İSTANBUL TECHNICAL UNIVERSITY ★ INSTITUTE OF SCIENCE AND TECHNOLOGY

COMPARISON OF DESIGN RULES WITH DYNAMIC MODELING OF A FULL SCALE WASTEWATER TREATMENT PLANT: PAŞAKÖY BIOLOGICAL NUTRIENT REMOVAL FACILITY

M.Sc. Thesis by
Berk GÜDER

Department : Environmental Engineering

Programme : Environmental Biotechnology

JANUARY, 2011

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TASARIM YÖNTEMLERİNİN TAM ÖLÇEKLİ ATIKSU ARITMA TESİSİ DİNAMİK SİMÜLASYON SONUÇLARIYLA KARŞILAŞTIRILMASI: PAŞAKÖY ATIKSU ARITMA TESİSİ

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FOREWORD

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I dedicate this thesis to my grandfather Orhan ÖZER

January 2011 Berk GÜDER

Environmental Engineer

TABLE OF CONTENTS

Page

ABBREVIATIONS

b_A : Endogenous decay rate for autotrophs
b_H : Endogenous decay rate for heterotrophs
AOR : Actual Oxygen Rate
ASM : Activated Sludge Model
BNR : Biological Nutrient Removal
BOD : Biochemical Oxygen Demand, mg/l
COD : Chemical Oxygen Demand, mg/l
C_S : Biodegredable COD concentration
DO : Dissolved Oxygen, mg/l
EBPR : Enhanced Biological Phosphorus Removal
f_E : Inert fraction of endogenous biomass
F/M : Food/Microorganisms ratio, d^{-1}
HRT : Hydraulic Retention Time, h
i_{NBM} : Nitrogen fraction of biomass
i_{NS} : Nitrogen fraction of soluble COD
i_{NX} : Nitrogen fraction of particulate COD
i_{PS} : Phosphate fraction of soluble COD
i_{PX} : Phosphate fraction of particulate COD
IR : Internal Recycle
k_{dn} : Endogenous decay coefficient
k_{dn} : Endogenous decay coefficient for nitrifying organisms
k_h : Maximum hydrolysis rate
K_La : Volumetric oxygen transfer coefficient
K_S : Half saturation constant for heterotrophic growth
k_h :Maximum hydrolysis rate
K_X : Half saturation constant for hydrolysis
K_O : Oxygen half saturation constant for heterotrophs
K_{OA} : Oxygen half saturation constant for autotrophs
ML(V)SS : Mixed Liquor (Volatile) Suspended Solids, mg/l
μ_{Amax} : Maxiumum autotrophic growth rate
N_{DN} : Denitrification capacity
N_{DP} : Denitrification potential
N_{OX} : Oxidized nitrogen
N_X : Nitrogen incorporated during heterotrophic growth
NR : Nitrified Recycle
OUR : Oxygen Uptake Rate
ORP : Oxidation Reduction Potential, mV
Q_{in} : Influent flowrate
Q_{IR} : Internal recycle flowrate
PAOs : Phosphorus Accumulating Organisms
RAS : Return Activated Sludge
S_A : Acetate COD, mgCOD/l
S_F : Fermentable COD, mgCOD/l
S_H : Rapidly hydrolyzable COD, mgCOD/l

SCFA : Short chain fatty acids, mgCOD/l
STD : Standart Deviation
S_{ND} : Soluble organic nitrogen, mgN/l
S_{NH} : Ammonia nitrogen, mgN/l
S_{NO} : Nitrate nitrogen, mgN/l
SRT : Sludge Retention Time, d
SRT_m : Minimum Sludge Retention Time, d
S_0 : Dissolved oxygen concentration
S_{PO4} : Ortho-phosphate
S_S : Readily biodegreadable COD, mgCOD/l
S_{Sini} : Initial readily biodegreadable COD, mgCOD/l
TCA : Tricarboxylic Acid
TKN : Total Kjeldahl Nitrogen, mg/l
TSS : Total Suspended Solids, mg/l
T-P : Total Phosphorus, mg/l
Y_H : Heterotrophic yield coefficient
Y_{NH} : Net heterotrophic yield coefficient
X_H : Active heterotrophic biomass
X_I : Particulate Inert COD
WWTP : Wastewater Treatment Plant
X_S : Slowly biodegreadable COD, mgCOD/l
X_{Sini} : Initial slowly biodegreadable COD, mgCOD/l
VFA : Volatile Fatty Acid, mg/l
V_A : Aeration volume
V_{AN} : Anaerobic volume
V_D : Denitrification volume
V_T : Total volume of aeration basin

LIST OF TABLES

LIST OF FIGURES

Page

COMPARISON OF DESIGN RULES WITH DYNAMIC MODELING OF A FULL SCALE WASTEWATER TREATMENT PLANT: PAŞAKÖY BIOLOGICAL NUTRIENT REMOVAL FACILITY

SUMMARY

Increasing pollution of water sources in world brings along strict discharge standarts together with necessity of efficient nutrient (nitrogen and phosphorus) removal from wastewaters. Compared to other treatment alternatives, the activated sludge technology is more convenient in terms of investment and operational costs. The discharge standarts imposed should only be secured by appropiate design, operation and refrofit of activated sludge plants.

Activated sludge models have been widely used for the design, control and optimization of activated sludge plants operated for organic carbon and nutrient removal. Hovewer, appropiate use of those models essentially requires (1) precise influent-wastewater characterization (2) wise use of calibration methodology for activated sludge models and as a result: (3) interpretation of results obtained from modeling study.

In the first part of the study, the wastewater characterization was determined at the operating conditions and the percentil values were calculated for the design. Then, the plant was designed with different (ATV-DVWK 131-E, University Cape Town and Water Environment Research Foundation) design methods and the results were compared.

In the second part of the study, the nutrient removal mechanism was evaluated for the plant was modelled in steady-state and dynamic conditions using plant-wide simulation model programme and the model results were compared with the discharge concentrations.

Last part of the study deals with the investigation of nutrient removal performance of extended aeration type full-scale activated sludge plant.

TASARIM YÖNTEMLERİNİN TAM ÖLÇEKLİ ATIKSU ARITMA TESİSİ DİNAMİK MODEL SİMÜLASYON SONUÇLARIYLA KARŞILAŞTIRILMASI: PAŞAKÖY ATIKSU ARITMA TESİSİ

ÖZET

Dünyadaki su kaynaklarının zamanla kirlenmesi, atıksu desarj limitlerinin arttırılması ve atıksulardan nutrient (azot, fosfor) gideriminin gerekliliğini de beraberinde getirmiştir. Diğer nütrient giderimi alternatiflerinin yanında aktif çamur teknolojisi ekonomik olması sebebiyle dünyada yaygın olarak kullanılmaktadır. Öngörülen desarj limitlerinin sağlanması da bu aktif çamur sistemlerinin şartlara uygun tasarımı, işletilmesi ve/veya optimizasyonu ile yerine getirilmektedir.

Aktif çamur modelleri artık aktif çamur sistemlerinin tasarımı, kontrolü ve optimizasyon çalışmalarında yaygın olarak kullanılmaktadır. Ancak, bu modellerin doğru ve verimli olarak kullanılabilmesi için (1) kesin model bazlı-atıksu karakterizayonu (2) arıtma tesislerine uygun modelleme yaklaşımı ve metodolojisinin uygulanması ve bunların sonucu olarak (3) elde edilen model sonuçlarının doğru olarak yorumlanması gerekmektedir.

Çalışmanın ilk aşamasında, işletme şartlarında atıksu karakterizasyonu belirlendi ve dizayn için persentil değerleri hesaplandı. Sonra, tesis farklı dizayn metodlarına göre (ATV-DVWK 131-E, Cape Town Üniversitesi ve Su Ortamında Araştırma Kuruluşu) tasarlanarak bunların sonuçları karşılaştırılmıştır.

Çalışmanın ikinci aşamasında, tam ölçekli tesis, simülasyon programı yardımıyla kararlı hal ve dinamik koşullarda modellendi ve model sonuçları desarj konsantrasyonlarıyla karşılaştırıldı.

Çalışmanın son aşamasında, uzun havalandırma tipi tam ölçekli aktik çamur tesisinin nütrient giderimi performansı modelleme çalışması ile incelenmiştir.

1. INTRODUCTION

1.1 Aim of study

The Biological Nutrient Removal (BNR) process is known to be the most convenient and economical alternative among other wastewater treatment methods to reduce the impact of nutrient (N, P) discharges. In this respect, the activated sludge technology has become more and more pronounced for meeting the strict nutrient and organic carbon discharge standarts imposed. However, the most efficient way of sustainable nutrient removal must be achieved by optimal design and control of such activated sludge systems. From an engineering point of view, an appropiate design, operation and retrofit of activated sludge plants reuires a beter understanding of the complex biological reactions taking place in activated sludge systems. The degree of complexity increases in paralel to the vast progress in biotechnology. In addition to that, the performance of activated sludge systems is generally influenced by dynamic conditions generally having an adverse impact on the effluent quality. Thus, the evaulation of those complex biological reactions together with dynamic factors requires computer aided model solutions to provide better insight in process dynamics and robust design and/or upgrade options for activated sludge plants.

Until recently, the activated sludge models have become convenient and popular tools design, operation and upgrade of activated sludge systems for organic carbon and nutrient removal. In reality, from a practial point of view, the model simulations enable to visualize a number of process scenarios under various conditions in a short period of time compared to a trial and error methodology. However, these activated sludge models strictly require (a) accurate model-based influent wastewater characterization (e.g. COD fractionation) and (b) dynamic model calibration to make the model mimic the actual behavior of the system. To optimize the removal of organic carbon, nitrogen and phosphorus in activated sludge plants, detailed model-based evaluations are useful since the process efficiency mainly depends upon the dynamics of the influent wastewater characteristics, the environmental factors and operating conditions.

The experimental data obtained from lab-scale setups, pilot-scale setups and full-scale plants provide ample information on process stoichiometry and kinetic to be used in activated sludge model itself.

In the first part of this thesis, the wastewater characterization of the plant was characterized and statistical analyzed for 2008. The generation of a mathematical approach was developed for the model-based COD characterization (i.e. readily and slowly biodegredable substrate) and the estimation of model parameters using batch respirometric data obtained with real wastewaters. After determining the wastewater characterization, the facility was designed with three different methods and the results were compared.

The second part deals with the steady-state and dynamic modelling of the full-scale plant. First, the organic carbon and nitrogen removal processes have been interpreted with the aid of calibrated model. Second, a stepwise model calibration methodology, model-based process analysis and a robust optimization methodology approach were proposed for Johannesburg configuration removing carbon, nitrogen and phosphorus simultaneously.

In the last part, the design and modeling results were evaulated for optimal process operation.

1.2 Scope of the Study

The scope of this thesis, modeling of the plant in steady-state and dynamic simulations. The results will be evaluated at the operating conditions and the affect of the results on design for optimal process operation. So, the differences between model results and design results will be compared and optimal operating conditions could be determined.

2. LITERATURE

2.1 Introduction

The nitrogen and phosphate originating from domestic and industrial discharges cause eutrophication problems in receiving water bodies. Generally, the eutrophication problem limits the potential use of receiving water due to diurnal algal activity unless the input of nutrients is reduced and/or controlled based on legislation. To protect water bodies from eutrohication, in Europe, the EEC Directive 91/271 (CEC, 1991) enforces discharge standarts with respect to total nitrogen and phosphate within sensitive areas. A-cost effective and sustainable nutrient discharge reduction into receiving waters can be guaranteed by appropiate utilization of the activated sludge process that has already been successfully applied for biological nutrient removal. Sustainable nutrient removal in an efficient way can be achieved by optimal design and control of activated sludge systems built for nutrient (N,P) removal. From an engineering point of view, depending upon environmental disturbances, an appropiate design and operation of activated sludge plants requires better understanding of the complex biological properties of activated sludge systems.

2.2 Processes selection for Nutrient Removal

2.2.1 Influent wastewater characteristics

Wastewater characterization, including rbCOD measurements, is essential to evaluate fully the design and performance of BPR systems. Biological phosphorus removal is initiated in the anaerobic zone where (acetate and propianate) is taken up by phosphorus-storing bacteria and converted to carbon storage products that provide energy and growth in the subsequent anoxic and aerobic zones. The rbCOD is the primary source of volatile fatty acids (VFAs) for the phosphorus-storing bacteria. The conversion of rbCOD to VFAs occurs quickly through fermentation in the anaerobic zone and 7 to 10 mg acetate results in about 1.0 mg P removal by enhanced phosphorus removal (Wentzel et al., 1989; Wentzel et al., 1990).

The more acetate, the more cell growth, and, thus, more phosphorus removal. Because of the need for organic material for nitrate removal, the amount of rbCOD relative to the amount of TKN in the effluent is also an important wastewater parameter.

The diurnal variation in the wastewater strength is also an important process consideration because the performance of phosphorus-storing bacteria depends on the availability of fermentation substrates, it is important to know if periods of low influent wastewater strength may affect BPR performance. For domestic wastewaters, the influent total BOD and rbCOD concentrations will vary with time over a 24-h period, with lower concentrations in the late evening and early morning hours. For smaller-sized communities, the variations are usually more pronounced and very little rbCOD may be present at certain times. During wet-weather conditions, especially in the winter, BPR may be difficult to achieve due to cold, low strength wastewater that does not readily become anaerobic. Extended periods of reduced rbCOD concentration have been reported to decrease BPR performance for a number of hours after the occurrence of low substrate concentration (Stephens and Stensel, 1998). The impact of continuous acatate for the plants where sludge fermentation has been done to produce additional VFAs has shown the benefit of a steady supply of rbCOD for biological phosphorus removal. In the modified Bardenpho trains at Kelowna, Canada, one train was fed fermentation liquor and the other train was used as the control. With continuous VFA addition, the effluent soluble phosphorus concentration decreased from 2.5 to 0.3 mg/l (Oldham and Stevens, 1985), and the VFA/P ratio was 6.7 g/g, an amount lower than the estimated 7 to 10 g/g. Based on these results, it appears that continuous acetate addition may provide more efficient biological phosphorus removal.

Ratios of wastewater organic matter to nutrient

The concentration of biodegradable organic matter relative to the nutrient concentrations in an influent wastewater can dramatically affect the performance of a BNR system. This is because of the key role biodegradable organic matter plays in nutrient removal. Nitrogen removal is accomplished when biodegradable substrate is used as the electron donor by denitrifying bacteria under anoxic conditions.

Phosphorus removal is accomplished when VFAs, which are either a part of the influent readily biodegradable substrate or are formed from it, are taken up and stored by PAOs in the anaerobic zone, thereby allowing them to increase the phosphorus content of the MLSS in the anaerobic zone (Grady et al., 1980). Table 2.1 provides general guidance concerning the amenability of various wasteaters (characterized in terms of the amount of nitrogen) to biological nitrogen removal. The values given can be used to screen candidate wastewaters to determine how difficult it may be to achieve good nitrogen removal.

Table 2.1: Relationship between expected biological nitrogen removal efficiency and influent organic matter to nitrogen ratios (Grady et al., 1980)

Nitrogen removal efficiency	COD/TKN	BOD_5/NH_3-N	BOD_5/ TKN
Poor	<5	<4	<2.5
Moderate	5-7	4-6	2.5-3.5
Good	7-9	6-8	3.5-5
Excellent	>9	>8	>5

A carbon limited wastewater is one in which insufficient organic matter is available to removal all of the phosphorus. As a sequence, phosphorus will be present in the process effluent at a concentration determined by the ralative concentrations of phosphorus and organic matter in the influent. A phosphorus limited wastewater is one in which more than sufficient organic matter is available to remove the phosphorus. Consequently, the effluent phosphorus concentration will generally be low when it is trated in a BPR process.

Table 2.2: Summary of observed influent BOD and COD to phosphorus removal ratios for different BPR processes (Metcalf Eddy,1999)

Type of BPR process	BOD/P ratio (g BOD/g P)	COD/P ratio (g COD/g P)	COD/TKN (g COD/g N)	SRT (d)
AO (Phoredox)	15-20	26-34	>12-15	2-5
A^2O	20-25	34-43	>12-15	5-25
Bardenpho	>25	>43	>11	10-20
Standart UCT (VIP)	15-20	26-34	>7-8	5-10
Modified UCT	20-25	34-43	>9	10-25

2.2.2 Plant configuration

2.2.2.1 AO (Phoredox) process

Enhanced biological phosphorus removal can be achieved with A/O type systems, widely known as Phoredox systems. This type of plant consists of anaerobic (A) reactor prior to the aerobic reactor (O) operated similar to predenitrification type plants. However, the nitrification is hindered by operating the plant under very low sludge ages (2 to 4 days).

Without any internal recycle stream, the first reactor becomes anaerobic since no nitrate is generated in the aerobic reactor (Barnard, 1974). The enhanced biological phosphorus removal, EBPR is promoted via uptake of volatile fatty acids in anaerobic reactor. The system is not operated for denitrification since there are no anoxic zones in the system layout.

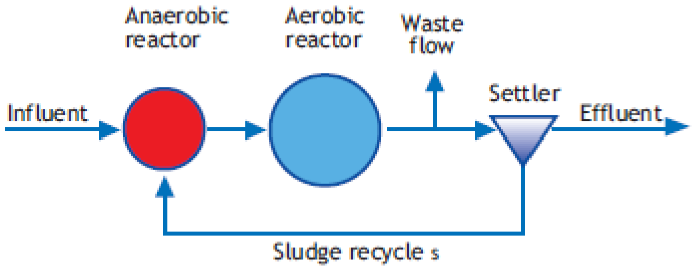


Figure 2.1: Plant layout for Phoredox (A/O) system

2.2.2.2 A^2O process

The difference of A^2O and the A/O process is the presence of an additional anoxic reactor between the anaerobic and aerobic reactor. This layout is the modification of the A/O system designed for denitrification together with EBPR. The anoxic zone reduces the nitrate load to the anaerobic compartment. The hydraulic retention time for the anoxic reactor is selected approximately around 1 hour (Metcalf and Eddy, 2003).

2.2.2.3 5-Stage modified Bardenpho process

The Bardenpho process can be modified for combined nitrogen and phosphorus removal. The staging sequence and recycle method are different from the A^2O process. The 5-stage system provides anaerobic, anoxic and aerobic stages for phosphorus, nitrogen and carbon removal. A second anoxic stage is provided for additional denitrification using nitrate produced in the aerobic stage as the electron acceptor, and the endogenous organic carbon as the electron donor. The final aerobic stage is used to strip residual nitrogen gas from solution and to minimize the release of phophorus in the final clarifier. Mixed liquor from the first aerobic zone is recycled to the anoxic zone. The 5-stage process uses a longer SRT (10 to 20 days) than the A^2O process, and thus increases the carbon oxidation capacity.

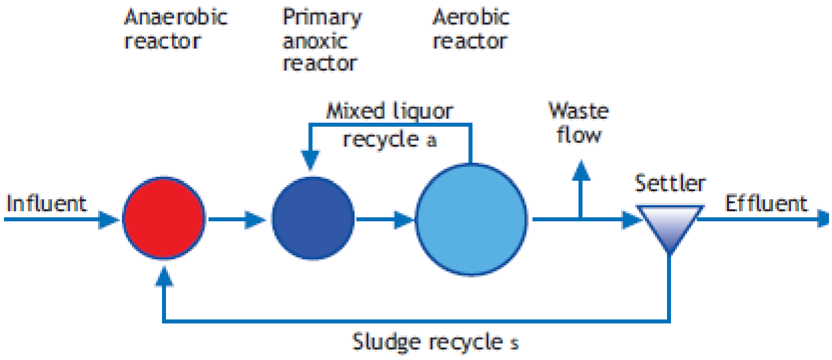


Figure 2.2: Plant layout for A^2O system

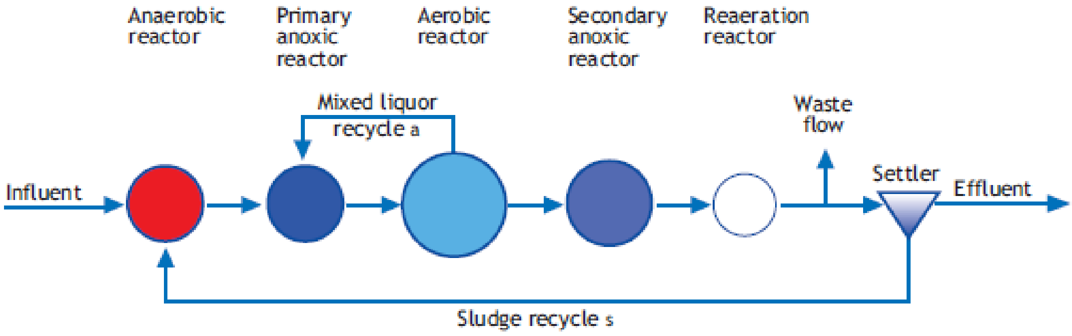


Figure 2.3: Plant layout for Bardenpho system

2.2.2.4 Johannesburg process

Another alternative is the johannesburg type activated sludge plant.In comparison to UCT, the RAS is diverted to a pre-anoxic reactor where the nitrate is denitrified by endogenous activity of biomass. The appropiate selection of sufficient reactor configuration and volume allows a reduced nitrate load in the anaerobic reactor. In the anaerobic reactor, higher MLSS concentration can be maintained compared to the UCT process.

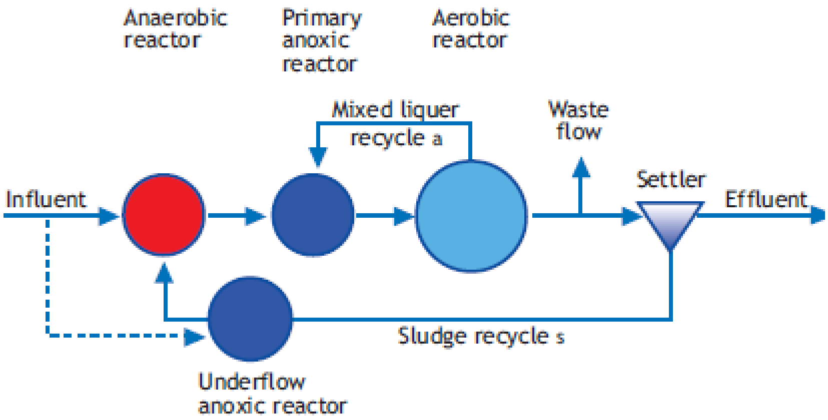


Figure 2.4: Plant layout for Johannesburg process

2.2.2.5 Standart UCT (VIP)

The UCT process was developed by the research group of the University of cape Town (South Africa) for EBPR from weak wastewaters. The objective was to minimize the amount of nitrate in the anaerobic reactor, which is critical for EBPR. In order to do so, the nitrate from the settler is recycled to the anoxic reactor ($2*Q_{influent}$). An additional denitrified recycle, DNR (see figure) is introduced from the anoxic to the anaerobic compartment. Since the mixed liquor is at a lower biomass concentration, the retention time for anaerobic reactor should be longer than that of the Phoredox systems. The retention time is generally in the range of 1-2 hours for the anaerobic reactor. In the Modified UCT process, the anoxic reactor is seperated into 2 anoxic compartments. The first and second compartments receive the return activated sludge RAS and nitrified recycle, respectively.

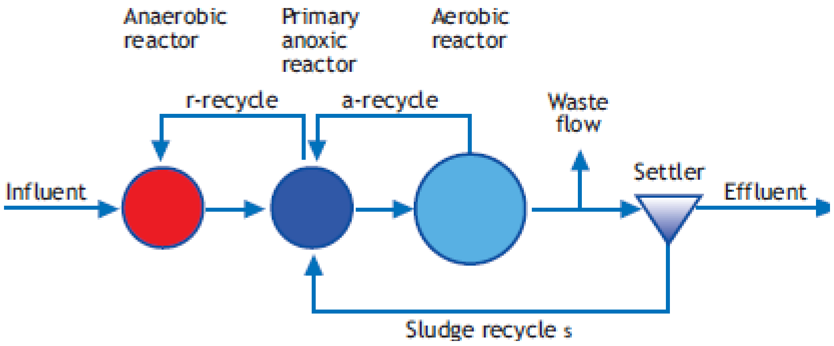


Figure 2.5: Plant layout for UCT (VIP)

2.2.2.6 Modified UCT

In the Modified UCT process, the anoxic reactor is seperated into 2 anoxic compartments. The first and second compartments receive the return activated sludge RAS and nitrified recycle, respectively.

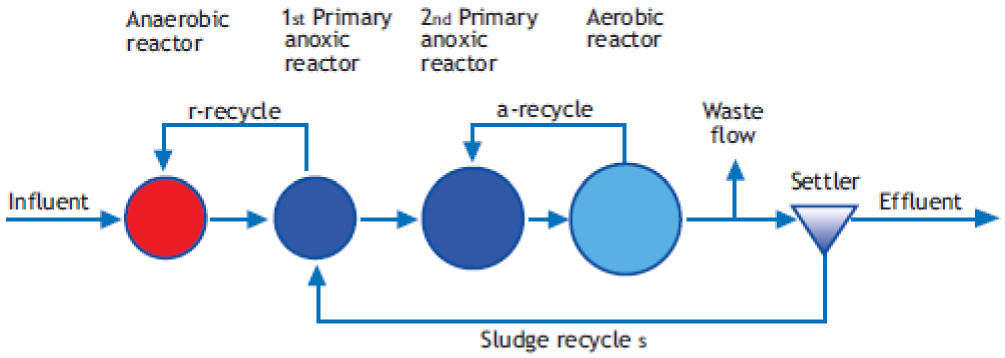


Figure 2.6: Modified UCT process

2.2.2.7 SBR with biological phosphorus removal

If sufficient nitrate is removed during the SBR operation, an anaerobic reaction period can be developed during and after the SBR fill period. An anoxic operating period is used after a sufficient aerobic time elapses for nitrification and nitrate production. Alternatively cyclic aerobic and anoxic periods can be used during the react period. The nitrate concentration is thus minimized before settling, and little nitrate is available to complete for rbCOD in the fill and initial ract period. Thus, anaerobic conditions occur in the fill and initial react period, so that rbCOD uptake and storage by phosphorus-accumulating bacteria can occur instead of rbCOD consumption by nitrate-reducing bacteria.

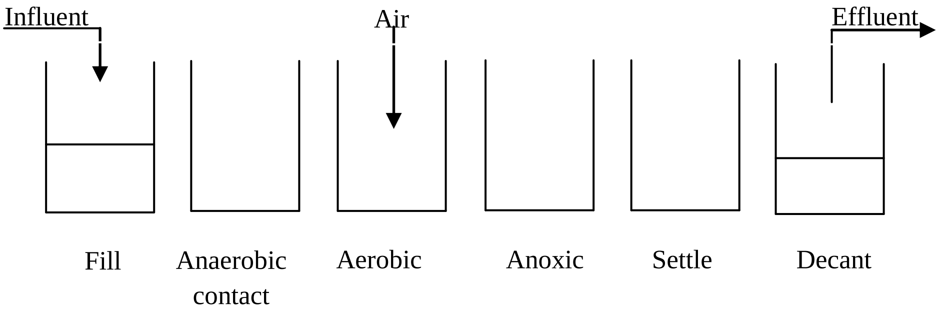


Figure 2.7: Layout of SBR

The design and operating parameters for commonly used nitrogen and biological phosphorus-removal processes are given in Table 2.3. According to the table, sludge retention time (SRT) covers a wide range for the processes of nitrogen and phosphorus removal. If only phosphorus removal is considered (e.g. A/O, A^2O processes), lower SRT values are selected in order to suppress nitrification since the nitrate consumes required VFA for EBPR. The mixed liquor suspended solids (MLSS) concentration to be maintained in biological reactor is generally around 2-4 mg/l for nutrient removal processes. For phosphorus removal, in parallel to reduced SRTs, lower HRTs are sufficient for A/O, A^2O processes to maintain desired MLSS in biological reactor. For nitrogen and phosphorus removal, SBRs are more flexible in terms of HRT selection. Broader return activated sludge (RAS) range of 25%-100% corresponds to A/O ve A^2O processes because of the fact that the nitrate load through the anaerobic compartment should be reduced for better EBPR efficiency. Mostly, a minimum RAS value 50% (sludge thickening factor) is preferred in order not to sludge blanket build up in the final clarifier. The Bardenpho, UCT and VIP processes require high internal recycle rates since the process train is composed of multi-staged reactors. In the operation, higher internal recycle rates are necessiated from aerobic to anoxic, anoxic to anaerobic reactors for those processes. In oxidation ditches, higher internal recycles are maintained in closed loop bioreactor with the aid of mixers.

Table 2.3: Design and operating parameters for commonly used nitrogen and biological phosphorus removal processes (Metcalf and Eddy, 2003)

Process	SRT (day)	MLSS (g/l)	HRT (h)			RAS (%)	IR (%)
			Anaerobic	Anoxic	Aerobic		
A/O	2-5	3-4	0.5-1.5	-	1-3	25-100	
A^2O	5-25	3-4	0.5-1.5	0.5-1	4-8	25-100	100-400
UCT	10-25	3-4	1-2	2-4	4-12	80-100	200-400 (anoxic) 100-300 (aerobic)
VIP	5-10	2-4	1-2	1-2	4-6	80-100	100-200 (anoxic) 100-300 (aerobic)
Bardenpho	10-20	3-4	0.5-1.5	1-3 (1st stage) 2-4 (2nd stage)	4-12 (1st stage) 0.5-1 (2nd stage)	50-100	200-400

*Sludge Retention Time (temperature dependent)

3. MATERIAL AND METHODS

3.1 Conceptual Approach

Within the scope of the work of wastewater treatment plant design and modeling, Paşaköy biological nutrient removal plant was selected treatings the wastewaters to protect the most important drinking water river basin, Ömerli, located in İstanbul. The steps of the study were summarized as follows;

- First, the influent wastewater characterization (T-P, T-N, COD, BOD, TSS and VSS) for the year of 2008 analysis results were used**.** Using these values, the facility was designed with different 3 methods (ATV-DVWK 131-E, Water Environment Research Foundation and South African).
- The wastewater characterization of Paşaköy WWTP was used to compare the calculations of the methods.
- Then, the facility was modelled with selected activated sludge model (Barker and Dold, 1997) under steady-state (yearly based average) and dynamic conditions (daily basis dynamics). In the following step, the modeling results were with experimental data design results.
- Finally, the calibrated model results were compared with process design results. The calibrated model was used to analyse the success of BNR operation perspective of process.

3.1.1 Characterization protocols

Driven by requirements of mathematical modelling of activated sludge systems, several systematic protocols for activated sludge model calibration were developed and include different wastewater characterization protocols. Four major protocols were developed by many research groups. The nature of these protocols range from simplified and rather practical, to those of increased complexity and more of academic and research interest.

- the STOWA protocol (Hulsbeek et al.,2002)

- the BIOMATH protocol (Vanrolleghem et.al., 2003)
- the WERF protocol for model calibration (Melcer et al., 2003)
- the Hochshulgruppe (HSG) guidelines (Langergraber et.al., 2004)

Due to its simlicity the STOWA calibration protocol was applied to Paşaköy WWTP for the characterization.

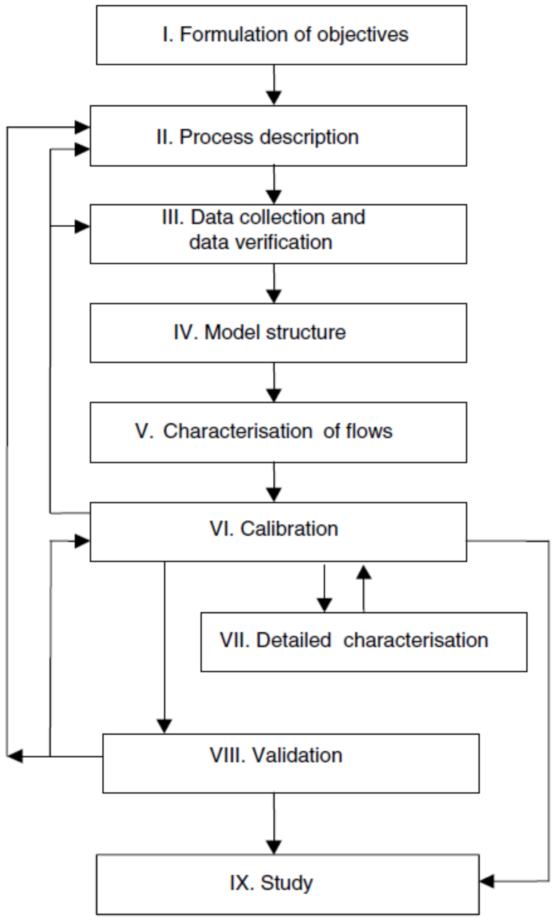


Figure 3.1: Main structure of the STOWA protocol (Hulsbeek et.al., 2002)

3.1.2 Process description

When the objectives of the study are clear, a definiton of the relevant process components can be made. Often it is not required to model the complete WWTP. Only those parts that fit within the described process dynamics are useful to consider in the model. If the distribution of sludge/water in systems with paralel lanes is well balanced, all streets can be modelled in the same way. If the distribution is not well balanced, each lane has to be modelled separately. In most cases it is only required to describe the activated sludge process (including secondary clarifiers). All in-and outgoing flows (i.e. influent, recirculation flows, internal flows from sludge treatment processes and effluent) must be defined.

3.1.3 Data collection and verification

In this phase the composition and the volume of the flows to the different process components, as well as the volume of the process components are defined. It is advised to gradually refine definitions, starting off with rough or approximating values and fine-tune them in the process. In general, many data of the WWTP are available (daily average concentrations, flow patterns). In protocol it is advised, to initially generate the compositions and the flows from the available data. If necessary, lacking relevant data can then be obtained with extra monitoring. After a first set of simulations a better directed monitoring programme can be developed. Concentrations which change most at spesific points in the treatment plant could be evaluated in detail whereas other values can be based on daily (flow proportional) averages.

3.1.4 Characterisation of the main flows

By using historical data and/or spesific measurements, the important process flows can be characterised. These flows include influent, effluent, centrate water from dewatering unit and internal recycle flows. The guidelines for influent characterization are the basis for the characterization (Roeleveld, 2001) of the different flows. If the model is used for a system choice, daily avarage concentrations of influent and effluent and the variations in the flow pattern are sufficient. If the model is used for optimisation, the development of control strategies spesific data from 4 or 2 hour composite samples is required.

3.1.5 Model Structure

Selected Model (Barker and Dold, 1997)

A mechanistic model for NDEBPR systems necessarily must account for a large number of biological processes to mimic the complex interactions that may affect the performance of a given treatment plant. Certain keys features of process behavior are discussed briefly before presenting the model in detail. The processes of the model were summarized as follows:

Anoxic Growth of PolyP organisms; A simplified representation of the behavioral patterns associated with phosphorus release and uptake in BEPR system**.** It is readily apparent that nitrate plays an important role in the performance of these systems. The current consensus interpretation is that nitrate entering the anaerobic zone will be used as an electron acceptor in the growth of non-polyP heterotrophs. As a result less substrate will be available for sequestration by the polyP organisms, with the net effect that the amount of P removal will be reduced. The amount of substrate available to the polyP organisms may be reduced to such an extent that these organisms are unable to sustain themselves in the system, and the capacity for BEPR is lost.

Denitrification by polyP organisms was excluded from the Wentzel *et al*., (1989a) model as observations showed minimal denitrification occured in laboratory systems comprising mainly polyP organisms. However, a recent review of experimental studies concerning denitrification behavior in BEPR activated-sludge systems (Barker and Dold, 1996a) indicates that P uptake by polyP organisms does occur in anoxic zones of nutrient removal systems. Based on the results of microbiological studies, as well as many continuous and batch reactor experimental studies, a number of conclusions were drawn;

- PolyP organisms are capable of concominant denitrification and P uptake.
- Nitrate can serve as an electron acceptor for the oxidation of stored PHB; however, not all polyP organisms capable of reducing nitrate appear able to use nitrite as an electron acceptor.
- Batch tests indicate that more stored carbon (PHB, PHV) is used for a given amount of P taken up when nitrate is the electron acceptor in place of oxygen.
- P uptake/PHB oxidation appear to occur simultaneously with P release/PHB storage when SCFAs are available under anoxic conditions. The relative rates of these processes, and the endogenous lysis of P, will determine whether or not a release or uptake of P is observed in an anoxic reactor.

For modeling denitrification by polyP organisms, it is assumed that a fraction (η_p) of the polyP organisms can use nitrate as an electron acceptor in the absence of oxygen for oxidation of stored PHB and uptake of phosphorus. In the model of Wentzel *et al.*, (1989a), there are four aerobic growth processes; the four permutations reflect switches between ammonia and nitrate as the nitrogen source for synthesis and at high and low soluble phosphorus concentrations. Strictly all four growth processes should be duplicated in the model for anoxic conditions. However, in an anoxic reactor of a continuous-flow system, growth is likely to occur in the presence of sufficient ammonia and soluble phosphorus. Therefore, in evaluating the possibility of anoxic growth, only the one aerobic growth process was duplicated for anoxic conditions, that is, growth with ammonia as the N source for synthesis and with no phosphorus limitation on the growth rate. The stoichiometric coefficient for phosphorus uptake ($f_{P,UPT}$), defined as the ratio of P taken up to PHB oxidized,was determined from batch experiments to be approximately 0.9 to 1.1 g P (Wentzel *et al.*, 1989a).

Conversion of Soluble Readily Biodegradable COD to SCFA; The principal linkage between the polyP and non-polyP heterotrophic organism masses in BEPR systems treating municipal wastewater is the conversion, by non-polyP, of complex readily biodegradable soluble COD (S_{BSC}) to SCFAs (S_{BSA}) under anaerobic conditions. This process is the source of SCFA to sustain polyP organism growth in the mixed-culture system as in the influent SCFA content usually is minimal.

A study of COD and nitrogen mass balances in activated-sludge systems (Barker and Dold, 1995) suggests that there is a significant ‘loss’ of COD in activated sludge systems incorporating anaerobic zones. Four different types of laboratory-scale system were studied: aerobic, anoxic, anoxic-aerobic, and anaerobic-anoxic-aerobic. The systems included a variety of configurations, with differing wastewater characteristics and operating parameters. The results suggest that although good COD balances are to be expected in aerobic and perhaps in anoxic-aerobic systems, systems incorporating anaerobic zones (such as EBPR systems) tend to exhibit low COD balances (<80%) (anoxic-only systems also appear to exhibit a loss of COD, but to a lesser extent). It would appear that this ‘loss’ of COD apparently is associated with the fermentation processes occurring in the anaerobic zone of BEPR systems treating municipal wastewater. Whether this COD loss is a direct result of fermentation (for example, through the generation of gas that evolves during the actual fermentation process) or an indirect result (for example, through the production of volatile compounds that are released from the system under aerated conditions) remains to be determined. Regardless, this ‘disapperance’ of some 20% of the influent COD in BEPR systems translates into reduced sludge production and oxygen demand compared to non-BEPR systems (Barker and Dold, 1996b). This feature alone strengthens the case for biological nutrient removal.

Initially, it was surmised by Barker and Dold (1995) that COD loss was induced by the inclusion of an anaerobic zone and that all COD loss was releated to the fermentation process in the anaerobic zone.

Sequestration of Short-Chain Fatty Acids by PolyP Organisms; In the anaerobic sequestration of SCFA by polyP organisms (for PHB storage, with associated phosphate release), it is assumed that the yield of PHB is Y_{HB} units of PHB (as COD) per unit SCFA COD taken up. A value of 0.89 g COD is suggested for Y_{PHB} based on the assumption that for an initial amount of 2.25 moles acetate, 2 moles enter the PHB formation pathway directly and 0.25 moles are directed to the TCA cycle (Wentzel *et al.*, 1986). That is, the model also incorporates COD loss in the sequestration reaction. This provides a second mechanism for ‘losing’ COD and allows the general model to mimic the COD balances (90%) observed in the enhenced cultures. It should be noted, however, that this COD loss is not suggested in the biochemical model because it is assumed that the available electrons from way from the TCA cycle (Barker and Dold, 1996b).

Nitrogen Source for Cell Synthesis; Reviewing the ASM1 model, Dold and Marais (1986) postulated that under certain circumstances, nitrate, instead of ammonia nitrogen, may serve as the nitrogen source for cell synthesis purposes. This postulate was confirmed from analysis of data collected over an extensive period, particularly in multiple series reactor configurations operated at long sludge ages that exhibited high nitrification rates.

Growth of non-PolyP Heterotrophs on Short-Chain Fatty Acid; For BEPR systems, it is necessary to distinguish between complex and SCFA readily biodegradable COD. Therefore, it is necessary to duplicate the four growth processes referred to above to account for possible growth on the two components of the readily biodegradable COD for the mixed-culture system. With regard to growth on SCFA, it is likely that only one the of four processes would be of consequence-anoxic growth with ammonia as the N source. This is because SCFAs are removed in the unaerated zones at the front end of the continuous-flow systems and do not enter the aerobic zones in appreciable concentrations.

Hydrolysis/Solubilization of Slowly Biodegradable COD; In the ASM1 model, the biodegradable material is divided into a readily biodegradable fraction (S_{ENM}). The readily biodegradable fraction is hypothesized to consist of material that can be absorbed readily by the organism and metabolized for energy and synthesis, whereas the slowly biodegradable fraction is assumed to be made up of particulute material and complex organic molecules that require extracellular enzymatic breakdown prior to absorbtion and utilization. In the ASM1 model, the rate of solubilization under anoxic conditions is assumed to be reduced by a factor η_{SOL} compared with the rate under aerobic conditions. Under anaerobic conditions the rate is assumed zero. Recent research on enzymatic hydrolysis (Dold *et al.*, 1991, and San Pedro *et al.*, 1994) indicates that hydrolysis does in fact occur under anaerobic conditions, and under anoxic conditions, the rate of hydrolysis appears similar to that under aerobic conditions. To provide flexibility in the model, η_{SOL} factors are incorporated: $\eta_{S,ANOX}$ and $\eta_{S,ANA}$. In addition, two hydrolysis efficiency factors, E_{ANOX} and E_{ANA}, have been included to allow for the possibility of COD loss during the breakdown of the enmeshed slowly biodegredable material to readily biodegradable material. That is, in hydrolysis of one COD unit slowly biodegradable COD, there is a production of E units of S_{BSC} and a loss of (1-E) units of COD. This third mechanism for COD loss allows the model to simulate the COD loss observed in anoxic-aerobic and anoxic-only systems.

Decay of PolyP Organisms Under Anoxic Conditions; The Wentzel enhanced culture model (Wentzel *et al.*, 1989a) did not consider anoxic behavior of PolyP organisms, and decay processes for these organisms were considered only for aerobic and anaerobic conditions. Processes for anoxic decay of PolyP organisms, stored polyphosphate lysis from anoxic decay have been included here (three additional processes).

Yield of Heterotrophs in Anoxic Growth; The ASM1 model assumes a single yield coefficient Y_H for non-polyP heterotrophs irrespective of whether oxygen or nitrate serves as the electron acceptor. It has long been surmised that the yield of organisms under anoxic conditions with nitrate as electron acceptor is lower than for aerobic growth. This does not have a significant effect on ASM1 model predictions of sludge production for ND systems treating municipal wastewater. For this situation, the influent TKN:COD ratio usually is relatively low (<0.12 g N/g COD), so the amount of nitrate generated is limited, and hence the organism mass generated in aerobic growth. In four of the systems, sludge production in the anoxic unit was significantly lower than in the corresponding aerobic one. It was concluded that the yield coefficient for anoxic growth was approximately 40% less than for aerobic growth.

Switching Functions for Phosphorus Limitation; Under aerobic (and anoxic) conditions, soluble phosphate serves as a P source for synthesis of the different organism masses in growth processes. Also, P is being taken up for EBPR. The Wentzel *et al.*, (1989b) model incorporates a switching function that causes the P uptake processes to switch off when soluble P becomes limiting. The same switching function thereshold concentration is used to switch off the growth processes when soluble P becomes limiting. Evidence from simulation of full-scale systems indicates that these processes have different threshold values for switching on and off at low P concentrations.

Releasable/Fixed Polyphosphate Components; Experimental observations indicate that stored polyphosphate is portioned between a low-and a high-molecular-weight fraction ($P_{PP\text{-}LO}$ and $P_{PP\text{-}HI}$), and that only the low-molecular-weight fraction can be released after being taken up (Mino *et al.,* 1984). In modeling the P uptake/release processes, it is assumed that only a fraction (f_{PP}) of the phosphate stored as poly-phosphate can be released in a subsequent anaerobic condition. The value for f_{PP} of 0.94 suggested here was selected on the basis of simulating behavior in a range of BEPR systems and is in accordance with values reported in the literature (Mino *et al.,* 1984). The remainder of the stored polyphosphate is termed fixed.

3.2 Treatment Plant Information

Paşaköy advanced biological treatment plant has design capacity of is 250,000 pe and 100,000 m^3/d of domestic wastewater and designed for removal of organic matter (COD), nitrogen (N) and phosphorus (P). For the first construction treatment plant based on pre-anoxic denitrification process and designed as A2/O. The revision for increasing phosphorus removal, bio-P by-pass line has been constructed and started up to operated as Johannesburg configuration. The main parts of treatment process and sludge treatment are consisting of:

- Coarse screens
- Inlet pumping station
- Fine screens
- Grit chamber
- Parshall flume
- Selector-distribution chamber
- Anaerobic tanks
- Anoxic/Oxic tanks
- Final clarifiers
- Sludge dewatering

The flow scheme of the Paşaköy Advanced Biological Wastewater Treatment Plant was given in Figure 3.2. The main parts of the Paşaköy Advanced Biological Wastewater Treatment Plant shown in Figure 3.2 were investigated in terms of project criterias. The main parts of the plant;

1) Grit chamber
2) Bio-P tanks
3) Aeration tanks
4) Final clarifiers
5) Sludge dewatering units

Grit chamber; as can be seen from the Table 3.1., grit chamber have been designed in accordance with the criteria.

Table 3.1: Grit chamber design parameters

Unit/Equipment	Size	Unit	*Recommended
Number of tanks	2	-	-
Width	4.3	m	2.5-7
Length	19	m	7.5-20
Water depth	4.6	m	2-5
Length/ Width ratio	4.4	-	3:1 to 5:1
Detention time	15	minute	3-10
Blower numbers	2+1	-	-
Blower capacity	300	m^3/h	-

* Metcalf and Eddy, 2003

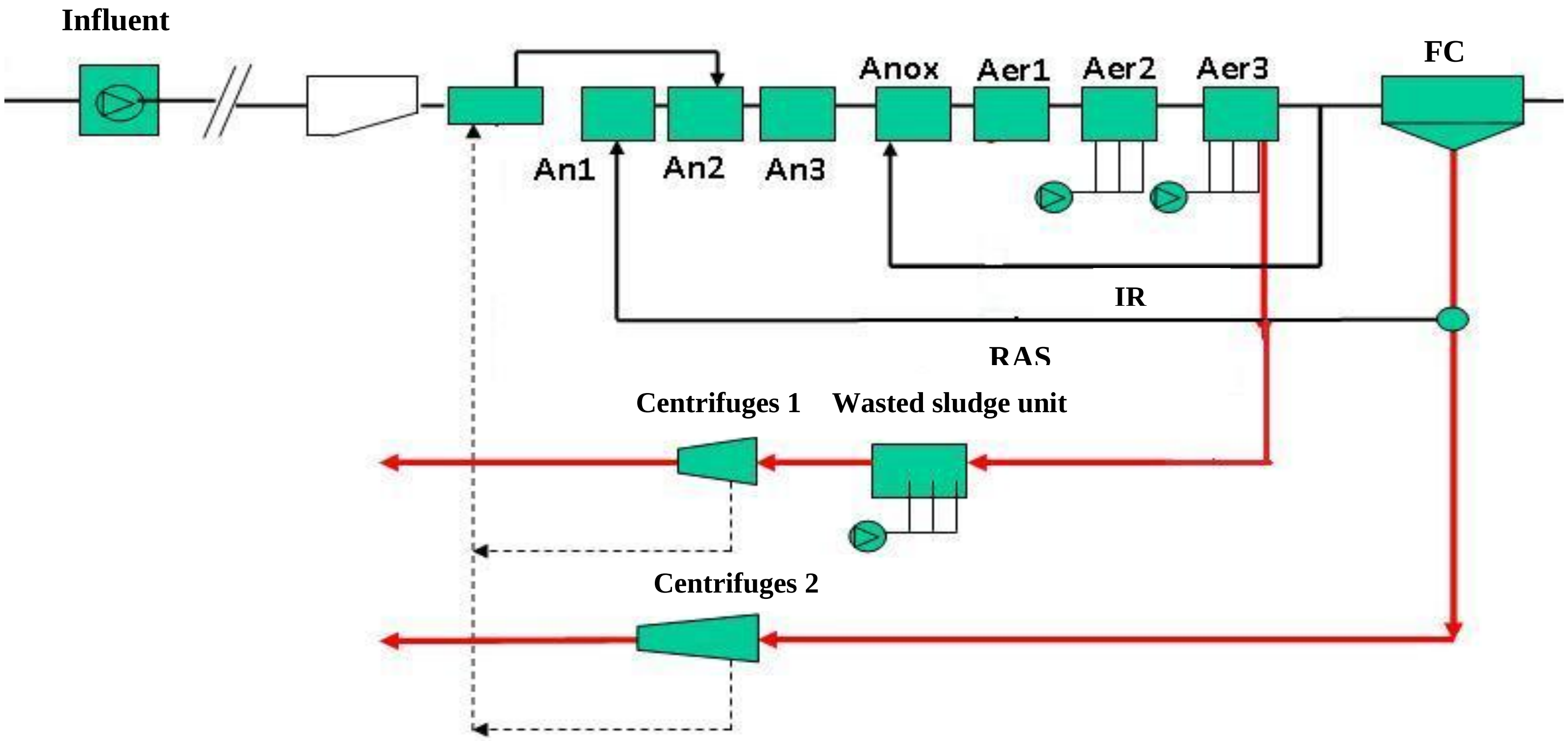


Figure 3.2: Flow scheme of Paşaköy WWTP (2008)

Bio-P tanks; bio-P tanks, three tanks connected in series as in project design but as a result of the revisions, the first tank is used for return activated sludge denitrification and the second tank is fed by influent wastewater. The design parameters of bio-P tanks are given in Table 3.2.

Table 3.2: Bio-P tanks design parameters

Unit/Equipment	Size	Unit
Tank configuration	Series	-
Number of tanks	3	-
Width	12	m
Length	48	m
Water depth	5.0	m
Hydraulic Retention time	2.5	h
Mixer numbers	6	-
Power mixer	1.7	w/m^3

Aeration tanks; aeration tanks is composed of four tanks and operated on the principle of pre-denitrification and simultaneous nitrification denitrification as illustrate in Figure 3.3. Tanks currently operated on the principle of pre-denitrification. After the Bio-P reactor, the mixed liquor is introduced to anoxic reactor with the anoxic fraction (V_D/V) of 25%. The oxygen distribution in the second anoxic reactor was manipulated to adjust the V_D/V ratio in the range of 25-50%. The other two reactors are kept as fully aerobic (DO=2.0 mg/L) to provide nitrification. The internal recirculation (Q_{IR}/Q) from the last aerobic to the first anoxic reactor was set to 3.8 in order to convey oxidized nitrogen for denitrification process as summarized in Table 3.3.

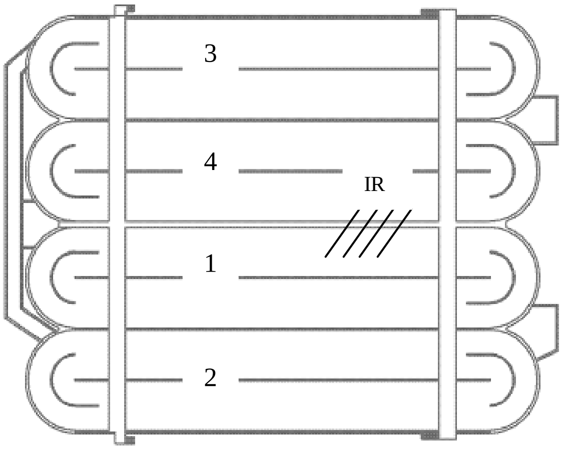


Figure 3.3: Aeration tanks

Table 3.3: Aeration tanks design parameters

Unit/Equipment	**Size**	**Unit**
Tank configuration	Sequence/Parallel	-
Number of tanks	4	-
Width	20	m
Length	110	m
Water depth	5.2	m
Total volume	44,000	m^3
Diffuser number	7,852	
1st tank	1,008	number
2nd tank	2,520	number
3rd tank	2,700	number
4th tank	1,624	number
Blower number	max. 51,000	Nm^3/h
Turbo (3)	4,950-11,000	Nm^3/h
Roots (2)	3,000-9,000	Nm^3/h
Internal recycle pump numbers	4	
Internal recycle pump capacity	3.8*Q	m^3/d

Final clarifiers; final clarifiers consists of four tanks which has 42 m diameter in a tank as given in Table 3.4. For return activated sludge, there are two pumping station each has five return acitvated sludge pumps.

Table 3.4: Final clarifiers project criterias

Unit/Equipment	**Size**	**Unit**
Number of tanks	4	-
Tank diameter	42	m
Total water depth	3.7	m
Total tank volume	20,495	m^3
Total surface area	5,538	m^2
Return sludge pump number	10	-
Return sludge pump capacity	540	m^3/h
Max. return sludge pump capacity	1.29*Q	m^3/d
WAS pump capacity	167	m^3/h
Max. WAS pump capacity	12,024	m^3/d

Sludge dewatering units; from the sludge treatment plant, decanter centrifuges with the addition of cationic polyelectrolyte is directly dewatered. The average percentage of solid matter for dewatered sludge is the order of 25% as shown in Table 3.5.

Table 3.5: Sludge dewatering equipments

Flotweg centrifuges	Capacity	17.5 m^3/h
	Capacity	350 kg solids/h
	Number	2
Westfalia centrifuges	Capacity	80 m^3/h
	Capacity	700 kg solids/h
	Number	2

3.2.1 Influent wastewater characterization

The most important point to multi-component modeling studies for activated sludge systems is determination of the wastewater characterization. In terms of modeling studies; to establish a balance between substrate, biomass and dissolved oxygen, it is necessary to determine the amount of COD as organic matter. However, COD parameter is not able to determine the organic matter as it's biodegradability and all modeling and design studies should be based on biodegradable organic matter, biological and inert fractions of biodegradable COD have to be determined.

Soluble inert COD (S_I), escapes with wastewater discharge without any biochemical reactions in the activated sludge system and the most important portion of the organic matter in the effluent wastewater.

Particular inert COD (X_I), is only wasted with sludge and accumulated depending on the sludge age.

The wastewater characterization under dry weather conditions of the facility for the year of 2008 is summarized in Table 3.6. The mean values is the results of the daily composite samples for collecting front part of the fine screens based on loads and the standart deviations and percentil values are the results of the statistical analysis. Especially, 85% percentil values widely used for design applications.

Table 3.6: Wastewater characterization of the facility for 2008

Parameter	COD (mg/l)	BOD_5 (mg/l)	TKN (mg/l)	T-P (mg/l)	TSS (mg/l)	VFA (mg/l)
Mean	520	282	59	7	380	20
STD	8	4,4	1	0,15	6	2
70% per.	582	330	65	9	470	27
85% per.	600	325	70	10	535	30

As the wastewater characterization generally evaluated for dry-weather, it suggests that in terms of COD medium-strong, nitrogen and phosphorus are strong domestic wastewater. According to the Table 3.7, the average COD/TKN and COD/TP ratios can be calculated as 8.0 and 66, respectively which are suitable for biological nutrient removal (Randall et al., 1992).

Table 3.7: The relation of parameters for the wastewater characterization of the facility

	COD/TKN	VFA/COD	NH_4-N/TKN	VFA/T-P	COD/ T-P
Mean	8.0	0.070	0.65	2.8	66
STD	0.12	0.002	0.11	0.1	1.2
70% per.	8.3	0.092	0.68	2.9	74
85% per.	8.5	0.112	0.70	3.0	86

The wastewater characterization of the facility shown in Figure 3.4 was statistical analyzed for winter and summer. The most of the days during winter months, TKN concentrations are approximately in the range of 40-60 mg/l, COD concentrations are 350-500 mg/l and T-P concentrations are 6-8 mg/l. The most of the days during summer months, TKN concentrations are approximately in the range of 60-80 mg/l, COD concentrations are 500-700 mg/l and T-P concentrations are 10-12 mg/l.

Average wastewater characterization for Paşaköy was compared with residential areas in Istanbul in Table 3.8 and some countries in Europe in Table 3.9. As the wastewater characterization compared for different residential areas for İstanbul, it has seen that in the table Paşaköy has the strongest wastewater characterization.

Table 3.8: Comparison of the wastewater characterization in Domestic Wastewater Treatment Plant at İstanbul by Orhon et al. (2000) and wastewater characterization for Ataköy by Okutman Tas et al. (2009)

Wastewater Treatment Plant	COD (mg/l)	BOD (mg/l)	TKN (mg/l)	T-P (mg/l)	TSS (mg/l)	COD/ TKN	NH_4/TKN	COD/T-P
Paşaköy	520	282	59	8.0	320	8.5	0.70	66
Kadıköy	450	220	49	8.1	310	9.2	0.62	55
Küçükçekmece	400	185	42	7.4	200	9.5	0.54	54
Baltalimanı	340	155	35	6.8	140	9.7	0.57	50
Ataköy	406	-	41	8.3	190	9.9	0.56	49

Table 3.9: Wastewater characterization in Europe (Pons et al., 2002)

Country	COD (mg/l)	TKN (mg/l)	T-P (mg/l)	TSS (mg/l)	COD/ TKN	COD/ T-P
Germany	548	59	8	208	9.3	68
France	634	52	9.3	302	12.2	68
Austria	526	44	7.1	-	12	74
Netherlands	450	42	6.7	237	10.8	67
Sweden	477	33	6.1	243	14.5	78
Turkey*	520	59	8.0	320	8.5	66

*Paşaköy Wastewater Treatment Plant

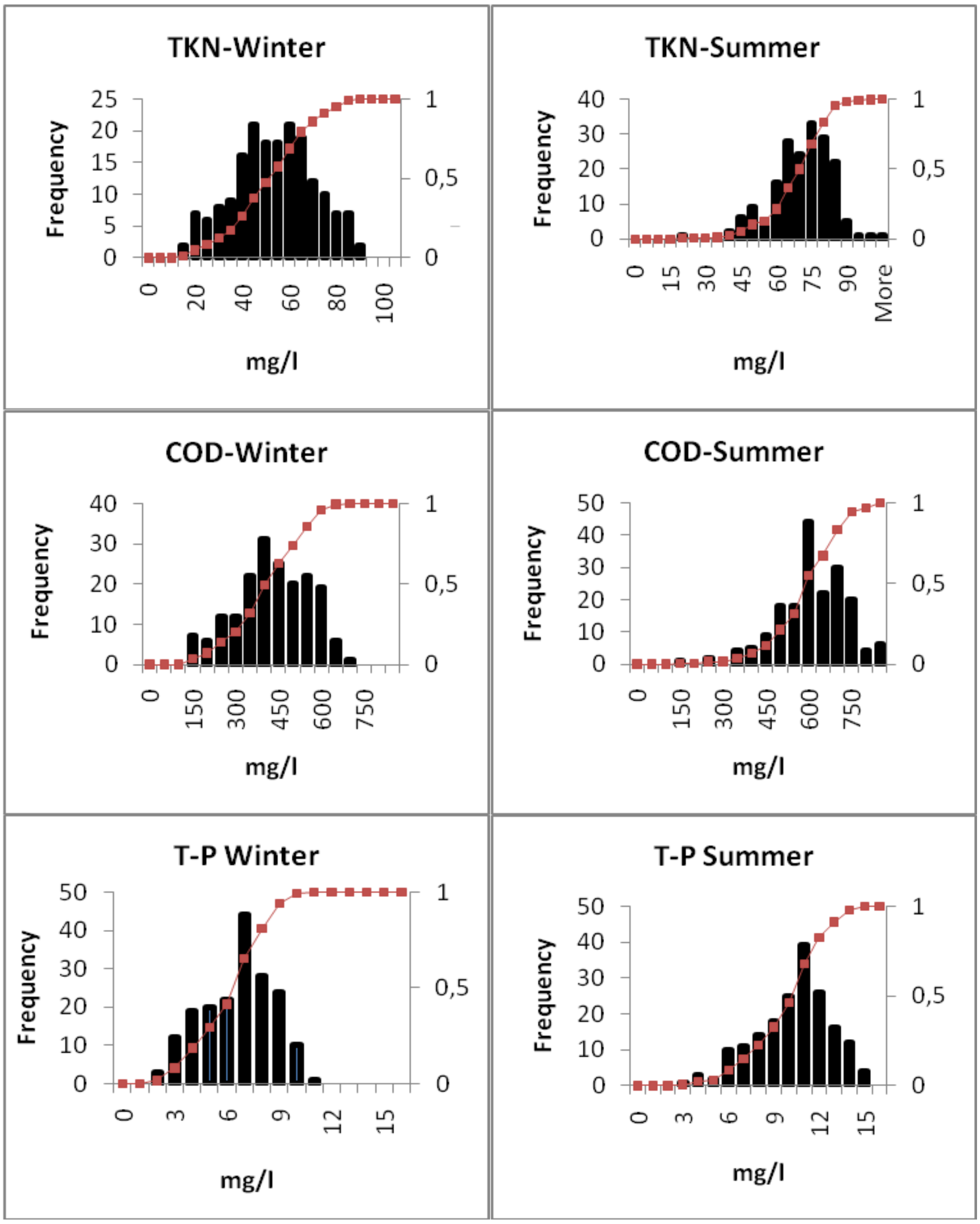


Figure 3.4: Histograms for influent parameters in summer-winter seasons in 2008

3.2.2 Design parameters using for design calculations

Design parameters using for design methods calculations for 2008 is presented in the Table 3.10.

Table 3.10: Design parameters of the plant

	Unit	**Measurement**
Influent Flowrate		
Dry weather flow, $Q_{DW,d}$	m^3/d	100,000
Dry weather flow, $Q_{DW,h}$	m^3/h	4,167
Peak Factor		1.25
Wet weather flow, $Q_{DW,d}$	m^3/d	125,000
Wet weather flow, $Q_{DW,h}$	m^3/h	5,208
Influent wastewater characteristics		
Total suspended solids, TSS	mgTSS/l	535
Biochemical oxygen demand, BOD_5	mg BOD/l	325
Total chemical oxygen demand, COD_T	mg COD/l	600
Particulate COD, X_{COD}	mg COD/l	420
Particulate inert COD, X_I	mg COD/l	50
Slowly biodeg. particulate COD, X_S	mg COD/l	370
Soluble COD, S_{COD}	mg COD/l	180
Soluble inert COD, S_I	mg COD/l	30
Readily biodegradable COD, S_S	mg COD/l	150
Fermentable COD, S_F	mg COD/l	120
Volatile fatty acid, S_{VFA}	mg COD/l	30
Total nitrogen, T-N	mg N/l	70
Ammonia, NH_4-N	mgN/l	49
Organic nitrogen	mgN/l	21
Nitrate	mgN/l	0
Total phosphorus, T-P	mgP/l	10
Influent Loads		
Total suspended solids loads	kg/d	53,500
BOD_5 loads	kg/d	32,500
COD loads	kg/d	60,000
T-N loads	kg/d	7,000
T-P loads	kg/d	1,000

3.3 The Algorithms of Different Design Methods

3.3.1 Design in ATV- DVWK 131E method

The Algorithm used for the design in ATV- DVWK 131E method is sumarized in Figure 3.5.

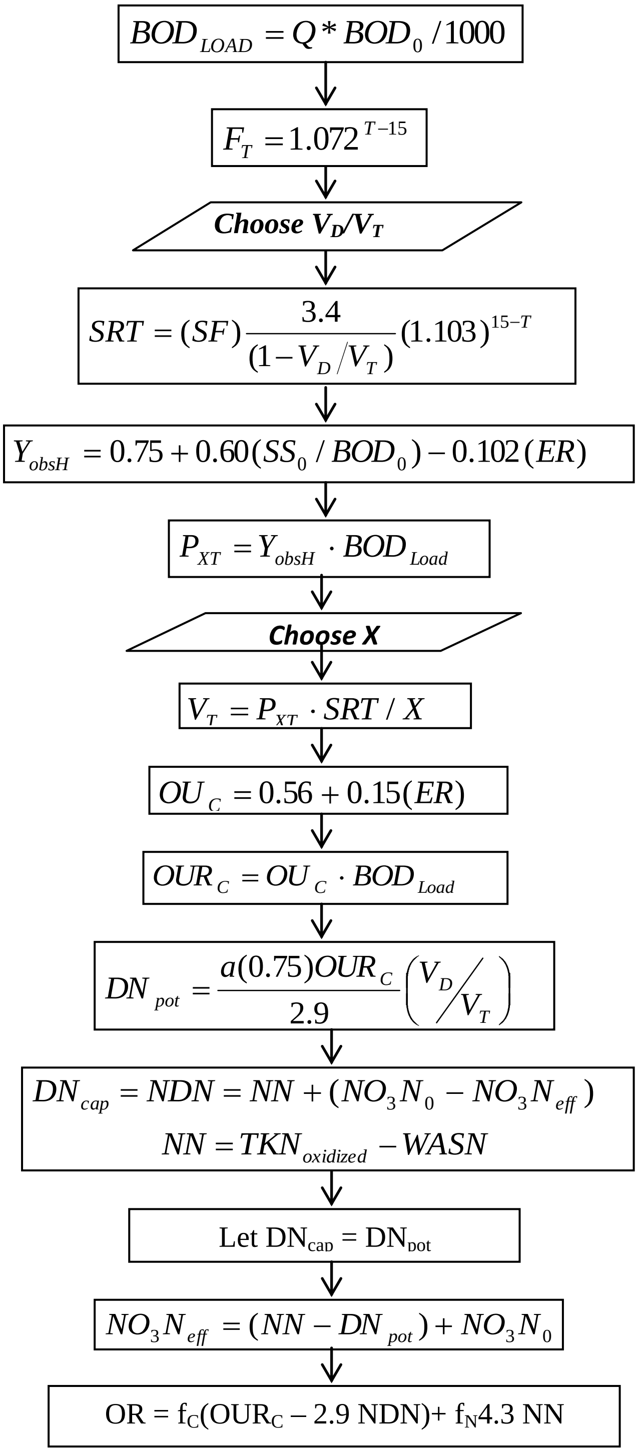


Figure 3.5: Algorithm in ATV- DVWK 131E method

Application on the Plants without nitrification; activated sludge plants without nitrification are dimensioned for sludge ages of four to five days as given in Table 3.11.

Table 3.11: Dimensioning sludge age in days dependent on the treatment target and the temperature as well as the size (ATV-DVWK 131-E, 2000)

	Size of the plant $B_{d,BOD5}$			
Treatment target	Up to 1.200 kg/d		Over 6.000 kg/d	
Dimensioning temperature	10 ^{0}C	12 ^{0}C	10 ^{0}C	12 ^{0}C
Without nitrification	5		4	
With nitrification	10	8,2	8	6,6
With nitrogen removal				
V_D/V_T = 0,2	12,5	10,3	10,0	8,3
0,3	14,3	11,7	11,4	9,4
0,4	16,7	13,7	13,3	11,0
0,5	20,0	16,4	16,0	13,2
Sludge stabilizaiton including N removal	25		Not recommended	

Application on the Plants with nitrification; the aerobic dimensioning sludge age to be maintained for nitrification ($t_{SS,aerob}$);

$$t_{SS,aerob} = SF * 3,4 * 1,103^{(15-T)} \quad \textbf{(3.1)}$$

The value of 3.4 is made up from the reciprocal of the maximum growth rate of the ammonia oxidants at 15^{0}C (2.13 d) and a factor of 1.6. Through the latter it is ensured that, with sufficient oxygen that and no other negative influence factors, enough nitrificants can be developed or held in the activated sludge.

Using the safety factor (SF) the following are taken into account;

- Variations of the maximum growth rate caused by certain substances in the wastewater, short-term temperature variations or/and pH shifts
- The mean effluent concentration of the ammonium
- The effect of variations of the influent nitrogen loads on the variations of the effluent ammonia concentration

Based on all experiences it is recommended, for municipal plants with a dimensioning capacity up to $B_{d,BOD} = 1,200\ kg/d$ (20,000 PT), to reckon with

SF=1.8 due to the more pronounced influent load fluctation and for $B_{d,BOD} > 6{,}000$ kg/d (100,000 PT) with $SF = 1.45$.

Application on the Plants with nitrification and denitrification; for nitrification and denitrification the total dimensioning sludge age($t_{SS,aerob,dim}$) ;

$$t_{SS,aerob,dim} = t_{SS,aerob} * \frac{1}{\left(1-\frac{V_D}{V_T}\right)} \tag{3.2}$$

Determination of the proportion of the reactor volume for denitrification; For designing of nitrogen removal systems, denitrified nitrate is, $S_{NO3,D}$ (mg/l) ;

$$S_{NO3,D} = C_{N,IAT} - S_{orgN,EST} - S_{NH4,EST} - S_{NO3,EST} - X_{orgN,BM} \tag{3.3}$$

The influent nitrate concentration ($S_{NO3,IAT}$) is in general, negligibly small.The concentration of organic nitrogen in the effluent can be set as $S_{orgN,EST}$=2 mg/l. To be on the safe side the ammonium content in the effluent for dimensioning is, as a rule, assumed as $S_{NH4,EST} = 0$.

Nitrogen requirement for biomass, $X_{orgN,BM}$ (mg/l);

$$X_{orgN,BM} = 0.04 \text{ to } 0.05 * C_{BOD,0} \tag{3.4}$$

$S_{NO3,D}/C_{BOD,IAT}$ gives the necessary denitrification capacity;

$$\frac{S_{NO3,D}}{C_{BOD,IAT}} = \frac{0.75 * OU_{C,BOD}}{2.9} * \frac{V_D}{V_{AT}} \tag{3.5}$$

With the relevant BOD_5 of the inflow to the biological reactor one obtains the ratio $S_{NO3,D}/C_{BOD,IAT}$ which gives the necessary denitrification capacity as given Table 3.12.

Table 3.12: Standart values for the dimensioning of denitrification for dry weather at temperatures from 10 ^{0}C to 12 ^{0}C and common conditions (ATV-DVWK 131-E, 2000)

V_D/V_{AT}	$S_{NO3,D}/C_{BOD,IAT}$	
	Pre-anoxic zone denitrification	Simultenous denitrication
0.2	0.11	0.06
0.3	0.13	0.09
0.4	0.14	0.12
0,5	0.15	0.15

Denitrification volumes smaller than $V_D/V_{AT} = 0.2$ and greater than $V_D/V_{AT} = 0.5$ are not recommended. For temperatures above 12 ^{0}C the denitrification capacity can be increased by capacity 1% per 1 ^{0}C.

If the dimensioning or re-calculation takes place on the basis of COD, one can reckon with $S_{NO3,D}/C_{COD,IAT} = 0.5 * (S_{NO3,D}/C_{BOD,IAT})$.

If the required denitrification capacity is larger than $S_{NO3,D}/ C_{BOD,IAT} = 0.15$, then a further increase of V_D/V_{AT} is not recommended. It is to be invastigated whether a volume reduction or partial by-passing or primary settling tank and/or, if applicable, a seperate sludge treatment are conducive to meeting the target. An alternative is to carry out the planning for the addition of external carbon.

Application for Phosphorus Removal; phosphorus removal can take place alone through simultenous precipitation, throgh excess biological phosphorus removal, as a rule combined with simultenous precipitation and through pre- or post precipitation.

Anaerobic mixing tanks for biological phosphorus removal are to be dimensioned for a minimum contact time of 0.5 to 0.75 hours, referred to the maximum dry weather inflow and the return sludge flow $(Q_{DW,h} + Q_{RS})$. The degree of the biological phosphorus removal depends, other than on the contact time, to a large extent on the ratio of the concentration of the readily biodegredable organic matter to the concentration of phosphorus. If, in winter, the anaerobic volume is used for denitrification, then during this period a lower biological excess phosphorus removal will establish.

For the determination of the phosphate to be precipitated a phosphorus balance, if necessary for different types of load, is to be drawn up:

$$X_{P,Prec} = C_{P,IAT} - C_{P,EST} - X_{P,BM} - X_{P,BioP} \quad \textbf{(3.6)}$$

- $X_{P,BioP} = 0.01 \text{ to } 0.015 * C_{BOD,IAT}$ or 0.005 to 0.007 $C_{COD,IAT}$ respectively with upstream anaerobic tanks
- if, with lower temperatures, $S_{NO3,EST}$ increases to ≥15 mg/l, it can be assumed: $X_{P,BioP} = 0.005 \text{ to } 0.01\ C_{BOD,IAT}$ or $0.0025 \text{ to } 0.005 * C_{COD,IAT}$ respectively with upstream anaerobic tanks

- in plants with pre-anoxic zone denitrification or step-feed denitrification,but without anaerobic tanks,an excess biological phosphorus removal of $X_{P,BioP} \leq 0.005 * C_{BOD,IAT}$ or $0.002 * C_{COD,IAT}$ respectively can be assumed.
- if, at low temperatures, the internal recirculations of pre-anoxic zone denitrification is discharged into the anaerobic tank, one can reckon with $X_{P,BioP} \leq 0.005 * C_{BOD,IAT}$ or $0.002 * C_{COD,IAT}$ respectively.

Determination of the sludge production; the sludge produced in an activated sludge plant is made up of organic matter resulting from degradation and stored solid matter as well as sludge resulting from phosphorus removal:

Sludge production for biological phosphorus removal is ($SP_{d,P}$);

$$X_{P,Bio,P} = 0.01 * C_{BOD,0} \tag{3.7}$$

$$SP_{d,P} = Q * 3 * X_{P,Bio,P} \tag{3.8}$$

Sludge production for carbon removal is, ($SP_{d,C}$);

$$SP_{d,C}(\text{COD based}) = Q * \left(\frac{X_{WAS,COD}}{0.8} * 1.45\right) + X_f)/1{,}000 \tag{3.9}$$

$$SP_{d,C}(\text{BOD based}) = B_{d,BOD} * (0.75 + 0.6 * \frac{X_{SS,IAT}}{C_{BOD,IAT}} - \frac{(1-0.2)*0.2*0.17*0.75t_{SS}*F_T}{1+0.17*t_{SS}*F_T} \tag{3.10}$$

Total sludge production is, (SP_d);

$$SP_d = SP_{d,C} + SP_{d,P} \tag{3.11}$$

Assumption of the sludge volume index and the mixed liquir suspended solids; The sludge volume index depends on the composition of the wastewater and the mixing characteristics of the aeration tank. A high fraction of readily biodegredable organic matter, as are contained in some commercial and industrial wastewater, can lead to higher sludge volume indices.

Table 3.13: Standart values for the sludge volume index (ATV-DVWK 131-E, 2000)

Treatment target	SVI (l/kg)	
	Favourable	Unfavourable
Without nitrification	100-150	120-180
Nitrification and denitrification	100-150	120-180
Sludge stabilization	75-120	100-150

If no usable data are available, the values listed in table are recommended for dimensioning taking into account critical operating conditions. The respectively lower values for the SVI can be applied, if

- primary settling is dispended with
- a selector or/an anaerobic mixing tank is placed upstream
- the biologiacal reactor is designed as a cascade (plug flow)

The concentration of mixed liquir suspended solids (SS_{AT}) is determined in the process of dimensioning the secondary settling tank.

Volume of the biological reactor; the required mass of suspended solids in the biological reactor is, ($M_{SS,AT}$);

$$M_{SS,AT} = t_{SS,Dim} * S_{P,d} \quad \textbf{(3.12)}$$

The volume of the biological reactor is, (V_{AT});

$$V_{AT} = \frac{M_{SS,AT}}{SS_{AT}} \quad \textbf{(3.13)}$$

Required recirculation and cycle time; the necessary total recirculation flow ratio (RC) for pre-anoxic zone denitrification results using $S_{NH4,N}$, the ammonium nitrogen concentration to be nitrified, as follows

The necessary total recirculation flow ratio is, (RC);

$$RC = \frac{S_{NH_4,N}}{S_{NO_3,EST}} - 1 \quad \textbf{(3.14)}$$

Oxygen transfer; the oxygen uptake is made up of the consumption for carbon removal (including the endogenous respiration)and, if necessary, the requirement for nitrification as well as the saving of oxygen from denitrification.

For carbon removal the following approach, using the Hartwig coefficients:

Oxygen transfer for carbon removal is, ($OU_{d,C}$)

$$OU_{d,C}(\text{BOD based}) = B_{d,BOD} * \left(0.56 + \frac{0.15 * t_{SS} * F_T}{1 + 0.17 * t_{SS} * F_T}\right) \quad \textbf{(3.15)}$$

$$OU_{d,C}(\text{COD based}) = Q * (C_{COD} - (S_I - X_{WAS,COD}))/1{,}000 \quad \textbf{(3.16)}$$

For nitrification the oxygen consumption is assumed to be 4.3 kg O_2 per kg oxidized nitrogen taking into account the metabolism of the nitrificants. Oxygen transfer for nitrification is, $(OU_{d,N})$;

$$OU_{d,N} = Q_d * 4.3 * (S_{NO3,D} - S_{NO3,IAT} + S_{NO3,EST})/1{,}000 \qquad \textbf{(3.17)}$$

For denitrification one reckons for carbon removal with 2.9 kg O_2 per kg denitrified nitrate nitrogen. For denitrification oxygen credit is, $(OU_{d,D})$;

$$OU_{d,D} = Q_d * 2.9 * S_{NO3,D}/1{,}000 \qquad \textbf{(3.18)}$$

The oxygen uptake rate for the daily peak is, (OU_h);

$$OU_h = \frac{f_C*(OU_{d,C}-OU_{d,N})+f_N*OU_{d,N}}{24} \qquad \textbf{(3.19)}$$

The peak factor f_C and f_N represents the ratio of the oxygen uptake rate for carbon and nitrogen removal in the peak hour to the avarage daily oxygen uptake rate.

Table 3.14: Peak factors of the oxygen uptake rate (ATV-DVWK 131-E, 2000)

	Sludge age (d)					
	4	6	8	10	15	25
f_C	1.3	1.25	1.2	1.2	1.15	1.1
f_N for $B_{d,BOD} \leq 1{,}200$ kg/d				2,5	2	1.5
f_N for $B_{d,BOD} \leq 6{,}000$ kg/d			2	1.8	1.5	

Dimensioning of the secondary settling tank; bases of the dimensioning are the maximum inflow rate with stormwater (Peak Wet Weather Flow Rate) Q_{ww},h, the sludge volume index SVI (l/kg) and the suspended solids concentration in the inluent to the secondary settling tanks SS_{EAT}.

For the design of secondary settling tanks the following are to be determined:

- Shape and dimensions of the secondary settling tank
- Permitted sludge storage and thickening time
- Return sludge flow rate as well as its control
- Type and method of operation of the sludge removal system
- Arrangement and design of the inlet and outlet

The following dimensioning rules apply for:

- Secondary settling tanks with lengths or diameters up to approximately 60 m
- Sludge volume index 50l/kg≤SVI≤200l/kg
- Diluted sludge volume DSV≤600l/m^3
- Return sludge flowrates

  Q_{RS}=0.75*$Q_{WW,h}$

  Q_{RS}=1.0* $Q_{WW,h}$
- Suspended solids concentration in the influent to the secondary settling tank SS_{EAT}≥1.0 kg/m^3

Permitted thickening time; the sludge volume index (table 2.1.5), together with the thickening time (t_{Th}) in the secondary settling tank determines the suspended solids concentration in the bottom sludge(SS_{BS}).

Table 3.15: Recommended t_{Th} in dependence on the degree of wastewater treatment (ATV-DVWK 131-E, 2000)

Type of wastewater treatment	**t_{Th} (h)**
Activated sludge plants without nitrification	1.5-2.0
Activated sludge plants with nitrification	1.0-1.5
Activated sludge plants with denitrification	2.0-(2.5)

An exceeding of the thickening time of t_{Th} = 2.0 h requires a very advanced denitrification in the biological reactor.

Suspended solids concentration in the return sludge; the achievable suspended solids concentration in the bottom sludge SS_{BS} can be estimated emprically in dependence on the sludge volume index SVI and thickening time. The suspended solids concentration in the bottom sludge, SS_{BS};

$$SS_{BS} = \frac{1{,}000}{SVI} * \sqrt[3]{t_{Th}} \tag{3.20}$$

The suspended solids concentration of the return sludge (SS_{RS}), as a result of the dilution with the short-circuit sludge flow, can be assumed in simplified form to be:

with scraper facilitities - $SS_{RS} \simeq 0.7 * SS_{BS}$

with scraper facilitities - $SS_{RS} \simeq 0.5 \text{ to } 0.8 * SS_{BS}$

The suspended solids concentration of the return sludge, (SS_{RS});

$$SS_{RS} = 0.7 * SS_{BS} \tag{3.21}$$

f_{obs} (dilution factor of return sludge) $= 0.7$

Return sludge ratio and suspended solids concentration in the influent to the secondary settling tank; the operating conditions in the aeration tank and in the secondary settling tank are influenced mutually through the dependence between the mixed-liquir suspended solids concentration of the return sludge SS_{RS} as well as the return sludge ratio RS. For the equilibrium state the following results from the suspended solids mass balance, neglecting $X_{SS,EST}$. Return sludge ratio, (RS);

$$RS = \frac{Q_{RS}}{Q} \tag{3.22}$$

The dimensioning of secondary settling tanks shall be based on a maximum return sludge flow rate $Q_{RS} = 0.75 * Q_{WW,h}$. The overall capacity of the return sludge pumps, including reserve, for operational reasons must be designed that the return sludge flow rate of $Q_{RS} = 1.0 * Q_{WW,h}$ can be achieved. The suspended solids concentration in the aeration tank, SS_{AT};

$$SS_{AT} = \frac{RS * SS_{RS}}{1 + RS} \tag{3.23}$$

Surface overflow rate and sludge volume surface loading rate; The surface overflow rate q_A is calculated from the permitted sludge volume loading rate q_{SV} and the diluted sludge volume DSV. The surface overflow rate, q_A;

$$q_A = \frac{q_{SV}}{DSV} = \frac{q_{SV}}{SS_{EAT} * SVI} \tag{3.24}$$

In order to keep the concentration of suspended solids $X_{SS,EAT}$ and the resulting COD and phosphorus concentration in the effluent of horizontal flow secondary settling tanks low, the following sludge volume loading rate q_{SV} shall not be exceeded:

$q_{SV} \leq 500\ l/(m^2/h)$ for $X_{SS,EST} \leq 20\ mg/l$

For mainly vertical flow secondary settling tanks, the following applies with the formation of a close sludge blanket or with an easily flocculating activated sludge:

$q_{SV} \leq 650l/(m^2/h)$ for $X_{SS,EST} \leq 20\ mg/l$

The surface overflow rate q_A shall not exceed 1.6 m/h with predominantly horizontal flow secondary settling tanks, and with predominantly vertical flow secondary settling tanks it shall not exceed 2.0 m/h.

Settling tank surface area; the required surface area of the secondary settling tank results as follows:

$$A_{ST} = \frac{Q_{WW.h}}{q_A} \tag{3.25}$$

Generally only with horizontal flow secondary settling tanks additional area for the inlet disturbance zone is required. The length of this disturbance zone is, an approximation, set equal to the side wall depth of the tank.

For vertical flow secondary settling tanks the affective surface area at the mid-point between inlet aperture and water level is to be set.

Settling tank depth; The various processes in secondary settling for the explained with the aid of functionally conditioned effective volumes.

The necessary depth of the secondary settling tank is made up from individual partial depths for the functional zones:

h_1: Clean water zone

h_2: Seperation zone

h_3: Density flow and storage zone

h_4: Thickening and sludge removal zone

The necessary depth of the secondary settling tank, h_T;

$$h_T = h_1 + h_2 + h_3 + h_4 \tag{3.26}$$

Clean water zone, h_1;

$$h_1 = \text{assume } 0.5 \text{ m}$$

The seperation/return flow zone must be dimensioned that the inflow, including the return sludge flow, referred to the free water volume, has a calculated detention time of 0.5 h. From this results; seperation zone, h_2;

$$h_2 = \frac{0.5 * q_A * (1+R)}{1 - DSV/1{,}000} \tag{3.27}$$

The density flow and storage zone must be dimensioned that under wet weather conditions $Q_{WW,h}$ the additional volume of sludge ($0.3 * SS_{EAT} * SVI$) with a concentration value of 500 l/m^3 expelled in 1.5 hours from the aeration tankcan be stored. In this period the activated sludge settles in the thickening zone and is assumed to be evenly distributed over the surface area A_{ST} of the secondary settling tank. The depth of the density flow and storage zone thus results as:

$$h_3 = \frac{1.5 * 0.3 * q_{SV} * (1 + RS)}{500} \qquad \textbf{(3.28)}$$

The thickening of the settled sludge takes place at the bottom of the tank in the thickening and sludge removal zone. There a sludge layer exists in which low rates to the sludge hopper occur. The thickening and sludge removal zone must be large enough so that the influent sludge load, having a suspended solids concentration of SS_{EAT}, within the thickening time t_{Th} can be thickened to the bottom sludge concentration SS_{BS}. With the assumption of an even distribution of the sludge mass over the surface of the secondary settling tank the height of the thickening and sludge removal zone results. The depth of thickening and sludge removal zone, h_4;

$$h_4 = \frac{SS_{EAT} * q_A * (1 + RS) * t_{Th}}{SS_{BS}} \qquad \textbf{(3.29)}$$

3.3.2 Design in Water Environment Research Foundation (WERF) method

The algorithm used for the design in WERF method is sumarized in Figure 3.6. Nitrification; perform the nitrification design following the steps below as for BOD removal except the design SRT must be determined. Determine the spesific growth rate **(μ_n)** for the nitrifying organisms. The nitrification rate will control the design because the nitrifying organisms grow more slowly than the heterotrophic organisms that remove organic carbon. The spesific growth rate (μ_n); is determined for the nitrifying organisms, (μ_n);

$$\mu_n = \left(\frac{\mu_{n,m} * N}{K_n + N}\right) * \left(\frac{DO}{K_o + DO}\right) - k_{dn} \qquad \textbf{(3.30)}$$

$$\mu_{n,m} = 0.75 * 1.072^{(T-20)} \qquad \textbf{(3.31)}$$

$$K_n = 0.74 * 1.053^{(T-20)} \qquad \textbf{(3.32)}$$

$$k_{dn} = 0.08 * 1.04^{(T-20)} \qquad \textbf{(3.33)}$$

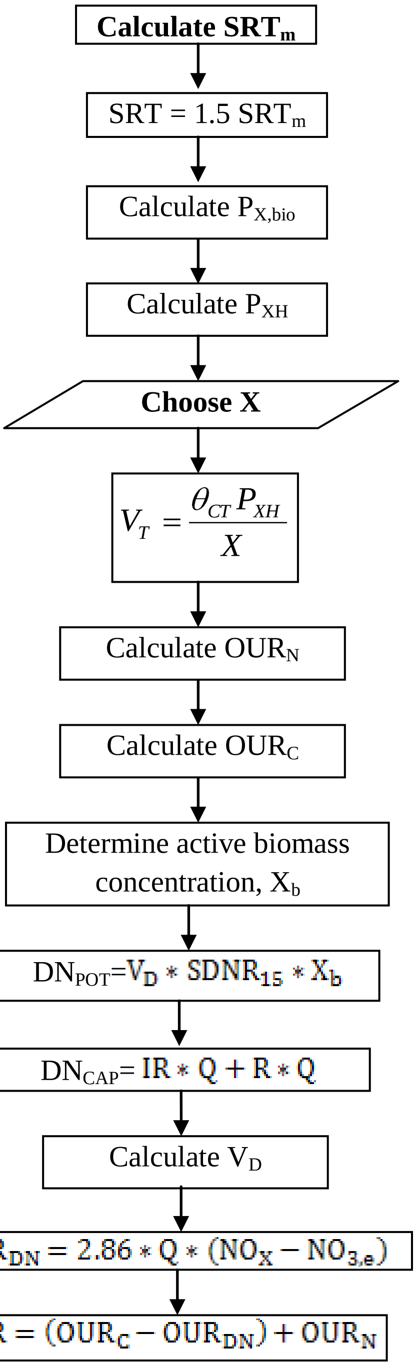


Figure 3.6: Algorithm in WERF method

Minimum sludge age for nitrification, $(\mathbf{SRT_m})$;is determined to achieve nitrification:

$$\mathrm{SRT_m} = \frac{1}{\mu_n} \qquad \textbf{(3.34)}$$

Total sludge age, (SRT); represents the avarage period time during which the sludge has remained in the system. The sludge age for the system design is selected with caution as variable nitrification growth rates have been observed at different sites, presumably due to the presence of inhibitory substances (Barker and Dold, 1997;

Fillos et al., 2000). For design sludge age, a factor of safety is used to increase the SRT above that calculated based on nitrification kinetics and the required effluent NH_4-N concentration. A factor of safety is used for two reasons: (1) to allow flexibility for operational variations in controlling the SRT, and (2) to provide for additional nitrifying bacteria to handle peak TKN loadings.

$$\mathrm{SF} = \mathrm{TKN\ peak}/\mathrm{TKN\ average}$$

$$\mathrm{SF} = \mathrm{assume}\ 1.5$$

$$\mathrm{SRT} = \mathrm{SRT}_m * \mathrm{SF} \tag{3.35}$$

For biomass production, $\mathbf{P_{X,bio}}$; first effluent BOD concentration is calculated. S effluent concentration, (S_{eff});

$$S_{eff} = K_s * \frac{(1+k_d*\mathrm{SRT})}{\mathrm{SRT}*(\mu_m-k_d)-1} \tag{3.36}$$

Biomass production, ($P_{X,bio}$);

- Heterotrophic biomass : $\frac{Q*Y*(S_0-S)}{(1+k_d*\mathrm{SRT})}$
- Cell debris : $\frac{f_d*k_d*Q*Y*(S_0-S)*\mathrm{SRT}}{(1+k_d*\mathrm{SRT})}$
- Nitrifying bacteria biomass : $\frac{Q*Y_n*NO_X}{(1+k_{dn}*\mathrm{SRT})}$

$$P_{X,bio} = \frac{Q*Y*(S_0-S)}{(1+k_d*\mathrm{SRT})} + \frac{f_d*k_d*Q*Y*(S_0-S)*\mathrm{SRT}}{(1+k_d*\mathrm{SRT})} + \frac{Q*Y_n*NO_X}{(1+k_{dn}*\mathrm{SRT})} \tag{3.37}$$

N oxidized to NO_3, (NO_X);

$$NO_X = \mathrm{TKN} - N_e - 0.12 * P_{X,bio}/Q \tag{3.38}$$

The total volatile mass of dry solids, (VSS); accounts for the heterotrophic biomass growth, cell debris from endogenous decay, nitrifying bacteria biomass, and nonbiodegradable volatile suspended solids and can be used to estimate sludge production. Concentration of VSS in the aeration basin, ($P_{X,VSS}$);

$$P_{X,VSS} = P_{X,bio} + Q * \mathrm{nbVSS} \tag{3.39}$$

Mass of VSS in the aeration basin, ($P_{X,VSS}$);

$$MX_{VSS} = P_{X,VSS} * \mathrm{SRT} \tag{3.40}$$

Concentration of TSS in the aeration basin, ($P_{X,TSS}$); the total mass of dry solids wasted/day includes TSS and not just VSS. The TSS includes the VSS plus inorganic solids. Inorganic solids in the influent wastewater $(\mathrm{TSS}_0 - \mathrm{VSS}_0)$ contribute to inorganic solids and are an additional solids production term. VSS fraction of the total biomass is about 0.85.

$$P_{X,TSS} = (P_{X,bio}/0.85) + Q * nbVSS + Q * (TSS_0 - VSS_0) \quad \textbf{(3.41)}$$

Mass of TSS in the aeration basin, ($P_{X,VSS}$);

$$MX_{TSS} = P_{X,TSS} * SRT \quad \textbf{(3.42)}$$

Volume of the aeration basin, V_T;

$$V_T = \frac{MX_{TSS}}{MLSS} \quad \textbf{(3.43)}$$

Hydraulic retention time, (r);

$$r = \frac{V}{Q+Q*R} \quad \textbf{(3.44)}$$

F/M ratio; a process parameter commonly used to characterize process designs and operating conditions is the food to microorganism ratio. Typical values for the BOD F/M ratio reported in the literature vary from 0.04 to 0.1.

$$F/M = \frac{Q*BOD_0}{V*MLSS} \quad \textbf{(3.45)}$$

Yield coefficient, (Y_{obs});

$$Y_{obs,TSS} = \frac{P_{X,TSS}}{bCOD\ removed} \quad \textbf{(3.46)}$$

When nitrification is included in the process, the total oxygen requirement will include the oxygen required (OR); for removal of carbonaceous material plus the oxygen required for ammonia and nitrite oxidation to nitrate. Oxygen uptake rate for carbon removal, (OUR_C);

$$OUR_C = Q * (S_0 - S) - 1.42 * P_{X,bio} \quad \textbf{(3.47)}$$

Oxygen uptake rate for nitrogen removal, (OUR_N);

$$OUR_N = Q * NO_X * 4.33 \quad \textbf{(3.48)}$$

Active biomass concentration, (X_b) in denitrification

$$X_b = \frac{Q*SRT}{V} * \frac{Y*(S_0-S)}{(1+k_d*SRT)} \quad \textbf{(3.49)}$$

IR ratio;

$$IR = \frac{NO_X}{NO_{3,e}} - 1 - R \quad \textbf{(3.50)}$$

$\mathbf{NO_3}$ fed to the anoxic tank, NO_X feed;

$$\text{Flowrate to the anoxic tank} = IR * Q + R * Q \quad \textbf{(3.51)}$$

$$NO_X\ \text{feed} = NO_{3,e} * \text{Flowrate to the anoxic tank} \quad \textbf{(3.52)}$$

Anoxic volume, (V_D);

$$V_D = r' * Q \tag{3.53}$$

$r' = \text{Hydraulic retention time}$

F/M_b ratio;

$$F/M_b = \frac{Q * S_0}{V_D * X_b} \tag{3.54}$$

The desktop design approach is based on using a spesific denitrification rate (SDNR), which is the nitrate reduction rate in the anoxic tank normalized to the MLSS concentration.

$$\text{Frac. of RBCOD} = \frac{S_{S,i}}{\text{RBCOD}} \tag{3.55}$$

$\text{SDNR}_b = 0.27\ d^{-1} \text{ at } 20\ C$

$$\text{SDNR}_{15} = \text{SDNR}_b * 1.026^{(T-20)} \tag{3.56}$$

NO_3 reduced, (**NO_r**);

$$NO_r = V_D * \text{SDNR}_{15} * X_b \tag{3.57}$$

Oxygen credit with denitrification, (OUR_{DN});

$$OUR_{DN} = 2.86 * Q * (NO_X - NO_{3,e}) \tag{3.58}$$

Actual Oxygen Requirement, (AOR);

$$OR = (OUR_C - OUR_{DN}) + OUR_N \tag{3.59}$$

P removed by BPR mechanism;

$$\text{Bio} - \text{P removal} = \frac{S_{VFA,i}}{\text{RBCOD/P}} \tag{3.60}$$

$\text{RBCOD/P} = \text{assume } 10$

P used for hetetrophic biomass, $\mathbf{P_{X,bio}}$;

$$P_{X,bio} = \frac{Q * Y * (S_0 - S)}{(1 + k_d * SRT)} + \frac{Q * Y_n * NO_X}{(1 + k_{dn} * SRT)} \tag{3.61}$$

P used for biomass growth;

$$\text{P used} = 0.015 \frac{gP}{\text{gbiomass}} * P_{X,bio} \tag{3.62}$$

Effluent soluble P;

$$\text{P removed} = \text{Bio} - \text{P removal} + \text{P used} \tag{3.63}$$

$$\text{Effluent soluble P} = TP_0 - \text{P removed} \tag{3.64}$$

P content of waste sludge;

$$T - \text{P in sludge} = \text{P removed} * Q/1{,}000 \tag{3.65}$$

Anaerobic volume, (V_{AN});

$$V_{AN} = \frac{Q * r}{24} \tag{3.66}$$

3.3.3 Design in South African method

The Algorithm used for the design in South African method is summarized in Figure 3.7.

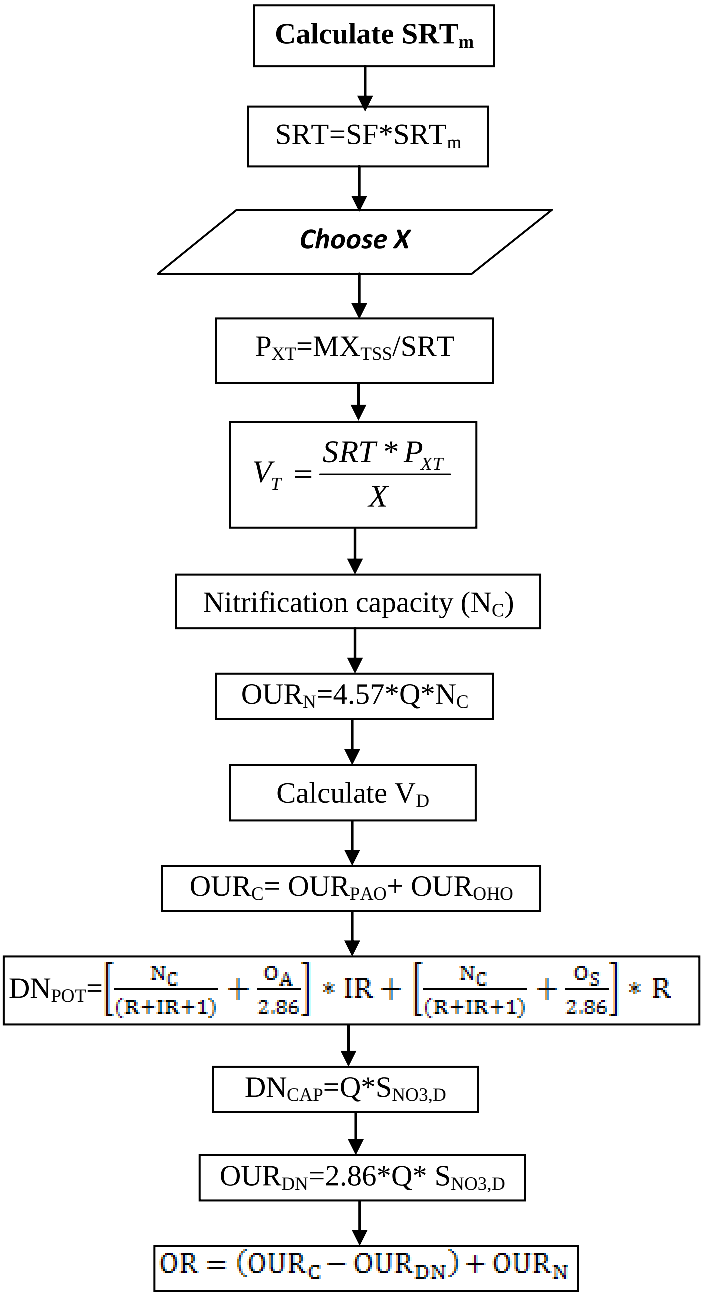


Figure 3.7: Algorithm in South African method

Minimum sludge age for nitrification, SRT_m; for the purposes of demonstrating nitrification under purely aerobic conditions, it will be accepted that the influent alkalinity is sufficiently high to maintain an effluent alkalinity above 50 mg/l as $CaCO_3$. Maximum spesific growth rate at 15 ^{0}C, $\mu_{A,m15;}$

$$\mu_{A,m15} = \mu_{A,m20} * \theta n^{(T-20)} \tag{3.67}$$

Endegenous respiration rate at 15 ^{0}C, b_{A15};

$$b_{A15} = b_{A20} * \theta_n{}^{(T-20)} \tag{3.68}$$

To calculate SRT_m, S_f is selected as 1.3. For the lowest design (15 ^{0}C), temperature, S_F is selected between 1.25-1.35.

$$SRT_m = S_f/(\mu_{Am} - b_A) \tag{3.69}$$

Design total sludge age, SRT;

$$SRT = 1.3 * SRT_m \tag{3.70}$$

Influent unbiodegredable soluble organic nitrogen concentration, (N_{ousi});

$$N_{ousi} = f_{N,ous} * C_{TKN,0} \tag{3.71}$$

Influent unbiodegredable particulate organics, (N_{oupi});

$$N_{oupi} = f_n * f_{s,up} * S_{ti}/f_{cv} \tag{3.72}$$

Influent biodegredable organic N concentration both soluble and particulate which is convertes to ammonia, (N_{obi});

$$N_{obi} = C_{TKN,0} * (1 - f_{N,a} - f_{N,ous}) - N_{oupi} \tag{3.73}$$

The unbiodegredable soluble organic nitrogen concentration in the effluent, (N_{ouse});

$$N_{ousi} = N_{ouse} \tag{3.74}$$

N concentration of the sludge, (N_s);

$$N_S = C_{COD} * f_n * (1 - f_{s,up} - f_{s,us}) * \frac{Y_H}{1+b_H * SRT} * (1 + f_h * b_H * SRT) + \frac{f_{s,up}}{f_{cv}} \tag{3.75}$$

From a TKN mass balance over the activated sludge system and SRT > SRT_m, the concentration of nitrate generated in the system N_C with respect to the influent flow is given by the influent TKN (N_{ti}) minus the soluble effluent TKN (N_{te}) and the concentration of influent TKN incorporated in the sludge wasted daily from activated sludge system (N_S). Nitirification capacity, (N_c);

$$N_c = N_{ti} - N_s - N_{ouse} \quad \textbf{(3.76)}$$

The effluent ammonia concentration, (N_{ae});

$$N_{ae} = \frac{K_n * (b_A + 1/SRT)}{\mu_{Am} * (1 - f_{xt}) - (b_A + 1/SRT)} \quad \textbf{(3.77)}$$

The effluent TKN concentration, (N_{te});

$$N_{te} = N_{ae} + N_{ouse} \quad \textbf{(3.78)}$$

The oxygen demand for nitrification, (OUR_N);

$$OUR_N = 4.57 * Q * N_c \quad \textbf{(3.79)}$$

The concentration of nitrate that an anoxic reactor can denitrify biologically is called that reactor’s denitrification potential. It is called a potential because whether or not it is achieved depends on the nitrate load on the anoxic reactor(s). Denitrification potential, (DN_{POT});

$$DN_{POT} = \left[\frac{N_C}{(R+IR+1)} + \frac{O_A}{2.86}\right] * IR + \left[\frac{N_C}{(R+IR+1)} + \frac{O_S}{2.86}\right] * R \quad \textbf{(3.80)}$$

IR ratio;

$$IR_{opt} = \left[-B + \sqrt{B^2 + 4 * A * C}\right]/2 * A \quad \textbf{(3.81)}$$

where;

$$A = O_A/2.86 \quad \textbf{(3.82)}$$

$$B = N_C - DN_{POT} + [(R+1) * O_A + R * O_S] \quad \textbf{(3.83)}$$

$$C = (R+1) * \left(DN_{POT} - R * \frac{O_S}{2.86}\right) - R * N_C \quad \textbf{(3.84)}$$

Denitrified nitrate, $S_{NO3,D}$;

$$S_{NO3,D} = (TKN_0 - TN_{eff} - \text{N concentration of the sludge}) \quad \textbf{(3.85)}$$

Oxygen credit with denitrification, OUR_{DN};

$$OUR_{DN} = Q * 2.86 * S_{NO3,D}/1{,}000 \quad \textbf{(3.86)}$$

To select the V_D/V_T ratio for anoxic volume the equation below is used;

$$\frac{V_D}{V_T} = 1 - S_f * (b_{A15} + \frac{1}{SRT})/\mu_{A,m15} \tag{3.87}$$

Anoxic volume, V_{AN};

$$V_{AN} = V_T * \frac{V_D}{V_T} \tag{3.88}$$

Fermentable COD available for conversion into VFAs after denitrification reactor in AN reactor, $\mathbf{S_{F,i,conv}}$; for Enhanced Biological Phosphorus Removal (EBPR),

$$S_{F,i,conv} = S_{F,i} - \left(8.6 * R * S_{NO3,s} + S_{NO3,i}\right) - 3 * \left(R * S_{O2,s} + S_{O2,i}\right) \tag{3.89}$$

Fermentable COD lost in the effluent of the last anaerobic reactor, $\mathbf{S_{F,Ann}}$; for EBPR,

$$S_{F,Ann} = \frac{S_{F,i,conv}}{(1+R)} / \left(1 + (k_{F,T} * (f_{AN} * MX_{OHO} / (N * Q * (1 + R))))\right)\text{^}n \tag{3.90}$$

VFAs stored by PAOs, $\mathbf{FS_{S,PAO}}$; for EBPR,

$$FS_{S,PAO} = Q * \left(S_{F,i,conv} - (1 + R) * S_{F,Ann}\right) + Q * S_{VFA,i} \tag{3.91}$$

Remaining biodegredable COD available to OHOs, $\mathbf{FCOD_{b,OHO}}$; for EBPR,

$$FCOD_{b,OHO} = FCOD_{b,i} - FS_{S,PAO} \tag{3.92}$$

Biomass equations corresponds to the biological mass present in the system as synthesized from the influent COD taking into account the cumulative effect of SRT.

Observed net yield coefficient for PAOs, $Y_{PAO,obs}$;

$$Y_{PAO,obs} = \frac{Y_{PAO}}{1+b_{PAO} * SRT} \tag{3.93}$$

Biological active mass for PAOs, MX_{PAO};

$$MX_{PAO} = Y_{PAO,obs} * FS_{S,PAO} * SRT \tag{3.94}$$

Endogenous mass for PAOs, $MX_{E,PAO}$;

$$MX_{E,PAO} = f_{XE,PAO} * b_{PAO} * MX_{PAO} * SRT \tag{3.95}$$

Observed net yield coefficient for Active mass (OHOs), $Y_{OHO,obs}$;

$$Y_{OHO,obs} = \frac{Y_{OHO}}{1+b_{OHO} * SRT} \tag{3.96}$$

Biological active mass for heterotrophs, MX_{OHO};

$$MX_{OHO} = Y_{OHO,obs} * F_{CODb,OHO} * SRT \tag{3.97}$$

Endogenous mass for OHOs, $MX_{E,OHO}$;

$$MX_{E,OHO} = fX_{E,OHO} * b_{OHO} * MX_{OHO} * SRT \tag{3.98}$$

Inert mass;

$$MX_{I} = fX_{I,COD,i} * F_{COD,i} * SRT / f_{CV} \tag{3.99}$$

P release, $\mathbf{S_{PO4,rel}}$;

$$S_{PO4,rel} = f_{PO4,rel} * FS_{S,PAO} / Q \tag{3.100}$$

ΔP by PAOs, $\mathbf{\Delta P_{PAO}}$;

$$\Delta P_{PAO} = f_{P,PAO} * MX_{PAO} / (SRT * Q) \tag{3.101}$$

ΔP by OHOs, $\mathbf{\Delta P_{OHO}}$;

$$\Delta P_{OHO} = f_{P,OHO} * MX_{OHO} / (SRT * Q) \tag{3.102}$$

ΔP by endegoneus mass, $\mathbf{\Delta P_{XE}}$;

$$\Delta P_{XE} = \Delta P_{XE,PAO} + \Delta P_{XE,OHO} \tag{3.103}$$

$$\Delta P_{XE,PAO} = f_{P,XE} * MX_{E,PAO} / (SRT * Q) \tag{3.104}$$

$$\Delta P_{XE,OHO} = f_{P,XE} * MX_{E,OHO} / (SRT * Q) \tag{3.105}$$

ΔP by influent inert mass, $\mathbf{\Delta P_{XI}}$;

$$\Delta P_{XI} = f_{P,XI} * MX_{I} / (SRT * Q) \tag{3.106}$$

Potential total P removal, $\mathbf{\Delta P_{SYS,POT}}$;

$$\Delta P_{SYS,POT} = \Delta P_{PAO} + \Delta P_{OHO} + \Delta P_{XE} + \Delta P_{XI} \tag{3.107}$$

Particulate P in the effluent, $\mathbf{X_{P,e}}$;

$$X_{P,e} = f_{P,TSS} * TSS_{e} \tag{3.108}$$

Effluent total P, $\mathbf{T_{P,e}}$;

$$T_{P,e} = T_{P,i} - \Delta P_{SYS,POT} + X_{P,e} \qquad \textbf{(3.109)}$$

Active fraction mass

$$MX_{bio} = MX_{PAO} + MX_{OHO} \qquad \textbf{(3.110)}$$

VSS sludge mass, $\mathbf{MX_{VSS}}$;

$$MX_{VSS} = MX_{PAO} + MX_{OHO} + MX_{E,PAO} + MX_{E,OHO} + MX_I \qquad \textbf{(3.111)}$$

FSS sludge mass, $\mathbf{MX_{FSS}}$;

$$MX_{FSS} = f_{FSS,OHO} * MX_{OHO} + f_{FSS,PAO} * MX_{PAO} + f_{X,FSS,i} * SRT \qquad \textbf{(3.112)}$$

Total mass, $\mathbf{MX_{TSS}}$;

$$MX_{TSS} = MX_{VSS} + MX_{FSS} \qquad \textbf{(3.113)}$$

VSS/TSS ratio for the sludge, $\mathbf{f_{VT}}$;

$$f_{VT} = MX_{VSS} / MX_{TSS} \qquad \textbf{(3.114)}$$

P content of TSS, $\mathbf{f_{P,TSS}}$;

$$f_{P,TSS} = ((f_{P,OHO} * MX_{OHO} + f_{P,XE} * \left(MX_{E,PAO} + MX_{E,OHO}\right) + f_{P,XI} * MX_I) / f_{VT} + (f_{P,PAO} * MX_{PAO}) / f_{VT,PAO} * (f_{P,FSS,i} * MX_{FSS}) / MX_{TSS} \qquad \textbf{(3.115)}$$

Process volume, $\mathbf{V_P}$;

$$V_P = MX_{TSS} / X_{TSS} \qquad \textbf{(3.116)}$$

The volume of anaerobic zone, $\mathbf{V_{P,AN}}$;

$$V_{P,AN} = f_{AN} * V_P \qquad \textbf{(3.117)}$$

Nitrogen requirement, $\mathbf{FN_{synth}}$;

$$FN_{synth} = f_{N,VSS} * MX_{VSS} / SRT \qquad \textbf{(3.118)}$$

$$TKN_{i,synth} = FN_{synth} / Q \qquad \textbf{(3.119)}$$

Oxygen demand by PAOs for synthesis and endogenous respiration;

$$FO_{2,PAO} = FO_{2,PAO,synth} + FO_{2,PAO,endo} \qquad \textbf{(3.120)}$$

$$FO_{2,PAO,synth} = FS_{S,PAO} * (1 - f_{CV} * Y_{PAO}) \quad \textbf{(3.121)}$$

$$FO_{2,PAO,endo} = FS_{S,PAO} * f_{CV} * \left(1 - f_{XE,PAO}\right) * b_{PAO} * Y_{PAO,obs} * SRT \quad \textbf{(3.122)}$$

Oxygen Demand by OHOs for synthesis and endogenous respiration;

$$FO_{2,OHO} = FO_{2,OHO,synth} + FO_{2,OHO,endo} \quad \textbf{(3.123)}$$

$$FO_{2,OHO,synth} = FCOD_{b,OHO} * (1 - f_{CV} * Y_{OHO}) \quad \textbf{(3.124)}$$

$$FO_{2,OHO,endo} = FCOD_{b,OHO} * f_{CV} * \left(1 - fX_{E,OHO}\right) * b_{OHO} * Y_{OHO,obs} * SRT \quad \textbf{(3.125)}$$

Total Oxygen Demand (carbonaceous), $\mathbf{FO_{2,C}}$;

$$FO_{2,C} = FO_{2,PAO} + FO_{2,OHO} \quad \textbf{(3.126)}$$

3.4 Modeling Approach

3.4.1 Plant layout

The plant was modelled by BIOWIN simulation programme using Johannesburg configuration according to the plant operational data of year 2008. The influent wastewater is introduced to second anaerobic reactor followed by anaerobic reactor3. The nitrification/denitrification processes are achieved in the following reactors of R1-R2**.** The reactors R3 and R4 were always kept as aerobic, however, the V_D/V ratio is adjusted depending on the oxygen concentration in R2, on the other hand conditions of R1 is set as completely anoxic corresponding to minimum V_D/V ratio of 25%. The plant layout in the model simulation shown in Figure 3.8.

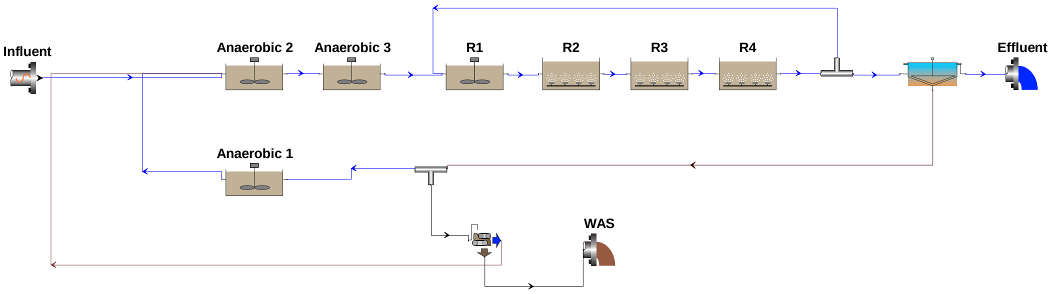


Figure 3.8: Plant layout for simulation in Biowin 3.0

3.4.2 Plant hydraulics

For the plant, under operating conditions, four aeration tank volumes considered seperately (11,000 m^3 for each tank) because of the operating conditions (anoxic-aerobic), but four final clarifiers and three anaerobic tank volumes considered a single volume (18,000 m^3 – 8100 m^3 for all tanks). All carousel reactors were regarded as completely mixed reactors since internal recirculation is much higher than that of influent flowrate (Insel et al., 2005).

3.4.3 Vesilind Settling model

Ekama et al., (1997) have reviewed the various types of settling that can occur in an SST, that is, discrete settling, hindered settling, zone settling, and compression. In grit and primary settling tanks, the settling process is dominated by discrete settling, in which the settling velocity is given by the Stokes settling velocity;

$$V_0 = [4gD_P(S_S - 1)/(3C_d)]^{1/2} \quad \textbf{(3.127)}$$

$C_d = 24/N_R$ for low particle Reynolds numbers ($N_R = D_P V_0/v<0.1$)

D_P = particle diameter (m of ft); and

S_s = particle spesific gravity (nondimesional)

Vesilind (1968) developed a settling velocity equation for zone settling;

$$V_S = V_0 e^{-kX} \quad \textbf{(3.128)}$$

k = constant (m^3/kg or L/mg)

X = the local concentration of the suspended solids (kg/ m^3 or mg/l)

V_S = zone settling (m/h)

V_0 =Stokes settling velocity (m/h)

3.4.4 Model implementation using operational data

For the steady-state conditions model implementation, the average influent data presented in the Table 3.16 for the plant for 2008 were used at operating conditions and the model calibrated.

For the dynamic simulation model implementation, daily sample results for determination of influent wastewater characterization and the other operating parameters (return sludge ratio, internal recycle ratio, sludge wasted, dissolved oxygen concentrations for each tank and oxidation reduction potential values for anaerobic tanks) were used and the model calibrated.

Table 3.16: Influent wastewater characterization used in steady-state simulation

Parameters	Abbreviation	Unit	Measurement
Sludge retention time	SRT	d	8.1
Influent flowrate	Q	m^3	100,000
Influent total COD	COD_0	mg/l	600
Influent TKN	TKN_0	mg/l	70
Influent total phosphorus concentration	$T\text{-}P_0$	mg/l	10
Influent nitrate concentration	$NO_3\text{-}N_0$	mg/l	BDL
Inorganic SS	Inor. SS	mg/l	270
Alkalinity	Alk.	mmol/l	8.78
pH			7.3

BDL: Below detection limit

3.4.5 Model calibration

In the absence of dissolved oxygen and oxidized nitrogen, the anaerobic hydrolysis process leads liberation of fermentable COD from slowly biodegradable substrate for nutrient removal. Being adjusted by anaerobic hydrolysis correction parameter, this process has a large impact on the PHA pool for an efficient EBPR since it generates readily biodegradable COD in bio-P reactors (Ekama and Wentzel, 1999). In this study, the contribution of anaerobic hydrolysis process to EBPR activity was found to be considerably low. This was also evidenced by low EBPR activity in connection with low anaerobic hydrolysis activity of ordinary heterotrophic organisms (OHOs) which was adjusted to 0.2. The activity of PAOs was also found to be low with respect to PHA sequestration rate assigned to 1.50 day^{-1} which is closer to the lower limit as given in Table 3.17.

In the calibration process, the half oxygen saturation constants for autotrophic and heterotrophic biomass were adjusted to higher levels reflecting relatively mass transfer limited conditions most probably due to elevated MLSS concentrations maintained in the reactor. Elevated half saturation parameters were also suggested for high MLSS concentrations i.e. in MBR operation (Insel et al., 2011). The anoxic hydrolysis reduction factor for hydrolysis were found to be in good agreement as suggested by Orhon et al. (1996) for similar activated sludge process in municipal wastewater treatment.

Table 3.17: Estimated model parameters

Parameter	**Unit**	**Calibrated**	**Default**	**Range***
Maximum growth rate for AOBs	day^{-1}	0.80	0.90	0.2-1.0
Ammonia oxidizer DO half saturation	mgO_2/L	0.40	0.25	0.002-0.5
Substrate half sat. constant for AOBs	mgN/L	1.00	0.70	0.1-1.0
Anoxic hydrolysis reduction factor	-	0.60	0.28	0.28-0.8
Anaerobic hydrolysis factor	-	0.20	0.50	0.1-0.5
Heterotrophic half saturation for DO	mgO_2/L	0.20	0.05	0.01-0.7
Poly-P half saturation	mgP/L	0.10	0.01	0.005-0.5
PAOs sequestration rate	day^{-1}	1.50	6.00	1.0-6.0

* Wentzel et al. (1992); Henze et al.(2000); Barker and Dold (1997); Murnleitner et al.(1997)

Settling model parameter; the settling model of Vesilind was used to estimate the effluent TSS concentration from the clarifier. The simulation showed that the TSS concentration always stayed below 30 mg/L during simulation period.

Table 3.18 : Vesilind settling model parameters

Parameter	**Abb.**	**Unit**	**Calibrated**
Max.Vesilind settling velocity	V_0	m/d	170
Vesilind hindered zone settling parameter	K	L/g	0.34
Specified TSS conc. for height calculation	-	mgTSS/l	2,500
Max. compactability constant	-	mg/l	15,000

4. RESULTS AND DISCUSSION

4.1 Design Results

4.1.1 Results of ATV-DVWK 131-E design

For this method, the results were calculated at 15 ^{0}C and 25 ^{0}C both based on COD and BOD (detail calculations described in the Appendix A) and shown in the table Table 4.1.

Table 4.1: Design results of the biological reactor (ATV-DVWK 131-E)

	Unit	Calculation	
Temperature	^{0}C	15	25
Sludge Age			
tss_{aerob}	d	4.9	2.2
tss_{dim}	d	8.1	3.3
Sludge Production			
MLSS	mgTSS/l	4,000	4,000
SP_d (BOD based)	kg/d	46,158	46,854
SP_d (COD based)	kg/d	40,598	41,394
Q_W	m^3/d	4,394	4,479
Nitrogen Balance			
Q_W N portion		0.025	0.025
WASN	kg/d	1,500	1,500
NN	kg/d	5,300	5,300
DN_{CAP}	kg/d	4,500	4,500
DN_{POT}	kg/d	4,478	4,316
Process Volume Calculations			
V_D/V ratio		0.39	0.40
Nitrification vol.	m^3	50,630	20,697
Denitrification vol.	m^3	32,370	13,798
Anaerobic volume	m^3	8,100	8,100
Total Volume	m^3	83,000	34,495
Air Demand			
OUR_C (BOD based)	kgO_2/d	35,523	33,378
OUR_C (COD based)	kgO_2/d	29,778	28,600
OUR_N	kgO_2/d	22,790	22,790
OUR_{DN}	kgO_2/d	12,441	12,441
*OR	kgO_2/d	66,384	64,970

*Based on COD

Design results of the final clarifier for ATV-DVWK 131-E design is shown in the Table 4.2.

Table 4.2: Design results of the final clarifiers (ATV-DVWK 131-E)

Parameter	**Unit**	**Calculation**	
Temperature	^0C	15	25
Return sludge parameters			
R		0.75	0.75
SVI	ml/g	100	100
T_{th}	h	2.3	2.3
SS_{BS}	mg/l	13,200	13,200
f_{obs}		0.7	0.7
SS_{RS}	mg/l	9,200	9,200
SS_{AT}	mg/l	4,480	4,480
Q_R	m^3/d	75,000	75,000
Final clarifier calculations			
q_A	m/h	1.08	1.08
Surface area	m^2	4,752	4,752
DSV	ml/l	425	425
q_{SV}	$l/(m^2/h)$	500	500
Final clarifier dimensions			
h_1 (clear water zone)	m	0.5	0.5
h_2 (seperation zone)	m	1.57	1.57
h_3 (storage zone)	m	0.68	0.68
h_4 (thickening zone)	m	1.31	1.31
Total water depth at 2/3 radius	m	4.06	4.06
Slope of basin	mm/m	60	60
Total water depth	m	3.6	3.6
Number of tank		4	4
Diameter of tank	m	41.14	41.14
Final clarifier retention time			
Volume of final clarifiers	m^3	17,017	17,017
Final clarifier retention time	h	2.2	2.2

4.1.2 Results of Water Environment Research Foundation (WERF) design

For this method, the results calculated at 15 ^{0}C and 25 ^{0}C (detail calculations described in the Appendix B) and the results presented in Table 4.3.

Table 4.3: Design results of the biological reactor (WERF)

	Unit	Calculation	
Temperature		15 ^{0}C	25 ^{0}C
Sludge Age			
tss_{aerob}	d	8	5.5
tss_{dim}	d	12	8.1
Sludge Production			
MLSS	mgTSS/l	4,000	4,000
SP_d	kg/d	48,430	50,888
F/M	d^{-1}	0.06	0.08
Q_W	m^3/d	5,241	5,496
Nitrogen Balance			
WASN	kg/d	1,273	1,524
NN	kg/d	5,527	5,276
DN_{CAP}	kg/d	4,727	4,476
DN_{POT}	kg/d	4,712	4,450
Process Volume Calculations			
V_D/V ratio		0.19	0.23
Denitrification vol.	m^3	27,917	24,167
Nitrification vol.	m^3	116,886	78,882
Anaerobic vol.	m^3	7,292	7,292
Total Volume	m^3	144,803	103,049
Air Demand			
OUR_C	kgO_2/d	36,834	33,971
OUR_N	kgO_2/d	23,932	22,894
OUR_{DN}	kgO_2/d	13,519	12,833
OR	kgO_2/d	47,246	43,885

4.1.3 Results of South African method

For this method, the results were calculated at 15 ^{0}C and 25 ^{0}C (detail calculations described in the Appendix C) and the results are presented in the Table 4.4.

Table 4.4: Design results of the biological reactor

	Unit	Calculation	
Temperature		15 ^{0}C	25 ^{0}C
Sludge Age			
tss_{aerob}	d	6	2
tss_{dim}	d	7.8	4.4
Sludge Production			
SP_d	kg/d	49,245	52,202
Q_W	m^3/d	5,330	5,682
Nitrogen Balance			
WASN	kg/d	1,829	2,078
NN	kg/d	5,016	4,940
DN_{CAP}	kg/d	4,216	4,149
DN_{POT}	kg/d	4,220	4,195
Process Volume Calculations			
MLSS	mgTSS/l	4,000	4,000
V_D/V ratio		0.16	0.25
Denitrification vol.	m^3	15,104	14,025
Nitrification vol.	m^3	80,654	43,149
Anaerobic vol.	m^3	9,576	5,717
Total Volume	m^3	95,758	57,174
Air Demand			
OUR_C	kgO_2/d	33,933	30,247
OUR_N	kgO_2/d	22,922	22,617
OUR_{DN}	kgO_2/d	12,057	11,864
OR	kgO_2/d	44,798	41,000

4.1.4 Comparison of design methods

4.1.4.1 At the fixed MLSS concentration

The facility designed with the different three methods (ATV-DVWK 131-E, WERF, and South African) and compared at 15 ^{0}C at the fixed MLSS concentration. The results presented in Table 4.5.

Table 4.5: Comparison of the different method results at the fixed MLSS

		ATV	South African	WERF
Aeration tank				
Min.Sludge age (SRT$_m$)	d	4.9	6	8
Total sludge age (SRT)	d	8.1	7.8	12
Mixed liq.sus.sol. (MLSS)	mgTSS/l	4,000	4,000	4,000
Sludge production				
Total sludge produc. (SP$_d$)	kg TSS/d	40,598	49,245	48,430
Sludge Waste (Q$_W$)	m^3/d	4,394	5,330	5,241
Process Volume				
Anoxic vol. fraction (V$_D$/V)		0.39	0.18	0.19
Total volume (V$_T$)	m^3	83,000	95,758	144,803
Aerobic volume (V$_A$)	m^3	50,630	78,522	116,686
Anoxic volume (V$_D$)	m^3	32,370	17,236	27,917
Anaerobic volume (V$_{AN}$)	m^3	8,100	9,576	7,292
Total Oxygen Requirement				
Oxygen req.for C (OUR$_C$)	kgO$_2$/d	29,778	33,933	36,834
Oxygen for N (OUR$_N$)	kgO$_2$/d	22,790	22,922	23,932
Oxygen credit DN (OUR$_{DN}$)	kgO$_2$/d	12,441	12,057	13,519
Total oxygen req. (OR)	kgO$_2$/d	66,387	44,797	47,246
Discharge concentrations				
Total Phosphorus (T-P)	mgP/l	3.4	1.7	5.4
Total Kjeldahl Nitro. (TKN)	mgN/l	2	2	1.6
Nitrate-Nitrogen (NO$_3$-N)	mgN/l	8	8	8
Total nitrogen (T-N)	mgN/l	10	10	9.6

The evaluation of process design prescriptions at fixed MLSS and different sludge age indicated that higher denitrification share (V_D/V) was required for ATV-131 compared to that of South African and WERF methods since the denitrification potential is accepted to be equally distributed and only a fraction (V_D/V) of biodegradable COD could be available for the denitrification. In this respect, the South African design yielded much smaller anoxic volume share ($V_D/V=0.18$) compared to ATV-131 protocol.

Similar effluent quality was obtained in ATV-131 and WERF designs except for the calculated effluent T-N concentration. The South African design method provided much lower level of effluent phosphate of 1.7 mg P/L compared to ATV-131 and WERF designs.

The total oxygen requirement (OR) remained around 70% of WERF and South African design methods compare with ATV-131 calculation.

4.1.4.2 Design at fixed MLSS and sludge age

The facility designed with the different three methods (ATV-DVWK 131-E, WERF, and South Africa) and compared at 15 ^{0}C at fixed MLSS concentration (4,000 mg/l) and sludge age (8.1 days). The results is presented in the Table 4.6.

The evaluation of process design prescriptions at fixed MLSS and sludge age indicated that higher denitrification share (V_D/V) was required for ATV-131 compared to that of South African and WERF methods since the denitrification potential is accepted to be equally distributed and only a fraction (V_D/V) of biodegradable COD could be available for the denitrification. On the other hand, the denitrification potential in WERF protocol is calculated by using a graphic normalized for different RBCOD fractions and F/M_b ratios of the anoxic reactor. In this respect, the WERF design yielded much smaller anoxic volume share ($V_D/V=0.17$) compared to ATV-131 protocol.

Similar effluent quality was obtained in ATV-131, WERF and South African designs except for the calculated effluent T-N concentration. The South African design method provided much lower level of effluent phosphate of 1.8 mg P/L compared to ATV-131 and WERF designs.

The total oxygen requirement (OR) remained around 65% of WERF and South African design method calculations compare with ATV-131.

Table 4.6: Comparison of the design methods results at fixed SRT and MLSS

		ATV	South Africa	WERF
Aeration tank				
Total sludge age (SRT)	d	8.1	8.1	8.1
Mixed liq.sus.sol. (MLSS)	mgTSS/l	4,000	4,000	4,000
Sludge production				
Total sludge produc. (SP_d)	kg TSS/d	40,598	49,023	50,888
Sludge Waste (Q_W)	m^3/d	4,394	5,305	5,507
Process Volume				
Anoxic vol. fraction (V_D/V)		0.39	0.17	0.23
Total volume (V_T)	m^3	83,000	99,271	103,049
Aerobic volume (V_A)	m^3	50,630	82,395	78,882
Anoxic volume (V_D)	m^3	32,370	16,876	24,167
Anaerobic volume (V_{AN})	m^3	8,100	9,927	7,292
Total Oxygen Requirement				
Oxygen req.for C (OUR_C)	kgO_2/d	29,778	34,179	33,841
Oxygen for N (OUR_N)	kgO_2/d	22,790	22,977	22,846
Oxygen credit DN (OUR_{DN})	kgO_2/d	12,441	12,091	12,802
Total oxygen req. (OR)	kgO_2/d	66,384	45,064	43,885
Discharge concentrations				
Total Phosphorus (T-P)	mgP/l	3.4	1.8	5.1
Total Kjelda. Nitro. (TKN)	mgN/l	2	2	2
Nitrate-Nitrogen (NO_3-N)	mgN/l	8	8	8
Total nitrogen (T-N)	mgN/l	10	10	10

4.2. Simulation Results

4.2.1. Steady state simulation results

Under steady-state simulation, design data used for modeling for each methods and the results of ATV 131 (2000), WERF and South African designs to compare the effluent quality, total sludge production, oxygen requirement, MLSS concentration at the same operating conditions using default and calibrated parameters (Table 3.17).

As shown in the Table 4.7, design results and the model results compared at fixed sludge age and different V_D/V_T ratios for the facility.

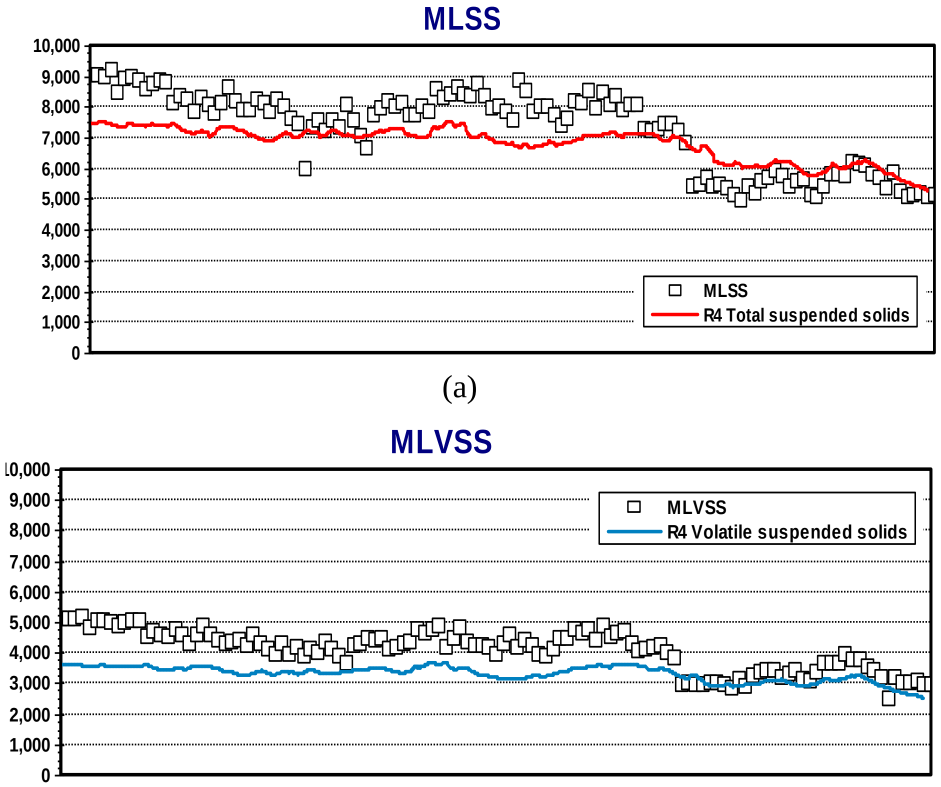
Table 4.7 given below summarizes the design and modeling results using the same wastewater treatment plant data. The information basically illustrates the differences in design and operation together with linking to certain process conditions depicted by estimated model parameters discussed in Table 3.17 above. In this context, compared to three design prescriptions, 25% lower sludge (in dry solids) was produced in real system for WERF and South African, and similar results found for ATV-131. On the other hand, modeling results for oxygen requirement was found to be 10% higher than the design results for WERF and South African on a yearly average loading basis. The effluent total nitrogen (TN) concentration was 35% lower than ATV-131 and South African design methods. Modeling results for effluent PO_4-P is higher than the design results. The low phosphate removal rate is attributed to low influent VFA concentration together with reduced anaerobic hydrolysis rate and poor PAO activities. In this respect, it was suggested to utilize new process options to boost EBPR by utilizing the slowly biodegradable COD with the fraction of 70% of the total influent COD.

Table 4.7: Steady-state simulation results at the design sludge ages (8.1 days)

Parameter	Unit	DESIGN RESULTS			MODELING RESULTS					
					Calibrated			Default		
		ATV	WERF	S.AFR.	ATV	WERF	S.AFR.	ATV	WERF	S.AFR.
Total sludge age, SRT	days	8.1	8.1	8.1	8.1	8.1	8.1	8.1	8.1	8.1
Anoxic volume ratio, V_D/V_T	-	0.39	0.23	0.18	0.39	0.23	0.18	0.39	0.23	0.18
Internal recycle, IR	-	5.6	4.8	4.9	5.6	4.8	4.9	5.6	4.8	4.9
MLSS concentration	$kgSS/m^3$	4.0	4.0	4.0	4.1	3.3	3.4	4.1	3.3	3.4
Oxygen requirement, OR_{ave}	kgO_2/d	66,384	43,885	45,064	45,524	47,448	47,472	47,088	49,128	51,504
Sludge production, SP_d	kgSS/d	40,590	50,888	49,023	40,905	40,901	41,147	41,713	41,152	41,661
Effluent quality										
NH_4-N	mgN/l	0.5	0.6	0.2	1.4	0.5	0.6	0.5	0.14	0.2
Soluble org.N	mgN/l	1.5	1.0	1.8	1.7	1.6	1.6	1.7	1.81	1.8
NO_X-N	mgN/l	8.0	8.0	8.0	3.5	7.0	4.2	6.3	12.88	20.5
T-N	mgN/l	10.0	9.6	10.0	6.6	9.1	6.4	8.5	14.69	22.5
PO_4-P	mgP/l	3.4	5.1	1.8	6.0	6.2	6.1	5.9	5.97	5.94

4.2.2. Dynamic simulation results

In the dynamic simulation, daily operational data in 2008 were used for modeling. The results of the simulation were evaluated with the plant data together with the dynamic simulation in Figures 4.1, 4.2 and 4.3. Simulation results shown in Figure 4.1a, MLSS concentrations under operating conditions is sligthly more than the model results. For the facility MLSS concentrations were kept in the range of 7,000-9,000 mg/l at high sludge age around 20 days. Especially during the summer months, there is a considerable difference between the model results MLSS concentrations and the plant aeration tank MLSS concentrations and it caused the operation of a high concentrations until november. For the facility MLVSS concentrations were kept in the range of 3,500-5,000 mg/l as plotted in Figure 4.1b.



(a)

(b)

Figure 4.1 : Dynamic simulation results of the (a) MLSS and (b) MLVSS concentrations

The average mass distribution in the wastewater treatment plant is illustrated in Figure 4.2 and the simulation results for ammonia nitrogen is shown in Figure 4.3.

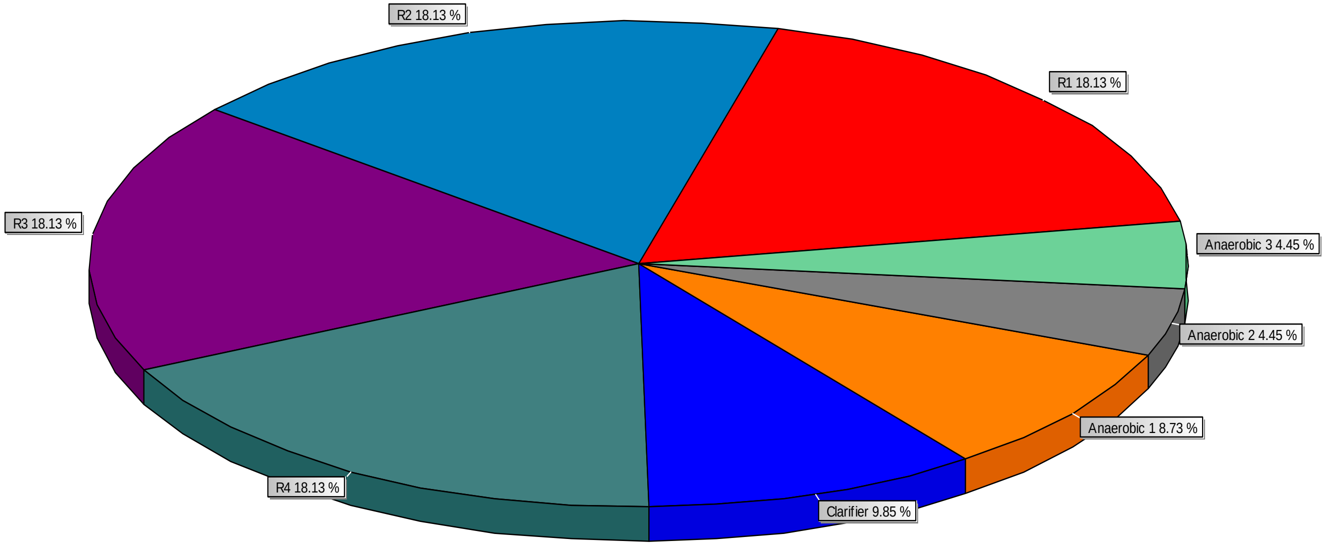


Figure 4.2: Average mass distribution in the wastewater treatment plant

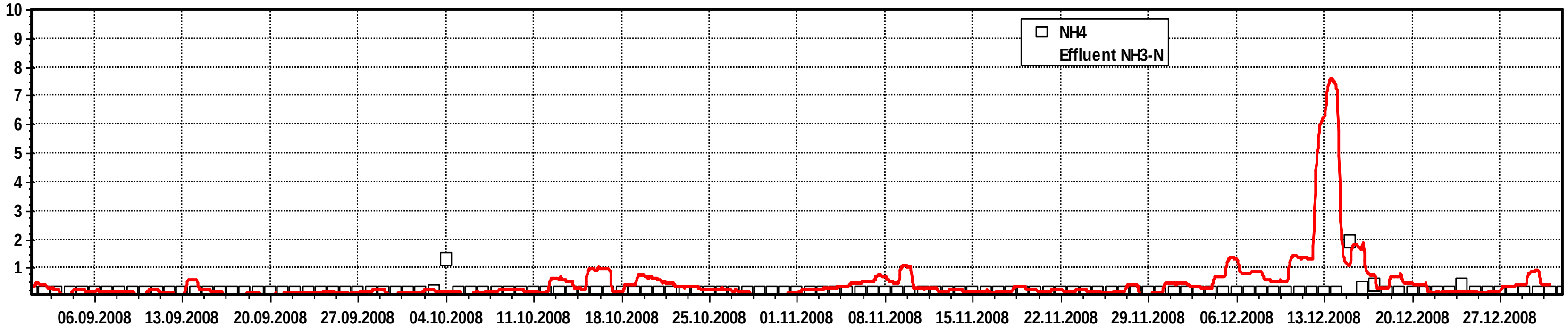


Figure 4.3: Dynamic simulation results of the NH4-N concentrations

Simulation results shown in Figure 4.3, for most days of the year, NH_4-N discharge concentrations and the model results are close to each other. From the results, NH_4-N removal for the facility is nearly 95-100%. A small hump at day 13 December simulated by the model is due to low oxygen concentration in the reactor

Simulation results shown in Figure 4.4, model results and the facility discharge concentrations for NO_3-N are close to each other. Effluent discharge concentration of nitrate is around 6-8 mgN/l. As seen in the figure, in some days, NO_3-N concentration exceeds 8 mg/l. Anoxic volume is insufficient in some cases. The other factor affecting NO_3-N removal is internal recycle ratio. In the facility, internal recycle is from aerobic reactor to anoxic reactor and this ratio is nearly 4 Q_{in}. As the internal recycle ratio increased, it affects NO_3-N removal positively. However, the nitrate recycle was found not to be sufficient to keep the first and second anoxic reactors anoxic.

Simulation results shown in Figure 4.5 for T-P concentrations, although the results in some months are very close each other, in some months there is small differences between each other because of the operating conditions. As described above, increasing of MLSS concentration leads to an increase in the sludge age. Under operating conditions system sludge age is nearly 20 days and this value is quite high for efficient phosphorus removal (Wentzel et. Al., 1990). Effluent phosphorus discharge concentration is around 3-5 mgP/l during the course of simulation period. The other important factor is influent COD/T-P ratio. As this ratio increases, it affects phosphorus removal positively. The average COD/TP ratio was around 75-85. The most important issue is the influent VFA concentration which is required for enhanced phosphorus removal. The average concentration in the influent is 30 mgCOD/l comparably lower than that of wastewater characterization in Europe.

As shown in the figure 4.6, as the influent total phosphorus concentration increase, effluent phosphorus concentrations increase in most days of the year. But in some days, observed the opposite cases because of effluent nitrate concentrations. Simulation results to evaluate the affect of NO_3-N on TP concentrations in the anaerobic and anoxic reactor were shown in figure 4.6 and 4.7.

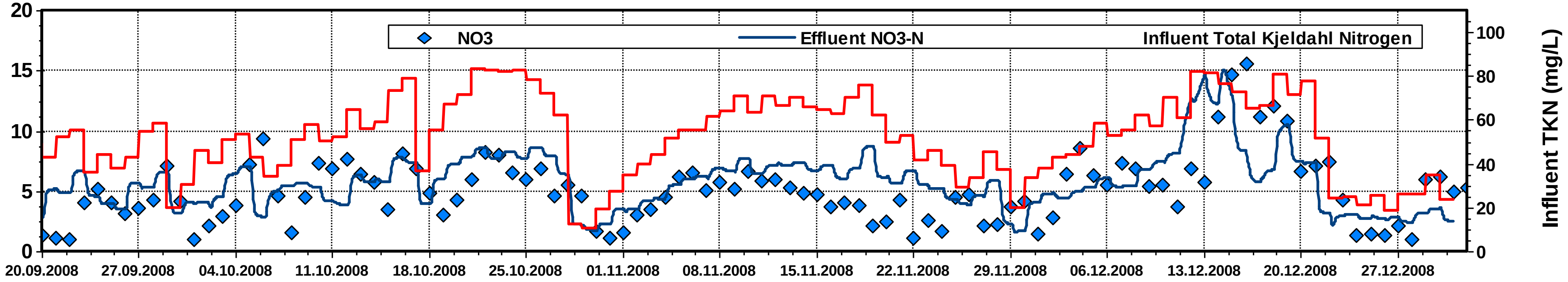


Figure 4.4: Influent TKN concentrations and dynamic simulation results for NO_3 concentrations

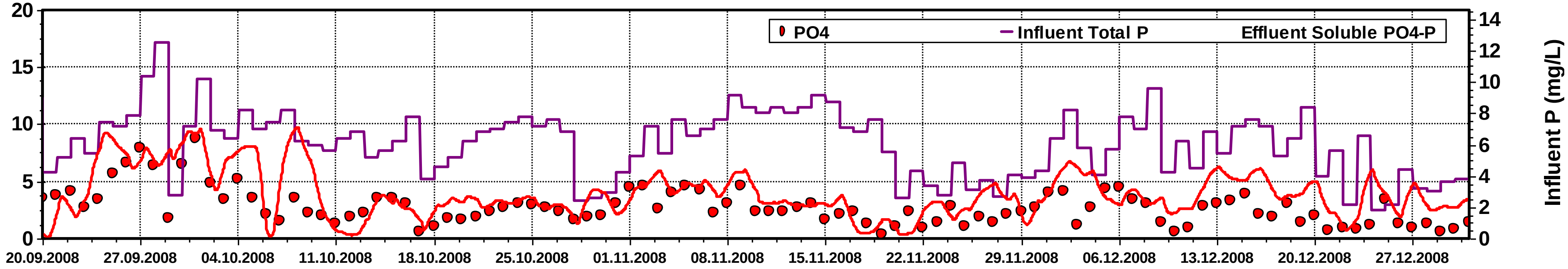


Figure 4.5: Dynamic influent TP concentration and simulation results of the PO4-P concentrations

NO_3-N concentration conveyed with return activated sludge enters into the anaerobic tank 1 and the concentrations seen in the Figure 4.7 is nearly 3-4 mg/l. After the revision of the facility, the anaerobic tank #1 was used as denitrification RAS. Accordingly, the phosphate is released in the absence of oxygen and nitrate. As a results phosphate is released due to anaerobic cleavage.

As shown in the figure 4.7, in September, NO_3-N concentrations in the anoxic tank (R1) is nearly zero and in the anoxic tank (R2) is low. During these days, it is clearly seen that in the figure 4.6 that the effluent phosphorus concentrations increased because of secondary phosphorus release.

On the other hand, the effluent TN level is far below the effluent limit of TN<10 mg/L. In this respect, the biodegradable COD used for excess denitrification can be directed to remove phosphate by advanced process control tools. As a result, much more EBPR can be achieved with securing optimal process conditions. Prefermentation of primary sludge could be a viable alternative for increasing the EBPR capacity.

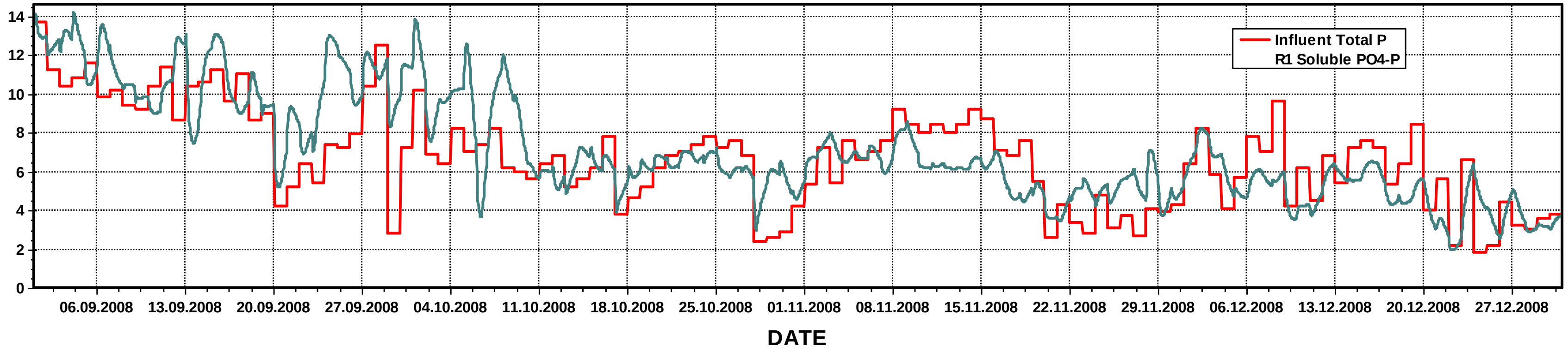


Figure 4.6 : Influent TP and dynamic simulation results of PO4-P concentrations in R1 reactor

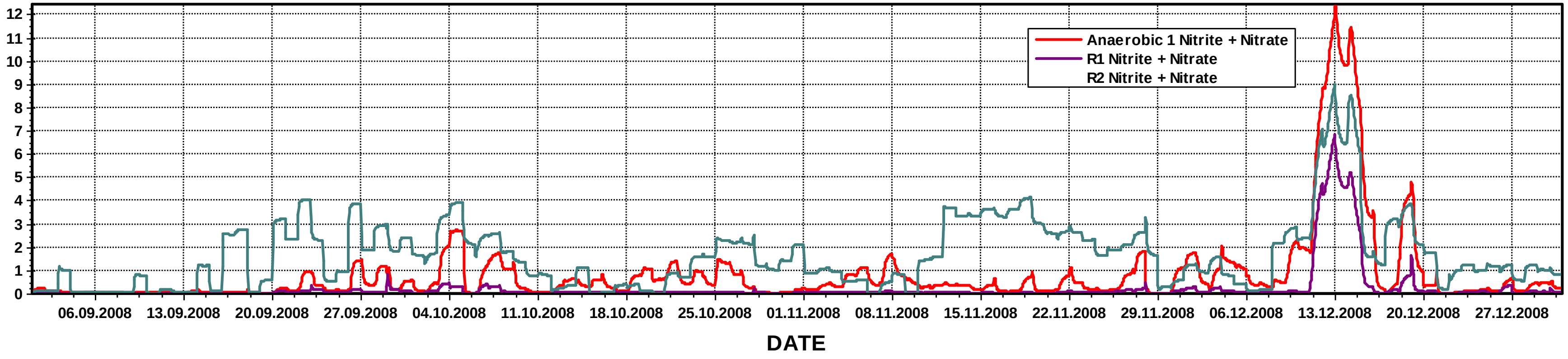


Figure 4.7 : Dynamic simulation results of the NO_X (NO_2+NO_3) concentrations in the anaerobic and anoxic reactor

5.CONCLUSION AND FUTURE PERSPECTIVES

The conclusions derived from the study were summarized accordingly:

- The design methods of WERF, ATV and South Africa for BNR activated sludge design at SRT of 8.1 days resulted in relatively close effluent TN concentrations. WERF and South African Methods yielded highest sludge production compared to ATV method.
- The effluent phosphate concentration for South Africa was found to be lower compared to other methods. The TP removals for ATV, WERF and South Africa were found to be 65%, 50% and 80%, respectively,
- The plant was modelled in steady-state and dynamic simulations. And the effluent data could be well characterized with the dynamic model simulations.
- Especially during the summer months, the SRT of the system was set to higher values around 20 days. This affected biological phosphorus removal negatively. The steady state simulations carried out with lower SRT yielded much lower effluent P concentrations. During the operation period, more sludge should be wasted and the system should be operated in shorter sludge age.
- The ammonia concentrations were kept below 1 mg/L which could also depicted by the model. The model parameters of nitrifiers could achieve successful nitrification under dynamic conditions at around 20 days of SRT.
- One important issue is that the nitrate in R1, R2 reactors are quite low that make these reactors as anaerobic. In this case, the denitrification potential can not efficiently be used. Second, secondary P release causes increase of effluent phosphate concentration.
- It is suggested that the reactor conditions prerequisite for optimal process (anoxic, aerobic) should be controlled with appropriate process control tools and process operation algorithms. The utilization of calibrated models could provide information on the optimal operational conditions to be secured by optimal instrumentation of wastewater treatment plant facility.

- Compared to suggested/default model parameters, the activated sludge yielded low EBPR capacity delineated with low substrate storage rate and low anaerobic hydrolysis rate. Higher half saturation model parameters govern relatively difficult mass transfer conditions at high MLSS concentrations in the system like MBR systems.
- The influent wastewater characterization (i.e VFA content) and anaerobic hydrolysis of slowly biodegradable organics play important role on the effluent quality. In that way, specific parameters resulted in having different picture in real life compared to that of design methods.
- The activated sludge modeling tools fills the gap between design methods and real operation by highlighting critical processes. In this respect, the design methods could also be adapted to local conditions with the aid of dynamic process modeling. Moreover, the robustness of suggested processes could be evaluated by implementing relevant control algorithms.

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APPENDICES

APPENDIX A : ATV-DVWK 131-E design method calculations

APPENDIX B : Water Environment Research Foundation (WERF) design method calculations

APPENDIX C : South Africa design method calculations

APPENDIX D : Kinetic coefficients and assumptions used in design methods

APPENDIX A (ATV-DVWK 131-E design method calculations)

All calculations at 15 ^{0}C

Sludge retention time (SRT)

The aerobic dimensioning sludge age to be maintained for nitrification is;

$t_{SS,aerob} = 3.4 * 1.103^{(15-T)} * SF_n$ $\qquad$ $SF_n = 1.45$

$t_{SS,aerob} = 4.93\ d$

For nitrification and denitrification the dimensioning sludge age results as follows;

$$t_{SS,dim} = \frac{t_{SS,aerob}}{\left(1 - \frac{V_D}{V_T}\right)}$$

$t_{SS,dim} = 8.1\ d$

Sludge production (SP_d)

COD based;

For sludge production (SP_d) and oxygen uptake rate for carbon (OUR_C) calculations based on COD in ATV-131, there are some assumptions which is different for wastewater characterization used in design.

$S_I = 0.05 * COD_T$

$S_I = 30\ mgCOD/l$

$X_f = B * X_{TSS}$

$$\left.\begin{array}{l} B = 0.2\ (\text{if primary sedimentation present}) \\ B = 0.3\ (\text{if not primary sedimentation}) \end{array}\right\}\ \text{Sand, grit etc in SS}$$

$X_f = 0.3 * 535$

$X_f = 161\ mgSS/l$

$X_{COD} = X_S + X_{i,COD}$

$X_{i,COD} = 0.25 * X_{COD}$

$X_{i,COD} = 105 \text{ mgCOD/l}$

$X_S = 420 - 105$

$X_S = 315 \text{ mgCOD/l}$

$X_{COD,BM} = Y_H/(1 + (b_H * F_T * SRT)) * X_S + S_S$

$Y_H = 0.67 \text{ gCOD/gCOD}_{deg}$

$b_H = 0.17 \text{ d}^{-1}$

$F_T = 1.072^{(T-15)} = 1$

$X_{COD,BM} = 0.67/(1 + 0.17 * 1 * 8.1)) * (315 + 150)$

$X_{COD,BM} = 131 \text{ mgCOD/l}$

$X_P = 0.2 * (b_H * F_T * SRT) * \ X_{COD,BM}$

$X_P = 0.2 * (1 + 0{,}17 * 1 * 8.1) * 131$

$X_P = 36 \text{ mgCOD/l}$

$X_{WAS,COD} = X_{i,COD} + X_{COD,BM} + X_P$

$X_{WAS,COD} = 105 + 131 + 36$

$X_{WAS,COD} = 272 \text{ mgCOD/l}$

$$SP_{d,COD} = Q * \left(\frac{X_{WAS,COD}}{0.8} * 1.45\right) + X_f))/1{,}000$$

$$\frac{gCOD}{gSS} = 1.45$$

Organics in wasted sludge = 0.8

$$SP_{d,COD} = 100{,}000 * \left(\frac{272}{0.8} * 1.45\right) + 161))/1{,}000$$

$SP_{d,COD} = 39{,}518 \text{ kg/d}$

$SP_{d,BioP} = COD_T * Q_{in} * 0.006 * 3/1{,}000$

$SP_{d,BioP} = 1{,}080 \text{ kg/d}$

Total sludge production based on COD;

$$SP_d = SP_{d,COD} + SP_{d,BioP}$$

$$SP_d = 39{,}518 + 1{,}080$$

$$SP_d = 40{,}598 \text{ kg/d}$$

BOD based;

Observed Hetetrophic Yield Coefficient ($Y_{obs,H}$);

$$Y_{obs,H} = (0.75 + 0.6 * \frac{X_{SS,IAT}}{C_{BOD,IAT}} - \frac{(1-0.2) * 0.2 * 0.17 * 0.75 * t_{SS} * F_T}{1 + 0.17 * t_{SS} * F_T}$$

$$Y_{obs,H} = (0.75 + 0.6 * \frac{535}{325} - \frac{(1-0.2) * 0.2 * 0.17 * 0.75 * 8.1 * 1}{1 + 0.17 * 8.1 * 1}$$

$$Y_{obs,H} = 1.38 \text{ mgVSS/mgBOD}$$

Sludge Production From Carbon Removal;

$$SP_{d,BOD} = B_{d,BOD} * Y_{obs,H}$$

$$SP_{d,BOD} = 32{,}500 * 1.38$$

$$SP_{d,BOD} = 44{,}696 \text{ kg/d}$$

Total sludge production based on BOD;

$$SP_d = SP_{d,BOD} + SP_{d,BioP}$$

$$SP_d = 44{,}696 + 1{,}463$$

$$SP_d = 46{,}158 \text{ kg/d}$$

Req. DN capacity

The daily nitrate concentration to be denitrified

$$S_{NO3,D} = C_{N,IAT} - C_{N,EST} - X_{orgN,BM}$$

$$S_{NO3,D} = 70 - 10 - 0.025 * 600$$

$$S_{NO3,D} = 45 \text{ mgN/l}$$

Req. denitrification capacity;

$$\mathrm{Req.DN_{CAP}} = \frac{\mathrm{S_{NO3,D}}}{\mathrm{C_{BOD}}}$$

$$\mathrm{Req.DN_{CAP}} = \frac{45}{325}$$

$$\mathrm{Req.DN_{CAP}} = 0.138$$

Mixed Liquir Suspended Solids (MLSS)

Bottom sludge suspended solids

$$\mathrm{SS_{BS}} = \frac{1{,}000}{\mathrm{SVI}} * \sqrt[3]{\mathrm{t_{Th}}}$$

$\mathrm{t_{Th}} = \text{select } 2.3 \text{ h}$

$$\mathrm{SS_{BS}} = \frac{1{,}000}{100} * \sqrt[3]{2}.3$$

$$\mathrm{SS_{BS}} = 13{,}200 \text{ mgSS/l}$$

Return sludge suspended solids

$$\mathrm{SS_{RS}} = \mathrm{f_{obs}} * \mathrm{SS_{BS}}$$

$\text{Dilution factor of return sludge} = \mathrm{f_{obs}} = 0.7$

$$\mathrm{SS_{RS}} = 0{,}7 * 13{,}200$$

$$\mathrm{SS_{RS}} = 9{,}240 \text{ mgSS/l}$$

Mixed liquir suspended solids

$$\mathrm{SS_{AT}} = \frac{\mathrm{R} * \mathrm{SS_{RS}}}{1 + \mathrm{R}}$$

$\mathrm{R} = \text{select } 0.75$

$$\mathrm{SS_{AT}} = \frac{0.75 * 9{,}240}{1 + 0.75}$$

$$\mathrm{SS_{AT}} = 4{,}000 \text{ mg/l}$$

Nitrogen balance

$\text{Waste activated sludge N portion} = 0.025$

Nitrogen incorporated to sludge;

$$\mathrm{WASN} = \mathrm{B_{d,COD}} * 0.025$$

$$\mathrm{WASN} = 60{,}000 * 0.025$$

$$\mathrm{WASN} = 1{,}500\ \mathrm{kg/d}$$

Nitrogen to be denitrified;

$$\mathrm{NN} = \mathrm{Q} * (\mathrm{C_{TKN}} - \mathrm{C_{TKN,eff}} - (\mathrm{C_{COD}} * 0.025))/1{,}000$$

$$\mathrm{NN} = 100.000 * (70 - 2 - (600 * 0.025))/1{,}000$$

$$\mathrm{NN} = 5{,}300\ \mathrm{kg/d}$$

Denitrification capacity;

$$\mathrm{DN_{cap}} = \mathrm{Q} * \mathrm{S_{NO3,D}}/1000$$

$$\mathrm{DN_{cap}} = 100{,}000 * 45/1{,}000$$

$$\mathrm{DN_{cap}} = 4{,}500\ \mathrm{kg/d}$$

Oxygen uptake for carbon (COD based);

$$\mathrm{OUR_C} = \mathrm{Q} * (\mathrm{C_{COD}} - (\mathrm{S_I} + \mathrm{X_{WAS,COD}}))/1000$$

$$\mathrm{OUR_C} = 100{,}000 * (600 - (30 + 272))/1000$$

$$\mathrm{OUR_C} = 29{,}778\ \mathrm{kgO_2/d}$$

Oxygen uptake for carbon (BOD based);

$$\mathrm{OUR_C} = \mathrm{B_{d,BOD}} * \left(0.56 + \frac{0.15 * \mathrm{t_{SS}} * \mathrm{F_T}}{1 + 0.17 * \mathrm{t_{SS}} * \mathrm{F_T}}\right)$$

$$\mathrm{OUR_C} = 32{,}500 * \left(0.56 + \frac{0.15 * 1 * 8.1}{1 + 0.17 * 8.1 * 1}\right)$$

$$\mathrm{OUR_C} = 35{,}523\ \mathrm{kgO_2/d}$$

Denitrification potential;

$$\mathrm{DN_{pot}} = \mathrm{a} * 0.75 * \frac{\mathrm{OU_{d,C}}}{2.9} * \frac{\mathrm{V_D}}{\mathrm{V_T}}$$

$$\mathrm{DN_{pot}} = 1.25 * 0{,}75 * \frac{35{,}523}{2.9} * 0.39$$

$$\mathrm{DN_{pot}} = 4{,}478\ \mathrm{kg/d}$$

Internal recycle

$$RC = \left(\frac{S_{inorgN}}{S_{NO3,EST}}\right) - 1$$

$$S_{inorgN} = C_{TKN,IAT} - S_{orgN,EST} - S_{NH4,EST} - X_{ORG,BM}$$

$$S_{inorgN} = 70 - 2 - 15$$

$$S_{inorgN} = 53 \text{ mgN/l}$$

$$RC = \left(\frac{53}{8}\right) - 1$$

$$RC = 5.6$$

Process volume calculations

Total process volume;

$$V_T = SRT * SP_d / SS_{AT}$$

$$V_T = 8.1 * \frac{40{,}598}{4{,}000} * 1{,}000$$

$$V_T = 83{,}000 \text{ m}^3$$

Nitrification volume;

$$V_A = V_T * (1 - \frac{V_D}{V_T})$$

$$V_A = 50{,}749 \text{ m}^3$$

Denitrification volume;

$$V_D = V_T * (\frac{V_D}{V_T})$$

$$V_D = 32{,}251 \text{ m}^3$$

Process Air Demand

Oxygen Uptake for nitrification;

$$OUR_N = 4.3 * NN$$

$$OUR_N = 4.3 * 5{,}300$$

$$\mathrm{OUR_N} = 22{,}790\ \mathrm{kgO_2/d}$$

Carbon removal by denitrification;

$$\mathrm{OUR_{DN}} = 2.9 * \mathrm{N_{DN}}$$

$$\mathrm{OUR_{DN}} = 2.9 * 4.290$$

$$\mathrm{OUR_{DN}} = 12{,}441\ \mathrm{kgO_2/d}$$

For carbon removal;

$$\mathrm{OUR_C} = \mathrm{Q} * (\mathrm{C_{COD}} - (\mathrm{S_I} + \mathrm{X_{WAS,COD}}))/1{,}000$$

$$\mathrm{OUR_C} = 100{,}000 * (600 - (30 + 272))/1{,}000$$

$$\mathrm{OUR_C} = 29{,}778\ \mathrm{kgO_2/d}$$

Total daily oxygen uptake;

$$OR = f_C \times (OUR_C - OUR_{DN}) + f_N \times OUR_N$$

$$OR = 1.2 \times (29{,}778 - 12{,}441) + 2.0 \times 22{,}790$$

$$OR = 66{,}384\, kgO_2 / d$$

Phosphorus removal, P removal;

$$\mathrm{X_{P,BioP}}(\text{enhanced uptake}) = 0.006 * \mathrm{C_{COD,IAT}}$$

$$\mathrm{X_{P,BioP}} = 0.006 * 600$$

$$\mathrm{X_{P,BioP}} = 3.6\ \mathrm{mgP/l}$$

$$\mathrm{X_{P,BM}}(\text{normal uptake}) = 0.005 * \mathrm{C_{COD,IAT}}$$

$$\mathrm{X_{P,BM}} = 3\ \mathrm{mgP/l}$$

$$\text{Effluent T} - \text{P concentration} = 10 - (3.6 + 3)$$

$$\text{Effluent T} - \text{P concentration} = 3.4\ \mathrm{mgP/l}$$

APPENDIX B (WERF) design method calculations

Specific growth rate for nitrification, (μ_n); (All calculations at 15 ^{0}C)

$$\mu_n = \left(\frac{\mu_{n,m} * N}{K_n + N}\right) * \left(\frac{DO}{K_0 + DO}\right) - k_{dn}$$

$$\mu_n = 0.132\ d^{-1}$$

Max. Spesific growth rate of nitrifying organisms, (μ_{nm});

$$\mu_{n,m} = 0.75 * 1.072^{(T-20)}$$

$$\mu_{n,m} = 0.53\ d^{-1}$$

Half velocity constant, K_n;

$$K_n = 0.74 * 1.053^{(T-20)}$$

$$K_n = 0{,}57\ mg/l$$

Endogenous decay coefficient for nitrifying organisms, k_{dn};

$$k_{dn} = 0.08 * 1.04^{(T-20)}$$

$$k_{dn} = 0.072\ d^{-1}$$

$$N = 0.50\ mg/l$$

$$DO = 2\ mg/l$$

$$K_0 = 0.5\ mg/l$$

(All kinetic coefficients presented in Appendix D)

Minimum sludge age for nitrification, (SRT_m);

$$SRT_m = \frac{1}{\mu_n}$$

$$SRT_m = \frac{1}{0.132}$$

$$SRT_m = 8\ d$$

Total sludge age, (SRT);

$\mathrm{SF} = \text{Safety factor}$

$\mathrm{SF} = \text{TKN peak/TKN average}$

$\mathrm{SF} = \text{ assume } 1.5$

$\mathrm{SRT} = \mathrm{SRT_m} * \mathrm{FS}$

$\mathrm{SRT} = 8 * 1.5$

$\mathrm{SRT} = 12 \text{ d}$

For biomass production, ($\mathbf{P_{X,bio}}$); first effluent BOD concentration is calculated;

S effluent concentration, (S_{eff});

$$\mathrm{S_{eff}} = \mathrm{K_s} * \frac{(1 + \mathrm{k_d} * \mathrm{SRT})}{\mathrm{SRT} * (\mu_m - \mathrm{k_d}) - 1}$$

$\mathrm{S_{eff}} = 1.03 \text{ mg BOD/l}$

Biomass production, ($P_{X,bio}$);

$$\mathrm{P_{X,bio}} = \frac{\mathrm{Q} * \mathrm{Y} * (\mathrm{S_0} - \mathrm{S})}{(1 + \mathrm{k_d} * \mathrm{SRT})} + \frac{\mathrm{f_d} * \mathrm{k_d} * \mathrm{Q} * \mathrm{Y} * (\mathrm{S_0} - \mathrm{S}) * \mathrm{SRT}}{(1 + \mathrm{k_d} * \mathrm{SRT})} + \frac{\mathrm{Q} * \mathrm{Y_n} * \mathrm{NO_X}}{(1 + \mathrm{k_{dn}} * \mathrm{SRT})}$$

$\mathrm{P_{X,bio}} = 10{,}608 \text{ kgVSS/d}$

N oxidized to NO_3, (NO_X);

$\mathrm{NO_X} = \mathrm{TKN} - \mathrm{N_e} - 0.12 * \mathrm{P_{X,bio}}/\mathrm{Q}$

$\mathrm{NO_X} = 70 - 2 - 0.12 * 10{,}608$

$\mathrm{NO_X} = 56 \text{ mgN/l}$

Mass of VSS and TSS

Concentration of VSS in the aeration basin, ($P_{X,VSS}$);

For this equation, the non biodegradable fraction of the VSS (nbVSS) was assumed as 35%.

$nbVSS = 0.35 \times 270 = 94.5 mgVSS/L$

$\mathrm{P_{X,VSS}} = \mathrm{P_{X,bio}} + \mathrm{Q} * \mathrm{nbVSS}$

$$P_{X,VSS} = 10{,}608 + 100{,}000 * 94.5/1{,}000$$

$$P_{X,VSS} = 20{,}058\ \mathrm{kg/d}$$

$$MX_{VSS} = P_{X,VSS} * SRT$$

$$MX_{VSS} = P_{X,VSS} * 12$$

$$MX_{VSS} = 239{,}288\ \mathrm{kg}$$

Concentration of TSS in the aeration basin, ($P_{X,TSS}$);

$$P_{X,TSS} = (P_{X,bio}/0.85) + Q * nbVSS + Q * (TSS_0 - VSS_0)$$

$$P_{X,TSS} = 48{,}430\ \mathrm{kg/d}$$

$$MX_{TSS} = P_{X,TSS} * SRT$$

$$MX_{TSS} = 48{,}430 * 12$$

$$MX_{TSS} = 579{,}212\ \mathrm{kg}$$

Volume of the aeration basin, (V_T);

*MLSS concentration= 4,000 mg/l (select as ATV-DVWK 131-E)

$$V_T = \frac{MX_{TSS}}{MLSS}$$

$$V_T = \frac{579{,}212}{4{,}000} \qquad V_T = 144{,}803\ \mathrm{m^3}$$

Hydraulic retention time (r);

$$r = \frac{V}{Q + Q * R}$$

$$r = \frac{144{,}803}{100{,}000 + 100{,}000 * 0.75} \qquad r = 19.9\ \mathrm{h}$$

F/M ratio;

$$F/M = \frac{Q * BOD_0}{V * MLSS}$$

$$F/M = \frac{100{,}000 * 32{,}500}{144{,}803 * 4{,}000}$$

$$F/M = 0.056\ \mathrm{d^{-1}}$$

Yield coefficient, (Y_{obs});

$$Y_{obs,TSS} = \frac{P_{X,TSS}}{bCOD\ removed}$$

$$Y_{obs,TSS} = 0{,}93\ \frac{gTSS}{gbCOD} * 1.6 \frac{gbCOD}{gBOD} = 1.49 \frac{gTSS}{gBOD}$$

Oxygen requirement, (OR);

Oxygen uptake rate for carbon, (OUR_C);

$$OUR_C = Q * (S_0 - S) - 1.42 * P_{X,bio}$$

$$OUR_C = 100{,}000 * (520 - 1.03) - 1.42 * 10{,}608$$

$$OUR_C = 36{,}834\ kgO_2/d$$

Oxygen uptake rate for nitrogen, (OUR_N);

$$OUR_N = Q * NO_X * 4.33$$

$$OUR_N = 100{,}000 * 56 * 4.33$$

$$OUR_N = 23{,}932\ kgO_2/d$$

Denitrification; Active biomass concentration, (X_b);

$$X_b = \frac{Q * SRT}{V} * \frac{Y * (S_0 - S)}{(1 + k_d * SRT)}$$

$$X_b = 834\ mg/l$$

IR ratio, (IR);

$$NO_{3,e} = \ 8\ mg/l$$

$$IR = \frac{NO_X}{NO_{3,e}} - 1 - R$$

$$IR = \frac{56}{8} - 1 - 0.75$$

$$IR = 5.25$$

NO_3 fed to the anoxic tank, (NO_X feed);

$$Flowrate\ to\ the\ anoxic\ tank = IR * Q + R * Q$$

$$\text{Flowrate to the anoxic tank} = 590{,}882\ \mathrm{m^3/d}$$

$$\mathrm{NO_X}\ \text{feed} = \mathrm{NO_{3,e}} * \text{Flowrate to the anoxic tank}$$

$$\mathrm{NO_X}\ \text{feed} = 4{,}727\ \mathrm{kg/d}$$

Anoxic volume, (V_D);

(to calculate anoxic volume, equalize $\mathrm{NO_X}$ feed to $\mathrm{NO_3}$ reduced)

$$r = 4.8\ \mathrm{h} = 0.2\ \mathrm{d}$$

$$V_D = r * Q$$

$$V_D = 27{,}917\ \mathrm{m^3}$$

F/M_b ratio;

$$\frac{F}{M_b} = \frac{Q * S_0}{V_D * X_b}$$

$$\frac{F}{M_b} = \frac{100{,}000 * 325}{27{,}917 * 834}$$

$$\frac{F}{M_b} = 1.40\ \mathrm{d^{-1}}$$

Spesific denitrification rate, ($SDNR_b$);

For $\mathrm{NO_3}$ reduced calculation, $SDNR_b$ is calculated for the selected temperature (15 ^{0}C). SNDR for 20 ^{0}C is selected from the figure below for F/M_b value and calculated for 15 ^{0}C using the temperature correction equation.

$$\text{Frac. of RBCOD} = \frac{S_{S,i}}{\text{RBCOD}}$$

$$\text{Frac. of RBCOD} = 0.23$$

$SDNR_b = 0.23\ \mathrm{d^{-1}}$ at 20 C (selected from the figure)

$$SDNR_{15} = SDNR_b * 1.026^{(T-20)}$$

$$SDNR_{15} = 0.20\ \mathrm{d^{-1}}$$

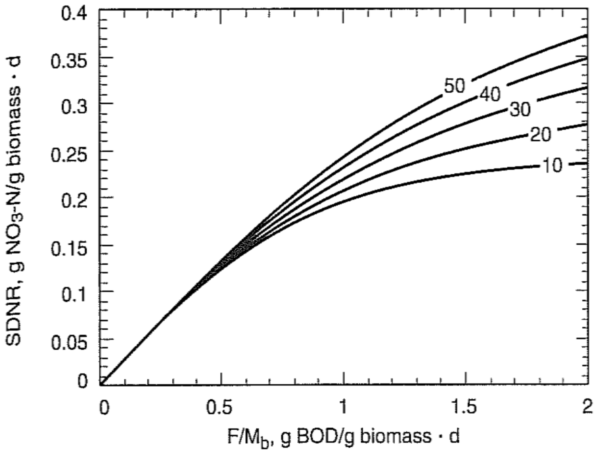

Figure B.1: Plot of spesific denitrification rates ($SDNR_b$) on biomass concentration at 20^0C (Metcalf&Eddy, 1997)

NO_3 reduced, (NO_r);

$$NO_r = V_D * SDNR_{15} * X_b$$

$$NO_r = 4{,}712 \text{ kg/d}$$

Oxygen credit with denitrification, (OUR_{DN});

2.86 = Oxygen equivalent of nitrate

$$OUR_{DN} = 2.86 * Q * (NO_X - NO_{3,e})$$

$$OUR_{DN} = 2.86 * 100{,}000 * (56 - 8)$$

$$OUR_{DN} = 13{,}519 \text{ kg/d}$$

Total Oxygen Requirement, (OR);

$$AOR = (OUR_C - OUR_{DN}) + OUR_N$$

$$AOR = (36{,}834 - 13{,}519) + 23{,}932$$

$$AOR = 47{,}246 \text{ kg } O_2/d$$

P removed by BPR mechanism;

RBCOD/P = assume 10

$$\text{Bio} - \text{P removal} = \frac{S_{VFA,i}}{RBCOD/P}$$

$$\text{Bio} - \text{P removal} = \frac{30}{10}$$

$$\text{Bio} - \text{P removal} = 3\ \text{mgP/l}$$

P used for hetetrophic biomass;

$$P_{X,bio} = \frac{Q * Y * (S_0 - S)}{(1 + k_d * SRT)} + \frac{Q * Y_n * NO_X}{(1 + k_{dn} * SRT)}$$

$$P_{X,bio} = 10{,}470\ \text{kg/d}$$

P used for biomass growth;

$$\text{P used} = 0.015 \frac{\text{gP}}{\text{gbiomass}} * P_{X,bio}$$

$$\text{P used} = 0.015 * 10{,}470$$

$$\text{P used} = 157\ \text{kg/d}$$

$$\text{P used} = 1.57\ \text{mg P/l}$$

Effluent soluble P;

$$\text{P removed} = \text{Bio} - \text{P removal} + \text{P used}$$

$$\text{P removed} = 4.57\ \text{mgP/l}$$

$$\text{Effluent soluble P} = TP_0 - \text{P removed}$$

$$\text{Effluent soluble P} = 10 - 4.57$$

$$\text{Effluent soluble P} = 5.43\ \text{mg P/l}$$

P content of waste sludge;

$$\text{T} - \text{P in sludge} = \text{P removed} * Q/1{,}000$$

$$\text{T} - \text{P in sludge} = 457\ \text{kg/d}$$

Anaerobic volume (V_{AN})

$$r' = \text{selected } 1\text{h}$$

$$V_{AN} = \frac{(Q + Q * R) * r'}{24} \qquad V_{AN} = 7{,}292\ \text{m}^3$$

APPENDIX C (South Africa design method calculations)

Nitrification

Minimum sludge age for nitrification, SRT_m;

Maximum spesific growth rate at 20 ^{0}C, $\mu_{A,m20;}$

$$\mu_{A,m20} = \text{select } 0.45\ d^{-1}$$

Temperature sensitivity for nitrification, θ_n;

$$\theta_n = 1{,}123$$

Maximum spesific growth rate at 15 ^{0}C, $\mu_{A,m15;}$

$$\mu_{A,m15} = \mu_{A,m20} * \theta n^{(T-20)}$$

$$\mu_{A,m15} = 0.252\ d^{-1}$$

Endegenous respiration rate at 20 ^{0}C, b_{A20};

$$b_{A20} = 0.04\ d^{-1}$$

Temperature sensitivity for endogenous respiration, θ_b;

$$\theta_b = 1.029$$

Endegenous respiration rate at 15 ^{0}C, b_{A15};

$$b_{A15} = b_{A20} * \theta_b{}^{(T-20)}$$

$$b_{A15} = 0.035\ d^{-1}$$

(All kinetic coefficients presented in Appendix D)

To calculate SRT_m, S_f is selected 1.3.

$$SRT_m = S_f/(\mu_{Am} - b_A)$$

$$SRT_m = 6.0\ d$$

Total sludge age, SRT;

For the lowest design temperature, S_f is selected between 1.25-1.35 and for the highest temperature S_f is selected between 1.5-2.5.

For 15 ^{0}C, S_f = selected 1.3

For 25 ^{0}C, S_f = selected 2

$SRT = S_f * SRT_m$

SRT for different temperatures;

For 15 ^0C = 7.8 d

For 25 ^0C = 4.4 d

Influent wastewater N concentrations;

Influent $NH_4 - N$ concentration (N_{ai});

$N_{ai} = f_{N,a} * N_{ti}$

$N_{ai} = 0.7 * 70$

$N_{ai} = 49 \text{ mgNH}_4 - N/l$

Unbiodegredable soluble organic nitrogen concentration (N_{ousi});

$N_{ousi} = f_{N,ous} * N_{ti}$

$N_{ousi} = 0.03 * 70$

$N_{ousi} = 2.1 \text{ mgN}/l$

Unbiodegredable particulate organics (N_{oupi});

$N_{oupi} = f_n * f_{s,up} * S_{ti}/f_{cv}$

$N_{oupi} = 6{,}1 \text{ mgN}/l$

Influent biodegredable organic N concentration (N_{obi}), both soluble and particulate which is convertes to ammonia;

$N_{obi} = N_{ti} * (1 - f_{N,a} - f_{N,ous}) - N_{oupi}$

$N_{obi} = 12.8 \text{ mgN}/l$

Nitrification process behaviour;

The unbiodegredable soluble organic nitrogen concentration (N_{ouse}) in the efluent;

$N_{ousi} = N_{ouse}$

$N_{ouse} = 2.1 \text{ mgN}/l$

N concentration of the sludge (N_s);

$$N_S = S_{ti} * f_n * (1 - f_{s,up} - f_{s,us}) * \frac{Y_{Hv}}{1 + b_H * SRT} * (1 + f_h * b_H * SRT) + \frac{f_{s,up}}{f_{cv}}$$

$$N_S = 17.8\ \text{mgN/l}$$

Nitirification capacity, (N_c);

$$N_C = N_{ti} - N_S - N_{te}$$

$$N_C = 70 - 17.6 - 2.2$$

$$N_c = 50.2\ \text{mg N/l}$$

The effluent ammonia concentration, (N_{ae});

$$N_{ae} = \frac{K_n * (b_A + 1/SRT)}{\mu_{Am} * (1 - f_{xt}) - (b_A + 1/SRT)}$$

$$N_{ae} = 0.17\ \text{mgN/l}$$

The effluent TKN concentration, (N_{te});

$$N_{te} = N_{ae} + N_{ouse}$$

$$N_{te} = 0.17 + 2.1$$

$$N_{te} = 2.27\ \text{mgN/l}$$

The oxygen demand for nitrification, (OUR_N);

$$OUR_N = 4.57 * Q_{in} * N_c/1{,}000$$

$$OUR_N = 4.57 * 100{,}000 * 50.2/1{,}000$$

$$OUR_N = 22{,}922\ \text{kgO}_2\text{/d}$$

Denitrification

Denitrification potential, (DN_{POT});

$$DN_{POT} = \left[\frac{N_C}{(R + IR + 1)} + \frac{O_{IR}}{2.86}\right] * IR + \left[\frac{N_C}{(R + IR + 1)} + \frac{O_R}{2.86}\right] * R$$

N_C = Nitrification capacity, mgN/l

R = Return sludge ratio

IR = Internal recycle ratio

O_{IR} = Dissolved oxygen in IR, mg O_2/l

O_R = Dissolved oxygen in R, mg O_2/l

$$DN_{POT} = \left[\frac{50.16}{(0.75 + 4.2 + 1)} + \frac{2}{2.86}\right] * 4.2 + \left[\frac{50.16}{(0.75 + 4.2 + 1)} + \frac{0}{2.86}\right] * 0.75$$

$$DN_{POT} = 44.67\ mg/l$$

IR ratio; $IR_{opt} = \left[-B + \sqrt{B^2 + 4 * A * C}\right] / 2 * A$

where;

$A = O_{IR}/2.86$

$B = N_C - DN_{POT} + [(R + 1) * O_{IR} + R * O_R]$

$$C = (R + 1) * \left(DN_{POT} - R * \frac{O_R}{2.86}\right) - R * N_C$$

$IR_{opt} = 4.2$

Selection of V_D/V_T ratio;

$$\frac{V_D}{V_T} = 1 - S_f * (b_{A15} + \frac{1}{SRT})/\mu_{A,m15}$$

$$\frac{V_D}{V_T} = 0.16$$

Anoxic volume, V_D;

$$V_D = V_T * \frac{V_D}{V_T}$$

$V_D = 95{,}758 * 0.16$

$V_D = 15{,}104\ m^3$

Denitrified nitrate, $S_{NO3,D}$;

$S_{NO3,D} = (TKN_0 - TN_{eff} - N\ \text{concentration the sludge})$

$S_{NO3,D} = (70 - 10.2 - 17.6)$

$S_{NO3,D} = 42.2\ mgN/l$

Oxygen credit with denitrification, $OUR_{DN;}$

$$OUR_{DN} = Q * 2.86 * S_{NO3,D}/1{,}000$$

$$OUR_{DN} = 100{,}000 * 2.86 * 42.2/1{,}000$$

$$OUR_{DN} = 12{,}057\ kgO_2/d$$

Enhanced Biological Phosphorus Removal (EBPR)

Fermentable COD available for conversion into VFAs after denitrification reactor in AN reactor;

$$S_{F,i,conv} = S_{F,i} - \left(8.6 * R * S_{NO3,s} + S_{NO3,i}\right) - 3 * \left(R * S_{O2,s} + S_{O2,i}\right)$$

$$S_{F,i,conv} = S_{F,i} - \text{COD for denitrification} - \text{COD forD.O}$$

$$\text{COD for denitrification} = 0\ \text{mg COD/l}$$

$$\text{COD for dissolved oxygen} = 0\ \text{mg COD/l}$$

$$S_{F,i,conv} = 120\ \text{mgCOD/l}$$

Fermentable COD lost in the effluent of the last anaerobic reactor;

$$S_{F,Ann} = \frac{S_{F,i,conv}}{(1+R)} / \left(1 + (k_{F,T} * (f_{AN} * MX_{OHO}/(N * Q * (1+R))))\right)\hat{}n$$

$$S_{F,Ann} = \frac{120}{1+0.75} / \left(1 + 0.0505 * \left(0.10 * \frac{55{,}597}{3 * 100{,}000 * (1+0.75)}\right)\right)\hat{}3$$

$$S_{F,Ann} = 48{,}8\ \text{mgCOD/l}$$

$$MX_{OHO} = \frac{Y_{OHO}}{(1 + b_{OHO,T} * SRT)} * FCOD_{b\_OHO} * SRT$$

$$MX_{OHO} = \frac{0{,}45}{1 + 0.202 * 7.8} * 45{,}536 * 7.8$$

$$MX_{OHO} = 55{,}597\ \text{kgCOD}$$

VFAs stored by PAOs;

$$FS_{S,PAO} = Q * \left(S_{F,i,conv} - (1+R) * S_{F,Ann}\right) + Q * S_{VFA,i}$$

$$FS_{S,PAO} = 100{,}000 * (120 - (1+0.75) * 48.8) + 100{,}000 * 30$$

$FS_{S,PAO} = 6{,}464\ kgCOD/d$

Remaining biodegredable COD available to OHOs;

$FCOD_{b,OHO} = FCOD_{b,i} - FS_{S,PAO}$

$FCOD_{b,OHO} = 52{,}000 - 6{,}464$

$FCOD_{b,OHO} = 45{,}536\ kgCOD/d$

Biomass equations

PAOs;

$Y_{PAO} = 0.45\ gAVSS/gCOD$

$$Y_{PAO,obs} = \frac{Y_{PAO}}{1 + b_{PAO} * SRT}$$

$Y_{PAO,obs} = 0.45/(1 + 0.034 * 7.8)$

$Y_{PAO,obs} = 0.34\ gAVSS/gCOD$

$MX_{PAO} = Y_{PAO,obs} * FS_{S,PAO} * SRT$

$MX_{PAO} = 0.34 * 6{,}464 * 7.8$

$MX_{PAO} = 17{,}256$ kgVSS in the system

Endogenous mass;

$MX_{E,PAO} = f_{XE,PAO} * b_{PAO} * MX_{PAO} * SRT$

$MX_{E,PAO} = 0.25 * 0.0337 * 17{,}256 * 7.8$

$MX_{E,PAO} = 1{,}342\ kgEVSS$

OHOs (Active mass);

$Y_{OHO} = 0.45\ gAVSS/gCOD$

$$Y_{OHO,obs} = \frac{Y_{OHO}}{1 + b_{OHO} * SRT}$$

$Y_{OHO,obs} = 0.45/(1 + 0.202 * 7.8)$

$Y_{OHO,obs} = 0.16\ gAVSS/gCOD$

$MX_{OHO} = Y_{OHO,obs} * F_{CODb,OHO} * SRT$

$MX_{OHO} = 0.16 * 45{,}536 * 7.8$

$MX_{OHO} = 55{,}597\ \text{kgVSS in the system}$

Endogenous mass;

$MX_{E,OHO} = fX_{E,OHO} * b_{OHO} * MX_{OHO} * SRT$

$MX_{E,OHO} = 0.20 * 0.202 * 55{,}597 * 7.8$

$MX_{E,OHO} = 20{,}757\ \text{kgEVSS}$

Inert mass;

$MX_I = fX_{I,COD,i} * F_{COD,i} * SRT/f_{CV}$

$$MX_I = 0.15 * 60{,}000 * \frac{7.8}{1.48}$$

$MX_I = 47{,}299\ \text{kgIVSS}$

Phosphorus removal;

P release;

$S_{PO4,rel} = f_{PO4,rel} * FS_{S,PAO}/Q$

$$S_{PO4,rel} = 0.5 * \frac{6{,}464}{100{,}000}$$

$S_{PO4,rel} = 32\ \text{mgP/l}$

ΔP by PAOs;

$\Delta P_{PAO} = f_{P,PAO} * MX_{PAO}/(SRT * Q)$

$$\Delta P_{PAO} = 0.38 * \frac{17{,}256}{7.8 * 100{,}000}$$

$\Delta P_{PAO} = 8.43\ \text{mgP/l}$

ΔP by OHOs;

$\Delta P_{OHO} = f_{P,OHO} * MX_{OHO}/(SRT * Q)$

$$\Delta P_{OHO} = 0.03 * \frac{55{,}597}{7.8 * 100{,}000}$$

$$\Delta P_{OHO} = 2.14\ mgP/l$$

ΔP by endegoneus mass;

$$\Delta P_{XE} = \Delta P_{XE,PAO} + \Delta P_{XE,OHO}$$

$$\Delta P_{XE,PAO} = f_{P,XE} * MX_{E,PAO}/(SRT * Q)$$

$$\Delta P_{XE,PAO} = 0.03 * 1{,}342\ /(7.8 * 100{,}000)$$

$$\Delta P_{XE,PAO} = 0.05\ mgP/l$$

$$\Delta P_{XE,OHO} = 0.03 * 20{,}757/(7.8 * 100{,}000)$$

$$\Delta P_{XE,PAO} = 0.80\ mgP/l$$

$$\Delta P_{XE} = 0.85\ mgP/l$$

ΔP by influent inert mass;

$$\Delta P_{XI} = f_{P,XI} * MX_{I}/(SRT * Q)$$

$$\Delta P_{XI} = 0.03 * 47{,}299/(7.8 * 100{,}000)$$

$$\Delta P_{XI} = 1.82\ mgP/l$$

Potential total P removal;

$$\Delta P_{SYS,POT} = \Delta P_{PAO} + \Delta P_{OHO} + \Delta P_{XE} + \Delta P_{XI}$$

$$\Delta P_{SYS,POT} = 8.43 + 2.14 + 0.85 + 1.82$$

$$\Delta P_{SYS,POT} = 13.24\ mgP/l$$

Particulate P in the effluent;

$$X_{P,e} = f_{P,TSS} * TSS_{e}$$

$$X_{P,e} = 4.96\ mgP/l$$

Effluent total P;

$$T_{P,e} = T_{P,i} - \Delta P_{SYS,POT} + X_{P,e}$$

$T_{P,e} = 10.0 - 13.24 + 4.96$

$T_{P,e} = 1.72\ \mathrm{mgP/l}$

VSS and TSS;

VSS and active fraction;

$MX_{bio} = MX_{PAO} + MX_{OHO}$

$MX_{bio} = 17{,}256 + 55{,}597$

$MX_{bio} = 72{,}853\ \mathrm{kgVSS}$

$MX_{VSS} = MX_{PAO} + MX_{OHO} + MX_{E,PAO} + MX_{E,OHO} + MX_{I}$

$MX_{VSS} = 17{,}256 + 55{,}597 + 1{,}342 + 20{,}757 + 47{,}299$

$MX_{VSS} = 142{,}252\ \mathrm{kgVSS}$

$f_{bio,VSS} = MX_{bio}/MX_{VSS}$

$f_{bio,VSS} = \dfrac{72{,}853}{142{,}252}$ $f_{bio,VSS} = 51\%$

FSS;

$MX_{FSS} = f_{FSS,OHO} * MX_{OHO} + f_{FSS,PAO} * MX_{PAO} + f_{X,FSS,i} * SRT$

$MX_{FSS} = 0.15 * 55{,}597 + 1.3 * 17{,}256 + 27{,}000 * 7.8$

$MX_{FSS} = 240{,}781\ \mathrm{kgFSS}$

TSS;

$MX_{TSS} = MX_{VSS} + MX_{FSS}$

$MX_{TSS} = 142{,}252 + 240{,}781$

$MX_{TSS} = 383{,}032\ \mathrm{kgTSS}$

f_{VT};

$f_{VT} = MX_{VSS}/MX_{TSS}$

$f_{VT} = 142{,}252 / 383{,}032$

$f_{VT} = 0.37\ \mathrm{gVSS/gTSS}$

P content of TSS;

$$
\begin{aligned}
f_{P,TSS} = ((f_{P,OHO} * MX_{OHO} + f_{P,XE} * \left(MX_{E,PAO} + MX_{E,OHO}\right) + f_{P,XI} * MX_{I}) / f_{VT} \\
+ (f_{P,PAO} * MX_{PAO}) / f_{VT,PAO} * (f_{P,FSS,i} * MX_{FSS}) / MX_{TSS}
\end{aligned}
$$

$f_{P,TSS} = 0.08\ gP/gTSS$

Process volume;

$X_{TSS} = \text{select } 4{,}000\ mgTSS/l$

$V_{P} = MX_{TSS} / X_{TSS}$

$V_{P} = 383{,}032\ /4{,}000$

$V_{P} = 95{,}758\ m^{3}$

The volume of anaerobic zone;

$V_{P,AN} = f_{AN} * V_{P}$

$V_{P,AN} = 0.1 * 95{,}758$

$V_{P,AN} = 9{,}576\ m^{3}$

Nitrogen requirement;

$FN_{synth} = f_{N,VSS} * MX_{VSS} / SRT$

$FN_{synth} = 0.10 * 142{,}252\ \ /7.8$

$FN_{synth} = 1{,}829\ kg/d$

$TKN_{i,synth} = FN_{synth} / Q$

$TKN_{i,synth} = 1{,}829\ /100.000$

$TKN_{i,synth} = 18\ mgN/l$

Oxygen demand by PAOs for synthesis and endogenous respiration;

$FO_{2,PAO} = FO_{2,PAO,synth} + FO_{2,PAO,endo}$

$FO_{2,PAO,synth} = FS_{S,PAO} * (1 - f_{CV} * Y_{PAO})$

$FO_{2,PAO,synth} = 6{,}464 * (1 - 1.48 * 0.45)$

$$FO_{2,PAO,synth} = 2{,}159\ kgO_2/d$$

$$FO_{2,PAO,endo} = FS_{S,PAO} * f_{CV} * \left(1 - f_{XE,PAO}\right) * b_{PAO} * Y_{PAO,obs} * SRT$$

$$FO_{2,PAO,endo} = 6{,}464 * 1.48 * (1 - 0.25) * 0.0337 * 0.33 * 7.8$$

$$FO_{2,PAO,endo} = 766\ kgO_2/d$$

$$FO_{2,PAO} = 2{,}925\ kgO_2/d$$

Oxygen Demand by OHOs for synthesis and endogenous respiration;

$$FO_{2,OHO} = FO_{2,OHO,synth} + FO_{2,OHO,endo}$$

$$FO_{2,OHO,synth} = FCOD_{b,OHO} * (1 - f_{CV} * Y_{OHO})$$

$$FO_{2,OHO,synth} = 45{,}536 * (1 - 1.48 * 0.45)$$

$$FO_{2,OHO,synth} = 15{,}209\ kgO_2/d$$

$$FO_{2,OHO,endo} = FCOD_{b,OHO} * f_{CV} * \left(1 - fX_{E,OHO}\right) * b_{OHO} * Y_{OHO,obs} * SRT$$

$$FO_{2,OHO,endo} = 45{,}536 * 1.48 * (1 - 0.20) * 0.202 * 0.14 * 7.8$$

$$FO_{2,OHO,endo} = 15{,}798\ kgO_2/d$$

$$FO_{2,OHO} = 15{,}209 \quad + 15{,}798$$

$$FO_{2,OHO} = 31{,}007\ kgO_2/d$$

O.D total (carbonaceous);

$$OUR_C = FO_{2,PAO} + FO_{2,OHO}$$

$$OUR_C = 2{,}925 + 31{,}007$$

$$OUR_C = 33{,}933\ kgO_2/d$$

Totall Oxygen Reqirement, (OR);

$$OR = (OUR_C - OUR_{DN}) + OUR_N$$

$$OR = (34{,}144 - 12{,}057) + 22{,}922$$

$$OR = 44{,}797\ kg\ O_2/d$$

APPENDIX D (Kinetic coefficients and assumptions used in design methods)

Table D.1: Assumptions used in ATV-131 design method

Considerations	**Symbol**	**Unit**	**Value**
Return sludge ratio	R		0.75
Anaerobic contact time	t	h	1
Sludge volume index	SVI	ml/g	100
Thickening time	t_{Th}	h	2.3
Dilution factor	f_{obs}		0.7

Table D.2: Kinetic coefficients used in WERF design method

Considerations	**Symbol**	**Unit**	**Value**
Max.specific growth rate of nitrifying bacteria	$\mu_{n,m}$	d-1	0.75
Half velocity constant	K_S	mg bCOD/l	20
Biomass yield	Y	g VSS/g bCOD	0.4
Endogenous decay coefficient	k_d	d-1	0.088
	f_d	g/g	0.15
Net biomass yield	Y_n	g VSS/g bCOD	0.12
Half saturation coefficient for DO	K_0	mg/l	0.5

Table D.3: Assumptions used in WERF design method

Considerations	**Symbol**	**Unit**	**Value**
Return sludge ratio	R		0.75
Anaerobic contact time	t	h	1
P content of hetetrophic biomass		g P/g biomass	0.015
RBCOD/P		g COD/g P	10
RBCOD/NO3		g COD/g N	6.6

Table D.4: Assumptions used in UCT design method

Description	Symbol	Unit	Value
Number of anaerobic zones	n	Reactors	3
Anaerobic mass fraction	f_{AN}	g VSS/g VSS	0.1
Sludge recycle	R		0.75
Dissolved oxygen in the sludge recycle	$S_{O2,s}$	g O_2/m^3	0
Nitrate concentration in the sludge recycle	$S_{NO3,s}$	g NO_3/m^3	0

Table D.5: Kinetic coefficients used in UCT design method

Description	Symbol	Unit	Value
First order fermentation rate constant	$k_{F,t}$	0.0505	m^3/g VSS.d
Anaerobic volume fraction	f_{AN}	0.10	-
Spesific end. mass loss rate of the OHO	$b_{OHO,t}$	0.24	d^{-1}
PAO spesific end. mass loss rate	$b_{PAO,t}$	0.04	d^{-1}
Biomass yield of PAO	Y_{PAO}	0.45	g VSS/g COD
Biomass yield of OHO	Y_{OHO}	0.45	g VSS/g COD
Fraction of endogenous residue of the OHOs	$f_{XE,OHO}$	0.20	g EVSS/g AVSS
Fraction of particulate inert COD	$f_{XI,COD,i}$	0.15	g COD/g VSS
Nitrogen content of active biomass	$f_{N,VSS}$	0.10	g N/g VSS
COD/VSS ratio of the sludge	f_{CV}	1.48	g COD/g VSS
Ratio of P release/VFA uptake	$f_{PO4,rel}$	0.50	g P/g COD
Fraction of P in the active PAO mass	$f_{P,PAO}$	0.38	g P/g VSS
Fraction of P in the active OHO mass	$f_{P,OHO}$	0.03	g P/g VSS
Fraction of P in the endogenous mass	$f_{P,XE}$	0.03	g P/g EVSS
Fraction of P in the inert mass	$f_{P,XI}$	0.03	g P/g IVSS
Fraction of P in the total suspended solids	$f_{P,TSS}$	0.124	g P/g TSS
Fraction of fixed suspended solids of OHOs	$f_{FSS,OHO}$	0.15	g FSS/g AVSS
Fraction of fixed suspended solids of PAOs	$f_{FSS,PAO}$	1.30	g FSS/g AVSS
Fraction of endogenous residue of the PAOs	$f_{XE,PAO}$	0.25	g EVSS/g AVSS
P fraction of influent FSS	$f_{P,FSS,i}$	0.02	g P/g VSS

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