

Ultrafiltration harvesting of microalgae culture cultivated in a WRRF: long-term performance and techno-economic and carbon footprint assessment

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Abstract: A cross-flow ultrafiltration harvesting system of a pre-concentrated microalgae culture was tested in an innovative anaerobic-based WRRF. The microalgae culture was cultivated in a membrane photobioreactor fed with the effluent from an anaerobic membrane bioreactor treating sewage. These harvested microalgae biomass was then anaerobically co-digested with primary and secondary sludge from the water line. Depending on the needs of this anaerobic co-digestion, the filtration harvesting process was evaluated intermittently over a period of 212 days for different operating conditions, mainly the total amount of microalgae biomass harvested and the desired final total solids concentration (up to 15.9 g·L⁻¹ with an average of 9.7 g·L⁻¹). Concentration ratios of 15-27 were obtained with average transmembrane fluxes ranged from 5 to 28 L·m⁻²·h⁻¹. Regarding membrane cleaning, both backflushing and chemical cleaning resulted in transmembrane flux recoveries that were, on average, 21% higher than those achieved with backflushing alone. The carbon footprint assessment shows promising results as the GHG emissions associated with the cross-flow ultrafiltration harvesting process could be less than the emissions savings associated with the energy recovered from the biogas production from the anaerobic valorisation of the harvested microalgae.

Keywords: Anaerobic digestion (AD); cross-flow; greenhouse gas (GHG) emissions; harvesting; Membrane photobioreactor (MPBR); microalgae; ultrafiltration (UF); water resource recovery facility (WRRF).

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1. Introduction

Worldwide industrial activities are facing great challenges to become more sustainable and “carbon free” by minimising emissions of greenhouse gases (GHG). Within industrial activities, large-scale wastewater treatment has scarcely considered sustainability in its processes so far, although its energy consumption is highly relevant, as it has prioritised effluents’ quality requirements. The energy consumption of urban wastewater treatment has been reported to account for above 1% of a country’s total energy consumption in Europe [1]. The current paradigm of water treatment needs to be shifted towards more sustainable and circular activities [2]. In fact, the new proposal for the updated wastewater treatment directive, which is currently being revised, aims to achieve energy neutrality by 2040 for all wastewater facilities above 10,000 p.e. [3]. For this reason, many researchers are on the lookout for alternative wastewater treatment systems that can be instrumental in the reduction of GHG emissions.

A green alternative whose interest has grown exponentially in the last decade is the cultivation of microalgae in combination with wastewater treatment. This alternative allows for the reduction of carbon emissions, the recovery of nutrients from water streams and the production of valuable microalgae biomass that contains approximately 50% of carbon, 10% of nitrogen and 1% of phosphorus [4–9]. Furthermore, wastewater valorisation through microalgae cultivation has been reported to reduce environmental impact and energy consumption compared to conventional treatment based on activated sludge systems [10].

Despite the many benefits of microalgae cultivation, current full-scale systems are scarce. One of the main issues of microalgae-based water resource recovery facilities (WRRFs) is the large surface areas that are required for successful cultivation of microalgae, as open reactors' depths are shallow (i.e., 15–40 cm), and hydraulic retention times (HRT) are typically around 3–10 days [11–15]. To improve this, increasing number of authors are testing different options to make the technology more efficient. For instance, some authors are trying to develop instrumentation, control, and automation (ICA) systems to improve microalgae performance [16]. Other reactor configurations have been proposed to increase the photosynthetic efficiency of microalgae, for instance by using thin-layer reactors to increase light availability [17,18]. For example, Morales-Amaral et al. [19] produced higher amounts of biomass in a 2-cm-deep thin-layer reactor than in a raceway pond (43% more). Other authors have incorporated membrane separation technologies to decouple the HRT from the solids retention time (SRT) in order to increase the nutrient loading to the treatment system while maintaining the microalgae biomass in the reactor for longer periods of time, giving them more time to grow and increase productivity [20–21]. Luo et al. [22] managed to reduce the HRT up to 1 d (decoupled from SRT of 9–30 d) using membrane photobioreactor technology. This implied an increase on the nitrogen and phosphorus yields of 25–100%. In addition, Gao et al. [23] were able to increase biomass productivity when reducing the HRT from 6 to 2 d (SRT= 21 d). To achieve this decoupling, the microalgae biomass needs to be separated from the wastewater stream, which is not a simple process and requires the use of harvesting systems that are generally energetically and economically demanding [24, 25]. A harvesting-step is also necessary to concentrate the microalgae biomass with the aim of obtaining added-value by-products, such as biogas from its anaerobic digestion [26], biocrude with high content of alkenes and alkanes [27], biodiesel [28], lipids [29], fatty acids methyl esters [30], nutraceutical applications as linoleic and linolenic acids [31], eicosapentaenoic acid (EPA) and docosahexanoic acid (DHA) [32] or pigments in a biorefinery approach [33]. Harvesting systems such as gravity sedimentation, coagulation-flocculation, electroflocculation, magnetic flocculation, bioflocculation, flotation, centrifugation and filtration have been widely reported (Table 1).

Table 1. Pros and cons of microalgae biomass separation technology.

Technology	Advantages	Disadvantages	References
Sedimentation	Simple. Low CAPEX and OPEX.	Poor settling capacity. Low effluent quality. Low biomass concentration. Biomass losses. Time consuming.	[24, 34, 37]
Coagulation- Flocculation	Proper settling rate. Proper effluent quality. Effective as pre-concentrating step.	High doses of chemical reagents. Possible photosynthesis inhibition from metallic flocculants. Metal presence in harvested biomass. Poor effluent quality for reuse.	[28, 34, 37–40]

Electroflocculation	Flocculants not required.	Metal presence in harvested biomass. Emerging technology (low TRL).	[25, 28, 41]
Magnetic flocculation	Fast, scalable, and efficient. Natural polymers can be added to coat magnetic particles.	Low efficiency for small particles. Low magnetic capacity lifetime. Emerging technology (low TRL).	[25, 38, 39]
Bioflocculation	Toxic chemicals are not required. Prevention of microalgae contamination. Bioflocculants are produced from biomass. Genetic engineering can enhance bioflocculants production.	Factors affecting bioflocculant release remain unclear. Emerging technology (low TRL).	[25, 39]
Flotation	Low CAPEX. Low HRT. High cost-effective method. Small footprint. Possible disruption of microalgae (pro for anaerobic valorisation).	Use of reagents can imply extra cost. Possible disruption of microalgae (con if integrity of biomass is needed).	[24, 39, 42]
Centrifugation	Quick and simple. Most algal cell types can be harvested. Can be used as second-step harvesting process.	High energy demand. High CAPEX. Shear stress to microalgae. Low EPS removal.	[25, 35, 36, 39, 43]
Filtration	High effluent quality. Small footprint. Can be combined with pre-harvesting steps. Moderately to high biomass concentration. Depending on the technology and configuration employed, can ensure the integrity of the microalgae or promote the hydrolysis of microalgae.	Membrane fouling and clogging. Membrane productivity in terms of CAPEX and OPEX. High energy requirements.	[21, 25, 26, 44–49]

Filtration seems to be one of the most promising harvesting methods as it allows to obtain high concentrations (concentration ratios up to a maximum of 150, with final concentrations reaching up TSS of 170 g·L⁻¹), high quality water effluents that could be reused for irrigation; and it can avoid damage to the microalgae cells so that they could be returned to the cultivation unit [2, 21, 45]. Both dead-end (no retentate stream) and cross-flow (CF) ultrafiltration (UF) (retentate and permeate streams) have been reported in literature for microalgae harvesting [21, 39, 50–52] and [21, 39, 44, 45, 47, 50–53], respectively. However, the major concern with membranes is related to fouling, which significantly increases operating costs due to physical cleaning (shear stress from cross-flow velocity in cross-flow filtration, air/gas-sparging in dead-end filtration, backflushing, etc) and chemical cleaning which in turn reduces membrane lifetime [54–55]. Consequently,

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the use of filtration as harvesting process needs further optimisation to enhance the feasibility of microalgae cultivation technology. Zhao et al. [21] gathered energy consumption data for filtration harvesting processes from 2003 to 2022, reporting values up to 3.5 kWh·m⁻³ of treated microalgae culture, with some peak values higher, but the majority falling between 0.2 and 1 kWh·m⁻³ for permeate fluxes between 17 and 45 LMH. Particularly, energy requirements for cross-flow ultrafiltration of microalgae are reported to be between 0.17–2.23 kWh·m⁻³ for initial concentrations of the culture between 1 and 2 g·L⁻¹ and final concentration ratios ranging from 25–113, and transmembrane pressure (TMP) ranging from 0.31 to 2 bar and cross-flow velocities (CFV) from 0.17 to 2 m·s⁻¹ [21, 44, 47, 50, 53, 56]. However, there is still scarce energy consumption data in membrane-based microalgae harvesting [21].

One advantage of harvesting by cross-flow ultrafiltration is that recent studies have reported that it can be used as a pre-treatment step for anaerobic valorisation of microalgae as it promotes their biodegradability (Table 1). For instance, Giménez et al. [26] concluded that the effect of the cross-flow ultrafiltration harvesting technique on the integrity of microalgae cell wall, in terms of viability and biodegradability, exhibited a noticeable effect that increased biodegradability. It was hypothesised that this effect was attributed to the induction of greater shear stress, which is not observed to the same extent in the case of dead-end ultrafiltration. The flexibility afforded by the difference in performance between the dead-end and cross-flow configurations in terms of cell wall integrity is one of the strengths of the filtration technology applied to microalgae harvesting, due to its ability to adapt to different potential uses of the harvested biomass.

The aim of this study is to evaluate the long-term performance of a cross-flow ultrafiltration process to harvest the microalgae biomass cultivated in a membrane photobioreactor for its further valorisation through anaerobic co-digestion. Specifically, this microalgae biomass is produced in an anaerobic-based WRRF platform for sulphate-rich sewage treatment [57]. In addition, this study presented a techno economic assessment and GHG emissions assessment at pilot-scale to facilitate the economic and environmental feasibility study of membrane-based harvesting. Specifically, for GHG emissions, the harvesting unit and a digestion unit for anaerobic valorisation of harvested microalgae were considered, considering different biogas valorisation scenarios.

2. Materials and Methods

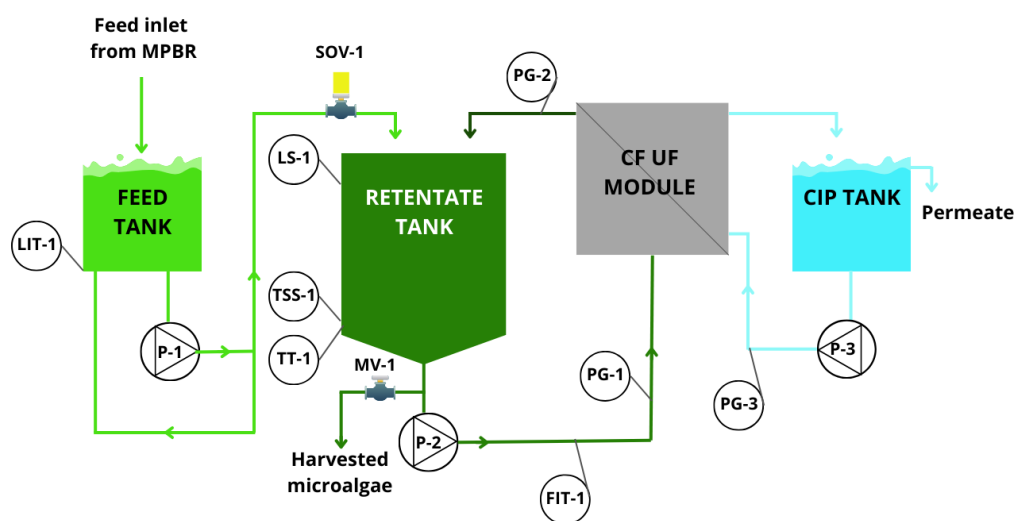
2.1. Description of harvesting pilot plant

The layout of the WRRF is shown in Supplementary materials (Figure S1), which basically consisted of a combination of membrane technologies with anaerobic processes for the valorisation of organic matter from wastewater in the form of biogas with anaerobic membrane bioreactor (AnMBR) technology, and nutrient recovery using microalgae cultivation with membrane photobioreactor technology (MPBR). This MPBR unit used membranes to decouple the SRT from the HRT, which was demonstrated to promote microalgae activity significantly [45]. This initial harvesting stage consisted of two membrane tanks with a set of hollow-fibre ultrafiltration membranes (KMS Puron® Koch Membrane Systems) with a total membrane area of 3.44 m² and an average pore size of 0.03 µm. The performance of this first stage has been previously analysed in González-Camejo et al. [58] The primary microalgae species in the culture were *Coelastrella* and *Desmodesmus*, representing over 99% of the total eukaryotic cells [59].

Anaerobic co-digestion using AnMBR technology was used for the energy recovery of all waste streams from the water line (i.e., primary sludge, AnMBR sludge, and harvested microalgae). Prior to this, a second harvesting stage, using cross-flow ultrafiltration, was used to achieve the desired levels of total suspended solids (TSS) in the microalgae waste (from the MPBR unit) to be fed to the anaerobic co-digestion unit. This pilot-scale facility was located at the Carraixet WWTP (Spain) and used as influent the effluent

from the pre-treatment stage of the full-scale plant. A more detailed description can be found elsewhere [57].

The cross-flow ultrafiltration pilot plant is shown in detail in Figure 1. The installation included a feed tank (0.7 m³), a retentate tank (0.04 m³), the cross-flow membrane module (in a vertical disposition, 1 m long), and, finally, the clean-in-place (CIP) tank (0.2 m³). This unit employed a tubular module with ultrafiltration membranes (HF 5.0-43-PM500, ROMICON® Koch Membrane Systems, USA). The pore size of the membranes had a molecular weight cut-off of 500 kDa, effectively retaining macromolecules. The total filtration surface covered 1 m², ensuring efficient separation and concentration of the microalgae biomass. The total cross-sectional area was 0.0003 m². This plant, which was operated in batch mode in cycles, allowed the removal of excess water, resulting in a highly concentrated algal biomass suitable for subsequent co-digestion processes [60].



CIP: clean-in-place; CF-UF: cross-flow ultrafiltration; FIT: liquid flow-rate transmitter; LIT: level indicating transmitter; LS: level switch; MV: manual valve; P: pump; PG: pressure gauge; SOV: on/off solenoid valve; TSS: total suspended solids; TT: temperature transmitter.

Figure 1. Flow diagram of the cross-flow ultrafiltration-based harvesting system.

2.2 Instrumentation and automation

The cross-flow harvesting system was automated to evaluate the filtration process properly (Figure 1). The following on-line sensors were installed: (i) one level transmitter in the feed tank LIT-1 (Waterpilot FMX167, Endress Hauser) to determine the feed tank level, LFT (m); (ii) one temperature probe in the retentate tank, TT-1 (RTD PT100 RS PRO, RS Components); (iii) one suspended solids sensor in the retentate tank, TSS-1 (SOLITAX ts-line sc LXV423.99.00100); (iv) one liquid flow meter in the retentate conduction (SI-TRANS FM MAG 1100 DN15, SIEMENS); (v) and three pressure gauges (PG-1 in the feed to the membrane, PG-2 in the retentate of the membrane and PG-3 in the backflushing pipe).

In terms of electromechanical equipment, there were 3 pumps: P-1 was the feed pump to the retentate tank in order to continuously replenish it with fresh culture to be concentrated and also to recirculate the fresh culture to the feed tank to guarantee proper mixing conditions; P-2 was the feed pump to the membrane module; whereas P-3 was the auxiliary pump to enable backflushing (Figure 1). Additionally, the plant was equipped with the following auxiliary equipment: an on/off solenoid valve (SOV-1) to allow the feed into the retentate tank; a level switch in the retentate tank, LS-1, acting on the solenoid valve SOV-1; and a valve (MV-1) to enable the collection of harvested microalgae after the completion of each batch's work cycle. The SOLITAX transmitter used in this study was equipped with colour correction and an automatic cleaning valve, which improved the

representativeness and accuracy of the measured and recorded data. Moreover, this sensor was installed with a 30-degree deviation from the regular perpendicular angle to the retentate tank surface to prevent probe fouling and reduce the noise of the signal. Furthermore, a comprehensive sensor cleaning protocol was implemented to improve and ensure stable data acquisition. The SOLITAX probe was calibrated once a week using laboratory data obtained from grab samples that were collected in duplicate from the retentate tank, at the beginning and end of each cycle, according to the Standard Methods [61]: method 2540 E. A linear correlation was established with the values recorded continuously by the SOLITAX probe ($R^2=0.9919$; $p\text{-value} < 0.05$; $n=38$).

All sensors were connected to a programmable logic controller (PLC) for process automation, data acquisition and control. The PLC was communicated with a PC on which supervisory control and data acquisition software (SCADA) was installed to receive, transform, and perform calculations with the obtained data, as well as enable visualisation of all relevant process parameters.

2.3. Pilot plant operation and monitoring

The harvesting plant was operated in batch mode in cycles. In each cycle, the pre-concentrated culture from the MPBR plant stored in the feed tank was pumped (P-1) to the retentate tank until it was full as detected by the level switch (LS-1) which closed the feed valve to the retentate tank (SOV-1). The P-1 pump remained on to ensure proper mixing conditions in the feed tank. The culture in the retentate tank was then continuously fed to the cross-flow membrane producing a permeate which was stored in the CIP tank and a retentate stream which returned to the retentate tank. As the volume of the retentate tank decreased (due to the accumulation of permeate in the CIP), the feed pump (P-1) was automatically activated to replenish the retentate tank. This operation was carried out continuously until the desired level of TSS concentration in the retentate tank was reached (over $8 \text{ g TSS}\cdot\text{L}^{-1}$). In some cycles the feed tank volume was exhausted, but the filtration cycle was not stop because the desired TSS concentration was not achieved. The required TSS concentration varied slightly depending on the requirements of the subsequent anaerobic co-digestion unit in the WRRF pilot plant.

To achieve this TSS concentration, the membrane was operated at a constant transmembrane pressure (TMP), whereas the transmembrane flux (J) remained variable. The TMP (bar) was calculated using Eq. 1 [62]:

$$TMP = \frac{P_1 + P_2}{2} - P_{\text{permeate}} \quad (1)$$

where P_1 represents the inlet gauge pressure (bar) to the membrane module, obtained by PG-1; P_2 is the outlet gauge pressure (bar), measured by PG-2 in the retentate stream; and P_{permeate} is the permeate gauge pressure (bar).

Cheryan [62] established that P_{permeate} should be zero when the permeate is open to the atmosphere, as in the present study. However, given the vertical configuration of the membrane and the outlet location at the top of the module, a constant P_{permeate} of 0.049 bar (0.5 m of water column) was considered.

Once the desired TSS concentration was reached, the harvested microalgae biomass was then removed, and the membrane was cleaned physically (backflushing) and sometimes chemically (every 6 operating cycles). In each cycle, the backflushing pump (P-3) was operated at its maximum capacity with a total backwash volume of 90 liters.

The cross-flow ultrafiltration plant was operated in batch cycles for 212 days (June to January), during which the WRRF pilot plant ran continuously. A total of 78 cycles were performed, two to three times per week (with some periods of non-operation), with an average duration of 25 hours per cycle. The cycle duration varied mainly according to the available pre-concentrated microalgae biomass from the MPBR pilot plant (total volume and TSS concentration), the TSS concentration required for the subsequent anaerobic co-

digestion unit in the pilot plant, and membrane permeability. This initial TSS concentration (TSS_i) was measured off-line in the feed tank and varied depending on the operation of the MPBR plant, which in turn depended on several variables [58, 63, 64]. Table 2 summarises the average operating and outdoor conditions over the 78 cycles.

Table 2. Average operating and outdoor conditions over the 78 cycles of cross-flow ultrafiltration.

Parameter	Unit	Mean ± SD
Temperature	°C	22 ± 6
Inlet gauge pressure (P1)	bar	2.11 ± 0.04
Outlet gauge pressure (P2)	bar	0.48 ± 0.02
Transmembrane pressure (TMP)	bar	1.25 ± 0.02
Initial volume in the feed tank (V _i)	L	450 ± 65
Initial TSS in the feed tank (TSS _i)	g·L ⁻¹	0.50 ± 0.12

For each operating cycle, the following parameters were calculated to fully monitor the filtration process:

- i. Cross flow velocity, CFV (m·s⁻¹), calculated as follows:

$$CFV = \frac{Q_{P-2}}{A_{CF}} \quad (2)$$

where Q_{P-2} is the feed flow rate to membrane module (m³·s⁻¹), A_{CF} is membrane cross-sectional area (m²).

- ii. Permeate flow rate, Q_p (L·h⁻¹), was calculated hydraulically using L_{FT} data from LIT-1, i.e., feed tank level variation (which is equivalent to CIP tank level variation):

$$Q_p = \frac{|\Delta L_{FT}| \cdot A_{FT}}{\Delta t \cdot 1000} \quad (3)$$

where ΔL_{FT} was the variation of L_{FT} (m) in a set period within the work cycle, Δt (h); and A_{FT} was the feed tank area (m²).

- iii. Transmembrane flux, J (LMH), was calculated as follows:

$$J = \frac{Q_p}{A_F} \quad (4)$$

where: Q_p was the permeate flow rate (L·h⁻¹), A_F was the filtration area of the membranes (m²).

- iv. Standardised transmembrane flux at 20 °C, J₂₀ (LMH), calculated as follows:

$$J_{20} = J \cdot e^{-0.0239 \cdot (T-20)} \quad (5)$$

where T (°C) is the temperature of the microalgae culture being concentrated in the harvesting system, measured by TT-1.

- v. Normalised transmembrane flux at 20 °C, J₂₀:J_{20,0},

$$J_{20}:J_{20,0} = \frac{J_{20}}{J_{20,0}} \quad (6)$$

where J_{20,0} is the initial J₂₀ obtained at the start of the entire experiment (32.7 LMH).

- vi. Membrane permeability standardised at 20 °C, K₂₀ (LMH·bar⁻¹) (Eq. 7):

$$K_{20} = \frac{J_{20}}{TMP} \quad (7)$$

- vii. Backflush flow rate, Q_{BF} (L·min⁻¹), was calculated as follows:

$$Q_{BF} = \frac{V_{BF}}{t_{BF}} \quad (8)$$

where V_{BF} (L) was the backflush volume measured in a time t_{BF} (min).

- viii. Transmembrane pressure during backflushing (TMP_{BF}) was calculated by Eq. 9:

$$TMP_{BF} = P_{BF} - \left(\frac{P_1 + P_2}{2} \right) \quad (9)$$

where P_{BF} (bar) is the discharge gauge pressure of pump P-3, measured by PG-3.

ix. Harvested microalgae culture biomass, $M_{TSS_{HV}}$ (g), calculated as follows:

$$M_{TSS_{HV}} = TSS_f \cdot V_{HV} \quad (10)$$

where TSS_f was the TSS recorded by the TSS-1 sensor at the end of each cycle ($g \cdot m^{-3}$), and

V_{HV} was the final volume in the retentate tank at the end of each cycle (m^3).

x. Harvesting rate HV_r ($g \cdot TSS \cdot m^{-2} \cdot h^{-1}$), calculated as follows:

$$HV_r = \frac{M_{TSS_{HV}}}{A_F \cdot t_{operation}} \quad (11)$$

where $t_{operation}$ is the cycle duration (h).

xi. Concentration ratio r (Eq. 12):

$$r = \frac{TSS_f}{TSS_i} \quad (12)$$

2.4. Energy and chemical reagents consumption

The energy consumption of the pumps P-1 and P-2 (W) was calculated adapting the equations proposed by Judd and Judd [65] and Ortiz-Tena et al. [66] (see Eq. 13 and Eq. 22). This energy model has already been applied to different UF membrane systems used in microalgae cultivation [58].

$$W_{P,j} = Q_{P,j} \cdot g \cdot \rho_{culture} \cdot \frac{\left\{ \left[\left(\frac{(L + L_{eq}) \cdot f \cdot v^2}{D \cdot 2 \cdot g} \right)_{asp} + \left(\frac{(L + L_{eq}) \cdot f \cdot v^2}{D \cdot 2 \cdot g} \right)_{imp} \right] + [Z_1 - Z_2] \right\}}{\eta_{pump,j}} \quad (13)$$

where $W_{P,j}$ is the power required for pump j (W), considering both the suction and discharge sections of the pump, using: the volumetric flow rate ($Q_{P,j}$ in $m^3 \cdot s^{-1}$), the acceleration of gravity (g in $m \cdot s^{-2}$), the density of the culture ($\rho_{culture}$ in $kg \cdot m^{-3}$), the length of the pipeline (L in m), the pressure drops due to accidents expressed as equivalent length (L_{eq} in m), the velocity of the culture (v in $m \cdot s^{-1}$), the friction factor (f , dimensionless), the cross-sectional diameter (D in m), the difference in elevation to overcome ($Z_1 - Z_2$, in m), and the efficiency of pump j ($\eta_{pump,j}$, 0.60 [57]).

To calculate the $\rho_{culture}$ ($kg \cdot m^{-3}$), the density of wet green microalgae biomass, $\rho_{a,w}$ ($kg \cdot m^{-3}$) was calculated using Eq. 14 [67]:

$$\rho_{a,w} = x_w \cdot \rho_w + (1 - x_w) \cdot \rho_{a,d} \quad (14)$$

where: x_w was the water mass content of wet green algal biomass, 0.82 [67]; $\rho_{a,d}$ was density of dry green microalgae biomass, $1,400 \text{ kg} \cdot m^{-3}$ [67]; ρ_w was the water density.

The density of the culture $\rho_{culture}$ can be obtained using Eq. 15:

$$\rho_{culture} = \frac{\rho_{a,w} + \rho_w}{\frac{TSS}{\rho_{a,w}} + 1} \quad (15)$$

where: TSS ($kg \cdot m^{-3}$) was microalgae concentration in the culture.

Regarding the viscosity of water, μ_w (Pa·s), it was calculated using Eq. 16 [68], valid between 0 and 40 °C, which aligned with the operating range of this study:

$$\mu_w = \mu_{w,20} \cdot 10^{\left(\frac{20-T}{T+96} \cdot (1.2364 - 1.37 \cdot 10^{-3} \cdot (20-T) + 5.7 \cdot 10^{-6} \cdot (20-T)^2) \right)} \quad (16)$$

where: $\mu_{w,20}$ was $0.001002 \text{ Pa} \cdot s$.

Mass fractional content in dry solids in the microalgae biomass, $w_{a,d}$, can be obtained by Eq. 17 [67]:

$$w_{a,d} = \frac{(1 - x_w) \cdot \rho_{a,d}}{\rho_{a,w}} \quad (17)$$

Microalgae volume fraction in the culture, $\Phi_{v,a}$, was determined by Eq. 18 [67]:

$$\Phi_{v,a} = \frac{TSS}{\rho_{a,w} \cdot w_{a,d}} \quad (18)$$

where TSS ($\text{kg} \cdot \text{m}^{-3}$) was microalgae concentration in the culture.

The viscosity of the microalgae culture can be calculated using Eq. 19, adapted from [67], which provides reliable values for the range $\Phi_{v,a} \leq 0.115$, well above the maximum value in this study:

$$\mu_{culture} = \mu_w \cdot \left(1 - \frac{\Phi_{v,a}}{\Phi_{v,a}^m}\right)^{-2} \quad (19)$$

where $\Phi_{v,a}^m$ was the maximum microalgae volume fraction, established in 0.637.

The friction factor (f) was calculated using the Swamee-James equation (Eq. 20):

$$f = \frac{0.25}{\left[\log_{10} \left(\frac{k/D}{3.7} + \frac{5.74}{\text{Re}^{0.9}}\right)\right]^2} \quad (20)$$

where Re is the dimensionless Reynolds number, k (m) is the internal roughness of the pipe, and D (m) is the cross-sectional diameter of the pipeline.

For the calculation of the power consumed by pump P-3 (W), used in the backflushing stages, Eq. 21 was employed:

$$W_{P,3} = \frac{Q_{BF} \cdot \text{TMP}_{BF}}{0.6 \cdot \eta_{P-3}} \quad (21)$$

Where: η_{P-3} was the efficiency of the pump P-3, with a value of 0.60 ;**Error! No se encuentra el origen de la referencia.**

The total energy consumption EC (kWh) was determined (Eq. 22) by aggregating the individual energy consumptions of each pump ($W_{P,j}$ in kW) multiplied by each operation time, t_j (h), for all three pumps:

$$EC = \sum (W_{P,j} \cdot t_j) \quad (22)$$

Based on the total energy consumption for each cycle, three ratios of interest have been defined:

- the energy consumption ratio of the harvesting system ($EC_m\text{-TSS}$, in $\text{kWh} \cdot \text{tTSS}^{-1}$) per tonne of harvested microalgae biomass ($M_{\text{TSS}_{HV}}$, in t) (Eq. 23);
- the energy consumption ratio of the harvesting system ($EC_v\text{-HV}$, in $\text{kWh} \cdot \text{m}^{-3}$) per treated volume of pre-concentrated microalgae culture, i.e., the initial volume of the feed tank (V_i , in m^{-3}) (Eq. 24);
- the energy consumption ratio of the harvesting system ($EC_v\text{-WRRF}$, $\text{kWh} \cdot \text{m}^{-3}$) per treated volume of water in the WRRF pilot plant (V_{WRRF} treated to generate V_{HV} , in m^{-3}) (Eq. 25):

$$EC_m\text{-TSS} = \frac{EC}{M_{\text{TSS}_{HV}}} \quad (23)$$

$$EC_v\text{-HV} = \frac{EC}{V_i} \quad (24)$$

$$EC_v\text{-WRRF} = \frac{EC}{V_{\text{WRRF}}} \quad (25)$$

The chemical cleanings of the membrane were effectively carried out using a dose of 200 ppm of sodium hypochlorite (NaClO) for 10 minutes, followed by rinsing with water only. As already mentioned, the chemical cleanings were performed every 150 hours of operation, i.e., 6 cycles (on average). The parameter $NaOClC_v\text{-WRRF}$ ($\text{g Cl}\cdot\text{m}^{-3}$) was defined to determine the sodium hypochlorite reagent consumption ratio per m^3 treated in the WRRF (Eq. 26).

$$NaOClC_v\text{-WRRF} = \frac{M_{\text{Cl}}}{V_{\text{WRRF}_{cc}}} \quad (26)$$

where: M_{Cl} (g Cl) is the NaClO mass used in chemical cleanings expressed as active chlorine, and $V_{\text{WRRF}_{cc}}$ (m^{-3}) is the amount of water treated in the WRRF since the last chemical cleaning.

Finally, the more relevant operating costs for the cross-flow ultrafiltration pilot plant were estimated, considering energy consumption and the use of sodium hypochlorite in chemical cleanings, $OPEX_{EC+Cl}$ ($\text{€}\cdot\text{m}^{-3}$ treated in WRRF). The cost for electricity was the average for the UE-27 in October 2023 ($0,178 \text{ €}\cdot\text{kWh}^{-1}$) [69]. The cost for hypochlorite in Europe was taken from [70] ($0.013 \text{ €}\cdot\text{gCl}^{-1}$).

2.5. GHG estimation.

To estimate the GHG emissions associated with the cross-flow ultrafiltration process, the methodology proposed by Parravacini et al. [71] was considered adapted similarly to previous works [60], and taking into account [72]. This methodology includes the estimation of both direct (GHG_{direct}) and indirect (GHG_{indirect}) emissions related to the energy consumption of the harvesting unit and the energy consumption and biogas production of the anaerobic digestion of the harvested microalgae under mesophilic conditions. The biogas production from the anaerobic digestion of harvested microalgae biomass was determined based on the experimental data from previous studies [26, 73]. In addition, the increase in methane yield due to the effect of cross-flow ultrafiltration on the microalgae valorisation (i.e., higher anaerobic biodegradability), according to **¡Error! No se encuentra el origen de la referencia.**, was also considered. In particular, this effect was taken into account by considering the enhancement of biogas production according to the concentration ratio (r) of the harvested microalgae, and it has been presented in Supplementary Materials (Figure S2). Methane yield correlated directly with the concentration ratio (r) for r values below 13.3. In contrast, for r values above 13.3 methane yield remained constant at $0.294 \text{ Nm}^{-3}\cdot\text{kgTSS}^{-1}$.

As for biogas valorisation, three scenarios were established: (i) Scenario 1, involving the use of high-efficiency cogeneration technology with a combined heat and power (CHP) system; (ii) Scenario 2, considering the upgrading to biomethane using conventional membrane technology and its subsequent injection into the grid; and (iii) Scenario 3, assessing the upgrading to biomethane using microalgae culture, which increased methane production during the anaerobic digestion (AD) stage due to the additional production of algal biomass at the upgrading unit, and subsequent injection into the grid, but this technology had not the possibility of heat recovery.

To determine the net energy demand per tonne of harvested microalgae biomass, E ($\text{kWh}\cdot\text{tTSS}^{-1}$) by means of (Eq. 27), was necessary to estimate: Q_{demand} was the total thermal energy demand ($\text{kWh}\cdot\text{tTSS}^{-1}$) (Eq. 28); W_{demand} was the total electrical energy demand ($\text{kWh}\cdot\text{tTSS}^{-1}$) (Eq. 29); and Q_{BM} was the biomethane production in Scenarios 2 and 3 ($\text{kWh}\cdot\text{tTSS}^{-1}$), calculated by (Eq. 30) if Q_{demand} was positive, and by (Eq. 31) if Q_{demand} was zero or negative.

$$E (\text{kWh} \cdot \text{tTSS}^{-1}) = Q_{\text{demand}} + W_{\text{demand}} + Q_{\text{BM}} \quad (27)$$

$$Q_{demand}(kWh \cdot tTSS^{-1}) = Q_{TOT} - Q_{recovered} \quad (28)$$

$$W_{demand}(kWh \cdot tTSS^{-1}) = W_{TOT} - W_{recovered} \quad (29)$$

$$Q_{BM}(kWh \cdot tTSS^{-1}) = (Q_{demand} - Q_{BG}) \cdot \varphi_{upgrading} \quad (30)$$

$$Q_{BM}(kWh \cdot tTSS^{-1}) = -Q_{BG} \cdot \varphi_{upgrading} \quad (31)$$

where Q_{TOT} is the heat required by the anaerobic digestion process per tonne of harvested microalgae biomass ($kWh \cdot tTSS^{-1}$); $Q_{recovered}$ is the heat recovered in the CHP system for Scenario 1, or the heat recovered in the upgrading stage for Scenario 2 ($kWh \cdot tTSS^{-1}$); W_{TOT} are the electrical consumptions of the equipment of the cross-flow ultrafiltration pilot plant, the anaerobic digester and the valorisation system of each scenario per tonne of harvested microalgae biomass ($kWh \cdot tTSS^{-1}$); $W_{recovered}$ is the electricity recovered in the CHP stage for Scenario 1 per tonne of harvested microalgae biomass ($kWh \cdot tTSS^{-1}$); Q_{BG} is the gross production of raw biogas per tonne of harvested microalgae biomass; $\varphi_{upgrading}$ is the efficiency of the upgrading process. Ratios, efficiencies, and emission factors are taken from [60, 71, 74–77].

Direct GHG emissions, indirect GHG emissions, and total GHG emissions per tonne of harvested microalgae biomass (M_{TSSHV} , in t) were calculated using the equations 32–35.

$$GHG_{direct,sc.1}(kgCO_2e \cdot tTSS^{-1}) = M_{BG} \cdot EF_{CH_4,sc.1} \cdot 28 \quad (32)$$

$$GHG_{direct,sc2\&3}(kgCO_2e \cdot tTSS^{-1}) = M_{BM} \cdot EF_{CH_4,sc,i} \cdot 28 \quad (33)$$

$$GHG_{indirect}(kgCO_2e \cdot tTSS^{-1}) = (Q_{BM} + Q_{demand}) \cdot EF_{natural\ gas} + W_{demand} \cdot EF_{electricity} \quad (34)$$

$$GHG_{total} = GHG_{direct} + GHG_{indirect} \quad (35)$$

where M_{BG} is the calculated gross production of raw biogas per tonne of harvested microalgae biomass ($kgCH_4 \cdot tTSS^{-1}$); M_{BM} is the production of biomethane per tonne of harvested microalgae biomass ($kgCH_4 \cdot tTSS^{-1}$); $EF_{CH_4,i}$ is the methane losses emission factor for each scenario; Q_{demand} is the total thermal energy demand per tonne of harvested microalgae biomass ($kWh \cdot tTSS^{-1}$); W_{demand} is the total electrical energy demand per tonne of harvested microalgae biomass ($kWh \cdot tTSS^{-1}$); $EF_{natural\ gas}$ is the specific emission factor for fossil natural gas from the grid in Europe; and $EF_{electricity}$ is the specific emission factor of European power companies.

3. Results

3.1. Filtration performance.

The harvesting pilot plant was in operation for a total of 1968 hours, intermittently distributed over a total of 78 cycles during a 212-day of operation of the WRRF pilot plant. Table 3 shows the average values of the main filtration performance parameters, where the resulting J and concentration ratios are in the range of other similar studies [21, 50].

Table 3. Average filtration performance over the entire operating period (78 cycles).

Parameter	Unit	Mean ± SD
TSS _f	g·L ⁻¹	9.7 ± 1.7
CFV	m·s ⁻¹	1.2 ± 0.3

J_{20}	LMH	16 ± 8
Normalised flux ($J_{20}:J_{20,0}$)	-	0.49 ± 0.23
K_{20}	LMH·bar ⁻¹	13 ± 6
TMP_{BF}	bar	1.21 ± 0.08
Q_{BF}	L·min ⁻¹	31 ± 8
HV_r	g·m ⁻² ·h ⁻¹	9 ± 3
r	-	21 ± 6
Cycle duration	h	25 ± 9

CFV: cross flow velocity; HV_r: biomass harvesting rate; J_{20} : transmembrane flux standardised at 20 °C; $J_{20,0}$: initial J_{20} ; K_{20} : membrane permeability standardised at 20 °C; Q_{BF} : backflush flow-rate; r: concentration ratio; TMP_{BF} : backflush stage transmembrane pressure; TSS_r: total suspended solids of the harvested biomass.

The filtration performance varied mainly according to the available pre-concentrated microalgae biomass from the MPBR pilot plant (TSS_i·V_i), the TSS concentration required for the subsequent anaerobic co-digestion unit (TSS_f), and the initial membrane permeability of each cycle depending on the membrane cleaning previously applied (physical or physical+chemical), in terms of normalised $J_{20}:J_{20,0}$. These operating conditions and the resulting filtration performance are shown in Figure 2a with the evolution of the TSS concentration in the retentate tank during each operating cycle, and in Figure 2b for 20 °C-standardised transmembrane flux normalised by the initial flux ($J_{20}:J_{20,0}$); concentration ratio (r) and the total mass culture to be harvested (total volume (V_i) per total suspended solids (TSS_i) in the feed tank). Both the total mass culture harvested (102–700 g) and the final TSS (8.0–15.9 g·L⁻¹) in each cycle varied significantly in some periods which strongly influenced the performance of the filtration cycle, in terms of the duration of the cycle and the final values obtained for the normalised transmembrane flux (Figure 2b) at the end of the cycle. For instance, the cycle with the highest TSS_f and total mass culture harvested (cycle at 149–150 operating days) lasted 74 hours and $J_{20}:J_{20,0}$ reached the lowest values of the operating period, i.e., 0.08.

Backflushing stages were performed at the end of each cycle, and in Figure 2b the dashed lines indicate the end of the cycles when chemical cleaning was also performed (backflush followed by chemical cleaning). Within each cycle, $J_{20}:J_{20,0}$ started with a relatively high value and decreased as the cycle progresses, due to the increase in TSS concentration in the retentate tank. After each BF stage, when considering consecutive cycles with similar TSS_i, $J_{20}:J_{20,0}$ at the beginning of the cycle partially recovered compared to the initial one of the previous cycle. However, full recovery was not achieved because backflushing only removed reversible fouling [58, 78]. Indeed, an accumulation of the irreversible fouling is indirectly observed with the decrease of $J_{20}:J_{20,0}$ at the beginning of the consecutive cycles between the dashed lines (between chemical cleanings). In contrast, when a backflush followed by a chemical cleaning was performed, a greater recovery of filtration capacity can be observed, expressed as $J_{20}:J_{20,0}$. The average of the initial values for $J_{20}:J_{20,0}$ obtained from cycles immediately after a combination of backflushing and chemical cleaning was 0.98 ± 0.01 , in contrast to the initial values from cycles after backflushing alone, which were 0.81 ± 0.10 .

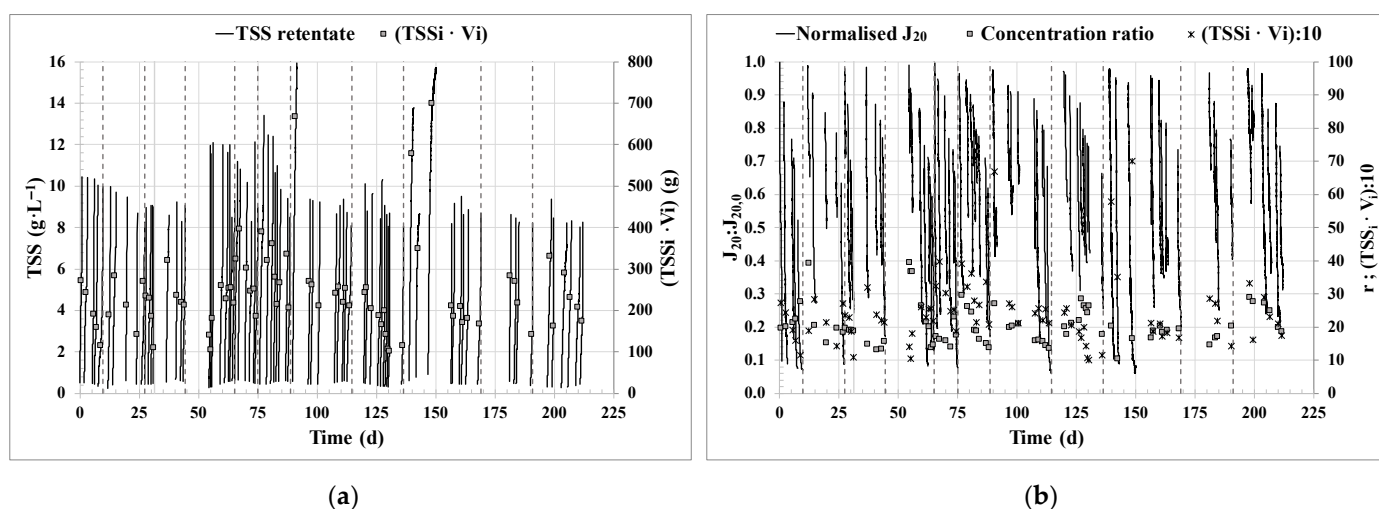


Figure 2. Evolution of filtration performance regarding: (a) TSS concentration in the retentate and biomass to be harvested (TSSi·Vi); (b) 20 °C- standardised transmembrane flux (J_{20}) normalised by the initial flux ($J_{20,0}$); concentration ratio (r) and biomass to be harvested (TSSi·Vi)/10. Dashed lines indicate the performance of chemical cleanings.

In Figure 3, detailed figures for the period between day 54 and day 78 of operation are presented (cycles 19 to 33), in an arbitrary manner, to observe the evolution of J_{20} : $J_{20,0}$, concentration ratio, biomass to be harvested, and TSS in the retentate stream over only a few batch-operating cycles. In Figure 3b, the partial effect of the backflushing at the end of each cycle on the J_{20} : $J_{20,0}$ recovery is evident, as well as the significantly higher effect of the chemical cleaning performed after cycles 26 and 32. Figure 3a shows that the evolution of the TSS concentration in the retentate tank had an inflection point in each cycle. This inflection point corresponds to the time when the feed tank volume was exhausted, removing the dilution effect and reducing the retentate volume and the transmembrane flux as seen by the slower decline of J_{20} : $J_{20,0}$ (Figure 3b). Cycles 22 to 26 (from day 58 of operation prior to chemical cleaning) shown in Figure 3 were operated for a similar total culture mass harvested but decreasing concentration ratio, however a gradual increase in cycle time was observed as the values of J_{20} : $J_{20,0}$ decreased due to the accumulation of irreversible fouling.

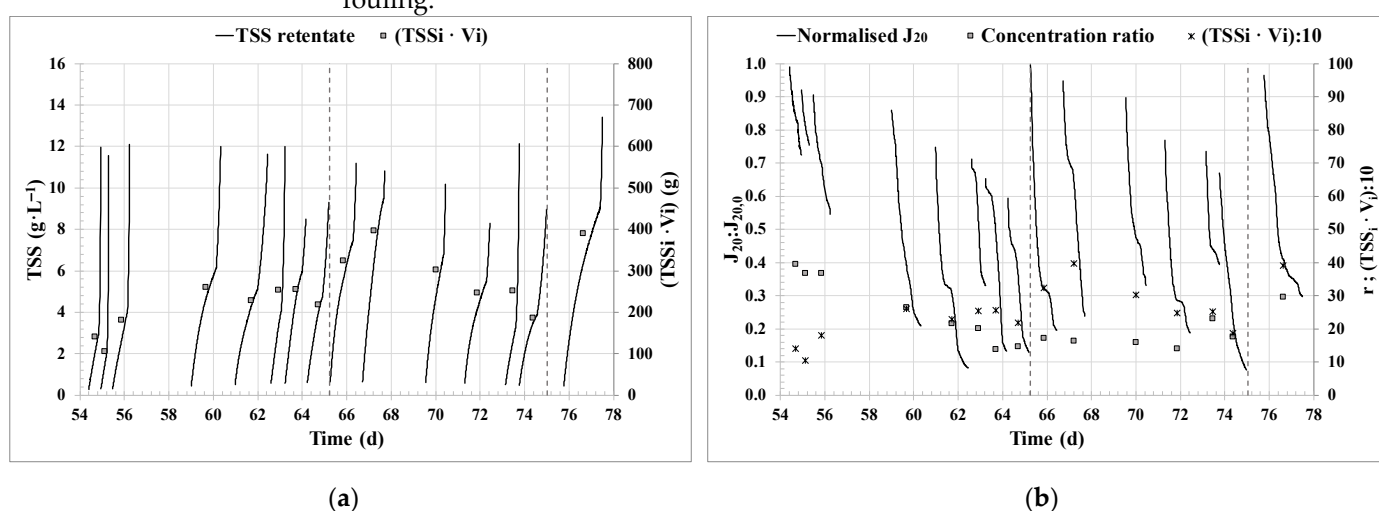


Figure 3. Details of the filtration performance for an aleatory period of 24 days of operation (cycles 19 to 33): (a) TSS concentration in the retentate and biomass to be harvested (TSSi·Vi); (b) 20 °C- standardised transmembrane flux (J_{20}) normalised by the initial flux ($J_{20,0}$); concentration ratio (r) and biomass to be harvested (TSSi·Vi)/10. Dashed lines indicate the performance of chemical cleanings.

One notable strength of the cross-flow ultrafiltration technology as second harvesting step lies in its robustness and reliability during operation, consistently ensuring the attainment of the $8 \text{ g}\cdot\text{L}^{-1}$ target. This target was necessary for the overall functionality of the WRRF pilot plant, specifically in the anaerobic digestion process [57]. In analogous experiments involving dead-end hollow-fibre ultrafiltration membrane modules as the second harvesting step, the $8 \text{ g}\cdot\text{L}^{-1}$ objective was not consistently achieved (data not shown). This discrepancy was likely attributed to variations in the TSS concentrations of the pre-concentrated feed stream or the specific of microalgae species present. However, it is well known that the energy demand of cross-flow operation is higher than that of dead-end operation. There is therefore room for improvement.

3.2. Techno-economic and carbon footprint assessment.

Table 4 shows the values of the energy ratios and chemical reagent consumption rates defined in equations 24 to 27 for the entire operating period.

Table 4. Table 4. Energy and chemical reagents consumption of the cross-flow ultrafiltration unit.

Parameter	Acronym	Unit	Mean $\pm$ SD
Energy consumption ratio per harvested algal biomass	EC_m_TSS	$\text{kWh}\cdot\text{kg}^{-1}$	1.51 ± 0.64
Energy consumption ratio per treated volume of pre-concentrated microalgae culture	EC_v_HV	$\text{kWh}\cdot\text{m}^{-3}$	0.76 ± 0.32
Energy consumption ratio per total treated volume in the WRRF	EC_v_WRRF	$\text{kWh}\cdot\text{m}^{-3}$	0.39 ± 0.16
Sodium hypochlorite consumption per total treated volume in the WRRF	$NaOClC_v_WRRF$	$\text{g Cl}\cdot\text{m}^{-3}$	0.97 ± 0.39
Operating expenses of energy and sodium hypochlorite per total treated volume in the WRRF	$OPEX_{EC+Cl}$	$\text{€}\cdot\text{m}^{-3}$	0.082 ± 0.034

The energy ratios obtained in this study for an average concentration ratio of the harvested biomass of 21 are within the normal range reported in the literature for this type of biomass and technology [21]. It should be noted that the present study is a pilot plant performance of almost 2,000 h of batch operation over a 7-month operation of the WRRF pilot plant. In particular, the obtained energy demand is also in the range of other membrane-based microalgae harvesting systems, i.e., $0.17\text{--}2.23 \text{ kWh}\cdot\text{m}^{-3}$ [21, 50]. Other harvesting technologies are even more energy intensive, such as centrifugation, which can consume up to $8 \text{ kWh}\cdot\text{m}^{-3}$ to achieve 90% of efficiency [21]. Important to note, is that there is room for improvement in the resulting filtration performance, for instance by means of optimising physical and chemical cleanings frequency.

Figure 4 shows, for a concentration ratio of 21, the results of the energy balance (Q_{demand} , W_{demand} and Q_{BM}), and the GHG assessment (GHG_{direct} , $GHG_{indirect}$) related to the harvesting unit and the anaerobic digestion unit for the harvested biomass for the three biogas valorisation scenarios considered.

As Figure 4a shows, Scenario 1 would not be able to cover all the electrical consumption (W_{demand}), but would produce a significant surplus of heat, Q_{demand} . In the case of Scenario 2, heat recovery from the conventional membrane upgrading would cover the heat requirements of the mesophilic digestion unit, and the substantial W_{demand} requirements would be greatly exceeded by the primary energy content of the biomethane injected into the grid, Q_{BM} . The difference between Scenarios 2 and 3 in terms of Q_{demand} can be explained by the lack of thermal recovery in microalgae-based harvesting systems. In Figure 4b, GHG emissions reflect the energy assessment due to the savings related to the fossil energy sources substituted by these renewable sources. Scenario 3 would be the most fa-

vourable one both in terms of energy recovery and GHG emissions savings. This improved performance could be primarily attributed to the additional biomethane production from CO₂ absorbed in the microalgae-based upgrading stage compared to both Scenarios 1 and 2, along with the low energy consumption of the upgrading process and the high methane recovery efficiency (exceeding 97.6%), compared to a combined efficiency of approximately 68% for co-generation in Scenario 1.

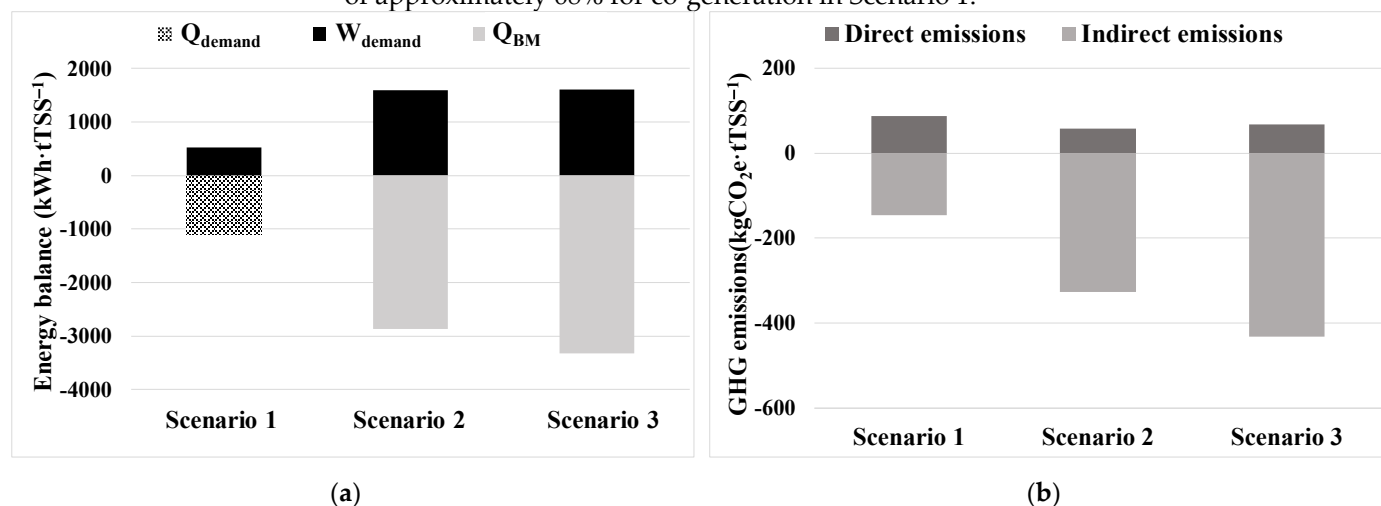


Figure 4. Energy balance and GHG assessment for a concentration ratio of 21 for three different biogas valorisation scenarios: Scenario 1: co-generation; Scenario 2: membrane upgrading; Scenario 3: microalgae upgrading: (a) energy balance in terms of Q_{demand}, W_{demand}, and Q_{BM}; (b) direct and indirect GHG emissions.

Figure 5 illustrates the net GHG emissions for all batch cycles vs. the concentration ratio (r), in the case of Scenario 3, chosen as an example since the trend is proportional for all scenarios. Given that most of the employed (r) values surpass the maximum methane yield obtained from previous results, i.e. the same methane yield is considered for r values above 13.3 [26, 73], the most favourable GHG emission values are observed for relatively low (r) values, around 13–17. These results are promising, as the associated emissions could remain below zero for cross-flow ultrafiltration membranes over a wide range of operating conditions, thereby resulting in net emissions savings.

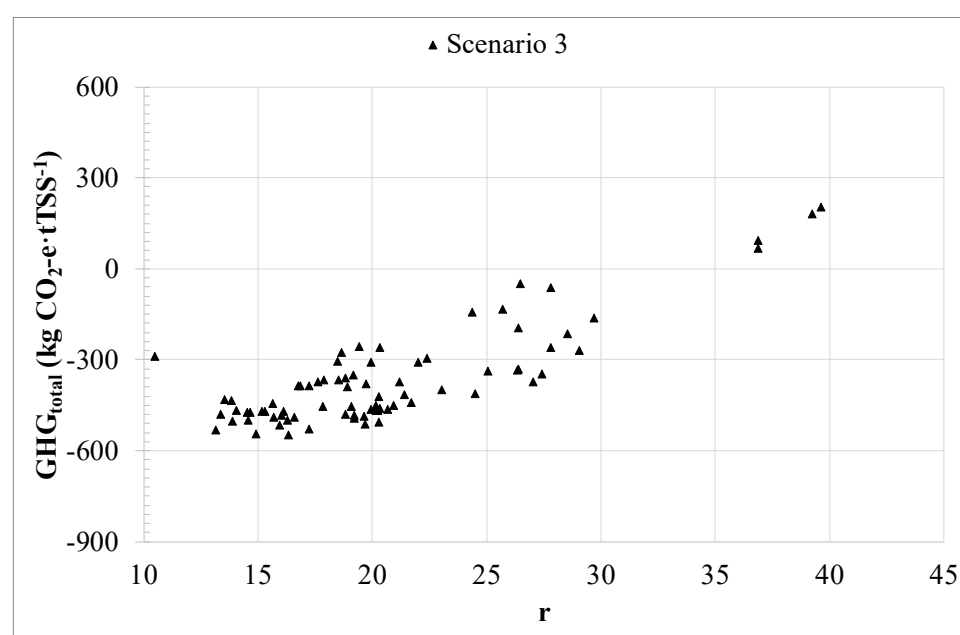


Figure 5. Total GHG emissions for Scenario 3 (microalgae upgrading to biomethane).

4. Conclusions

Cross-flow ultrafiltration appeared to be a robust technology for harvesting pre-concentrated microalgae biomass, which operated efficiently under variable operating conditions and was able to achieve biomass concentrations of up to $16 \text{ gTSS} \cdot \text{L}^{-1}$ (concentration ratio of up to 39.5). The performance of cross-flow ultrafiltration was comparable to other membrane-based harvesting systems in terms of transmembrane flux, concentration ratios and energy requirements. A backflushing step was performed after each batch cycle, and the periodic combination of backflushing with chemical cleaning was necessary for long-term operation. The recovery of transmembrane flux after physical and chemical cleaning was significant, 21% higher than that achieved with backflushing alone. Based on these results, further optimisation is required to improve membrane productivity in terms of CAPEX and OPEX by optimising physical and chemical cleaning. If the harvested microalgae biomass were valorised for biogas production, the direct and indirect GHG emissions associated with the harvesting process could be less than the emissions savings associated with the energy recovered through the production of this renewable energy source, especially if the biogas is upgraded through a microalgae cultivation process.

Supplementary Materials: The following supporting information can be downloaded at: www.mdpi.com/xxx/s1, Figure S1: Flow diagram of the WRRF scheme; Figure S2: Variation of biodegradability of the microalgae biomass and methane yield production with respect to concentration ratio (r).

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Glossary

A_{CF}	Membrane cross-sectional area
AD	Anaerobic digestion
A_F	Filtration area of the membranes
A_{FT}	Feed tank area
AnMBR	Anaerobic membrane bioreactor
CF	Cross-flow
CF-UF	Cross-flow ultrafiltration
CFV	Cross-flow velocity
CHP	Combined heat and power system

CIP	Clean-in-place	553
D	Cross-sectional diameter	554
E	Energy balance calculation	555
EC	Total energy consumption	556
EC _{m_TSS}	Energy consumption ratio of the harvesting system per harvested microalgae biomass	557
EC _{v_HV}	Energy consumption ratio of the harvesting system per treated volume of pre-concentrated microalgae culture	558 559
EC _{v_WRRF}	Energy consumption ratio of the harvesting system per treated volume of water in the WRRF	560 561
EF _{CH4}	Methane losses emission factor	562
EF _{electricity}	Specific emission factor of european power companies	563
EF _{natural_gas}	Specific emission factor for fossil natural gas from the grid, in Europe	564
f	Friction factor	565
FIT	Liquid flow-rate transmitter	566
g	Acceleration of gravity	567
GHG	Greenhouse gas	568
GHG _{direct}	Direct greenhouse gases emissions	569
GHG _{indirect}	Indirect greenhouse gases emissions	570
GHG _{total}	Total greenhouse gases emissions	571
HRT	Hydraulic retention time	572
HV _r	Harvesting rate	573
ICA	Instrumentation, control and automation	574
J	Transmembrane flux	575
J ₂₀	20 °C-standardised transmembrane flux	576
J _{20,0}	Initial 20°C-standardised transmembrane flux at the inception of the entire experiment	577
J _{20;J_{20,0}}	Normalised transmembrane flux at 20 °C	578
k	Internal roughness of the pipe	579
K ₂₀	20 °C-standardised permeability	580
l	Length of the pipeline	581
l _{eq}	Pressure drops due to accidents expressed as equivalent length	582
L	Level	583
LIT	Level indicating transmitter	584
LMH	Litter per square meter and hour	585
LS	Level switch	586
M _{BG}	Gross production of raw biogas, expressed as methane mass	587
M _{BM}	Biomethane production, expressed as methane mass	588
M _{Cl}	Sodium hypochlorite reagent mass used in chemical cleaning	589
M _{TSS_{HV}}	Harvested microalgae culture biomass	590
MPBR	Membrane photobioreactor	591
MV	Valve	592

NaOCl _{Cv_WRRF}	Sodium hypochlorite reagent consumption ratio per m ³ treated in the WRRF	593
OPEX _{EC+Cl}	Operating costs for energy consumption and sodium hypochlorite for the CF-UF pilot plant per m ³ treated in the WRRF	594 595
p.e.	Population equivalent	596
PG	Pressure gauge	597
P _j	Pressure at point j	598
P-j	Pump number j	599
PLC	Programmable logic controller	600
Q _{BG}	Gross production of raw biogas, expressed as primary energy	601
Q _{BM}	Biomethane production, expressed as primary energy	602
Q _{demand}	Total thermal energy demand	603
Q _j	Flow-rate for pump or stream j	604
Q _{recovered}	Heat recovered by the biogas valorisation system	605
Q _{TOT}	Heat required by the anaerobic co-digestion process	606
r	Concentration ratio	607
SCADA	Supervisory control and data acquisition software	608
SOV	On/off solenoid valve	609
SRT	Solids retention time	610
T	Temperature	611
TMP	Transmembrane pressure	612
TSS	Total suspended solids	613
TSS _f	Final TSS concentration for an operation cycle	614
TSS _i	Initial TSS concentration for an operation cycle	615
TT	Temperature transmitter	616
UF	Ultrafiltration	617
v	Velocity	618
V _{HV}	Final volume in the retentate tank for an operation cycle	619
V _{WRRF}	Volume of water treated in the WRRF to generate V _{HV}	620
V _{WRRFcc}	Volume of water treated in the WRRF since the last chemical cleaning	621
V _i	Initial volume in the feed tank for an operation cycle	622
w _{a,d}	Mass fractional content in dry solids in green microalgae biomass	623
W _{demand}	Total electrical energy demand	624
W _{P,j}	Power required for pump j	625
W _{recovered}	Electricity recovered by the biogas valorisation system	626
WRRF	Water resource recovery facility	627
x _w	Water mass content of wet green microalgae biomass	628
Z _j	Elevation of point j	629
ΔL _j	Variation of level j	630
Δt	Variation of time	631
Φ _{w,a} ^m	Maximum microalgae volume fraction in the culture	632

$\Phi_{w,a}$	Microalgae volume fraction in the culture	633
$\eta_{\text{pump},j}$	Efficiency of pump j	634
$\varphi_{\text{upgrading}}$	Efficiency of the upgrading process	635
μ	Viscosity	636
ρ	Density	637

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