Gareth Brown

Developing the next generation of treatment flowsheets for rural wastewater

School of Water Energy and Environment

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Abstract

Septic tank systems (STS) are one of the most common wastewater treatment systems in the world however these systems are becoming antiquated, struggling to meet tighter consent, needing frequent desludging and venting greenhouse gases to the environment. A flowsheet to tackle these issues was proposed consisting of an enhanced septic tank (EST) and a constructed wetland (CW). The proposed flowsheet was assessed by examining the underlying mechanisms, testing the flowsheet at pilot scale and assessing the cost and carbon implications of the flowsheet. A key aim of the thesis is to decrease the maintenance of the flowsheet to once every 5 to 7 years. The maintenance of septic tank is due to desludging, as the tank needs to be emptied when the sludge bed reaches a certain height. The desludging of a septic tank is dependent on the hydrolysis rate within the reactor. Batch studies into anaerobic digestion found that low operating temperatures in septic tanks impact the hydrolysis rate. However, temperature is not the dominant factor of hydrolysis and optimisation of mass transfer between hydrolytic microorganisms and degradable particulates can increase the hydrolysis rate by 200%. A pilot scale study investigated using baffles to promote hydraulic mixing within a septic tank. The presence of hydraulic mixing due to the baffles increased the hydrolysis rate constant of a septic tank from 0.0089 d^{-1} to 0.035d^{-1}, extending the time between desludging from 4.9 to 6.7 years. The proposed flowsheet is a lower cost treatment system than a conventional package treatment system (e.g. submerged aerated filter) over a 30 year life time and leads to a significant reduction in lifetime carbon emissions compared to a STS. The cost and carbon reduction of the flowsheet make the flowsheet a viable abatement technique.

Keywords:

Septic tanks, hydrolysis, anaerobic digestion, psychrophilic, nature-based solutions

Acknowledgements

First and foremost a big thank you to my supervisors; Bruce Jefferson, Jitka MacAdam and Gaby Dotro. To Bruce, thank you, without your seemingly endless stream of knowledge I would not have been able to get to the end of a thesis. We may not have always seen eye to eye, mostly on my unconventional naming schemes, but I appreciate everything you have done for me. Thank you Jitka for stepping up and helping through the final year. To Gaby thanking for helping guide me through my first few years at Cranfield, shaping my academic outlook and both academic and emotion support.

Additional I extend my thanks to my industrial sponsors Scottish Water, particularly Allan, Tamsyn and Kerry. I look forward to hearing from the demonstration system. I also thank Pablo for his MsC work on fluid dynamics and Amaka for here wetlands work with me.

I would also like to thank all the Cranfield technician team for the help and guidance. A big shout out goes to the King of the Pilot Hall Nigel Janes for all his help and support keeping not just mine but every pilot system funning. Additionally, Tony Beal for his help with constructing my wetland systems even if it did nearly lose a tractor, and Dave Darlow for support and an endless supply of sludge. I will miss the pilot hall coffee breaks we had.

To all my friend and colleagues at Cranfield, thank you for your help and support making my time in this indeterminate region of the country significantly more enjoyable. To the No PhD Club, Ben, Niall, Laura and Michele, and group Alumni Alex, thank you for helping us all stay sane during the times which shall not be named and also to the rest of the lunch group; Priya, Miles, Sopie, Rob, Tizzi and Tolu. Just keep swimming, you'll get there and can escape too.

Finally, to all my family for their love and support, and to my amazing partner Charlotte. Thank you for your patience, support and unofficial statistical consulting throughout this time I would not be here without you.

Table of Contents

List of Figures

List of Tables

List of Abbreviations

ABR	Anaerobic Baffled Reactor
AHF	Aerated Horizontal Flow
AP	Anaerobic Pond
ANOVA	Analysis of Variance
BOD	Biological Oxygen Demand
CFD	Computational Fluid Dynamics
COD	Chemical Oxygen Demand
CW	Constructed Wetland
D	Downcomer Baffle
DEWATS	Decentralised Wastewater Treatment System
DWF	Dry Weather Flow
EST	Enhanced Septic Tank
H_L	Left-Hand Horizontal Baffle
HLR	Hydraulic Loading Rate
H_R	Right-Hand Horizontal Baffle
HRT	Hydraulic Residence Time
K_c	Contois Constant
k_h	Hydrolysis Rate Constant
K_m	Michaelis Constant
LCA	Lifecycle Analysis
LCE	Lifetime Carbon Emissions
Max	Maximum
Min	Minimum
NBS	Nature-Based Solutions
NPV	Net Present Value
OLR	Organic Loading Rate
pCOD	Particulate Chemical Oxygen Demand

PE	Population Equivalent
pNPP	p-nitrophenol palmitate
PSD	Particle Size Distribution
Q_{av}	Average volumetric flowrate
Q_{pe}	Daily wastewater contribution per capita
rpm	Rotations per minute
RTD	Residence Time Distribution
SAF	Submerged Aerated Filter
SAR	Solids Accumulation Rate
sCOD	Soluble Chemical Oxygen Demand
STS	Septic Tank System
TCOD	Total Chemical Oxygen Demand
TSS	Total Suspended Solids
U	Upcomer Baffle
UASB	Upflow Anaerobic Sludge Bed
VF	Vertical Flow
VFA	Volatile Fatty Acids
v_p	Soil Percolation Rate
V_{max}	Maximum Rate
VSS	Volatile Suspended Solids
WLC	Whole life Costs

1. Introduction

1.1. The current state of rural wastewater treatment

Septic tanks were first introduced to the United Kingdom in the late 1800s, with the design undergoing limited changes since that time. A septic tank system (STS) consists of two primary elements: a septic tank and a drainage field or drainfield (Figure 1.1). The septic tank is a low-rate anaerobic reactor which has the primary goal of separating the solid and liquid components of domestic wastewater, the accumulated sludge is held within the septic tank until the tank is full and requires desludging. The liquid effluent of the septic tank flows into the drainage field, which disperses the wastewater across a field to slowly percolate into the soil.

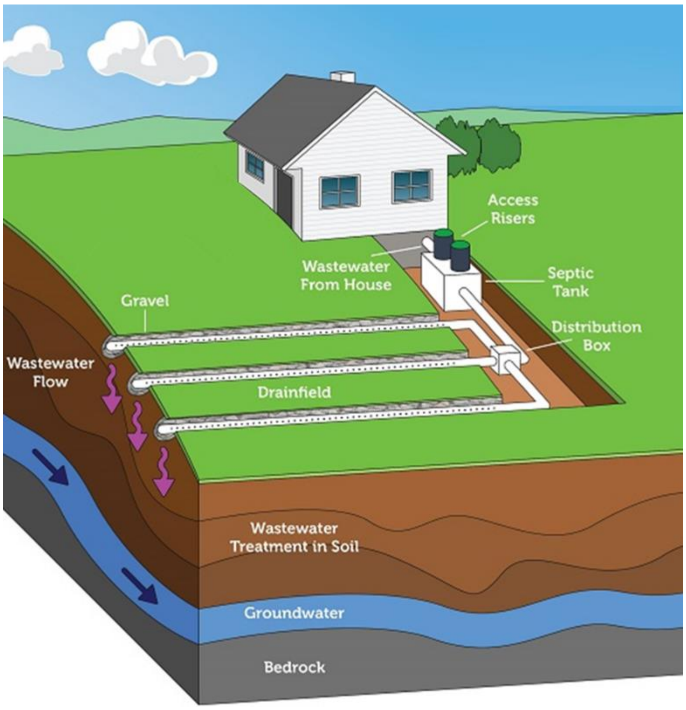

Figure 1.1: Diagram of a conventional STS (EPA, 2022)

The accumulation of faecal sludge in septic tanks is comparable to other onsite wastewater system such as Pit Latrines. STS and pit latrines treat approximately 43% of global wastewater, hence are vital systems for the treatment of domestic wastewater (UN Habitat & WHO, 2021). Both pit latrines and STS do not represent the final stage of faecal sludge management, instead providing the formation and storage of the sludge, when full the systems require desludging. Desludging of STS and pit latrines requires the suspension in operation and the pumping out of the sludge. The sludge is typically then transported to a larger works for final treatment and disposal.

Despite the ubiquity of STS around the world, these systems are flawed. An individual STS is unlikely to pose a significant risk to water quality due to the small quantity of wastewater discharged, however there is increasing evidence that cumulative STS can pose a significant risk to water quality. Recent findings indicate that in rural communities STS disproportionately effect surface water nitrogen, in comparison to fertiliser usage (Halliday et al., 2014). For instance, ammonia concentrations in ground and surface waters have been shown to more than double downstream of STS (Herren et al., 2021; Withers et al, 2012).

Beyond water quality, there are several other limitations to STS design. The sludge accumulation rate (SAR) and sizing of the septic tank means that it is recommended that a system is desludged annually (HM Government, 2015). However, actual operation of STSs is much more variable, of the over 1200 STS operated by Scottish Water the majority are desludged at least every 6 months. This is the primary maintenance cost of the systems, contributes to the systems carbon footprint and causes social disruption.

A septic tank is an anaerobic reactor, and hence produces methane. Methane is both a potential fuel source and a potent greenhouse gas. Methane has a warming potential 25 times greater than carbon dioxide over a 100-year time scale and is responsible for 30% of human activity induced temperature increases (IEA, 2022). Any methane generated by a septic tank is vented to the environment, therefore contributing to climate change. By 2030 the UK water sector is aiming to reach net zero operational emissions, therefore alongside improved treatment there is a carbon drive to improve STS.

1.2. Upgrading Septic Tank Systems

1.2.1. Desludging Frequency

The desludging frequency of a septic tank is governed by three key factors: solids removal rate, hydrolysis rate and tank volume. The minimisation of the solids removal rate within the anaerobic system would cause operational issues for latter treatment steps and therefore is not an effective means to reduce the maintenance requirements. Simply increasing the reactor volume corresponds with an increase system cost, hence the primary focus of optimisation is the hydrolysis rate in the anaerobic reactor. Hydrolysis is the first stage of anaerobic digestion and the microbial process in which solid particulates are degraded to their soluble constituents. This process is limited in low temperature environments in which septic tanks operated. Therefore, low temperature optimisation methods will need to be investigated.

1.2.2. Enhanced Treatment

Ammonia is primarily removed by nitrification, a two-step aerobic process mediated by autotrophic micro-organisms. Therefore, to improve the treatment of STS, an aerobic process is required. Aeration is an energy intensive process, consuming a significant portion of wastewater treatments energy demand. Alongside a high energy consumption, aerobic treatment processes often require significant maintenance. Nature based solutions (NBS) such as constructed wetlands provide a less energy and maintenance intense form of aerobic wastewater treatment. The two most prominent configurations of aerobic constructed wetlands are vertical flow wetlands (VF) and aerated horizontal flow wetlands (AHF). A VF provides an aerobic environment through cyclically loading and draining the treatment bed allowing for the aeration of the biofilm between feed doses. Dosing of a VF can be conducted passively or with pumping. In comparison, an AHF is a continuous flow system, with a coarser media bed than a VF, in which air is forced into the system to provide an aerobic environment. An AHF is a more compact treatment system, with a smaller footprint than a VF, however a VF can operate entirely passively unlike an AHF. The proposed flowsheet consists of an EST and a constructed wetland for secondary treatment.

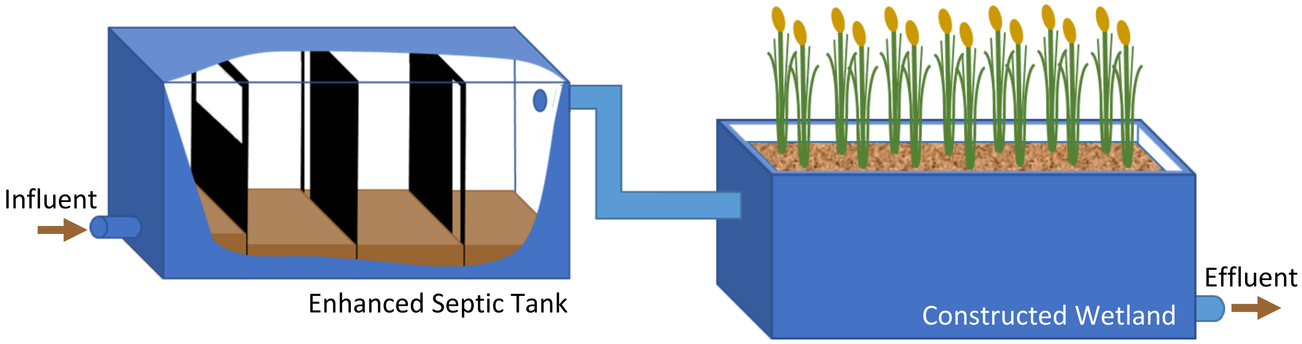

Figure 1.2: Proposed flowsheet

1.3. Project aims and objective

1.3.1. Aim

The overall aim of the thesis was to understand the impact of temperature and mixing on the processes that govern the sludge accumulation rate in septic tanks in order to develop alternative designs that could enable infrequent desludging. The associated ambition was to establish an alternative flowsheet based on this alternative design for onsite wastewater treatment that could improve treatment, reduce carbon footprint and provide the basis for additional added value.

1.3.2. Objectives

a. Review the current state of low temperature anaerobic treatment of onsite domestic wastewater
b. Assess the impact of sub 20^{o}C temperatures on hydrolysis rates in septic tank conditions and the impact of mixing.
c. Understand the impact of baffle arrangement within septic tanks to reduce the required desludging frequency
d. Establish the overall flowsheet effluent quality and relative effectiveness of aerated horizontal flow and vertical flow wetlands.
e. Determine the lifetime cost implications of improving septic tank flowsheets and whether this is a viable method to reduce the carbon footprint of rural wastewater treatment flowsheets.

1.4. Thesis Structure

The thesis is presented as a series of chapters formatted as papers for publication (Table 1.1). All chapters were written by Gareth Brown and have been edited by Prof. Bruce Jefferson. All experimental and pilot scale trials were undertaken by Gareth Brown. All modelling and statistical analysis was conducted by Gareth Brown with the exception of Computational fluid dynamics (CFD) modelling in Chapter 4 which was conducted by Pablo Andrés Piedra León as part of their MSc thesis.

Table 1.1: Thesis structure and journal submission plan

Chapter	Objective addressed	Title	Target Journal	Status
2	a	Onsite anaerobic reactor design and complexity, the impact on treatment and hydrolysis in temperate conditions: A systematic review.	Environmental Technology Reviews	In preparation
3	b	Impact of low temperatures on anaerobic hydrolysis in passive treatment systems such as septic tanks	Environmental Science: Water Research and Technology	In preparation
4	c	Enhancing the hydrolysis rate within septic tanks through baffling to extend desludging frequency	Water Research	Submitted
5	d	Constructed Wetlands for the Polishing of Enhanced Septic Tank Effluent: An Operational Case Study	TBC	In preparation
6	e	The whole life cost and carbon perspective of using nature-based solutions (NBS) to improve rural wastewater treatment based on septic tanks	H2Open	In preparation

A literature review was conducted to assess the current state of onsite anaerobic treatment at ambient temperatures, identify low complexity optimisation methods and highlight knowledge gaps in low temperature anaerobic digestion. This literature review presents a novel perspective to assess anaerobic reactors with a quantification of the reactor complexity. This literature review challenges the current design assumption that temperature is a dominant factor of hydrolysis rates in onsite reactor and highlights the lack of understanding of hydrolysis in low temperature systems for further investigation. This review entitled, Onsite anaerobic reactor design and complexity, the impact on treatment and hydrolysis in temperate conditions: A systematic review comprises chapter 2 of this thesis and is in preparation for submission to Environmental Technology Reviews.

Chapter 3 is world's first determination of hydrolysis rate constants below 20ºC for passive systems. The impact of temperature, seed and mixing on the hydrolysis rate constant was examined. Additionally established in these conditions that the first order model provided a good fit to the experimental data. The assessment of model fitting and experimental data show that mixing is a key optimisation method of hydrolysis which is further tested in chapter 4. This chapter is in preparation for submission to Environmental Science: Water Research and Technology, under the title Impact of low temperatures on anaerobic hydrolysis in passive treatment systems such as septic tanks

Chapter 4 examines differing baffle configurations, to manipulate flow hydraulic to maximise the hydrolysis rate in a reactor whilst maintaining treatment performance. Against a control septic tank, it was found that a combination of vertical and horizontal baffles could significantly reduce the desludging frequency of a septic tank. Computational fluid dynamic (CFD) modelling was used to assess the impact of the baffle on the flow conditions in the system. This chapter has been submitted for publication to Water Research under the title Enhancing the hydrolysis rate within septic tanks through baffling to extend desludging frequency.

Chapter 5 assess the performance of a pilot vertical flow and an aerated horizontal flow wetland treating septic tank effluent. Both systems showed strong removal and resilience and are both suitable technologies for use in the proposed flowsheet. This chapter is being prepared for submission, under the title Using constructed wetlands as a nature-based solution (NBS) for septic tank effluent.

Chapter 6 assesses the whole life cost and lifetime carbon emissions of the proposed flowsheet compared to an STS and a package treatment system. The proposed flowsheet was found to significantly reduce lifetime carbon emissions compared to both an STS and package treatment system. Although the proposed flowsheet typically has a higher whole life cost than an STS, it is a cost-effective abatement strategy alongside improving wastewater treatment and providing opportunities for delivering additional co-benefits like amenity value . This chapter is in preparation for submission to H2Open under the title The whole life cost and carbon perspective of using nature-based solutions (NBS) to improve rural wastewater treatment based on septic tanks

The overall implications of the research are presented in Chapter 7 alongside an assessment of the technical obstacles to wide scale adoption of the flowsheet for onsite wastewater treatment. Concluding remarks and recommendation for further work are provided in Chapter 8.

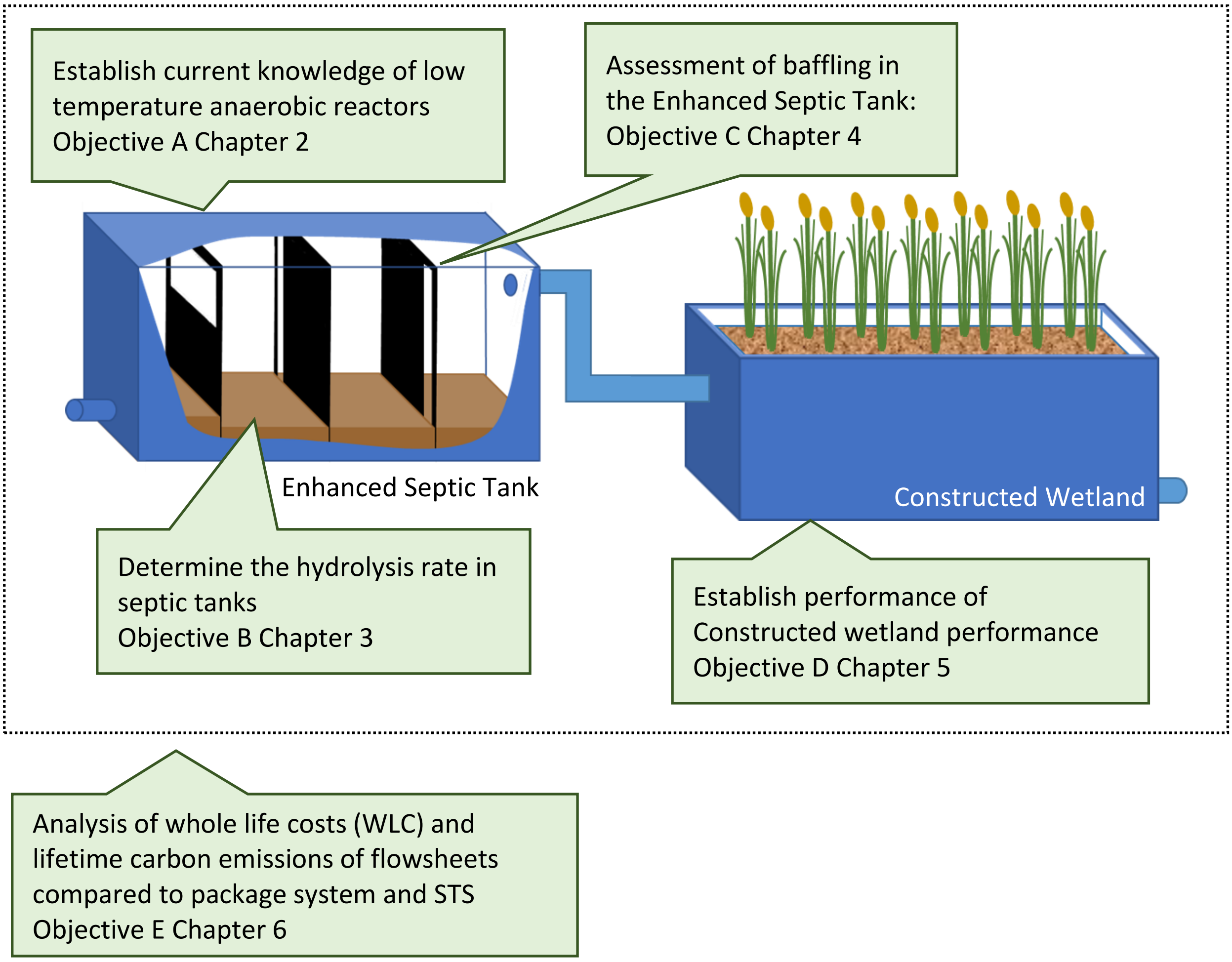

Figure 1.3: Thesis overview

1.5. References

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2. Onsite Anaerobic Reactor Design and Complexity, the Impact on Treatment and Hydrolysis in Temperate Conditions: A Systematic Review

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Abstract:

Onsite anaerobic treatment constitutes a sizeable portion of the world's wastewater treatment. These systems are predominantly operated at ambient temperatures and require simple designs. This study conducts a systematic review of four onsite anaerobic reactor designs and analyses the impact of operational conditions. Additionally, there is a novel insight into the complexity of the reactors, quantifying a previously qualitative parameter and providing an additional vector for optimisation. Broadly, at ambient temperatures the treatment of septic tanks, anaerobic ponds (AP), anaerobic baffled reactors (ABR) and upflow anaerobic sludge bed (UASB) reactors are with the same range. The greatest impact on the removal of TCOD and TSS was found to be temperature, with hydraulic residence time (HRT) having limited impact on TCOD removal. Complexity provides an assessment criterion for the efficiency of design modifications. The use of limited baffling was found to be an effective improvement for APs, ABRs and UASBs but excessive baffling in ABRs provided little additional benefit. There is limited reporting of the hydrolysis rate constant of onsite reactors despite being a promising optimisation criterion. Converse to removal of TCOD and TSS, temperature was not a dominant parameter for the hydrolysis rate constant of onsite reactors at ambient temperatures. Additionally, reactor modifications which increased complexity did not effectively increase the hydrolysis rate constant, unlike TCOD removal. These findings have the impact to reduce excessively complex designs for onsite anaerobic systems and provides a benchmark for the impact of complexity.

Keywords: Kinetics, Psychrophilic, Optimisation

2.1 Introduction

The estimated global discharge of domestic wastewater in 2020 was 271 billion m^3 (UN Habitat & WHO, 2021) of which 43% was discharged to non-sewered systems. Non sewered systems consist of both open defecation and onsite treatment such as septic tanks which represents 24% of global domestic wastewater treatment (UN Habitat & WHO, 2021). In most cases, onsite treatment technologies need to be installed and operated easily with minimal maintenance requirements. Example technologies include septic tanks, anaerobic ponds (AP), anaerobic baffled reactors (ABR) and upflow anaerobic sludge bed (UASB) reactors. The common features of anaerobic systems are that they require no aeration, produce less sludge than aerobic systems whilst still removing key containments, albeit with a lower removal efficiency than aerobic systems. The principal function of such technologies is to retain faecal sludge and store it until it can be collected and transported to a sludge treatment site.

The technologies work by two predominate mechanisms: solids separation for treatment and anaerobic digestion for minimising the sludge accumulation rate. The former is primarily associated with particle settlement and is related to the size and density of the particles as commonly defined by Stokes' law or through sludge blanket capture. Once settled the degradation of the resultant sludge proceeds by anaerobic digestion. The sludge accumulation rate (SAR) within a reactor is an equilibrium between the settling and digestion in the system.

Anaerobic digestion has four primary stages: hydrolysis, acidogenesis, acetogenesis and methanogenesis. Hydrolysis is the process by which complex material is degraded into soluble monomers and is most commonly considered the rate limiting step of anaerobic digestion (Gomec, 2010). Acidogenesis and acetogenesis convert organic monomers to volatile fatty acids (VFAs) and VFAs to acetate and hydrogen respectively (Siegrist et al., 2002). These are the two quickest stages of anaerobic digestion and are not considered rate limiting (Siegrist et al., 2002).

Optimising anaerobic digestion is most efficiently achieved by optimising the rate limiting step, hydrolysis. The predominant model for the rate of hydrolysis (r_h) is the first order model (Eastman & Ferguson, 1981);

$$r_h = \frac{d[S]}{dt} = -k_h[S]$$

Where; t is time (d), [S] is substrate concentration (mg l^{-1}) and k_h is the hydrolysis rate constant (d^{-1}). The substrate was initially modelled as particulate COD (Eastman & Ferguson, 1981) but can also be considered as volatile suspended solids (Vavilin et al., 2008). The expression is an empirical function with the hydrolysis rate constant representing an accumulative term for the potential factors that can impact hydrolysis (Eastman & Ferguson, 1981). Hydrolysis is conducted by extracellular enzymes, which are secreted by microorganisms to breakdown molecules too large to be degraded inside the cell (Burgess & Pletschke, 2008). The majority of hydrolytic enzymes are cell-bound (Burgess & Pletschke, 2008), therefore for hydrolysis to occur there needs to be sufficient contact between the microorganisms and substrate. Enzymatic reactions can be considered in three primary stages: attachment, degradation and release. Attachment is the process by which the enzyme contacts with the substrate and then binds to the enzyme reaction site, degradation is the process by which the substrate is broken down into the degradation compounds and finally, release is the detachment of the degradation compounds from the reaction site into the bulk medium.

The degradation of sludge in the first order kinetic equation, models the degradation as uniform between the macro components of the sludge. There is no provision for the composition and potential different rates for the degradation of lipids, proteins and carbohydrates, instead hydrolysis is measured through particulate COD or volatile suspended solids.

The rates of attachment and release are mass transfer limited and hence are controlled by the degree of mixing whereas the degradation step is governed by temperature. The impact of temperature is commonly modelled as an Arrhenius relationship although there is no specific consensus on either the activation energy or pre-exponential factor values that should be used (Burgess & Pletschke, 2008). Beyond temperature and mass transfer, the hydrolysis rate constant is also impacted by pH (Ristow, 2005) although the impact of pH also has a temperature related dimension adding a second layer of uncertainty (Vazquez et al., 2004). Other factors such as particle size (Aldin et al., 2011), hydraulic residence time (HRT) (Ristow, 2005) and VFA concentration (Vavilin & Angelidaki, 2005) have also been shown to impact the hydrolysis rate constant.

The importance of temperature and mixing is emphasised as most onsite anaerobic reactors are operated under ambient temperature conditions. This means that the biological processes operated under either psychrophilic (<20^{o}C) or low end mesophilic (20 – 45^{o}C) conditions (Lew et al., 2009). The difference between the two is reported to be related to the increased solubility of methane as temperature decreases and a change in the composition of the VFAs (Cruddas et al., 2014). Given that the temperature is unamended, the efficacy of the process becomes strongly related to the design of the reactor through it impact on mixing and reactor profiling in terms of compartmentalisation of different the biological functions.

The current review aims to examine the role of design complexity on the efficacy of onsite anaerobic reactors for domestic wastewater treatment to understand key design features and the potential to upgrade reactors such as septic tanks. The review was conducted using Scopus (www.Scopus.com), with the search terms for each reactor of study (Table 2.1) within title, abstract and keywords. In order to be eligible for inclusion, studies were required to be published after 1965 and prior to 26/04/2022, with results available in English and been published as an article, review or book chapter. The performance of reactors was analysed for systems which were: treating domestic wastewater, with available influent and effluent characterisation for at least one performance metric, the reactor HRT and operating temperature below 35^{o}C. Preliminary title and abstract screening was conducted to establish the eligibility of a study prior to analysis of the full paper (Table 2.1). The review specifically excludes anaerobic sludge reactors such as anaerobic digestion, anaerobic reactors coupled with membranes and systems that are operated at or above 35^{o}C. These systems are considered beyond the scope of this study and are extensively analysed in the current literature (Wang et al., 2018; Smith et al., 2012; Holm-Nielsen et al, 2009). Another anaerobic system such as Pit latrine and cesspools are considered out of scope as they lack an outflow for removal and do not treat domestic wastewater.

Table 2.1: Search terms and number of returned results from Scopus search

Search Term	Results	Screened	Analysed
“Septic Tank”	3188	61	21
“Anaerobic Pond” or "anaerobic lagoon"	726	40	20
"Anaerobic baffled reactor" or "Anaerobic baffle reactor"	680	57	33
"Upflow anaerobic sludge bed" or "Upflow anaerobic sludge blanket" or UASB	4814	156	73

2.2 Anaerobic Reactors

The most common onsite anaerobic reactors are: septic tanks, APs, ABRs and UASBs (Figure 2.1). A septic tank is a tank fed with domestic wastewater which flows horizontally through the system and solids settle within the tank to form a sludge bed. The tank is frequently separated into multiple chambers with vertical baffles to aid with settling and prevent sludge washout. The outlet of a septic tank has a scum guard to prevent the washout of floating solids. The sludge within a septic tank is fed batch such that when a determined volume of sludge has accumulated, typically 33-50% of the total tank volume, the system is halted, and the sludge is pumped out with a vacuum tanker for further treatment. British Standard guidelines (2015) recommends an HRT from 29 to 60 hours for septic tanks.

An AP or lagoon is typically an earthen anaerobic wastewater treatment system (van Haandel et al., 2006) where anaerobic conditions are maintained due the loading rate, enabling operation of AP as either open or covered systems. Wastewater is treated in a similar manner to a septic tank with a typical HRT of less than 24 hours (Mara, 2004). An AP also manages sludge in a similarly way to septic tanks, separating the HRT from the sludge retention time, and pumping out sludge when full. Both septic tanks and APs are designed to leave the sludge bed undisturbed minimising the resuspension of particles but limiting the mass transfer in the sludge bed.

An ABR is an adaptation of a septic tank in which baffles are used to divert the flow through the settled sludge to provide agitation in order to enhance mass transfer and flocculation (Reynaud & Buckley, 2016). The number of baffles can vary between 2 and 8, although it is limited to ensure the upflow velocity in any compartment does not exceed 1 m h^{-1} to minimise resuspension of solids (Gutterer et al., 2009). Typical design criteria are based on a limiting organic loading rate of 3 kg_{COD} m^{-3} d^{-1} and an HRT of 8 hours (Gutterer et al., 2009).

A UASB is a bottom fed reactor in which the flow rises through the tank. There is a high biomass concentration in the bottom section of the reactor, forming a sludge blanket. Post the sludge blanket, the structure of the reactor is typically designed to optimise for three phase separation, with gas deflectors on the sides of the reactor and a central gas collection system. Beyond the gas-liquid separator is a settling compartment, which can have an increased diameter for reduced fluid velocity, in which the remaining suspended solids are settling without gas uplift. Finally, there is a clarifying baffle to reduce sludge carry over (von Sperling & Chernicharo, 2005). Similarly, to an ABR, forcing flow through the sludge blanket increases the mass transfer in these regions, increases flocculation and allows for filtration. Sludge is periodically removed from the reactor in either a fed batch or semi-continuous manner. A fed batch operation is not dissimilar to the sludge management of STs, APs and ABRs, when the sludge level is too high the system is stopped and either pumped out or sludge is drained from a valve at the bottom of the reactor. Alternatively, a valve can be placed at a predetermined height and sludge can be discharged during operation when the sludge level is above the operational sludge height allowing for semi-continuous operation. UASBs are typically a more intensified process than the other listed systems with an HRT of 6 – 10 hours being standard (von Sperling & Chernicharo, 2005).

The differentiation between the reactors can be unclear, with studies with comparable systems listing them under difference names. For example, there is no fixed baffling to an ABR, septic tank or AP and so all can be an anaerobic reactor with baffles. In this scenario the design philosophy of the baffling is the key distinguishing feature, with the baffles in an ABR used to encourage mixing and sludge contact compared to in a septic tank where baffles are used to minimise mixing and encourage settling. To alleviate ambiguity of different definitions used within the literature, reactors systems are characterised as per the definitions below in preference of reported definition of reactor type given in the literature. In cases in which a system cannot be classified into a category, it will be characterised as within the reporting paper.

Septic Tank: An anaerobic reactor which provides settling and storage for sludge with a continuous liquid phase. Settled sludge remains settled and undisturbed during operation. Can be single chamber or broken into several chambers using vertical baffles.

Anaerobic Pond: An earthen anaerobic reactor which provides settling and storage for sludge with a continuous liquid phase. The reactor can be either covered or uncovered.

Anaerobic Baffled Reactor: An anaerobic reactor which uses a series of baffles to force wastewater under and over baffles, providing agitation to the sludge layer.

UASB: A single upflow anaerobic reactor with a settled sludge bed which the influent is feed through.

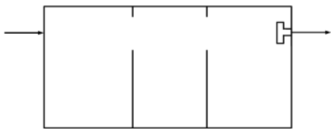
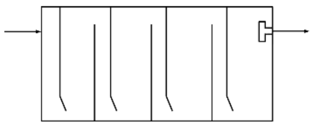
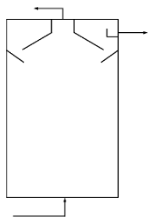
Septic Tank	Anaerobic Pond
	
Anaerobic Baffled Reactor	Upflow Anaerobic Sludge Bed
	

Figure 2.1: System diagrams of primary onsite anaerobic reactors

2.3 <u>Complexity</u>

The simplicity of a reactor design is often cited as a reason for considering it applicable (Gomec, 2010; Lettinga et al., 1993). The degree of complexity of a reactor design is rarely quantified although the idea aligns to the concept of design complexity which are heavily influenced by ideas of 'Inherent Safety' first described by Kletz (1978) in 'What you don't have can't leak', with the underlying principle being the minimisation of hazardous operations in a process plant rather than excessive control.

Complexity for a reactor system was defined by Ameri et al., (2008) as:

$$C = (dv + dr)ln(p + 2)$$

Where dv is the number of blocks, dr is the number of connections and p is the number of type of connections. A block in this study is considered as any element required to meaningfully depict the reactor on a system diagram such as; a pump, scum guard or baffle. A valve which is required purely for sampling purposes of an experimental regime is not counted, however a valve required for operation, such as a desludging valve in a UASB is considered. Connections represent the influent and effluent flows of the system. For example, considering a UASB (Figure 2.2). The number of connections is four, being the inlet, outlet, gas collection and sludge disposal, with all these connection types being mass flow connections giving a p of one. The UASB consists of a reactor tank, two gas deflectors, a clarifier baffle, a three-phase separator cone, a feed pump and a sludge control valve giving a total number of 7 blocks. This gives a total complexity of the system (Figure 2.2) of 12.2.

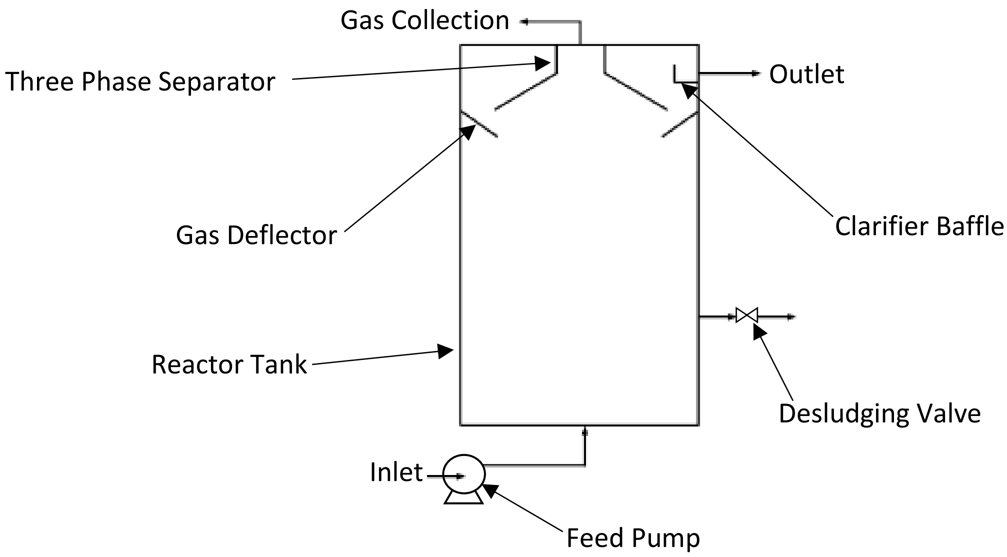


Figure 2.2: Conventional UASB and complexity quantification

2.4 Comparison of Reactor Types

2.4.1 Overall Reactor Performance

The median TCOD was 56%, 51%, 62% and 62% for septic tank, AP, and UASB reactors respectively. Equivalent data for TSS was 71%, 61%, 75% and 62% respectively. TSS removal was better than TCOD removal for the ABR, septic tank and AP systems reflecting the principal design feature is associated with separating solids. Whereas, the UASB systems had the same median levels for both TCOD and TSS reflecting the greater emphasis on reactivity and methane production. One-way ANOVA was conducted on the dataset (Supplementary Material 2.7.1), which found that the reactor type did have a significant impact on both the TCOD and TSS removal ($p<0.05$). Ad-Hoc testing between reactor categories found that for COD removal the systems could be divided into two categories the ST and AP group, and the ABR and UASB group ($p<0.05$). The ST and AP have no significant variation between each other but differ significantly from both the ABR and UASB with regards to COD removal, with the equivalent being true for the ABR and UASB respectively ($p<0.05$). Conversely for TSS removal the reactors could be divided into a group of ST and ABR with a second group of AP and UASB. This leads to the categorisation of the ABR being in the grouping with increased COD and TSS removal compared to a ST having increased TSS removal and lower COD removal, a UASB having increased COD removal but reduced TSS removal and finally the AP being in the lower grouping for both COD and TSS removal. However, this does not determine which variations between the reactors lead to this impact.

The removal of TCOD can be broken into two categories, particulate (pCOD) and soluble (sCOD). The removal of pCOD is influenced by the same factors as TSS removal, however sCOD is largely independent of settling. As expected, all reactor types had increased pCOD removal compared to sCOD removal. For example, median pCOD removal for the septic tanks and ABRs were 70% and 75% respectively and compared to median sCOD removal levels of 34% and 40% respectively. The median sCOD removal for the ABR and UASB were higher at 45% for the ABR and 62% for the UASB, congruent with the increased mass transfer delivered in these reactor designs. The higher pCOD removals reflect the strong link to suspended solids removal with the ratio between the two being close to parity in all reactor types.

The operation of UASBs is generally a more intensified process than the other reactors with a lower HRT and higher organic loading rates (OLR) (Table 2.2). In this respect UASBs achieve equivalent treatment to the other reactors with reduced sizing requirements, with a median HRT of 8 hours which is the below the 25th percentile for all other systems. The comparative median HRTs reported were 36 hours for the septic tank systems and 24 and 21 hours for the ABRs and APs respectively.

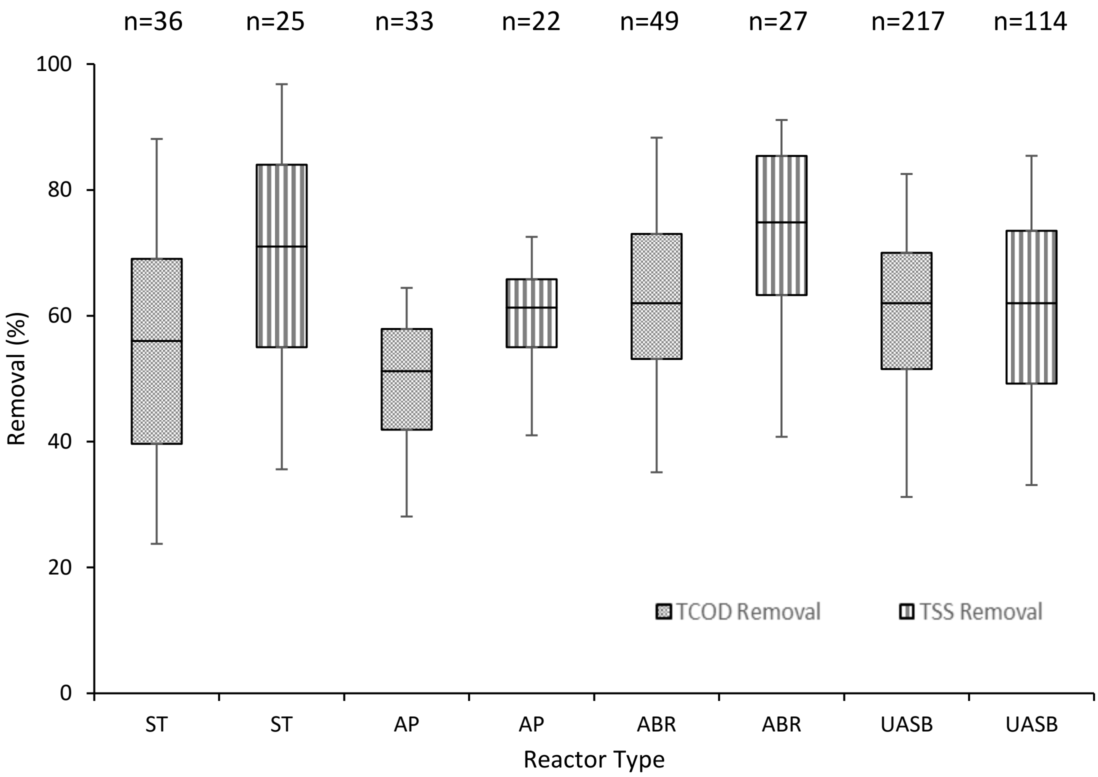


Figure 2.3: Box and whisker plot of the overall removal of literature reactors where boxes represent the interquartile range, with the divider being the median, whiskers extend to the 5th and 95th percentile.

The observed range of temperature operation for APs is limited with under 1 degree of variation from 25th to 75th percentile, this is partially due to 24 of the 39 operational data points coming from a single study (Peña et al., 2003). The other reactors had broader range of temperature operation, for instance an interquartile range of 12^{o}C for septic tanks and 9^{o}C for ABRs. The operational HRTs and loadings observed are broadly in-line with the expectation of a standard reactor for each respective technology type. Notably, the maximum HRT observed in a UASB of 132 hours is significantly greater than the typical system.

Table 2.2: Operational parameter of low temperature reactors from systematic review

		ST	AP	ABR	UASB
Temperature (^oC)	Min.	7.9	17	9	7.1
	25	15	24.5	15	19
	50	16	25	18	22.7
	75	27.4	25.2	24	26
	Max.	42.7	29.5	30	35
HRT (hours)	Min.	1.5	8	6	1
	25	24	16	12	6
	50	36	21	24	8
	75	81	48	48	15
	Max.	333	299	144	132
OLR (kg COD m^{-3} d^{-1})	Min.	0.043	0.045	0.043	0.07
	25	0.2	0.37	0.22	0.6
	50	0.4	0.63	0.41	1.11
	75	0.71	0.83	0.73	2
	Max.	1.7	1.4	2.1	17.2
Complexity (-)	Min.	4.4	3.3	5.5	5.5
	25	5.5	3.3	8.8	8.8
	50	6.6	4.4	12.1	9.9
	75	7.7	4.4	14.8	11.0
	Max.	11.1	6.6	20.9	15.4

2.4.2 Regression Analysis

To understand the relative impact of the different operating conditions on treatment a linear regression of the effluent TCOD and TSS was conducted (Figure 2.4). Within each reactor type, the impact of temperature, influent concentration, and HRT were analysed. Inter reactor analysis was conducted considering the previous factors and system complexity. Regression analysis was conducted using Matlab R2022b (www.matlab.mathworks.com), data was considered significant to a p value below 0.05.

The modelled equations were;

$COD_e = \alpha + \beta_1 COD_i + \beta_2 Temp + \beta_3 HRT + \beta_4 Complexity$

$TSS_e = \alpha + \beta_1 TSS_i + \beta_2 Temp + \beta_3 HRT + \beta_4 Complexity$

Where α is the intercept and β is the gradient for each factor, subfix e is effluent and i is influent. A factor is excluded from the linear regression model if found not to have a significant correlation ($p<0.05$). Normalised impact of each factor was examined by fixing all other modelled variables at the median value and varying the studied factor from the 5th to 95th percentile range. The normalised impact on removal was calculated by using influent and modelled effluent concentrations.

2.4.3 Organic Removal

When viewed across all reactor types, TCOD and TSS removal was shown to be most influenced by temperature (Figure 2.4). Temperature, normalised for other factors by linear regression, had a 32% and 25% increase in TCOD and TSS removal respectively from 10 to 30^{o}C. Comparatively, the impact of HRT across the 95% interval range of the dataset (4h to 105 h) was 8% and 23% for TCOD and TSS removal respectively. The impact of both temperature and HRT on TSS are compatible with a corresponding increase in viscosity and a decrease of water velocity. To further consider the impact of HRT on TCOD, the relative contribution of loading rate was considered after it was controlled for HRT. The overall impact of loading rate was a 1% change in both TCOD and TSS removal across the 95% range of influent range. This supports finding reported for ABRs treating domestic wastewater (Reynaud & Buckley, 2016) and reflects that the systems are overall not load limited, especially the unmixed systems (septic tanks and APs). All factors considered in the linear regression were found to have a significant relationship to effluent concentration, with the exception of complexity for TSS effluent which was thus excluded from the final linear regression model. This is supported by analysis of the individual reactor types where for the well mixed UASB systems, HRT rather than temperature is the dominant parameter for TSS removal. Beyond that the impact of temperature, HRT and OLR within the individual reactors are broadly in-line with the overall findings indicating similarity of design features that influence removal.

Controlled studies, examining the impact of an operational variable with all other factors fixed broadly align with the findings of the relationships from the linear regression (Abbasi et al., 2017; Serrano Leon et al., 2018). To illustrate, a septic tank and ABR operated at 27^{o}C showed increased COD removal of 12% and 17% as the HRT increased from 24 to 74 hours. Smaller increases in COD removal were reported for an ABR operated at 15^{o}C which increased by 3% between 12 to 24 hours and a further 3% between 24 to 48 hours (Haydar et al., 2018). At constant temperature and feed concentration, the impact of 7 HRTs ranging from 24 to 8 hours was investigated in an 82m^3 anaerobic pond (Peña et al., 2003). Peña et al., (2003) observed that COD removal increased with a reduction in HRT. Similarly, varying HRT from 12 to 36 hours in septic tanks at 16^{o}C resulted in change in COD removal of 3% (Chen et al., 2014).

Ceccoent et al., (2022a; 2022b) and Al Shayah and Mahmoud (2008) also analysed the impact of HRT on UASB systems and reported that increased HRT improved COD removal. In contrast, at 25^{o}C Chatterjee & Ghangrekar (2017) observed a decrease in COD removal in a UASB with increasing HRT from 8 to 24 hours. However, at 13^{o}C the impact of HRT was in-line with regression modelling in the same system (Chatterjee & Ghangrekar, 2017). The impact of temperature is consistent with the regression model across the individual studies. For instance, increasing temperature from 16 to 27^{o}C was reported to increase COD removal by 16% at an HRT of 48 hours (Mahmoud and Van Lier, 2011). However, a lower impact of temperature between 13 to 25^{o}C was reported for a UASB operating at an HRT of 4.7 hours where COD removal increased from 64 to 70% (Uemura and Harada, 2000).

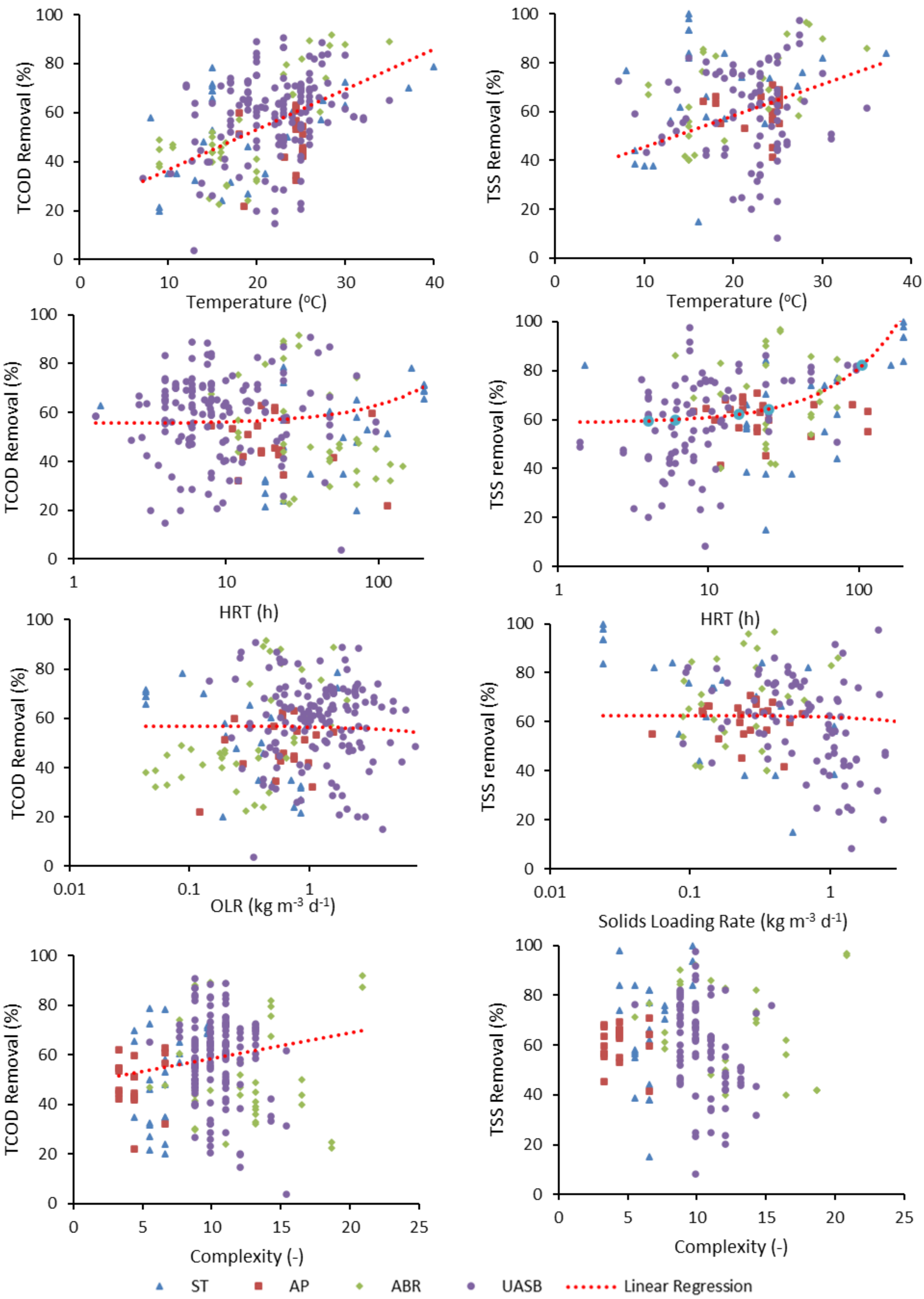


Figure 2.4: Impact of operational factors on removal (a-d) COD removal (e-h) TSS removal, a, e) Temperature b, f) HRT, c, g) Complexity, d, h) Loading rate Note: no significant relationship between TSS removal and complexity was found there is no fit of linear regression plotted.

2.4.4 Complexity

The second greatest impact, behind temperature, on COD removal was complexity, a measure of the intricacy of the reactor. The impact of increasing from the 5th to 95th percentile of complexity had a 10% increase on COD removal, 2.5% greater than the impact of HRT but 22% less than temperature. Notably, complexity was not found to have a significant impact on TSS removal, the only examined factor found to have a different relationship between TCOD and TSS (Figure 2.4). The relationship between COD removal and complexity does not imply that simply increasing the design complexity of a reactor increase performance rather that design alterations which add complexity are broadly successful at improving treatment.

The design variations and adaptations between and within reactors target different mechanisms. The addition of baffles in ABRs increases the sludge contact, hence mass transfer and flocculation compared to a conventional septic tank. The addition of media packing to a septic tank, such as tested by Chen et al., (2014) , does not increase the mass transfer but instead the settling by hindering particle travel. Complexity provides a parameter to assess the efficacy of these techniques.

The highest median complexity was observed for ABRs (Table 2.2) due to the presence of up to 17 baffles within a reactor. However, it is unclear what is the optimum number of baffles that are required for ABRs with studies ranging from 2 to 17. A negative impact of additional upflow baffles is broadly in-line with stokes' law as it will lead to an increase upflow velocity reducing settling. Therefore, within ABRs, the added complexity of additional baffles is not providing an effective improvement and thus the optimum number of baffles is likely to be towards the low end of the scale. However, comparing the impact of the baffling in ABRs to that of septic tanks, the median TCOD and TSS removal is 62% and 75% for ABRs and 56% and 71% for septic tanks (Table 2.2) so upflow baffling can lead to improved performance. Inclusion of upflow baffles in a low complexity system has been shown to provide an effective increase in treatment (Nasr & Mikhaeil, 2013). To illustrate, the work compared a baffleless septic tank and an ABR with two baffles both operating at 25ºC and across three different HRTs. The inclusion of baffles increased removal of TCOD and TSS by 7% and 12% respectively. This highlights the potential for baffling as a low complexity optimisation technique. The impact of horizontal and vertical baffles was examined by Cruddas et al., (2018, 2021). An AP with 2 horizontal baffles had a marginally higher TCOD and TSS removal than a ABR with 6 baffles, two of which were horizontal. This experimentation found that the profile of horizontal and vertical baffles had significantly different impact on treatment routes.

Das & Chaudhari (2009) investigated the impact of inclusion of a vertical baffle diagonally along the height of a UASB to promote mixing which corresponded to an increase in complexity of 1.1. The impact was an increase in TCOD of 12% irrespective of the three HRTs the UASB was operated at. Computational fluid dynamics, analysis of this system showed that the baffle provided an additional region of high mixing at the baffle whilst also lowering the flow velocity in the upper settling compartment of the reactor (Das et al., 2018).

Packing and filters are a common adaptation of anaerobic systems, observed for each reactor type excluding APs. A vertically baffled UASB was compared to UASB with filter media (Das & Chaudhari, 2015). Both adaptations provided the same increase to complexity and increased the TCOD removal at 4, 6 and 8 h HRTs. However, the vertically baffled UASBs had a greater impact on treatment than a UASB filter at all test conditions. Dohdoh et al., (2021) found that across five HRTs, a UASB filter provided an average 2% and 1% increase to TCOD and TSS removal respectively, below the expected impact of the complexity increase. This is therefore an inefficient adaptation for this system. Conversely, in a septic tank at 16ºC, Chen et al., (2014) found that a filter media provided at 42% TCOD and 17% TSS removal increase at an HRT of 36 hours, with less dramatic increases in removal at HRTs of 24 and 12 h.

The highest TSS removal in ABRs was observed in reactors will packed filters with TSS removals of approximately 90% (Khalekuzzaman et al., 2018; Sabry, 2010; Sharma et al., 2014; Singh et al., 2021). This was shown to occur irrespective of the number of baffles, with 90% removal reported with 2 baffles (Sharma et al., 2014) and 13 baffles (Khalekuzzaman et al., 2018) corresponding to complexity values of 8.8 and 20.9 respectively. Overall, this shows that a baffle can provide a low complexity modification for treatment improvement, but the benefits will occur through the addition of a small number of baffles only.

Complexity is a novel approach to assess the impact of reactor design on treatment performance. However utilising complexity as factor of analysis in the linear regression has the potential to bias the results. To assess the potential for bias and the utility of complexity as a factor of analysis the linear regression analysis of COD removal was considered both with and without complexity as a factor. It was not considered for TSS removal as complexity was not found to have a significant relationship and was therefore excluded from the final linear regression model. Comparing the variation in the modelled COD removal from the linear regression models to the literature reported values, it was found that the root square error of the dataset was greater without complexity, and thus the linear regression model considering complexity was a better fit to the data than without it. Thus, complexity was not biasing the findings of the linear regression.

The complexity modelled in this study is a structural complexity, that is, the complexity that arises from the system connections and interactions and is time independent (Wall, 2009). Although this measure of complexity benefits from ease of measurement, it is detached from the design goals of a system (Braha and Maimon 1998). For example, the current quantification method ignores the functional reasoning behind the component and will hence consider a baffle and pump of equivalent complexity. Although this can be helpful in the identification of excess components in a reactor, it provides limited information on the operational conditions of the reactor.

A dynamic complexity is a time-dependent characteristic arising from the various operational conditions of the system (Wall 2009). This measure can be directly related to the design goals but has a higher informational requirement

(Braha and Maimon 1998). At time of writing this study, there are no peer-reviewed dynamic complexities models which are suitable for wastewater. Although it is beyond the scope of this study to develop such a methodology, the development of such a method would allow for a more precise quantification of the complexity of anaerobic reactors and other wastewater systems. It is suggested that a dynamic complexity quantification could consider the potential failure states of a reactor, similar to what was outlined by Koolen (2001), and provide quantification through the operational responses to an issue and the number of diagnosis tests (or information) required to identify the operational response. The complexity could hence be calculated via information theory. For example, the number of operational responses for a septic tank are limited (i.e., primarily desludging) and it is hence low complexity however there are numerous potential variables and operational responses for a more complex system such as a UASB.

2.4.5 Methane Generation

Methane generation provides a more direct examination of operational parameters on anaerobic digestion. Methane data was predominantly reported for UASB reactors with no septic tanks and 3 APs studies in the study criteria found to have reported methane data (Table 2.3). Linear regression of the methane generation dataset was conducted analysing the impact of OLR, temperature and HRT against the methane generation ($g_{\text{COD CH4}}\ g_{\text{COD in}}^{-1}$). No significant ($P<0.05$) correlations with methane generation and temperature, HRT or OLR were found for the dataset of psychrophilic reactors (Supplementary Table A 2.7). This does not mean that these factors do not impact methane generation, simply no correlation was found in this dataset. For example, in a UASB at a fixed temperature and fed strength, Serrano Leon et al., (2018) found increasing the HRT from 6 to 36 hours at 19^{o}C increased the methane generation efficiency from 0.12 to 0.15 $l_{\text{CH4}}\ g_{\text{COD removed}}^{-1}$. Conversely at 17^{o}C with the same system minimal variation was observed varying HRT from 8 to 18 hours (Serrano León et al., 2018).

The simultaneous running of a 2 baffle AP and 6 baffle ABR, at an average temperature of 10.5^oC and HRT of 55 hours, found that the presence of upflow baffles increased the conversation efficiency from 0.16 l_{CH4} $g_{COD\ removed}^{-1}$ in the AP to 0.24 l_{CH4} $g_{COD\ removed}^{-1}$ in the ABR (Cruddas et al., 2018, 2021). Comparing the methane generation efficiency of the 7 ABR and 32 UASB with operating conditions, both the methane generation (0.23 and 0.08 l_{CH4} $g_{COD\ in}^{-1}$ for ABR and UASB respectively) and efficiency (0.31 and 0.13 l_{CH4} $g_{COD\ removed}^{-1}$ for ABR and UASB respectively) is greater in ABRs, with equivalent mean operational temperatures (19 and 22^oC for ABRs and UASBs respectively). The increase HRT and reduced loading rate of ABRs could be the reason for the increased methane generation, however a UASB at 35^oC with a high HRT and low OLR was found to operate in-line with the average UASB (Gao et al., 2020).

Table 2.3: Average methane production for onsite wastewater treatment systems (range of reported data)

	Number of systems	Average Methane Generation ($L\ g_{COD\ in}^{-1}$)	Average Methane Generation Efficiency ($L\ g_{COD\ removed}^{-1}$)
ST	0	-	-
AP	3	0.09 (0.07-0.11)	0.27 (0.16-0.45)
ABR	7	0.23 (0.11-0.5)	0.31 (0.24-0.4)
UASB	37	0.08 (0.007-0.13)	0.16 (0.012-0.44)

2.4.6 Sludge Accumulation

The SAR in reactors with fed batch characteristics is given by two ordinary differential equations based on a mass balance of the flows and first order hydrolysis kinetics. The equations are;

$$SAR = \frac{dV_s}{dt} = \frac{Q_{in}\,[TSS]_{in}R}{\rho_s} - \frac{k_h[S]V_s(1-f)}{\rho_s}$$

$$\frac{d[S]}{dt} = \frac{Q_{in}\,[TSS]_{in}R}{V_s}\left(\frac{[S]_{in}}{[TSS]_{in}} - \frac{[S]}{\rho_s}\right) - k_h[S](1 - \frac{[S](1-f)}{\rho_s})$$

Where V_s is the sludge volume (m^3), Q_{in} is the volumetric flowrate ($m^3\ d^{-1}$), [TSS] is the TSS concentration, R is the solid removal ratio, ρ_s is the sludge density (kg TSS m^{-3}), and f is the combined biomass yield and inert solid generation (-), an in subscript denotes influent characteristic. The derivation of these equations is in Supplementary Material 2.7.2. The SAR is a balance between the settling and digestion terms of the differential equation, the SAR will become constant when an equilibrium between the terms is reached. However measured SAR is not constant (Bounds, 1997; Gray, 1995; Mahon et al., 2022; Philip et al., 1993), this is frequently attributed to the acclimatisation period of systems such as septic tanks, being several years (Mahon et al., 2022). However, modelling the sludge accumulation, with a fixed hydrolysis rate constant in the range observed for operational system (Table 2.4), also leads to a sludge accumulation rate which decreases with time. This is because prior to reaching equilibrium, settling adds more substrate than is being removed at any given time, this leads to increasing substrate mass in the sludge bed. First order hydrolysis kinetics state the hydrolysis rate increases with increased substrate mass meaning more of the bed is digested, the hydrolysis rate increases until the substrate mass is sufficient that the hydrolysis rate is equal to the substrate settling rate producing a steady state. Therefore, a decreasing sludge accumulation rate is not sufficient evidence to show that acclimatisation of onsite anaerobic reactors takes years as this is also true for modelled systems with a constant degradation rate constant.

Therefore, the SAR is primarily a function of the removal ratio and the hydrolysis rate constant, on the basis the other factor is limited by fed characteristics or biological limits. For reactors with fed batch characteristics for solids, the rate of sludge accumulation is a key metric for determining the maintenance period. However, the reporting of sludge accumulation data for each system is limited, and less which can be transformed to consider the hydrolysis rate constant. No study within the review conditions was found which reported the hydrolysis rate constant of the study, and only 6% of the suitable studies could be used to calculate the hydrolysis rate constant (Table 2.4).

The hydrolysis rate constant of reactors is independent from the solids removal. Temperature and mass transfer are the two expected key parameters from a mechanistic perspective. The highest hydrolysis rate constants observed are for UASBs with high hydrolysis rate constants of 0.058 d^{-1} (Table 2.4). These are the reactors which are most optimised for mass transfer, directly feeding into the sludge bed. This is not in-line with observations of methane generation or generation efficiency in which ABRs outperformed UASBs, this is noteworthy as hydrolysis is expected to the rate limiting step and therefore the hydrolytic performance is expected to correlate to methane generation.

Al Jamal & Mahmoud (2009) found that reducing the HRT from 96 to 48 hours for a UASB increased the hydrolysis rate constant from 0.036 to 0.058 d^{-1}. Conversely, for ABRs decreasing the HRT from 24 to 8 hours decreased the hydrolysis rate constant from 0.030 to 0.0046 d^{-1} (Nasr et al., 2009), and decreasing the HRT from 72 to 24 hours decreased the hydrolysis rate constant from 0.0081 to 0.0025 d^{-1} (Nasr & Mikhaeil, 2013). There is not a clear mechanistic reason for a lowering of HRT decreasing the hydrolysis rate. Excessive baffling was not found to help TSS or TCOD removal, it is possible that in ABRs the systems where under stress at lower HRTs and a reduction in loading reduced the stress and increased hydrolysis. Unlike optimisation for treatment, the variation of HRT to optimise for hydrolysis is hence unclear.

Across the reported temperature range of 11 to 28^{o}C, the hydrolysis rate constant shows a slight negative correlation to temperature (Pearson = -0.29), this is contrary to the expectation of an Arrhenius relationship. This is likely due to the high variation in other factors (Table 2.4) but demonstrates that temperature does not dominant and is not the only parameter for optimisation of hydrolysis. In the limited dataset, temperature has a limited impact and showed much greater optimisation potential for removal than hydrolysis. The hydrolysis rate constant observed is independent from complexity (Pearson = -0.05) hence the optimisation methods observed (Table 2.4) are not impactful on hydrolysis. A low complexity method to improve the hydrolysis rate constant of anaerobic reactor could significantly impact the SAR and hence the maintenance requirements of the systems. This study highlights that is an avenue for further examination.

Table 2.4: Hydrolysis rate constants determined from available literature studies.

Reactor	Temp (^oC)	HRT (h)	OLR (kg COD m^{-3} d^{-1})	Complexity (-)	TSS R (%)	TCOD R (%)	Hydrolysis Rate Constant (day^{-1})	Reference
ST	15	165	0.087	6.5	82	78	0.016	(Lin et al., 2019)
	24	48	0.44	7.7	58	56	0.0063	(Nasr & Mikhaeil, 2013)
	27	24	0.89	7.7	55	53	0.0087	
	28	72	0.32	7.7	65	65	0.014	
AP	17	74	0.18	5.5	77	52	0.032	(Cruddas et al., 2014)
	11	55	0.20	5.5	71	47	0.016	(Cruddas et al., 2018, 2021)
ABR	24	8	2.1	14.3	69	68	0.0046	(Nasr et al., 2009)
	23	12	1.3	14.3	71	76	0.016	
	23	18	0.96	14.3	74	80	0.022	
	25	24	0.67	14.3	82	82	0.030	
	27	24	0.89	7.7	68	57	0.0025	(Nasr & Mikhaeil, 2013)
	24	48	0.44	7.7	72	64	0.0045	
	28	72	0.32	7.7	76	74	0.0081	
	15	28	0.33	18.7	42	25	0.0084	(Schalk et al., 2019)
	11	55	0.20	8.8	67	46	0.0087	(Cruddas et al., 2018, 2021)
UASB	17	48	0.45	8.8	74	51	0.058	(Al-Jamal & Mahmoud, 2009)
	17	96	0.23	8.8	78	54	0.036	

	23	5.7	2.1	8.8	46	48	0.024	(Elawwad & Hazem, 2017)

2.5 Conclusion

A review of onsite anaerobic reactors operated under ambient temperature conditions for wastewater treatment revealed similar levels of removal across all reactor types. The differences in design impacted the ability to deliver the level of treatment in smaller, more compact reactors as illustrated with UASBs. Across all reactor types, temperature was seen to be the most influential factor on TCOD and TSS removal with a secondary influence from HRT which was mostly observed with respect to TSS removal.

Consideration of design decisions on performance was viewed through the lens of complexity and revealed that targeted design amendments could yield significant improvements but there was generally a diminishing return on adding increasingly more complexity. Importantly this indicates that future design thinking should focus on simple modifications such as improving basic baffling. This is particularly relevant for septic tanks which provide a large proportion of all onsite sewage treatment. Future septic tank design with improved baffling arrangement offers the potential for enhanced removal. However, the connection with hydrolysis and the resultant sludge accumulation rate is currently poorly defined and represents an area for future focus given its link to the required frequency of desludging.

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Data underlying this paper can be accessed at: 10.17862/cranfield.rd.21731777

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2.7 Supplementary Material

2.7.1 ANOVA Methodology and Results

Analysis of Variance (ANOVA) testing was conducted in Excel. One-way ANOVA was chosen as it provides a method to determine a difference of a set of discrete variables on continuous data, such as the reactor type and COD removal.

For each reactor categorisation the mean COD and TSS removal were calculated. The Sum of the Squares for each reactor (SS_r) was then calculated by, subtracting the mean removal from each datapoint and squaring the resultant value. The sum of squares within (SSW) was calculated by adding all the SS_r for the given test. For example for the overall dataset test adding the SS_r for each reactor category. The sum of squares between (SSB) was calculated using the same methodology as for SSW, however using the overall mean removal of the relevant dataset rather than the localised mean. The degrees of freedom for the SSW (df_{ssw}) were calculated by subtracting the number of categories from the number of datapoints. The degrees of freedom for the SSB (df_{SSB}) were calculated by subtracting 1 from the number of categories analysed. The F ratio is calculated using the following equation;

$$F\ ratio = \frac{\left(\frac{SSW}{df_{SSW}}\right)}{\left(\frac{SSB}{df_{SSB}}\right)}$$

The significance of the F ratio is then given by the critical F value which is determined based on df_{ssw} and df_{SSB} and the significance level, in this study a $p<0.05$ was considered significant. Thus if the F ratio is greater than the critical F value the categories analysed are found to be significantly different.

The F ratios for the COD and TSS removal across the overall dataset are 8.06 and 3.30 respectively. With the critical F values being 2.63 and 2.65 respectively, hence there is significant variation between reactors for both COD and TSS removal.

Table A 2.1: F ratios between reactors with respect to COD removal

	ST	AP	ABR	UASB
ST	-	0.42	8.30	8.72
AP	0.42	-	15.65	15.39
ABR	8.30	15.65	-	0.48
UASB	8.72	15.39	0.48	-

Table A 2.2: Critical F values between reactors with respect to COD removal

	ST	AP	ABR	UASB
ST	-	f(1,67)=3.98	f(1,83)=3.96	f(1,255)=3.88
AP	f(1,67)=3.98	-	f(1,80)=3.96	f(1,241)=3.88
ABR	f(1,83)=3.96	f(1,80)=3.96	-	f(1,264)=3.88
UASB	f(1,255)=3.88	f(1,241)=3.88	f(1,264)=3.88	-

Table A 2.3: F ratios between reactors with respect to TSS removal

	ST	AP	ABR	UASB
ST	-	2.18	0.00	5.23
AP	2.18	-	3.24	0.21
ABR	0.00	3.24	-	6.34
UASB	5.23	0.21	6.34	-

Table A 2.4: Critical F values between reactors with respect to TSS removal

	ST	AP	ABR	UASB
ST	-	f(1,45)=4.06	f(1,50)=4.03	f(1,137)=3.91
AP	f(1,45)=4.06	-	f(1,47)=4.05	f(1,134)=3.91
ABR	f(1,50)=4.03	f(1,47)=4.05	-	f(1,134)=3.91
UASB	f(1,137)=3.91	f(1,134)=3.91	f(1,134)=3.91	-

2.7.2 Sludge Accumulation

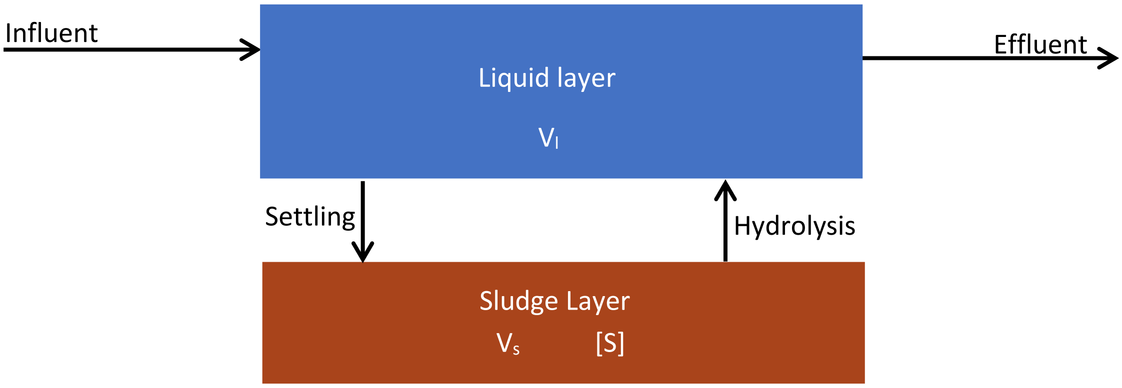

Figure A2.1: Simplified diagram of a septic tank

The model for the sludge accumulation rate is based on a mass balance on the sludge layer. The assumptions used to generate a sludge accumulation rate model are below.

Assumptions

1 The reactor volume is fixed

2 Hydrolysis is a first-order reaction

3 The sludge layer is well mixed, there is no spatial variation in concentration of sludge or constituent elements

4 Sludge concentration is constant with respect to time

5 Inert and biomass generation from hydrolysis is a fraction of hydrolysis and can be given by a combined term

Inlet= $Q_{in}([TSS]_{in} - [TSS]_{out}) = Q_{in}\ [TSS]_{in}R$ (1)

Where Q_{in} is the influent volumetric flowrate and [TSS] is the concentration of total suspended solids and R is the removal ratio.

Solids destruction = Hydrolysis = $k_h[VSS]V_sf_b$ (2)

Generation= $fk_h[VSS]V_s$ (3)

Where k_h is the hydrolysis rate constant, fb is the biodegradable fraction and f is the combined inert and biomass generation. Combining equations 1 to 3;

$$\frac{d}{dt}Sludge\ Mass = \frac{d}{dt}[S]V_s = Q_{in}\ [TSS]_{in}R - k_h[VSS]f_bV_s(1-f) \qquad (4)$$

$$\frac{dV_s}{dt} = \frac{Q_{in}\ [TSS]_{in}R}{[S]} - \frac{k_h[VSS]f_bV_s(1-f)}{[S]} \qquad (5)$$

Where [S] is the sludge concentration. To model the change in sludge volume the change of VSS with respect to time is required. This can be generated using the equations 1 -5, making two adjustments; the inlet of VSS can be given as a fraction of the inlet of the TSS and that the f term isn't required. Thus;

$$\frac{d[VSS]V_s}{dt} = f_{mvss}Q_{in}\ [TSS]_{in}R - k_h[VSS]f_bV_s \qquad (6)$$

Where f_{mvss} is the mass fraction VSS to TSS. Using the chain rule;

$$[VSS]\frac{dV_s}{dt} + V_s\frac{d[VSS]}{dt} = f_{mvss}Q_{in}\ [TSS]_{in}R - k_h[VSS]f_bV_s \qquad (7)$$

If $V_s > 0$

$$\frac{d[VSS]}{dt} = \frac{Q_{in}\ [TSS]_{in}R}{V_s}\left(f_{mvss} - \frac{[VSS]}{[S]}\right) - k_h[VSS]\left(1 - \frac{[VSS](1-f)}{[S]}\right) \qquad (8)$$

Finally;

$$Q_{in} = Q_{PE}PE \qquad (9)$$

Where Q_{PE} is the flowrate per population equivalent and PE is the population equivalent the system is serving. Equations 5 and 8 make a set of ODEs that can be solved using numerical integration. This is modelled using MATLAB and the inbuilt function ODE45, producing a model of the sludge volume over time.

2.7.3 Linear Regression Factors

Table A 2.5: Linear Regression Factors of COD

	Estimate	pValue
Intercept	249	$1.9\text{x}10^{-10}$
COD_i	0.44	$2.2\text{x}10^{-49}$
Complexity	-5.61	0.0080
Temp	-8.62	$3.6\text{x}10^{-12}$
HRT	-0.40	0.011
Excluding Complexity		
	Estimate	**pValue**

Intercept	178	1.8×10^{-10}
COD_i	0.45	3.6×10^{-50}
Temp	-8.163	4.1×10^{-11}
HRT	-0.33868	0.032

Table A 2.6: Linear Regression Factors of TSS

	Estimate	**pValue**
Intercept	104	2.2×10^{-8}
COD_i	0.38	2.7×10^{-24}
Complexity	-1.4	0.36
Temp	-3.5	3.1×10^{-4}
HRT	-0.62	4.3×10^{-7}
Excluding Complexity		
	Estimate	**pValue**
Intercept	88	9.6×10^{-5}
COD_i	0.38	6.6×10^{-25}
Temp	-3.5	3.5×10^{-4}
HRT	-0.60	7.3×10^{-7}

Table A 2.7: Beta values and significance levels of linear regression on methane dataset.

	Beta Value	**P Value**
OLR	0.0116	0.47
Temperature	0.006	0.061
HRT	-0.001	0.15

Table A 2.8: R^2 of linear regression to literature TCOD and TSS removal for operational factors as shown in Figure 2.4

	TCOD Removal	TSS Removal
Temperature	0.17	0.15
HRT	0.0026	0.076
Loading Rate	0.0044	0.13
Complexity	0.0057	-

3. Impact of Low Temperatures on Anaerobic Hydrolysis in Passive Treatment Systems Such as Septic Tanks

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Abstract

Psychrophilic anaerobic treatment of wastewater is an area of increasing consideration. The rate limiting step in such systems especially in passive variants, is hydrolysis. However, there is a genuine paucity of reported data sub 20°C limiting understanding of psychrophilic hydrolysis. In this study as series of batch tests from 5 to 37°C were conducted to determine the impact of temperature sub 20°C on hydrolysis. Additionally, the impact of seeding and mixing on the hydrolysis rate constant were investigated. For all test conditions both first order and Contois models showed a significantly better to fit the experimental data than the Michaelis-Menten model highlighting that the limiting condition to the hydrolysis rate is contact of the biomass to substrate rather than biomass growth. This finding was supported by the observation of an increase in first order hydrolysis rate constant from 0.0073 d^{-1} in an unmixed test at 10°C to 0.022 d^{-1} in a well-mixed test at 10°C. In fact, gentle mixing increased the hydrolysis rate constant above that observed at 37°C in an unmixed test. This study highlights that for the optimisation of anaerobic digestion in passive system in temperate conditions the focus should be on a systems mass transfer, rather than biological growth.

Keywords; Hydrolysis, Psychrophilic, Kinetics

3.1 Introduction

The treatment of domestic wastewater or faecal sludge under anaerobic conditions is achieved globally through a range of technologies such as septic tanks, anaerobic ponds, upflow anaerobic sludge blanket reactors (UASB) and anaerobic digesters (Chapter 2). Septic tanks are used throughout the world contributing to an estimated 24% of global wastewater treatment (UN Habitat & WHO, 2021). The main driver when using septic tanks is to provide retention and storage of faecal sludge whilst providing a basic level of treatment. The key maintenance requirement of such processes is periodic desludging, the frequency of which is determined by the combination of the storage volume, solids loading rate and the rate of biological breakdown of the captured solids through hydrolysis. Hydrolysis is a temperature impacted process such that the majority of applications of intensified anaerobic processes such as UASBs and anaerobic digestion are used in warm climates or heated. Whereas, septic tanks are operated under ambient temperatures which in a temperate climates means a temperature range between 5 and 20^{o}C (Cruddas et al., 2021).

The anaerobic biological processes that occur when treating complex organic material are commonly described by four stages: hydrolysis, acidogenesis, acetogenesis and methanogenesis (Batstone et al., 2002). Hydrolysis is generally considered the rate limiting step for degradation of particulate organic matter (Vavilin et al 2008) and suspended solids (Pavlostathis & Giraldo-Gomez, 1991) with the rate limitation particularly evident under low temperatures (Miron et al., 2000; Zeeman & Lettinga, 1999). In fact, hydrolysis remains the least defined stage and represents a combination of disintegration and solubilisation through enzymatic hydrolysis which can be considered by two different conceptual models related to the hydrolytic microorganisms:

1. The organisms first attach to a particle and then excrete enzyme in the vicinity to benefit from the soluble products formed (Vavilin et al., 1996).
2. The organisms release the enzymes into the bulk liquid and are adsorbed onto a particle (Jain et al., 1992).

Importantly, in most passive (or near passive) systems like septic tanks, the disintegration rate is likely to be rate limiting such that the influence of enzymatic hydrolysis of lipids, carbohydrates and proteins can be ignored (Feng et al., 2006). This is because the lack of mixing within the sludge layers restricts the available number of adsorption sites such that the enzyme concentration is in excess of the available adsorption sites (South et al., 1995). When the solids are covered by an even layer of hydrolytic organisms the surface will degrade at a constant depth per unit time dictated by the transfer of organisms onto the new layers (Vavilin et al, 1996).

The overall hydrolysis rate potentially depends on the feed concentration, temperature and the nature of both the seed and feed material (Gavala et al., 1999; Hills & Nakano, 1984) and this may impact hydrolysis rates when the make-up of the constituents is different. For instance, due to the impact of changing particle size, Hobson (1987) suggested that hydrolysis should be described relative to the available surface area rather than the concentration of substrate. Accordingly, a common way to attempt to enhance hydrolysis rates is to provide mechanical pre-treatment to change the available contact surface area through decreasing particle size, breakdown of polymer material or provide mixing to enable an increase rate of mass transfer. To illustrate, decreasing the particle size of animal manure from 840-590 mm to 590-350 mm in an anaerobic digester increased the biogas yield by 29% after 94 hours of treatment (Wen et al., 2004). The digestion of protein particles with a median particle size decreasing from 628 to 25 µm lead to 10-fold increase in hydrolysis rate constant (Aldin et al., 2011), with the hydrolysis rate constant having a logarithmic relationship to specific surface area. This is congruent with the expectations of a site limited reaction (Aldin et al., 2011).

The addition of hydrolytic enzymes has been shown to increase the degradation of waste activated sludge, primary sludge and mixed sludges in well mixed anaerobic systems (Rashed et al., 2010; Yang et al., 2010; Yu et al., 2013). However, investigations into the effect of biological and enzymatic additives into septic tanks have shown limited impact. For instance, trailing 48 septic tanks for a year, Pradhan et al., (2011) found that 3 commercially available additives had no significant impact on the sludge accumulation or treatment. The additives were not identified but purported to contain varying cultures of facultative anaerobes and proteases, amylases, lipases and cellulases. Additionally, Diak et al., (2012) found that two bench scale septic tanks dosed with 58 mg enzymes / L reactor and 116 mg enzymes / L reactor biweekly respectively, had no significant impact on sludge accumulation compared to an undosed control. The enzyme dose in this study (Diak et al., 2012) consisted of lipase (500 FIP/ g), amylase (1,000 BAU/g), protease (7,900 PC/g), cellulase (450 CU/g) and mixture of aerobic and anaerobic bacteria (2x10^9 CFU/g). Conversely, disturbing the formed sludge bed of a septic tank has been found to decrease sludge accumulation whilst increasing removal, implying increased hydrolysis (Almomani, 2016).

The impact of temperature in psychrophilic conditions is rarely considered with a genuine paucity of hydrolysis rate data below 20^{o}C which apply to ambient, passive systems such as septic tanks. To the authors knowledge the only available data on hydrolysis rate constants are at 10^{o}C in a mixed system (Ferreiro & Soto, 2003) and 15^{o}C in an unmixed passive system (Mahmoud et al., 2004). Previous related work has shown an increase in the production of colloids and non-degradable soluble compound when the temperature was decreased from 20^{o}C to 10^{o}C (Lew et al., 2009). This was interpreted into terms of a change in the relative rates of disintegration and solubilisation such that the proportion of solids converted in soluble organics decreased from 43% at 20^{o}C to 21% at 10^{o}C

3.1.1 Hydrolysis Models

The most common approach to hydrolysis modelling is to utilise first order kinetics of the form (Eastman and Ferguson, 1981):

$$\frac{d[S]}{dt} = -k_h[S]$$

Where t is time (d), [S] is the hydrolysis substrate concentration (mg l^{-1}) and k_h is the hydrolysis rate constant (d^{-1}) and represents an accumulative term for all the potential factors and so is an empirical expression of hydrolysis (Eastman & Ferguson, 1981). The applicability of first order kinetics has been shown to be appropriate when enzyme concentrations are in excess due to limited availability of solid surface adsorption sites (Sanders et al, 2003). The resultant hydrolysis rate does not depend on hydrolytic biomass concentrations and does not fit to a sigmoid type curve where the hydrolysis rate increases with time due to biological establishment (Vavilin et al, 2008). The suitability of the first order rate expression has been reported for the hydrolysis of primary sewage at 35 °C in terms of carbohydrate and protein hydrolysis by expressing the components as COD and relating the kinetics to degradable particles (Eastman and Ferguson, 1981). The use of batch tests to determine first order rate constants was deemed suitable as long as a well-balanced inoculum was used (Pablon Pereira et al, 2009).

In contrast, Miron et al, (2000) suggest that none of the individual components within primary sludge follow first order kinetics and so reporting a combined total hydrolysis rate for a UASB is an oversimplification. A number of alternative models are also commonly used including an additional empirical model with no mechanistic basis (Contois) and one based on theoretical consideration of the action of a single enzyme (Michaelis-Menten) (Table 3.1) (Wang & Li, 2014). The Contois model describes an inverse relationship between the microbial concentration and the limiting substrate concentration to the specific growth rate. Whereas, the Michaelis-Menten model envisages an enzyme reversibly binding to a substrate which is subsequently degraded. The difference between the two approaches can be framed with the question does instantaneous consumption depend only on substrate availability (Michaelis-Menten) or on microbial abundance as well (Contois)(Wang & Li, 2014). For both models, the product of the hydrolysis rate constant and biomass concentration is the maximum rate (V_{max}) for a given biomass concentration.

Both models were developed for soluble substrates (Tomei et al., 2008) but have frequently been successfully applied to insoluble particles for a variety of feed wastes such as municipal solid waste treatment (Krzystek et al., 2001), dairy manure digestion (Bhttacharya & Khai, 1987) and primary sludge (Lin, 1991). Both models simplify to a first order expression under specific limiting conditions. In the case of the Contois model this occurs when the biomass is sufficiently established and for the Michaelis-Menten model when K_m is sufficiently larger than the substrate concentration (Vavilin et al., 2008).

Table 3.1: Prominent hydrolysis rate models ([B] is biomass concentration, K_c is the Contois Constant and K_m is the Michaelis constant)

Name	Expression $\left(\frac{d[S]}{dt}=\right)$	Reference
First Order	$-k_h[S]$	(Eastman & Ferguson, 1981)
Contois	$-k_h[B]\frac{[S]}{K_c[B]+[S]}$	(Contois, 1959)
Michaelis-Menten	$-k_h[B]\frac{[S]}{K_m+[S]}$	(Monod, 1949)

The current work aims to understand the underrepresented impact of low temperatures on anaerobic hydrolysis of primary sludge in passive systems for temperatures below 20^{o}C. To do so a series of batch hydrolysis tests were conducted over the temperature range 5- 37^{o}C and the hydrolysis rate constant determined. This represents the first data reported on hydrolysis rate constants across a range of temperature below 20^{o}C. The work then modelled the hydrolysis rate against commonly used models to understand which was the most applicable and hence elucidate critical features related to passive hydrolysis.

3.2 Methodology

3.2.1 Hydrolysis Rate Tests

Hydrolysis rate tests were conducted in 1 l bottles, filled with 900 ml substrate and 100 ml headspace. The substate of the tests was designed to mimic septic tank conditions, with a recommended seed level of 7% (Cruddas et al., 2018) and 14% sludge level, approximately half the accumulated sludge volume for a septic tank desludged at 33%. Tests were seeded with 60 ml psychrophilic seed sludge, 120 ml primary sludge and 720 ml raw sewage. All tests were flushed with nitrogen gas to ensure anaerobic conditions. Two different sources of psychrophilic seed sludge were used during the investigation due to local availability across the duration of the work. For the experiments at 5, 15 and 37^{o}C, seed sludge for taken from a UASB operating at ambient temperatures. The UASB was a 70 l reactor with a flocculent sludge which had been operating at low temperatures for more than four years (Ribera-Pi et al., 2020). The remaining experiments had a seed sludge which was collected from an upflow region of a baffled septic tank (EST) operating at ambient temperatures (Chapter 2). Both reactor systems were operating from the same feed stock and equivalent temperatures.

To account for variability in biological growth batch tests were conducted in triplicate. The temperature of the tests was maintained with a water bath to an accuracy of ±0.1°C. Mixing tests were conducted with an overhead mixer, with the mixing intensified varied based on the mixer rotational speed recorded in rotations per minute (rpm). The rotational speed of the mixer was determined based on the mean hydraulic velocity for a hydraulically and well mechanically mixed system, approximately 6 s^{-1} and 80 s^{-1} respectively (Arnau et al., 2022). The operational conditions for each test is outlined in Table 3.2. Tests were sealed with a rubber bung. Sampling was conducted using a sampling tube from which passed through the bung to the mid-point of the 1 l bottles, the sampling tube was clip sealed when not in use. The mixer shaft for mixed tests was sealed using silicone grease. The UASB seeded tests were run for up to 90 days or until degradation plateaued. The EST seeded systems were run for 30 days as limited degradation beyond this point was observed in UASB seeded systems.

Table 3.2: Anaerobic degradation test conditions

Temperature (°C)	Seed	Mixing
5	UASB	None
7	Septic Tank	None
10	Septic Tank	None, 35 rpm and 85 rpm
15	UASB	None
20	Septic Tank	None
37	UASB	None

3.2.2 Analytical Procedures

For the initial 28 days of each batch test sampling was conducted 3 times a week, after 28 days tests were sampled weekly. Samples were collected of the mixed liquor at each sampling point. The mixed liquor was tested for; TSS, VSS, TCOD, sCOD and pH using standard methods (APHA, 2005). The pCOD was calculated by subtracting the sCOD of a sample from the TCOD. The three batches acted as triplicates for each experimental point. Unmixed tests were inverted twice prior to sampling to ensure a homogenous sample. Samples from mixed tests were assumed to be well mixed. Particle size distributions (PSD) were calculated using laser diffraction particle sizing (Mastersizer 3000, Malvern, UK)

3.2.3 Lipase Activity Assays

Lipase activity was measured as there is a well established methodology present in literature (Gessesse et al., 2003, Kim et al., 2012, Petropoulous et al., 2018) and this is often considered the limiting enzymatic stage of hydrolysis (Petropoulous et al., 2020) making it a viable proxy for enzymatic activity. The substrate for lipase activity was p-nitrophenol palmitate (pNPP), which results in a 1 mole release of p-nitrophenol dye for each mole of palmitate degraded (Petropoulos et al., 2018). A stock solution of pNPP was created by dissolving pNPP in isopropanol to a concentration of 20mM, which was stored at -20^{o}C. The buffer solution for the assay was 20mM Tris HCl buffer pH 8.0 (Sigma Aldrich, UK) containing 0.1%$_{(m/v)}$ Gum Arabic and 0.4%$_{(m/v)}$ Triton X-100 (Sigma Aldrich, UK) (Kim et al., 2012; Petropoulos et al., 2018).

Lipase activity was measured from samples from the batch tests, based on the method outlined by Hassard et al., (2018). Each assay was conducted in 1 ml microwells. The biomass sample was diluted in a 1:3 (v/v) ratio with the Tris Buffer solution. The assay consisted of 0.5 ml biomass sample and 0.5 ml substrate at varying of 250 µM, 200 µM, 150 µM, 100 µM, 50 µM and 25 µM. Each assay was run for 2 hours measuring the absorbance of p-nitrophenol (λ_{max}=410nm) every 30 seconds in a microplate reader (M2000 infinite pro, Tecan, Austria). Blanks of the biomass without substrate, substrate without biomass and deionised water (DI) were measured to account for the absorbance of the biomass and substrate and ensure there was no contamination. The absorbance was converted to molar concentration by subtracting the absorbance of the blanks and multiplying the result by the sample dilution factor and molar absorbance of p-nitrophenol. The molar absorbance of p-nitrophenol was calculated measuring the absorbance of the dye in DI at 8 concentrations from 10 to 500 µM, using linear regression the molar absorbance was calculated and calibrated using Beer's Law.

The kinetic constants of Michaelis-Menten (Table 3.1) were calculated from this data by assessing the initial gradient at each concentration assay to give the initial rate velocity (Hassard et al., 2018). The initial rate velocities at each concentration were fit to the Michaelis-Menten model using non-linear least squares fitting (Hassard et al., 2018), to give the V_{max} and K_m. The specific enzymatic activity is given by dividing the V_{max} by the VSS concentration of the sample (Hassard et al., 2018).

3.2.4 Kinetic Parameter Modelling

The first order hydrolysis rate constant was calculated for all conditions based on both pCOD and VSS degradation over time. Non-linear regression of the integrated form of the first order kinetics were fitted to the experimental data, using the initial concentration and hydrolysis rate constant as variables. The data points were calculated to 95% confidence using non-linear least squares.

Non-first order equations were analysed as a series of ordinary differential equations for the pCOD degradation and biomass concentration. The biomass for non-first equations (Table 3.1), was calculated based on the initial VSS and the biomass concentration over time was evaluated based on substrate degradation (Tomei et al., 2008; Vavilin et al., 1996);

$$\frac{d[B]}{dt} = Y\frac{d[S]}{dt} - b[B]$$

Where Y is the growth yield coefficient and b is the endogenous respiration rate. Y and b were assumed to be constant at 0.1 and 0.19 d^{-1} (Tomei et al., 2008). The equations were solved by numerical integration and mean squares fitting using Excel Slover (Microsoft). The hydrolysis rate constant, initial substrate concentration and kinetic constant were set as variables for the Contois and Michaelis-Menten models. The 95% confidence of the hydrolysis rate constants was calculated from the least squares fit. The fit of each model was assessed based on the R^2 fit compared to the experimental data points.

3.3 Results and Discussion

3.3.1 Impact of Temperature and Seed Type

For the UASB seeded system, the overall degradation of pCOD over the experimental timeframe was 34, 31 and 52% at temperatures of 5, 15 and 35°C respectively (Figure 3.1). The majority of the degradation occurred with the first 21 days with no observed change after 40 days indicating a linear rate followed by a plateau region which would be consistent with basis of the Contois and Michaelis-Menten models. However, analysis of the model fits revealed similarity for the first order and Contois models with R^2 of 0.65 and 0.64 at 5°C, 0.92 and 0.92 at 15°C and 0.81 and 0.85 at 37°C respectively. In contrast, the fit to the Michaelis-Menten model was poorer with R^2 values of 0.34, 0.42 and 0.69 at 5, 15 and 37°C respectively.

The subsequent experiments at 7, 10 and 20°C were run using the seed from the EST and were run for the 21 days over which the previous trials had shown the majority of the degradation. Across the 21 days the overall degradation of pCOD was 19, 18 and 26% for the 7, 10 and 20°C experiments (Figure 3.2). Comparison to the previous test revealed higher initial pCOD and lower overall levels of degradation indicating the impact of the seed material. Visual inspection often data indicated a closer fit to the models in general and a consistent linear nature over the initial 21 days. Analysis of the quality of model fit mirrored the observation for the tests with the UASB seed in that the better fits were observed with the first order and Contois models which were very similar and better than the fit to the Michaelis-Menten model. This was most apparent at 20°C where the R^2 fit of the models were 0.81, 0.81 and 0.48 for the first order, Contois and Michaelis-Menten models respectively. The consistent appropriateness of fit to the first order and Contois models with both seeds reflects the limiting conditions of the passive batch test where the active layer has an excess of hydrolytic micro-organisms and available enzyme due to mass transfer limitations. Hence the hydrolysis is limited due to the disintegration rather than solubilisation.

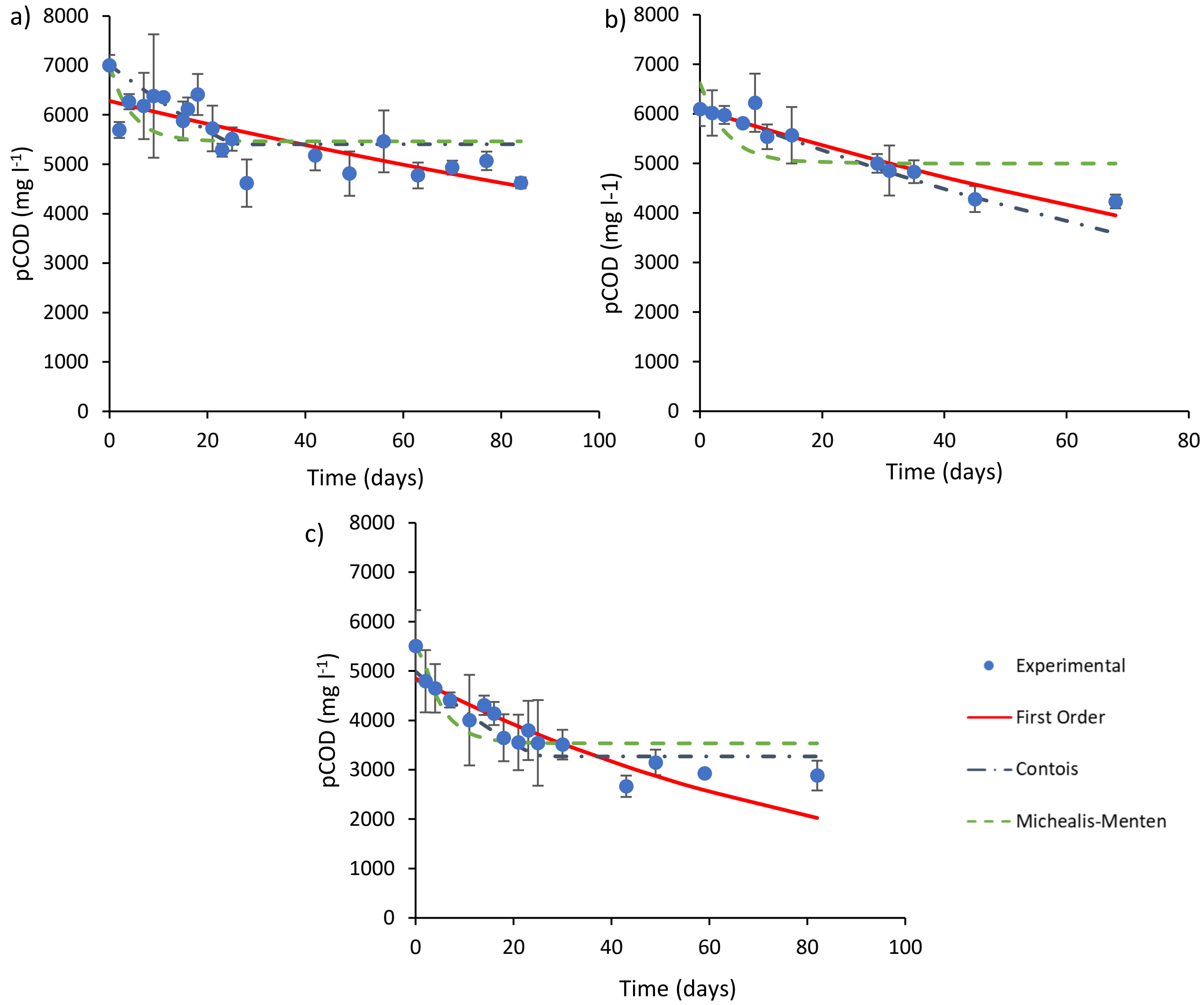

Figure 3.1: Experimental degradation in pCOD over time for unmixed UASB seeded systems at a) 5°C b) 15°C and c) 37°C and projected degradation from the first order, Contois and Michaelis-Menten models of hydrolysis

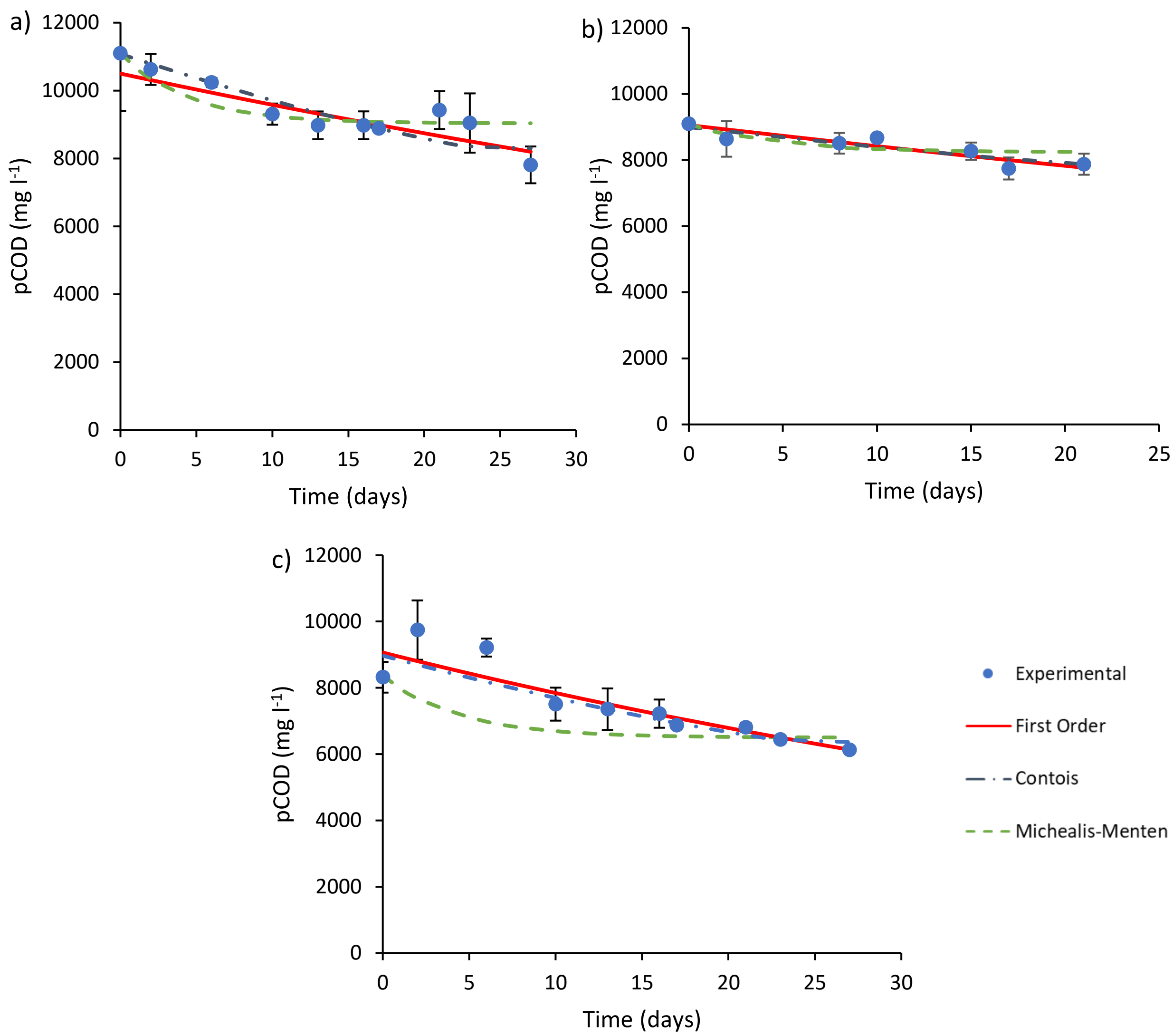

Figure 3.2: Experimental degradation in pCOD over time for unmixed EST seeded systems at a) 7ºC b) 10ºC and c) 20ºC and projected degradation from the first order, Contois and Michaelis-Menten models of hydrolysis

Further, the observation of the Contois model collapses to the first order model is in line with the expectation outlined in Vavilin et al., (2001) and indicates that the seeds were sufficiently stabilised. The fact that the Michaelis-Menten model did not fit well is in contrast to previous findings on a well-mixed system with secondary sludge at 37ºC and indicates the impact of mixing switching the controlling influences to both feed and biomass concentration. Interestingly, Tomei et al., (2008) found that sonication of the feed sludge did not significantly increase the rate implying that solubilisation rather than disintegration was the limiting factor of hydrolysis.

3.3.2 Impact of Mechanical Shear

The degradation of pCOD observed in the unmixed systems was compared to those with two different levels of mixing depicted by the rotational speed of the impeller (Figure 3.3). The total amount of pCOD degraded after 21 days of the batch test were 1219 mg l^{-1}, 2650 mg l^{-1} and 400 mg l^{-1} for the passive, 35 rpm and 85 rpm experiments respectively (Figures 3.2, 3.3). This suggests that inclusion of mixing is beneficial but can be excessive and results in inhibition of the microbial process. This is consistent with the fact that in the start-up of mesophilic anaerobic digesters, it has been found that reduced or intermittent mixing aids biological start-up (Hoffmann et al., 2008; McMahon et al., 2001) due to the sensitivity of methanogens (Lindmark et al., 2014). Accordingly, we posit that at low temperatures similar impact can occur with hydrolytic microorganisms through either direct damage or impact on the released enzymes (Lindmark et al., 2014). The regeneration time for hydrolytic microorganisms is generally considered slow (Lindmark et al., 2014) and will be further retarded by the low temperatures reinforcing the idea that whilst mixing is beneficial it must be carefully applied.

Previous analysis of the shear patterns and the distribution of low velocity gradients in such jars reveals a distribution of local velocity gradients and the associated time the water resides in the different shear zones (Bridgeman et al, 2008). Direct comparison between different systems is notoriously challenging but the system used in this paper is a direct comparison to the previous work. The previous work reported that at 30 rpm, the fluids experiences local shear rates of up to 150 s^{-1} but for less than 0.1 seconds and 100 s^{-1} for around 8 seconds. Whereas at 100 rpm (the closet to 85 rpm from the previous work), the fluid experiences local shear rates as high as 800 s^{-1} and spends 20 seconds at local rates above 250 s^{-1} (Bridgeman et al, 2008) indicating the difference in shear that the biology will experience as the rpm is increased.

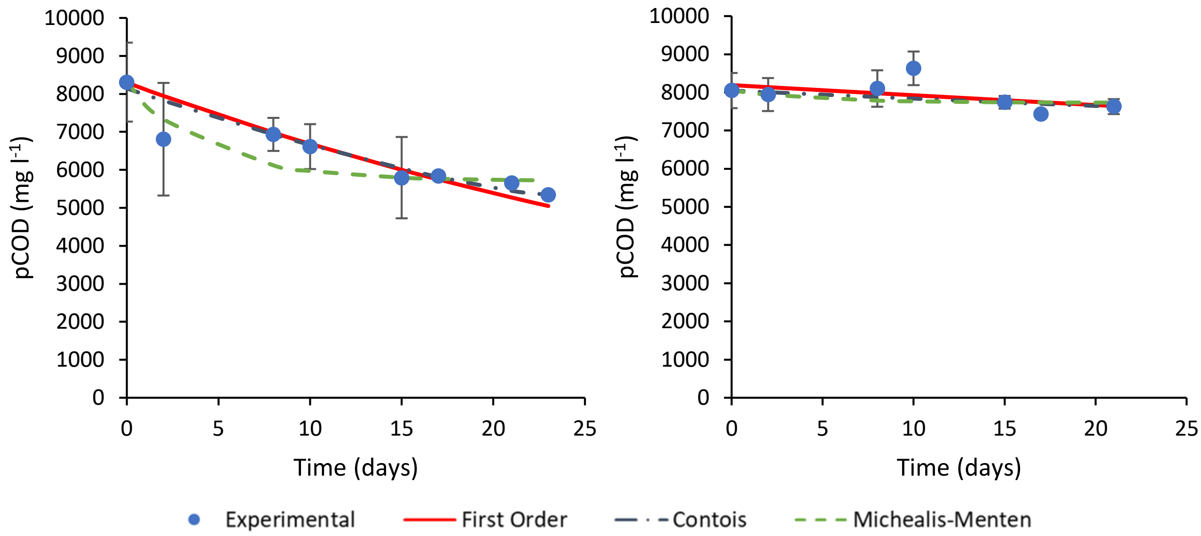

Figure 3.3: Experimental degradation in pCOD over time for EST seeded systems at 10ºC with a mixer speed of a) 35 rpm b) 85 rpm and projected degradation from the first order, Contois and Michaelis-Menten models of hydrolysis

3.3.3 Particle Size Distribution and Lipase Activity

To further explore the impact of temperature on hydrolysis, the particle size distributions were analysed at the start and end of the 5, 15 and 37ºC tests using the UASB seed (Figure 3.4). The impact of hydrolysis was to reduce the particle size of the solids in the feed samples which exhibited a bimodal profile with peaks at 95 and 600 µm. Post the hydrolysis tests, the profile became appreciably more mono modal with mode particle sizes of 80, 95 and 100 µm for 5, 15 and 37ºC respectively. In addition, the final particle size distributions shifted to smaller particles sizes the lower the temperature of the trial indicating the formation of more small particles. This is consistent with the work of Lew et al., (2009) who observed at 15 and 10ºC, hydrolysis of wastewater particle initially breaking down into colloidal particles (particle size <4.5 µm) and then being further degraded, compared to a 20ºC where a more significant fraction of wastewater particulates was degraded directly to soluble components. The current study used laser diffraction to measure the PSD and the inherent bias the instrument has towards larger particles masks the ability to see such changes in detail. However, the volume density at a particle size of 4.5 µm increased from 0.19% to 0.37% at 5°C compared to a decrease to 0.14% and 0.16% at 15°C and 37°C respectively. Overall, this supports the suggestion of a slowdown in degradation as the temperature decreases.

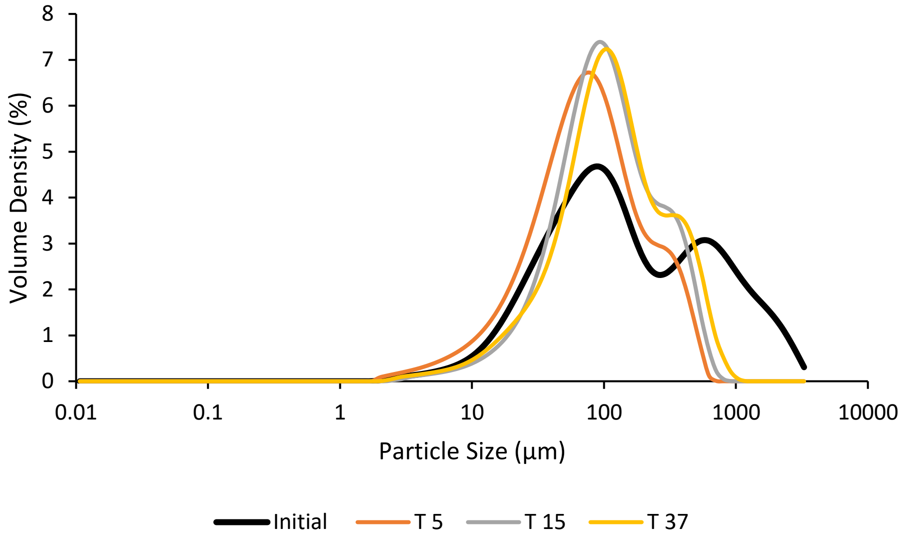

Figure 3.4: PSD of the tests at the initial condition and the end of experiments for the UASB seeded tests.

Lipase activity was used to monitor enzymatic processes in the UASB sludges and revealed that under low temperatures that the specific lipase activity decreased from the start of the test at 0.174 µM s^{-1} mg VSS^{-1} to 0.152 µM s^{-1} mg VSS^{-1} after 56 days. In contrast, at 37°C the specific lipase activity increased from 0.141 µM s^{-1} mg VSS^{-1} at the start to 0.252 µM s^{-1} mg VSS^{-1} after 30 days and to 0.344 µM s^{-1} mg VSS^{-1} after 82 days. The reduction in enzymatic activity and the increased production of small particles confirms the impacts of temperature. The reduced enzymatic activity will likely have less impact in unmixed system as they are in excess but would otherwise indicate a slowdown in hydrolysis is partially due to the impact on the enzymatic activity levels.

3.3.4 Hydrolysis Rate Constant

Conversion of the degradation data was based on the first order rate model due to the previous assessment of its appropriateness. The hydrolysis rate constant decreased with temperature with the UASB seed from 0.0143 ±0.0024 d^{-1} at

37^{o}C to 0.0063 ±0.0013 d^{-1} at 15^{o}C and 0.0038 ±0.0011 d^{-1} at 5^{o}C (Figure 3.5). The variation in seed sludge also appears to have had a notable impact on the hydrolysis rate constant with generally higher rate constants determined for the EST seed sludge tests. To illustrate, at 7^{o}C the first order rate constant was 0.0092 ±0.0030 d^{-1} which was higher than that observed for the UASB seed sludge at 15^{o}C. The highest rate constant observed in the study was at 10^{o}C with a mixer speed of 35 rpm, giving a hydrolysis rate constant of 0.0216 ±0.0063 d^{-1} which was higher than the unmixed at 35^{o}C. This suggests that mixing may be able to enhance the hydrolysis rate to compensate for reduction in temperatures. Support is provided for this notion from pilot trials of septic tanks with enhanced mixing developed by alternative baffle arrangement (Chapter 4). The most effective system was based on the front end of the EST which have two vertical baffles and resulted in a hydrolysis rate constant of 0.027 d^{-1} at a temperature of 14^{o}C. This compares to the passive batch tests with a rate constant of 0.0064 d^{-1} at 15^{o}C and 0.006 d^{-1} at 14^{o}C in a poorly mixed septic tank (Table 3.3).

Comparison of the current findings with the existing available data in the literature confirms the general point of a decrease in hydrolysis rate constant with temperature (Table 3.3). To establish if change in the rate constant is just directly a temperature impact, all rate constants were adjusted to 10^{o}C by using an Arrhenius relationship (Table 3.3). Temperature adjusted rate constants for the UASB trials were 0.0056, 0.0045 and 0.0022 d^{-1} for temperatures of 5, 15 and 37^{o}C. This compares to unadjusted values of 0.0038, 0.006 and 0.014 d^{-1} respectively and indicates the difference cannot be accounted for by temperature alone. A similar finding occurs in the well mixed trials from Ferreiro and Soto (2003) where the temperature adjusted hydrolysis rate constant is lower at 35 ^{o}C than at 10^{o}C or 20^{o}C (Table 3.3). Whereas in the other unmixed trail the temperature adjusted rate constant was approximately fixed at between 0.02 and 0.024 d^{-1} for temperatures between 15 and 35^{o}C (Mahmoud et al., 2004). Direct comparison of the absolute value shows no agreement congruent with differences in the seed material and feed make up. However, the ratio of the rate constants at the highest to lowest temperatures tested were similar at

3.7 in the current trial and 3.9 (Mahmoud et al., 2004) and 4.41 (Ferreiro and Soto, 2003). Overall, this indicates that temperature is not the controlling aspect to determine hydrolysis rate and that either appropriate selection of seed or mixing can have genuine impact on the ability to enhance hydrolysis in currently passive system operated at low temperature.

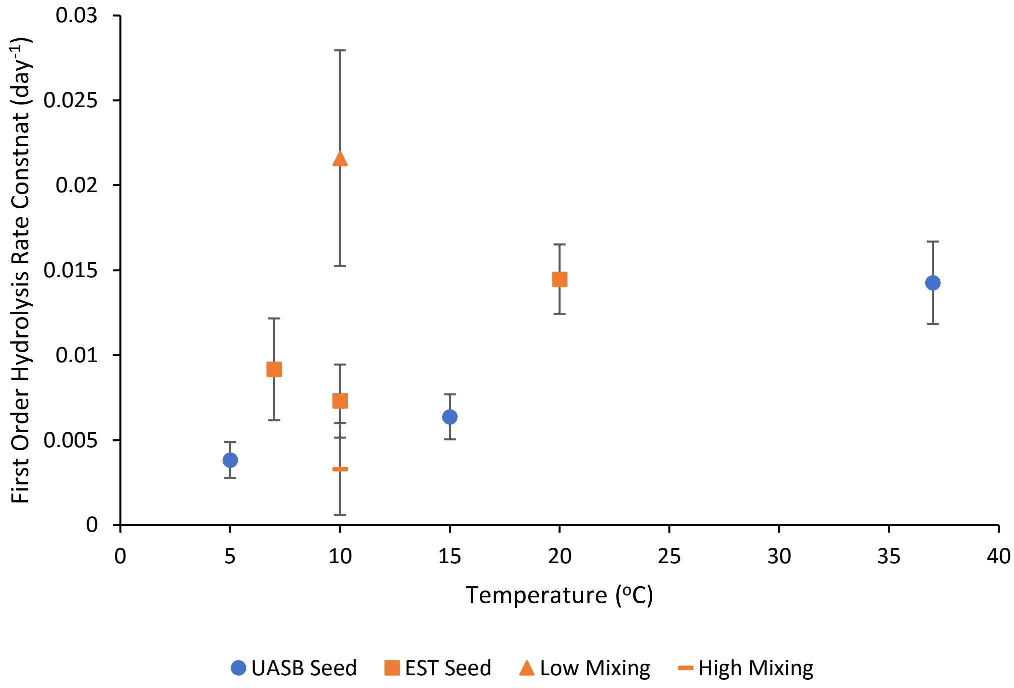

Figure 3.5: First order hydrolysis rate constants calculated for each test condition

Table 3.3: Reported first order hydrolysis rate constants of wastewater and sludge, and normalised hydrolysis rate constants to 10^{o}C

Temperature (^{o}C)	Hydrolysis Rate Constant (d^{-1})	Temperature adjusted hydrolysis rate (d^{-1})^{a}	Mixer Speed (rpm)	Reference
5	0.0038	0.0056	0	This Study
7	0.0092	0.012	0	This Study
10	0.0073	0.0073	0	This Study
10	0.022	0.022	35	This Study
10	0.0033	0.0033	85	This Study
10	0.0383	0.038	200	Ferreiro and Soto (2003)
14	0.006	0.0045	Low Hydraulic Mixing	Chapter 4
14	0.027	0.020	High Hydraulic Mixing	Chapter 4
15	0.0089	0.0061	Unmixed	Chapter 4
15	0.0064	0.0044	0	This Study
15	0.0289	0.020	-	Mahmoud et al. (2004)
18	0.035	0.020	Moderate Hydraulic Mixing	Chapter 4
20	0.014	0.0067	0	This Study
20	0.0953	0.046	200	Ferreiro and Soto (2003)
20	0.12	0.058	-	Elmitwalli et al (2013; 2001)
20	0.11	0.053	-	(Moser-Engeler et al., 1999)
21	0.03	0.013	80	(Ji et al., 2010)
25	0.07	0.024	-	Mahmoud et al. (2004)
25	0.004	0.0014	-	(Kassab et al., 2013)
25	0.006	0.0020	-	(Kassab et al., 2013)
35	0.113	0.020	-	Mahmoud et al. (2004)
35	0.169	0.030	200	Ferreiro and Soto (2003)
35	0.035	0.0062	250	Arnaiz et al., (2006)
37	0.014	0.0022	0	This Study

A – Based on Arrhenius equation calculated in (Mahmoud et al., 2004)

3.4 Conclusion

This study has established the first dataset concerning hydrolysis of sludge in a passive system akin to a septic tank below 20ºC. The experimental data fit to a first order kinetic model indicating that digestion of solids is limited by the disintegration of solids opposed to biological establishment, even in low temperatures. This is further highlighted by the observation of the Contois model approximating first order kinetics in most tests, showing the biomass is sufficiently established and not rate limiting. Importantly, the work revealed that mixing can enhance the hydrolysis rate and more than adjust for the impact of temperature. Further, until mixed, the addition of enzymatic supplements are unlikely to enhance the process as they are likely to already be in excess. Therefore, the impact of this study is that optimisation of solids destruction in passive anaerobic reactors should focus on the optimisation of the disintegration through physical processes opposed to biological establishment. The key is to optimise mass transfer to enhance hydrolysis without using excessive energy input which can inhibit hydrolysis.

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4. Enhancing the Hydrolysis Rate Within Septic Tanks Through Baffling to Extend Desludging Frequency

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Abstract

Septic tanks represent a significant sector of global wastewater treatment, however, require regular maintenance in the form of desludging. The current primary focus of study for low-rate anaerobic reactors is methane generation rather than hydrolysis which determines the desludging frequency. Three configurations of baffled enhanced septic tanks (EST) were tested to optimise the hydrolysis within septic tanks using baffles to encourage hydraulic mixing. The 3.6 m^3 ESTs were operated for 3 months with a hydraulic residence time (HRT) of 81.6 h. The treatment of the ESTs was equivalent to conventional septic tanks. The hydrolysis rate constant increased from an average of 0.0089 d^{-1} in a conventional septic tank to 0.035 d^{-1} in an EST by baffle optimisation. Increasing the hydrolysis rate constant increases the predicted desludging frequency of the ESTs from 4.9 to 6.7 years, allowing the ESTs to be maintained in tandem with natural-based solutions used for secondary treatment. The impact of this work has the potential to reduce the maintenance of septic tanks, lowering their carbon footprint, operational costs and social impact.

Keywords: septic tanks, hydrolysis, desludging

4.1. Introduction

Decentralised wastewater treatment systems (DEWATS) serve small rural and isolated communities using technologies such as septic tanks, pit latrines, anaerobic baffled reactors (ABR) and anaerobic ponds (AP) to provide sludge retention, biogas generation and treatment of the wastewater. DEWATS account for 45% of wastewater treatment globally of which approximately 50% is delivered by septic tanks (UN Habitat & WHO, 2021). For instance, Scottish Water alone operates over 1,200 septic tank sites ranging in population served from 5 to 1,000 PE with an estimated additional 160,000 households operating private septic tank systems across Scotland (O'Keeffe et al., 2015).

The traditional design of septic tanks is optimised for settling with the fluid flow through designed to minimise the disturbance to the sludge (Almomani, 2016), forming a continuous liquid phase and a fed-batch solids phase. To aid with settling, vertical baffles are used to divide the septic tank into 2 to 4 chambers. Septic tanks, like pit latrines, APs and ABRs, are low-rate anaerobic reactors enabling a degree of solids degradation through hydrolysis reactions that reduces the rate at which solids accumulate. The net solids accumulation rate (SAR) is a combination of the incoming solids loading rate, settling efficiency and the reduction in solids due to hydrolysis reactions. Literature on the sludge accumulation rate in septic tanks is limited but a number of empirical models are available (Bounds, 1997; Mahon et al., 2022). These models based the sludge accumulation rate based on surveys of operational tanks and their time between desludging. Theoretical models (e.g. Elmitwalli, 2013) have also been developed assuming a first order reaction rate for the solids destruction by hydrolysis;

$$r_h = \frac{d[S]}{dt} = k_h[S] \qquad \text{(Eastman \& Ferguson, 1981)}$$

Where [S] is the substrate concentration (mg l^{-1}), t is time (d) and k_h is the hydrolysis rate constant (d^{-1}). The hydrolysis rate alters as a function of temperature, pH (Gomec et al., 2002; Veeken et al., 2000), particle size (Aldin et al., 2011)and mixing (Halalsheh et al., 2011). For instance, k_h was reported to increase from 0.08 d^{-1} at 5^{o}C to 0.23 d^{-1} at 30^{o}C in a mixed systems treating concentrated domestic sewage (Elmitwalli et al., 2003). Reported rates in unmixed systems, akin to operation of DEWATS, reveals a lower k_h of 0.004 d^{-1} at 5^{o}C and 0.016 d^{-1} at 20^{o}C (Brown et al, 2021). No study was found analysing k_h from operational septic tanks however a rate constant of 0.016 d^{-1} was calculated from reported data of a system at 15^{o}C (Lin et al., 2019).

An unintended by-product of using anaerobic systems for DEWATS is the generation of methane, if not captured this can be a major greenhouse gas emission, a recent study of blackwater treatment with septic tanks in Vietnam revealed an average methane emission of 11.92 g $person^{-1}$ d^{-1} from the first chamber (Huynh et al., 2021). The majority of research in DEWATS is on the potential for methane generation, to illustrate, in the case of ABRs, a four chamber 900 l ABR operated at 18^{o}C generated an energy production of 0.45 kWh m^{-3} wastewater treated (Hahn & Figueroa, 2015; Pfluger et al., 2020). Similarly, a pilot AP study using 370 l reactors reported gas production rate of 5.4 l CH_4 m^{-3} wastewater treated at an average operational temperature of 10.5^{o}C(Cruddas et al., 2021). However, methane generation is not a primary function of DEWATS. The main function of DEWATS is sludge storage and wastewater treatment yet there is sparse literature on impact of DEWATS design on sludge degradation (Reynaud & Buckley, 2016). The primary maintenance requirement of septic tanks is desludging; therefore, reactor optimisation and design should consider the implications on maintenance of the systems. This study aims to examine the impact reactor design on desludging, given the importance for DEWATS.

Current UK regulations recommends annual desludging (HM Government, 2015) although actual rates can vary significantly, with numerous reports of unmaintained systems (Mahon et al., 2022). A recent survey of Scottish water on operated septic tank desludging revealed an average frequency of 6 months with <1% systems desludged between 12 and 24 months The process involves tanker visits to site and so is associated with costs, carbon footprint and social impacts. Consideration about the future of DEWATS within Scottish Water has led to the aim of upgrading existing septic tank design and coupling the systems with a nature-based solution (NBS). Such a flowsheet has the potential to provide a lower maintenance and more environmentally appropriate approach through increasing biodiversity and social amenity values. The concept aims to reduce the impact of desludging by extending the sludge storage to 5-7 years to coincide with the need to refurbish the NBS system (Dotro et al., 2012).

Desludging frequency is based on the maximum volume occupied by sludge and hence is a function of the overall reactor volume, the sludge fill level (between 30-50%) and the solids accumulation rate. Previous research has shown an enhancement in SAR from 0.77 cm d^{-1} to 0.24 cm d^{-1} in a septic tank trial through increasing sludge mixing by adjusting the inlet position to provide mixing of the sludge layer (Almomani, 2016). We posit in the current work that alteration of the baffling configuration can enable enhanced hydrolysis rates without a negative impact on the overall treatment. This is based on the concept of using a combination of vertical and horizontal baffles to develop an enhanced septic tank (EST) design that is made up of two functional sections, hydrolysis and settling, both of which are optimised by local fluid velocity profiles (Cruddas et al., 2018). However, each section benefits from very different profiles. In the hydrolysis section, high mixing velocities are preferred to ensure good distribution and mass transfer occur in order to ensure the microbiological activity is not mass transfer rate limited (Ma et al., 2019). In contrast, in the settling section, low velocity profiles are required without areas of high local velocity to both aid settling and prevent breakage of the biological aggregates (Ma et al., 2019).

The aim of the current work is to understand the potential of enhancing the hydrolysis rate in septic tanks through utilisation of a combination of vertical and horizontal baffles without negatively impacting treatment performance. This is utilised to establish the plausibility of a case for developing an EST design to reduce desludging frequencies of between 5-7 years. To achieve this a combination of computational fluid dynamics (CFD) and experimental trials was conducted on a series of baffled septic tanks treating municipal sewage to measure hydrolysis rates and removal performance. The measurement of gas generation was not feasible due to the requirement of regular sludge sampling. This work provides a novel insight into reactor designs impact on hydrolysis and removal.

4.2. **Methodology**

4.2.1. Experimental Set-up

Two pilot scale glass reinforced plastic reactors were used with internal dimensions of 3000 x 700 x 2000 mm (L x W x H), with a liquid height of 1700 mm and a working volume of 3.57 m^3 (Figure 4.1a). Both reactors were constructed with 7 moulded channels for the placement of baffles. The baffles were constructed from 20 mm thick polypropylene. Vertical baffles (Figure 4.1b) had a channel height of 670 mm and acted as either a upcomer (U) or downcomer (D) baffle with the channel being at the top or bottom of the reactor, respectively. The horizontal baffles (Figure 4.1c) had a channel width 15% of the total reactor, equal to 130mm (Cruddas et al., 2014), and were orientated to have the channel on the left (H_L) or right (H_R) side of the reactor.

Each EST consists of a hydrolysis section, which spans from the inlet to the first horizontal baffle, and a subsequent settling section. The settling region for each EST comprises of two horizontal baffles and aims to provide a region of uniform flow in which the solids that leave the hydrolysis region settle out of the liquid phase. The test reactors are labelled based on the target mixing intensity in the hydrolysis region, from low to high. Three configurations of baffles were utilised, which are schematised in Figure 4.2. EST-Low had only a upcomer baffle positioned 270 mm from the inlet, slot 3, and so generated the lowest design mixing intensity with a superficial upflow velocity of 0.37 m h^{-1} (Figure 4.2a). EST-Mid had both a upcomer and downcomer baffle to encourage two sludge mixing zones (Figure 4.2b). A narrower initial channel (200 mm, slot 2) increased the superficial upflow velocity to 0.5 m h^{-1} and the downcomer baffle was positioned 270 mm from the inlet, slot 3. EST-High has the same superficial upflow velocity as EST-Mid however the first horizontal baffle is positioned as a vertical baffle to increase the mixing in the second sludge mixing region (Figure 4.2c).

For EST-Low and EST-Mid the horizontal baffles were placed equidistantly in the reactor, in slots 5 and 7, as is recommended for anaerobic ponds (Vega et al., 2003). For EST-High the first horizontal baffle was positioned to increase mixing, in slot 4, and the second horizontal baffle was placed equidistant in the remaining volume, in slot 6. All EST configurations were bottom fed from the lateral mid-point, the outlet was positioned 590 mm from the left side of the reactor 343 mm from the top of the reactor, the outlets both had an external scum guard and overflow giving a total liquid height of 1700 mm.

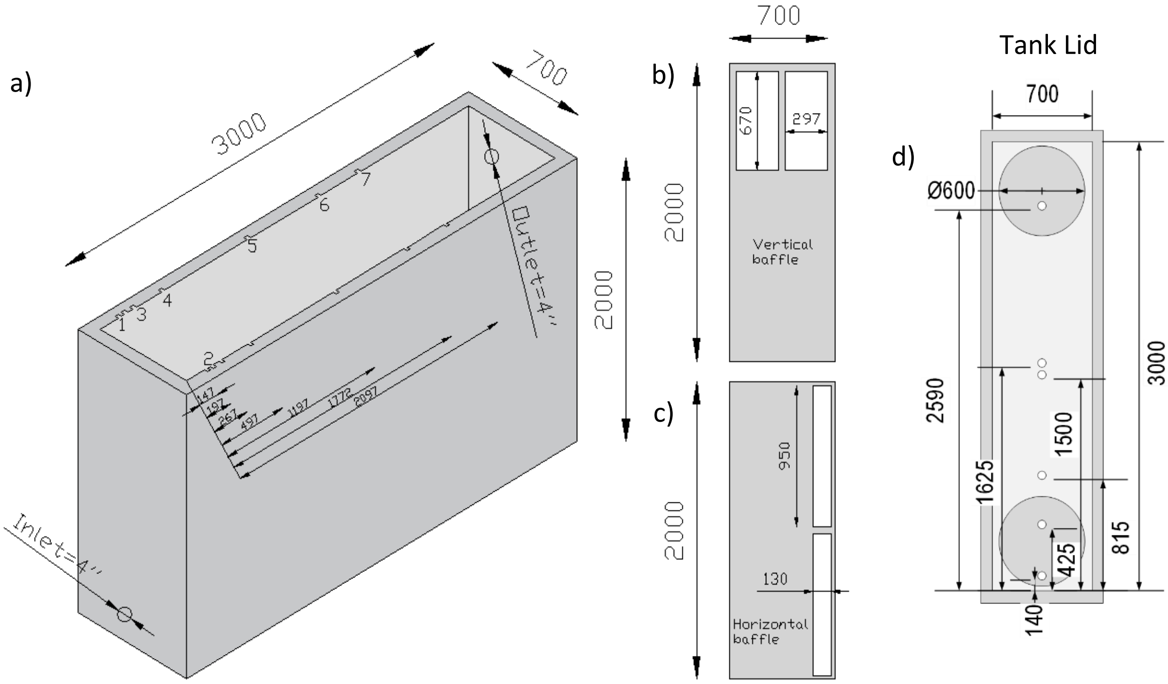

Figure 4.1: Diagram of the reactors a), vertical b) and horizontal c) baffles and tank lid d), all measurements in mm

The three EST configurations were run seperately alongside a control system with traditional septic tank style baffling. With EST-low and its control running from March to June 2021, EST-Mid and its control running from July to October 2020, EST-High and its control running from January to April 2022. The septic tank control was designed as a three-chamber system, with vertical baffles in slots 5 and 6 of the reactor (Figure 4.2a). The inlet of the control was positioned at a height of 1500 mm and the lateral mid-point, the outlet was the same as the EST reactors.

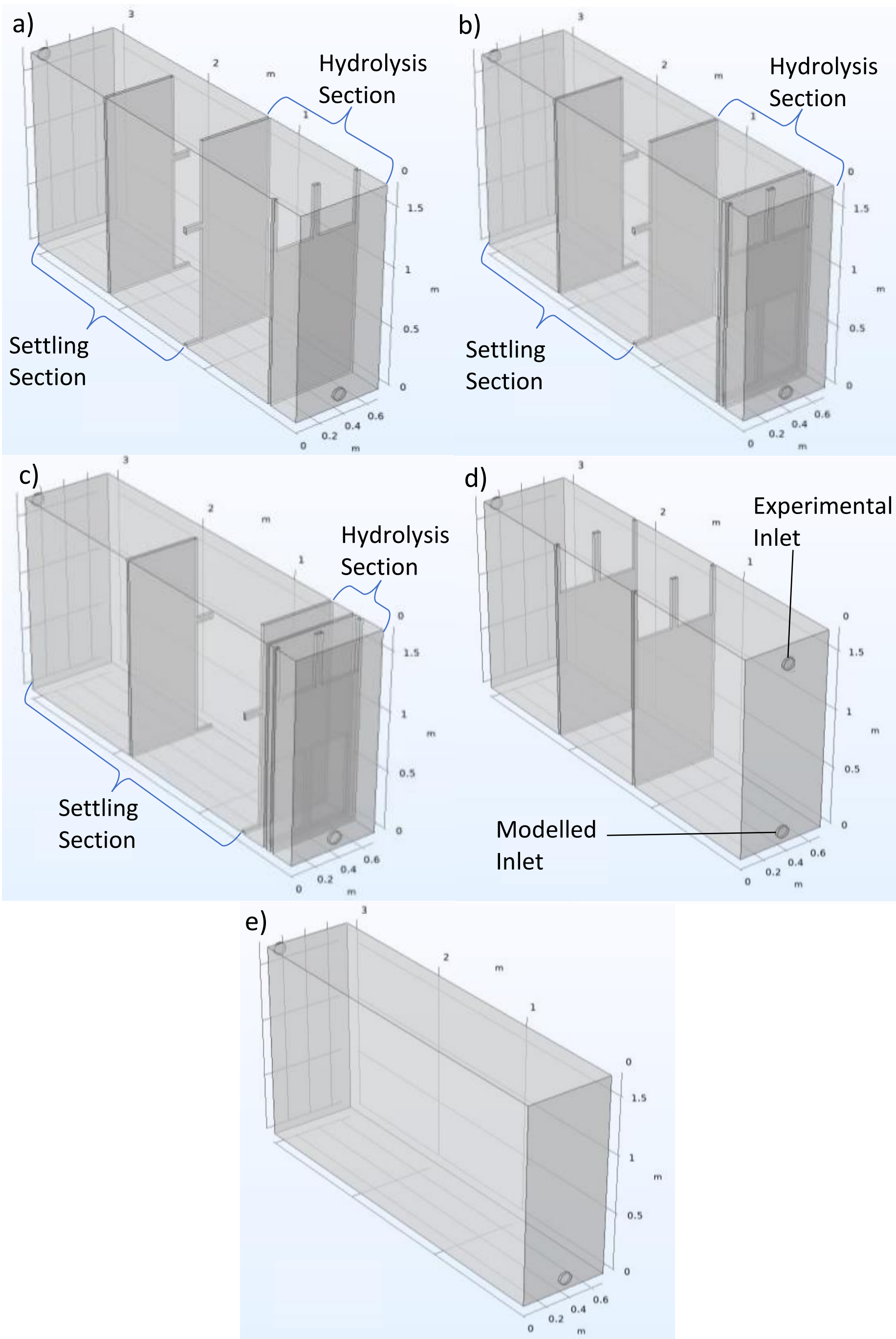

Figure 4.2: Baffled configuration diagram of experimental and modelled systems a) EST-Low, b) ESP-Mid, c) ESP-High, d) Control e) Unbaffled

The lid of both reactors was fitted with a rubber gasket and sealed using silicone sealant. There were six, 3-inch diameter sampling points on the lids of the reactors (Figure 4.1d) with ball valves to close when not in use. The sampling points were located respectively at 140 mm, 425 mm, 815 mm, 1500 mm, 1625 mm, 2590 mm from the inlet (Figure 4.2d). Prior to testing each reactor was filled with 250 l of septage and then fed wastewater at a rate of 0.51 m^3 d^{-1} for 7 days to establish anaerobic conditions. The reactors are designed to have a peak flow rate of 3 dry weather flow (DWF) representing an HRT of 51 h, as determined by;

$$n.DWF = n.Q_{PE} + R_i Q_{PE}$$

Where n is the DWF number (-), Q_{pe} is the daily wastewater contribution per capita (m^3 d^{-1} PE^{-1}) and R_i is the infiltration percentage (%). Flow is calculated where Q_{pe} is 0.15 m^3 d^{-1} PE^{-1} and 100% infiltration rate. Hence the reactors have a population equivalent of 2.8 PE, the average flow for a DEWATS is when n equals 1.25 hence the average flow (Q_{av}) is 1.05 m^3 d^{-1} with an HRT of 81.6 h. The reactors operated at Q_{av} for three months. The reactors were fed with sewage from Cranfield University, whose characteristics are reported in Table 4.2.

4.2.2. Analytical Methods

Influent and effluent grab samples were collected weekly and analysed for pH, TSS, VSS, TCOD and sCOD. Liquid and air temperatures were recorded at the time of sampling. TSS, VSS, TCOD and sCOD were measured in triplicate according to Standard Methods (APHA, 2005). The pCOD was calculated by subtracting sCOD from TCOD. TSS and sCOD were determined using 0.45 µm filters (Whatman, Maidstone, UK) and (Merck Millipore, Massachusetts, USA) respectively. Core samples were collected monthly from each reactor by using an 8 ft sludge sampler (Raven Environmental, Missouri, USA) at 5 points along the length of the reactor. Settled solids and scum fractions were decanted. The volume of sludge and scum was measured. TSS, VSS, TCOD and SCOD were measured in triplicate. The hydrolysis rate constants of the ESTs and their respective controls was calculated using a mass balance model of the system (Elmitwalli, 2013)(Supplementary Materials 2.7.2).

The solids concentration of each segment of the reactor was calculated based on applying the solids concentration in the core sample taken across the entire volume of the sampled segment. This was calculated based on dividing the TSS concentration of the core sample by the sample volume and multiplying the result by the volume of the section.

Table 4.1: Sewage characteristics for each configuration

	EST-Low Inlet	EST-Mid Inlet	EST-High Inlet
TSS ($mg\ l^{-1}$)	130 (±41)	120 (±36)	280 (±25)
VSS ($mg\ l^{-1}$)	120 (±45)	110 (±30)	250 (±35)
TCOD ($mg\ l^{-1}$)	220 (±94)	120 (±25)	420 (±77)
PCOD ($mg\ l^{-1}$)	190 (±79)	75 (±30)	300 (±78)
SCOD ($mg\ l^{-1}$)	65 (±12)	55 (±38)	98 (±41)
T_{in} (^{o}C)	14 (±1.8)	18 (±1.9)	14 (±1.0)
pH	7.6 (±0.09)	7.7 (±0.25)	7.36 (±0.16)

(± Standard Deviation)

4.2.3. Computational Fluid Dynamic Modelling

CFD modelling was conducted to study the impact of baffle positions on the hydrodynamics in the reactor. Five baffle configurations were modelled in total, by including the 4 test systems and a unbaffled tank (Figure 4.2). The unbaffled system was used to establish a baseline for the impact of baffles on the hydraulics. The reactors were modelled at 4 DWF conditions; 1 DWF, Q_{av}, 3 DWF and 6 DWF, which corresponds to HRTs of 102, 81.6, 51 and 28.8 h. To reduce the variability to solely the baffling each modelled reactor had the same inlet and outlet positioning. The CFD simulations were performed by using a finite element method through the use of COMSOL Multiphysics software, which described both fluid flow and solute transport processes in the reactor through the application of a finite element model. The fluid flow was described by the

Navier-Stokes equation (Plascencia-Jatoma et al., 2015), whilst the solute transport, which described the tracer dispersion in the reactor, was modelled by considering the advection-diffusion equation (Aouizerat-Elarby et al., 2000).

Laminar flow conditions were used throughout, as the equivalent Reynolds number never exceeded 2000. Steady state conditions were assumed for the resolution of the Navier-Stokes equation. For the model resolution, the reactor computational domain of Figure 4.2 was discretised by using a “coarse” mesh, with an average of 220,000 elements for the different configurations. Fluid flow modelled through the EST was considered as an incompressible Newtonian fluid. The inlet boundary condition was a constant flow velocity of; 1.24, 1.55, 2.46 and 4.39 mm s^{-1} for 1 DWF, Q_{av}, 3 DWF and 6 DWF respectively. The outlet boundary conditions were free discharge to atmospheric pressure. No slip conditions were assumed for reactor walls and baffles. A tracer residence time distribution (RTD) was modelled using a pulse injection and analysed using axial dispersion and tank in series models (Li et al, 2016). The CFD wasn’t experimentally validated, this would be needed to ensure appropriate application to real world systems.

The RTD analysis considered both a stationary study to obtain steam velocities and a time-dependent study for diluted species transport. The Peclet number of the reactors was calculated as the inverse of the dispersion number. The dispersion number was calculated from the variance in the RTD (Supplementary materials 4.6, Table A6.1).

4.3. Results and Discussion

4.3.1. Sludge Accumulation

The majority of solids accumulated in the hydrolysis sections for all EST configurations (Figure 4.3). To illustrate, 81%, 78%, 68% of the total sludge accumulated was associated with hydrolysis sections in configurations EST-Low, EST-Mid and EST-High respectively at the end of each trial. The solids concentration in the two compartments of EST-High’s hydrolysis section is greater than the other two ESTs with final concentrations of 5.4 and 11.6 g m^{-3}

compared to an average concentration of 2.5 and 1.0 g m^{-3} for EST-Low and Mid. When normalised just to length, the total solids accumulation in EST-High was 78% at a distance of 1200 mm, equivalent to the other configurations. Comparison to traditional baffling arrangement for septic tanks (control) revealed less controlled accumulation of the sludge and more even distribution along the length of the reactor.

The observed hydrolysis rate constants of the ESTs establishes that baffle arrangement can increase the hydrolysis rate (Figure 4.4). To illustrate, the average hydrolysis rate constant of the EST-High and EST-Mid configurations were 0.027 d^{-1} and 0.035 d^{-1} respectively compared to the controlled which averaged 0.0089 d^{-1} across the three trials. This is further supported by the fact the system with the lowest designed mixing, EST-Low delivered a lower hydrolysis rate constant of 0.0060 d^{-1} which is within the range observed for the control systems.

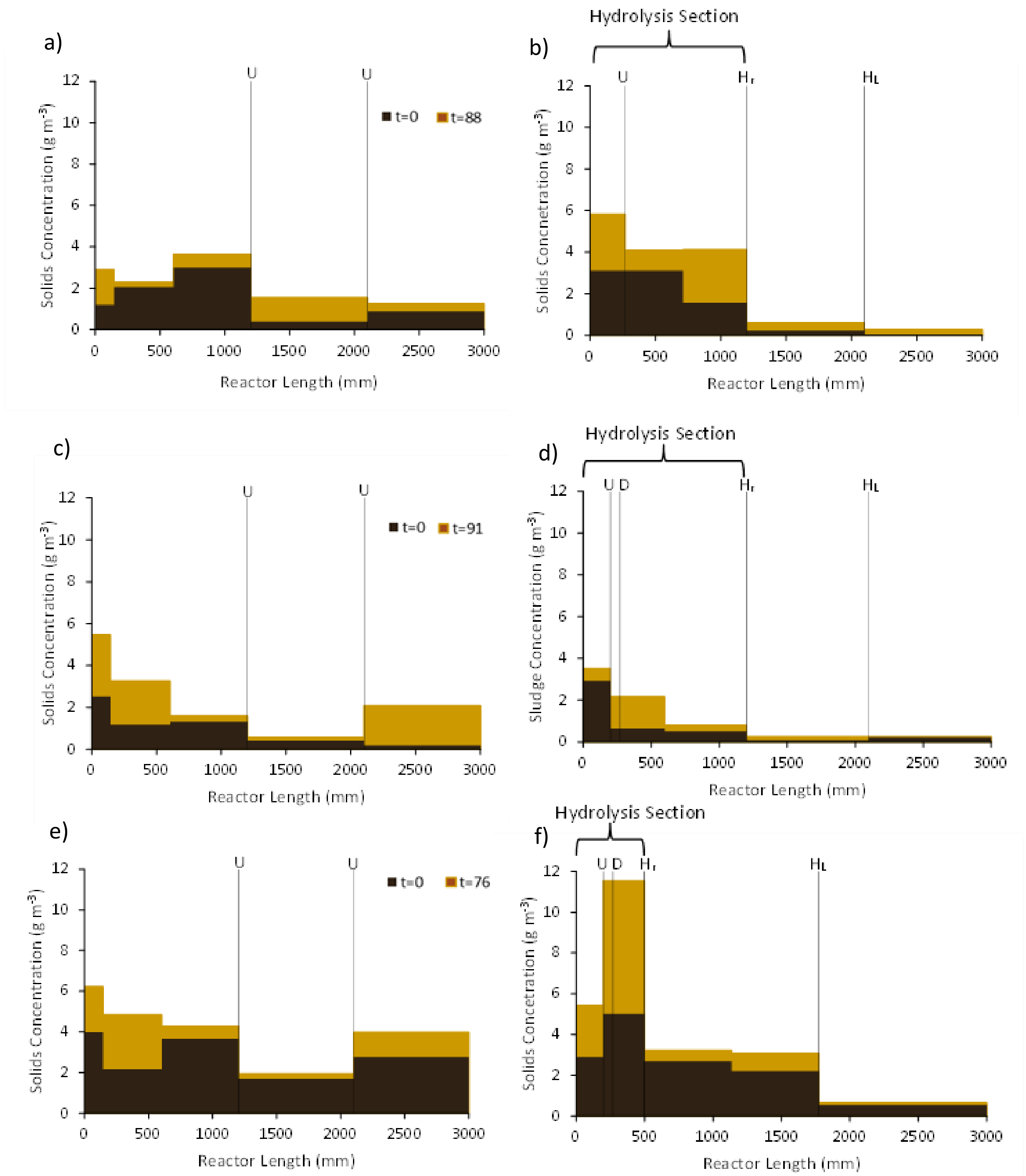

Figure 4.3: Sludge accumulation pattern for the operational period of the reactors based on the calculated solid concentration in each segment for a) Control EST-Low b) EST-Low c) Control EST-Mid d) EST-Mid e) Control EST-High f) EST-High

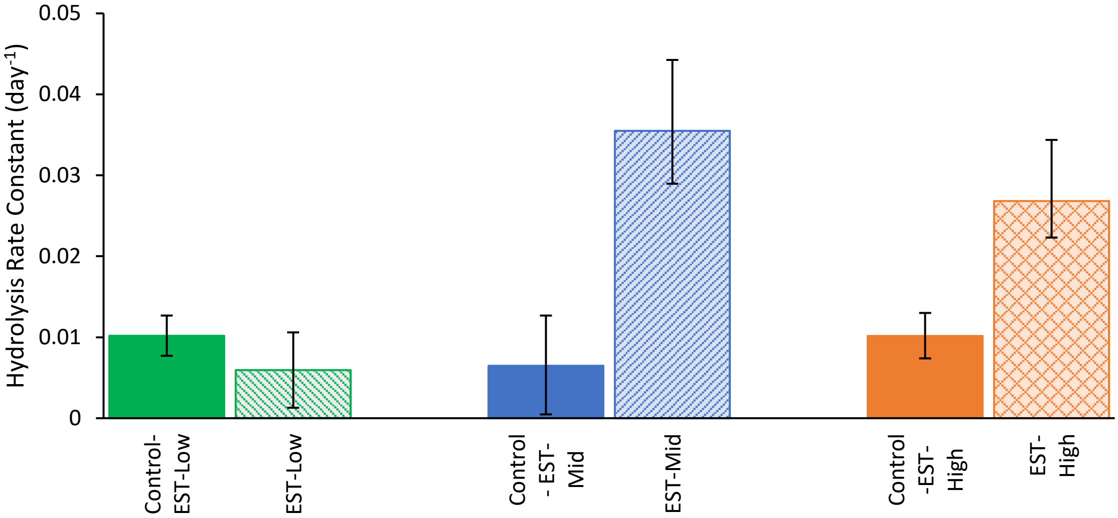

Figure 4.4: Hydrolysis rate constant of pilot systems, error bars are 95% confidence range

The hydrolysis rate constant of EST-Mid is greater than those of other reported ABRs or septic tanks (Table 4.2), whose rate constant was in the range of a horizontal AP and UASB operating at comparable HRT and temperature (HRT from 28 to 82 h and temperature from 10 to 19^{o}C). UASBs provide an intensified treatment process, with upflow equivalent to the hydrolysis section of the EST. Al Jamal & Mahmoud, (2009) operated two 0.8 m^3 UASB reactors concurrently with an HRT of 48 and 96 h, the removal of the systems was not significantly impacted by HRT variation. The increased hydrolysis rate of the 48 h UASB led to a reduction in TSS accumulation of 0.4 kg PE^{-1} yr^{-1}. However, the reduction in volume per capita meant the time for the system to reach 50% sludge volume was 3.63 years at 48 h and 4.43 years at 96 h. Baffling provides greater flexibility in optimisation consideration of upflow velocity than a UASB as the upflow velocity can be varied in a region without changing the overall system geometry or flowrate.

Table 4.2: Reported hydrolysis rate constants of low-rate anaerobic reactors operating from 10 to 19ºC and batch experiments conducted in Chapter 3

Reactor	Temp (ºC)	HRT (h)	Hydrolysis Rate Constant (day^{-1})	Reference
Control	15	81.6	0.0089 (0.005-0.013)	This Study
ESP-Low	14	81.6	0.0060 (0.0013-0.011)	This Study
ESP-Mid	18	81.6	0.035 (0.029-0.044)	This Study
ESP-High	14	81.6	0.027 (0.022-0.034)	This Study
Septic Tank	15	165	0.016	(Lin et al., 2019)
Vertically Baffled ABR	14.6	28	0.0084	(Schalk et al., 2019)
Horizontally Baffled AP	16.5	74	0.032	(Cruddas et al., 2014)
Horizontally Baffled AP	10.5	55	0.016	(Cruddas et al., 2018, 2021)
Horizontal-Vertical Baffled AP	10.5	55	0.0087	
UASB	17	48	0.058	(Al-Jamal & Mahmoud, 2009)
UASB	17	96	0.036	
Unmixed Batch Test	5	-	0.0038	Chapter 3
Unmixed Batch Test	7	-	0.0092	Chapter 3
Unmixed Batch Test	10	-	0.0073	Chapter 3
Mixed (35 rpm) Batch Test	10	-	0.022	Chapter 3
Mixed (85 rpm) Batch Test	10	-	0.0033	Chapter 3
Unmixed Batch Test	15	-	0.0064	Chapter 3
Unmixed Batch Test	20	-	0.014	Chapter 3
Unmixed Batch Test	37	-	0.014	Chapter 3

(95% confidence interval)

4.3.2. Pilot Plant Treatment Performance

Enhancing the hydrolysis of accumulated solids through alteration of the baffling arrangement was not detrimental to the treatment performance for any of the ESTs (Figure 4.5). This was observed for all parameters (TSS, VSS, TCOD, pCOD, sCOD) with the observed removal not being significantly different to those of the respective controls for EST-Low and EST-High and treatment was found to be significantly improved for EST-Mid (Figure 4.5). To illustrate in the case of EST-High, TSS and sCOD removal levels 75 ±4% and 29 ±18% for the control compared to 76 ±2% and 38 ±16% for EST-High. Further, typical reported removal rates for septic tanks treating domestic wastewater at ambient temperatures ranges from 51 to 85% and 35 to 70% for TSS and COD removal respectively (Kirjanova et al., 2014, Mesquira et al., 2021).

Table 4.3: Sewage and effluent characteristics for each configuration

	EST-Low			EST-Mid			EST-High		
	Inlet	Control Outlet	EST Outlet	Inlet	Control Outlet	EST Outlet	Inlet	Control Outlet	EST Outlet
TSS (mg l^{-1})	130 (±41)	62 (±23)	63 (±12)	120 (±36)	89 (±45)	41 (±15)	280 (±25)	69 (±19)	66 (±14)
VSS (mg l^{-1})	120 (±45)	55 (±12)	47 (±13)	110 (±30)	77 (±32)	33 (±13)	250 (±35)	55 (±12)	59 (±18)
TCOD (mg l^{-1})	220 (±94)	140 (±38)	130 (±46)	120 (±25)	110 (±28)	58 (±30)	420 (±77)	120 (±23)	120 (±55)
PCOD (mg l^{-1})	190 (±79)	70 (±32)	82 (±24)	75 (±30)	80 (±19)	23 (±19)	300 (±78)	48 (±17)	36 (±26)
SCOD (mg l^{-1})	65 (±12)	77 (±13)	76 (±12)	55 (±38)	51 (±12)	46 (±29)	98 (±41)	61 (±11)	69 (±22)
T_{in} (^{o}C)	14 (±1.8)			18 (±1.9)			14 (±1.0)		
pH	7.6 (±0.09)			7.7 (±0.25)			7.36 (±0.16)		

(±Standard Deviation)

The principal method of comparison between configurations was based on removal efficiency rather than absolute values to reflect the variation in influent levels observed across the trials (Table 4.1). To illustrate, influent TSS concentration was 130 ±31 mg l^{-1}, 120 ±36 mg l^{-1} and 280 ±25 mg l^{-1} during the trials for EST-Low, EST-Mid and EST-High respectively. The corresponding effluent concentrations were 63 ±12 mg l^{-1}, 41 ±15 mg l^{-1} and 66 ±14 mg l^{-1} indicating greater similarity than when viewed through removal efficiency lens. This is most notable when considering sCOD where the effluent levels were 76 ±12 mg l^{-1}, 46 ±29 mg l^{-1} and 69 ±22 mg l^{-1} for EST-Low, EST-Mid and EST-High respectively. The corresponding removal levels were -19%, 16% and 38% respectively.

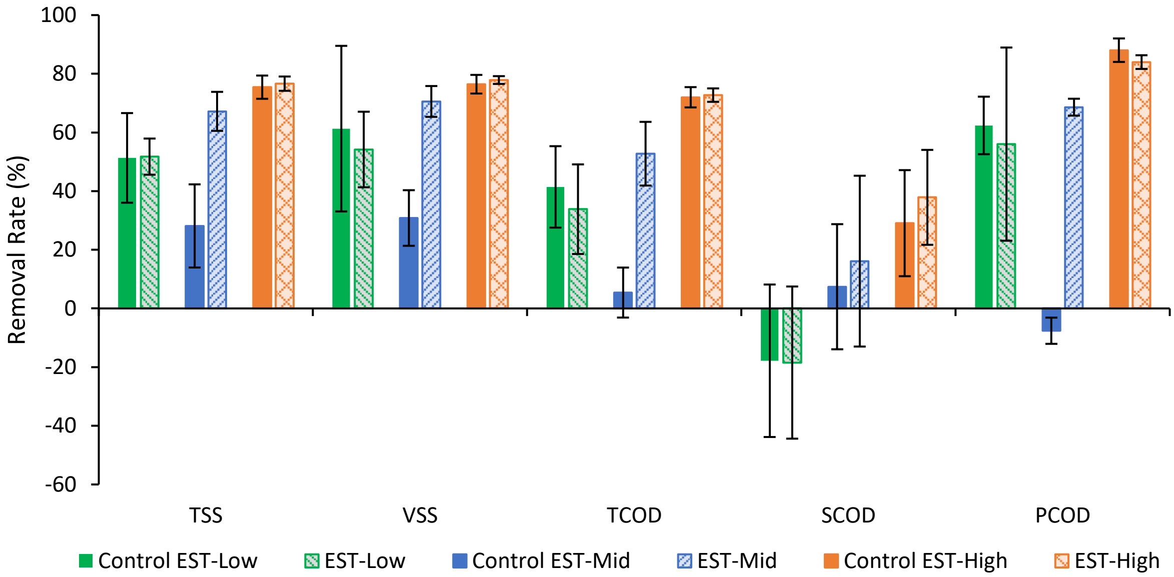

Figure 4.5: Removal levels of the ESTs and control systems

4.3.3. Hydraulic Analysis

Residence time distributions generated from the CFD model revealed a shift to a more general plug flow pattern through the use of enhanced baffling arrangements. For instance, the delay in the tracer peak increased from 0.29 normalised time for the unbaffled and control system at an HRT of 102 h to 0.5 normalised time for the ESTs. At the average flow, where HRT is 81.6 h, the estimated mean residence times for the ESTs were 66, 72 and 72 h for EST-Low, EST-Mid and EST-High respectively compared to 63 h for the Control configuration. Viewed in terms of dead space the impact of baffling configuration was further highlighted as the unbaffled system had lower dead space than a traditional septic tank (control) or EST-Low (Figure 4.6a). EST-Mid had the lowest dead space level at low and average flow, which varied between 5 and 9%. In comparison, EST-High had a stable dead space level across all HRTs at between 11 and 12%. This reinforces the notion that it is important to baffle appropriately for the targeted function.

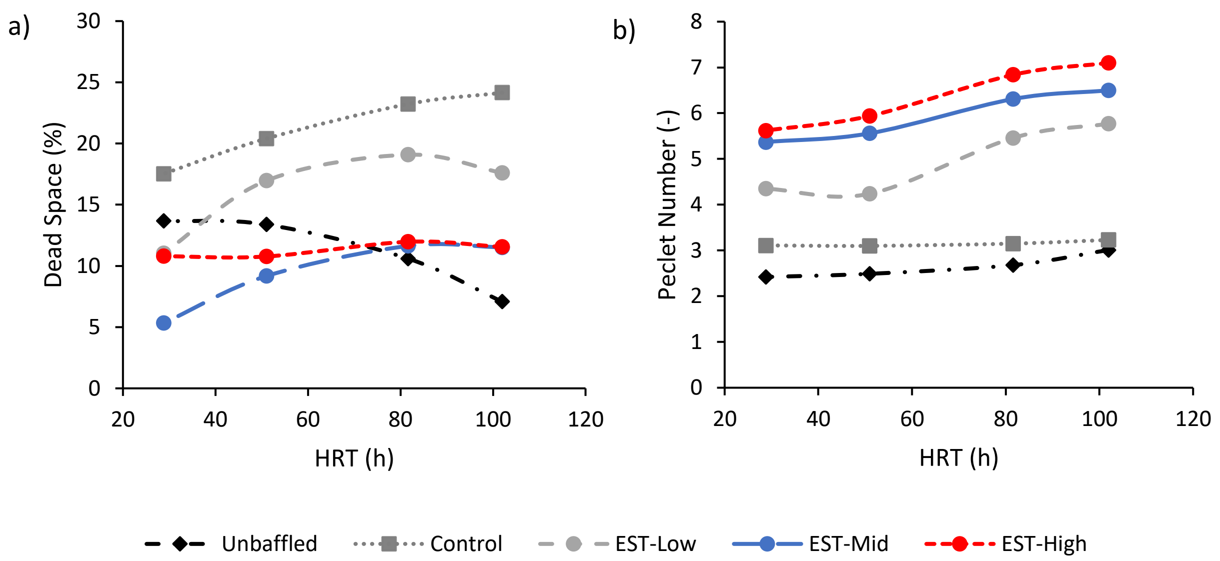

Figure 4.6: CFD analysis of the a) dead space against HRT b) number of tanks in series

The increase in the plug flow nature of the flow of the ESTs was confirmed by analysing the RTDs in terms of Peclet number. EST-Mid and EST-High remained in plug flow (Pe>5) (Li et al., 2016) for all studied HRTs (Figure 4.6b).

Comparatively, EST-Low deviated from plug flow conditions at higher flowrates with a Peclet number below 5 at HRTs below 70 h. Neither the unbaffled nor control were in plug flow for any HRT.

Further investigation into the local velocity distribution within the ESTs revealed how the changing baffle configuration impacted the total volume and position of the high velocity zones within the ESTs (Figure 4.7, shown in red). For instance, the absence of a downcomer baffle in EST-Low results in the high velocity extending only into the top of the second zone providing little hydrolytic mixing contrasted to the high hydrolytic mixing seen for the equivalent region in EST-Mid. Comparison of EST-High and EST-Mid illustrates that the proportion associated to hydrolytic mixing is controlled by the positioning on the first horizontal baffle (H_r) and hence this represents a key design feature that can be tailored to specific need. This is because the shortened hydrolysis region of EST-High did not produce additional benefit through more intense low velocities. Further, extending the region by pushing the first horizontal baffle further into the tank (as depicted by EST-Mid) had no detrimental impact on the efficacy of the settling zone.

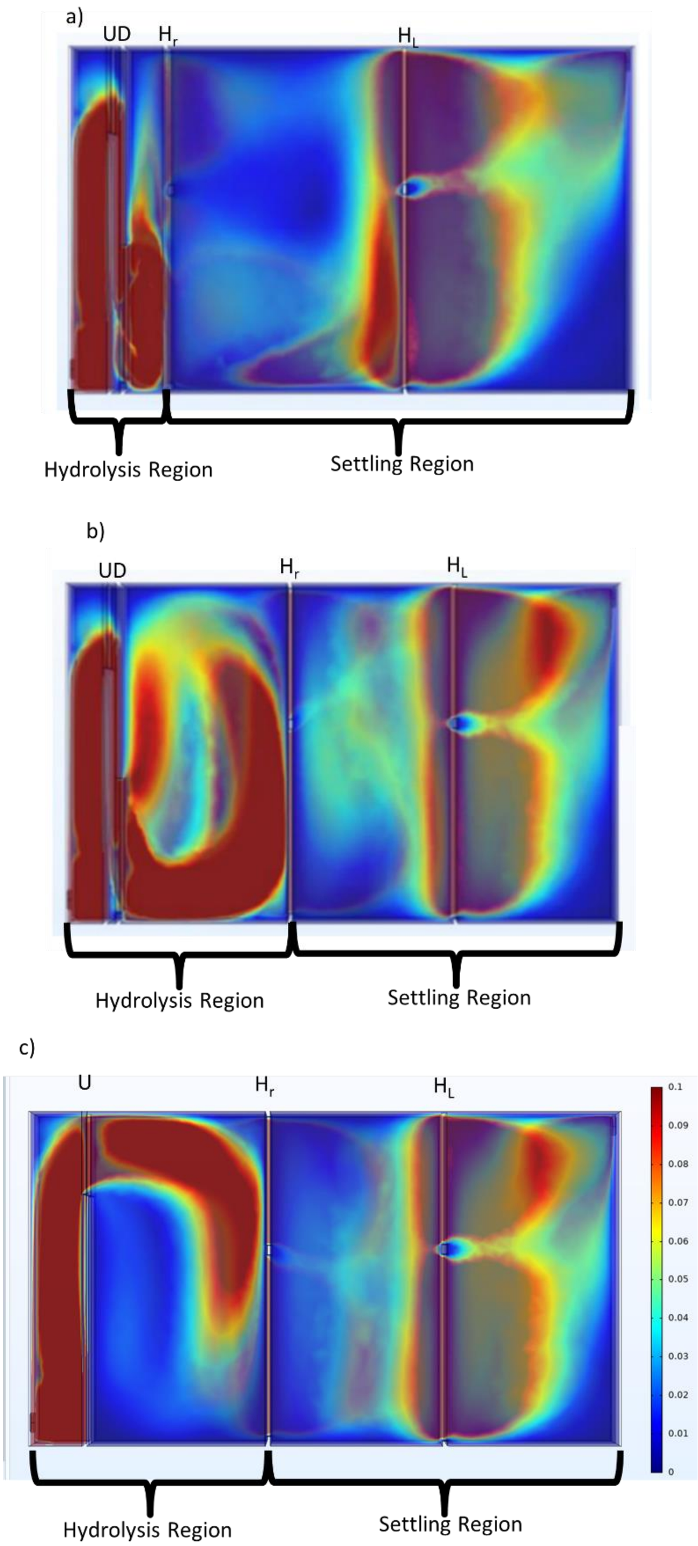

Figure 4.7: Lateral view for velocity colour scale in mm s^{-1} a) EST-high b) EST-Mid c) EST-Low at HRT = 28.8 h, U is upcomer baffle, D is dowmcomer baffle, H_R is right hand horionztal baffle and H_L is left hand horizontal baffle

4.3.4. Mass Balance Model

To understand the impact of the proposed enhanced baffling arrangements, the EST configurations were modelled to establish the likely desludging frequencies. Normalising for the variation in feed concentration, the systems were modelled with the influent solids expected as per British Water Standard (2013) and with the removal of EST-High configuration. Increasing the hydrolysis rate constant from the average control to EST-Mid increases the desludging frequency from 4.9 to 6.7 years (Figure 4.8). The modelled desludging frequency of the septic tank is in line with observations of operational septic tanks (Mahon et al., 2022). Importantly, the enhancement in hydrolysis rate delivered by the baffling arrangements in EST-Mid and EST-High resulted in desludging frequencies between 5-7 years. The benefit of enhancing the hydrolysis rate beyond around 0.025 days^{-1} on desludging frequency was negligible, for all fill levels, and hence the simple baffling arrangements proposed provide a degree of resilience. At this average hydrolysis rate, the desludging frequency can be extended through increasing the fill level in the tanks. Consequently, this can be tailored to maintenance requirements of the NBS flowsheet components providing a robust integrated system.

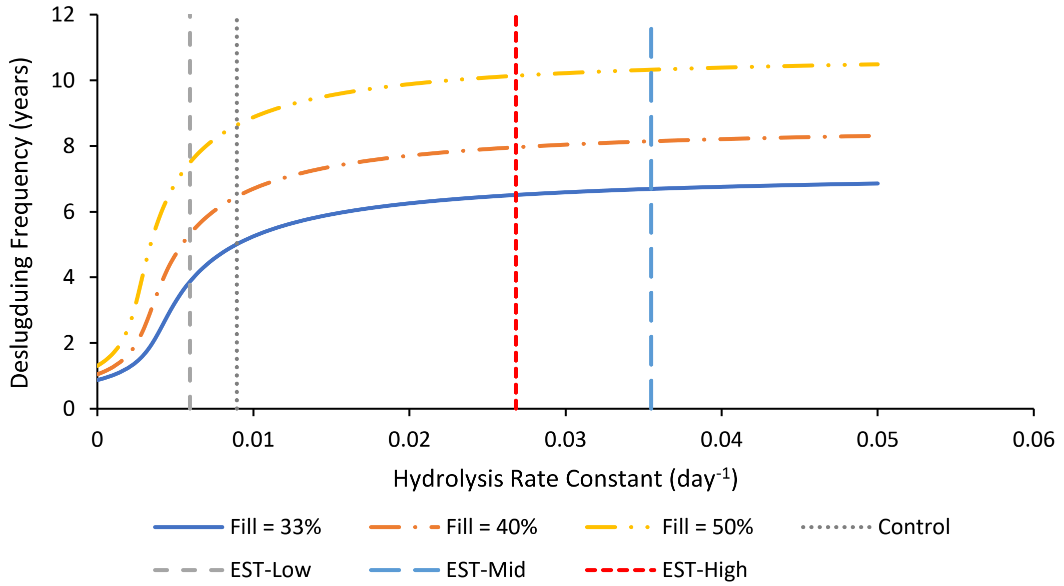

Figure 4.8: Projected desludging frequency of the pilot systems treating British Standard wastewater at 33, 40 and 50% fill levels.

The control systems in this study have a 3 DWF HRT of 48 hours compared to the 12 hours of Scottish Waters' design standard. At a fill rate of 33% the control systems in this study are projected to require desludging every 4.9 years. Comparatively the average desludging frequency of a Scottish Water operated septic tank is 6 months. Thus, the increase in sizing does provide a significant predicted increase in required desludging frequency, however the adjustment of baffles is shown to further increase the projected desludging frequency by several years (Figure 4.8). The design of the ESTs consists of a hydrolysis region and settling region, the impact of long-term accumulation on these sections is unknown. However, with the hydrolysis section being at the front of the reactor it will have a continual supply of undigested solids and more digested solids will be passed forward onto the settling region. Aided by the fact hydrolysis reduces particle sizes, hence the hydrolysis region will be able to subject the least digested material to the highest intensity mixing and mass transfer.

The work presented in the current study was conducted over a short operating periods of 90 days per systems and an average temperature of 14 and 18 °C. The actual temperatures within septic tanks are currently unknown but may be lower that the test temperatures for a proportion of the year. The next stage of the work is therefore to run prolonged trials across multiple years to confirm the translation of the current findings to real applications in terms of the average annual hydrolysis rate and the corresponding required desludging frequency.

This study considers the impact of baffling on the bulk hydrolysis rate constant and modelling this to determine the desludging frequency. However, in applying this to a longer-term study the impact of the macro-composition of sludge may be required. The three primary constituents of sludge are proteins, carbohydrates and lipids, with each having differing degradation pathways and rates. In ADM1 (Batstone 2002) the hydrolysis rate constants of are proteins, carbohydrates and lipids in mesophilic conditions are modelled as equivalent. Conversely Mahmoud et al., (2004) found at 25^{o}C the hydrolysis rate constants of proteins, carbohydrates and lipids were; 0.067, 0.10 and 0.11 d^{-1} respectively.

Investigating the impact of sludge retention time of hydrolysis, Miron et al., (2000) found that in a 25°C CSTR with a sludge retention time of 8 days the hydrolysis rate of proteins and carbohydrates were 0.428 and 0.806 d^{-1} respectively, however increasing the sludge retention time to 15 days led to the protein hydrolysis rate decreasing to 0.153 d^{-1} whereas for carbohydrates the hydrolysis rate increased to 0.842 d^{-1}. However there is limited knowledge of the macro composition or rate changes over the course of years or unmixed systems. On a bulk scale in septic tanks, it is frequently suggested that increasing sludge retention time leads to increased sludge degradation rates and hence hydrolysis rates (Mahon et al., 2022) based on observed sludge accumulation rates. However modelling in Chapter 2 showed this to be consistent with a fixed hydrolysis rate based on a bulk composition. There is a paucity of hydrolysis data at low temperatures (Chapter 2) and limited findings of the impact of sludge retention time and macro-composition and this an area which requires further study.

4.4. Conclusions

The work has successfully demonstrated the ability to raise the hydrolysis rate observed in septic tanks by alteration of the baffling arrangements. The outcome is an ability to design enhanced septic tanks that require infrequent desludging that can be match to the refurbishment needs of nature-based solutions. This synergy supports the development of a new flowsheet for small rural works that has the capability to deliver high levels of treatment with minimal maintenance requirements. The key is to utilise vertical baffling to provide a high velocity zone were the fresh feed contacts the existing sludge materials to enhance mass transfer and ultimately hydrolysis. Whilst further work is required to translate the findings to long term operation and the inherent variability in temperature of feed character. The potential for infrequent desludging without detrimental impacts on performance offer a fresh lens with which to view septic tanks and the value proposition they can offer for decentralised, rural wastewater treatment.

Acknowledgements

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Table A6.1: Description of the different parameters utilised in the RTD analysis

Name	Equation	Notes
Normalised time	$\theta = \frac{t}{HRT}$	t = sampling time
Normalised tracer concentration	$C_\theta = \frac{C(t)}{C_0}$	Co=initial tracer concentration
RTD function	$E(t) = \frac{C(t)}{\int_0^\infty C(t)\,dt}$	C(t)= measured tracer concentration at time t
Mean residence time	$\bar{t} = \frac{\int_0^\infty tC(t)dt}{\int_0^\infty C(t)\,dt} = \int_0^\infty tE(t)\,dt$	
Distribution variance	$\sigma^2{}_t = \frac{\int_0^\infty (t-\bar{t})^2 C(t)\,dt}{\int_0^\infty C(t)\,dt} = \int_0^\infty t^2 E(t)\,dt - \bar{t}^2$	
Dead space	$V_d = \left(1 - \frac{\bar{t}}{HRT}\right) * 100$	
Dimensionless variance of RTD	$\sigma^2 = \frac{\sigma_t^2}{\bar{t}^2}$	$\frac{D}{\mu L}$= dispersion number
	$\sigma^2 = 2\left(\frac{D}{\mu L}\right) - 2\left(\frac{D}{\mu L}\right)^2 * (1 - e^{-\frac{\mu L}{D}})$	
Peclet number	$Pe = \frac{\mu L}{D}$	Inverse to dispersion number
Number of tanks in series	$N = \frac{1}{\sigma^2}$	

5. Constructed Wetlands for the Polishing of Enhanced Septic Tank Effluent: An Operational Case Study

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Abstract

The study aimed to assess the performance of two pilot scale constructed wetlands treating enhanced septic tank effluent to establish is the combination was able to deliver treatment goals similar to those for small wastewater treatment works. To test this, two 6.25 m^2 wetlands, a vertical flow and an aerated horizontal flow wetland were operated for 22 months to establish treatment performance when treating the effluent from an enhanced septic tank. Both systems had average removals of TSS, BOD and NH_4 in the range of 80-95% demonstrating the suitability of the combination. However, the AHF performance was insufficient to meet the 95th percentile concentration for TSS. Diagnosis of the systems indicated it was due to operation at an elevated organic loading rate. In the future it is recommended that the organic loading of the system is maintained below at least 15 gBOD m^{-2} d^{-1} in the future.

Keywords: vertical flow wetland, aerated horizontal flow wetland.

5.1 Introduction

The eutrophication risk of effluent discharged from septic tanks into rural watercourses has been chronically underestimated (Withers et al., 2011), with the removal of septic tank systems having been shown to halve the ammonium concentration in a waterbody (Withers et al., 2012). A limiting factor for conventional septic tank system is the lack of aerobic treatment, meaning nitrification of ammonia is not possible. Post septic tank aerobic treatment can be delivered with high energy consuming aerobic processes (e.g. submerged aerated filters) but can also be treated with nature based solutions (e.g. constructed wetlands, CW) to provide a less energy intensive aerobic treatment solution. Two such CW systems are vertical flow (VF) and aerated horizontal flow (AHF) wetlands, both have been shown to successfully operate as secondary treatment of anaerobic effluent (Butterworth et al., 2016; Mander et al., 2014; Mietto & Borin, 2013).

The operation of a VF wetland provides an aerobic environment by cyclically dosing and draining of the treatment bed, allowing for passive aeration of the treatment biofilm between each treatment load. Each feed dose of the VF floods the surface of the wetland and then percolates through the media. Conversely, the AHF provides continuous flow maintained entirely below the surface of the wetland through which coarse bubbles are pumped into the bed. Forced aeration provides the conditions for nitrification to occur, the aeration of a AHF is typically significantly lower than is required for other forced aerated systems such as submerged aerated filters (Chapter 6). In spite of the reduced aeration requirements, the aeration infrastructure of AHFs often provides excessive aeration, within a gravel media the zone of influence of a bubble is 150-300 mm, and a lower aeration rate than is commonly applied can increase the oxygen transfer (Butterworth, 2014)

The VF and AHF wetland have different benefits, with the VF representing an entirely passive system which can operate without electricity. Comparatively an AHF has a higher energy demand but being a more intensified system has a lower associated footprint. This study aims to determine whether, VF and AHF are appropriate treatment solutions to achieve the target consents (Table 5.1) treating septic tank effluent. The consent levels represent a moderate level of consent, suitable for secondary discharge.

Table 5.1: Target consent for treatment flowsheet

Pollutant	95% Effluent Limit (mg l^{-1})
TSS	25
BOD	25
NH_4	15

5.2 Materials and Methods

5.2.1 Wetland Construction

The constructed wetlands were constructed in 2.5 m x 2.5 m x 1 m (L x W X H) glass reinforced plastic tanks. The VF wetland was designed based on constructed wetland association design guidelines (CWA, 2017). The primary treatment layer from 20-75 cm of the bed comprised sharp sand and 10 mm pea gravel mixed at a 2:1 ratio, the final 5 cm layer is 10 mm pea gravel. The treatment layer sits above a drainage layer of 20 -40 mm gravel for the bottom 15 cm and a 5 cm transition layer of 10 mm pea gravel. The hydraulic conductivity of the sand media was tested using the Grant test (CWA,2017) to ensure a hydraulic retention time between 30 and 60 s (CWA, 2017), the record hydraulic retention time of the treatment layer was 44 s. There were 4 equally spaced collection pipes in the VF wetland (figure 5.1a), at the opposing end to the outlet an aeration pipe arose from each collection pipe to 20 cm above the treatment bed. Two distribution pipes were installed on top of the treatment layer (Figure 5.1b), running the length of the wetland with 4 dosing points.

Figure 5.1: Construction stages of the VF wetland a) Drainage layer and collection pipes b) Treatment layer and distribution pipes

The AHF was designed based on wetlands outlined in Butterworth et al., (2016), the treatment bed was a height of 60 cm of 10 mm pea gravel. A collection pipe constructed from 1 inch PVC pipe, ran perpendicular to the flow direction the length of the reactor 5 cm from the base of the wetland. The distribution pipe was also constructed from 1 inch PVC pipe was 55 cm from the base of the wetland running perpendicular to the flow direction (Figure 5.2b). The distribution and collection pipes were surrounded by 5 cm of 20 – 40 mm gravel in each direction. The aeration of the AHF consisted of a 12 mm diameter low density polyethene pipe perforated with 2 mm holes at 300 mm intervals which snaked up and down the wetlands with each line separated by 300 mm with the initial and final line being 150 mm for the exterior wall giving a total of 5 aeration lines (Figure 5.2a). Both wetlands were planted with Typha Latifolia with 4 plants m^{-2} (Butterworth et al., 2016).

Figure 5.2: Construction stages of the AHF wetland a) Aeration piping and collection area, b) Treatment bed and distribution pipe

5.2.2 Wetland Operation

The wetlands were operated from July 2020 to May 2022, with a 2-month rest period, giving a total of 20 months of operation. Both wetlands were fed from combined effluent of two septic tanks treating raw domestic wastewater at ambient temperatures. The septic tanks had a hydraulic residence time (HRT) of 3.2 days operating a temperature from 8-25^{o}C and treated raw domestic wastewater. A detailed description of the septic tanks is provided in Chapter 4.

The VF wetland was designed to have a peak hydraulic loading rate (HLR) of 0.12 m^3 m^{-2} d^{-1} and an operational HLR of 0.075 m^3 m^{-2} d^{-1} (Table 5.2). The feed flowrate of 0.47 m^3 d^{-1} was feed from a 0.5 m^3 holding tank with a frequency of 8 doses of 59 l with a dose duration of 1 minute with a dose frequency every 4 hours. The AHF was designed based on a target ammonia load of 5 gNH_4 m^{-2} d^{-1} (Table 5.2) based on projected influent concentrations. This gave an operational flowrate of 0.60 m^3 d^{-1}, the AHF was fed continuously by peristaltic pump. The aeration was supplied from a compressed air system at a continuous fed of 30 l min^{-1}.

Table *5.2*: Wetland design envelope and mean operational loading (± standard deviation)

	VF		AHF	
	Design Parameter	Operational	Design Parameter	Operational
Hydraulic Loading (m^3 m^{-2} d^{-1})	0.075	0.075	0.4	0.15
Solids Loading (gTSS m^{-2} d^{-1})	8	5.7	18	7.3
Organic Loading (gBOD m^{-2} d^{-1})	10	13	15	17
Ammonia Loading (gNH_4 m^{-2} d^{-1})	15	2.3	5	3.0

5.2.3 Analytical Methods

Samples were collected for the influent and effluent of the wetlands from July 2020 and July 2021 were sampled weekly (excluding the Covid-19 related shutdown). Sampling was paused between July 2021 and February 2022 after which the wetlands were sampled monthly. The temperature and dissolved oxygen of the samples was measured at the point of sampling. BOD_5, NH_4 and TSS were measured in triplicate using Standard Methods.

5.3 Results and Discussion

The organic loading of the wetlands was beyond the target guidelines (Table 5.2) for both systems, with a higher than expected organic concentration in the fed. However, this did not appear to negatively impact operation with regards to organic removal with the wetlands having a mean effluent of 11±7 and 8±5 mg l^{-}

^{1} for the VF and AHF respectively (Table 5.3). This equates to a removal level of 93 and 84% for the VF and AHF respectively. The mean effluent and removal of TSS and NH_4 for the VF and AHF shows strong performance, with a TSS effluent of 5±3 and 12±12 mg l^{-1} for the VF and AHF respectively. For an AHF with equivalent loading treating septic tank effluent (Butterworth et al., 2016) the median effluent of TSS and NH_4 were 21 mg l^{-1} and 1.2 mg l^{-1}, equivalent to the performance of the study wetlands. Additionally, the treatment provided by the constructed wetlands is comparable to a package treatment system, which reports a mean effluent range of 16 mg l^{-1}, 11 mg l^{-1} and 8 mg l^{-1} for TSS, BOD and NH4 respectively (Tricel Environmental, 2022).

Table 5.3: Mean influent and effluent for the pilot constructed wetlands

	Influent	VF Effluent	AHF Effluent
TSS (mg l^{-1})	76 (±55)	5 (±3)	12 (±12)
BOD (mg l^{-1})	178 (±71)	11 (±7)	8 (±5)
NH_4 (mg l^{-1})	31 (±7)	0.5 (±0.8)	3 (±2)

Both wetlands operated as designed, without surface ponding. This is in-line with expectation that a CW only requires solids removal every five to seven years (Chapter 6). The plant growth on the VF was strong throughout the observed period, with full coverage of the surface (Figure 5.3a). The AHF however had growth issues in the later portion of the wetland with the *Typha* failing to establish (Figure 5.3b) and required replanting from reeds harvested from the VF wetland. It is posited that the aeration caused disturbance to the roots which hindered growth, which has been noted in other AHF systems (Butterworth et al., 2016).

a) b)

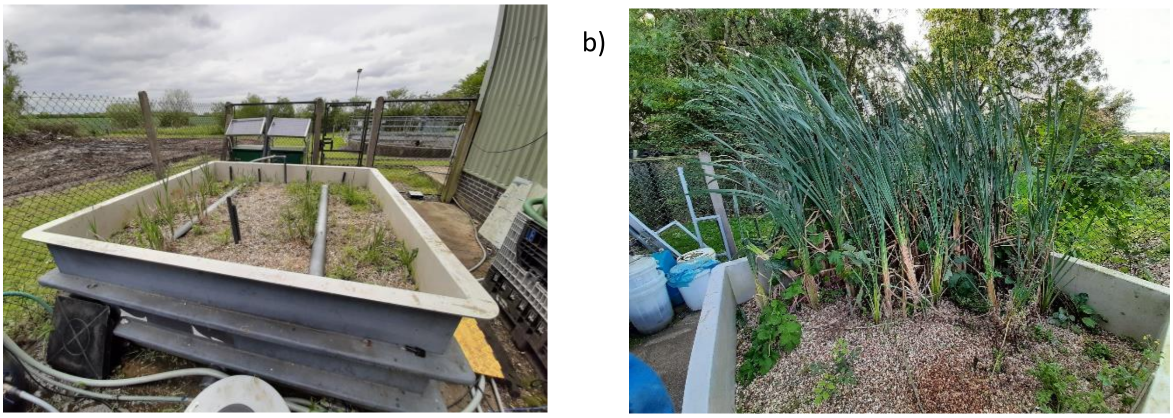

Figure 5.3: Surface condition of the wetlands a) The VF pictured in May 2021 b) The AHF pictured in October 2021

The consented effluent of the VF within this study is within the target range for all parameters (Table 5.4). However, the AHF is above the target consent for TSS with a 95th percentile effluent of 34 mg l^{-1} (Table 5.4). There was no observed surface channelling of the AHF which can be a common failure point of horizontal flow wetlands. The effluent percentile of TSS for the AHF highlights that the 95th percentile TSS effluent is not due to high range outliers but a more persistent issue within the system (Figure 5.4). A similar issue was encountered by Butterworth et al., (2016) for an AHF treating septic tank effluent showing high TSS effluent peaks having a 95th percentile TSS effluent of 45 mg l^{-1}. It is therefore recommended for this style of system to reduce the loading to ensure compliance, greater study is required to establish the optimum loading level.

Table 5.4: Consent effluents of the VF and AHF

	Target (mg l^{-1})	95th Percentile VF effluent	95th Percentile AHF effluent
TSS	25	15	34
BOD	25	22	14
NH_4	15	1.7	5.6

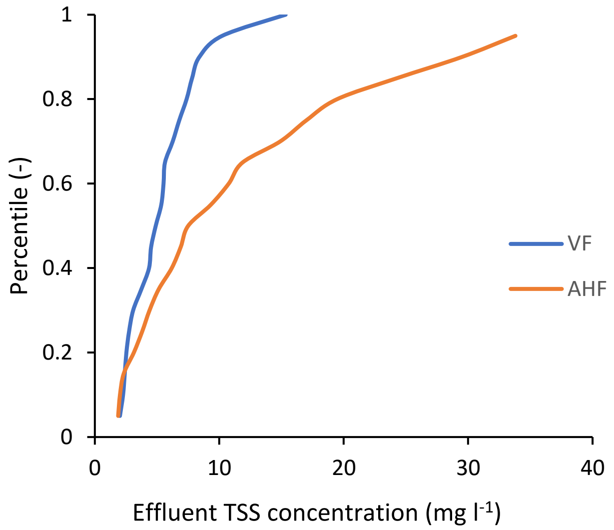

Figure 5.4: TSS effluent percentile graph for the VF and AHF

The high TSS samples recorded from the AHF were found during early operation, with two samples over 25 mg/L in July 2020, alongside 2 high effluent values in both the summer of 2021 and in spring 2022. The high effluent values do not align with a high influent sample, although due to the retention time of the wetland and the weekly sampling frequency this cannot be ruled out. Indeed, the retention time of the VF is significantly shorter so a high peak prior to sampling would be more noticeable on the AHF. The high samples recorded in the early operation of the wetland show that the system likely had not fully developed yet and should be considered with respect to the commissioning of the system.

5.4 Conclusion

Nature based solutions in the form of vertical flow or aerated horizontal flow wetlands provide an effective post treatment after a septic tank to enable effluent qualities congruent with small work discharge consents. Both systems could deliver substantially below the required standard for BOD and ammonia with the VF wetland effluent delivering a low ammonia of 0.5±0.8 mg l^{-1} and a BOD of 11±7 showing the potential for the technology combination. In comparison the AHF did not reach the target 95th percentile effluent consent of 25 mg l^{-1} TSS. Diagnosis of the trials indicated that this was related to organic loading rate the system ran at and it is recommended the organic loading of the system is maintained below at least 15 gBOD m^{-2} d^{-1} in the future.

Acknowledgments

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Data underlying this paper can be accessed at. 10.17862/cranfield.rd.21732836

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6. A Wholelife Cost and Carbon Perspective of Alternatives to Septic Tanks Utilising Nature-Based Solutions

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Abstract

Septic tank systems (STS) are a widely utilised treatment flowsheet for decentralised wastewater treatment in the UK. However, there is a growing consensus that these systems have a sizeable detrimental impact on the environment and there is a need for rural wastewater treatment flowsheets which have improved treatment capabilities. This study examines the lifetime cost and carbon emissions of using an enhanced septic tank nature-based solution (EST-NBS) to improve STS compared to a package treatment system (SAF). The wholelife cost (WLC) of the flowsheets and scope 2 cradle to grave lifetime carbon emissions (LCE) of the flowsheets were assessed. The EST-NBS flowsheets represent a lower cost improved treatment system than SAFs at population equivalents (PE) from 5 to 1000, projecting the system as the more cost-effective upgrade to an STS. An STS averages an LCE of over 4000 kg $CO_{2\text{-eq}}$ PE^{-1}, with all other considered flowsheets having lower emissions. The EST-NBS flowsheets had lower carbon emissions than SAFs. Even at low populations upgrading from an STS to an EST-NBS is a competitive abatement strategy, with costs of £260 $tCO_{2\ eq}^{-1}$ emissions avoided, at 1000 PE a NBS flowsheet has an abatement cost of -£17 $tCO_{2\ eq}^{-1}$. This study shows the potential of using NBS flowsheets in rural wastewater treatment providing both a carbon and cost incentive against traditional designs and abatement cost in line with technology such as thermal hydrolysis.

Keywords: septic tanks, nature-based solutions, WLC, LCA

6.1 Introduction

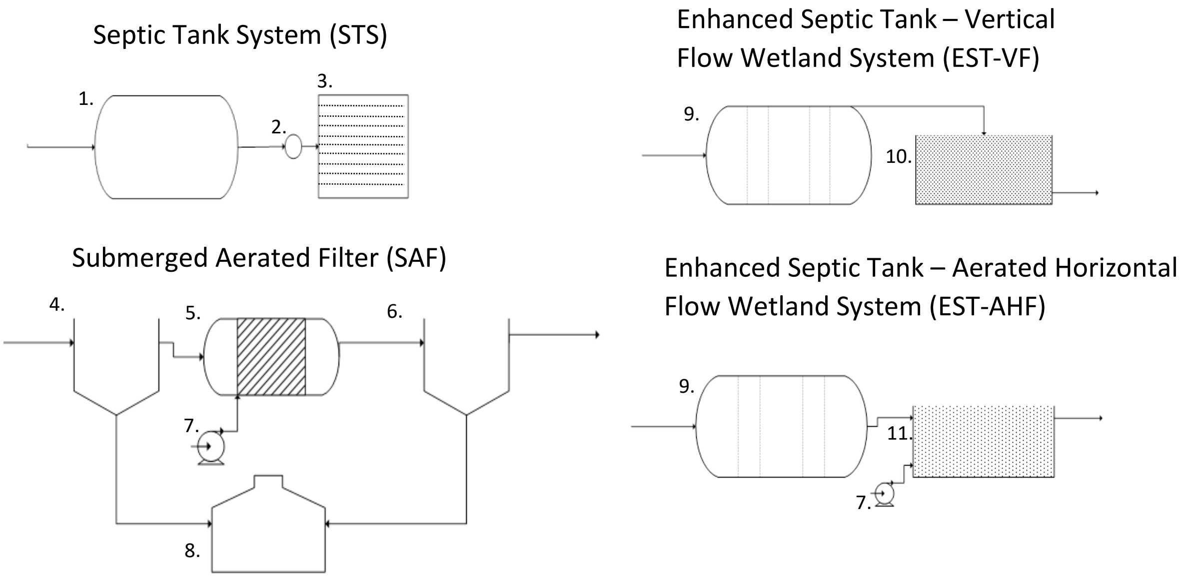
Septic tanks systems (STS) represent a quarter of all sewage treatment systems worldwide (UN Habitat and WHO, 2021) with similar estimates for STS usage in the US and Europe (Withers et al., 2014). For instance, in Scotland alone, around 160,000 properties are connected to STS (O'Keeffe et al., 2015), with over 1200 operated by Scottish Water. STS consist of two primary elements: the septic tank and a drainage field or soak away (Withers et al., 2012). Septic tanks are simple tanks with or without baffles, that function as low-rate anaerobic reactors which have the primary function of solids removal and retention until the collected solids (sludge) is transported away for further treatment at another facility. Recommended practice is to desludge the tanks annually (HM Government, 2015) although logistical constraints often mean they are desludged more frequently. A drainage field consists of a series of perforated pipes, atop a bed of media that distributes the septic tank effluent allowing it to slowly percolate through the drainage media and into the soil underneath (Figure 6.1).

There is growing consensus that the impact of STS on water quality is often understated (Dudley and May, 2007) and it is estimated that over 80% of STS in the UK are working inefficiently due to their age (May et al., 2015). Expected effluent quality after the septic tanks for a well functioningesti system is a total suspended solids of 80 mg l^{-1}, biological oxygen demand of 90 mg l^{-1}, with the ammonia remaining unchanged from the feed or slightly increasing with a typical level of around 35 mg l^{-1} (Table 6.1). Recent findings indicate that in rural communities STS disproportionately effect surface water nitrogen, in comparison to fertiliser usage (Halliday et al., 2014). For instance, ammonia concentrations in ground and surface waters have been shown to more than double downstream of STS (Herren et al., 2021; Withers et al, 2012). In addition, the anaerobic storage of sludge results in release of methane emissions with an estimated rate of 11 gCH_4 PE^{-1} d^{-1} equating to 0.11 tCO_2e y^{-1} over the 100 year time horizon and 0.34 tCO_2e y^{-1} over the 20 year time horizon

(European Commission, 2022). The regular frequency of desludging also adds further emissions and generates a nuisance impact on local residents. Accordingly, there is a growing need to update the flowsheet for onsite treatment beyond STS to one that improves treatment, reduces greenhouse gas emissions and reduces the frequency of desludging visits. The second ambition aligns to pledges made by the water industry to reach net zero either as full net zero emissions by 2040 (Scottish Water, 2020) or operational emissions by 2030 (Water UK, 2020a). Beyond the main aspirations there is an appetite to deliver co-benefits to extend the value proposition that any new solution can provide in terms of environmental or societal benefits.

6.1.1 Alternative Flowsheets

Two alternative approaches are considered (Figure 6.1). In the first approach, the primary stage is an enhanced septic tank (EST), which combines concepts from septic tanks, anaerobic ponds, and anaerobic baffled reactors (Chapter 4). The system uses baffles to increase digestion thus extending the time between desludging. The effluent from the septic tank is further treated in an aerobic nature-based solution configured as either a passively aerated vertical flow subsurface wetland (VF) or a forced aerated horizontal flow subsurface wetland (AHF) (Figure 6.1). Both are established technologies used in wastewater treatment and can deliver considerably enhanced effluent quality (Butterworth et al, 2016a). For instance, AHF wetlands have been reported to deliver median effluent qualities of 4 mg l^{-1} TSS, 4 mg l^{-1} BOD and 0.2 mg l^{-1} NH_3 when used as a tertiary treatment and 20 mg l^{-1} TSS, 5 mg l^{-1} BOD and 0.8 mg l^{-1} NH_3 when used as a post primary sedimentation treatment (Butterworth et al 2016b). Produced methane is collected for reuse or flaring to avoid methane emission into the atmosphere. Recent pilot trials of the NBS post an EST revealed the median effluent of the NBS post EST to be 5 m l^{-1} TSS, 11 mg l^{-1} BOD and 0.5 mg l^{-1} NH_4 and 12 m l^{-1} TSS, 8 mg l^{-1} BOD and 3 mg l^{-1} NH_4 for the VF and AHF respectively.



1. Septic Tank 2. Distribution System 3. Drainage Field 4. Primary Settling Tank 5. Submerged Aerated Filter 6. Secondary Settling Tank 7. Air Blower 8. Sludge Holding Tank 9. Enhanced Septic Tank 10. Vertical Flow wetland 11. Aerated Horizontal Flow wetland

Figure 6.1: Illustration of wastewater treatment flowsheets analysed

The other option is to use a commercially available package treatment process based on a submerged aerated filter (SAF) to provide aerobic treatment (Figure 6.1). The system consists of four primary elements: a primary settling tank, an aerated filter, secondary settling tank and sludge holding tank (Figure 6.1). The sludge from the primary and secondary settling tanks are stored in a sludge holding tank which requires desludging similarly to a septic tank. The certified effluent quality of package treatment SAFs of the style used in this study is 16 mg l^{-1} TSS, 11 mg l^{-1} BOD and 8 mg l^{-1} NH_4 (Tricel Environmental, 2022).

The aim of the current paper is to establish the financial and carbon basis for updating from septic tank systems to alternative flowsheets for population equivalents between 5 and 1000. To achieve this the four systems were designed at different population equivalents and the wholelife cost (WLC) and lifecycle carbon emissions (LCE) calculated. To aid in this goal the flowsheets will be considered from two case study perspectives: an old site and a new site. In addition, the paper will also consider the potential societal and ecological impacts of these flowsheets to provide an overall picture of the potential for using EST-NBS based systems.

6.2 Methodology

6.2.1 Design Case Studies

The first case study considered an existing site in which an STS is being replaced. The site aims to represent a ‘normal’ site, with dry soil conditions and a soil percolation rate (v_p) of 50 mm min^{-1}. The site is situated 40 miles from a large works where desludging tankers and site visits originate from. The site has sufficient topography for gravity feeding of the below ground assets.

The second case study considers the construction of a treatment system with equivalent ground and geographical conditions but is a new site. Therefore, the construction of auxiliaries such as security fencing and infrastructure such as roads need to be considered. Both case studies examined 15 population equivalent size flowsheets; 5, 10, 20, 30, 50, 75, 100, 150, 200, 300, 400, 500, 600, 800 and 1000 PE to assess the impact scale has on the significance of the key design choices.

6.2.2 Flowsheet Influent and Consent

British water code of practise flows and loads (British Water, 2013) were used for the influent loading of the flowsheets (Table 6.1). The design flow of the flowsheets is 0.375 m^3 d^{-1} PE^{-1} and a peak flow of 0.6 m^3 d^{-1} PE^{-1} equating to 3 dry weather flow (DWF) (Supplementary Material 6.6.1). The flowsheets for improved rural treatment works were designed to reach a secondary treatment level (Table 6.1). The comparative septic tank effluent is not discharged directly into the environment but through a drainfield, making the quantification of the final effluent challenging. Given that STS are unconsented systems (Withers et al., 2012) the STS flowsheets were not designed to treatment consent but instead government recommendation (Scottish Government, 2019) to reflect current practice.

Table 6.1: Flow and consent target for improved rural wastewater treatment flowsheets

Parameter	Influent Loading (g PE^{-1} d^{-1})	Effluent Consent Concentration (mg l^{-1})	Septic Tank Effluent (mg l^{-1})
Total Suspended Solids (TSS)	80	25	80
Biological Oxygen Demand (BOD_5)	60	25	90
Chemical Oxygen Demand (COD)	150	-	160
Ammonia (NH_4-N)	8	15	35

6.2.3 Flowsheet Design

STS in this study are designs to represent the current design flowsheet in the septic tanks were assumed to have a minimum hydraulic retention time (HRT) of 12 hours at peak flow, as per Scottish Water design standard. The tank is installed below ground as per the manufacturer's recommendations. All submerged tanks in this study were assumed to have a maximum design volume of 65 m^3. The drainage field is sized according to the Scottish Government guidance (2019) giving an area of 12.5 m^2 PE^{-1} for a field with a v_p of 50 mm min^{-1}. The drainage field is assumed to have a maximum length of 20 m (Scottish Government, 2019) and be constructed from perforated 13 mm low density polyethylene piping. The pipes are assumed to be installed in 0.8 m deep and 0.5 m wide trenches filled with a media of 4-10 mm pea gravel (Scottish Government, 2019). The trenches were assumed to be 2 m from any other drainage trenches.

The SAF package treatment plants were selected from available commercial units based on treatment hydraulic loading rates and that certified effluent was below the target range (Premier Tech Aqua, 2017; Tricel Environmental, 2022). The flowsheets were assumed to be installed as per the manufacturer's recommendation (Tricel Environmental, 2022). The effluent of the SAF is assumed to be discharged to surface waters.

The EST was designed to have a minimum HRT of 48 hours at peak flow, with 4 internal baffles (Chapter 4). EST systems were assumed to be constructed through earthworks, with a lined pit and soft cover made from synthetic rubber, using concrete posts for baffling. The wetlands were designed to current standard design guidance (Dotro et al., 2017) to mirror current practice in the UK (Butterworth et al., 2016b). The VF wetland area was determined to ensure they had a peak flow hydraulic loading of 0.12 m^3 m^{-2} d^{-1} with a square area and with a treatment depth of 0.8 m and freeboard of 0.3 m, the maximum length of a single VF wetland is 25 m. The media comprised of a main treatment layer of 55 cm of sharp sand and 10 mm shingle mixed at a 2:1 ratio with the final 5cm layer of 10 mm shingle. The treatment layer sits above a drainage layer of 20 -40 mm gravel for the bottom 15 cm and a 5 cm transition layer of 10 mm shingle (Jenkins, 2017). Distribution and collection pipes, constructed from PVC piping, are positioned every 1.25 m on the surface and bed of the wetland with 0.5 x 0.5 m concrete splash plates placed every 1.25 m (Jenkins, 2017). The system is dosed 8 times a day with 5 minutes doses delivered by a bell siphon within the EST.

The AHF wetlands design is based on a standard horizontal subsurface flow wetland used for tertiary treatment in the UK but with the adaptation of having a coarse bubble aeration grid installed on the flow of the wetland as is current practice (Butterworth et al., 2016b). Accordingly, the wetland was sized with a maximum treatment width of 25m and organic loading below 15 g_{BOD} m^{-2} d^{-1} with an air flow of 0.26 m^3 PE^{-1} h^{-1} (Butterworth et al., 2016b). The AHF was assumed to have a treatment height of 0.6 m and a freeboard of 0.3 m containing 4-10 mm pea gravel as the main treatment media and distribution layers of 100 mm stone for the first 0.5 m from the inlet and outlet, held in place with steel gabions.

Fences were assumed to be 1.8 m steel palisade fence posts with a 3 m wide access gate (Water UK, 2020b). Road infrastructure design was outsourced to quoting contractors consisting of two road designs, gravel and bitumen. Gravel roads were used when desludging visits were less than every 3 years (Jefferson, 2022) otherwise bitumen roads were assumed.

STS are not monitored beyond desludging. In contrast, the alternative flowsheets were assumed to be monitored monthly with electrical components inspected yearly. Replacement frequencies were assumed to be yearly for the filters, every 3 years for silencers, 5 years for bearings and 15 years for non-return and pressure relief valves. The operator time is contracted as £35 h^{-1} with inspections lasting one hour. Desludging was assumed to be conducted by diesel tankers, with the volume of sludge removed assumed to be 33% of the total tank volume for STSs, ESTs and SAFs. Inspection and maintenance vehicles are modelled as diesel vans with travel originating from a large works site.

6.2.4 Economic Assessment

Costs were calculated at June 2021 values, with costs adjusted by consumer price index (ONS, 2022). Flowsheets were calculated through component costing, with each element worth more than £50 quoted in triplicate. Where off the shelf products were not available cost curves were used to determine the value of the component. The system lifetime of the project was assumed to be 30 years (Gallagher and Gill, 2021). The land usage of sites in the old site case study all have a lower footprint than an STS so the purchasing of land is not considered. Due to the variability of pricing and lots available it is also not factor in for new site studies.

The net present value (NPV) of operational expenditure was calculated using;

$$NPV = \frac{R_t}{(1+i)^t}$$

Where R_t is the cash flow at time t, t is the year of operation and i is the discount rate. The discount rate represents the balance of interest and returns on investments against inflation, an average discount rate of 3.5% was used based on UK NET Zero and HM Treasury Greenbook guidance (Water UK, 2020a).

6.2.5 Lifecycle Analysis (LCA)

The Scope of the pledges given by the UK water companies for 2030 and 2040 are primarily Scope 2 with Scope 3 emissions only considered when a core activity is outsourced (Water UK, 2020a). Scope 2 emissions are defined as both the direct emissions from operation and construction of the flowsheet and indirect emissions from the consumption of electricity, heating or cooling for the operation of the flowsheet but not produced by the flowsheet (Greenhouse Gas Protocol, 2014). The LCA is analysed from cradle to grave, the system boundaries are outlined in Figure 6.2, (Resende et al.,2019). Embedded carbon and end of life calculations were conducted using OpenLCA 1.11.0 (GreenDelta), with the Ecoinvent v3.8 database (Ecoinvent). The lifecycle impact assessment model used for this study was ReCiPe 2016 Midpoint (H).

Flowsheet process emissions are derived by best available data in the literature (Table A 6.1) and biogas flaring was modelled as flaring of sour waste natural gas with biogenic carbon dioxide release from the Ecoinvent v3.8 database. Desludging tankers were modelled as unladen on the outward journey and the additional sludge weight is assumed to have a density of 1000 kg m^{-3}. Desludging tankers were modelled as the average carbon emissions per tonne mile of Euro 3 through to 6 lorries. Three lorry weight classes are considered, 3.5-7.5, 7.5-16 and 16-32 tonnes will be modelled assuming capacities of 2 m^3, 10.5 m^3 and 19 m^3 respectively. The carbon emissions of UK grid electricity up to 2040 are based on UK department for Business, Energy and Industrial strategy projections (2019). Post 2040, electrical grid emissions are based on sum of root square error with an exponential decay model, modelled using MATlab (Supplementary Material 6.6.2).

Uncertainty of the capital emissions was calculated using a Monte Carlo method with 1000 iterations (Heijungs, 2020), using the uncertainty within the Ecoinvent database. Due to the large variability in reported process emissions for each system, the uncertainty in process emissions was considered qualitatively.

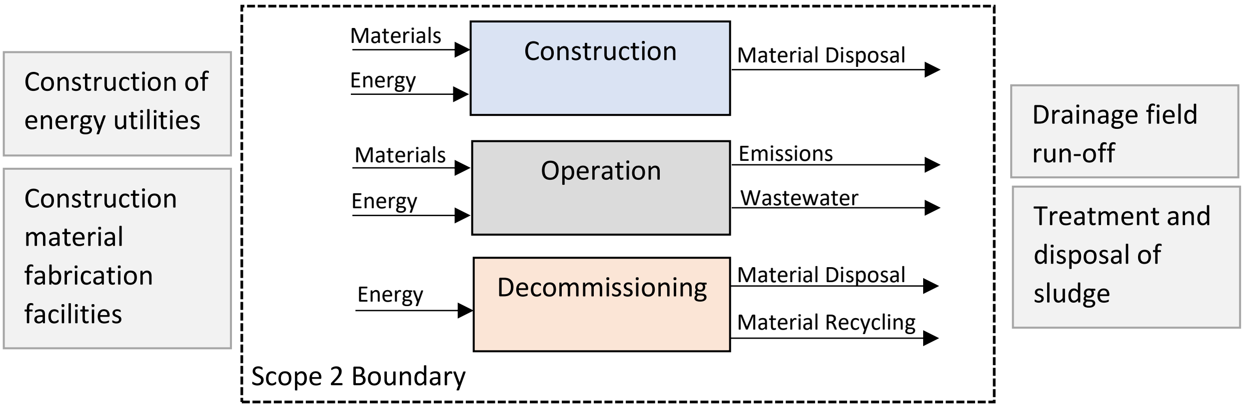


Figure 6.2: LCA system boundaries

6.3 Results and Discussion

6.3.1 Existing Site Study

6.3.1.1 Wholelife Cost (WLC)

At all examined populations the SAFs have the highest WLC per PE (Figure 6.3). To illustrate, in the case of 10 PE, the WLC of the SAF was £3160 PE^{-1} compared to £1100 PE^{-1} for the STS and £1650 PE^{-1} and £1750 PE^{-1} for the EST-VF and EST-AHF respectively. The difference in WLC per PE between the different options decreased as the scale increased due to economies of scale reducing the impact of fixed costs such as operator maintenance time. For example, at a PE of 1000, the WLC per PE was £748 PE^{-1}, £439PE^{-1}, £675 PE^{-1} and £370 PE^{-1} for the SAF, STS, EST-VF and EST-AHF respectively.

Comparison across the different alternative options shows that the EST-NBS flowsheets are a lower cost upgrade to STS than SAFs (Figure 6.3). The WLC per PE decreased for all options but at different rates such that the WLC per PE for the EST-AHF and the STS reached parity at sizes of 150 PE and above. The critical components that delivered the cost saving per PE in the EST-AHF were the operating costs of the AHF compared to the operating components in the STS as the capital cost per capita are approximately equal. Operating costs were 81% higher than those for the STS at small scales (Figure 6.3) but the difference decreased with increasing scale such that at 1000 PE the operating costs of the STS and EST-AHF are within 4%. The principle operating cost of the STS is desludging, which decreases from £776 PE^{-1} to £226 PE^{-1} as the scale increases from 10 to 1000 PE. In comparison, the main operating costs for the EST-AHF is the operating and maintenance costs associated with the AHF which decreases from £1350 PE^{-1} to £172 PE^{-1} from 10 to 1000 PE.

Similarly, the WLC per PE of SAFs and EST-VFs is comparable beyond 150 PE but higher than the STS. For instance, at 1000 PE the SAF and EST-VF represent an additional 70 and 54% WLC compared to the STS respectively. However, sub-150 PE, the EST-NBS flowsheets represent a notable saving compared to SAFs and hence offer a lower cost option for enhancing treatment effectiveness compared to a STS. To illustrate, at 100 PE the WLC of an EST-VF and EST-AHF are £786 PE^{-1} and £613 PE^{-1} compared to £1080 PE^{-1} for a SAF. Whilst the EST-VF is a more expensive option than the EST-AHF at larger scales, below 30 PE the EST-VF is the least cost option for improved treatment. This reflects the low operating cost of the VF system. At 10 PE the VF has a lifetime operating cost 25% lower than the AHF, however as the economies of scale reduce the impact of fixed factors such as operator transport and time, the increased capital cost of the VF system means that the EST-AHF is a lower cost alternative at larger scales.

Analysis of the cost breakdown reveals that the operating costs dominate the WLC at smaller scales but become a more minor component as scale increases reaching approximate parity with CAPEX at a scale of 30 PE, 150 PE, 75 and 1000 PE for the SAF, STS, EST-VF and EST-AHF respectively. To illustrate the contribution of operating costs, at a scale of 10 PE, was 79% for the SAF, 71% for the STS, 81% for the EST-AHF and 66% for the EST-VF (Figure 6.3). In comparison, at 1000 PE the OPEX contributes 30% of the WLC per PE for the SAF and 31% for the EST-VF.

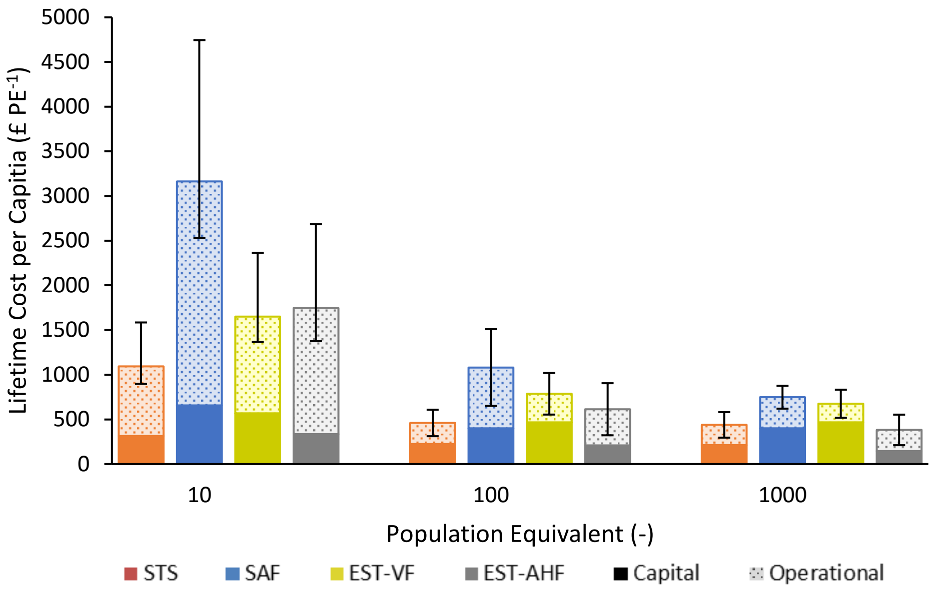


Figure 6.3: 30-year WLC per capita for existing site flowsheets at 10, 100 and1000 PE (error bars show discount rate from 0 to 6%)

The change in WLC per PE can be utilised to derive an economy of scale exponent assuming a standard power law fit. The different options group into two sets with the exponents of 0.61 and 0.66 for the SAF and EST-AHF and 0.80 and 0.82 for the STS and EST-VF. This compares to a to standard value of 0.66 reflecting the dominance of civil structures on the overall costs for the SAF and EST-AHF. Whereas the higher values of the STS and EST-VF indicate towards more modular components similar to the exponents reported for membrane-based systems (Jefferson, 2022). The land requirement of the proposed systems are smaller than that of current drainfield requirements. Current design guidelines for Scottish Water require that drainage fields are fenced off meaning the purchasing of additionally fencing is also not required. The SAF flowsheets has a greater difference in land footprint to that of the STS compared to the EST-AHF and EST-VT, with greater provision for use such as for solar panels.

6.3.1.2 Lifetime Carbon Emissions (LCE)

A different perspective of the flowsheets is observed when viewed through the lens of the lifetime carbon emissions (LCE). The EST-NBS options both reduced LCE substantially compared to either the SAF or the STS at all scales of operation (Figure 6.4). To illustrate with respect to a 10 PE scale, the LCE was 1740 and 1870 kg $CO_{2\ eq}$ PE^{-1} for the EST-VF and EST-AHF respectively and this compared to 3150 and 4190 kg $CO_{2\ eq}$ PE^{-1} for the SAF and STS respectively. Accordingly, adoption of the alternative options reduced the LCE compared to that of the existing STS by between 56 and 90% for the EST-AHF across all scales. The equivalent reduction across all scales was between 60 and 86% for the EST-VF and between 26 and 71% for the SAF. In terms of the EST-NBS options, the VF option has a lower LCE at small scales and reaches parity with AHF option at a scale of 100 PE. Beyond that the EST-AHF provides the lowest LCE as the capital components decreased more significantly with scale than that of the EST-VF. The reduction in capital emissions for an EST-VF is 60 kg $CO_{2\ eq}$ PE^{-1} across the scale of 10-1000 PE whereas the equivalent change for the EST-AHF is 160 kg $CO_{2\ eq}$ PE^{-1} (Figure 6.4)

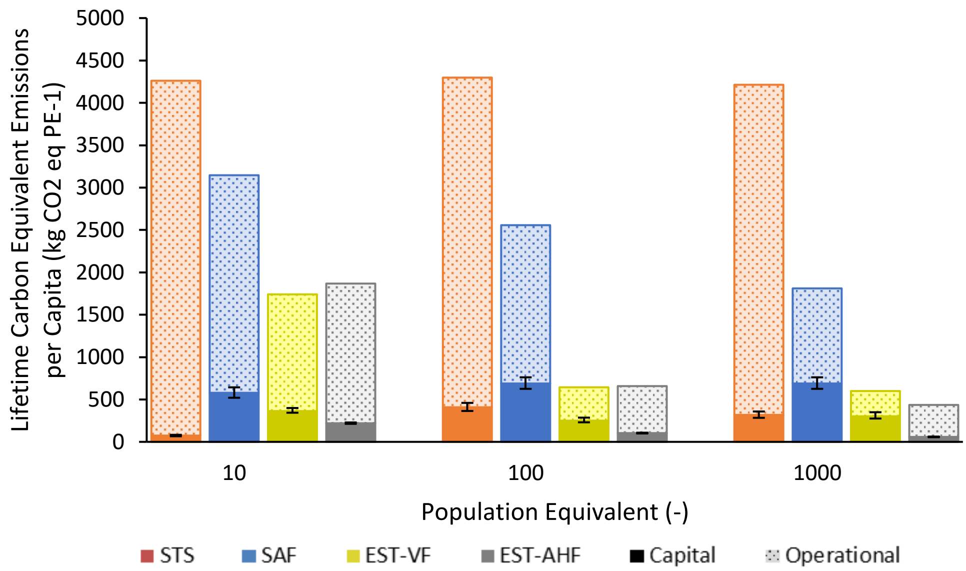


Figure 6.4: 30-year LCE per capita for existing site flowsheets at 10, 100 and 1000 PE (error bars show 95% confidence intervals from Monte Carlo analysis)

The LCE for the STS has a negligible scale impact ranging from 4260 kg $CO_{2\ eq}$ PE^{-1} at small scale to 4210 kg $CO_{2\ eq}$ PE^{-1} at larger scales due to the fact that the majority of the LCE is derived from process emissions of methane. This is in line with other studies into the lifetime carbon footprint of septic tank flowsheets which indicated that process emissions account for 90% of the total LCE

(Resende et al., 2019) compared to 85% in the current study. Whilst the STS directly vents methane into the atmosphere, the other options either prevent its formation by utilising aerobic conditions (SAF) or capture it (EST-VF; EST-AHF). The impact is to reduce the LCE associated with process emissions from 129 kg CO_{2eq} PE^{-1} yr^{-1} for a 100 PE in the case of an STS to 0.98, 0.88 and 0.80 kg CO_{2eq} PE^{-1} yr^{-1} for SAF, EST-VF and EST-AHF at the same scale. Over the 30-year lifetime of the assessment, the capture of methane in the EST-NBS systems avoids 3.8 t CO_{2eq} PE^{-1} compared to the STS, this is the equivalent to 3 flights from London to New York (Ecoinvent).

With respect to a centralised wastewater treatment, consisting of an activated sludge system and anaerobic digester, the operational emissions per capita are approximately 82 kg $CO_{2\ eq}$ PE^{-1} yr^{-1} (Piao et al., 2016), which is equivalent to a SAF flowsheet and greater than either EST-NBS flowsheet which are between 10 and 54 kg $CO_{2\ eq}$ PE^{-1} yr^{-1}. The operational emissions of an STS system is greater than that of a centralised works on a per capita basis.

6.3.1.3 Uncertainty Analysis of the LCE

The uncertainty in the capital emission of the LCE is low (Figure 6.4) with no significant impact on findings. On the other hand, the determination of process emissions remains the greatest area of uncertainty in the estimation of the LCE due to a lack of reported emission rates and the overall consistency of approach. To illustrate, in the case of the SAF a value of 64 $gCO_{2\text{-}eq}$ $m^{-3}{}_{\text{wastewater treated}}$ was derived from nitrous oxide emissions for SAFs from UK carbon accounting guidance (UK WIR, 2008) and methane emissions from sludge holding tanks of IPCC (IPCC, 2014). However, in a recent study from China, emissions ranged from 264 to 443 $gCO_{2\text{-}eq}$ $m^{-3}{}_{\text{wastewater treated}}$ with an average of 333 $gCO_{2\text{-}eq}$ $m^{-3}{}_{\text{wastewater treated}}$ (Hua et al., 2022). The consequence is a potential increase in process emission of five times, based on average and nine times, based on the maximum level reported which results in an additional 4 kg $_{CO2\ eq}$ PE^{-1} yr^{-1}, however this increase does not significantly impact the findings of this study in terms of technology comparison (Figure 6.5).

In the case of the AHF, there is no reported data on emissions post an anaerobic process. The closest approximated reported case is for treatment of fish farm effluent with an emission rate of 1.62 kg $CO_{2\ eq}$ PE^{-1} yr^{-1} (Maltais-Landry et al., 2009), the methane flux is in line with that reported for domestic AHF systems (Wang et al., 2008) . Reported data does exist for the vertical flow systems with projected process emissions ranging from 0.16 to 4.6 kg $CO_{2\ eq}$ PE^{-1} yr^{-1} (Mander et al., 2014). Nitrous oxide accounts for 10% of the process emissions of VF wetlands and 100% of AHF wetlands. This compares to 83% for the SAF revealing that management of the nitrogen treatment pathway is critical in minimising process emissions from such flowsheets. The general indications of how to minimise nitrous oxide emission are linked to avoiding overloading, running as close to steady state as possible and ensuring sufficient oxygen is distributed through the biofilm (Kampschreur et al., 2009). Comparing the operation of the NBS wetlands, the AHF runs in a continuous mode opposed to the batch nature of VFs. However, despite the expectation of an AHF having a lower N_2O emission rate than a VF, currently available literature gives a nitrous oxide emission rate for an AHF as 1.12g N_2O kg $TN_{treated}^{-1}$ (Maltais-Landry et al., 2009) compared to 0.18 g N_2O kg $TN_{treated}^{-1}$ for a VF (Mander et al., 2014). Measurement and establishing conditions to minimise overall process emissions remains an area of need and a recommendation for future work.

Interestingly, the estimation of methane emissions from septic tanks is consistent across different studies, with numerous studies reporting a value of 11 gCH_4 PE^{-1} d^{-1} (Diaz-Valbuena et al., 2011; Huynh et al., 2021; Truhlar et al., 2016). This equates to an overall emission rate of 112 kg $CO_{2\ eq}$ PE^{-1} yr^{-1}

The impact of possible variation in emission rate was established for the four technologies through a sensitivity analysis (Figure 6.5). The uncertainty of the actual emission rate is unlikely to impact the relative findings of the study. For instance, the other emissions for the SAF mean that considering all improved flowsheets to have the same process emissions would have negligible impact of the study’s findings in terms of the SAF (Figure 6.5). Further, the uncertainty in the process emission between the two NBS options means that they can be effectively viewed as similar. Overall, the EST-NBS systems can be viewed as able to deliver the lowest LCE of the four options considered irrespective of the uncertainty. For instance, for an EST-AHF to have a greater carbon footprint than SAFs this would require an additional 780 kg $CO_{2\text{-eq}}$ PE^{-1} at 1000 PE.

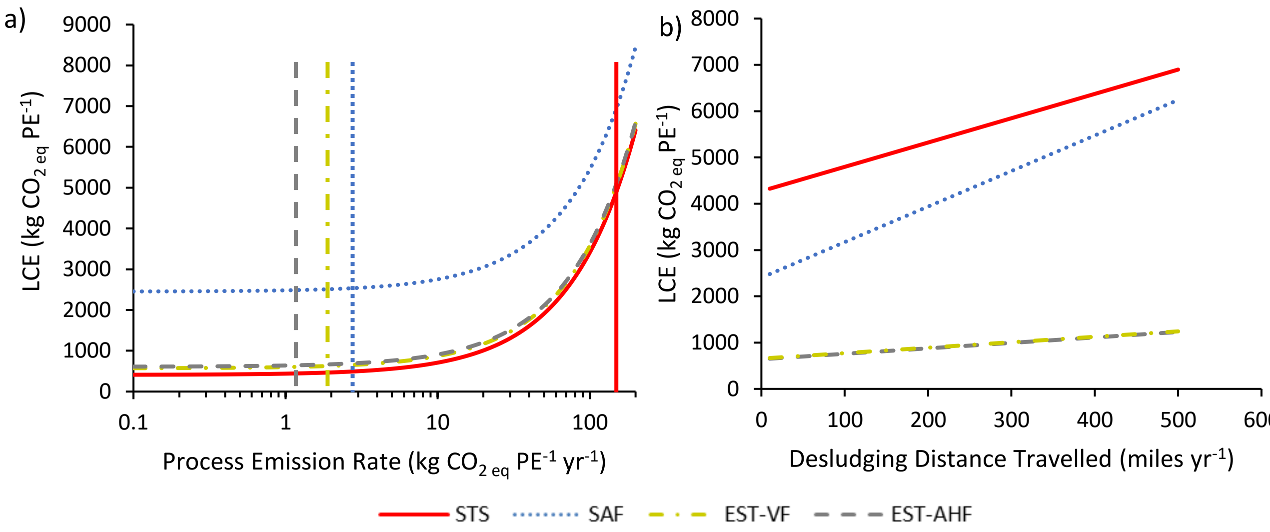


Figure 6.5: Impact of a) process emission rates and b) desludging miles travelled on LCE for the flowsheets at 100 PE (Vertical lines on Figure 6.5a represent the values used in the current study)

The other area of major uncertainty is related to the desludging frequency and distances travelled. The desludging frequency of the packaged treatment systems and the septic tanks is approximately 1 year, and this compares to the EST-NBS systems of 7 years. However, the desludging frequency of water utility operated septic tanks is known to vary from 3 to 24 months and is commonly accomplished through an integrated schedule to desludge multiple septic tanks in individual tankering trips. The contribution desludging makes to the operational emissions is 1%, 8%, 1% and 1% at 100 PE with a distance of 40 miles for the STS, SAF, EST-VF and EST-AHF respectively. If both the distance travelled and desludging frequency were double that of the modelled systems (or equivalent) there would be a 13%, 30% and 19% increase in the LCE of the STS, SAF and EST-NBS flowsheets respectively. This could be the case for particularly remote sites such as flowsheets situated on small islands where tankers are required to travel back to the mainland. However, although the desludging can have a significant impact on the LCE it does not impact the findings with respect to flowsheets with reasonable operational variation (Figureb). For example, with a travelling distance of 40 miles per trip it would require desludging every 60 days for the EST-NBS flowsheets to have an equivalent LCE to a SAF.

6.3.1.4 Abatement Potential

Over the predicted life span of 30 years, the relative cost of upgrading to an EST-VF at 10 PE is £560 PE^{-1} which produces a lifetime carbon reduction of 2520 kg $CO_{2\ eq}$ PE^{-1}, giving an abatement cost of £220 $tCO_{2\ eq}^{-1}$ over 30 years. Comparatively at 10 PE, the abatement cost of a SAF and EST-AHF is £1860 $tCO_{2\ eq}^{-1}$ and £260 $tCO_{2\ eq}^{-1}$. At larger populations, the abatement costs of the alternative options is reduced as the price differential from an improved flowsheet to STS decreases whilst the carbon reduction increases. For example, at the 1000 PE scale, the abatement cost of a SAF, EST-VF and EST-AHF are £92 tCO_2^{-1} and £65 tCO_2^{-1} and -£17 tCO_2 eq^{-1}. The negative number represents a cost-effective investment, providing an operational saving alongside carbon reduction, for example, a reduction in travel both saves money and fuel and the associated carbon emissions. The significances of the estimated abatement costs can be considered by framing the number within the general abatement costs published for selected technologies for the water sector (Table 6.2). Overall, this demonstrates that the proposed alternatives can provide a cost-effective means of carbon abatement whilst delivering a higher level of treatment. To illustrate, at 1000 PE an EST-AHF has a lower lifetime cost and emissions than STS meaning the abatement cost is -£17 tCO_2 eq^{-1}. This can be considered a cost-effective carbon reduction investment, in par or better than the investment/benefit associated with biomethane to grid injection schemes at larger treatment works. Whilst the scale of impact are not akin, this study showed there is a clear business case for investing in STS upgrading.

Table 6.2: The carbon reduction and the abatement cost of selected technologies for the UK Water Sector (Water UK, 2021)

Technology	Details	Abatement Cost (£ tCO_{2eq}^{-1})
Transport Reduction	Reduction in fleet travelling	-100 to -500
Solar Power	Reaching 20% of annual energy consumption via Solar Power	-100 to -500
Wind Power	Reaching 20% of annual energy consumption via Wind Power	-100 to -500
Biomethane	Grid injection of Biomethane	>10
Alternative Anaerobic Digestion	Upgrading conventional anaerobic digestion to thermal hydrolysis	>100
Natural Sequestration	Commitment to plant 11 million trees by 2030	>1000

6.3.2 New Site Study

In the case of new sites, additional consideration needs to be given for infrastructure costs associated to road and fencing. For instance, if the site requires regular desludging visits such as with the STS and SAF, a bitumen road is required (Jefferson, 2022). Whereas if the desludging frequency is less than every 3 years, then gravel road is considered suitable (Jefferson, 2022). The impact is seen through the difference in cost and emissions of the two options. Bitumen based roads have a cost of £130 m^{-1} and emissions of 66 $gCO_{2\,eq}$ m^{-1} $lane^{-1}$ compared to £26 m^{-1} and emissions of 4.9 $gCO_{2\,eq}$ m^{-1} $lane^{-1}$ for gravel-based roads (Espinoza et al., 2019). Consequently, there is a differential cost of £104 m^{-1} and an emission of 61.1 $gCO_{2\,eq}$ m^{-1} $lane^{-1}$. The exact impact will depend on the specific distances associated with a given site but applies for all options. Consequently, the cost and emissions for the NBS system will always be lower, further reinforcing the positive impact compared to a SAF or STS observed for the existing site case study.

The other major additional cost is associated with fencing which is dependent on the land footprint of the system. The footprint of the different options at 100 PE was 1260 m^2, 50 m^2, 589 m^2 and 360 m^2 PE^{-1} for the STS, SAF, EST-VF and EST-AHF respectively. This converts to a fencing cost of £10,000 for STSs and £2600, £4100 and £6000 for SAF, EST-VF and EST-AHF respectively. The lower costs for the EST-NBS over the STS represents the cost associated with the drainage field. This equates to a 22% increase in the lifetime costs of an STS, however for the improved systems the impact is lower being 3%, 8% and 7% for SAF, EST-VF, and EST-AHF respectively. Overall, fencing does not significantly change the assessment of costing compared to an existing system. For instance, at 100 PE an STS remains the lowest cost system, followed EST-AHF, then EST-VF and finally SAF.

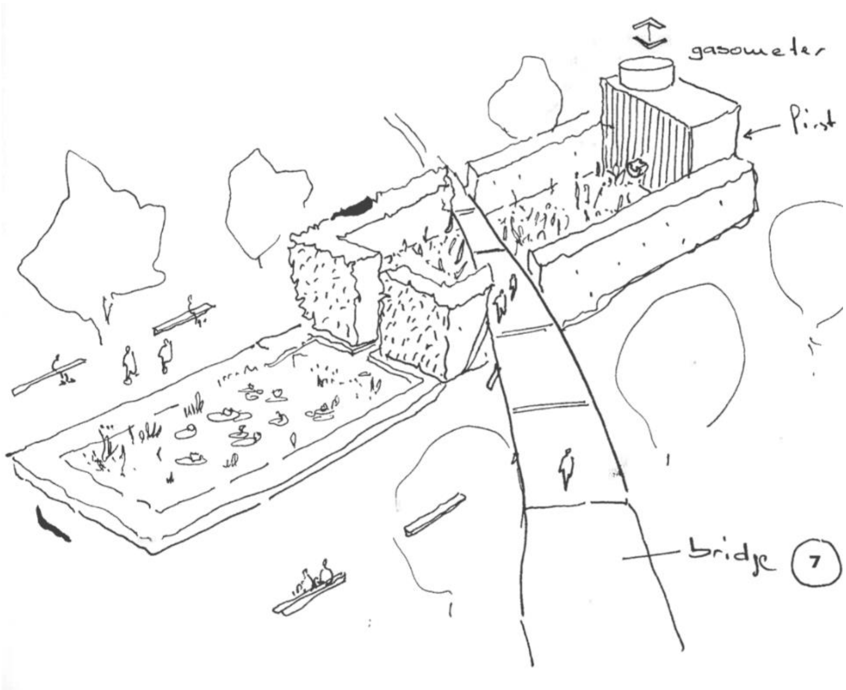
In more rural sites there is the possibility that an electrical connection is not possible. The SAF and EST-AHF are dependent on an electrical connection to operate, compared to an STS and EST-VF which can operate passively. The cost of installing a main electrical connection is £13,000 for a small business and can rise to up to £50,000 (Northern Power Grid, 2022). Onsite power generation could also be used but it is likely that a redundancy connection would be required. Therefore, in these scenarios, the adoption of a passive system such as the EST-VF would likely be beneficial.

6.3.3 Future Outlook

A key to the lower LCE for the EST-NBS options is the fact that no methane is vented into the atmosphere. However, if the gas from the EST systems were vented then the overall LCE would be unfavourable compared to the STS and show the importance of gas management. Instead of flaring, the produced methane could be utilised to provide heating for the EST or used within the local community providing the potential for further carbon offsets. A similar proposal for a sewer mining project to irrigate a park in Australia proposed the use of the gas produced to power several BBQ rings to enable people visiting the park to cook their lunch (Jefferson, 2022). Whilst there is currently insufficient knowledge to assess the feasibility of the different options, it provides an illustration of approaches to add additional value from the EST-NBS flowsheets.

Within this study the STS is considered as a baseline, currently operating reference frame. However as highlighted in Figure 6.4, the majority of the LCE of an STS are from the escape of methane, the installation of a flare or capture for this methane would reduce the carbon impact of STS by approximately 90%, at a much lower cost than the construction of a new works. This could be an option for reducing carbon impact in areas with less critical effluent quality requirements and where the current STS still has significant asset life before an intervention is needed. However, the benefits associated with lower environmental impact in terms of effluent quality would remain with the new flowsheets.

Further, the above assessment has been based on standard practice and has assumed a standard wetland design has been used. However, the adoption of the NBS component offers further opportunity concerning co-benefits such as biodiversity gain and social benefits such as improving mental health and resilience associated with people interacting with green infrastructure. Accordingly, the wetlands could be designed to be open to the public or protected using green security fencing solutions such as Hawthorn, which provide an analogous level of security to that of palisade fencing. The costs implication is an increase from £110 m^{-1} for steel palisade fence posts to £140 m^{-1} for Hawthorn. However, hedge rows have the potential for significant carbon storage (Axe et al., 2017), coupled with the improved aesthetic value this is an adjustment to flowsheets which warrants further investigation. Taking the concept further, the wetlands can be designed for enhanced aesthetic value with access for the public to spend time by the wetlands (Figure 6.6). This has the potential to reposition the treatment works in relation to the local community to add social value and engagement such that low level maintenance could be conducted by members of the local community. Whilst such considerations require development, they offer additional benefits to the basic items considered in this study and hence further enhance the case for adoption. Adaptations to the current flowsheet may be required to maximise the social utility of the space, for example, the LCA considers the utilisation of biogas flaring. Although this provides a good reduction in lifetime carbon emissions this might not be viable for a social space due to perception and safety risks. For the utilisation of



social space further consideration of the gas management is required, for example collection or offsite usage.

Figure 6.6: Treatment garden concept (Jefferson, 2022)

6.4 Conclusion

Nature based solutions to rural wastewater provide a lower cost method of improving treatment of septic tank systems compared to SAFs at both new and existing sites. In addition to being a cost-effective solution the NBS flowsheets have lower carbon emissions than more intensified SAF at populations from 5 to 1000 PE. The work shows the potential that additional focus on rural treatment can have on the carbon emissions of the UK water sector, comparable in cost effectiveness to upgrading large anaerobic digesters and biomethane grid injection. This is coupled with improved water treatment, reducing the eutrophication burden which STS impose. There is scope to expand the flowsheet to provide a social space and societal gain, alongside treatment and carbon reduction.

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6.6 Supplementary Material

6.6.1 Dry Weather Flow Calculation

Dry weather flow (DWF) is calculated based on;

$$n.DWF = n.Q_{PE} + IQ_{PE}$$

Where n is the DWF number, Q_{pe} is the daily wastewater contribution per capita (m^3 d^{-1} PE^{-1}) and I is the infiltration percentage. Flow is calculated where Q_{pe} is 0.15 m^3 d^{-1} PE^{-1} and 100% infiltration rate. Therefore 3 DWF is 0.6 m^3 PE^{-1} d^{-1}

Table A 6.1: Flowsheet Process Emissions

Process	Emission	Reference
Septic Tank		
Nitrous Oxide (g PE^{-1} d^{-1})	0.005	(Leverenz et al., 2010)
Methane (g PE^{-1} d^{-1})	11	(Leverenz et al., 2010)
Drainage Field		
Nitrous Oxide (g PE^{-1} d^{-1})	0.15	(Leverenz et al., 2010)
Methane (g PE^{-1} d^{-1})	-0.3	(Leverenz et al., 2010)
Discharge to Water Source		
Nitrous Oxide (kg N2O kg N discharged^{-1})	0.025	(Foley et al., 2010)
Methane (kg CH4 kg COD discharged^{-1})	0.0025	(Foley et al., 2010)
SAF		
Nitrous Oxide (kg N_2O kg TN^{-1})	0.002	(UK WIR, 2008)
Methane (mg CH4 Pe^{-1} d^{-1})	57	(UK WIR, 2008)
VF Wetland		
Nitrous Oxide (kg N_2O kg TN^{-1})	0.00018	(Mander et al., 2014)
Methane (kg CH4 kg TOC_{in}^{-1})	0.0128	(Mander et al., 2014)
AHF wetland		
Nitrous Oxide (kg N_2O kg TN^{-1})	0.00112	(Maltais-Landry et al., 2009)
Methane (mg CH4 Pe^{-1} d^{-1})	0	(Wang et al., 2008)

6.6.2 Matlab Code

```
beta0=[146.33;0.039];
modelfun=@(b,year)(b(1)*exp(-(year)*b(2)));
opts = statset('nlinfit'); %Fitting model
opts.RobustWgtFun = 'bisquare';
% Non linear Regression

[Cost_calc,R,J]=nlinfit(year,emission,modelfun,beta0,"Options",opts);
ci=nlparci(Cost_calc,R,"jacobian",J);
Power_order=[Cost_calc(2,1),ci(2,:)];
Initial=[Cost_calc(1,1),ci(1,:)];
```

7 Implications of the Work

The overall aim of the thesis was to understand the impact of temperature and mixing on the processes that govern the sludge accumulation rate in septic tanks in order to develop alternative designs that could enable infrequent desludging. The associated ambition was to establish an alternative flowsheet based on this alternative design for onsite wastewater treatment that could improve treatment, reduce carbon footprint and provide the basis for additional added value. The key discoveries of this thesis that we generated in response to the stated aim were:

- The provision of the world's first data set to determine hydrolysis rate constant of sewage sludge in sub 20 °C conditions.
- Under sub 20 °C conditions the hydrolysis of unmixed sludge is controlled by particulate disintegration processes and is effectively modelled using a first order kinetic model.
- The inclusion of two vertical baffles in the front end of a septic tank provides sufficient mixing within the sludge layer to significantly increase the hydrolysis rate of the system.
- Beyond a hydrolysis rate constant of 0.02 d^{-1}, further increase in the hydrolysis rate constant provides limited benefit to the desludging frequency.
- The proposed flowsheet significantly reduces the life cycle emissions of onsite treatment compared to both septic tank systems (STS) and package treatment systems.

The following is a discussion of the implications of the findings framed from questions raised by the industrial sponsor during the project duration.

What is the prospect for extended desludging frequencies on septic tank sites in Scotland?

Enhancing existing septic tank designs through appropriate baffling delivers a sufficient increase in hydrolysis rate constant, even under low temperature conditions, to enable extended desludging frequencies of up to seven years.

In Chapter 4 the adjustment of baffling in septic tanks was found to increase the hydrolysis rate constant from 0.0089 to 0.035 d^{-1} at systems operating at a mean temperature from 14 to 18^{o}C. A system with a hydrolysis rate constant of 0.035 d^{-1} and a hydraulic residence time of 48 hours at 3 DWF is projected to have desludging frequency of 6.7 years. A key objective of the proposed flowsheet is to increase the maintenance and hence desludging frequency of the flowsheet to at most once every 5 years, however there are uncertainties remaining on this objective.

The operational and sludge temperature of a septic tank is unclear and dependant on numerous factors. Although a septic tank operates in ambient conditions, the feed domestic wastewater is often heated, and water has a high heat capacity for heat retention. Septic tanks operating in Ireland were found to have a liquid temperature ranging from 13 to 15^{o}C during winter (February) operation with air temperature between -3 and 14^{o}C (Mahon et al., 2022). The feed temperature was consistently above 15^{o}C which allowed for the septic tanks to have a higher temperature than the surrounding soil of 7 to 8^{o}C (Mahon et al., 2022). This is in-line with the temperatures recorded in Chapter 4, however the average winter soil temperature from 1981 to 2010 in Northern Scotland is 0 to 4^{o}C (Met Office, 2022) with the annual soil temperature being below 4^{o}C in some regions (Met Office, 2022). Although not the only factor of hydrolysis, temperature can have a significant impact on the hydrolysis rate constant of a system. In Chapter 3 a reduction in temperature from 15 to 5^{o}C saw a 41% reduction in hydrolysis rate constant.

The feasible range of septic tank operational temperatures in a temperate climate is likely to be between 5 and 15^{o}C, in this range adjusting the hydrolysis rate constant of the pilot system reported in this thesis more than doubles the hydrolysis rate constant (Table 7.1). Applying the mass balance model to a conventional septic tank indicates that a 33% sludge fill level will be reached after 2.6 years at 5ºC and 5.0 years at 15ºC. In comparison, the proposed EST design extents the expected fill time of the system to 5.7 years at 5ºC and to 6.6 years at 15ºC. Beyond a hydrolysis rate constant of 0.02 d^{-1} increases in the hydrolysis rate constant have been shown to have a limited impact (Chapter 4, Figure 4.8). Accordingly, if the annual temperature within the sludge layer remains above 10-15ºC, mixing will have limited impact on the resultant desludging frequency.

Table 7.1: Arrhenius adjustment of hydrolysis rate constant of septic tank and enhanced septic tank systems reported in Chapter 4

	Hydrolysis Rate Constant at T=5ºC (day^{-1})	Hydrolysis Rate Constant at T=10ºC (day^{-1})	Hydrolysis Rate Constant at T=15ºC (day^{-1})
Conventional Septic Tank	0.0042	0.0061	0.0089
EST-Mid	0.013	0.020	0.029
EST-High	0.013	0.020	0.029

The increased hydrolysis rate constant of the proposed EST design is attributed to hydraulic mixing within the sludge layer. Equivalence was observed during the Jar test experiments (Chapter 3) which showed that mixing at an equivalent temperature can have a more significant impact on the hydrolysis rate constant than running at higher temperatures. Importantly the work showed the hydrolysis rate constant when mixed at 10ºC was equivalent or better to the rate constant in an unmixed system at 35ºC (Chapter 3, Table 3.3). Additionally, modelling of particulate degradation in passive and low temperature systems, found that disintegration is the limiting stage of hydrolysis as opposed to biological establishment and activity. The importance of this is that adding enzymatic products will not be effective as the systems is not enzyme limited and so enhancement must come from a combination of mixing and temperature. However, once mixed addition of enzyme products may provide additional enhancement.

Within a passive system such as a septic tank although the material may be sufficiently active and the contact of the biomass with newly settled substrate rich media is likely to be limiting. This is congruent with observations of passive systems with a long operational time, such as anaerobic ponds, in which there is highly degraded lower sludge layer however higher sludge regions are less active and degraded (Papadopoulos et al., 2003). The introduction of mixing within the sludge blanket provides the conditions for the establishment of active biomass within all sludge layers.

The final element of reaching an extended desludging frequency is increasing the tank volume, the recommended volume of an EST is 48 hours at 3 DWF compared to a design basis of 12 hours at 3 DWF for a conventional septic tank. Elmitiwalli (2013) found that tank volume increase was the primary method for reduction in desludging frequency. To reach a 4-year desludging frequency in a conventional septic tank, Mahon et al., (2022) recommends a 4.5 m^3 septic tank for 4 PE, this equates to an HRT of 45 hours at 3 DWF. This is in line with the pilot operation and modelling of a conventional septic tank in this thesis. The addition of volume has a two-factor impact on reducing desludging frequency by providing more space for sludge to be stored, this increase in store time then allows for more digestion of the settled sludge further decreasing the desludging frequency. The prediction of a conventional septic tank reaching 4 years with an increased volume highlights the dual role the mixing and tank volume are playing in the projection of the decreased desludging frequency requirement of an EST.

How does the new proposed flowsheet compare to a traditional small sewage works?

The assessment of the proposed flowsheet in Chapter 6 considered comparison to both STS and package treatment plants. The scale of the investigation went up to 1000 PE which goes beyond the threshold for more traditional small-scale treatment works systems, typically viewed between 50-2000 PE). Small-scale treatment works account for 75% of all treatment in the UK and therefore warrant an assessment alongside the proposed flowsheet.

A traditional small-scale works consists of three principal elements, firstly a settler as primary treatment to separate the sludge and liquid streams. The settled sewage is subsequently treated by an aerobic process such as a trickling filter or rotating biological contactor and then post-secondary treated effluent is then subjected to final settlement prior to release. The settled sludge from the primary and final settlers is held in a storage tank and then transported to a major works for further processing. The transported sludge will most commonly be added to an anaerobic digester at a large works for biogas generation and subsequently treated by processes such as dewatering.

The consideration of the proposed flowsheet can be considered in two parts with the EST and CW. In comparison to a small work the EST is the equivalent to the primary settler. A clear distinction between an EST and a primary settler is the holding of the solids. A primary settler operates semi-continuously with the settled sludge regularly emptied and storage in a holding tank, conversely the EST is a fed batch system with regards to sludge and the sludge is held for an extended period. Therefore, the footprint of a primary settler is significantly reduced compared to that of an EST. However, this is countered by the requirement for regular desludging of the sludge holding tank, typically weekly, compared to that of an EST which is desludging between 5 to 7 years.

The cost implication of the desludging frequency is dependent on factors such as road construction. The frequent desludging of a traditional small works will require the construction of bitumen road compared to the gravel required for the proposed flowsheet. This can have a significant lifetime cost implications with a bitumen road being £100 m^{-1} more expensive than a gravel track (Espinoza et al., 2019). However, it is likely the additional costs of the construction of a larger holding system will outweigh the cost savings of reduced desludging, as this was found for system smaller than 1000 PE.

From a carbon perspective, the use of an EST over a primary settler will likely lower the lifetime carbon emissions (LCE). For a septic tank system and SAF system process emissions were significant contributors to the LCE, a sludge holding tank provides the conditions for anaerobic digestion and hence fugitive methane emissions. For STS this was the primary emission source, although sludge holding tanks have a lower methane production rate than septic tanks (UK WIR, 2008) this is a mitigated release using an EST. Secondly, reduced desludging will avoid the emissions due to tankering. Comparing the WLC and LCE of a primary settler tricking filter (TF) system to an anaerobic pond TF system for 2000 PE, Cruddas et al., (2016) found the increased infrastructure of the anaerobic pond meant that the 20-year WLC of the AP system was marginally higher. However, there was a significant reduction in the LCE due to reduction in fugitive emissions and tankering. This is in-line with the finding of this thesis and shows the potential for low-rate anaerobic reactors for small scale treatment works alongside onsite treatment flowsheets.

The sludge for a small works is typically transported to a large works for anaerobic digestion and energy generation. However, if the sludge from the EST is a low energy product, this could negate the utility of anaerobic digestion of the sludge as it has already undergone this process for an extended period. Hence the sludge would only require post anaerobic digestion processing. This could either be conducted at a large works or more locally to the site. As desludging is expected to occur every 5 to 7 years the construction of a sludge processing system on a small-scale site is unlikely to be beneficial, a low-tech solution such as 5 days of composting could potential be applied (DEFRA, 2018). This could further reduce the need for tankering and road construction whilst providing a utility of the treatment system more locally.

The use of constructed wetlands as secondary treatment is an established technology (Dotro et al., 2017) and capable of meeting the consent standards of a conventional small works, although both flowsheets might require tertiary treatment for more stringent consent limits (Dotro et al., 2017). The economy of scale of the SAF and EST-AHF flowsheets up to 1000 PE were found to be similar at 0.61 and 0.66, on this basis there is not expected to be a point in which the per capita whole life cost of an EST-AHF is greater than that of a SAF. This is not the case for an EST-VF flowsheet which has a n-component of 0.82 and therefore is expected to converge with the WLC of a SAF system at scale not much greater than 1000 PE. Therefore, for small scale works an EST-AHF is a more suitable flowsheet than EST-VF, with an EST-AHF being comparable to a SAF and predicted to have a lower WLC.

With regards to treatment levels, the pilot system operated in Cranfield wastewater treatment works. This works is a conventional small works with a trickling filter system. Over the operational period of the pilots the mean effluent quality for Cranfield treatment works was 26 mg TSS $l^{-1,}$ 8.4 mg BOD l^{-1} and 1.6 mg NH_4 l^{-1} which is a higher concentration of all 3 contaminants compared to the VF and AHF systems (Chapter 5, Table 5.3). The mean TSS effluent of the Cranfield treatment works is above the target 95th percentile effluent quality of the proposed flowsheet. Hence the constructed wetlands can at least operation to a similar treatment level as a conventional treatment works.

The secondary treatment of a small-scale treatment works alongside a SAF could also be a TF system. A TF is a more passive technology than a SAF where anaerobic effluent is slowly fed over a porous media in which a biofilm degrades the contaminates. The more passive nature of a TF means that the power requirements are equivalent to that of an AHF (Butterworth et al., 2013). However, when utilised for either tertiary of secondary treatment, the cost of a TF is expected to be higher than an AHF for small wastewater treatment plants (Butterworth et al., 2013). This is due to the increased land and construction requirements. The LCE of a TF system is comparable to that of an AHF (Butterworth et al., 2013). Therefore the proposed flowsheet has viability as not just an onsite treatment system but also potential for small wastewater treatment plants.

7.1 References

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8 Conclusions

The overall aim of the thesis was to understand the impact of temperature and mixing on the processes that govern the sludge accumulation rate in septic tanks in order to develop alternative designs that could enable infrequent desludging. The use of a combination of vertical and horizontal baffles was used to increase the hydrolysis rate constant of a septic tank from 0.0089 d^{-1} to 0.035 d^{-1} compared to a conventional septic tank system. This was achieved whilst maintaining the treatment performance. The increase in hydrolysis rate constant is projected to decrease the desludging frequency from 4.9 to 6.7 years. A review into the literature of anaerobic treatment found that temperature had a dominant impact of the removal of TCOD and TSS of anaerobic reactors however the hydrolysis rate constant of these reactors was not significantly impacted by temperature. Batch tests into anaerobic digestion supported this assessment finding that gentle mixing increased the hydrolysis rate constant at 10^{o}C by 200% compared to an unmixed test. This was a greater increase than was observed by increasing the temperature from 15 to 37^{o}C. The mechanisms underpinning this finding was that hydrolysis is a limited by disintegration opposed to solubilisation and that the mass transfer within passive systems is limiting the assess to degradation sites opposed to a low degradation rate even in temperatures sub 20^{o}C. The mass transfer limitation of anaerobic digestion was utilised to increase hydrolysis rates and thus decrease desludging frequency by using baffling to provide hydraulic mixing of the anaerobic sludge increasing the mass transfer.

- Objective A: Review the current state of low temperature anaerobic treatment of onsite domestic wastewater

A systematic review of low temperature onsite anaerobic reactors found the treatment provide by septic tanks, APs, ABR and UASBs was approximately equivalent. The most influential factor on removal was temperature with a secondary influence from HRT observed for TSS removal. The assessment of reactor complexity highlighted that targeted design amendments could yield significant improvement but with diminishing returns and hence focus should be placed on limited simple modification.

- Objective B: Assess the impact of sub 20^{o}C temperatures on hydrolysis rates in septic tank conditions and the impact of mixing.

Low temperatures on anaerobic hydrolysis caused a significant reduction in the hydrolysis rate however was found to also impact the hydrolysis rate significantly. The biological establishment and activity in low temperatures was not found to be rate limiting in place the mass transfer and mixing was a more significant factor.

- Objective C: Understand the impact of baffle arrangement within septic tanks to reduce the required desludging frequency

The use of vertical baffles to increase sludge mixing was found to sufficiently increase the hydrolysis rate constant such that further hydraulic optimisation would have limited practical impact on the desludging frequency of septic tanks.

- Objective D: Establish the overall flowsheet effluent quality and relative effectiveness of aerated horizontal flow and vertical flow wetlands.

The mean effluent quality of the pilot VF and AHF were 5 mg TSS l^{-1}, 11 mg BOD l^{-1} and 0.5 mg NH_4 l^{-1} and 12 mg TSS l^{-1}, 8 mg BOD l^{-1} and 3mg NH_4 l^{-1} for the VF and AHF respectively. The VF showed effective performance in tested consented parameter however the AHF had an issue with high TSS effluents, and a load adjustment should be considered.

- Objective E: Determine the lifetime cost implications of improving septic tank flowsheets and whether this is a viable method to reduce the carbon footprint of rural wastewater treatment flowsheets.

The WLC of the proposed flowsheet is less than that of that of a conventional package treatment system used to improve onsite treatment at a scale from 5 to 1000 PE. The WLC of the proposed flowsheet is higher than an STS, with the exception of an EST-AHF at 1000 PE which is at parity for WLC. However, there is a significant reduction in LCE using the proposed flowsheet compared to an STS and package system. This reduction in WLC means that the adoption of the proposed flowsheet is a viable method to reduce the carbon footprint of rural wastewater treatment and is comparable in abatement cost to that of thermal hydrolysis.

8.1 Future Work

In Chapter 3 the important roles of disintegration and mixing on hydrolysis were highlighted. it was also observed in Chapter 4 that the hydraulic mixing in the enhanced septic tanks increased the hydrolysis rate. However there remains a significant question around the mechanics behind mixing which are aiding the hydrolysis rate constant. The mixing could be aiding the hydrolysis rate constant by physical action, in which the shear applied to the particles increases the disintegration of particles. Alternatively, the mixing could be aiding mass transfer as the contact between the biomass and particulates. A series of batch studies as described in Chapter 3 are recommended to study the balance of these mechanisms, alongside particle sizing and boundary layer thickness quantification. A study examining the hydrolysis in a series of mixed tests with variable mixing speed alongside the use of both raw and sonicated wastewater to alleviate the impact of particle size reduction. This would enable an assessment of the impact of particle size reduction and mass transfer from mixing.

As highlighted in Chapter 2 as the sludge retention time increases of a system the macro-composition of the sludge will change leading to potential variation in the hydrolysis rates as the desludging frequency is increased, further investigation is needed on the impact of the macro-composition of the sludge on hydrolysis rates and the impact of its changes as the sludge retention time increases.

As discussed in Chapter 6 there is sizeable uncertainty in the process emission of the constructed wetlands recommended for the treatment flowsheet, especially the AHF. To increase the understanding of the process emissions of an AHF the nitrogen pathways within the system need a greater understanding. The process emission of a system such as a CW are highly variable both annually and daily, therefore for an accurate assessment of the process emissions requires a long-term study with regular testing. The nitrogen pathways within a CW can be assessed by conducting seasonal mass balances on the system, to determine the transformations which the nitrogen species are undergoing. The assessment of gaseous emissions from VF wetlands is typically measured using a static chamber method (Mander et al., 2014), however for an AHF the gas emissions are less likely to be diffusive due to a gas flowrate and hence this method is less applicable (Pape et al., 2009). A more appropriate method for process emission quantification of AHFs is the dynamic chamber method (Pape et al., 2009) however the reporting of emissions from these methods is consistently different (Heinemeyer & McNamara, 2011). Therefore, for an accurate assessment of the process emissions, alongside a nitrogen mass balance, an assessment of the dynamic chamber process emissions of a VF wetland will allow for more accurate comparisons between the wetland systems.

The concept of a treatment garden is a promising one, however there are several hurdles which need addressing before implementation can be considered. The constructed wetlands are treating post-anaerobic effluent, which can have a high concentrations of hydrogen sulphate (H_2S) and volatile organics. These compounds can cause significant odours which would limit the aesthetic of a treatment garden. To resolve these issues three stages of future work are recommended. Firstly, there is a social element and therefore there needs to be an investigation into the social acceptability of a treatment garden and modes of implementation which would make the concept more acceptable. Secondly a quantification of the odour from the secondary wetlands. The odour of the system can be quantified using an odour panel or quantification of a surrogate such as H_2S. Comparably to the process emissions of the CWs the monitoring of the odour of the system needs to be a longer-term study than a spot test as variations in factors such as loading, temperature and plant establishment could significantly impact the odour of the system (Wheeler et al., 2007). The use of a surrogate of odour such as H_2S, means that a mass balance on sulphur could be constructed to highlight the pathways and variation of odour in the system allowing for a more targeted adaptation in the operation of the constructed wetland to minimise odour release. Within small and rural works there is also an increasing number of consents for Total Phosphorous (TP) and Total Nitrogen (TN) removal, which are not discussed in this thesis further work is needed to establish the performance of this flowsheet for TP and TN removal and the potential changes which would be required to accommodate these permits.

Finally with the consideration of odours, is the potential mitigation of odour release. This can also be designed in tandem with a potential reduction in process emissions. Biofilters are a biophysical process of gas scrubbing which uses counter-current water to gas flow and a natural media to remove containments from the gas stream (Lith et al., 1997; van Langenhove et al., 1986). The media of a biofilter is frequently organic media such as pine bark (Hernández et al., 2017; Luo & Lindsey, 2006). The use of a layer on top of a CW has been trialled to help insulate systems (Langergraber et al., 2009; Wu et al., 2011), but the use of a topping layer taking concepts from biofilters could be trialled to mitigate both odour and process emissions from CWs. The design of a gas biofilter layer would require the tackling of both mechanistic and operational issues. To assess the lifetime and hence long-term practicality of a gas biofilter, examination of the two key processes in a biofilter need consideration. Within a gas biofilter layer there are two key mechanisms; adsorption and biodegradation (van Langenhove et al., 1986). If the system is adsorption dominated removal is primarily achieved by the binding of pollutant gas to adsorption sites and therefore the media will become saturated and require changing. However, in a biodegradation dominated system the adsorption of pollutants is the rate limiting stage and the adsorbed compounds are subsequently degradation by microorganisms. Investigating these processes means that an assessment can be made on the lifetime, thickness and practicality of a gas biofilter layer. To examine these processes, it is recommended that small scale experiments are conducted to first assess the adsorption saturation of the media per unit surface area (van Langenhove et al., 1986). Then in a controlled tests operate the media in a biofilter, quantifying the pollutant gas loading, the removal rate, the pollutant concentration in the aqueous phase and degradation products. Finally, the media which has run in the biofilter can again be measured for the adsorption potential remaining. This will allow for an assessment of the adsorption sites, the saturation level after operation and the degradation rate. Finally, if the gas biofilter layer is considered a practical mitigation method the design can be operated on pilot system to assess the impact of treatment, field removal and wetness requirement.

8.1.1 Implementation of the Flowsheet at Full Scale

The trials in this thesis were conducted at pilot scale with the EST sized to treat the equivalent of 2.8 PE and the wetlands 1.1 and 1.6 PE for the VF and AHF respectively. This is smaller than the low end of the design envelope of 5 PE and the flowsheet aims for a maximum size of 1000 PE. Therefore, the learning taken from this thesis needs to be trailed at full scale for verification of the findings.

In collaboration with Scottish Water the key design points from this thesis are being applied to a 20 PE, full scale demonstration system. The design of the EST is based on EST-Mid, with 4 baffles, 1 upcomer, 1 downcomer and 2 horizontal baffles. The upcomer is spaced to have a peak upflow velocity of 0.5 m h^{-1}, the downcomer gives a upcomer to downcomer area ratio of 3:1 and the horizontal baffles are spaced equally in the remaining volume. The system is recommended to have a minimum HRT of 48 hours at 3 DWF. A monitoring scheme equivalent to that outline in Chapter 4 is recommended to allow for the assessment of the removal capabilities and sludge accumulation within the system.

The operation of the demonstration scale system will provide an assessment of the EST design for multiple seasons and provide a more accurate assessment of the desludging frequency based on a longer operational period. Ideally, the system would be monitored until the sludge volume reaches the desludging level, 33%, however a minimum of 1-year operational data is required. Additionally, a larger scale system can be monitored to assess the biogas generation capability of the system and a seasonal assessment of the potential energy generation and utilisation of the flowsheet.

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