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# IMPROVING BOD REMOVAL AT SNJ WASTEWATER TREATMENT PLANT BY BIOLOGICAL TREATMENT

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*June 2010*

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**Abbreviations**

| | |
|---|---|
| **AS** | Activated Sludge |
| **BOD** | Biochemical Oxygen demand |
| **COD** | Chemical Oxygen demand |
| **DO** | Dissolved oxygen |
| **MLSS** | Mixed liquor suspended solids |
| **MLVSS** | Mixed liquor volatile suspended Solids |
| **OUR** | Oxygen Utilization Rate |
| **RBCOD** | Readily biodegradable COD |
| **SBCOD** | Slowly biodegradable COD |
| **SBR** | Sequencing batch reactor |
| TOC | Total Organic Carbon |
| **TSS/VSS** | Total Suspended Solid, Volatile Suspended Solid |
| **WWTP** | Wastewater treatment plant |

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# ABSTRACT

SNJ uses chemical precipitation method to treat domestic wastewater. With regard to organic removal requirement; chemical treatment alone does not seem to be sufficient at SNJ/IVAR wastewater plant. This thesis is to assess the performance of the aerobic biological treatment on the wastewater of the plant in order to upgrade the existing plant to include biological treatment. The work was to conduct a laboratory scale SBR test for determination of the wastewater characteristics and the effect of temperature on the biological treatment. In addition to theoretical and experimental studies, the data was applied with a mathematical model of activated sludge used in modeling and design of biological treatment.

# 1 INTRODUCTION

In Stavanger, the Sentralrenseanlegg Nord-Jæren (SNJ) wastewater treatment plant (WWTP) is one of the largest wastewater plants in Norway. SNJ uses chemical precipitation method to treat domestic wastewater of 220000 pe (Asplan Viak, 2010) from the communities around Stavanger. In Norway, this method was originally intended for removal of phosphorus (P) and suspended solids (SS). The reason of applying chemical P-removal was often referred to as being the characteristics of Norwegian wastewater with dilute, low temperature and with main particulates organics fraction (Ødegaard, 1995 cited by Ydstebø, 2005).

After implementation of EU regulations, regulations on wastewater treatment set the requirements according to secondary treatment. Subsequently, focus has been shifted from removal of phosphorus and SS to removal of organic material ($BOD_5$, COD).

The treatment method applied by SNJ does not remove much of the soluble $BOD_5$. Therefore, the plant sometimes has difficulties to meet the EU secondary effluent discharge requirements which are:

- Maximum 25 mg/l $BOD_5$ and 125 mg/l COD
- Minimum 75 % $BOD_5$ and 70 % COD removal

The present project is an investigation of an alternative method for the chemical treatment in order to improve the BOD removal at SNJ plant to meet the discharge permits. Biological wastewater treatment is the most common method used for removal of dissolved BOD and was tested in this project

One aspect that is important for biological wastewater treatment is the effect of temperature, as the temperature in wastewater at SNJ can reach low values during winter such as 4-5ºC.

The objective of the study is to investigate the efficiency of substrate removal in biological treatment on the wastewater of the SNJ plant, based on theoretical and experimental studies. This will provide information for design of a better and acceptable treatment alternative for the plant.

The study aims to provide information on process rates and how temperature will affect the process rates. The data will be valuable for dynamic modeling with the software Aquasim which will issue information important for the recommendations of the design of the plant.

# 2 BACKGROUND AND LITTERATURE REVIEW

## 2.1 THE SNJ PLANT

Sentralrenseanlegg Nord-Jæren (SNJ) is located in Merkjavik in Randaberg municipality about 10 km north of Stavanger. It has been in operation since 1992.

### 2.1.1 Characteristics of the plant

The SNJ plant treats domestic wastewater from Randaberg, Stavanger, Sola, Gjesdal and Sandnes municipalities. The treatment is localized and constructed in the core of mountains as in cavern sites.

The treated wastewater is discharged to Håsteinfjorden.

The treatment plant North Jæren (SNJ) comprises an 8 km tunnel from Bjergsted in Stavanger, and a 4 km tunnel to discharge to Håsteinfjorden.

The inlet flow consists of wastewater basically from households with contribution from infiltration flow (surface water and infiltration of groundwater and water leaks) and also small part from industrial wastewater.

Generally, the wastewater from households is estimated around 200L / p.d (doc. Asplan viak)

The specific flows in 2009 corresponded to a total of 450 L / p.d, of which 250 L/ p d is the amount of industrial wastewater and infiltration water.

The processes (figure 2.1) involved in the sewage treatment at SNJ consist of:

Screening, chemical precipitation with ferric chloride, flocculation, sedimentation and treatment of sludge with anaerobic digestion.

- Screening:

The pre-treatment consists of mechanical screening and trapping of sand materials. In screening and sand trapping, coarse and large particles are separated from the wastewater with screens of 3mm opening while the sand is removed in grit chambers.

- Storage of chemical additives:

The chemical precipitation was originally based on ferric chloride and seawater. Ferric chloride is stored in 6 tanks of ca.70$m^3$ volume each. At present, seawater is not in use.

The ferric chloride and seawater can be added at the inlet of sand trap basin (figure 2.2). The dosage of ferric chloride is controlled by the incoming flow, the inlet pH and turbidity.

Addition of the chemical can also be made separately in the mixing chamber or flocculation chamber upstream of the sedimentation basin (figure 2.2) as the chemicals can react very quickly with the contaminants in the wastewater especially when mixed and thus form small particles.

- Flocculation:

In the flocculation chamber, smaller particles is grown to larger particles enhanced by mixing by bounding to each other forming flocs that will settle more rapid in the sedimentation basin compared to smaller particles.

- Sedimentation:

The sedimentation basins are located in 4 halls. Each hall consists of 2 parallel basins of 7.0 m wide and 67.6m long (surface: 473.2 $m^2$) and with water depth of 4.8m. The clarified water flows into effluent weirs to a channel between the 2 parallel basins and to the effluent pipe.

- Sludge disposal:

The sludge settled at the bottom of sedimentation basin is pumped out to the buffer tank in the sludge treatment plant. The sludge has a high solid content (TS= 5.0 to 5.5%). The floating materials are pumped either to the sludge treatment plant or withdrawn and dewatered before transfer to a sludge disposal facility.

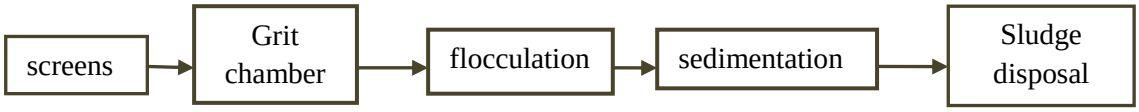


Figure 2.1 A simple schema of the treatment process at the SNJ plant

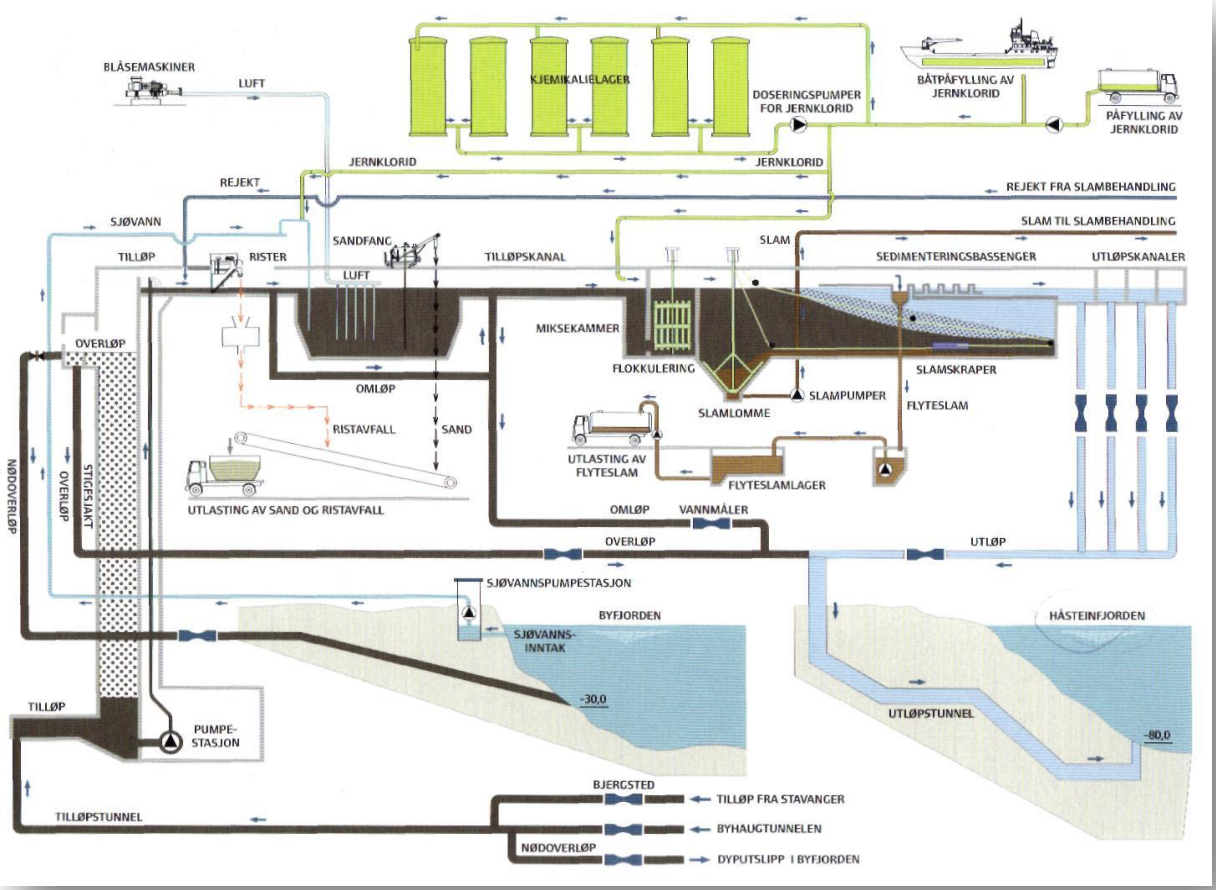


Figure 2.2 Process diagram of wastewater treatment at SNJ plant (Doc. IVAR)

### 2.1.2 Capacity of the plant

The plant is designed for 240,000 population equivalents (pe) (doc. IVAR). The estimated load was around 218300 pe in 2008, 221000 pe in 2009 and around 390000 pe previewed for 2050 (Asplan Viak, 2010).

Table 2.1 shows the number of population estimated in the communities from where the wastewater is generated.

Table 2.1 Number of populations in the municipalities

| Year | 2008 | 2009 | 2050 |
|---|---|---|---|
| Sandnes | 63000 | 64000 | 141800 |
| Stavanger | 121000 | 122700 | 180000 |
| Gjesdal | 3243 | 3303 | 17000 |
| Sola | 21446 | 21895 | 38000 |
| Randaberg | 9622 | 9774 | 15000 |
| **Sum population** | **218311** | **221672** | **391800** |

Source: Asplan Viak, 2010

*Sewage treatment plant*

The actual design for the system at SNJ are:

$Q_{dim}$ = 1500 l/s

$Q_{max,dim}$= 2500 l/s

$Q_{max}$= 4000 l/s

The Annual total flow reported by Asplan Viak (2010) was estimated to ca.37million $m^3$ in 2008 and 2009.

*The forecast design flow rates*

In 2050, the number of population in the municipalities related to the plant is expected to be ca.392000 (table 2.1). This estimation can be approximated to 400000 (Asplan Viak, 2010) which is an increase of 80% with regard to number of population in 2009.

With the number of residents (and certainly including visitors) previewed for 2050, the flowrate of the sewage treatment will be expected to increase up to 3.8 $m^3$/s (3800 l/s) and 5.2 $m^3$/s (5200 l/s) respectively for design flowrate and the maximum flowrate (Asplan Viak, 2010).

$Q_{dim}$(2050) = 2100 l/s

$Q_{max,dim}$ (2050) = 3800 l/s

$Q_{max}$ (2050) = 5200 l/s

This increased load has to be considered for the alternative of any new design of the plant in the future.

In estimation of design capacity in 2050, the load on the plant will correspond to 500000 pe (Asplan Viak, 2010).

The organic loading can be estimated as: 500.000 pe x 0.06kg $BOD_5$/pe.d = 30.000kg $BOD_5$/day for capacity in 2050.

*Sludge treatment*

The capacity of the plant for sludge treatment is summarized as:

Sludge production: 25t TS/day

Organic loading: 15t TS/day

Hydraulic loading: 500$m^3$/day

Sludge dewatering: 18t TS/day

## 2.2 OVERVIEW ON WASTEWATER TREATMENT

### 2.2.1 Chemical treatment

According to Metcalf and Eddy (1991) chemical precipitation often in combination with other type of treatment was used to enhance the degree of suspended solids and BOD removal for wastewater with large variations in the concentration. Since about 1970, the need to provide more complete removal of organic compounds and nutrients such as nitrogen and phosphorus in wastewater has brought to renewed interest in chemical precipitation. Today, it is most used to enhance phosphorus removal.

#### *2.2.1.1 Chemical removal performance*

The degree of clarification obtained depends on the quantity of chemicals used and the care with which the process is controlled (Metcalf and Eddy, 1991).

Removal of 80 to 90% of suspended solids yields removal of:

- 40-70% $BOD_5$
- 30-60% COD
- 80-90% of bacteria can be removed from chemical precipitation (Metcalf and Eddy,1991)

When sedimentation only is used (i.e.: without chemical addition), the removal efficiency is lower:

- about 50-70% of TS matter
- 30-40% organic matter (Metcalf and Eddy, 1991)

If we can compare that with the EU secondary treatment requirements of discharge, chemical precipitation does not fulfill the requirement of secondary effluent before discharge which is:

- Minimum 75 % $BOD_5$ and 70 % COD removal

Chemical precipitation involves addition of chemicals or coagulants to alter the physical state of dissolved and suspended matters for formation of aggregates and thereafter to facilitate their removal by sedimentation.

Colloidal and small particles settle very slowly and it will be very expensive to let them settle by themselves. Coagulation and flocculation are mostly used to remove settleable and non-settleable solids and also nutrients (nitrogen, phosphorus).

#### *2.2.1.2 Ferric chloride as coagulants*

The ability of an agent to coagulate water is related to its charge. The multivalent characteristic of cations in the coagulants strongly attracts them to charged colloidal particles and their relative insolubility ensures their removal to a high degree (Droste, 1997).

There are different coagulant salts that can be used in chemical precipitation. At the SNJ plant, they use iron salts (ferric-chloride: $FeCl_3$).

The characteristic of ferric chloride is to react with alkalinity or phosphates to form insoluble iron salts. The chemical reaction for the ferric coagulants, $FeCl_3$, involves precipitations in the form of ferric hydroxide. Complexes can be then formed (Sincero, 2003) that reacts with suspended solids, colloids and also phosphorus for the purpose of phosphorus removal.

The hydroxide can be provided either by the presence of calcium bicarbonate in the wastewater or by adding lime:

$$2FeCl_3 + 3Ca(HCO_3)_2 \longrightarrow 2Fe(OH)_{3_{(S)}} + 3CaCl_2 + 6CO_2$$

$with\ lime$:

$$2FeCl_3 + 3Ca(OH)_2 \longrightarrow 2Fe(OH)_{3_{(S)}} + 3CaCl_2$$

For phosphorus removal, the basic reaction involved in precipitation of phosphorous and iron is shown below (Metcalf and Eddy, 1991):

$$Fe^{+3} + H_nPO_4^{-3n} \leftrightarrow FePO_4 + nH^+$$

This reaction is just simple reaction as there are many competing reactions involved in the wastewater. Thus, this cannot be directly used for calculation of chemical dosages. Dosage should be done on the basis of a bench-scale test.

The complex $FePO_4$ or $Fe(OH)_3$ when produced are found in the sludge composition.

### *2.2.1.3 Effect of pH*

pH has an effect on the efficiency of the precipitation reactions and ferric salts work best in pH range of 4.5-5.5 (Droste, 1997).

Metcalf and Eddy (1991) reported that theoretically the minimum solubility of FePO4 is at pH 5.3 (figure 2.3). But practically, they have yielded good phosphorous removal in the range of 5.5 to 7.0.

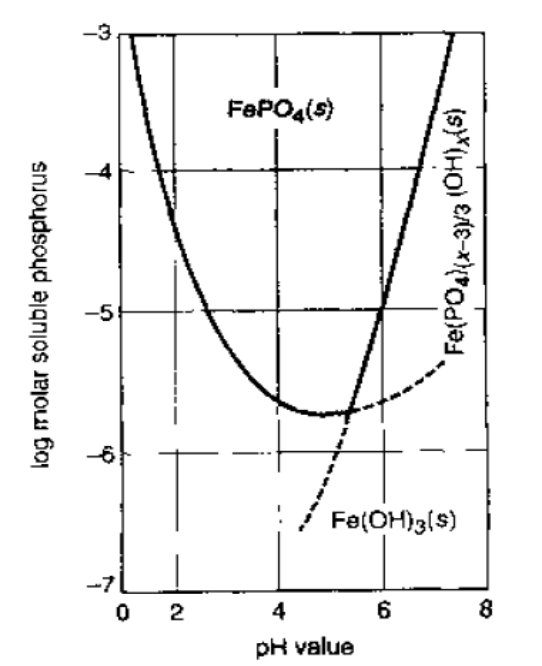


Figure 2.3 Concentration of ferric phosphate in equilibrium with soluble phosphorus and as function of pH (Metcalf and Eddy, 1991)

In Figure 2.3, the solid lines demarcate the concentration of residual soluble phosphate after ferric phosphate precipitation. Pure ferric phosphate is precipitated inside the solid lines and mixed complex polynuclear species are formed outside toward higher and lower pH values (dashed lines).

### *2.2.1.4 Influence of temperature*

For both iron and alum salts, it was found that a constant pH over temperature range of 5-20$^{o}$C produced the best coagulation-flocculation results (Droste, 1997). Thus, performance decreases with cold temperatures (< 5$^{o}$C).

Coagulants such as iron salts are often added to improve solid/liquid separation (Droste, 1997). Coagulants enhance the removal of solids and nutrients from wastewater. Nevertheless, they are inadequate to remove dissolved BOD and this is why biological treatment is needed at the SNJ plant to improve their effluent quality.

### 2.2.2 Biological treatment

The major objective in most biological treatment processes is the reduction of organic content (carbonaceous BOD) (Metcalf and Eddy, 1991) and in some cases removal of nutrients (N, P).

Biological treatment uses microorganisms which have ability to decompose dissolved and colloidal organic matters. Oxygen is supplied to the system to facilitate the degradation of organic compounds.

The most common form of biological treatment is activated sludge treatment that is based on microorganisms in suspension. Another system is biofilms that are based on microorganisms growing on surfaces.

#### *2.2.2.1 The Activated Sludge process*

The activated sludge process is used to improve the BOD removal of wastewater, by supplying enough oxygen and nutrients for the degradation of organic compounds (BOD).

Conventional activated sludge is composed of a bioreactor where bacteria transform the biodegradable substrate in the wastewater (organic compounds) into new cells in their metabolism. A clarifier is used to collect the sludge where the operation consists of the separation of the biomass and suspended solids from the wastewater by gravity settling. A fraction of the sludge is returned back to the aeration tank to increase the biomass concentration. This allows for reducing the required bioreactor volume and the retention time.

#### *2.2.2.2 Characteristics of aerobic activated sludge*

The biodegradable organic compounds serve as substrate for aerobic microorganisms. The substrate or electron donor is oxidized into $CO_2$, $H_2O$. Oxygen is used as electron acceptor resulting in energy generation for growth of microorganisms.

The growth rate is influenced by chemical and physical factors and to favour efficient substrate removal, i.e: high rates of growth, optimum requirements for growth must be provided.

The organic substrate serves both as energy source and carbon source. In addition, other nutrients such as nitrogen, phosphorous and others are also important, in the amounts shown in Table 2.2.

Table 2.2 Cell composition of microorganisms

| Percent of dry weight | | |
|---|---|---|
| Element | Range | Typical |
| Carbon | 45-55 | 50 |
| Oxygen | 16-22 | 20 |
| Nitrogen | 12-16 | 14 |
| Hydrogen | 7-10 | 8 |
| Phosphorus | 2-5 | 3 |
| Sulphur | 0,8-1,5 | 1 |
| Potassium | 0,8-1,5 | 1 |
| Sodium | 0,5-2,0 | 1 |
| Calcium | 0,4-0,7 | 0,5 |
| Magnesium | 0,4-0,7 | 0,5 |
| Chlorine | 0,4-0,7 | 0,5 |
| Iron | 0,1-0,4 | 0,2 |
| Others | 0,2-0,5 | 0,3 |

Source: Metcalf and Eddy, 1991

In domestic wastewater, nutrients such as N, P are supposed to be present in excess. However, in some industrial wastewaters with high concentration of BOD; N and P may be absent and must be added.

Temperature and pH are primary environmental variables. The optimum pH is around 6.5 to 8.5 (Water Environment Federation, 1994).

#### *2.2.2.3 Influence of temperature*

The temperature dependence of the biological reaction-rate constants is very important in assessing the overall efficiency of a biological treatment process. Temperature has significant influence on the reaction in the biological process and the biomass growth rate.

Higher temperature increases the rate of reaction and thus the growth rate increases. The growth rate of microorganism doubles approximately at each 10$^{o}$C up to a maximum temperature (Ydstebø, lecture 2008).

At low temperatures, growth is slow and most organisms found in aeration tank of activated sludge work best at moderate temperature (ca. 10 to 40ºC) (Water Environment Federation, 1994),

The effect of T on the reaction rate of a biological process is usually expressed as:

$\mu_{max\ T} = \mu_{max,20}.\theta^{(T-20)}$ (eq. 2.1)

Where

θ is a constant (temperature coefficient)

$k_{\_20}$ and $k_{\_T}$ are reaction rate at temperature 20 ºC and T and k can be μ_max or k_d

For activated sludge, a range from 1-1.08 has been reported for the temperature activity coefficients (Θ) (Metcalf and Eddy, 1991).

Specifically, for heterotrophic organisms: θ can be 1.07 for growth correction (μ_max) and 1.03 for decay correction (k_d) (Ydstebø, lectures 2009)

#### *2.2.2.4 Removal efficiency*

The removal efficiency varies according to the type of activated sludge used. It was reported for example that a conventional AS with plug-flow system, a continuous-flow stirred-tank reactor system and a SRB with intermittent flow stirred-tank reactor system can yield a BOD removal of 85-95% (Metcalf and Eddy, 1991).

#### *2.2.2.5 Substrate removal and biomass growth*

Substrate removal and biomass growth are interdependently related and the removal and consumption of substrate result in growth of new cells.

The Monod expression has been found to properly describe growth and substrate removal in many different types of biological treatment methods.

Every substance required by the microorganisms for their growth can limit the growth rate.

In aerobic growth on organic compounds (COD), $O_2$ or $NH_3$ may limit the growth rate (Ydstebø, lecture 2009).

The limiting compound can be determined by using the Monod equation for growth

$$\mu = \frac{\mu_{\max} \cdot C}{K_S + C} \quad \text{(eq. 2.2)}$$

Where ,

μ: Specific growth rate (1/d)

μ_max: Max specific growth rate (1/d)

Ks: Half saturation constant (mg/l)

C: Growth limiting compound (mg/l)

Mass balance for growth

Growth: $substrate = biomass + respiration$

$$\Delta COD = \Delta VSS + \Delta O$$

$$\Delta COD = Y.\Delta COD + (1 - Y).\Delta COD \quad \text{(eq. 2.3)}$$

To summarize the relations in the growth of biomass and substrate removal:

Growth: $\frac{dX}{dt} = \mu X = \frac{\mu_{max} C}{K_s + C} X$ (eq. 2.4)

Yield: $Y = \frac{\Delta X}{\Delta C} = \frac{dX/dt}{dC/dt}$ (eq. 2.5)

Substrate removal: $\frac{dC}{dt} = \frac{\mu X}{Y} = \frac{\mu_{max} C}{K_s + C} \frac{X}{Y}$ (eq. 2.6)

Oxygen consumption: $\frac{dO}{dt} = (1 - Y)\frac{dC}{dt} = (1 - Y)\frac{\mu X}{Y}$ (eq. 2.7)

X is used instead of VSS because X is referred as viable biomass. And substrate removal depends on the active biomass and a given amount of substrate can support only a given amount of active mass when the process is operated under starved conditions (Droste, 1997). Inappropriately, VSS can contain a significant amount of dead biomass along with active.

The part of substrate utilised in growth can be termed as growth yield, Y and expressed as the mass of organisms produced per mass of substrate removed.

Substrate is generally expressed in terms of COD, BOD (TOC). COD is the amount of a specified oxidant (ex: $Cr_2O_7^{2-}$) that reacts with the sample. The quantity of oxidant consumed is expressed in terms of its oxygen equivalence. Thus, COD can be defined as the chemical oxygen demand which is the amount of oxygen needed to complete oxidation of organic compounds in the sample.

BOD or Biochemical Oxygen demand is the amount of oxygen consumed by microorganisms for the degradation of organic compounds in the sample over a period of 5 days ($BOD_5$). This involves only the biodegradable organic compounds.

TOC or Total Organic Carbon is the amount of all carbon atoms bonded covalently in the organic compounds.

Based on the rate of their assimilation by organisms, substrates can be classified in 2 kinds:

- RBCOD, the easily or readily biodegradable which is directly utilised by the organisms as fast as they are available, for example: glucose.
- SBCOD or slowly biodegradable which is assimilated very slowly because hydrolysis is necessary and some COD originates from dead organisms as well. Thus, it may also depend on how fast the organisms die.

In an activated sludge, the organisms especially heterotrophs have 3 possibilities of getting the organic compounds necessary for their basal metabolism (Jenkins, 1978)

- Soluble organics from raw sewage, that is the soluble RBCOD
- Products of hydrolysis from biodegradable suspended organics that is particulate SBCOD converted to soluble substrate due to the action of exogenous enzymes.
- Internal degradation of cell structures referred to as endogenous respiration or decay

Decay

Biomass is continuously lost by decay. The endogenous respiration occurs when dead organisms is oxidized by the remaining living biomass. This requires oxygen consumption as well. But, a portion of the biomass is not oxidised and builds up as endogenous residue ($f_d$). The fraction oxidised is then (1- $f_d$).

Mass balance for decay

Decay: $\Delta X_H = \Delta X_E + \Delta O$

$$\Delta X_H = f_d.\Delta X_H + (1 - f_d).\Delta X_H \quad \text{(eq. 2.8)}$$

Growth $\frac{dX}{dt} = -k_d.X$ (eq. 2.9)

Endogenous residue production $\frac{dX_E}{dt} = f_d.k_d.X$ (eq. 2.10)

Oxygen consumption $\frac{dO}{dt} = (1 - f_d).k_d.X$ (eq. 2.11)

# 3 MATERIALS AND METHODS

A laboratory-scale experiment of biological wastewater treatment has been conducted to investigate the efficiency of BOD removal from wastewater and to provide information to the possible alternative for the actual treatment method on the plant.

So, to reach to that end, a batch SBR process was used for the lab-scale experiment. SBR can be operated to achieve BOD and SS removal, nitrogen reduction and phosphorus removal and it is especially suitable for small scale. The laboratory works were operated in cooperation with a master student colleague between beginning of January and 31st march.

## 3.1 EXPERIMENTAL CONDITIONS

As during winter, low temperature is an important issue in wastewater treatment and the experiments conditions were based on different temperatures:

- At 4ºC
- At 8ºC
- At room temperature, around 20ºC

The experiment was carried out in a digestor type SBR- Sequential Batch Reactor at 1.5L, 1.5L and 4.0L volume at 4, 8 and 20ºC respectively.

As the treatment was aerobic, aeration was used to keep the activated sludge in suspension and supply oxygen. The reactor was feed daily.

Table 3.1 summarizes the experimental conditions and operation

Table 3.1 Experimental conditions for aerobic reactors/SBR

| Experiment | Temperature (ºC) | Volume(l) | Oxygen (mg/l) | Feed frequency (day) | period | Loadings (l/d) |
|---|---|---|---|---|---|---|
| Batch 1 | 20 | 4 | 7.5-9.0 | 2<br>1<br>1 | 3/01/10-26/01/10<br>27/01/10-3/02/10<br>4/02/10-9/03/10 | 1-1.88<br>2-3<br>2.5 |
| Batch 2 | 4 | 1.5 | 10-12 | 1-2<br>1<br>1 | 3/01/10-26/01/10<br>27/01/10-1/02/10<br>2/02/10-26/02/10 | 0.38-0.60<br>1.2-1.4<br>1.25 |
| Batch 3 | 8 | 1.5 | | 1 | 4/03/10-18/03/10 | 1.25 |

Table 3.2 summarizes the aerobic digestion

Table 3.2 Experiments for decay determination

| Experiment (same batch as SBR) | Temperature (ºC) | Period | Feed |
|---|---|---|---|
| Batch 1 | 20 | 10/03/10-26/03/10 | 0 |
| Batch 2 | 4 | NA | |
| Batch 3 | 8 | 19/03/10-31/03/10 | 0 |

pH was maintained between 7.5 to 8.5.

At the beginning of the experiments, especially for 20 and 4ºC batchs, the feed volume and the wastage were not constant and the feed frequency was every 2 days. When more or less constant biomass concentration was attained, constant daily feeding was performed to keep SRT constant and have a steady state condition.

## 3.2 THE EXPERIMENTAL PROCEDURES

### 3.2.1 Experimental monitoring

As the system was SBR, the settling was operated in the same batch. The operating cycle includes filling, reaction, settling, decanting.

Raw wastewater was collected every 14 days from the inlet of SNJ plant after preliminary screening.

As the raw wastewater initially had low BOD concentration, it became difficult to generate sufficient sludge. The first cycles of the operation were devoted only to produce sludge and sugar was added to enhance the growth.

The growth of the biomass was initiated first by feeding every 2 days at the beginning of the experiment. The sludge was retained and half of the batch volume was decanted at the end of each cycle in order not to lose suspended solids in the effluent.

So, later from February, the feeding mode changed to daily feeding as the reactor contained a more concentrated and constant biomass.

Sludge was removed when the retention time was sufficient for the growth. The retention time, however, varied during the experiment probably because of the variation of the wastewater characteristics and operation.

In fact, the wastewater characteristics were not the same during the experiment, regarding the BOD content and the salinity.

### 3.2.2 SBR characteristics

OUR, TSS/VSS, TOC, COD were measured and monitored regularly in order to get data of the experiment.

The OUR was measured daily along with the TSS/VSS and TOC. COD was measured for every raw wastewater and non-systematically for some samples of effluent.

Those measurements were used for assessing kinetic parameters such as the maximum growth rate, the decay rate ($\mu_{max,}$ $k_d$) as well as the growth yield (Y).

### 3.2.3 Determination of decay by aerobic digestion

After terminating the aerobic activated sludge, the experiment was prolonged with an aerobic digestion test for determination of the endogenous respiration rate (decay rate). It was performed with the remaining sludge in the reactors for 12 days for 8ºC and 22 days for 20ºC. Digestion of sludge at 4ºC was not performed, because the temperature controlled room had limited period of 4ºC.

OUR, TSS/VSS over time and nitrate were analysed to determine the endogenous respiration rates. The decay rate was determined graphically with the OUR data or VSS.

### 3.2.4 Control

The pH was controlled and kept in the range of 7.5 and 8.5.

Usually pH goes down when nitrification occurs. In the 20ºC reactor, nitrification appeared for some periods of time.

Whenever the pH was reduced, a buffer such as sodium bicarbonate ($NaHCO_3$) was added to bring the pH back to optimum.

In our experiment, between 0.6 to 1g of $NaHCO_3$ were added to the mixed liquor to bring up the pH.

Nutrients, especially nitrogen and phosphorus are vital for bacteria growth. If the wastewater does not contain any nutrients, we need to add some. With domestic wastewater, adding nutrients is not supposed to be necessary as it already contains sufficient. However, it happened in our experiment

that the mixed liquor lacked phosphorous. So, a small amount of pre-prepared macronutrients $NaH_2PO_4.2H_2O$ was added to avoid phosphorus limitation.

### 3.2.5 Maintenance

The reactor are cleaned every second week to control biological growth on the walls. The rubber-type diffusers used in the aeration system may also have its pores plugged and biofilms growth. 5%HCl was used to clean the diffusers and the walls.

## 3.3 ANALYTICAL PROCEDURES

### 3.3.1 pH, temperature, DO, conductivity

pH, temperature, dissolved oxygen (DO), conductivity were measured to monitor the condition of the digestor. In addition, OUR (Oxygen utilization rate), solids TSS/VSS, TOC, COD, ortho-phosphate and nitrate were also analysed in order to follow the activity of bacteria, their growth and substrate removal.

pH, temperature, DO and conductivity were determined with a multimeter WTW multi 340i (for pH, oxygen (DO) and conductivity). An oxymeter Cellox 325 was used for recording the dissolved oxygen.

### 3.3.2 OUR

OUR is the Oxygen consumption Rate by bacteria. It is determined to monitor the bacteria activity by their oxygen consumption. A 250ml erlenmeyer flask is used for OUR measurements. It was applied during the daily monitoring of the mixed liquor and during the aerobic digestion tests.

The erlen meyer flask was filled completely with the mixed liquor. The probe of the oxymeter equipped with a rubber stopper is put in the flask. It had to be sealed completely so that no air entered the flask. The system is stirred while the reduction of DO is recorded until it reached about 2mg/l.

OUR measurements were done right after feeding with a high frequency during the beginning of the cycle when there was high concentration of substrate. These data was used for calculation of the maximum growth rate and to distinguish the readily biodegradable substrate (RBCOD.The liquid used for OUR measurements was put back in the reactor. The OUR was determined graphically by plotting the DO versus time.

### 3.3.3 TSS/VSS analysis

Filtration of the samples was performed with Whatman GF/C glass fibre filter with 1um pore size. The filter was dried at 105ºC and weighed before filtration and after drying. In a graduated cylinder, the sample to be filtered was measured. The volume depends on the amount of solids in the sample. The sample settled for a while in the graduated cylinder to make the filtration going faster, as the clear liquid on top of cylinder was filtered before the concentrated solids at the bottom.

TSS were determined after evaporating the solids and filter at 105ºC about 2 hours.

VSS were determined after burning the solids and filter at 550ºC for 30minutes.

The filtered liquid was preserved by acidification or freezing for further analysis such as minerals analysis, orthophosphate, nitrate, TOC and COD.

### 3.3.4 TOC

TOC was analysed with a Shimadzu 5000 A TOC analyzer. A filtered sample was injected and the organic carbon was oxidized catalycally after combustion at 680ºC to $CO_2$ and the measurement was based on $CO_2$ measurement by a non-dispersive infrared gas analysis.

### 3.3.5 COD analysis

For COD analysis, the closed Reflux colorimetric method was applied (Clesceri, Greenberg et al. 1998). In a HACH vial, 7.5 ml solution is prepared with 2.5 ml sample added to 1.5 ml of digestion solution and 3.5 ml sulphuric acid solution. After closing tightly the cap, the samples are digested at 150ºC about 2 hours. Then, the absorption of each sample was determined with the spectrophotometer HACH DR 2000 at 600nm wavelength.

### 3.3.6 Analyse of phosphate, nitrate and ammonia

Po4-P and NO3-N content of filtered samples are measured with an Ion Chromatograph (Dionex ICS-3000). Pre-treatment (filtration) was required by filtration through a 0.2um syringe filter to remove particles as the IC had a small diameter of the column.

For reactive P, standard solution is made by $K_2HPO_4$ to 100 mg/l P and calibration of the instrument is done within an appropriate range.

For dissolved N compounds such as ammonia ($NH_4$), nitrite ($NO_2$) and nitrate ($NO_3$), standards are made of $NH_4Cl$, $KNO_2$ and $KNO_3$ for ammonia, nitrite and nitrate respectively; within the range the sample concentrations are expected.

## 3.4 MATHEMATICAL MODELLING

The parameters determined from the experiment can be used in modeling. A mathematical model describing the process for biological wastewater treatment is required to determine the essentials for the design and control for future alternative to propose.

It can help optimizing the process design and control and determining the most favorable system.

The models are presented in matrix including kinetic rates and stoichiometry. The basic processes here are related to growth and decay of heteretrophs.

The activated sludge model is derived from the mass balances of different constituents. The model is used to describe the biological processes involved in the aeration tank of the system with return sludge. There are two options of sludge wasting as shown in figure 3.1; from the settler and directly from bioreactor. The system configuration on which the expressions are written is shown in Figure 3.1:

Accumulation = Inflow- outflow + reaction

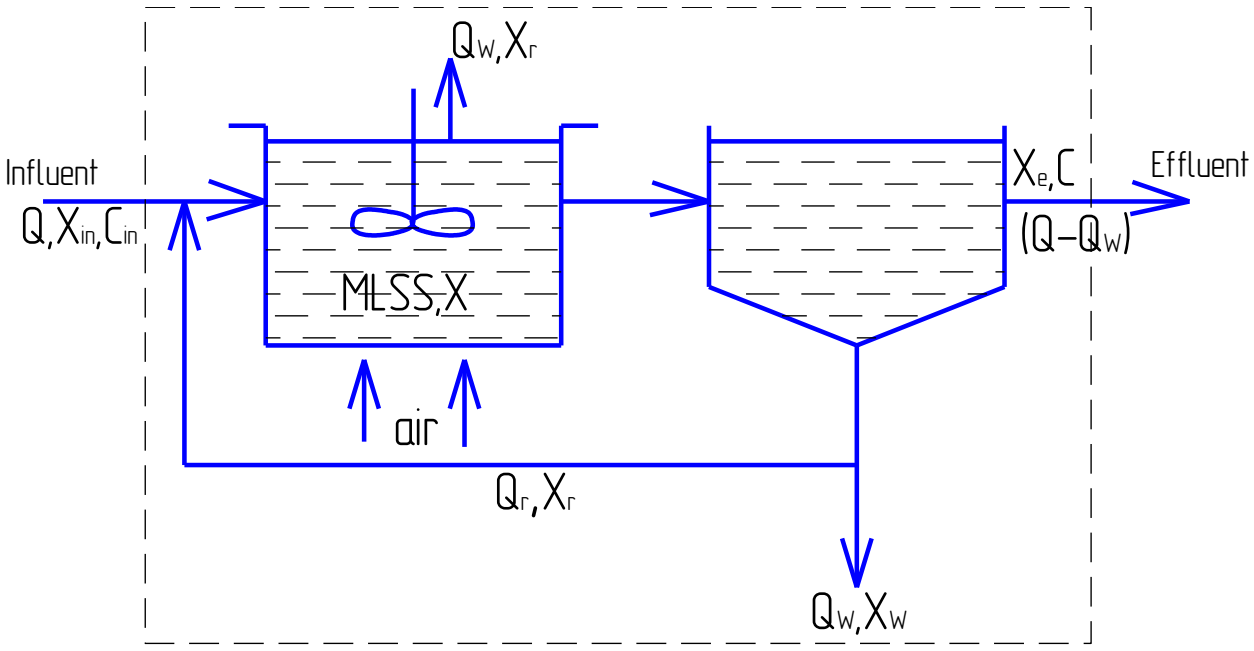


Figure 3.1 Schematic diagram of activated-sludge system for modelling

Variables are important as they significantly affect the process control and the performance of an aerobic biological treatment process. Those variables especially the retention time can be controlled through process design.

### 3.4.1 Biomass mass balance in bioreactor

By using the mass balance for biomass in a bioreactor (ideal CSTR or continuous stirred tank reactor), the relation between SRT and growth rate can be determined from:

$$V\frac{dX}{dt} = QX_{in} - QX_e - Q_wX_w + \mu XV - k_d\,XV \quad \text{(eq. 3.1)}$$

$$Q_wX_w = \mu XV - k_d\,XV$$

$$\frac{Q_wX_w}{XV} = \frac{1}{SRT} = \mu - k_d$$

$\mu$-$k_d$ is the net growth rate which is the inverse of retention time. SRT is the time the biomass remains in the reactor.

The previous expression can also be solved to result in the effluent substrate concentration C (COD) as function of the retention time.

$$\frac{1}{SRT} = \frac{\mu_{max}C}{K_s+C} - k_d\,,\ with\ \frac{\mu_{max}C}{K_s+C} = \mu \qquad \text{(eq. 3.2)}$$

$$\left(\frac{1}{SRT}+k_d\right)(K_s + C) = \mu_{max}C$$

$$C\left(\mu_{max} - \frac{1}{SRT} - k_d\right) = K_s\left(\frac{1}{SRT} + k_d\right)$$

$$C = \frac{K_s\left(\frac{1}{SRT}+k_d\right)}{\mu_{max}-\left(\frac{1}{SRT}+k_d\right)} \qquad \text{(eq. 3.3)}$$

This equation expresses the effect of the SRT on system performance. It is possible to regulate SRT to achieve good treatment efficiency that is determined by the effluent concentration, C (COD) in the reactor and effluent. To meet the requirement of a low effluent, a longer sludge retention time should be used.
SRT is determined in practice by controlling the sludge waste rate ($Q_{waste}\cdot X_r$).

Figure 3.2 shows the effluent substrate COD in function of the retention time and also in function of temperature.

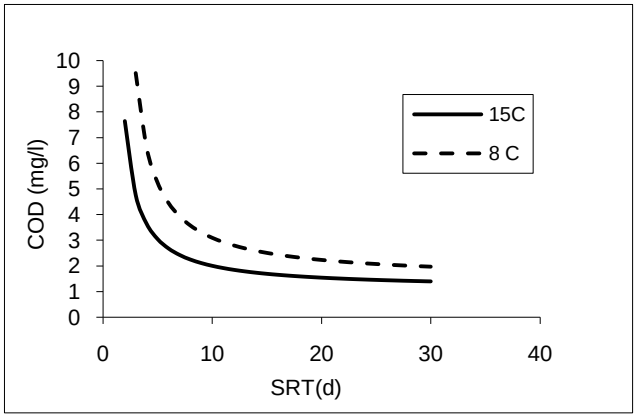


Figure 3.2 Effluent substrate concentration COD as function of SRT and temperature

If temperature is low, the maximum growth rate is lower and longer SRT is needed. For the same target effluent COD, SRT at 8ºC is longer than at 15ºC because growth is lower at lower temperature.

### 3.4.2 Mass balance for substrate

$$V\frac{dC}{dt} = QC_{in} - QC + \frac{\mu X}{Y}V \qquad \text{(eq. 3.4)}$$

Steady state:

$$C\ constant \Longrightarrow \frac{dC}{dt} = 0$$

$$0 = Q(C_{in} - C) - \frac{\mu X}{Y}V$$

$$Q(C_{in} - C) = \left(\frac{1}{SRT} + k_d\right)\frac{X}{Y}V \ \ with\ \mu = \frac{1}{SRT} + k_d$$

$$X_H = \frac{Q(C_{in}-C)Y}{V\left(\frac{1}{SRT}+k_d\right)} = \frac{Q(C_{in}-C)Y.SRT}{V(1+k_d\,.SRT)} \qquad \text{(eq. 3.5)}$$

This equation shows the SRT effect on biomass concentration. Long SRT result in low sludge waste, thus accumulation of high biomass concentration in the reactor.

The expression also shows the biomass concentration is function of substrate. The more substrate is removed $(C_{in} - C)$ the higher the biomass concentration$X_H$. High decay rate $k_d$ will decrease the biomass concentration.

According to the last equation, X is inversely proportional to the volume of reactor, V. At a specified SRT and organic loading, a certain mass of solids is generated and can be expressed as:

Biomass generation:

$$MX = V \cdot X = \frac{Q(C_{in} - C\ )Y \cdot SRT}{(SRT \cdot k_d + 1)} \quad \text{(eq. 3.6)}$$

### 3.4.3 Volume design

As mentioned earlier, X is inverse of the volume V. So, if the sludge is increased (biomass), the volume is reduced. The biomass can be increased by recycling it back to the bioreactor. Consequently, high substrate removal efficiency can be attained.

A given biomass concentration can determine the size of the bioreactor volume. For a design purpose, the biomass X is given, and the volume is determined based on the biomass generation expression:

$$V = \frac{MX}{X} \quad \text{(eq. 3.7)}$$

### 3.4.4 Mass balance endogenous residue

Decay is a loss of biomass due to death, endogenous respiration, cell lysis, maintenance and predation. There can be many approaches for description of decay in mathematical modeling. In activated sludge, the reduction of biomass by decay is balanced by accumulation of an unbiodegradable or inert particulate fraction and utilization of electron acceptor (Ydstebø, 2005).

The mass balance for endogenous residue can be generally given as:

Accumulation = inflow – outflow – waste

$$V.\frac{dX_E}{dt} = f_d.k_d.X.V - Q_w.X_{E_w} = 0\ (steady\ state) \quad \text{(eq. 3.8)}$$

$$X_E = f_d.k_d.X.\frac{V}{Q_w} = f_d.k_d.X.SRT \quad \text{(eq. 3.9)}$$

$$with\ SRT = \frac{VX_R}{Q_w X_w} = \frac{V}{Q_w}\ as\ X_R = X_w\ (same\ concentration)$$

Mass of endogenous residue:

$$MX_E = f_d.k_d.MX.SRT \quad \text{(eq. 3.10)}$$

This expression shows that the concentration of endogenous residue increases as the biomass increases.

### 3.4.5 Mass balance of inert residue

The mass balance for inert residue is given by:

$$V.\frac{dX_I}{dt} = Q.X_{Iin} - Q_w.X_{I_w} = 0\ (steady\ state) \quad \text{(eq. 3.11)}$$

$$\frac{V.X_I}{SRT} = Q_w.X_{I_w}$$

$$the\ influent\ inert,\ X_I = \frac{Q.X_{I_w}.SRT}{V} \quad \text{(eq. 3.12)}$$

Mass of unbiodegradable particulate:

$$MX_I = Q.COD_{up}.SRT \quad \text{(eq. 3.13)}$$

### 3.4.6 Sludge mass in bioreactor

Organics fractions: The MLVSS in the reactor is the addition of organic solids (active biomass and dead biomass)

Organic fractions = Biomass + Mass of Endogenous residue + Mass of Unbiodegradable organics

$$MLVSS = MX_H + MX_E + MX_I \quad \text{(eq. 3.14)}$$

The ratio MLVSS/MLSS is typically around 0.7-0.8

### 3.4.7 Model kinetic and stoichiometry

Table 3.3 present the matrix for process kinetics and stoichiometry of the aerobic system. Table 3.4 and table 3.5 show the different compounds and parameters applied in the aerobic carbon removal model.

Table 3.3 Kinetics and stoichiometry of an aerobic carbon removal system

| Variable / Process | $S_S$ | $S_O$ | $X_H$ | $X_S$ | $X_E$ | Rate equation (gCOD/l·d) |
|---|---|---|---|---|---|---|
| Growth of heterotrophs | $-\frac{1}{Y_H}$ | $-\frac{(1-Y_H)}{Y_H}$ | 1 | | | $\frac{\mu_{max}\cdot S_S}{K_S + S_S}\cdot X_H$ |
| Hydrolysis of SBCOD | 1 | | | -1 | | $k_H \frac{X_S / X_H}{K_X + X_S / X_H}\cdot X_H$ |
| Decay of heterotrophs | | $-(1 - f_d)$ | $-1$ | $f_d$ | $f_d$ | $k_d \cdot X_H$ |

Table 3.4 Compounds in the aerobic carbon removal model

| Description | Symbol | Unit |
|---|---|---|
| *Dissolved compounds* | | |
| RBCOD | $S_S$ | mgCOD/l |
| Dissolved oxygen | $S_O$ | mgO/l |
| *Particulate compounds* | | |
| Heterotrophic organisms | $X_H$ | mgCOD/l |
| SBCOD | $X_S$ | mgCOD/l |
| Inert residue from dead cells | $X_E$ | mgCOD/l |
| Inert particulate COD from influent | $X_I$ | mgCOD/l |

Table 3.5 Parameters in the aerobic carbon removal model

| Description | Symbol | Unit |
|---|---|---|
| *Stoichiometric parameters* | | |
| Growth yield for aerobic heterotrophic organisms | $Y_H$ | mgCOD/mgCOD |
| Unbiodegradable residue in cells | $f_d$ | mgCOD/mgCOD |
| *Kinetic parameters* | | |
| Maximum specific growth rate for heterotrophic organisms | $\mu_{max}$ | $d^{-1}$ |
| Hydrolysis rate | $k_H$ | $d^{-1}$ |
| Decay rate for heterotrophic organisms | k_dH | $d^{-1}$ |
| Half-saturation coefficient for RBCOD | K_S | $mgCOD_{Su}/l$ |
| Half-saturation coefficient for hydrolysis compounds | $K_X$ | mgCOD/mgCOD |

# 4 RESULTS AND DISCUSSION

## 4.1 HISTORY AND OVERVIEW OF THE EXPERIMENT

### 4.1.1 Operation conditions during experiment at 20ºC

#### *4.1.1.1 pH, temperature and conductivity*

The conditions of the experiment such as temperature, pH and conductivity are summarized in figure 4.1.

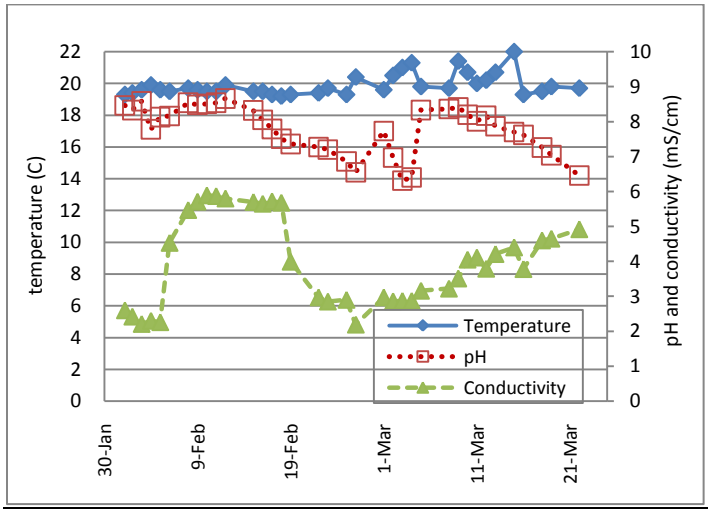


Figure 4.1 pH, temperature and conductivity at 20ºC

Along the period of experiment, the inlet wastewater varied not only with respect to organic content but also in ionic strength as we can observe in figure 4.1 for 20ºC and later in figure 4.3 for 4ºC. The ionic strength was quite high especially in between 8-18feb. This is believed to be caused by cold weather and application of salt to roads.

During the experiment, the temperature was almost constant, between 19-22ºC. The pH change is small along the experiment. It was kept at the range of 7.5 to 8.5 except for 2 or 3 days in between 25 february and 4 march where pH were recorded lower than 6.5, the minimum of the pH range reported for biological treatment in literatures. The pH drop was due to nitrification process in the reactor. The biological rates may be affected by that decrease in pH such as reduced growth rate. Therefore, as mentioned in methodology part (chap3), alkalinity was added to neutralize the pH.

The nitrification process can bring about pH drop because it consumes alkalinity. Hence, the 20ºC batch was seemingly affected by nitrification. Nitrification is a process by which ammonia is converted first to

nitrite by nitrifying bacteria *nitrosomonas* and then to nitrate by *nitrobacter*. In fact, during the process of transformation of ammonia to nitrate, $H^+$ is released. This engenders the decrease of pH.

$$NH_4^+ + 1.5O_2 \xrightarrow{nitrosomonas} NO_2^- + 2H^+ + H_2O$$

$$NO_2^- + 0.5O_2 \xrightarrow{nitrobacter} NO_3^-$$

In biological treatment process, the nitrogen is not removed by nitrification, but just converted to nitrate which will appear in the effluent. Therefore, the nitrate content of the effluent can be analyzed to check whether nitrification happened in the experiment. No significant amount of nitrate was determined during IC test meaning that nitrification can be neglected.

#### *4.1.1.2 MLVSS and SRT conditions*

The figure shows the condition of operation regarding MLVSS content and SRT along the experiment at 20ºC. MLVSS are the measured VSS content of the mixed liquor at the last hours of the end of each cycle. SRT was measured based on the total mass in the system over the total mass wasted in the decant and wasted sludge. So, a connection is laid between MLVSS content and the SRT.

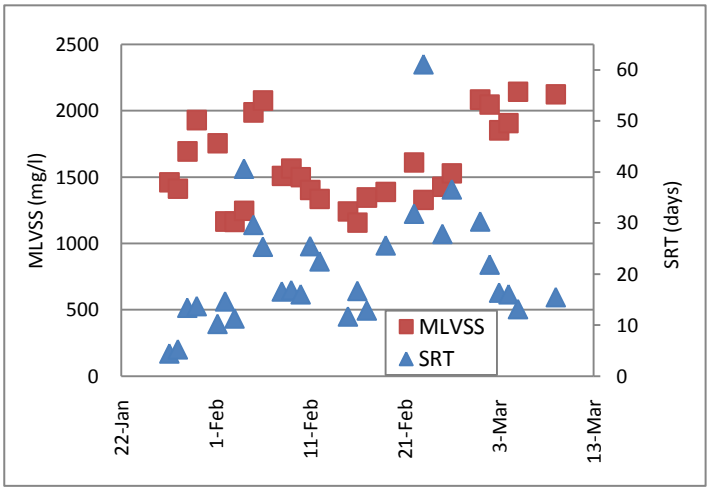


Figure 4.2 MLVSS and SRT during the experiment at 20ºC

The SRT and MLVSS varied along the period of experiment due to variation in operational procedures mainly sludge waste. The average SRT was estimated at 20.73 d varying between 4 and 60 days. This wide range of SRT is because we had 3 different wastewaters with regard to organic content (see later section 4.3). This could influence the operational procedures such as wasting of sludge as the response of the sludge growth could not be the same. This is also the case of the experiment at 4ºC (section 4.1.2.2).

There is effect of the SRT on biomass if we refer to the expression of biomass ($X_H$) (Chapter 3.4). The longer the SRT, the more sludge concentration is accumulated in the system. High SRT is met when the waste is too low. Therefore, at higher SRT, high MLVSS would be observed and lower biomass MLVSS corresponds to low SRT.

The reason for this high MLVSS up to February 6 was due to the high substrate concentration due to sugar addition.

### 4.1.2 Operation conditions at 4 and 8ºC

#### *4.1.2.1 pH, temperature and conductivity*

pH, temperature and conductivity from tests at 4º and 8ºC is presented in figure 4.3 and 4.4.

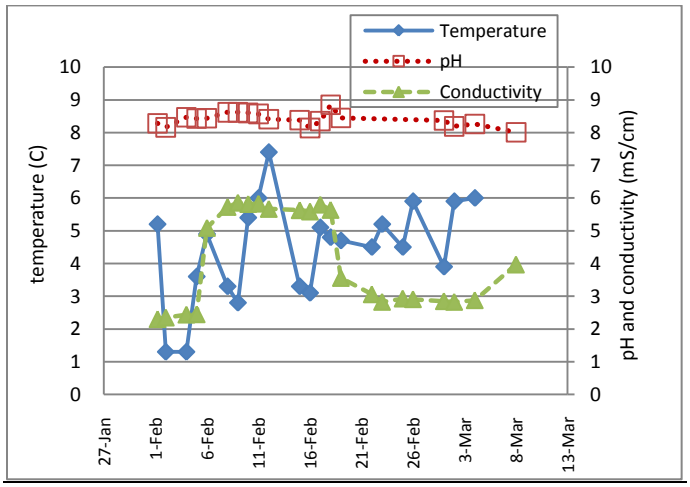


Figure 4.3 pH, temperature and conductivity at 4ºC

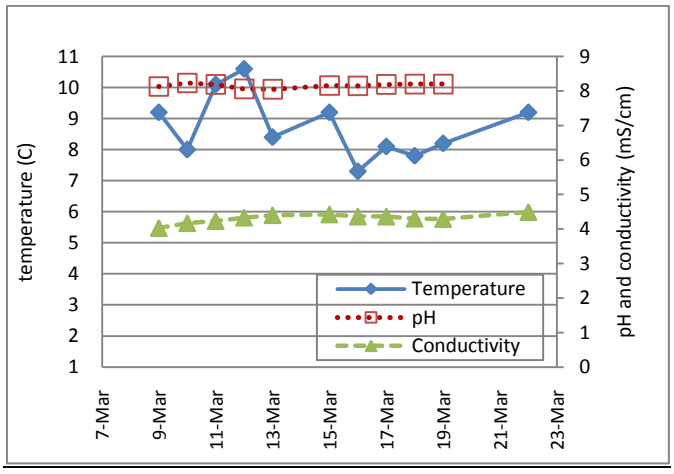


Figure 4.4 pH, temperature and conductivity at 8ºC

Temperature variation in 8ºC where relatively stable, between 7 and 10ºC. At 4ºC, temperature was unstable (between 1 to 6ºC) due to the uncontrolled variation of the room condition.

Unlike at 20ºC, pH was stable for experiment at 4 and 8ºC. pH was maintained higher than 8 and no decrease in pH was recorded. At lower temperature, the nitrification process did not occur and no nitrate was observed in the effluent.

According to U. Wiesman et al (2007), the nitrification rate is a function of temperature and at low wastewater temperatures in winter (< 10C), nitrification will normally not occur.

Due to low growth rate, long SRT is required for the growth of nitrifiers and at 8ºC, the SRT is short (about 5 days) which is not sufficient for nitrification.

At 4ºC, the wastewater used was the same as at 20ºC, the conductivity was also the same varying between 2 and 6 mS/cm corresponding to about 10 times more salt than regular wastewater.

At 8ºC, the conductivity did not vary because only one type of wastewater was used.

#### *4.1.2.2 MLVSS and SRT conditions*

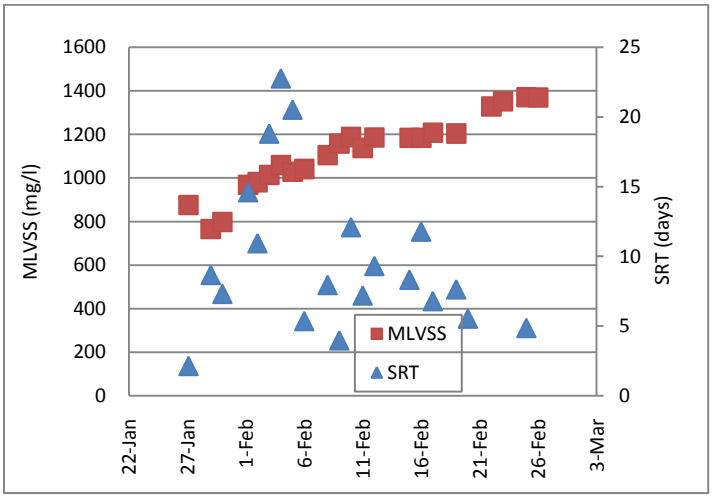


Figure 4.5 MLVSS and SRT during the experiment at 4ºC

At 4ºC, SRT was calculated to about 9.02d varying between 3 and 20 days.

At 4ºC, MLVSS increased gradually and stabilized at the end of the experiment. The growth at lower temperature is reduced, so it takes longer time for the reactor to build up sufficient biomass concentration.

At the end of the experiment, when SRT was low (around 5d), more concentrated and steady biomass was maintained. Thus, this low SRT was sufficient for biomass growth even at lower temperature.

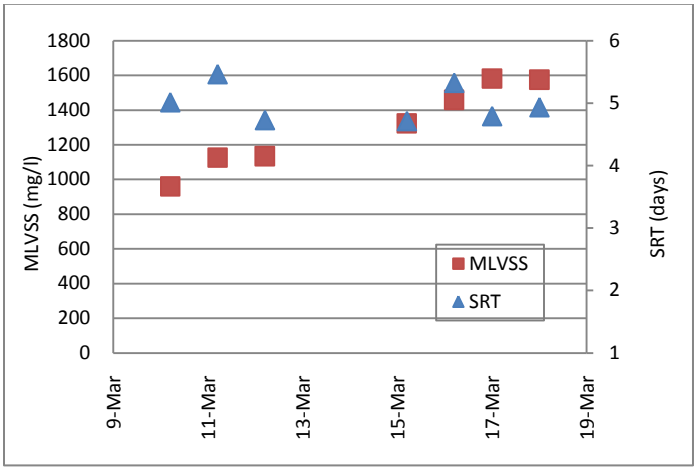


Figure 4.6 MLVSS and SRT during the experiment at 8ºC

An average SRT of 5d was estimated at 8ºC with a range between 4.72 and 5.47d.

A similar shape of the MLVSS curve as at 4ºC was observed. MLVSS builds up little by little until reaching a plateau where it was constant. The main difference is that the plateau was reached only 10days at 8ºC while it took almost 4weeks for 4ºC.

There are 2 explications for that: First, it is due to lower activity of the organisms at lower temperature. Second, at 8ºC, as there was already much sludge initially in the reactor (sludge remained from 4ºC experiment), it was not necessary to grow sludge from nothing like initially at 4ºC. A constant wasting was done from the beginning of the experiment, which may have speeded up the stabilization of the biomass concentration.

It can be observed from figure 4.6 that when the SRT was low (around 4 - 5.3d), the biomass kept on increasing until a steady concentration. As at 4ºC, this was sufficient SRT for 8ºC as MLVSS did not decline in this range.

The SRT was much lower at 4ºC and 8ºC compared to 20ºC; SRT was 5 and 9.02 against 20.73 found for 8ºC, 4ºC and 20ºC respectively. This is not common as SRT should normally decrease with increasing temperature. The explanation is the way of sampling and wasting of sludge during the experiment and there have been wasted relatively more at 4ºC and 8ºC than at 20ºC.

### 4.1.3 Conditions in a reaction cycle

#### *4.1.3.1 SBR test result at 20ºC*

The OUR, TOC and MLVSS during a reaction cycle is shown in figure 4.7.

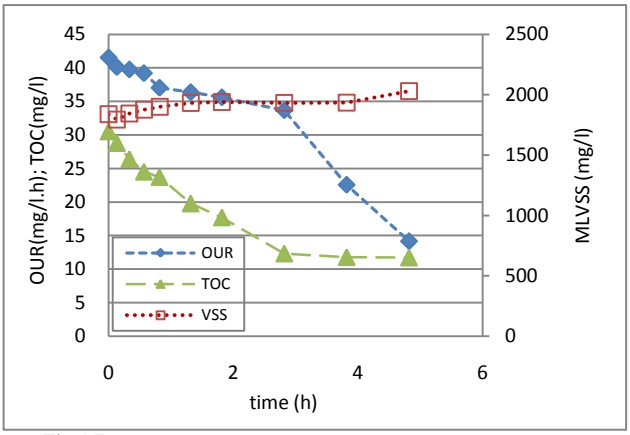


*Fig 4.7.a*

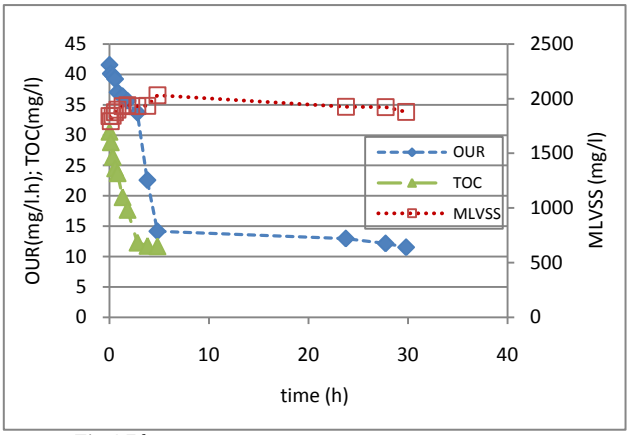


*Fig 4.7.b*

Figure 4.7 OUR, TOC and MLVSS during a reaction cycle at 20ºC (9th march)

At 20ºC, after addition of substrate, immediate high oxygen consumption is observed. The system responds to the substrate immediately with the highest OUR which corresponds to a rise in MLVSS concentration and immediate drop in substrate concentration. After a short discontinuity (fig 4.7a), a second drop of OUR occurs while the corresponding MLVSS continues growing at a more or less constant rate until reaching a peak point (figure 4.7b). This peak is reached only after 5-8 hours (5h in figure 4.7). The second drop corresponds probably to the exhaustion of substrate causing the MLVSS to gradually decrease.

The initial phase with rapid removal of substrate, rapid increase in MLVSS and high OUR corresponds to the uptake of easily biodegradable substrate. Only soluble organics in the substrate give rise to such immediate oxygen consumption (S.H. Jenkins, 1979).

As observed in fig 4.7b, where the second OUR drop stops (after 4-5hours), the biomass has reached its maximum, corresponding to that substrate was exhausted.

The initial phase of high OUR is related to assimilation of RBCOD and terminated at 0.5-1h while the 2nd OUR phase which corresponds to SBCOD was between 1-1.5 to 5-6 hours (example in fig 4.7a). The second phase is related to the use of the SBCOD resulting both in slower growth of biomass or apparently more constant concentration of MLVSS (fig 4.7a). The constant MLVSS concentration is due to net balance between SBCOD consumption (reduction of MLVSS) and the biomass increase (increase of

MLVSS) since MLVSS is the sum of biomass, unbiodegradable fractions and SBCOD. The observed constant removal of substrate is the removal of slowly biodegradable substrate.

The final phase is the decay or endogenous respiration phase. This phase started when the substrate was exhausted and the MLVSS starts to decrease. There is no more available substrate and the organisms are starving. Any activity (OUR) still observable in that phase is the substrate related to endogenous respiration where there is auto-degradation of the dead organisms. Died biomass are degraded by the remaining active biomass and unbiodegradable biomass also accumulates as inert residue.

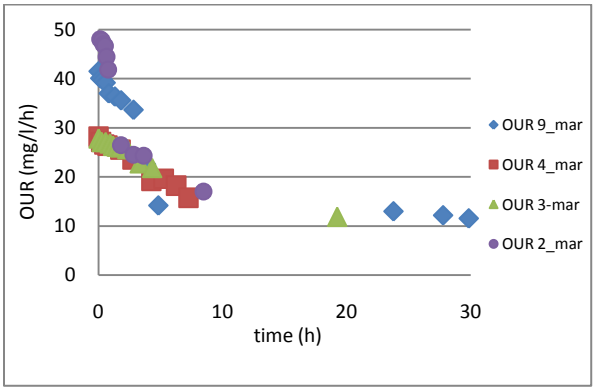


Figure 4.8 OUR profile during several reaction cycles at 20ºC

Figure 4.8 shows the OUR results obtained from different cycles at 20ºC. It demonstrates that the first rapid drop of OUR is within 0.5 to 1 h in most of the tests independently the initial OUR relating to the RBCOD. This is followed by the second drop of OUR at about 5-8 hours referring to the utilization of SBCOD.

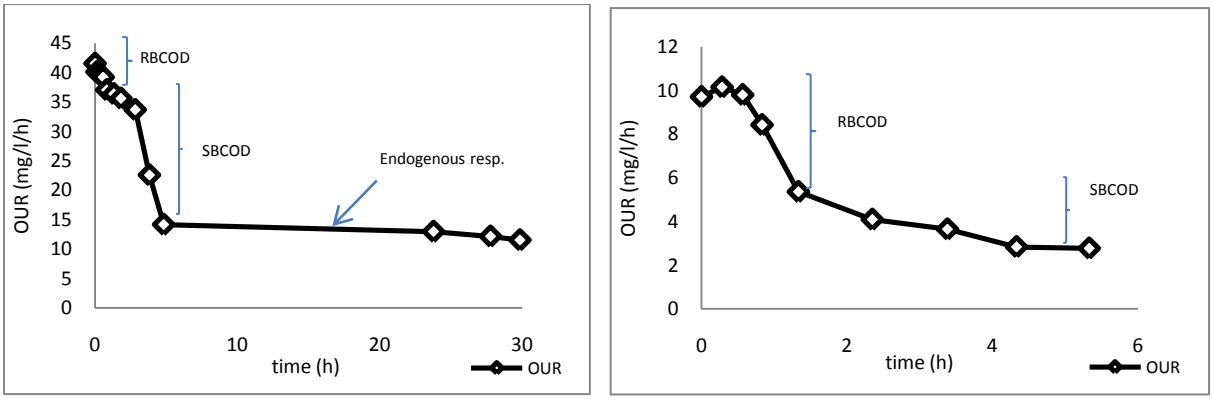


*fig 4.9.a* *fig 4.9.b*

Figure 4.9 Profiles of OUR during a reaction cycle at 20ºC (left) and at 4ºC (right)

Figure 4.9 represents a general feature of the OUR response in a biological reaction and 3 distinct phases are seen from the profile of OUR. Initial phase with a high oxygen consumption (fig 4.9a) is linked to

utilization of soluble substrate consumption, RBCOD. The soluble RBCOD passes directly through cell wall of the organisms and is metabolized at high rate (Ekama and Marais, 1984). Thus, RBCOD is used directly by organisms resulting in a rapid response and corresponding biomass increase (fig 4.7.b).

Phase 2 of the OUR represents utilization of the particulate slowly biodegradable substrate, SBCOD. The particulate slowly biodegradable substrate SBCOD cannot be metabolized directly by the organisms. The particulate COD requires adsorption, storage and extracellular breakdown prior to absorption by the cells (Ekama and Marais, 1984). This is why the growth rate of biomass is lower as substrates are not directly assimilated.

The last phase represents endogenous respiration where there is no external substrate left, but the activity is due to degradation of the dead cell compounds. The gradual decrease of OUR in this period is due to the gradual reduction in biomass as a result of decay.

At lower temperatures, the initial phase was longer than at 20ºC and terminated after 1- 2 hours. For some cycles, there is an increase in OUR in the initial phase as a result of rapid biomass growth. At 4ºC, the 2nd phase where SBCOD is degraded is much longer at lower temperatures than at 20ºC due to reduced hydrolysis rate. Thus, the change from the 2nd and the 3rd endogenous phase is not so distinguishable.

#### *4.1.3.2 SBR test result at 4ºC*

Figure 4.10 shows the OUR, TOC and MLVSS in a cycle at 4ºC

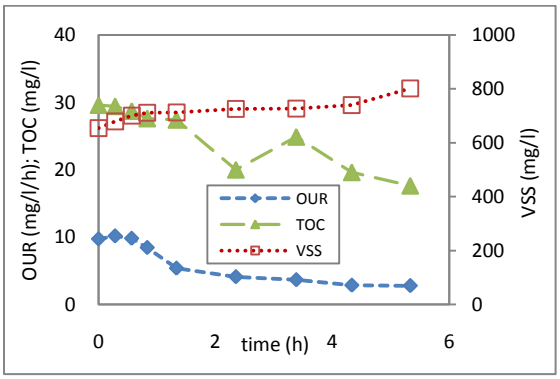


*fig 4.10.a*

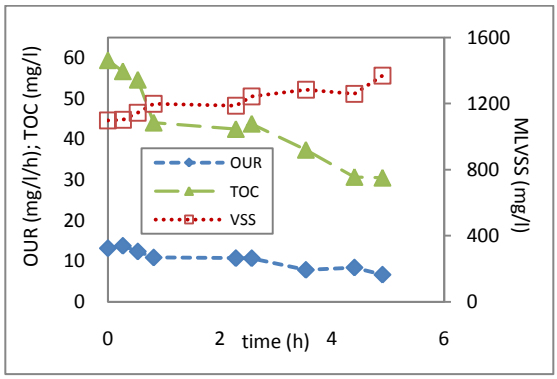


*fig 4.10.b*

Figure 4.10 OUR, TOC and MLVSS at 4ºC (fig4.10a, 18th and fig4.10b, 22nd feb)

As mentioned earlier, only two phases is observed (figure 4.10), since the growth rate is low and the transfer from phase 2 to 3 is not very distinguished.

A gradual increase in MLVSS and removal of substrate is observed representing the growth in the process. Unlike at 20ºC, at 4ºC, the decrease in OUR is not abrupt in the initial phase because at low temperature the organisms are less active than at higher temperature and thus, the substrate consumption rate is also lower. The second phase of OUR has a very slow decrease. The initial slight increase in OUR corresponds to immediate removal of substrate and increase in MLVSS. The peak of MLVSS is observed after 5 to 6 h when substrate reached low concentration or exhaustion.

Like figure 4.9b, in figure 4.10 has no clear endogenous phase is not yet present or it is not distinguishable from the 2nd lower phase of OUR. The corresponding MLVSS is at the maximum (increasing) at 5h where the endogenous phase has not yet started. Endogenous phase would follow after the 2nd phase and MLVSS would be expected to decrease (the relating data was not recorded).

#### *4.1.3.3 SBR test result at 8ºC*

Figure 4.11 represent the TOC, MLVSS and OUR profile from the experiment at 8ºC.

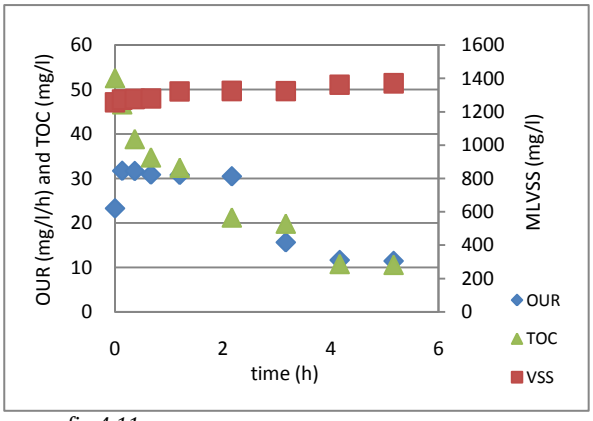


*fig 4.11a*

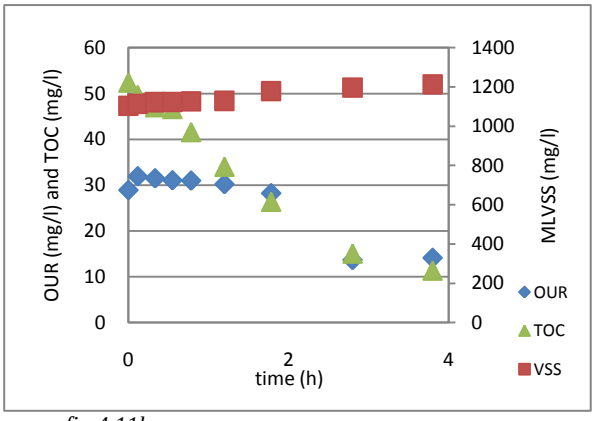


*fig 4.11b*

Figure 4.11 OUR, TOC and MLVSS at 8ºC (16 and 18 mar)

At 8ºC, more or less the same feature as in 4ºC is observed regarding the OUR profile and the MLVSS increase. The main difference is in OUR (figure 4.11). Figure 4.12 shows the OUR profile for 8ºC and the initial phase seemed to be longer compared to at 4ºC. Besides, the initial increase in OUR apparently is higher at 8ºC than at 4ºC. The initial increase of OUR could be because the utilization of substrate was not only direct but also through storage.

The second phase is a drop in OUR which reflect a continuous utilization of the stored substrate (from 3h in figure 4.12). At 5h, a peak of MLVSS is reached as seen at 4ºC, corresponding to the exhaustion of substrate. The endogenous respiration is not shown in figure 4.11 and fig 4.12.

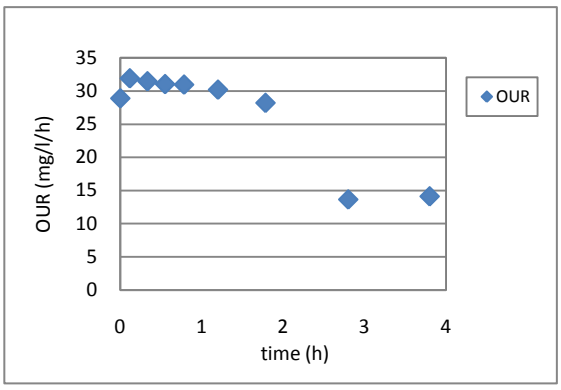


Figure 4.12 OUR profile during a reaction cycle at 8ºC

Compared to OUR profiles in figure 4.9 and figure 4.13 (in next section), the second phase at 8ºC is significantly longer than at 20ºC but a little longer than at 4ºC (around 2h). This strange behavior compared to 4ºC is incomprehensible since it would be expected that the behavior at 8ºC should be in between 20º and 4ºC

## 4.2 EFFECT OF TEMPERATURE ON PROCESS RATES

### 4.2.1 Effect of temperature on activity rate of microorganisms

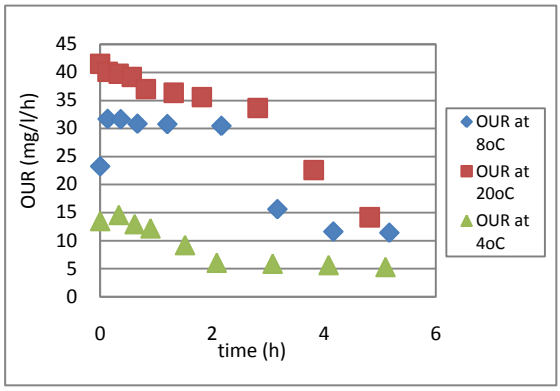


Figure 4.13 OUR at different temperature.

Figure 4.13 indicates OUR is highest at 20ºC during a reaction cycle. Thus, there is an influence of temperature on the removal rate of substrate which is linked to growth and OUR. The lower the temperature, the lower the OUR and growth rate due to slow uptake of the substrate.

Temperature can exert an effect on biological reactions in 2 ways: by influencing the rates of enzymatically catalyzed reactions and by affecting the rate of diffusion of substrate to the cells (C.P.L Grady and G.T.Daigger, 1999).

### 4.2.2 Effect of temperature on growth rate

Figure 4.14 displays the effect of temperature on maximum growth rate. Calculation of the maximum growth rate was based on the calculated SRT and $k_{d.}$

The maximum growth rate, $\mu_{max}$ is a function of temperature and so is the decay rate, $k_d$.

At 4ºC, determination of decay rate was based on temperature correction from 20ºC with a temperature coefficient Θ of 1.03 (Ydstebø, 2009). As a result, a $k_d$ of 0.07$d^{-1}$ and a $\mu_{max}$ of 0.77 $d^{-1}$ were evaluated.

$k_d$ at 20ºC was experimentally determined to 0.11 $d^{-1}$ and 0.12$d^{-1}$at 8ºC. The $\mu_{max}$ was determined at 2.53 $d^{-1}$ and 2.84 $d^{-1}$ at 20º and 8ºC respectively, which is strange as $\mu_{max}$ and $k_d$ at 8ºC is expected to be lower than at 20ºC.

To avoid error in calculation of parameters for the modeling part, a temperature correction was used for 8ºC, using the $k_d$ value evaluated at 20ºC. Thus, a $k_d$ of 0.08$d^{-1}$ at 8ºC was estimated.

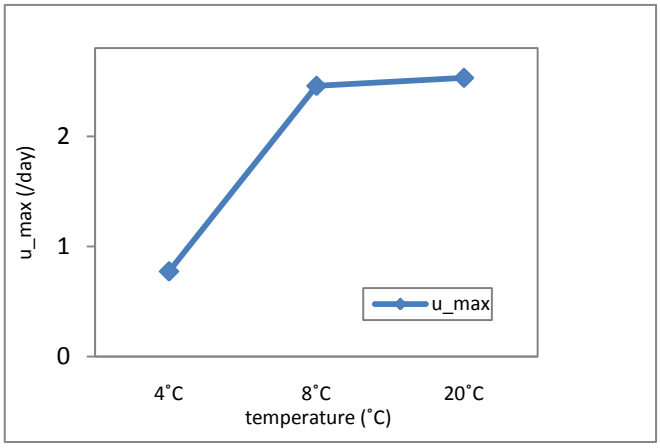


Figure 4.14 Effect of temperature on maximum growth rate

Figure 4.14 shows that the growth rate of microorganisms increases with temperature. The maximum growth rate at 8ºC and 20ºC are not significantly different as we found 2.46 and 2.53$d^{-1}$ respectively. This is not likely to be true as there should be big distinction between 8ºC and 20ºC and less difference between 4º and 8ºC. According to what has been mentioned earlier, the maximum growth rate of microorganism doubles approximately each 10ºC up to a maximum temperature (Ystebø, 2009 and Metcalf&Eddy, 1991). Thus, the values found from our calculation were not in accordance with that.

Probably, the inaccordance may arise from the fact that the experiment at 8ºC was very short, only 10days experiment for the main biological reaction. This may create ambiguity compared to results at 4ºC and 20ºC where there was much more data.

Moreover, another reason is probably the maximum growth rate calculation. There was big difference and inappropriate data in the SRT values as mentioned in section 4.1.2, especially at 20ºC which was overestimated.

## 4.3 TREATMENT PERFORMANCE

Table 4.1 represents the BOD and COD removal efficiency from the SBR lab-scale experiment.

Table 4.1 Inlet and effluent BOD and COD content of the wastewaters

| | $BOD_5$ (mg/l) | | $BOD_5$ | influent COD | | Effluent COD | | COD removal | |
|---|---|---|---|---|---|---|---|---|---|
| | inf_BOD | eff_BOD | removal | unfiltered | filtered | unfiltered | filtered | unfiltered | filtered |
| **ww 1** | 86 | N/A | N/A | 79 | 47.5 | N/A | N/A | N/A | N/A |
| **ww 2** | 61.40 | N/A | N/A | 168 | 92.25 | | | | |
| 20ºC | | | | | | 84.50 | 72.00 | 50% | 22% |
| 4ºC | | | | | | 81 | 70.00 | 52% | 24% |
| **ww 3** | 152.8 | | | 376.5 | 135.75 | | | | |
| 20ºC | | 7.5 | 95% | | | | | | |
| | | | | | | 70.00 | 54.00 | 81% | 60% |
| | | | | | | 78.00 | 62.00 | 79% | 54% |
| | | | | | | 53.50 | 46.50 | 86% | 66% |
| | | | | | | 51.50 | 48.00 | 86% | 65% |
| | | | | | | 76.50 | 58.50 | 80% | 57% |
| 4ºC | | 40.7 | 73% | | | 109.50 | 79.50 | 71% | 41% |
| | | | | | | 347.50 | 74.50 | 8% | 45% |

Table 4.1 shows that we had 3 different wastewaters during the experiment, the 2 first (ww1 and ww2) were quite low in $BOD_5$ content. That is why sugar addition was necessary to enhance the sludge growth at the beginning of the experiment.

The information about $BOD_5$ removal is available only for wastewater 3. As shown in table 4.1, 95% was the efficiency of $BOD_5$ removal for 20 ºC, 73% at 4ºC. Result at 8ºC was not recorded due to unavailability of the measurement apparatus.

As referred to the EU regulations mentioned in previous chapter, the effluent at 20ºC is according to the requirements with a high efficiency at 20ºC. The measurement at 4ºC was little lower but close to the effluent limit (73% vs. 75%). As a conclusion, the BOD efficiency removal at 4ºC was not poor but sufficient for the low temperature at the tested conditions.

Due to inaccuracy of the COD measurement in the beginning of the period, COD analyses were done for wastewater 2 and wastewater 3and their effluent were analyzed.

Table 4.2 represents the $BOD_5$, COD and TOC variations of some municipal wastewater plants reported by Droste (1997). It was noted that the data was from only a limited number of plants but the variation of the ratios is typical for treatment of domestic wastewater.

Table 4.2 Organic variation in treatment of municipal wastewater

| | $BOD_5$ (mg/l) | | COD (mg/l) | | TOC (mg/l) | | $BOD_5$/ TOC | | COD/TOC | |
|---|---|---|---|---|---|---|---|---|---|---|
| | average | range | average | range | average | range | average | range | average | range |
| Raw | 86 | 72-105 | 236 | 136-304 | 56 | 41-70 | 1.5 | 1.31-1.88 | 4.16 | 3.32-4.68 |
| Primary effluent | 58 | 46-68 | 204 | 146-299 | 52 | 44-61 | 1.11 | 1.00-1.33 | 3.9 | 3.19-5.85 |
| Final effluent | 15 | 11-20 | 84 | 77-95 | 35 | 33-40 | 0.44 | 0.20-0.69 | 2.4 | 2.02-2.58 |
| Ave. removal (%) | 83 | | 64 | | 38 | | | | | |

Source: Droste (1997)

It is shown that the average $BOD_5$ removal from the literature is a little lower but close to what we calculated, especially at 20ºC. Estimated COD removal from our experiments was better than reported in table 4.2.

COD removal in the experiment was estimated at 79-86% at 20ºC and ca. 71% at 4ºC. The latter is just around the threshold of permit for secondary effluent in EU regulations.

Thus, from the experiment results, the effluent quality at 20ºC and 4ºC meets the limit. For 4ºC, part of the results in effluent quality for $BOD_5$ removal was around the required level but close enough to expect that this can be improved by providing appropriate control and operation at 4ºC. In fact, during our experiment, the reactor at 4ºC was subjected to temperature changes due to uncontrolled temperature-room condition such as reported in section 3.1.2 (figure 4.3). This might have affected the treatment capacity of the reactor and the effluent quality.

Figure 4.15 represents the tendency of effluent (soluble COD effluent as from filtered samples) as function of time. The data was not available to the end of the cycle but only at 4-5 hour of test.

Substrate concentration was expressed as TOC, which was the method we used frequently. The COD equivalent of the substrate concentration was based on a COD/TOC ratio of 3; this ratio was chosen

appropriate for our data according to the variations in table 4.2.

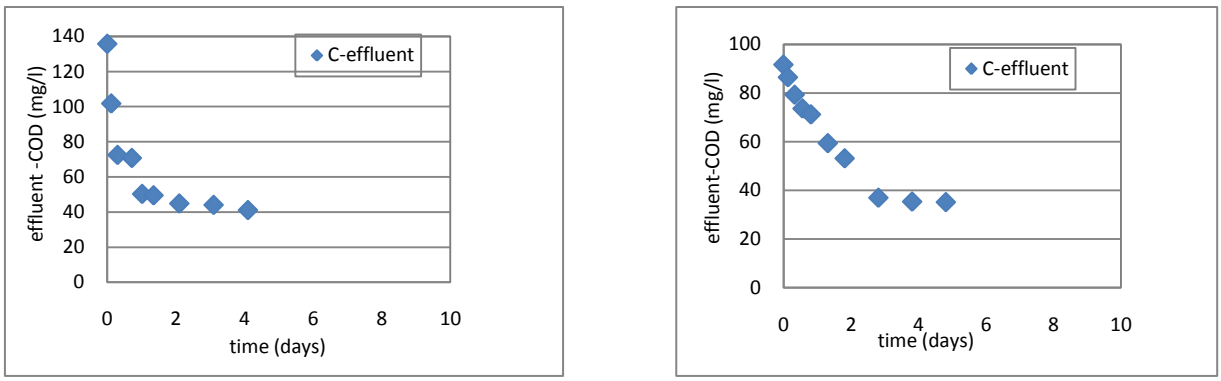


Figure 4.15 Substrate concentration effluent-COD as function of time at 20ºC (from filtered TOC multiplied 3 times)

Figure 4.15 demonstrates the soluble effluent COD. It can be deducted that the effluent COD decreases with time. At 4 to 5 hours, a lower and constant substrate concentration of about 30 to 40mg/l has been reached at 20ºC. The shape of the graph indicates that most of the biodegradable COD has been degraded and the remaining COD is unbiodegradable.

The measured total COD of influent wastewater varied, as can seen in table 4.1. Wastewater 1, 2 and 3 had a COD evaluated at an average of 79, 168 and 377 respectively. Since wastewater 3 was the one used during the most important experimental period, the value of 377 mg/l was used in the final estimation of the COD fractions.

Generally, according to Ydstebø (lecture, 2009), 5-10% of COD in wastewater is unbiodegradable soluble COD (which is from effluent filtered COD or from TOC), and 10-15% is unbiodegradable particulate COD which become part of VSS. Therefore, for this wastewater (ww3), unbiodegradable soluble COD was assumed to be 30-40mg/l according to TOC result and 40-60mg/l was assumed unbiodegradable particulate COD based on the total COD which is approximated to 400mg/l. Thus, for the total COD of 400mg/l the biodegradable part is considered to be around 300-330mg/l. This latter would be applied in for the dynamic simulation of the aerobic removal process in the following section.

## 4.4 DYNAMIC SIMULATION OF THE AEROBIC CARBON REMOVAL

The model given in the mathematical modeling chapter 3.4 was applied in the program Aquasim and different simulations were done to try to determine the agreement between our measured experimental data and theoretical predictions.

During simulations, the yield of 0.66 in COD units (0.45mgVSS/mgCOD*1.42COD/VSS) from literature (Ydstebø, lecture notes 2009) was applied. In addition, hydrolysis rate $k_H$ and maximum growth rate $\mu_{maxH}$ are the parameters to be estimated.

The prediction was performed in two steps:

- First, simulation with values from steady state calculation were applied
- Second, estimation of parameter $\mu_{maxH}$ and $k_d$ were performed with the aquasim based on experimental observations

### 4.4.1 Simulation with values of steady state model

When the estimated parameters from the experiments such as $\mu_{maxH}$, $k_d$, initial biomass concentrations and measured total COD were incorporated in the model, the simulations predicted higher OUR than the experiments. The model predicted about 80% higher OUR than the measured OUR (figure 4.17a) at 20ºC, 26% and ca. 50-60% increase in OUR at 8ºC and 4ºC respectively (fig 4.18a and fig 4.19a).

The input parameters are presented in table 4.3. Figure 4.17a, 4.18a and 4.19a show simulation with the estimated values from steady state. Prediction does not fit at all except for 20ºC. So, parameter estimation of $K_H$ and $\mu_{max}$ with the results from steady state were performed with the dynamic simulation.

At 4ºC, when the measured values (or from steady state) were used in the simulation, the model OUR was too high and far from the experimental (figure 4.18a). It could be that the biomass concentration in the system was a little lower than what was measured, 1300mg COD/l. If the latter was decreased, a better OUR prediction would be obtained (see section simulation with assumed values)

### 4.4.2 Parameter estimation with dynamic simulation

In order to reproduce the measured OUR, the hydrolysis rate and the maximum growth rate were estimated based on the measured values.

Table 4.3 summarizes the input parameters for the simulation and table 4.4 shows the estimated kinetic parameters for each temperature.

Table 4.3 Parameter inputs in the aerobic carbon removal model

| Description | Value | Unit |
|---|---|---|
| ***Dissolved compounds*** | | |
| RBCOD | 50 | mgCOD/l |
| Dissolved oxygen | >7 | mgO/l |
| ***Particulate compounds*** | | |
| Heterotrophic organisms | 800-1600* | mgCOD/l |
| SBCOD | 250-300 | mgCOD/l |
| Inert residue from dead cells | 140-160 | mgCOD/l |
| Inert particulate COD from influent | 200-250 | mgCOD/l |
| ***Stoichiometric parameters*** | | |
| Growth yield for aerobic heterotrophic organisms | 0.66 | mgCOD/mgCOD |
| Unbiodegradable residue in cells | 0.20 | mgCOD/mgCOD |
| ***Kinetic parameters*** | | |
| Half-saturation coefficient for RBCOD | 10 | mgCODSu/l |
| Half-saturation coefficient for hydrolysis compounds | 0.027 | mgCOD/mgCOD |

**from the steady state calculation for different temperature*

Table 4.4 Estimated parameters based on experimental observations

| Description | Values* | | | Unit |
|---|---|---|---|---|
| | 4º | 8º | 20º | |
| *Kinetic parameters* | | | | |
| Maximum specific growth rate for heterotrophic | 0.5 | 2.07 | 1.22 | $d^{-1}$ |
| Hydrolysis rate | 0.085 | 1.68 | 1.36 | $d^{-1}$ |
| Decay rate for heterotrophic organisms | 0.07* | 0.08* | 0.11* | $d^{-1}$ |

**from measurements and steady state ($k_d$ and $X\_H_{ini}$ values)*

The biomass concentration estimated from steady state and used in the simulation was 1300 mgCOD/l, 800 mgCOD/l and 1600 mgCOD/l at 4º, 8º and 20ºC respectively.

At 8ºC, $\mu_{max}$ was estimated higher than at 20ºC with 2.46 $d^{-1}$ estimated in steady state and 2.07 $d^{-1}$ after parameter estimation. Besides, with dynamic simulation (table 4.4), $\mu_{maxH}$ was estimated much higher than at 20º C (2.07$d^{-1}$ against 1.22 $d^{-1}$) which is not likely. This is because for the two evaluations, the biomass concentrations from steady state were very much different. The value for 20ºC was double of that at 8ºC. It is evident that this is wrong and calculation at 8ºC was underestimated as it is much lower

than at 4ºC. This inaccordance is probably due to error in experimental practice which resulted in overestimations of $\mu_{maxH}$ and $k_d$ at 8ºC.

Figure 4.16b, 4.17b and 4.18b indicates that the model predicted very close to measurements at 20ºC and 4ºC and while at 8ºC, the prediction was close but with more deviations.

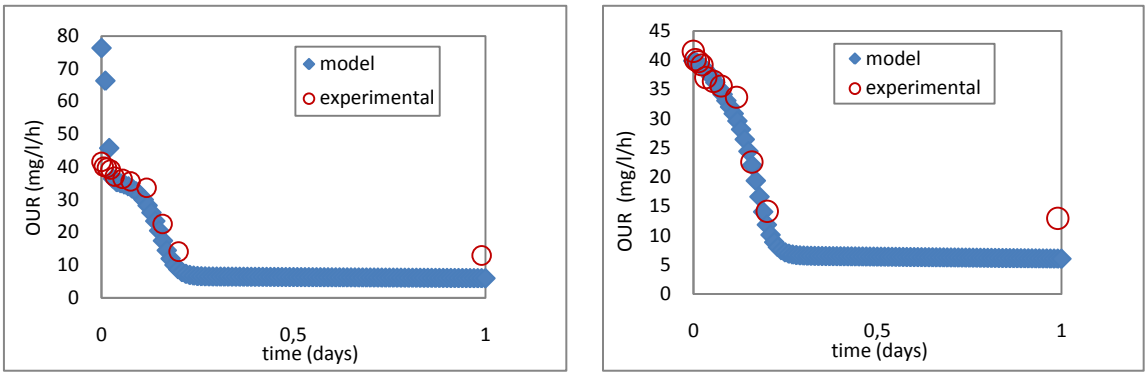


Figure 4.16 Simulation of OUR at 20ºC with the parameters from steady state model (fig 4.16a) and with estimated parameters based on the experimental results (fig 4.16b)

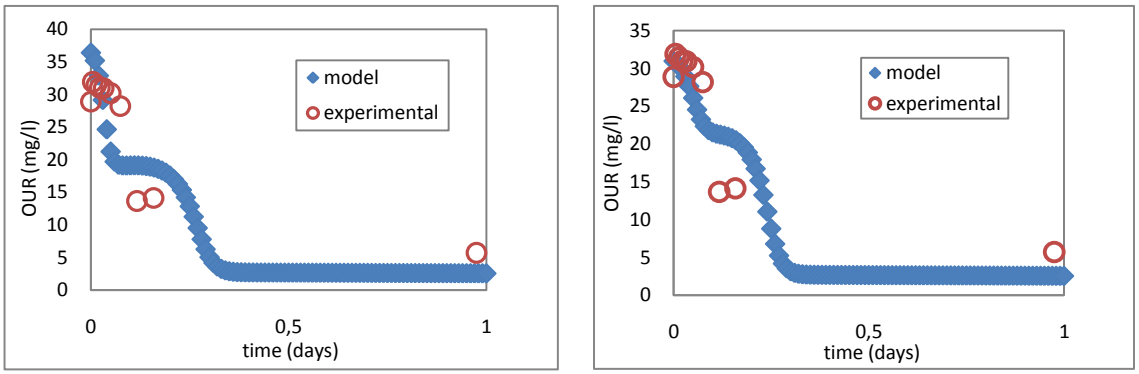


Figure 4.17 Simulation of OUR at 8ºC with the parameters from steady state model (fig 4.17a) and with estimated parameters based on the experimental results (fig 4.17b)

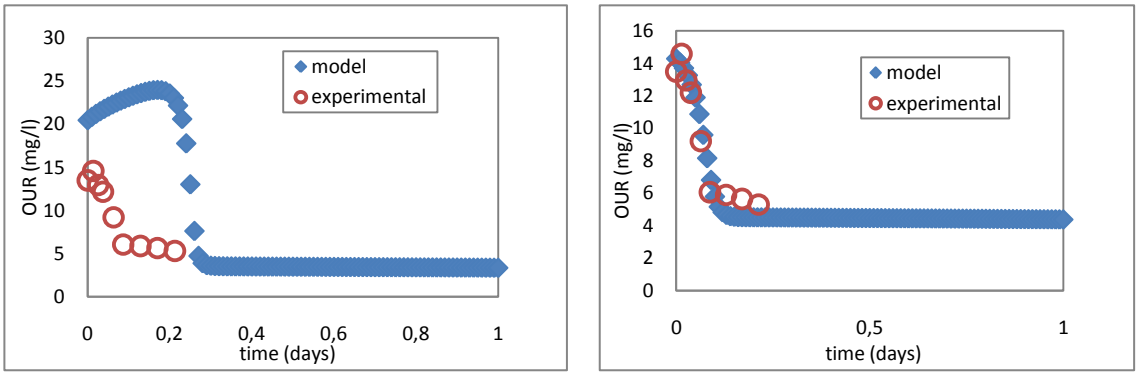


Figure 4.18 Simulation of OUR at 4ºC with the parameters from steady state model (fig 4.18a) and with estimated parameters based on the experimental results (fig 4.18b)

At 20ºC, predicted OUR fitted relatively close to the measured except for the last phase around 24h. This is probably because the experimental was estimated with higher $\mu_{max}$ , and overestimated compared to that predicted in the model. We measured steady state model with $\mu_{max}$ of 2.53$d^{-1}$ against 1.22 $d^{-1}$ for dynamic model.

### 4.4.3 Simulation with assumed parameters values

The relation between $\mu_{max}$, $K_{H,}$ $X_{Hini}$ and $k_d$ is the key for the simulation since the estimated parameters $\mu_{max}$, $K_H$ are inversely proportional to the initial biomass concentration $X_{Hini}$ . The expression for $\mu_{maxH}$ confirm this relation (chapter 3.4 and eq.4.1). The simulated parameters are also inversely dependant to $k_{d,}$ but to a small degree.

Therefore, by adjusting the parameters such as reducing decay rate, decreasing initial sludge concentrations (especially for 20º and 4ºC), the predicted and the experimental data correlate more closely.

The values were estimated by increasing or decreasing the initial biomass concentration with a decay rate of around 0.2$d^{-1}$, 0,1$d^{-1}$ and 0.05$d^{-1}$ at 20º, 8º and 4ºC respectively.

In the simulation, at constant $X_{Hini}$ and increasing $k_d$, the hydrolysis rate and the maximum growth rate are decreased. If $k_d$ increased from 0.2 to 0.22 at 20ºC, the estimated parameters which was 1.92$d^{-1}$ for both $K_H$ and $\mu_{max}$ was decreased to 1.91 $d^{-1}$and 1.89 $d^{-1}$ respectively.

Moreover, at constant $k_d$ and changing $X_{Hini}$, $K_H$ and $\mu_{max}$ change as well. $K_H$ and $\mu_{max}$ of 1.92$d^{-1}$ was reduced to 1.55$d^{-1}$ and 1.4$d^{-1}$ respectively when $X_{Hini}$ increased from 1000 to 1300mgCOD/l. Figure 4.16 shows this relation between biomass concentration and $\mu_{max}$ in addition to OUR.

Figure 4.19 shows the relation between the maximum growth rate and the biomass concentration.

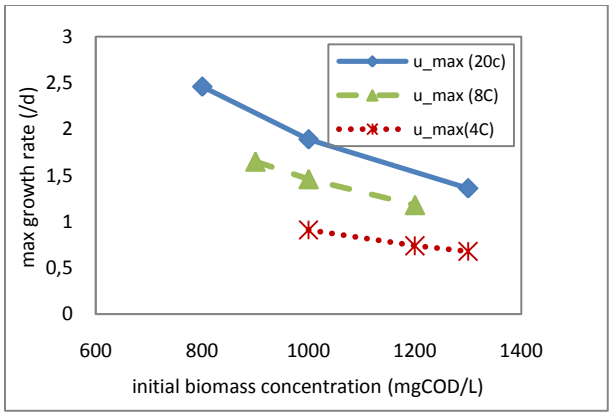


Figure 4.19 Estimated $\mu_{max}$ as function of biomass concentration

This relation is dictated by the expression $\mu_{max} = \frac{\text{OUR}.Y}{(1-Y)\,X_H}$ (eq. 4.1)

Figure 4.20, figure 4.21 and figure 4.22 present the simulated and measured OUR during a reaction cycle at 20ºC, 8ºC and 4ºC respectively. Assumed values other than values from steady state were used to fit better the model to the measured OUR.

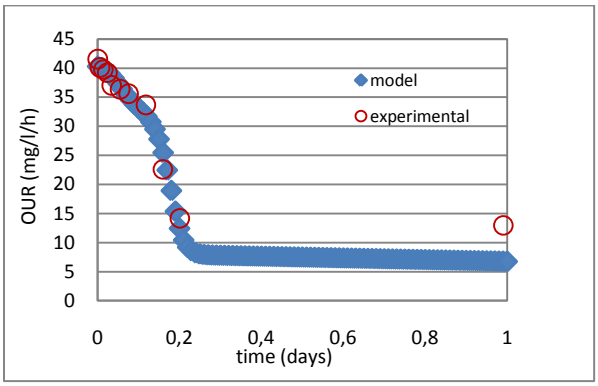


Figure 4.20 Experimental and simulated OUR in a cycle at 20ºC

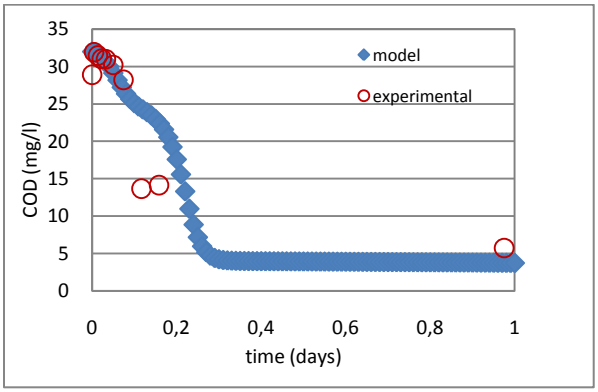


Figure 4.21 Experimental and simulated OUR in a cycle at 8ºC

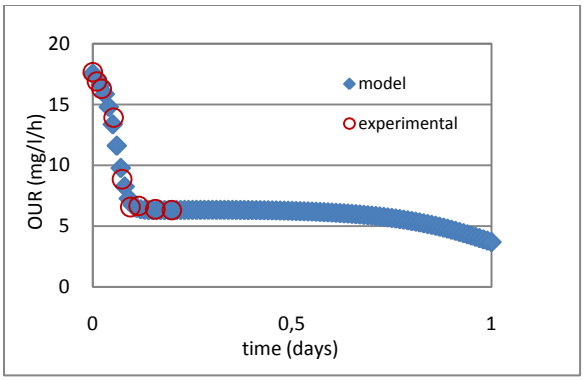


Figure 4.22 Experimental and simulated OUR in a cycle at 4ºC

As figure 4.22 shows, the model predicted very adequately OUR at 4ºC when the decay rate were changed to 0.05$d^{-1}$ and maximum growth rate of 0.76 $d^{-1}$ were also estimated. This was close to $k_d$ and $\mu_{max}$ of the estimated values from the steady state calculations which were 0.07 $d^{-1}$ of and 0.77 $d^{-1}$ respectively.

At 20ºC, there is no significant difference in the prediction of OUR from steady state (fig 4.16b) and that from assumed values (fig 4.20). For the latter, a decay rate of 0.2 $d^{-1}$ was applied. In contrast, at 8ºC, prediction with assumed $k_d$ value was much better than prediction of OUR from steady state values as shown in figure 4.21. The model predicted OUR well with small deviations. An assumed $k_d$ of 0.1 $d^{-1}$ predicted also better at 8ºC than the calculated from measurements which was 0.08 $d^{-1}$ (from 20ºC temperature correction).

Figure 4.23, figure 4.24 and fig 4.25 represent the simulated VSS along with growth and the unbiodegradable COD fractions at 20º, 8º and 4ºC respectively.

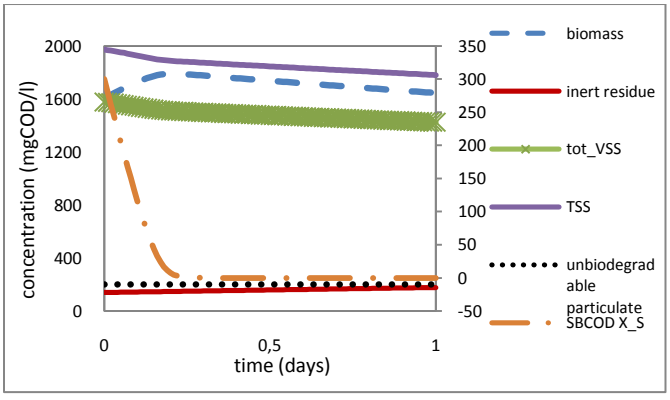


Figure 4.23 Simulation of MLVSS along with growth and the unbiodegradable COD fractions at 20ºC

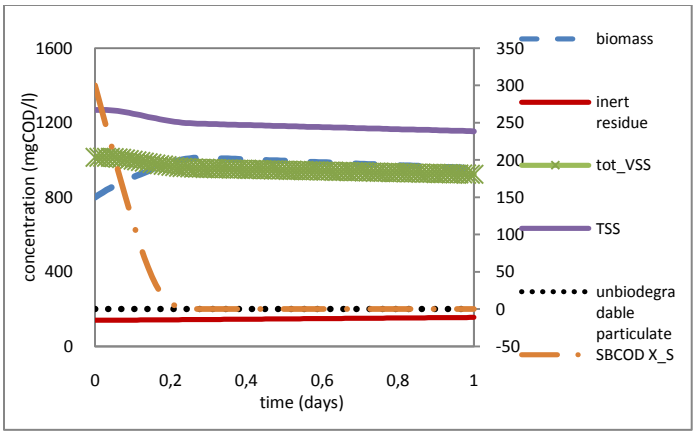


Figure 4.24 Simulation of VSS along with growth and the unbiodegradable COD fractions at 8ºC

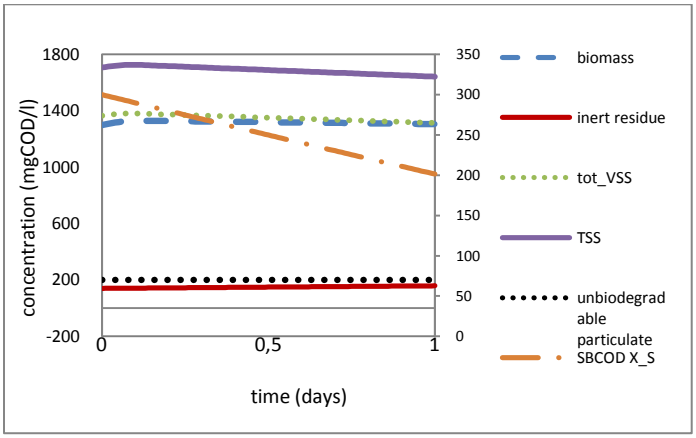


Figure 4.25 Simulation of VSS along with growth and the unbiodegradable COD fractions at 4ºC

Even if the measured MLVSS was not applied to the simulation, similarity can be inferred between the two approaches. If compared with the experimental observation (figure 4.7b, figure 4.10 and 4.11 for 20º, 4º and 8ºC respectively), the features are close.

Again here, 8ºC imitates the behavior at 20ºC with the simulation. The biomass growth is on exponential increase until substrate is used-up (figure 4.22 and figure 4.23). As a result, inert residue is building up. Afterwards, when some of the active biomass dies, a fraction will hydrolyze biodegradable COD and unbiodegradable residue will accumulate as product.

The simulations predicted that soluble substrate RBCOD was used up after 5.5h, 7h and after 24h while particulate biodegradable SBCOD substrate at about 6h, 7.3h and probably beyond 24h at 20º, 8º and 4ºC respectively. So, endogenous respiration dominates after 6, 7 at 20º, 8º and 4ºC.

As referred to the simulation, at 24h of the biological reaction at 4ºC, there are remaining particulate biodegradable COD which is of significant amount about 200 mgCOD/l. Correspondingly, the biomass

growth is still in the increasing phase (figure 4.25 and 4.28), meaning that at 4C, slowly biodegradable substrates are not yet depleted at 1 day of the test. This is probably why distinction of the endogenous phase was difficult in the results part (figure 4.9b and fig 4.10). Biomass growth would attain the maximum when the particulate biodegradable COD becomes very low.

This indicates that at cold temperature 4ºC, the break-down of substrates is very slow and takes longer time.

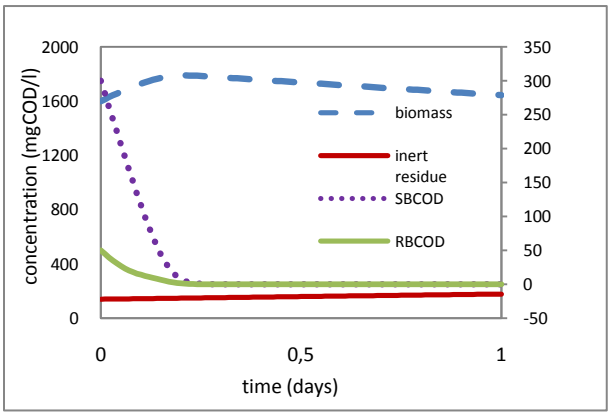


Figure 4.26 Simulation of soluble and particulate COD along with growth and inert fractions at 20ºC

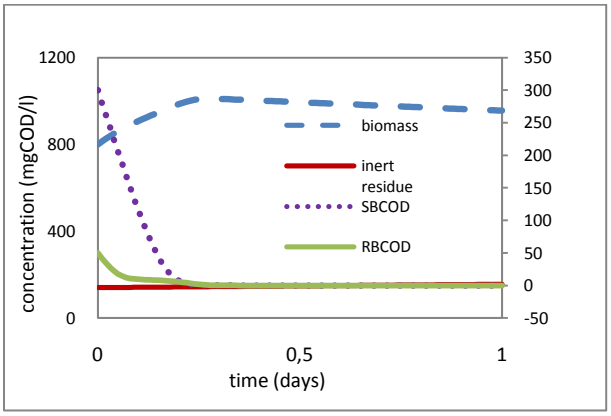


Figure 4.27 Simulation of soluble and particulate COD along with growth and inert fractions at 8ºC

At any temperature, soluble RBCOD is decreasing rapidly at initial period, the drop is steep until around 2h (0.09-0.1d) and then gradual decrease follows (more observed in figure 4.26 and 4.27). RBCOD is either present initially from the wastewater or from degradation of SBCOD. The first rapid decrease is certainly the result of the uptake of soluble RBCOD initially present in the influent wastewater. The second slower decrease corresponds to the combination of the remaining initial RBCOD and especially,

the soluble RBCOD resulting from the hydrolysis of particulate SBCOD. This second lower decrease differentiate the higher temperature to lower temperature depending on how fast or slow the SBCOD is degraded.

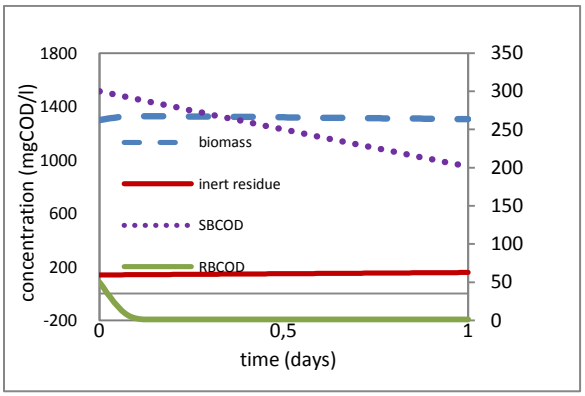


Figure 4.28 Simulation of soluble and particulate COD along with growth and inert fractions at 4ºC

From simulation, figure 4.23, figure 4.24 indicate that RBCOD (from influent and from hydrolysis) exhausted after ca.5.5h, around 6-7h for 20º and 8ºC respectively.

In a word, all of those parameters and information from dynamic simulation are valuable and useful for any alternative for the future design of the plant. The next part will then attempt to describe a simple design for the alternative for the actual treatment method of the plant.

# 5 DESIGN OF AS SYSTEM AND ALTERNATIVE TO SNJ TREATMENT PLANT

This chapter covers the design of an activated sludge system for the improvement of the actual treatment plant at SNJ/IVAR.

The study results will be considered in the following design of the proposed process.

As described in literature review part (chapter 2) an activated sludge comprises 2 main parts, an aeration tank and the settling tank (figure 3.3). The design of the main reactor, aeration basin will is (be most likely) the main aim of this part. Though, the possible option about what settling tank size to choose will be highlighted.

## 5.1 DESIGN CRITERIA OF THE BIOREACTOR

In the design of an activated sludge process some parameters should be taken into consideration such as retention time, biomass production, sludge production, oxygen requirement, and effluent characteristics.

The parameter SRT is of upmost importance for the design and the control of the process system because it will affect the other parameters. The design SRT must be determined and a safety factor (SF) should be included to account for changes in operating conditions (temperature, flowrate, etc...), or for instance any change in the influent loadings.

### 5.1.1 In situ and specific parameters

Some parameters characteristic of the plant such as the influent parameters and capacity deserve to be described or reminded because this will also constitute the basis of any calculation. As we described in part 2, ASPLAN VIAK (2010) has predicted for 2050 a design flowrate of 2.1$m^3$/s. In 2050, the organic loading is evaluated around 30000kg$BOD_5$/d, which is equivalent of 60000kg COD/d. In the calculation, the parameters applied are summarized as follows:

| | |
|---|---|
| COD_s (mg/l) | 30 |
| COD_p (mg/l) | 40 |
| COD_tot (mg/l) | 300-400 |
| Q ($m^3$/s) | 1.984 |
| MLSS (mg/l) | 3000-4000 |
| L (kgCOD/d) | 60000 |

As the calculation is based on 4ºC, $\mu_{max}$ and $k_d$ evaluated at 4ºC were used for the design. For 2050, the flowrate of 1.98$m^3$/s was evaluated. The total COD of 350mg/l total (average) was applied. As MLSS during the experiment was not as high as 4000mg/l, the average of 3500mg/l of typical MLSS for biological system was applied.

### 5.1.2 Design SRT

From chapter 1, SRT can be expressed by the relation between maximum growth rate and the decay rate as given in equation 1.15.

The selection of the design SRT is based on the evaluation of the lowest operating temperature of the experiment (4ºC). That is reasonable as well since low temperature will be the big challenge during winter and also to preview an yearly change of operating conditions. Thus, $\mu_{max}$ of 0.77 and $k_d$ of 0.07 from the experimental results at 4ºC will be accounted for SRT determination.

From the SRT expression, $SRT_{min}$ is obtained when growth rate is maximum ($\mu=\mu_{max}$)

$$SRT_{min} = \frac{1}{\mu_{max} - k_d} = \frac{1}{0.77 - 0.07} = 1.43 d^{-1}$$

This minimum required SRT which is very low because independent of C, the effluent COD. However the effluent quality is the basis of the evaluation of the design SRT as we have to meet low effluent.

The first step should be then defining the effluent target. The effluent requirement for Norway as we mentioned in the introduction part is 125mgCOD/l. This is high as far as quality of effluent. To have safety with the system performance, the total effluent-target COD can be set as 40 mg/l ($COD_{us}$ =30mg/l). The SRT related to this target will be then the minimum design SRT.

Figure 5.1 displays the effluent biodegradable COD and total effluent COD as function of sludge age. The expression used was given earlier in equation 1.16.

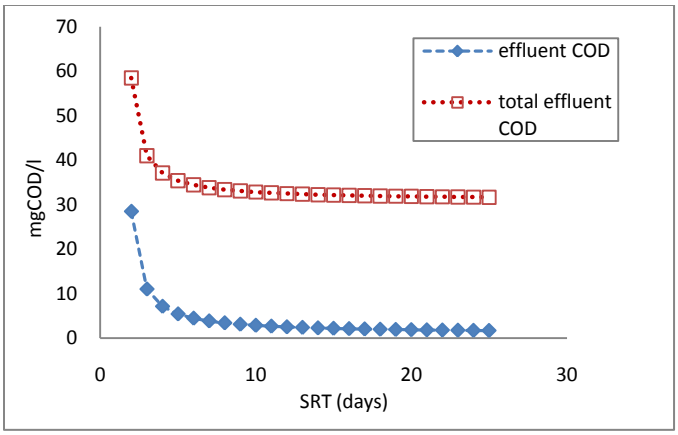


Figure 5.1 Effluent COD as function of retention time

Effluent biodegradable COD is decreased with increasing sludge age. From figure 5.1, when SRT is lower than around 3days, then the time is too short for the growth of heterotrophic organisms. So, this corresponds to minimum SRT to meet less than 40mg/l effluent COD.

Since it is obvious that having 0mg/l of TSS (100%separation) in the effluent is not possible, there should be expected COD in SS carried along with the effluent which we assume, for safety, the same as soluble COD (40mg/l). As a result, the overall total COD is expected to be between 70-80mg/l.

### 5.1.3 Biomass production

The biomass production and sludge fractions expressions were given in equations 1.19 and 1.27 and figure 5.2 presents the sludge fractions, biomass production over the sludge age.

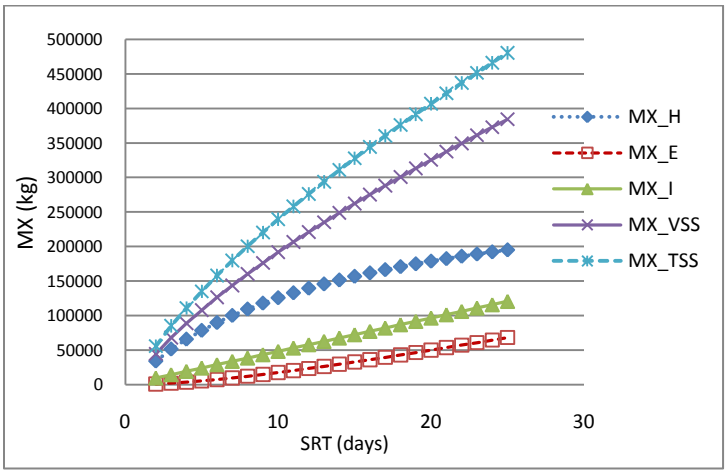


Figure 5.2 Sludge accumulation in the system as function of SRT

The fractions of sludge and biomass are increased at longer retention time.

The unbiodegradable fractions build up with the time. If inert solids are too high, there is a decrease in the capability of forming flocs, so wasting of sludge should be done periodically

### 5.1.4 Design volume of the reactor

The bioreactor volume can be determined using Eq.1.20. This equation shows that biomass production determines the required reactor volume.

The sludge production, total oxygen demand and volume are presented in figure 5.3

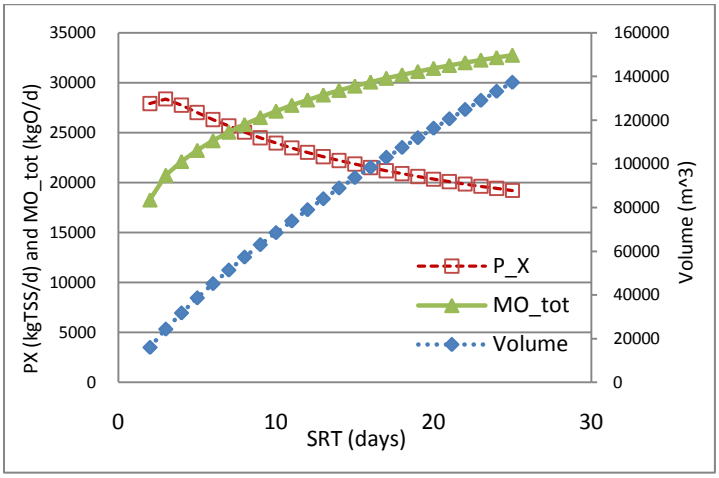


Figure 5.3 Sludge production, oxygen requirement and reactor volume as function of SRT

From figure 5.3, it indicates that the longer the SRT, the bigger the volume. For economical and space savings, the required aeration tank volume should be as minimum as possible corresponding to the lowest possible SRT. Therefore, 3-5 days is the minimum SRT for volume requirement.

Production of sludge is important to determine as it will affect also the capacity or design of the sludge disposal facilities. At the plant, the sludge will be reused for biogas production.

Sludge production (Px) decreases as retention time increases due to decay of biomass (figure 5.3). In the biogas plant, the more sludge implies the more energy generated by biogas. So, having the maximum sludge production possible would be beneficial. In figure 5.3, the maximum sludge, around 28000kgTSS/day is produced at about 3 days SRT.

### 5.1.5 Oxygen requirement

As microorganisms require oxygen for their growth and endogenous respiration, the plant will need the addition of the two oxygen requirements. Equation 2.7 and 2.11 describe the oxygen demand for growth

and for endogenous respiration respectively. The sum will give the total oxygen the plant needs to supply for an efficiency of the process.

In contrast to sludge production, oxygen demand increases with the sludge age due to increasing decay and oxidation of dead biomass (endogenous respiration) which require more oxygen. For the plant, to have the lowest operating cost possible with regard to oxygen requirement, as low SRT as possible is required.

A SRT of minimum 3days will satisfy this requirement for low oxygen demand (figure 5.3).

### 5.1.6 OUR

The OUR can be computed by dividing the oxygen demand by the aerator volume size. OUR at SRT of 6 days is predicted to be around 22.34mg/l/h which is much higher than the measurement during experiment (10-16mg/l/h). This is because the design MLSS concentration is much higher; 3500 mg/l vs 2000 mg/l.

A typical MLSS of 1500-3000mg/l and 2500-4000mg/l was reported by Metcalf and Eddy (1991) for conventional and complete-mix activated-sludge respectively. When the MLSS concentration is lowered to ca. 2000-2500mg/l, the OUR becomes close to that of the experiment (≈12-16mg/l/h). However, the volume size has increased from 45120 to 63170-78960$m^3$ which is almost an increase of 39-75% of the oxygen demand. Similarly, an MLSS of 4000mg/l would require lower reactor volume but higher OUR rate (25mg/l/h).

So, care should be taken when selecting the parameters design, because this may bring about higher operational cost.

### 5.1.7 Influence of loadings

Figure 4.5 demonstrates the influence of loadings or flowrate onto sludge production, biomass production and the reactor size.

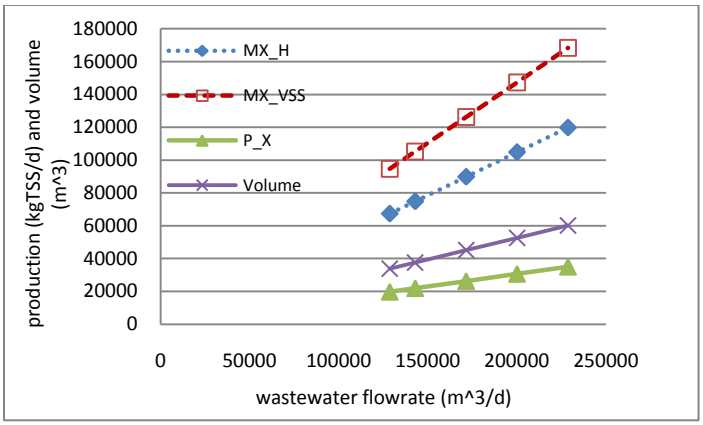


Figure 5.4 Biomass, sludge production and reactor volume as function of flow rate

As it is shown, MX_H, MX_VSS, sludge production and the volume of reactor increases with increasing flowrate. And the volume is also inversely proportional to the biomass concentration. To have a minimum volume possible is then to consider the biomass concentration, especially in activated sludge system and also to consider the flow rate.

### 5.1.8 Estimated design parameters

To recapitulate, the minimum SRT for sufficient effluent quality is 3days (sect.5.1.2). A SRT of 3-5days is the volume requirement (sect.5.1.4) and a minimum sludge age of 3 days for both sludge production and oxygen demand (5.1.5 and 5.1.6) are also required. Thus, to satisfy each requirement, the limit of 3 days would be favorable for both a better effluent quality (around limit of 40mg/l) and cost-effective reactor size and oxygen consumption.

The evaluated wastewater flowrate predicted for 2050 is around 171 428$m^3$/d or 1.98$m^3$/s which is close to what ASPLAN VIAK (2010) predicted (2.1$m^3$/s).

A safety factor is applied for more safe design especially when seasonal variation of conditions or loadings may happen during the year.

- SF = 2 is assumed ==> design SRT = 3*2 = 6 days.

Table present the parameters evaluated at SRT = 6 days

Table 5.1 Design parameters determined at 6 days sludge age

| SRT (d) | tot COD_e(mg/l) | Volume (m3) | P_X (kgTSS/d) | MO_tot (kgO/d) | Q (m3/d) |
|---|---|---|---|---|---|
| 6 | 34.43 | 45121.43 | 26320.83 | 24195.96 | 171428.6 |

The design volume for 2050 is predicted to be around 45120m$^3$ with a sludge production of approximately 26320 kg TSS/day and the oxygen demand of at about 24195kO/d.

The hydraulic retention time for the aeration tank can be determined using the information in table 5.1 and eq.2.13. $HRT = \frac{45121.43\ m^3 \times \frac{24h}{d}}{\frac{171428.6m^3}{d}} = 6.3h$

The design estimated for 2050 implies that an increase in organic loadings is predicted as actually the load is around 1.5m$^3$/s. Expansion of the area of treatment is also previewed as the volume of treatment will be increased.

If we assume the area occupied by the actual sedimentation basin will be the area the aeration tank will be installed. We can compute the ratio of increase of the area.

Actually, from the information given in chapter 2, the volume of the sedimentation basin can be estimated:

$V$ = 2 x 4 x 473.2 m$^2$x 4.8m = 18170.88 m$^3$

So, the total volume of the actual series of basin is not enough for the space of the design aeration tank. We need 2.5 times higher than actual volume.

## 5.2 DESIGN OF SETTLING TANK

Designing the settling tank is also a concern but no experimental data relating to settling characteristic is available for the wastewater. The settling capability of a treatment dictates its effluent performance. Clarifier design must provide adequate clarification of the effluent and solids thickening for the activated sludge solids (Metcalf and Eddy, 2003).

The following equation can help understand the relation between designing the aeration tank and the settling tank.
The area of the settling tank is given by the expression

$$A = \frac{Q}{v_s} \qquad \text{(eq. 5.1)}$$

And the volume of the settling tank ($V_s$) is: $V_s = \mathrm{A} \,.\, \mathrm{d}$ (eq. 5.2)

Where $V_s$ is the volume of the settling tank

A is the settling tank area (m$^2$)

Q is the flowrate (m$^3$/h)

$v_s$ is the settling velocity of solids or hydraulic rate of the settling tank (m/h)

d is the depth of the settling tank. Here, it is assumed to be 5m.

Velocity and surface are inversely proportional (eq.5.1). If settling velocity is decreased, the resulting surface area should be increased.

Metcalf and Eddy (2003) reported the typical value for the design of settling activated-sludge is an average of 400-800gal/ft$^2$.d. With the highest value, let us compute the relating typical settling velocity.

$v_s$ = 800gal/ ft$^2$.d * 0.0407 = 32.56m$^3$/m$^2$.d = 1.36 m/h.

If this value is to consider, knowing the organic loading previewed for 2050 (171428m$^3$/d), the corresponding design settling tank area is computed.

$$A = \frac{171428.6\ m^3/d}{32.56 m^3/m^2.d} = 5265 m^2$$

The design settling tank volume can also be evaluated

$$V_s = 5265 m^2 . 5m = 26325 m^3$$

Figure 5.5 shows the hydraulic rate of clarifier or settling velocity as function of MLSS concentration.

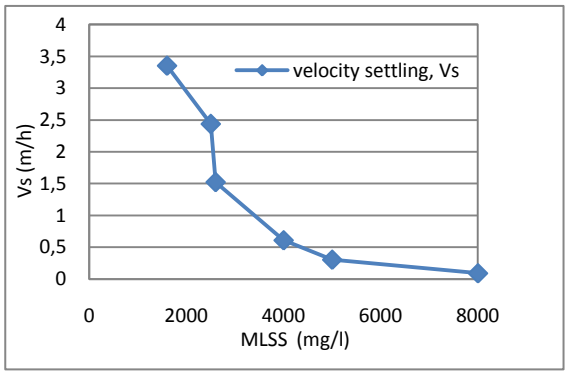


Figure 5.5 Settling velocity of solids as function of MLSS (adapted from example in Metcalf and Eddy, 1991)

When MLSS increases, the settling velocity decreases. This means smaller bioreactor and bigger settling tank.

This must be taken into consideration when relating SRT to MLSS

## 5.3 RECOMMENDATION

Two possible alternatives can be proposed for the alternative of the present treatment (fig 4.6):

1. Introduction of a complete biological activated sludge system for removal of $BOD_5$ and SS, without chemical treatment (fig 5.7).
2. Introduction of a combined process of complete biological activated sludge system for removal of $BOD_5$ and SS with chemical addition (ferric chloride)

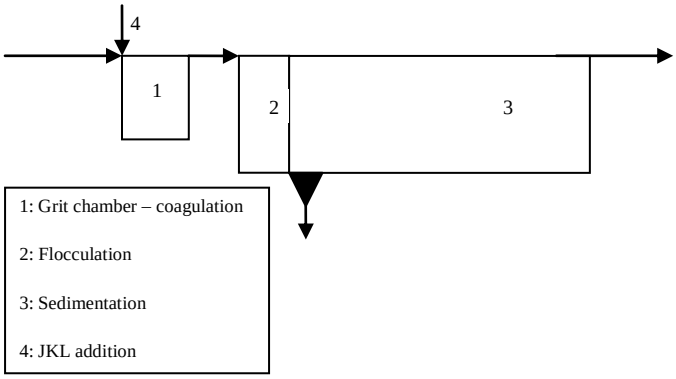


Figure 5.6 Present treatment at the plant

Both of the alternatives require area/volume expansion. This is because of the introduction of bioreactor and the settling tank. For 2050, as computed and shown in table 5.1, a reactor volume of about 45100$m^3$ is required with organic loading of 0.06kg $BOD_5$/pe.d (ASPLAN VIAK, 2010). The choice of the process depends on many factors including the compatibility with the existing treatment method, compatibility of the existing equipment and the existing space, the installation cost of any new equipment and the operating cost and other factors depending on the plant needs.

The first alternative is proposed if nutrient removal such as phosphorus is not included in treatment objective. It is suitable for low or medium wastewater.

The second alternative has additional treatment as chemical precipitation with addition of ferric chloride prior to the settling tank (fig 5.8) to optimize sludge-liquid separation thus to improve the SS removal. It can be suitable for high loading and can save area in the way that it eliminates the need of bigger clarifier if the activated sludge process alone was used.

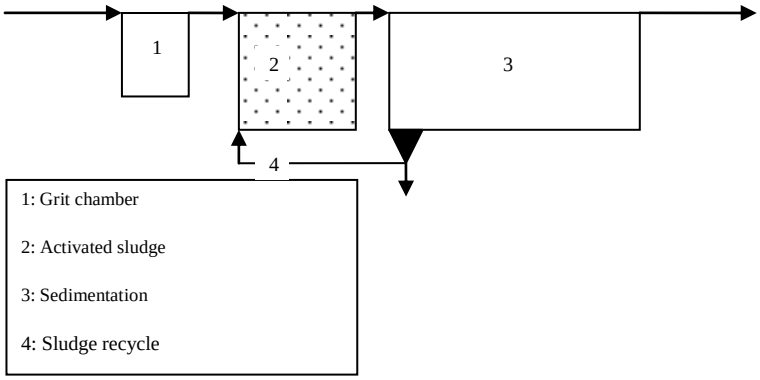


Figure 5.7 Activated sludge system without chemical treatment

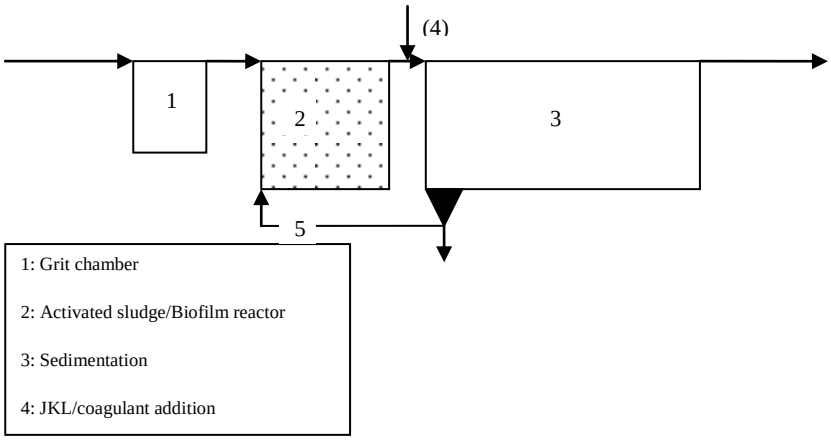


Figure 5.8 Activated sludge system combined with chemical treatment

# 6 CONCLUSION

In this thesis work, the performance of the biological wastewater treatment has been investigated. Theoretical and experimental approaches have been used. Measured and estimated values from steady state calculations in addition to estimated parameters from dynamic simulations served as basis for the design of the proposed system alternative to the actual treatment at the plant.

From this thesis work, we can conclude that:

- Temperature affects significantly the activity rate of microorganisms, the process rates and thus the system efficiency, especially at 20ºC and 4ºC. A decay rate of 0.11 $d^{-1}$ and 0.07$d^{-1}$ were determined at 20º and 4º respectively and a maximum growth rate of 2.53 $d^{-1}$ versus 0.77 $d^{-1}$. At 8ºC, overestimation of the calculated parameters and underestimation of the biomass concentration from steady state led to unreliability of the data.

- The treatment efficiency during the test was high at 20ºC and rather acceptable at lower temperature (4ºC). At 20ºC, the effluent quality met satisfactorily the discharge permit requirement with estimated $BOD_5$ removal of 95% and COD removal of 79-86% while at 4ºC, the results of 72-75% BOD removal and 71% COD removal were just above the standard limit.

- OUR behavior and other estimated parameters at 8ºC were expected to be closer to 4ºC. However, the lower temperature, 8ºC tended to imitate 20ºC with regard to not only OUR but also almost all experimental results and simulations. The reason for all this deviation is probably erroneous experimental procedures at 8ºC.

- Simulations based on steady state calculation predicted higher OUR than the experimental results. And estimation of the parameters $\mu_{max}$ and $k_d$ based on the experimental observations predicted accordingly.

- Design of activated sludge was based on the results and on the prediction of the year 2050 by ASPLAN VIAK.

- From the experimental study, activated sludge process is concluded as a possible alternative for the treatment of SNJ wastewater. Introduction of an activated sludge system at the plant seems to be a very good solution with or without the use of chemical addition, depending on the means and the needs of the plant.

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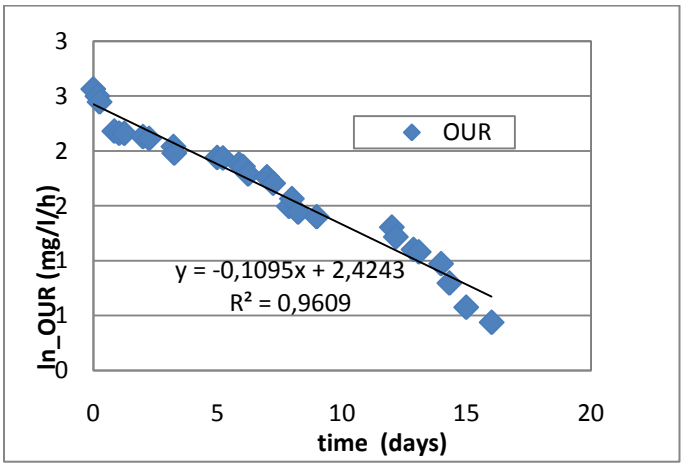
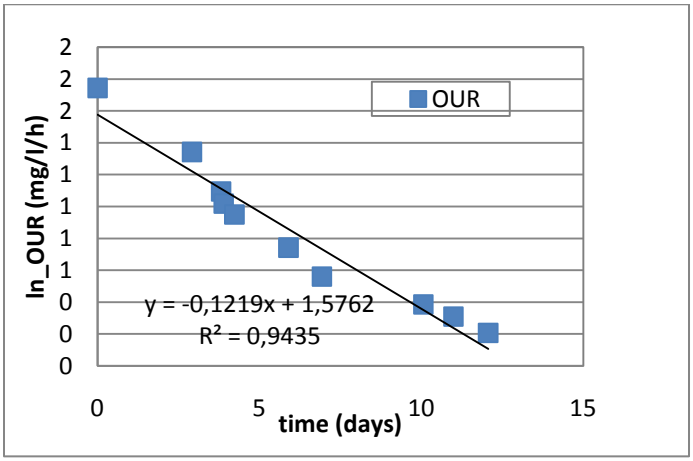
# APPENDIX I: Determination of decay rate $k_d$

- **At 20ºC**




- At 8ºC



## APPENDIX II : Steady state calculation of μ_max and X_H

| | 20ºC | 8ºC | 4ºC |
|---|---|---|---|
| k_d (/day) | 0.11 | 0.08 | 0.07 |
| OUR_mean (mgO/l/h) | 41.11 | 30.40 | 14.57 |
| Y (gVSS/gCOD) | 0.45 | 0.45 | 0.45 |
| Q (L/day) | 2.50 | 1.25 | 1.25 |
| V (L) | 4.00 | 1.50 | 1.50 |
| ΔCOD (mg/l) | 300.00 | 300.00 | 300.00 |
| SRT (d) | 20.73 | 5.00 | 9.02 |
| Y (gCOD/gCOD) | 0.66 | 0.66 | 0.66 |
| **X (mg/lVSS)** | **533.21** | **405.91** | **622.01** |
| **X (mg/lCOD)** | **757.16** | **576.40** | **883.26** |
| **μ_max (/d)** | **2.53** | **2.46** | **0.77** |

| | 20ºC | 8ºC (kd=0.08) | 8ºC (kd=0.12) | 4ºC |
|---|---|---|---|---|
| X (mg/lVSS) | 533.21 | 405.91 | 351.56 | 622.01 |
| XE (mg/lVSS) | 243.18 | 32.47 | 42.19 | 77.68 |
| XI (mg/lVSS) | 388.69 | 125.00 | 125.00 | 225.50 |
| X_H (mg/lVSS) | 1165.08 | 563.39 | 518.75 | 925.19 |
| **X_H (mg/lCOD)** | **1654.41** | **800.01** | **736.63** | **1313.77** |

## APPENDIX III : Design of activated sludge

| SRT | C_e | MX_H | MX_E | MX_I | MX_VSS | MX_TSS | V | P_X | MO_g | MO_e | MO_tot |
|---|---|---|---|---|---|---|---|---|---|---|---|
| (d) | mgCOD/l | kgVSS | kgVSS | kgVSS | kgVSS | kgTSS | m^3 | kgTSS/d | kgO/d | kgO/d | kgO/d |
| 2 | 28.500 | 34037.59 | 953.05 | 9657.95 | 44648.59 | 55810.74 | 15945.93 | 27905.37 | 15564.26 | 2706.67 | 18270.93 |
| 3 | 11.000 | 51449.82 | 2160.89 | 14486.92 | 68097.64 | 85122.05 | 24320.58 | 28374.02 | 16647.26 | 4091.29 | 20738.55 |
| 4 | 7.111 | 65785.71 | 3684.00 | 19315.90 | 88785.61 | 110982.01 | 31709.15 | 27745.50 | 16887.92 | 5231.28 | 22119.20 |
| 5 | 5.400 | 78457.14 | 5492.00 | 24144.87 | 108094.01 | 135117.52 | 38605.00 | 27023.50 | 16993.82 | 6238.91 | 23232.73 |
| 6 | 4.438 | 89821.18 | 7544.98 | 28973.84 | 126340.00 | 157925.00 | 45121.43 | 26320.83 | 17053.38 | 7142.58 | 24195.96 |
| 7 | 3.821 | 100091.89 | 9809.01 | 33802.82 | 143703.72 | 179629.65 | 51322.76 | 25661.38 | 17091.56 | 7959.31 | 25050.87 |
| 8 | 3.391 | 109427.62 | 12255.89 | 38631.79 | 160315.30 | 200394.12 | 57255.46 | 25049.27 | 17118.13 | 8701.68 | 25819.81 |
| 9 | 3.075 | 117953.83 | 14862.18 | 43460.76 | 176276.78 | 220345.97 | 62955.99 | 24482.89 | 17137.67 | 9379.69 | 26517.36 |
| 10 | 2.833 | 125773.11 | 17608.24 | 48289.74 | 191671.08 | 239588.85 | 68453.96 | 23958.89 | 17152.66 | 10001.48 | 27154.13 |
| 11 | 2.642 | 132970.76 | 20477.50 | 53118.71 | 206566.97 | 258208.72 | 73773.92 | 23473.52 | 17164.51 | 10573.84 | 27738.35 |
| 12 | 2.486 | 139618.60 | 23455.92 | 57947.69 | 221022.21 | 276277.76 | 78936.50 | 23023.15 | 17174.12 | 11102.47 | 28276.59 |
| 13 | 2.358 | 145777.61 | 26531.53 | 62776.66 | 235085.80 | 293857.25 | 83959.21 | 22604.40 | 17182.07 | 11592.24 | 28774.31 |
| 14 | 2.250 | 151500.00 | 29694.00 | 67605.63 | 248799.63 | 310999.54 | 88857.01 | 22214.25 | 17188.76 | 12047.28 | 29236.04 |
| 15 | 2.158 | 156830.74 | 32934.45 | 72434.61 | 262199.80 | 327749.75 | 93642.78 | 21849.98 | 17194.46 | 12471.18 | 29665.64 |
| 16 | 2.078 | 161808.78 | 36245.17 | 77263.58 | 275317.53 | 344146.92 | 98327.69 | 21509.18 | 17199.37 | 12867.03 | 30066.41 |
| 17 | 2.009 | 166468.09 | 39619.40 | 82092.56 | 288180.05 | 360225.06 | 102921.45 | 21189.71 | 17203.66 | 13237.54 | 30441.20 |
| 18 | 1.948 | 170838.35 | 43051.27 | 86921.53 | 300811.15 | 376013.94 | 107432.55 | 20889.66 | 17207.43 | 13585.07 | 30792.50 |
| 19 | 1.894 | 174945.70 | 46535.56 | 91750.50 | 313231.76 | 391539.70 | 111868.49 | 20607.35 | 17210.77 | 13911.68 | 31122.45 |
| 20 | 1.846 | 178813.19 | 50067.69 | 96579.48 | 325460.36 | 406825.44 | 116235.84 | 20341.27 | 17213.75 | 14219.22 | 31432.97 |
| 21 | 1.803 | 182461.24 | 53643.61 | 101408.45 | 337513.30 | 421891.62 | 120540.46 | 20090.08 | 17216.43 | 14509.32 | 31725.74 |
| 22 | 1.764 | 185908.04 | 57259.68 | 106237.42 | 349405.14 | 436756.43 | 124787.55 | 19852.57 | 17218.84 | 14783.41 | 32002.25 |
| 23 | 1.728 | 189169.80 | 60912.68 | 111066.40 | 361148.88 | 451436.10 | 128981.74 | 19627.66 | 17221.03 | 15042.78 | 32263.81 |
| 24 | 1.696 | 192261.05 | 64599.71 | 115895.37 | 372756.13 | 465945.16 | 133127.19 | 19414.38 | 17223.03 | 15288.60 | 32511.63 |
| 25 | 1.667 | 195194.81 | 68318.18 | 120724.35 | 384237.33 | 480296.67 | 137227.62 | 19211.87 | 17224.86 | 15521.89 | 32746.75 |