autotrophic biomass in the SBR at any given time, based on the growth kinetics and mass balance considerations. The initial volume (V_0) of the reactor selected for design shall be adequate to prevent solids carryover in the effluent during decanting, and the total reactor volume (V_T) shall be able to handle the peak flow. The initial volume that will prevent solids carryover can be expressed in terms of sludge volume index (SVI), desired safety factor (SF), MCRT (θ_c), growth yield coefficient (Y), influent flow rate (q), and influent and effluent BOD$_5$. The total time for fill and react is also limited by the number of cycles the reactor is required to perform in a day. These conditions provide the following inequality constraints.

$$V_0 \geq SF \bullet SVI \bullet q(S_{in} - S\;)Y \bullet \theta_c \bullet 10^{-6} \qquad (6.12)$$

$$t_b \leq t_{max} \qquad (6.13)$$

The batch time (effective cycle time duration, t_b) and the maximum allowable batch time (t_{max}) can be estimated using the equations (5.2) and (5.3). The most commonly used parameters for controlling the activated sludge process are hydraulic retention time (HRT), mean cell residence time (θ_c), food-to-microorganism ratio (F/M), and mixed liquor volatile suspended solids concentration (X). These control parameters provide process stability and have a specified operating range for the desired performance. The operating range for the control parameters is to be specified based on the desirable limits as recommended for the SBR process (U.S. EPA, 1999; Metcalf and Eddy, 2003). These control parameters and their operating ranges provide the following additional inequality constraints on the system.

$$HRT_{min} \leq HRT \leq HRT_{max} \tag{6.14}$$

$$(F/M)_{min} \leq (F/M) \leq (F/M)_{max} \tag{6.15}$$

$$\theta_{c,min} \leq \theta_c \leq \theta_{c,max} \tag{6.16}$$

$$X_{min} \leq X \leq X_{max} \tag{6.17}$$

Where

$$HRT = \frac{mV_T}{q_{avg}} \tag{6.18}$$

$$F/M = \frac{24 q_{avg} S_{in}}{m V_T X}$$ (6.19)

Suffix $_{min}$ and $_{max}$ are the minimum and maximum limits for the respective parameters and numerical values for these limits will have to be specified.

The effluent substrate concentrations such as biochemical oxygen demand (S), ammonia nitrogen (S_{NH}), and nitrate nitrogen (S_{NO}) at the end of cycle shall meet the desired effluent limitations, which provide an additional set of inequality constraints on the design decision variables. The effluent substrate concentration, S, S_{NH} and S_{NO} at the end of cycle are determined using the process model equations (3.5) through (3.24).

$$S \le S_{LIMIT}$$ (6.20)

$$S_{NH} \le S_{NH,LIMIT}$$ (6.21)

$$S_{NO} \le S_{NO,LIMIT}$$ (6.22)

Where

Suffix $_{LIMIT}$ is the effluent limitation for the respective parameters.

The constraints listed in the expressions (6.20) through (6.22) shall be satisfied under all influent flow rates including the peak influent flow rate, which means the set of expressions (6.20) through (6.22) will be evaluated twice, one for the average influent flow rate and the other for peak influent flow rate. Steps 10 and 11 outlined in the model-based design algorithm in Section 5.2 will be used for determining the duration of

the time phases to estimate the end of cycle effluent substrate concentration for the peak

flow conditions.

The influent flow rate will be low under the minimum flow conditions, which will take a

longer time to fill the SBR basin. The longer fill time will interfere with other phases of

the cycle such as settle, decant and sludge wasting, and impact the performance of the

SBR. Therefore, the fill time ($t_{f,\min}$) under minimum flow conditions shall not exceed

the batch time (t_b), the effective time cycle duration. This provides an additional

constraint on the design decision variables as expressed by the following expression.

$$t_{f,\min} \leq t_b \qquad\qquad\qquad (6.23)$$

The fill time ($t_{f,\min}$) for minimum flow conditions can be calculated using the equation

(5.10). All design constraints are shown in expressions (6.12) through (6.23). The

objective function (6.1) will have to be minimized subject to these constraints to produce

the optimal design.

6.2.3   Optimization Technique

The objective of the optimization is to find the minimum total cost of the system meeting

the process and operational constraints as listed in expressions (6.12) through (6.23). The

objective function listed in the equation (6.1) is a non-linear function, and is to be

minimized subject to the constraints (expressions (6.12) - (6.23)). The first step to solve

this type of problem (constrained non-linear programming) is to convert the constrained

problem into an unconstrained problem. The penalty function method (Rao, 1996; Edgar et al., 2001), which penalizes the infeasible solution, has been used in this problem for converting the constrained problem into an unconstrained problem. After formulating the penalty function, it is minimized using the selected optimization technique. The optimization technique that has been employed in this minimization problem is the same optimization technique, Simulated Annealing (SA), which has been used in model calibration. The advantages and disadvantages of using this optimization technique have already been discussed in Section 4.4. The penalty function is the only difference in the optimization procedure that was not presented earlier. As a result there will be two loops in the optimization technique presented in this section, one for penalizing the objective function for constraints violation and the other loop for implementing annealing schedule for minimization. A computer program has been developed in MATLAB to integrate both the process and optimization model, and solve the optimization problem using the Simulated Annealing (SA) algorithm. The MATLAB codes with explanations are listed in Appendix E.

6.3   Case Study

The optimization model developed above has been applied to the existing full-scale SBR system described in Chapter IV. The goal of this case study was to determine the optimum design factors (decision variables) such as the volume of the reactor, time duration of different phases (anoxic fill, aerated fill and react) of the cycle, mean cell residence time, and mixed liquor volatile suspended solids concentration to keep the total

cost of the system to a minimum. The resulting optimal cost was compared with the cost

of the existing plant which was originally designed by conventional design method.

Many cost parameters in the objective function are subject to variation depending on

geographical location and other factors. However, the sludge processing cost is subject

to more fluctuation than other cost parameters as it includes various costs associated with

labor, material, electric power, transportation hauling, permitting fee, and other

incidentals which are subject to variation. The reported range of costs for sludge

processing and end use are from a low of $144 to a high of $948 per dry ton (146 to 963

$/Metric Ton) (McMillan et al., 2000). Therefore, a sensitivity analysis has also been

performed to study the impact of variation in unit sludge processing costs on the optimum

design.

6.3.1   Optimal Design

The optimal design and operation of the SBR consists of finding the volume of the SBR

basin, air flow required for the basin and the amount of sludge needed to be wasted on a

cyclic basis that would keep the construction and operation costs to the minimum. This

will be accomplished by determining appropriate design factors such as the volume of the

basin, time duration of different phases (anoxic fill, aerated fill and react) of the cycle,

mean cell residence time, and mixed liquor volatile suspended solids concentration

through the optimization technique while satisfying all constraints related to process

modeling, operational criteria, influent conditions and effluent limitations. The

formulated optimization problem has 14 equipment parameters, 6 cost parameters, 15

kinetic and stochiometric parameters, 14 process variables and 13 constraints. Some

parameters are specific to site and operating conditions. Some parameters are specific to temperature. The influent conditions and the required effluent limitations used in the optimization problem are the same as the existing conventional design, and are shown in Table 6-1 (data were obtained from plant O&M manual) as follows:

Table 6-1: Design influent conditions and effluent limitations used in optimization model

Influent characteristics	
Influent flow, average (q_{avg}), m^3/h	836
Influent flow, peak (q_{peak}), m^3/h	1,672
Influent flow, minimum (q_{min}), m^3/h	473
Total BOD$_5$, mg/L	120
Soluble BOD$_5$ (S_{in}), mg/L	30
Particulate BOD$_5$ ($S_{P,in}$), mg/L	90
NH$_4$-N ($S_{NH,in}$), mg/L	17
Effluent limitations	
Effluent soluble BOD$_5$(S), mg/L	< 7.6
Effluent NH$_4$-N (S_{NH}), mg/L	< 1.5

The calibrated kinetic and stochiometric parameters for sampling episode 2 as shown in Table 6-2 were used in the optimization model. The upper and lower bounds for the key operating process variables that appear in constraints are shown in Table 6-3. The equipment and cost parameters used in the optimization model are shown in Table 6-4.

Table 6-2: Kinetic and stoichiometric parameters used in optimization model

Parameter	Unit	Value*
Y_H	mg VSS/mg BOD_5	0.55
K_P	mg BOD_5/mg VSS	0.095
K_N	mg N-NH_4/L	1
K_{O2}	mg/L	1
Y_A	mg VSS/ mg N-NH_4	0.15
ε	mg N-NH_4/mg BOD_5	0.05
K_D	mg N-NO_3/L	0.1
$K_{S,D}$	mg BOD_5/L	0.1
Y_D	mg VSS/ mg N-NO_3	0.512
ω	mg BOD_5/mg N-NO_3	4
$v_{S,max}$	mg BOD_5/mg VSS-h	0.7416
K_S	mg BOD_5/L	598.32
$v_{P,max}$	mg BOD_5/mg VSS-h	0.0158
$v_{N,max}$	mg N-NH_4/mg VSS-h	0.0183
$v_{D,max}$	mg N-NO_3/mg VSS-h	0.0004

* Values are for 20° C and are to be adjusted for temperature variations.

Table 6-3: Operating range for key process variables (US EPA, 1999; Metcalf and Eddy, 2003)

Parameter	Values	Unit
F/M ratio	0.05 – 0.30	
Hydraulic retention time, HRT	6 – 24	h
MCRT, θ_c	10 – 30	d
MLVSS at volume V_T, X	2000 - 3500	mg/L
Sludge volume index, SVI	150	mL/g
Safety factor, SF	2.50	
Maximum allowable batch time, t_{max}	3.30*	h

* Allows 4 time cycles a day including settle, decant and idle (Plant O&M manual).

Table 6-4: Equipment (Metcalf and Eddy, 2003; Sanitaire CD Catalog) and cost parameters

Parameter	Value	Unit	
Equipment parameters			
α	0.55		
β	0.98		
ϑ	1.024		
DO	2.0	mg/L	
P_{site}	14.29	psi	
P_{std}	14.70	psi	
C_{sat20}	10.39	mg/L	
SOTE	0.32		
$P_{del.}$	22.79	psi	
Blower Efficiency (eff)	0.70		
n	0.283		
R	8.314	kJ/k mole $^{\circ}$K	
Cost parameters			
Unit cost for reactor construction. c_v	350	\$/m^3	(Sedlak, 1991)
Blower installation cost, c_{bl}	1,310	\$/kW	(Sedlak, 1991)
Diffuser installation cost, c_{df}	1,060	\$/ kg O$_2$/h	(Sedlak, 1991)
Unit energy cost factor, c_e	0.07	\$/kW-h	
Unit sludge processing cost, c_s	0.75	\$/kg	(McMillan et al., 2000)
CF (For 25 years at 5% interest rate)	0.0709		

Note: Construction and installation cost shown in the Table were already adjusted for third quarter of 2008 price based on Engineering News Record (ENR) construction cost index.

Using the influent conditions and effluent limitations as listed in Table 6-1 and other

constraints and parameters listed in Tables 6-2 through 6-4, the optimization model that

was coded in MATLAB was run several times with random initial guesses for design

factors to determine the optimal design factors that keep the total cost of the SBR system

to a minimum. Most of the optimization model runs produced final results that

converged very close to the values reported in Table 6-5. The existing plant design data

are also shown in Table 6-5 for comparison. The results shown in Table 6-5 are for one

reactor for both sets of data, and the total cost for four reactors will have to be multiplied

by four.

Table 6-5: Comparison of results from optimization model with the existing plant design data

	Existing Plant	Optimization model
Initial reactor volume (V_0), m^3	3,430	2,027
Total reactor volume (V_T), m^3	4,684	3,281
Anoxic fill (t_{anf}), h	0.75	0.75
Anoxic fill + Aerated fill (t_{aef}), h	1.50	1.50
Anoxic fill + Aerated fill + React (t_r), h	3.30	3.20
MLVSS (X), mg/L	3,500	2,200
MCRT, (θ_c), d	25	16.5
F/M ratio	0.03	0.07
HRT, h	22.4	15.7
Waste sludge, kg/d	146	178
Standard oxygen requirement, kg/h	341	340
Amortized capital construction cost		
Reactor cost, $/year	116,000	81,500
Blower & diffuser, $/year	32,650	32,640
Operation and maintenance cost		
Aeration energy, $/year	19,400	18,600
Solids processing, $/year	40,000	48,900
Total cost (Capital and O&M), $/year	208,050	181,640

* Costs shown in Table 6-5 are for one reactor, and the total cost for four reactors will have to be multiplied by a factor of four.

The concentration profiles of effluent soluble BOD$_5$, NH$_4$-N, and NO$_3$-N corresponding

to optimal design factors under average flow conditions are shown in Figures 6-1, 6-2

and 6-3, respectively, along with the existing design. The concentration profiles of effluent soluble BOD_5, NH_4-N, and NO_3-N for optimal design factors under peak conditions are shown in Figures 6-4, 6-5 and 6-6, respectively.
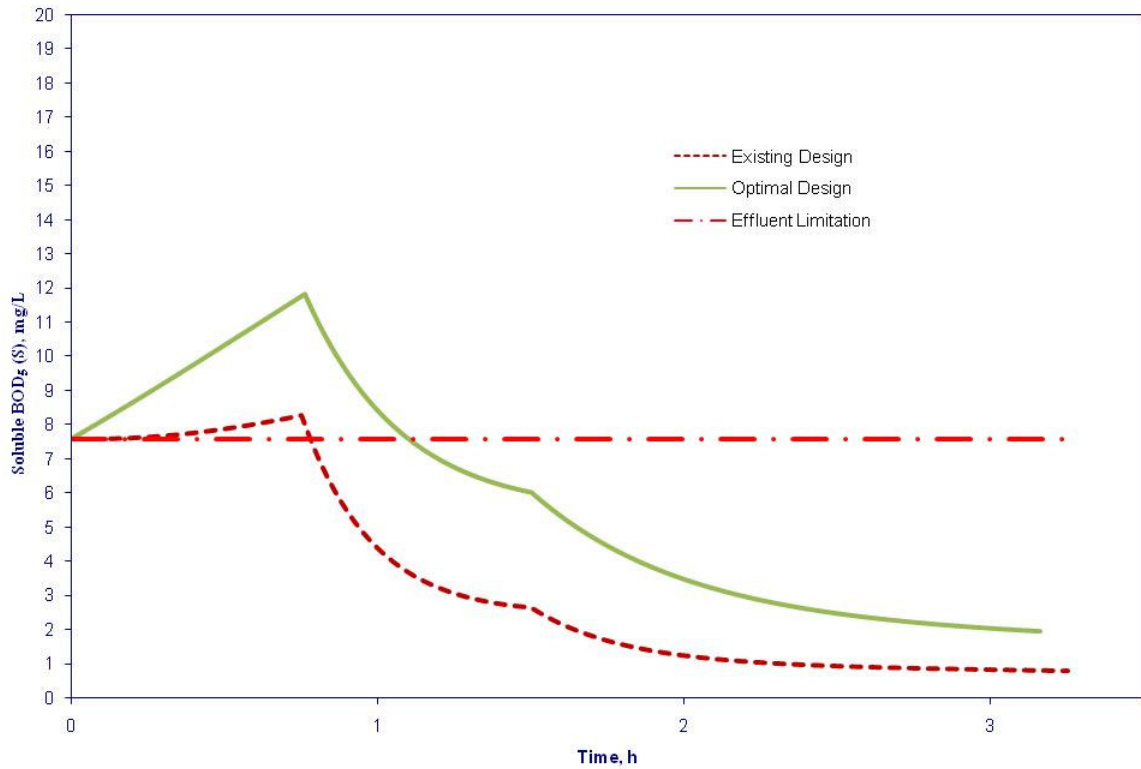
Figure 6-1: Comparison of effluent soluble BOD_5 concentration between optimal and existing design under average flow conditions.
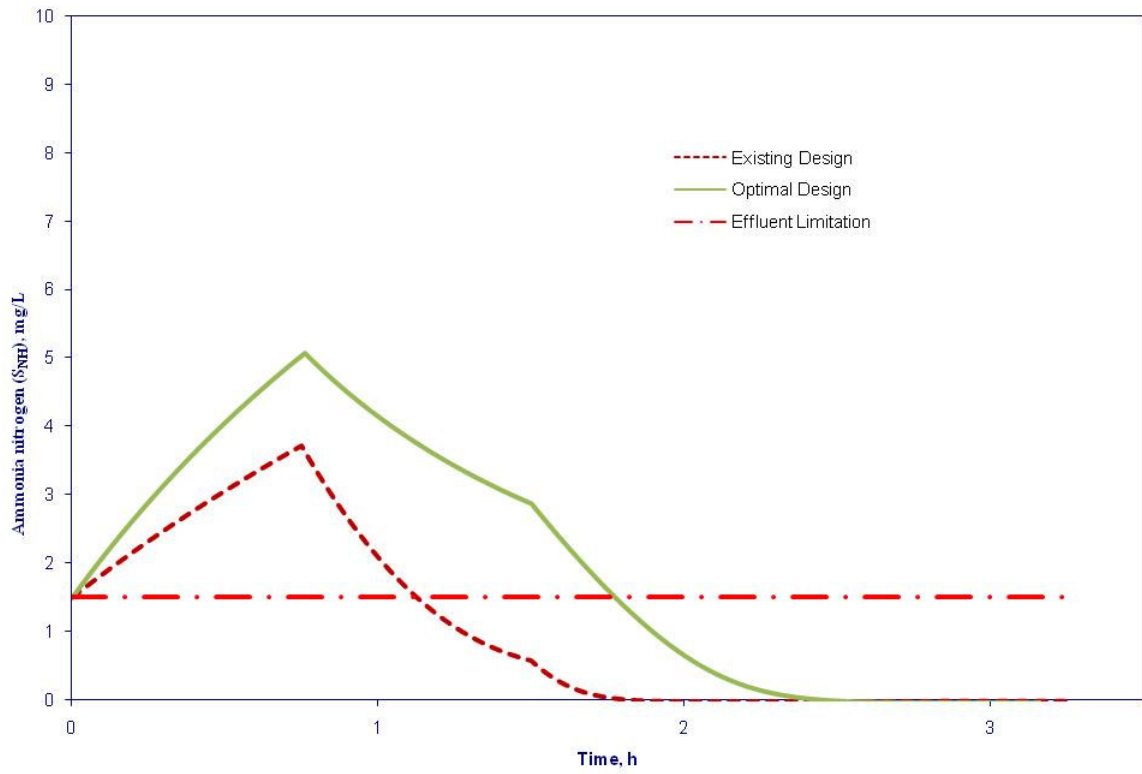
Figure 6-2: Comparison of effluent NH₄-N concentration between optimal and existing design under average flow conditions.
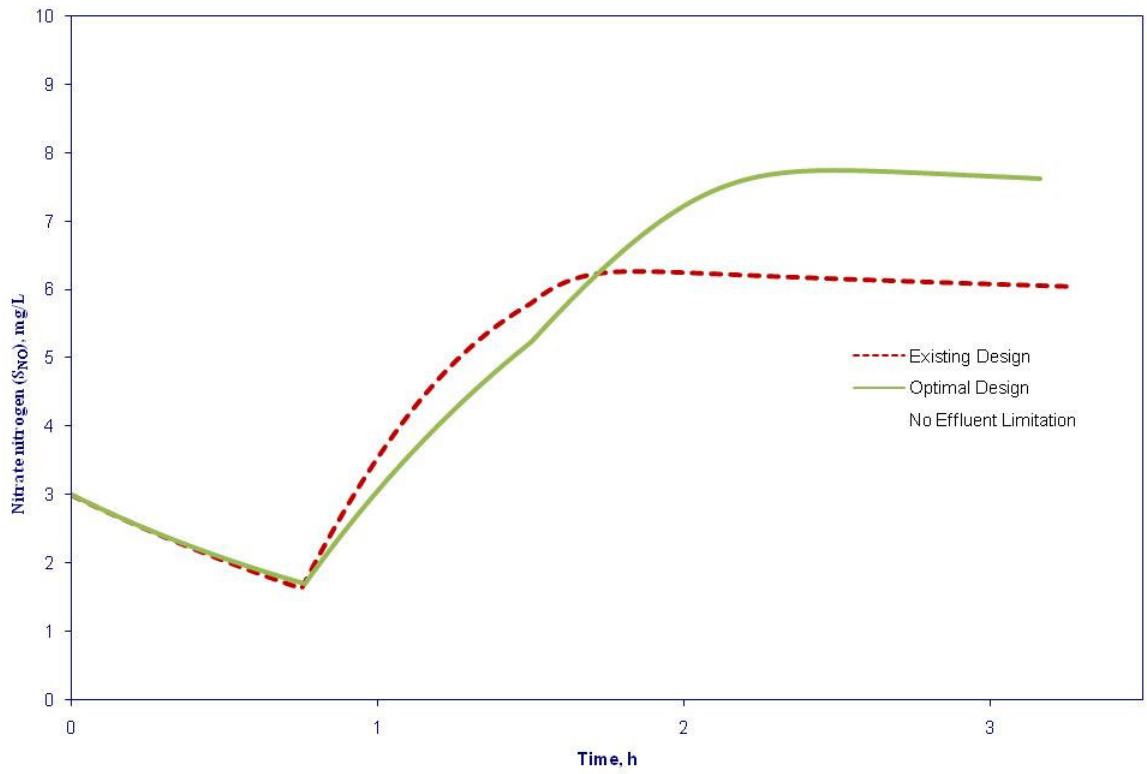
Figure 6-3: Comparison of effluent NO$_3$-N concentration between optimal and existing design under average flow conditions.
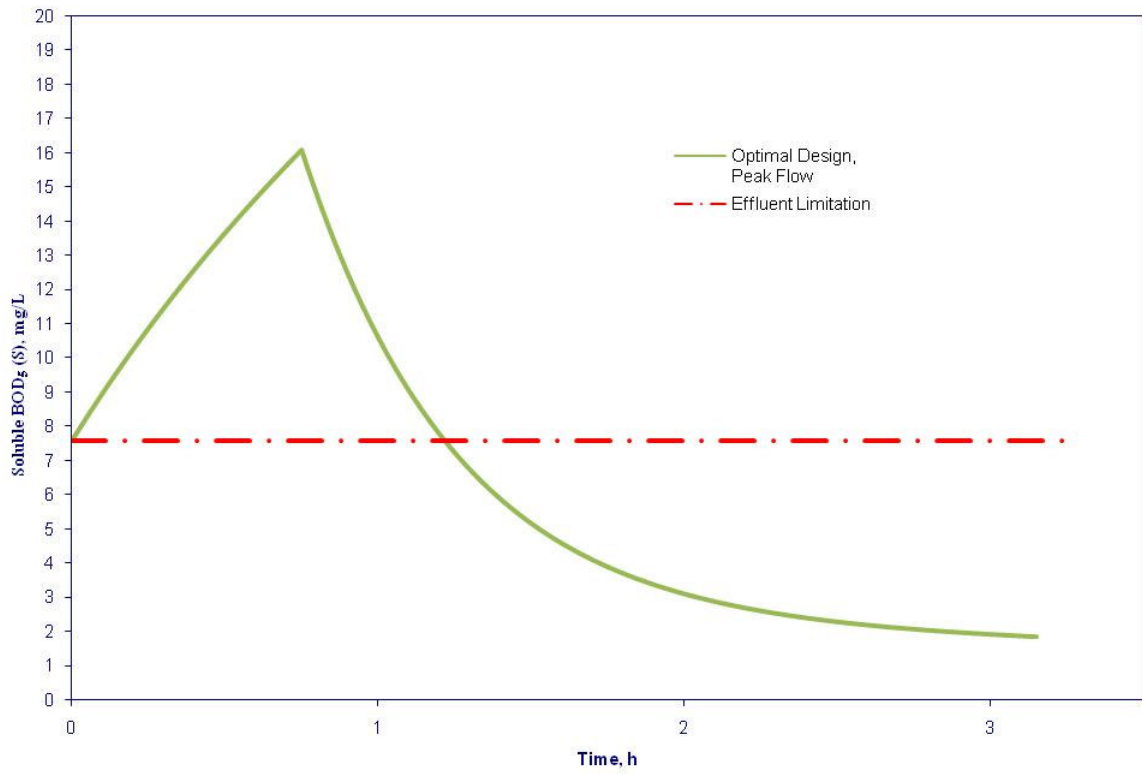
Figure 6-4: Effluent soluble BOD$_5$ concentration for optimal design under peak flow conditions.
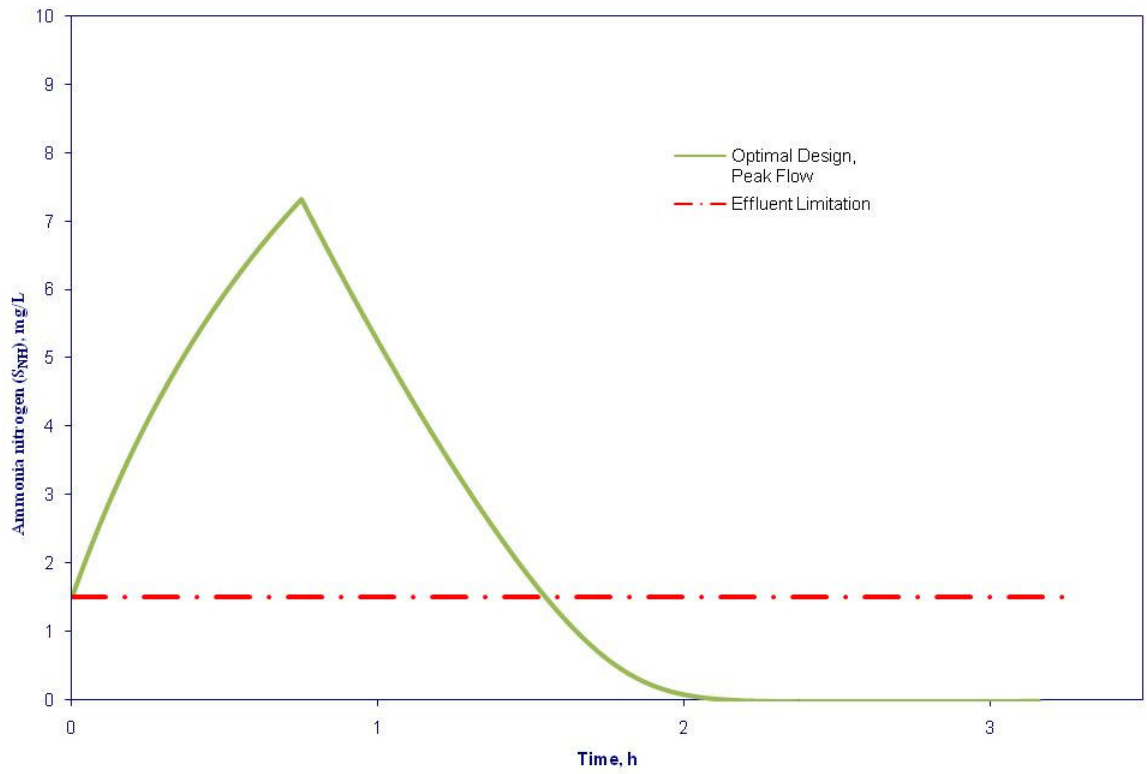
Figure 6-5: Effluent NH$_4$-N concentration for optimal design under peak flow conditions.
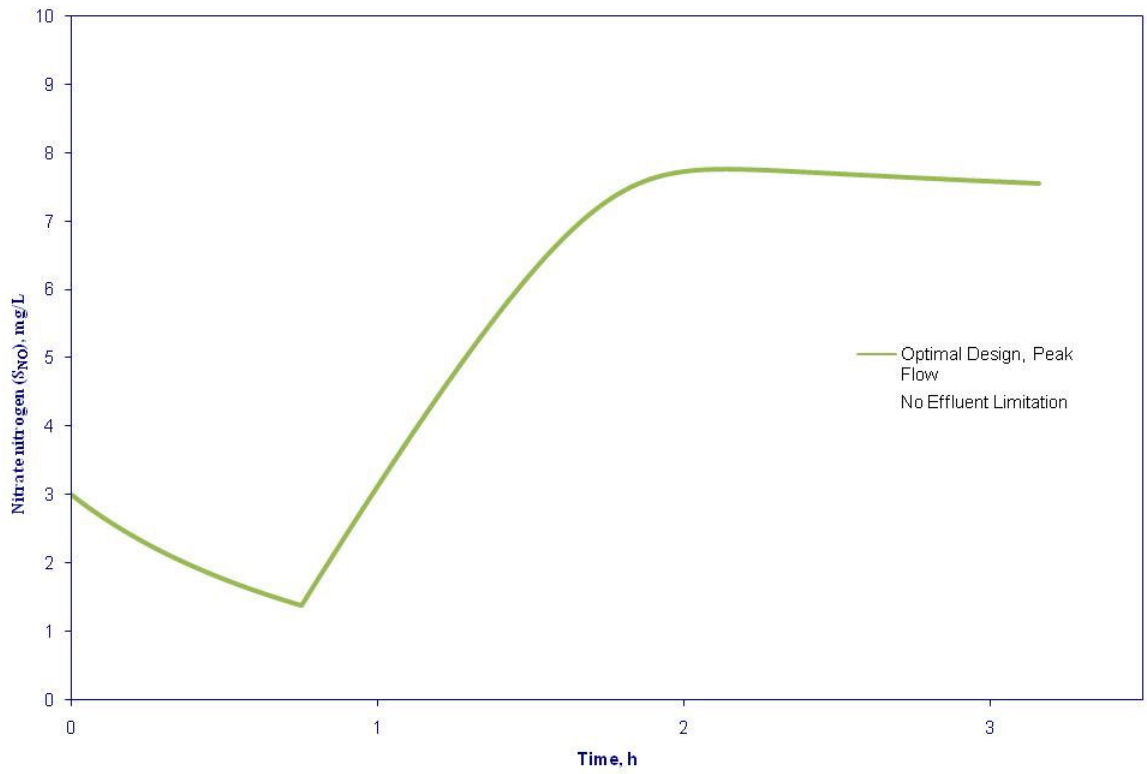
Figure 6-6: Effluent soluble NO₃-N concentration for optimal design under peak flow conditions.

The end of cycle concentrations for effluent soluble BOD_5, NH_4-N, and NO_3-N for optimal design under both average and peak flow conditions are lower than the effluent limitations. Comparison of the values in Table 6-5 indicates that the hydraulic retention time determined from the optimization model is 15.7 hours, which is 6.7 hours lower than the existing design. Due to this, optimal reactor volume is about 30 percent less than the existing plant reactor volume, resulting in annualized savings of about $34,500 in the capital construction cost for 25 years. The standard oxygen requirements for the existing and optimal design are 341 and 340 kg/h, respectively. Therefore, the installed capacity of the blower at the existing facility (75.86 kW) is almost the same as the blower capacity (75.73 kW) determined from the optimization model, and there was no appreciable variation in blower and diffuser installation cost between existing and optimal designs. Table 6-5 reveals that both the effective cycle duration and total aeration time from the optimal design is 0.10 hours less than the existing design. This is again a negligibly small difference. However, an appreciable difference was noticed in MCRT. The MCRT determined from the optimization model is 16.5 days compared to the existing design of 25 days, which drove the solids processing cost higher for the optimal design. However, the overall savings in total capital and operation and maintenance cost is about $26,410 annually (12.70 % less than the existing design) which is a significant saving for larger installations. It may be noticed in Figures 6-1 through 6-6 that end of cycle effluent substrate concentrations for optimal design under both average and peak flow conditions are lower than the effluent limitations. This suggests that an SBR can perform well under varying influent flow and shock loadings. Besides the optimal design satisfies all other design criteria such a hydraulic retention time, mean cell residence time, mixed liquor

suspended solids concentration and F/M ratio within the recommended range of design

values. Therefore, a conservative design approach to provide safety margin against

uncertainties in influent flow conditions is not required.

In the model-based design approach in Chapter V, the design factors such as MLVSS and

MCRT were retained at the same value for both model-based design and existing design,

which provided only the flexibility of varying the time phases of the cycle to reduce the

aeration energy cost. However, in the design optimization model all design factors were

allowed to vary to determine the optimal design factors for minimum total cost of the

system.   This has resulted in a MCRT of 16.5 days for optimal design which is lower

than the existing conventional design value. The impact on total cost is two-fold as can

be seen in Table 6-5. First, lower MCRT increases the solids production rate and thereby

increases the solids processing cost. Secondly, lower MCRT decreases oxygen

requirement resulting in reduced aeration energy cost. The inclusion of solids processing

cost in the optimization model moves the determination of MCRT towards a trade-off

between energy and solids processing costs. If solids processing cost is not included in

the optimization model, the optimal MCRT will always be the lowest value to keep the

energy cost low. This confirms the importance of including solids processing cost as an

integral part of operational cost optimization studies.

6.3.2   Cost Sensitivity Analysis

Although there are many cost parameters in the objective function, the unit cost for

sludge processing is subject to a wide variation depending on the type of processing and

the mode of disposal as discussed in the previous section. Therefore, a sensitivity

analysis has been performed to study the impact of possible variations in sludge

processing costs on total cost of the system, percent savings, and a key process control

parameter, MCRT. The optimization model was run several times by varying the unit

sludge processing cost from 0.20 $/kg ($ 200/ Metric Ton) to 1.0 $/ kg ($ 1000/ Metric

Ton) at the intervals of 0.20 $/kg, while retaining all other parameters as shown in Tables

6-1 through 6-4 as the same. The results showing the comparison of total cost for the

optimal design and the corresponding total cost of the existing design is presented in

Figure 6-7. The percent total cost savings over the existing design are shown in Figure 6-

8. Optimal MCRT for various unit sludge processing costs are shown in Figure 6-9.

Figure 6-7 reveals that the difference in total cost of the optimal design over existing

design is almost uniform with a constant annual savings of about $30,000. The percent

saving over the existing design varies from 12 percent to 18 percent as noticed in Figure

6-8. Figure 6-9 reveals that the optimal value for the mean-cell residence time increases

as the unit sludge processing cost increases. Increase in MCRT means reduced sludge

production. Increase in the optimal MCRT value with increase in unit sludge processing

cost moves the O&M cost relating to the sludge processing to a minimum by reducing the

sludge production rate. The sensitivity analysis clearly indicates that maintaining high or

low mean-cell residence for the optimal operation cost is dictated by the variations in

sludge processing cost. However, unit cost variation in sludge processing did not make

any appreciable change in total cost savings as the optimization model attempted to find a

balance among other parameters to keep the total cost to a minimum.
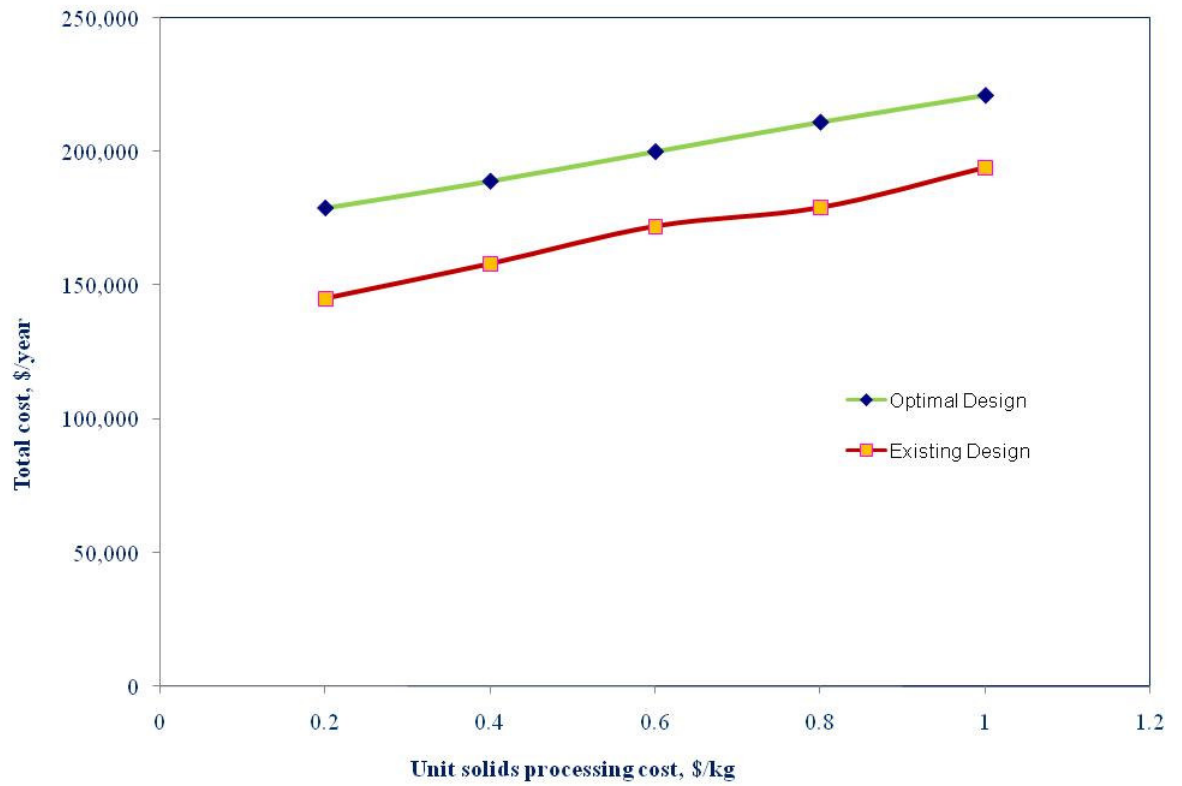
Figure 6-7: Comparison of total cost of optimal design with existing design at different unit sludge processing cost.
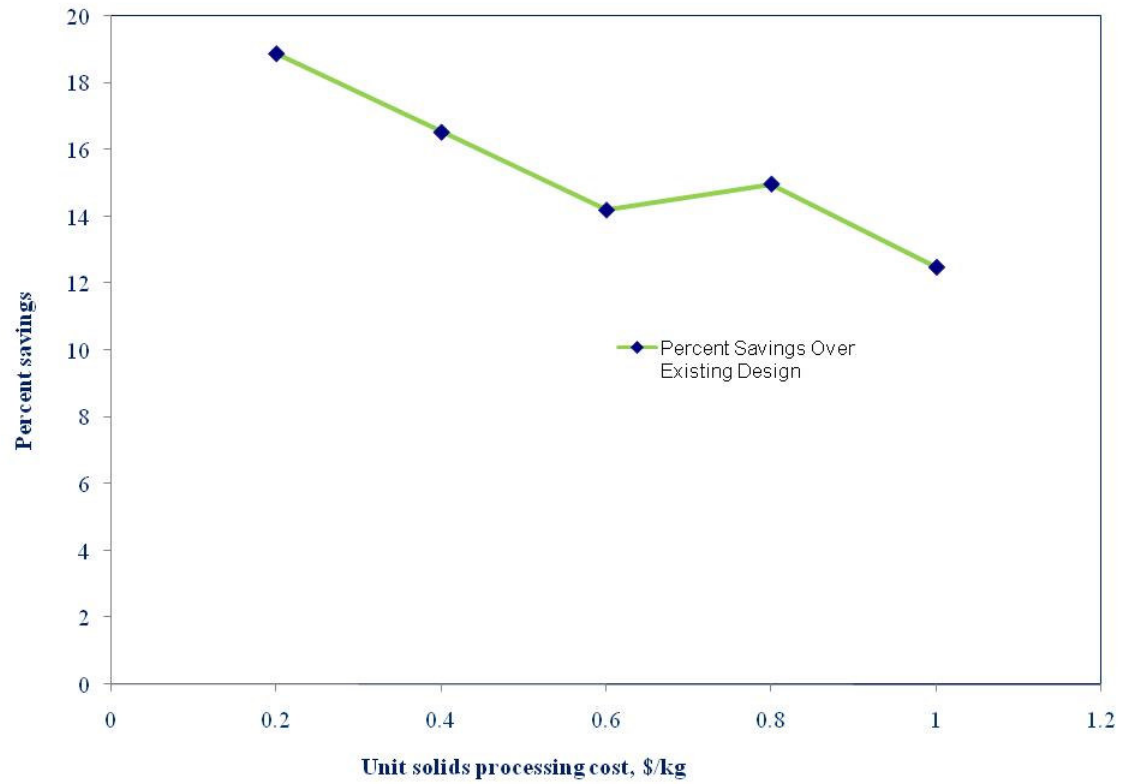
Figure 6-8: Percent cost savings optimal design over existing design at different unit sludge processing cost.
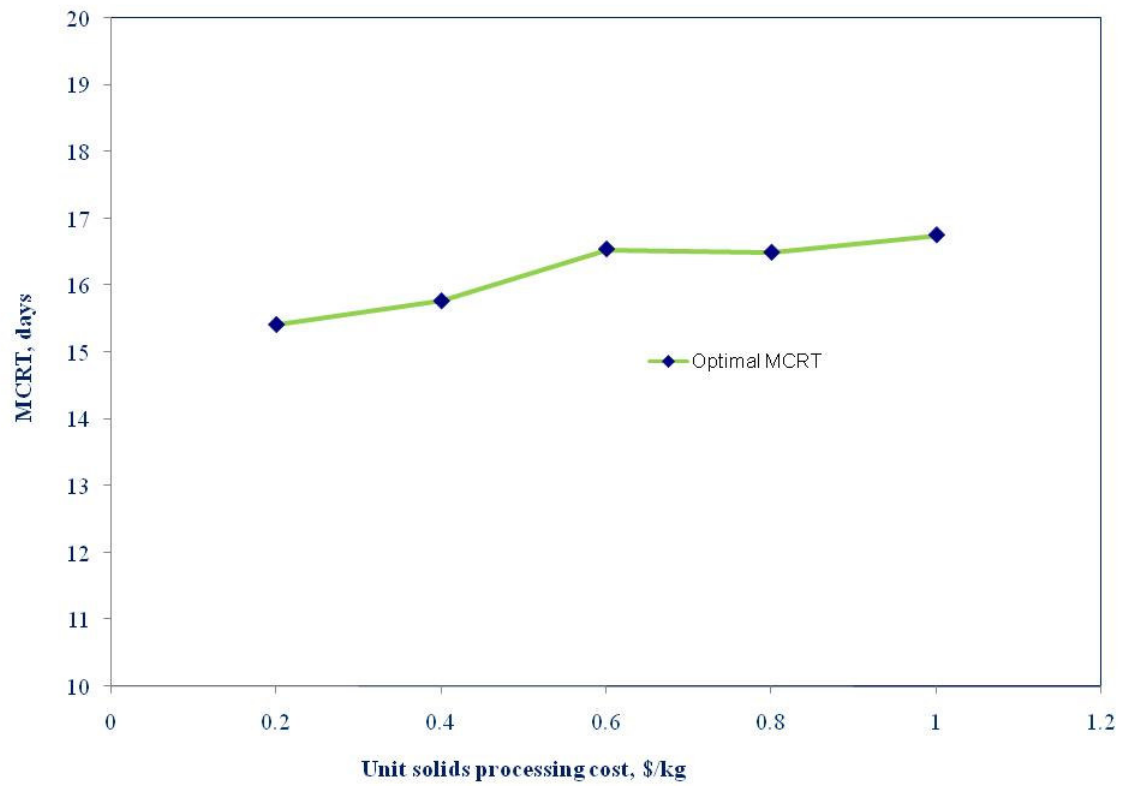
Figure 6-9: Optimal MCRT at different unit sludge processing cost.

6.4   Conclusions

In this chapter, an optimization model was developed for the design of an SBR for removal of bio-degradable organics and nitrogen. The developed optimization model was applied to an existing SBR system with a treatment capacity of 836 m^3/h (5.3 MGD) to obtain an optimum design, and the cost savings were compared. The optimization model produced a reactor volumetric reduction of about 30 percent and savings of about $26,410 annually in total construction and operation cost over 25 years life cycle of the plant. This represents a significant cost savings, and should encourage the designer to use design optimization models. The predicted end of cycle concentrations for effluent soluble BOD$_5$, NH$_4$-N, and NO$_3$-N for optimal design under peak flow conditions were 1.8 mg/L, 0.1 mg/L and 7.5 mg/L, respectively, much lower than the stipulated effluent limitations. In the conventional design method, conservative design values were used to account for influent variations, but this approach is not necessary for SBR systems. It is evident in Figures 6-1 through 6-6 that they perform well under varying influent conditions. This emphasizes that conservative design values used in the conventional design methods would result in more expensive designs.

Cost sensitivity analysis reveals that the use of appropriate unit cost factors for sludge processing is important for determining operational parameters such as MCRT. High sludge processing cost would require the plant to be designed for high mean-cell residence time. However, the variations in unit cost of sludge processing did not make any appreciable change in cost savings in this case.

CHAPTER VII

OPTIMAL OPERATIONAL STRATEGIES FOR ENERGY SAVINGS

7.1   Introduction

Design optimization models are useful and apply only to the design of new systems. In

this research a great deal of time has been spent on process model development, sample

data collection, calibration and verification of the existing SBR system. The existing SBR

system has already been constructed and had been in operation for about 15 years.

Therefore, not much could be done about changing the components that appear in the

capital construction cost such as reactor volume, diffuser or blower size. However,

changes that could be implemented to benefit the existing SBR system are implementing

suitable operational policies to reduce the operational cost of the existing system. Energy

and solids processing costs are the major operational costs for any wastewater treatment

facility. Therefore, in this chapter, a suitable operational policy has been developed for

reducing the energy cost of the SBR system. The operational policy will consist of a

chart that would provide the time duration of different phases of the cycle for a given

MLVSS concentration, F/M ratio, and wastewater temperature. This operational policy

will not make any major modifications to the operational procedures at the existing

system except adjusting the timing of the react phase in the control system.

This minor modification was intended to make a smooth and gradual transition from the existing practice, and to give ease and comfort to the operational staff.

7.2   Optimal Operational Strategies

Although the designed treatment capacity of the existing SBR system is 836 m³/h (5.3 MGD), the data collected during sampling as well as the plant records indicate that wastewater influent flow rate currently averages between 710 m³/h (4.5 MGD) and 790 m³/h (5 MGD). The maximum design MLVSS concentration for the SBR basin is about 3,200 mg/L. Collected sample data and the plant records indicate that MLVSS concentration in the SBR basins is usually maintained at a low level of around 1,000 mg/L to mitigate concerns of maintaining a low F/M ratio. A control system was set up to operate the SBR basins regularly on a daily basis with anoxic fill of 1.5 hours and react phase of 1.5 hrs totaling 3 hours and leaving 3 hours for other phases such as settle, decant, waste sludge and idle. Keeping anoxic fill at 1.5 hours will allow the plant to receive wastewater for 24 hours a day with 4 basins, each operated at four cycles per day. The blowers that supply air to the SBR basins are operated by constant speed motors which means the air flow rate to the SBR basins will be constant all the time and cannot be adjusted. Due to these operational constraints, the only flexibility allowing for minimizing aeration energy comes from adjusting the time cycle for reduced aeration time and adjusting the MLVSS concentration in the SBR basin. Therefore, the optimal operation strategy in this case involves selecting the react time through process simulation to meet the effluent limitations at the given MLVSS concentration.

Several process simulations were carried out using the calibrated model for an influent

flow rate of 790 m^3/h (5 MGD) for various MLVSS concentrations ranging from 930

mg/L to 2,800 mg/L at wastewater temperatures of 15, 20 and 25 degrees Celsius. The

wastewater temperature at the plant is usually expected to range between 15 and 25

degrees Celsius. The stoichiometric and kinetic parameters were adjusted for

temperature variations. For process simulations, the anoxic fill was kept at 90 minutes

and react time was varied in such a way that the end of cycle effluent substrate

concentration satisfies the plant effluent limitations. The results from the process

simulations are shown in Table 7-1 as an operational strategy chart. The operational

strategy chart provides the aerated react time for the given MLVSS concentration and

expected effluent substrate concentration. Knowing the wastewater temperature which

normally stays constant for the given season and MLVSS concentration in the SBR basin,

the plant operation staff can select the suitable aerated react time from the operational

strategy chart to meet the effluent limitations. The operational strategy chart also

provides the expected effluent quality and F/M ratio for the selected aerated react time

and MLVSS concentration. For example, if the wastewater temperature is 20 degrees

Celsius and the MLVSS concentration is 1460 mg/L, the required react time to meet the

specified effluent limitations is 66 minutes. The estimated F/M ratio is 0.08. The

expected effluent substrate concentrations of soluble BOD_5, NH_4-N, and NO_3-N as

shown in Figure 7-1 and tabulated in Table 7-1 are 4.7 mg/L, 0.97 mg/L and 3.57 mg/L,

respectively which are lower than the effluent limitations, BOD_5 of 7.6 mg/L and NH_4-N

of 1.5 mg/L.

Table 7-1: Optimal operational policies for existing SBR system

Influent Characteristics

Average Flow	=	790 m^3/h	5.0 MGD
Influent BOD5	=	120 mg/L	120 mg/L
Influent TKN	=	17 mg/L	17 mg/L

SBR Conditions

Length	=	35.4 m	116.0 ft
Width	=	35.4 m	116.0 ft
Initial Depth	=	3.0 m	9.9 ft
Initial Volume	=	3772 m^3	1.0 MG
Anoxic Fill Time	=	1.5 h	1.5 h
Final Depth	=	4.0 m	13.0 ft
Final Volume	=	4957 m^3	1.31 MG
Hydraulic Retention Time	=	25 h	25 h

Operational Strategy Chart

Wastewater Temperature °C	°F	Optimal React Time minutes	Optimal MLVSS mg/L	Optimal MLSS mg/L	Estimated F/M ratio	Expected Effluent Quality BOD$_5$ mg/L	NH$_4$-N mg/L	NO$_3$-N mg/L	Savings in energy cost* (%)
15	59	90	1860	2657	0.06	2.0	0.99	3.50	0
		84	2070	2957	0.06	1.9	0.90	3.58	7
		78	2270	3243	0.05	1.8	0.85	3.63	13
		72	2480	3543	0.05	1.7	0.84	3.65	20
		60	2790	3986	0.04	1.8	1.02	3.49	33
20	68	90	1050	1500	0.11	5.1	0.99	3.53	0
		78	1250	1786	0.09	4.8	0.92	3.60	13
		66	1460	2086	0.08	4.7	0.97	3.57	27
		60	1670	2386	0.07	4.4	0.87	3.66	33
		54	1880	2686	0.06	4.2	0.83	3.69	40
		48	2080	2971	0.06	4.2	0.87	3.67	47
		45	2280	3257	0.05	4.0	0.82	3.71	50
25	77	84	930	1329	0.12	7.1	0.17	4.29	7
		78	1140	1629	0.10	6.1	0.03	4.37	13
		60	1350	1929	0.08	6.3	0.12	4.33	33
		54	1550	2214	0.07	6.0	0.09	4.36	40
		48	1760	2514	0.07	5.8	0.09	4.37	47
		42	2080	2971	0.06	5.3	0.05	4.39	53
		36	2280	3257	0.05	5.6	0.12	4.36	60

* Percent savings in energy cost shown are comparison based on the current energy cost of $43,920 per year
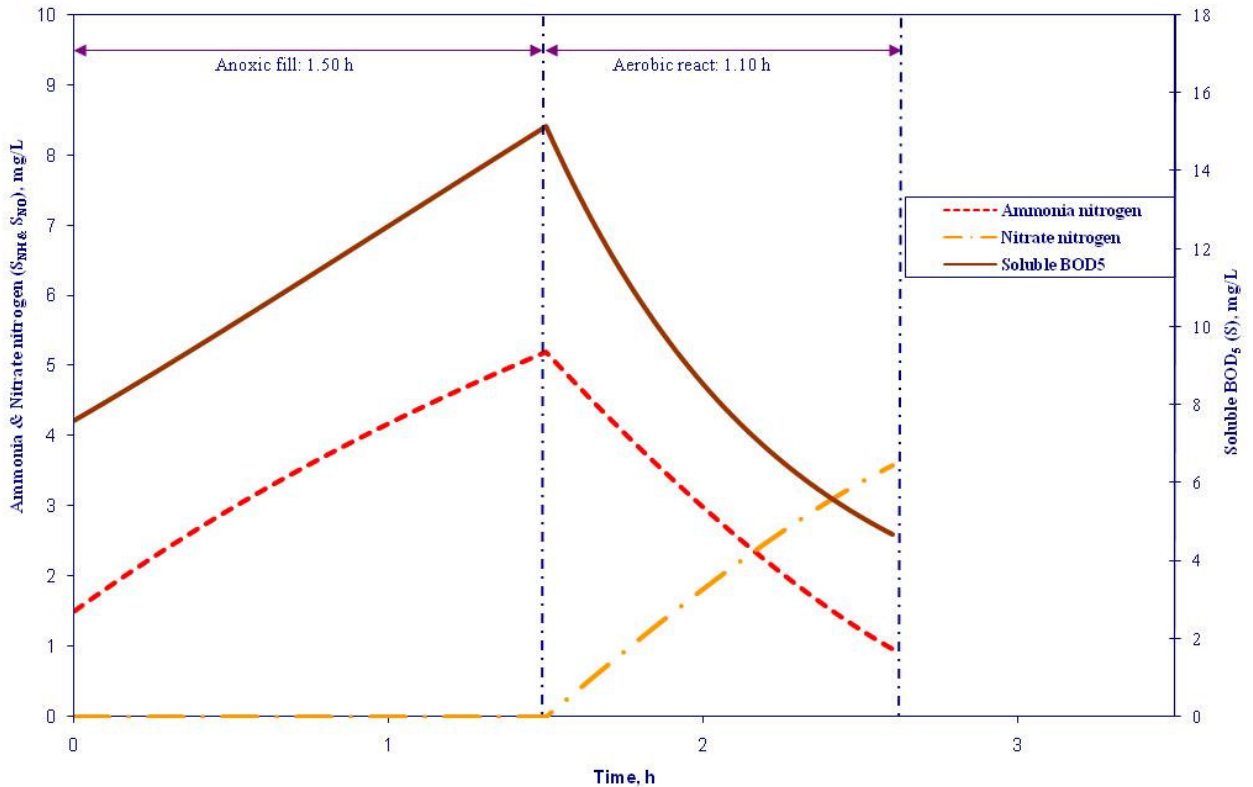
Figure 7-1: Concentration profile of soluble BOD$_5$, NH$_4$-N and NO$_3$-N for selected operational strategy.

Adjusting the aerated react time for SBR basin is simple and easy to implement, and does not constitute a major deviation from current operational practice. It may be noticed from the operation strategy chart that aerated react time could be reduced to a maximum of 36 minutes from the currently operating react time of 90 minutes depending on the solids concentration and wastewater temperature in the SBR basin. Every 5 minutes reduction in aerated react time would contribute to energy savings of about $610 per reactor per year resulting in $2,440 for four reactors at this plant.

7.3   Implementation

The City of Tahlequah is interested in implementing the results from this work that could benefit its SBR system in any manner. The operational strategy shown in Table 7-1 was discussed with the plant operational staff on September 16, 2008. This SBR system has been operated on the same time cycle (1.5 hours of anoxic fill and 1.5 hours of aerated react) for many years, and the MLVSS concentration was normally kept at a low level around 1,000 mg/L. The optimal operational policies shown in Table 7-1 suggest that the aerated react time could be reduced if the MLVSS concentration in the SBR basin is increased to a higher level. The plant operational staff is concerned about low F/M ratio and low dissolved oxygen concentration if higher MLVSS concentration is maintained in the basin. This concern was addressed subsequently by incorporating F/M ratio in the optimal operational policy chart which will alert the operational staff to avoid F/M ratio lower than 0.05. The diffusers and blowers were designed originally to provide sufficient oxygen and maintain DO of 2 mg/L. In fact, the capacity of the available blower is substantially higher, which was also confirmed during the design optimization. Addressing these two concerns made the plant staff more comfortable with implementing the changes. The plant staff are currently building up (increasing) the MLVSS concentration in one of the basins, on a trail basis, to the desired level at which they can reduce the react time. When the changes are made, the plant staff will also collect and analyze effluent samples for testing the effectiveness of these process modifications.

7.4 Conclusion

A design optimization model is useful for the design of new systems but does not provide much benefit to an existing system. However, in order to provide some practical benefits to the existing SBR system at the City of Tahlequah from which the data was collected for this research work, an optimal operational strategy was developed for reducing aeration energy cost. The developed operational strategy takes into account the current influent flow conditions and operational practices and intends not to make major, drastic modifications. The only modification required is changing the aerated react time in the SBR control system to a suitable value depending on the MLVSS concentration in the SBR basin. The plant staff is currently in the process of implementing the suggested operational strategies. Following the operational strategies outlined in this chapter could produce energy savings of $610 per reactor per year for reducing 5 minutes of aerated react time in a cycle. There are four reactors at this plant, and implementing the optimal operational policies could result in energy savings up to a maximum of $ 26,352 per year depending on the wastewater temperature and the ability to maintain higher MLVSS concentrations.

CHAPTER VIII

CONCLUSIONS AND FUTURE WORK

8.1   Conclusions

The following conclusions can be drawn based on the work reported in this investigation:

1.0   A process model with fewer parameters has been developed for removal of

organic and nitrogen substrates in the Sequencing Batch Reactor. The developed

process model was calibrated and validated with data obtained from the operation

of an existing full-scale 5.3 MGD (836 m^3/h) SBR system treating municipal

wastewater. In model calibration, an optimization technique called Simulated

Annealing was used to determine the unknown process sensitive model

parameters. Model predictions matched well with the measured data for NH_4-N

and NO_3-N. The model-predicted concentrations for NH_4-N were within 0.03

mg/L and 0.61 mg/L from experimental values, and the average deviation was

0.29 mg/L. The average deviation for NO_3-N from the experimental value was

0.23 mg/L. The model-predicted concentrations for BOD_5 were generally

between 0.5 mg/L and 1.8 mg/L except for two samples which are suspected

outliers. In general, model predictions were generally satisfactory, and described

reasonably well the trend of biodegradable organics removal, nitrification during

aeration, and denitrification during the anoxic period. The developed model, therefore, could be used successfully for design or process operation optimization.

2. A model-based design algorithm has been proposed for the design of SBR to demonstrate its cost effectiveness over the conventional design method. The proposed model-based design algorithm has been applied to a full-scale SBR system. Results from the proposed model-based design algorithm were compared with the existing design of a full-scale SBR system. Model-based design, which used the same design criteria that applied to the conventional design method, without any design optimization, produced a reactor volumetric reduction of about 33 percent and a total energy savings of $23,280 per year for the 836 m^3/h (5.3 MGD) SBR system.

3. An optimization model was developed for the design of the SBR for removal of organic and nitrogen substrates. The model-based design approach was used in the optimization model along with other considerations such as the construction and process operation costs, influent conditions, operational constraints, and effluent limitations. The developed model was applied to the existing full-scale SBR system (836 m^3/h capacity) to obtain an optimum design, and the cost savings were compared. The optimization model produced a reactor volumetric reduction of about 30 percent and annual savings of $26,410 in total construction and operation cost over 25 years life cycle of the plant.

4. A cost sensitivity analysis has been performed to study the impact of possible variations in sludge processing costs on total cost of the system, percent savings and a key process control parameter, MCRT. Cost sensitivity analysis reveals that

MCRT increases as the unit cost of sludge processing increases. However, this did not make any appreciable change in total cost savings as the optimization model attempted to find a balance among other parameters to keep the total cost to a minimum.

5. In order to provide some practical benefits to the existing SBR system, an optimal operational strategy was developed for reducing aeration energy cost with minimal modification to the current operating practice. The minimal modification requires only changing the aerated react time of SBR to a selected value from the operational strategy chart. Selection of this value is done based on the MLVSS concentration in the SBR basin and the wastewater temperature. Implementing the developed operational strategy could produce energy savings of $610 per reactor per year for reducing 5 minutes of aerated react time in a cycle. There are four reactors at this plant, and the energy savings could range from nothing to a maximum of $ 26,352 depending on the wastewater temperature and the ability to maintain higher MLVSS concentration.

The originality of this work lies in the holistic approach of producing optimal design information (volume of reactor, blower size, aeration time, length of time cycle) for a given set of constraints (process, influent and effluent constraints) taking into account both capital construction costs and operational costs. Although similar work has been done in the continuous activated sludge process, to date, cost optimization applied to design of a sequencing batch reactor for municipal wastewater treatment has not been presented. The SBR process is generally used for wastewater treatment in smaller

facilities. Results from this work substantiate that use of a model-based design approach or design optimization can reduce the SBR system cost by about 10 to 20 percent of the life cycle cost and also reduce the reactor volume by about 30 percent compared to the conventional design approach. These reductions in volume and cost are significant and can make the SBR system more attractive to larger installations.

8.2   Future Work

Sequencing Batch Reactors can also be designed for removal of phosphorus besides organic and nitrogen substrates. The removal of phosphorus requires anaerobic environment followed by aerobic environment for phosphorus release and uptake. However, the design of Sequencing Batch Reactors for removal of specific constituents from the wastewater is usually dictated by effluent limitations and cost economics. Although the City of Tahlequah has effluent limitations for phosphorus, the City installed a physico-chemical treatment system rather than operating their biological process for removal of phosphorus due to cost economics. Therefore, biological phosphorus removal could not be investigated in this work. If the concentration of phosphorus is high in the influent wastewater, it is worthwhile to extend the model developed in this work to include a phosphorus removal component into the model for future work.

In the objective function of the optimization model, the cost of sludge processing has been lumped by a unit cost factor ($ per kg of sludge processed) for simplification. In fact, sludge processing itself is an extensive process like the liquid treatment process, involving aerobic or anaerobic digestion in digesters, dewatering in dewatering

equipment coupled with polymer dosing for sludge conditioning and sludge pumps for transferring sludge from one unit to another. This provides a large scope for reduction of energy and chemical costs. Therefore, the process component of sludge treatment/processing could be considered for inclusion in the cost objective function for future work. Inclusion of the process components of solids processing would make the optimization model much more complex, but it would be worthwhile to study their influence on the design factors.

The environmental conditions such as dissolved oxygen (DO) concentration and MLVSS in the SBR basin for sampling episodes 1 and 2 are slightly different, resulting in different kinetic and stochiometric parameters. Obviously, the conditions that existed during sampling episode 1 were not near the design conditions. However, it is worthwhile to study particularly the effect of DO concentration in various phases of reaction on kinetic and stochiometric parameters and sludge settleability. The calibration technique already made available in this work can be easily used with additional data collection for determination of kinetic and stochiometric parameters. At the full-scale treatment system, the operational staff may not be willing to vary the DO concentration appreciably from design values due to concerns of potential effluent quality violations. Therefore, this study may have to be carried out at pilot scale in the laboratory.

The MATLAB programs developed in this study may have to be enhanced in the future to include graphical user interface for visualization and greater user friendliness.

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APPENDICES

Matlab Codes for Process Model

```
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% PROCESS MODEL FOR PREDICTING FINAL EFFLUENT SUBSTRATE AND BIOMASS
CONCENTRATIONS
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
Input known design parameters
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%


S_I(1,1) = 7.6; %% S0, Initial dissolved BOD5 conc. in tank, mg/L
S_I(2,1) = 1.5;  %% SNH0, Initial ammonia conc. in tank, mg/L
S_I(3,1) = 3.0; %% SNO0, Initial nitrate conc. in tank, mg/L
q = 836.0;       %% Influent average flow rate, m3/h
S_in(1,1) = 30.0; %% Sin, Influent dissolved BOD5 conc., mg/L
S_in(2,1) = 90.0; %% Spin, Influent particulate BOD5 conc., mg/L
S_in(3,1) = 17.0; %% SNHin, Influent TKN conc., mg/L
S_in(4,1) = 0.0;  %% SNOin, Influent nitrate conc., mg/L
Tw  = 20.0;       %% Wastewater temperature, degree celcius
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%  Z       = DESIGN DECISION VARIABLES
%%  Z       = [V0; tanf; tr; MCRT; MLVSS]
%%  Z(1,1)  = V0; Initial tank volume (m3)
%%  Z(2,1)  = tanf; Anoxic time interval (h)
%%  Z(3,1)  = tr; React time inteval (h)
%%  Z(4,1)  = MCRT; Mean cell residence time (d)
%%  Z(5,1)  = Initial MLVSS concentration (mg/L)
Z = [3430; 0.75; 1.75; 25; 4680]

%% Tank Initial Conditions


V0 = Z (1,1);                      %% Intial volume of the reactor, m3
Vf = Z(1,1)+(q*(24.0/(n*m)));   %% Final volume of the reactor, m3
XH0 = 0.80 * Z(5,1);            %% Initial hetro. MLVSS conc., mg/L
XA0 = 0.20 * Z(5,1);            %% Initial auto. MLVSS conc., mg/L
S0 = S_I(1,1);                  %% Initial dissolved BOD5 Conc. mg/L
SP0 = 0;                        %% Initial particulate BOD5 con. mg/L
SNH0 = S_I(2,1);                %% Initial ammonia concentration in mg/L
SNO0 = S_I(3,1);                %% Initial Nitrate concentration in mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
Time phasing


tanf = Z(2,1);                     %% Anoxic fill time interval, h
tr = Z(3,1);                       %% React time interval, h
tf = (24.0/(n*m));                 %% Total fill time, h
taef = tf – tanf;                  %% Aerobic fill time interval, h
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%


%% Influent Characteristics
XHin = 0;                %% Conc. of hetrotro. biomass in influent, mg/L
XAin = 0;                %% Conc. of autotro. biomass in influent, mg/L
Sin = S_in(1,1);         %% Influent dissolved BOD5 conc. in mg/L
```

```matlab
SPin = S_in(2,1);        %% Influent particulate BOD5 conc. in mg/L
SNHin = S_in(3,1);       %% Influent TKN concentration in mg/L
SNOin = S_in(4,1);       %% Influent nitrate concentration in mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Anoxic kinetic parmeters

vdm = 0.0004;            %% mg N-NO3 - (mg VSS day)-1
vdm = vdm *(1.12^(Tw-20.0)); %% Temperature correction
KD = 0.10;               %% mg N-NO3/L
KSD = 0.10;              %% mg BOD5/L
YD = 0.512;              %% mg VSS/mg N-NO3
w = 4.0;                 %% mg BOD5/mg N-NO3
vpm = 0.0158;            %% mg BOD5/mg VSS-h
vpm = vpm *(1.04^(Tw-20.0));    %% Temperature correction
KP = 0.095 ;             %% mg BOD5/mg VSS
KP = KP *(0.898^(Tw-20.0));     %% Temperature correction
e = 0.05;                %% mg N-NH4/mg BOD5

%% Aerobic kinetic parmeters
vnm = 0.0183;                    %% mg N-NO3 - (mg VSS day)-1
vnm = vnm *(1.12^(Tw-20.0));     %% Temperature correction
KN = 1.0;                        %% mg N-NH4/L
KO2 = 1.0;                       %% mg/L
YAuto = 0.15;                    %% mg VSS/mg N-NH4

vsm = 0.7416;                    %% mg BOD5 - (mg VSS day)-1
vsm = vsm *(1.02^(Tw-20.0));     %% Temperature correction
KS = 598.32;                     %% mg BOD5/L
YHAero = 0.55;                   %% mg VSS/mg BOD5
O2 = 2.0;
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Simulation data
N = 500;          % Simulation time intervals during anoxic fill cycle
NN = 500;         % Simulation time intervals for aerobic cycle
NNN = 500;        % Simulation time intervals for react cycle
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%%% Initialization of all variables

for i=1:N+NN+NNN+1;     %% Initialization of vector
    t(i)=0;
    V(i)=0;
    S(i)=0;
    SP(i)=0;
    SNH(i)=0;
    SNO(i)=0;
    XH(i)=0;
    XA(i)=0;
    TBOD5(i)=0;
end

S(1)=S0;
SP(1)=SP0;
```

```matlab
SNH(1)=SNH0;
SNO(1)=SNO0;
XH(1)=XH0;
XA(1)=XA0;
V(1)=V0;
t(1)=0;
TBOD5(1)=S(1)+SP(1);


%%%% Anoxic time cycle calculations

h = tanf/N;          % Time intervals

%%%% Following code solves equations (3.5) through (3.10)
%%%% using solution technique equations (3.31)

for i=1:N
    t(i+1)=t(i)+h;
    V(i+1)=V(i)+(q*h);
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    vd = vdm*SNO(i)*S(i)*(1.0/(KD+SNO(i)))*(1.0/(KD+S(i)));
    c1 = ((q/V(i))*(Sin-S(i))) - w*vd*XH(i) + vp*XH(i);
    S(i+1) = S(i) + h*c1;
    c2 = ((q/V(i))*(SPin-SP(i))) - vp*XH(i);
    SP(i+1) = SP(i) + h*c2;
    c3 = ((q/V(i))*(SNHin-SNH(i))) - e*w*vd*XH(i);
    SNH(i+1) = SNH(i) + h*c3;
    c4 = ((q/V(i))*(SNOin-SNO(i))) - vd*XH(i);
    SNO(i+1) = SNO(i) + h*c4;
    c5 = ((q/V(i))*(XHin-XH(i))) + vd*YD*XH(i);
    XH(i+1) = XH(i) + h*c5;
    XA(i+1) = XA(i);
    TBOD5(i+1)=S(i+1)+SP(i+1);
end

%%%% Aerobic filling time cycle calculations
hh = taef/NN;        % Time intervals

%%%% Following code solves equations (3.11) through (3.17)
%%%% using solution technique equations (3.31)

for i=N+1:N+1+NN
    t(i+1)=t(i)+hh;
    V(i+1)=V(i)+(q*hh);
    vs = vsm * S(i)*(1.0/(KS+S(i)));
    vn = vnm*SNH(i)*O2*(1.0/(KN+SNH(i)))*(1.0/(KO2+O2));
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    cc1 = ((q/V(i))*(Sin-S(i))) - vs*XH(i) + vp*XH(i);
    S(i+1) = S(i) + hh*cc1;
    cc2 = ((q/V(i))*(SPin-SP(i))) - vp*XH(i);
    SP(i+1) = SP(i) + hh*cc2;
    cc3 = ((q/V(i))*(SNHin-SNH(i))) - vn*XA(i) - e*vs*XH(i);
    SNH(i+1) = SNH(i) + hh*cc3;
    cc4 = ((q/V(i))*(SNOin-SNO(i))) + vn*XA(i);
    SNO(i+1) = SNO(i) + hh*cc4;
    cc5 = ((q/V(i))*(XHin-XH(i))) + vs*YHAero*XH(i);
    XH(i+1) = XH(i) + hh*cc5;
```

```matlab
    cc6 = ((q/V(i))*(XAin-XA(i))) + vn*YAuto*XA(i);
    XA(i+1) = XA(i) + hh*cc6;
    TBOD5(i+1)=S(i+1)+SP(i+1);
end

%%%% Aerobic react time cycle calculations
hhh = tr/NNN;          % Time intervals

%%%% Following codes solve equations (3.18) through (3.24)
%%%% using solution technique equations (3.31)

for i=N+NN+1:N+NN+NNN+1
    t(i+1)=t(i)+hhh;
    V(i+1)=V(i);
    vs = vsm * S(i)*(1.0/(KS+S(i)));
    vn = vnm*SNH(i)*O2*(1.0/(KN+SNH(i)))*(1.0/(KO2+O2));
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    ccc1 =  vp*XH(i) - vs*XH(i);
    S(i+1) = S(i) + hhh*ccc1;
    ccc2 = (-1.0)* vp*XH(i);
    SP(i+1) = SP(i) + hhh*ccc2;
    ccc3 = - vn*XA(i) - e*vs*XH(i);
    SNH(i+1) = SNH(i) + hhh*ccc3;
    ccc4 = vn*XA(i);
    SNO(i+1) = SNO(i) + hhh*ccc4;
    ccc5 = vs*YHAero*XH(i);
    XH(i+1) = XH(i) + hhh*ccc5;
    ccc6 = vn*YAuto*XA(i);
    XA(i+1) = XA(i) + hhh*ccc6;
    TBOD5(i+1)=S(i+1)+SP(i+1);
end
%{
%% For use if graphic display is required
figure(1)
plot (t,S)
figure(2)
plot (t,SNH)
figure(3)
plot (t,SNO)

fprintf(fid, '%12.8f\n', Z);
y = [t; S; SNH; SNO];
fprintf(fid, '%8.4f %12.8f %12.8f %12.8f\n', y);


%}


%%% Final results

S_Final(1,1) = S(N+NN+NNN);
S_Final(2,1) = SNH(N+NN+NNN);
S_Final(3,1) = SNO(N+NN+NNN);
X_Final = XH(N+NN+NNN) + XA(N+NN+NNN);
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
fclose(fid)
%}
```

APPENDIX – B

Sampling Analysis Data

FACILITY NAME: TAHLEQUAH WWTP
LOCATION: CITY OF TAHLEQUAH, OK

DATE: MAY 14, 2000

BATCH CYCLE INFORMATION

Time fill begins	=	8:50 AM
Time mixer turned on	=	8:50 AM
Time aerator turned on	=	10:20 AM
Time fill ends	=	10:20 AM
Time reaction phase starts	=	10:20 AM
Time reaction phase ends	=	11:50 AM

TANK INFORMATION

Tank ID	=	SBR 2
Tank initial height	=	9.89 FT
Tank final height when fill ends	=	12.30 FT
Flow rate	=	3.27 MGD
Filling/anoxic time	=	90 Min
Aerobic time	=	90 Min
Settling time	=	73 Min

Sample ID	Sample Location	Sample Type	Time of Sampling	Accurate Lab			OSU Stillwater			Wastewater Plant Lab			
				TBOD$_5$	Soluble BOD$_5$	TKN	NH$_3$-N	NO$_3$-N	MLVSS	DO	pH	Temp. °F	SVI
1	From Tank	Grab	8:50 AM	646	14.3		1.84	0.11	1830				
2	From Tank	Grab	9:15 AM		15.3		3.23	0.07		0.3	6.8	68.6	
3	From Tank	Grab	9:40 AM		14.7		4.77	0.06					
4	From Tank	Grab	10:05 AM		15.7		5.91	0.07					
5	From Tank	Grab	10:30 AM		16.1		6.77	0.08					
6	From Tank	Grab	10:55 AM		20.5		6.46	0.05	1540	2.2	6.92	67.5	
7	From Tank	Grab	11:20 AM		11.8		5.76	0.06					
8	From Tank	Grab	11:50 AM		6.8		4.57	0.07	1540	0.1	6.91	67.0	
9	From Tank	Grab	12:25 PM		3.7		4.1	1.38					321
10	Influent Line	Composite	For the entire fill period	175	38.1	26.2							

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FACILITY NAME: TAHLEQUAH WWTP
LOCATION: CITY OF TAHLEQUAH, OK

DATE: MAY 14, 2008

BATCH CYCLE INFORMATION

Time fill begins	=	2:50 PM
Time mixer turned on	=	2:50 PM
Time aerator turned on	=	4:20 PM
Time fill ends	=	4:20 PM
Time reaction phase starts	=	4:20 PM
Time reaction phase ends	=	5:50 PM

TANK INFORMATION

Tank ID	=	SBR 2
Tank initial height	=	9.59 FT
Tank final height when fill ends	=	12.10 FT
Flow rate	=	3.55 MGD
Filling/anoxic time	=	90 Min
Aerobic time	=	90 Min
Settling time	=	73 Min

Sample ID	Sample Location	Sample Type	Time of Sampling	Accurate Lab			OSU Stillwater			Wastewater Plant Lab			
				TBOD$_5$	Soluble BOD$_5$	TKN	NH$_3$-N	NO$_3$-N	MLVSS	DO	pH	Temp. °F	SVI
1	From Tank	Grab	2:55 AM	998	15.5		4.76	0.1	1350				
2	From Tank	Grab	3:15 AM		11.1		5.79	0.08		1.1	6.89	67.5	
3	From Tank	Grab	3:45 AM		3.8		6.53	0.08					
4	From Tank	Grab	4:05 AM		7.1		6.83	0.08					
5	From Tank	Grab	4:35 AM		14.3		7.02	0.08					
6	From Tank	Grab	4:55 AM		11.4		6.63	0.08	1360	0.9	6.79	68.1	
7	From Tank	Grab	5:20 AM		10.1		6.10	0.07					
8	From Tank	Grab	5:50 AM		1.8		6.07	0.08	1470	0.2	6.72	66.7	325
10	Influent Line	Composite	For the entire fill period	173	57.5	24.5							

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FACILITY NAME: TAHLEQUAH WWTP
LOCATION: CITY OF TAHLEQUAH, OK

DATE: MAY 15, 2008

BATCH CYCLE INFORMATION

Time fill begins	=	8:50 AM
Time mixer turned on	=	8:50 AM
Time aerator turned on	=	10:20 AM
Time fill ends	=	10:20 AM
Time reaction phase starts	=	10:20 AM
Time reaction phase ends	=	11:50 AM

TANK INFORMATION

Tank ID	=	SBR 2
Tank initial height	=	9.88 FT
Tank final height when fill ends	=	12.72 FT
Flow rate	=	4.56 MGD
Filling/anoxic time	=	90 Min
Aerobic time	=	90 Min
Settling time	=	73 Min

Sample ID	Sample Location	Sample Type	Time of Sampling	Accurate Lab TBOD$_5$	Soluble BOD$_5$	TKN	OSU Stillwater NH$_3$-N	NO$_3$-N	Wastewater Plant Lab MLVSS	DO	pH	Temp. °F	SVI
1	From Tank	Grab	8:50 AM	928	5.8		2.72	0.13	1510				
2	From Tank	Grab	9:15 AM		10.3		3.21	0.09		0.3	6.91	67.3	
3	From Tank	Grab	9:40 AM		9.8		3.75	0.07					
4	From Tank	Grab	10:05 AM		13.9		4.25	0.09					
5	From Tank	Grab	10:30 AM		9.2		4.54	0.06					
6	From Tank	Grab	10:55 AM		14.2		3.61	0.05	1510	0.1	6.99	66.8	
7	From Tank	Grab	11:20 AM		6.4		3.10	0.1					
8	From Tank	Grab	11:50 AM		8.8		2.01	0.06	1570	1.1	6.92	67.6	331
10	Influent Line	Composite	For the entire fill period	112	23.2	13.6							

Accurate
Labs & Training Center

Chain of Custody

Client Name- Tahlequah Public Works Authority

Project Name- INFLUENT, EFFLUENT

Lab Log # (Accurate Use Only)	Date Sample Taken	Time Sample Taken	Matrix (Water-W, Soil-S, Sludge-Sl, Other)	C O M P / G R A B	Client I.D. Sample Location	Field Results Time / pH / Temp °C / Cond.	Sample Preserv. & Container → No. of Container	Soluble BOD5 (P, none)	TBOD5 (P, none)	TKN (P, none)	H2SO4 (P)	H2SO4 (P)	H2SO4 (P)	none (P)
	5/14/08	9:00am-12:00pm	W	X	Raw Influent (Bottle)		3	X	X	X				
	5/14/08	9:00am-12:00pm	W	X	SBR 2		9	X						
	5/14/08	9:00am-12:00pm	W	X	SBR 2		1		X		S↓			
					*Please 5-flag	11°C/1Q								
					BID 5/15/08									

Comments:— Quote # VS- 2/7/2006 – 1276

Please fax results to Wanda Jones at (918) 431-1032. Fax is on from 8:00am to 3:30pm weekdays.
AS SOON AS POSSIBLE! Thank You. Please put P.O.# on invoice.
P.O.# 19198

Instrument Calibration

	Meter	Standards	Final Read.	Date , Time	Initials
pH		4 , 7 , 10			
Conductivity					

Sampled By: Larry Farguson

Company: TPWA

Relinquished By: (signature)	Date/Time 5/15/08 12:05pm	Received By: (signature)	Date/Time 5/15/08 12:05pm
Relinquished By: (signature)	Date/Time 5/15/08 2:5 pm	Received By: (signature)	Date/Time 5/15/08 1405
Relinquished to Lab By:	Date/Time	Received at Lab By:	Date/Time

Send Report To: ATTN: WANDA JONES
Tahlequah Public Works Authority
101 N. College P.O. Box 29
Tahlequah, OK. 74465

Address:

Phone #: (918) 456-9251   Fax #: (918) 431-1032

Send Invoice To: ATTN: MATT WILSON
and PSA Dewsberry
Ve1 M. Subramani
1350 S. Boulder
Suite 400
Tulsa OK 74119-3009

P.O. #-19198

Tahlequah Public Works Authority
625 W. Keetoowah
Tahlequah, OK 74464

Phone#: (918) 456-3591

Accurate, Inc.- 505 South Lowry Street   Phone: (405) 372-5300   ◊ Tulsa, OK   (918) 663-5400   ◊ FAX   (918) 663-6300
Stillwater, OK 74074   Fax: (405) 372-5396   ◊ Oklahoma City, OK   (405) 236-5333   ◊ Wichita, KS   (316) 683-1123

Accurate Labs & Training Center

Chain of Custody

Client Name- **Tahlequah Public Works Authority**

Project Name- **INFLUENT, EFFLUENT**

Lab Log # (Accurate Use Only)	Date Sample Taken	Time Sample Taken	Matrix (Water-W, Soil-S, Sludge-Sl, Other)	GRAB / COMP	Client I.D. Sample Location	Field Results (Time / pH / Temp °C / Cond.)	Soluble BOD5	TBOD5	TKN	No. of Container
	5/14/08	3:00pm-6:00pm	W	X	Raw Influent 1		X	X	X	3
	5/14/08	3:00pm-6:00pm	W	X	SBR 2		X	X		8
	5/14/08	3:00pm-6:00pm	W	X	SBR 2			X		1

Sample Preserv. & Container → P none / P none / P none / P H2SO4 / P H2SO4 / P none

* please J-flag
BD 5/15/08
11°C/1Q

Comments — Quote # VS- 2/7/2006 – 1276

Please fax results to Wanda Jones at (918) 431-1032. Thank You. Fax is on from 8:00am to 3:30pm weekdays. AS SOON AS POSSIBLE! Please put P.O.# on invoice.

P.O.# 19198

Sampled By: Larry Ferguson

	Signature	Date/Time
Relinquished By:	[signature]	5/15/08 12:05pm
Relinquished By:	[signature]	5/15/08 2:05pm
Relinquished to Lab By:		

Instrument Calibration

Meter	Standards	Final Read.	Date	Time	Initials
pH	4 , 7 , 10				
Conductivity					

Company: TPWA

	Signature	Date/Time
Received By:	[signature]	5/15/08 12:05pm
Received By:	[signature]	5/15/08 1405
Received at Lab By:		

Send Report To: ATTN: WANDA JONES
Tahlequah Public Works Authority
Address: 101 N. College P.O. Box 29
Tahlequah, OK. 74465
Phone #: (918) 456-9251   Fax #: (918) 431-1032

Send Invoice To: ATTN: MATT WILSON
P.O. #19198
Address: Tahlequah Public Works Authority
625 W. Keetoowah
Tahlequah, OK 74464
Phone #: (918) 456-3591   Fax #: (918)458-1916

Accurate, Inc.- 505 South Lowry Street   Phone: (405) 372-5300
Stillwater, OK 74074   Fax: (405) 372-5396

◇ Tulsa, OK   (918) 663-5400   (918) 663-6300
◇ Oklahoma City, OK (405) 236-5333   ◇ FAX   ◇ Wichita, KS (316) 683-1123

Accurate
Labs & Training Center

Chain of Custody

Client Name- **Tahlequah Public Works Authority**

Project Name- **INFLUENT, EFFLUENT**

Lab Log # (Accurate Use Only)	Date Sample Taken	Time Sample Taken	Matrix Water-W Soil - S Sludge-Sl Other	G R A B	C O M P	Client I.D. Sample Location	Field Results Time / pH / Temp °C / Cond.	Sample Preserv. & Container →	Analysis Requested → No. of Container →	Soluble BOD5 (P none)	TBOD5 (P none)	TKN (P none)	H2SO4 (P)	H2SO4 (P)	none (P)	(P)
	5/15/08	9:00am-12:00pm	W	X		Raw Influent 1			3	X	X	X				
	5/15/08	9:00am-12:00pm	W	X		SBR 2			8	X	X					
	5/15/08	9:00am-12:00pm	W	X		SBR 2			1		X					
						*Please 5-flag 8/5/13/08	11°C 11Q									

Comments: Quote # VS- 2/7/2006 – 1276

Please fax results to Wanda Jones at (918) 431-1032.
AS SOON AS POSSIBLE! Thank You.   Fax is on from 8:00am to 3:30pm weekdays.
P.O.# 19198   Please put P.O.# on invoice.

Instrument Calibration

	Standards	Final Read.	Date , Time	Initials
Meter pH	4 , 7 , 10			
Conductivity				

Sampled By: Larry Farguson

Company:TPWA

Relinquished By: _(signature)_	Date/Time 5/15/08 12:05pm	Received By: _(signature)_	Date/Time 5/15/08 12:05 ??
Relinquished By: _(signature)_	Date/Time 5/15/08 1:05pm	Received By: _Sarah Dixon_	Date/Time 5/15/08 1405
Relinquished to Lab By:	Date/Time	Received at Lab By:	Date/Time

Send Report To: ATTN: WANDA JONES
Tahlequah Public Works Authority
Address: 101 N. College P.O. Box 29
Tahlequah, OK. 74465

Phone #: (918) 456-9251   **Fax #:** (918) 431-1032

Send Invoice To: ATTN: MATT WILSON
Address: Tahlequah Public Works Authority
625 W. Keetoowah
Tahlequah, OK 74464

P.O. #19198

Phone #: (918) 456-3591   **Fax #:** (918)458-1916

Accurate, Inc.- 505 South Lowry Street   Phone: (405) 372-5300   ◊ Tulsa, OK   (918) 663-5400   ◊ FAX   (918) 663-6300
Stillwater, OK. 74074   Fax: (405) 372-5396   ◊ Oklahoma City, OK   (405) 236-5333   ◊ Wichita, KS   (316) 683-1123

130

Accurate, Inc.
Environmental & Laboratory Services

FACSIMILE

Date: _____

From: <u>Data Processing Dept.</u>

To: *(Company Name)* <u>Tahlequah</u>

Attention: <u>Wanda Jones</u> / Vel

Receiver Fax No. (918) 431-1032   / 918-587-007

No. of Pages: _____ *(Including Cover Transmittal Sheet)*

Comments:
Following is your report 8E150446

Thank You!

505 South Lowry • Stillwater, OK 74074 • Fax No. (405) 372-5396 • (405) 372-5300

Accurate
Environmental Labs

May 23, 2008

Client: City of Tahlequah PWA

101 N. College PO Box 29

Tahlequah, OK 74465

Requested By: Wanda Jones

nelap

National Environmental
Laboratory Accreditation
Program

LELAP CERT # 03039

Sample Project Name:	Influent, Effluent

Date Samples Received:	May 15, 2008	Time: 14:05	sample temp upon arrival at lab = 11°C - On Ice

Matrix: Water

Lab Log Numbers:

8E15046-01	8E15046-02	8E15046-03	8E15046-04
8E15046-05	8E15046-06	8E15046-07	8E15046-08
8E15046-09	8E15046-10	8E15046-11	8E15046-12
8E15046-13	8E15046-14	8E15046-15	8E15046-16
8E15046-17	8E15046-18	8E15046-19	8E15046-20
8E15046-21	8E15046-22	8E15046-23	8E15046-24
8E15046-25	8E15046-26	8E15046-27	8E15046-28
8E15046-29	8E15046-30	8E15046-31	

Work Order: 8E15046

Report # 8E15046-0523081418

EPA Lab ID#'s Stillwater OK00092   Tulsa OK00983   OKC OK00129   ICR OK 001

Oklahoma Certification: Stillwater WasteWater, DEQ 8316/ Drinking Water, DEQ D9602

Tulsa WasteWater, DEQ 9905 / Drinking Water, DEQ D9901

Oklahoma City WasteWater DEQ 7202

Kansas Certification: Stillwater NELAP CERT # E-10219

Louisiana Certification: LELAP # 03039

Method Reference: 40 CFR 136,261 Method for Chemical Analysis of Water and Waste
EPA-600/4-79-020, march, 1983. Test Methods for Evaluating Solid Wastes,
SW-846, Final Update III, Standard Methods 20th & 21st Edition for the
Examination of Water and Wastewater.

Analysis Reference: If qualifiers present in "Prep Info" or "Analysis Info", then analysis performed as
follows as follows: @= Tulsa Lab and * = OKC Lab. If no qualifiers present, then
analysis performed at Stillwater Lab.

Accurate Environmental Laboratories certify that the test results performed at the
Stillwater lab meet all requirements of NELAC. Any exceptions to this can be
found in the report footer or Quality Control Section of the report.

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Sample: _Raw Influent 1_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 5/14/08 12:00   **Lab Log#** 8E15046-01

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
BOD5 SM5210 B	Biochemical Oxygen Demand	175	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	38.1	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP
Kjeldahl Nitrogen SM4500NH3 B	Total Kjedahl Nitrogen	26.2	mg/L	2.00	05/20/08 09:45 CM	05/20/08 14:15 CM

Sample: _SBR2 (Bottle 2)_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 5/14/08 12:00   **Lab Log#** 8E15046-02

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	14.3	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: _SBR2 (Bottle 3)_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 5/14/08 12:00   **Lab Log#** 8E15046-03

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	15.3	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: _SBR2 (Bottle 4)_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 5/14/08 12:00   **Lab Log#** 8E15046-04

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	14.7	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: _SBR2 (Bottle 5)_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 5/14/08 12:00   **Lab Log#** 8E15046-05

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	15.7	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: _SBR2 (Bottle 6)_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 5/14/08 12:00   **Lab Log#** 8E15046-06

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	16.1	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: _SBR2 (Bottle 7)_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 5/14/08 12:00   **Lab Log#** 8E15046-07

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	20.5	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

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Sample: *SBR2 (Bottle 8)* **Location Code:** **PWSID#:**

Collection Type: Grab **Sample Time:** 5/14/08 12:00 **Lab Log#** 8E15046-08

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	11.8	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: *SBR2 (Bottle 9)* **Location Code:** **PWSID#:**

Collection Type: Grab **Sample Time:** 5/14/08 12:00 **Lab Log#** 8E15046-09

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	6.6	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: *SBR2 (Bottle 10)* **Location Code:** **PWSID#:**

Collection Type: Grab **Sample Time:** 5/14/08 12:00 **Lab Log#** 8E15046-10

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	3.7	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: *SBR2* **Location Code:** **PWSID#:**

Collection Type: Grab **Sample Time:** 5/14/08 12:00 **Lab Log#** 8E15046-11

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
BOD5 SM5210 B	Biochemical Oxygen Demand	848	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: *Raw Influent 1* **Location Code:** **PWSID#:**

Collection Type: Grab **Sample Time:** 5/14/08 18:00 **Lab Log#** 8E15046-12

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
BOD5 SM5210 B	Biochemical Oxygen Demand	173	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	57.5	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH
Kjeldahl Nitrogen SM4500NH3 E	Total Kjeldahl Nitrogen	24.5	mg/L	2.00	05/20/08 09:45 CM	05/20/08 14:15 CM

Sample: *SBR2 (Bottle 22)* **Location Code:** **PWSID#:**

Collection Type: Grab **Sample Time:** 5/14/08 18:00 **Lab Log#** 8E15046-13

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	15.5	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: *SBR2 (Bottle 23)* **Location Code:** **PWSID#:**

Collection Type: Grab **Sample Time:** 5/14/08 18:00 **Lab Log#** 8E15046-14

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	11.1	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

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Sample: _SBR2 (Bottle 24)_ **Location Code:** **PWSID#:**

Collection Type: Grab Sample Time: 5/14/08 18:00 Lab Log# 8E15046-15

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	3.8	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 25)_ **Location Code:** **PWSID#:**

Collection Type: Grab Sample Time: 5/14/08 18:00 Lab Log# 8E15046-16

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	7.1	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 26)_ **Location Code:** **PWSID#:**

Collection Type: Grab Sample Time: 5/14/08 18:00 Lab Log# 8E15046-17

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	14.3	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 27)_ **Location Code:** **PWSID#:**

Collection Type: Grab Sample Time: 5/14/08 18:00 Lab Log# 8E15046-18

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	11.4	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 28)_ **Location Code:** **PWSID#:**

Collection Type: Grab Sample Time: 5/14/08 18:00 Lab Log# 8E15046-19

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	10.1	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 29)_ **Location Code:** **PWSID#:**

Collection Type: Grab Sample Time: 5/14/08 18:00 Lab Log# 8E15046-20

Method/Parameter	Test	Result			PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	1.9	mg/L	J	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 21)_ **Location Code:** **PWSID#:**

Collection Type: Grab Sample Time: 5/14/08 18:00 Lab Log# 8E15046-21

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
BOD5 SM5210 B	Biochemical Oxygen Demand	998	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _Raw Influent 1_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-22

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
BOD5 SM5210 B	Biochemical Oxygen Demand	112	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	23.2	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP
Kjeldahl Nitrogen SM4500NH3 E	Total Kjeldahl Nitrogen	13.6	mg/L	2.00	05/20/08 09:45 CM	05/20/08 14:15 CM

Sample: _SBR2 (Bottle 42)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-23

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	5.8	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 43)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-24

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	10.3	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 44)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-25

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	9.8	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 45)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-26

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	13.9	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 46)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-27

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	9.2	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Sample: _SBR2 (Bottle 47)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-28

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	14.2	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

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Sample: _SBR2 (Bottle 48)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-29

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	6.4	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: _SBR2 (Bottle 49)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-30

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	8.8	mg/L	2.0	05/16/08 11:30 @SP	05/21/08 11:00 @SP

Sample: _SBR2 (Bottle 41)_   Location Code: PWSID#:

Collection Type: Grab   Sample Time: 5/15/08 12:00   Lab Log# 8E15046-31

Method/Parameter	Test	Result		PQL#	Prep Date & Analyst	Analysis Date & Analyst
BOD5 SM5210 B	Biochemical Oxygen Demand	928	mg/L	2.0	05/16/08 10:00 CM	05/21/08 10:00 TH

Notes and Definitions

J   Detected but below the Reporting Limit; therefore, result is an estimated concentration (CLP J-Flag).

PQL   Practical Quantitation Limit – the method detection Limit (MDL) adjusted for any dilutions or other changes made to the sample to deal with interferences/matrix effects

BPQL   Below Practical Quantitation Limit (if applicable).

The "prep date "of the analyte coincides with the 4th and 5th character of the appropriate QC batch.

Lab Manager

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Quality Control Data

Blank Data

QC Lab #	Test Group	Test	Result		PQL	Flags
S8E2123-BLK1	BOD5 SM5210 B	Biochemical Oxygen Demand	BPQL	mg/L	2.0	
S8E2318-BLK1	BOD5 SM5210 B	Biochemical Oxygen Demand	BPQL	mg/L	2.0	
S8E2124-BLK1	Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	BPQL	mg/L	2.0	
S8E2318-BLK1	Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	BPQL	mg/L	2.0	
S8E2141-BLK1	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	BPQL	mg/L	2.00	

Duplicate Sample Data

QC Lab #	Test Group	Test Name	Dup Result	Samp Result	% RPD	RPD Limit	Flags
S8E2123-DUP1	BOD5 SM5210 B	Biochemical Oxygen Demand	174	174	0.2	20	
S8E2123-DUP2	BOD5 SM5210 B	Biochemical Oxygen Demand	174	173	0.7	20	
S8E2318-DUP1	BOD5 SM5210 B	Biochemical Oxygen Demand	167	175	5	20	
S8E2318-DUP2	BOD5 SM5210 B	Biochemical Oxygen Demand	19.1	16.6	14	20	
S8E2318-DUP1	Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	BPQL	38.1	UDL	20	
S8E2318-DUP2	Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	BPQL		UDL	20	

Laboratory Control Sample Data

Lab QC#	Test Group	Test Name	LCS Result	Spike Level	Units	% Rec.	Control Limits	Flags
S8E2123-BS1	BOD5 SM5210 B	Biochemical Oxygen Demand	182	198.0	mg/L	92	84.6 - 115.4	
S8E2318-BS1	BOD5 SM5210 B	Biochemical Oxygen Demand	208	198.0	mg/L	105	84.6 - 115.4	
S8E2124-BS1	Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	182	198.0	mg/L	92	84.6 - 115.4	
S8E2318-BS1	Soluble BOD5 SM5210 B	Biochemical Oxygen Demand	208	198.0	mg/L	105	84.6 - 115.4	
S8E2141-BS2	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	20.2	20.00	mg/L	101	80 - 120	

Matrix Spike Data

QC Lab #	Test Group	Test Name	Sample Result	Units	Spike Result	Spike Level	% Rec.	Acceptance Limits	Flags
S8E2141-MS1	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	1.29	mg/L	20.5	20.00	96	80 - 120	

Matrix Spike Duplicate Data

QC Lab #	Test Group	Test Name	Sample Result	Spike Result	Spike Level	Units	% Rec.	Rec. Limits	% RPD	RPD Limit	Flags
S8E2141-MSD1	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	1.29	19.7	20.00	mg/L	92	80-120	4	20	

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Subramanian, Velmurugan

Attached you should find the data for the samples you brought in an 5/16. Please view the data and let me know if you need any reruns.

If you have any questions, please let me know.

Thanks and have a great day!

Barbara

Velmurugan Subramanian
5/17/2008
Form 2008-068

Water

Dilution Factor: NONE

SampleID	NH4-N
	ppm
1A	1.84
1B	3.23
1C	4.77
1D	5.91
1E	6.77
1F	6.45
1G	5.76
1H	4.57
1I	4.10
2A	4.78
2B	5.79
2C	6.53
2D	6.83
2E	7.02
2F	6.63
2G	6.10
2H	5.07
3A	2.72
3B	3.21
3C	3.75
3D	4.25
3E	4.54
3F	3.61
3G	3.10
3H	2.01
Duplicates	
1A	1.90
2C	6.57
3G	3.07

Dilution Factor: NONE

SampleID	NO3-N
	ppm
1M	0.11
1N	0.07
1O	0.06
1P	0.07
1Q	0.08
1R	0.05
1S	0.06
1T	0.07
1U	1.39
2M	0.10
2N	0.08
2O	0.06
2P	0.09
2Q	0.08
2R	0.08
2S	0.07
2T	0.08
3M	0.13
3N	0.09
3O	0.07
3P	0.09
3Q	0.06
3R	0.05
3S	0.10
3T	0.06
Duplicates	
1M	0.11
1U	1.36
3M	0.12

TAHLEQUAH PUBLIC WORKS AUTHORITY
Wastewater Treatment Facility
1410 E. Powell Road
Tahlequah, OK 74464
Phone: 918-456-9251
Fax: 918-431-1032

Date _5 May 08_   Fax _(918) 587-0071_

To _Vel M. Subramanian_

Pages _3_ (including coversheet)

Comments: _Here are the analyses_
for the tests on 5/14 & 5/15/08.

Larry Farguson

SAMPLING FORM

FACILITY NAME: J.M. Hicks WWTF
LOCATION: Tahlequah, OK

DATE: 5/14/08

BATCH CYCLE INFORMATION

Time fill begins
Time mixer turned on
Time aeration turned on
Time fill ends
Time reaction phase starts
Time reaction phase ends
Settling time

TANK INFORMATION

Tank ID = SBR #2
Tank initial height =
Tank final height when fill ends =

Sample ID	Sample Location	Sample Type	Time of Sampling and Sample Description	Accurate Lab			OSU Stillwater				Wastewater Plant Lab		
				TBOD₅	Soluble BOD₅	TKN	NH₃-N	NO₃-N	MLVSS	DO	pH	Temp. °F	SVI
1	From Tank	Grab	0:00 (Initial Sample, before filling begins)	21	22		2A 2M	XX	1350				
2	From Tank	Grab	0:25		23		2B 2N						
3	From Tank	Grab	0:50		24		2C 2O						
4	From Tank	Grab	1:15		25		2D 2P						
5	From Tank	Grab	1:40		26		2E 2Q						
6	From Tank	Grab	2:05		27		2F 2R	W1360	WW:1.1	6.89	67.5	
7	From Tank	Grab	2:30		28		2G 2S						
8	Effluent Line	Grab	2:55		29		2H 2T		0.9	6.79	68.1	
9	From Tank	Grab	3:20 (At End of react period)		81		2I 2U						
10	Effluent Line	Grab	4:20 (After settling from discent line)		90		2J 2V	2470	0.3	6.72	66.7	326
11	Influent Line	Composite	For the entire fill period	32	33	34	2K 2V						

Date: 5/15/08

Sample:	MLVSS	SVI	DO	pH	Temp °F
101	1570				
103	1510	—	—	—	—
104	1570	331	1.1	6.72	67.6

1580 mg/L

1815 mg/L MLVSS Inf

Aerobic 0.95 mg/L
Anoxic

Terry 67.4
pH 6.9

FACILITY NAME: J.M. Hicks WWTF
LOCATION: Tahlequah, OK

SAMPLING FORM

DATE: 5/14/08

BATCH CYCLE INFORMATION

Time fill begins =
Time mixer turned on =
Time aerator turned on =
Time fill ends =
Time reaction phase starts =
Time reaction phase ends =
Settling time =

TANK INFORMATION

Tank ID = SBR #2
Tank initial height =
Tank final height when fill ends =

Sample ID	Sample Location	Sample Type	Time of Sampling and Sample Description	Accurate Lab TBOD$_5$	Accurate Lab Soluble BOD$_5$	Accurate Lab TKN	OSU Stillwater NH$_3$-N	OSU Stillwater NO$_3$-N	OSU Stillwater MLVSS	Wastewater Plant Lab DO	Wastewater Plant Lab pH	Wastewater Plant Lab Temp. °F	Wastewater Plant Lab SVI
1	From Tank	Grab	0:00 (Initial Sample, before filling begins)	L	2	12	1A	1N	1630				
2	From Tank	Grab	0:25	3			1C	1O					
3	From Tank	Grab	0:50	4			1E	1P	DO: 0.31	6.80	66.6		
4	From Tank	Grab	1:15	5			1G	1Q					
5	From Tank	Grab	1:40	6			1I	1R					
6	From Tank	Grab	2:05	7			1K	1S					
7	From Tank	Grab	2:30	8			1M	15		2.52	6.92	67.5	
8	Effluent Line	Grab	2:55	9			1O	1T					
9	From Tank	Grab	3:20 (At End of react period)	10			1Q	1U	1540				
10	Effluent Line	Grab	4:00 (After settling from decant line)	11	13		1S	1T	1540	0.1	6.91	67.0	321
11	Influent Line	Composite	For the entire fill period	12	14		1J	1V					

157° mg/L

Accurate Environmental

FACSIMILE

Date:	~~6~~ 7/2/08
From:	KEN
To:	VEL
Attention:	
Receiver Fax No.	587-0071
No. of Pages:	11 *(Including This Cover Sheet)*

Comments:

Accurate
Environmental Labs

National Environmental
Laboratory Accreditation
Program
LELAP CERT # 03039

Sample Project Name:	Influent, Effluent
Date Samples Received:	June 19, 2008 Time: 14:30 sample temp upon arrival at lab = 1°C - On Ice
Matrix:	Water

Lab Log Numbers:

8F19038-01	8F19038-02	8F19038-03	8F19038-04
8F19038-05	8F19038-06	8F19038-07	8F19038-08
8F19038-09	8F19038-10	8F19038-11	8F19038-12
8F19038-13	8F19038-14	8F19038-15	8F19038-16
8F19038-17	8F19038-18	8F19038-19	8F19038-20
8F19038-21	8F19038-22	8F19038-23	8F19038-24
8F19038-25	8F19038-26	8F19038-27	

| Work Order: | 8F19038 |

Report # 8F19038-0702081157

EPA Lab ID#'s Stillwater OK00092   Tulsa OK00983   OKC OK00129   ICR OK 001

Oklahoma Certification: Stillwater WasteWater, DEQ 8316/ Drinking Water, DEQ D9602
Tulsa WasteWater, DEQ 9905 / Drinking Water, DEQ D9901
Oklahoma City WasteWater DEQ 7202

Kansas Certification: Stillwater NELAP CERT # E-10219

Louisiana Certification: LELAP # 03039

Method Reference: 40 CFR 136,261 Method for Chemical Analysis of Water and Waste
EPA-600/4-79-020, march, 1983. Test Methods for Evaluating Solid Wastes,
SW-846, Final Update III, 1998 Standard Methods (20th Edition) for the
Examination of Water and Wastewater.

Analysis Reference: If qualifiers present in "Prep Info" or "Analysis Info", then analysis performed as
follows as follows: @= Tulsa Lab and * = OKC Lab. If no qualifiers present, then
analysis performed at Stillwater Lab.

Accurate Environmental Laboratories certify that the test results performed at the
Stillwater lab meet all requirements of NELAC. Any exceptions to this can be
found in the report footer or Quality Control Section of the report.

Sample: *DRAFT: SBR2 Btl 1*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-01

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	14.0	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 2*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-02

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	16.8	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 3*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-03

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	19.3	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 4*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-04

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	19.3	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 5*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-05

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	23.2	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 6*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-06

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	21.4	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 7*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-07

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	17.2	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

505 S. Lowry Street   ■   Stillwater, OK 74074   ■   405-372-5300   ■   Fax: 405-372-5396

147

Sample: _DRAFT: SBR2 Btl. 8_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   **Lab Log#** 8F19038-08

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	25.6	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: _DRAFT: Raw Influent (Bottle 9)_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   **Lab Log#** 8F19038-09

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	37.5	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA
COD SM5220 D	Chemical Oxygen Demand	214	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA
Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	55.1	mg/L	2.00	06/23/08 08:55 CM	06/23/08 13:15 CM

Sample: _DRAFT: SBR2 Btl. 21_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   **Lab Log#** 8F19038-10

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	17.0	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: _DRAFT: SBR2 Btl. 22_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   **Lab Log#** 8F19038-11

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	17.3	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: _DRAFT: SBR2 Btl. 23_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   **Lab Log#** 8F19038-12

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	21.8	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: _DRAFT: SBR2 Btl. 24_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   **Lab Log#** 8F19038-13

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	15.0	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: _DRAFT: SBR2 Btl. 25_   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   **Lab Log#** 8F19038-14

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	26.1	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

505 S. Lowry Street   ■   Stillwater, OK 74074   ■   405-372-5300   ■   Fax: 405-372-5396

Sample: *DRAFT: SBR2 Btl 26*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   Lab Log# 8F19038-15

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	18.7	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 27*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   Lab Log# 8F19038-16

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	14.6	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 28*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   Lab Log# 8F19038-17

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	18.2	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: Raw Influent 29*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 2:50   Lab Log# 8F19038-18

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
COD SM5220 D	Chemical Oxygen Demand	150	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA
Soluble COD SM5220 D	Chemical Oxygen Demand	62.0	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA
Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	13.1	mg/L	2.00	06/23/08 08:55 CM	06/23/08 13:15 CM

Sample: *DRAFT: SBR2 Btl 31*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-19

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	15.4	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 32*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-20

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	18.1	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 33*   **Location Code:**   **PWSID#:**

Collection Type: Grab   Sample Time: 6/17/08 8:50   Lab Log# 8F19038-21

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	17.4	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

505 S. Lowry Street   ■ Stillwater, OK 74074   ■   405-372-5300   ■   Fax: 405-372-5396

Sample: *DRAFT: SBR2 Btl 34*   **Location Code:**   PWSID#:

Collection Type: Grab   Sample Time: 6/17/08 8:50   **Lab Log#** 8F19038-22

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	16.6	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 35*   **Location Code:**   PWSID#:

Collection Type: Grab   Sample Time: 6/17/08 8:50   **Lab Log#** 8F19038-23

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	23.5	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 36*   **Location Code:**   PWSID#:

Collection Type: Grab   Sample Time: 6/17/08 8:50   **Lab Log#** 8F19038-24

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	17.2	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 37*   **Location Code:**   PWSID#:

Collection Type: Grab   Sample Time: 6/17/08 8:50   **Lab Log#** 8F19038-25

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	20.4	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: SBR2 Btl 38*   **Location Code:**   PWSID#:

Collection Type: Grab   Sample Time: 6/17/08 8:50   **Lab Log#** 8F19038-26

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
Soluble COD SM5220 D	Chemical Oxygen Demand	25.6	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA

Sample: *DRAFT: Raw Influent 39*   **Location Code:**   PWSID#:

Collection Type: Grab   Sample Time: 6/18/08 8:50   **Lab Log#** 8F19038-27

Method/Parameter	Test	Result		PQL#	Prep Info	Analysis Info
COD SM5220 D	Chemical Oxygen Demand	119	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA
Soluble COD SM5220 D	Chemical Oxygen Demand	40.0	mg/L	5.0	06/20/08 12:00 @SA	06/20/08 15:00 @SA
Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	16.9	mg/L	2.00	06/23/08 08:55 CM	06/23/08 13:15 CM

Notes and Definitions

505 S. Lowry Street   ■   Stillwater, OK 74074   ■   405-372-5300   ■   Fax: 405-372-5396

150

A-01   High RPD due to matrix interference

PQL   Practical Quantitation Limit - the method detection Limit (MDL) adjusted for any dilutions or other changes made to the sample to deal with interferences/matrix effects

BPQL   Below Practical Quantitation Limit (if applicable).

Lab Manager

Quality Control Data

Blank Data

QC Lab #	Test Group	Test	Result		PQL	Flags
S8F2425-BLK1	Soluble COD SM5220 D	Chemical Oxygen Demand	BPQL	mg/L	5.0	
S8F2429-BLK1	COD SM5220 D	Chemical Oxygen Demand	BPQL	mg/L	5.0	
S8F2409-BLK1	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	BPQL	mg/L	1.00	

Duplicate Sample Data

QC Lab #	Test Group	Test Name	Dup Result	Samp Result	% RPD	RPD Limit	Flags
S8F2425-DUP1	Soluble COD SM5220 D	Chemical Oxygen Demand	22.0	14.0	44	20	A-01
S8F2425-DUP2	Soluble COD SM5220 D	Chemical Oxygen Demand	52.0	62.0	18	20	
S8F2429-DUP1	COD SM5220 D	Chemical Oxygen Demand	132	119	10	20	
S8F2409-DUP1	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	17.4	16.9	3	20	

Laboratory Control Sample Data

Lab QC#	Test Group	Test Name	LCS Result	Spike Level	Units	% Rec	Control Limits	Flags
S8F2425-BS1	Soluble COD SM5220 D	Chemical Oxygen Demand	1030	1000	mg/L	103	90 - 110	
S8F2425-BS2	Soluble COD SM5220 D	Chemical Oxygen Demand	90.7	100.0	mg/L	91	90 - 110	
S8F2429-BS1	COD SM5220 D	Chemical Oxygen Demand	1030	1000	mg/L	103	90 - 110	
S8F2409-BS1	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	20.1	20.00	mg/L	100	80 - 120	
S8F2409-BS2	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	19.8	20.00	mg/L	99	80 - 120	

Matrix Spike Data

QC Lab #	Test Group	Test Name	Sample Result	Units	Spike Result	Spike Level	% Rec	Acceptance Limits	Flags
S8F2425-MS1	Soluble COD SM5220 D	Chemical Oxygen Demand	15.4	mg/L	60.1	50.00	89	80 - 120	
S8F2429-MS1	COD SM5220 D	Chemical Oxygen Demand	119	mg/L	167	50.00	96	80 - 120	
S8F2409-MS1	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	1.28	mg/L	20.4	20.00	96	80 - 120	

Matrix Spike Duplicate Data

QC Lab #	Test Group	Test Name	Sample Result	Spike Result	Spike Level	Units	% Rec	Rec Limits	% RPD	RPD Limit	Flags
S8F2409-MSD1	Kjeldahl Nitrogen SM4500NH3 E	Total Kjedahl Nitrogen	1.28	20.9	20.00	mg/L	98	80-120	2	20	

505 S. Lowry Street   ■   Stillwater, OK 74074   ■   405-372-5300   ■   Fax: 405-372-5396

Accurate
Labs & Training Center

Chain of Custody

Client Name- Tahlequah Public Works Authority

Project Name- INFLUENT, EFFLUENT

Lab Log # (Accurate Use Only)	Date Sample Taken	Time Sample Taken	Matrix Water-W Soil- S Sludge-Sl Other	G R A B	C O M P	Client I.D. Sample Location	Field Results Time / . pH ./Temp °C / Cond.	Analysis Requested → No. of Container ↓	S-BOTP/D5/S COD	TBOD5/S T COD	TKN		
	6/17/08	2:50am	W	X		Raw Influent (Bott 29)		3	X	X	X		
	6/17/08	2:50am	W	X		SBR2 (Bottle 21-28)_		8	X	X			

Sample Preserv. & Container →

Comments: Quote # VS- 2/7/2006 – 1276
USE CONTAINER WITH "T" FOR TOTAL COD
Please fax results to Wanda Jones at (918) 431-1032
AS SOON AS POSSIBLE! Thank You. Fax is on from 8:00am to 3:30pm weekdays.
P.O.# 19389 Please put P.O.# on invoice. 1°C OY ICE

Sampled By: Larry Farguson

		Date/Time
Relinquished By:		6/18/08 12:30pm
Relinquished By:		6/18/08 12:30pm
Relinquished to Lab By:		Date/Time

Company: TPWA

		Date/Time
Received By:		6/18/08 12:30pm
Received By:		6/18/08 1450
Received at Lab By:		Date/Time

Instrument Calibration -

	Standards	Final Read.	Date , Time	Initials
Meter				
pH	4 , 7 , 10			
Conductivity				

Send Report To: ATTN: WANDA JONES
Tahlequah Public Works Authority
Address: 101 N. College P.O. Box 29
Tahlequah, OK. 74465

Phone #: (918) 456-9251   Fax #: (918)431-1032

Send Invoice To: ATTN: MATT WILSON   P.O. 19389

Address: Tahlequah Public Works Authority
625 W. Keetoowah
Tahlequah, OK 74464

Phone #: (918) 456-3591   Fax #: (918)458-1916

Accurate, Inc.- 505 South Lowry Street   Phone: (405) 372-5300   ◊ Tulsa, OK   (918) 663-5400   ◊ FAX   (918) 663-6300
Stillwater, OK 74074   Fax: (405) 372-5396   ◊ Oklahoma City, OK (405) 236-5333   ◊ Wichita, KS (316) 683-1123

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Accurate
Labs & Training Center

Chain of Custody

Client Name- **Tahlequah Public Works Authority**

Project Name- **INFLUENT, EFFLUENT**

Lab Log # (Accurate Use Only)	Date Sample Taken	Time Sample Taken	Matrix Water-W Soil-S Sludge-Sl Other	G R A B	C O M P	Client I.D. Sample Location	Field Results Time / pH / Temp °C / Cond.	No. of Container →	S-BOD5/S COD	BOD5/T COD	TKN
	6/17/08	8:50am	W	X		Raw Influent (Bottl 9)		3	X	X	X
	6/17/08	8:50am	W	X		SBR2 (Bottle 1-8)		8	X	X	X

Comments

Quote # VS- 2/7/2006 – 1276

USE CONTAINER WITH "1 T" FOR TOTAL COD

Please fax results to Wanda Jones at (918) 431-1032

AS SOON AS POSSIBLE! Thank You.   Fax is on from 8:00am to 3:30pm weekdays.

P.O.#   19389   Please put P.O.# on invoice.   1 °C OU ICE

Sampled By: Larry Farguson

	Date/Time		Date/Time
Relinquished By:	6/18/08 12:30am	Received By:	6/18/08 12:30am
Relinquished By:	6/18/08 14:30pm	Received By:	6/18/08 14:30pm
Relinquished to Lab By:	Date/Time	Received # Lab By:	Date/Time

Instrument Calibration -

	Meter	Standards	Final Read.	Date , Time	Initials
	pH	4 , 7 , 10			
	Conductivity				

Company: TPWA

Send Report To: ATTN: WANDA JONES
Tahlequah Public Works Authority

Address: 101 N. College P.O. Box 29
Tahlequah, OK. 74465

Phone #: (918) 456-9251   **Fax #:** (918) 431-1032

Send Invoice To: ATTN: MATT WILSON   P.O. 19389
Tahlequah Public Works Authority

Address: 625 W. Keetoowah
Tahlequah, OK 74464

Phone #: (918) 456-3591   **Fax #:** (918)458-1916

Accurate, Inc.- 505 South Lowry Street
Stillwater, OK 74074

Phone: (405) 372-5300   ◊ Tulsa, OK   (918) 663-5400   ◊ FAX   (918) 663-6300
Fax: (405) 372-5396   ◊ Oklahoma City, OK   (405) 236-5333   ◊ Wichita, KS   (316) 683-1123

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Chain of Custody

Accurate
Labs & Training Center

Client Name- **Tahlequah Public Works Authority**

Project Name- **INFLUENT, EFFLUENT**

Lab Log # (Accurate Use Only)	Date Sample Taken	Time Sample Taken	Matrix Water-W Soil - S Sludge-Sl Other	C O M P	G R A B	Client I.D. Sample Location	Field Results Time / pH / Temp °C / Cond.	No. of Container →	S. BOD/TDS / S COD	S. BOD/S T COD	TKN
	6/18/08	8:50am	W	X		Raw Influent (Bot 39)		3	X	X	X
	6/18/08	8:50am	W	X		SBR2 (Bottle 31-38)_		8	X	X	X

Sample Preserv. & Container →
KC = T LIME
KC = T LIME
500 mL

Analysis Requested →

Comments
Quote # VS- 2/7/2006 – 1276
USE CONTAINER WITH "T" FOR TOTAL COD.
Please fax results to Wanda Jones at (918) 431-1032.
AS SOON AS POSSIBLE! Thank You. Fax is on from 8:00am to 3:30pm weekdays.
P.O.# 19389   Please put P.O.# on invoice.   1°C ON ICE

Sampled By: Larry Farguson   **Company:**TPWA

Instrument Calibration -

	Standards	Final Read.	Date , Time	Initials
Meter	4 , 7 , 10			
pH				
Conductivity				

	Date/Time		Received By:	Date/Time
Relinquished By:	6/18/08 12:30 am			6/18/08 12:30 pm
Relinquished By:	6/18/18 12:30 pm		Received By:	6/18/08 1430
Relinquished to Lab By:	Date/Time		Received at Lab By:	Date/Time

Send Report To: ATTN: WANDA JONES
Tahlequah Public Works Authority
Address: 101 N. College P.O. Box 29
Tahlequah, OK. 74465

Phone #: (918) 456-9251   Fax #: (918) 431-1032

Send Invoice To: ATTN: MATT WILSON   P.O. 19389
Tahlequah Public Works Authority
Address: 625 W. Keetoowah
Tahlequah, OK 74464

Phone #: (918) 456-3591   Fax #: (918)458-1916

Accurate, Inc.- 505 South Lowry Street   Phone: (405) 372-5300   ◊ Tulsa, OK   (918) 663-5400   ◊ FAX   (918) 663-6300
Stillwater, OK 74074   Fax: (405) 372-5396   ◊ Oklahoma City, OK (405) 236-5333   ◊ Wichita, KS (316) 683-1123

Subramanian, Velmurugan

From:	OSU Soil Testing lab [soiltesting@okstate.edu]
Sent:	Thursday, July 03, 2008 9:11 AM
To:	Subramanian, Velmurugan
Subject:	Re: Sample data 2008-068
Attachments:	2008-090Sub.xls

Here you go....

Thanks
B

----- Original Message -----
From: Subramanian, Velmurugan
To: soiltesting
Sent: Wednesday, July 02, 2008 4:24 PM
Subject: RE: Sample data 2008-068

Barbara,

Do you have the results ready for the samples I delivered on June 19, 2008 ? Please send me the report if you have.
Thanks.

Vel Subramanian, P.E.,DEE
Project Engineer
PSA-Dewberry
1350 South Boulder, Suite 600
Tulsa, Oklahoma 74119-3216
918 587 7283
918 587 0071 fax
www.dewberry.com

Water

Dilution Factor: NONE

SampleID	NO3-N	NH4-N
	ppm	ppm
1	0.93	1.03
2	1.00	1.48
3	0.73	3.73
4	0.31	4.00
5	0.32	5.01
6	0.79	3.68
7	1.73	2.74
8	2.76	1.79
9	0.87	8.86
10	2.52	1.50
11	1.53	1.52
12	0.99	2.28
13	0.74	2.67
14	1.41	1.78
15	2.24	1.80
16	3.11	0.71
17	0.77	0.95
18	1.02	1.45
19	0.57	2.65
20	0.18	3.85
21	0.31	4.33
22	1.09	3.20
23	2.14	2.15
24	2.92	1.17

TAHLEQUAH PUBLIC WORKS AUTHORITY
Wastewater Treatment Facility
1410 E. Powell Road
Tahlequah, OK 74464
Phone: 918-456-9251
Fax: 918-431-1032

Date _6-20-08_   Fax _918-587-0071_

To _Vel M. Subramanian_

\# Pages _4_ (including coversheet)

Comments:_____

_____

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1st Cycle (morning)

FACILITY NAME: TAHLEQUAH WWTP
LOCATION: CITY OF TAHLEQUAH, OK

DATE: JUNE 16, 2008

BATCH CYCLE INFORMATION

Time fill begins	=	8:50 AM
Time mixer turned on	=	8:50 AM
Time aerator turned on	=	10:20 AM
Time fill ends	=	10:20 AM
Time reaction phase starts	=	10:20 AM
Time reaction phase ends	=	11:50 AM

TANK INFORMATION

Tank ID	=	SBR 2
Tank initial height	=	FT
Tank final height when fill ends	=	FT
Flow rate	=	0.00 MGD
Filling/anoxic time	=	90 Min
Aerobic time	=	90 Min
Settling time	=	73 Min

Sample ID	Sample Location	Sample Type	Time of Sampling	Accurate Lab TBOD₅	Accurate Lab Soluble BOD₅	OSU Stillwater TKN	OSU Stillwater NH₃-N	OSU Stillwater NO₃-N	OSU Stillwater MLVSS	OSU Stillwater DO	Wastewater Plant Lab pH	Wastewater Plant Lab Temp. °F	SVI
1	From Tank	Grab	8:50 AM	1	1		1	31	667				
2	From Tank	Grab	9:15 AM	2	2		2	32		1.5	7.1	70.5	
3	From Tank	Grab	9:40 AM 50	3	3		3	33					
4	From Tank	Grab	10:05 AM 1:15	4	4		4	34					
5	From Tank	Grab	10:30 AM 1:40	5	5		5	35					
6	From Tank	Grab	10:55 AM 2:05	6	6		6	36	980	1.2	7.0	71.3	
7	From Tank	Grab	11:20 AM 2:30	7	7		7	37					
8	From Tank	Grab	11:50 AM 3:00	8	8	8	8	38	1076	2.2	6.9	70.9	133
10	Influent Line	Composite	For the entire fill period	9	9	9							

Avg Temp 70.9° F
              21.6° C

200 mL
─────
1000 mL

MLVSS Avg: 907

159

2nd cycle (afternoon)

FACILITY NAME: TAHLEQUAH WWTP
LOCATION: CITY OF TAHLEQUAH, OK

DATE: JUNE / 8, 2008

BATCH CYCLE INFORMATION

Time fill begins	= 2:50 pm
Time mixer turned on	= 8:50 AM
Time aerator turned on	= 8:50 AM
Time fill ends	= 10:20 AM
Time reaction phase starts	= 10:20 AM
Time reaction phase ends	= 11:50 AM

TANK INFORMATION

Tank ID	= SBR 2
Tank initial height	= FT
Tank final height when fill ends	= FT
Flow rate	= 0.10 MGD
Filling/anoxic time	= 90 Min
Aerobic time	= 90 Min
Settling time	= 73 Min

Sample ID	Sample Location	Sample Type	Time of Sampling	Accurate Lab TBOD5	Soluble BOD5	TKN	OSU Stillwater NH3-N	NO3-N	MLVSS	Wastewater Plant Lab DO	pH	Temp. °F	SVI
1	From Tank	Grab	8:50 AM	31			17	47	847				
2	From Tank	Grab	9:15 AM 25		32		18	48		1.1	6.8	72.8	
3	From Tank	Grab	9:40 AM 50		33		19	49					
4	From Tank	Grab	10:05 AM 1:15		34		20	50					
5	From Tank	Grab	10:30 AM 1:40		35		21	51					
6	From Tank	Grab	10:55 AM 2:05		36		22	52	956	1.3	6.5	76.7	
7	From Tank	Grab	11:20 AM 2:30		37		23	53					
8	From Tank	Grab	11:50 AM 3:00		38		24	54	960	3.7	6.9	71.6	129
10	Influent Line	Composite	For the entire fill period	39	39	39	39						

AVG MLSS 72.47

22.5°C

190 ml / 1000 ml

MLSS mg: 924

160

1st cycle (morning)

FACILITY NAME: TAHLEQUAH WWTP
LOCATION: CITY OF TAHLEQUAH, OK

DATE: JUNE 19, 2008

BATCH CYCLE INFORMATION

Time fill begins	= 8:50am
Time mixer begins	= 2:50 PM
Time aerator turned on	= 2:50 PM
Time fill ends	= 4:20 PM
Time reaction phase starts	= 4:20 PM
Time reaction phase ends	= 5:50 PM

TANK INFORMATION

Tank ID	= SBR 2
Tank initial height	= FT
Tank final height when fill ends	= FT
Flow rate	= 0.00 MGD
Filling/anoxic time	= 90 Min
Aerobic time	= 90 Min
Settling time	= 73 Min

Sample ID	Sample Location	Sample Type	Time of Sampling	Accurate Lab TBOD₅	Soluble BOD₅	TKN	NH₃-N	NO₃-N	MLVSS	DO	pH	Temp. °F	SVI
1	From Tank	Grab	2:50 AM	21	21		8	39	817				
2	From Tank	Grab	3:15 AM 25	22	22		10	40		0.9	7.1	22.3	
3	From Tank	Grab	3:40 AM 50	23	23		11	41					
4	From Tank	Grab	4:05 AM 1:15	24	24		12	42					
5	From Tank	Grab	4:30 AM 1:40	25	25		13	43					
6	From Tank	Grab	4:55 AM 2:05	26	26		14	44 /1033	1.4	6.9	22.8		
7	From Tank	Grab	5:20 AM 2:20	27	27		15	45					
8	From Tank	Grab	5:50 AM 3:00	28	28		16	46 /1034	2.5	7.0	72.1	134	
10	Influent Line	Composite	For the entire fill period	29	29	29							

AVG Temp = 72.4°F / 22.4°C

AVG MLSS = 205 mL / 100 mL

Flow chart for calibration methodology using the Simulated Annealing optimization technique.

Set initial parameter vector X_i
Set $T = T_{High}$

Calculate objective function, Eq. (4.1) $f(X_i)$

Process model

Generate random change ΔX_i
$X_{i+1} = X_i + \Delta X_i$

$X_{i+1} < X_U$
$X_{i+1} > X_L$

No

Calculate $f(X_{i+1})$, Eq. (4.1)

Process model

$f(X_{i+1}) < f(X_i)$

No

Generate random number p

Accept $X_i = X_{i+1}$
Set $T = (0.8)T$

$e^{\frac{-(f(X_{i+1}) - f(X_i))}{T}} > p$

No

$T < T_{LOW}$
(OR)
$|f(X_{i+1}) - f(X_i)| < Limit$

No

Calibrated parameter vector X_i

Notations:
X = Unknown parameter vector
X_U = Upper bound for X
X_L = Lower bound for X
T = Annealing temperature
T_{High}, T_{LOW} = Upper and lower limits for T

SIMULATED ANNEALING

Algorithm:

1. Start with initial vector X_i , maximum temperature T_{High}, minimum temperature T_{Low}, Nmax. Set iteration i =1.

2. Generate random change, perturbation vector ΔX_i to create a new design vector

   $$X_{i+1} = X_i + \Delta X_i.$$

3. Calculate objective functions $f(X_i)$ and $f(X_{i+1})$. If $f(X_{i+1}) < f(X_i)$ and $f(X_{i+1})$, accept the new design vector.

4. If $f(X_{i+1}) > f(X_i)$, find the probability $e^{\frac{-(f(X_{i+1})-f(X_i))}{T}}$. If randomly generated number between 0 to 1 is less than $e^{\frac{-(f(X_{i+1})-f(X_i))}{T}}$, accept the new design vector. Otherwise, decrease the amount of random move and continue until Nmax.

5. If either step 3 or 4 is successful, reduce the temperature by 0.80 T. Continue steps 3 and 4 until any stopping criteria is met. The decrease in temperature, T reduces the probability of accepting higher objective function values than the existing lower objective function value. This is the key parameter for annealing schedule.

6. If stopping criteria satisfied, terminate. The resulting final design vector is the desired optimum vector with low objective function value.

Stopping Criteria used:

(1) $\left| f(X_{i+1}) - f(X_i) \right| < 0.001$

(2) $\left\| \dfrac{X^{*i+1} - X^{*i}}{X^{*i}} \right\| \leq 0.000001$

(3) $T < T_{LOW}$

Matlab Codes for Model Calibration

```
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%
%%  MATLAB MAIN PROGRAM FOR PROCESS CLIBRATION
%%  CASE STUDY: CITY OF TAHLEQUAH WASTEWATER TREAMENT PLANT
%%
%%  BY VELMURUGAN SUBRAMANIAN, MAY 2008
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%


%% Initial value of variables

%Z = [vsmax;Ks;YH;vpmax;KP;vnmax;KN;KO2;Yaut;Vdmax;KD;KSD;YD];


Zmin =  [0.1000; 20.000; 0.5500; 0.000050; 0.0950; 0.000050; 1.0000;
1.0000; 0.1500; 0.0100; 0.1000; 0.1000; 0.5120];
Zmax =  [10.000; 700.00; 0.5500; 5.000000; 0.0950; 3.400000; 1.0000;
1.0000; 0.1500; 0.1000; 0.1000; 0.1000; 0.5120];


fid = fopen('output901.txt', 'wt');

for i=1:13
   Z(i,1) = (Zmin(i,1) + (rand*(Zmax(i,1)-Zmin(i,1))));
 end

%% Calculates the maximum amount of allowed perturbation

for ii=1:13
    for jj=1:13
        if(ii==jj)
            D(ii,jj)= 0.005*(Zmax(ii,1)-Zmin(ii,1));
         else
            D(ii,jj)= 0;
         end
    end
end


%% Simulation Parmeters
NNmax = 15;              %% Inner equilibrium
Tmax = 0.08;              %% Outer equilibrium
Tmin = 0.008;           %% Stops outer equilibrium

%% Calculates the starting objective function value equation (4.1)

f_start = calibMSE_2cycle_13P_cy1_method10(Z)+
calibMSE_2cycle_13P_cy2_method10(Z)
Initial = f_start;
xdet_start = Z'*Z;
T = Tmax

fx = f_start; %% calibMSE_2cycle_13P_cy1_method10(Z)+
calibMSE_2cycle_13P_cy2_method10(Z);
```

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```matlab
while (T > Tmin)

    for NN=1:NNmax
        %% Performs step 2 and 3 of the simulated annealing algorithm

        [ZZ RR] = Random_generator(D,Z,Zmax,Zmin);
        fzz = calibMSE_2cycle_13P_cy1_method10(ZZ)+
      calibMSE_2cycle_13P_cy2_method10(ZZ);
        del=fzz -  fx;
        p_accept = exp(-((del)/(T)));
        r_num = rand;

        if (fzz < fx)
            Z = ZZ;
            fx = fzz;

        elseif (fzz > fx && p_accept > r_num)
            Z = ZZ;
            fx = fzz;

        end

    end

%% Calculating stopping criteria

    xdet = Z'*Z;
    test2 = abs(xdet_start - xdet);
    test1 = abs(f_start-fx);

%% Checking stopping criteria

    if ( test1 < 0.000001)
        break;
    end

    if ( test2 < 0.001)
        break;
    end

%% Temperature adjustment

    xdet_start = xdet;
    f_start = fx
    T = 0.80*T

end
Final = fx;

fprintf(fid, '%12.8f \n ', Initial);
fprintf(fid, '%12.8f \n ', Final);
fprintf(fid, '%12.8f\n', Z);

fclose(fid)
```

```matlab
%% FUNCTION THAT USES PROCESS MODEL DESCRIBED IN APPENDIX 1
%% AND CALCULATES RELATIVE MEAN SQUARE ERROR FOR CYCLE 1

function [ER] = calibMSE_2cycle_13P_cy1_method10(Z)


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% SEQUENTIAL BATCH REACTOR MODELING CALCULATING
%% MEAN SQUARE ERROR OF OBSERVED AND PREDICTED VALUES
%% VELMURUGAN SUBRAMANIAN
%% MAY 2008
%% FOUR SUBSTRATES MODEL FOR TAHLEQUAH SBR PLANT


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Tank Initial Conditions
%% Cycle 1 Sample collected on May 14, 2008

V0 = 3768;        % Intial volume of the reactor in m3
Vf = 4687;        % Final volume of the reactor in m3
XH0 = 1256;       % Initial hetrotrophic MLVSS concentration mg/L
XA0 = 314;        % Initial autotrophic MLVSS concentration mg/L
S0 = 14.30;         % Initial dissolved BOD5 Concentration mg/L
SP0 = 0;          % Initial particulate BOD5 concentration mg/L
SNH0 = 1.84;        % Initial ammonia concentration in mg/L
SNO0 = 0.11;        % Initial Nitrate concentration in mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Data that changes often
q = 613;                % Influent flow rate in m3/hr
tanf = 1.50;            % Anoxic fill time in hr
tr = 1.50;              % React time in hr

%% Influent Characteristics
XHin = 0;               % Concentration of hetrotrophic biomass in
influent in mg/L
XAin = 0;               % Concentration of autotrophic biomass in
influent in mg/L
Sin = 38.1; %41.0;        % Influent dissolved BOD5 concentration in
mg/L
SPin = 136.9; %56.6;     % Influent particulate BOD5 concentration in
mg/L
SNHin = 26.2; %21.1;    % Influent TKN concentration in mg/L
SNOin = 0.0;            % Influent nitrate concentration in mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Anoxic kinetic parmeters
%Z = [vsmax; Ks; YH; vpmax; KP; vnmax; KN; KO2; Yaut; Vdmax; KD; KSD;
YD];
%Z = [0.239; 80; 0.55; 0.153; 0.044; 0.156; 1; 1; 0.15; 0.0064; 0.1;
0.1; 0.512];

vdm = Z(10,1);
KD = Z(11,1);
KSD = Z(12,1);
```

```matlab
YD = Z(13,1);
w = 4.0;
vpm = Z(4,1);
KP = Z(5,1);
e = 0.05;

%% Aerobic kinetic parmeters
vnm = Z(6,1);
KN = Z(7,1);
KO2 = Z(8,1);
YAuto = Z(9,1);

vsm = Z(1,1);
KS = Z(2,1);
YHAero = Z(3,1);
O2 = 0.87;

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Simulation data
N = 5000;           % Simulation time intervals during anoxic fill cycle
NNN = 5000;         % Simulation time intervals for react cycle
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%%% Initialization of all variables

for i=1:N+NNN+1;    %% Intialization of vector
    t(i)=0;
    V(i)=0;
    S(i)=0;
    SP(i)=0;
    SNH(i)=0;
    SNO(i)=0;
    XH(i)=0;
    XA(i)=0;
    TBOD5(i)=0;
end

S(1)=S0;
SP(1)=SP0;
SNH(1)=SNH0;
SNO(1)=SNO0;
XH(1)=XH0;
XA(1)=XA0;
V(1)=V0;
t(1)=0;
TBOD5(1)=S(1)+SP(1);

%%%% Anoxic time cycle calculations

h = tanf/N;         % Time intervals

for i=1:N
    t(i+1)=t(i)+h;
    V(i+1)=V(i)+(q*h);
```

```matlab
        vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
        vd = vdm*SNO(i)*S(i)*(1.0/(KD+SNO(i)))*(1.0/(KD+S(i)));
        c1 = ((q/V(i))*(Sin-S(i))) - w*vd*XH(i) + vp*XH(i);
        S(i+1) = S(i) + h*c1;
        c2 = ((q/V(i))*(SPin-SP(i))) - vp*XH(i);
        SP(i+1) = SP(i) + h*c2;
        c3 = ((q/V(i))*(SNHin-SNH(i))) - e*w*vd*XH(i);
        SNH(i+1) = SNH(i) + h*c3;
        c4 = ((q/V(i))*(SNOin-SNO(i))) - vd*XH(i);
        SNO(i+1) = SNO(i) + h*c4;
        c5 = ((q/V(i))*(XHin-XH(i))) + vd*YD*XH(i);
        XH(i+1) = XH(i) + h*c5;
        XA(i+1) = XA(i);
        TBOD5(i+1)=S(i+1)+SP(i+1);
end

%%%% Aerobic react time cycle calculations
hhh = tr/NNN;        % Time intervals

for i=N+1:N+NNN+1
    t(i+1)=t(i)+hhh;
    V(i+1)=V(i);
    vs = vsm * S(i)*(1.0/(KS+S(i)));
    vn = vnm*SNH(i)*O2*(1.0/(KN+SNH(i)))*(1.0/(KO2+O2));
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    ccc1 =  vp*XH(i) - vs*XH(i);
    S(i+1) = S(i) + hhh*ccc1;
    ccc2 = (-1.0)* vp*XH(i);
    SP(i+1) = SP(i) + hhh*ccc2;
    ccc3 = - vn*XA(i) - e*vs*XH(i);
    SNH(i+1) = SNH(i) + hhh*ccc3;
    ccc4 = vn*XA(i);
    SNO(i+1) = SNO(i) + hhh*ccc4;
    ccc5 = vs*YHAero*XH(i);
    XH(i+1) = XH(i) + hhh*ccc5;
    ccc6 = vn*YAuto*XA(i);
    XA(i+1) = XA(i) + hhh*ccc6;
    TBOD5(i+1)=S(i+1)+SP(i+1);
end

%% Calculates relative mean square error for BOD5 (Term 1 of EQ (4.1))

ERRS = ((S(1)-14.30)/S(1))^2 + ((S(1388)-15.30)/S(1388))^2 + ((S(2778)-
14.70)/S(2778))^2 +...
        ((S(4165)-15.70)/S(4165))^2 + ((S(5553)-16.10)/S(5553))^2 +
((S(6941)-20.50)/S(6941))^2 +...
        ((S(8329)-11.80)/S(8329))^2 +((S(10000)-6.60)/S(10000))^2;
%%ERRS = 0.0;

%% Calculates relative mean square error for NH4-N (Term 2 of EQ (4.1))

ERRNH = ((SNH(1)-1.84)/SNH(1))^2 + ((SNH(1388)-3.23)/SNH(1388))^2 +
((SNH(2778)-4.77)/SNH(2778))^2 +...
        ((SNH(4165)-5.91)/SNH(4165))^2 + ((SNH(5553)-6.77)/SNH(5553))^2
+((SNH(6941)-6.45)/SNH(6941))^2 +...
```

```matlab
        ((SNH(8329)-5.76)/SNH(8329))^2 +((SNH(10000)-
4.57)/SNH(10000))^2;
 %%ERRNH = 0.0;

%% Calculates relative mean square error for NO3-N (Term 3 of EQ (4.1))

 ERRN0 = ((SNO(1)-0.11)/SNO(1))^2 + ((SNO(1388)-0.07)/SNO(1388))^2 +
((SNO(2778)-0.06)/SNO(2778))^2 +...
        ((SNO(4165)-0.07)/SNO(4165))^2 + ((SNO(5553)-0.08)/SNO(5553))^2
+((SNO(6941)-0.05)/SNO(6941))^2 +...
       ((SNO(8329)-0.06)/SNO(8329))^2 +((SNO(10000)-0.07)/SNO(10000))^2;

  if (ERRN0 > 1000)
      ERRN0 = 1000;
  end

%% Assigns weighing factor

ER =  (1.0* ERRS) + (2.0* ERRNH);  + (0.0 * ERRN0);




%% FUNCTION THAT USES PROCESS MODEL DESCRINED IN APPENDIX 1
%% AND CALCULATES RELATIVE MEAN SQUARE ERROR FOR CYCLE 2


function [ER] = calibMSE_2cycle_13P_cy2_method10(Z)

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% SEQUENTIAL BATCH REACTOR MODELING CALCULATING
%% MEAN SQUARE ERROR OF OBSERVED AND PREDICTED VALUES
%% VELMURUGAN SUBRAMANIAN
%% MAY 2008
%% FOUR SUBSTRATES MODEL FOR TAHLEQUAH CBR PLANT

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Tank Initial Conditions
%% Cycle 2 Sample collected on May 14, 2008 afternoon

V0 = 3768;        % Intial volume of the reactor in m3
Vf = 4610;        % Final volume of the reactor in m3
XH0 = 1114;       % Initial hetrotrophic MLVSS concentration mg/L
XA0 = 280;      % Initial autotrophic MLVSS concentration mg/L
S0 = 15.50;         % Initial dissolved BOD5 Concentration mg/L
SP0 = 0;         % Initial particulate BOD5 concentration mg/L
SNH0 = 4.78;      % Initial ammonia concentration in mg/L
SNO0 = 0.10;      % Initial Nitrate concentration in mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
% fid = fopen('output10.txt', 'wt');
```

```matlab
%Z = [vsmax;    Ks;        YH;      vpmax;    KP;       vnmax;     KN;
KO2;    Yaut;  Vdmax;   KD;        KSD;      YD   ];

Z = [0.5389; 600.0000; 0.5534; 0.027896; 0.0960; 0.000167; 1.0187;
0.9986; 0.1511; 0.0707; 0.0986; 0.0991; 0.0589];


%% Data that changes often
q = 561;            % Influent flow rate in m3/hr
tanf = 1.50;        % Anoxic fill time in hr
tr = 1.50;          % React time in hr



%% tf = (Vf-V0)/q;       % Time for filling in minutes
%% taef = tf - tanf;     % Aerobic fill time

%% Influent Characteristics
XHin = 0;                  % Concentration of hetrotrophic biomass in
influent in mg/L
XAin = 0;                  % Concentration of autotrophic biomass in
influent in mg/L
Sin = 57.50; %41.0;        % Influent dissolved BOD5 concentration in
mg/L
SPin = 115.5; %56.6;       % Influent particulate BOD5 concentration in
mg/L
SNHin = 24.50; %21.1;      % Influent TKN concentration in mg/L
SNOin = 0.0;               % Influent nitrate concentration in mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Anoxic kinetic parmeters

vdm = Z(10,1);
KD = Z(11,1);
KSD = Z(12,1);
YD = Z(13,1);
w = 4.0;
vpm = Z(4,1);
KP = Z(5,1);
e = 0.05;

%% Aerobic kinetic parmeters
vnm = Z(6,1);
KN = Z(7,1);
KO2 = Z(8,1);
YAuto = Z(9,1);

vsm = Z(1,1);
KS = Z(2,1);
YHAero = Z(3,1);
O2 = 0.73;


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Simulation data
N = 5000;           % Simulation time intervals during anoxic fill cycle
```

```matlab
%%NN = 100;         % Simulation time intervals for aerobic cycle
NNN = 5000;        % Simulation time intervals for react cycle
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%%% Initialization of all variables

for i=1:N+NNN+1;    %% Intialization of vector
    t(i)=0;
    V(i)=0;
    S(i)=0;
    SP(i)=0;
    SNH(i)=0;
    SNO(i)=0;
    XH(i)=0;
    XA(i)=0;
    TBOD5(i)=0;
end

S(1)=S0;
SP(1)=SP0;
SNH(1)=SNH0;
SNO(1)=SNO0;
XH(1)=XH0;
XA(1)=XA0;
V(1)=V0;
t(1)=0;
TBOD5(1)=S(1)+SP(1);


%%%% Anoxic time cycle calculations

h = tanf/N;        % Time intervals

for i=1:N
    t(i+1)=t(i)+h;
    V(i+1)=V(i)+(q*h);
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    vd = vdm*SNO(i)*S(i)*(1.0/(KD+SNO(i)))*(1.0/(KD+S(i)));
    c1 = ((q/V(i))*(Sin-S(i))) - w*vd*XH(i) + vp*XH(i);
    S(i+1) = S(i) + h*c1;
    c2 = ((q/V(i))*(SPin-SP(i))) - vp*XH(i);
    SP(i+1) = SP(i) + h*c2;
    c3 = ((q/V(i))*(SNHin-SNH(i))) - e*w*vd*XH(i);
    SNH(i+1) = SNH(i) + h*c3;
    c4 = ((q/V(i))*(SNOin-SNO(i))) - vd*XH(i);
    SNO(i+1) = SNO(i) + h*c4;
    c5 = ((q/V(i))*(XHin-XH(i))) + vd*YD*XH(i);
    XH(i+1) = XH(i) + h*c5;
    XA(i+1) = XA(i);
    TBOD5(i+1)=S(i+1)+SP(i+1);
end
%{
%%%% Aerobic filling time cycle calculations
%%% Aerobic filling is not used in this case for calibration
hh = taef/NN;       % Time intervals
```

```matlab
for i=N+1:N+1+NN
    t(i+1)=t(i)+hh;
    V(i+1)=V(i)+(q*hh);
    vs = vsm * S(i)*(1.0/(KS+S(i)));
    vn = vnm*SNH(i)*O2*(1.0/(KN+SNH(i)))*(1.0/(KO2+O2));
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    cc1 = ((q/V(i))*(Sin-S(i))) - vs*XH(i) + vp*XH(i);
    S(i+1) = S(i) + hh*cc1;
    cc2 = ((q/V(i))*(SPin-SP(i))) - vp*XH(i);
    SP(i+1) = SP(i) + hh*cc2;
    cc3 = ((q/V(i))*(SNHin-SNH(i))) - vn*XA(i) - e*vs*XH(i);
    SNH(i+1) = SNH(i) + hh*cc3;
    cc4 = ((q/V(i))*(SNOin-SNO(i))) + vn*XA(i);
    SNO(i+1) = SNO(i) + hh*cc4;
    cc5 = ((q/V(i))*(XHin-XH(i))) + vs*YHAero*XH(i);
    XH(i+1) = XH(i) + hh*cc5;
    cc6 = ((q/V(i))*(XAin-XA(i))) + vn*YAuto*XA(i);
    XA(i+1) = XA(i) + hh*cc6;
    TBOD5(i+1)=S(i+1)+SP(i+1);
end


%}


%%%% Aerobic react time cycle calculations
hhh = tr/NNN;         % Time intervals

for i=N+1:N+NNN+1
    t(i+1)=t(i)+hhh;
    V(i+1)=V(i);
    vs = vsm * S(i)*(1.0/(KS+S(i)));
    vn = vnm*SNH(i)*O2*(1.0/(KN+SNH(i)))*(1.0/(KO2+O2));
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    ccc1 =  vp*XH(i) - vs*XH(i);
    S(i+1) = S(i) + hhh*ccc1;
    ccc2 = (-1.0)* vp*XH(i);
    SP(i+1) = SP(i) + hhh*ccc2;
    ccc3 = - vn*XA(i) - e*vs*XH(i);
    SNH(i+1) = SNH(i) + hhh*ccc3;
    ccc4 = vn*XA(i);
    SNO(i+1) = SNO(i) + hhh*ccc4;
    ccc5 = vs*YHAero*XH(i);
    XH(i+1) = XH(i) + hhh*ccc5;
    ccc6 = vn*YAuto*XA(i);
    XA(i+1) = XA(i) + hhh*ccc6;
    TBOD5(i+1)=S(i+1)+SP(i+1);
end

%% Calculates relative mean square error for BOD5 (Term 1 of EQ (4.1))

ERRS = ((S(1)-15.50)/S(1))^2 + ((S(1388)-11.10)/S(1388))^2 +((S(2778)-
3.80)/S(2778))^2 +...
      ((S(4165)-7.10)/S(4165))^2 + ((S(5553)-14.30)/S(5553))^2
+((S(6941)-11.40)/S(6941))^2 +...
      ((S(8329)-10.10)/S(8329))^2 +((S(10000)-1.90)/S(10000))^2;
```

```matlab
%% Calculates relative mean square error for NH4-N (Term 2 of EQ (4.1))

ERRNH = ((SNH(1)-4.78)/SNH(1))^2 + ((SNH(1388)-5.79)/SNH(1388))^2
+((SNH(2778)-6.53)/SNH(2778))^2 +...
        ((SNH(4165)-6.83)/SNH(4165))^2 + ((SNH(5553)-7.02)/SNH(5553))^2
+((SNH(6941)-6.63)/SNH(6941))^2 +...
        ((SNH(8329)-6.10)/SNH(8329))^2 +((SNH(10000)-
5.07)/SNH(10000))^2;

%% Calculates relative mean square error for NO3-N (Term 3 of EQ (4.1))

 ERRN0 = ((SNO(1)-0.10)/SNO(1))^2 + ((SNO(1388)-0.08)/SNO(1388))^2
+((SNO(2778)-0.06)/SNO(2788))^2+...
        ((SNO(4165)-0.09)/SNO(4165))^2 + ((SNO(5553)-0.08)/SNO(5553))^2
+((SNO(6941)-0.08)/SNO(6941))^2+...
      ((SNO(8329)-0.07)/SNO(8329))^2 +((SNO(10000)-0.08)/SNO(10000))^2;

  if (ERRN0 > 1000)
     ERRN0 = 1000;
  end

%% Assigns weighing factor

ER =  (1.0 * ERRS)  + (2.0* ERRNH) + (0.0 * ERRN0);
```

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APPENDIX – D

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An optimization model for design and operation of sequencing batch reactor

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Abstract Sequencing batch reactors (SBR) are widely used in the treatment of municipal and industrial wastewater. However, in practice, their design is generally performed by the conventional design method. Using the conventional design method may not be cost effective for larger flows for two reasons: 1) it does not use a model-based approach for design, and 2) it does not take into account cost considerations. Most of the work found in literature focused on the operational optimization; very few studies applied optimization to the design of SBR. In this paper, an optimization model was developed for the design of SBR for removal of single pollutant, bio-degradable organics. The optimization model uses a model-based design approach and takes into consideration the construction and process operation costs, influent conditions, operational constraints, and effluent limitations. The developed model was applied to an existing wastewater treatment plant with a capacity of 7570 m^3/d to obtain the optimum design, and the cost savings were compared.

Keywords Sequencing batch reactor; optimal design; optimization; simulated annealing; model- based design.

INTRODUCTION

The sequencing batch reactor (SBR) has been widely used for treatment of municipal and industrial wastewater since its development by Irvine and co-workers (Dennis and Irvine, 1979). The SBR is perfectly suited for small wastewater flows (<10 MGD), while it performs satisfactorily even in large applications (Irvine et al., 1987). In United States most SBR installations are used for smaller wastewater treatment systems of less than two million gallons per day (US EPA, 1999). However, in recent years many larger installations are showing interest in the SBR system as it provides more operational flexibility to meet stringent effluent limits.

Currently, process designs for SBR are mostly performed by SBR equipment manufacturers using the conventional design approach based on mean cell residence time (MCRT), food to microorganism ratio (F/M ratio), mixed liquor suspended solids concentration (MLSS), and hydraulic retention time (HRT) (Geselbracht, 2007). The sizing of reactors and equipment based on this approach does not always result in a cost effective system as the designers usually use conservative design values rather than solving complex process models that predict the performance of the SBR. In the authors' experience, using the conventional design approach for SBR might result in a more economical process than other treatment options at smaller flows but not for the larger flows. Therefore, the optimization in design methodology for SBR is very important to make it more attractive to the larger installations.

Most of the works reported in the literature on SBR focused on treatability of wastes and experimental aspects of design and operation. Only a few researchers developed a systematic approach for the process analysis and dimensioning of SBR (Irvine and Ketchum, 1989; Ketchum, 1997; Artran et al.,

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2001). The application of cost optimization techniques to the design of SBR was not studied extensively although several authors focused on the optimization of SBR operation (Moreno, 1997; Katsogiannis et al., 1999; Coelho et al., 2000; Kim et al., 2001). These optimization studies were limited to minimizing the aeration energy consumption and batch time, and ignore solids processing costs. Solids processing is one of the major operational cost factors and should be given the same consideration as aeration energy cost. Recent work by Alhumaizi and Ajbar (2006) developed a design equation for cyclically operated biological reactors based on an unstructured first-order kinetic model that was used for optimization of volumetric efficiency and minimization of end-of-cycle pollutant concentration. However, the model ignores the costs associated with construction and operation.

The main objective of this work is to develop an optimization model to determine the optimal design parameters such as the reactor volume, batch time, mean cell residence time (MCRT), mixed liquor suspended solids concentration (MLSS), and waste sludge quantity to keep the total capital and operation cost of SBR system to a minimum while satisfying the effluent requirements and operational process stability criteria. The developed model was applied to a wastewater treatment plant with a capacity of 7570 m^3/d (2 MGD) to produce the optimal design, and the cost savings were compared with the existing design.

METHODS

Model Development
SBRs are a variation of the activated-sludge process. According to the U.S. EPA (1999), an SBR is no more than an activated-sludge plant that operates in time rather than space. SBR basically operates on a fill-and-draw basis. The reactor is filled during a discrete period of time and then operated as a batch reactor. After desired treatment, the mixed liquor is allowed to settle and the clarified supernatant is then drawn from the tank and discharged as a treated effluent (Irvine and Ketchum, 1989). The excess biomass produced in the tank is wasted as sludge for further treatment in the solids processing facility.

The biological process involved in the SBR can be modelled using activated sludge process models. These models range from simple to advanced, based on the components considered and hence the biochemical operations incorporated. While simple models have fewer parameters and are easier to apply, advanced models generally require software to solve the differential equations for dynamic simulations, and the level of complexity exceeds the capacity of the designer and the operator of an activated sludge system (Shahriari et al., 2006). The advanced models are much more complex for SBR considering the unsteady nature and different phases of the process. The most famous and recent advanced models are the family of models developed by the International Water Association task group on mathematical modelling for design and operation of biological wastewater treatment. These models are Activated Sludge Model No. 1 (ASM No. 1), Activated Sludge Model No. 2 (ASM No. 2), and Activated Sludge Model No. 3 (ASM No. 3). However, the advanced models have been successfully used by few researchers for process optimization (Artran et al., 2002; Sin et al., 2004). The simple models were developed based on substrate mass balances and bacterial growth kinetics (Dennis and Irvine, 1979; Irvine and Ketchum, 1989; Nakhla et al. 1997). In this work, a simple model for single substrate removal based on Monod-type kinetics has been used.

Objective function formulation
The objective function of the optimization model in this problem is the annualized cost for constructing the SBR system and the costs associated with its operation. The construction cost includes the cost of

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building the reactor (C_R), and the cost for installing blowers (C_B), and diffusers (C_D). The operation costs include the energy cost for aeration (C_{AE}) and the solids processing cost (C_{SP}). The resulting objective function to be minimized is expressed as follows:

$$Minimize \quad TC = C_R + C_B + C_D + C_{AE} + C_{SP} \tag{1}$$

The cost components listed in equation (1) are directly related to the volume of reactor, the amount of oxygen provided to the reactor and the amount of excess sludge wasted from the SBR system. The construction cost for building the reactor is a product of annualized cost factor (CF), unit cost for building the reactor (c_v, in \$/m^3), and the total volume of the reactor (V_T, in m^3).

$$C_R = CF \; V_T \; c_v \tag{2}$$

The oxygen required for the process is provided into the SBR basin through a diffused air system by blowers. The installation cost of the blower (C_B) and diffuser (C_D) is a function of process oxygen requirement (RO) and the amount of sludge wasted from the reactor (W_T). The following expressions are used to determine the process oxygen requirements and hence the blower and diffuser installation cost.

$$RO = Q(S_0 - S)(\frac{1}{f}) - 1.42 W_T \tag{3}$$

$$C_B = CF \; RO \, B_{HP} \, c_{bi} \tag{4}$$

$$C_D = CF \; RO \; c_{df} \tag{5}$$

Where RO is process oxygen requirement in kg/d, Q is influent flow rate in m^3/d, S_0 is influent substrate concentration in mg/L of BOD$_5$, S is effluent substrate concentration in mg/L of BOD$_5$, W_T is amount of sludge wasted in kg/d, f is conversion factor for BOD$_5$ to ultimate BOD, B_{HP} is blower capacity in kg O$_2$/d /kW, c_{bi} is blower installation cost in \$/kW , and c_{df} is diffuser installation cost in \$/ kg O$_2$/d.

The energy cost for providing aeration to the SBR basin is dependent on oxygen requirement, time of aeration (t_b , in h), and unit energy cost factor (c_e, in \$/kW-h).

$$C_{AE} = (365) RO \, B_{HP} \, t_b \, c_e \tag{6}$$

The solids processing cost for excess sludge produced in the process is a product of unit sludge processing cost (c_s, in \$/kg) and the amount of sludge wasted from the system (W_T, in kg/d). The excess sludge produced in the reactor must be wasted periodically to maintain a desired mean cell residence time (θ_c, in d) and mixed liquor suspended solids concentration (X , in mg/L).

$$W_T = \frac{V_T X}{\theta_c} \tag{7}$$

$$C_{SP} = (365)W_T c_s \qquad (8)$$

The objective function stated in equation (1) is evaluated using the listed equations (2) through (8). In the objective function above, the design decision variables are the volume of the reactor (V_T), effluent substrate concentration (S), the batch time (t_b), the mean cell residence time (θ_c), and the mixed liquor suspended solids concentration (X). The optimization model will determine optimum values for these design decision variables to keep the total cost to a minimum.

Constraints

The design decision variables shall meet several constraints to ensure the feasibility of process operation. These constraints arise from the process model equations, desired range for operational parameters, influent conditions, and effluent limitations. The process model equations determine the MLSS concentration (X) and effluent substrate concentration (S) in SBR at the given time, based on the growth kinetics and mass balance considerations. The process model equations listed below (Nakhla et al., 1997) provide equality constraints for the optimization problem.

During the fill time: $0 \le t \le t_f$

$$V = V_0 + \int_0^{t_f} Q\, dt \qquad (9)$$

$$\frac{dS}{dt} = \frac{Q}{V}(S_0 - S) + r_S S \qquad (10)$$

$$\frac{dX}{dt} = \frac{Q}{V}(X_0 - X) + r_X X \qquad (11)$$

During react time: $t_f \le t \le t_b$ $\qquad (12)$

$$\frac{dS}{dt} = r_S S \qquad (13)$$

$$\frac{dX}{dt} = r_X X \qquad (14)$$

$$r_S = -\frac{\mu_{max} X}{(K_S + S)Y} \qquad (15)$$

$$r_X = \frac{\mu_{max} S}{(K_S + S)} - k_d \qquad (16)$$

Where V_0 is initial volume of reactor in m^3 before the influent starts filling in, t_f is fill time in h, t_b is batch time that includes fill and react time, μ_{max} is maximum specific growth rate in h^{-1}, K_s is half-saturation coefficient in mg/L, Y is growth yield coefficient in mg/mg, and k_d is microbial decay coefficient in h^{-1}. The above set of differential equations can be solved using the finite difference method (Sewell, 1988) for a given fill and react time to determine effluent substrate concentration (S) and MLSS concentration (X).

The initial volume (V_0) of the reactor selected for the design shall be adequate to prevent solids carry over in the effluent during decanting, and the total reactor volume (V_T) shall be able to handle the peak

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flow. The initial volume that will prevent solids carry over can be expressed in terms of sludge volume index (SVI) and desired safety factor (SF). The total time for fill and react is also limited by the number of cycles the reactor is required to perform in a day. These conditions provide the following inequality constraints.

$$V_0 \geq SF\ SVI\ Q\ (S_0 - S\ )Y\theta_c\ 10^{-6} \tag{17}$$

$$Q_{avg} \leq Q \leq Q_{max} \tag{18}$$

$$t_f \leq t_b \leq t_{max} \quad for\ all\ Q \tag{19}$$

The most commonly used parameters for controlling the activated sludge process are food-to-microorganism ratio (F/M), MCRT (θ_c), and MLSS concentration (X). These control parameters provide process stability and have a specified operating range for the desired performance. The operating range for the control parameters is to be specified based on the desirable limits as recommended for the SBR process (US EPA, 1999; Metcalf and Eddy, 2003). These control parameters and their operating ranges along with the effluent limitations provide the following additional inequality constraints on the system.

$$(F/M)_{min} \leq (F/M) \leq \max(F/M)_{max} \tag{20}$$

$$\theta_{c,min} \leq \theta_c \leq \theta_{c,max} \tag{21}$$

$$X_{min} \leq X \leq X_{max} \tag{22}$$

$$S \leq S_{LIMIT} \quad for\ all\ Q \tag{23}$$

Optimization Method

The objective of the optimization is to find the minimum total cost of the system that meets the process and operational constraints, and effluent limitations as listed in the equations (9) through (23). The objective function listed in the equation (1) is a non-linear function, and is to be minimized subject to the constraints (equations 9-23). The first step to solve this type of problem (constrained non-linear programming) is to convert the constrained problem into an unconstrained problem. The penalty function method, which penalizes the infeasible solution, has been used in this study for converting the constrained problem into an unconstrained problem (Edgar et al., 2001). After formulating the penalty function, it is minimized using the selected optimization technique. The optimization technique that has been employed in this minimization problem is Simulated Annealing (SA), a meta-heuristics method. It is a random search technique which exploits an analogy between the way metal cools and freezes into a minimum energy crystalline structure (the annealing process) and the search for a minimum in a more general system. The algorithm employs a random search which not only accepts changes that decrease the objective function but also some changes that increase it with a specified probability. This feature allows the optimization to move towards the global minimum and avoids getting trapped in local minima. This optimization method is easy to implement with computers as it does not use the first derivative information of the objective function. For more details on the development of the Simulated Annealing (SA) algorithm and its implementation, the reader is directed to the references elsewhere (Brooks et al., 1995; Edgar et al., 2001).

The differential equations from the process modelling are solved using the finite difference method (Sewell, 1988). A computer program has been developed in MATLAB to integrate both the process and optimization model.

RESULTS AND DISCUSSION
The optimization model developed above has been applied to an existing SBR system to determine the optimal design parameters. The existing system was designed by conventional design methods to treat the wastewater flow of 7570 m³/d (2 MGD). The design information and reactor dimensions for the existing SBR system are shown in Table 1 as follows.

Table 1. Design data for the existing SBR system

Influent wastewater characteristics		
Average flow, Q_{avg}	7570	m³/d
Peak flow, Q_{max}	15140	m³/d
BOD₅, S_0	200	mg/L
Reactor data		
Number of reactors	2	
Total volume, V_T	3511	m³
Initial volume, V_0	2752	m³
Operational design data		
MCRT, θ_c	15.6	d
F/M ratio	0.070	
MLSS at volume V_T, X	3005	mg/L
Time for fill and react, t_b	2.76	h
Effluent requirement		
BOD₅, S	20	mg/L

The kinetic parameters, operational constraints, and cost factors used for the optimization model are shown in Table 2 as follows.

Table 2. Data used in the optimization model

Kinetic parameters			
Maximum specific growth rate, μ_{max}	0.125	h⁻¹	(Metcalf and Eddy, 2003)
Half-saturation coefficient, K_s	60	mg/L	(Metcalf and Eddy, 2003)
Growth yield coefficient, Y	0.60	mg/mg	(Metcalf and Eddy, 2003)
Microbial decay coefficient, k_d	0.0025	h⁻¹	(Metcalf and Eddy, 2003)
Operational Constraints			
F/M ratio	0.05 – 0.30		
MCRT, θ_c	10 – 30	d	
MLSS at volume V_T, X	2000 - 3500	mg/L	
Sludge volume index, SVI	140	mL/g	

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Safety factor, SF	1.25		
Maximum time for fill and react t_{max}	3.10*	h	
Cost factors			
Unit cost for reactor construction. c_v	200	$/m^3$	(Sedlak, 1991)
Blower installation cost, c_{bl}	550	$/HP	(Sedlak, 1991)
Diffuser installation cost, c_{df}	24	$/ kg O_2/d	(Sedlak, 1991)
Unit energy cost factor, c_e	0.07	$/kW-h	
CF (for 25 years at 5% interest rate)	0.0709		

* Allows 4 time cycles a day including settle, decant and idle

Using the influent conditions and effluent limitations as listed in Table 1 and other constraints as listed in Table 2, the optimization model was run to determine the optimal design parameters that keep the total cost of the SBR system to a minimum. The results from the optimization model are shown in Table 3 along with the existing plant data for comparison. The results shown in Table 3 are for one reactor.

Table 3. Comparison of results from optimization model with existing 2 MGD plant data

	Existing Plant	Optimization model
Initial reactor volume, m^3	2752	2323
Total reactor volume, m^3	3511	3082
Time for fill and react, h	2.76	3.10
Mean cell residence time, d	15.6	11.3
MLSS, mg/L	3005	2761
F/M ratio	0.070	0.090
Waste sludge, kg/d	676	753
Amortized capital construction cost		
Reactor cost, $/year	49,823	43,735
Blower & diffuser, $/year	129,230	104,710
Operation and maintenance cost		
Aeration energy, $/year	35,842	32,574
Solids processing, $/year	49,371	54,972
Total cost (Capital and O&M), $/year	264,260	235,990

* Costs shown in Table 3 are for one reactor, and the total cost would double for two reactors

The comparison of the values in Table 3 indicates that the total annual costs and the reactor volume determined from the optimization model are 10.6 percent and 12.2 percent, respectively, less than the existing plant designed using a conventional approach. This is a significant savings for larger installations. Table 3 also reveals that the optimal design value for MCRT is lower than the plant design value. Its impact on the total cost is two-fold as can be seen in Table 3. First, lower MCRT increases the solids production rate and thereby increases the solids processing cost. Secondly, lower MCRT decreases oxygen requirement resulting in reduced aeration energy cost and reduced installation

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cost for diffusers and blowers due to smaller equipment size. The inclusion of solids processing cost in the optimization model moves the determination of MCRT towards a trade-off between energy and solids processing costs. If solids processing cost is not included in the optimization model, the optimal MCRT will always be the lowest value to keep the energy cost low. This confirms the importance of including solids processing cost as an integral part of operational cost optimization studies.

CONCLUSION

The results of this study are encouraging and suggest that the life cycle cost of the SBR system can be reduced by about 10- 15%. Conservative design values are generally used to account for influent variations. For SBR systems this provides minimal benefits as they are already proven to perform well under varying influent conditions. Therefore, optimization model-based design should be used for SBR system design to make this option more attractive to large installations. The model presented above is for single substrate removal. However it can be readily extended to multiple components by including the mass balance and kinetics of the given substrate. Work is currently in progress to extend this model to include multiple components as well as anoxic cycle time. The extended model will be verified and calibrated with plant data before its use in the optimization model. This will make the model more practical and increase confidence of design engineers in moving towards model-based design and optimization.

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Matlab Codes for Optimization Model

```
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%
%%  MATLAB MAIN PROGRAM FOR SBR DESIGN OPTIMIZATION
%%  FOR REMOVAL OF BIODEGRADABLE ORGANICS AND NITROGEN
%%  CASE STUDY: CITY OF TAHLEQUAH WASTEWATER TREAMENT PLANT
%%
%%  BY VELMURUGAN SUBRAMANIAN, SEPTEMBER 2008
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%
fid = fopen('output9901.txt', 'wt');
%% Z        = DESIGN DECISION VARIABLES
%% Z        = [V0; tanf; tr; MCRT; MLVSS]
%% Z(1,1)   = V0; Initial tank volume (m3)
%% Z(2,1)   = tanf; Anoxic time interval (h)
%% Z(3,1)   = tr; React time inteval (h)
%% Z(4,1)   = MCRT; Mean cell residence time (d)
%% Z(5,1)   = Initial MLVSS concentration (mg/L)


%% Specify design range for decision variables


Zmin =      [2000; 0.75; 0.5; 10.0; 1800.0];
Zmax =      [4000; 1.50; 1.7; 30.0; 6000.0];


%% Randomly generate initial values for decision variables
%% This will generate different initial values for each run


for i=1:5
    Z(i,1) = (Zmin(i,1) + (rand*(Zmax(i,1)-Zmin(i,1))));
end


fprintf(fid, '%12.8f\n', Z);



%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%  Input known design parameters
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%


S_I(1,1) = 7.6; %% S0, Initial dissolved BOD5 conc. in tank, mg/L
S_I(2,1) = 1.5;  %% SNH0, Initial ammonia conc. in tank, mg/L
S_I(3,1) = 3.0; %% SNO0, Initial nitrate conc. in tank, mg/L
q = 836.0;      %% Influent average flow rate, m3/h
qp = 1672;      %% Influent peak flow rate, m3/h
qm = 473;       %% Influent minimum flow rate, m3/h
S_in(1,1) = 30.0; %% Sin, Influent dissolved BOD5 conc., mg/L
S_in(2,1) = 90.0; %% Spin, Influent particulate BOD5 conc., mg/L
S_in(3,1) = 17.0; %% SNHin, Influent TKN conc., mg/L
S_in(4,1) = 0.0;  %% SNOin, Influent nitrate conc., mg/L
Tw  = 20.0;       %% Wastewater temperature, degree celcius
```

```matlab
n = 4.0;              %% Number of reactors
m = 4.0;              %% Number cycles per day
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%


%% Total magnitude of change required in each variable

for ii=1:5
    for jj=1:5
        if(ii==jj)
            D(ii,jj)= 0.075*(Zmax(ii,1)-Zmin(ii,1));
         else
            D(ii,jj)= 0;
         end
    end
end


%% Simulation Parameters

Pmax = 50;
NNmax = 20;             %% Inner equilibrium
Tmax = 3000;            %% Outer equilibrium
Tmin = 0.1;             %% Stops outer equilibrium

%% Uses the process model to get the necessary input for objective
%% function calculation

[S_eff MLVSS] = sbrper_multi(qp,Tw,Z,S_I,S_in,n,m)

%% Calculates starting objective function value Eq. (6.1)

f_start = sbrobjective_cost_multi(q,Tw,Z,S_in,S_eff,MLVSS,n,m);

%% Evaluates the constraints Eq (6.12) through Eq. (6.23)

g1  = sbr_constraint1_multi(q,Z,n,m);
g2  = sbr_constraint2_multi(q,Z,S_in,S_eff,n,m,MLVSS);
g3  = sbr_constraint3_multi(q,Z,S_in,S_eff,n,m);
g4  = sbr_constraint4_multi(S_eff);
g5 = sbr_constraint5_multi(Z,n,m);
g6 = sbr_constraint6_multi(q,qp,Tw,Z,S_I,S_in,n,m);
g7 = sbr_constraint7_multi(q,qm,Z,n,m);


r = 1;

%% Calculates penalty function value

pf_start = f_start + r*(g1^2+g2^2+g3^2+g4^2+g5^2+g6^2+g7^2);
xdet_start = Z*Z';

%% Assigns temperature and penalty parameter

T = Tmax;
```

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```matlab
r = 1.0;

pfx = pf_start;
Initial_fx = f_start;    %% Report initial objective function value
Initial_pfx = pfx;   %% Report initial p objective function value
Ini_G = [g1; g2; g3; g4; g5; g6; g7]; %% Report initial constraints

P_pf_start = pfx;
T_pf_start = pfx;
for P=1:Pmax

%% Outer loop for simulated annealing and temperature adjustment

while (T > Tmin)


    %% Inner loop for minimizing penalty function and penalizing
    %% constraint violation

    for NN=1:NNmax

        [ZZ RR] = Random_generator_multi(D,Z,Zmax,Zmin);

        [S_eff MLVSS] = sbrper_multi(q,Tw,ZZ,S_I,S_in,n,m);
        fzz = sbrobjective_cost_multi(q,Tw,ZZ,S_in,S_eff,MLVSS,n,m)
        g1  = sbr_constraint1_multi(q,ZZ,n,m);
        g2  = sbr_constraint2_multi(q,ZZ,S_in,S_eff,n,m,MLVSS);
        g3  = sbr_constraint3_multi(q,ZZ,S_in,S_eff,n,m);
        g4  = sbr_constraint4_multi(S_eff);
        g5 = sbr_constraint5_multi(ZZ,n,m);
        g6 = sbr_constraint6_multi(q,qp,Tw,ZZ,S_I,S_in,n,m);
        g7 = sbr_constraint7_multi(q,qm,ZZ,n,m);
        pfzz = fzz + r*(g1^2+g2^2+g3^2+g4^2+g5^2+g6^2+g7^2);



        del=pfzz -  pfx;
        p_accept = exp(-((del)/(T)));
        r_num = rand;

        if (pfzz < pfx)
            Z = ZZ;
            pfx = pfzz;

        elseif (pfzz > pfx && p_accept > r_num)
            Z = ZZ;
            pfx = pfzz;

        end

    end

%%    xdet = Z'*Z;
%%    test2 = abs(xdet_start - xdet);
```

```matlab
        test1 = abs(T_pf_start - pfx);

        if ( test1 < 1.0)
            break;
        end

  %%  if ( test2 < 0.001)
    %%      break;
    %% end

  %%  xdet_start = xdet;
    T_pf_start =  pfx;
    T = 0.80*T


end

test2 = abs(P_pf_start - T_pf_start);

    if ( g1 == 0 && g2 == 0 && g3 == 0 && g4 == 0 && g5 == 0 && g6 == 0
&& g7 == 0 && test2 < 1.0)
        break;
    end
    P_pf_start = T_pf_start;
    r = 10.0*r;

end

Final_px = pfx;
Final_fx = fzz;
T
Final_r = r
Final_G = [g1; g2; g3; g4; g5; g6; g7]; %% Report final constraints


fprintf(fid, '%12.8f \n ', Initial_fx);
fprintf(fid, '%12.8f \n ', Initial_pfx);

fprintf(fid, '%12.8f\n', Ini_G);

fprintf(fid, '%12.8f\n', Z);
fprintf(fid, '%12.8f \n ', Final_px);
fprintf(fid, '%12.8f \n ', Final_fx);

fprintf(fid, '%12.8f\n', Final_r);
fprintf(fid, '%12.8f\n', Final_G);

fprintf(fid, '%12.8f\n', S_eff);
fprintf(fid, '%12.8f\n', MLVSS);

fclose(fid)
```

```matlab
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% FUNCTION FOR PREDICTING PROCESS PERFORMANCE
%% CALCULATES FINAL EFFLUENT SUBSTRATE AND BIOMASS CONCENTRATIONS
%% FOR USE IN CALCULATING OBJECTIVE FUNCTION AND CONSTRAINTS
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%


function [S_Final X_Final] = sbrper_multi(q,Tw,Z,S_I,S_in,n,m);

%% Tank Initial Conditions

V0 = Z (1,1);                %% Intial volume of the reactor, m3
Vf = Z(1,1)+(q*(24.0/(n*m))); %% Final volume of the reactor, m3
XH0 = 0.80 * Z(5,1);         %% Initial hetro. MLVSS conc., mg/L
XA0 = 0.20 * Z(5,1);         %% Initial auto. MLVSS conc., mg/L
S0 = S_I(1,1);               %% Initial dissolved BOD5 Concentration
mg/L
SP0 = 0;                     %% Initial particulate BOD5
concentration mg/L
SNH0 = S_I(2,1);             %% Initial ammonia concentration in mg/L
SNO0 = S_I(3,1);             %% Initial Nitrate concentration in mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%%%
%% Time phasing

tanf = Z(2,1);                   %% Anoxic fill time interval, h
tr = Z(3,1);                     %% React time interval, h
tf = (24.0/(n*m));               %% Total fill time, h
taef = tf - tanf;                %% Aerobic fill time interval, h
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%%%

%% Influent Characteristics
XHin = 0;                %% Conc. of hetrotro. biomass in influent, mg/L
XAin = 0;                %% Conc. of autotro. biomass in influent, mg/L
Sin = S_in(1,1);         %% Influent dissolved BOD5 concentration in
mg/L
SPin = S_in(2,1);        %% Influent particulate BOD5 concentration in
mg/L
SNHin = S_in(3,1);       %% Influent TKN concentration in mg/L
SNOin = S_in(4,1);       %% Influent nitrate concentration in mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%%

%% Anoxic kinetic parmeters

vdm = 0.0004;            %% mg N-NO3 - (mg VSS day)-1
vdm = vdm *(1.12^(Tw-20.0)); %% Temerature correction
KD = 0.10;              %% mg N-NO3/L
KSD = 0.10;             %% mg BOD5/L
YD = 0.512;             %% mg VSS/mg N-NO3
w = 4.0;                %% mg BOD5/mg N-NO3
vpm = 0.0158;           %% mg BOD5/mg VSS-h
vpm = vpm *(1.04^(Tw-20.0));    %% Temerature correction
KP = 0.095 ;            %% mg BOD5/mg VSS
KP = KP *(0.898^(Tw-20.0));     %% Temerature correction
e = 0.05;               %% mg N-NH4/mg BOD5
```

```matlab
%% Aerobic kinetic parmeters
vnm = 0.0183;                       %% mg N-NO3 - (mg VSS day)-1
vnm = vnm *(1.12^(Tw-20.0));        %% Temerature correction
KN = 1.0;                           %% mg N-NH4/L
KO2 = 1.0;                          %% mg/L
YAuto = 0.15;                       %% mg VSS/mg N-NH4

vsm = 0.7416;                       %% mg BOD5 - (mg VSS day)-1
vsm = vsm *(1.02^(Tw-20.0));        %% Temerature correction
KS = 598.32;                        %% mg BOD5/L
YHAero = 0.55;                      %% mg VSS/mg BOD5
O2 = 2.0;
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%%%

%% Simulation data
N = 500;          % Simulation time intervals during anoxic fill cycle
NN = 500;         % Simulation time intervals for aerobic cycle
NNN = 500;        % Simulation time intervals for react cycle
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%%%

%%% Initialization of all variables

for i=1:N+NN+NNN+1;    %% Intialization of vector
    t(i)=0;
    V(i)=0;
    S(i)=0;
    SP(i)=0;
    SNH(i)=0;
    SNO(i)=0;
    XH(i)=0;
    XA(i)=0;
    TBOD5(i)=0;
end

S(1)=S0;
SP(1)=SP0;
SNH(1)=SNH0;
SNO(1)=SNO0;
XH(1)=XH0;
XA(1)=XA0;
V(1)=V0;
t(1)=0;
TBOD5(1)=S(1)+SP(1);


%%%% Anoxic time cycle calculations

h = tanf/N;        % Time intervals

for i=1:N
    t(i+1)=t(i)+h;
    V(i+1)=V(i)+(q*h);
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
```

```matlab
    vd = vdm*SNO(i)*S(i)*(1.0/(KD+SNO(i)))*(1.0/(KD+S(i)));
    c1 = ((q/V(i))*(Sin-S(i))) - w*vd*XH(i) + vp*XH(i);
    S(i+1) = S(i) + h*c1;
    c2 = ((q/V(i))*(SPin-SP(i))) - vp*XH(i);
    SP(i+1) = SP(i) + h*c2;
    c3 = ((q/V(i))*(SNHin-SNH(i))) - e*w*vd*XH(i);
    SNH(i+1) = SNH(i) + h*c3;
    c4 = ((q/V(i))*(SNOin-SNO(i))) - vd*XH(i);
    SNO(i+1) = SNO(i) + h*c4;
    c5 = ((q/V(i))*(XHin-XH(i))) + vd*YD*XH(i);
    XH(i+1) = XH(i) + h*c5;
    XA(i+1) = XA(i);
    TBOD5(i+1)=S(i+1)+SP(i+1);
end


%%%% Aerobic filling time cycle calculations
hh = taef/NN;         % Time intervals

for i=N+1:N+1+NN
    t(i+1)=t(i)+hh;
    V(i+1)=V(i)+(q*hh);
    vs = vsm * S(i)*(1.0/(KS+S(i)));
    vn = vnm*SNH(i)*O2*(1.0/(KN+SNH(i)))*(1.0/(KO2+O2));
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    cc1 = ((q/V(i))*(Sin-S(i))) - vs*XH(i) + vp*XH(i);
    S(i+1) = S(i) + hh*cc1;
    cc2 = ((q/V(i))*(SPin-SP(i))) - vp*XH(i);
    SP(i+1) = SP(i) + hh*cc2;
    cc3 = ((q/V(i))*(SNHin-SNH(i))) - vn*XA(i) - e*vs*XH(i);
    SNH(i+1) = SNH(i) + hh*cc3;
    cc4 = ((q/V(i))*(SNOin-SNO(i))) + vn*XA(i);
    SNO(i+1) = SNO(i) + hh*cc4;
    cc5 = ((q/V(i))*(XHin-XH(i))) + vs*YHAero*XH(i);
    XH(i+1) = XH(i) + hh*cc5;
    cc6 = ((q/V(i))*(XAin-XA(i))) + vn*YAuto*XA(i);
    XA(i+1) = XA(i) + hh*cc6;
    TBOD5(i+1)=S(i+1)+SP(i+1);
end




%%%% Aerobic react time cycle calculations
hhh = tr/NNN;         % Time intervals

for i=N+NN+1:N+NN+NNN+1
    t(i+1)=t(i)+hhh;
    V(i+1)=V(i);
    vs = vsm * S(i)*(1.0/(KS+S(i)));
    vn = vnm*SNH(i)*O2*(1.0/(KN+SNH(i)))*(1.0/(KO2+O2));
    vp = vpm*(SP(i)/XH(i))*(1.0/(KP+(SP(i)/XH(i))));
    ccc1 =  vp*XH(i) - vs*XH(i);
    S(i+1) = S(i) + hhh*ccc1;
    ccc2 = (-1.0)* vp*XH(i);
    SP(i+1) = SP(i) + hhh*ccc2;
    ccc3 = - vn*XA(i) - e*vs*XH(i);
    SNH(i+1) = SNH(i) + hhh*ccc3;
```

```
    ccc4 = vn*XA(i);
    SNO(i+1) = SNO(i) + hhh*ccc4;
    ccc5 = vs*YHAero*XH(i);
    XH(i+1) = XH(i) + hhh*ccc5;
    ccc6 = vn*YAuto*XA(i);
    XA(i+1) = XA(i) + hhh*ccc6;
    TBOD5(i+1)=S(i+1)+SP(i+1);
end


%%% Final results

S_Final(1,1) = S(N+NN+NNN);
S_Final(2,1) = SNH(N+NN+NNN);
S_Final(3,1) = SNO(N+NN+NNN);
X_Final = XH(N+NN+NNN) + XA(N+NN+NNN);
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% SUB PROGRAM FOR ESTIMATING TOTAL ANNUALIZED CAPITAL & OPERATION COST
%% OBJECTIVE FUNCTION EQUATION (6.1)
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [TC] = sbrobjective_cost_multi(q,Tw,Z,S_in,S_eff,MLVSS,n,m)

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% COST & OTHER PARAMTERS - NEEDS TO BE MODIFIED AS NECESSARY
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
cv = 350;            %% Cost of reactor in $/m3
ce = 0.07;            %% Cost of energy in $/KW-Hr
cs = 0.75;            %% Sludge processing cost $/kg
irate = 0.05;         %% Interest rate in fraction
nyear = 25;           %% Number of years for annualization
cblower = 1310;        %% Blower cost$/kW
cdiffuser = 1060;     %% Diffuser cost / kg O2/h
f = 0.5;              %% 5day BOD to ultimate BOD factor

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% DIFFUSER AND BLOWER DATA - NEEDS TO BE MODIFIED AS NECESSARY
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
beta = 0.98;
Csat20 = 10.39;
Psite = 14.29;
Pstd = 14.7;
CsurfT = 8.83;
DO = 2.0;
alpha = 0.55;
theta = 1.024;
SOTE = 0.32;
Pdel = 22.79;
a_n = 0.283;
R = 8.314;
eff = 0.70;
Tair = 298;
Yield = 0.55;
```

```matlab
Kd = 0.05;
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

%% Calculation of annualized cost factor: CF
CF = ((irate*(1+irate)^nyear))/(((1+irate)^nyear)-1);

%% Calculation of reactor cost Eq. (6.2): CR

VT = Z(1,1)+(q*(24.0/(n*m)));
CR = CF*VT*cv;

%%  Calculation of O2 requirement Eq. (6.4) through Eq (6.6)
O2_BOD = (24.0/1000.0)*(1.0/n)*q*(1.0/f)*((S_in(1,1)+S_in(2,1))-
S_eff(1,1));
O2_N   = (24.0/1000.0)*(1.0/n)*4.57*q*(S_in(3,1)-S_eff(2,1));
S_Yield = Yield/(1.0+(Kd*Z(4,1)));
WT = (24.0/1000.0)*(1.0/n)*q*S_Yield*((S_in(1,1)+S_in(2,1))-
S_eff(1,1));
O2_Sludge = 1.42*WT;
AOR = O2_BOD + O2_N - O2_Sludge;    %% O2 requirement in kg/d per
reactor

%% Calculation of AOR to SOR using diffuser parmeters
Num_1 = (beta*Csat20*(Psite/Pstd)*(CsurfT/9.07))-DO;
Num_2 = alpha*(theta^(Tw-20.0));
AOR_to_SOR = (Num_1 * Num_2)*(1.0/Csat20);

SOR = AOR * (1.0/AOR_to_SOR);  %% Std O2 requirement in kg/d per
reactor
tae = ((24.0/(n*m))-Z(2,1)+Z(3,1)); %% Aeration time per reactor per
cycle
SOR = SOR * (1.0/(tae*m));        %% Std O2 requirement in kg/h per
reactor

%% Blower capacity estimate Eq. (6.7)
wair = (SOR/(60.0*60.0*0.232*SOTE));
Num_3 = (((Pdel/Psite)^a_n)-1.0);
Eb = wair*R*Tair*Num_3*(1.0/(29.7*a_n*eff));  %% Blower capacity in kW

CD = CF*SOR*cdiffuser;          %% Eq. (6.8) Diffuser installation cost
CB = CF*Eb*cblower;             %% Eq. (6.3) Blower installation cost
CAE = 365*Eb*tae*m*ce;          %% Eq. (6.9) Aeration energy cost
CSP = 365*WT*cs;                %% Eq. (6.11) Sludge processing cost
Capital = CD+CB;
OM = CAE+CSP;

TC = CR+CD+CB+CAE+CSP;          %% Total installation and operation cost
                               %% Eq. (6.1)


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
```

```matlab
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% EVALUATION OF CONSTRAINT 1: HYDRAULIC RETENTION TIME EQ. (6.14)
%% This function checks for violation of the design criteria for
%% Hydraulic Retention Time and penalizes the objective function
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [cg1] = sbr_constraint1_multi(q,Z,n,m)

taef = (24.0/(n*m));              %% Fill time at average flow
VT = Z(1,1)+(q*taef);              %% Total volume of each reactor
HRT = (n*VT/q);                   %% Hydraulic Retention Time

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%Internal Data to be modified if necessary
HRTmax = 24.00;                    %% Upper limit for HRT – 24 hrs
HRTmin = 6.00;                     %% Lower limit for HRT – 6 hrs
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

if (HRT < HRTmax && HRT > HRTmin)
    cg1 = 0;
elseif (HRT > HRTmax)
    cg1 = HRT – HRTmax;        %% Penalty for exceeding upper limit
else (HRT < HRTmin)
    cg1 = HRTmin – HRT;        %% Penalty for lower than lower limit
end

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%




%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% EVALUATION OF CONSTRAINT 2: F/M RATIO: EQUATION (6.15)
%% This function checks for violation of the design criteria for
%% F/M Ratio and penalizes the objective function
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [cg2] = sbr_constraint2_multi(q,Z,S_in,S_eff,n,m,MLVSS)

taef = (24.0/(n*m));              %% Fill time at average flow, m3/h
VT = Z(1,1)+(q*taef);              %% Total volume of each reactor, m3
S_total = S_in(2,1)+ S_in(3,1);   %% Total inf. BOD5 concentration, mg/L
                                  %% F/M Ratio
F_MRatio = (24.0*q*(S_total – S_eff(1,1)))/(n*VT*MLVSS);

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%%Internal Data to be modified if necessary
F_MRatiomax = 0.20;              %% Upper limit for F/M ratio – 0.20
F_MRatiomin = 0.05;              %% Lower limit for F/M ratio – 0.05
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

if (F_MRatio < F_MRatiomax && F_MRatio > F_MRatiomin)
    cg2 = 0;
elseif (F_MRatio > F_MRatiomax)
    cg2 = F_MRatio – F_MRatiomax;   %% Penalty for exceeding upper
limit
```

```matlab
else (F_MRatio < F_MRatiomin)
    cg2 = F_MRatiomin - F_MRatio;    %% Penalty for lower than lower
limit
end


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%




%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% EVALUATION OF CONSTRAINT 3: INITIAL VOLUME: EQUATION (6.12)
%% This function checks for adequacy of initial volume to prevent
%% solids carry over and penalizes the objective function accordingly
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [cg3] = sbr_constraint3_multi(q,Z,S_in,S_eff,n,m)

taef = (24.0/(n*m));              %% Fill time at average flow, m3/h
VT = Z(1,1)+(q*taef);             %% Total volume of each reactor, m3
S_total = S_in(2,1)+ S_in(1,1);   %% Total inf. BOD5 concentration, mg/L
S_eff(1,1)
S_rem = (S_total - S_eff(1,1));   %% BOD5 removed, mg/L

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% Internal data to be modified if necessary
SVI = 150.0;                      %% Sludge volume index, mL/g
SF = 2.5;                         %% Desired safety facor
Y = 0.55;                         %% Growth yield coefficient, mg/mg
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

V_ini = SF*SVI*(1/n)*q*24.0*S_rem*Y*Z(4,1)*(1.0/1000000.0);
                                  %% Calculate initial volume required

if (Z(1,1) == V_ini || Z(1,1) > V_ini )
    cg3 = 0;
else
    cg3 = V_ini - Z(1,1);    %% Penalty for inadequate volume
end


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%




%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% EVALUATION OF CONSTRAINT 4: EFFLUENT LIMITS: EQUATIONS (6.20)
%% THROUGH (6.22) FOR AVERAGE FLOW
%% This function checks for compliance with effluent limits
%% and penalizes the objective function accordingly
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [cg4] = sbr_constraint4_multi(S_eff)

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% Internal data to be modified if necessary
S_BODLimit = 7.6;                 %% Effluent limit for BOD5, mg/L
```

```matlab
S_NHLimit = 1.5;                  %% Effluent limit for NH4-N, mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%


if (S_eff(1,1) < S_BODLimit)
    cg4 = 0;
else
    cg4 = (S_eff(1,1)-S_BODLimit);  %% Penalty for BOD5 non compliance
end

if (S_eff(2,1) < S_NHLimit)
    cg4 = cg4 + 0;
else
    cg4 = cg4 + (S_eff(2,1)- S_NHLimit);
end                                 %% Penalty for NH4-N non compliance


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%




%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% EVALUATION OF CONSTRAINT 5: BATCH TIME: EQUATION (6.13)
%% This function checks for total batch time to be within
%% allowable batch time and penalizes the objective function
%% accordingly
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [cg5] = sbr_constraint5_multi(Z,n,m)

taef = (24.0/(n*m));            %% Fill time at average flow, h
t_b = taef + Z(3,1);        %% Total batch time, h

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% Internal data to be modified if necessary
t_bLimit = 3.2;                   %% Maximum allowable batch time, h
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

if (t_b < t_bLimit)
    cg5 = 0;
else
    cg5 = (t_b - t_bLimit);  %% Penalty for exceeding time limit
end


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%




%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% EVALUATION OF CONSTRAINT 6: EFFLUENT LIMITS AT PEAK FLOW: EQUATIONS
%% (6.20) THROUGH (6.22)
%% This function checks for compliance of effluent limits at peak flow
%% and penalizes the objective function accordingly
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [cg6] = sbr_constraint6_multi(q,qp,Tw,Z,S_I,S_in,n,m)
```

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```matlab
ZY(1,1) = Z(1,1);                    %% Initial volume
ZY(4,1) = Z(4,1);                     %% MCRT
ZY(5,1) = Z(5,1);                    %% MLVSS
ZY(3,1) = Z(3,1);                    %% React time

VT = Z(1,1)+(q*(24.0/(n*m)));    %% Total volume of each reactor
t_fp = (VT-Z(1,1))*(1/qp);       %% Fill time at peak flow

if (t_fp < ZY(2,1))             %% Reassigning anoxic time based on
    ZY(2,1) = t_fp;             %% fill time
else
    ZY(2,1) = Z(2,1);
end

[S_eff_pe MLVSS_pe] = sbrper_multi(qp,Tw,ZY,S_I,S_in,n,m);


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% Internal data to be modified if necessary
S_BODLimit = 7.6;                %% Effluent limit for BOD5, mg/L
S_NHLimit = 1.5;                 %% Effluent limit for NH4-N, mg/L
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

if (S_eff_pe(1,1) < S_BODLimit)
    cg6 = 0;
else
    cg6 = (S_eff_pe(1,1)-S_BODLimit);  %% Penalty for BOD5 non
compliance
end

if (S_eff_pe(2,1) < S_NHLimit)
    cg6 = cg6 + 0;
else
    cg6 = cg6 + (S_eff_pe(2,1)- S_NHLimit);
end                                  %% Penalty for NH4-N non compliance


%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%




%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% EVALUATION OF CONSTRAINT 7: FILL TIME DURING MINIMUM FLOW: EQ (6.23)
%% This function checks for fill time during minimum flow is less than
%% average flow batch time and penalizes the objective function
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [cg7] = sbr_constraint7_multi(q,qm,Z,n,m)

VT = Z(1,1)+(q*(24.0/(n*m)));    %% Total volume of each reactor
t_fm = (VT-Z(1,1))*(1/qm);       %% Fill time at peak flow

t_bLimit = (24.0/(n*m))+Z(3,1);  %% Batch time during average flow, h
```

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```matlab
if (t_fm < t_bLimit)
    cg7 = 0;
else
    cg7 = (t_fm - t_bLimit);   %% Penalty for exceeding time limit
end



%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%



%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%% FUNCTION FOR MAKING RANDOM MOVEMENTS IN DESIGN VECTOR FOR USE
%% IN SIMULATED ANNEALING
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

function [YY YR] = Random_generator_multi(DD,Y,Ymax,Ymin)

UU = [0; 0; 0; 0; 0];

for i=1:5
    A = 1.0;

    for k = 1:100
        UU(i,1) =  (-1.0+ (2.0*rand));
        YY (i,1) = Y(i,1) + A * DD(i,i)* UU(i,1);
        if ( YY(i,1) < Ymax(i,1) && YY(i,1) > Ymin(i,1))
            break;
        end

    end
end
YY;
YR = DD*UU;
YR = diag(YR);
```

VITA

Velmurugan Subramanian

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Doctor of Philosophy

Dissertation: SEQUENCING BATCH REACTOR FOR ORGANICS AND
NITROGEN REMOVAL: MODELING AND DESIGN OPTIMIZATION

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Pages in Study: 197   Candidate for the Degree of Doctor of Philosophy

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Scope and Method of Study: The overall research goal of this study is to develop an optimization model for the design of SBR with a new process model that describes the process adequately with as few parameters as possible to predict the performance well under all operating conditions. Developed process model was calibrated and validated prior to its use in the optimization model. The data for model calibration and validation were obtained from the operation of a full-scale 836 m^3/h (5.3 MGD) SBR system at the City of Tahlequah, Oklahoma. A calibration methodology was also presented for determination of unknown kinetic and stoichiometric parameters using an optimization technique called Simulated Annealing. Using the calibrated model, a model-based design methodology has been presented, and applied to the full-scale SBR for comparison with the existing conventional design. The proposed design methodology was, then, extended to design optimization model by including the capital and operating cost in the objective function. The objective function was then minimized with the same optimization technique, Simulated Annealing, subject to operational and process constraints. The results from the optimal design were, then, compared with the existing design. For the benefit of the existing SBR system, optimal operational strategies were also developed for energy savings.

Findings and Conclusions: The model predictions described reasonably well the trend of biodegradable organics removal, nitrification during aeration, and denitrification during the anoxic period. The model-predicted concentrations for NH_4-N were within 0.03 mg/L and 0.61 mg/L from experimental values, and the average deviation was 0.29 mg/L. The average deviation for NO_3-N from the experimental value was 0.23 mg/L. The model-predicted concentrations for BOD_5 were generally between 0.5 mg/L and 1.8 mg/L except for two samples which are suspected outliers. Results from this work substantiate that use of model-based design approach or design optimization can reduce the cost of SBR system by about 10 to 20 percent of the life cycle cost and also reduce the volume of the reactor by about 30 percent compared to the conventional design approach. These reductions in volume and cost are significant and can make the SBR system more attractive to larger installations.

ADVISER'S APPROVAL: Dr. John N. Veenstra