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Ultrafiltration after primary settler to enhance organic carbon valorization: Energy, economic and environmental assessment

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ABSTRACT

A membrane-based alternative consisting of introducing an ultrafiltration step after the primary settler of the Carraixet WWTP (Valencia, Spain) is analyzed. This alternative aim at increasing the organic carbon valorization in the anaerobic digestion and reducing the aeration requirements in the biological reactor. A comparison is made between the current WWTP and the membrane-based alternative based on their energy balances and applying the life cycle costing and life cycle assessment methodologies. Assessment of the membrane-based alternative shows a reduction in the energy demand of 0.053 kWh per kg of COD removed due to a decrease of the aeration needs and a higher methane production in the anaerobic digestion. However, a bypass stream of 25 % to the reactor is needed to guarantee enough COD for a proper nitrogen removal to accomplish with discharge limits. The improvement in the energy balance allowed a reduction in OPEX of €0.013 per kg of COD removed, while the implementation of the membrane system represents CAPEX of €0.105 per kg COD removed. Finally, membrane-based alternative shows a better environmental performance in all the categories evaluated with ReCiPe 2016. Specifically, lower N_2O emissions allowed a reduction of 0.3 kg CO_{2eq} per kg of COD removed.

1. Introduction

The ever-increasing demand for water also implies an increase in the production of wastewater and greater pressure on the treatment systems. It is thus relevant to explore and assess innovative solutions to enhance the sustainability of the whole system by ensuring adequate treatment capacity and improving the economic and environmental performance of wastewater treatment plants (WWTP) [1].

Conventional WWTPs based on aerobic processes (e.g., activated sludge) have been shown to be effective in removing traditional pollutants (organic matter and nutrients). However, the high energy consumption derived mainly from aerating the biological reactor is an important drawback [2]. Improving the efficiency of these facilities necessarily involves an organic carbon redirection to 1) reduce aeration needs and 2) enhance the energy balance.

Boosting anaerobic digestion (AD) in the sludge line over the aerobic biological reactor in the water line appears as a feasible option to meet these objectives. Its main advantages are [3]: i) lower sludge production due to the low anaerobic growth yield; ii) no aeration is required; iii) organic matter can be valorized through methane production; iv) reduction of net CO_2 emissions by means of energy recovery; and iv) increased nitrogen recovery potential in the sludge line. Reducing the organic load treated in the aerobic reactor would also lower the aeration requirements and would diminish N_2O emissions by reducing the nitrification-denitrification process [4]. However, as the capacity of primary settlers to capture particulate organic matter is limited, additional processes should be introduced in the water line to drive all the particulates towards AD. For this purpose, membrane technology has shown a promising performance, since it controlled pore size guarantees a selective suspended fraction retention [5]. Implementing membranes as the primary treatment before the biological reactor, known as direct membrane filtration (DMF), thus appears as a potential option for retrofitting existing WWTPs to maximize the recovery of organic matter and use it in AD.

Initially, most DMF studies focused on using membranes after primary settling to address the role of membrane fouling on the economics

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of the process. Indeed, severe fouling propensities have usually been reported (see, for instance, Kimura et al. [6]; Jin et al. [7]; and Gong et al. [8]), much more so than in other aerobic/anaerobic membrane bioreactors [9,10]. However, this problem can be controlled by applying effective anti-fouling strategies. For example, Abdessemed & Nezzal [11] used ultrafiltration (UF) membranes and coagulation by adding ferric chloride and adsorption with activated carbon, achieving a chemical oxygen demand (COD) effluent concentration below 15 $mg·L^{-1}$ under the best operating conditions. These authors highlighted the improved membrane performance after coagulation in terms of fouling propensity and treatment capacity. The studies by Sethi & Juby [12] are one of the few references to long-term evaluation (>1 year) of microfiltration (MF) technology in a pilot plant in which the primary clarifier supernatant was treated to achieve disinfected solid-free water. These authors showed it was possible to achieve long-term stable filtration without irreversible fouling. More recently, Hube et al. [13] assessed fouling and mitigation mechanisms of cross-flow MF and UF membranes for DMF of the supernatant from the primary settler of a municipal WWTP. The authors reported pore blocking and the formation of a cake layer during filtration in both MF and UF membranes. Although short-term physical cleaning was able to remove the cake-layer from UF membrane, its effectiveness was gradually reduced in MF membranes due to the residual particulates remaining inside the pores. The MF performance was improved by increasing the downtime for physical cleaning (i.e., shortening the length of the filtration stage), but it affected the efficiency of the process. Similar results were reported by Sanchis-Perucho et al. [14], who also concluded that MF membranes are not recommended when directly filtering municipal wastewater (MWW). In this latter study, UF and MF membranes were operated in treating both raw MWW and the primary settler supernatant and it was determined that the MF membrane suffer more severe fouling than UF membranes due to the former's more severe pore blockage propensity. These authors also found that the DMF of MWW is possible in the middle/long-term with UF membranes, achieving operating periods longer than 4 months without serious fouling issues when operating a UF membrane pilot-plant at a moderate transmembrane flux (about 10 LMH) and solids concentrations in the membrane tank around 6–11 $g·L^{-1}$ [14].

Concerning the effect of DMF on AD, some studies have reported the benefits of valorizing DMF concentrate over waste sludge. By way of example, Kimura et al. [15] studied a DMF process (MF membranes; 0.1-μm pore size) followed by an AD system treating raw urban wastewater. These authors reported a production of 390 mL $CH_4·g^{-1}$VS, which was higher than the methane produced from waste sludge in the wastewater facility (230 mL $CH_4·g^{-1}$VS). Moreover, energy consumption was significantly lower than the potential energy recovery of the DMF system, 0.013 and 0.15–0.30 $kWh·m^{-3}$, respectively. Sugiyama et al. [16] pretreated the influent wastewater with a moving bed biofilm reactor (MBBR) to reduce fouling and then applied MF and UF in parallel to conduct DMF. Results of methane production in AD were 372 ± 19 and 391 ± 10 mL $CH_4·g^{-1}$VS, respectively, similar to those obtained by Kimura et al. [15]. Energy surplus was also reported for the combination of MBBR+DMF + AD, resulting in 0.139 $kWh·m^{-3}$. On the other hand, Gong et al. [17] studied the preconcentration of the influent particulate organics with polyvinylidene fluoride hollow fiber membranes (0.02-μm pore size) and its subsequent valorization through AD, obtaining methane productions of 360–367 mL $CH_4·g^{-1}$VS, which is significantly higher than that obtained from excess sludge (i.e., 155–250 mL $CH_4·g^{-1}$VS), but lower than the methane obtained from primary sludge digestion (i.e., 415 mL $CH_4·g^{-1}$VS). The authors attributed the lower methane production from membrane concentrate than from primary sludge to the addition of polyaluminium chloride and powdered activated carbon to improve the performance of the membranes. Hafuka et al. [18] studied the digestibility of the recovered organic matter from DMF (flat sheet ceramic membranes; 0.1-μm pore size) and compared it with waste sludge. Results showed highest methane yields for DMF concentrate, i.e., 0.54 ± 0.05 mL $CH_4·g^{-1}$VS.

Despite the challenge of fouling, this technology is attractive due to its potential to recover resources from MWW, e.g., carbon, nutrients or reclaimed water. Direct filtration systems of raw MWW to concentrate organic matter with UF membranes in mid-term trials (up to one month) have been developed by, among others, Kimura et al. [6] and Lateef et al. [19], who found that 75 % of the organic matter present in raw MWW can be redirected towards energy recovery. Modise et al. [20] studied the performance of seven lab-scale MF membranes to mitigate pollution (organic matter, *E. coli*, Fecal Coliforms, and Enterococcus) from discharges into water bodies from the primary clarifier, highlighting the importance of pore size for microbial retention, and suggesting a maximum membrane cut-off value of 0.2 μm. Delgado Diaz et al. [21] studied the effect of a coagulation/sedimentation process prior to filtration of the primary clarifier supernatant with hollow fiber UF membranes (pore size 0.03 μm), demonstrating a high efficiency in the capture of organic matter (81–95 % in terms of COD. Tuyet et al. [22] applied DMF to raw MWW, obtaining a permeate with COD under the discharge limits (<50 $mg·L^{-1}$). The retentate was co-digested with kitchen waste and produced 64 $kWh·capita^{-1}·d^{-1}$. Nascimento et al. [10] used DMF for urban wastewater by means of UF and obtained a solid-free permeate, which favored its reuse potential.

This work evaluates UF membranes for treating the supernatant from the primary settler of the Carraixet WWTP (Valencia, Spain) to divert all the particulate and colloidal organic matter to the side-stream AD process, while reducing energy requirements for organic matter removal from the mainstream activated sludge (i.e., avoiding energy for organic matter oxidation). The current WWTP treatment scheme and the membrane-based alternative treatment were mathematically modelled and the economic and environmental outcomes of both treatment schemes were compared. This comparison included an energy balance, a life cycle costing (LCC) assessment and an environmental life cycle assessment (LCA).

2. Materials and methods

This study compared the economic and environmental issues in two scenarios: a) Scenario 0 represented the current Carraixet WWTP configuration, and b) Scenario 1 introduced a UF unit after the primary settler. Only the elements that differed in both scenarios were included in the assessment (membranes; membrane tanks; additional blowers and pumps; methane production; aeration requirements; chemical consumption; effluent quality; and sludge production and quality), since the variation of the common equipment and processes was considered negligible.

Both scenarios were assessed from an energetic, economic and environmental perspective. First, the Biological Nutrient Removal Model n° 2 (BNRM2) [23] implemented on DESASS 7.1. software [24] was calibrated by simulation and used to reproduce the current operation of the Carraixet WWTP (Scenario 0). Once the model was dynamically calibrated (i.e., represented the operation of the WWTP), the Scenario 1 simulations were carried out. The results were used to assess the energy, economic and environmental performance.

2.1. Current Carraixet WWTP

Carraixet WWTP is in Alboraya (Valencia, Spain) and serves a population equivalent (p.e.) of 133,000. The water line consists of pretreatment (grids, sieves, sand trap and degreaser), primary settler, activated sludge for biological nitrogen removal and phosphorus precipitation, coagulation-flocculation and disinfection (UV). After treatment, a fraction of the effluent is pumped into the surrounding irrigation channels for reuse on farmland. The sludge line includes a gravity thickener for primary sludge, a dissolved air flotation unit for secondary sludge, AD, and a centrifuge for sludge dewatering. Cogeneration for energy recovery completes the facility, as shown in Fig. 1.

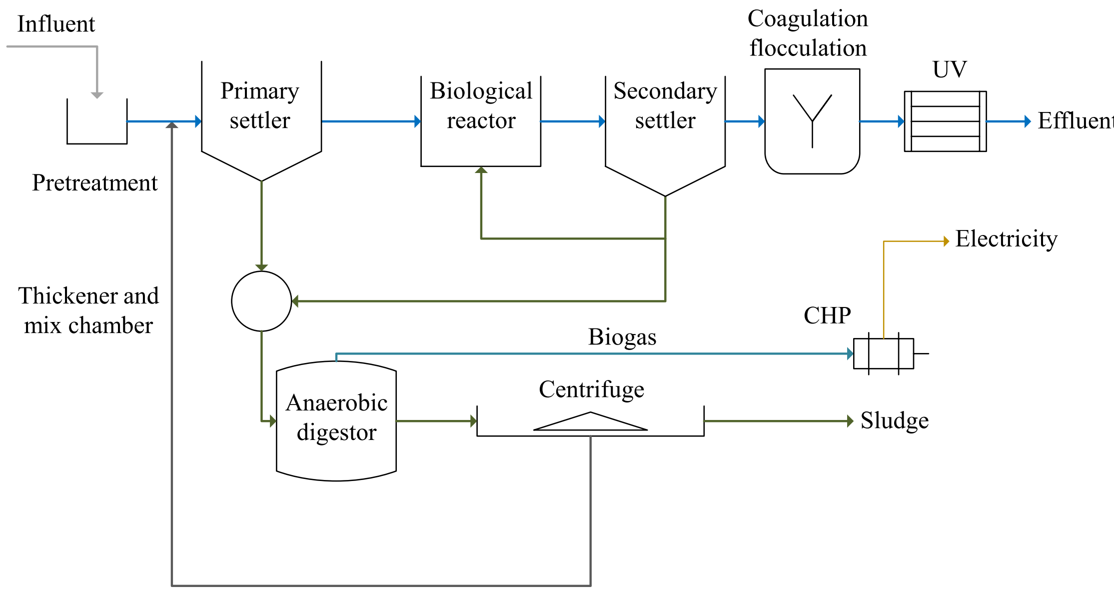

Fig. 1. Scenario 0: current Carraixet WWTP.

2.2. *WWTP membrane-based alternative*

Fig. 2 depicts the alternative layout of Scenario 1, based on including UF after the primary settler. By means of this membrane unit, only dissolved COD (COD_S) is treated in the activated sludge reactor, reducing the aeration needs associated with organic matter oxidation. All the particulate COD (COD_P) and a fraction of soluble COD (colloidal; with a particle size greater than the membrane pore size, i.e., 0.03 μm) is therefore sent to AD. It is important to highlight that Scenario 1 considers the possibility of a bypass stream from the outlet of the primary settler to the biological rector (dotted line) to provide the organic matter requirements for denitrification.

2.3. *Calibration and simulation*

Scenarios 0 and 1 were simulated on DESASS 7.1 software [24], which allows the complete wastewater treatment scheme to be evaluated, both in the water and sludge lines, based on the BNRM2 model [23]. BNRM2 considers a two-step nitrification process based on Jiménez Douglas [25] (for further details of model parameters, see Supplementary Material). As a previous step, a simulation-based calibration was conducted to adjust the model parameters to the WWTP performance.

The average filtration data from the UF pilot plant operated in the MEM4REC project were used for the assessment in Scenario 1. This pilot plant was equipped with a membrane module of 0.03-μm pore size and filtration area of 43.5 m^2 (PULSION® Koch Membrane Systems). Based on the project results, the net 20 °C-standardised transmembrane flux

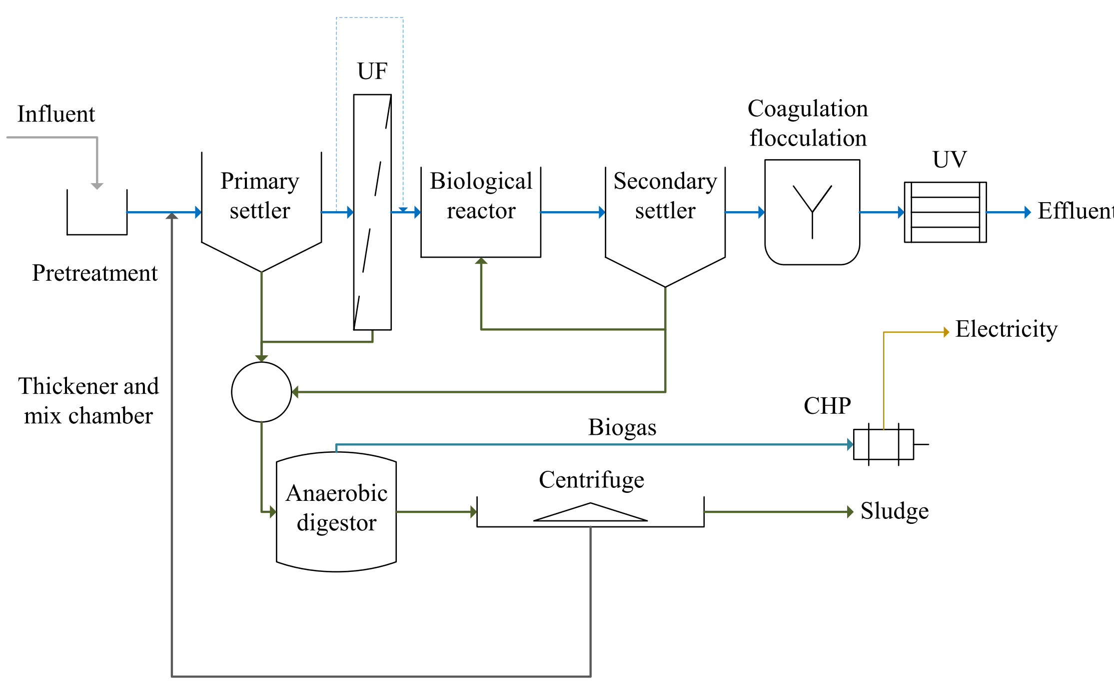

Fig. 2. Scenario 1: membrane-based alternative proposed in the MEM4REC project (bypass in dotted line).

(J_{20}) was set to 10 LMH and the specific ambient air demand per m^2 of membrane (SAD_m) to 0.1 $Nm^3 \cdot h^{-1} \cdot m^{-2}$. Average TMP during filtration of 250 mbar was considered. The membranes were chemically cleaned every 10 months, according to the results obtained by Sanchis-Perucho et al. [26] and Sanchis-Perucho et al. [14].

Table 1 shows the Carraixet influent quality. The simulations and filtration data provided the information required for the treatment, energy and environmental assessment: flows, composition, chemicals consumption, methane production, aeration requirements, etc.

2.4. Energy balance

The energy balance was calculated according to the energy model proposed by Judd & Judd [27] and Pretel et al. [28], included in DESASS 7.1. The energy consumption for aeration of the biological reactor was obtained directly by simulation, while that associated with the membrane scouring blowers, permeate pumps and bypass pumps was calculated by means of equations included in the Supplementary Material.

Energy recovery was estimated from the biogas flow rate, methane percentage in the biogas and methane calorific power. An average power efficiency of 35 % was selected, in accordance with the power efficiencies reported by Darrow et al. [29] for reciprocating engines used as combined heat and power (CHP) systems.

2.5. Life cycle costing

The economic evaluation of both scenarios was carried out by life cycle costing (LCC) to calculate the total annual costs (TAC). A functional unit of 1 kg of removed COD (COD_{rem}) was selected for the economic assessment. Therefore, the TAC (€·kg COD_{rem}^{-1}) of the whole process was calculated by adding the annual investment cost of the CAPEX to the annual OPEX, as shown in Eq. (1):

$$TAC = \frac{r\,(1+r)^t}{(1+r)^t - 1} \cdot CAPEX + OPEX \quad (1)$$

where r is the annual discount rate and t is the depreciation period in years.

LCC included construction and operation phases. The main considerations when calculating CAPEX and OPEX are summarized as follows:

CAPEX

- ➢ CAPEX in Scenario 0: Since Scenario 0 corresponds to the current operation of the treatment plant, its CAPEX is 0 €·kg COD_{rem}^{-1}.
- ➢ Depreciation: A depreciation period of 20 years for building works and 10 years for equipment and membranes was used to calculate the TAC. The annual discount rate was set to 3 % for the former and 5 % for the latter.
- ➢ Membrane tank: The membrane tank volume was estimated according to the dimensions of a commercial membrane unit (PULSION® Koch, 0.03-μm pore size).
- ➢ Pipelines: The maximum velocity of the liquid in pipes was set to 1.5 $m \cdot s^{-1}$ to calculate the pipe diameter, thus preventing excessive losses of mechanical energy.
- ➢ Equipment: Acquisition of blowers, pumps and UF membranes was included. Equipment expenditures were estimated by correlating the capacities (mainly based on flow rates) and unit costs provided by the manufacturers.

OPEX

- ➢ Power requirements: Both energy and power tariffs (Spanish) are included in the appraisals according to Somenergia.coop [30].
- ➢ Energy recovery from methane: Energy production from biogas generated in the AD was considered.
- ➢ Chemical reagents for membrane cleaning: Sodium hypochlorite and citric acid were considered for membranes chemical cleaning. 2000 ppm was adopted as the concentration of both sodium hypochlorite and citric acid, while the contact time was set to 6 h. A frequency of 1.2 chemical cleaning events per year was considered.
- ➢ Membrane replacement: The lifespan of UF membranes was used to calculate the membrane replacement costs. The lifespan of the UF membrane was calculated from the maximum total contact with chlorine permissible before membrane replacement: 500,000 ppm-hours cumulative, according to the supplier. Maximum lifespan is 12 years in any case.
- ➢ Equipment replacement and maintenance: The lifetime of blowers and pumps was established based on the total working hours in the manufacturers' recommendations. A maximum lifespan of 10 years was considered.
- ➢ Building works and construction maintenance: Maintenance of works and buildings were considered as a percentage of the associated CAPEX, according to the technical specifications for general WWTP constructions [31].
- ➢ Labor costs: labor costs were not included in this LCC since both scenarios were considered to be operated by the same staff.

Table 2 shows the unit costs and correlation functions used to assess the CAPEX and OPEX in the proposed scenarios.

2.6. Life cycle assessment

Environmental performance of both scenarios was calculated by means of a life cycle assessment (LCA) according to ISO 14040:2006. The life cycle inventory (LCI) of individual materials and processes (see Table 3) were compiled using the Ecoinvent 3.4 and Industry data 2.0 databases accessed via SimaPro v9.1.1 (PRé Consultants; The Netherlands). Life cycle impact assessment (LCIA) was based on the ReCiPe 2016 method and hierarchical approach, also accessed via SimaPro v9.1.1. The construction phase for additional infrastructure in

Table 1
Carraixet WWTP influent quality.

	Units	Concentration
Parameter		
COD_T	mg·L^{-1}	512
COD_S	mg·L^{-1}	104
P_T	mg-P·L^{-1}	8.8
N_T	mg-N·L^{-1}	57.6
NH_4^+-N	mg-N·L^{-1}	28.0
SO_4^{3-}-S	mg SO_4-S·L^{-1}	119.2

Table 2
Unit costs and correlation functions used to evaluate capital and operating expenses (CAPEX/OPEX) in both scenarios.

Unit costs and correlation function of capital and operating expenses	
Pipes, (final cost is function of diameter) €·m^{-1}	8–220
Polypropylene, €·m^{-2}	9.6
Ultrafiltration hollow-fiber membrane, €·m^{-2}	Membrane = −2.985·ln A[a] (m^2) + 68.159
Energy (function of time period), €·kWh^{-1}	0.231–0.340
Sodium hypochlorite, €·l^{-1}	0.72
Citric acid, €·kg^{-1}	0.75
Polyelectrolyte, €·kg^{-1}	2.35
Wasted sludge for farming, €·t^{-1}	4.81
Blower, €·unit^{-1}	Blower = 2152.4·ln Q[b] (m^3/h) -7849.6
Pumps, €·unit^{-1}	Pump = −0.6582·Q^2 [b] (m^3/h) + 263.54·Q[b] (m^3/h) + 1005.4

[a] Membrane surface.
[b] Flow.

Table 3
Materials and processes selected for the Life Cycle Inventory (LCI).

Construction materials, kg·kg COD_{rem}^{-1}	Chromium steel *(1 kg Steel, chromium steel 18/8 {GLO}| market for | APOS, S (Ecoinvent 3))* Cast iron (*1 kg Cast iron {GLO}| market for | APOS, S (Ecoinvent 3)*) Polyester resin (*1 kg Polyester resin, unsaturated {GLO}| market for | APOS, S (Ecoinvent 3)*) Polypropylene (*1 kg Polypropylene, granulate {GLO}| market for | APOS, S (Ecoinvent 3)*) PVC *(1 kg PVC pipe E (Industry data 2.0))* Transport *(1 tkm Transport, freight, lorry 16–32 metric ton, EURO6 {GLO}| market for | APOS, S (Ecoinvent 3)*	
Energy consumption, kWh·kg COD_{rem}^{-1}, *(1 kWh Electricity, medium voltage {ES}| market for | APOS, S (Ecoinvent 3))*		
Energy avoided (energy recovery from methane), kWh·kg COD_{rem}^{-1} *(1 kWh Electricity, medium voltage {ES}| market for | APOS, S (Ecoinvent 3))*		
Reagent consumption, kg·kg COD_{rem}^{-1}	Polyelectrolyte *(1 kg Polyacrylamide {GLO}| market for | APOS, S (Ecoinvent 3))* NaOCl *(1 kg Sodium hypochlorite, without water, in 15 % solution state {GLO}| market for | APOS, S (Ecoinvent 3))* Citric acid *(1 kg Citric acid {GLO}| market for | APOS, S (Ecoinvent 3))* Transport *(1 tkm Transport, freight, lorry 3.5–7.5 metric ton, EURO6 {GLO}| market for | APOS, S (Ecoinvent 3))*	
Discharge to water, kg·kg COD_{rem}^{-1}	Ammonium NT *(1 kg emissions NT to water)* Total phosphorous, PT*(1 kg emissions PT to water)*	
Discharge to soil, kg·kg COD_{rem}^{-1}	Disposal to agriculture *(1 kg Solid manure loading and spreading, by hydraulic loader and spreader {GLO}| market for | APOS, S (Ecoinvent 3))*	NH_4^+ to soil *(1 kg emissions NH_4^+ to soil through sludge spreading)* PO_4^{-3} to soil *(1 kg emissions PO_4^{-3} to soil through sludge spreading)* N-based fertilizer (*1 kg Nitrogen fertilizer, as N {GLO}| market for | APOS, S)* P-based fertilizer (*1 kg Phosphate fertilizer, as P2O5 {GLO}| market for | APOS, S (Ecoinvent 3))* N_2O to air *(1 kg emissions N_2O to air through sludge spreading)*
	Disposal to landfill *(1 kg Municipal solid waste {RoW}| treatment of, sanitary landfill | APOS, S (Ecoinvent 3))* Transport *(1 tkm Transport, freight, lorry 3.5–7.5 metric ton, EURO6 {GLO}| market for | APOS, S (Ecoinvent 3))*	

Scenario 1 and operation phase for both scenarios was considered. The dismantling phase was not included as it was considered negligible, according to Foley et al. [32]. In the operating phase, the following aspects were included in the LCA: i) biological and filtration performance, ii) sludge handling and disposal, iii) energy consumption, iv) chemicals consumption, v) energy use. As in LCC (see Section 2.5), a functional unit of 1 kg of COD_{rem} and a 20-year lifespan were set. The selected environmental categories were Global Warming Potential (GWP), Freshwater Eutrophication (FWE), Marine Eutrophication (ME) and Fossil Resource Scarcity (FRS). The results of other impact categories included in ReCiPe 2016 can be found in the Supplementary Material.

3. Results and discussion

The following sections give the results obtained from the simulation of Scenarios 0 and 1, as well as the comparison of the energy balance, LCC and LCA between the current wastewater treatment facility and the alternative proposed in the MEM4REC project for the Carraixet WWTP.

3.1. WWTP performance simulation

The simulation started with the calibration of the BNRM2 model to minimize the difference of the Carraixet experimental data performance and the model's outcome. Once the model had been calibrated, different operating alternatives were simulated with UF membranes after the primary clarifier. All the simulations were conducted at 20 °C.

3.1.1. Model calibration and simulation of scenario 0

The BNRM2 model was calibrated by the simulation-based calibration method. The model parameters were adjusted to minimize any differences between the real Carraixet WWTP performance and the simulation outcome. SRT was maintained at 7 days during the calibration, according to the average value maintained in the Carraixet activated sludge reactor.

Table 4 shows a comparison of the experimental values for different stream flows and the corresponding simulation after the model calibration. The agreement between the simulated results and the real plant data shows that the calibrated model could adequately reproduce the WWTP performance (further information on the model parameters are available in the Supplementary Material).

Finally, results also show that Carraixet WWTP would accomplish with the discharge criteria established in the DIRECTIVE (91/271/EEC) [33] for facilities with treatment capacities of >100,000 p.e. discharging to sensitive zone (<10 mg N·L^{-1} and < 1 mg P·L^{-1}).

3.1.2. Simulation of scenario 1

As already mentioned, an alternative configuration of the treatment plant (Scenario 1) was simulated to study the effect of incorporating the membranes in the primary treatment on the WWTP performance, as well as to compare this alternative with the current plant from an energy, economic and environmental point of view. For this, the objective of the simulations was to obtain a similar effluent quality as in Scenario 0 in terms of total nitrogen. This pollutant was the most complex for its removal from the studied system because of the low COD:N. Phosphorus was removed by chemical precipitation.

Table 5 shows the considered qualities of the primary settler effluent and UF permeate. The characteristics of the latter have been estimated considering the results obtained in the pilot plant with the operation

Table 4
Results of model calibration and simulation in Scenario 0 (current Carraixet WWTP).

	Unit	Carraixet WWTP	Simulation
Effluent			
COD_T	mg·L^{-1}	62 ± 5	53
COD_S	mg·L-1	–	44
NO_2^--N	mg N·L^{-1}	–	0.0
NO_3^--N	mg N·L^{-1}	–	7.3
NH_4^+-N	mg N·L^{-1}	1.4 ± 0.2	1.1
PO_4-P	mg P·L^{-1}	1.2 ± 0.2	0.8
N_T	mg N·L^{-1}	9.7 ± 0.4	9.3
P_T	mg P·L^{-1}	0.9 ± 0.2	0.9
Energy			
Aeration	kWh·d^{-1}	–	5860
Biogas			
Q_{biogas}	m^3·d^{-1}	–	4620
CH_4	%	–	61.4
Sludge			
M_{sludge}	kg·d^{-1}	–	8426
Chemicals			
$FeCl_3$	kg·d^{-1}	–	2200
Centrate			
COD_T	mg·L^{-1}	842.6 ± 25.3	857
X_{TSS}	mg·L^{-1}	555 ± 87	576
NH_4^+-N	mg N·L^{-1}	741.5 ± 35.1	781
N_T	mg N·L^{-1}	750.7 ± 37.2	798

Table 5
Primary settler effluent and UF permeate qualities.

Effluent	Primary settler effluent mg·L^{-1}	UF permeate mg·L^{-1}
COD_T	222	60
COD_S	104	60
DBO_5	95	27
X_{SST}	100	0
NH_4^+-N	28.0	27.8
PO_4-P	5.8	3.2
N_T	37.8	30.3
P_T	6.7	3.1

conditions mentioned in Section 2.3. These results can be found elsewhere [14,26].

UF membrane allowed retaining all suspended solids (X_{TSS}) and higher amounts of organic matter than primary settler, up to 73 % more. Indeed, particulate COD was totally retained and a higher fraction of COD_S (i.e., COD passing a 0.45-µm mess filter) was also derived to the AD. This is due to the pore size of the membrane (0.03 µm), which can prevent the passage of colloidal organic matter that would be considered soluble in a conventional suspended solids tests using 0.45-µm filters.

Table 6 summarizes the simulation outcomes and includes the values of the simulated facility operation and stream flows. The first subset of simulations is shown under the title of *Scenario 1a*, which comprises four sub-scenarios: a1, a2, a3 and a4. This set of simulations varied SRT (7 and 13 d) and dissolved oxygen (DO) (2.0, 1.0 and 0.5 mg O_2·L^{-1}) in the biological reactor. Sub-scenario 1-a1 was simulated with an SRT of 7 d and 2.0 mg·L^{-1} of DO in the reactor (the same operating conditions as in Scenario 0). In these conditions, the combined metabolism of the ammonia oxidizing bacteria (X_{AOB}) and nitrite oxidizing bacteria (X_{NOB}), both with significant populations in the reactor (61 and 35 mg COD·L^{-1}, respectively), raised the nitrate concentration in the effluent up to 21.8 mg N·L^{-1}. The lack of organic matter prevented Sub-scenario 1-a1 from denitrifying to remove enough nitrogen and, therefore, the total nitrogen in the effluent (22.9 mg N·L^{-1}) clearly exceeded the value obtained in Scenario 0 (9.3 mg N·L^{-1}).

Many factors have been identified in the literature that selectively inhibit or limit NOB growth, leading to nitritation/anammox or nitritation/denitrification. These factors include high temperature, high ammonium/free ammonia (FA) concentration, high nitrite/free nitrous acid (FNA) concentration, and low DO concentration [34]. The effect of temperature on suppressing NOB growth is generally more a disturbance than a real control parameter in most full-scale processes, especially when this partial nitrification is attempted in the water line of a WWTP. On the other hand, in low concentrated streams, such as urban wastewater, it is difficult to have strong FA and FNA inhibition of the NOB population without having high levels of ammonium and nitrite. Intermittent aeration with low DO is also proposed to control NOB growth through optimized conditions. During the MEM4REC project (data not shown), different control strategies have been tested to maintain the nitritation and it was observed that a proper DO control at low levels (around 0.5 mg·L^{-1}) and an influent feed control strategy to avoid the whole ammonium consumption in the reactor, were the key aspects to avoid the NOB growth.

Since Sub-scenario 1-a1 was unable to remove nitrogen due to the lack of COD, a new simulation was performed by reducing DO to 1.0 mg·L^{-1} (Sub-scenario 1-a2), with the aim of preventing the growth of X_{NOB}, much more dependent on oxygen [35], and boosting denitrification via nitrite, which requires less COD than the nitrate pathway. The simulation results showed that the X_{NOB} were indeed washed out, while the X_{AOB} were maintained. This was also reflected in the concentrations of nitrite (15.8 mg N·L^{-1}) and nitrate (1.3 mg·L^{-1}) in the effluent. It was also verified that this strategy reduced the total nitrogen in the effluent (17.8 mg N·L^{-1}), compared to Sub-scenario 1-a1, but by only 5.1 mg N·L^{-1}, this concentration still being much higher than in Scenario 0. SRT was then raised from 7 to 13 d to determine whether this increase improved denitrification (Sub-scenario 1-a3). However, the 13-d SRT

Table 6
Simulation results of scenario 1.

Operation	Unit	Scenario 0	Scenario 1a (without bypass)				Scenario 1b (with bypass)			
			a1	a2	a3	a4	b1	b2	b3	b4
SRT	d	7	7	7	13	13	7	13	13	13
DO	mg·L^{-1}	2	2.0	1.0	1.0	0.5	2.0	0.7	0.7	0.7
Bypass	%	0	0	0	0	0	25	15	20	25
Reactor										
X_{SST}	mg·L^{-1}	3734	1023	985	1756	1703	1803	2576	2838	3079
X_{AOB}	mg·L^{-1}	51	61	61	74	72	60	68	68	68
X_{NOB}	mg·L^{-1}	24	35	0	46	0	32	0	0	0
Effluent										
COD_T	mg·L^{-1}	53	24	23	24	23	27	26	27	28
COD_S	mg·L-1	44	20	20	20	20	20	20	20	20
NO_2^--N	mg N·L^{-1}	0	0.7	15.8	0.8	12.7	1.7	9.5	8.3	7.1
NO_3^--N	mg N·L^{-1}	7.3	21.8	1.3	20.0	1.2	17.1	0.8	0.7	0.6
NH_4^+-N	mg N·L^{-1}	1.1	0.1	0.4	0.1	1.1	0.1	0.6	0.6	0.7
N_T	mg N·L^{-1}	9.3	22.9	17.8	21.2	15.3	19.4	11.4	10.1	8.9
P_T	mg P·L^{-1}	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Energy										
Aeration	kWh·d^{-1}	4093	2821	1789	2523	1725	2996	2137	2207	2281
Biogas										
Q_{biogas}	m^3·d^{-1}	4382	6996	6079	6550	6501	6165	6218	6089	6282
CH_4	%	62.2	57.9	64.9	61.3	61.7	58.8	60.9	60.9	57.8
Q_{CH4}	m^3·d^{-1}	2726	4051	3945	4015	4011	3625	3787	3708	3630
Sludge										
M_{sludge}	kg·d^{-1}	8426	8496	8445	8355	8508	8643	8322	8295	8274

led to the appearance of X_{NOB} and therefore a higher effluent nitrate concentration (20.0 mg N·L^{-1}), which contributed to raising the concentration of total nitrogen in the discharge flow to 21.2 mg·L^{-1}. Sub-scenarios 1-a1 to a3 indicated that an increase in SRT allowed for increased denitrification. However, it also favored the growth of $X_{NOB,}$ whose final effect was to increase the concentration of total nitrogen in the effluent due to an organic matter deficit to achieve sufficient denitrification via nitrate. A simulation with high SRT and low OD was thus proposed to check whether the wash out of X_{NOB} occurred, so that denitrification could be promoted via nitrite, with a lower COD requirement than the nitrate route. This final simulation (Sub-scenario 1-a4) was carried out with an SRT of 13 d and reducing the DO concentration to 0.5 mg·L^{-1}. These operating conditions effectively removed X_{NOB} and reduced total nitrogen concentration (15.3 mg N·L^{-1}) in the discharge flow. However, the low organic load available for denitrification still kept nitrogen well above the Scenario 0 values.

It should be noted that in the sub-scenarios 1-a1 to a4 effluent COD_T (23–24 mg·L^{-1}) and COD_S (20 mg·L^{-1}) are lower than those obtained in Scenario 0 (53 and 44 mg·L^{-1}, respectively; see Table 4). This is because, while the inert soluble organic matter passed through the primary clarifier without changes in its concentration, some of it was effectively retained by the membrane, thanks to its pore size. The effluent in Scenario 1a thus presented a lower concentration for this inert organic matter, while its total and soluble quantity were reduced. There was also a reduction in the total suspended solids concentration (X_{TSS}) in the reactor in Scenario 1a (985–1756 mg·L^{-1}) compared to Scenario 0 (3734 mg·L^{-1}) as a result of avoiding the solids input from the primary treatment and the reduced available organic matter, due to the increase in the COD diverted to the sludge line, showing that under these conditions the reactor would be over dimensioned. In turn, the diverted COD was responsible for the higher biogas production (from 4620 m^3·d^{-1} in Scenario 0 to 6079–6996 m^3·d^{-1} in Scenario 1a) and reduced aeration requirements, as shown by the energy needs in Scenario 0 (5860 kWh·d^{-1}) and Scenario 1a (1725–2821 kWh·d^{-1}).

Since Scenario 1a could not reach a similar quality than Scenario 0 due to lack of organic matter, a new set of simulations was planned to include a bypass stream from the primary treatment effluent to the reactor to increase the available organic matter. Examples of this set are identified in Table 6 as *Scenario 1b* and include 4 sub-scenarios: b1, b2, b3 and b4.

Initially, SRT and DO were set to 7 d and 2 mg O_2·L^{-1}, respectively, as the facility operates as in Scenario 0, and the % bypass was adjusted to achieve the effluent quality in its current operation. Sub-scenario 1-b1 shows the results obtained at a 25 % bypass. Compared to sub-scenario 1-a1 with similar SRT and OD, this additional organic load boosted denitrification and reduced total nitrogen in the effluent by up to 17.1 mg N·L^{-1}. However, biogas production was reduced to 6165 m^3·d^{-1}, which was one of the objectives of the membrane-based alternative: re-using organic matter and recovering energy.

Based on the previous results, it was decided to simulate with higher SRT and lower DO to favor X_{NOB} wash-out for different bypass flows, specifically 15, 20 and 25 % of the influent flow. Sub-scenario 1-b2 increased SRT to 13 d, reduced OD to 0.7 mg·L^{-1} and lowered the bypass to 15 %. In this case, the nitrite pathway was enhanced compared to Sub-scenario 1-b1, N_T could be reduced to 11.4 mg N·L^{-1} and biogas production was increased up to 6218 m^3·d^{-1}. Finally, to meet N_T concentration in the effluent of Scenario 0, the bypass percentage was increased first to 20 % (Sub-scenario 1-b3) and then to 25 % (Sub-scenario 1-b4). In this latter simulation, as the result obtained for N_T was equal to that in Scenario 0 (8.9 mg N·L^{-1} both), it was considered that the performance of the two treatment schemes were similar and allowed their comparison. The following sections deal with the comparison of the energy balance, LCC and LCA between Scenario 0 and Sub-scenario 1-b4.

3.2. Energy balance

The results of the energy balance are shown in Table 7, which includes Scenario 0, Sub-scenario 1-b4, and Sub-scenario 1-a1 (UF membranes without bypass). As explained in Section 2, energy assessment only includes the elements that differ in scenarios. In Scenario 0 only the energy consumption of biological reactor aeration and energy recovery were calculated. In this case, the aeration requirements reached 0.223 kWh·kg COD_{rem}^{-1}, or 100 % of the total energy consumption considered. In Sub-scenario 1-b4, the new equipment for UF added extra energy expenses associated with: permeate pumps, blowers for membrane scouring and pumps to propel the bypass stream, where applicable. Of this equipment, in Sub-scenario 1-b4 the highest energy consumption was associated with the blowers (0.145 kWh·kg COD_{rem}^{-1}), which required 51.6 % of the energy consumption. This result is in line with other studies with UF membranes [36,37] which pointed out the significance of the blower energy expenditure for membrane scouring. Permeate pumps, with an energy consumptions of 0.015 kWh·kg COD_{rem}^{-1}, accounted for 5.2 % of the energy, while that of the bypass pumps was negligible (0.004 kWh·kg COD_{rem}^{-1}; 1.4 %). On the other hand, the oxygen requirements and their associated energy consumption for organic matter oxidation and nitrification in the aerobic reactor was reduced to 47.1 % of Scenario 0 (from 0.223 kWh·kg COD_{rem}^{-1} to 0.118 kWh·COD_{rem}^{-1}). As a result, energy consumptions in Scenario 0 and Sub-scenario 1-b4 account for 0.223 kWh·kg COD_{rem}^{-1} and 0.282 kWh·kg COD_{rem}^{-1}, respectively, with the membrane scouring requirements being responsible for the higher energy consumption.

With regard to energy recovery, Scenario 0 produced 0.550 kWh·kg COD_{rem}^{-1} while Sub-scenario 1-b4 increased energy recovery by up to 0.662 kWh·kg COD_{rem}^{-1}, or 20.4 % more, taking advantage of the additional organic matter available obtained through UF. As the net energy demand in Scenario 0 was −0.328 kWh·kg COD_{rem}^{-1} and −0.381 kWh·kg COD_{rem}^{-1} in Sub-scenario 1-b4, the alternative membrane-based scheme provided energy savings of 0.053 kWh·kg COD_{rem}^{-1}.

Finally, Sub-scenario 1-a1 shows the best energy performance, although the effluent does not meet discharge requirements. Since all the effluent from the primary settler is filtered by the membranes in this case, the greatest amount of organic matter derived to the AD is reached, achieving the highest energy recovery from biogas valorization (−0.745 kWh·kg COD_{rem}^{-1}). Furthermore, since a smaller amount of organic matter reaches the aerobic reactor, low energy needs for aeration are reached (0.043 kWh·kg COD_{rem}^{-1}). These two savings are capable of largely offsetting the greater energy consumption for permeate pumping (0.019 0.043 kWh·kg COD_{rem}^{-1}) and membrane sparging (0.194 kWh·kg COD_{rem}^{-1}). Therefore, locations without nutrients discharge restrictions are more favorable for the implementation of UF as primary treatment.

3.3. Life cycle costing

The main OPEX, CAPEX and TAC results are shown in Fig. 3. OPEX in Scenario 0 accounted for €-0.137 kg COD_{rem}^{-1}, while Sub-scenario 1-b4 showed similar values for operating expenditure: €-0.150 kg COD_{rem}^{-1}. This represents savings of €0.013 kg COD_{rem}^{-1} compared to Scenario 0. In Scenario 0, the main contributors to OPEX were reactor aeration (€0.116 kg COD_{rem}^{-1}) followed by chemical phosphorus precipitation (€0.024 kg COD_{rem}^{-1}), sludge conditioning (€0.006 kg COD_{rem}^{-1}) and sludge management (€0.002 kg COD_{rem}^{-1}). Power costs (€0.001 kg COD_{rem}^{-1}) contributed to a lesser extent and energy recovery allowed savings of €0.287 kg COD_{rem}^{-1}. On the other hand, membrane filtration in Sub-scenario 1-b4 appeared with two significant associated costs: €0.076 kg COD_{rem}^{-1} for membrane scouring and €0.025 kg COD_{rem}^{-1} for UF membrane replacement during the considered facility lifespan. Sub-scenario 1-b4 also had reactor aeration as a significant expenditure but it was reduced to €0.061 kg COD_{rem}^{-1} as COD and nitrogen removal in this unit was lowered to boost AD. Chemical precipitation was reduced to €0.011 kg COD_{rem}^{-1} (54 % lower than Scenario 0), but this saving was

Table 7
Comparison of energy consumption (EC), energy recovery (ER) and net energy demand (NED).

Scenario	Permeate pumping	Membrane scouring	Bypass pumping	Reactor aeration	Energy recovery	Energy consumption	Net energy demand
	kWh·kg COD_{rem}^{-1}						
Scenario 0	0.000	0.000	0.000	0.223	−0.550	0.223	−0.328
Sub-scenario 1-b4	0.015	0.145	0.004	0.118	−0.662	0.282	−0.381
Sub-scenario 1-a1	0.019	0.194	0.000	0.043	−0.745	0.256	−0.489

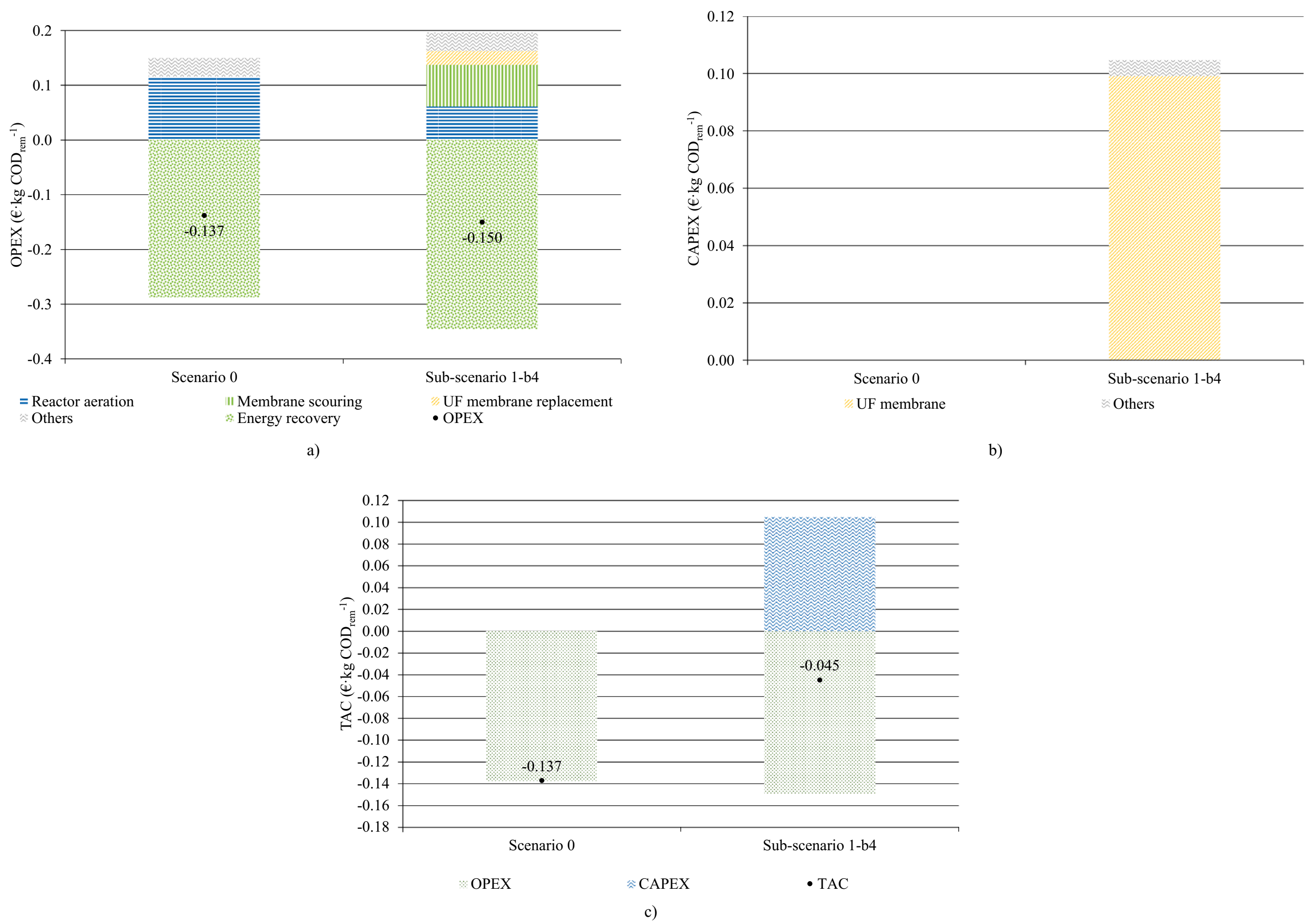

Fig. 3. Results of LCC for Scenario 0 and Sub-scenario 1-b4: a) OPEX; b) CAPEX; c) TAC.

almost counterbalanced by the inclusion of permeate pumping requirements (€0.008 kg COD_{rem}^{-1}). The costs associated with sludge conditioning and management did not change (€0.006 and €0.002 kg COD_{rem}^{-1}, respectively), since similar sludge production occurred in both scenarios. Scenario 0 produced 8426 kg·d^{-1} while sub-scenario 1-b4 produced 8274 kg·d^{-1}, a difference of 1.8 %. Maintenance of blowers and pumps, power cost (€0.001 kg COD_{rem}^{-1} each) and bypass pumping (€0.002 kg COD_{rem}^{-1}) contributed with low expenditure, while maintenance of pipes and membrane tanks showed negligible costs.

Regarding the overall energy cost of equipment operation (blowers and pumps), the result in Scenario 0 was €0.170 kg COD_{rem}^{-1} and higher in Sub-scenario 1-b4, €0.198 kg COD_{rem}^{-1}, which is consistent with the outcome of energy consumption (see Section 3.2). Considering energy recovery, revenues of €0.287 kg COD_{rem}^{-1} for Scenario 0 and €0.346 kg COD_{rem}^{-1} for Sub-scenario 1-b4 were obtained, and net energy profits of €0.170 kg COD_{rem}^{-1} and €0.198 kg COD_{rem}^{-1} were calculated, respectively. This means that Sub-scenario 1-b4 obtained a surplus of €0.028 kg COD_{rem}^{-1}. It is important to highlight that in this study it has been considered that the sale price of the energy surplus was the same as the purchase price of the electricity tariff.

Regarding CAPEX, only Sub-scenario 1-b4 showed values (€0.105 kg COD_{rem}^{-1}) since Scenario 0 represents a business-as-usual case with no investment. As Fig. 3b shows, UF membranes represent the main cost: €0.099 kg COD_{rem}^{-1} and 94.7 % of CAPEX. Pumps (€0.002 kg COD_{rem}^{-1}), pipes (€0.002 kg COD_{rem}^{-1}) and blowers (€0.001 kg COD_{rem}^{-1}) contribution is much lower and membrane tank construction is negligible.

Finally, Fig. 3c shows the TAC for both scenarios. As can be seen, the TAC in Scenario 0 is €0.092 kg COD_{rem}^{-1} lower than in Sub-scenario 1-b4. Considering that energy balance and OPEX in Sub-scenario 1-b4 are more favorable than Scenario 0, CAPEX shows that membrane cost is the main bottleneck for implementing the MEM4REC membrane-based alternative, since the operating savings do not offset the membrane investment of €0.099 kg COD_{rem}^{-1}, plus a membrane replacement cost of €0.025 kg COD_{rem}^{-1}.

Despite this, the improvement in the performance of the membranes, including their durability [38], together with their expected lower cost in the future in an energy context in which prices will remain high [39], could improve the economic sustainability of the MEM4REC alternative in the medium term over conventional treatment. In this regard, a 25 % increase in the membrane lifespan, going from 12 to 15 years, would mean reducing the TAC from €-0.045 kg COD_{rem}^{-1} to €−0.058 kg COD_{rem}^{-1}. A 20 % cheaper membrane would reduce the TAC to €-0.074 kg COD_{rem}^{-1}. A 10 % increase in the price of energy would mean that the TAC of Scenario 0 would be €-0.146 kg COD_{rem}^{-1} and of €-0.055 kg COD_{rem}^{-1} in case of Sub-scenario 1-b4. Increasing net transmembrane flux ($J_{20\ net}$) a 16 % (from 8.2 LMH to 9.5 LMH) by improving filtration productivity would reduce TAC to €-0.072 kg COD_{rem}^{-1}. This last increase would be justified by considering the use of more effective fouling control strategies during the direct filtration of urban wastewater. In fact, the results obtained during the MEM4REC project only included air scouring and backwashing as fouling controllers [26]. However, other studies have reported far better fouling reductions when applying alternative strategies such as chemical enhanced backwashing, coagulant dosing or membrane vibrations [6,19,40].

Given that all these changes are not mutually exclusive and could happen to a greater or lesser extent simultaneously, a combination, for example, of an increase in membrane durability of 10 % (to 13.2 years), a 20 % decrease in the cost of the membrane, an increase in the cost of energy of 10 %, and an improvement in $J_{20\ net}$ of 16 % (to 9.5 LMH) would allow the TAC of Sub-scenario 1-b4 to be reduced to €-0.111 kg COD_{rem}^{-1}. Under these conditions, the operating cost would allow the return period associated with the cost of the membranes to be 5.5 years, below their durability (13.2 years). In other words, the investment in membranes could be recovered during their lifespan.

The results obtained in this study were conditioned by the fact that the discharge from the treatment plant must achieve a nitrogen reduction compatible with the legislation on sensitive areas (DIRECTIVE (91/271/EEC), C [33]). This implies that part of the organic matter must be consumed in the nitrification-denitrification process, for which this study suggests a bypass flow. However, the results on biogas production and energy consumption obtained in Scenario 1a indicated that the energy consumption associated with biological reactor aeration and biogas production will be much more favorable in scenarios without a nitrogen discharge restriction, or one less stringent than in Scenario 1b. In Sub-scenario 1-a4, which presents effluent nitrogen concentrations close to the limit established in the directive for populations between 10,000 and 100,000 p.e. (i.e., 15 mg·L^{-1}), the calculated energy consumption in the biological reactor was 1725 kWh·d^{-1}, which represents a reduction of 24 % with respect to Sub-scenario 1-b4. Additionally, Sub-scenario 1-a4 presented a biogas production of 6501 m·d^{-1} compared to 6282 m^3·d^{-1} in Sub-scenario 1-b4, or 3.4 % more. The combined effect of reducing energy consumption in the reactor and increasing biogas production will improve the energy balance and in turn will reduce OPEX, thereby improving the competitiveness of the membrane-based alternative.

Finally, the combination of water treatment and water reuse for agricultural purposes also represents an opportunity to improve the feasibility of the proposed membrane-based alternative. On one hand,

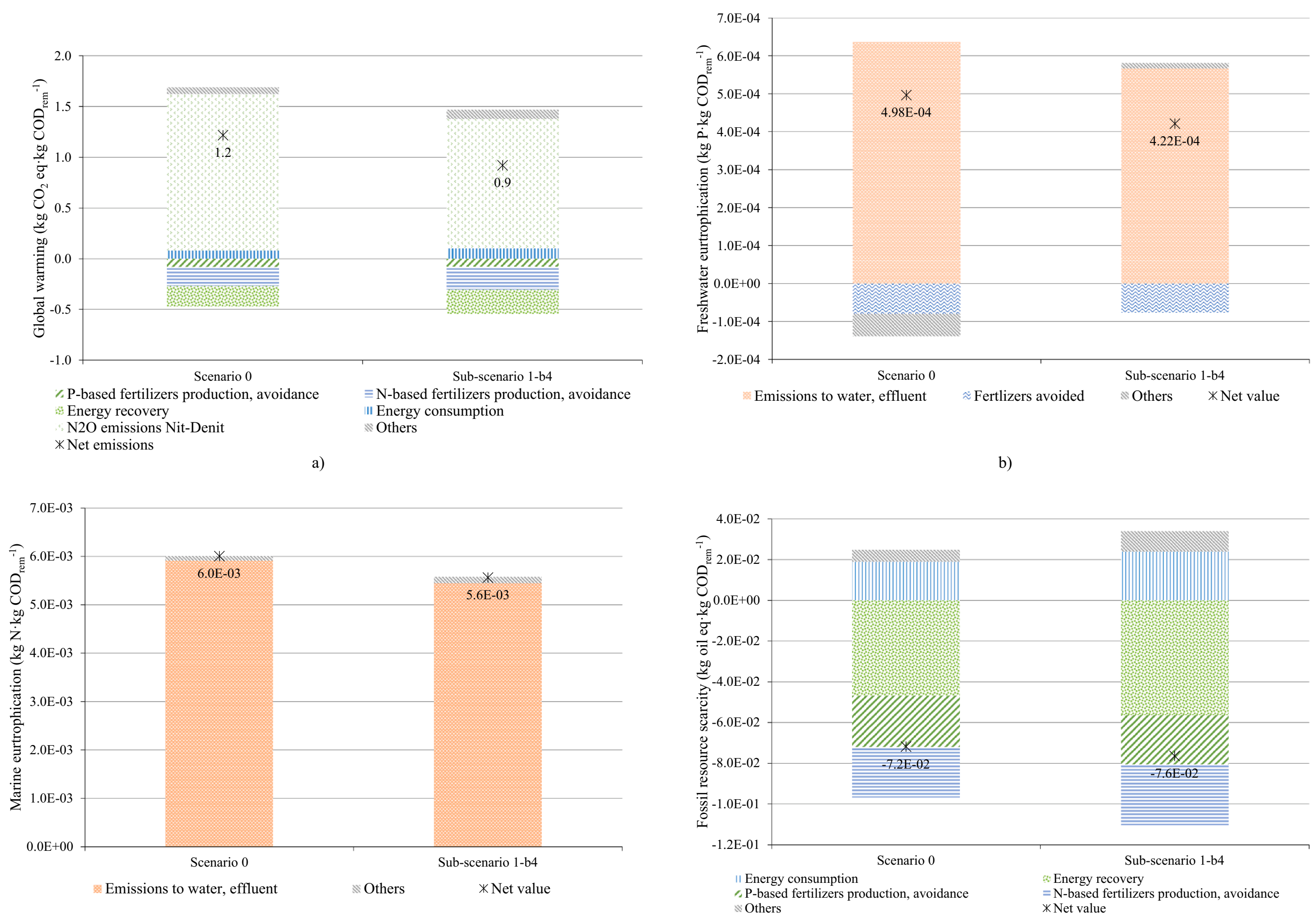

Fig. 4. Results of environmental assessment in both scenarios for a) global warming potential (GWP), b) freshwater eutrophication (FWE); c) marine eutrophication (ME); and d) fossil resource scarcity (FRS).

the cost of the treatment associated with nutrient removal would be reduced, while the nitrogen and phosphorus content of the effluent could be sold as substitutes for commercial fertilizers. In this regard, considering a urea price of €280 per ton and €480 per ton of triple superphosphate, WWTP effluent in Sub-scenario 1-a4 would contain €31,273 and €4905 per year of equivalent N-based and P-based fertilizers, respectively. There is therefore still room for improving the economic feasibility of the proposed membrane-based alternatives, especially in non-sensitive areas and locations with water reuse potential.

3.4. Life cycle assessment

The LCA results are shown in Fig. 4. The selected environmental impact of GWP, FWE and ME are commonly included in life cycle impact assessments of wastewater treatments. In this study, FRS was also considered as an impact category since resource depletion has already become a matter of worldwide concern.

Regarding GWP (Fig. 4a), estimation of N_2O emissions associated with nitrification-denitrification was the main contributor to GWP, accounting for 91 % and 87 % in Scenario 0 and Sub-scenario 1-b4, respectively, while indirect CO_2 emissions related to energy consumption were comparatively much lower (5 % in Scenario 0 and 7 % in Sub-scenario 1-b4). It should be remembered that we did not consider all the energy consumption in the scenarios, but only those that varied between scenarios. This result is in line with that of Daelman et al. [41], who reported 78 % of N_2O contribution to the total carbon footprint for a full-scale WWTP, considering all the contributions to GWP in the facility. It is also in line with that of Rodriguez-Caballero et al. [42], who obtained an average of 61 % for N_2O contribution in the La Roca del Vallès WWTP (Spain). These results highlight the crucial effect of N_2O emissions when assessing the carbon footprint of wastewater treatments, since the characterization factor for nitrous oxide is 298 kg of CO_2 per kg of N_2O. N_2O emissions were calculated as 4 % of the total nitrogen load entering the aerobic reactor, according to Kampschreur et al. [43]. However, the emissions factor reported by Kampschreur et al. [4] for different processes and facilities present a high variation (between 0 % and 95 % at lab-scale and in the range 0–15 % at full-scale) and are influenced by several conditions: nitrite concentration, dissolved oxygen and aeration rate. Accurate results would thus need specific emission data.

Energy recovery and the avoidance of fertilizer production through nutrient recovery by sludge spreading reduced the GWP by 0.269 kg $CO_2 \cdot kg\ COD_{rem}^{-1}$ (Scenario 0) and 0.305 kg $CO_2 \cdot kg\ COD_{rem}^{-1}$ (Sub-scenario 1-b4). Nutrient recovery showed a similar effect to energy recovery on reducing CO_2 emissions. Therefore, boosting the recovery of nitrogen and phosphorus from wastewater, not only in sludge but also in the effluent, and promoting their reuse as substitutes for mineral fertilizers appears to be an effective way of drastically reducing the water treatment carbon footprint. The GWP related to the construction and implementation of the membrane process showed a negligible contribution compared to the operation phase. Overall, Sub-scenario 1-b4 reduced CO_2 emissions more than Scenario 0, due to the reduction of the total nitrogen load entering the aerobic reactor (and therefore the amount of nitrogen that eventually formed N_2O during nitrification-denitrification).

Fig. 4b indicates direct emission of phosphorus to water bodies as the major contributor to FWE. The effluent mass flow of phosphorus in Scenario 0 and Sub-scenario 1-b4 involved an environmental load of $6.360 \cdot 10^{-4}$ kg $P \cdot kg\ COD_{rem}^{-1}$ and $5.668 \cdot 10^{-4}$ kg $P \cdot kg\ COD_{rem}^{-1}$, respectively. This process represented approx. 98 % in both scenarios. Moreover, avoiding mineral fertilizer production reduced net FWE by $8.011 \cdot 10^{-5}$ kg $P \cdot kg\ COD_{rem}^{-1}$ (Scenario 0) and $7.653 \cdot 10^{-5}$ kg $P \cdot kg\ COD_{rem}^{-1}$ (Sub-scenario 1-b4). Emissions to soil through sludge spreading did not add significantly to the environmental impact, while the construction phase effect was negligible.

Nitrogen emissions to water bodies were the main contributor to ME (Fig. 4c) with approx. 97–98 % of the total impact in both scenarios. The effluent of Scenario 0 had a slightly higher nitrogen load discharge (366 $kg \cdot d^{-1}$) than Sub-scenario 1-b4 (355 $kg \cdot d^{-1}$), so that the environmental impact in ME in Sub-scenario 1-b4 ($5.6 \cdot 10^{-3}$ kg $N \cdot kg\ COD_{rem}^{-1}$) was lower than in Scenario 0 ($6.0 \cdot 10^{-3}$ kg $N \cdot kg\ COD_{rem}^{-1}$). The contribution of emissions to soil, fertilizer avoidance through sludge spreading and the construction phase had a negligible effect on ME. The results of the ME and FWE impact categories confirmed nutrient recovery and reuse as crucial processes to improve the environmental performance of wastewater treatment. It would be necessary therefore to properly balance the use of sludge and water reuse for agriculture to avoid affecting water bodies due to leakage and runoff of nutrient compounds.

Fig. 4d shows the possible contribution of wastewater to reduce global resource depletion. The embedded energy and nutrient recovery can contribute to preserving conventional sources and turn wastewater into a supply stream for social and economic purposes. It is important to highlight that urban wastewater is constantly produced, since the domestic usage of water is legally prioritized. WWTPs could act as resource recovery facilities to regularly reintroduce materials (mainly nutrients and water) and energy into the economy. In this study, only the nutrient content in the sludge was considered for reuse, while water reclamation (with its nutrient content) was outside its scope. The environmental performance of the water treatment could thus be expected to improve by taking advantage of these resources, e.g. by means of fertigation of peri-urban crop fields [44].

The results of this study show that the implementation of UF membrane filtration systems in primary treatment can contribute to progress towards the objectives of sustainability and circular economy within the wastewater treatment sector. The incorporation of this technology allows upgrading existing WWTPs without carrying out significant modifications of the treatment scheme. On the other hand, future research and technological development in the field should focus on enhancing filtration productivity to improve DMF competitiveness, and on the treatment of the permeate through processes that reduce or eliminate the needs of bypassing a part of the influent COD to the biological reactor, e.g., SHARON or Anammox processes. On the other hand, alternative technologies such as ion exchange or electrodialysis are presenting promising for maximizing nutrient recovery, while membrane contactors and struvite crystallization have also shown good performances in order to obtain commercial fertilizer products [45,46].

4. Conclusion

A comparison between a current WWT facility (Carraixet WWTP, Valencia) and a membrane-based alternative consisting of implementing an UF module after the primary settler was carried out. The comparison was made based on an energy, economic and environmental assessment.

Process performance of the membrane-based layout was simulated for different OD and SRT in the biological reactor and bypass stream to provide adequate COD for nitrification. Equivalent effluent quality to the current facility was obtained with 0.7 $mg \cdot L^{-1}$ of OD, 13 d of SRT and 25 % of bypass. The energy balance shows a reduction of the net energy demand of 0.053 $kWh \cdot kg\ COD_{rem}^{-1}$ boosted by the reduction in the reactor aeration requirements, which decreased from 0.223 $kWh \cdot kg\ COD_{rem}^{-1}$ to 0.188 $kWh \cdot kg\ COD_{rem}^{-1}$, and the increase in energy recovery from 0.550 $kWh \cdot kg\ COD_{rem}^{-1}$ to 0.662 $kWh \cdot kg\ COD_{rem}^{-1}$.

Membrane-based alternative allowed reducing OPEX by €0.013 kg COD_{rem}^{-1}. The main contributor to direct operating cost was membrane scouring (€0.078 kg COD_{rem}^{-1}) followed membrane replacement (€0.025 kg COD_{rem}^{-1}) and reactor aeration (€0.006 kg COD_{rem}^{-1}), while energy recovery allowed operating savings of €0.035 kg COD_{rem}^{-1}. CAPEX was estimated in €0.105 kg COD_{rem}^{-1} and is dominated by membrane acquisition cost (€0.099 kg COD_{rem}^{-1}; 95 %). Finally, membrane-based TAC is €0.086 kg COD_{rem}^{-1} higher than current treatment cost and membrane cost appears as the main bottleneck for implementing the alternative scheme.

Conversely, LCA outcome demonstrate the environmental benefits of introducing UF membranes as primary treatment. Global warming potential can be reduced up to 0.3 kg CO_{2eq}·kg COD_{rem}^{-1} mainly due to a decrease in N_2O emissions and the improvement of energy recovery. Marine and freshwater eutrophication is also diminished due to the lower N and P discharge with the effluent. And the higher energy recovery contributes to preserve energy resources and hence fossil resource scarcity shows better results.

Therefore, taking advantage of the environmental benefits presented by the implementation of UF membranes as primary treatment necessarily involves improving their economic viability. To this end, some possibilities have been pointed out in this work. The improvement in the performance of the membranes and their cheaper costs, an energy context that favor energy recovery and its sell or considering nutrients and reclaimed water value can contribute to establish a favorable framework to boost this technology for WWTP retrofitting according to sustainable principles.

CRediT authorship contribution statement

A. Jiménez-Benítez: Conceptualization, Data curation, Formal analysis, Investigation, Methodology, Writing – original draft, Writing – review & editing. **P. Sanchís-Perucho:** Data curation, Formal analysis, Investigation, Methodology. **J. Godifredo:** Data curation, Formal analysis. **J. Serralta:** Conceptualization, Funding acquisition, Investigation, Methodology, Project administration, Writing – review & editing. **R. Barat:** Data curation, Formal analysis, Methodology, Writing – review & editing. **A. Robles:** Data curation, Formal analysis, Investigation, Methodology, Writing – review & editing. **A. Seco:** Formal analysis, Funding acquisition, Investigation, Project administration, Writing – review & editing.

Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Data availability

Data will be made available on request.

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Appendix A. Supplementary data

Supplementary data to this article can be found online at https://doi.org/10.1016/j.jwpe.2024.104892.

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