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Mass balance calculations of IVARs wastewater treatment plant

Enhanced biological phosphorus removal
based on an activated sludge system

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Preface

This master thesis is submitted to the Norwegian University of Science and Technology (NTNU) as a part of the master program of Civil and environmental engineering within the field of water supply and wastewater technology. This master thesis has been a collaboration between NTNU and IVAR, with professor Stein Wold Østerhus as the main supervisor. First I want to thank Stein Wold Østerhus and Blanca Magdalena Gonzalez Silvia for all their help and guidance, through many hours of discussion. This master thesis would not be a reality without their guidance and support. I also want to thank Leif Ystebø for sending real data to work with, and all the time he have spent answering all my questions and long e-mails.

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Abstract

This master thesis is written as a collaboration between NTNU and IVAR. IVAR's largest wastewater treatment plant has been the main focus, in Norwegian referred to as "Sentralrenseanlegg Nord-Jæren" (SNJ). The aim of this master was to make a theoretical mass balance of the existing biological treatment plant. IVAR SNJ recently changed from chemical precipitation to enhanced biological phosphorus removal in an activated sludge system. The reconstruction was fully implemented at IVAR SNJ in March 2018. A theoretical mass balance model could be a useful analyzing tool, that can emphasize possible limitations in the system to help reach a more optimized operation of the plant.

This thesis has been based on desktop work, where finding relevant literature to construct equations needed in the mass balance model have been a large part of the total workload. Each process has been examined, to find what is removed where, when and how. The existing system is complexed, because several of the process are dependent on the recycle of flows, hence it has also been very time consuming to make the excel-model functional. The last period have been used to test different scenarios in the model, with the purpose of imitating a normal operating situation at IVAR SNJ.

The wastewater entering IVAR SNJ is diluted, especially with low phosphorus and COD concentrations. Based on the results collected from the model phosphorus seems to be the main limitation for microbial growth at IVAR SNJ. The current biomass production at IVAR SNJ is assumed to be very low if existing, due to identical influent and effluent soluble phosphorus concentrations. From the model, approximately 70 % of total phosphorus entering the treatment plant is discharged into the sea, where the largest fraction follows the reject flow from the centrifugal dewatering process.

With the existing sludge treatment line phosphorus is not removed or utilized as a resource. To optimize the existing wastewater treatment plant the sludge treatment line need to change within a short period of time. The recommendation would be to implement an anaerobic mixer prior the anaerobic digestion (AD), to provoke release of luxury-P. By implementing this change the risk for unwanted struvite precipitation in the sludge treatment line would be drastically reduced, and phosphorus would be recovered and utilized as resource in IVARs fertilizer product, Minorga.

Sammendrag

Denne masteroppgaven er skrevet som et samarbeid mellom NTNU og IVAR. IVARs sentralrenseanlegg Nord-Jæren (SNJ) har vært hovedfokuset gjennom hele oppgaven, som er det største avløpsrenseanlegget i IVAR-regionen, som består av renseanlegg i 13 kommuner i Rogaland fylke. Formålet med denne masteren var å lage en teoretisk massebalanse over alle renseprosessene til IVAR SNJ. Sentralrenseanlegget har nylig blitt ombygget fra et avløpsanlegg basert på kjemisk felling til biologisk fosforfjerning i et aktivslamanlegg. Det biologiske renseanlegget som massebalansen baserer seg på ble implementert hos IVAR SNJ i mars 2018. En massebalansemodell kan brukes som et verktøy for å analysere det nye renseanlegget, slik at driften kan optimaliseres.

Prosessen har gått ut på finne nødvendig litteratur til alle renseprosessene, for å finne ut hva som fjernes hvor og på hvilken måte. Dette har vært en teoretisk oppgave som har basert seg på faglitteratur, og data fra IVAR. Mesteparten av tiden har blitt brukt på å finne og konstruere relevante formler som var nødvendig i massebalansen. I tillegg har mye tid blitt brukt til å få excel-modellen funksjonell, fordi anlegget består av mange renseprosesser som er avhengige av hverandre. Den siste tiden av masteren har gått med til å jobbe med den ferdige modellen ved å simulere ulike scenarioer, med formål om å etterligne en normal driftssituasjon hos IVAR SNJ.

Avløpsvannet som ankommer IVAR SNJ er utvannet, med spesielt lave fosfor og COD konsentrasjoner. Resultatene i modellen tilsier at fosfor er den begrensende faktor for mikrobiell vekst i dagens anlegg. Biomasseproduksjonen hos IVAR er antatt å være lav, da målt fosfor konsentrasjonen i utløpet er identisk lik innløpskonsentrasjonen. Fra modellen vises det at omtrent 70 % av all fosforen som ankommer anlegget blir sluppet direkte ut i Håsteinfjorden, hvor den største andelen følger rejektstrømmen fra sentrifugen.

Fosfor må gjenvinnes i slamlinjen for at renseanlegget skal utøve sin optimale funksjon. For å optimalisere dagens anlegg må slamlinjen hos IVAR SNJ endres innen kort tid. Det anbefales å tilføre en anaerob tank hvor luksus-P kan slippes ut før råtnetanken. Ved å implementere denne endringen vil man i større grad unngå uønsket struvittutfelling i anleggskomponenter, samtidig blir fosfor gjenvunnet og kan brukes som en ressurs i IVARs gjødselprodukt Minorga.

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Content

List of abbreviations

AD	Anaerobic digestion
aer	Aeration tank
AN1	First anaerobic reactor
AN2	Second anaerobic reactor
AN3	Third anaerobic reactor
bCOD	Biodegradable COD
BOD	Biochemical oxygen demand
bsCOD	Biodegradable soluble COD
CD	Centrifugal dewatering
COD	Chemical oxygen demand
CSS	Combined sewage system
DF	Drum filtration
DN	Denitrification
DO	Dissolved oxygen
fer	Fermentation
GAO	Glycogen-accumulating organisms
MLVSS	Mixed liquid volatile suspended solids
NIT	Nitrifiers
OHO	Ordinary heterotroph organisms
PAO	Polyphosphate accumulating organisms
pnbCOD	Particulate non-biodegradable COD
psbCOD	Particulate slowly biodegradable COD
HRT	Hydraulic retention time
TD	Thermal drying
TP	Treatment plant
TH	Thickener
tpCOD	Total particulate COD
tsCOD	Total soluble COD
TSS	Total suspended solids
R	Return sludge line
rbCOD	Readily biodegradable COD
SC	Screw compressor

sed	Sedimentation basin
SS	Screening station
SG	Sand and grease trap
snbCOD	Soluble non-biodegradable COD
ssbCOD	Soluble slowly biodegradable COD
SRT	Solid retention time
VFA	(Short-chain) Volatile fatty acid
VSS	Volatile suspended solids
WAS	Waste activated sludge
W	Waste sludge line

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1. Introduction

1.1 Background about IVAR

IVAR is a company owned by 13 municipalities in Rogaland, Norway. IVARs main tasks are within the fields of water, wastewater and renovation. This report focuses on IVARs largest wastewater treatment plant placed in Mekjarvik (SNJ), which treats wastewater from the municipalities: Sandnes, Stavanger, Gjesdal, Sola and Randaberg. IVARs wastewater treatment plant (SNJ) has recently changed from chemical precipitation to enhanced biological phosphorus removal based on an activated sludge system. This reconstruction will change the sludge characteristics significantly. The main reason for this drastically change is to achieve discharge requirements regarding organic material, because removal of soluble organic material is better achieved with biological treatment. The treated wastewater is released 1,6 kilometers from land, at a depth of 80 meters in a fjord called Håsteinfjorden. There is no requirement of removal of phosphors when the emission point is the sea, but IVAR want to recover phosphorus and utilize it as a resource in their fertilizer product, Minorga.

The reconstruction of IVAR SNJ happens in two steps, this thesis will focus on the first step. The first step entails the implementation of three process lines. Figure 1 shows the flow sheet of the existing treatment plant with only one process line, but there is actually three process lines in parallel. The process line includes the whole activated sludge system; AN1, AN2, AN3, Aeration tank and sedimentation basins. This system can handle expected load until 2035. The last step in the reconstruction is the implementation of a fourth process line, due to expectation of higher future loads in 2050. When the second step is implemented the treatment plant is designed for 500 000 PE. The chemical treatment plant was designed for 240 000 PE in comparison, hence it is a massive change. All processes prior the process lines are the same for step 1 and step 2, the only difference is the number of process lines, due to higher future load. The change from chemical precipitation to biological treatment happened gradually. The treatment plant was operating with chemical precipitation until June 2017. From June 2017 to August 2017 the treatment plant was operating with both chemical precipitation and the activated sludge system, 50/50. From August 2017 to March 2018 the treatment plant was operating with biological treatment only, with two process lines. From mid-March 2018 all three process lines was implemented.

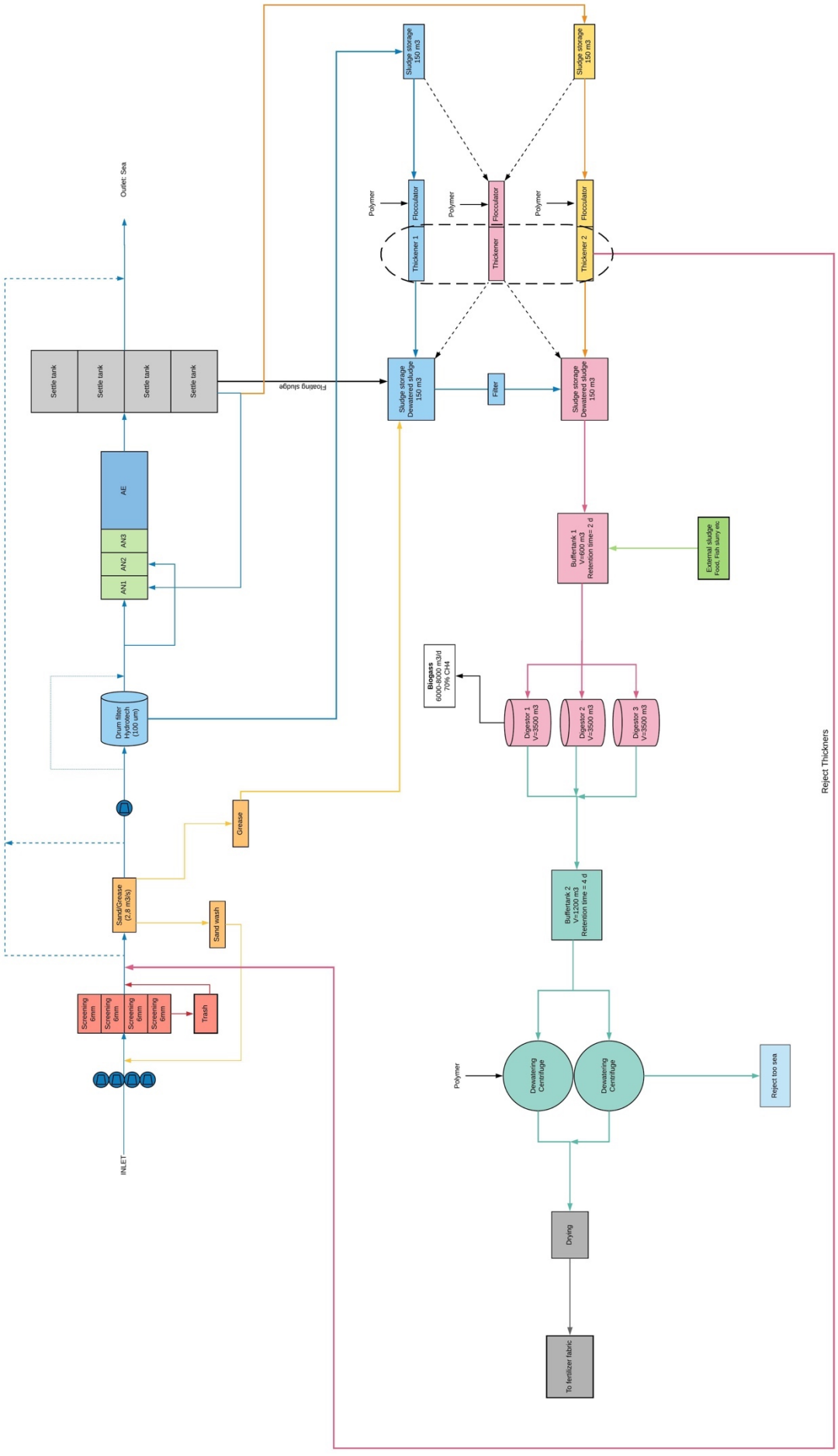

Figure 1:Flow sheet of IVAR SNJ

Since the actual startup was quite recent, there is a need to analyze the biological treatment plant more in detail, to be able to optimize the operation of the new system. A mass balance model of the existing treatment plant can be a useful tool, which can be used as a basis for decision making to secure optimal operation of the plant. In general biological sludge is less biodegradable than chemical sludge, because the biological sludge has been degraded during the activated sludge process by microorganisms (Carrere *et al.*, 2016). Chemical sludge will degrade similar to primary sludge, but in a slower rate (Tchobanoglous *et al.*, 2014). Waste activated sludge (WAS) consist of flocs of microbial biomass exopolymeric substances (EPS), which mainly are proteins and carbohydrates, and compounds that are not biodegradable (Carrere *et al.*, 2016).

1.2 Mass balance

A mass material balance analysis is a good way to analyze what is happens within a reactor as a function of time (Tchobanoglous *et al.*, 2014). Mass material balance is based on the principle that mass neither occurs nor disappears, but can change form. Mass can change from e.q solid form, to liquid or gas depending on the given process. For each process analyzed, equation (1-1) has been used as basis. Depending on the process and flow, one or several of the terms in equation (1-1) can be equal to zero.

$$\text{Accumulation} = \text{Inflow} - \text{Outflow} \pm \text{Generation} \qquad (1\text{-}1)$$

The main objective is to make a theoretical mass material balance of IVARs existing treatment plant. IVAR SNJ consist of physical and biological units processes that need to be included in the mass balance model.

1.3 Objectives

1. Make a theoretical mass balance model of IVARs existing wastewater treatment plant.
2. Test different scenarios in the mass balance model, and compare results.
3. Suggest recommended changes in the sludge treatment line.

2. Theory

The theory will look into wastewater characteristics for Norwegian wastewater and wastewater in general. Then each method at the treatment plant will be explained in detail, both the wastewater line and the sludge line. How they work, the main purpose, and challenges that may occur will be discussed.

2.1 Wastewater characteristics

IVAR SNJ receives and treats wastewater from six municipalities in Rogaland. The common sources of wastewater are domestic wastewater, industrial wastewater, infiltration/inflow (I/I) and stormwater (Tchobanoglous *et al.*, 2014). *Domestic wastewater* includes wastewater from residential areas, commercial, institutions and public facilities. Domestic wastewater are also called sanitary wastewater, and originates from water use in a household, school, hospital and workplaces such as cooking, dishes, laundry, showers and toilet (Ødegaard *et al.*, 2014). *Industrial wastewater* are wastewater produced by industrial processes, and do not include wastewater from cafeteria, offices and toilets. The composition of industrial wastewater varies a lot based on the given industry. Wastewater from industries can contain high concentration of toxic substances such as heavy metal, then the wastewater needs to be treated before released to the collection system. While other industries can produce wastewater with almost the same composition as sanitary wastewater only with higher concentration (Ødegaard *et al.*, 2014). *Infiltration/Inflow (I/I)* originates from unknown water sources and enters the collection system through holes, joints, manholes, overflow etc. Infiltrated water is mainly clean water, and will therefore dilute existing wastewater in the system. *Stormwater* derives from runoff from rainfalls and snowmelt (Tchobanoglous *et al.*, 2014). Stormwater will dilute the wastewater and are the main source for entry of sediments into the system.

2.1.1 The composition of wastewater

The composition of the inlet wastewater is important knowledge, because one should know what kind of pollutants that needs to be removed, and in what extent before selecting treatment processes. Wastewater consists of a variety of inorganic and organic settleable particles, suspended solids and dissolved substances.

2.1.1.1 Solids

Roughly one can distinguish between three different types of solids (1) ***Colloidal material*** with the particular size of 0,01-1 μm, (2) ***Suspended solid*** with a particular size bigger than 1 μm or 0,45 1 μm, and (3) ***Settleable particles*** with a particular size bigger than100 μm (Ødegaard *et al.*, 2014). There are many different measurable parameters used to describe the different solids in wastewater, here the most relevant will be explained.

Total solids (TS) include the weight of particles in suspension (TSS) and dissolved particles (TDS). TS is found by measuring the weight of the residual after a wastewater sample has been evaporated at a temperature around 103-105°C (Tchobanoglous *et al.*, 2014). *Total dissolved solids (TDS)* are the fraction that passed through a filter with opening of 1,2 μm. This definition also includes colloidal material, which actually are particles and not dissolved solids. The distinction between dissolved substances and particulate material is, in other words, slightly unclear. *Total suspended solid (TSS)* is particles in suspension per unit volume. These particles are in suspension due to low sedimentation rate. The concentration of TSS is found by measuring the weight of remaining substances on a filter with an opening of 1 μm or 0,45 1 μm (Ødegaard *et al.*, 2014, p. 478). TS, TSS and TDS contains both organic and inorganic substances. Inorganic substances in wastewater can be nutrients, gasses and metals, while organic substances normally consists of carbon, hydrogen and oxygen (Tchobanoglous *et al.*, 2014). *Volatile solids (VS)* represents the organic material, and is a measure of the solids removed when TS is heated to 500 °C. While *fixed solids (FS)* represents the inorganic material of TS, and is the amount of solids left after the incineration. Figure 2 gives an overview of the different fraction of TS.

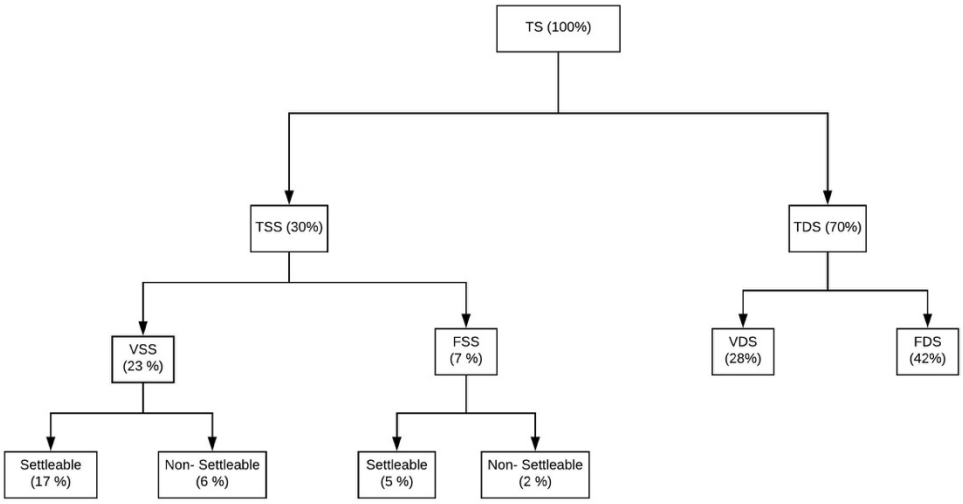

Figure 2: The solid distribution in wastewater based on total solids (Rossle and Pretorius, 2001)

2.1.1.2 Organic material

There are many different organic substances in wastewater. Generally organic materials in wastewater, consist of oxygen-consuming substances and organic micro-pollutants. Where oxygen-consuming substances are the main group, which consist of carbohydrates, proteins, fatty acids and detergents. Examples on micro pollutants are personal care products and pharmaceutical residues (Ødegaard *et al.*, 2014). Urea is also an important organic substance that mainly exist in raw wastewater. Urea is a neutral and non-toxic compound that originated from urine. In contact with water urea rapidly hydrolysis to ammonium (NH_4), which is an inorganic substance (Ødegaard *et al.*, 2014). The amount of soluble- and particulate organic material should be known, to understand what is removed where in the treatment plant. Normally one distinguish between easily biodegradable, slowly biodegradable and non-biodegradable organic material, based on the rate microorganisms degrade organic substances (Ødegaard *et al.*, 2014). There are two parameters frequently used to measure the content of organic material in wastewater: Biological oxygen demand (BOD) and chemical oxygen demand (COD).

Biochemical oxygen demand (BOD)

BOD_5 are a widely used parameter within the field of wastewater. The BOD_5 analysis measures the oxygen consumed by microorganisms during biochemical degradation of organic material during five days (Tchobanoglous *et al.*, 2014; Ødegaard *et al.*, 2014). This test gives an good indication of the amount of oxygen needed to stabilize the organic material present in the treatment plant, and the efficiency of biological treatment processes (Tchobanoglous *et al.*, 2014). The presents of microorganisms in the sample are crucial for the degradation to occur. Microorganisms will only degrade biodegradable organic material, first the easily biodegradable substances then the slowly biodegradable substances (Ødegaard *et al.*, 2014). Hence, the BOD parameter does not include the non-biodegradable substances present in the solution. The oxygen demand increases with a higher analysis period, if there is no time limitation the ultimate BOD (UBOD) is measured. UBOD gives the oxygen demand needed to degrade all biodegradable substances, readily and slowly biodegradable, present in the solution. The unit used for the BOD parameter is normally g O_2/m^3.

Chemical oxygen demand (COD)

The COD analysis is based on chemical oxidation reactions where the consumption of an oxidant in an acid solution are recalculated to the equivalent oxygen concentration (Ødegaard

et al., 2014). Normally potassium dichromate ($K_2Cr_2O_7$) is used as oxidant within the field of wastewater. A mix of the wastewater sample, the oxidant and sulfuric acid (H_2SO_4) are boiled, then the consumption of the oxidant are registered for each volume unit. The oxidant consumption are then recalculated to the oxygen consumption (g O_2/m^3). The COD parameter differs from BOD, because COD includes non-biodegradable organic material in addition to biodegradable organic material. Substances that are difficult to oxidize biologically are easier oxidized chemically. The COD test are completed after 2,5 hours compared to five days for the BOD test, which is a major advantage (Tchobanoglous *et al.*, 2014). Since COD are a rapid test, and the test is not depended on the presence of microorganisms in the solution it is easier to measure at several places in the treatment plant.

In biological treatment the different COD fractions is important, because microorganism will remove the favored organic substance first. Mainly one can divide the COD fraction into particulate and soluble COD. *The soluble COD* can be further fractionized to readily biodegradable COD (rbCOD), soluble slowly biodegradable COD (ssbCOD) and soluble non-biodegradable COD (snbCOD) (Tchobanoglous *et al.*, 2014). The rbCOD of the soluble COD consist of complex COD and volatile fatty acids (VFA). *The particulate COD* can be divided into particulate slowly biodegradable COD (psbCOD) and particulate non-biodegradable COD (pnbCOD).

Figure 3 gives an overview of the different COD fractions in wastewater, this COD-network has been made based on Pasztor, Thury and Pulai (2009) and Tchobanoglous *et al.* (2014)

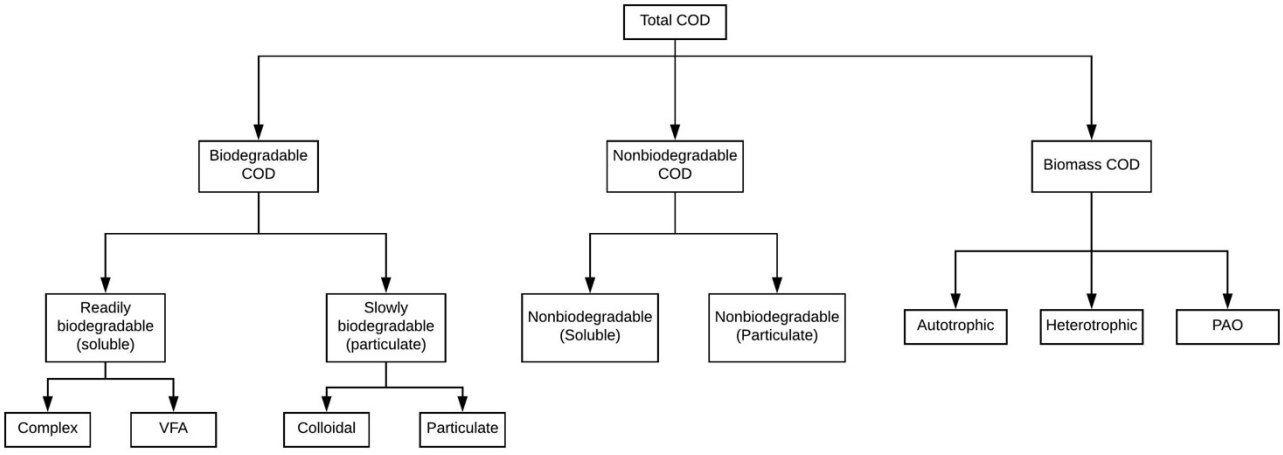

Figure 3: COD fractions in IVAR SNJs wastewater (Pasztor, Thury and Pulai, 2009; Tchobanoglous et al., 2014)

Interrelationship between COD and BOD

As explained above both COD and BOD are parameters used to measure gross amount of organic material in wastewater. COD includes more of the organic material compared with BOD, which makes the COD value higher than the BOD value. The interrelation between COD and BOD varies based on the composition of the wastewater. If the mass of organic biodegradable material is high, the BOD and COD ratio will have a higher value than if the wastewater consist high mass of organic non-biodegradable material. Typical values for the BOD/COD ratio in raw municipal wastewater are normally between 0,3-0,8 (Tchobanoglous *et al.*, 2014). At IVAR SNJ both soluble and particular COD and BOD are measured. Using the measured parameters in the inlet wastewater, one can calculate the ultimate BOD (UBOD) for both soluble and particular organic substances. The UBOD can be linked to the total amount of biodegradable COD (rbCOD, ssbCOD and psbCOD). The rbCOD are easily biodegradable, and are the substances the microorganism will degrade first. This thesis will assume that the rbCOD are equal to BOD_2. Equation (2-1) can be used to find UBOD and BOD_2, based on the measured BOD_5 value.

$$BOD_t = BOD_U\,(1 - e^{-kt}) \qquad (2\text{-}1)$$

Where:

BOD_U	Ultimate BOD concentration [g O_2/m^3]
BOD_t	BOD concentration after t days [g O_2/m^3]
k	First order reaction rate constant [d^{-1}] = 0,23 d^{-1} (Tchobanoglous *et al.*, 2014)
t	Time [d]

The first order reaction rate constant (k) for raw wastewater is normally in between 0,12- 0,46 d^{-1}, with the typical value of 0,23 d^{-1} (Tchobanoglous *et al.*, 2014).

Volatile suspended solids (VSS)

It is common to use volatile suspended solids (VSS) as a parameter for the amount of microorganisms (biomass). But VSS consists of other particulate organic material in addition to biomass as non-biodegradable organic material, and particulate slowly biodegradable organic material (Tchobanoglous *et al.*, 2014). In general, biomass obtain the largest fraction of VSS in wastewater (Ødegaard *et al.*, 2014), and especially in activated sludge plant where

the return sludge mainly consists of biomass. VSS is therefore a good parameter to observe biomass growth through the whole treatment plant (Tchobanoglous *et al.*, 2014). There is a interrelationship between particulate COD and VSS, because these parameters include the same particulate organic material. The correlation between particulate COD depends on the wastewater characteristics, so it varies from treatment plant to treatment plant.

2.1.1.3 Inorganic material

Most inorganic substances in wastewater exist is soluble, but there could also be some particular inorganic substances (Ødegaard *et al.*, 2014). Nutrients, pH and alkalinity and metals are examples of some interesting inorganic constituents in wastewater (Tchobanoglous *et al.*, 2014). Nutrient, pH and alkalinity will be further explained due to their importance within wastewater treatment.

Nutrients

The nutrients of interest within wastewater treatment are mainly phosphorus and nitrogen (Ødegaard *et al.*, 2014). These nutrients are often the limiting factor for growth of algae. High effluent concentration of nutrients can lead to algae bloom and eutrophication in the recipient. Eutrophication can give several unwanted effects in the water body, such as oxygen deficiency, unpleasant odor and taste, high turbidity and high concentration of algal toxins. Prevention of eutrophication is one of the main reason why wastewater is treated. Phosphorus is the limited nutrient in fresh water, and nitrogen in sea water. Phosphorus and nitrogen can also be bound organically and both can exist as soluble or particulate in wastewater, but the largest fraction of phosphorus and nitrogen in wastewater are soluble. The largest proportion of total phosphorus (Tot-P) are normally inorganic in wastewater.

pH and alkalinity

pH and alkalinity are important parameters within the field of wastewater treatment, because biological and chemical processes are depended on an optimal pH range to perform its function (Ødegaard *et al.*, 2014). The parameter alkalinity tells us the solutions ability to resist a change in pH, while pH is a parameter that tells us the amount of hydrogen ions in the solution. In general Norwegian wastewater have low alkalinity, meaning the initial pH easily could change as a result of processes that effects the alkalinity such as nitrification and denitrification. With a change in alkalinity it can be difficult to maintain the desired pH range.

2.1.1.4 Microorganisms

There are several microorganisms presents in raw municipal wastewater, which mainly originate from human waste. In biological treatment processes microorganisms is active in the removal of organic material, nitrogen and phosphorus (Tchobanoglous *et al.*, 2014). Microorganisms consists of carbon, hydrogen, oxygen, phosphorus and nitrogen, and to reproduce they need carbon, nutrients and metals (Dionisi, 2017).

One can divide microorganisms into heterotroph and autotroph based on which carbon source used for cell growth (Ødegaard *et al.*, 2014). Heterotrophs use organic material as their carbon source, while autotrophs use carbon dioxide as their (Tchobanoglous *et al.*, 2014). The energy needed for cell growth is either obtained from light or chemical oxidation reactions. Phototrophs use lights as energy source, while chemotrophs use energy generated from chemical oxidation reactions. Phototrophs and chemotrophs can either be heterotrophic or autotrophic. The chemical oxidation reactions performed by chemotrophs involves transfer of electron from an electron donor to an electron acceptor. The electron acceptor can either be internal and available within the cell during metabolism, or external like dissolved oxygen or nitrite/nitrate. The electron donor is mainly the substrate, organic carbon. Some microorganism can only meet their energy need in aerobic environment (obligate aerobes), while others only survive in anaerobic environment (obligate anaerobes). There are also microorganisms that can live in both anaerobic and aerobic environments (facultative anaerobes) (Tchobanoglous *et al.*, 2014).

Microorganisms prefer to convert the most biodegradable material first, then the slowly biodegradable organic material such as proteins (Ødegaard *et al.*, 2014). Last, microorganisms convert particular organic material that first needs to be hydrolyzed before it is taken up through the cell wall. Particular organic material is only converted to maintain life when there is no other food sources.

Temperature

Temperature is also an important parameter in biological treatment, because the growth rate of microorganism is depended on the temperature (Ødegaard *et al.*, 2014). Lower temperature result in lower growth rates, but if the temperature is too high microorganisms will die. To calculate temperature correlations for a given temperature depended parameter, the general equation (2-2) have been utilized.

$$r_{T1} = r_{T2} * \theta^{T1-T2} \quad (2\text{-}2)$$

Where:

r_{T1}	Conversion rate at temperature T1
r_{T2}	Conversion rate at temperature T2
θ	Temperature coefficient

Kinetics

Heterotrophs use organic material as a foundation to growth, and the conversion of organic material leads to cell growth (Ødegaard *et al.*, 2014). The growth rate are depended on the access to food (substrate), but also the oxygen concentration, pH, temperature and access to nutrients. In a treatment plant the access to substrates are limited, so the specific growth rate (μ) is depended on the given substrate concentration in the reactor, here referred to as biodegradable soluble COD (bsCOD). From equation (2-3) the growth rate of microorganisms can be calculated based on the specific growth rate and the concentration of microorganisms. Equation (2-4), also known as Monods equation, shows that the specific growth rate is depended on the substrate concentration.

$$\frac{dX}{dt} = \mu * X \quad (2\text{-}3)$$

$$\mu = \left(\frac{\mu_{max} * bsCOD}{K_S + bsCOD}\right) \quad (2\text{-}4)$$

Combining equation (2-3) and (2-4):

$$\frac{dX}{dt} = \left(\frac{\mu_{max} * bsCOD}{K_S + bsCOD}\right) X \quad (2\text{-}5)$$

Where:

$\frac{dX}{dt}$	Microbial growth rate from substrate utilization [g VSS /m^3*d]
μ	The specific growth rate [d^{-1}]
μ_{max}	Maximum specific growth rate (d^{-1})
bsCOD	bsCOD concentration [g bsCOD/m^3]
K_S	Half-saturation constant [g bsCOD/m^3]
X	Concentration of microorganisms [g VSS/m^3]

Growth of microorganisms ($\frac{dX}{dt}$) is direct proportional to the substrate utilization rate (r_{su}) and the synthesis yield (Y). The synthesis yield is a coefficient that gives the amount of biomass produced per substrate consumed, see equation (2-6).

$$Y = \frac{g\ VSS_{produced}}{g\ COD_{used}} \tag{2-6}$$

$$\frac{dX}{dt} = Yr_{su} \tag{2-7}$$

Microorganisms will not live forever, but in fact be reduced by death and predation also known as endogenous respiration. The net growth rate (r_{net}) for microbial growth will then be equal to equation (2-8). Where the k_dX term considers the biomass losses occurring.

$$r_{net} = Yr_{su} - k_dX \tag{2-8}$$

$$r_{net} = Y_{obs} * r_{su} \tag{2-9}$$

Where:

r_{net}	Net biomass growth rate [g VSS/m^3*d]
r_{su}	Substrate utilization rate [g bsCOD/m^3*d]
k_d	Endogenous decay coefficient [d^{-1}]
Y_{obs}	The observed yied [gVSS/g COD]

Endogenous respiration considers the decrease in biomass (Tchobanoglous *et al.*, 2014). There are several factors that result in the biomass losses, such as cell lysis due to death and stress, cell maintenance energy needs, and predation. Death of microorganisms can occur due to age, virus or environmental stress such as pH and temperature. Cell lysis results in a release of cellular substrate, which mainly are biodegradable. Reduction of biomass due to cell maintenance energy needs happens when there is a lack of substrate in the solution and the cellular carbon needs to be used to maintain life. Predation entails that microorganism eat each other when there is absence of substrate. The endogenous decay coefficient (k_d) in equation (2-8) account for all these factors, even though the biomass reduction will vary based on the given reduction factor. Equation (2-8) can be simplified by introducing the observed yield (Y_{obs}), this parameter is easily measured in treatment plant, and entails the biomass produced per COD removed for a given system, see equation (2-9).

2.2 Typical characterization of Norwegian wastewater

Depending on the treatment process it could be important to characterize the substances in the wastewater based on particle size and biodegradability, to understand what is removed where in the treatment plant. At IVAR SNJ particle size is important in the physical processes, while biodegradability is important in the biological processes.

Particle size

In the section about solids the methods used to find a distinction between dissolved solids, colloidal material, suspended solids and settleable particles were further explained. Table 1 are based on table 14.3 in Ødegaard *et al.* (2014), and present a rough fractionation of typical Norwegian wastewater with regard to particle size. In Table 1 the suspended fraction also includes settleable particles (>100 μm). From this table one can see that the largest fraction of organic material is particulate (suspended and colloidal), with approximately 75 % of the total. Nitrogen and phosphorus are mainly soluble with 72 % and 60 %, respectively. Nutrients are thereby hard to remove by physical unit processes alone.

Table 1: Fraction with regard to particle size (Ødegaard et al. 2014)

	Parameters (%)			
Fraction	COD	BOD_7	Tot-P	Tot-N
Suspended >1μm	57	53,3	20	16
0,1 μm < **Colloidal** < 1μm	18,6	23,3	20	12
Dissolved < 0,1 μm	24,4	23,3	60	72

The main part of dissolved organic material consists of carbohydrates, which are more biodegradable than protein and fats (Ødegaard *et al.*, 2014). The dissolved substances will more rapidly removed in an biological processes.

Biodegradability

As shown in Table 1 the organic material exists in all particle sizes, and each size have several different fraction of biodegradability. The fractionation of organic material regarding particle size and biodegradability are illustrated in Figure 4.

The ***suspended fraction*** of the organic material can be divided into biomass, slowly biodegradable- and non-biodegradable substances (Ødegaard *et al.*, 2014). The ***colloidal fraction*** of the organic material consists of slowly biodegradable substances and non-biodegradable substances. While the ***dissolved fraction*** consists of a readily biodegradable fraction and a non-biodegradable fraction. The difference between the slowly biodegradable and readily biodegradable is that readily biodegradable substances can be metabolized direct, while the slowly biodegradable substances needs to be hydrolyzed before it can be metabolized by microorganism Examples of organic substances that can be metabolized direct are small molecules of volatile fatty acids, carbohydrates, alcohols, peptones and amino acids (Henze, 1992). There is an difference between particulate and colloidal slowly biodegradable organic material, both needs to be hydrolyzed before metabolized by microorganisms, but the hydrolysis of the soluble/colloidal slowly biodegradable organic material are more rapid than the hydrolysis of the particulate slowly biodegradable material.

(Short chain) volatile fatty acids (VFA) are preferred as substrate for microorganisms performing the enhanced biological phosphorus removal (EBPR) process (Lie and Welander, 1997), and thereby an important parameter in the mass balance. There is an initial VFA concentration in the raw wastewater, but the potential VFA concentration of the wastewater are a sum of the initial VFA present in the solution and the VFA that can be formed through fermentation.

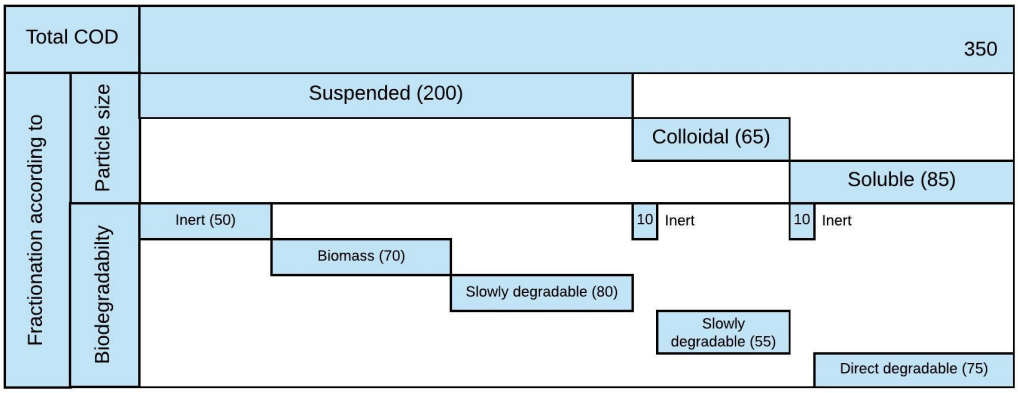

Figure 4: Fractionation of organic material in wastewater with respect to particle size and biodegradability (Ødegaard *et al.*, 2014)

Solid retention time (SRT)

The solid retention time, or sludge age, in the system will affect the biological treatment processes, and is an important design parameter for the activated sludge system. SRT is

determined by dividing the mass of solids in the reactor by the solids removed daily by the effluent and the waste sludge flow (Tchobanoglous *et al.*, 2014), see equation (2-10) Since the activated sludge system at IVAR SNJ consists of one anaerobic reactor were only the return sludge enters (AN1), SRT in the model have been calculated based on equation (2-11).

$$SRT = \frac{VX}{Q_{eff}X_e + Q_wX_R} \tag{2-10}$$

$$SRT_{IVAR} = \frac{(V_{tot,AN1}X_R) + (V_{tot,AN2} + V_{tot,AN3} + V_{tot,aer}) * X}{Q_{eff}X_e + Q_wX_R} \tag{2-11}$$

Where:

SRT	Solid retention time [d]
$V_{tot,ANn}$	Total volume of all three AN1, AN2, or AN3 reactors [m^3]
$V_{tot,aer}$	Total volume of all three aerations tanks [m^3]
Q_{eff}	The effluent flowrate from the treatment plant (Q_{eff}=Q_{in}-Q_w) [m^3/s]
X	Biomass concentration in the reactors [g VSS/m^3]
Q_w	Waste sludge flowrate [m^3/d]
X_{eff}	Biomass concentration in effluent [g VSS/m^3]
X_R	Biomass concentration in return line [g VSS/m^3]

Based on experiments performed by Erdal, Erdal and Randall (2006) it is recommend to have a sludge age, in a EBPR-process based on activated sludge, in the range between 16- 24 days, and 12- 17 days for 5 °C and 10 °C, respectively. In an activated sludge system there is a risk that microorganisms are washed out from the system faster than they reproduce, so it is important to operate with a sludge age higher than the minimum SRT for the given microorganism. SRT_{min} for EBPR is not equal to the SRT_{min} for nitrification. For the EBPR process the minimum sludge age will increase for decreasing temperatures (Mamais and Jenkins, 1992; Erdal, Erdal and Randall, 2006). The minimum sludge age needs to be found from experiments, because influent conditions will most likely affect the washout sludge age (Erdal, Erdal and Randall, 2006). From experiments performed by Mamais and Jenkins (1992) the minimum sludge age at 15 °C is found to be 2,7 days.

2.3 Wastewater treatment line

2.3.1 Screening

Screening is a physical unit process, and the main purpose with screening is to remove coarse material from the flow stream (Tchobanoglous *et al.*, 2014). This is material such as rocks, paper, leaves, branches, rags etc. Removal of coarse material prevents damage or clogs downstream in the treatment plant, and increases the efficiency and reliability in the next treatment processes At IVAR SNJ they have four screens with a filter opening of 6 mm, which goes into the category of fine screens. The capacity of each screen is 1 m^3/s. The substances removed by the screening station are sent to a screw compressor, where the total volume is decreased and is then transported to incineration. The reject water from the compressing process is sent back to the treatment plant right before the sand and grease removal. The screens used at IVAR are Huber belt screen EscaMax. The main principle is for the wastewater to flow through the screen while the solids removed remains at the screen (Huber Technologies, 2011). The solids travels upwards on the screening elements, and at the top of the belt the screen is continuously cleaned by a counter-rotating brush roller and an interior spray nozzle bar.

2.3.2 Aerated sand and grease removal

The second process in IVARs treatment line are aerated sand and grease removal, this is also a physical unit process. The main purpose of the sand and grease trap is to remove grit, oil and grease from the wastewater (Tchobanoglous *et al.*, 2014). The main principal is for grease to float to the surface and form a scum layer, and for sand to settle to the bottom of the basin. Grit consist of sand, gravel and heavy solid materials, and is removed to prevent accumulation of sediments in aeration tanks, digester, pipeline and channels and to protect mechanical equipment from abrasion and abnormal wear. While oil and grease are removed to preserve biological life in the recipient. Normally aerated sand and grease trap are divided into two zones by a longitudinally wall that is extended below the water surface. The purpose of the wall is to create a more quiet zone where grease can float towards the surface (Ødegaard *et al.*, 2014). The rising air bobbles will also collects grease and brings it to the surface in the grease channel where it is removed by a grease removal screw and sent to sludge treatment (Tchobanoglous *et al.*, 2014). Sand is collected at the bottom of the basin and transported with the help of airlift pumps to sand washers. The reject water from the sand washers is sent back to the treatment plant right before the screens. IVAR SNJ have one sand and grease trap with a total capacity of

2,8 m^3/s. If the flow is higher than the capacity, the surplus flow will be bypassed too the outlet. Table 2 gives volume and surface area of the sand and grease trap at IVAR SNJ.

Table 2: Volume and surface area of the sand/grease trap at IVAR SNJ

Design	**Sand**	**Grease**	**Total**
Volume [m^3]	937	513	1450
Surface area [m^2]	247	270	517

2.3.3 Drum filter

IVAR have 20 drum filters as primary treatment with an filter opening of 100 µm. Drum filter is a type of surface filtration, and the main purpose with surface filtration is to remove suspended particular material (Tchobanoglous *et al.*, 2014). The drum filter IVAR use is mechanical and self-cleansing, and the filter is backwashed every 1-2 minute to remove the remaining material on the filter. The removed substances and the backwash water is collected and transported to sludge treatment. Surface filter has become more popular lately, because it is a good replacement for depth filtration due to high effluent quality, smaller footprint and reduced maintenance requirements (Tchobanoglous *et al.*, 2014). Figure 5 shows how the filter operates; the wastewater to be filtered flows inside of the drum filter and is sent to the periphery of the drum while the drum filter rotates slowly (purple arrow). The wastewater then goes through the filter material and continues to the next treatment process (blue arrow). The substances remaining on the inside of the drum filter is collected by a collector inside of the drum and sent to sludge treatment (red arrow). A high pressure water spray is used to loosen and remove accumulated material left on the filter (Teknor; Tchobanoglous *et al.*, 2014)

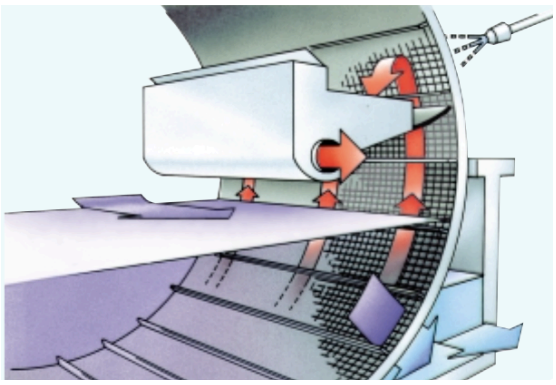

Figure 5: Hydrotech drum filter (Teknor)

2.3.4 Activated sludge

Activated sludge is a suspended growth process where bacteria float freely in the bioreactor and floc together in aggregates (Ødegaard *et al.*, 2014). The bacteria is maintained in liquid suspension by mixing (Tchobanoglous *et al.*, 2014). At IVAR SNJ the first three reactors in the activated sludge system is anaerobic, where mechanical pedals are used as mixers. The last reactor is aerobic, where oxygen is added to maintain a given concentration of dissolved oxygen in the reactor and to achieve mixing. It is important to maintain a suitable concentration of microorganisms in the activated sludge system, hence sludge is recycled in a return sludge line. The return sludge is sent to the first anaerobic tank (AN1). During the activated sludge process floc particles is formed with a size between 50-200 mm, this is needed to optimize the effectiveness at the particle separation step (Tchobanoglous *et al.*, 2014).

The activated sludge process is a complex microbial system, where different groups of heterotrophs and autotrophs bacteria can live side by side to perform different functions (Orhon, 2015). Function like removal of organic carbon, nitrification and denitrification and enhanced biological phosphorus removal. At IVAR SNJ the main purpose is to remove organic material and phosphorus.

2.3.5 Enhanced biological phosphorus removal

The overall reason to remove phosphorus from the wastewater is to prevent eutrophication in the receiving waterbody, especially when the emission point is freshwater (Tchobanoglous *et al.*, 2014). When the emission point is the ocean, as is the case for IVAR, there is no requirement to remove phosphorus in Norway (Miljødepartementet, 2004). IVAR has implemented enhanced biological phosphorus removal (EBPR) because phosphorus is a limited resource that they will utilize in their fertilize product, Minorga. There are two main advantages for using enhanced biological phosphorus removal (EBPR) instead of chemical phosphorus removal: EBPR produce less sludge, and phosphorus is easier to recover.

Basically EBPR involves uptake of phosphorus in biomass. The phosphorus rich biomass is removed by a particle separation step, which is sedimentation basins or clarifiers at IVAR SNJ. For an activated sludge system it is important that the particle separation step function well, because the system is depended on a high amount of biomass. If the separation step do not function well the discharge requirements would not be reached. The biomass removed by the clarifiers is return to the system and the surplus is sent to sludge treatment. Ordinary heterotroph

organisms (OHO) contain about 0,015 g P/g VSS (Tchobanoglous *et al.*, 2014). When new OHOs if formed due to cell growth they will remove soluble phosphorus from the solution. The expected removal efficiency of phosphorus when OHOs is the main microorganism present in the system is 10-20 %. To remove higher amount of phosphorus the presents of polyphosphate accumulating organisms (PAO) are important, because they can store approximately 0,38 g P/g VSS (Amy *et al.*, 2008).

PAOs are encouraged to grow and consume phosphorus in a systems that give PAOs a competitive advantage compared to other bacteria (Tchobanoglous *et al.*, 2014). The competitive advantage comes from being exposed to altering anaerobic and aerobic conditions. In an aerobic environment PAOs stores polyphosphate biochemically, which mean that soluble phosphorus is removed from the wastewater, and the largest fraction is temporary trapped (luxury-P) in the biomass. In the anaerobic reactors PAOs are able to transport and consume readily biodegradable COD (rbCOD) in the form of volatile fatty acids (VFA) by using the energy they have stored as polyphosphate in the previous aerobic reactor. To use the energy stored as polyphosphate there is a release of ortho-phosphate, which generates energy simultaneous with the consumption of VFA. The P-release leads to high amount of soluble ortho-phosphate in the anaerobic tank, which makes phosphorus more available for uptake in the aerobic reactor. OHOs will not be able to consume biodegradable COD in the anaerobic reactors, because they need an electron acceptor as oxygen or nitrate to provide energy for the consumption of organic carbon.

First anaerobic reactor (AN1)

At IVAR SNJ the return sludge enters AN1, which means the sludge have already been through the activated sludge system at least once before. The main purpose with this reactor is to hydrolysis slowly particulate biodegradable organic material to soluble organic compounds that is more biodegradable, and to remove nitrate through denitrification. Normally Norwegian wastewater have a small portion of readily biodegradable organic material in raw wastewater (Ødegaard *et al.*, 2014), so the hydrolysis in AN1 could be essential to achieve the wanted removal efficiency regarding phosphorus. The access to readily biodegradable organic material is important for the EBPR process, and are often the limiting factor in the biological phosphorus removal process (Tchobanoglous *et al.*, 2014). Longer hydraulic retention time (HRT) results in a higher production of readily biodegradable organic material in AN1. If nitrate follows the return sludge line, denitrification would occur in AN1. Denitrification involves the reduction

of nitrate (NO_3) to nitrogen gas (N_2) with the help from OHOs. In the absence of oxygen OHOs will use nitrate as an electron acceptor to be able to consume the organic material present in the reactor. It is important that denitrification occurs before AN2 and AN3, because denitrification entails consumption of readily biodegradable organic material that was supposed to be stored by PAOs (Ødegaard *et al.*, 2014). Nitrate can also affect PAOs metabolism negatively, which can lead to troubles connected to storage of polyphosphate.

Second and third anaerobic reactor (AN2 and AN3)

In the second and the third anaerobic reactors PAOs use the biochemical energy stored as polyphosphate to assimilate VFA, and to produce intracellular poly-β-hydroxyalkanoate (PHA) (Tchobanoglous *et al.*, 2014). PHA is stored in PAOs by using glycogen incorporated in their cells. Simultaneous with the assimilation of VFA orthophosphate is released. The concentration of orthophosphate in the reactor, and the PHA content in PAOs increases through AN2 and AN3. Due to the consumption of VFA the concentration of soluble COD decreases during the anaerobic reactors, meaning there also is removal of organic material in the anaerobic tanks (Ødegaard *et al.*, 2014). VFA originate from the raw wastewater, the rest is produced by bacteria in the anaerobic reactos through fermentation of influent rbCOD (Tchobanoglous *et al.*, 2014). If the hydraulic retention time in AN2 and AN3 is long enough particulate biodegradable COD can be hydrolyzed and fermented to VFA.

Glycogen-accumulating organisms (GAO) in the EBPR process is a major challenge, because they are huge competitors to PAOs (Tchobanoglous *et al.*, 2014). GAOs acts similar as PAOs by surviving under altering anaerobic and aerobic condition, and by consumption of ready biodegradable COD and production of PHA, but GAOs store and release glycogen instead of phosphorus (Shen and Zhou, 2016). Both PAOs and GAOs prefer VFA as their carbon source, so the selection of PAOs over GAOs are important for the EBPR process to work its purpose.

Aeration tank

In the aeration tank the stored PHA in PAOs is metabolized, this process gives energy and carbon source for new cell growth (Tchobanoglous *et al.*, 2014). The energy generated from the PHA oxidation causes polyphosphate synthesis to happen, with the uptake of soluble orthophosphate, and uptake of metal cations from the solution within the bacterial cell. Since there are more bacteria present in the solution due to cell growth the uptake of orthophosphate is higher than the release in the anaerobic reactors. The raw wastewater contributes with more

soluble phosphorus than released in the anaerobic zones, so the cell growth is crucial to remove enough phosphorus in biomass. PAOs are obligate aerobes meaning they need oxygen to grow. PAOs use oxygen as an electron acceptor to metabolize stored PHA, which is a necessary process for PAO to reproduce. The aerobic zone is thereby crucial for PAOs competitive advantage.

In addition to uptake of phosphorus, removal of organic material through aerobic degradation also occurs in this reactor. Higher removal of organic material was one of the main reason for the reconstruction at IVAR SNJ. Many microorganisms will be present in the activated sludge system in addition to PAOs. The aerobic degradation of organic material is mainly processed by OHOs. Ordinary heterotroph organisms degrade organic material to the end products carbon dioxide (CO_2) and Water (H_2O) in aerobic environment (Ødegaard *et al.*, 2014). This process provide OHOs with energy and carbon needed for production of new biomass. If all biodegradable organic material is converted the substances left are (1) CO_2 and H_2O, (2) non-biodegradable organic material, and (3) microorganism (biomass). Organic material are removed to prevent dissolved oxygen depletion in receiving waters (Tchobanoglous *et al.*, 2014).

IVAR SNJ is not designed for nitrogen removal, but there is a high possibility that some nitrogen removal will occur since the biological treatment consists of both aerobic and anaerobic reactors. The dissolved oxygen present in the aerobic tank will first be used by PAOs and OHOs to metabolize stored PHA and to degrade organic material, respectively. If the hydraulic retention time (HRT) long enough and the solid retention time (SRT) is high nitrification will occur. Nitrification is a biological two-step process: (1) ammonium is first oxidized to nitrite, (2) nitrite is so oxidized to nitrate (Ødegaard *et al.*, 2014). Nitrification is mainly performed by the autotroph microorganisms Nitrosomonas and Nitrobacter, here referred to as nitrifiers (NIT).

$$(1)\ \ NH_4^+ + \frac{3}{2}O_2 \rightarrow NO_2^- + 2H^+ + H_2O \qquad \text{Nitrosomonas}$$

$$(2)\ \ NO_2^- + \frac{1}{2}O_2 \rightarrow NO_3^- \qquad \text{Nitrobacter}$$

NIT are obligates aerobes, hence nitrification only happens in aerobic environment. NIT need oxygen to oxidize NH_4 and NO_2 to generate cell energy (Tchobanoglous *et al.*, 2014). Oxidation of ammonium to nitrite is the slowest and thereby limiting reaction, hence nitrite will almost immediately convert to nitrate (Amy *et al.*, 2008, p. 88). With this as a basis the denitrification calculations in the aeration tank has only been conducted for nitrate.

2.3.6 Sedimentation basin

The phosphorus rich biomass needs to be removed from the wastewater before it reaches the outlet. The particle separation process used at IVAR SNJ is sedimentation basins or clarifiers. The main function of this process is to remove particles, generated as flocs in the biological reactor, and remove other suspended solids in the wastewater before the treated wastewater is discharged into the sea. The process is quite simple; The main feature is to provide relatively stagnant conditions in the clarifier such that gravity settling of solid particles occurs (Tchobanoglous *et al.*, 2014). At IVAR SNJ there are four rectangular sedimentation basins after each process line, the wastewater is divided into four equal streams before the wastewater enters the basins. This is an continuously process where the wastewater is lead into the basin at one side and continues to the other side of the basin while particles sediments to the bottom of the basin (Ødegaard *et al.*, 2014). The sludge are collected from sludge pockets, and send back to AN1 as return sludge or sent to sludge treatment as surplus sludge.

2.4 Sludge treatment line

There are three types of sludge: Primary, biological and chemical sludge (Turoviskiy and Mathai, 2006). At IVAR SNJ primary and biological sludge are produced in the treatment plant and sent to sludge treatment. Primary sludge is the sludge removed by physical unit processes such as the mechanical drum filter at IVAR SNJ. Primary sludge is easier to dewater than biological sludge due to its composition of discrete particles and debris. The biological sludge at IVAR SNJ is the waste sludge produced in the activated sludge process. Biological sludge consist of light biological flocs and is thereby more difficult to dewater (Turoviskiy and Mathai, 2006).

2.4.1 Thickening

When sludge enters the sludge treatment line it mainly consist of water, approximately 99-98 % of the sludge is water. The water percentage depends on the sludge type. To make the sludge treatment processes as efficient as possible, it is favorable to reduce the sludge volume (Turoviskiy and Mathai, 2006). The thickening process removed water volume from the sludge, which will increase the solid concentration of the sludge. At IVAR SNJ they use ALDRUM G3 - rotary drum thickeners (RDT), which achieve the reduction in sludge volume by coagulation and flocculation of solids and drainage of free water by rotation of the internal screw. The organic cationic polymer added prior the thickening process at IVAR SNJ is CC floc 6145. The dosage added is different for primary and biological sludge. The dosage for primary sludge is approximately 1 kg polymer/ ton TS, and the dosage for biological sludge are 3 kg polymer/ ton TS.

2.4.2 Strainpress

The thickened primary sludge is sent through a sieve, before the primary and biological sludge is mixed in the storage tank for thickened sludge, see Figure 1. The Huber strainpress removes coarse material that have passed through the screening station, to avoid accumulation of large unknown items in the anaerobic digestion. The strainpress is pipe-shaped with openings of 5 mm that separates sludge and waste. The waste is transported to a compressor by a screw, where it is compacted and dewatered (Huber Technologies). This process has not been included in the mass balance model.

2.4.3 Anaerobic digestion

Anaerobic digestion is an anaerobic stabilization method that transform organic material to the end products methane and carbon dioxide in the absence of oxygen (Ødegaard *et al.*, 2014). In addition to biogas production there will be an organic residue that consists of inert material and the organic material that did not convert in the digester. The purpose with sludge stabilization is to remove odor, and to prevent the sludge to putrefaction when stored or used. The degradation of organic material in the digester will happen in four steps: (1) Hydrolysis, (2) acidogenesis, (3) acetogenesis, (4) and methanogenesis (Tchobanoglous *et al.*, 2014). Step two and three are often refereed as one step, fermentation. The digesters at IVAR SNJ operate in a mesophilic temperature, which range between 30 to 38 °C, and gas recirculation is used to ensure mixing in the digesters. Mixing prevents formation of a scum layer at the surface, accumulation of solids in the bottom and will ensure better contact between active biomass and sludge (Turoviskiy and Mathai, 2006).

Hydrolysis

Hydrolysis is a process where fermentative bacteria convert particulate biodegradable material to less complex compounds that is soluble. Soluble organic material can be metabolized directly by microorganisms (Amy *et al.*, 2008), so new biomass could be produced during the overall anaerobic degradation. In other words hydrolysis transform large complex molecules to smaller less complex compounds (Tchobanoglous *et al.*, 2014). Carbohydrates is transformed to sugar, proteins is converted to amino acids and lipids is hydrolyzed to long chain fatty acids (LCFA).

Acidogenesis

The second process in the overall anaerobic conversion of organic material is acidogenesis or fermentation. During fermentation the hydrolysis products sugar, amino acids and LCFA will be converted to VFA, CO_2 and hydrogen with the help of acidogenic microorganisms (Tchobanoglous *et al.*, 2014). Fermentation of amino acids and sugar results in acetate, propionate, butyrate, CO_2 and hydrogen. The fermentation product of LCFA is acetate, CO_2 and hydrogen. The bacteria performing the fermentation process use the substrate both as electron donor and electron acceptor. In addition to the different fermentation products cell growth will also occur during fermentation (Amy *et al.*, 2008). The acidogenesis process will result in a pH drop in the solution, but acidogenic microorganisms are active even at low pH. Actually CO_2 is the main consumer of alkalinity in a anaerobic digester, and not VFA as many

believes (Tchobanoglous *et al.*, 2014). A low pH can inhibit the methanogens microorganisms, that can stop the biogas production (Amy *et al.*, 2008).

Acetogenesis

Acetogenesis is a process that converts short chain fatty acids (SCFA) produced in the acidogenesis process even further down to the final product of fermentation which is acetate, hydrogen gas and carbon dioxide (Amy *et al.*, 2008). Since acetate is the final product, the acetate produced during acidogenesis will not be converted in this process. The conversion of butyrate, propionate and LCFAs palmitate to acetate requires a low hydrogen concentration. Acetogenesis process is therefore depended on the presence of hydrogen consuming organisms such as hydrogenotrophic methanogens.

Methanogenesis

The final step in the anaerobic degradation of organic material is methanogenesis, which is the conversion of acetate, carbon dioxide and hydrogen to methane and carbon dioxide (Tchobanoglous *et al.*, 2014). This process is carried out by methanogens. There are two groups of methanogenic organisms involved in this process, and they perform different tasks. The first group, aceticlastic methanogens, converts acetate to methane and carbon dioxide. The second group, hydrogenotrophic methanogens, will produce methane by using carbon dioxide as electron acceptor and hydrogen as electron donor. The latter group is needed for the acetogenesis process to occur. The total biogas production normally consist of 65 % methane and 35 % carbon dioxide (Tchobanoglous *et al.*, 2014), where about 70 % of the methane produced is related to the conversion of acetate by aceticlastic methanogens (Amy *et al.*, 2008). Aceticlastic methanogens have a very low growth rate, which is the main reason for the need of a long start-up time in the anaerobic digestion, and that high sludge concentration is favorable (Amy *et al.*, 2008).

2.4.4 Centrifugal dewatering

The sludge has a very different characteristic after leaving the digester compared with the composition entering the digester. Mainly because the solid concentration is reduced, which results in a more fluid sludge. The next process in the sludge treatment line is centrifugal dewatering, where the main principal is to reduce the water volume of the sludge. The sludge volume entering the centrifugal dewatering is reduced with approximately one fifth, which result in a non-fluid sludge (Turoviskiy and Mathai, 2006). At IVAR SNJ there are two

centrifuges in total, but only one in operation at the time. IVAR SNJ use Alfa Laval-ALDEC 556 G2, which is a solid-bowl centrifuge. The sludge is fed at a constant flowrate into the solid-bowl centrifuge, and the rotating bowl separates the sludge into a cake and a dilute liquid stream referred to as centrate (Tchobanoglous *et al.*, 2014). The reject flow, centrate, at IVAR SNJ is discharged directly into the sea. The solid concentration of the dense cake varies normally between 15-36 % (Turoviskiy and Mathai, 2006). The cationic organic polymer CC floc 6144 used at IVAR SNJ is assed prior the centrifugal dewatering, to improve the dewatering process. The polymer dosage at for the centrifuges is normally 5-7 kg/ ton TS.

2.4.5 Thermal drying

The last sludge treatment process is thermal drying. There are three different types of thermal dryers: Indirect dryers, direct dryers and combined dryers (Deng, Su and Yu, 2013). At IVAR SNJ they have two indirect dryers, where one is in operation at all times, and the last works as a spare. The heating media used is vapor, the vapor is produced by an oil and gas boiling system. Thermal drying is a very efficient method of further reduction of moisture content of the dewatered sludge cake (Fernandes *et al.*, 2007). The end product can be formed to pellets, the pellets have a solid content of approximately 85 % TS. At IVAR SNJ the dewatered sludge is heated to approximately 100 °C for 30 minutes in the thermal dryer, this process will ensure the pellets to be pathogen free and qualified to use in agricultural purposes.

This process is thereby a crucial process to ensure that the fertilizer product produces at IVAR, Minorga, is secure to use as a fertilizer. The thermal drying system is enclosed to reduce odor and particle emission to the atmosphere, and the evaporated liquid should be condensed and treated before released (Fernandes *et al.*, 2007). Indirect dryers produce less vapor, consume less energy and there is no contamination of heat media compared with direct dryers (Deng, Su and Yu, 2013). At IVAR SNJ the vapor generated during thermal drying is drawn off by a cyclone to separate dust particles, and then fed into a combined washing and heat recovery tower. The vapor is cooled down to condensate, and the heat energy generated from the condensation process will be fed into the heating system of the treatment plant through a heat exchanger. The condensate is returned to the boiling system to generate new vapor.

3 How to make the mass balance: The methodology

From the theory, one can see that several reactions and processes occur in a treatment plant, depending on the given process. These reactions are crucial for the treatment plant to meet the requirements. In this chapter the method behind the mass balance calculation will be explained. The assumptions made and the most relevant equations created for the each process will be the main focus in this chapter. Some general assumption and limitations that involves the processes are listed below:

- Steady state is assumed in the biological treatment processes
- The flow in and out of all biological reactors is constant ($Q_{in} = Q_{out}$), and the volume is always perfectly mixed.
    - Completely mixed flow reactor (CMFR).

Measured input parameters at IVAR SNJ

The input parameters are calculated based on a dataset of measurement sampled before the screens. The dataset are from 5. January 2017 until 17. January 2018 and are sampled once a week. Not all parameters have been weekly sampled, meaning there are some holes in the dataset for several of the parameters. For each parameter the average value have been calculated based on this dataset. Table 3 present the results of the calculations, these values are used as a basis for the mass material balance model.

Table 3: Average concentration of measured input parameters at IVAR SNJ

Input parameters	**Average concentration [g/m^3]**
TOT-COD	295
tsCOD	80
TOT-BOD	166
Soluble BOD	38,8
SS	192
Tot-P	3,64
Ortho-P	0,94

The inflow is an important parameter for the mass balance, the average flow reaching IVAR SNJ has been calculated based on a dataset from 1.January 2016 until 18. February 2018. This dataset present the inflow to IVAR SNJ every day in cubic meters per day. 2017 was a record year regarding rain events, meaning the inflow was higher this year compared to 2016 and the two months in 2018. $Q_{avg,1}$ in Table 4 represent the average inflow computed based on the dataset from 2016 and the two months in 2018, while $Q_{avg,2}$ represent the average flow calculated based on the whole dataset. $Q_{avg,2}$ will be used as the inflow in the mass balance model, because the concentrations measured corresponds to this flowrate.

Table 4: Average inflow to IVAR SNJ

Parameter	**Average flow [m^3/s]**
$Q_{avg,1}$	1,36 m^3/s
$Q_{avg,2}$	1,41 m^3/s

IVAR SNJ does not measure nitrogen, but this mass balance will include Tot-N, NH_4 and NO_3 as parameters based on theoretical values for Norwegian wastewater. Nitrogen can be removed through nitrification and denitrification in the biological treatment processes, and can interfere with the biological phosphorus removal. The different nitrogen parameters are therefore of relevance for this mass balance. Typical nitrogen concentration in raw wastewater in Norway are presented in Table 5 these values have a basis from Ødegaard *et al.* (2014)

Table 5: Typical nitrogen concentration in raw Norwegian wastewater

Nitrogen [g/m^3]	
Tot- N	24,8
NH4-N	19,1
NO3-N	0,3

Calculated input parameters

This mass balance will focus on TSS, COD, nitrogen and phosphorus as the main parameters. As explained in the theory it can be valuable to fractionize these parameter further down, because different parameters within the main parameters will be removed or converted in a given treatment process. The favorable organic fraction will be removed first in a biological process, like VFA in the AN2 and AN3. For example the hydrotech filter will mainly remove

large particular material, due to a filter opening of 100 μm. To find values for the different COD fractions the interrelationship between COD and BOD has been used. Measured soluble BOD has been used to calculate the fraction of soluble COD, and measured particular BOD has been used to calculate fraction of particular COD. Equations (3-1)-(3-5) have been used to calculate the different input parameters for COD. BOD_U dissolved, BOD_U particular and BOD_2 was calculated based on equation (2-1), with a k-value of 0,23 d^{-1}. The results are presented in Table 6.

The VFA fraction of the readily biodegradable substances are normally 50-70 % in raw wastewater (Henze, 1992). With this as a basis the VFA fraction have been set to 60 % of the readily biodegradable substances in the raw wastewater. The volatile fraction of the total suspended solids (TSS) in raw wastewater have been set to 76,7 % of TSS, based on table 3 presented by Rossle and Pretorius (2001).

Calculation of soluble COD:

$$rbCOD_{in,TP} = BOD_2 \quad (3\text{-}1)$$

$$ssbCOD_{in,TP} = BOD_{U,S} - BOD_2 \quad (3\text{-}2)$$

$$snbCOD_{in,TP} = tsCOD_{in,TP} - (rbCOD + ssbCOD) \quad (3\text{-}3)$$

Calculation of particular COD:

$$psbCOD_{in,TP} = BOD_{U,P} \quad (3\text{-}4)$$

$$pnbCOD_{in,TP} = tpCOD - BOD_{U,P} \quad (3\text{-}5)$$

Where:

BOD_2	Soluble biochemical oxygen demand after two days [g O_2/m^3]
$BOD_{U,P}$	Ultimate particulate biochemical oxygen demand [g O_2/m^3]
$BOD_{U,S}$	Ultimate soluble biochemical oxygen demand [g O_2/m^3]
$pnbCOD_{in,TP}$	Particulate non-biodegradable COD entering the treatment plant [g/m^3]
$psbCOD_{in,TP}$	Particulate slowly biodegradable COD entering the treatment plant [g/m^3]
$rbCOD_{in,\ TP}$	Readily biodegradable COD entering the treatment plant [g/m^3]
$ssbCOD_{in,TP}$	Soluble slowly biodegradable COD entering the treatment plant [g/m^3]

$snbCOD_{in,TP}$ Soluble non-biodegradable COD entering the treatment plant [g/m^3]

Table 6: Calculated BOD values

$BOD_{U,S}$	56,8 g O_2/m^3
$BOD_{U,P}$	186,1 g O_2/m^3
Soluble BOD_2	20,93 g O_2/m^3

The calculated concentration for all parameters entering IVAR SNJ is presented in Table 7, where the measured and assumed concentrations have been used as a basis in the calculations, see Table 3 and Table 4.

Table 7: Calculated concentrations entering IVAR SNJ in g/m^3

INPUT PARAMETERS	
TOT-COD	295
tsCOD	80
rbCOD	20,93
VFA	12,56
ssbCOD	35,84
snbCOD	23,22
tpCOD	215
psbCOD	186,1
pnbCOD	28,86
TSS	192
VSS	147,2
Tot-P	3,64
Ortho-P	0,94
Tot-N	24,8
NH4	19,1
NO3	0,3
O_2	1

3.1 Wastewater treatment line

This part will look into each process in the treatment plant, and discuss assumptions made and equations used to make the mass balance model.

3.1.1 Screening station

At IVAR SNJ wastewater is collected before the screen to measure the concentration of different parameters in the inlet wastewater, in this part of the treatment plant there normally is several large items in suspension that is not relevant for sampling. If trash like rags and Q-tips are collected in the measured sample the result can give the SS parameter an unrealistic concentration. At IVAR SNJ most of the substances removed by the screens do not reach the sampler, because the suction hose will be clogged before large items reach the sampler. Due to the sampling technique the measured input concentrations is set as the output concentrations at the screen station. Calculations done are thereby summarized to the influent concentrations to the screening station, and is made based on data from IVAR SNJ. The trash load daily removed is approximately 500 kg, this trash load normally contain of 40 % TS. It is assumed that only particulate matter will be removed by the screens, due to the screen opening of 6 mm. With this as a basis the trash load will actually contain of 40 % TSS directly. Equation (3-6) has been used to find the TSS concentration removed by the screens.

$$TSS_{r,SS} = \frac{M_{trash} * f_{TS,SS}}{Q} \quad (3\text{-}6)$$

Where:

$TSS_{r,SS}$	TSS concentration removed by the screens [g TSS/m^3]
M_{trash}	The daily trash load removed by the screens [g/d]
$f_{TS,SS}$	The fraction of the daily trash load that is TSS = 0,4
$Q_{in,SS}$	The flowrate entering the screening station [m^3/d]

One can divide TSS into settleable (> 100 μm) and non-settleable TSS, and these fractions consist of one volatile and one fixed fraction. At the screening station mainly settleable TSS will be removed, due to the large filter opening. Approximately 77,3 % of the settleable TSS concentration is settleable VSS (Rossle and Pretorius, 2001), so the VSS concentration removed by the screens is given by equation (3-7).

$$VSS_{r,SS} = 0{,}773 * TSS_{r,SS} \quad (3\text{-}7)$$

The interrelationship between VSS and particulate COD used in this thesis is a pCOD/VSS ratio (f_{cv}) of 1,48 g pCOD/g VSS. This value is selected with a basis in the paper written by Gori *et al.* (2011) and the COD/VSS ratio used by Amy *et al.* (2008). It assumed that a percentage of psbCOD and pnbCOD removed is related to the given percentage of the total particulate COD. The particulate slowly biodegradable COD, and the particulate non-biodegradable COD removed by the screens can be calculated as shown in equation (3-8) and (3-9), respectively.

$$\mathrm{psbCOD_{r,SS}} = \mathrm{f_{cv}} * \mathrm{VSS_{r,SS}} * \left(\frac{\mathrm{psbCOD_{eff,SS}}}{\mathrm{tpCOD_{eff,SS}}}\right) \tag{3-8}$$

$$\mathrm{pnbCOD_{r,SS}} = \mathrm{f_{cv}} * \mathrm{VSS_{r,SS}} * \left(\frac{\mathrm{pnbCOD_{eff,SS}}}{\mathrm{tpCOD_{eff,SS}}}\right) \tag{3-9}$$

Where:

$\mathrm{psbCOD_{r,SS}}$	psbCOD concentration removed by the screens [g COD/m^3]
$\mathrm{pnbCOD_{r,SS}}$	pnbCOD concentration removed by the screens [g COD/m^3]
f_{cv}	Particulate COD/VSS ratio [g COD/g VSS]
$\mathrm{VSS_{r,SS}}$	The VSS concentration removed by the screens [g VSS/m^3]
$\mathrm{psbCOD_{eff,SS}}$	psbCOD concentration leaving screening station [g COD/m^3]
$\mathrm{tpCOD_{eff,SS}}$	Total particulate COD concentration leaving screening station [g COD/m^3]
$\mathrm{pnbCOD_{eff,SS}}$	pnbCOD concentration leaving screening station [g COD/m^3]

The effluent concentration is used as a basis because the effluent concentration is set equal to the measured and calculated concentrations based on data from IVAR SNJ. Meaning the substances removed by the screen will be added to the inlet concentrations in the model.

3.1.2 Aerated sand and grease removal

In the aerated sand and grease removal, obviously sand and grease is removed. The removed sand is assumed to affect the TSS parameter directly, and only the fixed portion of the TSS parameter. Data from IVAR gives that approximately 500 kg sand is removed daily, with a TS of 70-80 %. Same approach used for the screening station was used for the sand and grease removal process. IVAR SNJ did not have any data about the daily grease load removed, so it is assumed a daily grease load of 100 kg/d, with 60 % TS, and that 95 % is VSS in the mass balance model. These values can easily be changed in the model, when data is collected. By assuming these values it will not affect the result in any matter, because it have neglectable

impact in the calculations. The only difference between the sand and grease removed is that the grease load will mainly affect particulate parameters containing organic material meaning TSS, VSS, pnbCOD and psbCOD.

3.1.3 Drum filters

The filter opening of the hydrotech drum filters is 100 μm, and the drum filters are backwash frequently by process water (treated wastewater). The filter removes substances with a larger fraction size than the filter opening, but can also remove smaller substances if there is cake build-up on the filter material. For the hydrotech drum filter used at IVAR SNJ the cake formed on the filters is minimal if it exists at all, due to the frequent backwash. Cake formation has been neglected in the mass balance, meaning the substances removed in the drum filter have fraction size equal or larger than 100 μm. The substances removed by the drum filter is particulate, meaning TSS, VSS, psbCOD, pnbCOD, particulate nitrogen and particulate phosphorus is the parameters reduced due to the drum filter. The Norwegian requirement for primary treatment is assumed to be reached by the drum filter, which is an SS-reduction of 50 % (Miljødepartementet, 2004). It is assumed that 77,3 % of the removed TSS is VSS based on table 3 by Rössle and Pretorius (2001). The amount of TSS and VSS removed by the drum filter is given by equation (3-10) and (3-11). The psbCOD and pnbCOD removed is calculated based on the same approach used for the screens.

$$TSS_{r,DF} = (0{,}5 * TSS_{in,DF}) * Q_{in,DF} \quad (3\text{-}10)$$

$$VSS_{r,DF} = (0{,}773 * TSS_{r,DF}) \quad (3\text{-}11)$$

Where:

$TSS_{r,DF}$	TSS concentration removed by drum filters [g TSS/m^3]
$VSS_{r,DF}$	VSS concentration removed by drum filters [g VSS/m^3]
$Q_{in,DF}$	Flowrate entering the drum filters [m^3/d]

3.1.4 Biological treatment

There are several processes that need to be taking into account in each biological reactor at. In an activated sludge system there is a mixture of microorganisms presents. One can divide the microorganisms of relevance into three population groups (Amy *et al.*, 2008):

1. Nitrifiers (NIT)
2. Ordinary heterotrophic organisms (OHO)
3. Phosphorus accumulating organisms (PAO)

All microorganisms will accumulate phosphorus for cell growth. The contribution of phosphorus removal by NITs can be neglected compared to OHOs and PAOs, as this population constitutes only a small part of the total biomass (Amy et al., 2008). Figure 6 gives an overview of the activated sludge system at IVAR SNJ, and will be used as a basis in the mass balance calculations for the different biological reactors. Where Q_{in} is the flowrate entering the activated sludge system, Q_R the return sludge flow, Q_w is the waste flow and Q_{eff} is the effluent flowrate. X represent the different concentrations in the given flow stream.

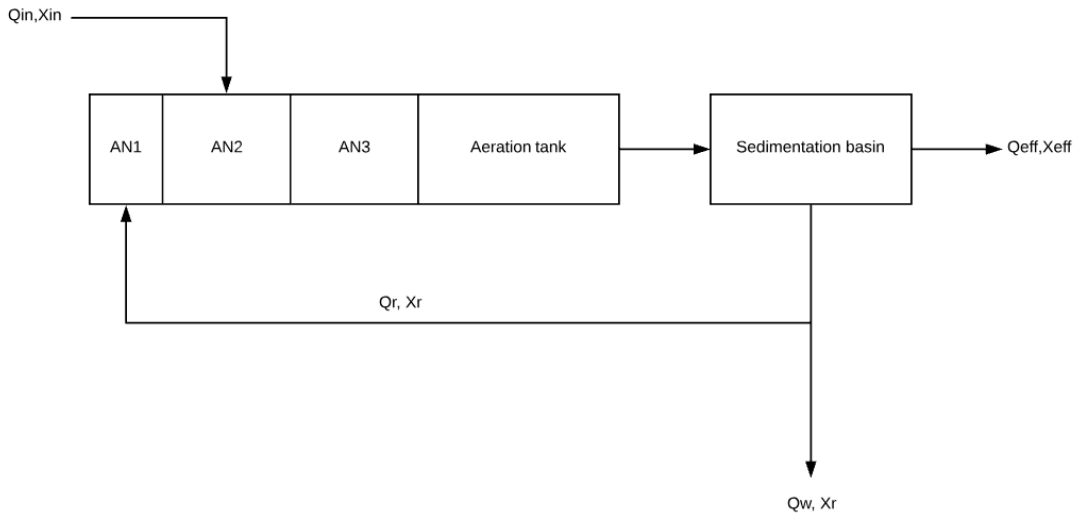

Figure 6: The activated sludge system at IVAR SNJ

Completely mixed continuous-flow reactor (CMFR)

For the biological treatment processes it is assumed the reactors to be completely mixed flow reactors (CMFR. In a CMFR the masses is completely mixed instantaneously and uniformly throughout the reactor (Tchobanoglous *et al.*, 2014). The hydraulic retention time in a CMFR will depend on the boundary volume and the flow entering the reactor. The concentrations in a

CMFR are the same as the effluent concentrations. A schematic sketch of a CMFR is given in Figure 7. The boundary volume in a CMFR is constant, because inflow and outflow is equal.

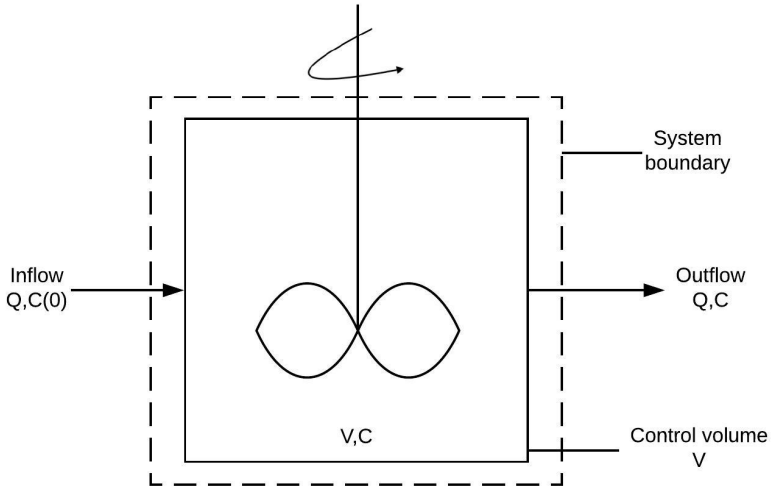

Figure 7: Sketch over a completely mixed flow reactor (CMFR)

Steady state

Equation (1-1) can be rewritten with symbols based on Figure 7, see equation (3-12). When steady state is assumed the accumulation term can be set equal to zero ($\frac{dC}{dt} = 0$), see equation (3-13). The steady state simplification is based on long-term operation, when the concentrations do not change with time (Tchobanoglous *et al.*, 2014).

$$\frac{dC}{dt}V = QC_0 - QC \pm rC \tag{3-12}$$

$$0 = QC_0 - QC \pm rC \tag{3-13}$$

3.1.4.1 First anaerobic tank

Even though IVAR SNJ have the possibility to send raw wastewater into AN1 in addition to the return sludge, this mass balance will make calculations based on the return sludge only. The return sludge mainly consists of biomass, and accumulated particulate non-biodegradable material which is not removed in the biological treatment processes. The particulate slowly biodegradable organics present in AN1 originated from death of microorganisms, also called endogenous respiration. Denitrification will occur when nitrate are present in the return sludge and when the particulate slowly biodegradable organic material have been solubilized. OHOs can use nitrate as electron acceptor when there is absent of oxygen, and readily biodegradable organic material as electron donor. Since the return sludge is the only input into AN1, the amount nitrate converted to nitrogen gas trough denitrification are depended on the readily biodegradable organic material available after hydrolysis. With this as a basis the process order occurring in AN1 are ass listed below:

1. **Hydrolysis**
2. **Fermentation**
3. **Dissolved oxygen (DO)**
4. **Denitrification**

The return sludge line

Concentrations in return sludge line

To perform calculations for the different processes in AN1, the amount of substances entering the reactor is of big interest. The equations used to calculate the input entering AN1 in the mass balance model is presented in Table 8. Data from IVAR SNJ gives that there is approximately 2 % TS in the return sludge, which gives a TSS concentration of 20 000 g/m^3. Based on table 4.2 in Amy *et al.* (2008) the VSS/TSS ratio ($f_{vss,R}$) for settled activated sludge has been set equals to 0,83 g VSS/g TSS.

All soluble material have been set equal to the concentrations in the "output return sludge" in the model. The particulate COD has been calculated based on the pCOD/VSS ratio selected for the system. It is assumed that 25 % of the solids in return sludge line consist of non-biodegradable material (Tchobanoglous *et al.*, 2014). Then two new parameters can be introduced: F_{psbCOD} and F_{snbCOD}.

$$F_{psbCOD} = f_{cv} * 0{,}75 = 1{,}11 \quad (3\text{-}14)$$

$$F_{pnbCOD} = f_{cv} * 0{,}25 = 0{,}37 \quad (3\text{-}15)$$

Where:

F_{psbCOD} 1,11 g psbCOD removed per g VSS removed

F_{pnbCOD} 0,37 g pnbCOD removed per g VSS removed

Table 8: The formulas used to calculate the input to AN1

Parameters	Formulas
TOT-COD	$tsCOD_R + tpCOD_R$
$tsCOD_R$	$rbCOD_R + ssbCOD_R + snbCOD_R$
$rbCOD_R$	$rbCOD_{R,output}$
VFA_R	$VFA_{R,output}$
$ssbCOD_R$	$ssbCOD_{R,output}$
$snbCOD_R$	$snbCOD_{R,output}$
$tpCOD_R$	$psbCOD_R + pnbCOD_R$
$psbCOD_R$	$VSS_R * F_{psbCOD}$
$pnbCOD_R$	$VSS_R * F_{pnbCOD}$
TSS_R	$0{,}02 * \rho_w * 1000\ g/kg$
VSS_R	$TSS_R * f_{VSS,R}$
$TotP_R$	$PO4_R + VSS_R * [(f_{A,PAO} * f_{PAO}) + f_{OHO} * (0{,}75 - f_{A,PAO})]$
$PO4_R$	$PO4_{R,output}$
$TotN_R$	$(VSS_R * 0{,}75 * f_n) + NH4_R$
$NH4_R$	$NH4_{R,output}$
$NO3_R$	$NH4_{R,output}$
$O2_R$	$O2_{R,output}$

TotN will include assimilated nitrogen and the soluble ammonium ions. Where the nitrogen content in biomass (f_n) is assumed to be 0,1 g N/gVSS based on Amy *et al.* (2008), and since 25 % of the particulate material is non-biodegradable it is assumed that 75 % of VSS_R is biomass. The active fraction of PAOs ($f_{A,PAO}$) in the return sludge line varies based on the scenarios tested. The total phosphorus content in PAOs (f_{PAO}) has been set to 0,38 g P/g VSS (Amy *et al.*, 2008). For other microorganism it is assumed a phosphorus content of 0,015 g P/g

VSS (Amy *et al.*, 2008), which equals to the phosphorus content in OHOs (f_{OHO}). Since 75 % of VSS_R is assumed to be biomass the fraction of other microorganism is depended on the selected fraction of active PAOs.

The flowrate in return sludge line (Q_R)

The return flow (Q_R) is an important parameter in the mass balance model, and is needed to perform the calculations in AN1. Low Q_R results in higher hydraulic retention time, which leads to more rbCOD available for denitrification or higher phosphorus release in the AN2 and AN3, and thereby higher P-removal. The simplification made to calculated the return flow is to neglect hydrolysis and fermentation in AN1 and AN2. The system boundary used to find Q_R is presented in a sketch in Figure 8.

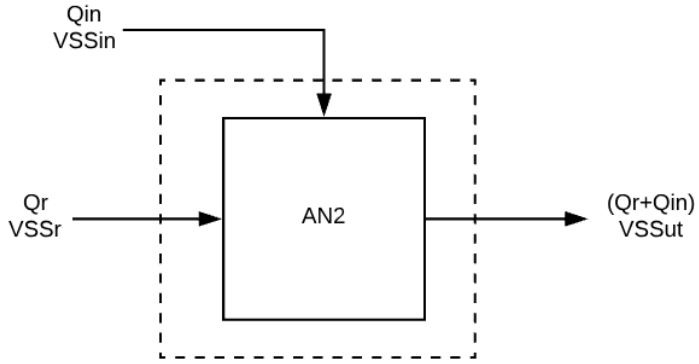

Figure 8: The system boundary used to find Q_R.

The return flow can be calculated based on the mass balance of the biomass, see equation (3-16) and (3-17). Where VSS_{AN2} presented the biomass concentration selected in the reactor.

$$Q_R VSS_R + Q_{in} VSS_{in} = VSS_{AN2}(Q_R + Q_{in}) \tag{3-16}$$

$$Q_R = \frac{Q_{in}(VSS_{AN2} - VSS_{in})}{(VSS_R - VSS_{AN2})} \tag{3-17}$$

1. Hydrolysis in AN1

The first anaerobic tank can be compared with an anaerobic digester due to processing of return sludge, but AN1 has lower temperature and much shorter hydraulic retention time (HRT). One assumption made is that only hydrolysis and acidogenesis fermentation occur in AN1, due to short HRT and low temperature. Figure 9 illustrates the hydrolysis products based on COD

fractions. The percentages in the figure have the basis from Gori *et al.* (2011) and Souza *et al.* (2013). Slowly particulate biodegradable COD are converted to particulate COD (15%) and soluble COD (85%) during hydrolysis (Gori *et al.*, 2011). The percentage converted to rbCOD, VFA and ssbCOD where found based on the faction converted to amino acids, sugar and long chain fatty acids (LCFA) (Souza *et al.*, 2013). The percentage converted to sugar and amino acid is assumed to be readily biodegradable COD, and long chain fatty acids (LCFA) assumed to be ssbCOD. For the fractions converted to snbCOD, psbCOD, pnbCOD values presented by Souza *et al.* (2013) has been used directly.

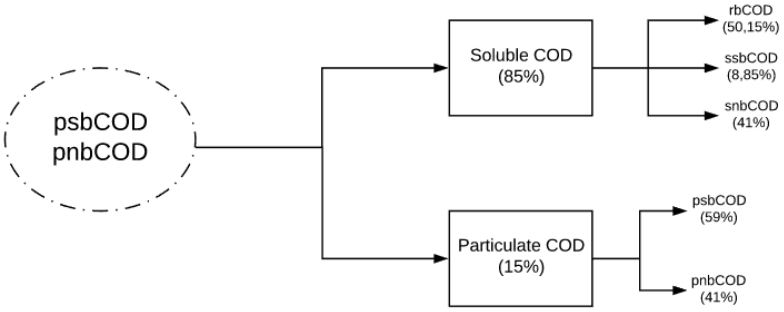

Figure 9: Hydrolysis products based on COD fraction

First order hydrolysis rate constant

To be able to calculate the amount organic material converted by hydrolysis, the hydrolysis rate constant (k_{hyd}) is an important parameter. The hydrolysis conversion rate and the readily biodegradable organic substances available for fermentation are two rate-limiting concepts in anaerobic processes (Tchobanoglous *et al.*, 2014). The first order hydrolysis rate constant is depended on temperature and the complex organic material that is hydrolyzed (carbohydrates, protein or fat). Higher temperature results in a higher k_{hyd} value. For simplification one k_{hyd} has been used in the model, which includes all carbohydrates, protein and fat. Since the hydrolysis rate constant is temperature depended, equation (2-2) has been used in the mass balance to find a correlation for a given temperature in the reactor. θ has been set to 1,14 based on the paper by Zhang *et al.* (2016), where different hydrolysis rates was measured at given temperatures both for cellulose (carbohydrate) and tributyrin (fat). The hydrolysis rate constant for cellulose has been used as a basis in this model, due to reasonable values compared with the hydrolysis rate constants summarized in table 2 by Mani, Sundaram and Das (2016). The hydrolysis rate constant at 15 °C temperature is $K_{hyd,15} = 0{,}03\ d^{-1}$ (Zhang *et al.*, 2016), which is the temperature selected in the mass balance model.

Mass balance calculations: Hydrolysis

COD mass balance

As Figure 9 illustrates the biodegradable particulate COD hydrolyzed are converted to particulate and soluble COD. The psbCOD concentration hydrolyzed is given by equation (3-18).

$$\mathrm{psbCOD_{hyd,AN1}} = \frac{\mathrm{k_{hyd} * V_{tot,AN1} * psbCOD_{in,AN1}}}{\mathrm{Q_R}} \quad (3\text{-}18)$$

Where:

$\mathrm{psbCOD_{hyd,AN1}}$	psbCOD hydrolyzed in AN1 [g COD/m^3]
$\mathrm{V_{tot,\ AN1}}$	The total volume of all three AN1 reactors [m^3]
$\mathrm{k_{hyd}}$	The first order hydrolysis rate constant [d^{-1}]
$\mathrm{psbCOD_{in,AN1}}$	psbCOD entering AN1 [g COD/m^3]

With equation (3-18) as a basis the different COD fraction formed can be calculated, see Table 9. 85 % of $\mathrm{psbCOD_{hyd}}$ are assumed to be converted to soluble COD (f_s), and 15 % to particulate COD (f_p). The total COD mass before and after hydrolysis is assumed to be identical, because particulate COD will mainly transform to simpler, more biodegradable COD-fractions. The only reduction in total COD concentration is assumed to occur when COD is utilized for cell growth.

Table 9: Equations used to calculate the different COD fractions formed through hydrolysis

HYDROLYSIS IN AN1		
Main parameters	**Under parameters**	**Formulas**
Soluble COD	**$\mathrm{tsCOD_{hyd,AN1}}$**	$\mathrm{psbCOD_{hyd,AN1} * f_s}$
	$\mathrm{rbCOD_{hyd,AN1}}$	$\mathrm{tsCOD_{hyd,AN1}} * 0{,}515$
	$\mathrm{ssbCOD_{hyd,AN1}}$	$\mathrm{tsCOD_{hyd,AN1}} * 0{,}0885$
	$\mathrm{snbCOD_{hyd,AN1}}$	$\mathrm{tsCOD_{hyd,AN1}} * 0{,}41$
Particulate COD	**$\mathrm{tpCOD_{hyd,AN1}}$**	$\mathrm{psbCOD_{hyd,AN1}} * \mathrm{f_p}$
	$\mathrm{psbCOD_{hyd,AN1}}$	$\mathrm{tpCOD_{hyd,AN1}} * 0{,}59$
	$\mathrm{pnbCOD_{hyd,AN1}}$	$\mathrm{tpCOD_{hyd,AN1}} * 0{,}41$

Biomass mass balance

Hydrolysis results in a reduction in biomass (VSS) because the particulate slowly biodegradable COD hydrolyzed in AN1 are biomass. The reduction in biomass is given by equation (3-19).

$$VSS_{hyd,AN1} = \left(\frac{psbCOD_{hyd}}{f_{cv}}\right) \quad (3\text{-}19)$$

2. Acidogenesis fermentation in AN1

The second process in the overall anaerobic degradation is fermentation. The organic material ready for fermentation in AN1 have been set equal to the hydrolysis products produced in the same reactor, see equation (3-23). Both hydrolysis and fermentation are time-depended, but the acidogenesis fermentation are the most rapid conversion process in anaerobic degradation of organic material (Amy *et al.*, 2008). With this as a basis it is reasonable to think that all organic material hydrolyzed in AN1, could be fermented.

Mass balance calculation: Acidogenesis fermentation

Fermentation will mainly affect the biodegradable soluble COD (bsCOD) in the tank; in AN1 this means the hydrolysis products produced. The particulate COD before and after acidogenesis is assumed to be equal in the model, with the exception of the COD utilized for cell growth during fermentation. Even though this is not the reality, because some CO_2 will be produced in this process.

COD mass balance

Some of the biodegradable soluble COD available in AN1 will be converted to VFA during fermentation. The COD mass balance for fermentation is represented in equation (3-20).

$$bsCOD_{in,AN1} Q_R = (bsCOD_{res,AN1} Q_R) + (r_{fer} V_{tot,AN1}) \quad (3\text{-}20)$$

Where:

$bsCOD_{in,AN1}$	Biodegradable soluble COD available in AN1[g bsCOD/m^3]
$bsCOD_{res,AN1}$	The bsCOD concentration not fermented in AN1 [g bsCOD/m^3]
Q_R	The return flow [m^3/d]
$V_{tot,AN1}$	The total volume of all AN1 tanks [m^3]

Where the conversion rate for fermentation, r_{fer}, are given by equation (3-21).

$$r_{fer,AN1} = -k_{fer,T} * bsCOD_{in,AN1} * X_{OHO,active,AN1} \quad (3\text{-}21)$$

$$X_{OHO,active,AN1} = VSS_R * f_{A,OHO} \quad (3\text{-}22)$$

Where:

r_{fer}	Conversion rate of fermentable organic material [g COD/m^3 *d]
$k_{fer,T}$	First order fermentation rate constant at T degrees [m^3/g VSS*d]
$bsCOD_{in,AN1}$	The bsCOD concentration available for fermentation in AN1 [g COD/m^3]
$X_{OHO,active,AN1}$	Concentration of active OHOs in AN1 [g VSS/m^3]
VSS_R	VSS concentration in the return sludge line [g VSS/m^3]
$f_{A,OHO}$	Fraction of active OHOs in the return sludge line

Since the fraction of active OHOs, $f_{A,OHO}$, is unknown for the system this value is set to be 21,4 % of VSS_R in the model. This percentage is selected based on the assumption that the active PAOs consist of 40 % of the active biomass, or 15 % of VSS, and that NITs consist of 3 % of the active biomass in the system (Amy *et al.*, 2008). If one assume the active biomass to only consist of PAOs, OHOs and NITs as a simplification, the fraction of OHOs can be calculated to be 21,4 % of VSS_R. This fraction has been kept constant in the model for all scenarios, even when the fraction of PAOs have changed. Fermentation is the only process that includes the active OHOs concentration in the mass balance model, so this assumption will most likely not affect the results in a large extent.

The bsCOD concentration available for fermentation in AN1 is given by equation (3-23).

$$bsCOD_{in,AN1} = tsCOD_{hyd,AN1} * (f_{rbCOD} + f_{ssbCOD}) - rbCOD_{DO,AN1} \quad (3\text{-}23)$$

Where:

$tsCOD_{hyd,AN1}$	The total soluble COD concentration after hydrolysis in AN1 [g/m^3]
f_{rbCOD}	The fraction of tsCOD converted to rbCOD during hydrolysis [g/m^3]
f_{ssbCOD}	The fraction of tsCOD converted to ssbCOD during hydrolysis [g/m^3]

The bsCOD concentration converted to VFA in AN1 are given by equation (3-24), the final $bsCOD_{fer}$ value used in the mass balance is depended on the bsCOD available in the solution, and found with a basis in Table 10.

$$\mathrm{bsCOD_{fer,AN1}} = \frac{r_{fer} * V_{tot,AN1}}{Q_R} \quad (3\text{-}24)$$

Table 10: $bsCOD_{fer}$ are depended on the bsCOD available in the solution

If	Then
$bsCOD_{fer,AN1} > bsCOD_{in,AN1}$	$bsCOD_{fer,AN1} = bsCOD_{in,AN1}$
$bsCOD_{fer,AN1} < bsCOD_{in,AN1}$	$bsCOD_{fer,AN1} = bsCOD_{fer,AN1}$

Biomass mass balance

The increase in biomass due to fermentation (VSS_{fer}) is given by equation (3-25).

$$\mathrm{VSS_{fer,AN1}} = \left(Y_{fer} * \mathrm{bsCOD_{fer,AN1}}\right) \quad (3\text{-}25)$$

Where:

$VSS_{fer,AN1}$ The increase in biomass due to fermentation in AN1 [g VSS/m^3]

Y_{fer} The synthesis yield for acidogenic microorganisms [g VSS/g COD_{fer}]

3. Dissolved oxygen (DO) in AN1

If dissolved oxygen follows the return sludge OHOs will remove COD in the anaerobic reactors, by using oxygen as an electron acceptor and organic carbon as the electron donor. This process will occur before denitrification, because oxygen is a more "wanted" electron acceptor compared to nitrate.

COD removal due to dissolved oxygen:

The COD removed are either oxidized or incorporated in new cells, see equation (3-26). In this mass balance it is assumed that only rbCOD is removed when oxygen is used as electron acceptor in the anaerobic reactors. The amount of rbCOD removed is dependent on the amount of oxygen dissolved and rbCOD available in the reactor. It is assumed that 1 g COD is oxidized for every gram oxygen removed, see equation (3-27). To find the amount of rbCOD consumed due to DO in the anaerobic reactor the following approach have been used, based on equation (7-129) formed by Tchobanoglous *et al.* (2014).

$$\mathrm{rbCOD_{DO}} = \mathrm{rbCOD_{ox}} + \mathrm{rbCOD_{cells}} \quad (3\text{-}26)$$

Where:

$rbCOD_{DO}$	rbCOD concentration removed due to DO [g rbCOD/m^3]
$rbCOD_{ox}$	rbCOD concentration removed by oxidation [g rbCOD/m^3]
$rbCOD_{cells}$	rbCOD concentration incorporated in cells [g rbCOD/m^3]

$$rbCOD_{ox} = DO \tag{3-27}$$

$$rbCOD_{cells} = f_{cv} * Y_{OHO} * rbCOD_{DO} \tag{3-28}$$

If equation (3-28) is included in equation (3-26), the COD oxidized can also be calculated as shown in equation (3-29).

$$rbCOD_{ox} = rbCOD_{DO}(1 - f_{cv} * Y_{OHO}) \tag{3-29}$$

Where:

DO	Dissolved oxygen removed [g O_2/m^3]
f_{cv}	tpCOD/VSS ratio = 1,48 g COD/g VSS
Y_{OHO}	Synthesis yield for OHOs [g VSS/g COD]= 0,45 (Tchobanoglous *et al.*, 2014)
$rbCOD_{DO}$	rbCOD consumed due to DO [g COD/m^3]

If equation (3-27) is set equal to equation (3-29) the following equation can be formed:

$$rbCOD_{DO}(1 - f_{cv} * Y_{OHO}) = DO \tag{3-30}$$

To find the amount of rbCOD consumed due to DO in the anaerobic reactor, one can divide DO on $rbCOD_{DO}$, which result in equation (3-31).

$$F_{DO} = \frac{rbCOD_{DO}}{DO} = \frac{1}{1 - f_{cv} * Y_{OHO}} = 2{,}99\ \frac{g\ COD}{g\ O_2} \tag{3-31}$$

The only difference between equation (3-13) and equation (7-129) formed by Tchobanoglous *et al.* (2014) is the f_{cv} value. The amount of rbCOD that can potentially be removed due to DO in AN1 is given by equation (3-32).

$$rbCOD_{pot,DO,AN1} = F_{DO} * DO_{in,AN1} \tag{3-32}$$

Where:

$rbCOD_{pot,DO,AN1}$	The potential rbCOD removal in AN1 due to DO [gCOD/m^3]
$DO_{in,AN1}$	The dissolved oxygen concentration entering AN1 [g O_2/m^3]
F_{DO}	The amount of rbCOD consumed per DO removed [g rbCOD/g O_2]

The rbCOD actually consumed in AN1 because of DO ($rbCOD_{DO,AN1}$) is depended on the rbCOD available in the reactor. If the dissolved oxygen concentration entering the reactor is high, the potential rbCOD removal in the reactor would be high, but it is not possible to removed more rbCOD than what is available, see Table 11.

Table 11: The rbCOD consumed due to DO is depended on the rbCOD available in the reactor

If	**Then**
$rbCOD_{pot,DO,AN1} \geq rbCOD_{ava,DO,AN1}$	$rbCOD_{DO,AN1} = rbCOD_{ava,DO,AN1}$
$rbCOD_{pot,DO,AN1} < rbCOD_{ava,DO,AN1}$	$rbCOD_{DO,AN1} = rbCOD_{pot,DO,AN1}$

The rbCOD available in AN1 is the rbCOD formed in the reactor through hydrolysis and fermentation, and is given in equation (3-33) and (3-34).

$$\mathrm{rbCOD_{ava,DO,AN1}} = \mathrm{rbCOD_{fer,AN1}} \tag{3-33}$$

$$\mathrm{rbCOD_{fer,AN1}} = \mathrm{if}\,[\mathrm{bsCOD_{fer,AN1}} < \mathrm{rbCOD_{hyd,AN1}}; \mathrm{rbCOD_{hyd,AN1}}; \mathrm{bsCOD_{fer,AN1}}] \tag{3-34}$$

Increase in biomass (VSS) due to DO:

The increase in biomass due to presents of DO is given by equation (3-35).

$$\mathrm{VSS_{DO}} = \mathrm{Y_{OHO}} * \mathrm{rbCOD_{DO}} \tag{3-35}$$

Where:

VSS_{DO}	The increase in biomass due to dissolved oxygen [g VSS/m^3]
Y_{OHO}	The synthesis yield for OHOs [g VSS/g COD]
$rbCOD_{DO}$	rbCOD concentration utilized for cell growth due to DO [g COD/m^3]

4. Denitrification (DN) in AN1

Readily biodegradable COD will be consumed by OHOs when nitrate are present in the anaerobic reactors. In the denitrification process OHOs will use nitrate as electron acceptor and

rbCOD as electron donor, this process will lead to the transfer of nitrate to nitrogen gas, the nitrogen gas will then escapes to the atmosphere (Tchobanoglous *et al.*, 2014). The utilization of rbCOD in the denitrification process happens simultaneous with the hydrolysis of slowly biodegradable COD. Influent rbCOD results in faster denitrification rates, compared to slowly biodegradable COD (Amy *et al.*, 2008). This thesis has made the assumption that only rbCOD will be removed due to denitrification in the anaerobic reactors. The denitrification process in AN1 is depended on the hydrolysis process, because the return sludge mainly consist of biomass. The oxygen equivalent for nitrate (NO_3) is 2,86 g O_2/g NO_3-N, and the oxygen equivalent for nitrite (NO_2) is 1,71 g O_2/g NO_2-N (Tchobanoglous *et al.*, 2014). When nitrate is used as an electron acceptor more COD is needed to remove nitrate from the solution, compared to dissolved oxygen as an electron acceptor. Similar to the DO in anaerobic reactor, COD removed is either oxidized or incorporated in cells. The same approach used to find F_{DO} was used to find F_{DN} only with a higher oxygen equivalent, because nitrate is the electron acceptor not oxygen. The rbCOD removed per nitrate removed (F_{DN}) is given by equation (3-36).

$$F_{DN} = \frac{2{,}86}{1 - f_{cv} * Y_{OHO,DN}} \tag{3-36}$$

The yield for denitrification in anaerobic tanks have been set to 0,32 g VSS produced per g COD removed with an basis in Tchobanoglous *et al.* (2014). Then the F_{DN} parameter is calculated to be 5,43 g rbCOD/g NO_3.

Mass balance calculations: Denitrification

Nitrogen mass balance

The nitrate removed from the solution is of great interest in the mass balance model, because the efficiency of the EBPR is negatively affected by the denitrification process. The potential nitrate removal through denitrification can be calculated based on equation (3-37).

$$NO3_{pot,AN1} = \frac{1}{F_{DN}} * rbCOD_{ava,DN,AN1} \tag{3-37}$$

$NO3_{pot}$	The potential nitrate removed through denitrification in AN1 [g NO3/m^3]
$rbCOD_{ava,DN,AN1}$	rbCOD concentration available for DN in AN1 [g COD/m^3]
F_{DN}	The amount of rbCOD consumed due to denitrification [g rbCOD/g NO3]

F_{DN}= 5,43 g rbCOD/g NO3

The nitrate concentration removed through denitrification is depended on the rbCOD available in the reactor, which is given by equation (3-38). The nitrate removed in AN1 ($NO3_{DN,AN1}$) is given by the dependency presented in Table 12.

Table 12: NO3 removed in AN1 depends on the $rbCOD_{ava}$ and influent NO3

If	Then
$NO3_{pot,AN1} \geq NO3_{in,AN1}$	$NO3_{DN,AN1} = NO3_{in,AN1}$
$NO3_{pot,\ AN1} < NO3_{in,AN1}$	$NO3_{DN,AN1} = NO3_{pot,AN1}$

$$\mathrm{rbCOD_{ava,DN,AN1}} = \mathrm{rbCOD_{fer,AN1}} - \mathrm{rbCOD_{DO,AN1}} \quad (3\text{-}38)$$

Where:

$rbCOD_{ava,DN,AN1}$	rbCOD available for denitrification in AN1 [g COD/m^3]
$rbCOD_{fer,AN1}$	rbCOD concentration in the reactor after fermentation [g COD/m^3]
$rbCOD_{DO,AN1}$	rbCOD removed due to DO in AN1 [g COD/m^3]

COD mass balance

In the COD removed due to denitrification is calculated based on equation (3-39).

$$\mathrm{rbCOD_{DN,AN1}} = 2{,}86 * \mathrm{NO3_{DN,AN1}} \quad (3\text{-}39)$$

Where

$rbCOD_{DN,AN1}$	The rbCOD removed through denitrification in AN1 [g COD/m^3]
2,86	gram COD consumed per gram nitrate removed
$NO3_{DN,AN1}$	Nitrate concentration removed in denitrification in AN1 [g NO3/m^3]

Biomass (VSS) mass balance

Denitrification leads to a increase in biomass, due to cell growth. The net increase in biomass is calculated based on equation (3-40).

$$\mathrm{VSS_{DN}} = \mathrm{Y_{OHO,DN}} * \mathrm{rbCOD_{DN,AN1}} \quad (3\text{-}40)$$

Where:

VSS_{DN}	The increase in biomass due to denitrification [g VSS/m^3]

$Y_{OHO,DN}$ Synthesis yield for OHO when nitrate is the electron acceptor [g VSS/gCOD]

Cell growth leads to an increase in particulate COD

In AN1 fermentation, DO and DN results in production of new biomass (VSS), this will affect the particulate COD concentration. The increase in particulate COD due to cell growth is calculated as shown in equation (3-42) and (3-42). This needs to be done for all processes that leads to an increase in biomass in all reactors.

$$\mathrm{psbCOD_{increase}} = \mathrm{VSS_{new}} * \mathrm{F_{psbCOD}} \quad (3\text{-}41)$$

$$\mathrm{pnbCOD_{increase}} = \mathrm{VSS_{new}} * \mathrm{F_{pnbCOD}} \quad (3\text{-}42)$$

Where:

$psbCOD_{increase}$ Increase in psbCOD due to cell growth [g COD/m^3]

$pnbCOD_{increase}$ Increase in psbCOD due to cell growth [g COD/m^3]

VSS_{new} The increase in VSS due to cell growth [g VSS/m^3]

3.1.4.2 Second and third anaerobic tank

The second and third anaerobic reactor is identical in size and have the same treatment purpose. The main difference in AN2 and AN3 compared to AN1 is the composition of the inlet wastewater entering the reactors. In AN2 a mixture sludge leaving AN1 and the raw wastewater from the drum filters enters the reactor. The composition of the raw wastewater is more biodegradable compared to the sludge from AN1, which mainly consist of biomass. There are several processes that occur in this tank, which is listed below.

1. ***Dissolved oxygen (DO)***
2. ***Denitrification (DN)***
3. ***Hydrolysis***
4. ***Fermentation***
5. ***P-release***

Calculation of inlet concentration entering AN2

The masses from AN1 is small compared to the masses in the raw wastewater, because the return flow is very small compared to $Q_{in.}$ Equation (3-43) gives the general equation of how the different inlet concentration entering AN2 have been calculated in the model.

$$C_{in,AN2} = \frac{(C_{eff,AN1} * Q_R) + (C_{eff,in} * Q_{in})}{(Q_R + Q_{DF})} \quad (3\text{-}43)$$

Where:

$C_{in,\ AN2}$	The concentration entering AN2 [g/m^3]
$C_{eff,AN1}$	The concentration leaving AN1 [g/m^3]
C_{in}	The concentration entering process lines [g/m^3]
Q_{in}	The flow rate entering the process lines [m^3/d]

1. Dissolved oxygen (DO) in AN2 and AN3

The same equations used in AN1 was also used for AN2 and AN3 for COD removal due to the presence of dissolved oxygen. For reactor AN2 the COD removal due to DO has been divided into two time steps.

COD removal due to DO in AN2

1) <u>Step 1: COD removal occurs immediately</u>

In the first time step DO is removed due to influent rbCOD. The actually rbCOD ($rbCOD_{DO1,AN2}$) and DO ($DO_{1,AN2}$) removal in the first step is depended on the rbCOD entering AN2 ($rbCOD_{in,AN2}$).

$$rbCOD_{pot1,AN2} = F_{DO} * DO_{in,AN1} \quad (3\text{-}44)$$

$$DO_{1,AN2} = \frac{1}{F_{DO}} * rbCOD_{DO1,AN2} \quad (3\text{-}45)$$

$$rbCOD_{ava,DO1,AN2} = rbCOD_{in,AN2} \quad (3\text{-}46)$$

$$DO_{res,AN2} = DO_{in,AN2} - DO_{1,AN2} \quad (3\text{-}47)$$

Still dissolved oxygen in the reactor?

- **Yes**: $DO_{res,AN2} > 0$
  - COD and DO removal will occur when rbCOD are formed through hydrolysis and fermentation. Follow step 2.
- **No**: $DO_{res,AN2} = 0$
  - Effluent dissolved oxygen concentration equals to zero.

2) Step 2: COD removal occur when rbCOD are formed through hydrolysis and fermentation

The residue dissolved oxygen concentration in AN2 ($DO_{res,AN2}$) will first be removed when rbCOD is made available through hydrolysis and fermentation.

$$rbCOD_{pot2,AN2} = F_{DO} * DO_{res,AN2} \quad (3\text{-}48)$$

$$DO_{2,AN2} = \frac{1}{F_{DO}} * rbCOD_{DO2,AN2} \quad (3\text{-}49)$$

$$rbCOD_{ava,DO2,AN2} = rbCOD_{fer,AN2} \quad (3\text{-}50)$$

Where:

$rbCOD_{ava,DO1,AN2}$ rbCOD available in the first DO step [g COD/m^3]

$rbCOD_{DO1,AN2}$	rbCOD removed in the first DO step [g COD/m^3]
$DO_{res,\ AN2}$	The DO residue after the firste DO step [g O_2/m^3]

COD removal due to DO in AN3

If there still is dissolved oxygen in the solution when the wastewater reach the third anaerobic reactor, the rbCOD concentration entering AN3 is zero, due to the presents of dissolved oxygen in the previous anaerobic reactors. For DO and rbCOD to be removed in AN3 rbCOD needs to be formed through hydrolysis and fermentation, hence the similar calculations performed for AN1 have been used for AN3.

2. Denitrification (DN) in AN2 and AN3

Also for denitrification the calculations for AN2 and AN3 differs, due to variance in the materials entering the reactor. If not all rbCOD is consumed in step 1 due to DO in AN2, and there is nitrate present in the reactor the denitrification process could also occur in two different time steps. For AN3 the denitrification is dependent on the rbCOD formed through hydrolysis and fermentation similar to in AN1. The nitrate concentration present in the anaerobic reactors at IVAR SNJ will in general be low, because even though nitrate can be produced in the aeration tank the largest nitrate mass will follow the treated wastewater and be discharged in the sea, because Q_{eff} is higher than Q_R.

Denitrification in AN2

1) <u>Step 1: Denitrification happens immediately</u>

Nitrate removed by the inlet rbCOD concentration is calculated based in the the following equations.

$$NO3_{pot1,AN2} = (\frac{1}{F_{DN}} * rbCOD_{ava,DN1,AN2}) \quad (3\text{-}51)$$

$$rbCOD_{ava,DN1,AN2} = rbCOD_{in,AN2} - rbCOD_{DO1,AN2} \quad (3\text{-}52)$$

$$NO3_{res,AN2} = NO3_{in,AN2} - NO3_{DN1,AN2} \quad (3\text{-}53)$$

- Still nitrate in the reactor?
  - **Yes**: $NO3_{res,AN2} > 0$

- Denitrification will occur when rbCOD are formed through hydrolysis and fermentation. Follow step 2.

- **No:** $NO3_{res,AN2} = 0$
  - Effluent nitrate concentration equals to zero.

2) <u>Step 2: Denitrification occurs when rbCOD are formed through hydrolysis and fermentation</u>

Nitrate removed by the rbCOD available after hydrolysis and fermentation is given by the following equations.

$$NO3_{pot2,AN2} = (\frac{1}{F_{DN}} * rbCOD_{ava,DN2,AN2}) \quad (3\text{-}54)$$

$$rbCOD_{ava,DN2,AN2} = rbCOD_{fer,AN2} - rbCOD_{DO2,AN2} \quad (3\text{-}55)$$

Where:

$NO3_{DN1,AN1}$	Nitrate concentration removed in the first DN step [g $NO3/m^3$]
$NO3_{res,AN2}$	The nitrate residue in the solution after the first DN step in AN2 [g $NO3/m^3$]
$rbCOD_{ava,DN1,AN2}$	rbCOD available for DN in the first DN step in AN2 [g COD/m^3]

3. Hydrolysis in AN2 and AN3

The same approach and constants used in AN1 have also been used for AN2 and AN3 to find the amount of psbCOD hydrolyzed. The main difference between AN1 and AN2/AN3 is the hydraulic retention time (HRT). The HRT are shorter in AN2 and AN3 due to higher flow, which means that less psbCOD is hydrolyzed in these tanks compared to AN1.

4. Acidogenesis fermentation in AN2 and AN3

The same approach used in AN1 have also been used for AN2 and AN3 to find the amount of bsCOD converted to VFA through fermentation. The main difference is that fermentation can start immediately in AN2, because there is rbCOD and ssbCOD present in the inlet of the reactor. The hydrolysis started in AN1 will continue in AN2 and AN3, meaning some of the psbCOD not hydrolyzed in AN1 will be hydrolyzed in the next anaerobic reactors. The fermentation process and the hydrolysis process will run side by side in AN2, since both hydrolysis product and psbCOD are present in these reactors. The biodegradable soluble COD available for fermentation in AN2/AN3 is given by equation (3-56). The two last term in equation (3-56) is only relevant for reactor AN2.

$$bsCOD_{in,ANn} = (rbCOD_{in,ANn} + ssbCOD_{in,ANn}) + tsCOD_{hyd,ANn} * (f_{rbCOD} + f_{ssbCOD}) - rbCOD_{DO1,AN2} - rbCOD_{DN1,AN2} \quad (3\text{-}56)$$

The active fraction of OHOs in AN2 and AN3 is calculated as shown I equation (3-57), since $Q_{in,AN2} = Q_{in,AN3}$ this equation is used for both reactors.

$$X_{OHO,active,AN2} = \frac{X_{OHO,active,AN1} * Q_R}{Q_{in,AN2}} \quad (3\text{-}57)$$

5. Release of phosphorus by PAOs in AN2 and AN3

As explain in the theory PAOs use the energy stored as polyphosphate to utilize VFA and produce PHA in anaerobic environments. In this process orthophosphate are released in the solution. The amount of orthophosphate released in the given tank (AN2 and AN3) is of interest for the mass balance calculation, to understand how the EBPR process is working. For every mole VFA stored by PAO, one mole P is released to generate enough energy to store VFA as PHA. 1 mole P/mole COD equals 0,5 gP/g COD (Amy *et al.*, 2008). The potential release of orthophosphate is given by equation (3-58):

$$P_{rel,pot,ANn} = f_{PO4} * M_{VFA(ava,PAO,ANn)} \quad (3\text{-}58)$$

Where:

$P_{rel,pot,ANn}$	The potential P-release in AN2 or AN3 [g P/d]
f_{PO4}	Ratio P released/VFA uptake = 0,5 g P/g VFA
$M_{VFA(ava,PAO,ANn)}$	The VFA mass available for PAOs in AN2 or AN3 [g VFA/d]

The amount of VFA available ($VFA_{ava,PAO}$) for uptake by PAOs is given by equation (3-59) and (3-60).

$$M_{VFA(ava,PAO,AN2)} = (VFA_{fer,AN2} - VFA_{DO2,AN2} - VFA_{DN2,AN2}) * Q_{in,AN2} \quad (3\text{-}59)$$

$$M_{VFA(ava,PAO,AN3)} = (VFA_{fer,AN3} - VFA_{DO,AN3} - VFA_{DN,AN3}) * Q_{in,AN3} \quad (3\text{-}60)$$

Where:

$VFA_{fer,ANn}$	The VFA concentration formed through fermentation [g VFA/m^3]
$VFA_{DO2,AN2}$	VFA concentration utilized due to DO step 2 in AN2 [g VFA/m^3]
$VFA_{DN2,AN2}$	VFA concentration utilized during DN step 2 in AN2 [g VFA/m^3]
$VFA_{DO,AN3}$	VFA concentration utilized due to DO in AN3 [g VFA/m^3]
$VFA_{DN,AN3}$	VFA concentration utilized during denitrification in AN3 [g VFA/m^3]
$Q_{in,ANn}$	The flow entering AN2/AN3 [m^3/d]

If the mass of active PAOs in the return sludge is too low to consume the available VFA, the P-release will be lower than the potential P-release given by equation (3-58). So the mass of active PAOs available for P-release needs to calculated, see equation (3-61).

$$M_{A,PAO} = VSS_R * f_{A,PAO} * Q_R \quad (3\text{-}61)$$

Where:

$M_{A,PAO}$	The mass of active PAOs [g PAOs/d]
VSS_R	The total biomass concentration in the return line [g VSS/m^3]
$f_{A,PAO}$	The active PAO fraction of the total biomass

PAOs can only release the luxury uptake of phosphorus, because the residue is assimilated in the biomass. The total phosphorus content in PAOs has been set too 0,38 g P/g VSS. Normal phosphorus content in activated sludge systems, without EBPR, range between 0,01- 0,03 g P/g (Amy *et al.*, 2008). So if one assumes that phosphorus assimilated by PAOs is equal to the phosphorus content in OHOs, which is set to 0,015 g P/g VSS. The luxury uptake of phosphorus by PAOs equals to 0,365 g P/g VSS. The maximum amount of phosphorus active PAOs can release in AN2 and AN3 is given by equation (3- 62).

$$P_{rel,MAX} = M_{A,PAO} * f_{luxP} \quad (3\text{-}62)$$

Where:

$P_{rel,MAX}$	The maximum phosphorus that can be released in the anaerobic tanks [g P/d]
f_{luxP}	The luxury uptake of phosphorus [g P/g VSS] = 0,355 g P/g VSS

The actually phosphorus release is both AN2 and AN3 is then given by equation (3-63):

$$P_{rel,ac} = \min\left(P_{rel,MAX}; P_{rel,pot,tot}\right) \quad (3\text{-}63)$$

Where:

$P_{rel,ac}$	The total amount of phosphorus actually released in AN2 and AN3 [g P/d]
$P_{rel,pot,\ tot}$	The sum of $P_{rel,pot,\ AN2}$ and $P_{rel,pot,\ AN3}$ calculated based on equation (3-58) [g P/d]

The total amount of VFA actually consumed by PAOs in AN2 and AN3 needs to be calculated, which is given by equation (3-64).

$$M_{VFA,PAO} = \min\left(M_{VFA,max,PAO}; M_{VFA(ava,PAO,tot)}\right) \quad (3\text{-}64)$$

Where:

$$M_{VFA,max,PAO} = \frac{P_{rel,max}}{f_{PO4}} \quad (3\text{-}65)$$

$$M_{VFA(ava,PAO,tot)} = M_{VFA(ava,PAO,AN2)} + M_{VFA(ava,PAO,AN3)} \quad (3\text{-}66)$$

$M_{VFA,PAO}$	The actual VFA mass consumed by PAOs [g VFA/d]
$M_{VFA,\ max,PAO}$	The maximum VFA mass that can be removed by PAOs [g VFA/d]
$M_{VFA,ava,\ PAO,tot}$	The total VFA mass available to be consumed by PAOs [g VFA/d]

3.1.4.3 Aeration tank

Organic material and phosphorus are removed in the aeration tank due to aerobic degradation and uptake of phosphorus by PAO and OHO for cell growth. Transfer of ammonium to nitrate, can also occur in this reactor. The processes of interest in the aerobic tank are listed below:

1. ***COD removal***
2. ***Phosphorus removal***
3. ***Nitrification***

1. COD removal in aeration tank

COD removed in the aeration tank are either oxidized to CO_2 and H_2O or utilized for cell growth. COD incorporated in cells can be oxidized later on by endogenous respiration (Tchobanoglous *et al.*, 2014). The COD fraction first removed depends on the biodegradability of the organic matter, as is the case for all the biological treatment processes. The preferred degradation order in both anaerobic and aerobic degradation of organic material are: (1) rbCOD, (2) ssbCOD and (3) psbCOD. For COD removal in aeration tank it is assumed that only the bsCOD is removed, meaning particulate biodegradable COD will not be included in the calculation in the model.

Mass balance calculations: COD removal in the aeration tank

COD mass balance

The equations used to calculate COD removal in the aeration tank has it basis from the approach presented by Tchobanoglous *et al.* (2014). The COD mass balance in the aeration tank is given by equation (3-67):

$$\frac{dS}{dt}V_{tot,aer} = Q_{in,aer}bsCOD_{in,aer} - Q_{in,aer}bsCOD_{eff,aer} - r_{su}V_{tot,aer} \quad (3\text{-}67)$$

Where:

$\frac{dS}{dt}$	Change in substrate concentration in the reactor [g COD/m^3*d]
$bsCOD_{in,aer}$	bsCOD concentration entering the aeration tank [g COD/m^3]
$bsCOD_{eff,aer}$	bsCOD concentration leaving the aeration tank [g COD/m^3]
$Q_{in,aer}$	Flowrate entering the aeration tank [m^3/d]
r_{su}	Substrate utilization rate [g COD/m^3*d]

$V_{tot,aer}$ Total volume of all three aeration reaactors [m^3]

The substrate utilization rate, r_{su}, is conducted from equation (2-5) and (2-7) which results in equation (3-68).

$$r_{su} = \frac{\mu_{max}}{Y_{OHO}}\left(\frac{bsCOD_{eff,aer}}{K_s + bsCOD_{eff,aer}}\right)X \tag{3-68}$$

The biomass concentration in the reactors (X) can be found from the ***biomass mass balance*** of the whole system, see equation (3-69) and Figure 6. In the biomass mass balance AN1 is not included in the calculations, because AN1 differs from the other reactors in biomass concentration.

$$\frac{dX}{dt}V_{tot} = Q_{in}X_{in} - Q_{eff}X_{eff} - Q_wX_R + r_{net}V_{tot} \tag{3-69}$$

Where:

V_{tot} The total volume of all three AN2, AN3 and Aer reactors [m^3]

Q_{in} Flowrate entering process lines [m^3/d]

X_{in} Biomass concentration entering process lines [g VSS/m^3]

Q_{eff} The effluent flowrate from sedimentation basin ($Q_{eff} = Q-Q_w$) [m^3/d]

X_R Biomass concentration in return line [g VSS/m^3]

X_{eff} Biomass concentration in effluent [g VSS/m^3]

r_{net} Net rate of biomass production [g VSS/m^3*d]

Assuming steady state ($\frac{dX}{dt} = 0$), and that the influent biomass concentration can be neglected. In addition to this implement equation (2-8) for r_{net}, equation (3-70) can then be simplified to:

$$\frac{Q_{eff}X_{eff} + Q_wX_R}{V_{tot}X} = \frac{Yr_{su}}{X} - k_{d,OHO} \tag{3-70}$$

The left term in equation (3-70) equals to the inverse of the solid retention time (SRT), see equation (2-10). r_{su} can also be calculated as the amount of substrate in the reactor divided by the reactor volume which gives equation (3-71).

$$r_{su} = \frac{Q_{in}(bsCOD_{in} - bsCOD_{eff})}{V_{tot}} \quad (3\text{-}71)$$

Combining equation (3-70) and (3-71) gives the following result:

$$\frac{1}{SRT} = \frac{YQ_{in}(bsCOD_{in} - bsCOD_{eff})}{XV_{tot}} - k_{d,OHO} \quad (3\text{-}72)$$

Solving equation (3-72) with regard to X, and implementing the hydraulic retention time (HRT) as the volume divided on the flow ($\tau = V/Q$), gives:

$$X = \frac{SRT}{\tau}\left[\frac{Y(bsCOD_{in} - bsCOD_{eff})}{1 + k_{d,OHO}SRT}\right] \quad (3\text{-}73)$$

Now, back to the ***COD mass balance***, equation (3-67). Also here steady state is assumed ($\frac{dS}{dt} = 0$). If r_{su} is set equal to equation (3-68) the following equation can be formed:

$$Q(bsCOD_{in} - bsCOD_{eff}) - \frac{\mu_{max}}{Y}\left(\frac{bsCOD_{eff}}{K_s + bsCOD_{eff}}\right)XV_{tot} = 0 \quad (3\text{-}74)$$

If equation (3-74) is rearranged and the biomass concentration in the reactor (X) is set equal to equation (3-73) the results is:

$$bsCOD_{in} - bsCOD_{eff} = \frac{\tau * \mu_{max}}{Y}\left(\frac{bsCOD_{eff}}{K_S + bsCOD_{eff}}\right) * \left(\frac{SRT}{\tau}\right)\left[\frac{Y(bsCOD_{in} - bsCOD_{eff}}{1 + k_{d,OHO}SRT}\right] \quad (3\text{-}75)$$

By rearranging equation (3-75) the equation for $bsCOD_{eff}$ can be formed, see equation (3-76).

$$bsCOD_{eff} = \frac{K_s[1 + k_{d,OHO}SRT]}{SRT(\mu_{max} - k_{d,OHO}) - 1} \quad (3\text{-}76)$$

Equation (3-76) includes all COD removed in the aeration tank, both COD oxidized and COD incorporated in cells. From equation (3-76) one can see that the effluent bsCOD concentration

depends on SRT and kinetics for growth and endogenous respiration. The influent bsCOD concentration will not affect the effluent bcCOD concentration. All parameters in equation (3-76) are fixed parameters, selected in advance, with the exception of SRT which is calculated in the mass balance model based on equation (2-11). Typical kinetics constants for COD removal in the aeration tank are given in Table 13, collected from Tchobanoglous *et al.*(2014). If the biodegradable soluble COD entering the aeration tank is lower than the calculated bsCOD effluent, then the $bsCOD_{eff,aer}$ will be set equal to bsCOD entering the reactor in the mass balance model.

Table 13: Typical kinetics constants for growth and endogenous respiration for COD removal (Tchobanoglous et al., 2014, p. 755)

Parameter	Typical values at 20°C	θ-value	Values at 15°C
$\mu_{max,COD}$	6 d^{-1}	1,07	4,3 d^{-1}
$K_{s,COD}$	8 g COD/m^3	1	8 g COD/m^3
Y_{OHO}	0,45 g VSS/ g COD_{ox}		0,45 g VSS/g COD_{ox}
$k_{d,OHO}$	0,12 d^{-1}	1,04	0,098 d^{-1}

2. Phosphorus removal

There are mainly four different processes to take into account when it comes to phosphorus removal in the activated sludge system: (1) Phosphorus is removed from the solution by PAOs (2) and by OHOs. In addition to phosphorus is removed due to (3) endogenous respiration (4), and accumulation of influent non-biodegradable particulate mass in the system. The equations used for phosphorus removal have its basis from the example presented in table 7.6 by Amy *et al.* (2008). The phosphorus removed is orthophosphate. The phosphorus removed is transformed from liquid to solid form.

Phosphorus removed by PAOs

In the aeration tank active PAOs will oxidize stored PHA, which generates the energy needed for uptake of phosphorus. The main difference between OHOs and PAOs when it comes to P-removal is the fraction of phosphorus in the active biomass, where PAOs have a higher P fraction than OHOs. The P fraction in active PAOs has been set to 0,38 g P/g VSS. The phosphorus concentration released in AN2 and AN3 will be removed by "the same" PAOs that released them. The phosphorus removed in the aeration tank by PAOs is related to the amount of new PAOs generated in the aeration tank and the "old" PAOs that released phosphorus in

AN2 and AN3. The net new biomass produced in the aeration tank is related to the amount of VFA consumed by PAOs in AN2 and AN3, because the active PAOs that consumed VFA in the anaerobic reactors will be able to reproduce. The total phosphorus removed by PAO is given by equation (3-77).

$$P_{r,PAO} = \left(\frac{f_{PAO} * M_{PAO,new}}{Q_{in,aer}}\right) + \left(\frac{P_{rel,ac}}{Q_{in,AN2}}\right) \quad (3\text{-}77)$$

Where:

$P_{r,PAO}$	Phosphorus removed by PAOs [g P/m^3]
f_{PAO}	Fraction of P in active PAOs [g P/g VSS] = 0,38 g P/g VSS
$M_{PAO,new}$	The production of new active mass of PAOs [g VSS/d]
$P_{rel,ac}$	The phosphorus concentration released in AN2 and AN3 [g P/d]
$Q_{in,AN2}$	The flow entering AN2/AN3 [m^3/d]
$Q_{in,aer}$	The flow entering the aeration tank [m^3/d]

The net new active mass of PAOs is calculated based equation (3-78).

$$M_{PAO,new} = Y_{obs,PAO} * \left(\frac{M_{VFA,PAO}}{Q_{in,aer}}\right) \quad (3\text{-}78)$$

By using observed yield instead of synthetic yield in equation (3-61), the reduction in the active PAOs mass due to endogenous respiration is included. The VFA concentration utilized by PAOs in AN2 or AN3 is given by equation (3-64). The general formula to calculate the observed yield are given by equation (3-79), which is formed based on equation (2-8), (2-9), (3-71) and (3-73):

$$Y_{OBS,i} = \frac{Y_i}{1 + k_{d,n} * SRT} \quad (3\text{-}79)$$

Where:

$Y_{OBS,i}$	The observed yield for microorganism i [g VSS/g COD$_r$]
$k_{d,i}$	The endogenous decay coefficient for microorganism i [d^{-1}]
Y_i	The synthesis yield for microorganisms i [g VSS/COD]

Phosphorus removed by OHOs

OHOs also needs phosphorus for cell growth, but in much smaller degree than PAOs. The fraction of P for active OHOs is set to 0,015 g P/ g VSS. Almost the same approach used to find phosphorus removed by PAOs has been used for OHOs.

$$P_{r,OHO} = \frac{f_{OHO} * M_{OHO,new}}{Q_{in,aer}} \quad (3\text{-}80)$$

$P_{r,OHO}$ Phosphorus removed by OHOs [g P/m^3]

f_{OHO} Fraction of P in active OHOs [g P/g VSS] = 0,015 g P/g VSS

$M_{OHO,new}$ The production of new active mass of OHOs [g VSS/d]

The active mass of OHOs that will remove P is calculated based on equation (3-81):

$$M_{OHO,new} = Y_{obs,OHO} * [(bCOD_{in} * Q_{in}) - M_{VFA,PAO}] \quad (3\text{-}81)$$

The observed yield for OHOs is calculated based on equation (3-79).

Where:

$bCOD_{in}$ The biodegradable COD concentration entering the process lines [g/m^3] ($bCOD_{in} = rbCOD_{in} + ssbCOD_{in} + psbCOD_{in}$)

Phosphorus removed due to endogenous respiration

Endogenous respiration is the loss of biomass due to death and predation, this process will leave a non-biodegradable residue in the solution. The residue from endogenous respiration can contribute to phosphorus removal, and needs to be taking into account in the mass balance. The endogenous residue mass of interest originates from PAOs and OHOs, and is calculated based on equation (3-83) and (3-84). The phosphorus removed from the system is calculated based on equation (3-82).

$$P_{r,end} = \frac{f_{P,end} * (M_{end,PAO} + M_{end,OHO})}{Q_{in,aer}} \quad (3\text{-}82)$$

Where:

$P_{r,end}$ Phosphorus removed due to endogenous residue mass [g P/m^3]

$f_{P,end}$ Fraction of non-biodegradable mass that is phosphorus [g P/g VSS]

$M_{end,PAO}$ Mass of endogenous residue in the system due to PAO [g VSS]

$M_{end,OHO}$ Mass of endogenous residue in the system due to OHO [g VSS]

$$M_{end,PAO} = f_{end,PAO} * k_{d,PAO} * M_{PAO,new} \quad (3\text{-}83)$$

$$M_{end,OHO} = f_{end,OHO} * k_{d,OHO} * M_{OHO,new} \quad (3\text{-}84)$$

Where:

$f_{end,PAO}$ Fraction of endogenous particulate residue of PAOs = 0,2 (Amy *et al.*, 2008)

$f_{end,OHO}$ Fraction of endogenous particulate residue of OHOs = 0,25 (Amy *et al.*, 2008)

Phosphorus removed due to influent pnbCOD

Due to the return sludge line there will be an accumulation of particulate non-biodegradable organic material in the system, the only escape is through the waste sludge line and the effluent of the treatment plant. Since it is important to maintain a high biomass concentration in activated sludge system the highest pnbCOD mass will follow the return sludge line. This pnbCOD mass also contributes to removal of phosphorus from the solution.

$$P_{r,inert} = \frac{f_{p,inert} * M_{inert}}{Q_{in}} \quad (3\text{-}85)$$

Where:

$P_{r,inert}$ Phosphorus removal due to influent inert mass [g P/m^3]

$f_{p,inert}$ The phosphorus fraction of the non-biodegradable mass [g P/g VSS]

M_{inert} The non-biodegradable mass in the system originated from the influent [g VSS]

$$M_{inert} = \frac{pnbCOD_{in} * Q_{in}}{f_{cv}} \quad (3\text{-}86)$$

Where:

$pnbCOD_{in}$ The pnbCOD concentration entering the biological treatment [g/m^3]

f_{cv} pCOD/VSS ration of the sludge = 1,48 g COD/g VSS

It is not possible to remove more phosphorus than what is available, so the phosphorus actually removed are given by the following equation.

$$P_{r,ac} = \min (P_{r,tot}; PO4_{,in,aer}) \quad (3\text{-}87)$$

Where:

$P_{r,ac}$	The amount of soluble phosphorus actually removed from the solution [g P/m^3]
$P_{r,tot}$	The total phosphorus concentration that can be removed from the solution $P_{r,tot} = P_{r,PAO} + P_{r,OHO} + P_{r,end} + P_{r,inert}$
$PO4_{in,aer}$	The influent Orthophosphate concentration entering aeration tank [g P/m^3]

3. Nitrification in aeration tank

Nitrification is the conversion of ammonium to nitrate by nitrifiers (NIT) with the presence of oxygen. For denitrification to happen in the anaerobic reactors, nitrification needs to occur in the aeration tank. It is recommended to have a long sludge age (SRT) for biological treatment plant with nitrogen removal. If SRT is equal or lower than the minimum SRT nitrifiers is washed out of the system, and nitrification will not occur (Amy *et al.*, 2008). Nitrogen can be completely removed from the solution through nitrification/denitrification, and soluble nitrogen can be transformed to particulate nitrogen by assimilation in new biomass. Influent soluble non-biodegradable nitrogen will remain unchanged throughout the whole treatment plant. When SRT is less or equal to the minimum SRT the effluent ammonia nitrogen concentration ($NH4_{eff}$) is equal to the ammonia nitrogen concentration entering the aeration tank ($NH4_{in,aer}$). Equation (3-88) shows how SRT_{min} is calculated; this equation has its basis from equation (5-12) from Amy *et al.* (2008).

$$SRT_{min,NIT} = \frac{1}{\left(1 + \frac{K_{S,NIT}}{NH4_{in,aer}}\right)\mu_{Max,NIT} - k_{d,NIT}} \quad (3\text{-}88)$$

Oxidation of ammonium is the limiting process in the two-step biological conversion of ammonium to nitrate. The amount of ammonium oxidized by NITs is direct linked to the amount of nitrate in the aeration tank. When steady state is assumed NITs can be seen as catalyzer for the nitrification process, and the nitrogen stored by NIT can then be neglected (Amy *et al.*, 2008), meaning that all ammonium oxidized will be converted to nitrate.

Nitrogen removed by biomass (VSS):

The biomass (VSS) in a biological reactor contains of active biomass, endogenous residue and non-biodegradable particulate organics which all comprise of nitrogen and phosphorus. The amount of influent ammonia nitrogen removed from the system is assumed to be removed

mainly by PAOs and OHOs, due to the cell growth. The equation used in the mass balance model to find assimilated nitrogen is given by equation (3-89) and the dependency presented in Table 14.

$$N_{sludge} = \frac{f_n * (M_{OHO,new} + M_{PAO,new})}{Q_{in,aer}} \quad (3\text{-}89)$$

Table 14: Nitrogen incorporated in biomass is depended on NH4 available

If	Then
$N_{sludge} \geq NH4_{in,aer}$	$N_{sludge} = NH4_{in,aer}$

Where:

N_{sludge} Nitrogen assimilated in new biomass [g N/m^3]

f_n Nitrogen content in new biomass [g N/m^3] = 0,1 g N/g VSS (Amy *et al.*, 2008)

Mass balance calculations: Nitrification

Nitrifiers (NIT) mass balance

$$M_{NIT} = \begin{pmatrix}\text{Mass of}\\ \text{new NIT}\end{pmatrix} - \begin{pmatrix}\text{Redused NIT mass}\\ \text{due to endogenous decay}\end{pmatrix} - \begin{pmatrix}\text{Mass of NIT}\\ \text{in waste flow}\end{pmatrix} \quad (3\text{-}90)$$

Where

M_{NIT} The change in NIT mass in the activated sludge system [g VSS]

Equation (3-90) can be written with symbols and parameters as following:

$$M_{NIT} = \left[\left(\frac{\mu_{max,NIT} * NH_{4,eff}}{K_{s,NIT} + NH_{4,eff}}\right) X_{NIT} V * \tau\right] - \left[k_{d,NIT} X_{NIT} V * \tau\right] - \left[X_{NIT} Q_w \tau\right] \quad (3\text{-}91)$$

Equation (3-92) can be conducted from equation (3-91), which has its basis from calculations performed by Amy et al. (2008).

$$NH4_{eff,aer} = \frac{K_{s,NIT}(k_{d,NIT} + \frac{1}{SRT})}{\mu_{max,NIT} - k_{d,NIT} - \frac{1}{SRT}} \quad (3\text{-}92)$$

Where:

$NH4_{eff}$	The effluent ammonium concentration [g NH_4/m^3]
$K_{S,NIT}$	Half saturation constant for ammonium [g NH_4/m^3]
$k_{d,NIT}$	Endogenous decay coefficient for nitrifiers [d^{-1}]
$\mu_{max,NIT}$	Maximum specific growth rate for nitrifiers [d^{-1}]
SRT	Solid retention time [d]

Table 15:Kinetics values selected for autotrophic nitrifers

Parameters	**Value at 20 °C**	**θ**	**Value at 15 °C**	**Reference**
$K_{S,NIT}$	1 g NH4/m^3	1,123	0,56 g NH4/m^3	(Amy *et al.*, 2008)
$k_{d,NIT}$	0,04 d^{-1}	1,029	0,035 d^{-1}	(Amy *et al.*, 2008)
$\mu_{max,NIT}$	0,7 d^{-1}	1,072	0,5 d^{-1}	(Tchobanoglous *et al.*, 2014)

As one can see from equation (3-92) the effluent ammonia nitrogen concentration is only depended on kinetics of NITs, and SRT. For a given sludge age and kinetics values the effluent concentration will remain the same at all times. Table 15 gives the kinetics properties used for NITs in the mass balance model. With e.g a sludge age of 10 days, the effluent ammonium concentration is calculated to be 0,21 g NH4/m^3. The ammonia nitrogen concentration entering the treatment plant is 19,1 g/m^3, see Table 7, which means that 98,9 % of influent ammonium is either assimilated in biomass or converted to nitrate. 98,9 % seems to be unlikely high, since IVAR SNJ promotes the EBPR-process. The dissolved oxygen in the aeration tank is first utilized by PAOs and OHOs, before NITs can convert ammonium to nitrate. In the mass balance calculations it is assumed that 70 % of the ammonia nitrogen available after cell growth is converted to nitrate if the sludge age is higher than SRT_{min}. This percentage is selected based on Norwegian regulations for nitrogen removal (Miljødepartementet, 2004). It can be argued that also 70 % is too high, but more realistic than 98,8 %. The effluent ammonia nitrogen concentration in the mass balance is then calculated based on equation (3-93).

$$NH4_{eff,aer} = \text{if}\,[SRT = SRT_{min}; NH4_{in,aer} - N_{sludge}; (NH4_{in,aer} - N_{sludge}) * 0{,}3 \qquad (3\text{-}93)$$

The ammonium concentration available for nitrification is then given by equation (3-94):

$$NH4_{ava,NIT} = NH4_{in,aer} - N_{sludge} - NH4_{eff,aer} \qquad (3\text{-}94)$$

Where:

$NH4_{ava,NIT}$	The ammonia nitrogen concentration available for nitrification [g NH4/m^3]
$NH4_{in,aer}$	Ammonia nitrogen concentration entering the aeration tank [g NH4/m^3]
$NO3_{in,aer}$	The nitrate concentration entering the aeration tank [g NO_3/m^3]

The effluent nitrate concentration, $NO3_{eff,aer}$, is given by equation (3-95). It is assumed that all ammonia nitrogen available for nitrification will be converted to nitrate, since NITs is assumed to only catalyze the nitrification process and because oxidation of ammonia nitrogen is the limiting process in nitrification.

$$NO3_{eff,aer} = (NO3_{in,aer} + NH4_{ava,NIT}) \tag{3-95}$$

3.1.5 Sedimentation basins (Clarifier)

All solids in a EBPR process based on activated sludge will become settleable solids, due to the efficient bio-flocculation of organic activated sludge mass (Amy *et al.*, 2008). Which means that all solids in theory can be removed by settling. In the mass balance calculations it is assumed a solid capture of 99 %, meaning that 1 % of the solids will follow the effluent flow, see equation (3-99). The masses removed from the system are either removed by the effluent or the waste sludge line. Figure 6 gives a schematic overview over the activated sludge system at IVAR SNJ. All mass flow need to correspond to each other, and all formulas for the flowrates including the mass of biomass in the given flows is given below.

$$Q_{eff,sed} = Q_{in} - Q_w \tag{3-96}$$

$$Q_w = \frac{M_{VSS,w}}{VSS_R} \tag{3-97}$$

$$Q_{R,Out} = Q_{in,sed} - Q_{eff,sed} \tag{3-98}$$

$$M_{VSS,eff} = M_{VSS,in,sed} * (1 - f_{sc,sed}) \tag{3-99}$$

$$SP_{net} = \left(M_{VSS,in,sed} - M_{VSS,in,AN2}\right) + M_{VSS,in} - M_{hyd,AN1} = M_{VSS,eff} + M_{VSS,w} \tag{3-100}$$

$$M_{VSS,w} = SP_{net} - M_{VSS,eff} \tag{3-101}$$

$$M_{hyd,AN1} = M_{VSS,in,AN1} - M_{VSS,eff,AN1} \quad (3\text{-}102)$$

Where:

$M_{VSS,eff}$	Mass of VSS in effluent [g/d]
$M_{VSS,eff,AN1}$	Mass of VSS leaving the first anaerobic reactor [g/d]
$M_{hyd,AN1}$	Mass of VSS reduced in AN1 due to hydrolysis [g/d]
$M_{VSS,in,AN1}$	Mass of VSS entering the first anaerobic reactor [g/d]
$M_{VSS,in,AN2}$	Mass of VSS entering the second anaerobic reactor [g/d]
$M_{VSS,in}$	Mass of VSS entering the process lines [g/d]
$M_{VSS,in,sed}$	Mass of VSS entering settling tank [g/d]
$M_{VSS,w}$	Mass of VSS in the waste sludge line [g/d]
$Q_{eff,sed}$	Flowrate leaving the sedimentation basins [m^3/d]
Q_{in}	Influent flowrate entering process-lines [m^3/d]
$Q_{R,out}$	Return flowrate in the output [m^3/d]
Q_w	Waste flowrate [m^3/d]
SP_{net}	Net sludge production, removed from the system by the effluent and waste line [g/d]
VSS_R	VSS concentration in the return/waste sludge line [g/m^3]

The input to AN1 has been held constant in the model, with the exception of the soluble fractions, which has been set equal to the output return parameter. This result in a small difference between particulate material in the input and output return sludge line, especially particulate COD and Tot-P, which is a weakness with the model. To be able to control the settling properties in the clarifier, and thereby the solid concentration in the return sludge line this was best solution to the problem, because if the input and output is set equal to each other there was to many dependencies in the system, which made the model “crash”. The relationship has been focused on the biomass (VSS), where a given TS percentage in the return sludge line has been set. For the normal situation the solid concentration in the return sludge line is to 2 %. The biomass in the waste sludge line has been determent based on the amount of biomass in the effluent and the net sludge production, see equation (3-100). Where the net sludge production is equal to the new biomass generated in the reactors summarized with the VSS entering the process lines from the raw wastewater, minus the decrease in VSS due to hydrolysis in AN1. Several of the parameters in these equations are depended on each other, so manual calculations and iterations had been used to make this calculation possible.

3.2 Sludge treatment line

The sludge treatment line at IVAR SNJ consist of several processes that reduce the sludge volume by removing water, the anaerobic digester which stabilize the sludge, and theremal drying that sanitized and dewaters the sludge. All relevant calculation and assumption made will be explained under each given process.

3.2.1 Storage tanks

The storage tanks main function is to store the sludge before it is transported to the next process, one assumption made is that no biological processes will occur in the storage tanks. This will actually not be the case in reality due to long hydraulic retention time, but the sludge stored is not complexly mixed, which will result in very complex calculations, and the effect on the different parameters is thereby difficult to find. Hence, the storage tanks have not been implemented to perform any treatment process. This thesis will assume that the inlet concentration is equal to effluent concentration for the different storage tanks.

3.2.2 Thickener

Thickening the sludge before it is feed to the digester is beneficial, because it reduces both the biomass volume and the digester volume (Turoviskiy and Mathai, 2006). Normally the primary sludge have higher solid content than the biological sludge, because the flocs generated in the activated sludge system is large and light, while the solids removed by the primary treatment are denser. Thickener 1 thickens primary sludge in the mass balance calculation, and thickner 2 biological sludge. The sludge flow leaving the given thickener is given by the equation (3-103).

$$Q_{eff,THn} = \frac{M_{cake,THn}}{S_{SL,THn} * \rho_w * P_{S,THn}} \tag{3-103}$$

Where:

$Q_{eff,THn}$	The sludge flow leaving thickener n [m^3/d]
$M_{cake,THn}$	The solid cake formed during thickening [kg/d]
	$M_{Cake} = M_{TSS,THn} * f_{SC,THn}$
$M_{TSS,THn}$	The mass of TSS entering thickener n [kg/d]
$S_{SL,THn}$	Specific gravity of the sludge leaving thickener n
ρ_w	The density of water = 1000 kg/m^3

$P_{S,THn}$ Fraction of solid in thickened sludge in thickener n

The specific gravity of sludge (S_{SL}) and the fraction of solid (P_{SmTHn}) after thickening will differ for primary and biological sludge. The specific gravity of sludge needs to be calculated based on the specific gravity of solid, see equation (3-104) and (3-105). Specific gravity of soldis before and after the thickening process is equal, because the VSS and FSS fraction of TSS will not change. The specific gravity solids (S_S) is calculated based on the influent FSS and VSS as shown in equation (3-104), which has its basis from equation (13-1) from Tchobanoglous *et al.* (2014).

$$S_S = \frac{1}{\left(\frac{\left(\frac{FSS}{TSS}\right)}{S_f}\right) + \left(\frac{\left(\frac{VSS}{TSS}\right)}{S_V}\right)} \quad (3\text{-}104)$$

Where:

S_S Specific gravity of solids

$\left(\frac{FSS}{TSS}\right)$ The fixed fraction of TSS

$\left(\frac{VSS}{TSS}\right)$ The volatile fraction of TSS

S_f Specific gravity of fixed solids = 2,5 (Tchobanoglous *et al.*, 2014)

S_v Specific gravity of volatile solids = 1 (Tchobanoglous *et al.*, 2014)

The specific gravity of the sludge will change due to the thickening process, to calculate the sludge volume leaving the thickener the specific gravity of the thickened sludge needs to be calculated, see equation (3-105).

$$S_{SL} = \frac{1}{\left(\left(\frac{P_{s,THn}}{S_S}\right) + \left(\frac{P_{w,THn}}{\rho_w}\right)\right)} \quad (3\text{-}105)$$

Where:

S_{SL} Specific gravity of sludge

ρ_w Density of water = 1

$P_{w,THn}$ Water fraction of thickened sludge in thickener n

The cationic organic polymer added before flocculation is assumed to affect TSS, VSS, tpCOD and psbCOD, because the polymer is biodegradable. The contribution is minimal compared to the high solid concentrations, and could probably be neglected. The polymer is added to achieve the desired solid capture. Typical performance data for a rotary drum thickener is given by Table 16.

Table 16: Typical performance data for RDT (Turoviskiy and Mathai, 2006)

Sludge type	Feed solid	Water removed	Thickened solids	Solids capture
Primary	3-6 %	40-75 %	7-9 %	93-98 %
WAS	0,5-1,0 %	70-90 %	4-9 %	93-99 %

The total reject from TH1 and TH2 is sent back to the treatment plant, right before the sand and grease trap. This will have a small effect on the masses entering the sand and grease trap, and the masses entering the process lines. The flowrate from the reject is very small compared with the flow treated at IVAR SNJ, so the mass contribution is quite small. Same as with the return flow in the activated sludge system, the reject has been included in the model by iteration and manual calculations.

3.2.3 Anaerobic digestion (AD)

Methanogens are sensitive to changes in pH and temperature, and can easily be inhibited even at small changes (Turoviskiy and Mathai, 2006). This will affect the methane production, and can in worst case result in collapse and no methane production. It is assumed that the temperature in the digester is 35 °C at all times, but the pH will depend on the ammonia nitrogen solubilized in the digester. If the ammonia nitrogen concentration in the digestion reach a threshold value it will result in total inhibition of methanogens microorganisms. This mass balance assume a continuous sludge feeding, and that methanogens will not have the opportunity to recover if the ammonia nitrogen concentration exceed the threshold value. In other words it is assumed maximum methane production, or no methane production. The biological sludge wasted from the EBPR process contains high levels of phosphorus, due to the high fractions of PAOs. About 55 % of the influent VSS is destructed in the anaerobic digesters at IVAR SNJ. PAOs and OHOs are both obligate aerobes, which mean they can only meet the energy need in aerobic environments. Due to exposure to anaerobic conditions over a long period of time it is assumed that the net VFA consumption by OHOs or PAOs equals to zero. It is also assumed that the effluent nitrate and dissolved oxygen concentration equals to zero,

and that all luxury uptake of phosphorus will be released in the digester, and that all COD utilized for these processes will be available for biogas production.

Release of nitrogen and phosphorus in anaerobic digestion

Release of assimilated nitrogen and phosphorus

Assimilated nitrogen and phosphorus will be released in the digester due to the destruction of biomass. The effluent soluble P and N concentration is thereby very high compared to the influent soluble concentrations.

Assimilated nitrogen

Ammonia nitrogen exists either as ammonium ions (NH_4^+) or as ammonia gas (NH_3) in a water solution, depending on the solutions pH and temperature (Tchobanoglous *et al.*, 2014). Consequently NH_4^+ have been used in this mass balance, because at pH 7-8 ammonium ions is dominant, see figure 2-15 presented by Tchobanoglous *et al.* (2014). Generally microorganisms can grow in the pH-range from 5-9, where neutral pH is the most optimal pH (Ødegaard *et al.*, 2014). The amount of ammonia nitrogen released in the digester is given by equation (3-106).

$$\mathrm{NH_{4,rel,AD}} = \mathrm{VSS_{des}} * \mathrm{f_{n,AD}} \tag{3-106}$$

Where:

$NH_{4,rel,AD}$	The release of ammonia nitrogen in AD [g NH_4/m^3]
VSS_{des}	Destructed biomass (VSS) [g VSS/m^3]
$f_{n,AD}$	The nitrogen content in the biomass entering AD [g N/ g VSS]

The nitrogen content in the biomass entering AD had been calculated as shown in equation (3-107), since both primary and biological sludge are input in AD the nitrogen content in the biomass will not be equal as the first assumed in the activated sludge system.

$$\mathrm{f_{n,AD}} = \frac{(\mathrm{TotN_{in,AD}} - \mathrm{NH4_{in,AD}} - \mathrm{NO3_{in,AD}})}{\mathrm{VSS_{in,AD}}} \tag{3-107}$$

Where:

$TotN_{in,AD}$	Total nitrogen concentration entering AD [g N/m^3]
$NH4_{in,AD}$	Ammonia nitrogen concentration entering AD [g $NH4/m^3$]
$NO3_{in,AD}$	Nitrate concentration entering AD [g $NO3/m^3$]

$VSS_{in,AD}$ VSS concentration entering AD [g VSS/m^3]

A optimal total ammonia nitrogen (TAN) concentration is needed to ensure growth, but if the TAN concentration exceeds a threshold value it will result in direct inhibition of microbial activity (Rajagopal, Massé and Singh, 2013). The threshold value for total inhibition due to TAN concentrations will vary from digester to digester based on how they are operated. It is difficult to separate inhibition due to high TAN concentration and due to organic loading rate, because the two parameters are related (Moestedt *et al.*, 2016). Rajagopal, Massé and Singh (2013) reviewed several studies on inhibition due to TAN concentration, and with this as a basis the ammonia nitrogen threshold value used in this mass balance model has been set to 3000 g NH_4/m^3. If the TAN concentration exceeds this values all methanogenesis will be inhibited, and there will not be any biogas production in the digester.

Assimilated phosphorus

Assimilated phosphorus will also be release due to destruction of biomass in the digester, this release is calculated based on equation (3-108).

$$PO4_{rel,ass,AD} = VSS_{des} * f_{P,ass} \quad (3\text{-}108)$$

Where:

$PO4_{rel,ass,\ AD}$ The release of assimilated phosphorus in AD [g P/m^3]

$f_{P,ass}$ The phosphorus content assimilated in biomass = 0,015

Release of luxury phosphorus

In addition to the release of assimilated phosphorus, PAOs will release the luxury uptake of phosphorus. To calculate the amount of phosphorus released due to PAOs, the mass of active PAOs needs to be calculated. The fraction of active PAOs in the waste sludge line is given by the ($f_{A,PAO}$) selected for the given scenario. It is assumed to be zero PAOs in the biomass originated from primary sludge. The mass of PAOs entering the digester is then given by equation (3-109).

$$M_{PAO,AD} = (VSS_{eff,TH2} * f_{A,PAO}) * Q_{eff,TH2} \quad (3\text{-}109)$$

Where:

$M_{PAO,AD}$ The active mass of PAOs entering the anaerobic digester [g VSS/d]

$VSS_{eff,TH2}$	The VSS concentration leaving TH2 [g VSS/m^3]
$f_{A,PAO}$	The fraction of active PAOs in the biomass = 0,15
$Q_{eff,TH2}$	The flow leaving the "biological" thickener [m^3/d]

Now the phosphorus released due to luxury uptake by active PAOs can be calculated, see equation (3-110).

$$PO4_{rel,PAO,AD} = \frac{M_{PAO,AD} * f_{luxP}}{Q_{AD}} \quad (3\text{-}110)$$

Where:

$PO4_{rel,PAO,\ AD}$	The release of luxury P in the anaerobic digester [g P/m^3]
Q_{AD}	The flow entering the anaerobic digester [m^3/d]

Biogas production

The methane production is calculated based on equation (3-111), which has its basis from example 7-10 presented by Tchobanoglous *et al.* (2014).

$$CH_{4,AD} = (COD_{CH4} * Q_{AD}) * f_{CH4} \quad (3\text{-}111)$$

Where:

$CH4_{AD}$	Methane production [l /d]
COD_{CH4}	The COD available for methane production [g/m^3]
Q_{AD}	The flow entering/leaving AD [m^3/d]
f_{CH4}	The CH_4 equivalent of converted COD [L CH_4 /g COD]

$$COD_{CH4} = VSS_{des} * f_{cv} \quad (3\text{-}112)$$

The CH_4 equivalent of converted COD (f_{CH4}) is found based on the following approach. The amount of oxygen needed to oxidize methane to carbon dioxide and water are given by the chemical reaction below.

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$$

From this reaction the amount of COD needed per mole methane produced can be found. The molar weigh of O_2 is 32 g/mole, and for each mole methane there is a need of 2 moles O_2, which

gives the value of 64 g COD/mole CH_4. The volume of gas per mole of gas at a given temperature can be calculated based on the ideal gas low, see equation (3-113).

$$V = \frac{nRT}{P} \tag{3-113}$$

Where:

V	Gas volume [L]
n	Number of mole = 1 mole
R	The gas constant = 0,082057 atm*L/mole*K
T	Temperature in kelvin = °C + 273,15 K
P	Pressure = 1 atm

At a temperature of 35 °C the volume of 1 mole gas equals to 25,3 liters. The CH_4 equivalent of converted COD at 35 °C is then 0,4 l CH_4/g COD, see equation (3-114).

$$f_{CH4} = \frac{25{,}3 \text{ liter}}{64 \frac{\text{g COD}}{\text{g CH4}}} = 0{,}4 \frac{\text{l } CH_4}{\text{g COD}} \tag{3-114}$$

The total biogas production is calculated with a basis in the amount of methane produced, where it is assumed that methane accounts for 65 % of the total biogas production.

$$\text{Biogas}_{AD} = \frac{CH_{4,AD}}{0{,}65} \tag{3-115}$$

3.2.4 Centrifugal dewatering

The anaerobically digested sludge is very fluid, due to the destruction of biomass. Since the sludge has a more fluid characteristic there is a need to reduce the sludge volume by removing free water. The same calculation approach used for the thickener is also used for the centrifugal dewatering, only with a higher solid capture and different polymer dosage. The solid capture utilized for the centrifugal dewatering mass balance calculations is 97 %. See Table 17 for typical performance data for a solid-bowl centrifuge handling a mixture of primary- and waste activated sludge (WAS) that has been through an anaerobic digestion process.

Table 17: Typical dewatering performance for solid bowl for anaerobically digested sludge (Tchobanoglous et al., 2014)

Sludge type	Feed solids [%]	Cake solids [%]	Solid capture [%]
Primary + WAS	2-4	22-35	95 +

3.2.5 Thermal drying

Indirect thermal drying is the last sludge treatment process at IVAR SNJ. Since the sludge is heated to 100 °C for about 30 minutes, some highly volatile organic compounds will evaporate, in addition to ammonia gas. It is assumed a solid capture of 99 %, and the particulate material "lost" leaves the thermal dryer as dust. Ammonium ions are in a dynamic equilibrium with ammonia gas in aqueous solutions (Maurer and Müller, 2012), see the equilibrium equation below. The amount of ammonia and ammonium in a solution is depended on the given pH and temperature (Tchobanoglous *et al.*, 2014). An increase in pH and temperature results in more ammonia (Rajagopal, Massé and Singh, 2013).

$$NH_4^+ \rightarrow NH_3 + H^+$$

There are several volatile components release from municipal sewage sludge during the drying process, where ammonia is the primary gas released (Liu *et al.*, 2015). Since the sludge already has been stabilized and reduced, with the release of biogas, in the digester the release of other volatile compounds than NH_3 in the thermal dryer has been neglected.

Ammonia emission

There are high amounts of free ammonia nitrogen in the digested sludge due to the destruction of biomass. The digested sludge is then dewatered in the centrifuge, where approximately 90 % of the free ammonia nitrogen at IVAR SNJ follows the centrate due to the reduction of water volume based on the calculations performed in mass balance model. Since the digested sludge is dewatered before dried the NH_3 emission in the thermal dryer is lower than what it possibly could be. Liu *et al.* (2015) examined the characteristics of ammonia emission during thermal drying of lime sludge and raw sludge, where the impact of temperature and time where thoroughly investigated. The raw sludge examined had not been digested, and the initial ammonia concentration was quite low. In general an increase in temperature had higher effect on the NH_3 emission than an increase in time. In the temperature range 100-130 °C water evaporated, and free ammonia was released.

In the mass balance model it is assumed that 92 % of free ammonia nitrogen entering the thermal dryer is removed by NH_3 emission. This value has been selected with a basis in experiments performed by Rabah and Darwish (2012), Pantelopoulos, Magid and Jensen (2016) and Maurer and Müller (2012). Rabah and Darwish (2012) performed a test on how initial temperature affected NH_3 emission, where a heat exchanger was used to heat the water from 22 to 54-59 °C for 15 and 9 minutes. The overall nitrogen removal was then 90,28 %. Pantelopoulos, Magid and Jensen (2016) examined NH_3 emission for digested and dewatered animal waste in a thermal dryer operating at different temperatures. The main goal was to reach a solid content of 85 % with different temperature and pH ranges and to measure the nitrogen removal. At a natural pH of 9,5 and a temperature of 100 °C the NH_3 emission was approximately 93 %. Maurer and Müller (2012) measured the ammonia nitrogen losses to be 91,7 % for dewatered digestate. The feedstock was a mixture of residues from food, feed production, pig and cattle slurry. None of the experiments on ammonia emission in a thermal dryer was performed on digested, dewatered sewage sludge. If IVAR performs test on the NH3 emission in the thermal dryer, this value can easily be changed in the model. The sludge volume leaving the thermal dryer is calculated based on the same principal as for the thickener and centrifuge.

3.3 Testing Mass-balance-model

After the mass balance calculations were performed, and the excel sheet was functional, it was of interest to test outputs from the model by changing some of the parameters. In addition to testing different scenarios, a simulation of how IVAR SNJ is operating today was performed.

3.3.1 Normal operation at IVAR SNJ

The simulation of the normal operation at IVAR SNJ was performed in two different scenarios. The inlet concentrations used as input in scenario Normal 1 were based on the average concentrations measured in the period mid-March to end of May 2018, see Table 19. For the second scenario, Normal 2, the average concentrations given in Table 7, and repeated in Table 19 were used as input in the model. All three process lines were first in operation at IVAR SNJ from mid-March 2018. Comparing results from the model with measured values at IVAR SNJ is of interest both to test the model and to reveal potential problems with today´s operation. Scenario Normal 1 was compared with measured values in the effluent, due to the recent implementation of all three process lines at IVAR SNJ.

Today IVAR SNJ operate with a solid retention time of 4-5 days, and a solid concentration in the return sludge line of approximately 2 %. SRT is calculated in the model based on the biomass concentration in AN1 and the other reactors, all reactor volumes and the total biomass removed from the system by the effluent and the waste sludge line, see equation (2-11). For the model to "match" the real system as good as possible the biomass concentration selected in the system were based on a wanted SRT of 4,5 days. The fraction of active PAOs in the return sludge line was selected in the model such that steady state was achieved. Table 18 and Table 19 presents the model-input for both scenarios, Normal 1 and Normal 2, to create a good comparison of the real operation at IVAR SNJ. By comparing the concentrations of the substances entering IVAR SNJ in Normal 1 and Normal 2 one can see that the inlet concentrations are more diluted in the second scenario, which is due to less precipitation in this period compared to 2017.

Table 18:Selected operation parameters for IVAR SNJ

Parameters	Normal 1	Normal 2
$Q_{in,TP}$ [m^3/s]	1,369	1,41
TSS_R [%]	2	2
$f_{A,PAO}$ [%]	3,83	4,17
SRT [d]	4,57	4,48
X [g VSS/m^3]	2300	1250
$TotP_{in,TP}$ [g P/m^3]	5	3,94
$PO4_{in,TP}$ [g P/m^3]	2	0,94

Table 19:Inlet concentration used in scenario Normal 1 and Normal 2

Parameters	Normal 1	Normal 2
TOT-COD	449	295
tsCOD	101	80
rbCOD	23,93	20,93
VFA	13,92	12,56
ssbCOD	39,72	35,84
snbCOD	38,08	23,22
tpCOD	348	215
psbCOD	314,6	186,1
pnbCOD	33,38	28,86
TSS	303	192
VSS	232,3	147,2
Tot-P	5	3,64
Ortho-P	2	0,94
Tot-N	24,8	24,8
NH4	19,1	19,1
NO3	0,3	0,3
O_2	1	1

3.3.2 Different scenarios

In total four main scenarios were tested in the model, and each scenario was tested for two different biomass concentrations in the reactors of 3000 and 4000 g VSS/m^3. In total eight scenarios was tested in the model. The concentrations presented in Table 19 under scenario "Normal 2" were used as input for all these scenarios, except of phosphorus. The inlet phosphorus concentration was increased to 17 g/m^3 for Tot-P and 7 g/m^3 for ortho-P. The intension by increasing the inlet phosphorus concentration to "unrealistic high" concentrations was to find other limitations in the system when phosphorus was no longer a limitation. For scenario 1 and scenario 2 the fraction of PAOs was selected based on when steady state were achieved, while for scenario 3 and 4 the fraction of active PAOs were set to 1 %, even though steady state was not achieved. For scenario 1 and 3 the solid concentration in the return sludge line is set to 2 %, while in scenario 2 and 4 it is reduced to 1 %. All input parameters for all scenarios are presented in Table 20.

Table 20: Input parameters for Scenario 1-1 to 4-2

Parameters	**1-1**	**1-2**	**2-1**	**2-2**	**3-1**	**3-2**	**4-1**	**4-2**
TSS_R [%]	2	2	1	1	2	2	1	1
$f_{A,PAO}$ [%]	31,5	31	25,5	22,5	1	1	1	1
SRT [d]	8	10,35	7	9,54	8,2	10,4	7	9,53
X [g VSS /m^3]	3000	4000	3000	4000	3000	4000	3000	4000
$TotP_{in,TP}$ [g P/m^3]	17	17	17	17	17	17	17	17
$PO4_{in,TP}$ [g P/m^3]	7	7	7	7	7	7	7	7

In,TP= entering treatment plant

4. Results and discussion

The result from the mass balance model are divided into two parts: The normal operation situation at IVAR SNJ and result from the difference scenarios.

4.1 Normal operation at IVAR SNJ

4.1.1 Wastewater treatment line

Comparison of measured and simulated effluent concentrations (Normal 1)

Outputs from scenario Normal 1 were compared with measurement performed at IVAR SNJ in the period from mid-March to end of May 2018. Only Normal 1 was compared with real data from IVAR SNJ, because this scenario is created for the same period, which gives a better basis for comparison.

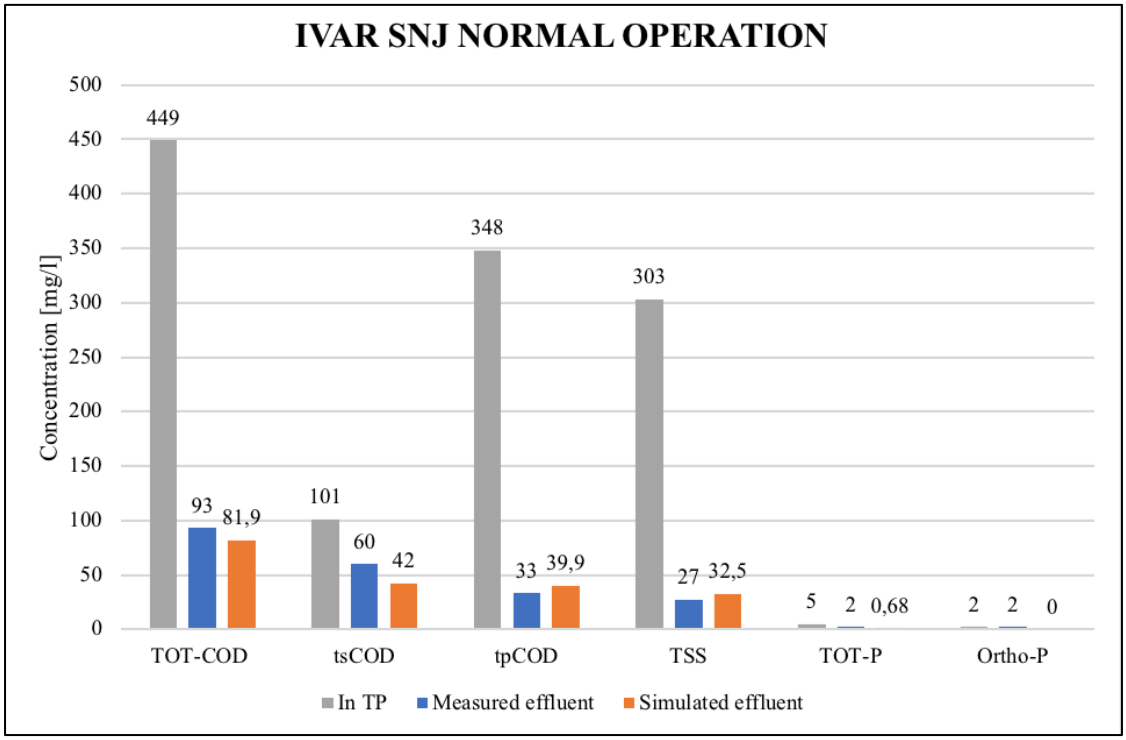

Figure 10: Comparison of measured and simulated effluent concentrations at IVAR SNJ based on scenario Normal 1

Measured (blue) and simulated (orange) concentrations in the effluent are presented in Figure 10 together with the influent concentrations (grey). From the figure one can see that simulated particulate, TSS and tpCOD, are slightly higher than the measured values. At IVAR SNJ they operate with 1500-2000 g TSS/m^3 in the reactors to maintain a sludge age around 4-5 days. While in the model, TSS in the reactors needs to be approximately 2800 g TSS/m^3 (2300 g VSS/m^3) to maintain the same sludge age. If the TSS concentration entering the clarifier is

assumed to be 2000 g TSS/m^3 at IVAR SNJ, and measured effluent is 27 g TSS/m^3, see Figure 10, the real solid capture of the clarifier can be calculated to be 98,83 %. This percentage is calculated based on flowrates collected from the model for this scenario, where the flowrate entering and leaving the clarifier were 1.595 m^3/s and 1.375 m^3/s, respectively. Based on this the solid capture selected in the model, $f_{SC,sed}$= 0,99, is actually a good "match" to the real situation at IVAR SNJ.

If TSS concentration in the reactors is set to 2000 g/m^3, X= 1660 g VSS/m^3, and all other input parameters are equal to the values given in Table 18 and Table 19, the sludge age for the system is calculated to be 3,6 days in the model. For a sludge age in this low range there is a risk for washout of PAOs (Amy *et al.*, 2008). Since the solid capture selected for the model is a good match to reality, but SRT from the model differs from the reality it might indicate that the production of new biomass is lower at IVAR SNJ than in the model, see equation (2-10). The measured orthophosphate concentration in the effluent is equal to the influent orthophosphate concentration, see Figure 10, which supports the statement of low biomass production at IVAR SNJ.

For both simulations of IVAR SNJ, phosphorus is actually the limitation for cell growth in the system. The model has not included nutrients as a possible limitation for biomass production, which is a weakness with the model. The biomass generated in the model for both scenarios is actually higher than possible, because there is not enough phosphorus to produce this amounts of biomass. This can also be the reason for a lower sludge age in the model than in reality when X is set to 1660 g VSS/m^3 in the model, but the measured influent and effluent Ortho-P at IVAR SNJ still indicates that there is a problem related to biomass production at IVAR SNJ. For scenario Normal 2 one can achieve the same removal efficiency of 100 % if the fraction of PAOs is reduced to zero. Meaning that influent ortho-P will be removed with or without PAOs. For this scenario, it would be difficult to promote growth of PAOs, or other microorganisms for that matter, due to the low influent ortho-P concentration.

Diluted wastewater entering IVAR SNJ

The wastewater used as input in scenario Normal 2 was more diluted compared to the input in scenario Normal 1, see Table 19. Table 21 presents some result for both scenarios, and the differences between them. These scenarios have been tested to see the difference in performance due to a change in influent concentrations and flow. A lower biomass

concentration, X= 1250 g VSS/m^3, is needed in scenario Normal 2 to maintain the sludge age of 4,5 days. If X is set equal to 2300 g VSS/m^3 in scenario Normal 2 the sludge age of the system is calculated to be 6,58 days. The reason is a lower sludge production in this scenario, less food is available for microorganisms in the system. More sludge is sent to sludge treatment line in scenario Normal 1 compared with Normal 2 ($M_{VSS,w}$), which result in a higher biogas production. A lower biomass concentration in the reactors result in a lower return flow in scenario Normal 2, but still less sludge is sent to sludge treatment. Diluted influent concentration result in less biomass production, and thereby a lower sludge production, which makes the biological treatment less efficient. For both scenarios the phosphorus content of the sludge is about 2 %, normally for a EBPR-process the phosphorus content should be as high as 3-6 % (Bi, Guo and Chen, 2013).

Table 21: Conditions and result for the simulation of scenario Normal 1 and Normal 2

Scenarios		**Normal 1**	**Normal 2**
Conditions	$Q_{in,TP}$ [m^3/s]	1,369	1,41
	TSS_R [%]	2	2
	$f_{A,PAO}$ [%]	3,83	4,17
	SRT [d]	4,57	4,48
	X [g VSS/m^3]	2200	1250
	$TotP_{in,TP}$ [g/m^3]	5	3,94
	Ortho-$P_{in,TP}$ [g/m^3]	2	0,94
Results	$M_{totP,in,TP}$ [kg/d]	626,2	480
	$M_{totP,in}$ [kg/d]	432,3	312
	$M_{VSS,in}$ [kg/d]	14630	9619
	$M_{totP,w}$ [kg/d]	351,7	268
	$M_{VSS,w}$ [kg/d]	13942	10111
	Q_w [m^3/d]	840	609
	Q_R [m^3/d]	18218	9402
	$M_{totP,eff}$ [kg/d]	80,6	44,5
	$M_{totP,centrate}$ [kg/d]	370	274
	$M_{NH4,centrate}$ [kg/d]	681	546
	P-content in sludge	0,02	0,02

in,TP = entering treatment plant
w= Waste sludge line
eff = Leaving sedimentation basin/Clarifier
Centrate = Reject from centrifugal dewatering

The low concentrations entering IVAR SNJ can be argued to be related to infiltration of extraneous water, and stormwater in the sewage collection system, which dilutes the incoming wastewater. From the result given in Table 21 one can see that wastewater characteristics affects the biological treatment processes. Growth of biomass is either limited by the nutrients, here phosphorus, or organic carbon concentrations. The hydraulic retention time (HRT) in each bioreactor decreases with increased inflow, which results in less soluble COD available through hydrolysis in the anaerobic reactors. The Norwegian sewage collection system consists of a combination of combined sewage systems (CSS), where both stormwater and wastewater are handled and transported in the same system, and separated sewage systems, where wastewater and stormwater are handled in two separated systems (Ødegaard *et al.*, 2014).

As an example the sewage collection system in the municipality of Stavanger consist of 47 % CSS, where approximately 20,4 million m^3 of the wastewater sent to IVAR SNJ is in fact clean water (Stavanger Kommune, 2011), either rainwater transported in the combined sewage system or inflow of extraneous water. The "clean" water transported in the wastewater collection system of Stavanger municipality constitute of approximately 62 % of all water transported in the system, where 4 mill. m^3 will not reach IVAR SNJ due to local discharge, combined sewage overflow or exfiltration, see Figure 11. For separated system the main problem connected to extraneous water is faulty connections, where stormwater from e.g. a residential area is connected to the wastewater pipe and not the stormwater pipe (Beheshti, Sægrov and Ugarelli, 2015). A reduction in infiltration/inflow of extraneous water into the sewage system would be very beneficial for the biological treatment at IVAR SNJ, resulting in higher concentration by the inlet of the treatment plant, and thereby more efficient treatment processes (Lindholm and Bjerkholt, 2011).

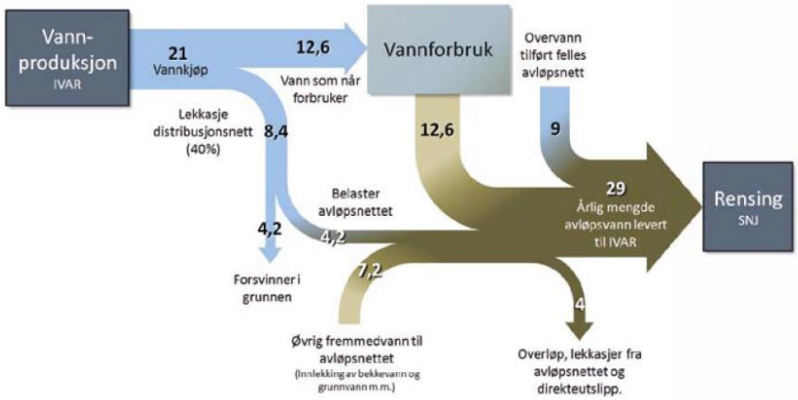

Figure 11: Water balance from the municipality of Stavanger (Stavanger Kommune, 2011).

Based on statistics, the total population in Stavanger, Sandnes, Sola, Randaberg and Gjesdal is 258 724 people (Sentralbyrå, 2018). If the average water consumption for a person is assumed to be 150 l/pe*d in Norway (Ræstad *et al.*, 2010), the inflow to IVAR SNJ should be approximately 0,45 m^3/s. For a rough comparison, the average flowrate entering IVAR SNJ used in the scenario Normal 2 is 1,41 m^3/s, which indicates that approximately 70 % of the wastewater entering IVAR SNJ is "clean water" that don´t need to be treated.

4.1.2 Sludge treatment line

The centrate is the reject from the centrifugal dewatering process, and is directly discharged into the ocean at IVAR SNJ. Tot-P discharged at IVAR SNJ is a sum of $M_{totP,eff}$ and $M_{totP,centrate}$, which is presented for both scenario in Table 21. Based on these results 72 and 66 % of Tot-P entering the treatment plant are discharged in scenario Normal 1 and Normal 2, respectively. Where 60 % and 57 % of Tot-$P_{in,TP}$ are discharged by the centrate. The largest fraction, more than 95 %, of the total phosphorus and nitrogen that follows the reject flow from the centrifugal dewatering is soluble, due to the destruction of biomass in the anaerobic digestion. Discharge of soluble nitrogen and phosphorus will increase the risk of eutrophication in the recipient, and these resources is "lost" when directly discharged without any form of recover.

One can conclude that IVAR need to change their sludge treatment line, as phosphorus is removed in the EBPR-process and later discharged in the sludge treatment line. With the existing sludge treatment line the reconstruction have worsen the situation, where less phosphorus is recovered and used in the fertilizer product, Minorga. More phosphorus is discharged, because phosphorus is incorporated in microorganisms and not chemically bound, which makes it more easily solubilized in the anaerobic digestion. There is a higher risk for struvite precipitation in the anaerobic digestion, pipes or the next processes in line. For unwanted struvite or Magnesium ammonium phosphate ($MgNH_4PO_4$) to be formed there is a need for magnesium (Parkin and Owen, 1986), which can originate from destruction of biomass in the anaerobic digestion or infiltration/inflow of seawater in the sewage collection system. Struvite can cause maintenance problems, such as clogged pipes and problems connected to heat exchangers, because struvite is difficult to remove when first attached (Parkin and Owen, 1986). To prevent unwanted struvite precipitation it would be beneficial to provoke struvite precipitation prior the anaerobic digestion.

The sludge treatment line at IVAR SNJ need to change in a short period of time, to prevent all these disadvantages. Figure 12 and Figure 13 presents two different alternatives for a new sludge treatment line based on the existing sludge treatment line, where P and N recover are included by controlled struvite precipitation.

Alternative 1: Release of luxury P prior anaerobic digestion

In the first alternative, it is suggested to use buffer tank 1 as an anaerobic mixer, to release luxury phosphorus prior the anaerobic digestion. By utilizing the existing system there is no need for a total reconstruction of the sludge line, which is more cost-efficient and space-saving. The hydraulic retention time in buffer tank 1 is normally two days, which should give enough time for VFA production through hydrolysis and fermentation. Primary sludge is more biodegradable than biological sludge (Carrere *et al.*, 2016), hence the VFA production in the anaerobic mixer would be dependent on the mass of primary sludge. The amount of luxury P released would depend on the VFA made available in the tank, one could also add external acetate in the tank to increase the P-release. One big advantage with this alternative is that unwanted struvite precipitation would be reduced due to the separation of P and N release. Assimilated phosphorus would still be released in the anaerobic digestion, but the amount of P solubilized in the digester would be drastically reduced if this system was implemented at IVAR SNJ. The struvite formed from the two different reject flows, centrate and from the anaerobic mixer, could be sent to IVARs fertilizer fabric and implemented in Minorga.

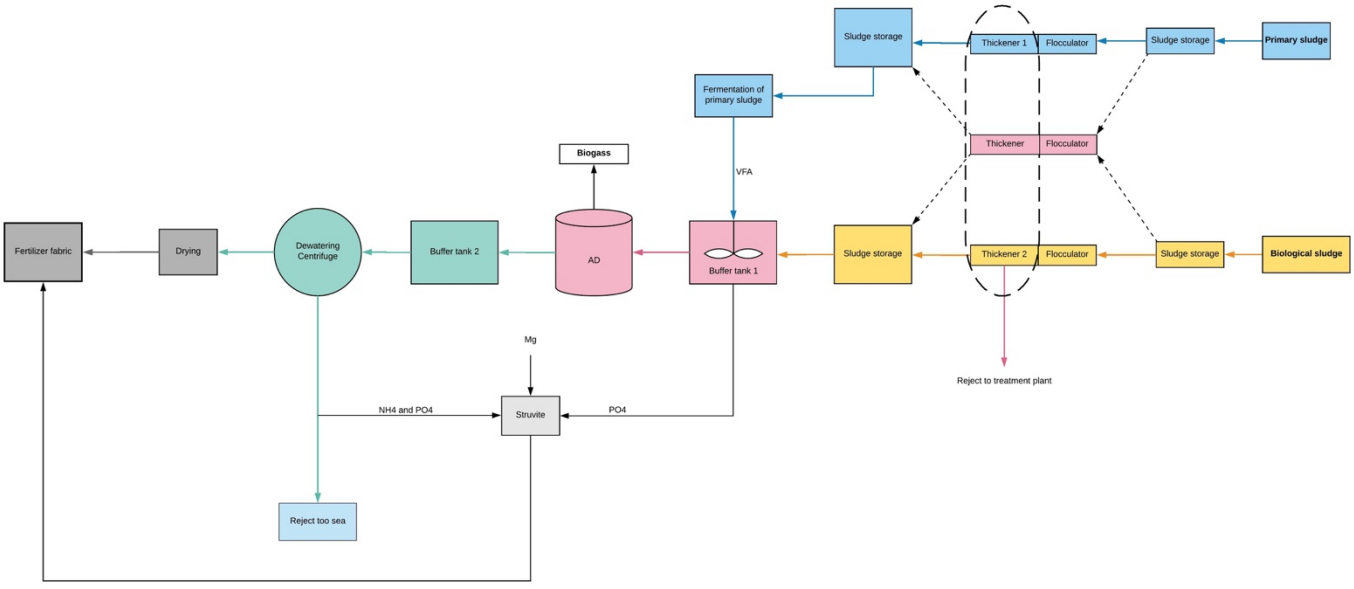

Figure 12: Alternative 1 for a new sludge treatment line

Alternative 2: Controlled struvite precipitation before discharge of centrate

The second alternative requires no change in the existing sludge treatment line, the only difference is that the centrate is collected before discharged to provoke controlled struvite precipitation. Where the struvite formed would be sent to the fertilizer fabric and used as a resource in Minorga. This alternative would improve the existing system, because phosphorus and nitrogen would be recovered and used as a resource. The main disadvantage with this alternative is that unwanted struvite precipitation would still be a problem at IVAR SNJ, especially in the pipes and processes in line after the anaerobic digestion.

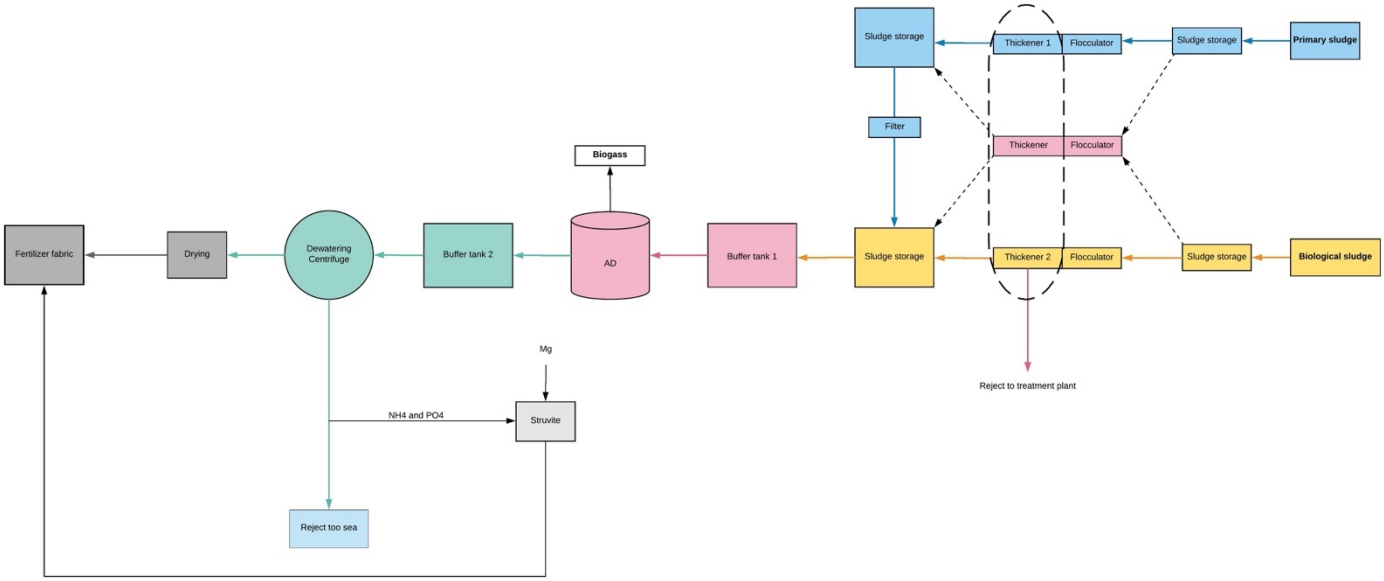

Figure 13: Alternative 2 for a new sludge treatment line

4.2 Different scenarios

The difference between the scenarios tested is presented in Table 22 together with some outputs from the model. Scenario 1-1 to 2-2 were based on the actually fraction of active PAOs needed in the return sludge line to reach steady state; $M_{totP,in} = M_{totP,eff} + M_{totP,w}$. For scenario 3-1 to 4-2 the sum of the effluent phosphorus and the phosphorus sent to sludge treatment, $M_{totP,w}$, are lower than the mass of Tot-P entering the process lines. The reason for this result is a higher mass of tot-P in the "output return sludge line" compared with the assumed input used in the model. This is a weakness in this model, because more phosphorus should have been wasted when the fraction of active PAOs is reduced to 1 %.

Table 22: Conditions and results for scenario 1-1 to 4-2

Scenarios		1-1	1-2	2-1	2-2	3-1	3-2	4-1	4-2
Conditions	$Q_{in,TP}$ [m^3/d]	1,41	1,41	1,41	1,41	1,41	1,41	1,41	1,41
	TSS_R [%]	2	2	1	1	2	2	1	1
	$f_{A,PAO}$ [%]	31,5	31	25,5	22,5	1	1	1	1
	SRT [d]	8	10,35	7	9,54	8,2	10,4	7	9,53
	X [g/m^3]	3000	4000	3000	4000	3000	4000	3000	4000
	$TotP_{in,TP}$ [g/m^3]	17	17	17	17	17	17	17	17
	Ortho-$P_{in,TP}$ [g/m^3]	7	7	7	7	7	7	7	7
Results	$M_{totP,in,TP}$ [kg/d]	2071	2071	2071	2071	2071	2071	2071	2071
	$M_{totP,in}$ [kg/d]	1515	1506	1507	1492	1492	1491	1492	1490
	$M_{totP,w}$ [kg/d]	872	571	576	119	116	76	93	21
	$M_{VSS,w}$ [kg/d]	6911	4590	5506	1273	6727	4533	5506	1273
	Q_R [m^3/s]	0,306	0,44	0,79	1,3	0,306	0,44	0,79	1,3
	$M_{totP,eff}$ [kg/d]	643	935	931	1373	411	298	427	643
	$M_{totP,centrate}$ [kg/d]	804	552	554	177	153	129	140	94
	$M_{NH4,centrate}$ [kg/d]	423	335	360	211	405	333	370	211
	P-content in sludge [%]	10	10	8,6	8	1,5	1,4	1,4	1,3

The EBPR-Process

Biomass concentration of 3000 g VSS/m^3 in the reactor

Effluent orthophosphate concentrations for all scenarios are presented together with the selected orthophosphate concentration entering the treatment plant in Figure 14. Scenario 1-1 (blue) result in a removal efficiency of 91,3 %, which is a high percentage based on the diluted wastewater used as input and the high phosphorus concentration needed to remove. Wastewater treatment plant designed for phosphorus removal is required to remove 90 % of the phosphorus entering the treatment plant (Miljødepartementet, 2004).

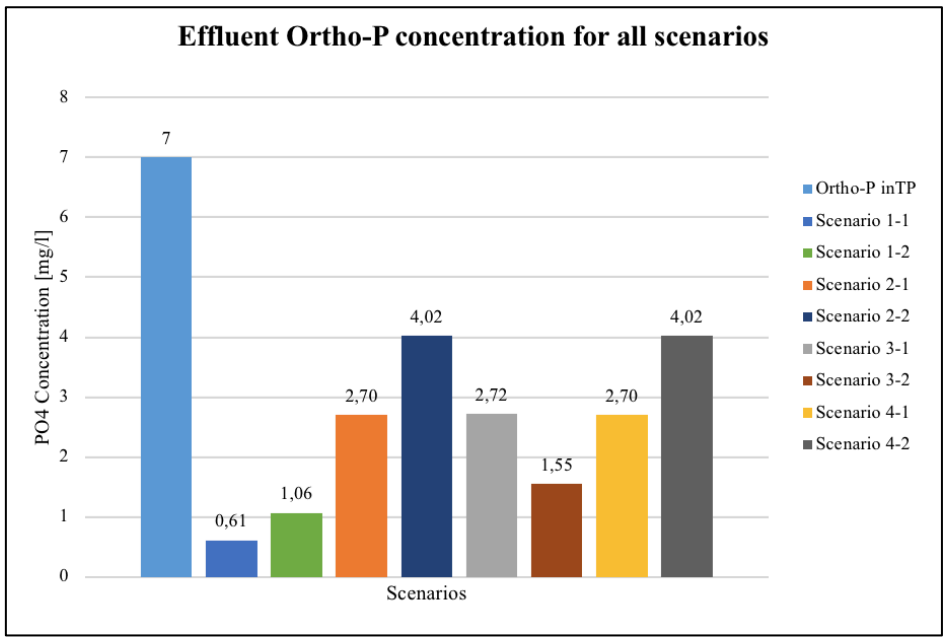

Figure 14: Effluent ortho-phosphate concentration for all scenarios

Scenario 2-1 (orange) , and 4-1 (yellow) results in identical effluent concentrations, and both scenarios are based on the solid concentration of 1 % in the return sludge line. VFA available for PAOs in AN2 and AN3 is the limitation for both scenarios. The reduction in the active fraction of PAOs from 25,5 to 1 % has no impact on the effluent ortho-phosphate concentration, meaning that less than 1 % of the PAOs consume VFA and releases P in AN2 and AN3. If the nitrate concentration in the return sludge line is set to zero for scenario 2-1, all ortho-P are removed through the EBPR-process. This can indicate that the nitrification rate is unrealistically high in the model, and needs to be further checked based on real data from IVAR SNJ. In scenario 3-1 (grey) all active PAOs consume VFA and release ortho-P in the anaerobic reactors, so the active PAO fraction is the limited factor for further P-removal in the aeration tank for this scenario.

Biomass concentration of 4000 g VSS/m^3 in the reactors

In scenario 1-2 (green) and 2-2 (dark blue) the effluent orthophosphate concentration is higher than in scenario 1-1 (blue) and 2-1 (orange), where VFA is the limitation for further P-removal for all these scenarios. The main reason for less VFA available for PAOs in scenario 1-2 and 2-2 than in scenario 1-1 and 2-1, is a higher nitrate concentration in the inlet of AN2. To maintain a biomass concentration in the reactors of 4000 g VSS/m^3, the return flow need to be increased, which results in more nitrate in the system. Meaning more rbCOD is utilized in denitrification in AN2, and less VFA is available for PAOs. The increase in biomass

concentration actually makes the EBPR-process less efficient when VFA is the limiting factor, because more nitrate needs to be denitrified. Also for a biomass concentration of 4000 g VSS/m^3 scenario 2-2 and 4-2 (dark grey) gives the same effluent orthophosphate concentration, and VFA is still the limitation for higher P-removal.

In both scenario 3-1 (grey) and 3-2 (red) the fraction of PAOs is the limitation for further P-removal. A lower orthophosphate concentration is achieved for scenario 3-2, when the biomass concentration in the reactors is increased to 4000 g VSS/m^3. The reason for a higher P-removal in scenario 3-2 compared with 3-1, is due to a higher mass of active PAOs in the system, because the return flow is higher for scenario 3-2 to maintain the wanted biomass concentration. When mass of active PAOs is the limitation for P-removal an increase in the biomass concentration in the reactors is positive.

Sludge production

The sludge production is presented in mass of volatile suspended solids, $M_{VSS,w}$, in Figure 15. The biomass concentration of 3000 g VSS/m^3 result (naturally) in a higher sludge production than X= 4000 g VSS/m^3, because less VSS is recycled in the return sludge line. The combination of X= 3000 g VSS/m^3 and 2 % TSS in the return sludge line results in the highest sludge production, see scenario 1-1 (blue) and 3-1 (grey). Higher VSS mass sent to sludge treatment, corresponds to higher biogas production in the anaerobic reactor. The lowest sludge production is achieved when TSS_R= 1 % and X= 4000 g VSS/m^3, for scenario 2-2 (dark blue) and 4-2 (dark grey).

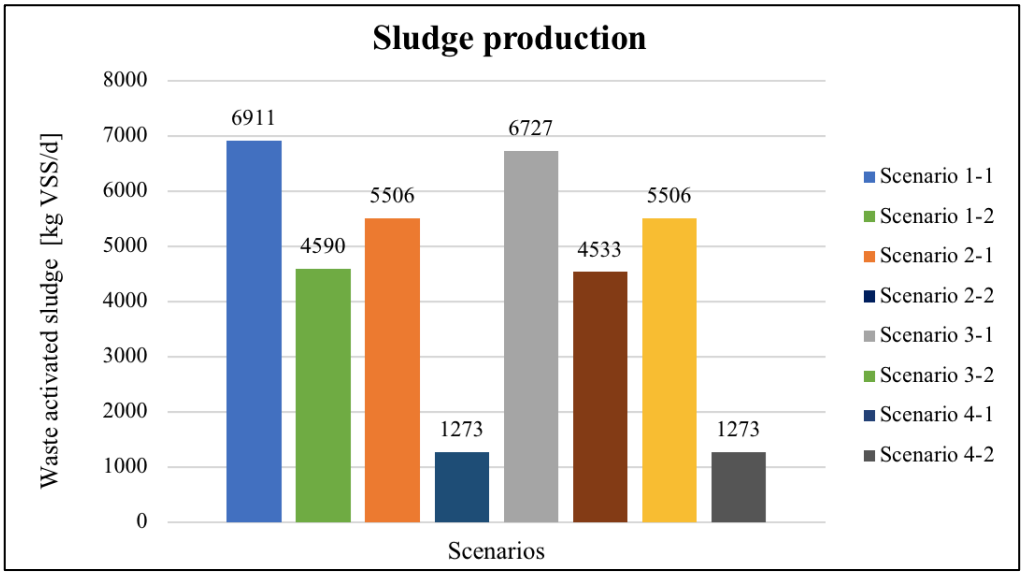

Figure 15: Sludge production based on VSS for all scenarios

Summary scenarios

From these findings it would be recommended to operate with a solid concentration in the return sludge line of 2 % or higher to provide enough food to the microorganisms in the system. The settling properties in the clarifier are of huge importance for the EBPR- process to function optimal at IVAR SNJ, due to diluted wastewater. If nitrification in the aeration tank could be avoided in the aeration tank it would improve the system. If not, IVAR should consider adding external carbon in AN1 to denitrify all recycled nitrate in this reactor, to avoid nitrates negative impact on the EBPR-process. The favorable biomass concentration in the reactors based on these scenarios is 3000 g VSS/m^3, based on the mass sludge sent to sludge treatment.

5. Limitations in the model

Limitation and weaknesses in the mass balance model will be discussed.

Cell growth through fermentation

A mistake has been made in the mass balance model, which unfortunately was detected too late in the process to change the error in the result presented in this thesis. The model have included cell growth due to fermentation, where all VFA produced ($bsCOD_{fer}$) have been used as input in the calculation of the production of new biomass in all three anaerobic reactors. The VFA produced during fermentation have also been used in the calculations if dissolved oxygen and/or nitrate are present in the anaerobic reactors. Hence, cell growth due to fermentation have been based on a VFA mass that already have been consumed by other microorganisms in the model. This error have now been fixed in the model by neglecting the cell growth due to fermentation. It is assumed that the actual biomass production through fermentation is a very small fraction of the VFA produced, and thereby could be neglected from the calculations. As an example: For the simulation of the normal operation at IVAR SNJ (normal 2), the sludge production ($M_{VSS,w}$) would be reduced with 2,6 % if the biomass produced during fermentation was neglected. This error will not affect the final conclusion made in this thesis, because the error is relative. For all scenario tested and presented in the results, the VSS mass sent to sludge treatment is slightly higher than what is should, but not in a large extent.

Time dependent rates: Hydrolysis and fermentation

Both the hydrolysis rate and the fermentation rate constants is depended on time, the hydraulic retention time in the given anaerobic tank is therefore of big interest. It is assumed in the mass balance calculations that the biological reactors at IVAR SNJ behave as a completely mixed flow reactor (CMFR). The psbCOD and bsCOD used as input in the calculations of hydrolysis and fermentation, should therefore be the concentration in the reactor. Since CMFR is assumed, the concentration in the reactor is equal to the effluent concentration. For simplification the concentration entering the reactor have been used as input in the mass balance calculations. The difference between the psbCOD entering the anaerobic reactors and the psbCOD in the reactor is relatively small, especially for AN1 where the inlet concentrations is relatively high.

Selected constant values

The first fermentation rate constant, k_{fer}, used in the mass balance calculations is given for a temperature of 20 °C (Amy *et al.*, 2008), the temperature used in the mass balance model is 15 °C. Since a temperature correlation was not given by Amy *et al.* (2008), the rate constant for 20 °C was used directly in the mass balance calculations. For a temperature of 15 °C this rate constant would probably be slightly lower, but for a temperature difference of 5 °C it is assumed to be okay in this mass balance model.

The active concentration of OHOs ($X_{OHO,active}$) used to calculate the amount of VFA produced in the fermentation process have been set to 21,4 % in the model for all scenarios. This percentage was assumed based on a fraction of active PAOs in the return sludge line of 15 %, with the simplification that the active biomass only consists of OHOs, PAOs and NITs. Which is not the reality in a wastewater treatment plant. It would be recommended to change this percentage based on real data, or find a link between the fraction of active PAOs and fraction of active OHOs in return sludge line that is valid for all situations. This parameters only affects the VFA produced during fermentation in the model. The fermentation process in the model is limited by the biodegradable soluble COD available in the reactor and in the reactors, for especially AN2 and AN3 are the masses fermented very small due to short hydraulic retention time.

Nitrification in the aeration tank

Nitrification will first occur after OHOs and PAOs have utilized oxygen for their own cell growth. The nitrification process will be limited by the hydraulic retention time, SRT, and the amount of biodegradable organic material present in the aeration tank. In the model it is assumed that 70 % of the ammonium available after growth of OHOs and PAOs will be converted to nitrate. This percentage is most likely too high, because the activated sludge system at IVAR SNJ operates with a low sludge age in a relatively cold climate (Amy *et al.*, 2008). Some nitrification will most likely happen at IVAR SNJ nevertheless, because the amount of bsCOD entering the aeration tank, based on the mass balance model, is low. If a high fraction of the influent ammonium is converted to nitrate at IVAR SNJ, they should consider adding more rbCOD in the first anaerobic tank to make the bio-P reactors, AN2 and AN3, more efficient regarding phosphorus removal.

Cell growth in the activated sludge system

As already mentioned the model have not included nutrients, N and P, as a possible limitation for cell growth, so the biomass produced in the model is higher than what it would be in reality when nutrients is the limitation for microbial growth. To make the model better this limitation should be included.

Difference between return sludge input and return sludge output

To make the model work its purpose it was found necessary to held the particulate material psbCOD, pnbCOD, TSS, VSS, Tot-N and Tot-P entering AN1 constant to be able to test different scenarios. The soluble fraction used as input have been set equal to the output return sludge line box. For the different scenarios when steady state was not achieved, especially for particulate COD and phosphorus, the return output was slightly different compared with the return input. The variance is small for particulate COD, but for Tot-P there seems to be an accumulation of particulate phosphorus in the output when the selected active fraction of PAOs is lower than what is should have been to be able to reach steady state. In general the model should be made more user-friendly and automatic, and not be based on the need of manual changes.

Sludge storage

In the mass balance model it is assumed that the concentrations entering the sludge storage tanks is equal to the concentrations leaving the storage tanks. This assumption is probably sufficient for the different storage tank before and after the thickeners, but the retention time in the first and second buffer tank is approximately 2 and 4 days, respectively. If the sludge is stored longer than 2 or 3 days, it will deteriorate, become odorous and be more difficult to dewater (Tchobanoglous *et al.*, 2014). Based on this fact the simplification made in the model will not match the reality for buffer tank 1 and 2, because anaerobic degradation of organic material will occur in these tanks, and these processes should be included in the mass balance model. If IVAR change their sludge treatment line, and change buffer tank 1 to an anaerobic mixer for luxury-P release, a new mass balance analysis should be performed for the new system.

Anaerobic digestion

For the calculations of VSS destruction in the anaerobic digestion it is used a percentage of 55 %, which is based on the destruction of biomass for the previous treatment plant at IVAR SNJ

with chemical sludge. Since biological sludge is less biodegradable and mainly is composed of bacteria (Carrere *et al.*, 2016), the percentage of VSS destruction is probably too high to match the real situation. This value can be changes in the model when the biological treatment plant have been in operation for a while, and IVAR have collected updated data. The biomass production in the anaerobic digestion could be increase by adding e.g a thermal pretreatment to ease cell disruption (Carrere *et al.*, 2016).

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6. Conclusion

The aim of this master thesis was to make a theoretical mass balance model identical to IVAR´s wastewater treatment plant (SNJ). IVAR SNJ recently changed from chemical precipitation to enhanced biological phosphorus removal (EBPR) based on activated sludge. The theoretical mass balance model created can be used as an analyzing tool, that emphasize possible limitations in the system. Hence, the model can be used as a basis for decision making to secure optimal operation of the plant.

The main limitation for microbial growth at IVARs wastewater treatment plant seems to be the access to phosphorus. Infiltration/inflow of extraneous water, and combined sewage systems (CSS) are most likely the reason for the diluted wastewater entering IVAR SNJ. A reduction in the amount of clean water in the sewage system would be beneficial for both the municipalities and IVAR SNJ. When the influent orthophosphate concentration was set to 0,94 g P/m^3 in the model, based on the real situation in 2017, there was no need for PAOs to achieve a removal efficiency of 100 %. This indicates that there is a need for a higher phosphorus concentration to create a beneficial environment for growth of PAOs, and other microorganisms for that matter.

From the comparison of measured and simulated effluent concentration, it can be concluded that the EBPR- process at IVAR SNJ is not optimal at the moment. The average effluent soluble phosphorus concentration measured at IVAR SNJ was equal to the concentration entering the treatment plant, which indicates that the production of new biomass is low or not existing. The new biological treatment plant, with three process lines, was first implemented in March 2018, hence this can be related to startup problems.

Based on the results from the scenarios tested for “unrealistic high” influent phosphorus concentrations, it would be recommended to have a solid concentration of 2 % or higher in the return sludge line. This is to make sure there is enough biodegradable COD in the activated sludge system, and for the EBPR-process. If COD is the limitation for further P-removal, it would be recommended to maintain a biomass concentration in the reactors around 3000 g VSS/m^3, because a higher concentration makes the P-removal less efficient due to more nitrate in the system. If a higher biomass is wanted one should consider adding external carbon in AN1, to prevent the negative impact nitrate has on the EBPR-process in AN2 and AN3.

With the existing sludge treatment line phosphorus is not removed or utilized as a resource. From the simulation of the normal operation at IVAR SNJ it was found that approximately 70 % of influent Tot-P follows the effluent and centrate flow, which is directly discharged in the ocean. Hence, the sludge treatment line need to change before the new treatment plant works its potential purpose. The recommendation would be to implement an anaerobic mixer prior the anaerobic digestion (AD), to provoke release of luxury-P. By changing buffer tank 1 to an anaerobic mixer the change could be less time and space consuming. The release of luxury-P is depended on the access to volatile fatty acids (VFA), which could be produced through hydrolysis and fermentation of primary sludge and be added externally. Struvite could then be formed under controlled conditions, by mixing the reject from the anaerobic mixer and the centrifugal dewatering (centrate). The recovered phosphorus and nitrogen can then be utilized in IVAR fertilizer product, Minorga. By implementing this change the risk for unwanted struvite precipitation in the sludge treatment line would be drastically reduced.

7. Further work

The mass balance model is created in excel, and the user interface is unfortunately not very user-friendly. It would be recommended to either get familiar with the model as it is, or to create the same model in excel or in another program with an easier user interface.

Recommended improvements in the model

- Include nutrients, phosphorus and nitrogen, as a possible limitation in the production of new biomass.
- The actual fraction of active PAOs in the system should be calculated automatically in the model.
- The masses used as input in AN1 should automatically be equal to the output.
- Include SRT_{min} for the EBPR-process as a limitation. The minimum sludge age should be found at IVAR SNJ based on real data.
- Find a link between the fraction of active PAOs and active OHOs to calculate the actual fraction of active OHOs ($f_{A,OHO}$) in the return sludge line.
- Include the relationship between NITs, OHOs and PAOs in the nitrification calculations, so nitrification first occurs after cell growth of PAOs and OHOs, if the hydraulic retention time is long enough.
    - Could alternatively change the nitrogen removal percentage in the mass balance model based on real data from IVAR SNJ.
- Include the treatment processes occurring in buffer tank 1 and buffer tank 2 in the model.

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NOMENCLATURE

Symbol	Description	Units
$bCOD_{in}$	The biodegradable COD entering the process lines	g bCOD /m^3
$Biogas_{AD}$	Biogas production in AD	m^3/d
BOD_2	Biochemical oxygen demand after 2 days	g O_2/m^3
BOD_t	Biochemical oxygen demand after t days	g O_2/m^3
BOD_U	Ultimate biochemical oxygen demand	g O_2/m^3
$BOD_{U,P}$	Ultimate particulate biochemical oxygen demand	g O_2/m^3
$BOD_{U,S}$	Ultimate soluble biochemical oxygen demand	g O_2/m^3
$bsCOD_{eff}$	Effluent bsCOD concentration	g COD /m^3
$bsCOD_{eff,aer}$	bsCOD concentration leaving the aeration tank	g bsCOD/m^3
$bsCOD_{fer,AN1}$	bsCOD converted to VFA in AN1 through fermentation	g VFA/m^3
$bsCOD_{in}$	bsCOD concentration entering the process lines	g bsCOD/m^3
$bsCOD_{in,AN1}$	The bsCOD concentration available in AN1	g bsCOD/m^3
$bsCOD_{in,aer}$	bsCOD concentration entering aeration tank	g bsCOD/m^3
$bsCOD_{in,AN1}$	bsCOD concentration available for fermentation in AN1	g bsCOD/m^3
$bsCOD_{in,ANn}$	bsCOD available for fermentation in AN2 or AN3	g bsCOD/m^3
$bsCOD_{res,\ AN1}$	The bsCOD concentration not fermented in AN1	g bsCOD/m^3
$C_{eff,AN1}$	The concentration leaving AN1	g/m^3
C_{in}	The concentration entering the process lines	g/m^3
$C_{in,AN2}$	The concentration entering AN2	g/m^3
$CH4_{AD}$	Methane production in AD	l CH4 /d
COD_{cell}	COD concentration incorporated in cells	g COD/m^3
COD_{CH4}	The COD available for methane production	g COD/m^3
COD_{ox}	COD oxidized	g COD/m^3
DO	The dissolved oxygen concentration	g O_2/m^3
dS/dt	Change in substrate concentration in the reactor	g COD/m^3*d
dX/dt	Microbial growth rate from substrate utilization	g VSS /m^3*d
F_{DN}	The amount of rbCOD consumed through denitrification	g rbCOD/g NO_3-N_r
F_{DO}	The amount of rbCOD removed due to dissolved oxygen	g rbCOD/g O_2
F_{pnbCOD}	The new biomass converted to pnbCOD	g pnbCOD/g VSS
F_{psbCOD}	The new biomass converted to psbCOD	g psbCOD/g VSS

$f_{A,OHO}$	The active OHO fraction in biomass (VSS)	-
$f_{A,PAO}$	The active PAO fraction in biomass (VSS)	-
f_{CH4}	The CH_4 equivalent of converted COD	l CH_4 /g COD
f_{cv}	Particulate COD / VSS ratio	g tpCOD/g VSS
$f_{end,OHO}$	Fraction of endogenous particulate residue of OHOs	-
$f_{end,PAO}$	Fraction of endogenous particulate residue of PAOs	-
$f_{fer,growth}$	COD fraction utilized for cell growth during fermentation	-
f_n	The nitrogen content in the biomass	g N/ g VSS
$f_{n,AD}$	The nitrogen content in the biomass entering AD	g N/ g VSS
f_{luxP}	The luxury uptake of phosphorus	g P/g VSS
f_{OHO}	Fraction of P in active OHOs	g P/g VSS
f_p	Particulate fraction after hydrolysis	-
$f_{P,ass}$	The phosphorus content assimilated in the biomass	g P/g VSS
f_{PAO}	Fraction of P in active PAOs	g P/g VSS
$f_{P,end}$	Fraction of non-biodegradable mass that is phosphorus	g P/g VSS
$f_{p,inert}$	The phosphorus fraction of the non-biodegradable mass	g P/g VSS
f_{PO4}	Ratio P release/VFA uptake	g P/g VFA
$f_{SC,sed}$	Solid capture in sedimentation basin (Clarifier)	-
$f_{SC,\ THn}$	Solids captured in thickener n	-
f_s	Soluble fraction after hydrolysis	-
f_{snbCOD}	Percentage of tsCOD converted to snbCOD by hydrolysis	-
$f_{snb\text{-}N}$	Fraction of influent TotN that is inert soluble organic	-
f_{ssbCOD}	Percentage of tsCOD converted to ssbCOD by hydrolysis	-
FSS/TSS	The fixed fraction of TSS	-
$f_{TS,G}$	TS fraction of the grease load removed	-
$f_{TS,S}$	The TS fraction of the daily sand load removed	-
$f_{TS,SS}$	The TS fraction of the daily trash load removed	-
f_{rbCOD}	Fraction of tsCOD converted to rbCOD by hydrolysis	-
$f_{VSS,R}$	VSS fraction of TSS in return sludge line	-
$f_{VSS,G}$	VSS fraction of TSS in grease trap	-
$k_{d,fer}$	Endogenous decay coefficient for fermentation	d^{-1}
$k_{d,i}$	Endogenous decay coefficient for microorganism i	d^{-1}
k_{hyd}	Hydrolysis rate constant	d^{-1}

K_S	Substrate half-saturation constant	g COD/m^3
$K_{S,NIT}$	Half saturation constant for ammonium	g NH_4/m^3
$K_{S,NO3}$	Nitrate half-saturation constant	g NO_3-N /m^3
$k_{fer,T}$	First order fermentation rate constant at T degrees	m^3/g VSS*d
$K_{s,fer}$	Substrate half-saturation constant for fermentation	g COD/m^3
$M_{A,PAO}$	The mass of active PAOs	g VSS /d
$M_{cake,THn}$	The solid cake formed during thickening	kg TSS /d
$M_{eff,AN1}$	Mass of VSS leaving AN1	g VSS/d
M_{eff}	Mass of VSS in the effluent	g VSS/d
$M_{end,OHO}$	Mass of endogenous residue in the system due to OHO	g VSS/d
$M_{end,PAO}$	Mass of endogenous residue in the system due to PAO	g VSS/d
$M_{hyd,AN1}$	Mass of VSS reduced in AN1 due to hydrolysis	g VSS/d
$M_{in,AN1}$	Mass of VSS entering AN1	g VSS/d
$M_{in,AN2}$	Mass of VSS entering AN2	g VSS/d
M_{in}	Mass of VSS entering the process lines	g VSS/d
M_{inert}	Non-biodegradable mass originated from the influent	g VSS/d
$M_{in,sed}$	VSS mass entering settling tank	g VSS/d
M_{NIT}	The change in NIT mass in the activated sludge system	g VSS
$M_{OHO,new}$	The production of new active mass of OHOs	g VSS/d
$M_{PAO,AD}$	The active mass of PAOs entering the anaerobic digester	g VSS/d
$M_{PAO,new}$	The production of new active mass of PAOs	g VSS/ d
$M_{TSS,THn}$	The mass of TSS entering thickener n	Kg TSS/d
$M_{VFA(ava, PAO,ANn)}$	The VFA mass available for PAOs in AN2 or AN3	g VFA/d
$M_{VFA,ava, PAO,tot}$	The total VFA mass available to be consumed by PAOs	g VFA/d
$M_{VFA,PAO}$	The actual VFA mass consumed by PAOs	g VFA/d
$M_{VFA, max,PAO}$	The maximum VFA mass that can be removed by PAOs	g VFA/d
M_{Trash}	The daily trash load removed by the screens	G TSS/d
$M_{VFA,PAO}$	The amount of VFA consumed by active PAOs	g VFA/d
M_w	VSS mass in the waste sludge line	g VSS/d
n	Number of mole	mole
NH4	Ammonia nitrogen concentration	g NH4/m^3

$NH4_{ava,NIT}$	The ammonia nitrogen concentration available for nitrification	g $NH4/m^3$
$NO3_{DN}$	Nitrate removed through denitrification	g $NO3/m^3$
$NH4_{eff,aer}$	Ammonia nitrogen concentration leaving aer	g NH_4/m^3
$NH4_{in,AD}$	Ammonia nitrogen concentration entering AD	g $NH4/m^3$
$NH4_{in,aer}$	Ammonia nitrogen concentration entering aeration tank	g $NH4/m^3$
$NH_{4,rel,AD}$	The release of ammonia nitrogen in anaerobic digestion	g NH_4/m^3
$NO3_{pot}$	The potential nitrate concentration removed	g $NO3\text{-}N/m^3$
$NO3_{pot,AN1}$	The potential nitrate concentration removed through DN	g $NO3\text{-}N/m^3$
NO3	Nitrate concentration	g $NO3\text{-}N/m^3$
$NO3_{DO,ANn}$	Nitrate removed in anaerobic reactor n due to DO	g $NO3\text{-}N/m^3$
$NO3_{DN,Ann}$	Nitrate removed through DN in anaerobic reaction n	g $NO3\text{-}N/m^3$
$NO3_{DN,AN1}$	Nitrate concentration removed in denitrification in AN1	g $NO3\text{-}N/m^3$
$NO3_{eff,aer}$	Nitrate concentration leaving aeration tank	g $NO3\text{-}N/m^3$
$NO3_{in,AD}$	Nitrate concentration entering AD	g $NO3/m^3$
$NO3_{in,aer}$	Nitrate concentration entering aeration tank	g $NO3\text{-}N/m^3$
N_{sludge}	Nitrogen concentration assimilated in biomass	g N/m^3
P	Pressure	atm
PO_4	Ortho-phosphate concentration	g P/m^3
$PO4_{in,aer}$	Orto-phosphate concentration entering aeration tank	g P/m^3
$PO4_{rel,ass,AD}$	The release of assimilated phosphorus in AD	g P/m^3
$PO4_{rel,PAO,AD}$	The release of luxury P in AD	g P/m^3
$P_{r,ac}$	The phosphorus actually removed from the solution	g P/m^3
$P_{rel,ac}$	The total amount of phosphorus actually released	g P/d
$P_{rel,pot,\ ANn}$	The potential P-release in AN2 or AN3	g P/d
$P_{rel,MAX}$	The maximum amount of phosphorus possible to release	g P/d
$P_{r,end}$	Phosphorus removed due to endogenous residue mass	g P/m^3
$P_{r,inert}$	Phosphorus removal due to influent inert mass	g P/m^3
$P_{r,OHO}$	Phosphorus removed by OHOs	g P/m^3
$P_{r,PAO}$	Phosphorus removed by PAOs	g P/m^3
$P_{r,tot}$	The phosphorus concentration possible to remove	g P/m^3
P_S	Fraction of solid in sludge	-
$P_{S,THn}$	Fraction of solid in thickened sludge in thickener n	-

$pnbCOD_{eff,SS}$	pnbCOD concentration leaving screening station	g pnbCOD/m^3
$psbCOD_{eff,SS}$	psbCOD concentration leaving screening station	g psbCOD/m^3
$pnbCOD_{in}$	The pnbCOD concentration entering the process lines	g pnbCOD/m^3
$psbCOD_{hyd}$	The amount of psbCOD solubilized	g COD/m^3
$psbCOD_{hyd,AN1}$	The amount of psbCOD hydrolyzed in AN1	g psbCOD/m^3
$psbCOD_{in,AN1}$	psbCOD concentration entering AN1	g psbCOD/m^3
$P_{w,THn}$	Water fraction of thickened sludge in thickener n	-
Q_{AD}	Flowrate entering the anaerobic digestion	m^3/d
Q_{in}	Flowrate entering the process lines	m^3/d
$Q_{eff,sed}$	Flowrate leaving sedimentation basin	m^3/d
$Q_{eff,THn}$	The sludge flow leaving thickener n	m^3/d
$Q_{eff,TH2}$	The flow leaving the thickener 2	m^3/d
$Q_{in,aer}$	Flowrate entering the aeration tank	m^3/d
$Q_{in,ANn}$	The flow entering AN2 or AN3	m^3/d
$Q_{in,DF}$	Flowrate entering drum filters	m^3/d
$Q_{in,SS}$	The flowrate entering the screening station	m^3/d
Q_R	Return flowrate	m^3/d
$Q_{R,out}$	Return flowrate in the output of the model	m^3/d
Q_w	Waste sludge flowrate	m^3/d
R	The gas constant = 0,082057 atm*L/mole*K	atm*L/mole*K
$rbCOD_{ava,DN,ANn}$	rbCOD available for DN in AN, AN2 or AN3	g rbCOD/m^3
$rbCOD_{ava,DO,ANn}$	rbCOD available for DO in AN1, AN2 or AN3	g rbCOD/m^3
$rbCOD_{cells}$	rbCOD incorporated in cells	g rbCOD/m^3
$rbCOD_{DO}$	rbCOD utilized due to the presents of dissolved oxygen	g rbCOD/m^3
$rbCOD_{DN}$	rbCOD utilized through denitrification	g rbCOD/m^3
$rbCOD_{hyd}$	rbCOD available after hydrolysis	g rbCOD/m^3
$rbCOD_{in,ANn}$	rbCOD concentration entering AN1, AN2 or AN3	g rbCOD/m^3
$rbCOD_{OX}$	rbCOD concentration removed through oxidation	g rbCOD/m^3
$rbCOD_{pot,DO}$	The potential rbCOD concentration removed due to DO	g rbCOD/m^3
$rbCOD_{DN}$	rbCOD concentration removed through denitrification	g rbCOD/m^3
r_{net}	Net biomass growth rate	g VSS/m^3*d
r_{su}	Substrate utilization rate	g COD/m^3*d
r_{T1}	Conversion rate at temperature T1	

r_{NO3}	Nitrate consumption rate	g NO_3-N/m^3*d
r_{fer}	Conversion rate of fermentable organic material	g COD/ m^3 *d
S_f	Specific gravity of fixed solids	-
SP_{net}	The net sludge production removed from system	g VSS/d
SRT	Solid retention time	d
$SRT_{min,NIT}$	Minimum solid retention time for nitrifiers	d
S_s	Specific gravity of solids	-
$ssbCOD_{in,ANn}$	ssbCOD concentration entering AN2 or AN3	g ssbCOD/m^3
S_{SL}	Specific gravity of sludge	-
$S_{SL,THn}$	Specific gravity of the sludge leaving thickener n	-
S_v	Specific gravity of volatile solids	-
T	Temperature in kelvin	K
TotN	Total nitrogen concentration	g N/m^3
$TotN_{in,AD}$	Total nitrogen concentration entering AD	g N/m^3
TotP	Total phosphorus	g P/m^3
$tpCOD_{eff,SS}$	tpCOD concentration leaving the screening station	g COD/m^3
$tsCOD_{hyd}$	The total soluble COD concentration after hydrolysis	g tsCOD/m^3
$TSS_{in,DF}$	TSS concentration entering the drum filters	g TSS/m^3
$TSS_{r,SS}$	TSS concentration removed in the screening station	g TSS/m^3
$TSS_{r,DF}$	TSS concentration removed by drum filters	g TSS/m^3
$VFA_{DN2,AN2}$	VFA concentration utilized during DN step 2 in AN2	g VFA/m^3
$VFA_{DN,AN3}$	VFA concentration utilized during denitrification in AN3	g VFA/m^3
$VFA_{DO2,AN2}$	VFA concentration utilized due to DO step 2 in AN2	g VFA/m^3
$VFA_{DO,AN3}$	VFA concentration utilized due to DO in AN3	g VFA/m^3
$VFA_{fer,ANn}$	The VFA concentration formed through fermentation in AN1, AN2 or AN3	g VFA/m^3
V_{tot}	The total volume of all three AN2, AN3 and Aer reactors	m^3
$V_{tot,aer}$	Total volume of all three aeration tanks	m^3
$V_{tot,ANn}$	Total volume of all three AN1, AN2 or AN3 reactors	m^3
VSS_{AN2}	VSS concentration in reactor AN2	g VSS/m^3
VSS_{des}	Destructed biomass in anaerobic digestion	g VSS/m^3
VSS_{DN}	The increase in biomass due to denitrification	g VSS/m^3
VSS_{DO}	The increase in biomass due to dissolved oxygen	g VSS/m^3

$VSS_{eff,TH2}$	The VSS concentration leaving thickener 2	g VSS/m^3
$VSS_{fer,AN1}$	The increase in biomass due to fermentation in AN1	g VSS/m^3
$VSS_{hyd,AN1}$	The reduction in VSS due to hydrolysis in AN1	g VSS/m^3
VSS_{in}	VSS concentration entering process lines	g VSS/m^3
$VSS_{in,AD}$	VSS concentration entering AD	g VSS/m^3
VSS_{new}	The increase in VSS due to cell growth	g VSS/m^3
VSS_R	Biomass concentration in the return sludge line	g VSS/m^3
$VSS_{r,DF}$	VSS concentration removed by drum filters	g VSS/m^3
$VSS_{r,SS}$	VSS concentration removed by the screens	g VSS/m^3
VSS/TSS	The volatile fraction of TSS	-
X	Biomass concentration in the reactors	g VSS/m^3
X_{eff}	Biomass concentration in effluent	g VSS/m^3
X_{in}	Biomass concentration entering the process lines	g VSS/m^3
$X_{OHO,active}$	The concentration of active OHOs in the reactor	g VSS/m^3
X_{NIT}	Concentration of nitrifiers in the reactor	g VSS/m^3
$X_{PAO,active}$	The active PAO concentration	g PAO/m^3
$X_{OHO,\ active}$	The active heterotrophic bacteria concentration	g OHO/m^3
X_R	VSS concentration in return/waste sludge line	g VSS/m^3
Y	The synthesis yield	g VSS/g COD
Y_i	The synthesis yield for microorganism i	g VSS/g COD
$Y_{OBS,i}$	The observed yield for microorganism i	g VSS/g COD
$Y_{OHO,DN}$	Synthesis yield for OHO when nitrate is the electron acceptor	g VSS/gCOD
Y_{fer}	The synthesis yield of acidogenic	g VSS/g COD
μ	The specific growth rate	d^{-1}
μ_{max}	Maximum specific growth rate	d^{-1}
$\mu_{max,NIT}$	Maximum specific growth rate for nitrifiers	d^{-1}
θ	Temperature coefficient	-
τ	Hydraulic retention time	d
ρ_w	Density of water	kg/m^3