Investigation of a Rotating Hollow Fibre Membrane Bioreactor System for Wastewater Treatment

Zur Erlangung des akademischen Grades eines

DOKTORS DER INGENIEURWISSENSCHAFTEN (Dr.-Ing.)

von der KIT Fakultät für Bauingenieur-, Geo- und Umweltwissenschaften

des Karlsruher Instituts für Technologie (KIT)

genehmigte

DISSERTATION

von

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Tag der mündlichen Prüfung: 26. Juli 2024

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Karlsruhe 2024

Abstract

The biological wastewater treatment by microorganisms represents a pivotal phase in the operation of wastewater treatment plants. The membrane bioreactor system has emerged as an alternative to the conventional activated sludge system for wastewater treatment where the space is limited, as it eliminates the need for a secondary clarifier while achieving high nutrient removal efficiencies and biomass retention. However, membrane bioreactor systems face certain drawbacks. As the membrane bioreactor operates at a higher mixed liquor suspended solids concentration, the tendency for membrane fouling is increased, and the oxygen transfer into the sludge is depleted.

This dissertation presents an innovative approach to address the dual challenges of oxygen transfer and membrane fouling in membrane bioreactor systems. The approach involves the development and testing of a pilot-scale rotating hollow fibre membrane module. The study comprises a series of experiments conducted in both batch and continuous flow pilot plants, focusing on oxygen transfer rates and membrane fouling behaviour under various operational conditions.

Three peer-reviewed publications form the core of this dissertation, detailing the performance of the novel hollow fibre membrane bioreactor design. The first publication investigates the impact of membrane rotation on fine-bubble aeration, revealing significant improvements in oxygen transfer rates and energy efficiency. The second publication examines the influence of solid holdup on oxygen transfer in the rotating hollow fibre membrane bioreactor under different aeration conditions, highlighting the consistent impact of solids on oxygen transfer depletion. The third publication analyses the effect of operational parameters on transmembrane pressure and filtrate quality, demonstrating the potential of the rotating membrane module to mitigate fouling and enhance system efficiency.

Key findings from the study include:

- The rotating hollow fibre membrane bioreactor module significantly improves oxygen transfer efficiency, with the improvement offsetting the additional energy required for rotation, suggesting potential cost savings in full-scale applications.
- The solid holdup universally impacts oxygen transfer depletion, regardless of reactor type, diffuser setup, and rotational speed, emphasising the necessity of incorporating solid holdup or liquid holdup into the calculations of the α-factor.
- The rotating module reduces the increase in transmembrane pressure, allowing for extended operation periods (up to 48 days) with satisfactory performance, even without aeration during warmer seasons.

- The novel design of the hollow fibre membrane bioreactor shows high feasibility for integration into existing conventional activated sludge systems without major modifications, potentially reducing the total membrane area needed and, consequently, capital expenditure.

The dissertation concludes with a discussion on the broader implications of the rotating hollow fibre membrane bioreactor system, including its potential to operate at higher mixed liquor suspended solid concentrations, its advantages in preventing silting, and its adaptability to different operational conditions. Recommendations for future research include the optimization of operational parameters and the exploration of alternative membrane materials to enhance sustainability and performance.

Kurzfassung

Die biologische Reinigung von Abwasser stellt das Kernstück der kommunalen Abwasserreinigung auf Kläranlagen dar. Das Membranbioreaktorsystem erweist sich hierfür als vielversprechende Alternative zum konventionellen Belebtschlammsystem, insbesondere aufgrund des begrenzten Platzbedarfs. Es macht den Einsatz eines Nachklärbeckens überflüssig und zeichnet sich gleichzeitig durch einen hohen Nährstoffeliminationsgrad sowie einen hohen Biomasserückhalt aus. Allerdings sind auch gewisse Nachteile von Membranbioreaktorsystemen zu berücksichtigen. Da der Membranbioreaktor mit höheren Trockensubstanzgehalten betrieben wird, ist eine Tendenz zum Membranfouling sowie eine Verringerung des Sauerstoffeintrags zu beobachten.

Die vorliegende Dissertation präsentiert einen innovativen Ansatz zur Bewältigung der doppelten Herausforderungen eines effizienten Sauerstoffeintrags und einer Verminderung von Membranverblockung und -verschlammung in Membranbioreaktorsystemen. Der Ansatz umfasst die Entwicklung und Erprobung eines rotierenden Hohlfaser-Membranmoduls im Pilotmaßstab. Die Studie umfasst eine Reihe von Experimenten, die sowohl in Batch- als auch in kontinuierlichen Durchfluss-Pilotanlagen durchgeführt wurden. Im Rahmen der Studie wurde der Fokus auf die Untersuchung der Sauerstoffeintragsraten sowie des Verschmutzungsverhaltens der Membran unter verschiedenen Betriebsbedingungen gelegt.

Die Dissertation basiert auf drei begutachteten Veröffentlichungen, welche die Leistungsfähigkeit des neuartigen Hohlfaser-Membranbioreaktordesigns beschreiben. Die erste Veröffentlichung analysiert die Auswirkungen der Membranrotation auf die feinblasige Belüftung und zeigt signifikante Verbesserungen des Sauerstoffeintrags und der Energieeffizienz. Die zweite Veröffentlichung analysiert den Einfluss des Feststoffrückhalts (Volumen der Flocken) auf den Sauerstoffeintrag im rotierenden Hohlfaser-Membranbioreaktor unter verschiedenen Belüftungsbedingungen. Dabei wird die konsistente Auswirkung der Feststoffe auf den resultierenden Sauerstoffeintrag hervorgehoben. Die dritte Veröffentlichung analysiert die Auswirkungen der Betriebsparameter auf den Transmembrandruck und die Qualität des Filtrats und demonstriert das Potenzial des rotierenden Membranmoduls zur Verringerung von Membranverblockung und -verschlammung.

Zusammenfassend lassen sich die wichtigsten Ergebnisse der Dissertation wie folgt zusammenfassen:

- Die rotierende Hohlfaser-Membranbioreaktoreinheit weist eine signifikante Verbesserung des Sauerstoffeintrags auf, wobei die Effizienzsteigerung den

zusätzlichen Energiebedarf für die Rotation kompensiert. Dies deutet auf potenzielle Kosteneinsparungen in großtechnischen Anwendungen hin.

- Der Feststoffrückhalt beeinflusst den Sauerstoffeintrag unabhängig vom Reaktortyp, der Art der Belüfter und der Rotationsgeschwindigkeit. Dies unterstreicht die Notwendigkeit, den Anteil der Feststoffphase bzw. den Anteil der resultierenden Flüssigkeitsphase in die Berechnung des α-Faktors einzubeziehen.
- Das rotierende Modul reduziert den Anstieg des Transmembrandrucks und ermöglicht längere Betriebszeiten (bis zu 48 Tage) mit zufriedenstellender Leistung, selbst ohne Belüftung in wärmeren Jahreszeiten.
- Das neuartige Design des Hohlfaser-Membranbioreaktors eröffnet die Möglichkeit einer hohen Integration in bestehende Belebtschlammanlagen, ohne dass größere Modifikationen erforderlich wären. Infolgedessen würde sich die benötigte Membranfläche und damit die Investitionskosten reduzieren.

Die Dissertation schließt mit einer Diskussion über die weitreichenden Implikationen des Hohlfaser-Membranbioreaktorsystems. Dabei werden insbesondere das Potenzial des Systems bei höheren Trockensubstanzgehalten, seine Vorteile bei der Vermeidung von Membranverblockung und -verschlammung sowie seine Anpassungsfähigkeit an verschiedene Betriebsbedingungen erörtert. Für zukünftige Forschungsarbeiten werden folgende Empfehlungen ausgesprochen: Die Betriebsparameter sollten optimiert und alternative Membranmaterialien untersucht werden, um die Nachhaltigkeit und Leistung des rotierenden Hohlfaser-Membranmoduls weiter zu verbessern.

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List of Abbreviations

AGS	aerobic granular sludge
ASP	alkali/surfactant/polymer
BOD	biological oxygen demand
CAPEX	capital expenditure
CAS	conventional activated sludge
CB	coarse bubble
CEB	chemically enhanced backflush
CIA	cleaning in air
CIP	cleaning in place
COD	chemical oxygen demand
CP	concentration polarization
EPS	extracellular polymeric substances
F/M	food-to-microorganism
FB	fine bubble
HF MBR	hollow fibre membrane bioreactor
HFV	hydrostatic floc volume
k_La	oxygen transfer coefficient
LMH	litres per square meter per hour
MBR	membrane bioreactor
MC	maintenance cleaning
MCRT	mean cell residence time
MF	Microfiltration
MLSS	mixed liquor suspended solids
MLVSS	mixed liquor volatile suspended solids
OTR	oxygen transfer rate
PE	population equivalent
PES	polyether sulfone
PFAS	per- and polyfluoroalkyl substances
PVDF	polyvinylidene difluoride
SAE	standard aeration efficiency
SBR	sequencing batch reactor
sCOD	soluble chemical oxygen demand
SMP	soluble microbial products
SOTR	standard oxygen transfer rate
SRT	solids retention time
SVI	sludge volume index
TSS	total suspended solids
UF	ultrafiltration
WWTP	wastewater treatment plant
αSAE	standard aeration efficiency for activated sludge

1. Introduction

The biological wastewater treatment by microorganisms is a crucial phase in the operation of wastewater treatment plants (WWTP). A common method employed in the biological treatment stage involves the use of suspended biomass, where microorganisms (in the form of suspended biomass), wastewater, and air are mixed together in a tank. This mixture is referred to as activated sludge. The typical design based on the suspended biomass concept is known as the conventional activated sludge (CAS) system.

The majority of microorganisms in the CAS system require oxygen to degrade pollutants, requiring significant energy input to introduce oxygen into the reactor through various aeration systems. This energy requirement represents a substantial portion of the operating costs in wastewater treatment, accounting for 50% to 60% of the energy consumption of the entire wastewater treatment plant (Chen et al. 2022c). In order to meet the oxygen demand of the micro-organisms and ensure effective wastewater treatment, while considering the economic factor, the optimum design of the aeration system must be achieved through efficient energy use. This is evidenced by numerous studies conducted to enhance energy efficiency using various approaches, such as those conducted by Baquero-Rodríguez et al. (2018), Lozano Avilés et al. (2019), Muloiwa et al. (2023) and Oulebsir et al. (2020).

In this context, the aerobic membrane bioreactor (MBR) system has emerged as a promising alternative for wastewater applications with limited space, as it eliminates the need for a secondary clarifier while achieving high nutrient removal efficiency and biomass retention (Judd 2016). Brepols et al. (2008) provided an example of a ratio of two in a comparison of the required area between CAS and MBR. Moreover, a number of studies have demonstrated that MBR effluent often surpasses CAS systems in terms of quality, particularly in the reduction of pollutants such as chemical oxygen demand (COD), ammonium, and suspended solids, including challenging substances like pharmaceuticals (Chandrasekeran et al. 2007; Sriboonnak et al. 2022; Rahman et al. 2023; Ayyoub et al. 2022; Li et al. 2022). Furthermore, the MBR system is capable of operating at higher mixed liquor suspended solids (MLSS) concentrations than the CAS system, which reduces the volume of the reactor and the production of sludge (Judd 2011). However, this also entails certain drawbacks. Higher MLSS concentrations result in increased membrane fouling and a decline of oxygen transfer into the sludge (Germain and Stephenson 2005; Lee and Kim 2013).

This study established a novel concept of rotating hollow fibre (HF) MBR, which was then implemented in a pilot-scale membrane module for testing in two different reactors: a batch reactor and a continuous flow reactor. The batch reactor was employed for oxygen transfer experiments without any filtration process. The oxygen transfer experiments were conducted on clean water and activated sludge with three different MLSS concentrations. In addition, a series of experiments were also conducted on aerobic granular sludge. Meanwhile, the continuous flow reactor was employed for experiments in which the membrane module was continuously in operation, filtering the wastewater, which in this study was the return sludge from biological tanks of a WWTP.

The central argument of this thesis is based on three peer-reviewed scientific publications (Chapters 3, 4 and 5), which present the results and discussions of experiments conducted on the aforementioned rotating HF MBR. The primary focus of this study is on the performance of the membrane module in addressing two critical issues in MBR systems: oxygen transfer and membrane fouling.

The work presented was conducted in accordance with the following five research objectives:

1. **Evaluate the theoretical development on oxygen transfer and membrane fouling within MBR system.**

   Two main drawbacks in MBR applications are limited oxygen transfer and enhanced membrane fouling, prompting a significant number of studies focused on these issues. Various strategies have been proposed by many researchers to optimise the aeration system for MBRs and mitigate membrane fouling. Thus, Chapter 2 addresses the theoretical background and development to provide perspective on these issues. Furthermore, at the beginning of Chapters 3, 4, and 5, a concise literature review is provided, which addresses the topic of each chapter.

2. **Determine the impact of the rotational speed of the membrane module, airflow rate and aeration type on oxygen transfer.**

   The principal concept of the membrane module employed in this study is its rotatable feature, which generates a dynamic shear force within the tank. This may result in alterations to the behaviour of rising air bubbles during the aeration process. Consequently, it is necessary to determine the impact of membrane rotation at different speeds, in conjunction with various aeration types and airflow rates. Chapter 3 presents the results and analysis of initial oxygen transfer experiments with fine bubble aeration on clean water and activated sludge within the typical range of MLSS concentrations. Chapter 4 presents further oxygen transfer experiments with fine and coarse bubble aerations on different MLSS concentrations of activated sludge. In Chapter 6, the results from additional experiments on aerobic granular sludge are presented and compared to the results from clean water and activated sludge experiments.

3. **Assess the feasibility of additional energy generated by membrane rotation.**

   The addition of further equipment to an existing system, such as the rotating device in this study, frequently raises a question regarding the necessity of the additional energy expenditure. Therefore, it is necessary to quantify whether the additional energy consumed by the rotating device is technically and economically viable. This can be achieved by providing the oxygen transfer coefficient and aeration efficiency when rotation is applied to the system and then comparing these results to experiments without any rotation. This is discussed in detail in Chapter 3 and 4.

4. **Identify the influence of suspended solids on oxygen transfer.**

   Suspended solids, in the form of MLSS concentration and viscosity, are frequently considered the primary factors influencing oxygen transfer in activated sludge. However, these parameters fail to account for the three-phase nature of activated sludge. Henkel (2010) introduced the hydrostatic floc volume (HFV) method to estimate liquid and solid holdup in sludge and their influence on oxygen transfer. Therefore, this study will assess whether suspended solids or solid holdup can explain the effect of solids in sludge on oxygen transfer within the studied MBR system, and the results are discussed in Chapter 4.

5. **Observe the filtrating performance and membrane fouling behaviour of the rotating HF MBR during the filtration operation.**

   The fouling of membranes results in a reduction in performance, which is manifested by an increase in the transmembrane pressure and a reduction in the flux. Consequently, it is essential to assess the performance of the pilot-scale membrane module utilised in this study, including the determination of the optimal operational parameters and cleaning strategies. Furthermore, the efficiency of solid removal is evaluated by measuring the quality of the filtrate. These topics are addressed in Chapter 5.

2. Theoretical Background and Development on Oxygen Transfer and Membrane Fouling in MBR System.

2.1. MBR as Alternative for CAS

The practice of water reuse is becoming increasingly prevalent worldwide, as societies demand advanced wastewater treatment processes to meet the standards required for reuse. A lower food-to-microorganism (F/M) ratio leads to higher removal efficiency of biological oxygen demand (BOD), which is measured as BOD_5. This can be achieved by increasing the MLSS concentration in the biological reactor. The MLSS can be concentrated in the biological reactor by extracting the treated effluent using membrane filtration (Hai et al. 2011).

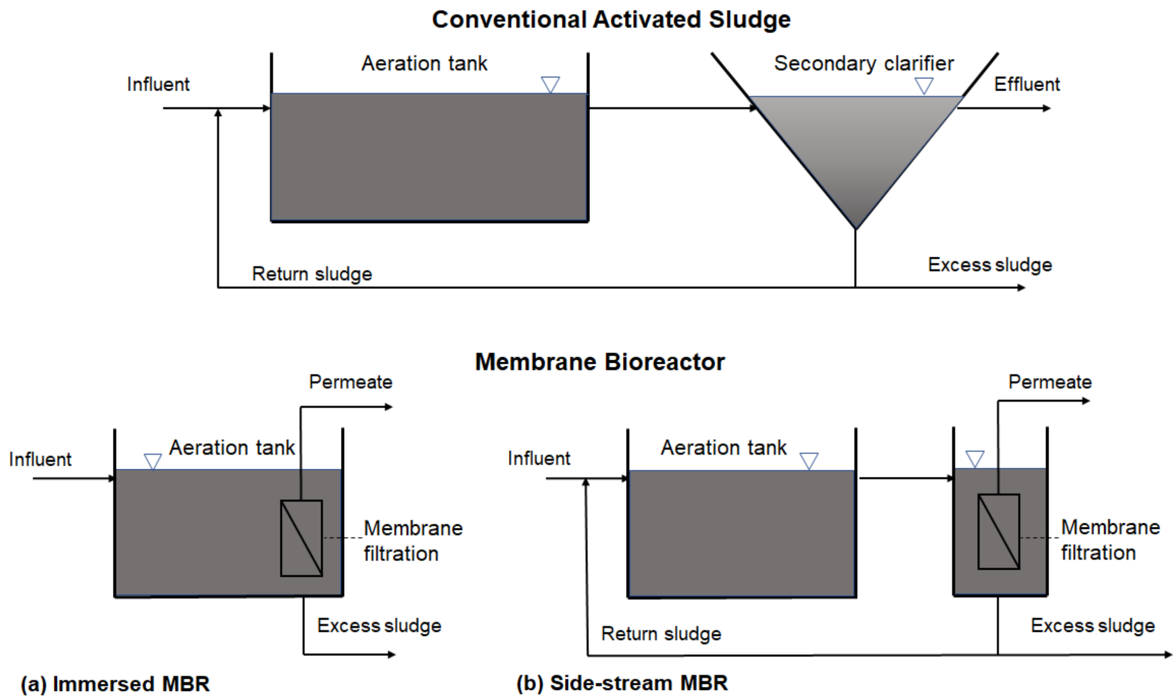


Figure 2.1. Scheme of the CAS system (top) and two different configurations of MBR (bottom), adapted from Judd (2011) and Ladewig and Al-Shaeli (2017).

By combining membrane technology with biological treatment, MBR system represents an alternative to CAS system. In MBR system, porous membrane filtration replaces the secondary clarifier which typically used in CAS system, as illustrated in Figure 2.1. The pore diameter of the membranes utilised in MBRs allows the rejection of activated sludge flocs, free-living bacteria, and even large viruses or particles (Park et al. 2015). It also reduces the required size of the aeration tank and increases the efficiency of the biotreatment process.

The pore size of the microfiltration (MF) and ultrafiltration (UF) membranes used in the MBR system is approximately 0.1 to 10 μm and 0.01 to 0.15 μm, respectively (Nasir et al. 2022). At these sizes, MBR enables the removal of suspended and colloidal materials, resulting in a clarified and substantially disinfected effluent (Judd 2011). The quality of the effluent from the MBR system is comparable to that obtained from membrane filtration following the CAS system but in a single step, compact unit (Judd 2016; Hai et al. 2019; Pandey and Kant Singh 2014). A number of studies have demonstrated that MBR is capable of eliminating at least 90% of COD, BOD_5 and total suspended solids (TSS) (Ayyoub et al. 2022; Khastoo et al. 2021; Yang et al. 2020; Gil et al. 2010). Moreover, the MBR system is also capable of retaining all microorganisms (Ng and Hermanowicz 2005), including those with poor settling properties. Consequently, the system is not affected by the sedimentation behaviour of sludge flocs.

Early MBR systems were operated with very high solids retention time (SRT) of up to 100 days and MLSS concentrations as high as 30 g/L (Ladewig and Al-Shaeli 2017). This extended SRT allows for the retention of slow-growing microorganisms, including those that can grow on synthetic chemicals. Consequently, less energy is available for cell production, ultimately leading to reduced sludge production in MBR systems compared to CAS systems (Hai et al. 2019). Nevertheless, recent trends in MBR operation favour lower MLSS concentrations between 10 - 15 g/L, in order to decrease membrane fouling tendency and membrane cleaning intensity (Ladewig and Al-Shaeli 2017; Le-Clech et al. 2006).

2.2. Configuration of MBR

The configuration of the membrane is of critical importance in determining the overall performance of the process. According to Judd (2011), the optimal configuration of the membrane should aim to achieve the following aspects:

- A high membrane area to module bulk volume ratio, which is known as packing density,
- a high degree of turbulence to promote mass transfer on the feed side,
- a low energy expenditure per unit of product water volume,
- a low cost per unit membrane area,
- a design that facilitates cleaning, and
- a design that permits modularisation.

There are two principal process configurations of MBR, namely submerged (or immersed) and side stream (Judd 2011; Park et al. 2015; Ladewig and Al-Shaeli 2017). In the submerged configuration, biodegradation and biomass separation occur within a single reactor. In contrast, the side stream configuration separates biomass in an additional tank using a pressure-driven membrane (see Figure 2.1). The submerged MBR configuration is the most widely used for biomass rejection due to its low specific energy demand, rendering it the most economically viable option for large-scale applications (Judd 2011).

2.3. The Fundamental of Oxygen Mass Transfer in Water

In both the CAS and MBR systems, the aeration system performs two functions: oxygen transfer and mixing. While mixing is crucial for ensuring the full utilisation of the activated sludge reactor volume and uniform dispersion of dissolved oxygen throughout the mixed liquor, oxygen transfer is still regarded as the primary function subjected to biological treatment (Mueller et al. 2002). The following section will discuss the fundamental kinetics of oxygen mass transfer into water and wastewater, as well as the various measurement methods employed.

2.3.1. The kinetics of oxygen mass transfer

The release of air to water results in the formation of air bubbles. These bubbles act as carriers for gases, which are driven by Brownian motion from the gas phase to the liquid phase until equilibrium is reached and the gases become dissolved in the water (Mueller et al. 2002; Henkel 2010). The fundamental theory employed for the calculation of oxygen transfer into water is the Two-Film Theory, as proposed by Lewis and Whitman (1924). This theory states that the transfer rate can be expressed in terms of an overall transfer coefficient and resistances on either side of the interface. The basic equation for this theory is as follows:

$$\frac{dc}{dt} = k_L a \, . \left(c_S - c_{(t)}\right) \qquad \text{(Equation 2.1)}$$

where *dc/dt* representing the increase of oxygen concentration in respect of time (g/(m^3·h)); $k_L a$ representing the oxygen transfer coefficient (h^{-1}); c_s representing the saturation concentration of oxygen (g/m^3); *c(t)* representing the oxygen concentration at time t (g/m^3).

The oxygen transfer coefficient is the characteristic value for the aeration device set to a specific condition in a specific tank and is calculated from a oxygen supply test. The physical properties of the wastewater, such as temperature, influence the aeration process. An increase in wastewater temperature will result in a decrease in the absolute value of oxygen uptake rate. The effect of temperature on the oxygen transfer coefficient is expressed through the φ factor. For a standard temperature of 20 °C, the oxygen transfer coefficient is a function of the specific measured temperature and can be expressed by the following formula.

$$k_L a_{20} = k_L a_T \, . \varphi^{(T-20)} \qquad \text{(Equation 2.2)}$$

Typical values for φ are between 1.015 and 1.040 (Judd 2011). In DWA-M 209 (2007), the φ value is set at 1.024.

Furthermore, the presence of impurities in wastewater results in a reduction in the aeration coefficient compared to that observed in clean water. This reduction is quantified by the alpha (α) factor, which represents the ratio of the aeration coefficient in wastewater to that in clean water.

$$\alpha - \mathrm{factor} = \frac{k_L a_{\mathrm{wastewater}}}{k_L a_{\mathrm{clean\ water}}} \quad \text{(Equation 2.3)}$$

Another critical parameter in oxygen transfer is the standard oxygen transfer rate (SOTR). SOTR is defined as the oxygen transfer rate in kilograms in clean water in a specific tank volume at 20°C, 1 atm, and with initial DO concentrations of zero within one hour. The SOTR formula is as follows.

$$\mathrm{SOTR} = \frac{V . k_L a_{20} \cdot c_{S,20}}{1000} \quad \text{(Equation 2.4)}$$

The presence of salts and particulates in wastewater has been demonstrated to impact the oxygen transfer rate (Judd 2011). Neutral salts have been shown to reduce both the oxygen saturation value and the oxygen solubility (DWA-M 209 2007; Drewnowski et al. 2019). Consequently, in wastewater, SOTR must consider the salt factor, defined as the β factor, and is calculated using the following formula.

$$\alpha\mathrm{SOTR} = \frac{V . \alpha k_L a_{20} \cdot \beta c_{S,20}}{1000} \quad \text{(Equation 2.5)}$$

The β factor can be calculated using the total dissolved solids (TDS) concentration (in g/m^3) according to the following formula, as proposed by Rosso et al. (2008).

$$\beta = 1 - 0.01 * \frac{TDS}{1000} \quad \text{(Equation 2.6)}$$

The dual purpose of aeration in an immersed membrane bioreactor is to both clean the membrane and provide dissolved oxygen, thus maintaining a viable biomass. This necessitates a compromise between mixing, which requires larger bubbles, and oxygen dissolution, which is more efficient with smaller bubbles. Consequently, only a small proportion of the oxygen in the supplied air, which may be as low as 10%, is utilised by the biomass, with this proportion decreasing with increasing biomass concentration (Judd, 2007). This efficiency is quantified by the standard aeration efficiency (SAE), which is defined as the rate of oxygen transfer per unit of power input (P), expressed in kilograms of oxygen per kilowatt-hour. SAE can be specified for clean water or for activated sludge (αSAE) and calculated using the following formulas (DWA-M 209 2007).

$$SAE = \frac{SOTR}{P} \quad \text{(Equation 2.7)}$$

$$\alpha SAE = \frac{\alpha SOTR}{P} \quad \text{(Equation 2.8)}$$

2.3.2. Oxygen transfer measurement

Oxygen transfer measurements can be conducted in either clean water or activated sludge, with or without flow. The accurate measurement of the K_La value under operational conditions is crucial for the determination of accurate design α-factors and for the long-term monitoring of aeration system performance (Capela et al. 2004). These measurements are essential for evaluating the efficiency of aerators and the overall aeration system. The three principal methods for measuring oxygen transfer are the off-gas method (also known as the steady-state method), the absorption method and the desorption method (Henkel 2010; DWA-M 209 2007). The latter two are also referred to as the dynamic method.

The off-gas method entails the continuous measurement of oxygen transfer rates in activated sludge systems by estimating the steady-state oxygen transfer coefficient through the application of the gas phase mass balance of the aerated volume. This method entails comparing the concentrations of oxygen and carbon dioxide from ambient air with those of the off-gas (Henkel 2010). In a study conducted by Capela et al. (2004), it was found that under constant process conditions, the mean oxygen transfer coefficients obtained from all methods were similar. It is recommended that a method be selected based on the constraints of the site, the cost, and the limitations of the method. Krause et al. (2003) demonstrated that both non-steady state and steady state methods are applicable in MBR systems with high MLSS concentrations, yielding comparable results for oxygen transfer rates.

Given that it can be applied in any type of aeration system (Capela et al. 2004), the dynamic method is a widely used approach for measuring oxygen transfer (Gourich et al. 2006). The effective concentration range for this analysis is typically narrow, between 5 and 8 mg/L. In the absorption method, oxygen saturation concentration is depleted by adding sodium sulfite or aerating with pure nitrogen gas in clean water measurements. In activated sludge systems, microbial respiration is employed to reduce the oxygen concentration. In order to prevent stressing aerobic bacteria and ensure the reliability of the analysis, depletion must cease before a concentration of 0 mg/L is reached. Consequently, the analysis concentration range is limited to 5-8 mg/L. Subsequently, the aerator is initiated, and the oxygen concentration is continuously monitored until it reaches a stable state. Conversely, in the desorption method, the oxygen concentration is initially increased artificially, typically using hydrogen peroxide or pure oxygen gas, to a concentration of at least 10 mg/L above the initial concentration. This is because the guideline set forth in DWA-M 209 (2007) stipulates that the difference between the saturated DO concentration and the initially elevated DO concentration should reach at least 8 mg/L in order to obtain a reliable result. This is then continuously measured until it stabilises. Figure 2.2 illustrates the difference plot between the desorption and absorption methods.

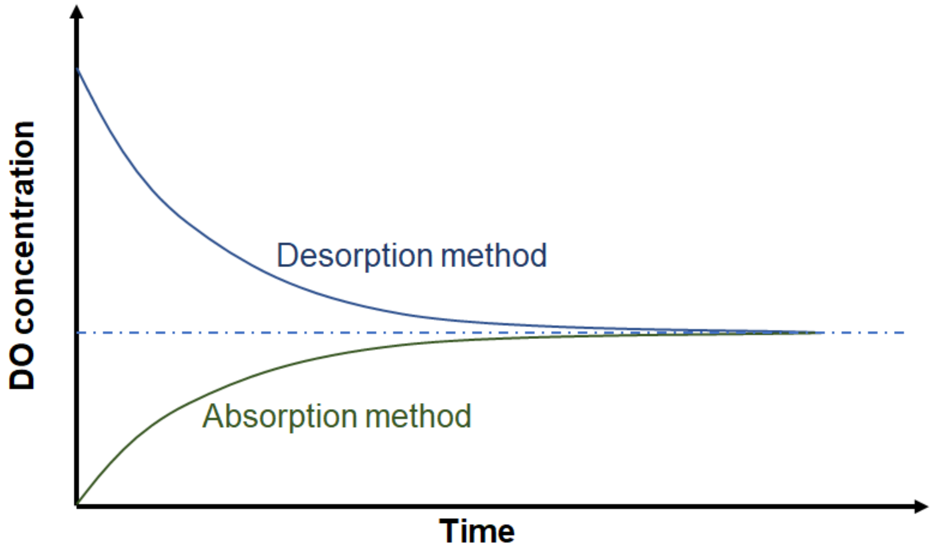


Figure 2.2. Illustrative plot of the desorption and absorption method, adapted from Wagner et al. (1998) and Henkel (2010).

2.4. Oxygen Transfer in MBR System

In an aerobic MBR system, a bubble column reactor is employed to bring the liquid phase (activated sludge) and the gas phase (air) into contact, thereby initiating mass transfer between the phases. The characteristic of air bubbles plays a pivotal role in influencing the efficiency (Germain and Stephenson 2005). The introduction of air into the reactor is typically achieved through via immersed air-bubble diffusers or surface aeration (Judd 2011). The use of fine bubble aerators offers the advantage of superior oxygen transfer within the reactor, while the use of coarse bubble aerators provides better cleaning effects through air scouring and the agitation of fibres (Judd 2011; Hai et al. 2019). However, the efficiency of fine bubbles can be compromised by bubble coalescence, resulting in k_La values comparable to those of coarse bubbles (Germain and Stephenson 2005), especially if the bubble column reactors are equipped with additional mixing devices, such as impellers, to further enhance mass transfer efficiency.

In order to gain a deeper understanding of the processes occurring within a bubble column reactor, it is essential to understand factors that affect the oxygen transfer in MBR system, which are divided in three main categories: biomass characteristics, aeration characteristics and operational characteristics.

2.4.1. Biomass characteristics

The MLSS concentration has frequently been identified as the primary factor affecting oxygen transfer within activated sludge systems, including MBR systems. A number of studies have demonstrated that the oxygen transfer coefficient and the α-factor decrease as solids concentration or viscosity increases. This, in turn, leads to a higher energy demand to sufficiently supply oxygen to the biomass (Germain and Stephenson 2005; Krampe and Krauth 2003; Kim et al. 2019; Germain et al. 2007; Judd 2011). Although a trend was identified whereby the α-factor decreased as the MLSS concentration increased, when the results of several studies were compared, no consistent value of the α-factor was found for the same MLSS concentration. This discrepancy can be attributed to the fact that the measurements were conducted under varying operational conditions.

The reduction in oxygen transfer with increasing MLSS concentration can be attributed to a number of factors, which are interrelated. As MLSS concentration increases, apparent viscosity also rises (van der Roest et al. 2002; Hai et al. 2019; Durán et al. 2016). Viscosity plays a crucial role in determining the oxygen transfer coefficient. As viscosity increases, the liquid film becomes thicker, thereby increasing the resistance to oxygen transfer from the gas to the liquid phase. Furthermore, the presence of high MLSS in conjunction with elevated viscosity can result in the coalescence of bubbles. This coalescence results in a reduction of the specific interfacial area (a) (Martín et al. 2007), which in turn leads to a decline in the overall oxygen transfer coefficient (k_La). This phenomenon is particularly noticeable in fine bubble aeration systems. Furthermore, sludge is defined as a non-Newtonian pseudoplastic fluid, meaning that an increase in shear stress leads to a decrease in viscosity. Consequently, increasing the aeration rate has two beneficial effects on oxygen transfer in MBR systems. Firstly, it increases the amount of oxygen available in the system. Secondly, it reduces the viscosity of the biomass by increasing the shear stress, as observed by G Germain and Stephenson (2005) and Günder (2001). Nevertheless, it is important to note that Pollice et al. (2006) discovered that the apparent viscosity increases at a slower rate than the solid concentration. Consequently, the increase of the MLSS concentration in the MBR may result in a negligible increase in viscosity.

Due to the high variability in predicting the α-factor using the MLSS concentration, Henkel et al. (2009b) demonstrated a better correlation between the α-factor and the mixed liquor volatile suspended solids (MLVSS) concentration. Their findings indicated that the MLVSS concentration uniformly influences the α-factor in both fine bubble and coarse bubble systems. This is further supported by Henkel et al. (2011), who demonstrated that the MLVSS concentration controls the free water content (liquid holdup) and the floc volume (solid holdup) in the sludge. The latter is defined as the HFV. As the HFV increases, the interfacial area between the liquid phase and the bubble is reduced by floc friction, leading to a decrease in oxygen transfer. Their results indicated that the reduction in the oxygen transfer rate is attributable to the increase of the floc volume, rather than an increase in viscosity or MLSS concentration.

The effect of solids concentration on the oxygen transfer coefficient is also dependent on particle size, which within MBR systems is typically broad, ranging from a few micrometres to 500 µm. The findings of Germain and Stephenson (2005) indicated that the effects of particle size and solids concentration on oxygen transfer are interrelated. When the solids concentration was increased with fine particles (up to 0.01 mm), the oxygen transfer coefficient initially increased to a certain point, stabilised, and then decreased with further solids loading. Conversely, an increase in solids concentration with larger particles (around 1–3 mm) led directly to a decrease in oxygen transfer coefficient. This second phenomenon is more likely to occur in MBRs, where the mean particle size is generally above 0.01 mm, and some particles easily reach a diameter of 0.5 mm. Nevertheless, Koide et al. (1984) observed no modification in oxygen transfer coefficient when increasing the diameter of gel particles from 1.88 to 3.98 mm in a bubble column. It is also important to note that particle size can potentially change during the biological process as a result of aeration. An increase in aeration intensity to increase the DO concentration leads to a higher mixing intensity, which in turn induces floc breakup and results in the formation of particles of different sizes (Germain and Stephenson 2005). This is corroborated by the findings of Abbassi et al. (2000), who demonstrated that an increase in airflow and DO concentration had a comparable effect on floc break-up and particle size.

Furthermore, soluble microbial products (SMP) and extracellular polymeric substances (EPS) are additional parameters frequently considered when determining oxygen transfer in activated sludge systems. For DO to reach the active sites on the bacterial cell membrane, it must first penetrate the liquid film around the flocs (SMP) and then diffuse through the floc matrix (EPS). It is noteworthy that the findings of Germain et al. (2007) indicated that only the COD fraction of SMP exerts an influence on oxygen transfer, whereas the carbohydrate and protein fractions do not. This is attributed to the presence of surfactants in the biomass. Meanwhile, EPS, originating from bacterial metabolism and incoming wastewater compounds, constitute the largest fraction of the floc in activated sludge and exert a significant influence on its properties (Henkel 2010; Li and Ganczarczyk 1990). A study by Steinmentz (1996) demonstrated that elevated EPS concentrations are associated with reduced α-factor, suggesting that EPS may influence oxygen transfer. However, the precise mechanism remains uncertain, given that EPS is primarily situated within the floc and does not directly interact with bubbles (Henkel 2010).

In addition, the presence of surfactants in wastewater poses a significant challenge to aeration processes. According to a study by Wagner and Pöpel (1996), surfactants have a positive effect by inhibiting the coalescence effect, but they significantly reduce the liquid film coefficient (k_L) value. On the one hand, the diffusion coefficient is reduced due to the increased resistance at the bubble interface caused by the adsorbed surfactants. On the other hand, the interface changes from mobile to more rigid, which decreases the surface renewal rate, reduces the turbulence of the bubble wake and consequently the coalescence tendency (Henkel 2010). And when these two effects are combined, the presence of surfactants reduces the oxygen transfer coefficient. This is supported by several other studies by Chen et al. (2007), Gillot (2000b), Painmanakul et al. (2005) and Rosso et al. (2006). However, a study by Campbell and Wang (2020) showed that the influent sodium dodecyl sulphate, a commonly used surfactant, had no direct effect on oxygen transfer performance. They found that sludge morphological parameters such as sludge volume index (SVI) and apparent viscosity were more important for oxygen transfer than the influent surfactant.

2.4.2. Operational characteristic

Operational characteristics such as SRT, temperature, and salinity also play an important role in oxygen transfer. The SRT, also known as sludge age or mean cell residence time (MCRT), is a crucial factor in determining the treatment grade, biomass concentration, and oxygen requirements in wastewater treatment. Increasing SRT allows slow-growing microorganisms to establish, enhancing complex substance degradation. Conversely, decreasing SRT leads to the accumulation of slowly degradable substances on floc surfaces (Henkel et al. 2009a). At high SRT, there is a more pronounced degradation of substances, particularly surfactants present in the liquid phase, which have a negative impact on oxygen transfer (Gillot and Héduit 2008; Rosso et al. 2008). However, increasing SRT by raising MLVSS concentration reduces the free water content, thereby reducing the α-factor (Henkel et al. 2009a).

Moreover, oxygen transfer, like any diffusion-based process, is temperature-dependent. The diffusion rate increases with rising temperatures, and consequently, the liquid film coefficient, which depends on the diffusion rate through the liquid film at the interface, also increases with higher temperatures. Higher temperatures also lead to lower viscosity, which means an increased oxygen transfer rate (OTR) (Dong and Hung 1985). However, higher temperatures result in a lower oxygen saturation limit for water and reduced oxygen solubility (Jenkins 2014). Nevertheless, this reduction in oxygen solubility is insignificant in comparison to the increase in the k_La value, which typically results in a higher OTR at elevated temperatures (Judd 2011; Vogelaar 2000).

The presence of salts in water, particularly calcium and magnesium salts, has been demonstrated to enhance the impact of surfactants on oxygen transfer (Lynch and Sawyer

1960). Furthermore, in the event of coalescence, the elevated concentration of salts in wastewater in comparison to that of clean water should result to an increase in gas holdup. This increase in gas holdup enhances the superficial area "a," thereby increasing the oxygen transfer coefficient (k_La) (Henkel 2010; Zannotti and Giovannetti 2015).

2.4.3. Aeration characteristic

In MBR systems, aeration is arguably the most critical unit operation, providing oxygen to microorganisms for their metabolism and to the membrane surface for fouling control. Fine bubble aeration, with its extensive surface area, is advantageous for efficient oxygen transfer to cells. In contrast, coarse bubble aeration, which produces large-sized bubbles, is effective for vibrating and scouring the membrane bundles (Park et al. 2015), while still providing oxygen transfer into the biomass (Judd 2011; Cornel et al. 2003). The aeration in a MBR system is comprised of blowers that compress air and transmit it through pipes to air diffusers, which are typically situated at the base of the reactor. The size of the bubbles plays a pivotal role in this process. In the majority of cases, alterations in k_La are attributable to alterations in the air bubbles (Campbell and Boyle 1989).

The materials used in the construction of fine bubble diffusers in wastewater treatment include ceramics, porous plastics, and perforated membranes. Ceramics are composed of alumina, aluminium silicate, and silica, and feature interconnected passageways that give rise to a distinctive bubble pattern. Porous plastics offer several advantages, including a lighter weight, resistance to breakage, and cost-effectiveness. However, they have lower strength and environmental resistance. Perforated membranes, created by punching holes in a membrane, are resistant to fouling but experience changes in physical properties over time. The bubble diameters for fine bubble aeration and coarse bubble aeration range from 1.5 to 3.0 mm and from 20 to 40 mm (Mueller et al. 2002).

As outlined by Mueller et al. (2002), there are five principal types of fine bubble diffusers based on shape: plates, tubes, panels, domes, and discs The former two are widely used in wastewater treatment.. The flux ranges for these diffusers are 0.09–0.18, 0.16–15.7, 0.007–0.111, 0.32–7.8, and 0.27–9.4 $m^3_N/h/m^2$, respectively. In a study conducted by Cheng et al. (2016), the effect of four different fine diffuser shapes (I-shape, C-shape, S-shape and Disc-shape) on oxygen transfer efficiency was observed. The result of this study revealed that the I-shaped diffuser exhibited the highest efficiency.

In terms of the orifice size of the aeration system, a smaller orifice size results in a decrease in bubble size, which in turn increases the K_La and standard oxygen transfer rate due to a greater surface area per unit volume and increased contact time. This is evidenced by a study by Da Silva et al. (2019), which found that small bubbles produced by the aerator with smaller pores (16–40 μm) were more suitable for MBR systems by maintaining higher stability, in addition to offering superior antifouling activity, when compared to aerators with larger pores (100–160 μm). The superiority effect of smaller bubble size on oxygen transfer is a general consensus, as evidenced by other studies, including those by Herrmann-Heber et al. (2021) , Lu et al. (2022), Suwartha et al. (2020) and Wang et al. (2020). However, Wild et al. (2023) demonstrated that improved oxygen transfer is attributed to an increase in the liquid mass transfer coefficient (k_L), which is triggered by larger bubbles rather than the specific interfacial area of bubble. This is supported by findings from Collivignarelli et al. (2019), who showed that the fine bubble system did not show significant advantages over the coarse bubble system in terms of oxygen transfer efficiency.

Furthermore, within the aeration system, the airflow rate exerts a significant influence on the shape, rise velocity, and turbulence of the bubbles. In the case of fine bubble diffusers, the size of the bubbles is determined by the airflow rate and the orifice size. As the airflow rate increases, larger bubbles are formed. An increase in the airflow per diffuser results in larger bubble diameters and a smaller area-to-volume ratio, which in turn leads to a decrease in oxygen transfer efficiency (OTE) (Campbell and Boyle 1989; Déronzier et al. 1998; Gillot 2000a; Iranpour and Stenstrom 2001). Nevertheless, elevated aeration rates typically result in elevated oxygen transfer rates OTR (JI and Zhou 2006) due to the turbulence generated by the mixing process, which reduces the boundary layer depth and improves the oxygen transfer coefficient, and the increase in the amount of air in contact with water per unit time.

Other factors to consider in aeration characteristic are diffuser density and submergence (depth). The diffuser density refers to the area covered by diffusers relative to the total tank floor area. A higher density of diffusers with uniform spacing in a full coverage configuration has the effect of reducing bubble velocity to its terminal velocity, which in turn maximises bubble retention time and improves tank agitation and mixing. This, in turn, enhances OTE significantly (Al-Ahmady 2006; Campbell and Boyle 1989) and increases specific oxygen absorption (Wagner and Pöpel 1998). Furthermore, the impact of depth on aeration tanks with regard to oxygen transfer has been investigated. Some studies have indicated that increasing depth enhances the oxygen transfer process (Fernández-Álvarez et al. 2014; Al-Ahmady 2006; Sandberg 2010). This is because a greater reactor depth increases the contact time between water and air bubbles, thereby improving OTE. However, deeper tanks may result in a reduction in the α-factor due to the accumulation of contaminants on bubble surfaces. Nevertheless, it is also important to consider the findings of a study by Fernández-Álvarez et al. (2014), which concluded that, based on a global cost analysis comparing the savings in aeration equipment with the increasing civil work costs for increasing depths, the optimum depth is 6.5 m. Beyond this depth, the savings are minimal.

2.4.4. Interrelation between factors affecting oxygen transfer

The preceding explanations indicate that several factors from three main categories—biomass characteristics, aeration characteristics, and operational characteristics—are partially intercorrelated. This is depicted in Figure 2.3. Consequently, it is important to note that modifying one factor can lead to changes in other factors. Therefore, system modification and optimisation must be conducted by thoroughly considering all influencing factors.

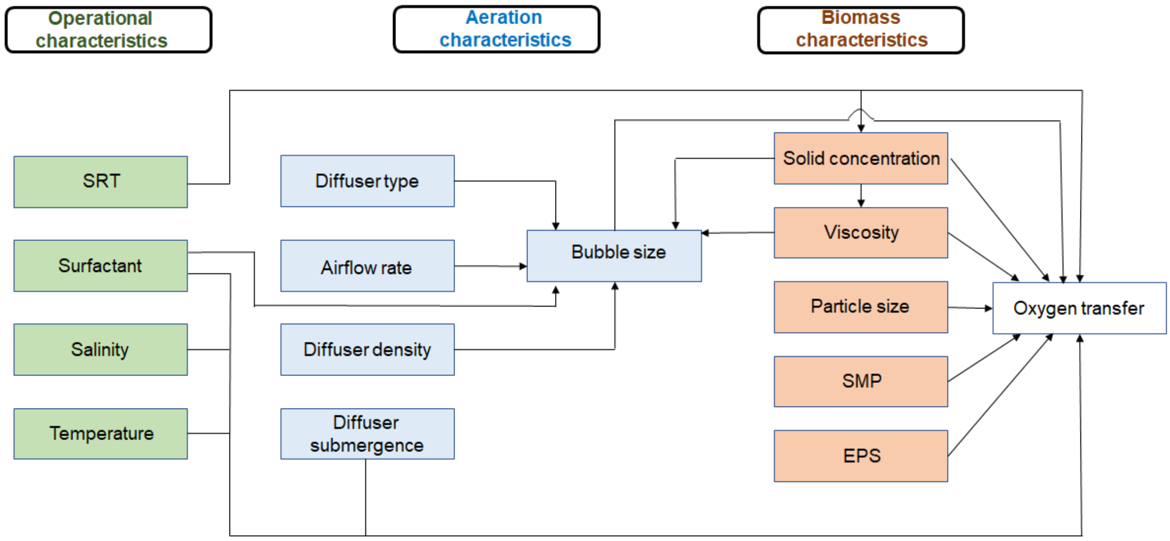


Figure 2.3. Interrelation between factors affecting oxygen transfer in MBR system

2.5. Membrane Fouling in MBR System

The phenomenon of membrane fouling can be observed through a reduction in permeation flux or an elevation in transmembrane pressure (TMP). Membrane fouling occurs when TMP increases to maintain a constant flux, or when permeate flux declines under the constant pressure mode (Park et al. 2015). It represents a significant challenge in MBR operation, in addition to the transfer of oxygen. The most significant impact on the efficiency of an MBR system can be achieved by addressing or mitigating this issue (Al-Asheh et al. 2021). Fouling occurs when the performance of the membrane deteriorates due to the deposition of dissolved and/or suspended substances on its surface, in its openings, or within its pores (Iorhemen et al. 2016). The deterioration of performance is manifested in a reduction in operating flux, an increase in feed pressure, a reduction in productivity, an increase in system downtime, and a requirement for frequent membrane maintenance and replacement. Therefore, controlling membrane fouling is essential for the stable operation of an MBR. Given the critical importance of membrane fouling to MBR performance, a significant amount of research has been conducted on membrane materials and the development of new processes to understand and reduce membrane fouling (Judd 2011), in addition to studies on the mechanism and mitigation of membrane fouling.

2.5.1. Mechanism and classification of membrane fouling

The explanation of the sources and mechanism responsible for membrane fouling is a topic that has been extensively discussed in numerous literatures. Radjenović et al. (2008) identified four primary factors for membrane fouling: the adsorption of macromolecular and colloidal matter, biofilm growth on the membrane surface, precipitation of inorganic matter, and membrane aging. In practice, these mechanisms often occur simultaneously, with the specific fouling mechanism dictating the location of fouling. Furthermore, Tchobanoglous (2003) and Iorhemen et al. (2016) described several mechanisms that can lead to fouling. The first mechanism is the sorption of foulants with a smaller size than the membrane pore size, which results in pore narrowing. The second mechanism is the deposition of foulants having a similar size to the pore size, which results in pore plugging. The final mechanism is the formation of a cake layer by foulants on the top of the surface of the membrane. In addition to these mechanisms, Radjenović et al. (2008) proposed intermediate clogging, which is caused by the blockage of pores by foulants having a smaller size than the pore size.

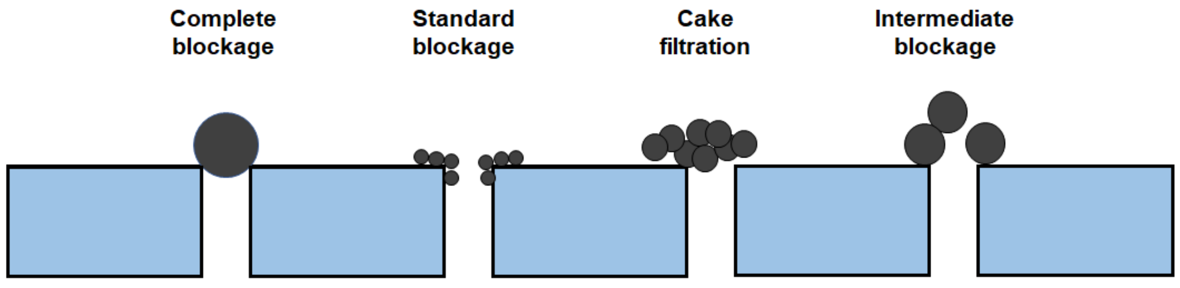


Figure 2.4. Types of membrane blockage, adapted from Radjenović et al. (2008)

With regard to the extent of membrane pore blockage, Radjenovic et al. (2008) identified four distinct types of blockage that result in membrane fouling (see Figure 2.4), which are:

- Complete blockage. This occurs when pores are fully occluded by particles, with no particle superimposition.
- Intermediate blockage. This occurs when pores are occluded by particles with particle superimposition.
- Standard blockage. This type of blocking occurs when particles smaller than the membrane pore size deposit onto the pore walls, thus reducing the pore size.
- Cake filtration. This occurs when particles larger than the membrane pore size deposit onto the membrane surface.

Furthermore, Gkotsis et al. (2014) defined membrane fouling as the collective resistance at the membrane-solution interface. This resistance can be increased by a number of factors, including the concentration of rejected solutes in the vicinity of the membrane surface; the precipitation of sparingly soluble macromolecular polymers and inorganic compounds; and the accumulation of retained solids on the membrane, resulting in the formation of a cake layer. All of these factors contribute to membrane fouling, with the first two factors being exacerbated by concentration polarization (CP), which refers to the tendency of solutes to accumulate at the membrane-solution interface within a concentration boundary layer or liquid film during crossflow operation.

In accordance with these factors, Du et al. (2020) classified membrane fouling into three types: internal fouling, external fouling, and CP. Internal fouling, or pore blocking, is defined as the deposition of solutes and colloidal particles within the membrane pores. External fouling is characterised by the deposition of particles, colloids, and macromolecules on the membrane surface, resulting in the formation of a gel layer (due to pressure differences affecting macromolecules and colloids) or a cake layer (from solids accumulation).

Another crucial distinction in membrane fouling is the differentiation between reversible and irreversible fouling. In this study, four distinct categorisations for membrane fouling based on the capability of the membrane to revert to its original state according to Du et al. (2020) and Judd (2011) are utilised: reversible, residual, irreversible, and permanent fouling, as depicted in Table 2.1. Reversible or removable fouling, such as cake fouling, is caused by loose contamination and can be eliminated using physical processes such as backwashing or intermittent membrane operation in crossflow filtration. Both residual and irreversible fouling result from pore blockage and firmly adhering contamination (Meng et al. 2009). While residual fouling can be removed by low-dose intermediate cleaning using chemically enhanced backflush (CEB), irreversible fouling requires highly concentrated chemical cleaning for removal. Meanwhile, permanent or non-recoverable or irreparable fouling cannot be removed by any means and is considered permanent contamination. Figure 2.5 illustrates the development of fouling over time.

Table 2.1. The rate and onset of fouling, as described by Judd (2011) and Du et al. (2020)

Fouling type	Rate of fouling (mbar/min)	Onset of fouling
Reversible fouling	0.1 - 1	10 minutes
Residual fouling	0.01 - 0.1	1 – 2 weeks
Irreversible fouling	0.001 - 0.01	6 – 12 months
Permanent fouling	0.0001 - 0.001	Several years

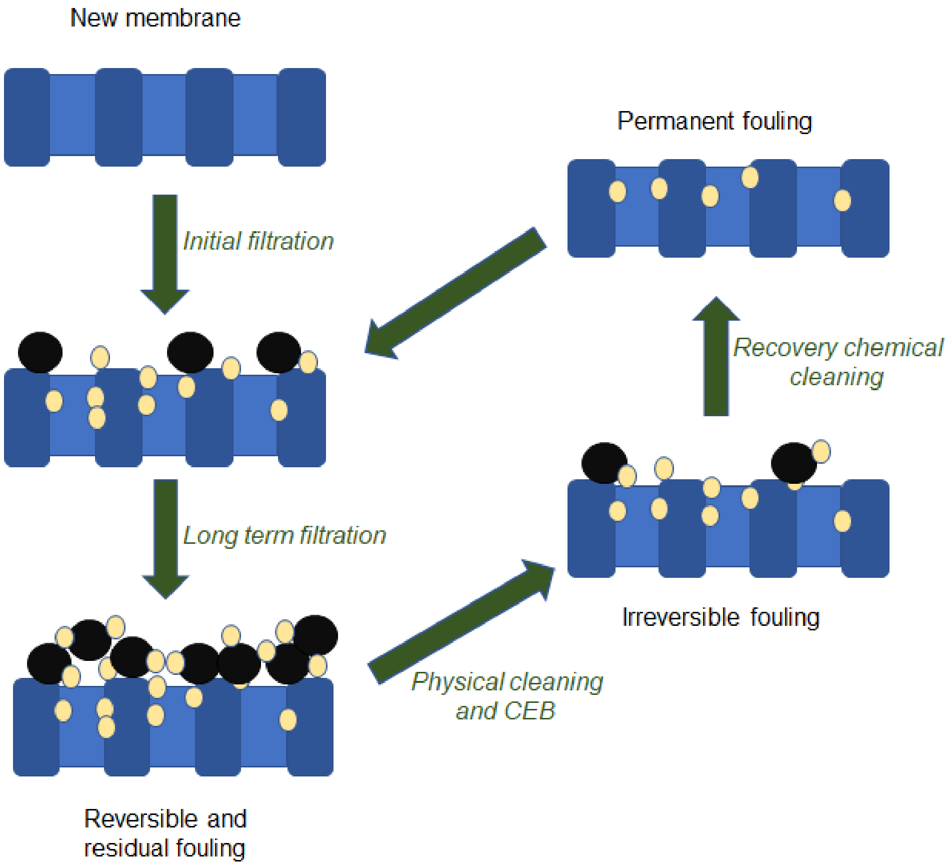


Figure 2.5. The development of fouling over time, adapted from Ladewig and Al-Shaeli (2017) according to definition by Judd (2011).

Judd (2011) described the fouling process as occurring in three stages: conditioning fouling, steady fouling, and TMP jump. In the initial stages of fouling, EPS and SMP interact with the membrane, causing flocculant material to adhere without the formation of a conventional cake layer. In the steady fouling stage, the temporary attachment of flocculent material and the presence of soluble microbial products lead to pore blocking and uneven fouling due to the inconsistency of the flow distribution. Finally, in the TMP jump stage, an increase in permeation in less fouled regions causes flux to exceed the critical threshold, resulting in a sudden rise in transmembrane pressure. This is due to the accelerated growth of the fouling rate.

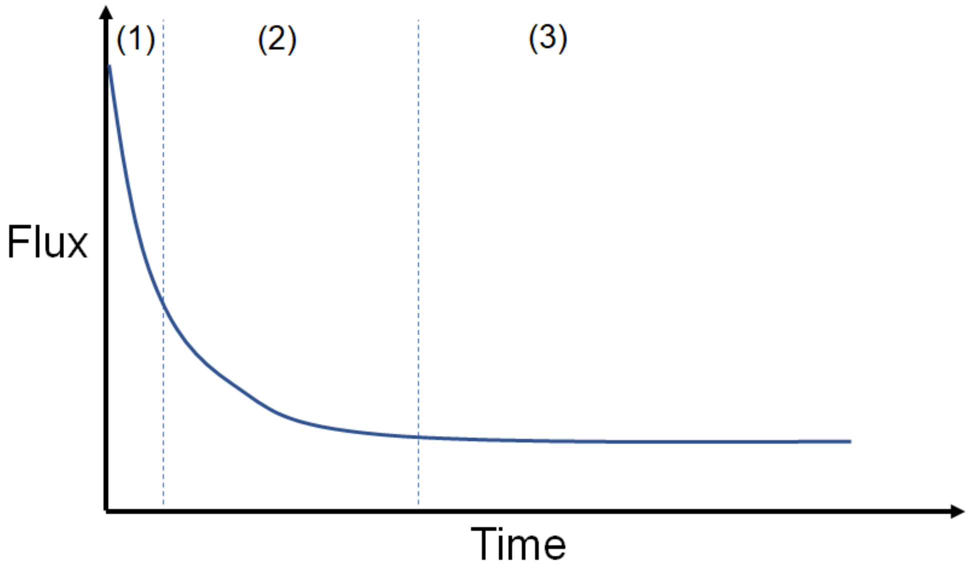


Figure 2.6. A Schematic presentation of the three stages in flux decline due to fouling, adapted from Abdelrasoul et al. (2013)

In accordance to this, Abdelrasoul et al. (2013) explained flux decline as the results of membrane fouling development, as shown in the Figure 2.6. The flux decline process initiates with a rapid initial decline in permeate flux, which is attributed to the rapid blocking of membrane pores (stage 1). This is followed by stage 2, which is characterized by a prolonged period of gradual flux decrease. This is due to the formation and growth of a cake layer on the membrane surface, which creates additional resistance to permeate flow. Finally, the process reaches a steady state flux once the cake layer reaches a point where its thickness stabilises and no longer significantly impacts the flux. The rapid initial drop in permeate flux is due to the quick blocking of membrane pores, which initially allows for maximal flux because the pores are clean and open. Flux declines as the pores become blocked by retained particles, with the degree of blockage depending on the shape and size of both particles and pores. Following pore blockage, a further decline in flux is caused by the formation and growth of a cake layer on the membrane surface, which adds resistance to permeate flow. This process continues to decrease flux over time as the layer thickens.

2.5.2. Factors influencing membrane fouling

Fouling is the result of the interaction between foulants, which include particulates, colloidal particles, biomacromolecules, and other substances present in separation solutions, and the membrane surface. These foulants can be classified as organic, inorganic, or biological, and they can exist in various forms (Huang et al. 2012; Le-Clech et al. 2006; Rana and Matsuura 2010; Ladewig and Al-Shaeli 2017). The foulants interact with the membrane surface in both physical and chemical ways, with the chemical interactions leading to the degradation of the membrane material. A considerable number of studies have demonstrated the effects of numerous factors on membrane fouling. For instance, as listed by Du et al. (2020), Judd (2011) and Ladewig and Al-Shaeli (2017). And most of these factors are interconnected. This study will concentrate on selected main factors, with a particular focus on biomass characteristics and operational conditions.

Some studies have indicated that an elevated MLSS concentration may accelerate membrane fouling and reduce membrane permeability, such as evidenced by Lee and Kim (2013) and Mohan and Nagalakshmi (2024). However, Gkotsis and Zouboulis (2019) argued that there is only a weak correlation between MLSS concentration and membrane fouling, particularly within moderate concentration ranges (e.g., 3.5–8.5 g/L). Nevertheless, it is advisable to avoid values above 20–30 g/L, as they can cause accelerated fouling by increasing viscosity, which hinders air diffusion (aeration) in the mixed liquor. This result is supported by the findings of Wu and Huang (2009), who found that the correlation between MLSS and decreased membrane permeability exists only at MLSS concentrations above 10 g/L, with no correlation below this concentration. Furthermore, Rosenberger et al. (2005) observed that the influence of MLSS on fouling is inconsistent. They noted that MLSS below 6 g/L had a minimal effect on fouling, no effect between 8 and 12 g/L, and increased fouling at levels above 15 g/L.

Furthermore, it has been demonstrated that an increase in the MLSS concentration will typically result in an increase in the viscosity of the mixed liquid (Du et al. 2020). As previously observed by Moreau et al. (2009), MLSS is the main factor influencing sludge viscosity. However, they found that sludge viscosity did not significantly impact fouling reversibility or membrane performance in terms of permeability and filterability. However, a study by Pradhan et al. (2022) demonstrated that a threefold increase in viscosity due to the addition of glycerol resulted in a threefold higher TMP development. This is corroborated by the findings of Ueda et al. (1996), which demonstrated that an increase in sludge viscosity resulted in a significant rise in suction pressure within the MBR system. Moreover, Trussell et al. (2007) also identified an association between sludge viscosity and decreased permeability. It is noteworthy that below a certain range of MLSS concentration, which is between 10 and 17 g/L, sludge viscosity remains low and increases gradually; however, above this range, viscosity increases exponentially (Le-Clech et al. 2006; Itonaga et al. 2004).

A number of studies have also identified EPS and SMP as the cause of membrane fouling (Huang et al. 2000; Lesjean et al. 2005; Liu et al. 2005; Ladewig and Al-Shaeli 2017; Chang et al. 1999). Gkotsis and Zouboulis (2019) demonstrated that SMP, as the soluble fraction of EPS, is the most significant foulant. Of particular importance is the carbohydrate fraction of SMP (SMPc), which exhibits hydrophilic and gelling properties that facilitate membrane attachment. Furthermore, the findings of Hwang et al. (2012) demonstrated that EPS in the form of polysaccharides exerts a profound impact on filtration performance. They create a more compact filter cake with enhanced resistance and adsorb onto membrane pore walls, reducing pore size and causing fouling, which is supported by the findings from Banti et al. (2020). EPS and SMP promote microbial adhesion to the membrane surface due to their adhesive properties. These substances act as building materials for biofilms and sludge flocs, thereby reducing membrane permeability (Malaeb et al. 2013; Meuler-List 2020; Iorhemen et al. 2016). The increase of EPS can be caused by the chosen disinfection methods (free chlorine, chloramine, and ozone). This is demonstrated in the study by Chen et al. (2022b), who observed an increase in EPS due to disinfection and a fouling potential on UF and RO membranes.

The floc size also affects membrane fouling in MBR system. A reduction in floc size results in an increase in both the specific energy barrier and the specific interaction energy, making smaller sludge flocs more adhesive and thus facilitating fouling, as described in the results of a study by Shen et al. (2015). Furthermore, they showed that smaller flocs increase the hydraulic resistance of the cake layer and the resistance caused by osmotic pressure. Consequently, increasing floc size can reduce fouling. It triggered some studies with objective of achieving larger flocs, such as using aerobic granulation (Tay et al. 2007; Zhao et al. 2014), the addition of activated carbon (Baêta et al. 2016; Remy et al. 2009) and the addition of zeolite (Rezaei and Mehrnia 2014; Park et al. 2016).

Furthermore, several studies have demonstrated the influence of the pH value on membrane fouling. Zhang et al. (2014) investigated the effect of pH on membrane fouling in a submerged MBR and indicated that lower pH levels led to higher adherence of sludge flocs on the membrane surface, which in turn decreased the repulsive energy barrier and facilitated foulant attachment. The results are corroborated by findings from Nazmkhah et al. (2022), which demonstrated that the application of sweep coagulation and a neutral pH effectively reduced SMP. Other studies, such as those conducted by Sanguanpak et al. (2015), Sweity et al. (2011) and Wang et al. (2012) , have reached similar conclusions, identifying that a reduction in the pH of the mixed liquor can lead to an increase in the rate of membrane fouling in MBR.

In accordance with the aforementioned findings, Sweity et al. (2011) investigated the effect of pH on pore blockage of a UF membrane by EPS. Their findings indicated that the adsorption of EPS on the membrane surface increases at low pH values due to the generally negatively charged membrane surface, such as with polyvinylidene difluoride (PVDF) membranes. However, the fouling rate was significantly lower at a pH of 6.3 compared to a pH of 8.3. Additionally, they observed a correlation between increased pH and reduced membrane permeability. Wang et al. (2012) further supported these findings, showing that the tendency of EPS to flocculate was highest at a pH of 4.8, thereby reducing the fouling rate. Conversely, Sanguanpak et al. (2015) observed that in an immersed MBR process with landfill leachate, fouling was most severe at a low pH of 5.5 due to increased EPS formation. However, this organic fouling could be effectively removed using chemical cleaning.

A number of studies have indicated that increased membrane fouling is associated with decreasing temperatures (Chen et al. 2022a; Gao et al. 2022; Ma et al. 2013; Theuri et al. 2023; van den Brink et al. 2011). In addition, Alresheedi and Basu (2019) demonstrated that increased fouling irreversibility and reduced effectiveness of backwashing and chemical cleaning were observed at lower temperatures in ceramic ultrafiltration membranes. A further study by Ren et al. (2019) on ultrafiltration of alkali/surfactant/polymer (ASP) flood wastewater revealed that lower temperatures led to an increase in feed water viscosity and a reduction in diffusion capacity. Furthermore, the increase in fouling can result in a decrease in flux, as demonstrated by Yu et al. (2022), who observed a 36 % reduction in critical flux at lower temperatures when comparing solution temperatures of 15 °C and 35 °C. However, in contrast to this result, a study by Chu et al. (2016) on reverse osmosis (RO) membrane filtration demonstrated that lower feed water temperature (19 °C vs. 25 °C) resulted in milder membrane fouling. A mixed result was obtained by Al-Mutwalli et al. (2022), who conducted experiments on ceramic ultrafiltration membranes for the treatment of acid whey. The study observed that higher temperatures resulted in increased membrane fouling for disc modules, whereas tubular modules experienced less fouling at higher temperatures. Nevertheless, the consensus is that lower temperatures have a negative effect on membrane filtration, as summarised by Xu et al. (2023a), who conducted a review of the effects of temperature on membrane filtration. The study concluded that low temperatures have a detrimental effect on the membrane structure, resulting in increased fouling and reduced chemical cleaning efficacy, which in turn impairs the performance of membrane filtration.

Salinity is also defined as a factor influencing membrane fouling in wastewater treatment. While salt has a detrimental impact on biological systems, its presence in MBR systems leads to electrostatic attraction and chemical precipitation on membrane surfaces (Elimelech et al. 1997). For instance, Jang et al. (2013) demonstrated that high salinity, characterised by NaCl concentrations of 5 to 20 g/l, increased membrane fouling by promoting pore closure. Moreover, Di Bella et al. (2013) demonstrated that high salinity has a detrimental effect on biokinetics and fouling, primarily due to irreversible deposits in the cake layer resulting from deteriorated sludge properties and high EPS concentration. Additionally, the ionic composition of wastewater significantly affects floc structure and strength, contributing to fouling.

Sludge bulking can also lead to membrane fouling, as demonstrated by Wu et al. (2023). Their study investigated the fouling behaviour of an MBR system during periods of normal operation and sludge bulking, and explored the underlying mechanisms of these different behaviours.

The study found that under normal conditions, the MBR could operate stably for approximately 60 days without the need for membrane cleaning. Nevertheless, during the sludge bulking phase, it was necessary to perform daily membrane cleaning in order to maintain the operational stability of the system.

2.5.3. Fouling mitigation strategies

The complex nature of membrane fouling necessitates the implementation of a variety of strategies to mitigate its effects. Various strategies exist to address membrane fouling, which can aim to mitigate fouling formation and to target fouling removal. Reduction strategies aim to slow fouling formation, with a common approach being the application of shear force to the membrane surface. In submerged MBR plants, this is often achieved through the use of membrane aeration units positioned beneath the membrane, which disrupt solids deposition. Increasing turbulence within the MBR via mechanical movement, including concepts such as moving or rotating membrane modules, has also demonstrated potential in reducing fouling (Wu et al., 2008; Ruigomez et al., 2022). Ladewig and Al-Shaeli (2017) proposed a number of different strategies to mitigate fouling, including: pre-treatment of the feed, optimisation of operating conditions, cleaning procedures and the modification of the surface of the membrane.

The pre-treatment of feed material represents a pivotal aspect of the mitigation of membrane fouling. This process employs a combination of physical and chemical techniques to remove particulates and macromolecules that clog pores or deposit on the membrane surface (Ladewig and Al-Shaeli 2017). Physical methods, such as prefiltration, centrifugation, and heat treatment followed by settling, are effective in removing suspended particles, especially in dairy plants (Aptel and Clifton 1986). The addition of ferric chloride and alum can improve permeability and effluent quality in MBR systems. These substances are capable of forming large aggregates from initial molecules, thereby reducing membrane fouling (Holbrook et al. 2004). Furthermore, a study by Dong et al. (2007) showed that coagulation could effectively remove the hydrophobic organics, resulting in an increase in flux. Compared to alum, ferric chloride is a more potent coagulant enhancing the formation of rigid flocs and iron-oxidising bacteria, however, it is also more costly (Park et al. 2005).

Another strategy for mitigating membrane fouling is to enhance the operating conditions at the membrane surface. This can be achieved by increasing the cross-flow velocity and shear stress. This approach enhances mass transfer efficiency and turbulence, thereby reducing concentration polarization. One method for achieving this is through the use of aeration, which has the effect of scouring the membrane surface. Furthermore, this can be achieved by optimising the SRT. To enhance filtration performance, a long SRT is employed, which minimises the formation of SMP and EPS by creating starved conditions (Judd 2011).

However, it is important to note that an excessively long SRT can lead to severe membrane fouling due to the accumulation of MLSS or increased sludge (filamentous) production (Liu and Liu 2006). Another method to mitigate membrane fouling is to determine the sustainable flux, which varies between different MBR systems. An MBR system is economically viable only if it maintains a reasonable flux rate without significant fouling. In order to manage membrane fouling, the majority of MBR systems operate at low fluxes. In general, the sustainable flux is considered to be subcritical flux by default. In MBR systems, the sustainable flux, or sometimes referred to as the sustainable critical flux, is defined as the flux at which the TMP increases gradually at an acceptable rate, allowing long-term operation (Xu et al. 2023b). Therefore, in order to assess the sustainable flux, extended filtration periods are required (Ladewig and Al-Shaeli 2017).

Furthermore, the cleaning process is of great importance for the regeneration of membranes and the maintenance of their effectiveness. Cleaning should be performed when the flux decreases slightly and the transmembrane pressure increases significantly (Ladewig and Al-Shaeli 2017). There are three main forms of cleaning: physical, chemical, and a combination of both. Physical and chemical cleaning methods are employed to remove foulants, such as microbial flocs, from the membrane surface. Physical cleaning is conducted by backflushing and relaxation (allowing the membrane to idle for a short time) during membrane is in operation. Meanwhile, chemical cleaning is performed when the membrane is not in operation and is conducted by two different cleaning protocols: maintenance cleaning and recovery cleaning (Judd 2011; Park et al. 2015). The objective of maintenance cleaning is to maintain membrane permeability and thereby reduce the frequency of intensive cleaning. This is achieved by either maintaining the membrane in situ, conducting a clean in place (CIP) procedure, or draining the membrane tank (referred to as cleaning in air, or CIA). Recovery cleaning, or intensive cleaning, is either conducted ex situ or in the drained membrane tank, allowing the membranes to be soaked in the cleaning reagent, with sodium hypochlorite being a widely used reagent (Judd 2011). It is important to note, that chemical cleaning should be kept to a minimum, as repeated chemical cleaning can shorten membrane life, degrade the membrane, and the disposal of spent cleaning reagents causes environmental problems (Ladewig and Al-Shaeli 2017; Yamamura et al. 2007a; Yamamura et al. 2007b; Yamamura et al. 2007c).

Another approach to prevent fouling in MBR systems and enhance membrane permeability is chemical modification of the membrane surface. Membranes with hydrophobic properties are more prone to severe fouling, prompting numerous studies on making membranes more hydrophilic (Ladewig and Al-Shaeli 2017). For example, the addition of a coating layer on the membrane surface has been demonstrated to enhance antifouling performance in a number of studies, including those by Kasemset et al. (2016), who added a polydopamine layer to a polysulfone membrane. In addition, Li et al. (2018) added a dopamine coating to a PVDF membrane, while Kumar and Jaafar (2018) added a titanium dioxide electrospun nanofiber coating to a PVDF membrane. Alternatively, polymer blending is a potential solution. This process involves physically mixing two or more compounds into a single solution using the same solvent (Díez and Rosal 2020). This polymer blending has led to several positive results, including an improvement in membrane permeation flux, hydrophilicity, porosity, and permeability (Mujtaba et al. 2019).

3. A New Concept of a Rotating Hollow Fibre Membrane Module: Impact of Rotation on Fine-bubble Aeration.

This chapter has been published as an article in Water Science and Technology:

Mahdariza, Fathul; Domingo Rimoldi, Ignacio; Henkel, Jochen; Morck, Tobias (2022): A New Concept of a Rotating Hollow Fibre Membrane Module: Impact of Rotation on Fine-Bubble Aeration. In Water Science and Technology: A Journal of the International Association on Water Pollution Research 85 (9), pp. 2737–2747. DOI: 10.2166/wst.2022.144.

Abstract: A new concept of a rotating membrane module in a membrane bioreactor (MBR) system was tested for its effect on oxygen transfer in clean water and wastewater. The membrane module consists of horizontally aligned hollow fibres connected to the vertically positioned permeate tube which rotates. The results indicated that oxygen transfer can be improved by up to 50% at the highest applied rotational speed (50 rpm) and that the additional energy demand required for the rotation can be compensated by the enhanced oxygen transfer. However, at the highest rotational speed (50 rpm), the fine bubbles bypassed the MBR module, and, consequently, could not contribute to any cleaning effect. The α-factors at different rotational speeds showed similar results. This indicates that the depletion was caused neither by surfactants nor by viscosity phenomena but rather by the floc/solid holdup of the sludge.

3.1. Introduction

The use of MBR systems becomes a growing interest due to tighter environmental regulations, water scarcity and the rising need of water reuse (Judd 2016; Paul and Jones 2016). The MBR system offers the ability to achieve high nutrient removal efficiency and complete biomass retention at a small footprint (Hoinkis et al. 2012; Meng et al. 2012). However, MBR systems also suffer from membrane fouling and clogging, which impacts the performance of the system significantly and makes frequent cleaning procedures inevitable. In addition, MBR systems are still considered expensive in comparison to CAS systems. The average energy consumption of immersed MBRs per m^3 of treated municipal wastewater ranges between 0.8 - 1.1 kWh/m3 with the lowest reported value being 0.4 kWh/m^3 (Judd 2011; Krzeminski et al. 2017).

For both MBR and CAS systems, aeration is responsible for more than 50% of the energy demand (Helmi and Gallucci 2020; Judd 2011). The overall aeration demand is typically higher for MBR than for CAS systems because the higher sludge concentration (floc / solid holdup) lowers the oxygen transfer rate and an additional air cross flow is required to reduce membrane clogging and fouling (Henkel et al. 2011).

Several studies were conducted to improve the performance of MBR in terms of fouling reduction and aeration efficiency (Helmi and Gallucci 2020; Judd 2016). One possible method is the introduction of a dynamic shear force through rotation (Jaffrin 2008; Jiang et al. 2013; Ruigómez et al. 2016; Wu et al. 2008; Zsirai et al. 2016). Rector et al. (2006) did experiments on a rotating membrane bioreactor using a HF module. They concluded that the rotation of HF membrane modules provides an efficient method of prompting turbulent flow and higher dispersion in an MBR system. Paul and Jones (2016) also showed that rotating MBR systems had better performance than static MBR when both systems were operated under similar conditions.

This paper introduces a novel membrane configuration where horizontally aligned HFs are connected to a vertical permeate tube. The membrane module rotates in the activated sludge reactor. The rotation creates a permanent shear force at the fibre and eliminates the need of a uniformly distributed aeration device below the membrane unit. In comparison to current membrane systems, the fibre length is relatively short (< 20 cm) which is expected to improve the backwash and cleaning process significantly.

The study was conducted to investigate the impact of different rotation velocities on oxygen transfer in a conventional activated sludge plant. Hence a fine bubble aeration system and activated sludge from a nearby conventional activated sludge plant were used. In order to determine the impact of sludge flocs on aeration performance also clean water tests were performed to evaluate the ratio of sludge to clean water oxygen transfer coefficients (α-factor). Finally, the rotation energy was calculated and put into relationship with the oxygen transfer measurements.

3.2. Material and Method

3.2.1. Membrane prototype

The membrane module (Figure 3.1) consists of 1,950 HF membranes with one side potted horizontally into the permeate tube and sealed at the other side. The material of HF membrane is PVDF with the pore size of 0.3 µm. The total surface area of the membrane module is 2.25 m^2, which leads to a packing density of 81 m^2/m^3. Due to the horizontal membrane alignment and the applied rotation, all membranes are in contact with the air bubbles even though only one segment of the membrane prototype is aerated.

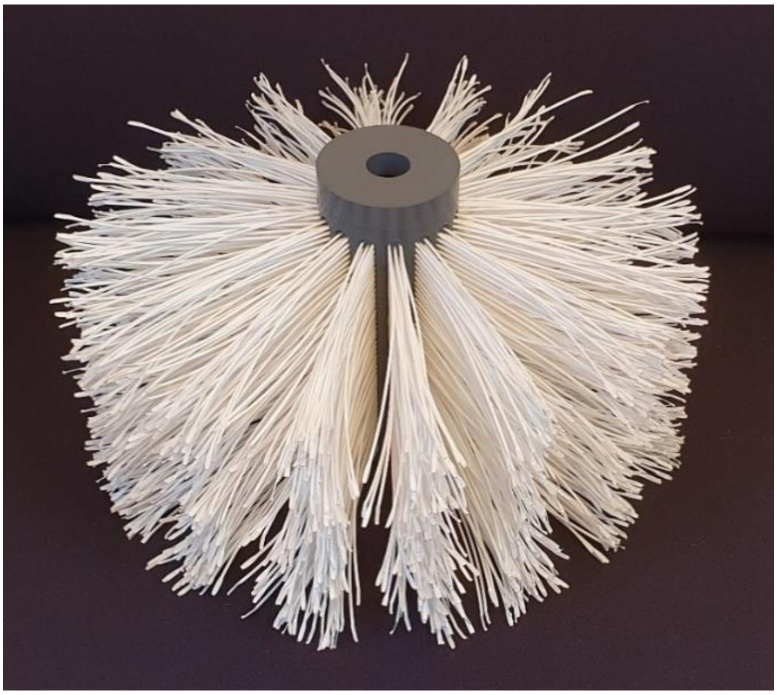


Figure 3.1. Rotatable membrane prototype system with 1,950 horizontally aligned hollow fibres

3.2.2. Experimental setup

The measurements were carried out in a transparent tank filled with clean water (softened tap water) and activated sludge with a volume of 1 m^3. The softened water was chosen to prevent any hardness precipitation during the experiments. The activated sludge was obtained from the Wastewater Treatment Plant in Königsbach, Germany (55,000 PE). The experiments were performed in batch mode. No feed or filtrate were added nor taken during the experiment. A Flexnorm 500 diffuser (OTT System GmbH & Co. KG, Langenhagen, Germany) was used as fine bubble air diffuser. The airflow rates and dissolved oxygen (DO) concentrations were measured with a thermal flow sensor TA Di 21.6 GE (Hoentzsch GmbH, Waiblingen-Hegnach, Germany) and three oxygen electrodes (PRONOVA Analysentechnik GmbH & Co., Bad Klosterlausnitz Germany), respectively. In addition, the ambient parameters (air temperature, pressure and partial humidity) were documented through a weather station. Figure 3.2 shows the setup of the pilot plant.

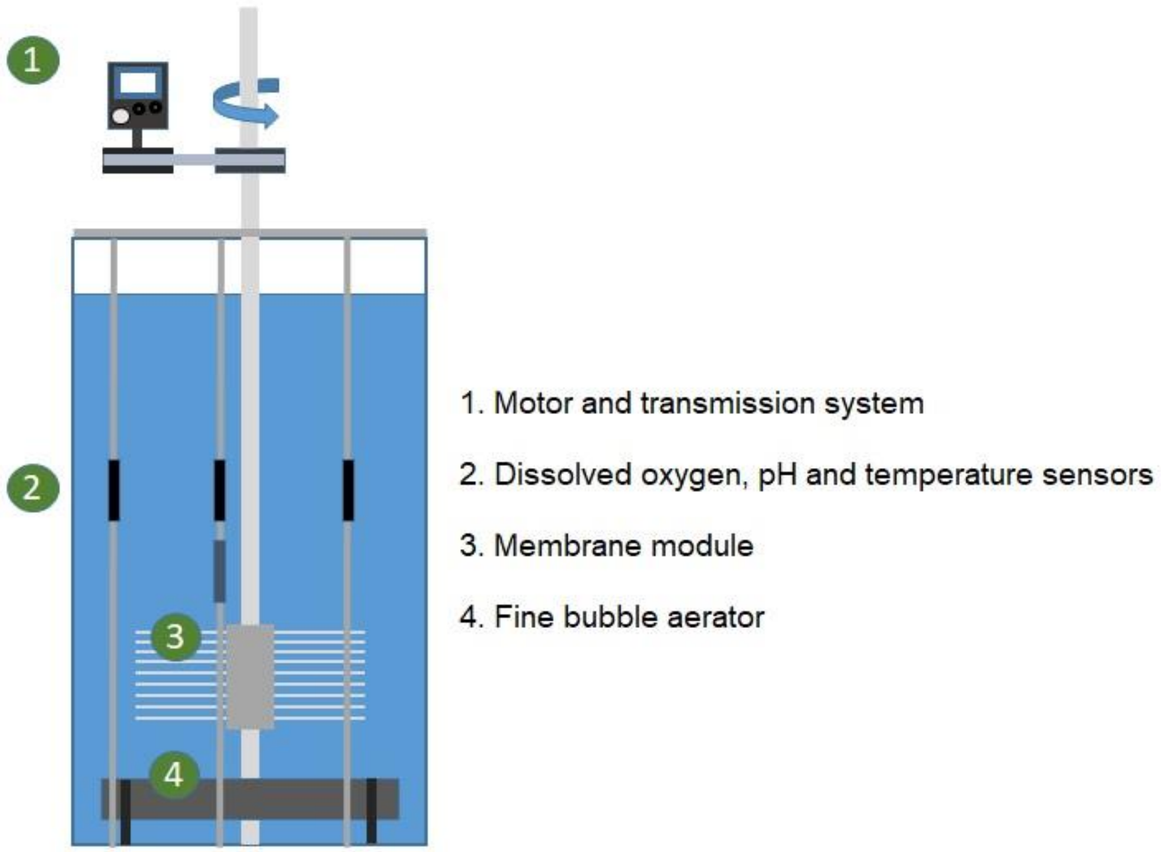


Figure 3.2. Setup of the pilot plant of the rotatable membrane prototype system

3.2.3. Sludge characteristics

Table 3.1 and Table 3.2 below show the sludge characteristics and an example of the floc sedimentation test to define the HFV for the activated sludge during the experiment days. After an initial decrease in MLVSS and HFV during the first 12 hours following activated sludge collection, both values remained relatively stable during experiments, indicating that their impact on the measured oxygen transfer coefficient was stable.

Table 3.1. Sludge characteristics for activated sludge experiments at 0 and 50 rpm.

Working day	Sampling time	MLSS (g/L)	MLVSS (g/L)	HFV (ml/L)	Conductivity (µS/cm)
1	-	-	-	-	-
	Afternoon	4.50	3.15	180	1,540
2	Morning	3.88	2.71	170	1,555
	Afternoon	3.58	2.47	150	1,563
3	Morning	3.50	2.40	160	1,565
	Afternoon	3.60	2.50	175	1,577
4	Morning	3.70	2.50	175	1,585
	Afternoon	3.60	2.40	175	1,590

Table 3.2. Floc sedimentation test for activated sludge experiment.

Elapsed time	Volume ratio (mL/L)	
0 h	1,000	
0 h 30 m	300	
16 h 10 m	175	
24 h 0 m	175	(HFV)
39 h 20 m	175	

3.2.4. Oxygen transfer measurements

As the oxygen concentration in the sludge ranged from 7.2 - 9.9 mg/L and the saturated concentration was up to 10.13 mg/L O_2 during the experiments, the off-gas method was not appropriate in that condition (Krause et al. 2003). Therefore, the desorption method described by Wagner et al. (1998) and DWA-M 209 (2007) was chosen for calculating the k_La value. In this method, the oxygen transfer rate is determined from the decrease in the DO concentration when the water is diffused with normal air. The DO concentration in clean water and in activated sludge was artificially increased above saturated concentration in advance using pure oxygen aeration into the water. Five different rotational speeds (0, 10, 30, 40 and 50 rpm) for clean water and three different rotational speeds (0, 30 and 50 rpm) for activated sludge were tested at three different airflow rates (1, 2, 4 m^3/h). The k_La value was normalized to the standard conditions (k_La_{20}) at a water temperature of 20 ˚C and an atmospheric pressure of 1013 hPa, and with the correction factor for a salt concentration of 1 g/L, due to significant differences between the salt content of clean water and wastewater (DWA-M 209 2007).

During activated sludge experiments, it is compulsory to maintain constant respiration rates. This was achieved by aerating the activated sludge for at least 12 hours before the experiment started. It nullified the impact of degradable surfactants on oxygen transfer. Consequently, the results should mainly reflect the impact of activated sludge flocs on oxygen transfer depletion.

3.2.5. Calculating the α-factor

Three oxygen sensors recorded the change of the oxygen concentration in the sludge at a constant airflow rate and a constant sludge concentration. From these records, the k_La was determined by non-linear regression and an average k_La was calculated. Subsequently, the airflow rate was changed and the procedure was repeated at the same sludge concentration. Three airflow rates were chosen for each experiment series with the same sludge concentration. Afterwards, the three average k_La_{20} values were plotted against the specific airflow rate and trends were defined by polymeric trendlines.

This procedure was performed in clean water and activated sludge. Finally, the comparison between the k_La_{20} value of activated sludge and clean water defined as the α-factor was calculated by dividing the trendline equation at a certain sludge concentration by the equation obtained during the clean water experiment. As x-variable, the three applied specific airflow rates were inserted into the equation and as a result, three α-factors were received for each sludge concentration. This modus operandi was applied since it was not possible to have identical airflow rates for each experiment series. Humidity, air pressure, air temperature and the air blower influenced the airflow rate, which was transformed to standard conditions.

3.2.6. Calculating the energy demand

The energy consumption was calculated from the electrical demand of the blower and the reading from the rotational device. The blower energy consumption followed the equation introduced by the blower manufacturer: Power (W) = 35 x volume flow. According to the reading from the rotational device, it consumed 16.52 W at 30 rpm and 17.27 W at 50 rpm.

3.3. Results and Discussions

3.3.1. Clean water experiments

The results in clean water showed that both specific airflow rates and the rotational speeds of the membrane prototype influenced the k_La_{20} value in direct proportion, see Figure 3.3.

The oxygen transfer rate rises steadily because with increasing airflow rate, the gas holdup increases. As the airflow rates increase, the efficiency decreases slightly due to the creation of bigger bubbles at the flexible orifice and bubble coalescence nearby the diffuser, which is typical of membrane diffusers (Painmanakul et al. 2004). Most likely, the effect of coalescence is additionally favoured through the contact of the bubbles with the rotating membrane module, which acts like a barrier for the uprising bubbles and facilitates the collision of the bubbles. As the bubble size was not investigated in this study, there is no final proof of this assumption.

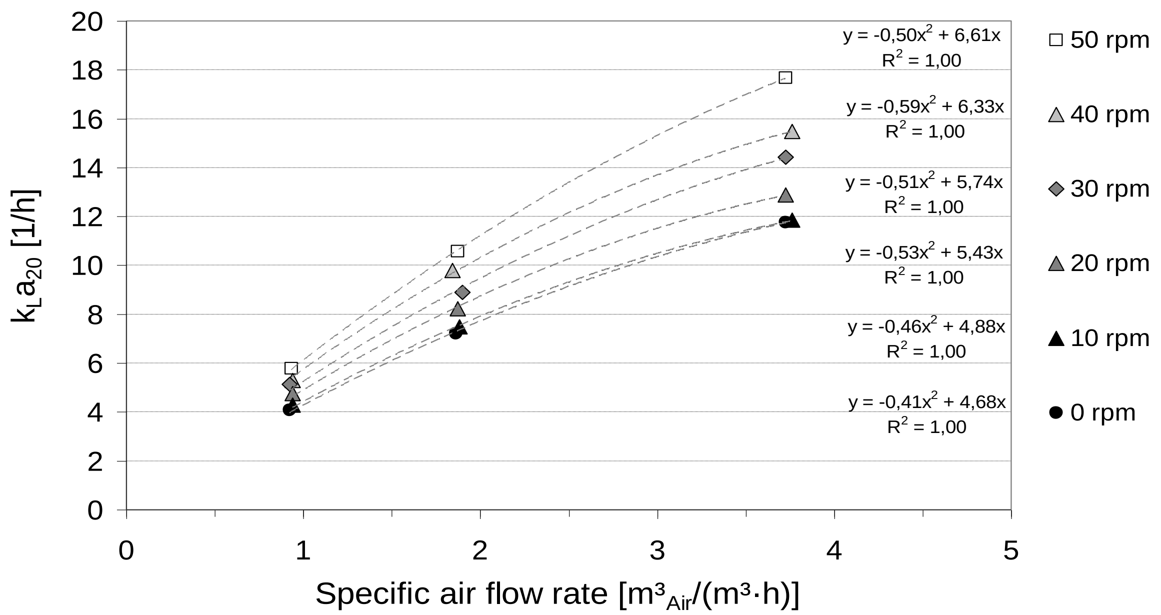


Figure 3.3. The obtained standard oxygen transfer coefficient values by different specific airflow rates and rotational speeds for fine bubble aeration in clean water experiments.

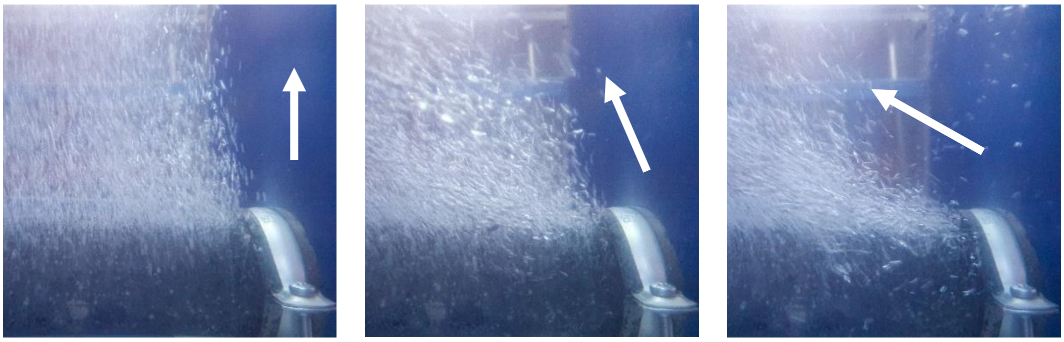
Table 3.3. Improvement of standard oxygen transfer coefficient values due to different rotational speeds for fine bubble aeration in clean water experiments.

Rotational speeds (rpm)	**k_La_{20} value improvement in comparison to no membrane rotation (%)**			
	At specific airflow rates of			**Average improvement**
	1 $m^3/m^3.h$	**2 $m^3/m^3.h$**	**4 $m^3/m^3.h$**	
10	4 %	3 %	0 %	2 %
20	15 %	13 %	9 %	12 %
30	22 %	22 %	22 %	22 %
40	34 %	33 %	31 %	33 %
50	43 %	45 %	52 %	47 %

Furthermore, the rotation of the membrane module leads to an improvement in oxygen transfer, as shown in Table 3.3. With the exception of the 10 rpm experiment, every increase of the rotational speed by 10 rpm raised the oxygen transfer by roughly 10%. The biggest improvement was achieved at the highest specific airflow rate of 4 $m^3/(m^3 \cdot h)$ and 50 rpm with an increase in k_La_{20} of 52% compared to the experiment without rotation. With increasing airflow rate, the improvement in oxygen transfer rate decreased slightly for the same rotational speed except for the experiment at 50 rpm where it increased from 43% to 52%.

The following observations of the effect of rotation of the membrane module were made:

- At a rotational speed of 10 rpm, no effect on bubble formation at the orifice was visible. The bubbles rose straight to the top of the reactor. With increasing rotational speed, this pattern changed. Figure 3.4 shows that at a rotational speed of 30 rpm, the increased fluid force led to an angular raise of the bubbles, which was even more pronounced at 50 rpm. The increased fluid flow force at the orifice stimulated a faster bubble detachment and clearly changed the bubble raise behaviour. Both effects had a positive impact on oxygen transfer because the interfacial area and the residence time of the bubbles were increased.



10 rpm 30 rpm 50 rpm

Figure 3.4. The effect of cross flow at fine bubble aeration for 1 m^3/h

- A second effect was visible at the level of the membrane device. At 0 rpm and 10 rpm, the bubble entered the membrane device and little to no effect on the bubble rising behaviour was visible. With increasing rotational speed, the bubbles experienced an additional horizontal acceleration which led to an increase in the bubble residence time and improved oxygen transfer. However, at a high rotational speed (50 rpm), the effect of horizontal acceleration during bubble formation became the determining factor, thus the majority of bubbles did not rise through the membrane device but bypassed the module. On this account, the enhancement of k_La_{20} with increasing airflow rate at 50 rpm was caused by the reduction of coalescence phenomena triggered by the rotating membrane module. This effect should not prevent the primary intention of crossflow aeration to provide sufficient shear forces to membrane fibres to reduce fouling / clogging of the module. So far open is the question whether the higher rotational speed can compensate/ offset the need for an additional air crossflow to suppress fouling and clogging of the module.

3.3.2. Activated sludge experiments

Additional experiments were performed to investigate the effect of the rotating membrane module on oxygen transfer in activated sludge from a municipal wastewater treatment plant.

Similar to the results obtained in clean water, the results of the oxygen transfer experiments for activated sludge at 0, 30 and 50 rpm also showed that k_La_{20} values improved in direct proportion as both specific airflow rates and the rotational speeds increased (see Figure 3.5). Similar improvement rates caused by rotation as in clean water were measured as well (Table 3.4) due to the same dependencies provided in the activated sludge system.

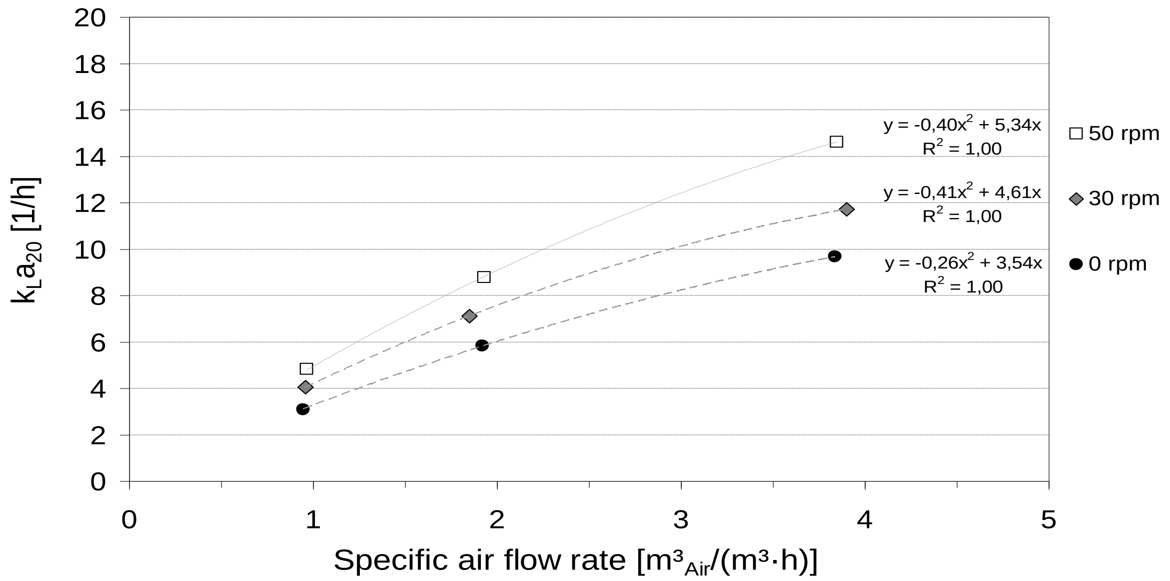


Figure 3.5. Standard oxygen transfer coefficient values for fine bubble aeration in activated sludge experiments

Table 3.4. Improvement of standard oxygen transfer coefficient values due to different rotational speeds for fine bubble aeration in activated sludge experiments

Rotational speeds (rpm)	αk_La_{20} value improvement in comparison to no membrane rotation (%)			
	At specific airflow rates of			Average improvement
	1 $m^3/m^3\cdot h$	2 $m^3/m^3\cdot h$	4 $m^3/m^3\cdot h$	
30	28 %	25 %	19 %	24 %
50	51 %	50 %	50 %	50 %

Table 3.5 shows the obtained α-factors at different specific airflow rates. The α-factors ranged from 0.77 - 0.81, almost independently from rotational speeds and specific airflow rates.

Table 3.5. The obtained α-factors for fine bubble aeration

HFV (mL/L)	MLSS (g/L)	MLVSS (g/L)	Rotational speed (rpm)	α-factor at specific airflow rate		
				1 $m^3/(m^3\cdot h)$	2 $m^3/(m^3\cdot h)$	4 $m^3/(m^3\cdot h)$
175	3.6	2.4	0	0.77	0.78	0.82
175	4.3	2.7	30	0.80	0.80	0.80
150	3.9	2.5	50	0.81	0.81	0.81

A slight increase with rotational speed can be observed for specific airflow rates of 1 and 2 $m^3/(m^3\cdot h)$, while with 4 $m^3/(m^3\cdot h)$, no connection to rotational speed was detected. This observation suggests that the effect that was responsible for the decrease in oxygen transfer by around 20% in activated sludge must act nearly independently from the rotational speed and the applied airflow rate.

Germain et al. (2005) and Krause (2005) argued that one effect that impacts oxygen transfer depletion in activated sludge could be attributed to the increase in apparent viscosity. With this assumption, an improvement in the α-factor with increasing rotational speed can be expected. Because activated sludge is described as a non-Newtonian pseudoplastic fluid (Rosenberger 2003; Yang et al. 2009), an increasing shear stress should have decreased the apparent viscosity of the sludge and thus improved the α-factor significantly. Such effect could not be observed during experiment series with the rotatable membrane module system.

Several other authors (Garrido-Baserba et al. 2020; Rosso et al. 2006; Wagner and Pöpel 1996) argued that dissolved organic matter like surfactants mainly contribute to the depletion of the α-factor in activated sludge. As in the case of apparent viscosity, a positive impact of rotation on the alpha factors can be expected if surfactants are the major contributors to the decrease observed. The higher turbulence introduced by the rotation which clearly led to an improvement in oxygen transfer should have also increased the surface renewal rate of the bubbles and as such decreased the effect of surfactants significantly. Again, such an effect could not be observed during experiment series with the rotatable membrane module system, particularly because the activated sludge was stabilized before the oxygen transfer tests.

Another factor that influences oxygen transfer in a three-phase system is the floc / solid holdup (Deckwer 1992; Henkel et al. 2011; Mena et al. 2005). Already van der Kroon (1968) demonstrated that aluminium hydroxide and activated sludge flocs steadily decreased oxygen transfer with increasing concentration. Henkel (2010) demonstrated that despite different reactor configurations (bubble column, airlift reactor), diffuser systems (fine bubble, coarse bubble, combination of both) and sludges / slurries of different origins (e.g. iron hydroxide flocs, municipal activated sludge, greywater activated sludge), the α-factor showed similar values if the floc volume was used as comparative parameter. Because the different rotational velocities did not show a significant impact on the obtained α-factors, the floc volume seems to be the main driver of oxygen transfer depletion according to present study. The rotational speed indeed impacted the bubble formation and gas holdup which lead to an increase in oxygen transfer compared to the results obtained without rotation, but it did not impact the reduction of the interfacial area of the bubble caused by the sludge flocs.

3.3.3. Analysis of energy demand

In order to investigate if the energy demand required by the rotation of the membranes could be compensated by the improved oxygen transfer rate, the SAE was plotted against the specific airflow rates as shown in Figure 3.6.

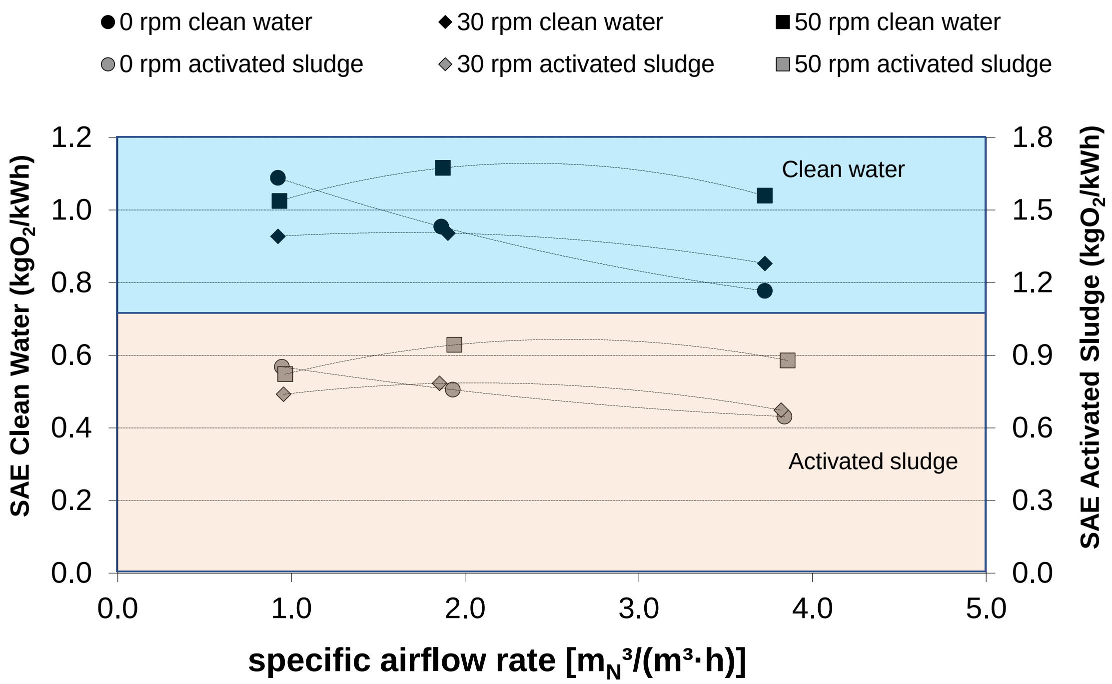


Figure 3.6. Standard aeration efficiency at different specific airflow rates and rotational speeds

Except for clean water experiments at the lowest airflow rate, the SAE was equal or higher compared to the experiments without membrane module rotation. An improvement of 24% at 2 $m^3/(m^3 \cdot h)$ and 36% at 4 $m^3/(m^3 \cdot h)$ was achieved at a rotational speed of 50 rpm if matched against the activated sludge results at 0 rpm. Consequently, the additional energy required for rotation was overcompensated by the improved oxygen transfer efficiency in the activated sludge experiments.

3.4. Conclusions

A set of experiments was conducted with a new rotating type of membrane module suitable for membrane bioreactor applications to study its effect on oxygen transfer and energy consumption. The rotation of the membrane module showed the following effects:

a. oxygen transfer was significantly improved by the application of rotation (~50% at 50 rpm),
b. the rates of improvement in oxygen transfer for clean water and activated sludge were similar,
c. the α-factors showed comparable values for all experiments independent of the rotational speed,

d. an energy comparison indicates that the additional energy demand for rotation can be compensated by the better oxygen transfer efficiency,
e. at the highest rotational speed (50 rpm), the rotation induced such high shear forces to the bubble formation that the majority of bubbles bypassed the membrane module,
f. compared to full-scale systems using fine bubble aeration, the SAE was low (< 1,2 kgO_2/kWh vs. 4 kgO_2/kWh), which can be attributed to the pilot-scale blower and motor for rotation. Consequently, a transfer of these results to full-scale applications is not feasible and requires further investigations.

The proposed membrane module design encourages the possible reduction of total membrane area needed in comparison to existing membrane design available in the market. According to Lo et al. (2015), who studied cost estimation for small membrane bioreactor, the membrane cost contributes to approximately 50-64% to total capital expenditure (CAPEX). Therefore, if the proposed membrane module design can reduce the total membrane area needed by 50%, it has potential to reduce total CAPEX by approximately 30%.

Furthermore, the challenge for a wider use of MBR technology is its integration into an existing CAS system without major modification. The membrane module design in present study shows high feasibility to meet these expectations. Further experiments include the testing of the rotating module under process conditions (filtration, cleaning, fouling behaviour) and the study of the impact of rotation to a coarse bubble aeration system at high sludge concentrations.

3.5. Acknowledgements

The authors want to thank the Ministry of Finance of the Republic of Indonesia for the LPDP scholarship provided to Fathul Mahdariza (Ref. Number: S-1289/LPDP.4/2018). The research did not receive any other specific grants from funding agencies in the public, commercial or not-for-profit sectors.

Moreover, the authors want to thank deeply the following people at Institute for Water and River Basin Management (KIT, Germany) for valuable technical and laboratory assistance: Sven Urbach, Tobias Mumbauer, Manfred Lösche, Patrick Dufner and Maik Erhard.

4. The Impact of Solid/Floc Holdup on Oxygen Transfer in a Rotating Hollow Fibre Membrane Bioreactor under Endogenous Conditions.

This chapter has been published as an article in Water Science and Technology:

Mahdariza, Fathul; Georg, Wilhelm; Wille, Ernst-Marius; Morck, Tobias (2023): The Impact of Solid/Floc Holdup on Oxygen Transfer in a Rotating Hollow Fibre Membrane Bioreactor under Endogenous Conditions. In Water Science and Technology: A Journal of the International Association on Water Pollution Research 88 (5), pp. 1232–1245. DOI: 10.2166/wst.2023.265.

Abstract: A set of oxygen transfer experiments in clean water and three different activated sludge concentrations were conducted with fine and coarse bubble aeration in a rotating hollow fibres membrane bioreactor to observe the impact of different rotation speeds on the oxygen transfer rate. The results showed that with increasing membrane rotational speed the oxygen transfer coefficient enhanced while the α-factor showed similar values at comparable sludge concentrations and solid / floc holdups. The highest improvement rates occurred during the experiments with coarse bubble aeration at 50 rpm and the lowest specific airflow rate. The solid / floc holdup appears to universally impact oxygen transfer depletion regardless what reactor type, diffuser set up and membrane rotational speed was used in the wastewater experiments.

4.1. Introduction

Gas transfer measurements are routine in the field of multi-phase flow studies. If the impact of particles is investigated the solid holdup is introduced, which describes the fraction of solids within the total volume of the suspension (Sun and Furusaki 1989). The aeration of activated sludge belongs to the biggest applications in the field of multi-phase flow studies and it consumes 50–60% of the energy consumption in a WWTP (Chen et al. 2022c). One fundamental problem for slurries that consist of activated sludge flocs or hydroxide flocs is the determination of the solid / floc holdup and its impact on oxygen transfer. Surrogate parameters used to express the impact of the sludge concentration on oxygen transfer are the mixed liquor (volatile) suspended solids concentration (MLSS, MLVSS) (Capodici et al. 2019; Germain et al. 2007; Günder 1999; Henkel 2010; Kayser 1967; Kim et al. 2020; Krause 2005; Wolfbauer et al. 1977) or the apparent viscosity of the slurry (Ali et al. 2022; Campbell et al. 2019; Durán et al. 2016; Krampe 2001; Krause 2005; Rosenberger 2003). However, these parameters lack describing that activated sludge is a three-phase mixture, consisting of flocs (gel/solid phase), free available water (liquid phase) and air bubbles (gaseous phase). The sensors that measure the oxygen concentration in activated sludge measure the oxygen concentration in the liquid phase whose viscosity does not change. It is the number of flocs that increases with increasing sludge concentration and it is the floc that interacts with the bubble and the impurities of the water.

Henkel (2010) developed a method called hydrostatic floc volume to approximate the free water content/liquid holdup respectively the floc volume/solid holdup of activated sludge and iron hydroxide slurries (see 4.2. Materials and Methods). By comparing oxygen transfer experiments using different concentrations of activated sludge flocs and iron hydroxide flocs it was shown that both slurries follow the same pattern of oxygen transfer depletion if the results were compared against the HFV, independently if coarse bubble or fine bubble aeration systems were used. These results could not be explained with the common theory that the apparent viscosity is triggering oxygen transfer depletion.

This study tested the effect of a novel rotating MBR device and the impact of different rotational speeds on oxygen transfer rate at various sludge concentrations. It is known that an additional agitation device typically improves oxygen transfer in slurry systems compared to systems without agitation (Barrera-Cortés et al. 2006; Di Palma and Verdone 2009; Kubsad et al. 2004; Mahdariza et al. 2022). Fine and coarse bubble aeration devices were tested because both are also used in practice for crossflow aeration in traditional MBR systems.

The experiment series also allowed us to recheck the impact of the solid holdup/floc volume and apparent viscosity on oxygen transfer in activated sludge. Following the current theory of the impact of apparent viscosity, it was expected that with increasing rotational speed the α-factor would improve and that coarse bubble aeration would show higher α-factors compared to fine bubble aeration.

4.2. Material and Methods

4.2.1. Oxygen mass transfer kinetics in clean water and activated sludge

If air is released to the water it will turn into air bubbles, and then all gases present in the air bubble will start moving from the gas to the liquid phase, driven by the Brownian motion, until they reach a state of equilibrium and become dissolved in water. The basic theory used for the calculation of oxygen transfer into water is the Two-film Theory from Lewis and Whitman (1924), which states that the transfer rate can be expressed in terms of an overall transfer coefficient and resistances on either side of the interface. Only if the liquid phase is completely mixed and any concentration gradient in the liquid phase is eliminated by turbulence, the basic equation is as follows:

$$\frac{dc(t)}{dt} = k_L a \cdot \left(c_L{}^* - c(t)\right) \qquad \text{(Equation 4.1)}$$

with *dc(t)/dt* representing the increase of oxygen concentration in respect of time (g/(m³·h)); k_La representing the oxygen transfer coefficient (h^{-1}); c_L* representing the oxygen saturation concentration (g/m³); *c(t)* representing the oxygen concentration at time *t* (g/m³)

However, when reactive particles/flocs are present, the standard diffusion equation needs to be adapted. The overall volume of the solids reduces the volume of the liquid phase and thus cannot be neglected. According to Sun and Furusaki (1989), the standard equation changes into:

$$\frac{dc(t)}{dt} = \frac{k_L a}{\varepsilon_L} \cdot \left(c_L{}^* - c(t)\right) - r \qquad \text{(Equation 4.2)}$$

and

$$\varepsilon_L + \varepsilon_P = 1 \qquad \text{(Equation 4.3)}$$

With ε_L representing the liquid holdup (-); ε_p representing the solid/floc holdup (-); *r* representing the reaction term (mol/(L·s))

A consequence of this equation is that at a theoretical solid/floc holdup of 1, the k_La is 0.

To describe the practically observed differences in clean water and wastewater oxygen transfer studies the α-factor was introduced (Kayser 1967):

$$\alpha - factor = \frac{k_L a_{wastewater}}{k_L a_{clean\ water}} \qquad \text{(Equation 4.4)}$$

The α-factor is used to estimate the required standard oxygen transfer rate (SOTR), which is one of key parameters in wastewater engineering.

4.2.2. Experimental setup

The experiments used the same membrane module from previous experiments (Mahdariza et al. 2022), which consists of 1,950 HF membranes with one side potted horizontally into the permeate tube and sealed at the other side (Figure 4.1). The material of the HF fibre is PVDF with a pore size of 0.3 µm. The total surface area of the membrane module is 2.25 m², which leads to a packing density of 81 m²/m³. Due to the horizontal membrane alignment and the applied rotation, all membranes are in contact with the air bubbles even though only one segment of the membrane prototype is aerated.

Figure 4.1. Rotatable hollow fibre membrane module

The measurements were carried out in a transparent tank filled with clean water (softened tap water) and activated sludge with a volume of 1 m^3. The softened drinking water was chosen to prevent any hardness precipitation during the experiments and water storage. The 3-5 g/L MLSS concentration activated sludge (AS I) was obtained from the WWTP in Königsbach, Germany (55,000 PE). Meanwhile, the activated sludge with 7-9 g/L MLSS (AS II) and 11-13 g/L MLSS (AS III) were obtained from the WWTP in Kassel, Germany (340,000 PE). The sludge thickening process was completed on the treatment facility before being transported to the pilot plant for the oxygen transfer experiments. The experiments were performed in batch mode, which means no feed or filtrate were added nor taken during oxygen transfer measurements. A Flexnorm 500 diffuser (OTT System GmbH & Co. KG, Langenhagen, Germany) was used as fine bubble air diffuser, while a self-made coarse bubble air diffuser using a PVC pipe was used as coarse bubble aerator with two 3-mm-holes. The airflow rates and dissolved oxygen (DO) concentrations were measured with a thermal flow sensor TA Di 21.6 GE (Hoentzsch GmbH, Waiblingen-Hegnach, Germany) and three oxygen electrodes with 10-15 s response time (PRONOVA Analysentechnik GmbH & Co., Bad Klosterlausnitz Germany), respectively. In addition, the ambient parameters, i.e., air temperature, pressure and partial humidity were documented through a weather station. Figure 4.2 shows the setup of the pilot plant.

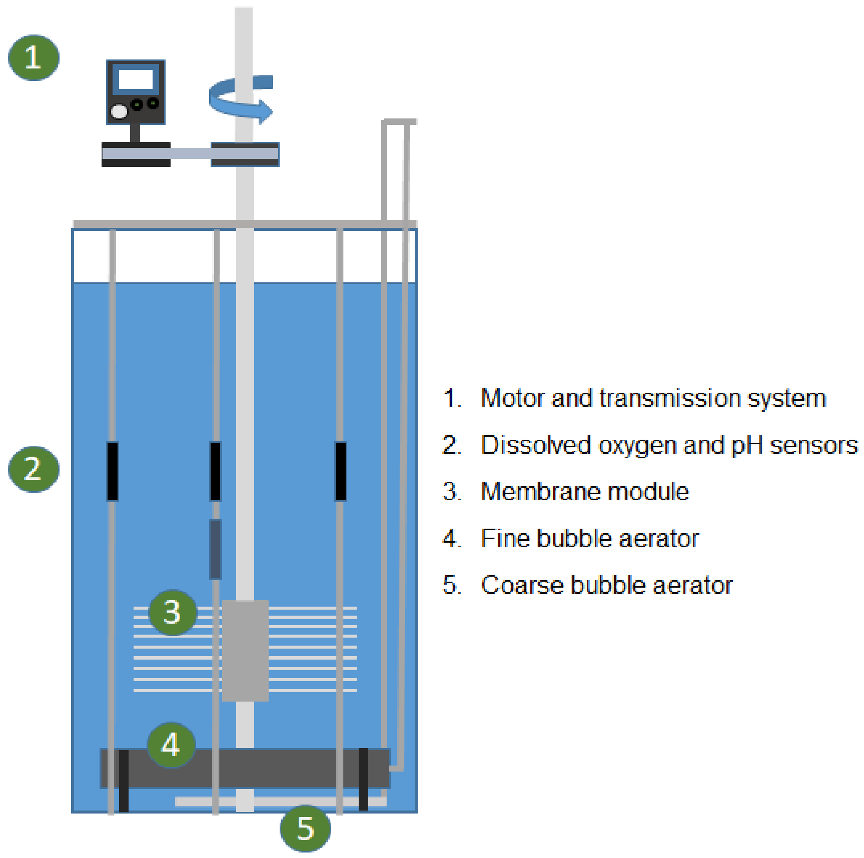


Figure 4.2. Setup of the pilot plant of a rotatable membrane prototype system (Mahdariza et al. 2022)

4.2.3. Sludge thickening

Since the MLSS concentration of the raw activated sludge from WWTP Kassel varied from 2 to 4 g/L, sludge thickening processes were conducted in advance for AS II and AS III experiments. The raw activated sludge was filled into two containers and sedimented for one hour. Afterwards, the 60 - 80% supernatant was taken out and the sedimented sludge from both tanks were collected into one tank. The process was repeated until the targeted amount of 1 m^3 thickened sludge was obtained. However, due to the variation of raw sludge for different collecting time in addition to the presence of some floating sludge as a result of the respiration process during sedimentation, the obtained thickened sludge MLSS concentration had some small variation over all experimental weeks.

4.2.4. Sludge characteristic

HFV measurements were conducted before and after the experiment. A 1 L sample of the activated sludge was transferred into a 1 L measuring cylinder. Different to the sludge volume index (SVI) measurement developed by Dick and Vesilind (1969) which measures the settled flocs after 30 minutes, the sample was left until the settled floc volume remained constant.

In addition to HFV, the MLSS, MLVSS, soluble chemical oxygen demand (sCOD), temperature and conductivity were also measured. Table 4.1 shows an example of measured sludge characteristic and HFV values during one week of coarse bubble experiments on 11-13 g/L MLSS sludge for 30 and 50 rpm. As shown, the values were relatively stable during the week of experiments. Still, a trend in decreasing MLSS, MLVSS and HFV is recognizable with increasing days.

Table 4.1. Sludge characteristics during one experimental week with 11-13 g/L MLSS concentration

Experiment day	**Sampling time**	**MLSS (g/L)**	**MLVSS (g/L)**	**HFV (mL/L)**	**Cond. (µS/cm)**	**sCOD (mg/L)**	**Temp. (˚C)**
1	Morning	12.00	7.98	380	1,040	77.0	9.0
	Afternoon	12.10	8.01	360	1,068		11.3
2	Morning	11.80	7.73	370	1,112	86.8	9.3
	Afternoon	11.90	7.88	370	1,135		10.2
3	Morning	11.40	7.39	340	1,204	88.5	9.4
	Afternoon	11.40	7.49	340	1,263		10.9
4	Morning	11.20	7.27	350	1,340	90.7	9.6
	Afternoon	11.30	7.37	340	1,383		10.5

It is worth noting that each experimental week did not have identical initial MLSS value, since the MLSS concentration of raw wastewater which was thickened did not have the same value as well. During all activated sludge experiments, the prepared activated sludge was aerated for at least 12 hours in order to maintain constant respiration rates and to nullify the potential impact of impurities on oxygen transfer. Therefore, the results mainly reflect the impact of activated sludge on oxygen transfer depletion. Furthermore, Campbell et al. (2020) highlighted the effect of filamentous organisms on oxygen transfer efficiency. However, the SVI values for sludge experiments in this study were below 150 mL/g, which means that the presence of filamentous organisms is limited, hence, it can be assumed that the distortion in oxygen transfer due to filamentous organisms is neglectable. Some key properties of AS I, AS II and AS III are listed in Table 4.2**.**

Table 4.2. Sludge characteristics during all experimental weeks

		Clean water	AS I	AS II	AS III
Fine bubble					
MLSS	g/L	---	3.9(±0.3)	8.9(±0.9)	12.4(±0.8)
MLVSS	g/L	---	2.5(±0.1)	5.7(±0.7)	8.2(±0.5)
sCOD	mg/L	---	---	44(±2)	44(±5)
Conductivity	µS/cm	811(±0)	1,481(±126)	1,170(±80)	852(±213)
pH	-	6.9(±0.0)	6.9(±0.0)	7.4(±0.2)	7.3(±0.2)
Temperature	°C	17.7(±0.4)	15.6(±0.9)	14.3(±1.6)	10.2(±0.6)
Loss on ignition	%	---	35(±2)	36(±1)	34(±0)
SVI	mL/g	---	---	78(±3)	92(±4)
HFV	mL/L	---	167(±13)	268(±18)	354(±13)
Endogenous respiration	mgO_2/ (gMLVSS·h)	---	---	1.6(±0.1)	2.6(±0.4)
Coarse bubble					
MLSS	g/L	---	3.7(±0.1)	8.0(±0.3)	11.5(±0.4)
MLVSS	g/L	---	2.6(±0.1)	5.0(±0.2)	7.6(±0.3)
sCOD	mg/L	---	----	---	74(±19)
Conductivity	µS/cm	811(±0)	1,392(±293)	1,003(±142)	1,277(±79)
pH	-	6.9(±0.0)	6.9(±0.0)	6.3(±0.5)	6.6(±0.3)
Temperature	°C	18.0(±0.5)	14.4(±2.5)	19.2(±1.2)	9.6(±0.9)
Loss on ignition	%	---	32(±1)	37(±1)	34(±1)
SVI	mL/g	---	---	82(±2)	83(±7)
HFV	mL/L	---	179(±15)	265(±10)	347(±10)
Endogenous respiration	mgO_2/ (gMLVSS·h)	---	---	2.2(±0.5)	2.0(±0.2)

4.2.5. Oxygen transfer measurement and calculation

In this study, the desorption method (DWA-M 209 2007; Wagner et al. 1998) using pure oxygen was selected for calculating the k_La value to guarantee comparable results to previously conducted fine bubble experiments on clean water and 3-5 g/L MLSS concentration of activated sludge (Mahdariza et al. 2022). Three different rotational speeds (0, 30, 50 rpm) were tested at three different airflow rates (1, 2, 4 m^3/h). However, for coarse bubble experiments at 7-9 g/L (AS II) and 11-13 g/L (AS III) MLSS concentration, additional experiments with airflow rate of 5 m^3/h were conducted, because at an airflow rate of 1 m^3/h, not sufficient oxygen could be transferred to satisfy oxygen consumption caused by endogenous respiration.

The oxygen concentration was recorded by three oxygen sensors during experiment at a constant airflow rate and constant membrane module rotational speed. In order to fulfil the requirement according to guideline from DWA-M 209 (2007), during all experiments, the airflow rate was maintained to have fluctuation less than ± 10% and the temperature difference between the beginning and the end of experiment did not exceed 2 °C. The decrease of recorded oxygen concentration was then determined by non-linear regression to produce average k_La value from all three sensors. The obtained k_La value was normalized to the standard conditions (k_La_{20}) at a water temperature of 20 ˚C and an atmospheric pressure of 1,013 hPa, and with the correction factor for a salt concentration of 1 g/L, due to significant differences between the salt content of clean water and wastewater (DWA-M 209 2007).

This procedure was performed in clean water and activated sludge with three different MLSS concentrations. Afterwards, the polynomial trend lines of calculated k_La_{20} values from different specific airflow rates for each membrane module rotational speed were generated. Finally, the comparison between the k_La_{20} value of activated sludge and clean water at specific airflow rate defined as the α-factor was calculated by dividing the trendline equation at a certain sludge concentration by the equation obtained during the clean water experiment.

4.3. Results and Discussions

Four experiment series (clean water, AS I, AS II and AS III) were performed with coarse and fine bubble diffusers at different sludge concentrations to investigate the impact of rotation and solid/floc holdup on oxygen transfer. Sampling was executed after the sludge was aerated overnight, before and after the experiments, which also ensured that all sludges had the same conditions of endogenous respiration. In Table 4.3 and Table 4.4, the results of oxygen transfer experiments are summarized. The table also incorporates the results of Mahdariza et al. (2022).

4.3.1. Impact of rotation and airflow rate on oxygen transfer

Generally, coarse bubble experiments show lower oxygen transfer rates at the same specific airflow rate and sludge concentration compared to fine bubble experiments. The bigger bubble size of coarse bubble aeration (16–20 mm) compared to fine bubble aeration (2–3 mm) leads to a lower interfacial area and therefore lower oxygen transfer rates at the same specific airflow rate.

All results have in common that with increasing airflow rate and increasing rotational speed the k_La value increases, except for the experiment with coarse bubble aeration at 30 rpm and 12 g/L MLSS concentration at a specific airflow rate of 5 $m^3_N/(m^3 \cdot h)$. The exception for coarse bubble aeration can be explained by the higher solid holdup/floc volume during the transfer experiments with rotation compared to no rotation, which has an additional negative effect on oxygen transfer.

Table 4.3. Coarse bubble experiment results

Experiments	**MLSS**	**MLVSS**	**HFV**	**Specific airflow 2 $m^3_N/(m^3\cdot h)$**			**Specific airflow 4 $m^3_N/(m^3\cdot h)$**			**Specific airflow 5 $m^3_N/(m^3\cdot h)$**		
	(g/L)	**(g/L)**	**(mL/L)**	**k_La_{20} (1/h)**	**Imp.* (%)**	**α (-)**	**k_La_{20} (1/h)**	**Imp. (%)**	**α (-)**	**k_La_{20} (1/h)**	**Imp. (%)**	**α (-)**
0 rpm clean water				2.64			5.23			6.52		
30 rpm clean water				3.39	+28		5.96	+14		6.94	+06	
50 rpm clean water				6.10	+131		9.16	+75		9.56	+47	
0 rpm AS I	3.70	2.50	155	2.28		0.87	4.31		0.82	5.23		0.80
30 rpm AS I	3.87	2.60	193	2.71	+19	0.80	4.88	+13	0.82	5.78	+10	0.83
50 rpm AS I	3.63	2.53	178	4.44	+94	0.73	6.83	+58	0.75	7.27	+39	0.76
0 rpm AS II	8.14	5.19	275	1.87		0.71	3.97		0.76	5.10		0.78
30 rpm AS II	8.07	5.07	265	2.33	+24	0.69	4.38	+10	0.74	5.30	+04	0.76
50 rpm AS II	7.71	4.85	254	3.95	+111	0.65	6.14	+55	0.67	6.57	+29	0.69
0 rpm AS III	11.30	7.36	340	1.41		0.54	3.48		0.67	4.76		0.73
30 rpm AS III	12.10	8.01	360	1.88	+33	0.55	3.76	+08	0.63	4.70	-1	0.68
50 rpm AS III	11.33	7.42	343	3.11	+120	0.51	5.32	+53	0.58	6.10	+28	0.64

(*) Improvement compared to 0 rpm.

Table 4.4. Fine bubble experiment results

Experiments	**MLSS (g/L)**	**MLVSS (g/L)**	**HFV (mL/L)**	**Specific airflow 2 $m^3_N/(m^3 \cdot h)$**			**Specific airflow 4 $m^3_N/(m^3 \cdot h)$**			**Specific airflow 5 $m^3_N/(m^3 \cdot h)$**		
				k_La_{20} (1/h)	**Imp.* (%)**	**α (-)**	**k_La_{20} (1/h)**	**Imp. (%)**	**α (-)**	**k_La_{20} (1/h)**	**Imp. (%)**	**α (-)**
0 rpm clean water				4.12			7.36			11.17		
30 rpm clean water				5.25	+27		9.40	+28		14.39	+29	
50 rpm clean water				6.52	+58		11.61	+58		17.50	+57	
0 rpm AS I	3.60	2.40	175	3.20		0.78	5.92		0.80	9.92		0.89
30 rpm AS I	4.30	2.70	175	4.23	+32	0.80	7.53	+27	0.80	11.38	+15	0.79
50 rpm AS I	3.90	2.50	150	4.92	+54	0.76	9.05	+53	0.78	14.93	+50	0.85
0 rpm AS II	9.20	6.23	274	2.71		0.66	4.85		0.66	7.48		0.67
30 rpm AS II	8.47	5.38	275	3.43	+27	0.65	6.07	+25	0.65	9.00	+20	0.63
50 rpm AS II	8.61	5.43	255	4.14	+53	0.64	7.37	+52	0.63	11.06	+48	0.63
0 rpm AS III	13.27	8.82	367	2.35		0.57	4.12		0.56	5.95		0.53
30 rpm AS III	11.87	7.84	350	3.40	+45	0.65	5.99	+45	0.64	8.73	+47	0.61
50 rpm AS III	12.10	8.05	345	3.88	+66	0.60	6.97	+69	0.60	10.76	+81	0.61

(*) Improvement compared to 0 rpm.

The biggest improvement was observed for coarse bubble aeration at a rotational speed of 50 rpm and a specific airflow rate of 2 $m^3_N/(m^3 \cdot h)$. The achieved oxygen transfer coefficients are nearly as high as for the fine bubble experiments at the same airflow rate and rotational speed. This can be explained by the impact of the membrane module. For fine bubble aeration, the rotation at 50 rpm caused such a high circular fluid flow force to the bubbles that they were bypassing the membrane fibres (Mahdariza et al. 2022). However, coarse bubble formation and rising behaviour are governed by the liquid inertia and gas momentum forces and are only little impacted by the fluid flow forces. Consequently, the coarse bubbles were still rising straight up to the membrane module. Once the bubbles hit the rotating fibres they disintegrated, forming fine bubbles and now the fluid force evenly distributed these bubbles in the reactor. This effect decreased with increasing airflow rate for coarse bubble aeration because at the higher gas holdup and heterogeneous flow regime, a portion of the fine bubbles again coalesced and formed larger bubbles.

The even increase in oxygen transfer rates for fine bubble experiments with increased rotational speed at a specific airflow rate can mainly be explained by smaller bubble formation at the orifice due to increased liquid flow forces and the change in flow pattern from straight upwards to more circular caused by the rotation of the module. Both effects increased the gas holdup and consequently lead to a steady increase in oxygen transfer at the chosen airflow rates.

4.3.2. Impact of rotation and solid / floc hold up on oxygen transfer

The effect of increased airflow rate and rotational speed had only little effect on the α-factor at comparable floc volumes and sludge concentrations (see Table 4.3 and Table 4.4).

The results indicate that the α-factor from all applied airflow rates for each MLSS concentration and membrane rotational speed presented in Figure 4.3 and Figure 4.4 follows the same pattern no matter which rotational speed or which aeration system is used. Neither for fine bubble aeration nor for coarse bubble aeration, it could be observed that an increase in rotational speed improved the α-factor. All measurements show quite similar values. Only for coarse bubble aeration at a rotational speed of 50 rpm a clear decrease in α-factor is measured.

This result is contrary to most of the current literature where oxygen transfer depression with increasing sludge concentration is mainly explained by the effect of apparent viscosity on activated sludge (Campbell et al. 2019; Durán et al. 2016; Krampe and Krauth 2003). Accordingly, the non-Newtonian pseudoplastic fluid properties of activated sludge should have caused an increase in α-factor with increasing rotational speed and higher α-factors should have been observed for coarse bubble aeration. The results of Figure 3.3 and Figure 3.4 contradict this theory.

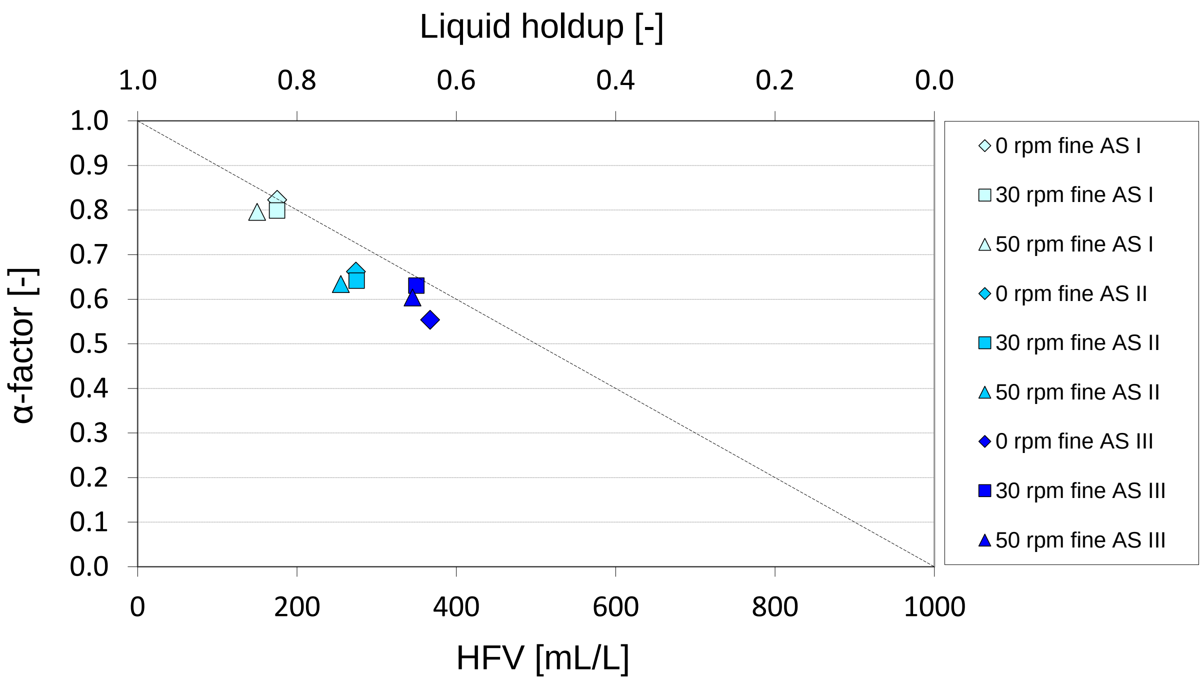


Figure 4.3. Average α-factors for fine bubble aeration at various rotational speed under endogenous conditions

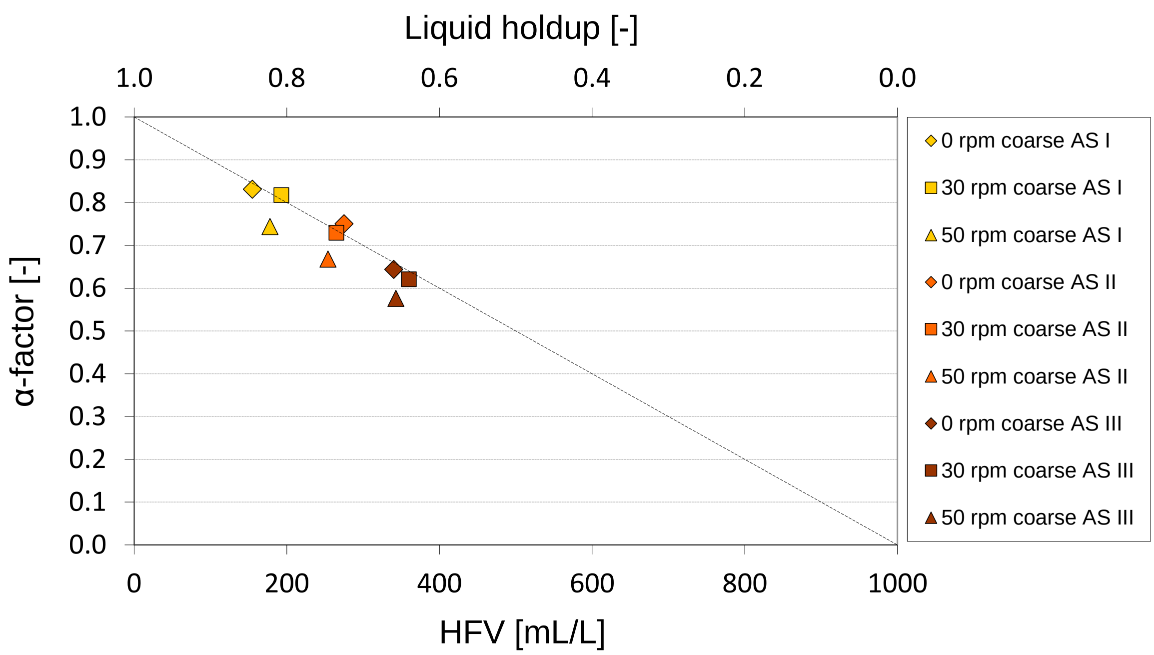


Figure 4.4. Average α-factors for coarse bubble aeration at various rotational speed under endogenous conditions.

A similar conclusion was drawn by Henkel et al. (2011) when comparing fine bubble and coarse bubble aeration systems using iron hydroxide flocs and activated sludge flocs. Again, the non-

Newtonian pseudoplastic fluid properties of the activated sludge should have theoretically caused higher α-factors for coarse bubble aeration due to the shear-thinning effect. No significant difference in the α-factor could be determined by Henkel et al. (2011) if the solid holdup/floc volume (HFV) was used to correlate the results.

Based on the results of this study and Henkel et al. (2011), activated sludge flocs are behaving similarly to solid particles on oxygen transfer. Studying activated sludge under the microscope shows that the sludge floc creates its own cluster and clearly separates from the free water content (Campbell 2020; Mesquita et al. 2013). This is corroborated by the structure of granular activated sludge flocs with spherical solids/particles.

By using Equation 4.2 and correcting the oxygen transfer results in these experiments by the reduced liquid holdup, which can be estimated by using the HFV, the corrected α-factors are in the range of 0.9–1.0. The results are compared to obtained α-factors from Equation 4.4, as shown in Table 4.5.

Table 4.5. The comparison of obtained α-factors from two different equations

	AS I			**AS II**			**AS III**		
α-factor fine bubble									
rpm	0	30	50	0	30	50	0	30	50
$(k_La_{waste}/\ k_La_{clean})$	0.82	0.80	0.80	0.66	0.64	0.63	0.55	0.63	0.60
solid / floc holdup ε_P	0.18	0.18	0.15	0.27	0.28	0.26	0.37	0.35	0.35
$(k_La_{waste}/\ ((1-\varepsilon_P)\ k_La_{clean}))$	1.00	0.97	0.94	0.91	0.88	0.85	0.88	0.97	0.92
α-factor coarse bubble									
rpm	0	30	50	0	30	50	0	30	50
$(k_La_{waste}/\ k_La_{clean})$	0.83	0.82	0.74	0.75	0.73	0.67	0.64	0.62	0.58
solid / floc holdup ε_P	0.16	0.19	0.18	0.28	0.27	0.25	0.34	0.36	0.34
$(k_La_{waste}/\ ((1-\varepsilon_P)\ k_La_{clean}))$	0.98	1.01	0.91	1.04	0.99	0.90	0.98	0.97	0.88

Consequently, oxygen transfer coefficients into the liquid phase achieved during stabilized activated sludge experiments are comparable to the clean water experiments. This is supported by experiments from Henkel (2010), Kayser (1967) and Steinmentz (1996), who measured and investigated the impact of the pure liquid phase on the α-factor of activated sludge plants without deriving a significant impact of the wastewater effluent or MBR filtrate.

In summary, despite the usage of different reactor geometries, different aeration devices (fine bubble disk aerator, fine bubble tube aerator, coarse bubble two aeration holes, coarse bubble 10 aeration holes) and different slurries (MBR sludge, thickened activated sludge from different plants, iron hydroxide slurry), the α-factor decreases in the same pattern with decreasing liquid holdup and increasing floc/solid holdup by using HFV as reference value (Figure 4.5).

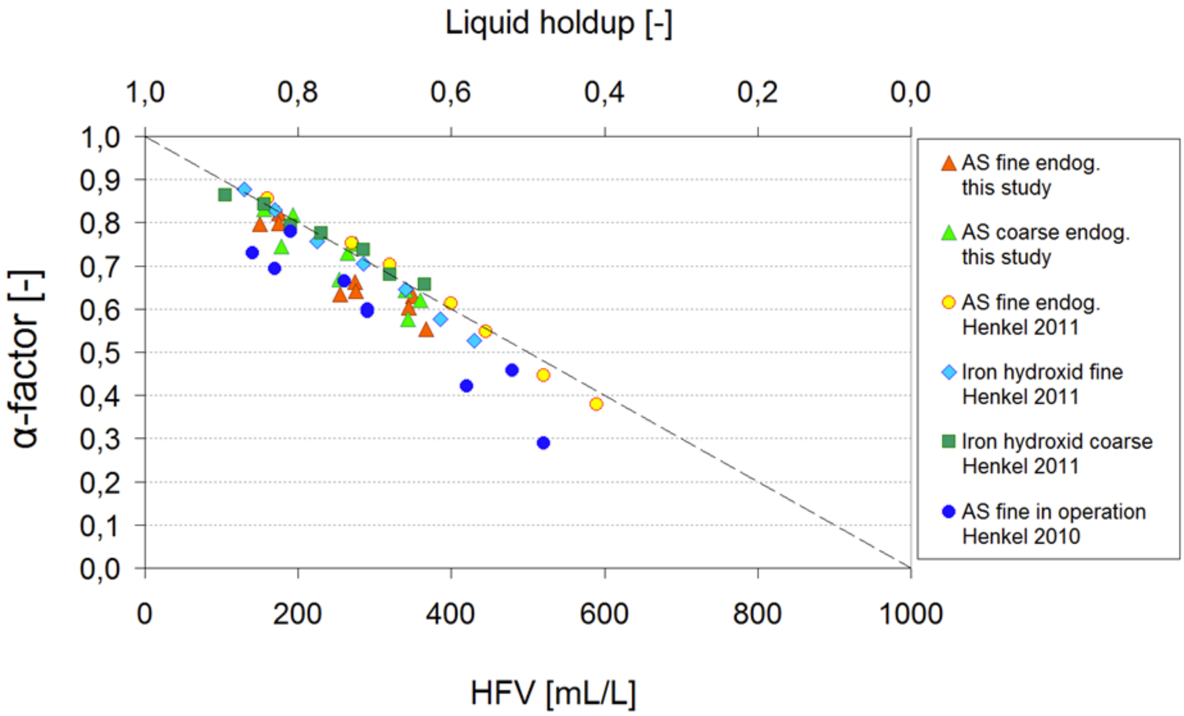


Figure 4.5. Summary of HFV experiments with fine and coarse bubble aeration under endogenous conditions and in operation

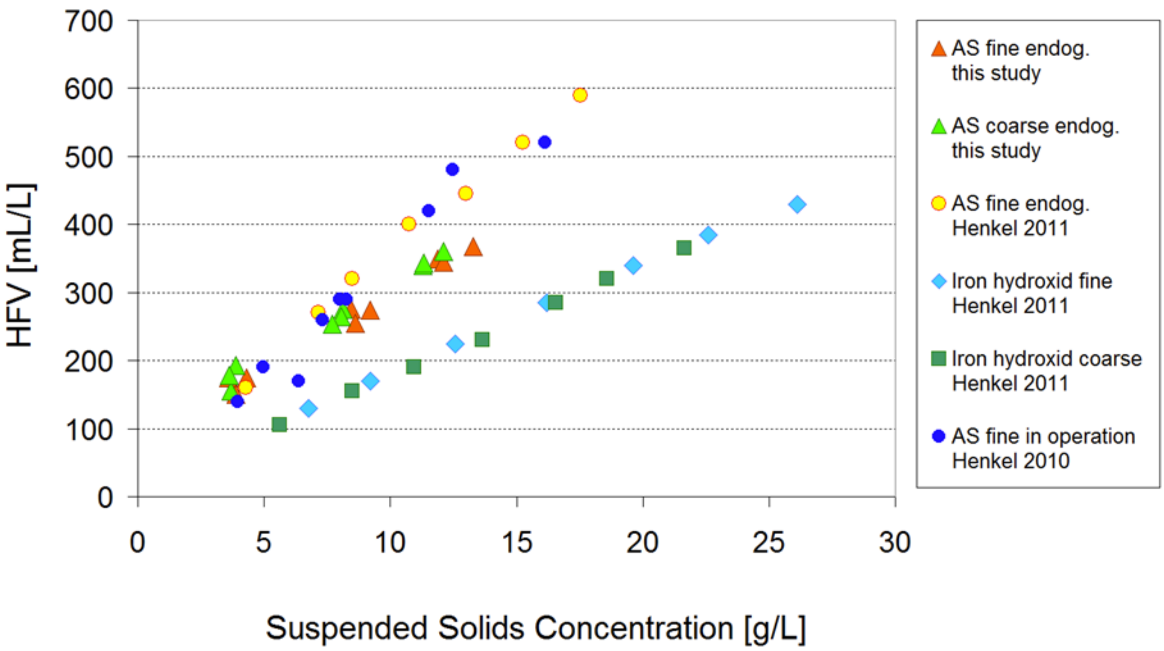


Figure 4.6. Suspended solids concentration vs. HFV under endogenous conditions and in operation.

To the best of the authors' knowledge, a consistent survey of these influencing factors on oxygen transfer has not been published yet. Due to the fact that the suspended solids concentration (MLSS) is still used as the main parameter to compare oxygen transfer results, the impact of activated sludge flocs on oxygen transfer is until now not included in the majority of the studies. Several authors aimed to disclose the impact of activated sludge flocs by running experiments with different materials, e.g., aluminum hydroxide, activated carbon, peat and bentonite clay, using similar suspended solid concentrations and comparing the oxygen transfer results (van der Kroon 1968; Steinmentz 1996; Henkel 2010; Blanco Zúñiga et al. 2021). However, the suspended solids concentration does not reflect the volumetric fraction of the different materials. Iron hydroxide flocs occupy for example a different volume at the same suspended solids concentration compared to activated sludge flocs, as shown in Figure 4.6.

But also activated sludge taken from a wastewater plant that operates without primary sedimentation shows a different floc volume to suspended solids ratio compared to activated sludge from a wastewater plant with primary sedimentation due to higher content of silt, clay, and sand (Henkel et al. 2011). In addition, the SRT impacts the floc volume as it influences the organic content of the activated sludge floc. Plants running at higher SRTs typically show a lower loss on ignition of the sludge compared to plants that operate at low SRT (Foladori et al. 2010). Finally, a finding from Wu et al. (2021) presented the influence of floc size and circularity on oxygen uptake rate, which is supported by a study by Burger et al. (2017) who showed that filamentous bacteria influence floc morphology, impacting oxygen transfer but also the free available water content and solid/floc holdup. These diverse impact factors on floc volume are not reflected by the MLSS concentration and thus apparently different α-factors were obtained in the past by using only MLSS concentration as the reference value.

4.3.3. Practical implication

The still-existing lack of a common understanding of which parameters rule oxygen transfer in activated sludge is mainly caused by the fact that important parameters like the impact of the flocs (MLSS, MLVSS, HFV) and the impact of impurities (surfactants, polymers, adsorbed organics) are discussed independently although they are interconnected, e.g., by Schwarz et al. (2021).

The floc volume (solid holdup) is directly linked to the mass of sludge in the system and because of this, it also governs parameters like the F/M ratio or the SRT and it is responsible for the amount of adsorbed organic matter to the floc and the dissolved impurities in the sludge, which can additionally impact oxygen transfer (Gillot and Héduit 2008; Rosso et al. 2008; Schwarz et al. 2021; Bencsik et al. 2022).

Acknowledging these interdependencies, the worst oxygen transfer conditions occur where a high load of impurities that influence oxygen transfer (high F/M ratio, low SRT) and high floc volumes (solid holdup, TSS concentration) jointly appear (Schwarz et al. 2023). This is for example the case for sequencing batch reactors that do not use additional agitation. Just after the sedimentation phase when aeration is used to expand the settled sludge bed (high floc concentration) and still the amount of adsorbed organic to the floc is high (high F/M ratio), the lowest α-factors are observed. This has been confirmed by Cecconi et al. (2020) and Strubbe et al. (2023), who reported α-factor values as low as 0.2 in such applications. This is even lower than the typically measured α-factor of 0.3–0.4 in the raw wastewater influent (Kayser 1967; Henkel 2010) or 0.40–0.45 in the activated sludge plants running at an SRT of 2.0 (Kroiss and Klager 2018; Schwarz et al. 2021). On the contrary, aerobic stabilization plants, which use low TSS concentrations (low floc volume) and low F/M ratio (high SRT) specifically at the end of the aeration basin, show α-factors as high as 0.85 (Gillot and Héduit 2008; Schwarz et al. 2021), as long as no filamentous bacteria occur (Campbell and Wang 2020).

4.4. Conclusions

Oxygen transfer experiments were conducted with a new rotating type of HF membrane module using fine and coarse bubble aeration with different airflow rates and membrane rotational speeds.

- For both fine and coarse bubble experiments oxygen transfer coefficients rise with increasing rotational speed of the membrane at the same solid/floc holdup and sludge concentration.

- The improvement of oxygen transfer rate at 30 rpm is on average higher for fine bubble aeration (25%) compared to coarse bubble aeration (10%). At 50 rpm, the highest improvement rate could be observed for coarse bubble aeration at the lowest airflow rate tested (100%). However, with increasing airflow rate, this improvement rate decreases again significantly for coarse bubble aeration while for fine bubble aeration, it stays nearly constant.

- Despite the very distinct impact of rotation and airflow rate on oxygen transfer in activated sludge, the α-factors showed quite similar values for both fine and coarse bubble aeration at comparable sludge concentrations and solid/floc holdups.

- The solid holdup or liquid holdup has so far not been considered in the calculations of the α-factor to describe the impact of the activated sludge floc on oxygen transfer in wastewater engineering. However, the results in this study and previous studies indicate the need to do so, as it appears to universally impact oxygen transfer no matter what reactor type (bubble column, airlift reactor), diffuser setup (disk aerator, tube aerator, fine bubble, coarse bubble) and rotational speed (30 rpm, 50 rpm) was used in the wastewater experiments. Practically, the individual solid/floc holdup can be correlated to the MLSS concentration of each WWTP.

- The study could not confirm that coarse bubble aeration compared to fine bubble aeration systems generally create higher α-factors and that the α-factor generally increases with increasing turbulence (Stenstrom and Gilbert 1981). Consequently, these statements cannot be generalized for the impact of the solid holdup and the liquid holdup in floc suspensions.

4.5. Acknowledgement

The authors want to thank the Ministry of Finance of the Republic of Indonesia for the LPDP scholarship provided to Fathul Mahdariza (Ref. Number: S-1289/LPDP.4/2018). The research did not receive any other specific grants from funding agencies in the public, commercial or not-for-profit sectors.

The authors would like to thank Dr.-Ing. Jochen Henkel for very fruitful discussion during this research. We are grateful to Dipl.-Ing. Uwe Neuschäfer, Dietmar Landgrebe and their team at the WWTP Kassel for providing technical support during pilot plant experiments. In addition, the authors want to thank deeply Dr.-Ing. Ursula Telgmann, Andrea Brandl and Monika Degenhardt (University of Kassel, Chair of Urban Water Engineering) for their excellent laboratory work in addition to Dipl.-Ing. Ralf Feldner (University of Kassel, VPUW) and Lukas Marmucki for valuable technical assistance. Finally, we would like to thank Dr.-Ing. Stephan Fuchs, Sven Urbach, Tobias Mumbauer, Patrick Dufner and Maik Erhard (Karlsruhe Institute of Technology, Department of Aquatic Environmental Engineering) for assistance and technical support during the first phase of experiments.

5. Enhancing Membrane Bioreactor Efficiency: The Impact of Rotating Membrane Modules and Aeration Strategies on Transmembrane Pressure

This chapter has been published as an article in Water Science and Technology:

Mahdariza, Fathul; Georg, Wilhelm; Pronold, Henri; Morck, Tobias (2024): Enhancing Membrane Bioreactor Efficiency: The Impact of Rotating Membrane Modules and Aeration Strategies on Transmembrane Pressure. In Water Science and Technology: A Journal of the International Association on Water Pollution Research 90 (3), pp. 985–994. DOI: 10.2166/wst.2024.242.

Abstract: The study analyses the performance of a pilot plant using a rotating hollow fibre (HF) membrane bioreactor (MBR) system. The experiments evaluated the effect of operational parameters such as rotational speed, aeration strategies and maintenance cleaning procedures on the efficiency of the system, in particular transmembrane pressure (TMP) and filtrate quality. The results indicate that the rotating membrane module reduces TMP increase and can operate for 48 days with satisfactory performance, even without aeration. This has the potential to significantly improve efficiency, resulting in significant energy savings. In addition, two maintenance cleaning methods, clean in air and clean in place, were tested and found to be efficient for weekly maintenance cleaning. It was observed that operating without aeration during colder seasons may not be effective. Therefore, adaptive strategies are needed to address seasonal temperature variations.

5.1. Introduction

The MBR system, integrating biological processes with membrane filtration, provides a robust approach to wastewater treatment (Judd 2016). Its compact design, achieved by eliminating the need for a secondary settling tank, confers a distinct advantage over CAS systems. The MBR process offers additional benefits, including the capability to operate at higher MLSS concentrations, an extended sludge age, and reduced sludge production compared to CAS methods (Barreto et al. 2017; Pollice et al. 2008; Visvanathan et al. 2000). These benefits have encouraged the adoption of MBR technology across 200 countries by 2016, with the global market for MBR systems experiencing an annual growth rate of 15% (Judd 2016).

Nevertheless, fouling is a significant challenge associated with MBR systems, which is characterized by the accumulation of disruptive deposits on the membrane surface or within the pores. These deposits originate from retained salts, macromolecules, colloids, and particles (Ladewig and Al-Shaeli 2017). To deal with fouling, operational strategies such as chemical and mechanical cleaning strategies to prevent the membrane from fouling. However, these countermeasures add complexity to the system, resulting in increased capital and operational costs. The main cost drivers are the periodic need to replace membrane modules and the implementation of anti-fouling strategies, which collectively raise the financial burden of utilizing MBR technology (Judd 2011; Rahman et al. 2023).

The phenomenon of fouling represents the primary challenge encountered in an MBR system. As a result, solving or reducing the problem has the greatest impact on the efficiency of an MBR plant (Al-Asheh et al. 2021). Thereby, fouling is a process by which the membrane experiences a loss of performance due to the deposition of dissolved and/or suspended matter on the membrane surface, openings or within the pores (Iorhemen et al. 2016). This leads to a decrease in flux, which is the quantity of material passing through a unit area of membrane per unit time, measured in litres per m^2 per hour (or LMH) (Judd 2011). The fouling depends on a large number of factors such as the cleaning strategy, the operating conditions, the specific properties of the wastewater, and the membrane used (Al-Asheh et al. 2021). A common strategy to reduce membrane fouling is by using coarse bubble (CB) aeration under the membrane module. Yet the increased oxygenation leads to increased foam formation and increased ongoing energy requirements, which can be almost double that of the CAS process (Al-Asheh et al. 2021; Iorhemen et al. 2016; Judd 2016). The use of CB aeration compromises oxygen transfer efficiency, which could be significantly improved by using fine bubble (FB) aeration (Henkel et al. 2009b; Mahdariza et al. 2023; Zuo et al. 2024).

Another possibility is to increase the shear force to prevent the attachment of biofilm to the membrane surface. A potential solution is to introduce a rotational membrane, which can minimize the formation of reversible fouling (Jiang et al. 2012; Rector et al. 2006; Wu et al. 2008; Zuo et al. 2010). Furthermore, a novel pilot-plant scale prototype of rotating HF MBR modules was built and studied in batch process (Mahdariza et al. 2022; Mahdariza et al. 2023). In contrast to conventional HF MBR modules, the new concept applies a continuing sheer force by rotation to the new arrangement of HF membrane modules. The results showed that the additional energy required for rotation can be overcompensated by the improved oxygen transfer efficiency driven by rotation.

In this study, a series of experiments was conducted with the rotating HF membrane module in operation. The objective was to evaluate the performance of the pilot plant under various operational parameters, with a focus on TMP and filtrate quality.

5.2. Material and Methods

The rotating HF membrane module discussed in this study is the same as the one utilized in prior research by Mahdariza et al. (2022) and Mahdariza et al. (2023). The module (Figure 5.1) comprises 1,950 HF membranes constructed from PVDF, each with a pore size of 0.3 µm. Every individual fibre is sealed at one end and attached at the other end to the permeate tube, arranged horizontally. The total membrane area of the module is 2.25 m^2, yielding a packing density of 81 m^2/m^3. The unique feature of this setup is the rotation of the HF membrane module, coupled with the horizontal positioning of the membranes, which ensure that air bubbles effectively contact all fibres during the aeration process.

Figure 5.1. The rotatable HF membrane module.

The experiments were carried out at the WWTP Kassel, Germany (340,000 PE), utilizing return sludge as the inflow source. The MLSS concentration of the return sludge ranged from 5 to 7 g/L, with an average solid retention time of 15 days. This sludge was introduced into the pilot plant without additional screening at a flowrate of 1 L/s. The 1 m^3 capacity tank typically held 0.9 m^3 of sludge until it reached two designated overflow channels. Filtration was achieved using a membrane module, which was facilitated by a positive displacement pump (Flojet, Xylem Water Solutions Deutschland GmbH, Langenhagen, Germany). An identical pump model was employed for the backwashing process. For aeration, a Flexnorm 500 diffuser (OTT System GmbH & Co. KG, Langenhagen, Germany) was employed to generate fine bubbles, while a custom-built coarse bubble diffuser, constructed from PVC piping with three 3-mm holes, served for coarse aeration. The rotation of HF membrane module was powered by a motor (MSF-Vathauer Antriebstechnik GmbH & Co KG, Detmold, Germany). The setup of the pilot plant is depicted in Figure 5.2.

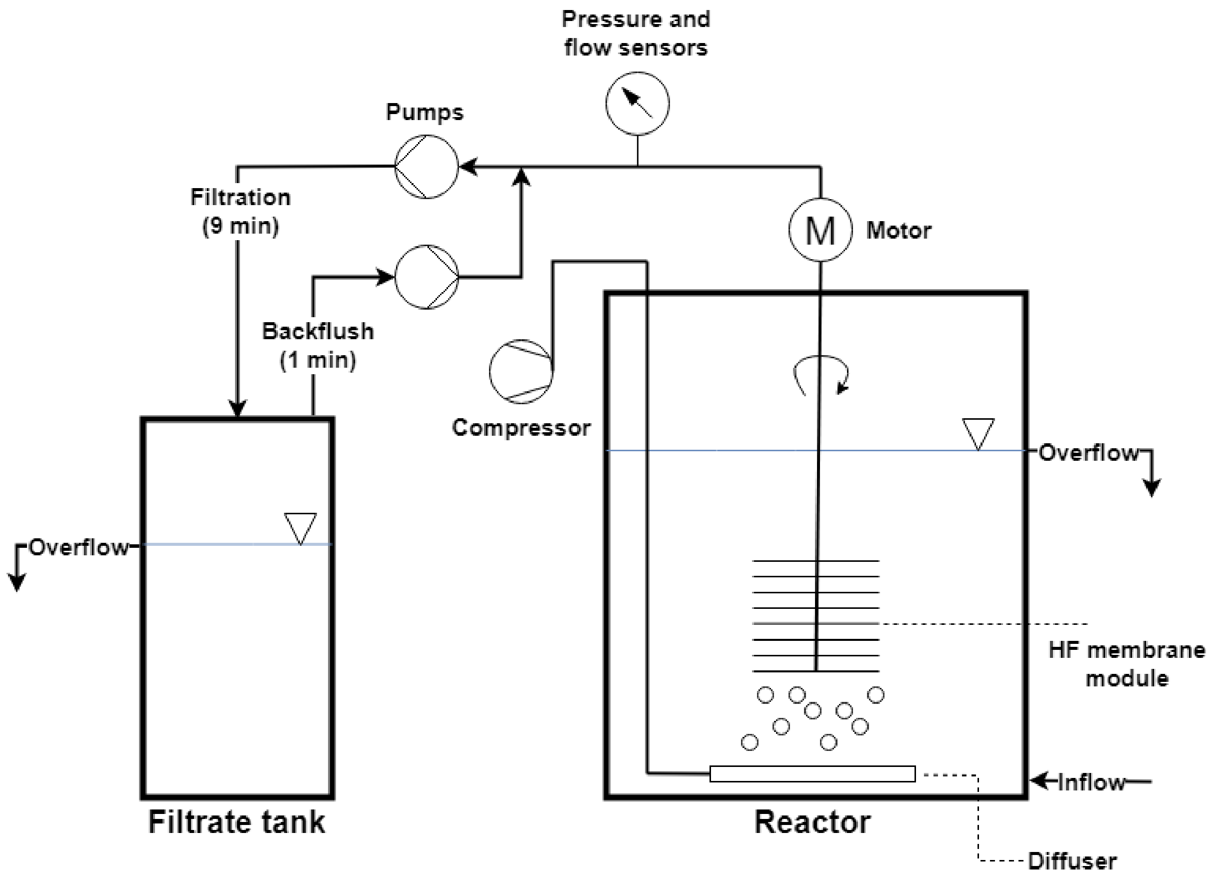


Figure 5.2. Setup of the pilot plant.

The research explored a range of operational variables, such as diverse module rotational speeds and aeration systems, to assess their impact on system performance. After each experimental session, the system underwent maintenance cleaning (MC), which involved chemically enhanced backflush with 20 L of sodium hypochlorite (NaOCl). This procedure was essential for maintaining the efficiency of the system.

In the first phase of the experiment, two different ratios of filtering to backflushing processes and fluxes were examined. The outcomes of this phase guided the subsequent experimentation during the second phase, which focused on testing different aeration strategies and membrane module rotations. Each experimental set was scheduled to run for approximately one week (6-7 days), after which MC of the membrane module was conducted. This phase also aimed to evaluate the effectiveness of MC by exploring different concentrations of sodium hypochlorite. Following the methodologies outlined by Judd (2011) and Wang et al. (2014), two distinct approaches were compared: cleaning in air (CIA), involving emptying the reactor, and cleaning in place (CIP), where the reactor remained filled with wastewater. The final phase extended over 7 weeks, during which the pilot plant operated to assess the feasibility of the MBR module in practical applications.

5.3. Results and Discussions

5.3.1. The measurement of the quality of sludge and filtrate

A series of parameters were subjected to periodic laboratory sampling during all experiments conducted within this study. In addition to monitoring the quality of the input sludge, it was necessary to observe any changes in the filtrate quality that may have occurred when different operational setups were applied during the operational period of the pilot plant.

Table 5.1. The characteristic of sludge and filtrate during experiment

Parameters	First Phase		Second Phase		Third Phase	
	Sludge	Filtrate	Sludge	Filtrate	Sludge	Filtrate
MLSS conc. (g/L)	5.9(±0.4)	-	6.2(±0.4)	-	7.0(±2.2)	-
Turbidity (NTU)	-	0.6(±0.1)	-	0.8(±0.2)	-	0.4(±0.1)
pH	6.9(±0.1)	7.0(±0.0)	6.9(±0.1)	7.0(±0.1)	6.8(±0.1)	6.8(±0.1)
Conductivity (µS/cm)	1,203(±54)	1,230(±58)	1,165(±153)	1,206(±146)	734(±211)	849(±215)
sCOD (mg/L)	50.0(±3.5)	21.6(±2.6)	48.8(±2.4)	21.0(±1.8)	24.8(±8.1)	14.5(±3.9)

As illustrated in Table 5.1, the pilot plant demonstrated effective performance in terms of sCOD removal and filtrate turbidity even in the absence of pre-screening. Despite the MLSS concentration exceeding 8 g/L on several occasions during the 48-day operation period (third phase), the turbidity of the filtrate remained consistently below 1 NTU. Furthermore, the pilot plant demonstrated the ability to maintain an average sCOD removal of 54%, which aligns with reported values for microfiltration MBR systems from other studies, which range between 25% and 98% (Ahn and Song 1999; Baek and Pagilla 2006; Deowan et al. 2019; Kabuba et al. 2023; Lin et al. 2012; You et al. 2007).

5.3.2. The impact of filtration-to-backflush time ratio and flux on TMP

In the initial phase of the experiment, the investigation focused on the impact of the filtration-to-backflush ratio and the flux on the increase of TMP. The objective was to achieve a TMP increase that would allow stable operation with one MC per week.

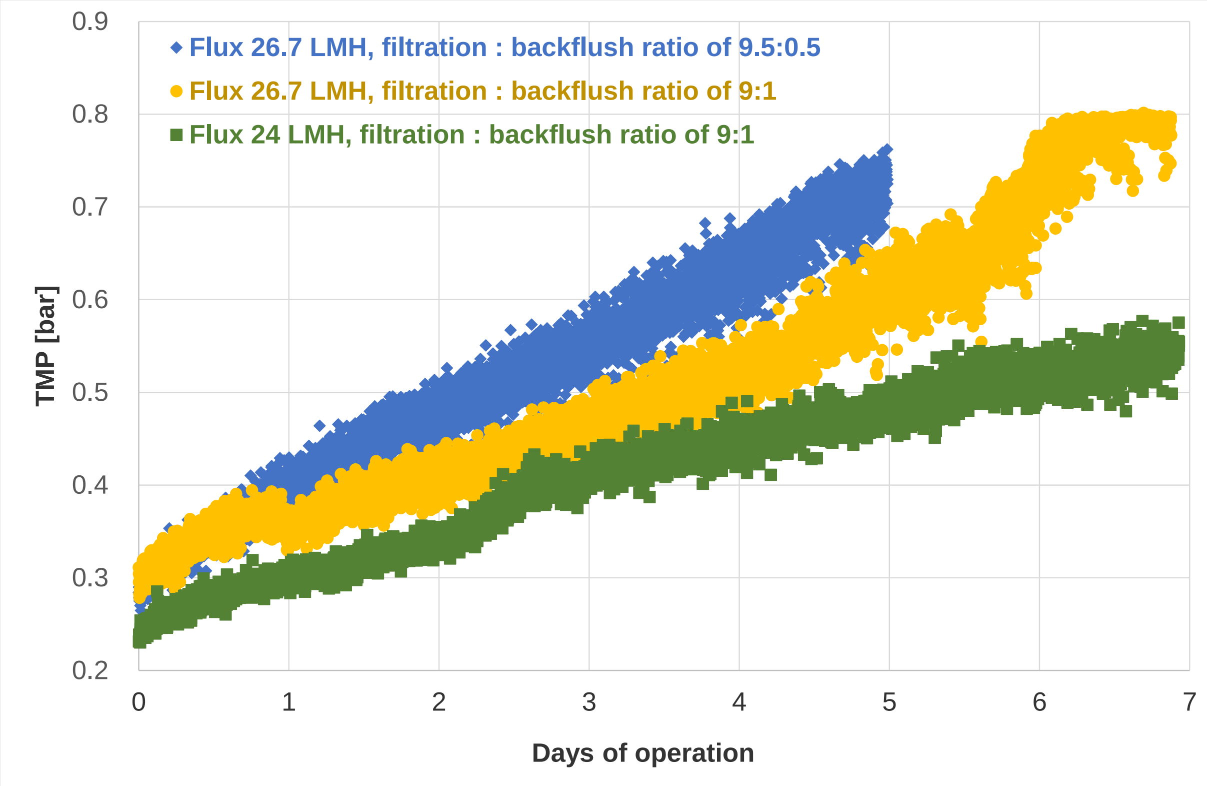


Figure 5.3. TMP increase across varied filtration to backflush ratios and filtration fluxes.

Setting the filtration flux to 26.7 LMH (filtration flowrate 1 L/min) with CB aeration and 20 rpm module rotation, the results indicated that with a filtration backwash ratio of 9:1 minute the TMP increase was too steep to achieve the desired outcome, as shown in Figure 5.3. A reduction in filtration time and an increase in backwash time improved the performance. Finally, the desired outcome was achieved by employing a filtration backwash ratio of 9:1 minute at a flux of 24 LMH (filtration flowrate 0.9 L/min). In support of this, Jiang et al. (2005) also demonstrated that operating the MBR system at a flux of less than 25 LMH resulted in more stable long-term performance. This flux value is commonly observed in MBR systems (Judd 2016).

5.3.3. The impact of aeration strategies and module rotation on TMP

In the subsequent stage of the investigation, the focus shifted to assessing the impact of three different aeration strategies: without aeration, FB aeration, and CB aeration. This assessment was further enhanced by adjusting the rotational speeds of the membrane module to 0, 20, and 30 rpm. The limitation to 30 rpm was based on insights from a previous study on the same membrane system conducted by Mahdariza et al. (2022). The study identified that SAE peaked at this speed, with efficiency declining at higher rotation speeds.

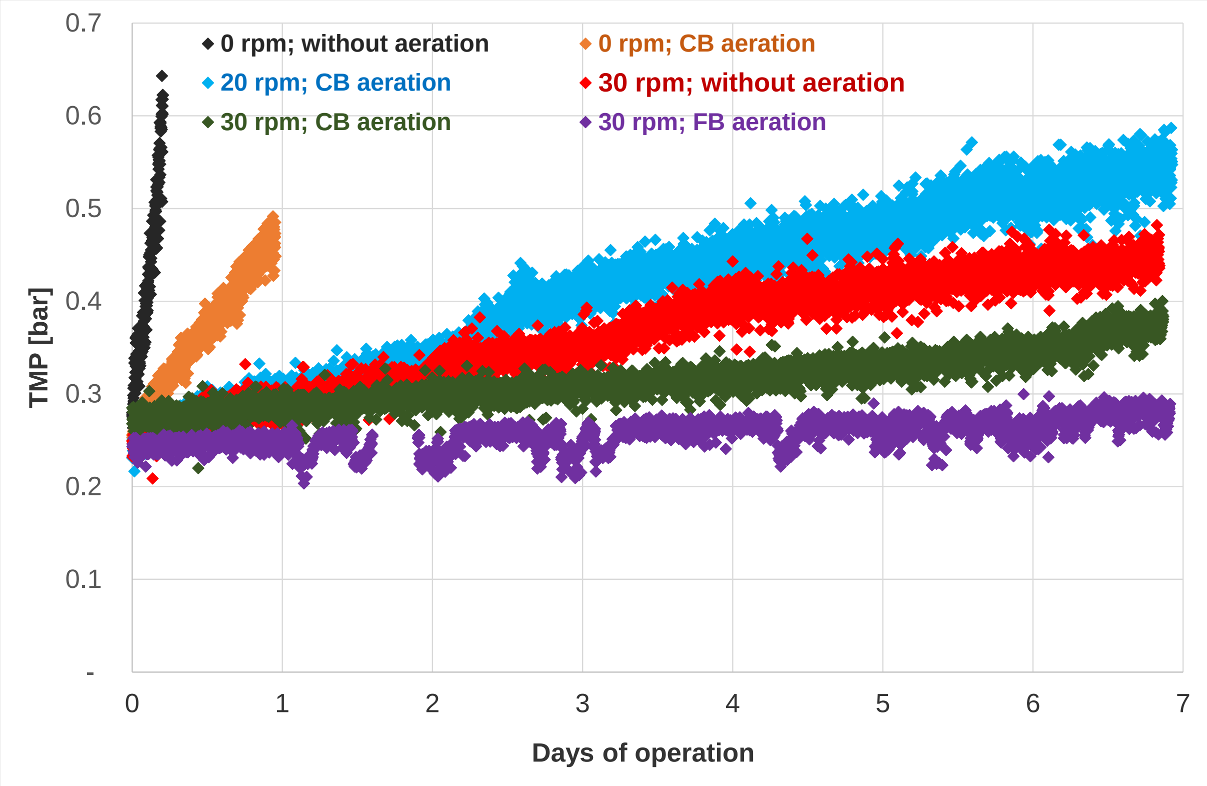


Figure 5.4. TMP increase across varied aeration strategies and membrane rotational speeds (Mahdariza et al. 2024).

To mitigate the influence of sludge temperature on the outcomes, the experiments for this second phase were scheduled within the same season during spring. This approach ensured comparable results. Figure 5.4 illustrates that two out of six experiments experienced early termination when membrane rotation was not employed, due to a rapid increase in TMP exceeding anticipated levels. In the experiment without aeration and without module rotation, this occurrence was somewhat expected, given that the interaction between foulant and membrane surface was not disrupted. In another experiment, in which only aeration was applied without module rotation, only a portion of the membrane module was scoured, as the design of the membrane module did not allow for all membrane fibres to interact with air bubbles in the absence of rotation. However, after the implementation of membrane rotation, the rise of TMP was effectively controlled even in scenarios without any aeration.

A number of studies have highlighted the enhanced physical cleaning benefits of CB aeration. These include studies by Braak et al. (2017), Judd (2005), Phattaranawik et al. (2007) and Zhao et al. (2021). Furthermore, a study conducted by Jones (2017) demonstrated that the rotational mechanisms in a rotating MBR system contributed to a mere 12% of fouling prevention by removing the cake, with the majority of the removal achieved through air scouring. However, the findings from this study revealed no substantial differences in the increase of TMP among the various aeration strategies when membrane rotation was implemented. This indicates that the efficacy of membrane cleaning and fouling prevention may not be as significantly influenced by the type of aeration employed as assumed. This thereby emphasizes the pivotal role of membrane rotation in maintaining optimal membrane performance. In addition to facilitating oxygen transfer, membrane rotation has been demonstrated to increase shear force, thereby limiting the build-up of a cake layer on the surface of the membrane. This finding is consistent with previous research on this specific rotating HF membrane module (Mahdariza et al. 2022; Mahdariza et al. 2023).

5.3.4. Maintenance cleaning strategy

Furthermore, the influence of MC on TMP reduction was inspected through experiments employing three different concentrations of NaOCl solution, while maintaining the solution temperature between 30 to 38 °C. In each cleaning cycle, a 5 L volume of NaOCl solution was introduced to the membrane module four times, each followed by a 5-min soaking period.

Table 5.2. TMP before and after maintenance cleaning (Mahdariza et al. 2024).

NaOCl solution concentration (ppm)	cleaning in air (CIA)			cleaning in place (CIP)		
	TMP before cleaning (bar)	TMP after cleaning (bar)	TMP reduction (bar)	TMP before cleaning (bar)	TMP after cleaning (bar)	TMP reduction (bar)
250	0.58	0.32	0.26	0.55	0.36	0.19
500	0.56	0.30	0.26	0.54	0.41	0.13
1,000	0.69	0.33	0.36	0.37	0.30	0.07

Table 5.2 demonstrates that increasing the concentration of NaOCl did not result in a proportional decrease in TMP within the CIP approach. In contrast, the CIA approach demonstrated a direct correlation between increased NaOCl concentration and TMP reduction, highlighting its effectiveness. It is important to note the variation in initial TMP values prior to MC in different experimental setups. Despite the observed variations, the data obtained suggest that the CIA method is more effective in reducing TMP when compared to the CIP method.

This finding is consistent with the results from Brepols et al. (2008), which demonstrated that CIA achieved twice the permeability recovery compared to CIP with the same NaOCl dosage. However, it is noteworthy that the overall CIA process in this pilot plant required an additional hour compared to CIP, which was attributed to the time necessary for emptying and refilling the tank. Therefore, the CIP method is considered to be sufficiently effective for routine weekly maintenance cleaning.

5.3.5. Further experiment without aeration

Based on these results, a longer period of experiment was conducted without aeration and relying solely on membrane rotation to increase the shear force. The pilot plant was successfully operated for 48 days as shown in Figure 5.5. In addition, the pilot plant was operated without any chemical cleaning for 2 weeks during the second experiment of this phase. This finding demonstrates the potential for energy savings by avoiding the need to increase aeration and potentially eliminating it entirely without compromising filtrate quality.

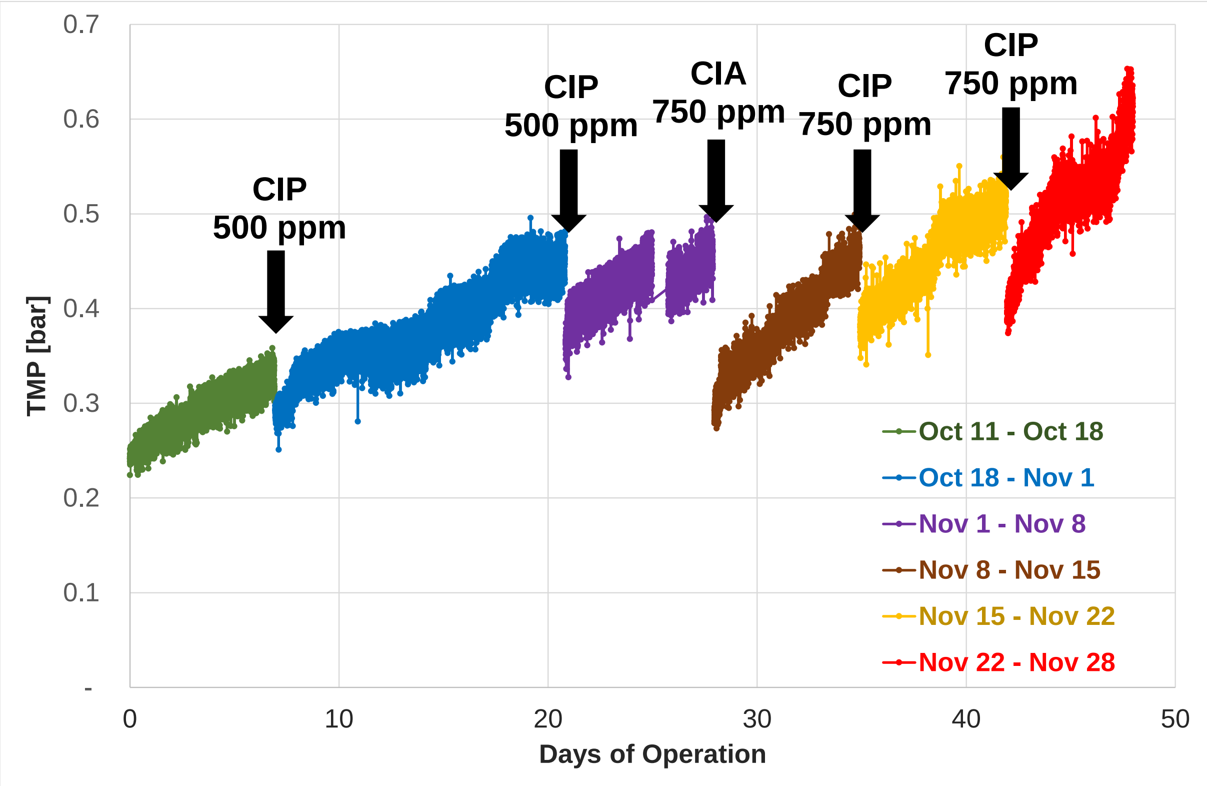


Figure 5.5. TMP increase during 48 days operation with membrane rotation and without aeration.

In terms of MC, the pilot plant was effectively operated with four times CIP and one-time CIA over the course of this period. An operational pause on the 26th day, necessitated by pump repairs at the WWTP, briefly halted inflow to the MBR pilot plant. As seasonal temperatures began to fall, leading to cooler wastewater, there was a noticeable acceleration in the increase of TMP after the 28th day of operation. The experiment showed that a higher concentration of NaOCl for CIP was necessary due to this condition. By the end of the experiment, TMP levels had risen to over 0.6 bar. This result emphasizes the need for increased NaOCl concentrations and more frequent MCs during colder months, which is particularly important for operations without aeration.

5.3.6. The impact of sludge temperature on membrane fouling

Several studies have found a correlation between decreasing temperatures and increased membrane fouling. This is attributed to the enhanced release of soluble microbial products and extracellular polymeric substances by filamentous bacteria (Iorhemen et al. 2016; Ma et al. 2013; van den Brink et al. 2011). Therefore, temperature differentials are considered a significant factor that could either exacerbate or alleviate fouling during filtration and backwash operations. To investigate this hypothesis, the study repeated two experiments under varying sludge temperatures caused by seasonal changes. Figure 5.6 illustrates the comparison of these experiments, demonstrating the effect of temperature on TMP increase.

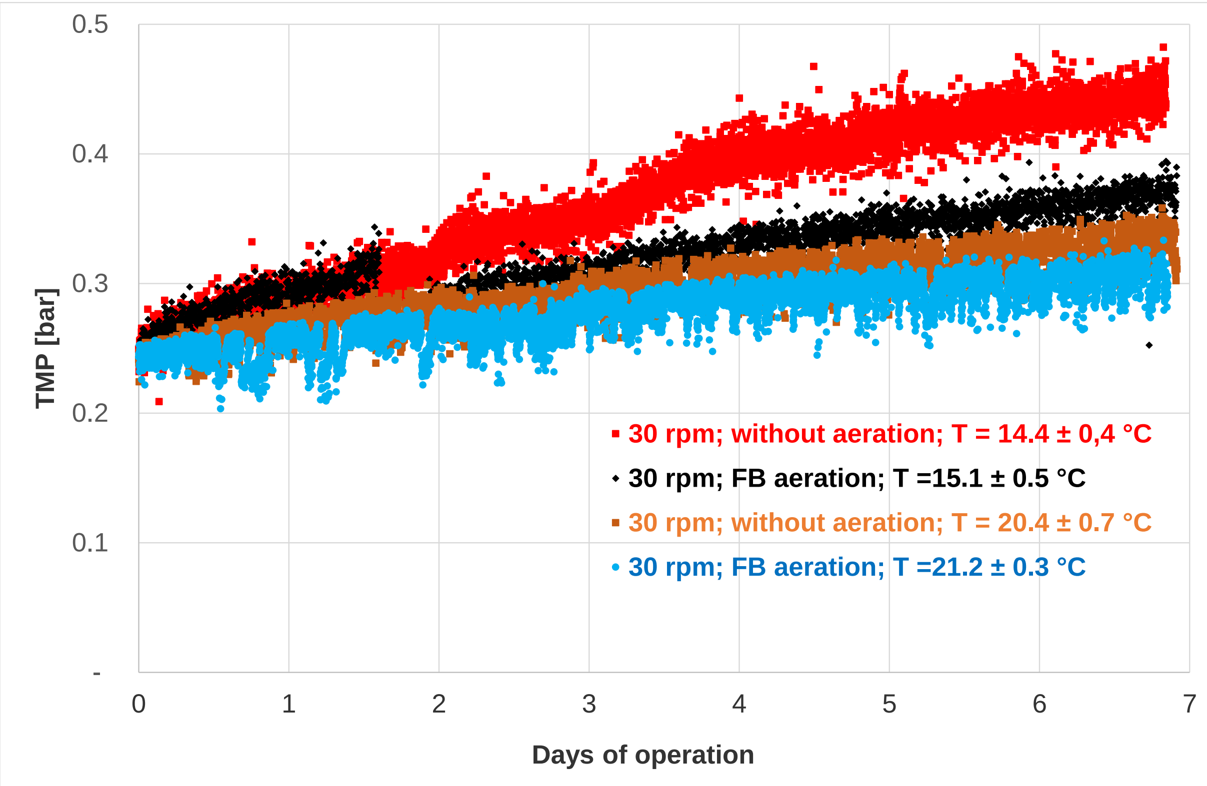


Figure 5.6. TMP increase for different sludge temperature.

This outcome demonstrates that experiments conducted with sludge at a temperature 6 °C lower resulted in a faster increase in TMP. Hence, for this specific rotating membrane module configuration, it is not recommended to use a configuration without aeration during colder seasons. To overcome the challenges posed by lower temperatures, it is suggested to adopt one or more alternative strategies. Itokawa et al. (2008) recommended doubling the frequency of MC in colder months compared to summer. Other strategies, such as the use of aeration systems, reducing operational flux, and applying higher concentrations of NaOCl for MC, can also significantly mitigate fouling rates and ensure optimal membrane system performance during colder months.

5.4. Conclusions

This study evaluated the performance of a pilot plant MBR system utilizing a rotating HF membrane module by measuring the increase in TMP and the quality of the filtrate during the filtration operation. The operational variables of filtration flux, filtration-to-backflush time ratio, aeration strategies, and membrane rotational speeds were found to exert an influence on the dynamics of TMP during membrane filtration. Furthermore, two distinct methodologies for conducting MC were employed, and their efficacy in reducing TMP was evaluated. The most significant findings of this study are as follows:

a. Based on the initial phase of the experimental series, the optimal configuration for the pilot plant to operate for 7 days of filtration was a flux of 24 LMH and a filtration-to-backflush time ratio of 9:1 min. The pilot plant also demonstrated the capacity to operate at elevated filtration flux; however, it is not recommended for extended periods of operation.

b. The pilot plant exhibited the ability to operate and achieve the desired level of TMP increase through membrane module rotation even in the absence of aeration. This was corroborated by a subsequent extended period of operation utilizing this configuration. The results indicated that membrane rotation had a more pronounced effect on the control of fouling than the type of aeration employed.

c. Both FB and CB aerations performed in the same manner for the TMP increase behaviour during the experiments. This suggests that FB aeration can be used to enhance oxygen transfer without compromising fouling mitigation and should the membrane module be integrated as a submerged MBR system into an existing aeration tank of a CAS system.

d. Temperature effects on fouling dynamics are correlated with an accelerated TMP increase at lower temperatures. Consequently, it is recommended that MC protocols and operating strategies be adjusted during colder seasons.

Further research could be directed towards optimizing the operating parameters, such as varying the membrane rotational speeds and applying relaxation, to improve the performance of this rotating HF MBR module for its application in wastewater treatment.

5.5. Acknowledgement

The authors would like to express their sincere appreciation to the Ministry of Finance of the Republic of Indonesia for granting the LPDP scholarship to Fathul Mahdariza (ref. number: S-1289/LPDP.4/2018). Special recognition is also given to Dipl.-Ing. Klaus Strätz (Enwat GmbH) and Andreas Scharf (frapp GmbH) for their role in the initial construction and automation of the pilot plant. The authors are deeply indebted to Dr.-Ing. Jochen Henkel for his invaluable contributions, manifested in insightful discussions that greatly enriched the research. Appreciations are also due to Dipl.-Ing. Frank Koch, Dietmar Landgrebe and the dedicated team at WWTP Kassel for their unwavering technical support. Further acknowledgment is extended to Dr.-Ing. Ursula Telgmann, Andrea Brandl, and Monika Degenhardt from Chair of Urban Water Engineering, the University of Kassel, for their meticulous laboratory work. The authors are equally appreciative of the contributions made by Dipl.-Ing. Ralf Feldner and Lukas Marmucki, for their technical assistance. Lastly, the authors express profound gratitude to PD Dr.-Ing. Stephan Fuchs and the esteemed team at the Department of Water Quality Management, Karlsruhe Institute of Technology, whose assistance during the initial stage of this study was invaluable.

6. General Discussion and Outlook

In the field of wastewater treatment, the MBR system offers a number of advantages that have led to its emergence as a promising alternative to the CAS system. The initial concept of the MBR system was to be capable of functioning at significantly higher SRT and MLSS concentrations, with the objective of enhancing the efficacy of the treatment process. Nevertheless, the MBR system encountered significant challenges, including poor oxygen transfer and membrane fouling. This prompted the question of the suitability and sustainability of MBR applications. Another consequence is that, in recent times, MBR systems have been observed to operate at a safer MLSS concentration of 10–15 g/L, which represents a notable reduction from the MLSS concentrations employed in early MBR systems, which were typically in the range of 30 g/L (Ladewig and Al-Shaeli 2017). As a result, a substantial number of studies have been conducted with the objective of identifying solutions to these two principal drawbacks.

This chapter presents a general discussion based on the results described in Chapters 3, 4, and 5. The discussion is focused on oxygen transfer and fouling phenomena during MBR operation. Additionally, the results of an additional experiment are included. Finally, the chapter addresses the future development for this particular membrane module.

6.1. Perspectives on Oxygen Transfer

This study introduces a novel concept of an MBR system featuring a rotatable HF membrane module. The objective of this novel concept is to enhance oxygen transfer while simultaneously mitigating membrane fouling. The incorporation of additional equipment, such as a rotating device for agitation, frequently prompts inquiries regarding the necessity of the additional energy expenditure. However, the rotating HF MBR module system employed in this study demonstrates that the improved oxygen transfer can compensate for the additional energy consumed by the motor for rotation. This is evidenced by the obtained k_La_{20} and SAE values, as discussed in Chapters 3 and 4.

The results of this study indicate that the experiments with fine bubble aeration on activated sludge with MLSS concentration in the range of 11-13 g/L, employing 30 rpm membrane rotation, produced comparable oxygen transfer coefficients to those observed in experiments on activated sludge with MLSS concentration in the range of 3-5 g/L, which is the typical range for CAS systems, in the absence of rotation (see Table 4.4). A similar comparison was obtained for the experiment with coarse bubble aeration on activated sludge with an MLSS concentration in the range of 11-13 g/L, with 50 rpm membrane rotation (see Table 4.3). The results demonstrate that the introduction of rotation represents a promising solution with regard to oxygen transfer, with the potential to enhance the operational MLSS concentration to a level comparable to that observed during the early development stage. It is noteworthy that an attempt was made to thicken the sludge in order to achieve an MLSS concentration above 15 g/L for subsequent experiments. However, due to the nature of activated sludge in the WWTP, it was not possible to obtain a consistent value after the thickening process. Nevertheless, it would be of interest to conduct further experiments with this value range of MLSS concentration on the pilot plant in this study in the future.

Furthermore, this investigation examined the effect of solid concentration on oxygen transfer. It is widely acknowledged that the MLSS concentration exerts a direct influence on oxygen transfer in an activated sludge system, as described in Chapter 2.4.1. The findings of this study align with the prevailing view that higher MLSS concentrations result in lower oxygen transfer coefficients. However, this study demonstrated that the HFV, or solid holdup, provides a more accurate explanation of the correlation between activated sludge floc and α-factor (see Figure 4.5), which is supported by the findings of Henkel et al. (2011). Other studies have defined the relationship between MLSS concentration and α-factor as an exponential correlation (Muller et al. 1995; Germain et al. 2007; Günder 2001; Krampe and Krauth 2003; Xu et al. 2017). However, the results of the clean water and activated sludge experiments in this study indicated that solid holdup appears to have a nearly linear impact on the α-factor, regardless of the reactor type, diffuser system, or membrane rotational speed.

The observed outcome provides further motivation for conducting additional experiments investigating oxygen transfer in aerobic granular sludge (AGS). The AGS was obtained from the Nereda® reactor at the WWTP in Altena, Germany (35,000 PE). The average measured MLSS concentrations for AGS during fine and coarse bubble experiments were 6.6 g/L and 5.7 g/L, respectively. No sludge thickening was conducted for the AGS. Furthermore, in anticipation of the higher tendency of AGS to settle, the experiments were conducted at an airflow rate of 2 m^3/h and higher in order to avoid sludge sedimentation during the experiments, in order to satisfy the requirements stated in DWA-M 209 (2007). As with the findings of the experiments on clean water and activated sludge described in Chapters 3 and 4, the AGS experiments also demonstrated that an increase in airflow rates and membrane rotational speeds led to a higher oxygen transfer coefficient. The highest improvement (60 %) occurred in coarse bubble experiment at specific airflow rate of 2 $m^3_N/(m^3.h)$ and rotational speed of 50 rpm, as shown in Table 6.1.

Table 6.1. Results from oxygen transfer experiment on aerobic granular sludge

Experiments	**MLSS**	**MLVSS**	**HFV**	**Specific airflow 2 $m^3_N/(m^3 \cdot h)$**			**Specific airflow 3.5 $m^3_N/(m^3 \cdot h)$**			**Specific airflow 5 $m^3_N/(m^3 \cdot h)$**		
	(g/L)	**(g/L)**	**(mL/L)**	**k_La_{20} (1/h)**	**Imp.* (%)**	**α (-)**	**k_La_{20} (1/h)**	**Imp. (%)**	**α (-)**	**k_La_{20} (1/h)**	**Imp. (%)**	**α (-)**
Fine bubble:												
0 rpm	6.05	4.65	185	5.24		0.73	8.10		0.74	10.06		0.76
30 rpm	6.95	5.33	210	6.55	+26	0.79	9.96	+25	0.77	12.07	+22	0.76
50 rpm	6.91	5.31	200	7.93	+53	0.71	12.08	+51	0.72	14.67	+48	0.73
Coarse bubble:												
0 rpm	5.12	4.11	173	3.85		0.91	5.19		0.90	6.57		0.90
30 rpm	6.62	5.52	180	4.25	+13	0.81	5.60	+08	0.82	6.75	+03	0.83
50 rpm	5.34	4.15	173	6.15	+60	0.74	7.77	+50	0.76	9.19	+40	0.79

(*) Improvement compared to 0 rpm.

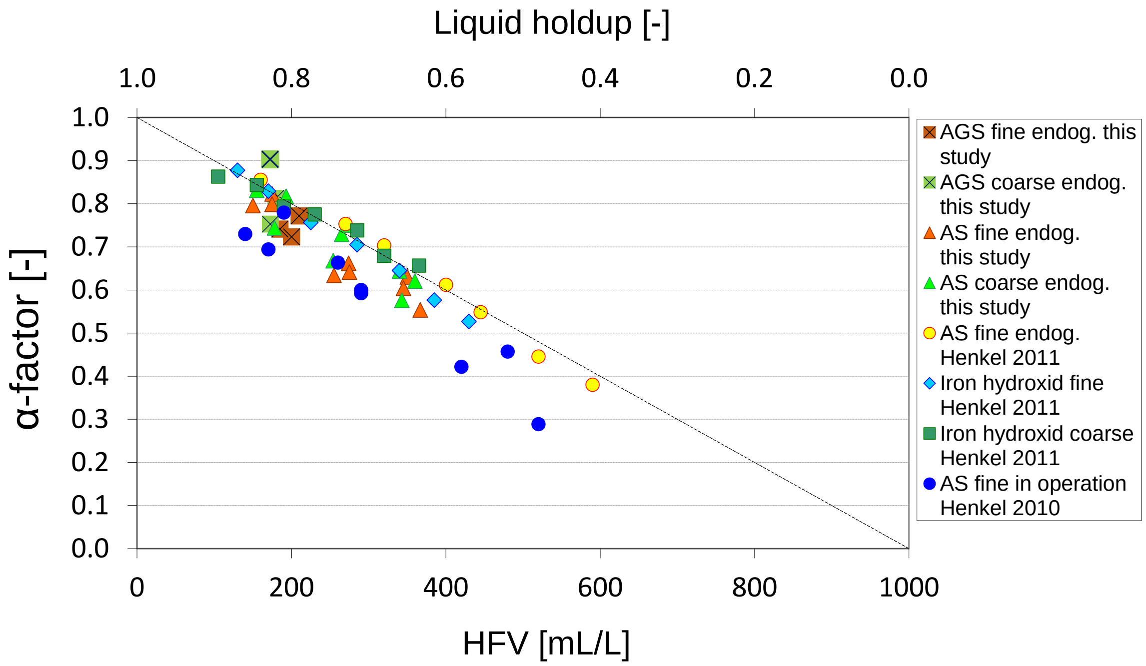


Figure 6.1. The correlation between HFV (solid holdup) and α-factor for clean water, activated sludge and AGS experiments.

Moreover, the results of the AGS experiments are comparable to those of the activated sludge experiments, with no clear correlation observed between rotational speed and the α-factor. However, when the α-factor results from the AGS experiments are integrated with those from other experiments on activated sludge, they follow the pattern previously shown in Figure 4.5, as depicted in the Figure 6.1.

The results of the AGS experiments once again emphasise the necessity of incorporating solid holdup or liquid holdup into the evaluation of α-factors in order to accurately describe the impact of activated sludge floc on oxygen transfer in wastewater treatment.

6.2. Perspectives on Membrane Fouling and Maintenance

In Chapter 5, it was demonstrated that the HF MBR module in this study is capable of operating without any air scouring during the hot months. This result represents a significant saving for side-stream configuration of MBR, where typically an additional aeration system is needed to be installed in the MBR tank and run throughout the year. Should this HF MBR system be implemented in the immersed configuration, the utilisation of fine bubble aeration instead of coarse bubble aeration could prove advantageous in enhancing oxygen transfer within the aeration tank, given that the fouling behaviour observed with this HF MBR module is comparable between fine bubble and coarse bubble aeration (see Figure 5.4).

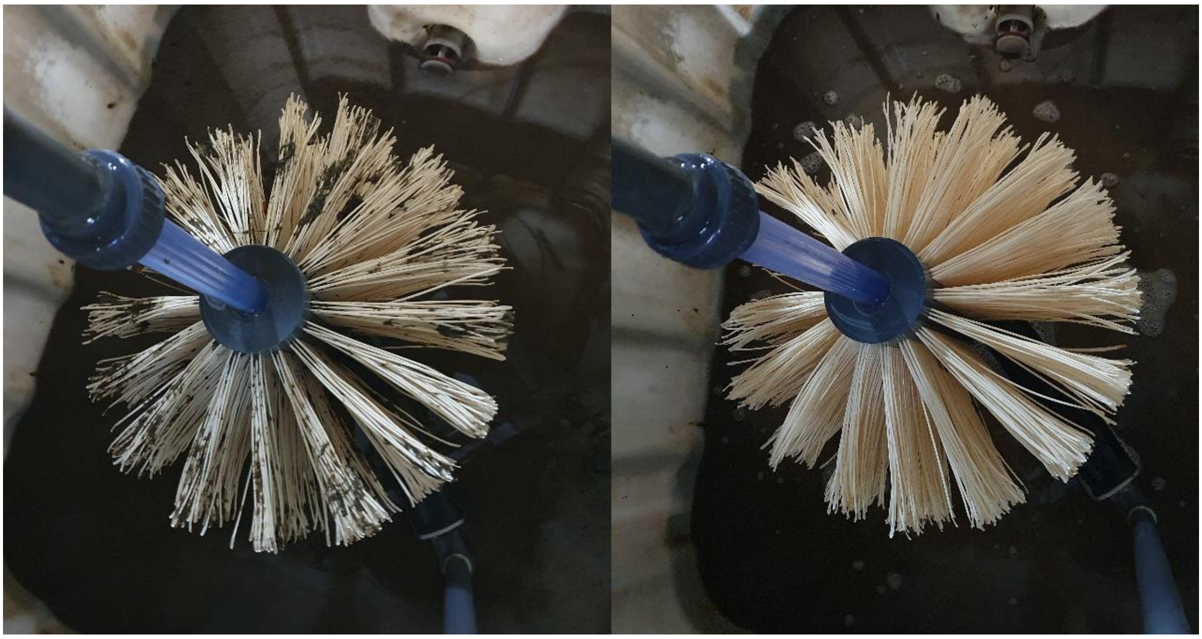


Figure 6.2. HF membrane module following one week of operation (left) and subsequent water spraying and CIP cleaning (right)

The HF MBR module has the advantage of preventing silting, which is often observed in rack-bundled membrane modules and preventing the attachment of waste, such as hairs or small plastics, which have been frequently found in return sludge during filtration experiments, due to the absence of pre-screening in the pilot plant. The less compact design of the module allows less attachment to the membrane surface, thus facilitating the possibility of cleaning the membrane surface using water spray when clean in air method is conducted for maintenance cleaning, as illustrated in Figure 6.2.

Furthermore, the design of this HF MBR system enables the rapid identification of any leakage spots in the event of a fibre being broken or sealing failure, as illustrated in Figure 6.3. The application of backflushing, which is conducted using water if the tank is drained, or using air if the membrane is submerged with clean water, allows for the rapid identification of any leakage spots. This will be of significant benefit to operators. Moreover, the removal of a single broken fibre will not affect the performance of the membrane module, given that in this pilot-scale module, a single fibre contributes to only 0.05 % of the total membrane area.

Figure 6.3. Leakage identification on HF membrane module

6.3. Future Development

The HF membrane module in this study is a prototype, constructed for research purposes and intended for operation in a pilot-scale reactor. Assuming that the membrane module operates at a flux of 24 L/(m^2.h) with a filtration backflush ratio of 9:1, which corresponds to a net flux of 19.2 L/(m^2.h), this pilot-scale membrane module has a filtering capacity of 1.04 m^3/day. However, the distinctive design of this membrane module permits further scale up, as illustrated in Figure 6.4. Furthermore, there is potential for further enhancement of the design to achieve a higher packing density than the current value of 81 m^2/m^3. This is considerably lower than several commercial membrane filters, which have a packing density in the range of 200-600 m^2/m^3 (Judd 2011). Increasing the packing density would result in an increased filtering capacity. Nevertheless, it is worth noting that increasing packing density can result in a non-uniform permeate profile along the fibre length due to the spatial distribution of the cake along the fibre over time. A very high number of packing density can also result in a dramatic decrease in flux. These observations were made by Günther et al. (2010; 2012).

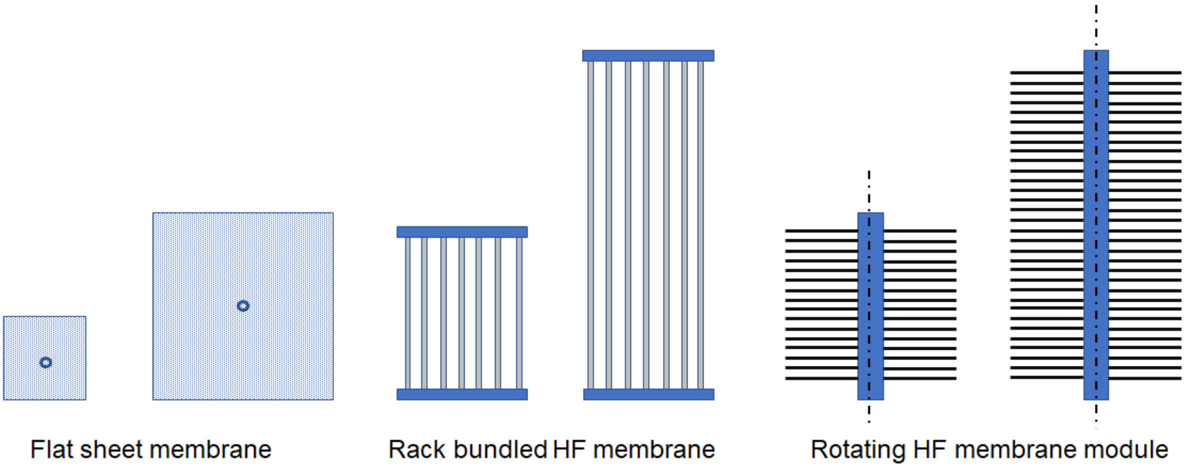


Figure 6.4. Potential scale-up for membrane module

Furthermore, the outcomes of the oxygen transfer experiment on AGS indicate a potential for the implementation of this rotating HF membrane module in a sequencing batch reactor (SBR) system employing AGS. A study by Campo et al. (2021) demonstrated the advantage of combining AGS and MBR. The study found that AGS technology can mitigate membrane fouling due to its high settleability and strong microbial structure. Nevertheless, it is important to note that granule breakage remains a significant concern, as it can lead to increased pore-blocking. Furthermore, a study by Zhang and Jiang (2019) investigated the membrane fouling mechanisms of AGS with different sizes, identifying a critical AGS size (1–1.2 mm) where fouling was most severe. The study found that larger AGS sizes (>1.2 mm) resulted in increased flux and reduced fouling due to a loose cake layer and high permeability. In contrast, smaller AGS sizes (<1 mm) exhibited higher flux and lower fouling due to less EPS formation. The results of the aforementioned studies, in conjunction with the findings of the present study, which demonstrated a positive impact of the rotating module on the enhancement of oxygen transfer on AGS, collectively indicate that the integration of AGS with MBR, specifically with the rotating HF membrane module employed in the present study, is a viable proposition.

With regard to the improvement of membrane material, the current module prototype employs PVDF as its material. However, there is an increasing concern that PVDF may leach components or degrade over time, contributing to per- and polyfluoroalkyl substances (PFAS) contamination in the environment. It is therefore recommended that alternative materials, such as polyether sulfone (PES) or ceramic, be used in future prototypes, and that filtration experiments be conducted once again to observe their fouling behaviour.

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Appendix

A.1. Supplementary Material to Chapter 3

Chapter 3 and the supplementary material have been published in article In Water Science and Technology:

Mahdariza, Fathul; Domingo Rimoldi, Ignacio; Henkel, Jochen; Morck, Tobias (2022): A New Concept of a Rotating Hollow Fibre Membrane Module: Impact of Rotation on Fine-Bubble Aeration. In Water Science and Technology: A Journal of the International Association on Water Pollution Research 85 (9), pp. 2737–2747. DOI: 10.2166/wst.2022.144.

A.1.1. Code used in MATLAB® software for $K_L a_T$ calculation

```
numData = xlsread('Name of excel file')

%The name of the excel file with the oxygen concentration data should be %write here.
% The excel file and this script should be located in the same folder.

x = numData(:,1);
% The first column is time

y = numData(:,2);
% The second column is oxygen concentration

f = @(b,x) -b(1).*exp(-b(2).*x)+b(3);
% Objective Function

co_est = 25; % initial values to start with the iteration process
cs_est = 9,5;
kat_est = 0,004;

B = fminsearch(@(b) norm(y - f(b,x)), [cs_est-co_est; kat_est; cs_est])
% Estimate Parameters

figure
plot(x, y, 'pg')
hold on
plot(x, f(B,x), '-r')
hold off
grid
xlabel('x')
ylabel('f(x)')
text(27, 105, sprintf('f(x) = %.1f\\cdote^{%.3f\\cdotx}%+.1f', B))

R = f(B,x)- y
figure
hold on
plot(x, R, '-r')
hold off
grid
xlabel('x')
ylabel('f(x)')
```

```
X = [B(2)];
disp('k.at[1/s]=')
disp (X)
Y = [B(3)];
disp('Cs[mg/l]=')
disp (Y)
```

A.1.2. Example of the course of DO concentration and residual value curves

The given example is for the experiment in clean water at rotational speed of 30 rpm and airflow rate of 2 m^3/h.

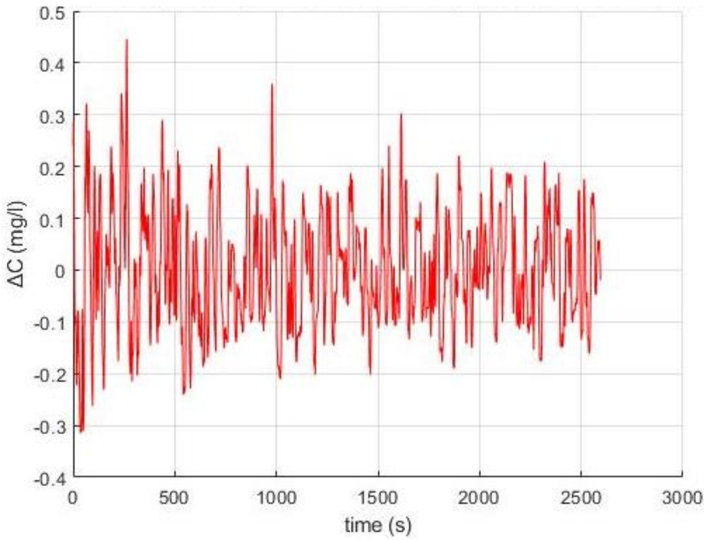

Figure A.1. The residual value curve for the experiment in clean water at rotational speed of 30 rpm and airflow rate of 2 m^3/h.

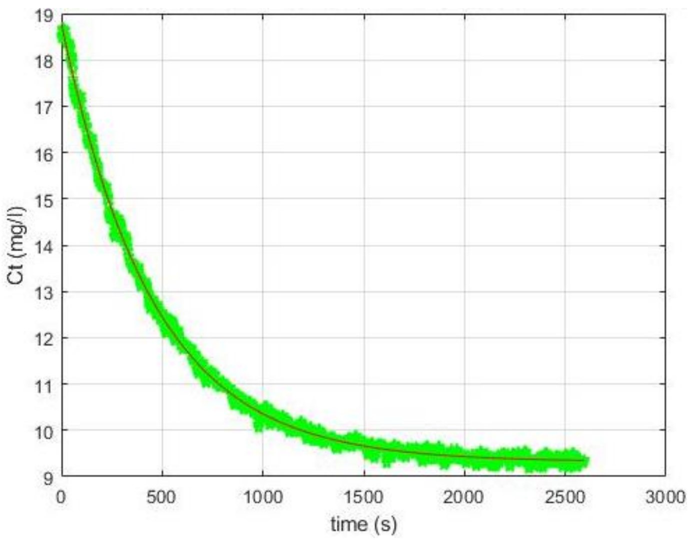

Figure A.2. The course of DO concentration for the experiment in clean water at rotational speed of 30 rpm and airflow rate of 2 m^3/h.

A.1.3. The recorded ambient conditions and obtained oxygen transfer values

Table A.1. The recorded ambient conditions and obtained oxygen transfer values (K_La_T) for all experiments.

Type of water	Rotational speed (rpm)	Average airflow rate (m^3/h)	Ambient pressure (hPa)	Ambient temperature (˚C)	Ambient humidity (%)	Average water temperature (˚C)	Obtained k_La_T (1/h)		
							Sensor 1	Sensor 2	Sensor 3
Clean water	0	0.99	1,008	18.5	36	17.5	3.69	3.60	3.64
		2.00	1,008	18.5	36	17.5	6.40	6.47	6.48
		4.00	1,008	18.5	36	17.5	10.55	10.48	10.50
	10	1.01	1,015	17.8	47	17.4	3.80	3.82	3.85
		2.02	1,012	17.4	50	16.6	6.60	6.49	6.50
		4.04	1,012	17.4	50	16.6	10.42	10.30	10.11
	20	1.01	1,006	15.5	47	15.7	4.11	4.06	4.01
		2.01	1,006	15.5	47	15.7	7.04	6.94	7.15
		4.00	1,006	15.5	47	15.7	11.00	11.04	11.05
	30	0.99	995	18.4	39	17.5	4.67	4.52	4.56
		2.04	1,008	18.5	36	17.5	7.95	7.94	7.96
		4.00	995	18.4	39	17.5	12.86	12.92	12.88
	40	0.99	995	20.7	50	19.0	Sensor was in maintenance	4.88	4.91
		1.98	995	20.7	50	19.0		9.05	9.06
		4.04	995	20.7	50	19.0		14.31	14.34
	50	1.00	1,002	19.5	32	18.2		5.27	5.25
		2.01	1,002	19.5	32	18.2		9.51	9.69
		4.00	1,002	19.5	32	18.2		16.05	16.05
Activated sludge	0	0.99	1,015	17.5	46	16.6	2.85	2.90	2.85
		2.02	1,008	18.5	36	16.6	5.20	5.17	5.23
		4.02	1,008	18.5	36	16.8	8.91	8.81	8.90
	30	1.00	1,011	17.8	47	15.6	3.60	3.64	3.58
		1.94	1,015	16.7	41	14.2	6.13	6.03	6.04
		4.00	1,011	17.8	47	15.6	10.01	9.87	10.07
	50	1.01	1,021	21.3	36	15.3	4.06	4.08	4.07
		2.03	1,021	21.3	36	14.9	7.76	7.81	7.72
		4.04	1,021	21.3	36	14.8	13.12	12.86	12.91

A.1.4. The obtained standard oxygen transfer value

The following Table A.2 shows the obtained k_La_{20} value for clean water experiment, and the data are plotted in the Figure 3.3.

Table A.2. The obtained k_La_{20} value for clean water experiments.

Rotational speed (rpm)	Average airflow rate during experiment (m^3/h)	K_La_{20} value with salt correction factor (1/h)
0	0.99	4.02
	2.00	7.18
	4.00	11.64
10	1.01	4.24
	2.02	7.42
	4.04	11.74
20	1.01	4.70
	2.01	8.15
	4.00	12.77
30	0.99	5.08
	2.04	8.82
	4.00	14.29
40	0.99	5.24
	1.98	9.69
	4.04	15.33
50	1.00	5.74
	2.01	10.47
	4.00	17.51

The following Table A.3 shows the obtained k_La_{20} value for activated sludge experiment, and the data are plotted in the Figure 3.5.

Table A.3. The obtained k_La_{20} value for activated sludge experiments.

Rotational speed (rpm)	Average airflow rate during experiment (m^3/h)	K_La_{20} value with salt correction factor (1/h)
0	0.99	3.10
	2.02	5.62
	4.02	9.54
30	1.00	4.05
	1.94	7.05
	4.00	11.21
50	1.01	4.54
	2.03	8.74
	4.04	14.63

A.1.4. The obtained Standard Aeration Efficiency (SAE)

The following Table A.4 shows the obtained SAE value and the data are plotted in the Figure 3.6.

Table A.4. The obtained SAE values for all experiments.

Type of water	Rotational speed (rpm)	Average airflow rate during experiment (m^3/h)	Standard Aeration Efficiency (kgO_2/kWh)
Clean water	0	0.99	1.09
		2.00	0.95
		4.00	0.78
	30	0.99	0.93
		2.04	0.94
		4.00	0.85
	50	1.00	1.03
		2.01	1.12
		4.00	1.04
Activated sludge	0	0.99	0.85
		2.02	0.76
		4.02	0.65
	30	1.00	0.74
		1.94	0.78
		4.00	0.67
	50	1.01	0.82
		2.03	0.94
		4.04	0.88