Development of a Dynamic Simulation Model for EqualizationTanks

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Dissertation submitted in partial fulfilment of the requirements for the degree of Master in Science: Civil Engineering

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Abstract

The influent to a water and resource recovery facility (WRRF) generally exhibits significant diurnal variations in flow rate and load concentration. This makes determining the operating parameters and subsequently the overall operation of plants difficult, especially in developing countries due to the lack of highly skilled operators. Hence, there is an incentive for the control and operation of WRRFs in developing countries to be improved.

Flow equalization tanks were identified as a potential method to attenuate the diurnal variations in flow rate and load concentration into plants. The main aim of this research was to develop a viable dynamic simulation model for the operation of flow equalization tanks, within a plant-wide framework (to allow for the evaluation of design and control strategies). The next aim was to determine the benefits of equalization tanks towards design and optimised operation of future WRRFs via scenario analyses. Finally, the effects of the equalization tank on the performance of various unit processes in a WRRF were to be investigated.

The model was developed in three steps; i) the development of the required equations to model equalization tanks, ii) mass balance throughout the model for internal consistency and iii) scenario analyses to determine if the model generated reasonable and scientifically sound outputs. The model was developed using Microsoft Excel Visual Basic (VBA) and WEST®. Two scenarios were considered to assess the equalization tank modelled. Scenario One involved the comparison of the capital cost, unit process sizes and total footprint of a balanced sludge age Modified Ludzack-Ettinger (MLE) system with and without an equalization tank. Scenario Two compared the plant performance of the MLE system designed in Scenario One with and without a flow equalization tank.

A dynamic simulation model replicating equalization tanks was successfully developed. From scenario analyses, it was determined that using an MLE system and only considering equalization of flow, there was a reduction in the size of several unit processes by 8-9% (primary settling tank, biological reactors, secondary settling tank, flotation unit, anoxic-aerobic digester), due to the less conservative design values that could be used as the variations of the influent were decreased. Despite this, a 13% overall increase in the footprint of the WRRF was observed due to the addition of the equalization tank. The attenuation of diurnal flow variations also resulted in reduction of various plant parameters by up to 50% (flow, OUR, VSS flux). Finally, there was a 10% improvement in the performance of various unit processes due to the presence of the equalization tank.

In conclusion, the inclusion of equalization tanks in WRRFs has significant positive effects. These results were obtained with equalization of flow only. Some other limitations were experienced during the project resulting in the following recommendations: further research will be needed to validate and calibrate the model, As the model was not successfully incorporated in a plant-wide framework, further developments in that direction are required, as well as including the equalization of load in the model.

Key words: equalization tank, water resource and recovery facility, Modified Ludzack-Ettinger system, dynamic model.

Acknowledgements

GOD – For giving me the strength to complete this project.

Dr David Ikumi – For guiding me and giving me sufficient support to learn throughout this project and complete it.

Professor George Ekama – For assisting me and supporting me throughout this project.

Chris Brouckaert – For assisting me in the modelling in Microsoft Excel and WEST®.

My parents; Mr & Mrs Fotso and siblings – For your continuous love, and support in all forms throughout this project. Without you this would not be possible.

Mastercard Foundation Scholars Program – For the immense financial support and development throughout the course of this project.

The Civil Engineering Department at the University of Cape Town – For making me feel at home.

My friends – For the constant support throughout this project.

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Acronyms and symbols

Abbreviation	**Description**
AAD	Anoxic Aerobic Digestion
AD	Anaerobic digestion
ADWF	Average dry weather flow
AS	Activated sludge
BOD	Biological oxygen demand
BSM2	Benchmark Simulation Model 2
COD	Chemical oxygen demand
DO	Dissolved oxygen
DWS	Department of Water and Sanitation
FSA	Free and saline ammonia
ISS	Inorganic suspended solid
MLE	Modified Ludzack-Ettinger
OHO	Ordinary heterotrophic organisms
OUR	Oxygen utilization rate
PAO	Phosphorus accumulating organisms
PDWF	Peak dry weather flow
PS	Primary sludge
PST	Primary settling tank
PWM_SA	Plant-wide dynamic model of South Africa
PWWF	Peak wet weather flow
S.A.	South Africa
SOUR	Specific oxygen utilisation rate
SRT	Sludge retention time
SST	Secondary settling tank
TKN	Total Kjeldahl Nitrogen
UCT	University of Cape Town
VFA	Volatile fatty acid
VSS	Volatile suspended solids
WAS	Waste activated sludge
WRRF	Water and resource recovery facilities
WWTP	Wastewater treatment plant

Symbol	Description	Unit
$\boldsymbol{\alpha}$	A weighting factor (for load and flow error)	Unitless
$\boldsymbol{\beta}$	A weighting factor (for volumetric error)	Unitless
$\boldsymbol{\delta v}$	Volume differential within the limits at which the volumetric error attains a value (%)	Unitless
$\boldsymbol{\omega}$	A weighting factor (for rate of change of flow rate error)	Unitless
Δ	Change applied at each interval	Ml/d
$\mathbf{\Delta P_{OHO}}$	P removal due to OHOs	gP/m^3
$\mathbf{\Delta P_{PAO}}$	P removal due to PAOs	gP/m^3
$\mathbf{\Delta P_{SYS,\ ACT}}$	Actual total P removal for the system	gP/m^3
$\mathbf{\Delta P_{SYS,\ POT}}$	Potential total P removal by the system	gP/m^3
$\mathbf{\Delta P_{XE}}$	P removal due to endogenous residue mass	gP/m^3
$\mathbf{\Delta P_{XI}}$	P removal due to influent inert mass	gP/m^3
$\mathbf{\Delta t}$	Time taken for an interval to elapse	hr
$\boldsymbol{\mu}_{\mathbf{AmT}}$	Maximum specific growth rate constant	/d
$\mathbf{b_{AD}}$	Acidogen endogenous respiration rate	/d
$\mathbf{b_{AT}}$	Specific endogenous respiration rate for nitrifiers at T^{o}C	/d
$\mathbf{b_{H}}$	Specific rate of endogenous mass loss of OHOs	/d
$\mathbf{b_{OHO,T}}$	OHO specific endogenous mass loss rate constant at temperature T^{o}C	gEVSS/gVSS.d
$\mathbf{b_{PAO,T}}$	PAO specific endogenous mass loss rate constant at temperature T^{o}C	gEVSS/gVSS.d
$\mathbf{C_0}$	Influent concentration into balancing tank	mg/l
$\mathbf{C_1}$	Balancing tank concentration	mg/l
$\mathbf{C_t}$	Balancing tank concentration at time t	mg/l
$\mathbf{C_{t-1}}$	Balancing tank concentration at time t-1	mg/l
$\mathbf{d}$	Time taken by one interval	hr
$\mathbf{D_{p1}}$	Total denitrification potential in the primary anoxic reactor	mgN/l
$\mathbf{D_{p1RBCOD}}$	Denitrification potential of the influent RBCOD in the primary anoxic reactor	mgN/l
$\mathbf{D_{p1SBCOD}}$	Denitrification potential of the influent SBCOD in the primary anoxic reactor	mgN/l
$\mathbf{D_{p3}}$	Total denitrification potential of the secondary anoxic reactor	mgN/l
$\mathbf{D_{p3SBCOD}}$	Denitrification potential of SBCOD in the secondary anoxic reactor	
$\boldsymbol{\frac{dv}{dt}}$	Rate of change of tank volume	Ml/d
$\mathbf{E}$	Proportion of biodegradable particulate organics utilized conserved as biomass	Sludge produced/COD utilized
$\mathbf{E_e}$	Equalization error	Unitless
$\mathbf{E_f}$	Flow error	Unitless

$\mathbf{E_{ld}}$	Load error	Unitless
$\mathbf{E_{lm}}$	Penalty error for tank volume limit	Unitless
$\mathbf{E_s}$	Penalty error for rate of change of tank outflow rate	Unitless
$\mathbf{E_t}$	Total error expression	Unitless
$\mathbf{F}$	Mean influent flow rate from (t-1) to t	m^3/d
$\mathbf{F_i}$	Influent flow rate into balancing tank	m^3/d
$\mathbf{F_o}$	Effluent flow rate out of balancing tank	m^3/d
$\mathbf{f_{AD}}$	Acidogen biomass unbiodegradable fraction	Unitless
$\mathbf{f_{AN}}$	Anaerobic mass fraction	gVSS/gVSS
$\mathbf{f_{cv}}$	COD to VSS ratio of the unbiodegradable particulate organics	mgCOD/mgVSS
$\mathbf{f_{FSS,OHO}}$	Fraction of fixed (inorganic) suspended solids of OHOs	gFSS/gAVSS
$\mathbf{f_{FSS,PAO}}$	Fraction of fixed (inorganic) suspended solids of PAOs	gFSS/gAVSS
$\mathbf{f_H}$	Unbiodegrable fraction of the OHOs	mgCOD/mgCOD
$\mathbf{f_{iOHO}}$	Inorganic content of OHOs	mgISS/mgCOD
$\mathbf{f_n}$	Nitrogen content of VSS	mgN/mgVSS
$\mathbf{FO_{2,OHO}}$	Daily mass of oxygen consumed by OHOs	gO_2/d
$\mathbf{FO_{2,PAO}}$	Daily mass of oxygen consumed by PAOs	gO_2/d
$\mathbf{FO_C}$	Daily carbonaceous oxygen demand	gO_2/d
$\mathbf{FO_D}$	COD flux in denitrification	kgCOD/d
$\mathbf{FO_n}$	Daily mass of oxygen required for nitrification	kgO_2/d
$\mathbf{FO_T}$	Total daily mass of oxygen required	kgO_2/d
$\mathbf{f_{P,FSS}}$	P fraction of fixed (inorganic) suspended solids mass	gP/gFSS
$\mathbf{f_{P,OHO}}$	Fraction of OHO active mass that is P	gP/gAVSS
$\mathbf{f_{P,PAO}}$	Fraction of PAO active mass that is P	gP/gAVSS
$\mathbf{f_{P,TSS}}$	P fraction of total suspended solids mass	gP/gTSS
$\mathbf{f_{P,XE}}$	Fraction of endogenous residue mass that is P	gP/gEVSS
$\mathbf{f_{PS'up}}$	Unbiodegradable particulate fraction of the primary sludge	mgCOD/mgCOD
$\mathbf{f_{S'bs}}$	Fraction of total COD that is readily biodegradable	mgCOD/mgCOD
$\mathbf{f_{s'up}}$	Fraction of total COD which is unbiodegradable particulate	mgCOD/mgCOD
$\mathbf{f_{S'us}}$	Fraction of total COD which is unbiodegradable soluble	mgCOD/mgCOD
$\mathbf{FS_{bi}}$	Daily flux of influent biodegradable COD	mgCOD/d
$\mathbf{FS_{b,OHO}}$	Daily mass of biodegradable substrate available to OHOs	gCOD/d
$\mathbf{FS_{bs,PAO}}$	Daily mass of substrate stored by PAOs in the anaerobic reactor	gCOD/d
$\mathbf{FS_{ti}}$	Daily mass of influent COD	gCOD/d
$\mathbf{f_{x1}}$	Primary anoxic sludge mass fraction	Unitless
$\mathbf{f_{x3}}$	Secondary anoxic sludge mass fraction	Unitless
$\mathbf{f_{XE,OHO}}$	Fraction of endogenous particulate residue of OHOs	gEVSS/gAVSS
$\mathbf{f_{XE,PAO}}$	Fraction of endogenous particulate residue of PAOs	gEVSS/gAVSS
$\mathbf{FX_{FSS,i}}$	Daily mass of influent FSS	gFSS/d

$\mathbf{f_{XI,COD,i}}$	Fraction of influent COD that is particulate inert	gCOD/gCOD
$\mathbf{FX_{Ii}}$	Daily flux of influent particulate unbiodegradable COD	mgCOD/d
$\mathbf{FX_{IOi}}$	Daily flux of influent particulate unbiodegradable matter	mgISS/d
$\mathbf{FX_{t}}$	Daily flux of total solids produced	mgTSS/d
$\mathbf{K_2}$	Second specific rate of denitrification in primary anoxic reactor	$mgNO_3^-$ N/mgOHOVSS.d
$\mathbf{K_3}$	Specific denitrification rate in secondary anoxic reactor of NDBEPR system on SBCOD at temperature T	$mgNO_3^-$ N/mgOHOVSS.d
$\mathbf{K_m}$	Kinetic constant	gCOD/gCOD biomass.d
$\mathbf{K_{nT}}$	Half saturation constant for nitrifiers at T°C	mgN/l
$\mathbf{K_s}$	Kinetic constant	gCOD/l
$\mathbf{L_i}$	Influent load	kg/d
$\mathbf{MX_{BH}}$	Mass of OHO in the bioreactor	mgVSS
$\mathbf{MX_{E,OHO}}$	Mass of endogenous residue in the system	gEVSS
$\mathbf{MX_{E,PAO}}$	Mass of PAO endogenous residue	gEVSS
$\mathbf{MX_{EH}}$	Mass of endogenous residue in the bioreactor	mgVSS
$\mathbf{MX_{FSS}}$	Mass of fixed suspended solids in the system	gFSS
$\mathbf{MX_{I}}$	Mass of inert organic matter in the system coming from the influent	gIVSS
$\mathbf{MX_{OHO}}$	Mass of OHOs in the whole NDBEPR system	gAVSS
$\mathbf{MX_{PAO}}$	Biological active mass of PAO	gAVSS
$\mathbf{MX_{TSS}}$	Mass of total suspended solids in the system	gTSS
$\mathbf{MX_{v}}$	Mass of volatile suspended solids in the bioreactor	mgVSS
$\mathbf{N_{ae}}$	Effluent ammonia concentration	mgN/l
$\mathbf{N_{ai}}$	Influent ammonia concentration	mgN/l
$\mathbf{N_{c}}$	Nitrification capacity of the reactor	mgN/l
$\mathbf{N_{ne}}$	Effluent nitrate concentration	mgN/l
$\mathbf{N_{obpi}}$	Influent biodegradable particulate organic nitrogen concentration	mgN/l
$\mathbf{N_{obsi}}$	Influent biodegradable soluble organic nitrogen concentration	mgN/l
$\mathbf{N_{oupi}}$	Influent unbiodegradable particulate organic nitrogen concentration	mgN/l
$\mathbf{N_{ouse}}$	Unbiodegradable soluble organic N concentration in the effluent	mgN/l
$\mathbf{N_{s}}$	Concentration of TKN in influent required for sludge production	mgN/l
$\mathbf{N_{te}}$	Effluent TKN concentration	mgN/l
$\mathbf{p_{CO2}}$	Partial pressure of CO_2 in the gas phase (mol fraction)	Unitless
pH	Negative $\log_{10}$ of the (H^+) activity	Unitless
$\mathbf{pH_{i}}$	Influent pH	Unitless
pK'a	Negative $\log_{10}$ of acid dissociation constant	Unitless

pK'_{C1}	Negative $\log_{10}$ of 1st carbonate system apparent dissociation constant	Unitless
pK'_{C2}	Negative $\log_{10}$ of 2nd carbonate system apparent dissociation constant	Unitless
$pK'HCO_2$	Negative $\log_{10}$ of the apparent Henry's law constant	Unitless
Q	Flow into AD	m^3/d
Q_i	Influent flow rate	l/d
Q_{in}	Influent flow to the balancing tank	m^3/d
Q_e	Effluent flow from the balancing tank	m^3/d
Q_w	Waste flow rate from the reactor	l/d
Q_m	Methane gas flow per litre influent	l methane/d/l influent flow/d
R	Retention time of the sludge in the AD	d
r_h	Monod hydrolysis rate	gCOD/l.d
R_s	Sludge age in the bioreactor	d
R_{sm}	Minimum sludge age for nitrification	d
s	Sludge recycle ratio based on influent flow	$m^3.d/m^3.d$
S_{bi}	Influent biodegradable COD concentration	$gCOD/m^3$
S_{bp}	Residual biodegradable COD concetration	gCOD/l
$S_{bp,hydrolysed}$	Hydrolysed biodegradable particulate COD in the AD	gCOD/l
S_{bpi}	Hydrolysable COD concentration from the influent	gCOD/l
S_{bsAci}	Dissociated acetate species concentration in the influent	gCOD/l
S_{bsHAci}	Undissociated acetate species concentration in the influent	gCOD/l
S_{bsai}	VFA concentration in the influent	mgCOD/l
$S_{bsf,ANn}$	Concentration of fermentable COD in the effluent of the nth anaerobic reactor	gCOD/l
S_{bsfi}	Fermentable readily biodegradable COD concentration in the influent	mgCOD/l
$S_{bsf,i,conv}$	Fermentable COD available for conversion per volume of influent	mgCOD/l
S_f	Safety factor	Unitless
S_m	Effluent methane concentration	gCOD/l
$S_{NO3,i}$	Nitrate concentration in the influent to anaerobic reactor	$gNO_3\text{-}N/m^3$
$S_{NO3,s}$	Nitrate concentration in the sludge recycle to anaerobic reactor	gNO_3^-N/m^3
$S_{O2,i}$	Oxygen concentration in the influent to anaerobic reactor	gO_2/m^3
$S_{O2,s}$	Oxygen concentration in the sludge recycle to the anaerobic reactor	gO_2/m^3
S_{te}	Effluent total COD concentration	$gCOD/m^3$
S_{ti}	Influent concentration of total COD	$gCOD/m^3$
S_{upi}	Unbiodegradable COD concentration in the influent	$gCOD/m^3$
t	Time step	d

$\mathbf{T_i}$	Total number of intervals	Unitless
$\mathbf{T_{P,e}}$	Effluent total phosphorus concentration	gP/m^3
$\mathbf{T_{p,i}}$	Influent total phosphorus concentration	gP/m^3
$\mathbf{TSS_e}$	Total suspended solids concentration of the effluent	$gTSS/m^3$
V	Volume	m^3
$\mathbf{V_{lb}}$	Lower normal tank volume limit (% of tank volume)	Unitless
$\mathbf{V_{lu}}$	Upper normal tank volume limit (% of tank volume)	Unitless
$\mathbf{V_p}$	Equalization tank hold-up (% of tank volume)	Unitless
$\mathbf{V_t}$	Volume of balancing tank at time t	m^3
$\mathbf{V_{t-1}}$	Volume of balancing tank at time t-1	m^3
$\mathbf{X_{EH}}$	Endogenous residue concentration	mgVSS/l
$\mathbf{X_I}$	Unbiodegradable particulate organic volatile suspended solid concentration	mgVSS/l
$\mathbf{X_{Ii}}$	Unbiodegradable particulate organics concentration in the influent expressed as VSS	mgVSS/l
$\mathbf{X_{IOi}}$	Influent inorganics concentration	mgISS/l
$\mathbf{X_V}$	Volatile suspended solid concentration	mgVSS/l
$\mathbf{Y_{AD}}$	Pseudo-acidogen yield coefficient	gCOD biomass/gCOD organics hydrolysed
$\mathbf{Y_H}$	COD yield of OHOs	mgCOD/mgCOD
$\mathbf{Y_{OHO}}$	Biomass yield of OHOs	gAVSS/gCOD
$\mathbf{Y_{PAO}}$	Biomass yield of PAOs	gAVSS/gCOD
$\mathbf{Z_{AD}}$	Concentration of acidogen biomass	gCOD/l
$\mathbf{Z_E}$	Acidogen endogenous residue concentration	gCOD/l

1. Introduction

This project focuses on the development and assessment of a dynamic simulation model for equalization tanks. This is a model that predicts the required outflows from an equalization tank of a specific volume, to reduce variations from the input of the tank and hence improve waste water treatment plant (WWTP) operations.

1.1 Background and motivation to the research

The conventional arrangement of water and resource recovery facilities (WRRFs) is such that the influent coming from the sewer network is discharged into its inlet headworks and pumped to the primary settling tank (PST), whereby settleable material (i.e., the PST underflow; primary sludge) is collected and sent for treatment to an anaerobic digestion (AD) system, and the suspended and dissolved material (i.e., the PST overflow; settled wastewater) is pumped to the biological reactors of an activated sludge (AS) system (see Figure 1).

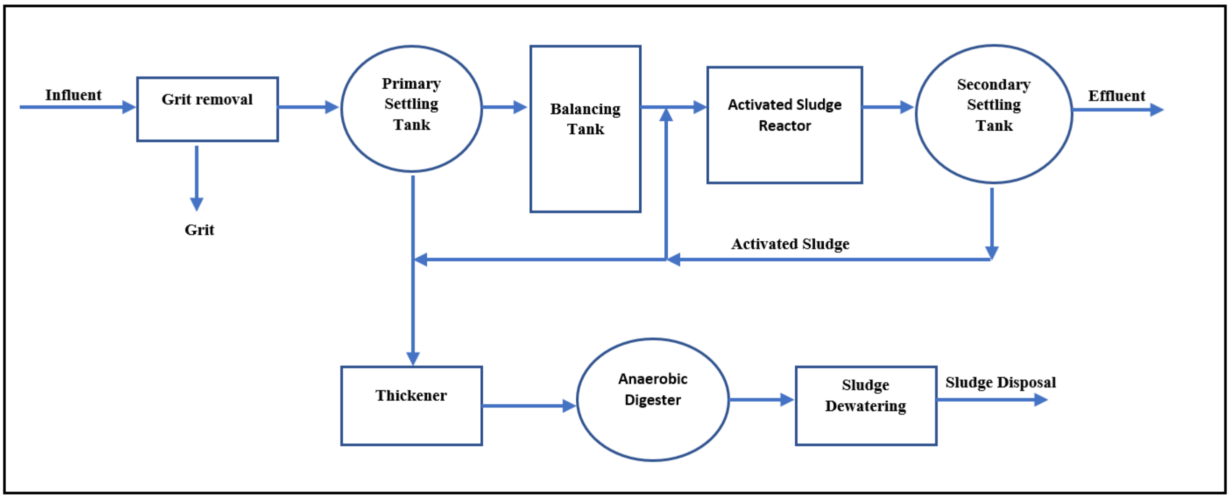

Figure 1: Schematic of a WWTP with a balancing tank

In these AS reactors, it is crucial to know the amount of dissolved oxygen required by the various microorganisms mediating the biological treatment process as well as the number of days required for this biomass to complete the treatment process. This amount of time dictates the period required to keep the solid material (including the treatment mediating biomass) in the system (i.e., the solid retention time or sludge age) and is controlled by the quantity of sludge washed-out from the reactors every day. The amount of time that the liquids remain in the system (hydraulic retention time) is usually shorter due to the final treatment process of secondary settling tanks (SST) that separate reactor output solids from the aqueous phase to create the treated effluent (SST overflow) while the solids (SST underflow, containing the biomass) are recycled to the reactors and only exit via the daily waste flow (Ekama & Wentzel, 2008).

To be able to determine the operating parameters of the plant such as the amount of dissolved oxygen to be provided or the sludge age, the waste coming into the reactor must be characterized, and this involves numerous measurements and assumptions. This characterization is a crucial step for the adequate performance and operation of the plant and is sensitive to variations in the influent flow and concentrations (Ekama & Wentzel, 2008). Dold (1982) points out that the influent to a wastewater treatment plant generally exhibits wide diurnal variations in both flow rate and concentration, and consequently in the loading rate (the product of flow and concentrations).

Dold (1982) realised that in South Africa (S.A.) (and various other developing countries), in-plant control is not suited to nutrient removal processes due to the lack of technological advancements (which include accurate and reliable monitoring equipment) and worker expertise required for complex in-plant control strategies.

Since then, Dold (1982) successfully created a control strategy to enable easier operations at WWTPs, but this strategy uses outdated computer programs.

With the advent of plant-wide models and the paradigm shift of converting wastewater treatment plants to WRRFs, there is an incentive for the control and operation of the WRRFs in developing countries to be improved. This research project contributes towards such progress through the creation of a viable computer model for equalization tanks, within a plant-wide framework, to allow for the evaluation of design and control strategies that could be realized with the addition of this unit process to the WRRF configuration.

1.2 Scope and limitations of research

This project resulted in the development of the primary phase of the dynamic simulation model. Hence, the model provided the basic requirements, i.e. flow equalization. Other considerations will be included sequentially in future projects, to ensure the full scope of the model is achieved. Due to this being the primary phase of development, the scope and limitations of this research were as follows;

- No calculations on the hydraulic performance of the tank and how the outflow was achieved were included. These calculations will only be deemed important once a valid model has been developed.
- Only the equalization of the influent wastewater flow was investigated, not the equalization of the loading rate.
- Only in-line equalization was considered as this was considered better than side-line equalization from preliminary results of Dold (1982).
- The tank was considered to be completely mixed with no biological process happening as well as no evaporation. This is because investigation on the biological process will be assessed in future projects.
- The model was developed in Microsoft Excel (with Visual Basics) and WEST® (Van Hooren *et al.*, 2003). This was to provide the model in a more recent and widely used software package.
- The model balanced the flows if a tank volume was provided, it did not predict the best tank size.

- The assessment of the equalization tank model was done in steady state and dynamic state (in Microsoft Excel and WEST® respectively) with a Modified Ludzack-Ettinger (MLE) configuration selected as the connected AS system.

1.3 Research problem, questions and hypotheses

1.3.1 Research problem

Diurnal variations in influent causes problems in the AS performance as well as the operation of the biological treatment process (Mikola, 2013). When considering performance, the varying loads and flows can cause "overloading" of the reactive capacity of the organism mass or the physical design provisions. Such overloading often results in poor nitrification, poor effluent quality standards, overflowing in the SST and inadequate oxygenation during peak load periods; hence, poor sludge settling properties (Dold,1982). These variations equally increase the capital cost of WWTPs as the unit processes have to be designed using the peak flow rate (Dold, 1982).

There is no commonly accepted method for assessing the efficiency of the flow equalization (Mikola, 2013), due to the wrong conclusions put out by researchers on the effects of an equalization tank (Dold, 1982). From this, it can be inferred that the low presence of equalization tanks at different WWTPs can be due to difficulty in their operation, in-plant control strategies (Dold,1982) and a lack of understanding of the kinetic behaviour of the AS process; hence, the impact of flow equalisation on overall system performance (Dold, 1982).

The purpose of this research was to develop a dynamic simulation model of equalization tanks using Dold's (1982) approach as a guide. This model was to be incorporated in the University of Cape Town's (UCT) Plant-Wide Dynamic Model (PWM_SA) system, with outflows being predicted and the effects of these balanced outflows on other unit processes and on the general performance of the WWTP determined. Hence, this can be used as significant motivation for the inclusion of equalization tanks in S.A. and other countries at a later stage. To fulfil this, data from benchmark simulation models (Nopens *et al.*, 2010), was used to develop the model and test its robustness.

1.3.2 Research questions and hypotheses

The research questions and hypotheses are as follows;

1- Can a complete dynamic simulation model replicating equalization tanks be developed?
    - Hypothesis: A dynamic simulation model replicating balancing tanks to a satisfactory level can be developed (Dold, 1982).
2- Can the Benchmark Simulation Model 2 (BSM2) be used to satisfactorily calibrate and validate the model?
    - Hypothesis: The BSM2 can be used to satisfactorily calibrate and validate the model (Nopens *et al.*, 2010).
3- Can significant benefits of balancing tanks for WRRFs be identified via the model?

  - Hypothesis: Significant benefits of balancing tanks for WRRFs can be identified using the model (Dold, 1982).

4- Can the model be made compatible with the existing UCT plant wide model?

  - Hypothesis: The complete dynamic simulation model can be made compatible to the existing UCT plant wide model as it is based on similar concepts (Ikumi, 2011).

1.4 Research objectives

The overarching aim of this research was the development of a dynamic simulation model for equalization tanks. The sub-objectives were as follows;

- Develop a complete dynamic simulation model for equalization tanks based on the investigations by Dold (1982).
- Calibrate and validate the model using data from the International Water Association (IWA) BSM2 with 609 days of data (Nopens *et al.*, 2010).
- Make the model compatible with the existing UCT plant wide model (i.e., PWM_SA; Ikumi *et al.*, 2014).
- Determine the benefits of equalization tanks towards design and optimised operation of future WRRFs via scenario analysis. By simulating a scenario that exhibits the extent to which the inclusion of an equalization tank in a system wide setup can be useful.
- Depict the effects of a functional equalization tank on a WWTP. The effects on the operation of unit processes downstream and the effects on the performance of the WWTP.

1.5 Overview of thesis project

The Introduction section of the thesis starts with a background and motivation to the research, providing an argument as to why the research is relevant. It is followed by the scope and limitations of the research which provide the delimitations which confined this research project. The research problem is then developed and supported by research questions and hypotheses to guide the research. This is followed by the research objectives which highlight what the research is expected to achieve.

The Literature Review section discusses on research on the most prevalent type of treatment technologies in S.A. and the unit processes that consume the most electricity. It continues with an overview of WWTP biological processes, AS and anaerobic digestion (AD) stoichiometry and the stoichiometric link of equalization tanks to AS and AD. The section further elucidates on what the UCT plant wide model is and why it is important to have an equalization tank model that fits in this plant wide model. Then it points out research on the reasons to use equalization tanks and the various types of equalization tanks. It ends on the equalization algorithm developed by Dold (1982).

The Model Development section subsequently describes the platforms used to develop the model and the steps followed to develop the model.

The Results and Discussion section reports on the performance of the equalization tank and includes two scenarios which asses the equalization tank. The first is a steady state scenario whereby an MLE system is designed with and without the inclusion of an equalization tank and the unit process sizes and the capital costs are compared. The second scenario involves a dynamic simulation of the plant designed in the first scenario and the performance of various key unit processes in the plant are assessed and compared (with and without an equalization tank).

The thesis ends with a Conclusion section which summarises the findings obtained and possible further projects on equalization tank models.

2. Literature review

Due to the recent increase in demand and the lack of electricity generating infrastructure, there has been a rapid increase in electricity cost. Energy will potentially remain a high-cost item in the future for municipalities which operate waste water treatment processes (Burgess, 2013). Burgess (2013) equally points out that the power outages that were experienced between November 2007 and January 2008 as well as in 2013 were mainly due to demand outstripping supply. Hence it is imperative to keep the energy usage low at WRRFs while maintaining the effluent standards imposed by the Department of Water and Sanitation (DWS). As an example of the effects of electricity costs increases in WWTPs, Burgess (2013) states the following “It is estimated that by 2020 the cost of electricity for the treatment of wastewater in Johannesburg will have risen from R97 million per annum to more than R300 million per annum, making the existing wastewater treatment operation possibly unaffordable. Failure of the wastewater treatment operations would thus have a devastating effect on the economy, environment, health services and social activities of the city”. An equalization tank is a valuable addition to these treatment systems and can provide the control options required to strategize and optimize energy use for efficient operation of the WRRF.

However, to understand the relevance of equalization tanks within the WRRF, it is essential to identify the treatment technologies used to treat wastewater in S.A. (As seen in Table 1), knowing that balancing tanks are mainly used with AS systems.

Table 1 : Treatment technology types in South Africa (Scheepers and van de Merwe-Botha, 2013)

		Water Treatment Technology Types															
		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
Province	Technology per province	Activated Sludge	Activated Sludge & BNR	Activated Sludge & Diffused Air	Activated Sludge & Extended Aeration	Activated Sludge & MBR	Activated Sludge & SBR	Activated Sludge & Mechanical Aeration	Aerated Lagoons/Oxidation Ponds	Anaerobic Contactor	Anaerobic Ponds/Facultative Ponds	Biological (Trickling) Filters	Passive Ditch	Rotating Biological Contactor	Package Plants	Other	Unknown or not specified
Limpopo	75	13	3	0	0	0	0	0	44	0	0	14	0	0	1	0	0
Mpumalanga	94	39	4	0	0	0	0	0	24	0	1	22	0	2	0	0	2
Gauteng	69	24	19	0	1	0	1	0	2	0	0	19	0	0	0	3	0
North West	42	19	4	0	2	0	0	0	12	0	0	5	0	0	0	0	0
Free State	134	36	3	0	0	0	0	1	49	0	14	27	1	0	0	0	3
Northern Cape	75	11	1	0	1	0	1	0	52	0	2	6	0	0	0	1	0
KwaZulu Natal	158	58	2	17	5	1	1	4	38	0	7	21	0	1	2	1	0
Eastern Cape	146	45	4	0	0	0	1	0	61	0	2	19	2	2	6	2	2
Western Cape	182	65	1	4	0	0	2	2	57	0	3	12	3	3	3	1	26
Totals	975	310	41	21	9	1	6	7	339	0	29	145	6	8	12	8	33
	975	1-7 : Activated sludge and variations							8 & 10 : Ponds and lagoons			11 : Biofilter	9 & 12 - 16 : Other				
	975	395							368			145	67				

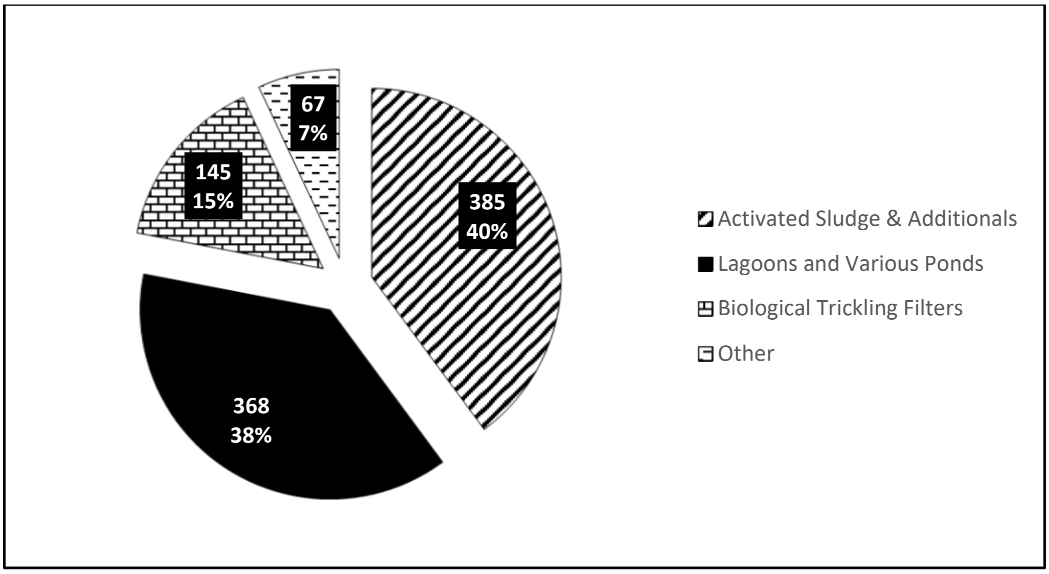

Figure 2: Treatment technology types in South Africa (Scheepers & van de Merwe-Botha, 2013)

Figure 2 shows that most treatment plants in S.A. use a form of AS technology.

Scheepers & van de Merwe-Botha's (2013) investigations indicate that there is a higher preference of more complex and potentially costly technology like AS in S.A. as opposed to the lower to medium level technologies to treat waste. They report that this might be due to the stricter effluent standards imposed by the DWS, but it might not always be the case. It is possible that there is not enough attention directed towards investigating low to medium level alternatives. It was noted that sustainability of higher-level technologies like the AS may overwhelm some municipalities, resulting in poor effluent, due to a lack of skills, cost recovery and power consumption amongst other factors (Scheepers & van de Merwe-Botha, 2013). Can the installation of an equalization tank help with such factors, improve operation of high-level technology treatment types used and improve effluent quality and equally reduce the power consumption? This is a research question that requires investigation. The development of a mathematical model that accurately depicts the balancing tank within a plant wide setting shall be able to identify the potential for energy saving, effluent quality improvement and provide expert guidance towards the design and operation of future waste treatment systems.

Scheepers & van de Merwe-Botha (2013) provide a summary of the technology level trends from their study as shown in Figure 3.

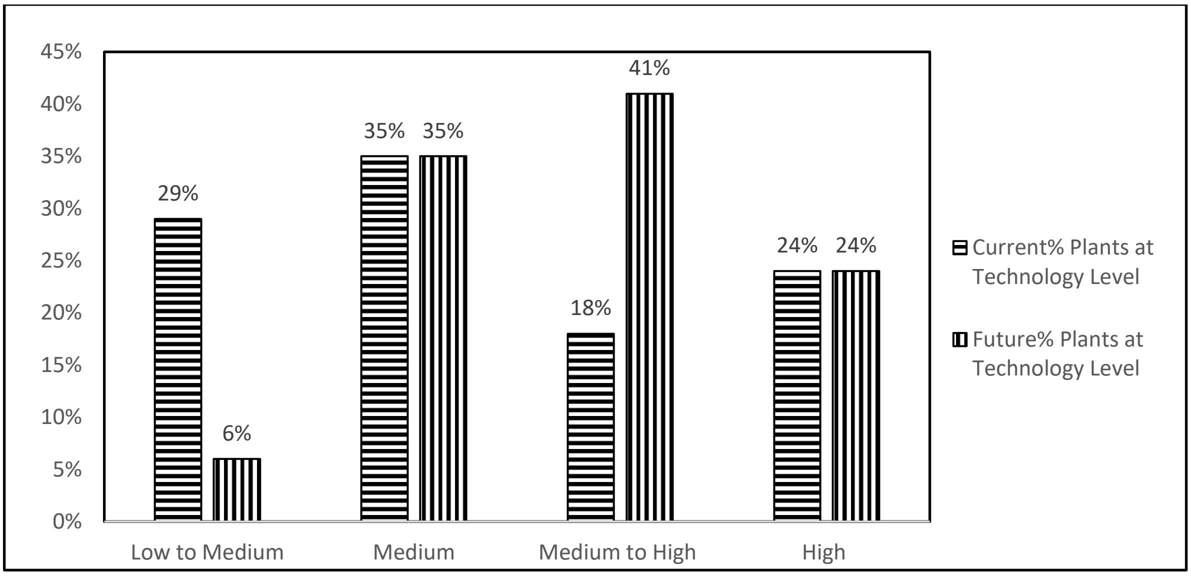

Figure 3: Technology level trends (Scheepers and van de Merwe-Botha, 2013)

In the study it was highlighted that upgrading to a higher technology level is the main trend with respect to wastewater treatment. This has serious implications, not just from a performance point of view (with respect to some municipalities), but also from an energy and cost point of view. Scheepers & van de Merwe-Botha (2013) pointed out that high-level technologies use more energy and therefore are costlier, with energy-intensive technologies being opted for to lower energy technologies; hence, treatment of wastewater will be more expensive in the foreseeable future due to higher energy requirements and tariff increases. This indicates that energy conservation measures will be crucial in wastewater treatment plants in the near future.

Table 2 summarizes the typical energy requirements of AS treatment plants of different sizes.

Table 2: Energy requirement for typical activated sludge treatment plants (Scheepers and van de Merwe-Botha, 2013)

Energy requirement for activated sludge treatment plant					
Process Unit	kWh/day				
	0.5 Ml/d	2 Ml/d	10 Ml/d	25 Ml/d	100 Ml/d
Pumping	28.53	114.12	507.71	1205.28	4336.37
Inlet Works	6.74	26.95	48.64	89.82	334.18
Primary Clarifiers	1.98	7.93	42.63	102.37	409.47
Aeration	70.27	281.08	1453.73	4705.55	17908.18
Secondary Clarifiers	1.98	7.93	42.63	102.37	409.47
Disinfection	0.13	0.53	2.73	17.83	70.01
Sludge Management	159.3	637.18	1424.77	1194.72	4379.96
Lights & Buildings	26.42	105.67	218.61	528.34	1585.03
Totals	295.34	1181.37	3741.45	7946.28	29432.67
Consumption ratio (kWh/Ml)	590.69	590.69	374.15	317.85	294.33

Overall aeration is the most energy intensive function in plants of 10Ml/d and above. Therefore, for such plants, electricity cost is the most significant cost in wastewater treatment plants in 1st world countries, and with the electricity cost steadily increasing in S.A., electricity costs are equally steadily becoming the most expensive part of running WWTPs with aeration tanks (Scheepers & van de Merwe-Botha, 2013).

Table 3: Running cost of a typical plant in a 1st world country (Scheepers and van de Merwe-Botha, 2013)

Description	**Percentage**
Wastewater discharge fee (similar to the SA "Waste Discharge Charge System Levy")	18%
Electricity cost	27%
Chemical cost	6%
Staff cost	18%
Maintenance and replacement cost	10%
Sludge disposal and transport	13%
Administration cost	9%

2.1 WWTP biological processes

An illustration of a WWTP can be seen in Figure 4. The main processes of concern for this research are the secondary settling tank performance, as well as the biological process performances in the AS reactor, due to the adverse effects variations in daily influent flow and environmental conditions have on them.

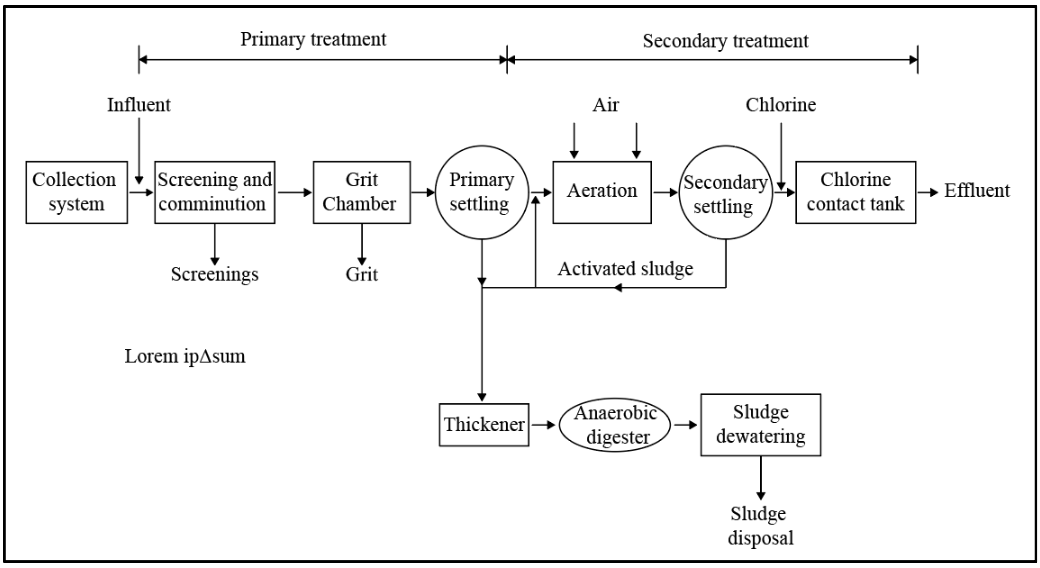

Figure 4: Wastewater treatment process schematic (Spellman, 2003)

The biological processes occurring in the WWTP are primarily the organics, nitrogen and phosphorus removal.

Organic removal is the primary function of an AS system. Ordinary heterotrophic organisms (OHOs) are responsible for breaking down the organic material for metabolic purposes. Dissolved oxygen acts as an electron acceptor in this process, and hence adequate provision of dissolved oxygen is crucial (Ekama & Wentzel, 2008a).

Nitrification forms part of the process through which NH_3 is removed from the wastewater, and nitrification is mediated by nitrifiers. Nitrification is the biological process in which Free and Saline Ammonia (FSA) is oxidized to nitrite and nitrate in the aerobic reactor (Ekama & Wentzel, 2008b). Numerous factors influence nitrification, but the relevant parameters to this research are as follows;

- Dissolved oxygen (DO) concentration: Low DO concentrations have a negative effect on the nitrification rate. At very low DO concentrations, the growth rate of nitrifiers can significantly reduce, resulting in poor nitrification and FSA in the effluent.
- Cyclic flow and loading rates: Nitrification efficiency of the AS system decreases significantly under cyclic flows and loading rates. Varying daily inflow causes the effluent NH_3 to increase, as the nitrifiers cannot operate above their maximum activity rate, and hence not all NH_3 is oxidized (Ekama & Wentzel, 2008b).

Phosphorus removal is required to prevent eutrophication downstream of the WWTP. It is mediated by Phosphorus accumulating organisms (PAOs). The PAOs are aerobes and require dissolved oxygen to be able to take up phosphorus (P) in the aerobic reactor and subsequently be flushed out of the treatment plant (resulting in Phosphorus removal) (Henze *et al.*, 2008). Hence, knowledge on how much dissolved oxygen is to be supplied in the reactor is important.

2.2 Activated sludge and anaerobic digestion stoichiometry

It is important that the link between the biological processes occurring in various unit processes of the WWTP and the equalization tank effects is established. The equalization tank mainly affects the influent flow and hence the flux (loads) of materials entering the WWTP; hence, these effects can be simulated using appropriate models.
For this, stoichiometric equations for steady state models of the biological processes happening in the WWTP process have already been developed (Ekama *et al.*, 2011).

Such equations relate the organic loads, flows and concentrations to the treatment performance (AS, AD, oxygen utilisation, organic removal) and effluent components from wastewater treatment plants. As the equalization tank directly impacts the flow and consequently the loading rate into WWTP, its impacts on the treatment performance, effluent, and design of various unit processes can easily be monitored using these stoichiometric equations and models. One of the goals of this research is to be able to depict such effects on WWTP and determine how beneficial they can be for the future design and operation of WWTP in S.A. and around the world.

The possible impact of the balancing tank on the organic, nitrogen, phosphorus removal and AD is discussed in this section. The AS and AD stoichiometry models are listed in the Table 4 (definitions of terms in Appendix A).

Table 4: AS and AD stoichiometry equations

Equation	**Equation Number**	**Reference**
Organic Removal		
$FS_{ti} = Q_i S_{ti}$	(1)	Ekama and Wentzel (2008a)
$FS_{bi} = FS_{ti}(1 - f_{S'us} - f_{S'up})$	(2)	
$FX_{Ii} = \frac{FS_{ti} f_{S'up}}{f_{cv}}$	(3)	
$FX_{IOi} = Q_i X_{Ioi}$	(4)	
$MX_{BH} = FS_{bi} \frac{Y_H R_s}{(1 + b_H R_s)}$	(5)	
$MX_{EH} = f_H b_H MX_{BH} R_s$	(6)	
$MX_I = R_s \frac{FX_{Ii}}{f_{cv}}$	(7)	
$MX_V = MX_{BH} + MX_{EH} + MX_I$	(8)	
$MX_{IO} = FX_{IOi} R_s + f_{iOHO} MX_{BH}$	(9)	
$MX_t = MX_V + MX_{IO}$	(10)	
$FO_c = FS_{bi}[(1 - f_{cv} Y_H) + (1 - f_H) b_H \frac{Y_H f_{cv} R_s}{(1 + b_H R_s)}$	(11)	
$FX_t = \frac{MX_t}{R_s}$	(12)	
$S_{te} = S_{use} + f_{cv} X_v$	(13)	
Nitrogen Removal		
$N_a = N_{ae} = \frac{K_{nT}(b_{AT} + \frac{1}{R_s})}{\mu_{AmT} - (b_{AT} + \frac{1}{R_s})}$	(14)	Ekama and Wentzel (2008b)
$N_{ae} = \frac{K_{nT}(b_{AT} + \frac{1}{R_s})}{\mu_{AmT}(1 - f_{xt}) - (b_{AT} + \frac{1}{R_s})}$	(15)	
$N_{ae} = \frac{K_{nT}}{(S_f - 1)}$	(16)	
$N_{ae} = N_{ai} + N_{obsi} + N_{obpi} - (N_s - N_{oupi})$	(17)	
$N_{te} = N_{ouse} + N_{ae}$	(18)	
$N_s = f_n \frac{MX_v}{Q_i R_s}$	(19)	
$N_{te} = N_{ti} - N_s$	(20)	
$N_{ne} = N_c = N_{ti} - N_{te} - N_s$	(21)	
$D_{p1RBCOD} = \frac{f_{Sb's} S_{bi}(1 - f_{cv} Y_H)}{2.86}$	(22)	

$D_{p1SBCOD} = \frac{K_2 f_{x1} S_{bi} Y_H R_s}{(1 + b_h R_s)}$	(23)	
$D_{p1} = D_{p1RBCOD} + D_{p1SBCOD}$	(24)	
$D_{p3SBCOD} = \frac{K_3 f_{x3} S_{bi} Y_H R_s}{(1 + b_h R_s)}$	(25)	
$D_{p3} = 0 + D_{p3SBCOD}$	(26)	
$FO_n = 4.57 N_c Q_i$	(27)	
$FO_d = 2.86(N_c - N_{ne})Q_i$	(28)	
$FO_t = FO_c + FO_n - FO_d$	(29)	
Phosphorus Removal		
$MX_{PAO} = \frac{Y_{PAO}}{(1 + b_{PAO,T} R_s)} FS_{bs,PAO} R_s$	(30)	Henze *et al.* (2008)
$MX_{E,PAO} = f_{XE,PAO} * b_{PAO,T} * MX_{PAO} * R_s$	(31)	
$MX_{OHO} = \frac{Y_{OHO}}{(1 + b_{OHO,T} R_s)} FS_{b,OHO} R_s$	(32)	
$MX_{E,OHO} = f_{XE,OHO} * b_{OHO,T} * MX_{OHO} * R_s$	(33)	
$MX_I = \frac{f_{XI,COD,i} FS_{ti} R_s}{f_{cv}}$	(34)	
$S_{bsf,i,conv} = S_{bsf,i} - 8.6\left(sS_{NO3,s} + S_{NO3,i}\right) - 3.0(sS_{O2,s} + S_{O2,i})$	(35)	
$S_{bsf,ANn} = \frac{\frac{S_{bsf,i,conv}}{(1+s)}}{[1 + k_{F,T} \frac{MX_{OHO}}{Q_i(1+s)} \frac{f_{AN}}{N}]^n}$	(36)	
$FS_{bs,PAO} = Q_i\left[S_{bsf,i,conv} - (1+s)S_{bsf,ANn}\right] + Q_i S_{bsai}$	(37)	
$\Delta P_{PAO} = f_{P,PAO} \frac{MX_{PAO}}{R_s} \frac{1}{Q_i}$	(38)	
$\Delta P_{OHO} = f_{P,OHO} \frac{MX_{OHO}}{R_s} \frac{1}{Q_i}$	(39)	
$\Delta P_{XE} = f_{P,XE} \frac{(MX_{E,PAO} + MX_{E,OHO})}{R_s} \frac{1}{Q_i}$	(40)	
$\Delta P_{XI} = f_{P,XI} \frac{MX_{I,i}}{R_s} \frac{1}{Q_i}$	(41)	
$\Delta P_{SYS,POT} = \Delta P_{PAO} + \Delta P_{OHO} + \Delta P_{XE} + \Delta P_{XI}$	(42)	
$\Delta P_{SYS,ACT} = \min\,(\Delta P_{SYS,POT}; T_{P,i})$	(43)	
$MX_{VSS} = MX_{PAO} + MX_{OHO} + MX_{E,PAO} + MX_{E,OHO} + MX_I$	(44)	
$MX_{FSS} = f_{FSS,OHO} MX_{OHO} + f_{FSS,PAO} MX_{PAO} + FX_{ISS,i} R_s$	(45)	
$MX_{TSS} = MX_{VSS} + MX_{ISS}$	(46)	

$f_{p,TSS} = \frac{\frac{f_{P,OHO}MX_{OHO}}{f_{VT}}}{MX_{TSS}} + \frac{\frac{f_{P,XE}(MX_{E,OHO} + MX_{E,PAO})}{f_{VT}}}{MX_{TSS}} + \frac{\frac{f_{P,XI}MX_{I,i}}{f_{VT}}}{MX_{TSS}} + \frac{\frac{f_{P,PAO}MX_{PAO}}{f_{VT,PAO}}}{MX_{TSS}} + \frac{f_{P,FSS,i}MX_{FSS}}{MX_{TSS}}$	(47)	
$X_{P,e} = f_{P,TSS}TSS_e$	(48)	
$T_{P,e} = T_{P,i} - \Delta P_{SYS,ACT} + X_{P,e}$	(49)	
$FO_{2,PAO} = (1 - f_{CV}Y_{PAO})FS_{bs,PAO} + f_{CV}(1 - f_{E,PAO})b_{PAO,T}MX_{PAO}$	(50)	
$FO_{2,OHO} = (1 - f_{CV}Y_{OHO})FS_{b,OHO} + f_{CV}(1 - f_{E,OHO})b_{OHO,T}MX_{OHO}$	(51)	
$FO_C = FO_{2,OHO} + FO_{2,PAO}$	(52)	
Anaerobic Digestion		
$R = \frac{V}{Q}$	(53)	Sötemann *et al.* (2005)
$S_{bpi} = \left(1 - f_{PS'up}\right)S_{ti} - S_{bsai}$	(54)	
$S_{upi} = f_{PS'up}S_{ti}$	(55)	
$S_{bp} = \frac{K_s(\frac{1}{R} + b_{AD})}{Y_{AD}K_m - (\frac{1}{R} + b_{AD})}$	(56)	
$S_{bp\ hydrolysed} = S_{bpi} - S_{bp}$	(57)	
$Z_{AD} = \frac{Y_{AD}(S_{bpi} - S_{bp})}{[1 + b_{AD}R(1 - Y_{AD}\{1 - f_{AD}\})]}$	(58)	
$Z_E = f_{AD} * b_{AD} * Z_{AD} * R_S$	(59)	
$r_h = \frac{K_m S_{bp}}{[Z_{AD}\left(K_s + S_{bp}\right)]}$	(60)	
$S_m = (1 - Y_{AD})Rr_h + S_{bsai}$	(61)	
$Q_m = [(1 - Y_{AD})Rr_h + S_{bsai}]\frac{24}{64}$	(62)	
$S_{te} = S_{up} + Z_{AD} + S_{bp} + Z_E + S_m$	(63)	

$C_xH_yO_zN_a + \left(2x + a - z - \frac{ED_S}{D_B}(n + 2k - m) - \frac{(1-E)D_S}{4}\right)H_2O \rightarrow \left(x - a - \frac{ED_S}{D_B}(n - k) - \frac{(1-E)D_S}{8}\right)CO_2 + \left(\frac{(1-E)D_S}{8}\right)CH_4 + \left(\frac{ED_S}{D_B}\right)C_kH_lO_mN_n + \left(a - \frac{nED_S}{D_B}\right)NH_4^+ + (a - n\frac{ED_S}{D_B})HCO_3^-$	(64)	
$CH_3COOH \longrightarrow CH_4 + CO_2$ and	(65)	
$CH_3COO^- + H_2O \longrightarrow CH_4 + HCO_3^-$	(66)	
$S_{bsHAci} = \frac{S_{bsai}}{(1 + 10^{pH_i - pK'_a})}$	(67)	
$S_{bsAci} = \frac{S_{bsai}}{(1 + 10^{pK'_a - pH_i})}$	(68)	
$p_{CO2} = \frac{[HCO_3^-](1 + 10^{pK'_{c1} - pH} + 10^{pH - pK'_{c2}})}{10^{-pK'_{HCO2}}(1 + 10^{pH - pK'_{c1}} + 10^{2pH - pK'_{c1} - pK'_{c2}})}$	(69)	

Equations 1-13 in Table 4 represent the stoichiometric equations for organic removal in case P is not removed. The mathematical equations to obtain various parameters such as biomass mass, dissolved oxygen utilized, chemical oxygen demand (COD) in the effluent, etc., are as shown in these equations.

Equations 14-29 in Table 4 are the stoichiometric equations in case there is significant nitrogen removal. Parameters such as N in the effluent, denitrification potential, dissolved oxygen used for nitrification, etc., can be determined using such mathematical equations.

Equations 30-52 in Table 4 are the stoichiometric equations if P-removal is present in the AS. The mathematical equations to obtain various parameters such as P removed via biomass, P in the effluent, mass of biomass in the AS reactor, etc., are obtained using the aforementioned equations.

Equations 53-69 in Table 4 are the stoichiometric equations representing the processes taking place in the AD. Parameters like COD hydrolysed, biomass concentration in the AD reactor, methane production, etc., can be obtained using these equations.

2.2.1 Stoichiometric link of equalization tanks to AS and AD

A balancing tank ideally stores influent flows and pumps them out to the A.S at a constant flow rate (Dold, 1982). This provides several benefits (i.e., better operation of the WWTP, better removal of organic material, and improved performance of SSTs, which subsequently improve the effluent quality).

2.2.1.1 Impact of balancing tank on organic matter removal

With the organic matter removal stoichiometry presented in Table 4, important parameters can be adequately predicted as well as the effect of the balancing tank on such parameters (Ekama *et al.,* 2011).

In the presence of a balancing tank, the diurnal cyclic variations into the WWTP will be significantly dampened (i.e the influent flows and fluxes predicted from Equation 1-4 will be significantly dampened). The reactor solids masses (i.e, biomass, unbiodegradable particulates and inorganic solid; predicted using Equations 5-10) will be higher and remain more constant than if the flow variations are not dampened as there will be a more constant supply of organic matter and dissolved oxygen. The oxygen demand (predicted using Equation 11) will be easily predicted, and its supply will be better, improving organic matter removal. Due to the increase in reactor masses, the total sludge (predicted from Equation 12) will equally increase. As there is better organic removal, the effluent COD concentration (predicted from Equation 13) will be lower.

2.2.1.2 Impact of balancing tank on nitrogen removal

The presence of a balancing tank equally has an impact on nitrogen removal. Due to the balancing tank stabilising the flow into the AS reactor, the nitrification process can be performed more efficiently, resulting in less N in the effluent as the N utilisation by the biomass would be higher. The denitrification potential improves as there is a constant COD in the reactor; hence, reducing the N content in the effluent.

The dissolved oxygen demand required for nitrification (predicted from Equation 27) can be better predicted due to the constant flow into the reactor. Similarly, the dissolved oxygen recovered via denitrification can be predicted accurately with constant flows.

2.2.1.3 Impact of balancing tank on phosphorus removal

As previously described, if the reactor masses increase, they will be better predicted if a balancing tank is present. This means the PAOs and OHOs as well as their endogenous residue and the inert masses will increase which indicates that the P-removal potential by the system increases (from Equation 42). As a consequence of this, the P in the effluent will reduce (Equations 48 & 49). The dissolved oxygen utilized by the PAOs, can equally be predicted better and hence provided efficiently, which has a positive impact on the performance of the PAOs.

2.2.1.4 Impact of balancing tank on Anaerobic Digestion (AD)

With the presence of a balancing tank, the retention time of the waste in the AD can be more efficiently predicted (Equation 53). There will equally be an effect on the COD hydrolysed and hence the methane produced by the AD which is not known upfront. The prediction of the methane produced will thus be precise. The stoichiometric predictions to obtain the reactants from the AD will equally be improved; hence, the pH can equally be predicted more accurately. All these because the influent flows are more stabilised and hence more easily predictable at various time intervals.

2.3 UCT plant wide model

To be able to investigate the fate of the different components of the wastewater fed into WWTP, finding the most cost-effective methods for the operation and design of WWTP, to minimize the energy consumption and cost while improving effluent quality and maximizing nutrient recovery, the use of plant-wide models is necessary (Ikumi, 2011). These models can virtually replicate the system because they contain sets of simultaneous equations that represent the biological, physical and chemical processes occurring in the system, coded (from theoretical knowledge of plant procedures) into computer simulation programmes. Therefore, critical data that could be applied in long term planning and management of recoverable resources could be produced using such models (Ikumi *et al.*, 2014).

In this research, the UCT three phase (aqueous-gas-solid) plant-wide dynamic model (PWM_SA) is considered (Ikumi *et al.*, 2014). The progress up to date in this model includes the completion of steady-state and three-phase kinetic simulation models for nitrification denitrification (ND) AS, combined biological N and P removal AS and anoxic-aerobic (AAD) or AD of primary and waste AS with interlinking non-reactive thickening physical unit operations.

The PWM_SA to date is calibrated for;

- Hydrolysis kinetics for PS and WAS (Ikumi *et al.*, 2014)
- Kinetics of P uptake and P release in AS and anoxic aerobic digestion (AAD) systems and multiple mineral precipitation (Solon *et al.*, 2017)

The PWM_SA does not yet have a rigorous compatible model for a balancing tank, and hence the effects of such a unit process cannot yet be investigated at a plant-wide level. This research involved developing a balancing tank model and making it compatible with the PWM_SA to allow for system-wide exploration of operational strategies.

2.4 Reasons for using balancing tanks

Dold (1982), Mikola (2013) and Foess *et al.* (1977) agree that the influent to a WRRF exhibits a wide diurnal cyclic variation, both in flow rate and concentration (hence consequently in loading rate). These variations also happen from day to day, between weekends and weekdays, and from season to season. It has been observed that these diurnal cyclic loading rate variations can range from 4 to 6 times to less than a quarter of the average daily loading rate; hence, the variations can be significant, as illustrated in Figure 5.

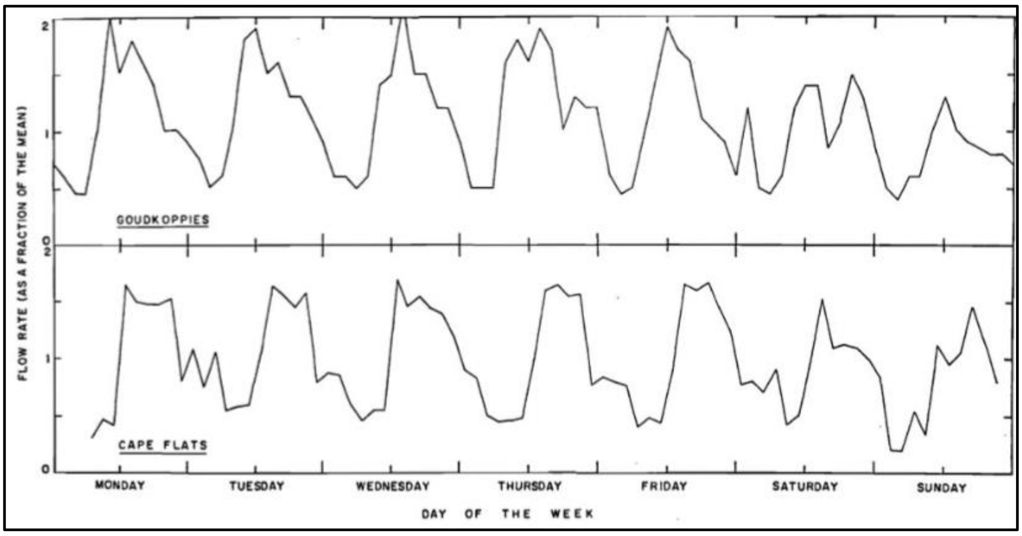

Figure 5: Raw wastewater flow variations (Dold, 1982)

2.4.1 Effects of daily influent flow variations

The effects of variations in the daily influent flow are as follows;

- In terms of design, Dold (1982) found that influent flow and concentration variations increase the capital cost of the WWTP as the unit processes will have to be designed for peak flow rate, resulting in oversized unit processes.
- With respect to WWTP performance, the varying loading rates and flows can cause hydraulic disturbances, which are extended to the secondary settling tanks, leading to sludge overflow to the effluent. The variations can equally cause inadequate oxygenation which can result in poor effluent quality due to reduced nitrification and COD removal and poor settling properties of the sludge (Mikola, 2013 & Dold, 1982).
- The operation becomes complicated and requires sophisticated methods to control aeration, i.e. the aeration will have to be constantly adjusted to provide the required oxygenation rates. Over aeration would lead to energy wastage and poor sludge performance including sloughing while under aeration would result in poor performance via the growth of undesirable filamentous organisms that cause sludge settling problems (Mikola, 2013). This is especially important for conditions where the effluent NH_3 requirements are strict and cannot fluctuate (e.g. if the plant is connected to a water reclamation system).

Including equalization tanks in WRRFs might help in having less varying influent flows and concentrations (Dold, 1982). Flow equalization comes with associated load equalization, but studies by Dold (1982), Foess *et al.* (1977) and Mikola (2013) proposed that this associated load equalization might not always be sufficient to significantly decrease the operational difficulties experienced at WWTPs; hence, adequate control strategies also need to be developed. This research is limited to the equalization of the inflows.

2.4.2 Assessing equalization tanks and benefits of equalization tanks

Traditionally, flow equalization was the main objective when using equalization tanks, but after research, no specific benefits could be seen via the use of such tanks in WRRFs. Dold (1982) notes that this is because it depends on the parameters analysed when trying to identify the benefits of equalization tanks, and this was one of the main reasons why research on equalization tanks has not been very extensive. Benefits of equalization tanks could be identified if:

- Relevant process parameters were measured (not COD, because the effluent concentration is virtually insensitive to dynamic loading conditions as most of the organic load stays in the reactor, is completely utilized or is wasted. Under steady state, only the soluble unbiodegradable organics escape via the liquid component with which effluent COD is measured). Benefits will be seen only if under cyclic conditions, the SST is overloaded for at least a portion of the day, otherwise no benefit will be identified.
- The operating parameters used during the evaluation make sense; if you consider systems with short sludge ages, the response of the process under variable input conditions is significantly dampened; hence, in such cases the efficiency of the flow equalization cannot be demonstrated, as opposed to systems with long sludge ages.
- The equalization facilities are adequately operated hence resulting in a good degree of flow equalization (Dold, 1982).

Foess *et al.* (1977) for example could not obtain significant improvements in the biological oxygen demand (BOD) of one of the plants used for the research, because BOD is not a good parameter to consider towards determining the benefits of equalization tanks. A similar assertion can be observed with COD.

The above points being considered, some benefits of equalization tanks are as follows:

- Improved performance of SSTs due to more constant solids loading rates as a result of the decrease in hydraulic disturbances. This also implies that SST will need to be also designed for a substantially lower flow rate. With flow equalization it is also likely that the flow to the plant can be increased without additional provision of settling capacity or the process can be operated at longer sludge ages if greater efficiency of nitrification is needed (Dold, 1982 & Mikola, 2013).
- Flow equalization equally minimizes denitrification challenges in existing plants whereby nitrification is an objective (if there is cyclic influent flow, more solids pass through the PST; hence, there is more sludge in the reactor, affecting the performance of denitrification, therefore more constant loading results in less sludge in the reactor) (Mikola, 2013).
- The decrease in hydraulic loading because of the use of an equalization tanks also results in an improved performance of the PST (Dold, 1982).
- Simplified control of in-plant flow rate dependent operations such as chemical dosing and recycle pumping (Mikola, 2013).
- Simplified control of aeration due to attenuations of influent load rate variations. This will reduce both capital costs for aeration and the operational problems in matching the oxygenation intensity to the actual dissolved oxygen requirements (Dold, 1982).
- Foess *et al.* (1977) showed that the addition of an equalization tank causes a marginal increase in power consumption of between 2 and 6%; hence, the overall effect of the tanks was beneficial.

- There was improved biological performance, as the sludge age and the aeration requirements can be better predicted and hence, COD removal, nitrification and denitrification will be better (Dold, 1982).

Although it is possible to reduce the influent flow rate fluctuations, it is impossible to obtain complete equalization; hence, it is crucial to have an adequate way to operate an equalization tank (Dold, 1982).

2.5 Types of equalization tanks

To be able to model equalization tanks, it is important to know the various types that exist. Equalization tanks can be operated in the constant volume mode whereby the tank is always full and hence the flow fluctuations are not reduced, because the influent flow equals the outflow. Concentration fluctuations are attenuated due to the time the waste spends in the tank (retention time). Equalization tanks can also be operated in the variable mode whereby the outflow rate from the tank is regulated, and the volume in the tank varies; hence, there can be flow and concentration equalization (Dold, 1982).

The variable volume mode has 2 types of physical configurations, the in-line arrangement, whereby all the influent flow to the process passes through the equalization tank and the side-line arrangement where only a portion of the influent flow passes via the equalization tank, while the rest flows directly to the downstream unit processes. The side-line arrangement is further divided into flow “splitting” (fixed fraction from the influent flow is diverted to the equalization tank and the rest passes directly to the downstream process) and flow “topping” (flows above the average daily flow are diverted to the equalization tank, while the remainder passes directly to the downstream process) (Foess *et al.*, 1977 & Dold, 1982).

Foess *et al.* (1977) found that both in-line and side-line equalization systems have been highly effective in levelling influent flow variations. However, while side-line equalization had only limited effectiveness in levelling the concentration of wastewater parameters, in-line equalization has better effectiveness due to continual blending. Mass equalization from both in-line and side-line systems are limited and they are mainly due to equalization of flow rate. Additional investigations are required to conclusively determine which type of equalization configuration is the best, but in this research only in-line equalization tanks shall be considered as initial results from Dold (1982) indicate the in-line equalization configuration provides better equalization.

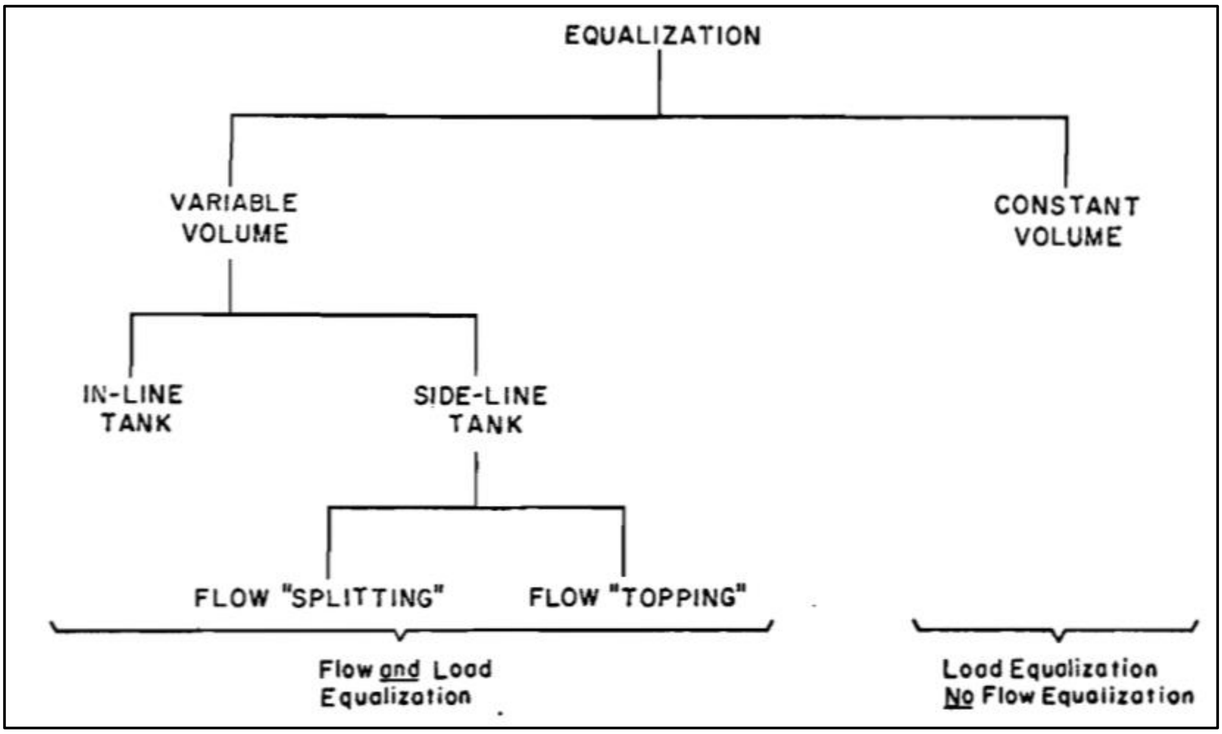

Figure 6: Hierarchy of equalization (Dold, 1982)

2.6 Control strategies for equalization tanks

Due to variations of the influent flows and loads, operation of WWTPs has been a challenge (Dold, 1982). The research aim from Dold (1982) was to develop a strategy (that will ensure optimal utilization of the available capacity of a tank, taking into consideration the variability of the daily influent cyclic flow) for automatic control of equalization basins which will overcome problems inherent in manually-controlled basins. That is, an "online" intelligent control strategy where the optimal outflow rate from the tank is specified from the microprocessor on the basis of the application of some optimization procedure. For this, he proposed an equalization algorithm, with an error function which uses a weighting factor to determine which component should be optimized (flow or concentration). To be able to develop a control strategy using such an algorithm, predictions of the influent patterns at any time "t" for the ensuing 24hrs are required. The predictions are based on historical flow and concentration as well as differences between actual and historical flow rates for the period before the prediction. The predicted influent values are used to determine the optimum simulated tank outflow profile for the next 24hrs using 30min intervals, the same process is repeated at the start of each control interval. The algorithm differentiates between influent patterns for weekdays and weekends; hence, the strategy reduces the effect of the transition from weekday to weekend and vice versa.

2.6.1 Development of the equalization algorithm

The control strategy used as a background for this research is that developed by Dold (1982).

It is important to first note that the essence of the problem of equalization is as follows; "Under the cyclic inputs of flow & load, determine the appropriate outflow rate at any time such that the flow and/or load will be optimally equalized." (Dold, 1982). This research focused on equalisation of flow. Dold (1982) considered two aspects for the development of the equalization algorithm;

- The development of an iterative procedure whereby, given some starting condition, optimal equalized flow and loading rate patterns can be determined.
- Formulation of the error expression.

To achieve such, an iterative procedure must be implemented.

2.6.1.1 Development of the iterative procedure

First, a mathematical analysis of the equalization tank response was performed. The relationship between inflow, outflow and tank volume response was derived from a material balance, as well as the concentration response.

(i) Outflow rate and volume response

Dold (1982) recognised that the relationship between inflow, outflow and tank volume is derived from a material balance as can be seen in Equation 70.

$$\begin{aligned}&[Rate\ of\ flow\ into\ tank]\\&\quad=[Rate\ of\ flow\ out\ of\ tank]+[Rate\ of\ evaporation\ from\ tank]\\&\quad+[Rate\ of\ accumulation\ in\ tank]\end{aligned} \qquad (70)$$

Assuming the rate of evaporation from the tank is negligible; the following was obtained (Equation 71);

$$\frac{dv}{dt}=F_0-F_1 \qquad (71)$$

Where;
V = Tank volume (m^3)
F_0 = Influent flow rate (m^3/d)
F_1 = Effluent flow rate (m^3/d)
t = Time (d)

Dold (1982) found that the analysis of the tank response was stable when an effluent flow profile was specified over the day, and knowing the influent flow pattern, the tank volume profile could be computed by integration.

(ii) Concentration response

The associated effluent concentration profile can be obtained once the flow and volume behaviour has been established. A mass balance can be used to determine this (see Equation 72).

$$\begin{aligned}&[Rate\ of\ mass\ flow\ into\ tank]=[Rate\ of\ mass\ flow\ out\ of\ tank]+\\&\qquad[Rate\ of\ mass\ removal\ by\ biological\ reaction]+\\&\qquad[Rate\ of\ mass\ accumulation\ in\ tank]\end{aligned} \qquad (72)$$

Assuming the rate of mass removal by biological reaction is negligible, the following can be obtained (Equation 73);

$$\frac{dC_1}{dt} = \frac{F_O(C_O - C_1)}{V} \quad (73)$$

Where;

C_0 = Influent concentration into balancing tank (mg/l)
C_1 = Balancing tank concentration (effluent concentration) (mg/l)
V = Volume of tank (m^3)

By approximating equations (75) and (76) by finite difference, Dold (1982) achieved (Equation 74 & 75);

$$V_t = V_{t-1} + \Delta t(F_0 - F_1) \quad (74)$$

And

$$C_t = C_{t-1} = \frac{\Delta t}{V_t}[F_0(C_{0t} - C_t)] \quad (75)$$

Where;

V_t = Volume of tank at time t (m^3)
V_{t-1} = Volume of tank at time (t-1) (m^3)
Δt = Length of time of interval (d)
F_0 = Mean influent flow rate from (t-1) to t (m^3/d)
F = Mean effluent flow rate from (t-1) to t (m^3/d)
C_t = Balancing tank concentration at time t (mg/l)
C_{t-1} = Balancing tank concentration at time (t-1) (mg/l)

(iii) Optimum tank outflow rate profile

Dold (1982) used an iterative process to generate the diurnal equalization tank effluent flow rate profile. This method gradually improves the outflow rate profile in a step-by-step fashion, until no more improvements can be made with regards to an error criterion which was developed. The extended step-by-step process as well as the mechanism to make changes on the outflow rate profile was reported in Dold (1982). Dold (1982) used a method called the fast convergence method to optimize the outflow profile, which required fewer computations to get the optimal values, compared to other methods tested. The prescribed methodology from Dold (1982) served as the background of this research. This also involved the development of an error expression.

2.6.1.2 Development of the error expression

The error expression was used as the criterion for evaluating the progress when optimizing the tank outflow profile. Dold (1982) considered various aspects he deemed as of high importance for this expression. It is important to note that modifications can be made when necessary on this error expression in the developed computer model to ensure the model is efficient.

When developing the error expression, the main objective is to quantify the various factors that will contribute to a “satisfactory” performance of the equalization facility. These factors may affect the equalization facility differently from plant to plant; hence, should be taken into consideration in the error expression (Dold, 1982).

The factors considered by Dold (1982) which were equally tested in the model are as follows;

A. An equalization error is associated to any deviation in the tank outflow rate and the effluent mass loading rate. To account for the relative importance attached to these two parameters, a weighting factor has been incorporated Dold (1982) (See Equation 76).

$$E_e = \alpha E_f + (1 - \alpha) E_{ld} \tag{76}$$

Where:

E_e = Equalization error
α = A weighting factor (Constant)
E_f = Flow error
E_{ld} = Load error

B. A large penalty error is imposed when the upper and lower tank volume limits are exceeded, i.e., the physical constraints of the system are not obeyed. To avoid instabilities in the development of the optimal tank outflow rate profile, this error was made to be continuous and to increase gradually with changes in volume (Dold, 1982).

For the upper limit (Subjective, but will initially be taken as 95% of tank volume) - Equation 77;

$$E_{lm} = \beta \sum_{i=1}^{n-1} [V_p - (V_{lu} - \delta_v)]^6 \ for\ all\ i : V_p > (V_{lu} - \delta_v) \tag{77}$$

For the lower limit (Subjective, but will initially be taken as 5% of tank volume) – Equation 78;

$$E_{lm} = \beta \sum_{i=1}^{n-1} [V_p - (V_{lb} - \delta_v)]^6 \ for\ all\ i : V_p < (V_{lb} + \delta_v) \tag{78}$$

Where;

E_{lm} = Penalty error for tank volume limit
V_p = Equalization tank hold-up (% of total tank volume)
V_{lu} = Upper normal tank volume limit (% of tank volume)
V_{lb} = Lower normal tank volume limit (% of tank volume)
δ_v = Volume differential within the limits at which E_{lm} attains a value (%)
β = A weighting factor (constant)

C. An error was equally associated to the rate of change of the tank outflow rate (Equation 79). This was to ensure that the optimal tank outflow rate profile has no "spikes", and to avoid rapid changes in the tank outflow rate over a small range, as this would have effects on downstream processes such as secondary clarifiers (Dold,1982).

$$E_s = \omega \sum_{i=1}^{n} \left| \frac{dF}{dt} \right| \tag{79}$$

Where;

E_s = Penalty error for rate of change of tank outflow rate.
ω = A weighting factor (constant) included so that the effect of E_s does not override the effect of E_e.
n = Number of simulation points over the day from 00h00 to 24h00, inclusive.

From points A,B, & C, the total error expression can be expressed as follows (Equation 80);

$$E_t = E_e + E_{lm} + E_s \tag{80}$$

E_t above was accepted for further development by Dold (1982), for various forms of input flow and load patterns, and hence shall be used in the conceptualization of the model. E_t ensured;

- Steady convergence to the optimal tank outflow profile and a relatively “smooth” outflow rate profile, and
- An associated tank hold-up profile which respected the tank volume physical constraints.

It is important to note that preliminary values for the weighting factors which worked for most of the scenarios tested by Dold (1982) were used and are as shown in the Table 5.

Table 5: Initial values of weighting factors (Dold, 1982)

Symbol	Description	Value
α	Weighting factor for equalization error	0.5
β	Weighting factor for tank volume limit penalty error	$2*10^{-6}$
ω	Weighting factor for rate of change of tank outflow rate penalty error	50

In case a situation arises where E_t does not ensure steady convergence and a suitable tank profile, for a selected influent flow and mass loading rate pattern, the weighting factors in the error expression will have to be adjusted. The effects of the weighting factors on the optimization of the influent flow and load patterns had to be tested during model evaluation.

3. Model development

3.1 Introduction

The WWTP equalization tank model was developed in three steps; the first step was to develop all the required equations necessary to develop an equalization tank model. The second step involved assessing the mass balance (i.e., model verification for internal consistency) throughout the equalization tank with the final step involving the assessment of the whole model as to whether it generates reasonable and scientifically sound outputs via a scenario analysis.

3.1.1 Introduction to platforms used in developing the model

3.1.1.1 Microsoft Excel Visual Basics (VBA)

Due to the iterative nature of the optimisation process, VBA was deemed to be the most efficient tool to replicate the process and is the most common interface which can be used with the Microsoft Excel interface.

3.1.1.2 WEST®

To be able to determine the impact of the equalization tank in a plant-wide setting, WEST® was deemed the best tool, as multiple unit processes of a WWTP have already been coded into the software, and they can all be connected to simulate WWTPs quite accurately.

The software can determine the improvement in performance (effluent quality, aeration prediction, electricity consumption) of a WWTP in case an equalization tank is included in its configuration.

3.1.2 System description

As described in Section 1.1 the system in which the equalization tank model was incorporated is as shown in Figure 1. The equalization tank's role is to equalize the flows throughout the day, to ensure better organic removal performance.

3.1.3 Assumptions considered

The following assumptions were considered in the development of the model;

- It was assumed that the tank content is completely mixed; hence, the concentrations at all points in the tank are equal (the potential for imposing of computational fluid dynamics for unmixed conditions could be considered in subsequent work).
- It was assumed that no biological nor chemical process would occur in the equalization tank.
- Only the physical process (flow) was considered and was balanced. Therefore, a balance of mass was assumed to be just a consequence of the flow balance. A flow (water) balance is ensured as well as a mass balance of the material present in the tank.

3.2 Step 1: Development of the optimization model

3.2.1 Fundamentals of the model

The Dold (1982) modelling approach was used as a basis for the development of this model, but not exclusively (see Section 2.6.1). Error functions were used to guide the model towards optimal balanced outflows. Additions to the Dold (1982) model have been done at various steps described in the subsequent sections.

3.2.1.1 Modifications to Dold (1982) optimization method

A few modifications were made to Dold (1982) method, to ensure that the model runs smoothly on VBA. These modifications were as follows;

- Mechanism for making a single change to the outflow rate profile, and
- Modification of the fast convergence method.

(i) Mechanism for making a single change to the outflow rate profile

The mechanism for making a single change to the outflow rate profile by Dold (1982) is as follows (copied directly from Dold (1982));

- "Identify the interval under consideration. Assume that the interval is that from time tn to time (tn+24/N), in this case time (tn+2) in Figure 7. This is the [tn(N/24)+1]th interval in the day. The flow rate at the end of the interval is either increased or decreased, as required, by an amount delta (an increase is illustrated in this case)."
- "Increment the flow rate over this interval linearly from point A (the beginning of the interval) to point B (the end of the interval), where the increment at point B is delta."
- "Decrement the flow rate linearly from that at point B (the end of the interval) to the existing flow rate at point C (the end of the next interval)."
- "Make an adjustment to the outflow rate profile in order to maintain a material balance, i.e. to ensure that the mean outflow over the day remains at 1.0. The effect of steps 2 and 3 is to increase the outflow rate over two intervals from time tn to time [tn+2*(24/N)], in this case (tn+4). To maintain a material balance the incremental flow volume must be decremented over the remaining (N-2) intervals. The method used is to linearly decrease flow rate from point C to the midpoint of the remaining (N-2) intervals, i.e (N-2)/2 intervals forward of point C, in this case (12-2)/2 = 5 intervals of 2 hours, i.e 10 hours forward of time (tn+4), to point D; then increase linearly from point D back to point A. The exact decrement at point D depends on delta, and is chosen to satisfy the material balance, i.e Area 1 equals Area 2 in Figure 7."

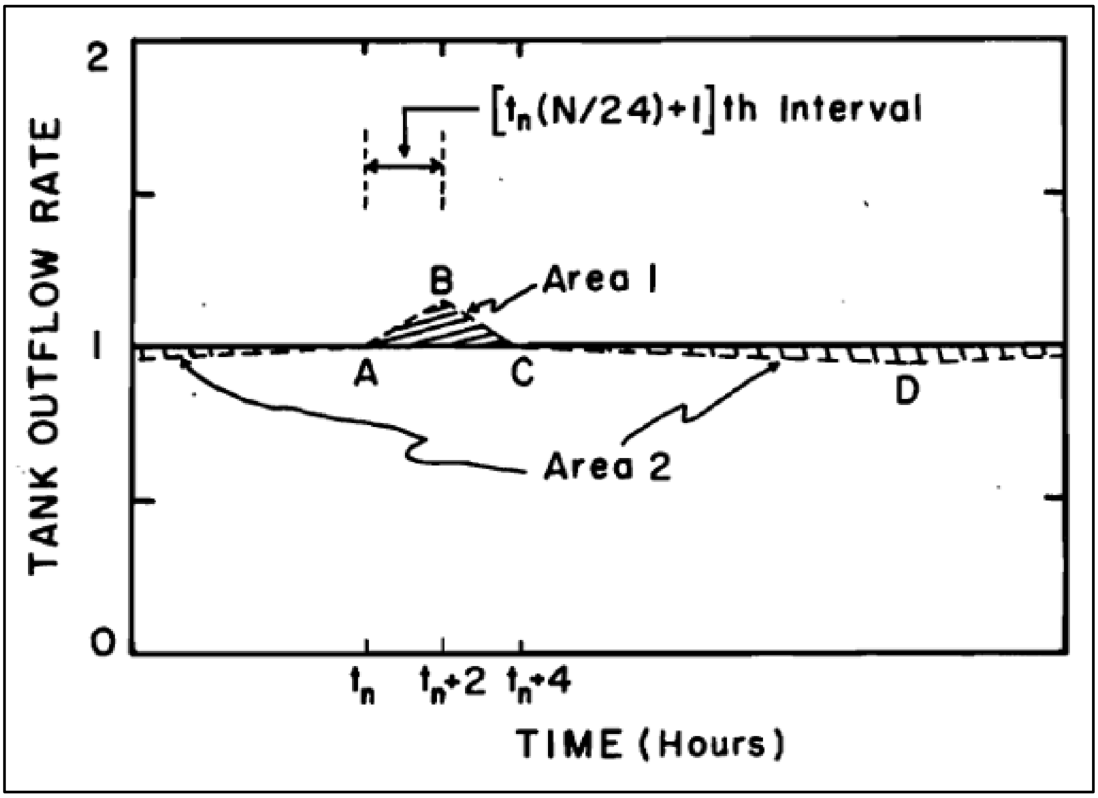

Figure 7: Method of adjustment of the outflow profile (Dold, 1982)

The mechanism for making a single change to the outflow rate profile in this research was as follows;

- Identify the interval under consideration. The flow rate at the end of the interval was either increased or decreased by delta. As illustrated in Figure 9, interval N (N=2) was considered. Intervals instead of time were used the x-axis.
- The flow rate was increased linearly from interval N-2 to N, where the increment at N was delta.
- The flow rate was decreased linearly from interval N to the existing flow rate at interval N+2.
- An adjustment was then made to the outflow rate profile to maintain a suitable material balance, which differed from Dold (1982), to allow VBA to function smoothly. An interval was identified (N+1)+0.5*(Total intervals-N-1), which was at the midpoint between interval N+2 and the total number of intervals considered. At that interval, the outflow rate was (in this case) decreased by delta, to ensure an adequate material mass balance.

(ii) Modification of the Fast Convergence Method

The optimisation method used was a modified version of the fast convergence method as follows;

- At each interval, if the direction of delta applied caused a decrease in the total error, delta was applied repeatedly at that interval until the error increased, which indicated that the minimum error at that interval had been crossed. Delta was then applied in the opposite direction once, to obtain that minimum error.
- The optimization continued until the total error did not decrease after optimization.

To achieve optimization, some key parameters used must be highlighted.

3.2.1.2 The fundamental parameters used

The fundamental parameters used in this model were as follows;

- The starting volume, which depends on the retention time selected.
- The number of intervals a day will be broken down into, knowing that each interval will be optimized.
- The number of days the optimization will be done for.
- The upper limit of the tank allowed (typical value is 95% of retention time).
- The lower limit of the tank allowed (typical value is 5% of retention time).
- The error coefficients which enable to determine which error weighs the most (typical values considered by Dold (1982) were; α=0.5, β=2.0E-06, ω=50).
- The value Delta (Δ) applied at each interval during the optimisation (This was chosen by the operator).

3.2.1.3 Influent data

Having set the number of intervals, the period (time) for each interval can be determined. Measured influent values were provided as input to the model at each of those intervals (i.e., the inflow, and various influent concentrations, as well as the influent loads). From these inputs, the influent flow and loading rate errors were computed, by the model, using Equations 81 to 82.

$$E_f = \sum_{i=1}^{n-1} \left(\frac{F_i}{\overline{F}} - 1\right)^2 \Delta t \tag{81}$$

Where;

E_f = Flow error

F_i = Influent flow (Ml/d)

$\overline{F}$ = Mean daily influent flow (Ml/d)

Δt = Time taken for an interval to elapse (hr)

$$E_{ld} = \sum_{i=1}^{n-1} \left(\frac{L_i}{\overline{L}} - 1\right)^2 \Delta \tag{82}$$

Where;

E_{ld} = Load error

L_i = Influent load (kg/d)

$\overline{L}$ = Mean daily influent load (kg/d)

Δt = Time taken for an interval to elapse (hr)

Overall, the equalization error was a weighted sum of the individual flow and loading rate errors described. The equalization error was as in Equations 83 & 84.

$$E_e = \frac{\{\alpha \sum_{i=1}^{n-1} (\frac{F_i}{\overline{F}} - 1)^2 \Delta t + (1-\alpha) \sum_{i=1}^{n-1} (\frac{L_i}{\overline{L}} - 1)^2 \Delta t\}}{(n-1)} \tag{83}$$

$$E_e = \alpha E_f + (1 - \alpha) E_{ld} \tag{84}$$

The daily volume of liquid and the daily mass of the COD in the influent can be computed using a mass balance at a later stage. The Simpson rule was used to get an average for the flow and the mass daily (see Equation 85).

$$Average = \frac{d}{3} * [First\ value + Last\ value + 4 \times \sum Even\ intervals + 2 \times \sum Odd\ \ intervals] \tag{85}$$

Where;
d= Time taken by one interval (hr)

3.2.1.4 Outflow Data

The following were computed; effluent flow, tank concentrations (COD, VFA, TOC, TKN, FSA, TP, OP), tank volume, fractional volume, rate of change of the tank volume (dv/dt), load out, flow error (E_f), load error (E_{ld}), volumetric error (E_{lm}), flow rate change error (E_s). Table 6 highlights the significance of individual parameters in the outflow data.

Table 6: Significance of outflow data

Output	Definitions	Significance as selected output variables
Effluent flow	Flow out of the balancing tank at any point in time	Used to determine if the balancing tank performed or not, when compared with the inflow
Tank concentrations	Concentrations out of the tank and in the tank (as we assume completely mixed tank)	Important values to know as the flow out of the tank moves to the next unit processes, and their treatment potentials can be assessed after flow balancing
Tank volume	Volume of liquid in the tank at any point in time	Knowing the tank volume at each time period helps determine if the balancing tank is over stressed or not
Fractional volume	This is the volume of liquid in the tank at any point in time divided by the total volume	Gives an indication as to how full the tank is and can help in determining at which interval more or less outflow is needed (between 0 and 1)
Rate of change of the tank volume (dv/dt)	This is the rate at which the liquid is pumped out of the tank.	Helps in determining if the outflow profile is smooth or “spiky”.
Load out	This is the mass of solid out the tank	Helps in knowing the mass that will be treated in the subsequent unit processes

Flow error (E_f)	This is an error associated to the magnitude of flow out of the tank	Helps in computing the total error, and hence directs the optimisation process.
Load error (E_{ld})	This is an error associated to the magnitude of the load out of the tank	Helps in computing the total error, and hence directs the optimisation process.
Volumetric Error (E_{lm})	This is an error associated to the level of the liquid in the tank	Helps in computing the total error, and hence directs the optimisation process.
Flow rate change error (E_s)	This is an error associated with the rate at which the liquid is pumped out of the tank	Helps in computing the total error, and hence directs the optimisation process.

Initially, the effluent flow for all intervals considered was taken as the mean influent flow. Optimisation was then done (discussed in Section 3.2.2) to obtain the final (optimized) effluent flows.

To obtain the COD and all other tank concentrations at each interval, a similar formula to that used in obtaining COD was used (see Equation 86) as illustrated in Figure 8.

$$COD_{out\ at\ t} = \frac{V_{t-1} * COD_{out\ at\ t-1} + Q_{in} * COD_{in\ at\ t} * \Delta t}{V_t + Q_e * \Delta t} \qquad (86)$$

Where;

$COD_{out\ at\ t}$ = COD concentration at time t (mg/l)
V_{t-1} = Volume of liquid in balancing tank at time t-1 (Ml)
Q_{in} = Influent flow to balancing tank (m^3/d)
Δt = Change in time between tow intervals (hr)
Q_e = Effluent flow from balancing tank (m^3/d)

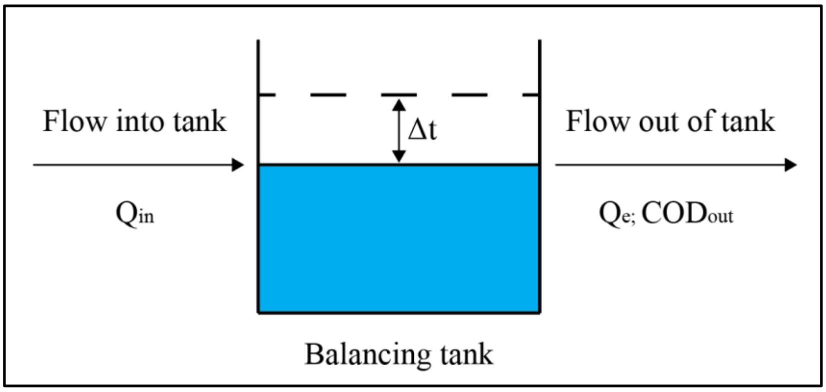

Figure 8: Flow into and out of balancing tank

The various tank concentrations were assumed to be zero at interval 0. The volume of the balancing tank at interval 0 was considered to be the max volume (given, or obtained from computed retention time), multiplied by the starting volume percentage. The volume of the tank for the next intervals was obtained by the difference between the flow into the tank and out of the tank during the time interval considered as shown in Equation 87.

$$V_t = V_{t-1} + \left(Q_{in\ at\ t} - \frac{Q_{out\ at\ t} + Q_{out\ at\ t-1}}{2}\right) * \Delta t \tag{87}$$

Where;
V_t = Balancing tank volume at time t (Ml)
V_{t-1} = Balancing tank volume at time t-1 (Ml)
$Q_{in\ at\ t}$ = Flow into the tank at time t (Ml/d)
$Q_{out\ at\ t}$ = Flow out of the tank at time t (Ml/d)
$Q_{out\ at\ t-1}$ = Flow out of the tank at time t-1 (Ml/d)
Δt = Time taken for an interval to elapse (hr)

Therefore, fractional volume of the balancing tank was obtained using Equation 88:

$$Fractional\ volume = \frac{Volume\ at\ current\ interval}{Max\ tank\ volume} \tag{88}$$

Where dV/dt represented the rate of change of the tank volume, considering the flow into and out of the tank over a specific time interval such that Equation 89 is achieved.

$$\frac{dv}{dt} = Q_{in\ at\ t} - \frac{(Q_{(out\ at\ t)} + Q_{(out\ at\ t-1)})}{2} \tag{89}$$

The load out of the tank at each interval was obtained from Equation 90:

$$Load\ out = Effluent\ flow * Tank\ concentration \tag{90}$$

The errors considered at each interval with respect to the outflow were the flow error (E_f), the load error (E_{ld}), the volumetric error (E_{lm}) and the rate of change of outflow error (E_s).

The expressions for E_f and E_{ld} are highlighted in Equations 81 and 82. The volumetric error was thus expressed as Equations 91 & 92.

Upper Limit;

$$E_{lm} = \beta \sum_{i=1}^{n-1} \{V_p - (V_{lu} - \delta v)\}^6\ For\ all\ i\ such\ that\ V_p > (V_{lu} - \delta v) \tag{91}$$

Lower Limit

$$E_{lm} = \beta \sum_{i=1}^{n-1} \{V_p - (V_{lb} + \delta v)\}^6\ For\ all\ i\ such\ that\ V_p < (V_{lb} + \delta v) \tag{92}$$

Where;
E_{lm} = Volumetric error
V_p = Equalization tank hold-up (% of total tank volume)
V_{lu} = Upper normal tank volume limit (% of total tank volume)
V_{lb} = Lower normal tank volume limit (% of total tank volume)
δv = Volume differential within the limits at which Elm attains a value (5% results in satisfactory behaviour)

For any rate of change in the outflow error, Equation 93 was used:

$$E_s = \omega \sum_{i=1}^{n} \left| \frac{dF_i}{dt} \right| \tag{93}$$

Where;
E_s = Penalty error for rate of change of tank outflow rate
ω = A weighting factor included so that the effect of E_s does not override the effect of E_e
F_i = Tank outflow rate (Ml/d)

t = time step (d)

At the end of the optimisation, the total error was computed. Therefore, the optimisation ensured that this total error moved towards its minimum value possible, in order to have a balanced flow-out of the balancing tank.

$$E_t = E_e + E_{lm} + E_s \tag{94}$$

3.2.2 Optimisation Process

Dold (1982) considered 2 types of optimization methods; i.e i) the method of steepest descent and ii) the fast convergence approach. Dold (1982) determined that the fast convergence approach is the more efficient optimisation approach. The methodology for the fast convergence approach and the method of steepest descent can be found in Dold (1982).

The method used in this research was a modification of the fast convergence method proposed by Dold (1982) (Section 3.2.1.1). This was made to ensure accurate optimisation at acceptable times. Before applying the optimisation method, the following were established;

- Influent data (flows and concentrations) at each interval.
- Parameters described in Section 3.2.1.2.
- Initial effluent flow at each interval, such that it was the main daily inflow.

Knowing these parameters, the optimisation methodology was as follows;

1. Two conditions were placed on the whole optimisation code to perform the optimisation routines until they were met. The first condition was for the code to run as long as the fractional volume at any interval was out of the range 0-1, and the second was for the code to run as long as the outflow at any interval was greater than or less than a specified fraction of the influent flow.
2. There were 3 main optimisation routines that ran successively, the first was the volumetric error optimisation routine, the second was the flow rate change error optimisation routine and the third was the fractional volume optimisation routine.
3. The optimisation for the volumetric error routine was set to run from N= 2 to Ti-6 (where N was the interval under consideration and Ti was the total interval). It was realized that applying a change at the first and last two intervals cannot be done accurately with the code, and all intervals after Ti-6 (i.e Ti-5 to Ti-2) needed to be done separately, to allow smooth optimization.
4. The initial volumetric error was hereafter recorded.
5. A value of +Δ was added to the first interval effluent outflow considered (i.e N=2). A value of 0.5Δ was then added to the immediate adjacent intervals (in this case intervals 1 and 3). An equal but opposite value was added at the (N+1)+(0.5)*(Ti-N-1)th interval (-Δ), and a value of -0.5Δ was added to its immediate adjacent intervals to ensure that a mass balance was maintained.
6. The new volumetric error due to that change in effluent flow at the interval was hereafter recorded.
7. If the new error was less than the initial error, which implied that the optimisation was in the correct direction, hence +Δ was added in the same direction as previously described

and -Δ in the same direction and interval as previously described, until one of the following conditions was reached; i) the error after adding +Δ was bigger than the previous error or ii) the effluent flow became less than 80% of the minimum daily inflow or iii) the effluent flow became greater than 120% of the maximum daily inflow (note that the 80% and 120% were not fixed values, and can be adjusted accordingly depending on how smooth the operator wants the flow to be).

8. At that specific moment no further optimisation in that direction was possible, and the code moves to the next interval. Before moving to the next interval, -Δ was applied at the interval (-0.5Δ at its immediate adjacent intervals) and +Δ at interval (N+1) + (0.5)*(Ti-N-1), (+0.5Δ at its adjacent intervals) to get to the least error, prior to the application of the second condition highlighted in Step 7.
9. In case the initial volumetric error (from Step 4) was smaller than the new volumetric error (Step 6), the Δ values applied were removed to obtain the initial effluent flow, as no optimisation can be done in that direction.
10. In case of Step 9, a value of -Δ was then added at the first interval (-0.5Δ added to its immediate adjacent intervals) with a value of +Δ being added at the (N+1) + (0.5)*(Ti-N-1)th interval (+0.5Δ added to its immediate adjacent intervals).
11. The new volumetric error value was then computed and compared to the initial volumetric error.
12. If the new error was less than the initial error, it implied that the optimisation was in the correct direction; hence, -Δ was added in the same direction as before and +Δ in the same direction and interval as before, till one of the following conditions was reached; i) the error after adding -Δ was bigger than the previous error or ii) the effluent flow became less than 80% of the minimum daily inflow or iii) the effluent flow becomes greater than 120% of the maximum daily inflow.
13. At that specific moment no more optimisation in that direction was possible, and the code moved to the next interval. Before moving to the next interval, +Δ was applied at the interval (+0.5Δ at its immediate adjacent intervals) and -Δ at interval (N+1) + (0.5)*(Ti-N-1), (+0.5Δ at its adjacent intervals) to get to the least error, just before the second condition in Step 12.
14. In case the initial volumetric error (from Step 4) was smaller than the new volumetric error (Step 11), the Δ values applied were removed to obtain the initial effluent flow, as no optimisation can be done in that direction.
15. For intervals Ti-5 to Ti-2, a similar approach was used, the only difference was in the method of applying a change to the outflow profile. If +Δ was applied at an interval, 0.5Δ was applied at the adjacent intervals, and -Δ is applied at interval Ti-1. The areas were still identical but in the opposite directions; hence, conservation of mass was determined to be preserved.
16. This code then ran to the last interval (N=Ti-2).
17. The same optimisation process described above for the volumetric error was used for the flow rate change error.

18. For the fractional volume optimisation, intervals N = 2 to Ti-6 were considered as before. A condition was placed to overarch this code, as long as the fractional volume at any interval was not within the range 0-1, the code ran.
19. A similar optimisation approach as to the first two optimisation routines was used, the only difference was at the level of the conditions to be respected.
20. The initial volumetric fraction at the first interval was then recorded.
21. A value of +Δ was applied at the first interval effluent flow (and 0.5 Δ was applied at the immediate adjacent intervals). An equal but opposite value was added at the (N+1)+(0.5)*(Ti-N-1)th interval (and -0.5 Δ was added to its immediate adjacent intervals) to ensure that the mass balance was maintained.
22. The new volumetric fraction was compared to 0 and 1. If it was less than 0 and greater than 1, the next step was performed.
23. The Δ values were applied in the same direction till all of the following conditions were met; the volumetric fraction value was between 0.05 and 0.95 (exclusive), the outflow at the interval was less than 120% of the maximum daily inflow and greater than 80% of the minimum daily inflow (Note that these values can be changed depending on how much the operator wants the tank to be stressed).
24. In case the new volumetric fraction was between 0 and 1 (exclusive), -Δ was applied at the first interval (and -0.5 Δ was applied at the immediate adjacent intervals) and +Δ was applied at interval (N+1)+(0.5)*(Ti-N-1) (and -0.5Δ was applied at the immediate adjacent intervals). This was done to move back to the initial volumetric fraction value.
25. Then the same process was performed with the opposite Δ values.
26. For intervals Ti-5 to Ti-2, a similar approach was used, the only difference was in the method of applying a change to the outflow profile. If +Δ was applied at an interval, 0.5Δ was applied at the adjacent intervals, and -Δ was applied at interval Ti-1. The areas were still identical but in the opposite directions; hence, conservation of mass was preserved.
27. These 3 optimisation routines were performed in a loop till the initial conditions in Step 1 were met.

Figure 9 illustrates adjustments to the outflow profiles generated.

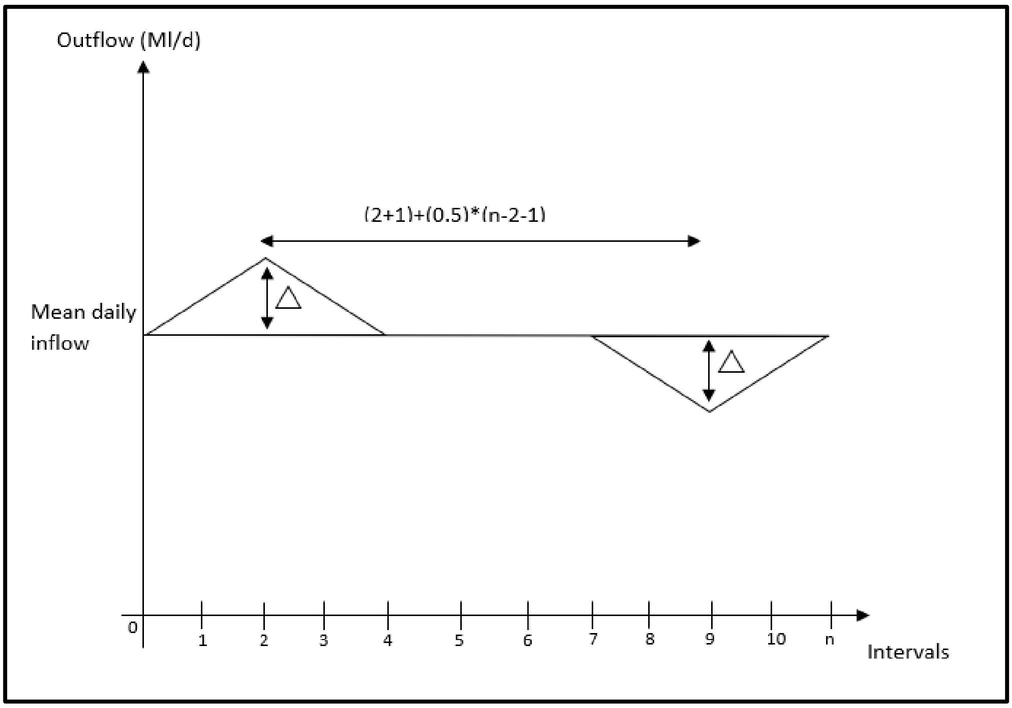

Figure 9: Adjustment to the outflow profile

Step 2: Mass Balance Check

To ensure the tank was internally consistent (i.e., respected conservation of masses and flows), a COD mass balance and a water balance over the tank were performed at the end of the optimisation. As for the influent flow, the Simpson's rule was applied on the whole data set (output values at every interval) to obtain the volume of water flowing out of the tank and being conserved in the tank, and also to compute the mass of COD leaving the tank and the mass staying in the tank (Only the flow and COD were considered over the tank, because the main focus of the tank was on flow equalization). Equations 95 & 95 indicate the balances used for water & COD.

$$Water\ Balance = \frac{Volume_{out\ of\ tank} + Volume_{conserved\ in\ tank}}{Volume_{into\ tank}} x100 \tag{95}$$

$$COD\ mass\ Balance = \frac{COD_{Flux\ out\ of\ tank} + COD_{Flux\ conserved\ in\ tank}}{COD\ _{Flux\ into\ tank}} x100 \tag{96}$$

3.3 Step 3: Model Check

The model was checked via scenario analyses. The impact of the inclusion of an equalization tank was assessed in an MLE system at steady state (using Microsoft Excel) and at dynamic state (using WEST® Environment).

4. Results and Discussion

This chapter reports on the performance of the equalization tank and two (n = 2) scenarios that were assessed to determine the potential benefits of including an equalization tank in an MLE system.

The first scenario considered steady state, to obtain the best design unit process sizes (smallest for optimum operation) and eventually the capital costs with and without an equalization tank. These were then compared to determine a cost-effective option.

The second scenario considered a dynamic state. The unit processes sizes obtained from the first scenario were then used, and under dynamic simulation, with the performance of various key unit processes in the plant being assessed and compared with and without an equalization tank.

4.1 Performance of the equalization tank

The flow into the WWTP was the only parameter evaluated using the equalization tank in this study. Different initial tank volumes were tested to find the optimum and the following were obtained;

- Balanced influent flow with 40% initial equalization tank volume
- Balanced influent flow with 60% initial equalization tank volume
- Balanced influent flow with 70% initial equalization tank volume

These specific volumes were those successfully modelled by the VBA model. The chemical and biological components were assumed to be completely mixed and similar influent concentrations were fed to the system, whether the flow was balanced or not.

The effect of the equalization tank can be appreciated when looking at the standard deviation of the flow data for both the unbalanced and balanced flow. A summary (Table 7) highlights standard deviation for a 7-day flow data set for the various scenarios.

Table 7: Standard deviation of various influent flows

Flow	Initial volume	Standard deviation
Unbalanced	100%	6.6
Balanced	40%	3.9
Balanced	60%	3.3
Balanced	70%	3.3

As the flow variability is significantly reduced, the operation was deemed better under such circumstances. Figure 10 provides a visual representation of the impact of the equalization tank on the flow.

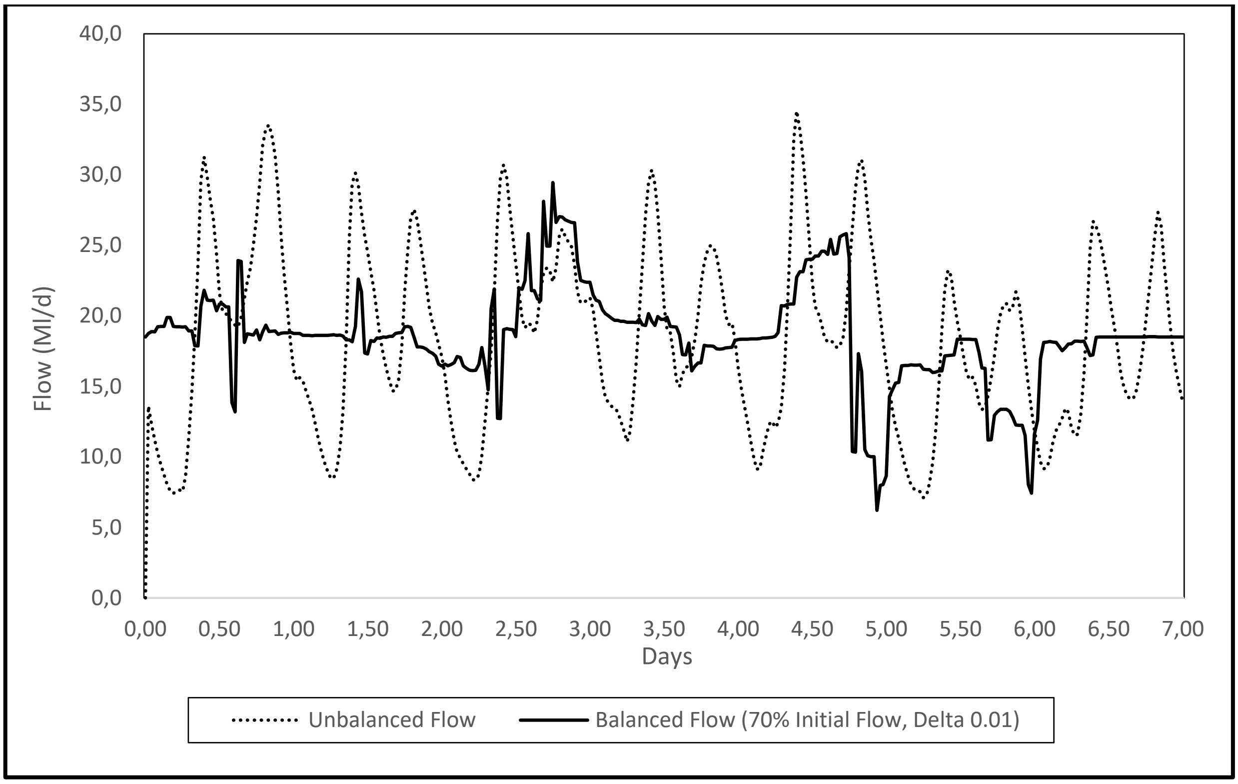

Figure 10:Performance of equalization tank

4.2 Scenario One

4.2.1 Introduction

This scenario involved the comparison of the capital cost, unit process sizes and total footprint of a balanced sludge age MLE system with and without a flow equalization tank. The flows used are random 1-day BSM2 flows as well as their characteristics. All parameters used can be found in Appendix C (Ekama, 2018). The equalization tank model was developed in Visual Basic (VBA) (Appendix B). The equalization tank used for this scenario was the 70% initial volume balancing tank as it had the best flow buffering as seen in Table 7.

Two plants were designed, one with 8-unit processes (P_1, excluding equalization tank) and one with 9-unit processes (P_2, including equalization tank designed in this project) which will be called P2. The 8-unit processes were similar for both plants and had the following units:

- Primary settling tank.
- AS reactor including nitrification and denitrification in a balanced sludge retention time (SRT) MLE system.
- Secondary settling tank.
- Gravity thickening of primary sludge (PS).
- Anaerobic digestion for stabilizing PS.
- Anoxic-Aerobic digestion for stabilizing WAS.
- Aeration in the anoxic-aerobic digester and activated sludge systems.

4.2.2 Objectives

The objective of this scenario analysis was to:

- Discuss the advantages and disadvantages of including an equalization tank to a plant without one, with respect to the capital cost, unit process sizes, effluent quality and ease of operation.

4.2.3 Phases layout

4.2.3.1 P1 (unbalanced flow design)

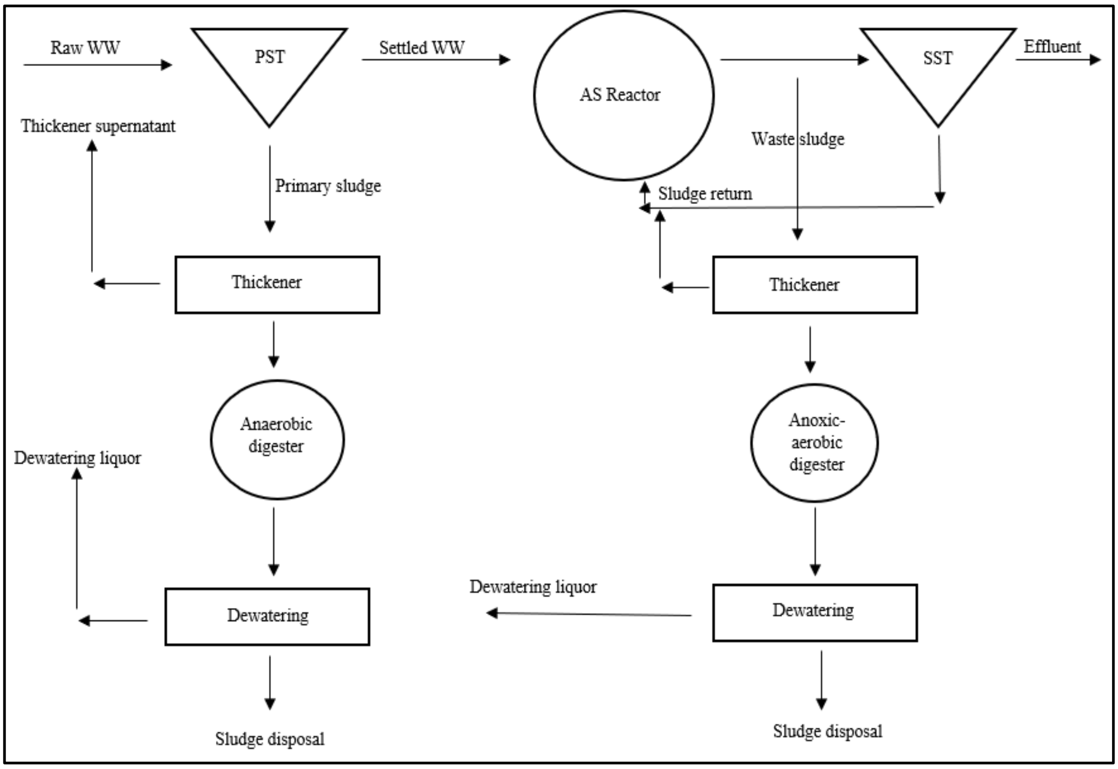

Figure 11: P1 WWTP Layout

4.2.3.2 P2 (balanced flow design)

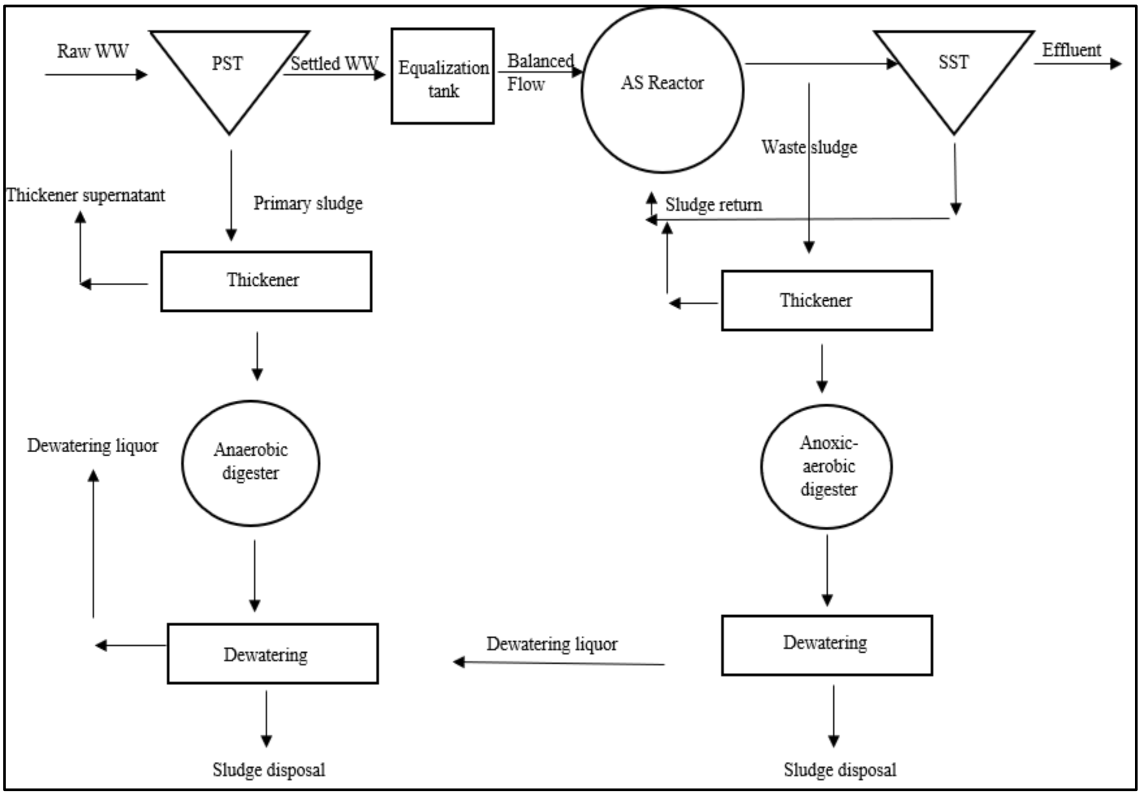

Figure 12: P2 WWTP Layout

4.2.4 Design limitations

The limitations of this scenario analysis were;

- The running costs which were not considered in the decision making of this design due to lack of detail to do so.

4.3 Scenario One: Results and discussion

4.3.1 Wastewater inflow

A visual depiction of the settled wastewater diurnal flow for the balanced and unbalanced flow can be seen in Figure 13.

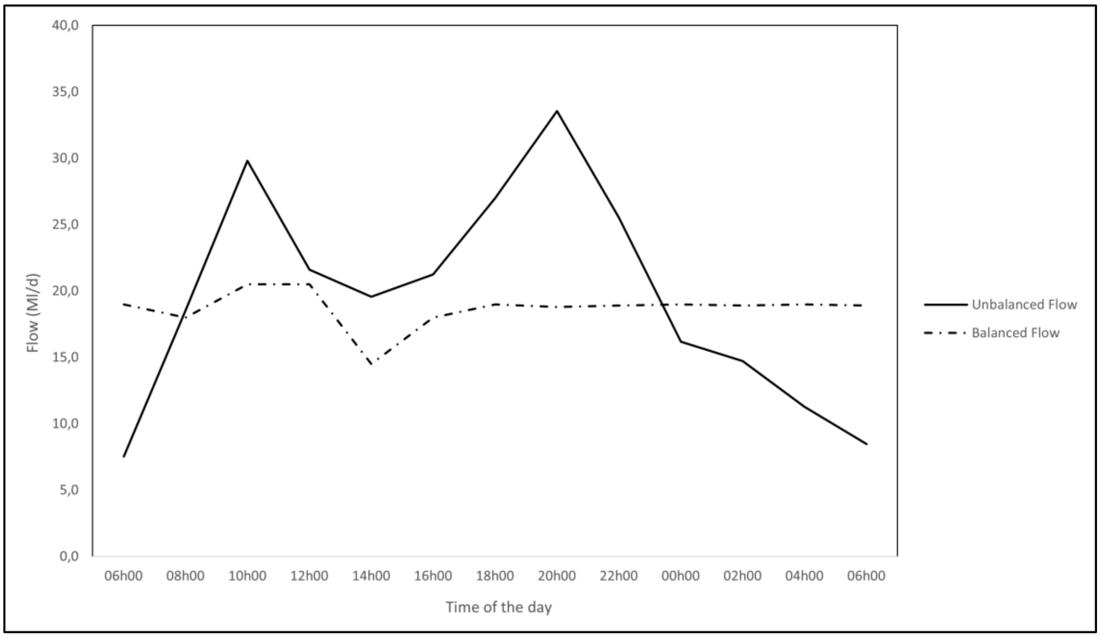

Figure 13: Diurnal Flow Variation Form

The settled wastewater and primary sludge (PS) were completely characterised and are presented in a summary table (see Appendix D**)**. Note that all the values computed in the characterisation were used for the design of these plants.

4.3.2 Unbalanced flow (P1) system design

4.3.2.1 Primary sedimentation

While considering the overflow rate at an average dry weather flow (ADWF) and peak wet weather flow (PWWF), a single PST was designed, of diameter 32.2m at the critical overflow rate of 2.4m/h for PWWF. A summary table for the design can be found in Appendix D.

4.3.2.2 AS reactor

The AS reactor designed was an MLE system with nitrification and denitrification (but excluding P removal). The design was done at 11.2days (balanced sludge age) and 16°C, which is the minimum average winter temperature and could thus result in the largest reactor volumes.

The AS reactor had one module of 7345m^3 in volume and a reactor TSS concentration of 4.9gTSS/l. The number of reactor modules was identical to that of the SST module to ensure optimal cost and design.

Due to the relatively high TKN/COD ratio of 0.12, the a_{opt} computed was 5.96 as this was less than the practical a-recycle of 6:1 resulting in the anoxic reactor slightly loaded less than its denitrification potential. This recycle ratio was kept for denitrification, which resulted in a 95.1% N removal and an effluent alkalinity of 78.84mg/l as $CaCO_3$; hence, no dosage of lime will be required, as the pH will stay above **7**.

To maintain the sludge age of 7.8 days, a daily waste flow of 0.66Ml/d is required, preferably wasted directly from the reactor as it facilitates control of sludge age and is required to avoid possible sludge rise in the SST. This waste flow amounts to a WAS flux of 3215kgTSS/d.

A summary table of the design components of the P1 AS reactor can be found in Appendix D.

4.3.2.3 Secondary sedimentation

A single SST of area 1051m^2 (37m diameter) was designed for the PWWF (932m^3/h). At ADWF, peak dry weather flow (PDWF) and PWWF, the s-recycle can be operated at 1:1 with respect to the ADWF. A summary table of this design can be found in Appendix D.

4.3.2.4 Gravity thickening of primary sludge

To be able to thicken the primary sludge (PS) to a maximum TSS of 60gTSS/l, a gravity thickener was sized, for a solids' loading rate of 80kg/m2, inflow of 183.3m3/d and sludge mass of 8449kgTSS/d which is the PS TSS flux. An area of 177m2 (15m diameter) was obtained. Having a waste flow of 141m3/d and a supernatant flow of 42m3/d. A summary table of the design can be found in Appendix D.

4.3.2.5 Dissolved air flotation (DAF) unit for WAS thickening

To ensure that the WAS is thickened in the absence of anaerobic conditions (which can exist in gravity thickeners), which can result in the release of P during the thickening and make the WAS unsuitable for wastage on drying beds, a DAF unit was designed to thicken the WAS to 60 gTSS/l. The solids' loading rate used was 34.5 kgTSS/m^2/d and the inflow was 656m^3/d, resulting in a DAF unit of area 93m^2 (11m diameter). The waste flow from the DAF unit was 54m^3/d and the supernatant flow was 603m^3/d. A summary table of the design can be found in Appendix D.

4.3.2.6 Anaerobic digestion for stabilizing PS

The supernatant flows were calculated for the DAF and gravity thickener units, but for the PS AD, it was assumed that the PS flow and concentrations are similar as those going into the AD, as the supernatant flows were deemed negligible.

To reduce the amount of primary sludge wasted by converting its biodegradable component into useful gas (methane), an anaerobic digester was designed. This digester had a volume of 1833m^3 and removed 52% TSS (62% VSS) and 63% COD, with this COD being converted to methane (2499m^3/d, 91.7% of COD utilized converted to methane) and CO_2 (1953 m^3/d). The digester pH was 6.82, lower than the optimum pH for the growth of methanogens (7-8) which may result in the digester failure with minimal breakdown of COD. To ensure that the pH stayed in that range, alkalinity must be added in the feed to the digester. The quantity is dependent on the anticipated organic acid production capacity of the sludge feed.

The potential of energy generation of the digester from methane was investigated and it was found that 469kW could be generated by the PS AD (including the efficiency), which meant that it could fully power the MLE AS aeration requirements of P1. The power requirements to run the AD itself was estimated at 220kW (including heat loss and power required to increase sludge temperature). This implied that the AD power generated can be used to be self-sustaining and 100% of the MLE AS aeration requirements of P1.

The AD is followed by a dewatering unit that thickens the sludge to 200 gTSS/l. A summary table of the design can be found in Appendix D.

4.3.2.7 Anoxic-aerobic digestion for stabilizing WAS

Flow-through anoxic-aerobic digesters were designed so the WAS specific dissolved oxygen utilization rate (SOUR) (including nitrification) was 1.5gO/(gVSS.h) for a stable WAS and maximum oxygen utilization rate (OUR) of 250mgO/l.h, to allow the disposal of the sludge onto drying beds. Digesters (n = 2) in series were found to be the best option with a total retention time of 19.8 days. A total volume of 1060m^3 (530m^3 for one) was provided for both digesters. The total oxygen demand required was 2012kgO/d. The flux of WAS discharged to the drying beds was completed as 1982kgTSS/d at a flow of 54m^3/d, with a VSS/TSS ratio of 0.69.

A summary table of the design can be found in Appendix D.

4.3.2.8 Aeration in the anoxic-aerobic digester and activated sludge systems

The peak OUR of the AS was obtained at 24°C and was 133mgO/l.h for the whole reactor. The AS reactor was compartmentalized (n = 4) and the cheapest combination for the computed aeration requirement for each compartment was provided (including efficiency) as follows:

- Compartment 1: Required-30kW Provided-36kW (1 unit of 40kW)
- Compartment 2: Required-23kW Provided-36kW (1 unit of 40kW)
- Compartment 3: Required-19kW Provided-36kW (1 unit of 40kW)
- Compartment 4: Required-19kW Provided-36kW (1 unit of 40kW)

The peak OUR of the anoxic-aerobic digester was obtained at 16°C to be 102.1mgO/l.h for the first digester and 32.7mgO/l.h for the second digester. The cheapest combination for the computed aeration requirement for each digester was provided (including efficiency) as follows:

- Digester 1: Required-55kW Provided-58.5kW (1 unit of 65kW)
- Digester 2: Required-19kW Provided-36kW (1 unit of 40kW)

A summary table of the design can be found in Appendix D.

4.3.3 Balanced flow (P2) system design

4.3.3.1 Primary sedimentation

The same inflow conditions as for P1 were considered, a PST (n = 1) being designed, of diameter 35.1m at the critical overflow rate of 2.4m/h for PDWF. This PST was placed before the equalization tank and was allowed for settled water to be used for the AS design of this system. A summary table for the design can be found in Appendix D.

4.3.3.2 Flow balancing

To ensure the P2 design was operated efficiently with minimal fluctuations of concentrations and flows and consequently aeration fluctuations (not easily dealt with, resulting in operation efficiencies), an equalization tank of volume 4.63Ml and retention time of 6 days, and providing a mean daily inflow of 18.7Ml/d (9% decrease from P1 of 20.3Ml/d) was designed. This was to accommodate short-term wet weather conditions. Details of the equalization tank can be seen in Appendix B.

4.3.3.3 AS reactor

The AS reactor had one module of 6752m^3 (which is an 8% decrease in the AS reactor volume compared to P1 without an equalization tank) with a reactor TSS concentration of 4.9gTSS/l. The number of reactor modules was identical to that of the SST module to ensure optimal cost and design. From the stoichiometric equations defined in Section 2, it was observed that as the inflow into the AS reactor decreases due to the equalization tank, there is less organic matter present; hence, there is less biomass, endogenous residue and inorganic particulates resulting in less total solids in the reactor and consequently a smaller volume (as the TSS concentration stays the same); an assertion estimated using Equation 97.

$$V_r = \frac{MX_t}{X_t} \qquad (97)$$

Where;

V_r = Reactor volume (l)

MX_t = Mass of total suspended solids in the reactor (gTSS)

X_t = Concentration of total suspended solids in the reactor (gTSS/l)

The a_{opt}, Nitrogen removal, effluent alkalinity and pH were identical to those of P1 as the inflow had no effect on these parameters.

To maintain the sludge age of 7.8 days, a daily waste flow of 0.6Ml/d (9% decrease from P1) was required. Preferably, wasted directly from the reactor as it facilitates control of sludge age and is required to avoid possible sludge rise in the SST. This waste flow amounted to a WAS flux of 2954kgTSS/d (8% decrease from P1). These differences were present due to the 8% decrease in the AS reactor volume.

A summary table of the design components of the P2 AS reactor can be found in Appendix D.

4.3.3.4 Secondary sedimentation

A single SST (n = 1) of area 966m^2 (35m diameter) (8% decrease from P1) was designed for PWWF (857m^3/h), an 8% decrease from P1 design. For ADWF, PDWF and PWWF, the s-recycle can be operated at 1:1 with respect to the ADWF. These differences were caused by the 9% decrease in mean inflow due to the equalization tank. A summary table of this design can be found in Appendix D.

4.3.3.5 Gravity thickening of primary sludge

The gravity thickener for P2 was identical to that of P1, because its inflow is from the PST underflow, which is before the equalization tank.

4.3.3.6 Dissolved air flotation (DAF) unit for WAS thickening

A DAF unit was designed to thicken the WAS to 60 gTSS/l. The solids loading rate was 34.5 kgTSS/m^2/d and the inflow was 603m^3/d, an 8% decrease from P1design resulting in a DAF unit of area 86m^2 (10m diameter), which was a reduction of 7.5% from P1. The waste flow from the DAF unit was 49m^3/d, also 9% smaller than that of P1 albeit with the supernatant flow of 554m^3/d (8% difference from P1). A summary table of the design can be found in Appendix D. Due to the reduced AS reactor volume and consequently lower waste flow, the DAF unit equally received less inflow resulting in a smaller area unit.

4.3.3.7 Anaerobic digestion for stabilizing PS

As the equalization tank is placed after the PST, the AD design of P2 and P1 were identical and unaffected by the flow balancing. A summary table of the design can be found in Appendix D.

If the equalization tank is placed before the PST, it can be noticed that the methane produced reduces by 24% from 2498m^3/d to 1893m^3/d due to less COD in the influent (8333kgCOD/d for P2 and 10550kgCOD/d for P1).

4.3.3.8 Anoxic-aerobic digestion for stabilizing WAS

Two digesters in series were found to be the best option with a total retention time of 19.8 days. A total volume of 974m^3 (487m^3 for one) representing an 8% decrease from P1, was provided for both digesters. The total oxygen demand required was 1576kgO/d and was a 22% decrease from the P1 design. The flux of WAS discharged to the drying beds was 1821kgTSS/d (8% decrease from P1) at a flow of 49m^3/d (9% decrease from P1), with a VSS/TSS ratio of 0.69. Due to less biomass in the AS reactor (hence less TSS), less WAS was estimated to be wasted in P2 compared to P1; hence, resulting in smaller anoxic-aerobic digesters with an oxygen demand reduction of 22%.

A summary table of the design can be found in Appendix D.

4.3.3.9 Aeration in the anoxic-aerobic digester and activated sludge systems

The peak OUR of the AS was obtained at 24°C and was 105mgO/l.h for the whole reactor. Due to the presence of the flow equalization tank, this peak OUR was similar to the average OUR, as there was minimal diurnal flow fluctuation experienced by the system. The AS reactor was compartmentalized into 4 and aeration requirement for each compartment was provided (including efficiency) as follows:

- Compartment 1: Required-22kW Provided-36kW (1 unit of 40kW)
- Compartment 2: Required-17kW Provided-36kW (1 unit of 40kW)
- Compartment 3: Required-14kW Provided-36kW (1 unit of 40kW)
- Compartment 4: Required-14kW Provided-36kW (1 unit of 40kW)

The aeration requirements in each compartment was approximated as 26% lower than those from the P1 system, due to the equalization tank which reduced the flow variations throughout the day. But the aeration provided was similar in each compartment as the minimum aeration components used for this purpose in this design were 40kW (Appendix C). This meant there was no cost savings in the aeration operations as the difference was insignificant.

The peak OUR of the anoxic-aerobic digester was obtained at 16°C to be 102.1mgO/l.h for the 1st digester and 32.7mgO/l.h for the second digester. The aeration requirement for each digester was provided (including efficiency) as follows:

- Digester 1: Required-53.1kW Provided-58.5kW (1 unit of 65kW)
- Digester 2: Required-17kW Provided-36kW (1 unit of 40kW)

Although the amount of WAS in P2 was lower than that of P1, their aeration requirements for the digesters were similar, because the amount of biodegradable COD moving to the WAS in both systems was determined to be similar.

A summary table of the design can be found in Appendix D.

4.4 Scenario One: Cost, footprint, and effluent quality evaluations

4.4.1 Cost analysis

Table 8 lists costs of P1 and P2.

Table 8: Cost analysis of P1 in comparison to P2

Unit Operation	P1	P2	Difference (%)	Unit
Equalization Tank	0	1390	N/A	*1000Rand
PST	4028	3515	13	*1000Rand
SST	2356	2237	5	*1000Rand
Gravity Thickener	799	799	0	*1000Rand
Biological Reactor	3513	3294	6	*1000Rand
Mechanical Aeration AS	1217	1217	0	*1000Rand
Mechanical Aeration Anoxic-Aerobic Digester	702	702	0	*1000Rand
Anoxic-Aerobic Digester Volume	525	489	7	*1000Rand
Anaerobic Digester PS	2511	2511	0	*1000Rand
Flotation Unit	1841	1761	4	*1000Rand
Total	17492	17915	2.4	*1000Rand

The costs of most unit processes decrease or are constant with the presence of an equalization tank. The inclusion of an equalization tank after considering the reduction in sizes of the other unit processes resulted in a 2.4% increase in capital cost. Although there was a slight reduction in aeration due to the equalization tank, the aeration costs were identical for P1 and P2. This was because the minimal commercial sizes of the aeration units were 40kW.

The operational costs of running the plant were not computed, but due to numerous reductions in various parameters as can be seen in Table 9, the positive effects (reduction in operational costs and operation process) of the inclusion of an equalization tank can possibly outweigh its initial capital cost during long term use. Mikola (2013), reporting on a research study from a Finnish plant, opined that including an equalization process can decrease sludge production by 5-10%, resulting in savings of 10 000 – 20 000 € (R175000 – R330000, as of February 2020) in sludge treatment yearly. Hence, future research on the reductions of operational costs with the inclusion of an equalization tank should be carried-out to determine the number of years needed to recover the capital cost of the equalization tank.

Table 9: Improvements of MLE system with equalization tank

Parameter	P1	P2	Difference (%)	Unit
Mean daily inflow	20.3	18.7	7.9	Ml/d
Daily waste flow	0.66	0.6	9.1	Ml/d
WAS flux	3215	2954	8.1	kgTSS/d
Waste flow from flotation unit	54	49	9.3	m3/d
Total oxygen demand in anoxic-aerobic digester	2012	1576	22	kgO/d
Flux discharged to drying beds	1982	1821	8	kgTSS/d
Discharge flow from anoxic-aerobic digester	54	49	9	m3/d
Aeration in AS reactors (required)	91	67	26	kW
Aeration in AS reactors (provided)	160	160	0	kW
Aeration in anoxic-aerobic digester (required)	74	70.1	5	kW
Aeration in anoxic-aerobic digester (provided)	105	105	0	kW

4.4.2 Footprint analysis

Table 10 lists the surface areas and volumes of P1 and P2 unit processes.

Table 10: Volumes and Surface Areas of Unit Processes

Unit Operation	P1	P2	Difference (%)	Unit
Flow Balancing	0	4630	N/A	m^3
PST	812	649	20	m^2
SST	1052	966	8	m^2
Gravity Thickener	177	177	0	m^2
Biological Reactor	7348	6752	8	m^3
Anoxic-Aerobic Digester	1060	974	8	m^3
Anaerobic Digester PS	1833	1833	0	m^3
Flotation Unit	93	86	8	m^2

To obtain the footprint of each plant (P1 & P2), the surface areas of each unit process must be obtained. The depth of the digesters and the equalization tank was taken as 4m, and the depth of the biological reactor was taken as 3m, giving us the total areas listed in Table 11.

Table 11: P1 & P2 Footprints

Unit Operation	P1	P2	Difference (%)	Unit
Flow Balancing	0	1158	N/A	m^2
PST	812	649	20	m^2
SST	1052	966	8	m^2
Gravity Thickener	177	177	0	m^2
Biological Reactor	2449	2251	8	m^2
Anoxic-Aerobic Digester	265	244	8	m^2
Anaerobic Digester PS	458	458	0	m^2
Flotation Unit	93	86	8	m^2
Total	5306	5988	13	m^2

It was observed that in a similar pattern to the costs, the sizes of the unit processes all decreased or were equivalent when adding an equalization tank. The inclusion of the equalization tank increased the footprint by 13%.

4.4.3 Effluent quality

Table 12 enlists the effluent quality from each design phase.

Table 12: Effluent quality summary table

Effluent parameter	P1	P2	Units
Suse	57.50	57.50	mgCOD/l
Cuse	19.18	19.18	mgC/l
Nouse	1.00	1.00	mgN/l
Nae	2.52	2.52	mgN/l
Nne	7.11	7.11	mgNO_3-N/l
Pouse	0.00	0.00	mgP/l
Pae	13.72	13.72	mgP/l
COD	57.50	57.50	mgCOD/l
TKN	3.52	3.52	mgTKN/l
TP	13.72	13.72	mgTP/l
TOC	19.18	19.18	mgTOC/l
TN	10.62	10.62	mgN/l

Overall, the effluent quality from both systems was identical. This was because of a number of reasons:

- P1 and P2 are both similar MLE systems, with P2 only having an equalization tank as an extra unit.
- The equalization tank modelled equalizes the flow only, and assumes the tank is completely mixed; hence, P1 and P2 had similar loading rates by the end of the day; however, with different outflow profiles.
- The model used for this scenario is a steady state model; hence, all biological processes are assumed to be fully completed, which is impractical in municipal WWTP.

The effluent quality of both systems fall below those required by the South African wastewater effluent standards general limits and were hence deemed satisfactory, except the Ortho-Phosphates (13.72mgP/l) that are higher than the general limits (10mg/l).

Hence, if only the effluent quality is to be considered, there were minimal differences between both systems, evaluated.

4.5 Scenario One: Summary

The main aspects that were assessed from this scenario were the differences in capital cost, size of plants (footprint) and effluent quality.

The capital cost of the P2 system (R17 915 000) is 2.4% higher than that of the P1 system (R17 492 000). The operational costs of the plant as well as the ease of operation are benefits of including the equalization tank that need further assessment in future research to provide a more compelling argument in the inclusion of an equalization tank in a WWTP (from a financial point of view).

From the resources used in this research, including an equalization tank to a system does provide benefits as there was a reduction in unit size processes, aeration demand, sludge production.

4.6 Scenario Two

4.6.1 Introduction

This scenario reports on the comparison of the plant performance of the MLE system designed in Scenario One with and without a flow equalization tank. The only difference was that the dewatering liquor was not recycled to the influent. The parameters used in this system can be found in Appendix E. A steady state simulation of 100 days was run, followed by a 7-day dynamic simulation, both in the WEST® environment. The simulations were run using random 7-day BSM2 data (Nopens *et al.*, 2010).

For the unbalanced flow scenario, the 7-day BSM2 data was fed as the influent to the MLE system in the WEST® environment unaltered, and results obtained at different unit operations were collected. These results were then compared to those obtained when the flows fed as the influent to the MLE system were balanced (with equalization tank initially full to 40%, 60% and 70%).

4.6.2 Objectives

The objective of this scenario analysis was to:

- Discuss and compare the performance of the AS system with and without an equalization tank in the P2 design plant under a dynamic simulation.

4.6.3 Plant Layout

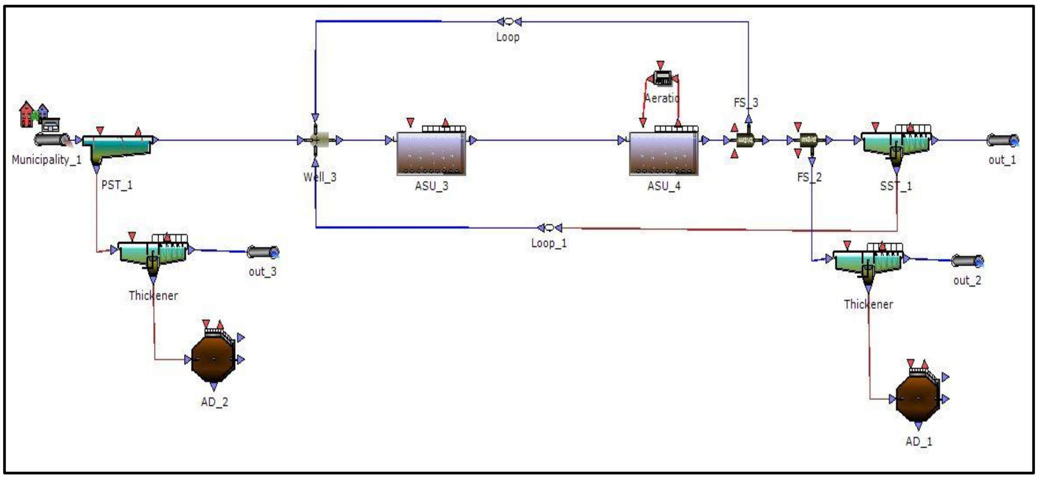

Figure 14: Plant Layout WEST® Environment

4.6.4 Limitations

The limitations of this scenario analysis were:

- Ease of plant operations after including an equalization tank which could not be assessed.
- The equalization tank model could not be coded into the WEST® environment; hence, to assess its effect on the plant, the equalized values from the VBA were used as influent input flow data points in the WEST® environment.
- The effect of the equalization tank was only assessed on an MLE system plant layout.

4.7 Performance of the AS system

The performance of various AS properties could be assessed (OUR, aeration energy, COD removal, nitrification, waste generated), by inputting balanced flows as the influent flow to the system. The AS results from the balanced flow of an equalization tank of 6hr retention time at various initial volumes (percentage of the tank that is full before balancing starts) are presented.

4.7.1 OUR

The magnitude of the OUR depends on various factors such as; the concentration of COD and total kjedahl number (TKN) in the influent to the system, the system volume, and the aerobic mass fraction (which is the fraction of the system sludge mass in the aerobic reactor) (Ikumi,2011). The cyclic nature of OUR in this case is a consequence of the cyclic nature of the unbalanced flow into the WWTP. Such a pattern causes the operation in WWTPs to be tedious, and if aeration is inadequate, this might result in poor WWTP performance as described in Section 2. The OUR profile only exists in the aerobic reactor, as the anoxic and anaerobic reactors do not require dissolved oxygen present. Figure 15 shows the OUR when unbalanced influent flows were used, for the system.

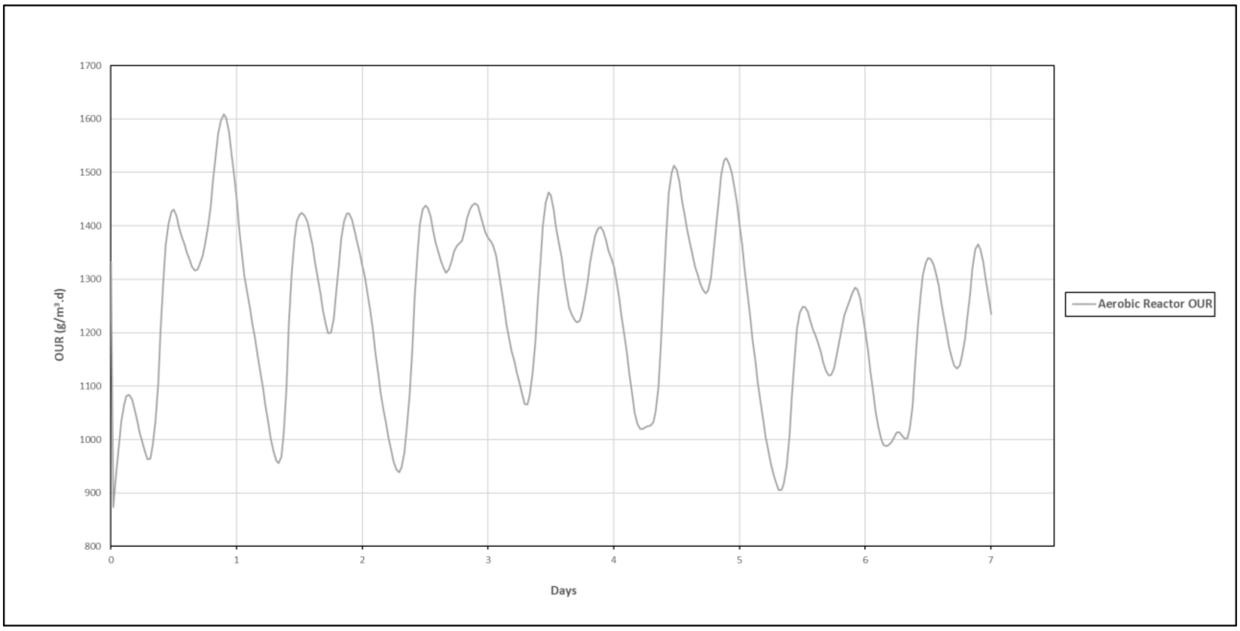

Figure 15:OUR for Unbalanced Influent Flow

After the flow was balanced, it was observed that the variability of the OUR significantly decreased. Figures 16 & 17 show the OUR when the flow was balanced at an initial tank volume of 40% and 70%, respectively.

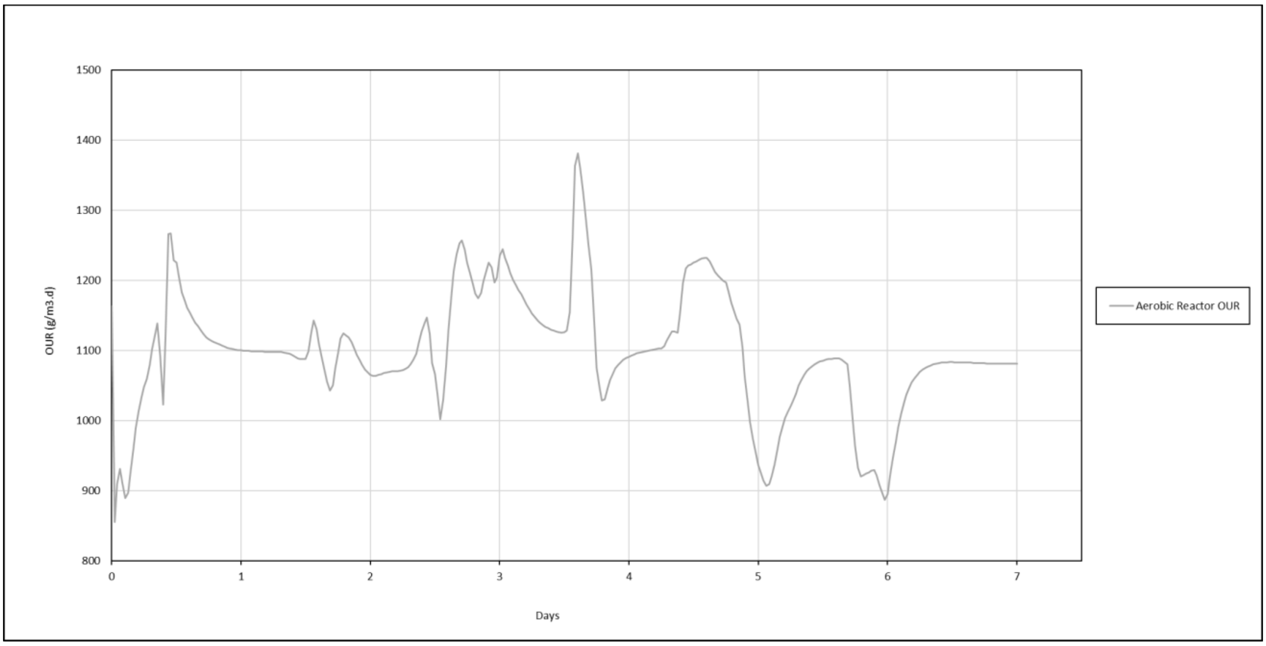

Figure 16:OUR for Balanced Influent Flow (40% Initial Volume)

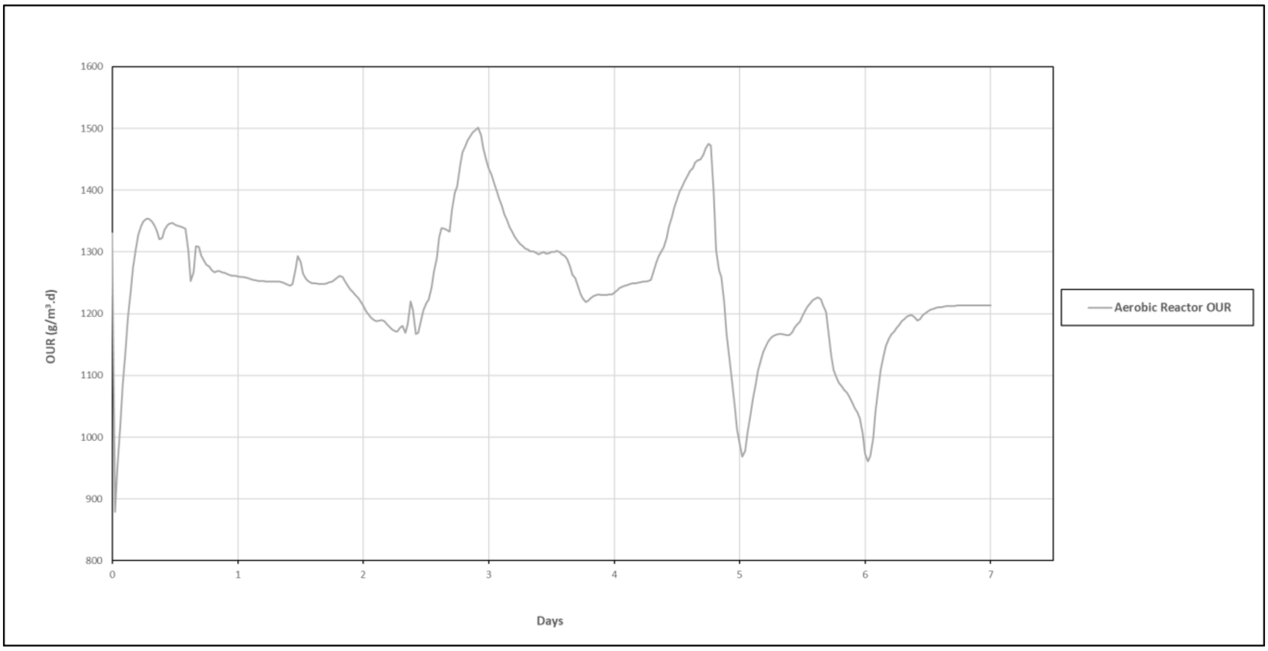

Figure 17:OUR for Balanced Influent Flow (70% Initial Volume)

The decreased variability in OUR was quantified using the standard deviation as illustrated in Figure 18.

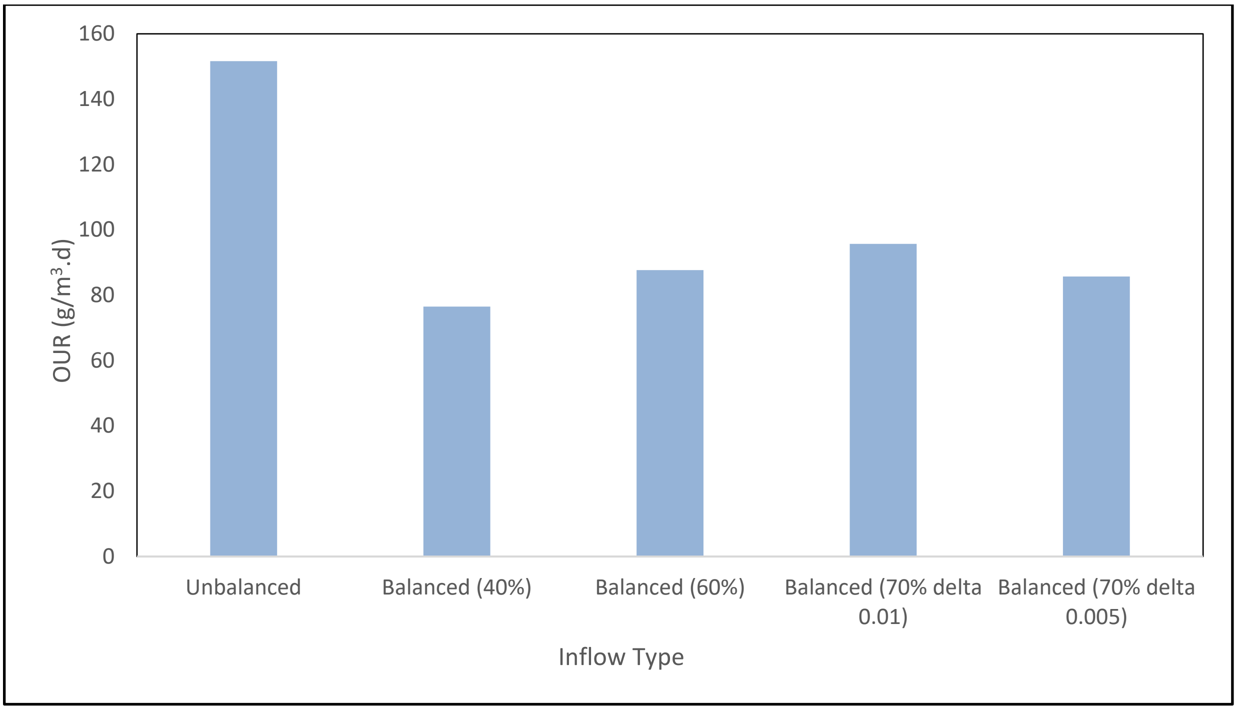

Figure 18:OUR Standard Deviation

In summary, the OUR variability is significantly reduced, by up to **50%** when the flow is balanced. It was therefore prudent to assume that this might result in better operation at plants, as operators only need to vary how much dissolved oxygen they provide through a smaller range.

4.7.2 Aeration energy

The aeration energy required in the aerobic reactor (to pump oxygen) was obtained from Equation 98.

$$Aeration\ Energy = Oxygen\ saturation\ concentration * Oxygen\ transfer\ coefficient * Reactor\ volume \qquad (98)$$

The aeration energy required for the unbalanced influent flow can be seen in Figure 19.

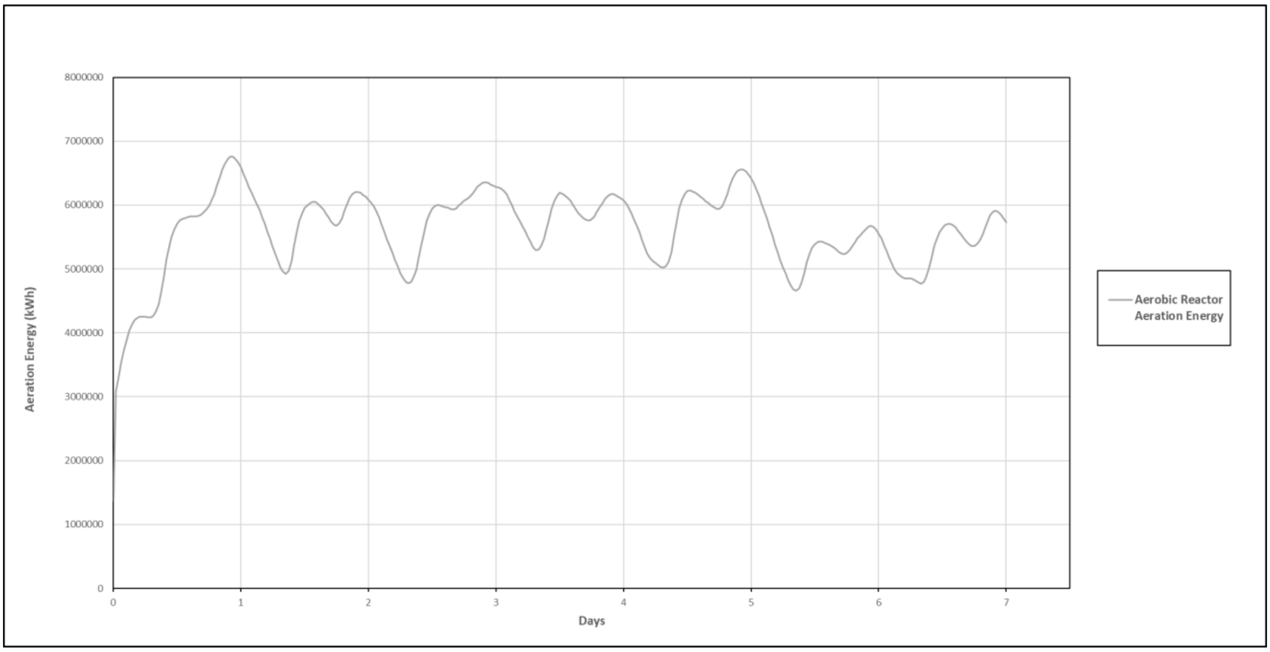

Figure 19:Aeration Energy for Unbalanced Influent Flow

The aeration energy profile varied unpredictably, which eventually renders the operation of plants complex. This can result in either over aeration or under aeration, which can have significant impacts on costs (increased if there is over aeration), and WWTP performance (poor if there is under aeration, as the organisms do not have sufficient dissolved oxygen to breakdown the biomass, resulting in poor effluent quality).

When the flow is balanced, the aeration energy increases in value, with its mean values shown in Table 13.

Table 13: Mean Aeration Energy values

Aeration Energy (kWh)		
Mean	**Status**	**Reactor**
5180305	Unbalanced	Aerobic
5201631	Balanced (40%)	Aerobic
5214640	Balanced (60%)	Aerobic
5211941	Balanced (70%)	Aerobic

The aeration energy increases (insignificantly, be less than 1%) because the mean oxygen to be provided throughout each day is increased, as the flow does not fluctuate as much; hence, more aeration energy on average is required (due to a higher mean COD and TKN mass into the reactor). This is a positive result because this aeration energy can be provided in a more efficient way in the plant avoiding potential poor performance of the plant that occurs when the flow varies more significantly. The variability of the aeration energy also reduces significantly when the influent flow is balanced, as can be seen in Table 14.

Table 14: Standard Deviation of Aeration Energy

Aeration Energy (kWh)		
Standard Deviation	**Status**	**Reactor**
587614	Unbalanced	Aerobic
427714	Balanced (40%)	Aerobic
458078	Balanced (60%)	Aerobic
462106	Balanced (70%)	Aerobic

The standard deviation showed that the variability of the aeration energy had been significantly buffered, which might help in making the operation of the WWTP better and more efficient. The aeration energy profiles when the influent flow is balanced can be seen in Figures 20 & 21.

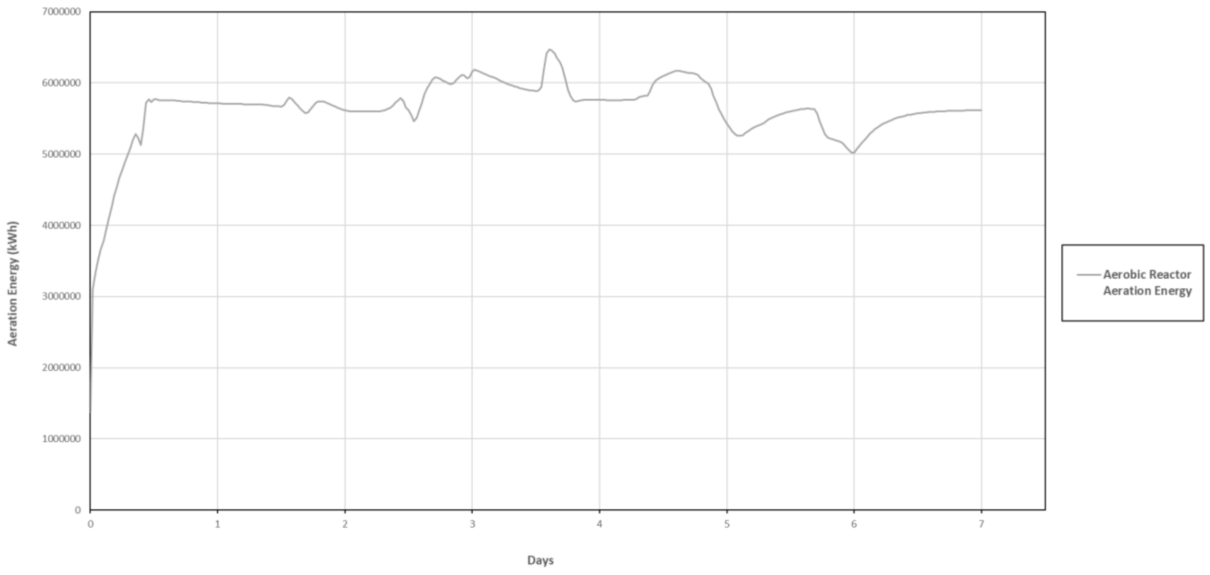

Figure 20:Aeration Energy Profile (Balanced Influent Flow 40% Initial Volume)

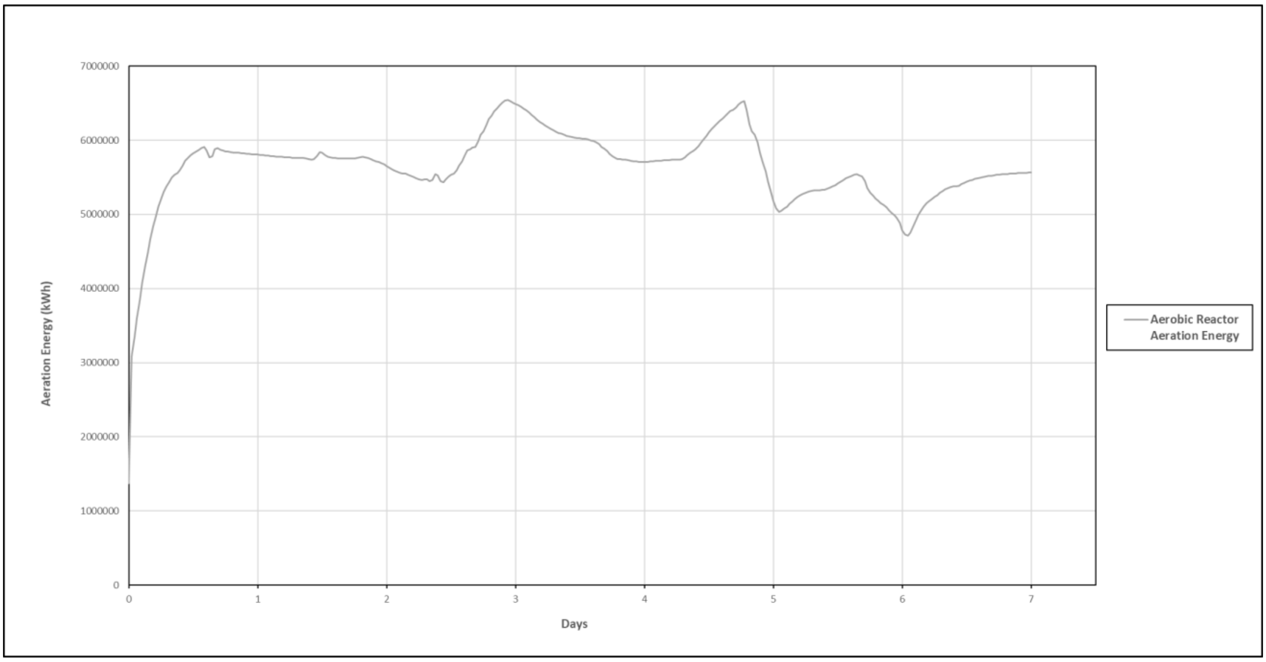

Figure 21:Aeration Energy (Balanced Influent Flow 70% Initial Volume)

In summary, the flow balancing causes the fluctuation in aeration energy throughout the modelling time to decrease in a similar way as the OUR, because the COD and TKN loads fed into the aerobic reactor are more constant than with the unbalanced flow, which brings in COD and TKN loads that vary largely in direct proportion to the inflow profiles. Although the aeration energy variation decreases, the mean aeration energy does not change significantly with the presence of an equalization tank; hence, the operation costs were observed not to decrease.

4.7.3 COD removal

The effluent COD from the reactors for the 7-day simulation are displayed in this section. For the unbalanced flow, the effluent COD varied (Figure 22). The reactors (n = 2) had similar concentrations due to the recycles in the system.

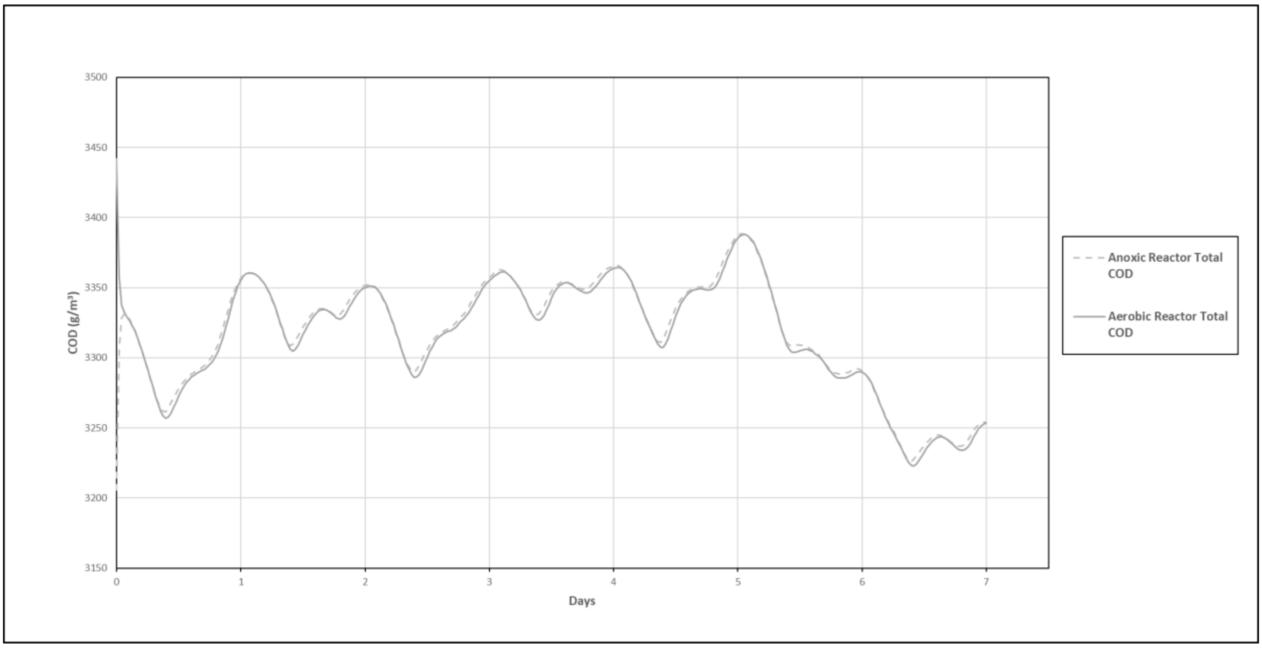

Figure 22:Total COD with Unbalanced Influent Flow

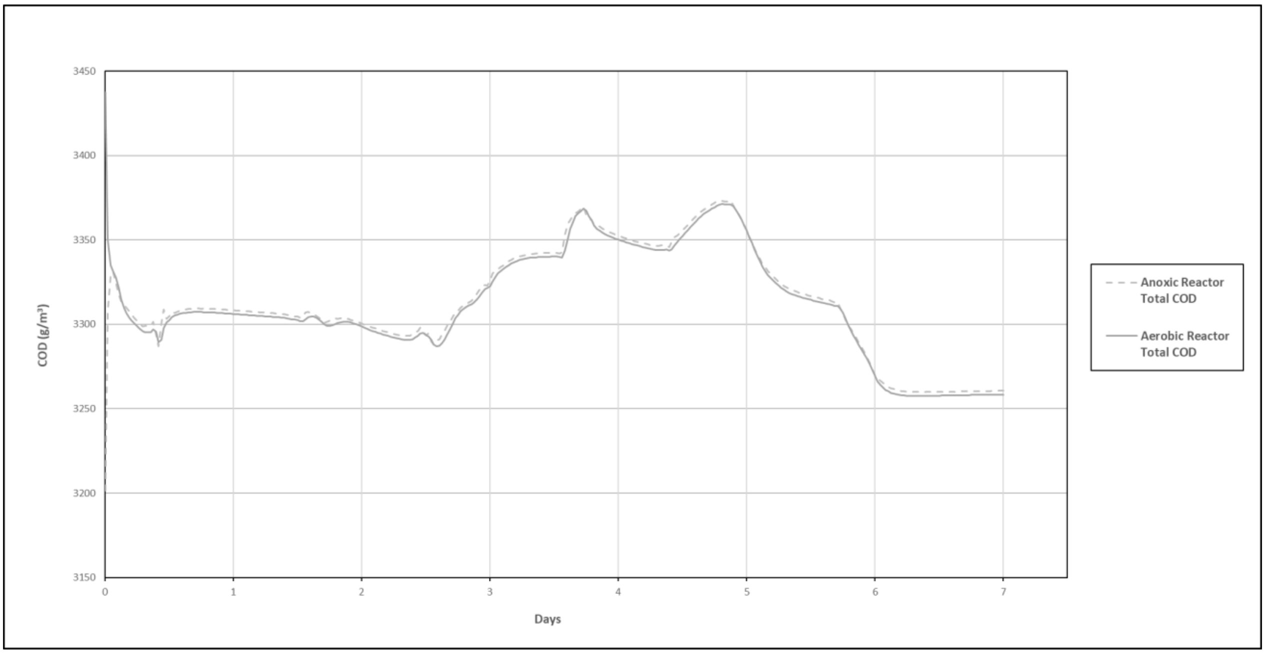

Figure 23:Total COD with Balanced Influent Flow (40% Initial Volume)

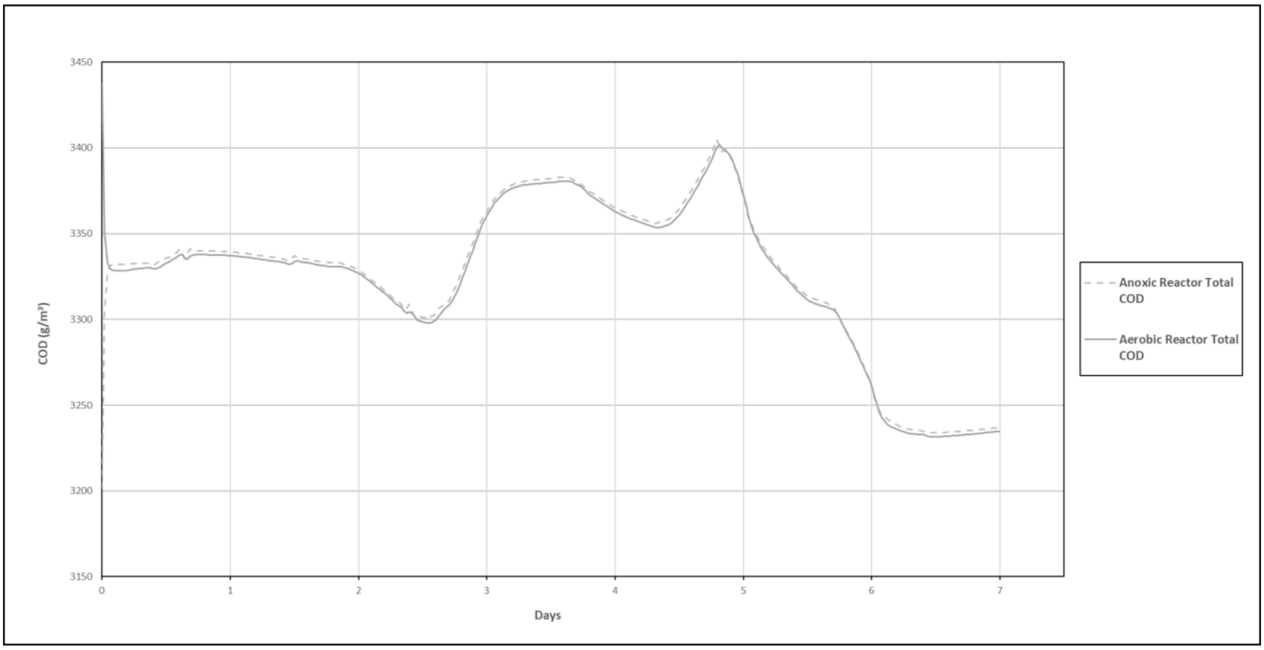

Figure 24:Total COD for Balanced Influent Flow (70% Initial Volume)

From Figures 23 and 24, it was observed that the concentrations of the COD after the flow was balanced were similar. The mean total COD (in the aerobic reactor) was 3317 g/m^3, for an unbalanced influent flow while it was 3312 g/m^3 for balanced influent flow (40% initial volume) and 3326g/m^3 for the balanced influent flow (70% initial volume). The effluent COD was obtained from the stoichiometry formulae shown in Section 2 (Equation 13).

$$S_{te} = S_{use} + f_{cv}X_v \tag{13}$$

S_{use} is soluble and unbiodegradable; hence, it was unaffected by the equalization tank and it was the same for the balanced and unbalanced flows. The only variable in this equation is the concentration of the solids in the reactor X_v (see Equation 8).

$$MX_V = MX_{BH} + MX_{EH} + MX_I \tag{8}$$

The only term in Equation 8 that is the same for unbalanced and balanced flows is the MX_I (the unbiodegradable particulates), MX_{EH} which is the endogenous residue of the biomass is strictly dependent on MX_{BH} (Equation 6).

$$MX_{EH} = f_H b_H MX_{BH} R_s \tag{6}$$

Hence, the main component that can alter the COD concentration in the reactor is the growth of biomass (MX_{BH}), and this can be quantified as in Equation 5.

$$MX_{BH} = FS_{bi} \frac{Y_H R_s}{(1+b_H R_s)} \tag{5}$$

If there is any change in the amount and activity of the biomass that grows in the reactor, COD flux out the reactors has to be investigated.

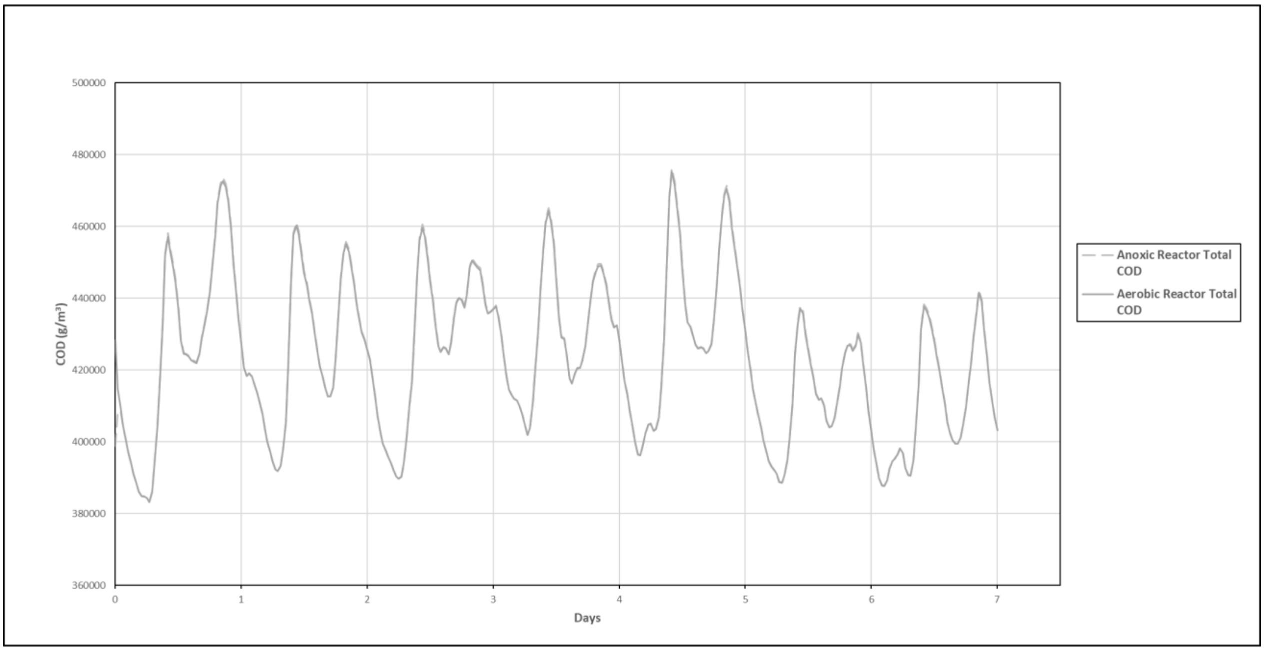

Figure 25:COD Flux for Unbalanced Influent Flow

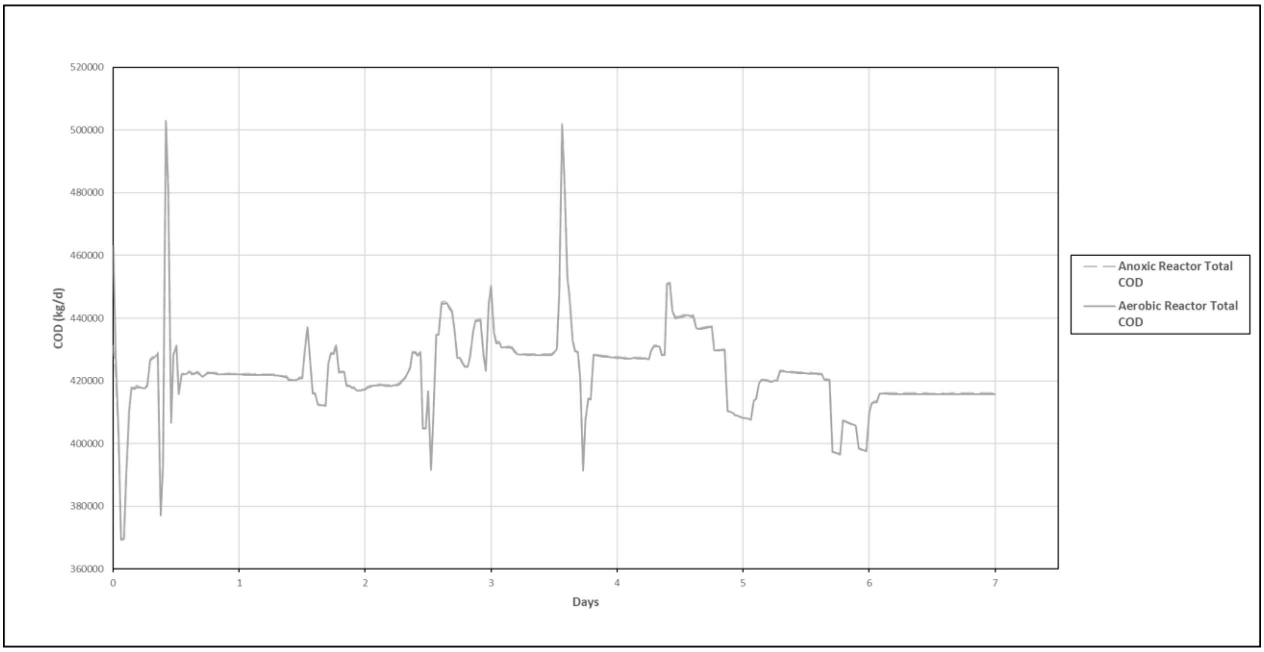

Figure 26:COD Flux for Balanced Influent Flow (40% Influent Flow)

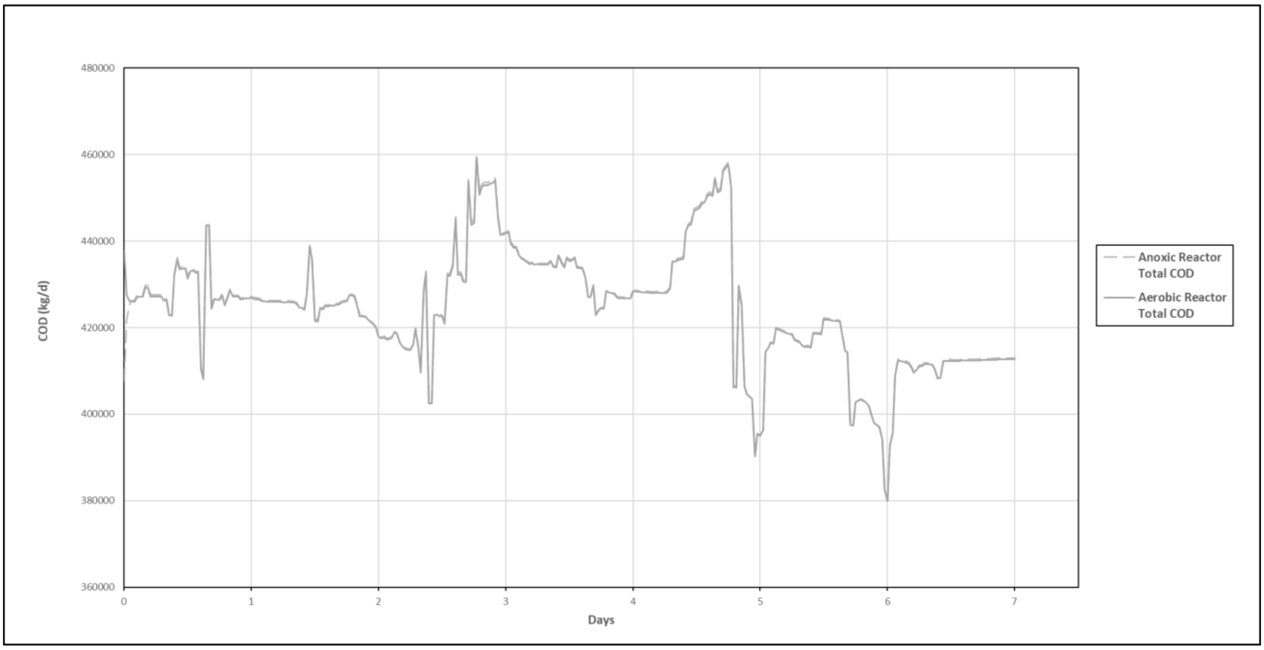

Figure 27: COD Flux for Balanced Influent Flow (70% Influent Flow)

Figures 25, 26 and 27 show the total COD fluxes from the anoxic and aerobic reactors for the unbalanced, and balanced influent flow at 40% and 70% initial equalization tank volume. The mean fluxes for these scenarios in the aerobic reactors were 423241kg/d, 422663kg/d and 424349kg/d, respectively. Hence, the COD flux difference was less than 1% for these cases, showing no major change due to flow equalization. The differences in the anoxic reactor were similar due to the recycles.

Generally, the effluent standards depend on the COD leaving the SST, which is the unbiodegradable COD, and any particulate matter depending on the performance of the SST (the performance of the SST was investigated and reported in Section 4.7.7).

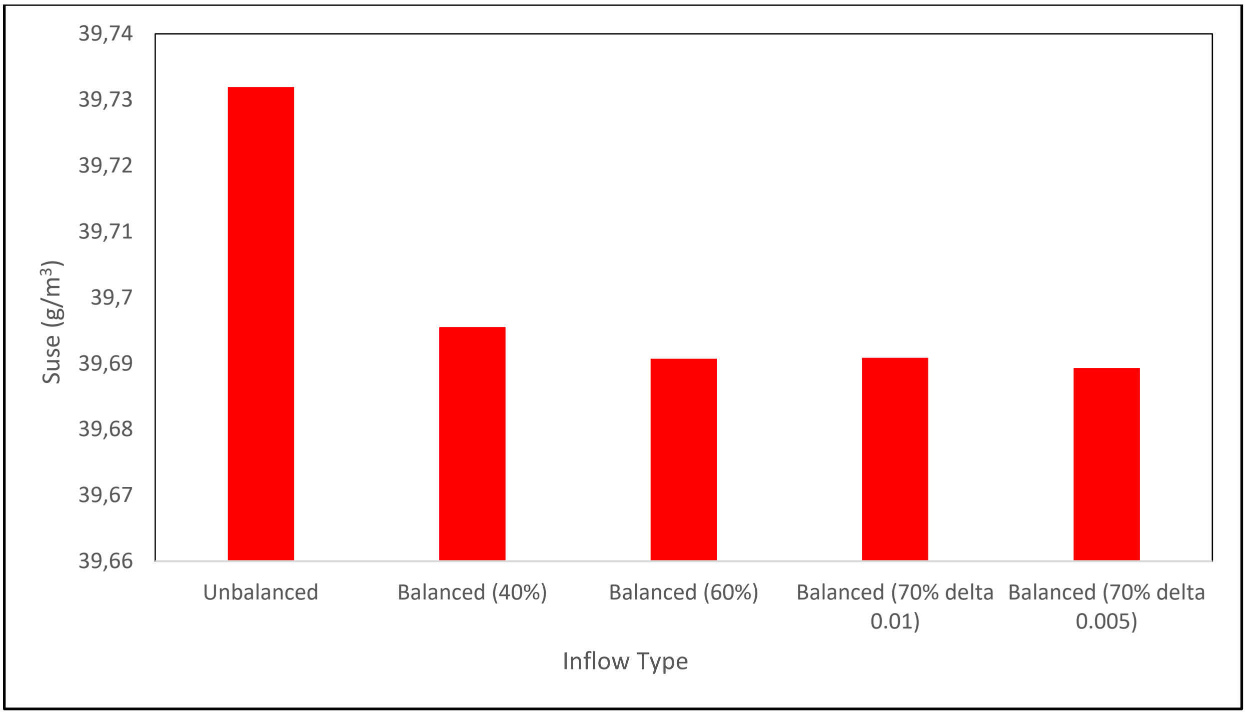

Figure 28:Unbiodegradable Soluble COD Mean

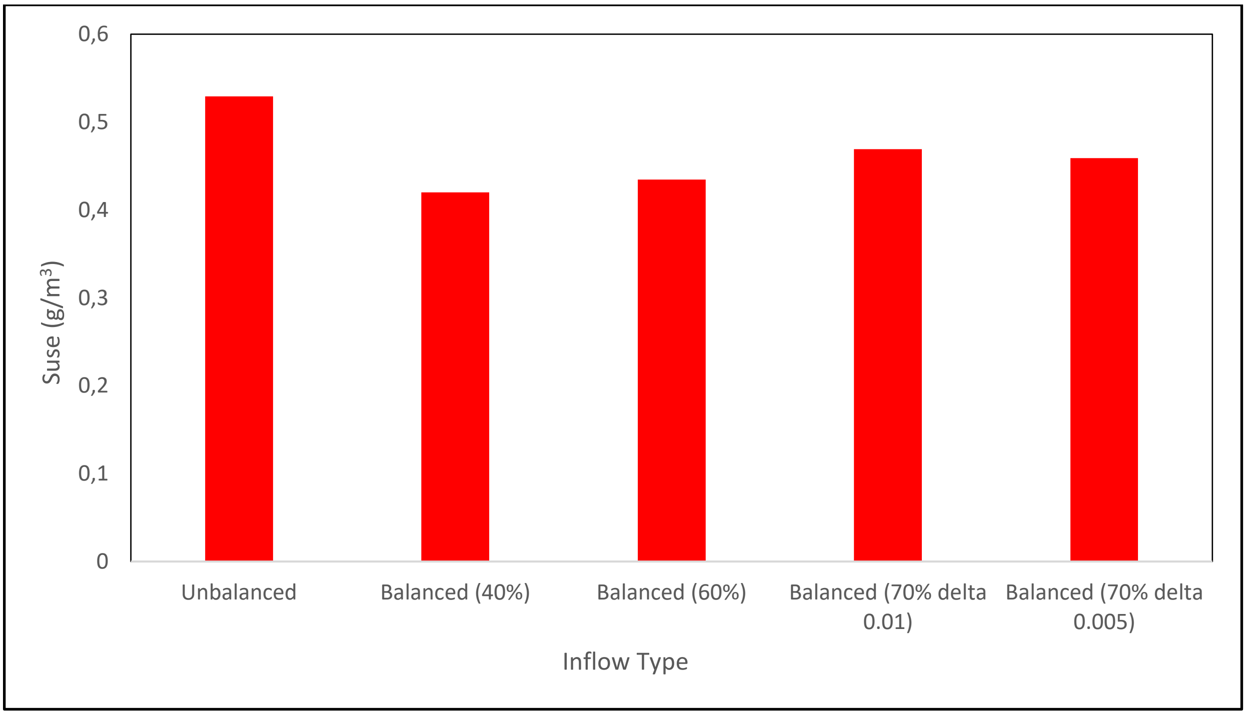

Figure 29:Unbiodegradable Soluble COD Standard Deviation

Figure 28 and 29 further elaborated that there is no difference in the variation of the S_{use} in the effluent and that its concentration is similar whether the flow is balanced or not. This was in line with the observations from Dold (1982), showing that the equalization tank only has any measurable effect whereby particulate matter are of concern. This effect is due to the variation of the effluent COD flux (the biomass is still fed at the same amount of total biodegradable material; hence, it is expected that its total mass is not to differ regardless of the variation of the influent flow) as can be seen in Table 15.

Table 15: VSS Flux Standard Deviation

VSS Flux (kg/d)		
Standard Deviation	**Status**	**Reactor**
14293	Unbalanced	Aerobic
9209	Balanced (40%)	Aerobic
8929	Balanced (60%)	Aerobic
8993	Balanced (70% delta 0.01)	Aerobic

In summary, with the presence of an equalization tank, there was no significant effect on the unbiodegradable soluble COD, and the particulate mass in the effluent varied less than with the unbalanced flow, but the mean mass was approximately similar (within 1%).

4.7.4 Nitrate generation

To achieve N removal, nitrification (i.e., the aerobic conversion of influent TKN to nitrate), and denitrification must happen (i.e., anoxic conversion of nitrates to nitrogen gas) (Ikumi, 2011). Figure 30 shows the nitrate concentration in the reactors when the influent flow was not balanced.

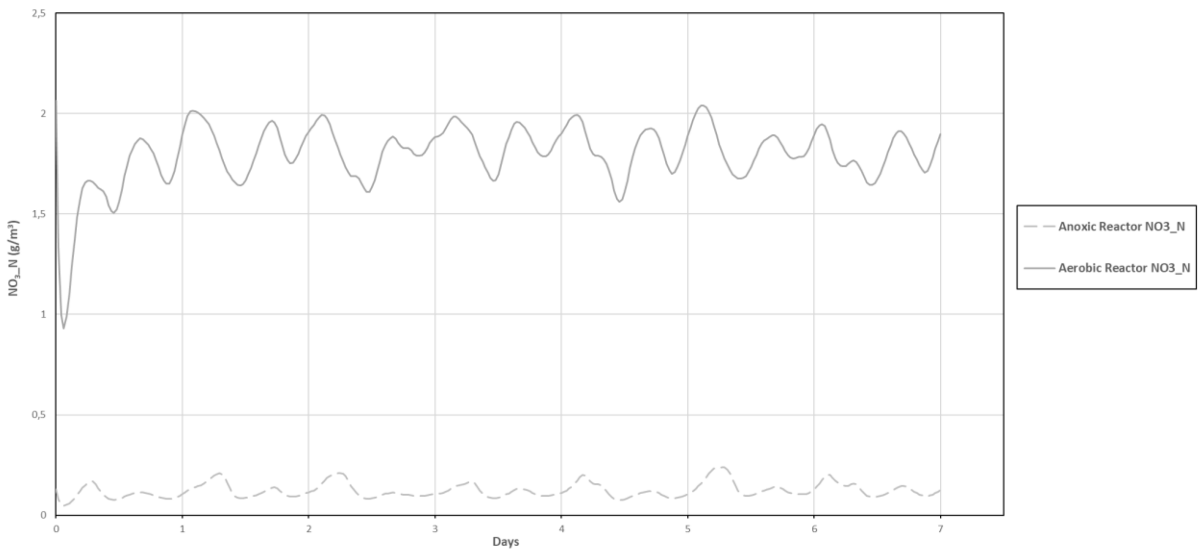

Figure 30:Effluent Nitrate Concentration (Unbalanced Influent Flow)

Due to nitrification happening in the aerobic reactor, the highest concentration of NO_3 was prevalent. The NO_3 concentration varied highly due to the daily varying influent supply.

It can be seen in the Figures 31 & 32 that there is a slight improvement (2%) in the nitrification process when the flow is balanced, with the mean concentrations of nitrate produced increasing from 1.80 g/m^3 to 1.83 g/m^3 (for balanced flow with 40% and 70% initial volume, 2% difference). The nitrate fluxes equally increased if the flow was balanced, from a mean mass of 229 kg/d (for unbalanced influent flow) to 234 kg/d (for 40% initial volume balanced flow) and to 233 kg/d (for 70% initial volume balanced flow). This improvement was because the nitrifiers are better able to oxidize ammonia as they are provided with a more stable and adequate amount of dissolved oxygen.

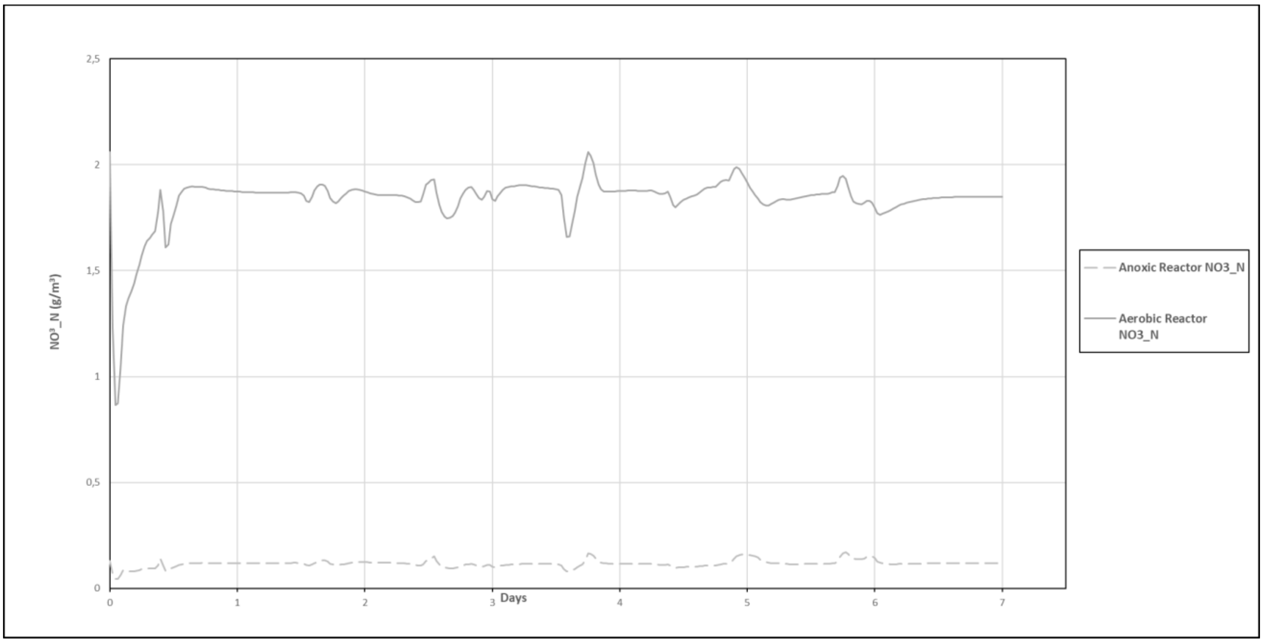

Figure 31:Effluent Nitrate Concentration (Balanced Influent Flow, 40% Initial Volume)

The fluctuations of the nitrate concentration were similar for the unbalanced and balanced influent flows because nitrate is soluble and not affected by the equalization tank as can be seen in Figure 33.

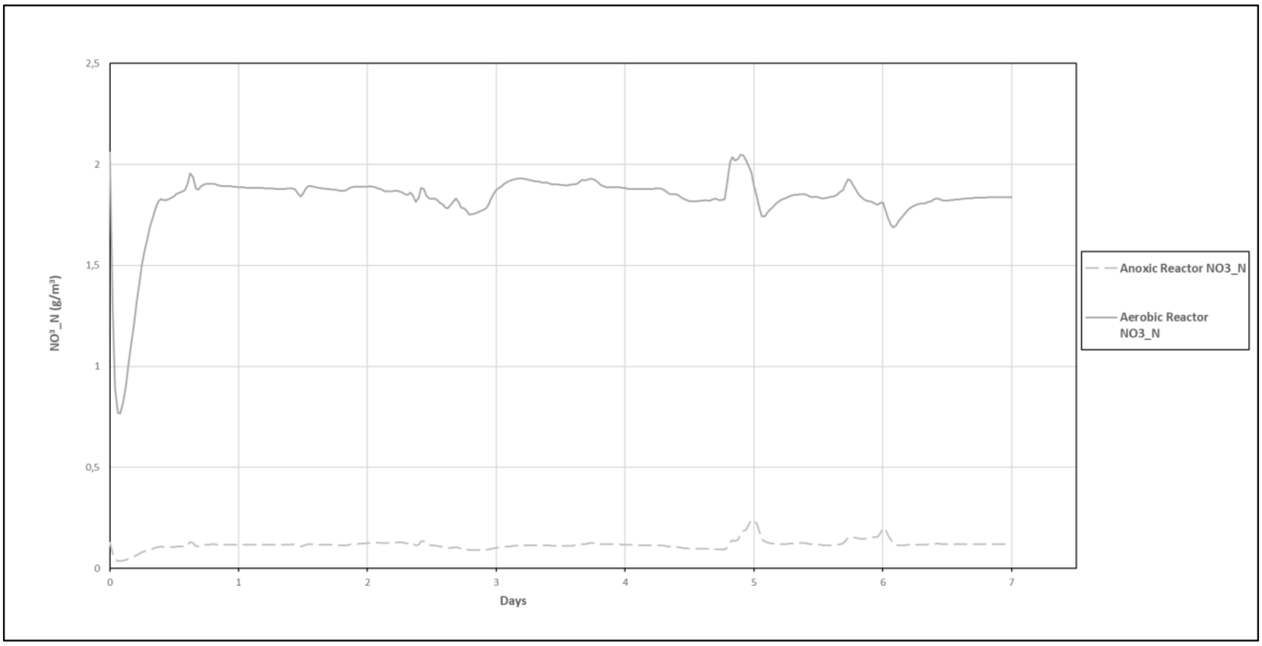

Figure 32:Effluent Nitrate Concentration (Balanced Influent Flow, 60% Initial Volume)

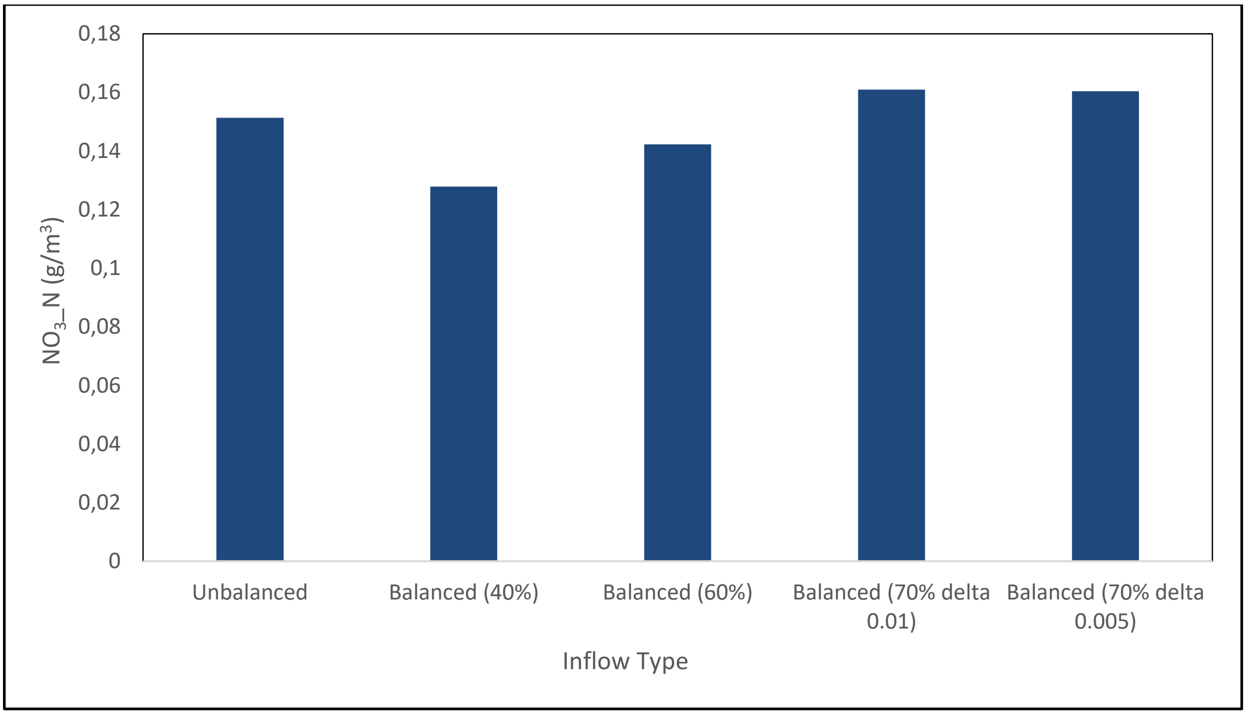

Figure 33:Effluent Nitrate Concentration variation with Balanced and Unbalanced flows

In summary, the nitrification process of the aerobic reactor slightly improved in the presence of a balancing tank due to better provisioning of the required dissolved oxygen.

4.7.5 Sludge generation

The sludge wasted from the reactors is directly related to the COD flux. Figures 34 & 35 show the profile of the VSS flux out of the reactors for the 7-day simulation for the unbalanced influent flow and balanced influent flows.

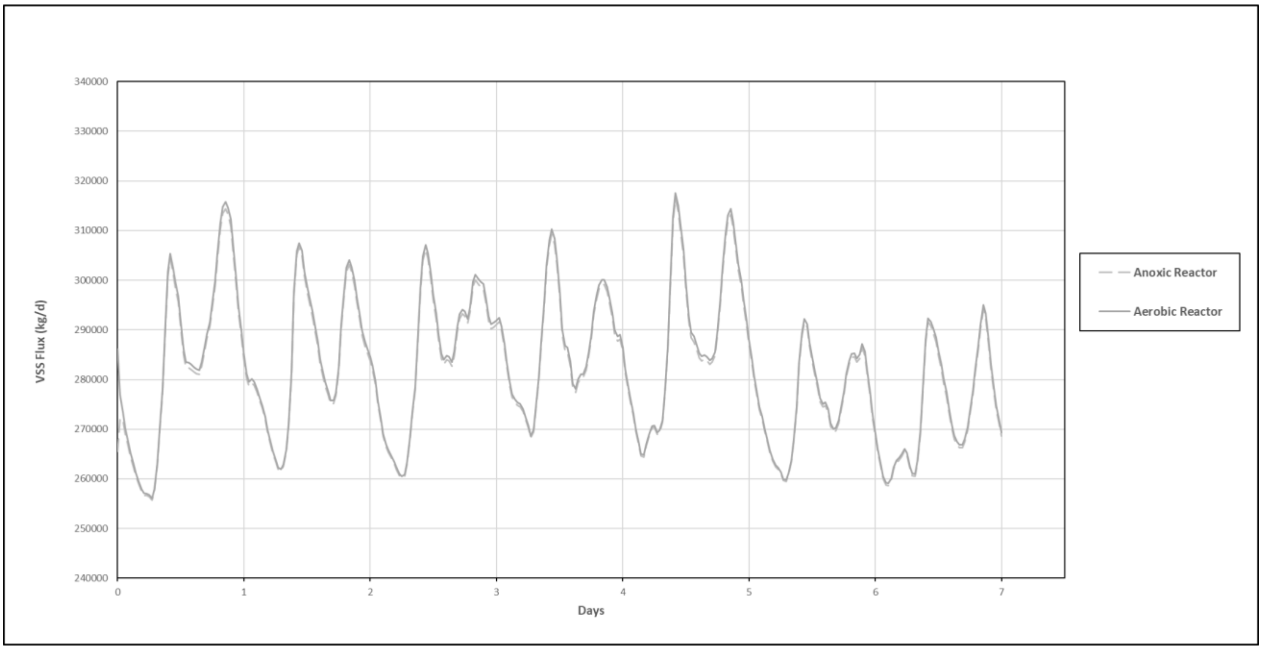

Figure 34:VSS Flux (Unbalanced Influent Flow)

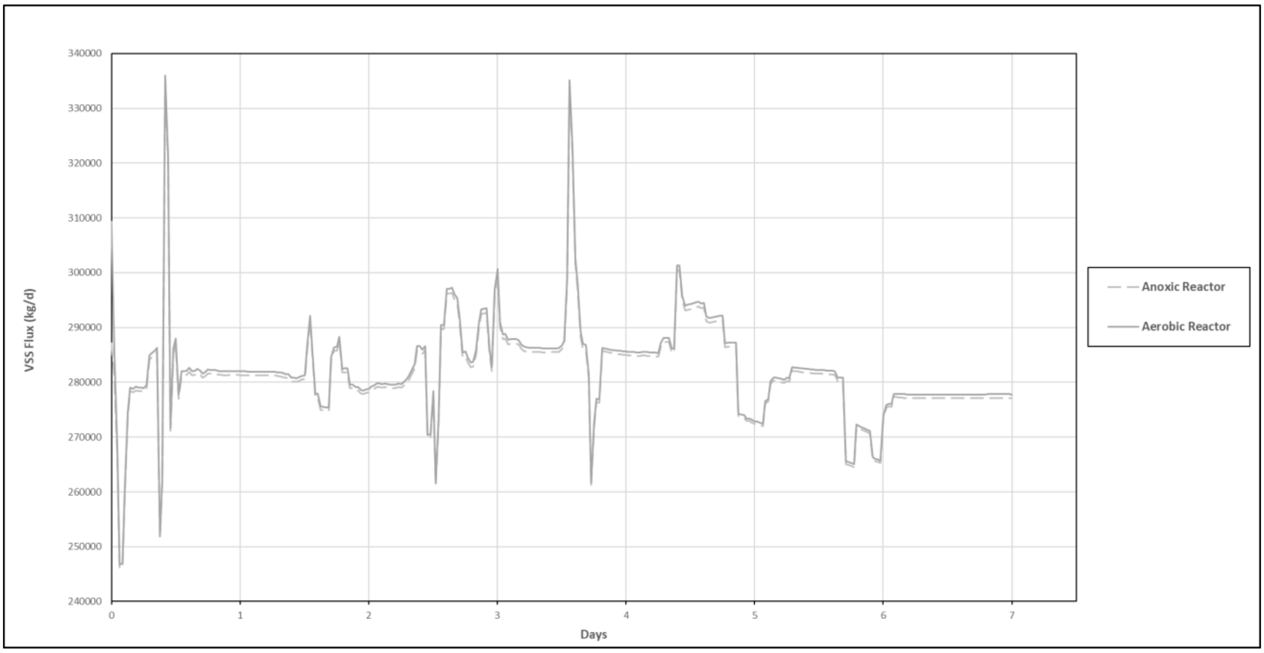

Figure 35:VSS Flux (Balanced Influent Flow, 40% Initial Volume)

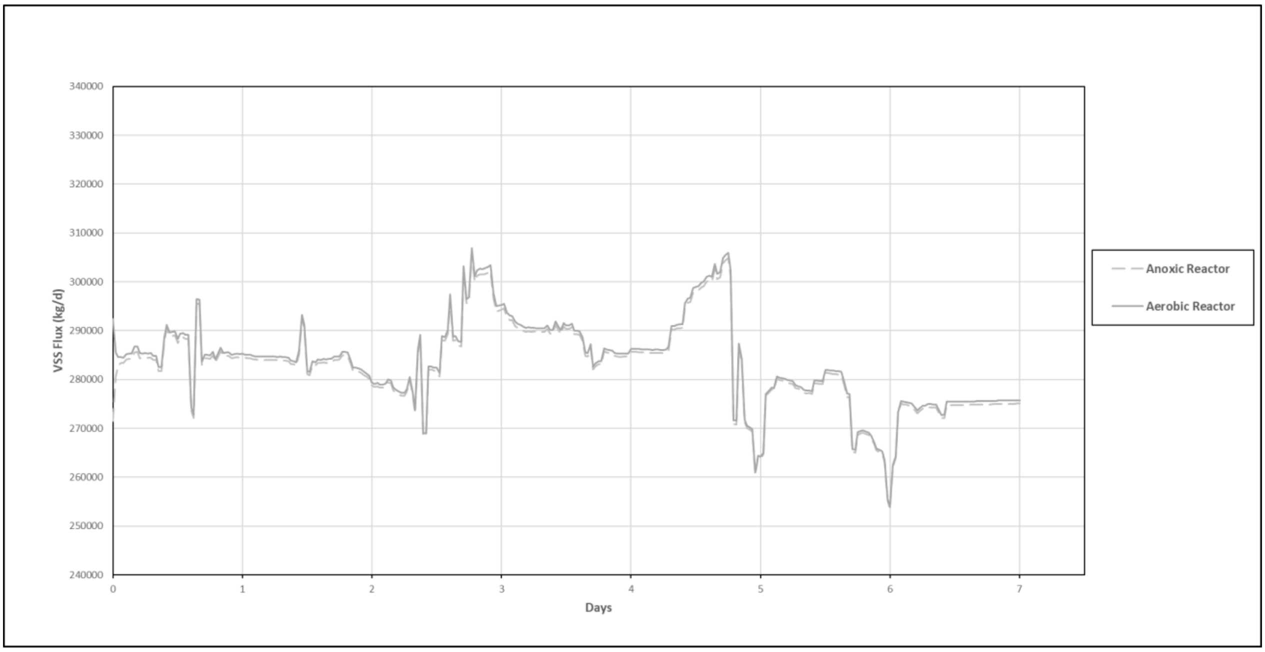

Figure 36:VSS Flux (Balanced Influent Flow, 70% Initial Volume)

Similar to the COD flux, the volatile suspended solids (VSS) flux did not change significantly (by less than 1%) when the flow was balanced, as there was no increase in COD, more biomass could not be formed. The mean mass of VSS generated for the various flow can be seen in Figure 37.

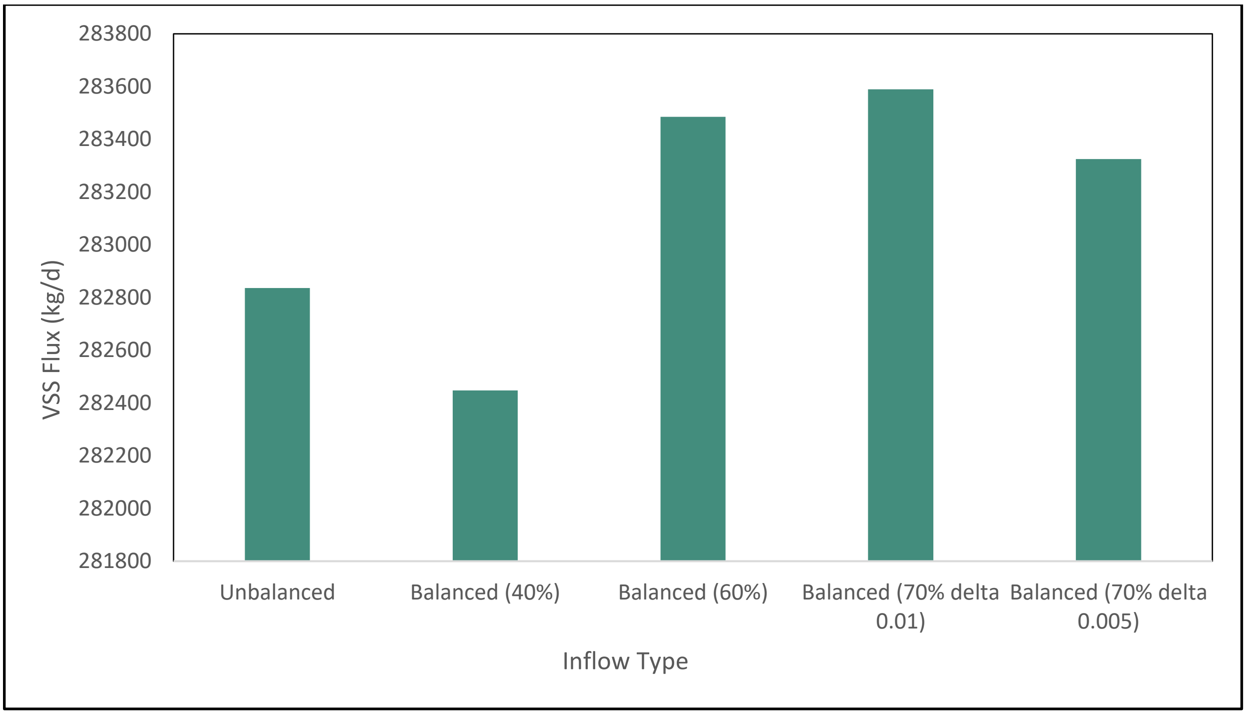

Figure 37: Mean VSS Flux from Aerobic Reactor

Overall, the variability of the VSS flux generated was significantly decreased (by 35%-43%) when the influent flow was balanced, as can be seen in Figure 38.

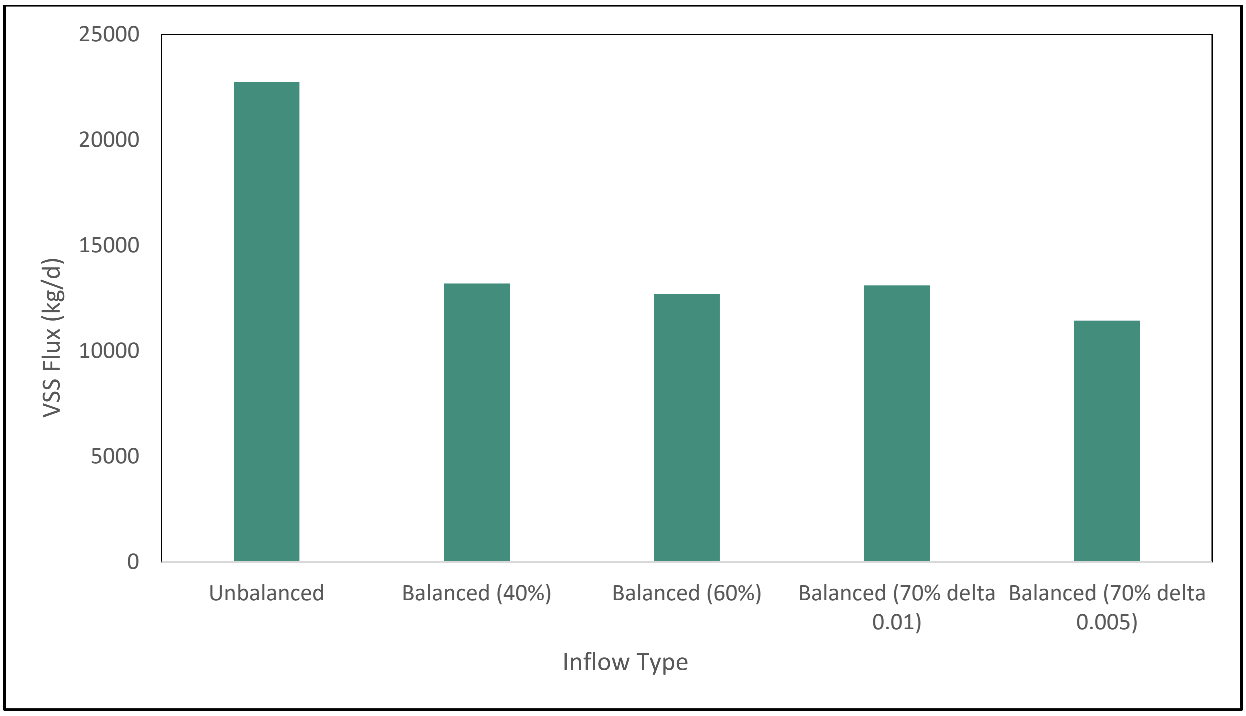

Figure 38: Standard Deviation of VSS Flux in Aerobic Reactor

In summary, the balancing tank significantly reduced the variation of the VSS, as it is largely composed of solids.

4.7.6 Primary sludge anaerobic digestion

The performance of the AD investigated was in relation to the methane (CH_4) production and pH of the digester. The mean CH_4 produced for the unbalanced and balanced influent flows were quasi-identical, having differences of less than 1%. The mean CH_4 flux with the unbalanced flow was 874 m^3/d while the mean with the balanced flow was 875m^3/d. The pH of the AD digesters with and without balanced flows were identical (6.2), indicating that the equalization tank in this scenario had minimal effect on the AD performance.

4.7.7 Secondary settling tank (SST) performance

The SST performance investigated was the flow into the SST. The inflow into the SST with and without the equalization tank presented focused on were within 1% (37139m^3/d and 37145m^3/d, respectively). The equalization tank with this layout did not have a significant effect on the inflow into the SST; hence, on the overall SST performance.

4.8 Scenario Two: Summary

The main aspects assessed in this scenario were the performance of the AS system, AD and SST. It was seen in the above sections that including an equalization tank has insignificant effects on the overall performance of the AS system. It only reduced the flow variations which might potentially improve plant operations (these were not assessed in this study). A summary of the primary findings were:

- The OUR variability was reduced by close to 50% with the presence of an equalization tank. Hence, it was assumed that the operation of the plant might be easier as the dissolved oxygen to be provided varies over a smaller range. This agrees to Dold (1982) findings.
- The aeration energy equally reduced, but the mean aeration energy only changed within 1%; hence, the operation cost of aeration was determined to not alter in any significant manner.
- All unbiodegradable soluble material were not affected by the presence of an equalization tank as they do not settle.
- The nitrification process improved by 2% in the presence of an equalization tank, due to better provision of dissolved oxygen, and perhaps reduced variability of OUR.
- The VSS flux did not change significantly (by less than 1%) in the presence of an equalization tank, but its variability was significantly improved (by 35% - 43%).
- The anaerobic digester and SST performances were not altered by the presence of an equalization tank.

Overall, most of the literature reviewed focuses on how well flow is equalized, and does not take into consideration the impact of the equalization tank on the whole plant system; hence, some of these findings could not be directly compared to the literature reviewed.

5. Conclusion

The aim of this project was to develop and assess a dynamic simulation model for equalization tanks. The outcome of this project is the primary phase of the dynamic simulation model which provides only basic requirements, i.e. flow equalization. Two of the research questions set at the start of the project were answered (which were; i) can a complete dynamic simulation model replicating equalization tanks be developed? and ii) can significant benefits of equalization tanks for WRRFs be identified via the model?) while the other two (which were; i) can the BSM2 be used to satisfactorily calibrate and validate the model? and ii) can the model be made compatible with the existing UCT plant-wide model?) are recommendations for further projects.

A dynamic simulation model replicating equalization tanks was successfully developed, and using an MLE system as case scenario, it was determined that when only considering flow equalization, the main benefits obtained are the reduction in size of some unit processes by 8-9% (primary settling tank, biological reactors, secondary settling tank, flotation unit, anoxic-aerobic digester) and the significant reduction in variability of various plant parameters by up to 50% (flow, OUR, VSS flux) which can potentially improve the ease of operation in plants. The model could not be calibrated nor validated and could not be made compatible with the existing UCT plant wide model due to challenges faced in the project that are described later in this section.

Using Dold (1982) as a guideline, a dynamic simulation model for an equalization tank was developed in Microsoft Excel (which will help ease further developments associated with this work, as this package is widely used as opposed to the techniques used by Dold (1982)). This model optimizes historical flow data to get balanced outflows for an equalization tank of a given volume. The model is functional and the equalization tank results in various degrees of flow equalization depending on the initial tank volume, with the best equalization results being experienced at 70% initial tank volume (with a standard deviation for flow reduced from 6.6 with unbalanced flow to 3.3). The impact of an equalization tank was tested on an MLE system in the WEST simulation platform and the following was observed:

- In an MLE system, the presence of an equalization tank has significant positive effects (about 10% improvement) on the performance of various unit processes.
- The presence of the equalization tank reduces the unit process design sizes of the MLE system (Close to 10% reduction in sizes).
- With the addition of an equalization tank in the MLE system used in the scenario analysis, the capital cost of the plant increased by just 2.4%, with the plant footprint substantially increasing by 13% (from 5306m^2 to 5988m^2).
- The presence of an equalization significantly decreases the variation of flow, aeration energy, and OUR. Hence the ease of operation of plants can be positively impacted with the presence of an equalization tank..

These results best depict the benefit of an equalization in a WRRF.

It was suspected that the reason the equalization tank does not have more significant positive effect (i.e improvement of more than 10% in unit sizes) is because the PST already provides a form of equalization, imparting benefits of plant equalization (with respect to size of unit processes mainly). Further assessment on the impact of plants without PST present can be done in the future to confirm this.

Some challenges and limitations experienced during the project were as follows:

- The equalization tank model can only cope with up to 7-day historical inflow data at once. This is to be improved in further stages of the development of this model.
- The validation and calibration of the model were not performed. The development of the equalization tank model was deemed sufficient for this project, and the validation and calibration were of the model will be performed in a further research.
- The main challenge experienced is that the model could not be incorporated in the UCT plant wide system in the WEST® simulation environment. This was due to the inability of the coding environment to process step-wise integration. To circumvent this challenge, the model was run as a macro with influent historical data in Microsoft Excel, and the resulting balanced outflow were put in the WEST® environment as the influent into an existing MLE plant system. This method could replicate the effect of the balancing tank.

Despite these challenges, the first phase of the development of a dynamic simulation model was successfully achieved in this project.

Further developments of the model required are as follows (which can be done in subsequent projects include):

- Validation and calibration of the model.
- Making the model cope with data of more than 7-days.
- Successfully incorporating it in the UCT plant wide system, to contribute towards holistic decision making for optimized operation of water and resource recovery facilities.
- Including equalization of load in the model.

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Appendix A Definition of literature review terms

Term	Definition	Unit
$[HCO_3^-]$	Bicarbonate concentration	mol/l
ΔP_{OHO}	P removal due to OHOs	gP/m^3
ΔP_{PAO}	P removal due to PAOs	gP/m^3
$\Delta P_{SYS, ACT}$	Actual total P removal for the system	gP/m^3
$\Delta P_{SYS, POT}$	Potential total P removal by the system	gP/m^3
ΔP_{XE}	P Removal due to endogenous residue mass	gP/m^3
ΔP_{XI}	P Removal due to influent inert mass	gP/m^3
μ_{AmT}	Maximum specific growth rate of nitrifiers at T°C	d^{-1}
2.86	Oxygen equivalent of nitrate	Unitless
24	Gas volume at 20°C	l/mol
3.0	Mass of COD removed per unit of oxygen utilized	$gCOD/gO_2$
64	COD mass per mol of methane	gCOD/mol
8.6	Mass of COD removed per unit of nitrate denitrified	$gCOD/NO_3$-N
b_{AD}	Acidogen endogenous respiration rate	d^{-1}
b_{AT}	Specific endogenous respiration rate for nitrifiers at 20°C	d^{-1}
b_H	Specific rate of endogenous mass loss of OHOs	d^{-1}
$b_{OHO,T}$	OHO specific endogenous mass loss rate constant at temperature T°C	d^{-1}
$b_{PAO,T}$	PAO specific endogenous mass loss rate at temperature T°C	gEVSS/gVSS.d
D_B	4k+l-2m-3n	e^-eq/mol biomass
$D_{p1RBCOD}$	Denitrification potential of the influent RBCOD in the primary anoxic reactor	mgN/l
D_S	4x+y-2z-3a	e^-eq/mol biodegradable organics
E	$\frac{V(Z_{AD}+Z_E)}{R_sQ_i(S_{bpi}-S_{bp})} = \frac{Y_{AD}(1+f_{AD}*b_{AD}*R_s)}{[1+b_{AD}R_s(1-Y_{AD}\{1-f_{AD}\})]}$	Unitless

f_{AD}	Acidogen biomass unbiodegradable fraction	Unitless
f_{AN}	Anaerobic mass fraction (gVSS/gVSS)	Unitless
fcv	COD to VSS ratio of the sludge	Unitless
$f_{FSS,OHO}$	Fraction of FSS in the OHO active biomass	gFSS/gVSS
$f_{FSS,PAO}$	Fraction of FSS in the PAO active biomass	gFSS/gVSS
f_H	Unbiodegradable fraction of the OHOs	mgCOD/mgCOD
f_{iOHO}	Inorganic content of OHOs	mgISS/mgCOD
f_n	Nitrogen content of VSS	mgN/mgVSS
$FO_{2,\ OHO}$	Daily mass of oxygen consumed by OHOs	gO_2/d
$FO_{2,\ PAO}$	Daily mass of oxygen consumed by PAOs	gO_2/d
FO_C	Daily carbonaceous oxygen demand	gO_2/d
FO_c	Daily flux of oxygen utilised	mgO_2/d
FO_n	Flux of oxygen required for nitrification	kgO_2/d
FO_t	Total flux of O_2 required including the oxygen recovered by denitrification	kgO_2/d
$f_{P,FSS}$	P fraction of fixed (inorganic) suspended solids mass	gP/gFSS
$f_{P,OHO}$	fraction of OHO active mass that is P	gP/gAVSS
$f_{P,PAO}$	Fraction of PAO active mass that is P	gP/gAVSS
$f_{P,TSS}$	P fraction of total suspended solids mass	gP/gTSS
$f_{P,XE}$	fraction of endogenous residue mass that is P	gP/gEVSS
$f_{P,XE}$	fraction of inert mass that is P	gP/gEVSS
$f_{PS'up}$	Unbiodegradable particulate fraction of the primary sludge	Unitless
$f_{S'bs}$	RBCOD fraction with respect to influent biodegradable COD	Unitless
$f_{s'up}$	Particulate unbiodegradable fraction of total influent COD	Unitless
$f_{s'us}$	Soluble unbiodegradable fraction of total influent COD	Unitless
$FS_{b,OHO}$	Daily mass of biodegradable substrate available to OHOs	gCOD/d
FSbi	Daily flux of influent biodegradable COD	mgCOD/d

$FS_{bs,PAO}$	Daily mass of readily biodegradable substrate stored by PAOs in the anaerobic reactor	gCOD/d
FS_{ti}	Daily mass of influent COD	gCOD/d
f_{x1}	Primary anoxic sludge mass fraction	Unitless
f_{x3}	Secondary anoxic sludge mass fraction	Unitless
$f_{XE,OHO}$	Fraction of endogenous particulate residue of OHO	gEVSS/gAVSS
$f_{XE,PAO}$	Fraction of endogenous particulate residue of PAOs	gEVSS/gAVSS
$FX_{FSS,i}$	Daily mass of influent FSS	gFSS/d
$f_{XI,COD,i}$	Fraction of influent COD that is particulate inert	gCOD/gCOD
FX_{Ii}	Daily flux of influent particulate unbiodegradable COD	mgCOD/d
FX_{IOi}	Daily flux of influent particulate inorganic matter	mgISS/d
FX_{t}	Daily flux of total solids produced	mgTSS/d
K_2	Second specific rate of denitrification in primary anoxic reactor	$mgNO_3$-N/mgOHOVSS.d
K_3	Specific rate of denitrification in secondary anoxic reactor	$mgNO_3$-N/mgOHOVSS.d
K_m	Kinetic constant	gCOD organics/gCOD biomass.d
K_{nT}	Half saturation constant for nitrifiers at T^oC	mgN/l
K_s	Kinetic constant	gCOD/l
MX_{BH}	Mass of OHO in the bioreactor	mgVSS
$MX_{E, PAO}$	Mass of PAO endogenous residue in the system	gEVSS
$MX_{E,OHO}$	Mass of OHO endogenous residue in the system	gEVSS
MX_{EH}	Mass of endogenous residue in the bioreactor	mgVSS
MX_{FSS}	mass of fixed suspended solids in the system	gFSS
MX_{I}	mass of inert organic matter in the system coming from the influent	gIVSS
MX_{IO}	Mass of influent particulate inorganic matter in the bioreactor	mgISS

MX_{OHO}	OHO active biomass	gAVSS
MX_{PAO}	Mass of PAO in the system	gAVSS
MXt	Mass of TSS in the reactor	gTSS
M_{XV}	Mass of volatile suspended solids in the bioreactor	mgVSS
MX_{VSS}	VSS mass in the system	gVSS
N	Total number of anaerobic reactors of equal volume in the series n = 1, 2…N	Unitless
N_a	Bulk liquid ammonia concentration	mgN/l
N_{ae}	Effluent ammonia concentration	mgN/l
N_{ai}	Influent ammonia concentration	mgN/l
N_c	Nitrification capacity	mgN/l
N_{ne}	Effluent nitrate concentration	mgN/l
N_{obpi}	Influent biodegradable particulate organic nitrogen	mgN/ l
N_{obsi}	Influent biodegradable soluble organic nitrogen	mgN/l
N_{oupi}	Influent unbiodegradable particulate organic nitrogen	mgN/l
N_{ouse}	Effluent unbiodegradable soluble organic nitrogen	mgN/l
N_s	Concentration of TKN in influent required for sludge production	mgN/l
N_{te}	Effluent TKN concentration	mgN/l
p_{CO2}	Partial pressure of CO_2 in the gas phase	mol fraction
pH	Negative log10 of the (H^+) activity	Unitless
pH_i	Influent pH	Unitless
pK'a	Negative $\log_{10}$ of acid dissociation constant	Unitless
pK'_{C1}, pK'_{C2}	Negative $\log_{10}$ of 1st and 2nd carbonate system apparent dissociation constant (where “apparent” means corrected for ionic strength effects)	Unitless
pK'_{HCO2}	Negative $\log_{10}$ of the apparent Henry’s law constant	Unitless
Q_i	Influent flow rate	m^3/d
Q_m	Methane gas flow per litre influent	l methane/d/l influent flow/d
R	Retention time	d
r_h	Monod hydrolysis rate	gCOD/l.d

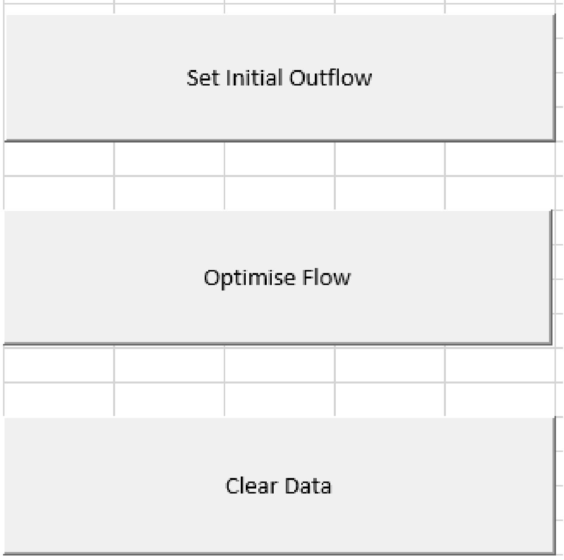
R_s	Sludge retention time	(d)
s	Sludge recycle ratio to anaerobic reactor based on influent flow	Unitless
Sbi	Influent biodegradable COD	mgCOD/l
S_{bi}	Influent biodegradable COD	mgCOD/l
S_{bp}	Residual biodegradable COD concentration	gCOD/l
S_{bpi}	Hydrolysable COD concentration from the influent	gCOD/l
S_{bsAci}	Dissociated acetate species concentration in the influent	gCOD/l
S_{bsai}	Concentration of VFA in the influent	gCOD/m^3
S_{bsai}	Volatile fatty acid concentration in the influent	gCOD/)
$S_{bsf,ANn}$	Concentration of fermentable COD in the effluent of the nth anaerobic reactor	gCOD/m^3
$S_{bsf,i}$	Fermentable COD influent concentration	gCOD/m^3
$S_{bsf,i,conv}$	Fermentable COD available for conversion per volume of influent	gCOD/m^3
S_{bsHAci}	Undissociated acetate species concentration in the influent	gCOD/l
S_m	Effluent methane concentration	gCOD/l
$S_{NO3,i}$	Nitrate concentration in the influent to anaerobic reactor	gNO_3-N/m^3
$S_{NO3,s}$	Nitrate concentration in the sludge recycle to anaerobic reactor	gNO_3^-N/m^3
$S_{O2,i}$	Oxygen concentration in the influent to anaerobic reactor	gO_2/m^3
$S_{O2,s}$	Oxygen concentration in the sludge recycle to the anaerobic reactor	gO_2/m^3
S_{te}	Effluent COD	(mgCOD/l
Sti	Influent total COD	mgCOD/l
S_{upi}	Unbiodegradable COD concentration in the influent	gCOD/l
$T_{P,e}$	Effluent total P concentration	gP/m^3
$T_{p,i}$	Influent total P concentration	gP/m^3
TSS_e	Total suspended solids concentration of the effluent	gTSS/m^3
V	Volume of the AD	m^3

X_{Ii}	Influent unbiodegradable matter in activated sludge	mgVSS/l
X_{IOi}	Influent inorganics concentration	mgISS/l
X_t	TSS concentration is reactor	mgTSS/l
X_v	Effluent VSS concentration	mgVSS/l
Y_{AD}	Pseudo-acidogen yield coefficient	gCOD biomass/gCOD organics hydrolysed
Y_H	Yield of OHOs	0.45mgVSS/mgCOD
Y_{OHO}	OHO yield	gAVSS/gCOD
Y_{PAO}	PAO biomass yield	gAVSS/gCOD
Z_{AD}	Concentration of acidogen biomass	gCOD/l
Z_E	Acidogen endogenous residue concentration	gCOD/l

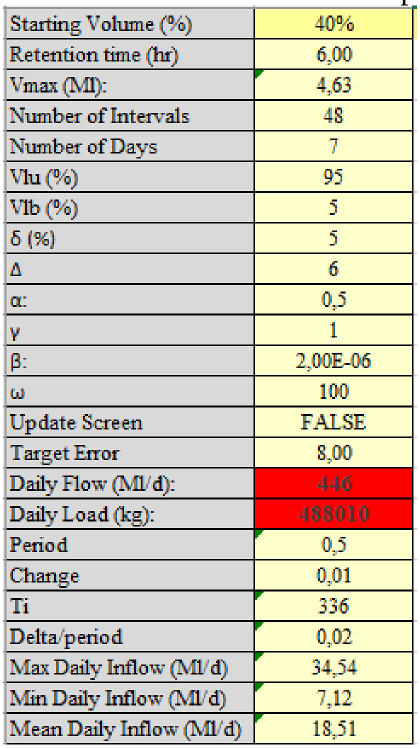
Appendix B Equalization tank model details

The model is made of 8 distinct sheets namely;

- **Buttons**: Here the model can be run by a simple click of a button, masking all the background procedure.



- **Parameters**: All the major parameters used to run the model are presented in this sheet and used in the macro for the optimization.



Starting Volume (%)	40%
Retention time (hr)	6,00
Vmax (Ml):	4,63
Number of Intervals	48
Number of Days	7
Vlu (%)	95
Vlb (%)	5
δ (%)	5
Δ	6
α:	0,5
γ	1
β:	2,00E-06
ω	100
Update Screen	FALSE
Target Error	8,00
Daily Flow (Ml/d):	446
Daily Load (kg):	488010
Period	0,5
Change	0,01
Ti	336
Delta/period	0,02
Max Daily Inflow (Ml/d)	34,54
Min Daily Inflow (Ml/d)	7,12
Mean Daily Inflow (Ml/d)	18,51

- **24hr Influent**: This sheet contains the influent flow of 1-7 days.
- **Historical Data**: The 24hr influent data is interpolated to obtain influent data for every interval, present in this sheet.
- **Outputs**: This worksheet contains the effluent flow as well as the tank concentration for every interval, as well as the error functions at each of these intervals.

- **Errors**: This worksheet has the sum of all the errors.
- **Graphs**: This worksheet contains graphs of the tank profile and inflow and outflow profile throughout the duration of the model.

Visual Basics Macro Code

```
Sub start()

Dim inp As Worksheet
Dim outp As Worksheet
Dim err As Worksheet
Dim par As Worksheet

Set inp = ThisWorkbook.Sheets("Historical data")
Set outp = ThisWorkbook.Sheets("Outputs")
Set err = ThisWorkbook.Sheets("Errors")
Set par = ThisWorkbook.Sheets("Parameters")

Dim rowid As Integer
Dim store As Double
Dim iter As Integer
Dim mean As Double
Dim store2 As Double
Dim store3 As Double
Dim store4 As Double
Dim store5 As Double
Dim store6 As Double
Dim intervals As Integer
Dim days As Integer
Dim N As Integer
Dim volume_style As String
Dim cycle As Integer

volume_style = par.Cells(1, 3)
outp.Range("K2").Value = par.Cells(1, 2).Value * par.Cells(3, 2).Value
If volume_style = "Use this Value" Then
cycle = 5
Else
cycle = 4
End If
If cycle = 4 Then
err.Cells(5, 2) = err.Cells(5, 4).Formula
Else
err.Cells(5, 2) = 0
End If
days = par.Cells(5, 2).Value
intervals = par.Cells(4, 2).Value
mean = par.Cells(16, 2).Value / (24)

For N = 1 To days
For rowid = intervals * (N - 1) + 1 To intervals * N

If inp.Cells(rowid + 2, 3).Value = vbNullString Then '
   MsgBox "Please enter a value of influent concentration for every interval", , "Error"
   End
ElseIf inp.Cells(rowid + 2, 2).Value = vbNullString Then
   MsgBox "Please enter a value of influent flow for every interval", , "Error"
   End

Else
'  Do nothing
End If
Next
Next


outp.Range("B2:B" & (intervals * days + 2)).Value = inp.Cells(1, 16).Value

End Sub
Sub optimise()

Dim inp As Worksheet
Dim outp As Worksheet
Dim err As Worksheet
Dim par As Worksheet

Set inp = ThisWorkbook.Sheets("Historical data")
Set outp = ThisWorkbook.Sheets("Outputs")
Set err = ThisWorkbook.Sheets("Errors")
```

```
Set par = ThisWorkbook.Sheets("Parameters")

Dim rowid As Integer
Dim store As Double
Dim iter As Integer
Dim mean As Double
Dim store2 As Double
Dim store3 As Double
Dim store4 As Double
Dim store5 As Double
Dim store6 As Double
Dim intervals As Integer
Dim days As Integer
Dim N As Integer
Dim acc As Integer
Dim counter As Integer
Dim target As Single
Dim volume_style As String
Dim cycle As Integer
Dim volume As Double
Dim X As Integer
Dim Ti As Integer
Dim Y As Double
Dim Pd As Double
Dim StartTime As Double
Dim MinutesElapsed As String


volume = par.Cells(3, 2)

volume_style = par.Cells(1, 3)

If volume_style = "Use this Value" Then
days = par.Cells(5, 2).Value

intervals = par.Cells(4, 2).Value

mean = inp.Cells(1, 16).Value

Delta = par.Cells(19, 2).Value

Ti = par.Cells(20, 2).Value

Pd = par.Cells(18, 2).Value

End If

'*******OVERARCHING OPTIMIZATION CODE************

StartTime = Timer

'Line1:
store3 = err.Cells(6, 2).Value

Do While outp.Cells(339, 33).Value <> 0 Or outp.Cells(339, 21).Value <> 0

Do While outp.Cells(339, 33).Value <> 0 Or outp.Cells(339, 21).Value <> 0

Debug.Print err.Cells(6, 2).Value

'****************************Error Optimization code, volumetric error**************************

For N = 2 To Ti - 6

'******Applying +Delta********

store = err.Cells(3, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta

store2 = err.Cells(3, 2).Value
```

```
If store2 < store Then
Do Until store2 > store Or outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value
store = err.Cells(3, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta
store2 = err.Cells(3, 2).Value
Loop
Debug.Print store2
Do Until outp.Cells(N + 2, 2).Value < 1 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value > 1 * par.Cells(23, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta
Debug.Print store2
Loop
End If

If store2 > store Then
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta
End If
'*****Applying -delta*******
store = err.Cells(3, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta
store2 = err.Cells(3, 2).Value

If store2 < store Then
```

```
Do Until store2 > store Or outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value
Debug.Print store2
store = err.Cells(3, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta
store2 = err.Cells(3, 2).Value

Loop
Debug.Print store2
Do Until outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta
Debug.Print store2
Loop
End If
If store2 > store Then
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta

End If
Next

'*****************Last Intervals******************
For N = Ti - 5 To Ti - 2

'******Applying +Delta********
store = err.Cells(3, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta
```

```
store2 = err.Cells(3, 2).Value

If store2 < store Then
Do Until store2 > store Or outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value
store = err.Cells(3, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta
store2 = err.Cells(3, 2).Value

Loop
Debug.Print store2
Do Until outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta
Debug.Print store2
Loop
End If

If store2 > store Then
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta
End If
'*****Applying -delta*******
store = err.Cells(3, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta
store2 = err.Cells(3, 2).Value
'Debug.Print store2 - store
If store2 < store Then
Do Until store2 > store Or outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value
Debug.Print store2
store = err.Cells(3, 2).Value
'Debug.Print store
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
```

```
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta
store2 = err.Cells(3, 2).Value
Loop
Debug.Print store2
Do Until outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta
Debug.Print store2
Loop
End If
If store2 > store Then
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta
End If
Next

Debug.Print err.Cells(6, 2).Value
'****************************Error Optimization code, Flow rate change error**************************
For N = 2 To Ti - 6
'******Applying +Delta********
store = err.Cells(4, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta

store2 = err.Cells(4, 2).Value
'Debug.Print store2 - store
If store2 < store Then
Do Until store2 > store Or outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value
store = err.Cells(4, 2).Value
'Debug.Print store
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
```

```
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta

store2 = err.Cells(4, 2).Value

'Debug.Print store2

'Debug.Print store2 - store

Loop
Debug.Print store2

Do Until outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta ' 1 step back to get min error

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta

Loop

Debug.Print store2

End If

If store2 > store Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta

End If

'*****Applying -delta*******

store = err.Cells(4, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta

store2 = err.Cells(4, 2).Value

'Debug.Print store2 - store

If store2 < store Then

Do Until store2 > store Or outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value

store = err.Cells(4, 2).Value

'Debug.Print store

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
```

```
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta

store2 = err.Cells(4, 2).Value

'Debug.Print store2

'Debug.Print store2 - store

Loop
Debug.Print store2

Do Until outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta          '1 step back to get min error

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta

Debug.Print store2

Loop


End If

If store2 > store Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta

End If

Next

store4 = err.Cells(6, 2).Value

Debug.Print err.Cells(6, 2).Value



'******************Last Intervals******************

For N = Ti - 5 To Ti - 2

'******Applying +Delta********

store = err.Cells(4, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta

store2 = err.Cells(4, 2).Value


If store2 < store Then
```

```
Do Until store2 > store Or outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value
store = err.Cells(4, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta


store2 = err.Cells(4, 2).Value


Loop
Debug.Print store2

Do Until outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta

Loop

Debug.Print store2

End If

If store2 > store Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta


End If

'*****Applying -delta*******

store = err.Cells(4, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta

store2 = err.Cells(4, 2).Value

'Debug.Print store2 - store

If store2 < store Then

Do Until store2 > store Or outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value

store = err.Cells(4, 2).Value

'Debug.Print store

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta


store2 = err.Cells(4, 2).Value

'Debug.Print store2
```

```
'Debug.Print store2 - store

Loop
Debug.Print store2

Do Until outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value Or outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta

Debug.Print store2

Loop


End If

If store2 > store Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta


End If

Next

store4 = err.Cells(6, 2).Value

Debug.Print err.Cells(6, 2).Value

Loop


'***************Making Sure the Fractional Volume is between 0 and 1***********

Do Until outp.Cells(Ti + 3, 21).Value = 0

For N = 2 To Ti - 6

store = outp.Cells(N + 2, 12).Value

If store < 0 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta

store2 = outp.Cells(N + 2, 12).Value

Debug.Print store2

If store < store2 And store2 < 1 Then

store = store2

Debug.Print store - store2

Do Until 0.01 < outp.Cells(N + 2, 12).Value And outp.Cells(N + 2, 12).Value < 0.99 And outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value And outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
```

```
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta
store2 = outp.Cells(N + 2, 12).Value
Loop
Debug.Print store2
ElseIf store > store2 Then
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta
Debug.Print outp.Cells(N + 2, 12).Value
End If
End If
store = outp.Cells(N + 2, 12).Value
Debug.Print store
If store < 0 Then
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta
store2 = outp.Cells(N + 2, 12).Value
Debug.Print store2
If store < store2 And store2 < 1 Then
store = store2
Debug.Print store - store2
Do Until 0.01 < outp.Cells(N + 2, 12).Value And outp.Cells(N + 2, 12).Value < 0.99 And outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value And outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta
store2 = outp.Cells(N + 2, 12).Value
Loop
ElseIf store > store2 Then
```

```
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta
store2 = outp.Cells(N + 2, 12).Value
Debug.Print outp.Cells(N + 2, 12).Value
End If
End If
store = outp.Cells(N + 2, 12).Value
If store > 1 Then
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta
store2 = outp.Cells(N + 2, 12).Value
Debug.Print store2
Debug.Print store
If store > store2 And store2 > 0 Then
store = store2
Debug.Print store - store2
Do Until 0.01 < outp.Cells(N + 2, 12).Value And outp.Cells(N + 2, 12).Value < 0.99 And outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value And outp.Cells(N + 2, 2).Value > 0.8 * par.Cells(23, 2).Value
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta
store2 = outp.Cells(N + 2, 12).Value
Loop
ElseIf store < store2 Then
outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta
outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta
outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta
outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta
store2 = outp.Cells(N + 2, 12).Value
```

```
Debug.Print store2

End If

End If

store = outp.Cells(N + 2, 12).Value
Debug.Print store

If store > 1 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta

store2 = outp.Cells(N + 2, 12).Value

Debug.Print store2

If store > store2 And store2 > 0 Then

store = store2

Debug.Print store - store2

Do Until 0.01 < outp.Cells(N + 2, 12).Value And outp.Cells(N + 2, 12).Value < 0.99 And outp.Cells(N + 2, 2).Value < 1.2 * par.Cells(22, 2).Value And outp.Cells(N + 2, 2).Value > 0.8
* par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value + Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value + 1 / 2 * Delta

store2 = outp.Cells(N + 2, 12).Value

Loop


ElseIf store < store2 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 2, 2).Value - Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 3, 2).Value - 1 / 2 * Delta

outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value = outp.Cells(((N + 2) + (0.5 * (Ti - N - 2))) + 1, 2).Value - 1 / 2 * Delta

store2 = outp.Cells(N + 2, 12).Value

Debug.Print outp.Cells(N + 2, 12).Value

End If

End If

Next

'*************Next Intervals****************

For N = Ti - 5 To Ti - 2
```

```
store = outp.Cells(N + 2, 12).Value

If store < 0 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta

store2 = outp.Cells(N + 2, 12).Value

Debug.Print store2

If store < store2 And store2 < 1 Then


Do Until 0.01 < outp.Cells(N + 2, 12).Value And outp.Cells(N + 2, 12).Value < 0.99 And outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value And outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta

store2 = outp.Cells(N + 2, 12).Value

Loop


ElseIf store > store2 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta

Debug.Print outp.Cells(N + 2, 12).Value

End If
End If

store = outp.Cells(N + 2, 12).Value

If store < 0 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta

store2 = outp.Cells(N + 2, 12).Value



If store < store2 And store2 < 1 Then


Do Until 0.01 < outp.Cells(N + 2, 12).Value And outp.Cells(N + 2, 12).Value < 0.99 And outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value And outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta

store2 = outp.Cells(N + 2, 12).Value

Loop
```

```
ElseIf store2 > store2 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta

store2 = outp.Cells(N + 2, 12).Value

Debug.Print outp.Cells(N + 2, 12).Value

End If
End If

store = outp.Cells(N + 2, 12).Value

If store > 1 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta

store2 = outp.Cells(N + 2, 12).Value

If store > store2 And store2 > 0 Then

store = store2

Do Until 0.01 < outp.Cells(N + 2, 12).Value And outp.Cells(N + 2, 12).Value < 0.99 And outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value And outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta

store2 = outp.Cells(N + 2, 12).Value

Loop

ElseIf store < store2 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta

store2 = outp.Cells(N + 2, 12).Value

End If
End If

store = outp.Cells(N + 2, 12).Value

If store > 1 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta

store2 = outp.Cells(N + 2, 12).Value

If store > store2 And store2 > 0 Then
```

```
store = store2

Do Until 0.01 < outp.Cells(N + 2, 12).Value And outp.Cells(N + 2, 12).Value < 0.99 And outp.Cells(N + 2, 2).Value > 1.2 * par.Cells(22, 2).Value And outp.Cells(N + 2, 2).Value < 0.8 * par.Cells(23, 2).Value

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value - Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value - 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value - 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value + Delta

store2 = outp.Cells(N + 2, 12).Value

Loop

ElseIf store < store2 Then

outp.Cells(N + 2, 2).Value = outp.Cells(N + 2, 2).Value + Delta

outp.Cells(N + 1, 2).Value = outp.Cells(N + 1, 2).Value + 1 / 2 * Delta

outp.Cells(N + 3, 2).Value = outp.Cells(N + 3, 2).Value + 1 / 2 * Delta

outp.Cells(Ti - 1, 2).Value = outp.Cells(Ti - 1, 2).Value - Delta

store2 = outp.Cells(N + 2, 12).Value

End If
End If

Next

Loop

Loop


'If outp.Cells(339, 21).Value <> 0 Then GoTo Line1


'Determine how many seconds code took to run
MinutesElapsed = Format((Timer - StartTime) / 86400, "hh:mm:ss")

'Notify user in seconds

MsgBox "This code ran in " & MinutesElapsed & "minutes", vbInformation
'Next

'***************************************************************************
Exit Sub
'//////////////////////////////////
'Overflow:
'MsgBox "Tank Volume exceeded boundaries. Please run simulation again with a smaller step size."        'This is the message box I was referring to'
'End
'//////////////////////////////////

End Sub


Sub inflow()
Dim par As Worksheet
Dim inp As Worksheet
Dim profile As Worksheet
Dim rowid As Integer
Dim intervals As Integer
Dim N As Integer
Dim days As Integer

Set par = ThisWorkbook.Sheets("Parameters")
Set inp = ThisWorkbook.Sheets("Historical data")
Set profile = ThisWorkbook.Sheets("24hr Influent")
intervals = par.Cells(4, 2).Value
days = par.Cells(5, 2).Value


If intervals = 12 Then GoTo twelve_intervals
If intervals = 24 Then GoTo twenty_four_intervals
If intervals = 48 Then GoTo forty_eight_intervals
'////////////////////////////////////////////////////////////
twelve_intervals:
MsgBox "Input values will use the values for every 2nd hour.", , "Warning!"
```

```
For N = 1 To days
For rowid = intervals * (N - 1) + 1 To intervals * N

    inp.Cells(rowid + 2, 2).Value = (profile.Cells(2 * rowid + 1, 2).Value + profile.Cells(rowid + 1, 2).Value) / 2
    inp.Cells(rowid + 2, 4).Value = (profile.Cells(2 * rowid + 1, 4).Value + profile.Cells(rowid + 1, 4).Value) / 2
    inp.Cells(rowid + 2, 5).Value = (profile.Cells(2 * rowid + 1, 5).Value + profile.Cells(rowid + 1, 5).Value) / 2
    inp.Cells(rowid + 2, 6).Value = (profile.Cells(2 * rowid + 1, 6).Value + profile.Cells(rowid + 1, 6).Value) / 2
    inp.Cells(rowid + 2, 7).Value = (profile.Cells(2 * rowid + 1, 7).Value + profile.Cells(rowid + 1, 7).Value) / 2
    inp.Cells(rowid + 2, 8).Value = (profile.Cells(2 * rowid + 1, 8).Value + profile.Cells(rowid + 1, 8).Value) / 2
    inp.Cells(rowid + 2, 9).Value = (profile.Cells(2 * rowid + 1, 9).Value + profile.Cells(rowid + 1, 9).Value) / 2
    inp.Cells(rowid + 2, 10).Value = (profile.Cells(2 * rowid + 1, 10).Value + profile.Cells(rowid + 1, 10).Value) / 2

Next
Next
'inp.Protect
Exit Sub
'//////////////////////////////////////////////////////////////
twenty_four_intervals:
For N = 1 To days
For rowid = intervals * (N - 1) + 1 To intervals * N

    inp.Cells(rowid + 2, 2).Value = profile.Cells(rowid + 1, 2).Value
    inp.Cells(rowid + 2, 4).Value = profile.Cells(rowid + 1, 4).Value
    inp.Cells(rowid + 2, 5).Value = profile.Cells(rowid + 1, 5).Value
    inp.Cells(rowid + 2, 6).Value = profile.Cells(rowid + 1, 6).Value
    inp.Cells(rowid + 2, 7).Value = profile.Cells(rowid + 1, 7).Value
    inp.Cells(rowid + 2, 8).Value = profile.Cells(rowid + 1, 8).Value
    inp.Cells(rowid + 2, 9).Value = profile.Cells(rowid + 1, 9).Value
    inp.Cells(rowid + 2, 10).Value = profile.Cells(rowid + 1, 10).Value
Next
Next

'inp.Protect
Exit Sub
'//////////////////////////////////////////////////////////////
forty_eight_intervals:

MsgBox "Input values will be interpolated.", , "Warning!"
For N = 1 To days
For rowid = intervals * (N - 1) + 1 To intervals * N
    If Int(rowid / 2) = rowid / 2 Then
        inp.Cells(rowid + 2, 2).Value = profile.Cells(rowid / 2 + 1, 2).Value
        inp.Cells(rowid + 2, 4).Value = profile.Cells(rowid / 2 + 1, 4).Value
        inp.Cells(rowid + 2, 5).Value = profile.Cells(rowid / 2 + 1, 5).Value
        inp.Cells(rowid + 2, 6).Value = profile.Cells(rowid / 2 + 1, 6).Value
        inp.Cells(rowid + 2, 7).Value = profile.Cells(rowid / 2 + 1, 7).Value
        inp.Cells(rowid + 2, 8).Value = profile.Cells(rowid / 2 + 1, 8).Value
        inp.Cells(rowid + 2, 9).Value = profile.Cells(rowid / 2 + 1, 9).Value
        inp.Cells(rowid + 2, 10).Value = profile.Cells(rowid / 2 + 1, 10).Value
    ElseIf rowid = 1 Then

        inp.Cells(rowid + 2, 2).Value = (profile.Cells((rowid + 1) / 2 + 1, 2).Value + profile.Cells(24 * days + 1, 2).Value) / 2
        inp.Cells(rowid + 2, 4).Value = (profile.Cells((rowid + 1) / 2 + 1, 4).Value + profile.Cells(24 * days + 1, 4).Value) / 2
        inp.Cells(rowid + 2, 5).Value = (profile.Cells((rowid + 1) / 2 + 1, 5).Value + profile.Cells(24 * days + 1, 5).Value) / 2
        inp.Cells(rowid + 2, 6).Value = (profile.Cells((rowid + 1) / 2 + 1, 6).Value + profile.Cells(24 * days + 1, 6).Value) / 2
        inp.Cells(rowid + 2, 7).Value = (profile.Cells((rowid + 1) / 2 + 1, 7).Value + profile.Cells(24 * days + 1, 7).Value) / 2
        inp.Cells(rowid + 2, 8).Value = (profile.Cells((rowid + 1) / 2 + 1, 8).Value + profile.Cells(24 * days + 1, 8).Value) / 2
        inp.Cells(rowid + 2, 9).Value = (profile.Cells((rowid + 1) / 2 + 1, 9).Value + profile.Cells(24 * days + 1, 9).Value) / 2
        inp.Cells(rowid + 2, 10).Value = (profile.Cells((rowid + 1) / 2 + 1, 10).Value + profile.Cells(24 * days + 1, 10).Value) / 2

    Else
        inp.Cells(rowid + 2, 2).Value = (profile.Cells((rowid + 1) / 2 + 1, 2).Value + profile.Cells(CInt((rowid + 1) / 2), 2).Value) / 2
        inp.Cells(rowid + 2, 4).Value = (profile.Cells((rowid + 1) / 2 + 1, 4).Value + profile.Cells(CInt((rowid + 1) / 2), 4).Value) / 2
        inp.Cells(rowid + 2, 5).Value = (profile.Cells((rowid + 1) / 2 + 1, 5).Value + profile.Cells(CInt((rowid + 1) / 2), 5).Value) / 2
        inp.Cells(rowid + 2, 6).Value = (profile.Cells((rowid + 1) / 2 + 1, 6).Value + profile.Cells(CInt((rowid + 1) / 2), 6).Value) / 2
        inp.Cells(rowid + 2, 7).Value = (profile.Cells((rowid + 1) / 2 + 1, 7).Value + profile.Cells(CInt((rowid + 1) / 2), 7).Value) / 2
        inp.Cells(rowid + 2, 8).Value = (profile.Cells((rowid + 1) / 2 + 1, 8).Value + profile.Cells(CInt((rowid + 1) / 2), 8).Value) / 2
        inp.Cells(rowid + 2, 9).Value = (profile.Cells((rowid + 1) / 2 + 1, 9).Value + profile.Cells(CInt((rowid + 1) / 2), 9).Value) / 2
        inp.Cells(rowid + 2, 10).Value = (profile.Cells((rowid + 1) / 2 + 1, 10).Value + profile.Cells(CInt((rowid + 1) / 2), 10).Value) / 2
    End If
Next
Next
inp.Protect
End Sub
Sub clear()
    Application.ScreenUpdating = True
    Dim outp As Worksheet
    Dim inp As Worksheet
    Dim prof As Worksheet
    Dim id As Integer
    Dim par As Worksheet
    Dim graphs As Worksheet
    Dim err As Worksheet
```

```
    Dim start As Worksheet
    Dim act As Worksheet
    Set start = ThisWorkbook.Sheets("Buttons")
    Set err = ThisWorkbook.Sheets("Parameters")
    Set graphs = ThisWorkbook.Sheets("Graphs")
    Set par = ThisWorkbook.Sheets("Parameters")
    Set outp = ThisWorkbook.Sheets("Outputs")
    Set inp = ThisWorkbook.Sheets("Historical data")
    Set prof = ThisWorkbook.Sheets("24hr Influent")
    Set act = ThisWorkbook.Sheets("Actual flows")

    outp.Unprotect
    inp.Unprotect
    prof.Unprotect

    par.Cells(9, 2).Value = 6

      outp.Range("B2:B338").Value = vbNullString
      outp.Range("D2:J2").Value = vbNullString
      inp.Range("B3:B338").Value = vbNullString
      inp.Range("D2:J338").Value = vbNullString
      prof.Range("B2:B338").Value = vbNullString
      prof.Range("D2:J338").Value = vbNullString


    start.Unprotect
    err.Unprotect
    graphs.Unprotect
    outp.Unprotect
    inp.Unprotect
    par.Unprotect

    Application.ScreenUpdating = True

  End Sub
```

Water and Mass Balance over the equalization tank

After balancing the flow at 70% initial tank volume, the water and mass balanced were both satisfactory at 99.99%.

The Simpson rule was used to get the flow averages.

$$\frac{d}{3} = [Firsi + Last + 4 \times \sum Even + 2 \times \sum Odd]$$

Table 16: Water Balance over equalization tank

Term	Values	Units
d	0,5	
Last value rowid	338	
Last value	18,5	
Area under curve	3106,5	Ml
Volume conserved in tank	0,5	Ml
Water Balance	99,99%	

Where water balance is obtained using the following

$$\frac{Volume_{out} + Volume\ in\ tank}{Volume_{in}} \times 100$$

COD Mass Balance

Table 17: COD Mass balance over equalization tank

Term	Value	Unit
d	0,5	
Last value rowid	338	
Last value	0	
Area under curve	3370906,07	kg
Mass conserved in tank	106000	kg
COD Mass Balance	100%	

Where COD balance is obtained from;

$$\frac{COD\ Flux_{out} + COD\ mass\ in\ tank}{COD\ Flux_{in}} \times 100$$

Appendix C Scenarios parameters , inflow data and capital cost functions

Table 18: Parameters used in Scenario 1

Parameter	Value	Unit
Minimum average temperature	16	oC
Maximum average temperature	24	oC
Max specific growth rate of nitrifiers	0.45	/d
Factor of safety for nitrification	1.25	
PST underflow rate	0.893*ADWF	
Balanced Sludge retention time	11.2	days
Maximum TSS	6%	
Maximum OUR of aerators	250	mgO/(l.h)
Diluted Sludge Volume Index	120	ml/g
1DFT correction factor	0.8	
PWWF/ADWF ratio	1.1	
Maximum practical a-recycle ratio	6.0	
Altitude	200	m
Reactor dissolved oxygen concentration (Activated Sludge)	2.0	mgO/l
Shaft to water standard oxygen transfer rate	2.6	kgO/kWh
Alpha (For aeration power)	0.85	
Beta (For aeration power)	0.95	
Line to shaft efficiency (Aeration)	0.9	
Max PST overflow rate at ADWF	1.2	m/h
Max PST overflow rate at PWWF	2.4	m/h
Recommended Primary sludge loading rates	80-120	kgTSS/m2d
Specific oxygen utilization rate of WAS	1.5	gO/(kgVSS.h)
Maximum specific hydrolysis rate	3.43	gCODBPO/(gCOD Acidogens)
Half saturation coefficient	6.76	gCOD/l
AD SRT	10	days
Biomass Yield (YAD)	0.113	gCOD biomass/gCOD organics
Endogenous respiration rate (bAD)	0.041	gCOD biomass/gCOD biomass.d
Biomass unbiodegradable fraction (fAD)	0.000	COD unbiodeg biomass/gCOD biomass
Methane gas properties of AD	10	kWh/Nm^3
Thermal efficiency of gas engine	45%	
Heat loss from AD	30	W/m^3 AD volume
Digested PS dewatered to	20% TSS	

The inflow data can be found in the table below;

Table 19: Inflow data

Interval	Inflow (BSM2 Data) (Ml/d)	COD (mg/l)
0	14,5	0
1	13,6	1150
2	12,3	1108
3	11,2	900
4	10,3	739
5	9,4	600
6	8,7	443
7	8,1	446
8	7,6	449
9	7,4	700
10	7,6	844
11	7,7	900
12	7,5	1161
13	8,6	1200
14	11,3	1309
15	14,7	1330
16	18,5	1341
17	23,7	1400
18	29,6	1478
19	31,3	1405
20	29,8	1372
21	28,3	1300
22	26,8	1267
23	24,4	1240
24	21,6	1214
25	20,3	1150
26	20,2	1108
27	19,9	900
28	19,6	739
29	19,3	600
30	19,2	443
31	19,9	446
32	21,3	449
33	22,4	700
34	23,4	844
35	25,0	900

36	27,0	1161
37	29,5	1200
38	32,1	1309
39	33,3	1330
40	33,5	1341
41	32,8	1400
42	31,4	1478
43	28,7	1405
44	25,6	1372
45	22,8	1300
46	20,4	1267
47	18,2	1240
48	16,2	1214
49	15,4	1150
50	15,6	1108
51	15,4	900
52	14,7	739
53	14,0	600
54	13,3	443
55	12,4	446
56	11,3	449
57	10,4	700
58	9,6	844
59	9,0	900
60	8,5	1161
61	8,5	1200
62	9,2	1309
63	10,7	1330
64	13,2	1341
65	18,1	1400
66	25,6	1478
67	29,4	1405
68	30,1	1372
69	29,3	1300
70	27,3	1267
71	25,7	1240
72	24,5	1214
73	23,2	1150
74	21,8	1108
75	20,2	900

76	18,5	739
77	17,2	600
78	16,3	443
79	15,4	446
80	14,7	449
81	14,7	700
82	15,5	844
83	17,8	900
84	21,7	1161
85	24,7	1200
86	27,0	1309
87	27,6	1330
88	26,8	1341
89	25,4	1400
90	23,8	1478
91	22,1	1405
92	20,6	1372
93	19,5	1300
94	18,8	1267
95	18,0	1240
96	17,1	1214
97	15,7	1150
98	14,0	1108
99	12,5	900
100	11,2	739
101	10,3	600
102	9,9	443
103	9,5	446
104	9,1	449
105	8,8	700
106	8,4	844
107	8,4	900
108	8,7	1161
109	10,1	1200
110	12,5	1309
111	15,0	1330
112	17,5	1341
113	21,6	1400
114	26,8	1478
115	29,7	1405

116	30,7	1372
117	29,8	1300
118	27,7	1267
119	25,6	1240
120	23,8	1214
121	21,7	1150
122	19,8	1108
123	19,2	900
124	19,5	739
125	19,3	600
126	18,8	443
127	19,7	446
128	21,8	449
129	23,0	700
130	23,4	844
131	23,2	900
132	22,4	1161
133	23,4	1200
134	25,6	1309
135	26,2	1330
136	25,6	1341
137	25,3	1400
138	24,9	1478
139	23,6	1405
140	21,7	1372
141	20,9	1300
142	21,0	1267
143	21,1	1240
144	21,3	1214
145	20,4	1150
146	18,8	1108
147	17,1	900
148	15,3	739
149	14,2	600
150	13,8	443
151	13,6	446
152	13,5	449
153	13,2	700
154	12,6	844
155	11,9	900

156	11,1	1161
157	11,8	1200
158	14,1	1309
159	16,9	1330
160	20,1	1341
161	23,7	1400
162	27,2	1478
163	29,4	1405
164	30,3	1372
165	29,3	1300
166	27,3	1267
167	24,1	1240
168	20,6	1214
169	19,0	1150
170	18,8	1108
171	17,5	900
172	15,4	739
173	15,0	600
174	15,9	443
175	16,4	446
176	16,4	449
177	17,1	700
178	18,3	844
179	20,1	900
180	22,3	1161
181	23,7	1200
182	24,6	1309
183	25,0	1330
184	24,9	1341
185	24,2	1400
186	23,1	1478
187	21,6	1405
188	20,0	1372
189	19,4	1300
190	19,5	1267
191	18,3	1240
192	16,4	1214
193	14,8	1150
194	13,7	1108
195	12,5	900

196	11,2	739
197	10,1	600
198	9,1	443
199	9,2	446
200	10,3	449
201	11,4	700
202	12,3	844
203	12,5	900
204	12,1	1161
205	12,4	1200
206	13,5	1309
207	16,2	1330
208	20,5	1341
209	26,0	1400
210	32,5	1478
211	34,5	1405
212	33,2	1372
213	31,1	1300
214	28,7	1267
215	25,5	1240
216	22,3	1214
217	20,7	1150
218	20,2	1108
219	19,4	900
220	18,5	739
221	18,1	600
222	18,2	443
223	18,1	446
224	17,7	449
225	17,9	700
226	18,5	844
227	20,3	900
228	23,2	1161
229	26,2	1200
230	29,1	1309
231	30,6	1330
232	31,1	1341
233	29,7	1400
234	27,4	1478
235	25,4	1405

236	23,9	1372
237	22,0	1300
238	20,1	1267
239	18,2	1240
240	16,5	1214
241	14,9	1150
242	13,3	1108
243	12,2	900
244	11,3	739
245	10,4	600
246	9,4	443
247	8,6	446
248	8,0	449
249	7,7	700
250	7,7	844
251	7,5	900
252	7,1	1161
253	7,3	1200
254	8,1	1309
255	9,6	1330
256	11,7	1341
257	14,9	1400
258	19,1	1478
259	21,9	1405
260	23,2	1372
261	22,8	1300
262	21,1	1267
263	19,6	1240
264	18,4	1214
265	17,2	1150
266	15,9	1108
267	15,5	900
268	15,6	739
269	15,0	600
270	13,8	443
271	13,3	446
272	13,6	449
273	14,3	700
274	15,6	844
275	17,2	900

276	18,9	1161
277	20,1	1200
278	20,8	1309
279	20,9	1330
280	20,4	1341
281	20,7	1400
282	21,7	1478
283	21,0	1405
284	19,1	1372
285	17,1	1300
286	15,1	1267
287	13,3	1240
288	11,9	1214
289	10,7	1150
290	9,6	1108
291	9,2	900
292	9,2	739
293	9,9	600
294	11,1	443
295	11,9	446
296	12,2	449
297	12,8	700
298	13,4	844
299	13,2	900
300	12,0	1161
301	11,5	1200
302	11,6	1309
303	13,0	1330
304	15,9	1341
305	19,8	1400
306	24,7	1478
307	26,7	1405
308	26,3	1372
309	25,6	1300
310	24,6	1267
311	23,2	1240
312	21,9	1214
313	20,4	1150
314	19,1	1108
315	17,5	900

316	15,9	739
317	14,9	600
318	14,3	443
319	14,1	446
320	14,1	449
321	14,7	700
322	15,9	844
323	17,5	900
324	19,4	1161
325	21,5	1200
326	23,6	1309
327	25,6	1330
328	27,4	1341
329	26,4	1400
330	23,9	1478
331	21,4	1405
332	19,2	1372
333	17,4	1300
334	16,0	1267
335	14,8	1240
336	13,9	1214

The cost function and the table below (Ekama, 2018) were used to cost the various unit processes in Scenario 1.

$$Cost = Constant \times Parameter^{Power}$$

Table 20: Capital cost functions for various Unit Operations (Ekama,2008)

Unit Operation	Parameter	Units	Minimum Size	Maximum Size	Constant	Power
Flow balancing tank	Volume	Megalitre	2	16	350	0.9
PST, SST and gravity thickener	Diameter	Metre	15	40	30	1.212
Biological reactors	Volume	Megalitre	1	16	770	0.761
Aeration	Power	kW	30	110	40	0.55
Anaerobic Digester	Volume	Megalitre	1	10	1500	0.85
Flotation	Diameter	Metre	10	35	150	1.05

Appendix D Scenario 1 summary tables

Wastewater characterisation

The summary tables for the wastewater characterisation for the unbalanced flow can be seen below;

Table 21: COD Characterisation (Unbalanced Flow)

Chemical Oxygen Demand (COD) Characterisation					
Parameter	**Symbol**	**Raw**	**Settled**	**PS**	**Units**
Total Influent COD	S_{ti}	1122,49	614,00	57555,95	mg/l
Influent biodegradable soluble	S_{bsi}	235,50	235,50	235,50	mg/l
VFA	S_{bsai}	50,00	50,00	50,00	mg/l
FBSO	S_{bsfi}	185,50	185,50	185,50	mg/l
Influent biodegradable particulate	S_{bpi}	683,57	301,35	43102,66	mg/l
Settleable	$S_{bpi,Set}$	382,22	0,00		mg/l
Non-Settleable	$S_{bpi,NSet}$	301,35	301,35		mg/l
Influent unbiodegradable soluble	S_{usi}	57,50	57,50	57,50	mg/l
Influent unbiodegradable particulate	S_{upi}	145,92	19,65	14160,29	mg/l
Settleable	$S_{upi,Set}$	126,28	0,00		mg/l
Non-Settleable	$S_{upi,NSet}$	19,65	19,65		mg/l
Influent total biodegradable	S_{bi}	919,06	536,85	43338,16	mg/l
Influent total unbiodegradable	S_{ui}	203,43	77,15	14217,79	mg/l
Influent total particulate	S_{pi}	829,49	321,00	57262,95	mg/l
Influent total soluble	S_{si}	293,00	293,00	293,00	mg/l

Table 22: Organic Carbon Characterisation (Unbalanced Flow)

Organic Carbon Characterisation					
Parameter	**Symbol**	**Raw**	**Settled**	**PS**	**Units**
Total Influent TOC	C_{ti}	383,10	205,00	20149,01	mg/l
Influent biodegradable soluble	C_{bsi}	80,32	80,32	80,32	mg/l
VFA	C_{bsai}	18,75	18,75	18,75	mg/l
FBSO	C_{bsfi}	61,57	61,57	61,57	mg/l
Influent biodegradable particulate	C_{bpi}	232,56	98,63	15096,75	mg/l
Settleable	$C_{bpi,Set}$	133,93	0,00		mg/l
Non-Settleable	$C_{bpi,NSet}$	98,63	98,63		mg/l
Influent unbiodegradable soluble	C_{usi}	19,18	19,18	19,18	mg/l
Influent unbiodegradable particulate	C_{upi}	51,04	6,87	4952,75	mg/l
Settleable	$C_{upi,Set}$	44,17	0,00		mg/l
Non-Settleable	$C_{upi,NSet}$	6,87	6,87		mg/l
Influent total biodegradable	C_{bi}	312,88	178,95	15177,07	mg/l
Influent total unbiodegradable	C_{ui}	70,22	26,05	4971,93	mg/l
Influent total particulate	C_{pi}	283,60	105,50	20049,51	mg/l
Influent total soluble	C_{si}	99,50	99,50	99,50	mg/l

Table 23: Nitrogen Characterisation (Unbalanced Flow)

Nitrogen Characterisation					
Parameter	**Symbol**	**Raw**	**Settled**	**PS**	**Units**
Total Influent TKN	N_{ti}	89,90	71,90	2087,58	mg/l
Influent inorganic N (FSA)	N_{ai}	57,36	57,36	57,36	mg/l
Influent organic N	N_{oi}	32,54	14,54	2030,22	mg/l
Influent organic biodegradable soluble	N_{obsi}	5,24	5,24	5,24	mg/l
Influent organic biodegradable particulate	N_{obpi}	16,45	6,97	1067,85	mg/l
Settleable	$N_{obpi,Set}$	9,47	0,00		mg/l
Non-Settleable	$N_{obsi,NSet}$	6,97	6,97		mg/l
Influent organic unbiodegradable soluble	N_{ousi}	1,00	1,00	1,00	mg/l
Influent organic unbiodegradable particulate	N_{oupi}	9,85	1,33	956,13	mg/l
Settleable	$N_{oupi,Set}$	8,53	0,00		mg/l
Non-Settleable	$N_{oupi,NSet}$	1,33	1,33		mg/l
Influent total organic biodegradable	N_{obi}	21,69	12,22	1073,09	mg/l
Influent total organic unbiodegradable	N_{oui}	10,85	2,33	957,13	mg/l
Influent total particulate	N_{tpi}	26,30	8,30	2023,98	mg/l
Influent total soluble	N_{tsi}	63,60	63,60	63,60	mg/l

Table 24: Phosphorus Characterisation (Unbalanced Flow)

Phosphorus Characterisation					
Parameter	**Symbol**	**Raw**	**Settled**	**PS**	**Units**
Total Influent P	P_{ti}	20,07	16,44	422,93	mg/l
Influent inorganic P	P_{ai}	13,72	13,72	13,72	mg/l
Influent organic P	P_{oi}	6,35	2,72	409,21	mg/l
Influent organic biodegradable soluble	P_{obsi}	1,32	1,32	1,32	mg/l
Influent organic biodegradable particulate	P_{obpi}	2,57	1,07	168,86	mg/l
Settleable	$P_{obpi,Set}$	1,50	0,00		mg/l
Non-Settleable	$P_{obsi,NSet}$	1,07	1,07		mg/l
Influent organic unbiodegradable soluble	P_{ousi}	0,00	0,00	0,00	mg/l
Influent organic unbiodegradable particulate	P_{oupi}	2,46	0,33	239,03	mg/l
Settleable	$P_{oupi,Set}$	2,13	0,00		mg/l
Non-Settleable	$P_{oupi,NSet}$	0,33	0,33		mg/l
Influent total organic biodegradable	P_{obi}	3,88	2,39	170,18	mg/l
Influent total organic unbiodegradable	P_{oui}	2,46	0,33	239,03	mg/l
Influent total particulate	P_{tpi}	5,03	1,40	407,89	mg/l
Influent total soluble	P_{tsi}	15,04	15,04	15,04	mg/l

The summary tables for the wastewater characterisation for the balanced flow (70% initial volume) can be seen below;

Table 25: COD Characterisation (Balanced Flow)

Chemical Oxygen Demand (COD) Characterisation					
Parameter	**Symbol**	**Raw**	**Settled**	**PS**	**Units**
Total Influent COD	S_{ti}	1054,32	614,00	49922,27	mg/l
Influent biodegradable soluble	S_{bsi}	235,50	235,50	235,50	mg/l
VFA	S_{bsai}	50,00	50,00	50,00	mg/l
FBSO	S_{bsfi}	185,50	185,50	185,50	mg/l
Influent biodegradable particulate	S_{bpi}	624,26	301,35	36461,36	mg/l
Settleable	$S_{bpi,Set}$	322,91	0,00		mg/l
Non-Settleable	$S_{bpi,NSet}$	301,35	301,35		mg/l
Influent unbiodegradable soluble	S_{usi}	57,50	57,50	57,50	mg/l
Influent unbiodegradable particulate	S_{upi}	137,06	19,65	13167,91	mg/l
Settleable	$S_{upi,Set}$	117,41	0,00		mg/l
Non-Settleable	$S_{upi,NSet}$	19,65	19,65		mg/l
Influent total biodegradable	S_{bi}	859,76	536,85	36696,86	mg/l
Influent total unbiodegradable	S_{ui}	194,57	77,15	13225,41	mg/l
Influent total particulate	S_{pi}	761,32	321,00	49629,27	mg/l
Influent total soluble	S_{si}	293,00	293,00	293,00	mg/l

Table 26: Organic Carbon Characterisation (Balanced Flow)

Organic Carbon Characterisation					
Parameter	**Symbol**	**Raw**	**Settled**	**PS**	**Units**
Total Influent TOC	C_{ti}	383,10	205,00	20149,01	mg/l
Influent biodegradable soluble	C_{bsi}	80,32	80,32	80,32	mg/l
VFA	C_{bsai}	18,75	18,75	18,75	mg/l
FBSO	C_{bsfi}	61,57	61,57	61,57	mg/l
Influent biodegradable particulate	C_{bpi}	235,66	98,63	15443,85	mg/l
Settleable	$C_{bpi,Set}$	137,03	0,00		mg/l
Non-Settleable	$C_{bpi,NSet}$	98,63	98,63		mg/l
Influent unbiodegradable soluble	C_{usi}	19,18	19,18	19,18	mg/l
Influent unbiodegradable particulate	C_{upi}	47,94	6,87	4605,66	mg/l
Settleable	$C_{upi,Set}$	41,07	0,00		mg/l
Non-Settleable	$C_{upi,NSet}$	6,87	6,87		mg/l
Influent total biodegradable	C_{bi}	315,98	178,95	15524,17	mg/l
Influent total unbiodegradable	C_{ui}	67,12	26,05	4624,84	mg/l
Influent total particulate	C_{pi}	283,60	105,50	20049,51	mg/l
Influent total soluble	C_{si}	99,50	99,50	99,50	mg/l

Table 27: Nitrogen Characterisation (Balanced Flow)

Nitrogen Characterisation					
Parameter	**Symbol**	**Raw**	**Settled**	**PS**	**Units**
Total Influent TKN	N_{ti}	89,90	71,90	2087,58	mg/l
Influent inorganic N (FSA)	N_{ai}	53,53	53,53	53,53	mg/l
Influent organic N	N_{oi}	36,37	18,37	2034,05	mg/l
Influent organic biodegradable soluble	N_{obsi}	9,07	9,07	9,07	mg/l
Influent organic biodegradable particulate	N_{obpi}	17,05	6,97	1134,85	mg/l
Settleable	$N_{obpi,Set}$	10,07	0,00		mg/l
Non-Settleable	$N_{obsi,NSet}$	6,97	6,97		mg/l
Influent organic unbiodegradable soluble	N_{ousi}	1,00	1,00	1,00	mg/l
Influent organic unbiodegradable particulate	N_{oupi}	9,25	1,33	889,12	mg/l
Settleable	$N_{oupi,Set}$	7,93	0,00		mg/l
Non-Settleable	$N_{oupi,NSet}$	1,33	1,33		mg/l
Influent total organic biodegradable	N_{obi}	26,12	16,05	1143,93	mg/l
Influent total organic unbiodegradable	N_{oui}	10,25	2,33	890,12	mg/l
Influent total particulate	N_{tpi}	26,30	8,30	2023,98	mg/l
Influent total soluble	N_{tsi}	63,60	63,60	63,60	mg/l

Table 28: Phosphorus Characterisation (Balanced Flow)

Phosphorus Characterisation					
Parameter	**Symbol**	**Raw**	**Settled**	**PS**	**Units**
Total Influent P	P_{ti}	20,07	16,44	422,93	mg/l
Influent inorganic P	P_{ai}	12,84	12,84	12,84	mg/l
Influent organic P	P_{oi}	7,23	3,60	410,10	mg/l
Influent organic biodegradable soluble	P_{obsi}	2,20	2,20	2,20	mg/l
Influent organic biodegradable particulate	P_{obpi}	2,72	1,07	185,61	mg/l
Settleable	$P_{obpi,Set}$	1,65	0,00		mg/l
Non-Settleable	$P_{obsi,NSet}$	1,07	1,07		mg/l
Influent organic unbiodegradable soluble	P_{ousi}	0,00	0,00	0,00	mg/l
Influent organic unbiodegradable particulate	P_{oupi}	2,31	0,33	222,28	mg/l
Settleable	$P_{oupi,Set}$	1,98	0,00		mg/l
Non-Settleable	$P_{oupi,NSet}$	0,33	0,33		mg/l
Influent total organic biodegradable	P_{obi}	4,92	3,27	187,82	mg/l
Influent total organic unbiodegradable	P_{oui}	2,31	0,33	222,28	mg/l
Influent total particulate	P_{tpi}	5,03	1,40	407,89	mg/l
Influent total soluble	P_{tsi}	15,04	15,04	15,04	mg/l

Unit Processes for P1 and P2

Primary Sedimentation

Table 29: Primary Settling Tank (Unbalanced Flow/P1)

Parameter	Value	Unit
Overflow rate at ADWF	1,2	m/h
Overflow rate at PWWF	2,4	m/h
ADWF	20,5	Ml/d
PDWF	33,5	
PWWF/PDWF ratio	1,395	Given
PWWF	46,79	Ml/d
Max Area ADWF	712,70	m^2
Max Area PWWF	812,41	m^2
Design area	812,41	m^2
Diameter	32,16	m
Min Diameter	15	m
Max Diameter	40	m
Diameter Check	Ok	
Diameter Provided	32,16	m
Number of PST	1	
Cost	4027,64	x1000R
Actual Design Area	406,20	m^2

Table 30: Primary Settling Tank (Balanced Flow/P2)

Parameter	Value	Unit
Overflow rate at ADWF	1,2	m/h
Overflow rate at PWWF	2,4	m/h
ADWF	18,69	Ml/d
PDWF	21,10	
PWWF/PDWF ratio	1,395	Given
PWWF	29,44	Ml/d
Max Area ADWF	649,06	m^2
Max Area PWWF	511,10	m^2
Design area	649,06	m^2
Diameter	28,75	m
Min Diameter	15	m
Max Diameter	40	m
Diam Check	Ok	
Diameter Provided	28,75	m
Number of PST	1	
Cost	3515,38	x1000R
Actual Design Area	649,06	m^2

Activated Sludge Reactors

P1 Design

The biological reactor was designed at **16°C** which is the minimum winter temperature of the locality and at **7.8d** which. This allows the largest reactor sizes to be designed to account for the slow growth of nitrifiers at low temperatures. To design the biological reactor, the active sludge constants in the table below were used;

Table 31: Activated Sludge Constants

Activated Sludge Constants	Values	Units
Yhv	0.45	mgVSS/mgCOD
fh (Unbiodegradable fraction of active organisms)	0.2	
fcv	1.481	mgCOD/mgVSS
bh20 (endogenous respiration rate)	0.24	/d
bH16	0.214	/d
fn	0.1	mgN/mgVSS
fp	0.025	mgP/mgVSS
Sf (Nitrification safety factor)	1.25	
µAm20	0.45	/d
µAm16	0.283	/d
Kn20	1	mgFSA/l
Kn16	0.629	mgFSA/l
bA20	0.04	/d
bA16	0.0357	/d
YA	0.1	

The material loads on the plant were as follows;

Table 32: Material loads on plant

Materials Loads on Plant	Values	Units	Formula
Qi	20,3	Ml/d	
FSti	12490,2	kgCOD/d	Fsti=Qi*Sti
FNti	1462,6	kgN/d	Fnti=Qi*Nti
FPti	334,4	kgP/d	Fpti=Qi*Pti
FSupi	399,7	kgCOD/d	Fsupi=fs'up*Fsti
FSusi	1169,8	kgCOD/d	Fsusi=fs'us*Fsti
FXIi (Unbio Part Org load as VSS)	269,9	kgVSS/d	FXIi=(fs'up/fcv)*FSti
FSbi	10920,8	kgCOD/d	Fsbi=(1-fs'up-fs'us)*Fsti
FXIoi	587,8	kgISS/d	FXIoi=Qi*Xioi
FNai	1166,8	kgN/d	Fnai=Nai*Qi or fN'a*Fnti
FNoi	295,8	kgN/d	Fnoi=(Nti-Nai)*Qi or (1-fN'a)Fnti
FNoupi	27,0	kgN/d	Fnoupi=fn*(fs'up/fcv)*Fsti
FNousi	20,3	kgN/d	Nousi*Qi or Qi*(Nte-Nae)
FNobi	248,5	kgN/d	Qi(Nti-Nai-Nousi-Noupi)

Knowing the sludge age as well as the material loads on the plant, it was possible to obtain the masses, concentrations and active fraction of WAS with respect to the TSS and VSS in the activated sludge reactor as shown below;

Table 33: Activated sludge reactor masses and concentrations

Activated Sludge Reactor	Values	Units	Formula
Rs	11,2	d	
YH*Rs/(1+bh*Rs)	1,48		
Biodegradable COD remaining	0	mgCOD/l	Sb=YH*Rs/(1+bH*Rs)*(1/Kv)
Total filtered effluent COD (Suse)	57,50	mgCOD/l	Ste=Suse+Sb
Change in Bio COD flux	10920,78	kgCOD/d	FdelSb=Fsbi-Fsbe
MXBH	16200,12	kgVSS	MXBH=FdelSb*[YH*Rs/(1+bH*Rs)]
MXEH	7768,12	kgVSS	MXEH=fh*bh*Rs*MXBH
MXI	3022,62	kgVSS	MXI=FXIi*Rs
MXv	26990,86	kgVSS	MXv=MXBH+MXEH+MXI
MXIo	9013,77	kgISS	MXIo=0.15*MXBH+FXIoi*Rs
MXt	36004,64	kgTSS	MXt=MXv+MXIO
VSS/TSS ratio of sludge	0,75	VSS/TSS	MXv/MXt
Xt	4,9	gTSS/l	
Vr	7,35	Ml	v=Mxt/Xt
Vr per person	0,00001	l/person	
Rhn	0,36	d	Rhn=V/Qi
XBH	2204,73	mgVSS/l	XBH=MXBH/V
XEH	1057,19	mgVSS/l	
XI	411,36	mgVSS/l	
XV	3673,28	mgVSS/l	
XIO	1226,72	mgISS/l	
Xt	4900	mgTSS/l	
fav	0,60		
fat	0,45		
FNs	240,99	kgN/d	FNs=fn*FXv
Ns	11,85	mgN/l	Ns=FNs/Qi

The sludge wasted from the reactor was as follows;

Table 34: Sludge wastage

Sludge wastage	Values	Units
Qw	0.656	Ml/d
FXv	2409.9	kgVSS/d
FXt	3214.7	kgTSS/d

The nitrification calculations were as follows;

Table 35: Nitrification

Nitrification	Values	Units	Formula
SRTmin (smaller than actual Rs, hence ok)	4.044	d	SRTm=1/uAmT(1-fxt)-bAT
fxm	0.276		fxm= 1-Sf(bAt+1/Rs)/µamT
Nae	2.52	mgN/l	knT/(Sf-1)

A summary table of denitrification is found below;

Table 36: Denitrification summary table

Denitrification	Values	Units	Formula
fSb's	0,439		Sbsi/Sbi
Take fx1	0,448		
fx1min	0,1		fx1min=(1+bnRs)/(K1T*Yh*Rs)fsb's(1-fcvYh)/2.86
Nitrification capacity (Nc)	56,5		Nti-Ns-Nouse-Nae
Fon	5256,1		4.57*Qi*Nne
Sbi	536,8	mgCOD/l	Sbsi+Sbpi
DP1	53,9	mgNO3-N/l influent	Sbi*fSb's*((1-fcv*Yhv)/2.86)+(Sbi*K2*fx1*Yhv*Rs)/(1+bh*Rs)
Computing aopt			
A	0,7		Oa/2.86
B	4,3		Nc-Dp1+{(s+1)Oa+sOs}/2.86
C	50,7		(s+1)*(Dp1-sOs/2.86)-sNc
aopt	6,0		aopt=[-B+sqrt(B^2+4AC]/(2A)
aopt used	6,0		
Nnemin=Nneaopt	7,1	mgNO3-N/l	Nc/(a-recycle+1+s-recycle)
Ntemin	3,5	mgN/l	Nae+Nouse
Total Nitrogen in effluent	10,6	mgN/l	Nne+Nae+Nouse
%N removal	95,1	%	((Nti-TN)/Nti)*100
Anoxic reactor volume	3291,3	m^3	fx1*Vr
Aerobic reactor volume	4056,6	m^3	(1-fx1)*Vr
Oxygen recovered	2875,9	kgO/d	2.86*(Nc-Nne)*Qi
Fon	5256,1	kgO/d	4.57*Qi*Nc
Foc	7751,4	kgO/d	
Total oxygen demand	10131,6	kgO/d	Foc+Fon-Fod
OUR aerobic reactor	104,1	mgO/ (l.h)	
Influent Alk (Given)	300,0	mg/l as CaCO3	
Alk taken up (via Nitrification)	403,7	mg/l as CaCO3	7.14*(Nai-Nae)
Alk recovered (via denitrification)	176,5	mg/l as CaCO3	3.57*(Nc-Nne)

Alk taken up (via N in biomass and end res.)	37,6	mg/l as CaCO3	3.57*(Ns-Noupi)
OrgN Alk recovered	43,6	mg/l as CaCO3	3.57*(Nobpi+Nobsi)
Effluent Alk	78,8	mg/l a CaCO3	Inf Alk+Alk Rec-Alk Taken up+OrgN component-Alk N in bio & ER
Influent pH	7,2		

The following denitrification constant were used;

Table 37: Denitrification constants

Denitrification Constants	Values	Units
K_{220}	0.101	mgNO3-N/(mgOHOVSS.d)
Theta K_2	1.08	
K_{2T}	0.0742	mgNO3-N/(mgOHOVSS.d)
O_a	2	mgO/l
O_s	1	mgO/l
S recycle flow	1	
K_{120}	0.72	mgNO3-N/(mgOHOVSS.d)
Theta K_1	1.2	
K_{1T}	0.347	mgNO3-N/(mgOHOVSS.d)

100% mass balances for COD, N, P and TOD were obtained over the reactor.

Table 38: COD mass balance

COD Mass Balance	Values	Units	Formula
COD mass in	12490,24	kgCOD/d	Qi*Sti
Qe	19,69	Ml/d	Qi-Qw
Ste wasted	1169,78	kgCOD/d	Qi*Suse
MLSS COD wasted	3569,06	kgCOD/d	(MXv*fcv/Rs)
FOC	7751,41	kgO/d	
COD mass Out	12490,24	kgCOD/d	Eff+Sludge Waste+Foc
%COD balance	100	%	Out/In * 100

Table 39: Nitrogen mass balance

Nitrogen Mass Balance	Values	Units	Formula
Nitrogen mass in (Fnti)	1462,620	kgN/d	Qi*Nti
Fnouse	20,342	kgN/d	Qi*Nouse
Fnae	51,162	kgN/d	Qi*Nae
Fnne	144,562	kgN/d	Qi*Nne
FNs	240,990	kgN/d	FXv*fn
FN2	1005,564		Qi*N2
Nitrogen mass out	1462,62	kgN/d	Sum of all the above
% Nitrogen balance	100	%	Out/In *100

Table 40: Phosphorus mass balance

Phosphorus Mass Balance	Values	Units	Formula
P mass in (Fpti)	334,43	kgP/d	Qi*Pti
Fpouse	0,00	kgP/d	Qi*Pouse
FPs	60,25	kgP/d	FXv*fp
Ps	2,96	mgP/l	FPs/Qi
Pti	16,44	mgP/l	Given
Pte	13,48	mgP/l	Pti-Ps
Pouse	0,00	mgP/l	Given
Pae	13,48	mgP/l	Pte-Pouse
FPte	274,18	mgP/l	Qi*Pte
P mass out	334,43	kgP/d	Fpte+FPs
P mass Balance	100	%	Out/In * 100

Table 41: TOD mass balance

TOD Mass Balance	Values	Units	Formula
TOD in	19174,41	kgO/d	Qi(Sti+4.57Nti)
TOD out via Qw	48,27	kgO/d	Qw(Ste+4.57Nte)
TOD out via Qw	4670,38	kgO/d	Qw(CODsludge+4.57Ns)
TOD out via Qe	1448,28	kgO/d	Qe(Ste+4.57Nte)
TOD out via N2 gas	2875,91	kgO/d	Qi(Nc-Nne)*2.86
TOD out via oxygenation	10131,57	kgO/d	Foc+Fon-Fod=Fot
TOD out (Total)	19174,41	kgO/d	
TOD mass balance	100	%	Out/In * 100

P2 Design

The results from the AS reactor of the P2 design are as follows;

Table 42: Material loads for P2 design

Materials Loads on Plant	Values	Units
Qi	18,69	Ml/d
FSti	11477,50	kgCOD/d
FNti	1344,03	kgN/d
FPti	307,31	kgP/d
FSupi	367,28	kgCOD/d
FSusi	1074,93	kgCOD/d
FXIi (Unbio Part Org load as VSS)	247,99	kgVSS/d
FSbi	10035,29	kgCOD/d
FXIoi	540,17	kgISS/d
FNai	1072,18	kgN/d
FNoi	271,85	kgN/d
FNoupi	24,80	kgN/d
FNousi	18,69	kgN/d
FNobi	228,36	kgN/d

Table 43: Activated sludge reactor masses and concentrations for P2 design

Activated Sludge Reactor	Values	Units
Rs	11,2	d
YH*Rs/(1+bh*Rs)	1,48	
Biodegradable COD remaining	0	mgCOD/l
Total filtered effl COD (Suse)	57,50	mgCOD/l
Change in Bio COD flux	10035,29	kgCOD/d
MXBH	14886,56	kgVSS
MXEH	7138,26	kgVSS
MXI	2777,54	kgVSS
MXv	24802,36	kgVSS
MXIo	8282,91	kgISS
MXt	33085,27	kgTSS
VSS/TSS ratio of sludge	0,75	VSS/TSS
Xt	4,9	gTSS/l
Vr	6,75	Ml
Vr per person	0,00001	l/person
Rhn	0,36	d
XBH	2204,73	mgVSS/l
XEH	1057,19	mgVSS/l
XI	411,36	mgVSS/l
XV	3673,28	mgVSS/l
XIO	1226,72	mgISS/l
Xt	4900	mgTSS/l
fav	0,60	
fat	0,45	
FNs	221,45	kgN/d
Ns	11,85	mgN/l

Table 44: Sludge wastage for P1 design

Sludge wastage	Values	Units
Qw	0.603	Ml/d
FXv	2214.5	kgVSS/d
FXt	2954	kgTSS/d

Table 45: Denitrification summary table for P2 design

Denitrification	Values	Units
fSb's	0.439	
Take fx1	0.448	
fx1min	0.0549	
Nitrification capacity (Nc)	56.54	
Fon	4829,90	

Sbi	536,85	mgCOD/l
DP1	53,947	mgNO3-N/l influent
Computing aopt		
A	0,699	
B	4,340	
C	50,655	
aopt	5,956	
aprac	6	
aopt (After comparison with aprac)	5,956	
Nnemin=Nneaopt	7,11	mgNO3-N/l
Ntemin	3,52	mgN/l
Total Nitrogen in effluent	10,62	mgN/l
%N removal	95,111	%
Anoxic reactor volume	3024,4	m^3
Aerobic reactor volume	3727,7	m^3
Oxygen recovered	2642,73	kgO/d
Fon	4829,90	kgO/d
Foc	7122,90	kgO/d
Total oxygen demand	9310,07	kgO/d
OUR aerobic reactor	104,06	mgO/(l.h)
Influent Alk (Given)	300,00	mg/l as CaCO3
Alk taken up (via Nitrification)	403,68	mg/l as CaCO3
Alk recovered (via denitr)	176,47	mg/l as CaCO3
Alk taken up (via N in biomass and end res.)	37,56	mg/l as CaCO3
OrgN Alk recovered	43,61	mg/l as CaCO3
Effluent Alk	78,84	mg/l a CaCO3
Influent pH	7,20	

100% mass balances were obtained for the COD, N, P and TOD.

Secondary settling tank

The SST design involved a cost optimisation process, to get the reactor TSS concentration at minimum cost and respecting the constraint that each reactor module has the same number of SSTs. The procedure was as follows;

The flux rating (0.8), DSVI (120ml/g), PWWF/PDWF ratio (1.395), reactor cost equation (770(reactor volume)$^{0.761}$) and SST cost equation (30(SST diameter)$^{1.212}$) were used.

For the P1 design, T=the PDWF was obtained from the diurnal flow data (**889.9m^3/h**), and hence the PWWF could be obtained from 1.395*PDWF (**932.4m^3/h**). For the P2 design, the ADWF was used (**778.9m^3/h**) The SSVI was obtained from 0.67*DVI (**80.4ml/l**) and hence from the 3 equations; $\frac{V_o}{n} = 67.9e^{-0.016SSVI}, n = 0.88 - 0.393log\frac{V_o}{n}, V_o = \frac{V_o}{n}n$

V_o = **7.121m/h** and n=**0.38m^3/kg** were obtained. Starting at Xt of 1gTSS/l and knowing the MXt from the reactor, the corresponding total reactor volume could be computed from

MXt/Xt. Knowing the PWWF, the total SST area was computed as follows; $\frac{QiPWWF\ (\frac{m^3}{h})}{V_o e^{-nXt} * flux\ rating}$. The diameter could be obtained from there. Knowing the allowed maximum volume of 1 reactor (16000m^3) and the total reactor volume for the Xt given, the N^o of modules could be obtained from $\frac{Total\ reactor\ volume}{\max volume\ of\ 1\ reactor}$ and the volume of each reactor could be obtained $\frac{Total\ reactor\ volume}{No\ of\ modules}$. In a similar way, the N^o of SSTs is found as $\frac{Total\ SST\ area}{Maximum\ area\ limit}$ with the limit being **1256.64m^2** (from 40m diameter). The area of each SST is $\frac{Total\ SST\ area}{No\ of\ SSTs}$. The cost of the reactors and SSTs can then be computed for that specific concentration. This was done iteratively at increments of 0.05gTSS/l till the concentration at which the reactor modules had the same number of SSTs was obtained, at minimum cost.

P1 Design

Reactor concentration – **4.9gTSS/l** No of SSTs provided – **1**

Volume of 1 reactor – **7347.8m^3**No of reactor modules provided – **1**

Diameter of 1 SST – **36.6m**

P2 Design

Reactor concentration – **4.9gTSS/l** No of SSTs provided – **1**

Volume of 1 reactor – **6752.1m^3**No of reactor modules provided – **1**

Diameter of 1 SST – **35.1m**

The flux theory was used to obtain the SST recycle ratio and concentrations of the SST underflow. From the flow data, the ADWF, PDWF, PWWF and min DWF were obtained. Then the max overflow rate at PWWF and the minimum SST area were obtained.

Anoxic-aerobic digester

P1 design

To enable the WAS to be discharged to drying beds, an anoxic-aerobic digester was provided, allowing the WAS to be stable by respecting its SOUR (1.5gO/kgVSS.h) and its corresponding effluent active fraction found to be **0.11**. The digester designed was divided into sub-reactors and it was observed that having 3 sub-reactors provides the least digester volume (**852m^3**). This option was abandoned because the OUR in the 1st sub-reactor was greater than the max OUR for the aerators, **149.1mgO/l.h>125mgO/l.h** (half of 250 because of denitrification requiring the aeration cycle to be 3h on and 3h off). A digester with **2 sub-reactors** in series was adopted because it had the lowest area of the viable options (1 reactor only – **2183 m^3**, 2 sub-reactors in series – **1060m^3** total).

Fully aerobic digesters were designed identically as the anoxic-aerobic digesters (apart from the max OUR kept at 250mgO/l.h as no denitrification was required and the aeration cycle had no interruptions). Fully aerobic digesters were discarded because they had higher oxygen demand and hence OUR (as no oxygen was recovered). The table below provides the OUR values;

Table 46: Aerobic Digestion Summary Table

Completely Aerobic Digestion				
Number of Aerobic compartments	**Volume of 1 digester (m^3)**	**Total Volume (m^3)**	**Oxygen Demand (Total kgO/d)**	**OUR (highest, in 1st reactor mgO/l)**
1	2182.7	2182.7	2011.9	38.4
2	530.1	1060.1	2011.9	119.7
3	283.9	851.6	2011.9	174.9

Table 47: Anoxic-aerobic summary table

Aerobic-Anoxic Digestion					
Number of Aerobic-Anoxic compartments	**Volume of 1 digester (m^3)**	**Total Volume (m^3)**	**Oxygen Demand (Total kgO/d)**	**OUR (highest, in 1st reactor mgO/l)**	**fave actual**
1	2182.7	2182.7	1714.96	32.7	0.11
2	530.1	1060.1	1714.96	102.21	0.11
3	283.9	851.6	1714.96	149.1	0.11

A mass balance over the digester was performed, giving 100% balances for TOD, COD, N & P.

Table 48: Summary table of double anoxic-aerobic digester

Double anoxic-aerobic digester			
Item	**Value**	**Unit**	**Source**
Maximum OUR in the 1st or only reactor	250	mgO/l.h	Design Parameter
Thickener maximum conc	6%	TSS	Design Parameter
Qw	656.1	m3/d	From AS Design
FXt	3214.7	kgTSS/d	From AS Design
Xt (before thickening)	4.9	kgTSS/m3	FXt/Qw
fati	0.45	-	From AS Design
favi	0.60	-	From AS Design
fi - VSS/TSS ratio	0.75	gVSS/gTSS	From AS Design
Required SOUR (including nitrification)	1.5	mgO/gVSS.h	Design Parameter
fave - Effluent active fraction (incl. Nit)	0.11	-	24*SOUR/[1000*(fcv + 4.57*fn)*(1-fH)*bH]
Alpha	8.42	-	1/fave +fH-1
Beta	0.87	-	1/fai +fH-1
Number of reactors	2	-	Design spec

Retention time for the whole digester	19.8	d	(1/bHT)*(alpha/beta -1) for a single or (N/bHt)*{(alpha/beta)^(1/N)-1} for multiple
fsvr1	0.3	-	favi*(1-fH)*(1-beta/alpha) or {favi*(1-fH)*bH*(Rhn/N)}/[1+bH*(Rhn/N)}]^n
fsvr2	0.1	-	{favi*(1-fH)*bH*(Rhn/N)}/[1+bH*(Rhn/N)}]^n
fsvrn	0.431		Sum of the indivual fractions
FXv	2409.9	kgVSS/d	Tut 6
FOt1 (With Denitrification)	1298.49	kgO/d	FXv *(fcv+(4.57-2.86)*fn)*fsvr1
FOt1 (Without Denitrification)	1523.29	kgO/d	FXv *(fcv+(4.57)*fn)*fsvr1
FOt2 (With Denitrification)	416.47	kgO/d	FXv *(fcv+(4.57-2.86)*fn)*fsvr2
FOt2 (Without Denitrification)	488.57	kgO/d	FXv *(fcv+(4.57)*fn)*fsvr2
FOD excluding recovered by denitrification	2011.86	kgO/d	FXv *(fcv+4.57*fn)*fsvr
FOD including recovered by denitrification	1715	kgO/d	FXv *(fcv+(4.57-2.86)*fn)*fsvrn or sum of the FOts
Max Xt after thickening in g TSS/l	60	gTSS/l	% * 1000
Qi (feed flow at the max Xt)	53.58	m3/d	FXt/Xt
Volume	1060.1	m3	Rhn * Qi
Volume (of a single reactor)	530.1	m3	Rhn * Qi for a single or Vd/N for multiple reactors
OUR in the 1st reactor (at 6%)	102.1	mgO/l.h	(FOt* 10^6)/(Vd*10^3*24)
Max OURt with anoxic-aerobic conditions	125	mgO/l.h	
fi effluent - VSS/TSS ratio	0.69		Xve/Xte

Table 49: Mass balance over anoxic-aerobic digester

TOD	**Value**	**Unit**
TOD in as VSS	4670.38	kgTOD/d
TOD out as VSS	2658.52	kgTOD/d
TOD out as Oxygen Demand	1714.96	kgO/d
TOD out as N2 gas	296.90	kgO/d
Total TOD out	4670.38	kgTOD/d
Total TOD out/TOD in	100	
COD	**Value**	**Unit**
COD in	3569.06	kgCOD/d
COD out	2031.61	kgCOD/d
FOt	1537.45	kgO/d
Balance	100	
N	**Value**	**Unit**
Nin	240.99	kgN/d
Nout	137.18	kgN/d
FN2	103.81	kgN/d
Balance	100	
P	**Value**	**Unit**
Pin	60.25	kgP/d

Pout	60.25	kgP/d
Balance	100	

Table 50: Effluent from anoxic-aerobic digester

Units	Filt Influent-MLE Effluent	Effluent filtered	Effluent particulate	Effluent unfiltered
kgCOD/m3	0.06	0.06	37.92	37.98
kgVSS/m3	0	0	25.6	25.6
kgISS/m3	0	0	11.39	11.39
kgTSS/lm3	0	0	36.99	36.99
kgFSA/m3	0	0	0	0
kgOrgN/m3	0.001	0.001	2.56	2.56
mgTKN/l	0.001	0.001	2.56	2.56
mgNO3/l	0.01	0.01	0.000	0.01
mgTN-N/l	0.011	0.011	2.56	2.57
mgOP-P/l	0.013	0.50	0	0.5
mgOrgP/l	0.000	0.000	0.64	0.64
mgTP-P/l	0.01	0.5	0.64	1.14

The waste flow rate and flux of WAS discharged to drying beds was obtained by mass balance over the digester as follow;

Table 51: Sludge waste to the drying beds

Parameter	Value	Units	Formula
Waste flow rate (at max Xt)	53.6	m^3/d	(Qi*Xti)/Xte
Flux of WAS discharged to drying beds	1981.9	kgTSS/d	Qw*Xte

P2 Design

The results for the P2 design are presented in summary tables;

Table 52: Summary table of double anoxic-aerobic digester for P2 Design

Double digester		
Item	**Value**	**Unit**
Temperature of the AS system	16	degrees
Maximum OUR in the 1st or only reactor	250	mgO/l.h
Thickener maximum conc	6%	TSS
Qw	602.9	m^3/d
FXt	2954	kgTSS/d
Xt (before thickening)	4.9	kgTSS/m^3
fati	0.45	-
favi	0.60	-
fi - VSS/TSS ratio	0.75	gVSS/gTSS
fn	0.1	mgN/mgVSS

fp	0.025	mgP/mgVSS
fcv	1.48	mgCOD/mgVSS
fh	0.2	-
bH20	0.24	/d
bhT	0.21	/d
Required SOUR (including nitrification)	1.5	mgO/gVSS.h
fave - Effluent active fraction (incl. Nit)	0.11	-
Alpha	8.42	-
Beta	0.87	-
Number of reactors	2	-
Retention time for the whole digester	19.8	d
fsvr1	0.3	-
fsvr2	0.1	-
fsvrn	0.43	
FXv	2214.5	kgVSS/d
FOt1 (With Denitrification)	1193.2	kgO/d
FOt1 (Without Denitrification)	1399.8	kgO/d
FOt2 (With Denitrification)	382.7	kgO/d
FOt2 (Without Denitrification)	448.96	kgO/d
FOD excluding recovered by denitrification	1848.7	kgO/d
FOD including recovered by denitrification	1575.9	kgO/d
Max Xt after thickening in g TSS/l	60	gTSS/l
Qi (feed flow at the max Xt)	49.2	m^3/d
Volume	974.2	m^3
Volume (of a single reactor)	487.1	m^3
OUR in the 1st reactor (at 6%)	102.1	mgO/l.h
Max OURt with anoxic-aerobic conditions	125	mgO/l.h
fi effluent - VSS/TSS ratio	0.69	

100% mass balances for TOD, COD, N and P were obtained.

Table 53: Effluent from double anoxic-aerobic digester for P2 design

Units	**Filt Influent-MLE Effluent**	**Effluent filtered**	**Effluent particulate**	**Effluent unfiltered**
kgCOD/m3	0.06	0.06	37.9	37.9
kgVSS/m3	0	0	25.6	25.6
kgISS/m3	0	0	11.4	11.4
kgTSS/lm3	0	0	36.99	36.99
kgFSA/m3	0	0	0	0
kgOrgN/m3	0.001	0.001	2.56	2.56
kgTKN/m3	0.001	0.001	2.56	2.56
kgNO3/m3	0.01	0.01	0	0.01
kgTN-N/m3	0.01	0.01	2.56	2.57
kgOP-P/m3	0.013	0.5	0	0.5
kgOrgP/m3	0	0	0.64	0.64

kgTP-P/m3	0.01	0.5	0.64	1.14

Table 54: Sludge waste to the drying beds for P2 design

Parameter	Value	Units
Waste flow rate (at max Xt)	49.23	m^3/d
Flux of WAS discharged to drying beds	1821.12	kgTSS/d

Aeration requirements for AS reactor and anoxic-aerobic digester

P1 Design

To determine the aeration requirements of the AS reactor, it was important to first determine the peak OUR in the MLE system. This is due to the cyclic variations of flow into the system. As the reactor was compartmentalized into **4** equal compartments in series, OUR was split into 33%, 25%, 21% and 21% in compartments 1,2,3 and 4 respectively. The highest peak OUR for the AS reactor was at **24°C** (**132.95mgO/l.h**).

Table 55: AS Peak OUR at 16°C

Parameter	Value	Unit	Formula
Peak TOD load	37024.5	kg/d	Max TOD in a day
Average TOD load	19174.4	kg/d	Simpson's rule on COD+4.57TKN for whole day
a_L	0.931		(Peak TOD/Average TOD)-1
Average OUR MLE	104.1	mgO/l.h	Computed above
Peak OUR	131.2	mgO/l.h	(1+0.28*a_L)*Average OUR in MLE
Peak FOt	133.0	kgO/h	Peak OUR*Aerobic volume

Table 56:AS Peak OUR at 24°C

Parameter	Value	Unit	Formula
Peak TOD load	37024.5	kg/d	Max TOD in a day
Average TOD load	19174.41	kg/d	Simpson's rule on COD+4.57TKN for whole day
a_L	0.93		(Peak TOD/Average TOD)-1
Average OUR MLE	105.46	mgO/l.h	
Peak OUR	132.95	mgO/l.h	(1+0.28*a_L)*Average OUR in MLE
Peak FOt	134.83	kgO/h	Peak OUR*Aerobic volume

For the aeration requirements of the anoxic-aerobic digester, the peak OUR in each digester was obtained at **16°C** and not **24°C** like for the AS aeration.

Table 57: Anoxic-aerobic digester peak OUR at 16ºC

Parameter	Value	
Digester #	1	2
Peak OUR (mgO/l.h)	102.1	32.7

Table 58: Anoxic-aerobic digester peak OUR at 24ºC

Parameter	Value	
Digester #	1	2
Peak OUR (mgO/l.h)	88	27.1

To compute the power requirements for the digester and the A.S reactor, various parameters were computed, presented in the tables below;

Table 59: Data provided to compute the power requirements for aeration

Data Provided	AS	Anoxic-aerobic Digester	Units
Altitude	200	200	m
Reactor DO concentration	2	1	mgO/l
RStd	2.6	2.6	kgO/kWh
Alpha	0.85	0.55	
Beta	0.95	0.75	
Line to shaft efficiency	0.9	0.9	
pstd	17.51	17.51	mmHg
PStd	760	760	mmHg
CS20	9.07	9.07	mgO/l
Theta	1.012	1.012	
Min Temp	16	16	ºC
Max Temp	24	24	ºC

Table 60:Parameters computed to determine the power requirements for aeration

Parameters	AS	Anoxic-aerobic Digester	Units
P_{site}	742.14	742.14	mmHg
$p_{site(16)}$	13.67	13.67	mmHg
$p_{site(24)}$	22.43	22.43	mmHg
CSstd (16)	9.83	9.83	mgO/l
CSstd (24)	8.42	8.42	mgO/l
Cssite (16)	9.65	9.65	mgO/l

Cssite (24)	8.16	8.16	mgO/l
KLa16/KLa20	0.95	0.95	
KLa24/KLa20	1.05	1.05	
Ract/RStd (16oC)	0.64	0.36	
Ract/RStd (24oC)	0.57	0.33	

The aeration size units as well as their efficiency were considered, and to provide the aeration required, the cheapest combination of number of units was provided.

Table 61: Power requirements for AS aeration

Aeration and Power Requirement for Activated Sludge				
Compartment #	1	2	3	4
Volume (m^3)	1014.2	1014.2	1014.2	1014.2
OUR percentages	0.33	0.25	0.21	0.21
OUR (mgO/lh) 24°C	43.87	33.24	27.92	27.92
Fotmax (kgO/h) 24°C	44.49	33.71	28.31	28.31
Ract(24)	1.47	1.47	1.47	1.47
Max Power Required (24) - kW	30.27	22.93	19.26	19.26
Manual Aeration Size (kW)	1 of 40	1 of 40	1 of 40	1 of 40
Provided Aeration considering efficiency (kW)	36	36	36	36

Table 62: Power requirements for Aerobic-Anoxic Digestion

Aeration and Power Requirements for Aer-Anox Digestion		
Digester #	1	2
Volume (m^3)	530.1	530.1
OUR(mgO/lh)	102.1	32.7
Fotmax (kgO/h) 16°C	54.1	17.4
Ract(16)	0.937	0.937
Max Power Required (16) - kW	57.73	18.52
Manual Aeration size (kW)	1 of 65	1 of 40
Provided Aeration considering efficiency	58.5	36.0

The design formulae for aeration were as follows;

$$P = 10^{(2.88117-0.000053407*Altitude)},\ P_{20}(1.0639)^{(T-20)}\ ,\ C_{Ssat} = C_{S20}\frac{51.6}{(31.6+T)}$$

$$C_{Ssite} = C_{Sstd} = \frac{(P_{site}-p_{site})}{(P_{std}-p_{std})},\ K_{LaT} = K_{La20}(\theta)^{(T-20)},\ \frac{R_{Act}}{R_{Std}} = \frac{\alpha\theta^{(T-20)}[\frac{(P_{site}-p_{site})51.6}{(760-17.51)(31.6+T)}\beta 9.07-C_L}{9.07}$$

$$Power = \frac{FO_{tmax}}{R_{Act}}$$

P2 Design

Table 63: A Peak OUR at 16°C for P2 Design

Parameter	Value	Unit	Formula
Peak TOD load	37024.5	kg/d	Max TOD in a day
Average TOD load	19174.4	kg/d	Simpson's rule on COD+4.57TKN for whole day
a_L	0.931		(Peak TOD/Average TOD)-1
Average OUR MLE	104.1	mgO/l.h	Computed above
Peak OUR	104.1	mgO/l.h	(1+0.28*a_L)*Average OUR in MLE
Peak FOt	96.98	kgO/h	Peak OUR*Aerobic volume

Table 64:AS Peak OUR at 24°C for P2 Design

Parameter	Value	Unit	Formula
Peak TOD load	37024.5	kg/d	Max TOD in a day
Average TOD load	19174.41	kg/d	Simpson's rule on COD+4.57TKN for whole day
a_L	0.93		(Peak TOD/Average TOD)-1
Average OUR MLE	105.46	mgO/l.h	
Peak OUR	105.46	mgO/l.h	(1+0.28*a_L)*Average OUR in MLE
Peak FOt	98.28	kgO/h	Peak OUR*Aerobic volume

For the aeration requirements of the anoxic-aerobic digester, the peak OUR in each digester was obtained at **16°C** and not **24°C** like for the AS aeration.

Table 65: Anoxic-aerobic digester peak OUR at 16°C for P2 Design

Parameter	Value	
Digester #	1	2
Peak OUR (mgO/l.h)	102.1	32.7

Table 66: Anoxic-aerobic peak OUR at 24°C for P2 Design

Parameter	Value	
Digester #	1	2
Peak OUR (mgO/l.h)	88	27.1

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Table 67: Power requirements for AS aeration for P2 Design

Aeration and Power Requirement for Activated Sludge				
Compartment #	1	2	3	4
Volume (m^3)	931.9	931.9	931.9	931.9
OUR percentages	0.33	0.25	0.21	0.21
OUR (mgO/lh) 24ºC	34.8	26.36	22.15	22.15
Fotmax (kgO/h) 24ºC	32.43	24.57	20.37	20.37
Ract(24)	1.47	1.47	1.47	1.47
Max Power Required (24) - kW	22.1	16.7	14.04	14.04
Manual Aeration Size (kW)	1 of 40	1 of 40	1 of 40	1 of 40
Provided Aeration considering efficiency (kW)	36	36	36	36

Table 68: Power requirements for Aerobic-Anoxic digestion for P2 Design

Aeration and Power Requirements for Aer-Anox Digestion		
Digester #	1	2
Volume (m^3)	487.1	487.1
OUR(mgO/lh)	102.1	32.7
Fotmax (kgO/h) 16ºC	49.72	15.95
Ract(16)	0.937	0.937
Max Power Required (16) - kW	53.1	17.01
Manual Aeration size (kW)	1 of 65	1 of 40
Provided Aeration considering efficiency	58.5	36.0

Gravity Thickener

P1 Design

A solids loading rate of 80kg/m^2.d was used, thickening the WAS sludge mass of 8449kgTSS/d to 60gTSS/l.

Table 69: Gravity thickener summary table for P1 Design

Parameters	Value	Unit	Formula
Sludge Mass	8449	kg/d	WAS Flux
Solids Loading Rate	80	kg/m^2.d	Given
TSS concentration out	60	g/l	
Qi	183.3	m3/d	PS Flow
Qw	140.8	m3/d	(PS TSS*PS Flow)/ (TSS out)
Qe (Thickener Supernatant)	42.5	m3/d	Qi-Qw
Area	176.7	m^2	Sludge Mass/SLR
Diameter	15	m	

P2 Design

Table 70: Gravity thickener summary table for P2 Design

Parameters	Value	Unit	Formula
Sludge Mass	8449	kg/d	WAS Flux
Solids Loading Rate	80	$kg/m^2.d$	Given
TSS concentration out	60	g/l	
Qi	183.3	m3/d	PS Flow
Qw	140.8	m3/d	(PS TSS*PS Flow)/ (TSS out)
Qe (Thickener Supernatant)	42.5	m3/d	Qi-Qw
Area	176.7	m^2	Sludge Mass/SLR
Diameter	15	m	

Dissolved Air Flotation Unit

P1 Design

Table 71: DAF summary table for P1 Design

Parameters	Value	Unit
as	0,034	
dw	1,69	m
Qs	34,52	kgTSS/m2.d
Cfloat	6	%
Cf	60	gTSS/l
Vs	36	m/d
VL	259,19	m/d
Qi (PWWF)	656,061	m3/d
Qi	656,061	m3/d
Qw	53,578	m3/d
Qe (Thickener Supernatant)	602,483	m3/d
A	93,13	m2
Diameter	10,89	m

P2 Design

Table 72: DAF summary table for P2 Design

Parameters	Value	Unit
as	0,034	
dw	1,69	m
Qs	34,52	kgTSS/m2.d
Cfloat	6	%
Cf	60	gTSS/l
Vs	36	m/d

VL	259,19	m/d
Qi (PWWF)	602,866	m3/d
Qi	602,866	m3/d
Qw	49,234	m3/d
Qe (Thickener Supernatant)	553,632	m3/d
A	85,58	m2
Diameter	10,44	m

Anaerobic Digestion of Primary Sludge

P1 Design

Table 73: PS AD for P1 Design

Parameter	Value	Unit
Methane Gas Properties	10	kWh/Nm3
Thermal efficiency of gas engine	0.45	%
Heat loss from AD	30	W/m3 AD volume
Digested PS dewatered to	20	%TSS
Energy Generated by Methane at 100% efficiency	1041.13	kW
MLE Aeration Energy required P202	1582.44	kW
MLE Aeration Energy required P201	160.0	kW
Proportion of Energy Supplied to P202 Aer.	29.61	%
Proportion of Energy Supplied to P201 Aer.	100	%
Estimated AD volume	1832.9	m3
Ambient Temperature	16	oC
Mesophilic Temperature	37	oC
Temperature increase required	21	oC
Inflow	183.29	m3/d
Sludge Specific Heat	4.18	kJ/kgoC
m3/kg ratio	1000	
Energy Required to heat AD	16089599.14	kJ/d
Energy Required to heat AD	186.22	kJ/s
Conversion to KiloWatts	186.22	kW
Heat Loss from AD	54.99	kW
Power produced by AD with efficiency	468.51	kW
Proportion of power required by AD itself	51.5	%
Proportion of power left considering P201 and heat loss and eff.	14.4	%

P2 Design

Table 74: PS AD for P2 Design

Parameter	Value	Unit
Methane Gas Properties	10	kWh/Nm3
Thermal efficiency of gas engine	0,45	%
Heat loss from AD	30	W/m3 AD volume
Digested PS dewatered to	20	%TSS
Energy Generated by Methane at 100% efficiency	788,64	kW
MLE Aeration Energy required P202	1582,44	kW
MLE Aeration Energy required P201	160,00	kW
Proportion of Energy Supplied to P202 Aer.	22,43	%
Proportion of Energy Supplied to P201 Aer.	100	%
Estimated AD volume	1669,284	m3
Ambient Temperature	16	oC
Mesophilic Temperature	37	oC
Temperature increase required	21	oC
Inflow	166,93	m3/d
Sludge Specific Heat	4,18	kJ/kgoC
m3/kg ratio	1000	
Energy Required to heat AD	14652974,2	kJ/d
Energy Required to heat AD	169,59	kJ/s
Conversion to KiloWatts	169,59	kW
Heat Loss from AD	50,08	kW
Power produced by AD with efficiency	354,89	kW
Proportion of power required by AD itself	61,899	%
Proportion of power left considering P202 and heat loss and eff.	-407,797	%
Proportion of power left considering P201 and heat loss and eff.	-6,984	%

Appendix E Parameters used in WEST®

Table 75: Initial Values for WEST Simulation

Name		Unit	Default Value	Lower bound	Upper bound	Fixed
Category: Parameters						
Group:						
f_XU_Bio_lysis	**0,08**	**dUnit/dUnit**	**0,08**	**-INF**	**+INF**	
Group: Composition parameters						
ISS_BM	**0,15**	**g/gCOD**	**0,15**	**0,1425**	**0,1575**	
Group: Kinetic						
b_ANO	**0,15**	**1/d**	**0,15**	**0,12**	**0,18**	
b_OHO	**0,62**	**1/d**	**0,62**	**0,496**	**0,744**	
b_PAO	**0,04**	**1/d**	**0,04**	**0,032**	**0,048**	
K_F	**4**	**g/m3**	**4**	**2**	**6**	
K_IPP	**0,02**	**g/m3**	**0,02**	**0,01**	**0,03**	
k_M_BInf_OHO_hy	**3**	**gCOD/(gCOD*d)**	**3**	**2,4**	**3,6**	
k_M_BOrg_OHO_h	**3**	**gCOD/(gCOD*d)**	**3**	**2,4**	**3,6**	
b_PHA	**0,04**	**1/d**	**0,04**	**0,032**	**0,048**	
K_MAX	**0,34**	**g/m3**	**0,34**	**0,272**	**0,408**	
K_NO	**0,5**	**gNO3-N/m3**	**0,5**	**0,25**	**0,75**	
K_O2	**0,2**	**g/m3**	**0,2**	**0,1**	**0,3**	
K_O2_ANO	**0,2**	**g/m3**	**0,2**	**0,1**	**0,3**	
K_O2_PP	**0,1**		**0,1**	**0,05**	**0,15**	
K_ALK	**0,1**	**g/m3**	**0,1**	**0,05**	**0,15**	
K_S_ALK_ANO	**0,1**	**g/m3**	**0,1**	**0,05**	**0,15**	
K_PS	**0,2**	**g/m3**	**0,2**	**0,1**	**0,3**	
K_X	**0,1**	**gCOD/gCOD**	**0,1**	**0,05**	**0,15**	
K_S_Glyco	**0,001**		**0,001**	**0,0005**	**0,0015**	
b_PP	**0,04**	**1/d**	**0,04**	**0,032**	**0,048**	
K_S_NH_ANO	**0,3**	**g/m3**	**0,3**	**0,15**	**0,45**	
K_NH	**0,05**	**g/m3**	**0,05**	**0,025**	**0,075**	
K_PHA	**0,01**	**g/m3**	**0,01**	**0,005**	**0,015**	
K_P	**0,01**	**g/m3**	**0,01**	**0,005**	**0,015**	
K_PP	**0,01**	**g/m3**	**0,01**	**0,005**	**0,015**	
K_A	**4**	**g/m3**	**4**	**2**	**6**	
K_S_F_OHO_ferm	**20**	**g/m3**	**20**	**10**	**30**	
KLa_CO2	**0**	**1/d**	**0**	**0**	**20**	
KS_fPP_PAO_PHAs	**0,0001**		**0,0001**	**-INF**	**+INF**	
f_gly_PAO_max	**0,32**	**dUnit/dUnit**	**0,32**	**0,304**	**0,336**	
mu_ANO	**0,55**	**1/d**	**0,55**	**0,275**	**0,825**	
mu_OHO_max	**6**	**1/d**	**6**	**4,8**	**7,2**	
mu_PAO	**1**	**1/d**	**1**	**0,8**	**1,2**	
n_NO_Het	**0,3**	**-**	**0,3**	**0,24**	**0,36**	
n_OHO_BInf_ferm	**0,1**	**-**	**0,1**	**0,08**	**0,12**	
n_OHO_BInf_hyd	**0,6**	**-**	**0,6**	**0,48**	**0,72**	
n_OHO_BOrg_fer	**0,1**	**-**	**0,1**	**0,08**	**0,12**	

n_OHO_BOrg_hyd	0,6	-	0,6	0,48	0,72
Q_OHO_F_VFA_fer	15	1/d	15	7,5	22,5
Q_Glycogen	0,2	1/d	0,2	0	20
Q_PAO_PO4_PPsto	4,5	1/d	4,5	2,25	6,75
Q_PAO_PP_PHAsto	1,5		1,5	0,75	2,25
Group: kinetics					
kdis_cal	0		0	-INF	+INF
Group: stoichiometry					
i_H_SF_mol_perC	1,942	dUnit/dUnit	2,01	1,005	2,1945
i_H_SU_mol_perC	1,833	dUnit/dUnit	1,646	0,823	2,469
i_H_XBInf_mol_pe	1,623	dUnit/dUnit	2,19	1,095	3,285
i_H_XUInf_mol_pe	1,534	dUnit/dUnit	1,482	0,741	2,223
i_H_XUOrg_mol_p	1,534	dUnit/dUnit	1,482	0,741	2,223

Table 76: Parameters for Aeration Control

Name		Unit	Default Value	Lower bound	Upper bound Fixed
Category: Manipulated Variables					
Group: Operational					
K_P	25		25	-INF	+INF
T_I	0,1 d		0,1	0	+INF
u0	50		50	-INF	+INF
u_Max	1000		1000	-INF	+INF
u_Min	0		0	-INF	+INF
y_S	2		2	-INF	+INF

Table 77: Parameters for Aerobic Reactor

Name		Unit	Default Value	Lower bound	Upper bound Fixed
Category: Manipulated Variables					
Group: Operational					
Kla	0	1/d	0	0	5000
Temp	20	degC	15	-273,15	+INF
Category: Parameters					
Group:					
f_XU_Bio_lysis	0,08	dUnit/dUnit	0,08	-INF	+INF
Group: Aeration					
OTR_Energy	1800	g/kWh	1800	-INF	+INF
Group: Composition parameters					
ISS_BM	0,15	g/gCOD	0,15	0	1
Group: Conversion factors					
F_TSS_COD	0,75	-	0,75	0	1
Group: Dimension					
Vol	3384,9	m3	1000	0	+INF
Group: Kinetic					
A_20	1	1/d	1	0	20
A_35	1	1/d	1	0	20
A_55	1	1/d	1	0	20
KA_H_N	1E-07	g/m3	1E-07	0	100
KI_H_N	0,00085	g/m3	0,00085	0	100
K_A	4	g/m3	4	0	100
K_ALK	0,1	g/m3	0,1	0	100

K_A_PO4	5		5	-INF	+INF
K_F	4	g/m3	4	0	100
K_IPP	0,02	g/m3	0,02	0	100
K_MAX	0,34	g/m3	0,34	0	100
K_NH	0,05	g/m3	0,05	0	100
K_NO	0,5	gNO3-N/m3	0,5	0	2
K_O2	0,2	g/m3	0,2	0	100
K_O2_ANO	0,2	g/m3	0,2	0	100
K_P	0,01	g/m3	0,01	0	100
K_PHA	0,01	g/m3	0,01	0	100
K_PP	0,01	g/m3	0,01	0	100
K_PS	0,2	g/m3	0,2	0	100
K_S_ALK_ANO	0,1	g/m3	0,1	0	100
K_S_F_OHO_ferm	20	g/m3	20	0	100
K_S_NH_ANO	0,3	g/m3	0,3	0	100
K_X	0,1	gCOD/gCOD	0,1	0	100
K_destruct	5	1/d	5	0	20
PCO2_atm	0,0003467	Pa	0,0003467	0	+INF
Q_Glycogen	0,2	1/d	0,2	0	20
Q_OHO_F_VFA_fer	20	1/d	20	0	20
Q_PAO_PO4_PPsto	4,5	1/d	4,5	0	20
Q_PAO_PP_PHAsto	1,5	1/d	1,5	0	20
S_O_Sat	8,9	g/m3	8,9	0	+INF
Th_d_XG_20	1,12	1/d	1,12	0	20
Th_d_XG_35	1,12	1/d	1,12	0	20
Th_d_XH_20	1,12	1/d	1,12	0	20
Th_d_XH_35	1,12	1/d	1,12	0	20
Th_d_XN_20	1,072	1/d	1,072	0	20
Th_d_XN_35	1,072	1/d	1,072	0	20
Th_h_20	1,116	1/d	1,116	0	20
Th_h_35	0,05	1/d	0,05	0	20
Th_m_XH_20	1,072	1/d	1,072	0	20

Table 78: Parameters Anaerobic Reactor

Name		Unit	Default Value	Lower bound	Upper bound Fixed
Category: Manipulated Variables					
Group: Operational					
Kla	0	1/d	0	0	5000
Temp	20	degC	15	-273,15	+INF
Category: Parameters					
Group:					
f_XU_Bio_lysis	0,08	dUnit/dUnit	0,08	-INF	+INF
Group: Aeration					
OTR_Energy	1800	g/kWh	1800	-INF	+INF
Group: Composition parameters					
ISS_BM	0,15	g/gCOD	0,15	0	1
Group: Conversion factors					
F_TSS_COD	0,75	-	0,75	0	1
Group: Dimension					
Vol	712,6	m3	1000	0	+INF
Group: Kinetic					

A_20	1	1/d	1	0	20
A_35	1	1/d	1	0	20
A_55	1	1/d	1	0	20
KA_H_N	1E-07	g/m3	1E-07	0	100
KI_H_N	0,00085	g/m3	0,00085	0	100
K_A	4	g/m3	4	0	100
K_ALK	0,1	g/m3	0,1	0	100
K_A_PO4	5		5	-INF	+INF
K_F	4	g/m3	4	0	100
K_IPP	0,02	g/m3	0,02	0	100
K_MAX	0,34	g/m3	0,34	0	100
K_NH	0,05	g/m3	0,05	0	100
K_NO	0,5	gNO3-N/m3	0,5	0	2
K_O2	0,2	g/m3	0,2	0	100
K_O2_ANO	0,2	g/m3	0,2	0	100
K_P	0,01	g/m3	0,01	0	100
K_PHA	0,01	g/m3	0,01	0	100
K_PP	0,01	g/m3	0,01	0	100
K_PS	0,2	g/m3	0,2	0	100
K_S_ALK_ANO	0,1	g/m3	0,1	0	100
K_S_F_OHO_ferm	20	g/m3	20	0	100
K_S_NH_ANO	0,3	g/m3	0,3	0	100
K_X	0,1	gCOD/gCOD	0,1	0	100
K_destruct	5	1/d	5	0	20
PCO2_atm	0,0003467	Pa	0,0003467	0	+INF
Q_Glycogen	0,2	1/d	0,2	0	20
Q_OHO_F_VFA_fer	20	1/d	20	0	20
Q_PAO_PO4_PPsto	4,5	1/d	4,5	0	20
Q_PAO_PP_PHAsto	1,5	1/d	1,5	0	20
S_O_Sat	8,9	g/m3	8,9	0	+INF
Th_d_XG_20	1,12	1/d	1,12	0	20
Th_d_XG_35	1,12	1/d	1,12	0	20
Th_d_XH_20	1,12	1/d	1,12	0	20
Th_d_XH_35	1,12	1/d	1,12	0	20
Th_d_XN_20	1,072	1/d	1,072	0	20
Th_d_XN_35	1,072	1/d	1,072	0	20
Th_h_20	1,116	1/d	1,116	0	20
Th_h_35	0,05	1/d	0,05	0	20
Th_m_XH_20	1,072	1/d	1,072	0	20

Table 79: Parameters Anoxic Reactor

Name		Unit	Default Value	Lower bound	Upper bound Fixed
Category: Manipulated Variables					
Group: Operational					
Kla	0	1/d	0	0	5000
Temp	20	degC	15	-273,15	+INF
Category: Parameters					
Group:					
f_XU_Bio_lysis	0,08	dUnit/dUnit	0,08	-INF	+INF
Group: Aeration					

OTR_Energy	**1800**	**g/kWh**	**1800**	**-INF**	**+INF**
Group: Composition parameters					
ISS_BM	**0,15**	**g/gCOD**	**0,15**	**0**	**1**
Group: Conversion factors					
F_TSS_COD	**0,75**	**-**	**0,75**	**0**	**1**
Group: Dimension					
Vol	**3028,6**	**m3**	**1000**	**0**	**+INF**
Group: Kinetic					
A_20	**1**	**1/d**	**1**	**0**	**20**
A_35	**1**	**1/d**	**1**	**0**	**20**
A_55	**1**	**1/d**	**1**	**0**	**20**
KA_H_N	**1E-07**	**g/m3**	**1E-07**	**0**	**100**
KI_H_N	**0,00085**	**g/m3**	**0,00085**	**0**	**100**
K_A	**4**	**g/m3**	**4**	**0**	**100**
K_ALK	**0,1**	**g/m3**	**0,1**	**0**	**100**
K_A_PO4	**5**		**5**	**-INF**	**+INF**
K_F	**4**	**g/m3**	**4**	**0**	**100**
K_IPP	**0,02**	**g/m3**	**0,02**	**0**	**100**
K_MAX	**0,34**	**g/m3**	**0,34**	**0**	**100**
K_NH	**0,05**	**g/m3**	**0,05**	**0**	**100**
K_NO	**0,5**	**gNO3-N/m3**	**0,5**	**0**	**2**
K_O2	**0,2**	**g/m3**	**0,2**	**0**	**100**
K_O2_ANO	**0,2**	**g/m3**	**0,2**	**0**	**100**
K_P	**0,01**	**g/m3**	**0,01**	**0**	**100**
K_PHA	**0,01**	**g/m3**	**0,01**	**0**	**100**
K_PP	**0,01**	**g/m3**	**0,01**	**0**	**100**
K_PS	**0,2**	**g/m3**	**0,2**	**0**	**100**
K_S_ALK_ANO	**0,1**	**g/m3**	**0,1**	**0**	**100**
K_S_F_OHO_ferm	**20**	**g/m3**	**20**	**0**	**100**
K_S_NH_ANO	**0,3**	**g/m3**	**0,3**	**0**	**100**
K_X	**0,1**	**gCOD/gCOD**	**0,1**	**0**	**100**
K_destruct	**5**	**1/d**	**5**	**0**	**20**
PCO2_atm	**0,0003467**	**Pa**	**0,0003467**	**0**	**+INF**
Q_Glycogen	**0,2**	**1/d**	**0,2**	**0**	**20**
Q_OHO_F_VFA_fer	**20**	**1/d**	**20**	**0**	**20**
Q_PAO_PO4_PPsto	**4,5**	**1/d**	**4,5**	**0**	**20**
Q_PAO_PP_PHAsto	**1,5**	**1/d**	**1,5**	**0**	**20**
S_O_Sat	**8,9**	**g/m3**	**8,9**	**0**	**+INF**
Th_d_XG_20	**1,12**	**1/d**	**1,12**	**0**	**20**
Th_d_XG_35	**1,12**	**1/d**	**1,12**	**0**	**20**
Th_d_XH_20	**1,12**	**1/d**	**1,12**	**0**	**20**
Th_d_XH_35	**1,12**	**1/d**	**1,12**	**0**	**20**
Th_d_XN_20	**1,072**	**1/d**	**1,072**	**0**	**20**
Th_d_XN_35	**1,072**	**1/d**	**1,072**	**0**	**20**
Th_h_20	**1,116**	**1/d**	**1,116**	**0**	**20**
Th_h_35	**0,05**	**1/d**	**0,05**	**0**	**20**
Th_m_XH_20	**1,072**	**1/d**	**1,072**	**0**	**20**

Table 80: Parameters PST

Name	Unit	Default Value	Lower bound	Upper bound	Fixed
Category: Manipulated Variables					
Group: Operational					
Q_Under	**150 m3/d**	**10**	**0**	**+INF**	
Category: Parameters					
Group: Conversion factors					
F_TSS_COD	**0,75 -**	**0,75**	**0**	**1**	
Group: Energy					
F_Energy_FlowRat	**0,04 kWh/m3**	**0,04**	**-INF**	**+INF**	
Group: Settling					
f_ns	**0,6 -**	**0,6**	**0**	**1**	

Table 81: Parameters SST

Name	Unit	Default Value	Lower bound	Upper bound	Fixed
Category: Manipulated Variables					
Group: Operational					
Q_Under	**15000 m3/d**	**10**	**0**	**+INF**	
Category: Parameters					
Group: Conversion factors					
F_TSS_COD	**0,75 -**	**0,75**	**0**	**1**	
Group: Energy					
F_Energy_FlowRat	**0,04 kWh/m3**	**0,04**	**-INF**	**+INF**	
Group: Settling					
f_ns	**0,005 -**	**0,005**	**0**	**1**	