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Life cycle costing of AnMBR technology for urban wastewater treatment: A case study based on a demo-scale AnMBR system

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ABSTRACT

This study aims at assessing the economic performance of a projected full-scale anaerobic membrane bioreactor (AnMBR) for urban wastewater (UWW) treatment at ambient temperature. To this aim, data from an AnMBR demonstration plant (industrial prototype, TRL 6) was used, which was operated for 3 years treating real UWW, allowing gathering a robust set of information for scaling-up to full scale. The obtained results revealed that reactor mixing (0.056–0.124 kWh·$kgCOD_{rem}^{-1}$; 34–57%), and membrane scouring (0.048–0.120 kWh·$kgCOD_{rem}^{-1}$; 22–48%) were the main contributors to the total energy demand; while net energy productions between 0.210 and 0.645 kWh·$kgCOD_{rem}^{-1}$ were achieved. Capital expenditure was highly influenced by UF membranes (€0.029–0.073 $kgCOD_{rem}$ $^{-1}$; 31–49%), combined heat and power technology for energy recovery (€0.012–0.023 $kgCOD_{rem}^{-1}$; 8–24%), and reactor construction (€0.07–0.014 $kgCOD_{rem}^{-1}$; 8–13%); while the main contributors to operating expenditure were energy requirements (€0.042–0.069 $kgCOD_{rem}^{-1}$; 41–46%), membrane replacement (€0.011–0.028 $kgCOD_{rem}^{-1}$; 9–17%), and discharge fee (€0.010–0.020 $kgCOD_{rem}^{-1}$; 9–12%). Total annualized costs showed high variability, between €− 0.003 and 0.188 $kgCOD_{rem}^{-1}$. Results presents AnMBR as a competitive technology for UWW treatment compared to conventional aerobic technologies (e.g., CAS). Membrane fouling control; hydraulic retention time; biogas requirements for reactor mixing and membrane stirring; and energy recovery efficiency were identified as key parameters for improving economic sustainability of AnMBR technology.

1. Introduction

Water management is one of the fields in which new technologies, concepts and approaches are being actively investigated to successfully face the climate change threat, meet Circular Economy (CE) principles and accomplish Sustainable Development Goals [32]. It is therefore crucial to progress towards converting former wastewater treatment plants (WWTP) into full water resource recovery facilities in which urban wastewater (UWW) is regarded as a raw material and no longer as a waste product [13]. This paradigm shift would take advantage of the value embedded in wastewater and boost the economic framework linked to its management.

Among the possibilities for improving the economic performance of wastewater treatment facilities, the combination of anaerobic digestion (AD) and membrane bioreactors (MBR), resulting in the so-called anaerobic membrane bioreactor (AnMBR), is gaining momentum [20, 27,36]. Reduced growth rates of anaerobic microorganisms can be compensated, even when operating at low reactor temperatures, since an AnMBR can decouple the sludge retention time (SRT) from the hydraulic retention time (HRT) by membrane filtration. The additional benefits of AnMBR also include:

Abbreviations: AD, anaerobic digestion; AnMBR, anaerobic membrane bioreactor; CAPEX, capital expenditure; CAS, conventional activated sludge; CE, circular economy; CHP, combined heat and power; COD, chemical oxygen demand; DM, degassing membranes; HRT, hydraulic retention time; J_{20}, 20°C-standardized transmembrane flux; LCC, life cycle costing; MBR, membrane bioreactor; MLTSS, mixed liquor total suspended solids; NED, net energy demand; OLR, organic loading rate; OPEX, operating expenditure; SA, sensitivity analyses; SGD_P, sparging gas demand per m^3 of permeate; SO_4^{2-}-S, sulfate; SRT, sludge retention time; TAC, total annualized cost; TEC, total energy consumption; TER, total energy recovery; UA, uncertainty analyses; UWW, urban wastewater; WWTP, wastewater treatment plant.

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- **Effluent quality:** membrane pore size can ensure microbial quality in the effluent [42], favoring the production of fit-for-purpose reclaimed water.
- **Organic matter valorization:** anaerobic metabolism converts biodegradable organics into methane, which acts as an energy carrier and enhances the energy balance of wastewater treatment [44].
- **Biosolids production and stabilization:** anaerobic metabolism reduces sludge waste management needs due to the low growth yield of anaerobic biomass, while onsite sludge stabilization can be achieved by operating at a high SRT [39].
- **Nutrient recovery:** anaerobic systems conserve most of the mineralized nutrients embedded in the effluent, which favors their recovery [33] or direct reuse through fertigation [19].
- **Reduced footprint:** the MBR configuration eliminates settling units, which reduces the investment costs associated with land for new facilities and favors the retrofitting of existing plants with space constraints [47].

However, AnMBR technology for UWW treatment still faces some operating issues that need to be properly solved. The main bottlenecks for widening the implementation of AnMBR are: sensitivity to process dynamics and development of appropriate control strategies [37]; membrane fouling [16]; and loss of dissolved methane with the effluent [46], which worsen the energy balance and increase the carbon footprint.

Despite the aforementioned challenges, AnMBR has continuously improved its performance, e.g., in terms of effluent quality or maintenance of competitive transmembrane fluxes [34]. However, from an economic perspective, energy demand needs to be further optimized for enhancing its feasibility. Membrane fouling control through gas scouring has been reported as one of the main contributors to energy costs in AnMBR systems [20,29]. Chemical demands for membrane cleaning and membrane replacement also contribute to operating expenses (OPEX) of AnMBR technology [22]. Sludge handling and management is also becoming an emerging global problem [10], so that the reduced sludge production of AnMBRs, jointly with the energy produced through biodegradable organics, can offset the operating costs related to membrane operation and favor the financial framework of AnMBRs against conventional aerobic treatments.

Robust comparisons between different treatment processes and/or operating conditions still represent a complex challenge, especially in the emerging technologies, due to the wide range of assumptions considered, different influent/effluent quality and methodologies applied. In this regard, Ferrer et al. [7] proposed a method of AnMBR design and assessment and obtained a total annual cost (TAC) including CAPEX and OPEX of between €0.101 and €0.097 per m^3 for sulfate-rich UWW and between €0.097 and €0.070 per m^3 for low-sulfate UWW. Using the same method, Pretel et al. [30] calculated a TAC of €0.126–0.169 per m^3 for an AnMBR combined with CAS or MBR-based post-treatment for nutrient removal. These values are in line with those reported by Smith et al. [43] for an AnMBR treating medium- and high-strength wastewater (around €0.124 per m^3). These authors also studied the influence of sludge fate (landfill, incineration, and land application) and concluded that AnMBR costs are lower than high rate activated sludge+AD and CAS+AD when sludge is landfilled, lower than CAS+AD when it is incinerated and lower than AeMBR+AD in any scenario. Xu et al. [48] reported significant differences between AeMBR and AnMBR when water and nutrient reuse is considered: in a discharge scenario, the net present value (NPV) of an AeMBR was $11 million lower than AnMBR, but in a reuse scenario fertilizer savings would allow a NPV $41 million lower for AnMBR. Furthermore, tertiary treatments for disinfection are unlikely to be necessary when ultrafiltration membranes are used, contributing to avoiding or reducing the costs associated with conventional disinfection processes, e.g., UV radiation, chlorination, etc. In this regard, boosting water reuse can also provide revenues for water management systems that could make AnMBR economic performance more attractive.

These results confirm a promising trend, especially when they are compared with the values reported for conventional and aerobic treatments. For example, Iglesias et al. [17] analyzed the costs of full-scale MBRs and extended aeration (EA) facilities in Spain. These authors calculated an OPEX for the studied MBR in the range of €0.20–0.45 per m^3 and an average of €0.22 per m^3 in case of EA. When EA was combined with reclaimed treatments, operating costs rose to €0.31-€0.40 per m^3. Total costs for general MBR technology at more than 50% of its treating capacity was estimated to be €0.20–0.50 per m^3. On the other hand, Xiao et al. [47] reported an OPEX of $0.11–0.20 per m^3 for aerobic MBR in China. EPSAR, the Public Entity for Sewage Treatment in the *Comunitat Valenciana* (Spain), reported €0.33 per m^3 as the average OPEX for all the technologies implemented in this region, based on different aerobic treatments for different treatment capacities [5].

This work aimed to assess the economic performance of a full-scale AnMBR for UWW using Life Cycle Costing (LCC) methodology, which has been widely applied to economic appraisals considering all the costs of constructing, operating, maintaining and disposing of products, processes and services [23]. The results were obtained using the data from an AnMBR demonstration plant treating UWW under different operating conditions within the LIFE MEMORY Project [20,36]. Given the size of the demonstration plant (equipped with commercial ultra-filtration hollow-fiber membranes, industrial pumping and auxiliary elements and control and monitoring devices) and the long operating period (around 3 years with different ambient conditions), data obtained from the plant can be considered as a robust set of information for scaling-up AnMBR to full-scale facilities and favor the adoption of AnMBR and its advantages over conventional UWW treatments.

2. Materials and methods

2.1. AnMBR plant

The demonstration plant comprised a 40 m^3 anaerobic reactor (AnR) (34.4 m^3 working volume) connected to three x 0.8 m^3 membrane tanks (MT-X) (0.7 m^3 working volume), equipped with one ultrafiltration membrane system (PURON® PSH41, 0.03-μm pore size and 41 m^2 total filtration area). The total system filtration area is 123 m^2. The plant also includes a rotofilter (RF, 1.5-mm screen size), an equalization tank (ET, 1.1 m^3), a clean-in-place (CIP) tank (0.37 m^3), and degassing membranes (DM) to recover the methane dissolved in the effluent (further details of DM in Sanchis-Perucho et al. [38]).

For energy calculations, data regarding flow rate, liquid level, pressure (transmembrane, inlet and outlet for blowers, and gas), temperature and methane production was obtained from the available plant instrumentation. Data obtained from the DM unit (transmembrane pressure, flow rates, and methane recovery efficiency) were also used to estimate the dissolved methane recovery potential of the system, according to Sanchis-Perucho et al. [38]. Adiabatic indexes and blower and pump efficiencies were estimated from the standard values for this equipment. Further details of the plant can be found in Robles et al. [34]. The flow diagram and an overview of the AnMBR are provided in e-supplementary materials.

2.2. AnMBR operation

The AnMBR prototype treated effluent from the pre-treatment step (screening and sand removal) of the "Alcázar de San Juan" WWTP. After pre-treatment in the RF and homogenization in the ET, the wastewater was pumped to the AnR. The sludge was continuously recycled through the external membrane tanks (immersed configuration), where the final effluent was obtained by vacuum filtration. A fraction of the biogas produced in the AnR was constantly recirculated from the headspace to the bottom of the reactor through coarse bubble diffusers for stirring purposes. This mixing strategy favored the stripping of dissolved gases from the liquid phase for their collection. Another fraction of the biogas produced was driven to the bottom of the membrane tanks for membrane scouring.

The membrane operating strategy followed a sequence of 300-s basic F-R cycle (280 s filtration and 20 s relaxation), 40 s of back-flush every 5 F-R cycles, 20 s of ventilation every 15 F-R cycles and 40 s of degasification every 50 F-R cycles.

The selected average operating conditions and characterization of the five scenarios evaluated in this study are provided in e-supplementary materials (see Table S1) and further details can also be found in Robles et al. [36] and Jiménez-Benítez et al. [20]. Influent quality was characterized by a high COD (755 ± 224–1403 ± 532 mg·L^{-1}) and COD:SO_4^{2-}-S ratios between 4.80 ± 2.86 and 8.70 ± 3.69 due to dairy- and wine-industry contributions. SRT of the demo-plant was set to 70 days, based on previous studies on AnMBR technology for urban wastewater treatment at ambient temperature. By achieving high concentrations of biomass in the reactor, an appropriate level of biodegradability can be ensured even at low temperatures. This has been validated in previous studies (e.g., Giménez et al. [12]). However, to prevent the excessive concentrations of suspended solids in the mixed liquor that could negatively affect the filtration process [35], HRT was set at values above 20 h. This allowed maintaining mixed liquor solids concentrations compatible with competitive filtration process performances.

Scenarios I to V were used to scale and estimate the performance of a full-scale AnMBR system that would treat the influent to the "Alcázar de San Juan" WWTP (i.e., 16,000 m^3·d^{-1}). From this scaling, both the energy balance and the LCC were evaluated.

Regarding the biological process, an average COD removal efficiency of 87.2 ± 6.1% was achieved while effluent COD concentration remained below its discharge limits according to European regulations (<125 mg·L^{-1}) despite treating medium-high organic load rates (OLR) (0.60 ± 0.20–1.28 ± 0.38 kg COD·m^{-3}·d^{-1}). Biogas production varied widely throughout the studied periods (1359 ± 999–5558 ± 3204 STP L·d^{-1}), but the methane percentage remained high and stable (75 ± 2–77 ± 2%).

The filtration process performance demonstrated that it was possible to operate the membranes at competitive 20ºC-standardized gross transmembrane fluxes ($J_{20\ gross}$), 15–25 L·m^{-2}·h^{-1}, while maintaining low membrane fouling propensities (0.3–4.9 mbar·d^{-1}) by applying specific gas demands per volume of permeate produced (SGD_P) of 10–20 $Nm^3_{biogas}·m^{-3}_{permeate}$ (see Jiménez-Benítez et al. [20]). These SGD_p were slightly lower than those reported for aerobic MBR by Iglesias et al. [17] (18.3–24.32 $Nm^3_{biogas}·m^{-3}_{permeate}$) and to those applied by Pretel et al. [31] to an AnMBR (14–32 $Nm^3_{biogas}·m^{-3}_{permeate}$), both also equipped with hollow fiber membranes.

2.3. Energy assessment

The energy balance was calculated according to the equations proposed in Judd and Judd [21] and Pretel et al. [29] (see e-supplementary materials). The equipment considered for power energy requirements ($W_{elements}$) included rotofilter, pumps (feeding, recycling and permeating), blowers and a sludge dewatering system. Heat energy requirements (Q) were not considered in the study since the full-scale AnMBR was assumed to operate at ambient temperature, as per the demo-scale AnMBR.

Combined Heat and Power (CHP) technology was considered for energy recovery from biogas. Since none of the available CHP systems is suitable for low biogas productions, W_{biogas} was estimated from the biogas flow rate Q_{biogas} (L·d^{-1}), methane percentage in the biogas $\%_{CH4}$, and methane calorific power CV_{CH4} (kJ·m^{-3}). In this case, the widely used reciprocating engines were selected as CHP. These systems can achieve high power and overall CHP efficiencies (24–36% and 66–71%, respectively) according to Darrow et al. [4]. Eq. 1 gives the energy from the biogas recovered in terms of power (W_{biogas} in kW):

$$W_{biogas} = \frac{Q_{biogas} \bullet \%_{CH4} \bullet CV_{CH4} \bullet \%_{power\ efficiency\ CHP}}{1000 \cdot 24 \cdot 3600} \quad (1)$$

An average $\%_{power\ efficiency\ CHP}$ of 35% was selected in this study. Table S1 (in e-supplementary materials) shows the biogas flow rate and the biogas methane percentage for each scenario. The concentration of dissolved methane in the effluent was calculated from Henry's Law. Saturation conditions were considered for all evaluated temperatures (see Giménez et al. [11]), since coarse bubble diffuser stirring favors methane stripping from the liquid phase thus avoiding oversaturation. The percentage of recovered methane was calculated following previous studies carried out in the demonstration plant [38].

2.4. Life cycle costing

The economic feasibility of the projected full-scale AnMBR was assessed by means of the experimental data provided by the prototype through LCC, which was used to calculate the total annual costs (TAC), defined as the sum of the capital and operating expenses (CAPEX and OPEX, respectively) as in Eq. 2:

$$TAC = \frac{r\ (1 + r)^t}{(1 + r)^t - 1} \cdot CAPEX + OPEX \quad (2)$$

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

A functional unit of 1 kg of removed COD ($kgCOD_{rem}$) was selected for the economic assessment, according to the main pollutant target of AnMBR technology. A depreciation period of 20 years for building works (annual discount rate 3%) and 10 years for equipment (annual discount rate 5%) was used to calculate the CAPEX. For OPEX, the energy costs considered both biogas and dissolved methane recovery by DM. Chemicals for membrane cleaning (sodium hypochlorite and citric acid) were included in OPEX appraisals. Since chemical cleaning degrades membranes, the UF membrane lifespan was calculated considering the cleaning frequency requirement obtained during the demo-scale AnMBR operation and the permissible maximum total contact with chlorine before membrane replacement, i.e., 500,000 ppm-hours cumulative, or a maximum of 10 years, according to the supplier. The lifetime of blowers and pumps was based on the total working hours recommended by the manufacturers (50,000 and 75,000 h, respectively) or also set to a maximum of 10 years. When possible, equipment costs were estimated by correlations between capacities (mainly based on flow rates) and the unit costs provided by the manufacturers, while labor costs were not included. LCC was composed of construction and operating costs over 20 years. Dismantling was not included.

Further considerations when calculating CAPEX and OPEX can be found in the e-supplementary materials. Table 1 shows the unit costs and correlation functions used to assess CAPEX and OPEX in the proposed scenarios.

2.5. Sensitivity and uncertainty analyses

Sensitivity (SA) and uncertainty analyses (UA) were carried out to determine the most significant parameters and the related uncertainty in the obtained results. In this regard, for the purpose of SA, the Standardized Regression Coefficients (SRC) method was applied to select the most influential parameters. A threshold of 0.1 (absolute value) for the standardized regression slope (b_i) was chosen to select the significant factors. The UA was conducted by the Monte Carlo method to evaluate the propagation of uncertainty in the results, which were assessed by means of i) the 5th and 95th percentiles and ii) the empirical cumulative distribution function (eCDF). 2000 model runs were performed for both SA and UA. The sampling matrix was generated by the Latin hypercube sampling (LHS) method with a variability of 5% for the membrane and equipment cost quoted by manufacturers and 10% for the rest of

Table 1
Unit costs and correlation functions used to evaluate capital and operating expenses (CAPEX/OPEX) in the scenarios evaluated.

Unit costs and correlation function of capital and operating expenses		
Pipes, (final cost is function of diameter) $€\cdot m^{-1}$ [a]		8–220
Concrete wall/Floor&cover, $€\cdot m^{-3}$		350/130
Steel cover, $€/m^2$ [b]		51.11
Ultrafiltration hollow-fiber membrane, $€\cdot m^{-2}$ [c]		$€\cdot m^{-2} = -2.985\cdot \ln A\ (m^2) + 68.159$
Energy	Energy term $€\cdot kWh^{-1}$ [d]	0.231–0.340
	Power term $€\cdot kW^{-1}\cdot year^{-1}$ [d]	1.57–24.73
Sodium hypochlorite (NaOCl Cl active 5%), $€\cdot L^{-1}$		0.72
Citric acid (Citric acid 1-hidrate), $€\cdot kg^{-1}$		2.96
Polyelectrolyte, $€\cdot kg^{-1}$		2.35
Wasted sludge for farming, € per t		4.81
Wasted sludge for landfill, € per t		30.05
Wasted sludge for incineration, € per t		250
Blower, $€\cdot unit^{-1}$ [e]		$€\cdot unit^{-1} = 970.66\cdot Q\ (Nm^3\cdot h^{-1})^{0.6008}$
Water pumps, $€\cdot unit^{-1}$ [f]		$€\cdot unit^{-1} = -0.6582\cdot Q^2\ (m^3\cdot h^{-1}) + 263.54\cdot Q\ (m^3/h) + 1005.4$
Sludge pumps, $€\cdot unit^{-1}$ [g]		$€\cdot unit^{-1} = 1071.1\cdot \ln Q\ (m^3\cdot h^{-1}) + 19{,}643$
Rotofilter, $€\cdot unit^{-1}$ [h]		$€\cdot unit^{-1} = 13.831\cdot Q\ (m^3\cdot h^{-1}) + 2220.8$
Thick-bubble diffuser with 4" Membrane Disc Aeration Head (Barmatec S.L.), $€\cdot unit^{-1}$		16.94
Dewatering system (centrifuge), $€\cdot unit^{-1}$ [i]		$€\cdot unit^{-1} = 7246.1\cdot \ln Q\ (m^3\cdot h^{-1}) + 93{,}099$
CHP, reciprocating engine	$€\cdot unit^{-1}$ [j]	$€\cdot unit^{-1} = -2\cdot 10^{-9}\cdot kW^3 + 5\cdot 10^{-5}\cdot kW^2 - 0.4842\cdot kW + 2708.9$
	O&M $€\cdot kWh^{-1}$ [j]	$€\cdot kWh^{-1} = 0.0192\cdot e^{-0.0001\cdot kWh}$
PermSelect® Silicone Membrane Module, $€\cdot m^{-2}$		37.10

[a] Frans Bonhomme España
[b] BEDEC (iTeC)
[c] Koch Membrane System
[d] Energy and power tariff 6.1 TD. Entidad Pública de Saneamiento de Aguas Residuales de la Comunidad Valenciana (EPSAR). Public Entity for Sewage Treatment of Comunitat Valenciana
[e] Pedro Gil S.L.; Tebyc S.L.; FPZ
[f] Sugein S.L.; Tebyc S.L.
[g] Bombas Ideal S.A.
[h] Drainfilter S.L.
[i] Hutchison Hayes Separation Inc.
[j] U.S. Department of Energy: Combined Heat and Power Technology-Fact Sheet Series

parameters considered (further details can be found in e-supplementary materials). OPEX, CAPEX and TAC were used as outputs for SA and UA assessment.

3. Results and discussion

3.1. Energy assessment

Total energy consumption (TEC), total energy recovery (TER) and net energy demand (NED) of the projected full-scale AnMBR plant were evaluated for each scenario (see Table 2 and Fig. 1). These results are discussed in the following sections.

3.1.1. Total energy consumption

shows that the main contributors to the TEC in all the scenarios were reactor mixing (34–57%) and membrane scouring (22–48%). Energy inputs related to methane recovery blowers (0.3–0.5%), rotofilter (0.5–0.7%) and waste sludge pumping (0.1%) were negligible. Moderate energy inputs were required for sludge dewatering (5–9%), feed pumping (1–2%), permeate pumping (2–8%) and sludge recycling

Table 2
Total annualized cost (TAC) comparison of different treatment technologies for urban wastewater.

Technology	Scale	Membrane configuration	TAC ($€\cdot kgCOD_{rem}^{-1}$)	Ref
AnMBR	Pilot+simulation	HF	0.14–0.20	Ferrer et al. [7]
ANMBR	Pilot+simulation	HF	0.18–0.43	Pretel et al. [28]
MBR	Pilot+simulation	HF	0.21	Pretel et al. [30]
MBR	Full	HF	1.74–3.33	Bertanza et al. [2]
CAS	Full	-	1.18–2.51	Bertanza et al. [2]
CAS	Pilot+simulation	-	0.14	Pretel et al. [30]
CAS	Full+simulation	-	0.93–0.95	Jafarinejad [18]
SBR	Full+simulation	-	1.07–1.13	Jafarinejad [18]
EAAS	Full+simulation	-	0.95–1.04	Jafarinejad [18]

AnMBR: anaerobic membrane bioreactor; MBR: membrane bioreactor; CAS: conventional activated sludge; SBR: sequential batch reactor; EAAS: extender aeration activated sludge; HF: hollow fiber

(3–4%).

Scenario II required the lowest energy consumption (TEC of 0.166 $kWh\cdot kgCOD_{rem}^{-1}$), mainly due to its comparatively low sludge dewatering, reactor mixing, and membrane scouring requirements (0.009 $kWh\cdot kgCOD_{rem}^{-1}$, 0.056 $kWh\cdot kgCOD_{rem}^{-1}$ and 0.080 $kWh\cdot kgCOD_{rem}^{-1}$, respectively). These results were influenced by the highest COD removal efficiency (equivalent to 20631 $kgCOD\cdot d^{-1}$ at industrial scale) and the lowest HRT (25 ± 1 h) within the operating conditions evaluated in this work. On the other hand, Scenario III required the highest energy consumption (TEC of 0.273 $kWh\cdot kgCOD_{rem}^{-1}$) due to its lower COD removal capacity (12,428 $kgCOD\cdot d^{-1}$ at industrial scale), which increased the energy requirements per kg of COD removed, especially for membrane scouring (0.120 $kWh\cdot kgCOD_{rem}^{-1}$) and reactor mixing (0.097 $kWh\cdot kgCOD_{rem}{}^{-1}$), even though its HRT (26 ± 2 h) was almost the same as that of Scenario II.

The permeate pumping results also highlighted the importance of maintaining not only low fouling rates (FRs), but also low TMP. For example, although Scenarios III and IV operated at low FR (0.3 and 0.5 $mbar\cdot day^{-1}$, respectively), the average TMP in Scenario IV (87 mbar) was significantly lower than in Scenario III (462 mbar), resulting in a reduction of the energy consumption from 0.022 $kWh\cdot kgCOD_{rem}^{-1}$ to 0.005 $kWh\cdot kgCOD_{rem}^{-1}$ (77% less), associated with permeate pumping. It is important to consider that filtration performance is influenced by several inter-related biological, physical and chemical factors (e.g., temperature, shear rate, SRT, HRT, extracellular polymeric substances and soluble microbial products concentrations, operation mode, etc.) [26]. The higher temperature (27 ± 1 °C) and lower mixed liquor total suspended solids (MLTSS) concentration (10.4 ± 0.8 $g\cdot L^{-1}$) in Scenario IV could have led to better sludge filtration properties compared to Scenario III (temperature 19 ± 1°C and 11.3 ± 1.0 $g\cdot L^{-1}$ MLTSS). Additionally, both Scenarios I and IV operated at similar temperature (27 ± 1°C) and lower MLTSS in Scenario I than in Scenario IV (8.4 ± 0.5 and 10.4 ± 0.8 $g\cdot L^{-1}$, respectively), but TMP was significantly higher in Scenario I (318 mbar) than in Scenario IV (87 mbar). Energy consumption for permeate pumping in Scenario I (0.010 $kWh\cdot kgCOD_{rem}^{-1}$) was hence double that of Scenario IV (0.005 $kWh\cdot kgCOD_{rem}^{-1}$) even though COD removal in Scenario IV (10,815 $kgCOD_{rem}\cdot d^{-1}$) was significantly lower than in Scenario I (19,138 $kgCOD_{rem}\cdot d^{-1}$).

Membrane scouring has been classically identified as a key parameter of AnMBR technology [40]. Scenarios I and II performed with

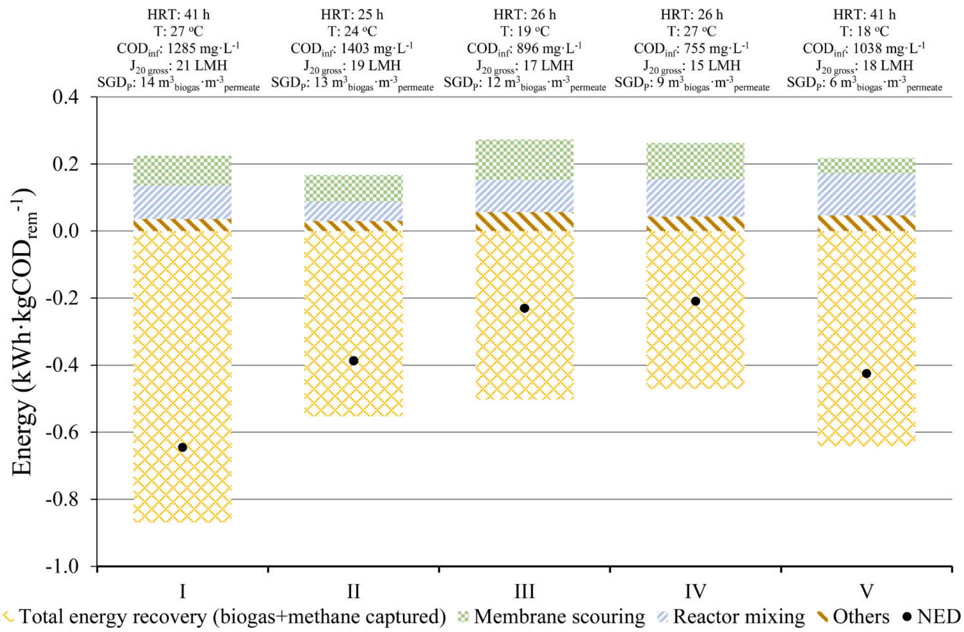


Fig. 1. Total energy consumption (TEC), total energy recovery (TER) and net energy demand (NED) for the projected full-scale AnMBR plant. “Others” refers to: rotofilter; feeding and permeate pumping; sludge waste, recycling and dewatering and methane recovery blowers.

similar high COD removal (19138 and 20631 $kgCOD_{rem}^{-1}\cdot d^{-1}$, respectively) and were operated with similar SGD_p (14 and 13 $Nm^3_{biogas}\cdot m^{-3}_{permeate}$, respectively). Both showed therefore low energy consumption associated with membrane scouring: 0.089 $kWh\cdot kgCOD_{rem}^{-1}$ in Scenario I and 0.080 $kWh\cdot kgCOD_{rem}^{-1}$ in Scenario II. On the other hand, Scenarios II and III were both operated at a similar average SGD_p of 12 and 13 $Nm^3 biogas\cdot m^{-3}_{permeate}$ respectively but had different COD removal performance (20631 and 12428 $kgCOD_{rem}\cdot d^{-1}$, respectively). Therefore, Scenario II showed lower energy consumption regarding membrane scouring (0.080 $kWh\cdot kgCOD_{rem}^{-1}$) than Scenario III (0.120 $kWh\cdot kgCOD_{rem}^{-1}$). Finally, Scenario V was operated with the lowest SGD_p (6 $Nm^3_{biogas}\cdot m^{-3}_{permeate}$) and performed with an intermediate COD removal rate (15376 $kgCOD_{rem}\cdot d^{-1}$). Scenario V therefore had the lowest energy expenditure for membrane scouring (0.048 $kWh\cdot kgCOD_{rem}^{-1}$).

These results confirm that the challenging issue of optimizing energy consumption in AnMBR performance depends on multiple factors. Biological conditions in the reactor should ensure high COD removal, not only for accomplishing the regulations, but also to better reach high energy efficiencies. Optimizing HRT to reduce reactor mixing expenditures and designing a comprehensive filtration strategy able to maintain low TMP are crucial to reducing TEC. This strategy should also consider appropriate combinations of $J_{20\ gross}$ and SGD_p.

3.1.2. Total energy recovery

Table 2 shows the TER when energy is recovered from both: only biogas; and biogas and dissolved methane captured from the effluent. The results confirmed the potential of AnMBR for organic matter valorization via methane production, although the energy production values varied widely. Scenarios I and V showed the highest energy recoveries (−0.827 and −0.581 $kWh\cdot kgCOD_{rem}^{-1}$, respectively), while III (−0.429 $kWh\cdot kgCOD_{rem}^{-1}$) and IV (−0.386 $kWh\cdot kgCOD_{rem}^{-1}$) had lower energy productions. Scenario II showed an intermediate result (−0.513 $kWh\cdot kgCOD_{rem}^{-1}$).

These energy outcomes were influenced by both the influent characteristics and the operating conditions (see Table S1 in e-supplementary materials). In Scenarios I and II, the high $COD:SO_4^{2}$-S ratios (8.7 ± 3.7 and 8.2 ± 4.4) and temperatures (27 ± 1°C and 24 ± 2°C) resulted in organic matter conversions with methane yields (Y^{CH4}) of 169.0 ± 95.1 L $STP\cdot kgCOD_{rem}^{-1}$ (Scenario I) and 109.6 ± 70.0 L $STP\cdot kgCOD_{rem}^{-1}$ (Scenario II). Scenario IV showed a reduced methane yield of 70.2 ± 36.0 L $STP\cdot kgCOD_{rem}^{-1}$, even though it operated at a high temperature (27 ± 1 °C), due to its low $COD:SO_4^{2}$-S ratio (4.8 ± 2.9) and OLR (0.71 ± 0.23 $kgCOD_{in}\cdot m^{-3}\cdot d^{-1}$). The comparison between Scenarios IV and V also depicted the importance of the $COD:SO_4^{2}$-S ratio: both scenarios operated at similar OLR (0.71 ± 0.23 and 0.60 ± 0.20 $kgCOD_{in}\cdot m^{-3}\cdot d^{-1}$, respectively) but the lower $COD:SO_4^{2}$-S ratio in Scenario IV (4.80 ± 2.86) than in Scenario V (6.94 ± 2.83) led the former to reduced methane productions (methane yield of 70.2 ± 36.0 L $STP\cdot kgCOD_{rem}^{-1}$) despite its high operating temperature (27 ± 2°C).

Hence, increasing the $COD:SO_4^{2}$-S ratio or increasing the OLR available for methanogenic organisms appears to be a suitable approach to improve energy recovery. For this, additional co-substrates could be explored, such as adding food waste, as proposed by Moñino et al. [25] or Galib et al. [9]. Operating at high temperatures also favors the conversion of organic matter into methane while reducing its losses in the effluent. Both effects contribute to improving energy recovery.

Recovering dissolved methane increased TER by 0.040–0.086 $kWh\cdot kgCOD_{rem}^{-1}$, thus improving the energy balance while reducing the carbon footprint related to fugitive CH_4 emissions. The energy demands associated with the methane recovery blowers appeared to be negligible. These results confirm that degassing membranes are an attractive technology for enhancing AnMBR performance.

3.1.3. Net energy demand

As Table 2 shows, the best NED performance was achieved in Scenario I (−0.645 $kWh\cdot kgCOD_{rem}^{-1}$). In this case, the relatively high reactor mixing energy costs (0.099 $kWh\cdot kgCOD_{rem}^{-1}$) due to the high HRT (41 ± 13 h) could be offset by the highest TER (−0.870 $kWh\cdot kgCOD_{rem}^{-1}$) driven by its operation at higher ambient temperatures (27 ± 1 °C), OLR (0.75 ± 0.25 $kgCOD\cdot m^{-3}\cdot d^{-1}$) and $COD:SO_4^{2}$-S ratio (8.70 ± 3.69) than the other scenarios. Scenarios II to V also performed as net energy producers with values between − 0.210 $kWh\cdot kgCOD_{rem}^{-1}$ (Scenario IV) and − 0.425 $kWh\cdot kgCOD_{rem}^{-1}$ (Scenario V).

By way of comparison, Hernández-Sancho et al. [15] reported an average NED of 1.684 ± 1.428 $kWh\cdot kgCOD_{rem}^{-1}$ based on data provided by the EPSAR for 177 WWTP in Valencia (Spain), mainly obtained from conventional aerobic-based facilities. More recently, Silva and Rosa [41] evaluated 17 Portuguese WWTPs based on extended aeration and CAS, and found an average of 1.3 kWh·kgCOD$_{rem}^{-1}$ and a 25–75 percentile range between 0.5 and 1.8 kWh·kgCOD$_{rem}^{-1}$. Gurung et al. [14] calculated the NED for the CAS-based Finnish Mikkeli WWTP as 0.81 kWh·kgCOD$_{rem}^{-1}$, and identified AnMBR and co-digestion of sludge and organic waste as suitable technological options for retrofitting the facility. Vaccari et al. [45] assessed a total of 267 activated sludge WWTPs in Italy and obtained a range between 3.2 kWh·kgCOD$_{rem}^{-1}$ (treatment capacity < 2000 person-equivalent) and 0.85 kWh·kgCOD$_{rem}^{-1}$ (treatment capacity > 100,000 person-equivalent). A similar study conducted in Greece [3] concluded that WTTPs with above 100,000 person-equivalent treatment capacities operated at a NED of 0.35 kWh·kgCOD$_{rem}^{-1}$ and 0.95 kWh·kgCOD$_{rem}^{-1}$ below 10,000 person-equivalent. Barillon et al. [1] reported net energy demands of six full-scale MBR with two membrane configurations (flat sheet and hollow fiber). The results indicated that energy consumption was approximately within the range of 0.7–2.1 kWh·kgCOD$_{rem}^{-1}$. Pretel et al. [30] simulated an aerobic MBR treating urban wastewater and obtained net energy demands between 0.99 and 1.02 kWh·kgCOD$_{rem}^{-1}$. Finally, Longo et al. [24] gathered energy consumption data of 601 WWTP with different treatments schemes and reported average NED of 2.91, 1.30, and 0.57 kWh·kgCOD$_{rem}^{-1}$ for MBR, extended aeration, and CAS, respectively.

It is important to emphasize that the above-mentioned energy consumption from aerobic-based WWTPs also included nitrification, not occurring in an AnMBR treatment. According to Fraia et al. [8], energy consumption for nitrogen removal would be 6.08 kWh per kg of influent total nitrogen (kWh·TN_{in}) in conventional WWTP with high nitrogen removal efficiency (> 89.2%). Considering the influent nitrogen load in the five scenarios (35.6–54.7 mg N·L^{-1}), the aforementioned removal efficiency for this nutrient would be equivalent to an increase in energy consumption in the 0.202–0.370 kWh·kgCOD$_{rem}^{-1}$ range. NED for AnMBR with nitrogen removal would therefore still be lower than the energy requirements for conventional treatments.

The NED results obtained illustrate the potential of AnMBR systems for recovering resources from organic matter, even when treating sulfate-rich UWW, and for improving the energy efficiency of treatment facilities. The main drivers of the NED results are the OLR and COD removal capacity, HRT, temperature and the COD:SO_4^{2-}-S ratio. In this sense, a previous study focused on the biological performance of the demo-scale AnMBR [36] conducted a mass balance and concluded that the influent COD transformed into biogas was between 12% and 44%, while dissolved methane account for 5–9%. COD consumed by sulfate-reducing bacteria was calculated as 23–42% and the organic matter removed via waste sludge 18–31%. Therefore, there is still room to improve the recovery percentage and, consequently, the energy balance of AnMBR technology.

3.2. *Life cycle costing*

3.2.1. *Capital expenditure*

Fig. 2 summarizes the CAPEX results obtained for the projected full-scale AnMBR system (detailed results and relative weights can be found in the e-supplementary material).

These results showed the lowest value for Scenarios II (€0.080 kgCOD$_{rem}^{-1}$). Scenarios I (€0.096 kgCOD$_{rem}^{-1}$) and V (€0.109 kgCOD$_{rem}^{-1}$) presented similar investment cost, while Scenario IV obtained the highest CAPEX (€0.147 kgCOD$_{rem}^{-1}$).

UF membranes, anaerobic reactor building, blowers, and reciprocating engines were the main contributors to CAPEX. The reactor construction involved an investment of between €0.007 kgCOD$_{rem}^{-1}$ (Scenario II) and €0.014 kgCOD$_{rem}^{-1}$ (Scenario V), representing around 8–13% of the CAPEX. Since reactor volume is determined by the HRT selected in each scenario, there is clear room for optimization of savings in this constructive element. In this regard, Scenario II required the lowest reactor investment per kg of COD removed due to the combination of the highest COD removal (20631 kgCOD·d^{-1}) and the lowest HRT (25 ± 1 h).

In terms of equipment, the results obtained show that UF membranes were among the main contributors to CAPEX. The membrane acquisition cost varied from €0.029 to €0.073 kgCOD$_{rem}^{-1}$ (between 31% and 49% of CAPEX). Strategies for maximizing filtration process productivity (e.g., optimizing transmembrane fluxes, gas sparging intensities for membrane scouring and downtime for physical membrane cleaning) are required to reduce membrane needs and minimize CAPEX. Enhancing material effectiveness to preventing membrane fouling and therefore maintaining high fluxes; improving membrane scouring procedures; revamping module configurations; and membrane market trends (i.e., reduction of membrane acquisition costs) would also help to reduce this item.

Investment in reciprocating engines also appeared as a relevant cost.

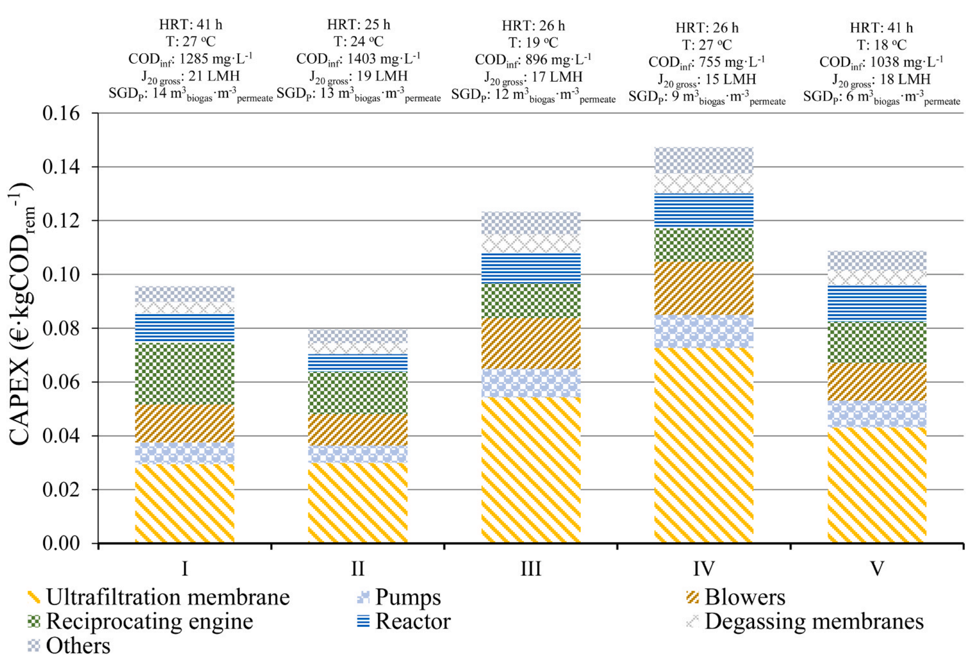


Fig. 2. Allocation of CAPEX for each scenario. "Others" refers to: pumps; degassing membranes; diffusers; pipes; sludge centrifuge; membrane tanks and rotofilter.

Scenario I showed the highest CAPEX allocation for this equipment (€0.023 $kgCOD^{-1}_{rem}$) due to the highest biogas production (5239 ± 4435 $m^3 \cdot d^{-1}$), which entails equipment with greater capacity. Scenario IV required the lowest capital expenditure for this unit (€0.012 $kgCOD^{-1}_{rem}$) because of its low biogas production (1359 ± 999 $m^3 \cdot d^{-1}$). Investment in CHP varied between 8% (Scenario IV) and 24% (Scenario I) of CAPEX.

Blowers also showed a significant contribution to CAPEX. In this case, Scenarios III and IV had the highest expenditures, with €0.019 and €0.020 $kgCOD^{-1}_{rem}$, respectively. Despite their low operating HRT (26 ± 2 h in both scenarios) and intermediate SGD_P (12 and 9 $Nm^3_{biogas} \cdot m^{-3}_{permeate}$, respectively), their comparatively low organic matter removal resulted in a high blower cost per kg of COD removed. On the other hand, Scenario II with similar SGD_P (13 $Nm^3_{biogas} \cdot m^{-3}_{permeate}$) than Scenario III but the highest COD removal showed the lowest blower cost (€0.012 $kgCOD^{-1}_{rem}$). Investment in blowers was between 13% and 15% of CAPEX.

Lastly, pumps (8–9%), DM (4–6%), centrifuge for sludge dewatering (2–3%), rotofilter (2–3%), and pipes (1%) contributed less to CAPEX while membrane tanks (0.2–0.3%) and diffusers (0.1%) were negligible in all scenarios.

Reducing CAPEX is thus crucial to improving investment expenditure competitiveness for AnMBR compared to aerobic MBR and CAS. The strategy to reduce CAPEX involves optimizing the HRT to adjust total reaction volume, increasing the filtration productivity to reduce the membrane needs, properly selecting the CHP power to avoid oversizing and optimizing the capacities of the blowers.

3.2.2. Operating cost

As can be seen in Fig. 3, OPEX varied between €− 0.099 $kgCOD^{-1}_{rem}$ (Scenario I) and €0.041 $kgCOD^{-1}_{rem}$ (Scenario IV) (detailed results and relative weights can be found in e-supplementary material). One of the main contributors to OPEX was TEC (€0.042–0.069 $kgCOD^{-1}_{rem}$), representing 41–46% of the total operating cost, excluding energy recovery.

UF replacement was also a significant OPEX cost. In this regard, Scenario IV, with the lowest $J_{20\ gross}$ (15 LMH) required the largest membrane area and therefore showed the highest replacement cost (€0.028 $kgCOD^{-1}_{rem}$) while Scenario I with the highest $J_{20\ gross}$ (21 LMH) showed the lowest UF replacement cost (€0.011 $kgCOD^{-1}_{rem}$). This expense was between 9% and 17% of OPEX, excluding energy recovery. It is important to highlight that FR directly affected UF membrane replacement cost since membrane lifespan and the associated replacement frequency is determined by the maximum total contact with chlorine permissible during membrane cleaning. However, in this study all the scenarios required one replacement over the considered lifespan of the facility so that there were no differences in this concept.

The contribution of reciprocating engines to OPEX was high in Scenarios I (€0.017 $kgCOD^{-1}_{rem}$; 14%), II (€0.012 $kgCOD^{-1}_{rem}$; 12%) and V (€0.011 $kgCOD^{-1}_{rem}$; 9%), in which a significant amount of biogas was produced. However, overall CHP contributed with savings of €0.099–0.181 per $kgCOD_{rem}$ when CAPEX and OPEX are balanced with cost reductions due to energy recovery. Moreover, these revenues do not consider the heat recovered and the environmental benefits of avoiding methane release, which is an additional value to take into account. Further improvement in CHP efficiencies would also improve AnMBR economic and environmental performance. Revenues could also be increased if energy recovery from biogas were to be conducted in the tariff periods with the highest energy costs and not proportionally distributed throughout the day, as considered in this study. However, this optimization strategy would entail an additional infrastructure (e.g., gasometer) and economic appraisals to assess its feasibility.

The fee for effluent discharging to surface water significantly contributed to the operating costs (€0.010–0.020 $kgCOD^{-1}_{rem}$; 9–12% of OPEX), excluding energy recovery. Reducing or even removing this charge in case of water (and nutrient) recovery could be a proper governance instrument for promoting this practice, taking advantage of the economic, environmental, and social benefits of water (and nutrient) recovery. By way of example, Jiménez-Benítez et al. [19] evaluated two case studies of fertigation with AnMBR effluent in Spain and Italy, showing up to 100% in mineral fertilizers savings, and obtaining reductions of 75% (Spain) and even a carbon sink effect (Italy) when CO_2 emissions associated with savings on fertilizer production are considered. For this, appropriate reclaimed water management tools should be implemented, e.g., according to the Regulation (EU) 2020/741 on the minimum requirements for water reuse [6].

Finally, it is important to highlight the reduced OPEX associated with sludge conditioning (€0.002–0.003 $kgCOD^{-1}_{rem}$; 2%) and sludge disposal (€0.002–0.004 $kgCOD^{-1}_{rem}$; 2%), which confirm the potential benefits of

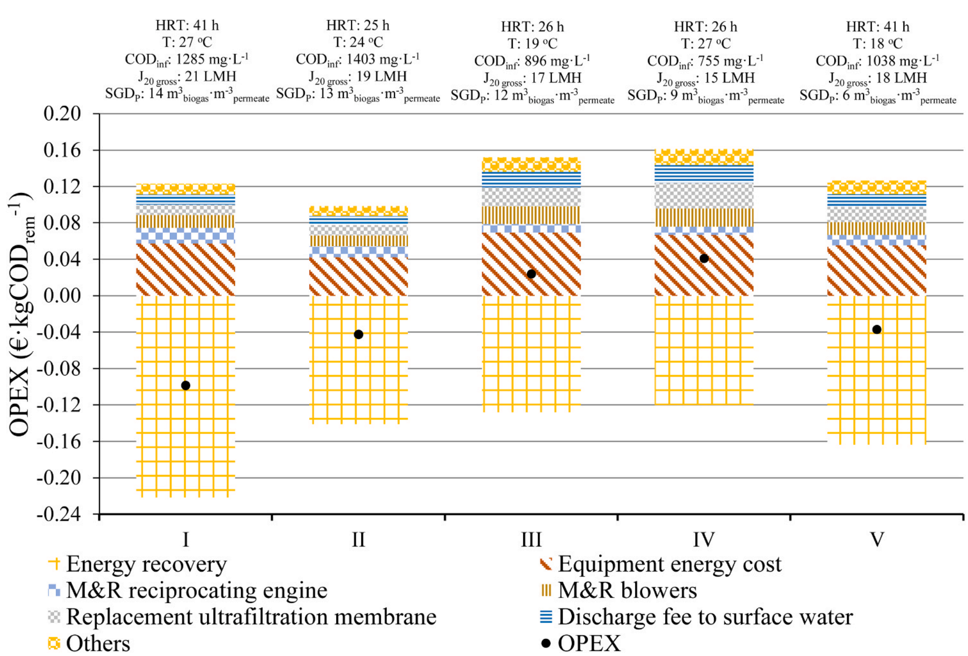


Fig. 3. Allocation of OPEX for each scenario. “Others” refers to: power cost; membrane cleaning; sludge conditioning; sludge disposal; M&R pumps; M&R CHP; M&R blowers; M&R civil works; M&R pipes and diffusers; M&R rotofilter; discharge fee to surface water; and degassing membrane replacement.

AnMBR as an alternative to aerobic treatments (e.g., CAS, EA) to address waste sludge management.

These results are in agreement with the trends found by Judd [22] in which equipment energy needs, UF membrane replacement and discharge fees are the main costs in the OPEX.

3.2.3. Total annual cost

Fig. 4 shows that TAC varied from €− 0.003 $kgCOD_{rem}^{-1}$ (Scenario I) to €0.188 $kgCOD_{rem}^{-1}$ (Scenario IV). The results from Scenario I are explained by a low CAPEX (€0.096 $kgCOD_{rem}^{-1}$) mainly due to the lower UF membrane cost (€0.029 $kgCOD_{rem}^{-1}$) resulting from applying high $J_{20\ gross}$ (21 LMH); and the lowest OPEX (€−0.099 $kgCOD_{rem}^{-1}$) thanks to its high energy recovery (€−0.222 $kgCOD_{rem}^{-1}$). On the other hand, Scenario IV had the highest CAPEX (€0.147 $kgCOD_{rem}^{-1}$) influenced by its lower $J_{20\ gross}$ (15 LMH) and therefore high UF membrane cost (€0.073 $kgCOD_{rem}^{-1}$); and OPEX (€0.041 $kgCOD_{rem}^{-1}$) influenced by its high total maintenance and replacement costs (€0.068 $kgCOD_{rem}^{-1}$) and equipment energy costs (€0.067 $kgCOD_{rem}^{-1}$).

Table 2 shows TAC results reported in bibliography for different treatment technologies for urban wastewater. It is important to highlight the difficulties involved for achieving robust comparisons due to the variety of functional units, assumptions and hypotheses, or treatment schemes within the reviewed studies. Therefore, the comments presented here should be considered as indicative. Ferrer et al. [7] modelled an AnMBR with 50,000 $m^3 \cdot d^{-1}$ of treatment capacity. These authors obtained a TAC between €0.14 $kgCOD_{rem}^{-1}$ (low-sulfate influent and considering methane captured from the effluent) and €0.20 $kgCOD_{rem}^{-1}$ (sulfate-rich influent without methane capture). Pretel et al. [28] simulated an AnMBR treating sulfate-rich and low-sulfate wastewater at 15 and 30 °C and obtained TAC in the range €0.18–0.43 $kgCOD_{rem}^{-1}$. With regards to MBR technology, Pretel et al., [30] obtained a TAC of €0.21 $kgCOD_{rem}^{-1}$, which are significantly lower than those obtained by Bertanza et al. [2] (€1.74–3.33 $kgCOD_{rem}^{-1}$). In the case of CAS, Pretel et al., [30] calculated a TAC of €0.14 $kgCOD_{rem}^{-1}$, while Bertanza et al. [2] reported much higher TAC for the same technology (€1.18–2.51 $kgCOD_{rem}^{-1}$). Lastly, Jafarinejad [18] modelled three different treatments schemes: CAS, sequencing batch reactor and extended aeration activated sludge. The author obtained TAC of €0.93–0.95, €1.07–0.13, and €0.95–1.04 $kgCOD_{rem}^{-1}$, respectively. According to these values, the AnMBR system evaluated in the present study would be competitive even in the case of Scenario IV, which showed the highest TAC, i.e., €0.188 $kgCOD_{rem}^{-1}$.

Finally, CAPEX appears as the main contributor to TAC in Scenarios II to V, which is consistent with the results reported by Smith et al. [43]. Scenario I is especially relevant, in which the high energy production offset the other operating costs and CAPEX, bringing the TAC close 0 (€−0.003 $kgCOD_{rem}^{-1}$). Optimization strategies should thus be used to address both CAPEX and OPEX, since both items showed a potential for reducing TAC and contributing to the economic sustainability of AnMBR-based wastewater treatment.

3.3. Sensitivity analysis

By way of example, the results from the SA of Scenario I are shown in Fig. 5, providing the main influencing parameters ($b_i \geq 0.1$).

3.3.1. OPEX output

Fig. 5a shows the SA results for OPEX. The positive main influencing parameters were HRT, blower cost, biogas for reactor mixing, SGD_P applied to membrane tanks and the AnMBR lifespan. Most of these were related to the biogas sparging needs to mix the reactor or stir the membranes. Reducing gas sparging intensities would thus entail significant savings in OPEX. Optimizing HRT would also reduce reactor volume and hence mixing requirements. Finally, extending the AnMBR lifespan by 10%, from 20 to 22 years, would increase the replacement cost associated with membranes and equipment and thus also the OPEX. In Scenario I, it would be needed to replace both the UF membranes and the equipment (lifespan of 10 years) once throughout the lifespan set for the AnMBR plant (i.e., 20 years). By applying an uncertainty of 10%, AnMBR lifespan would increase from 20 to 22 years in the upper boundary. Thus, new membranes and equipment would be necessary for years 21 and 22. However, the total cost of these new devices would not be distributed in a decade, but only among the two additional years, increasing therefore the total replacement cost for this scenario. In this respect, the second replacement for membranes and equipment would be on service only a 20% of their potential durability.

In relation to the parameters with negative influence, increasing the membranes' useful life and effective filtration time (i.e., operating

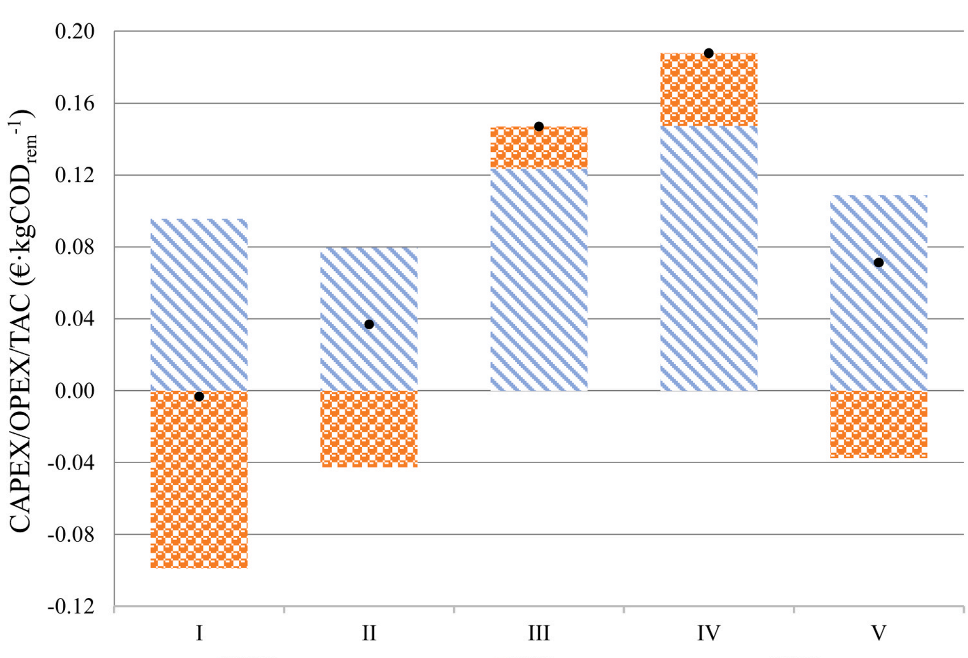


Fig. 4. Allocation of total annual cost (TAC) for each scenario.

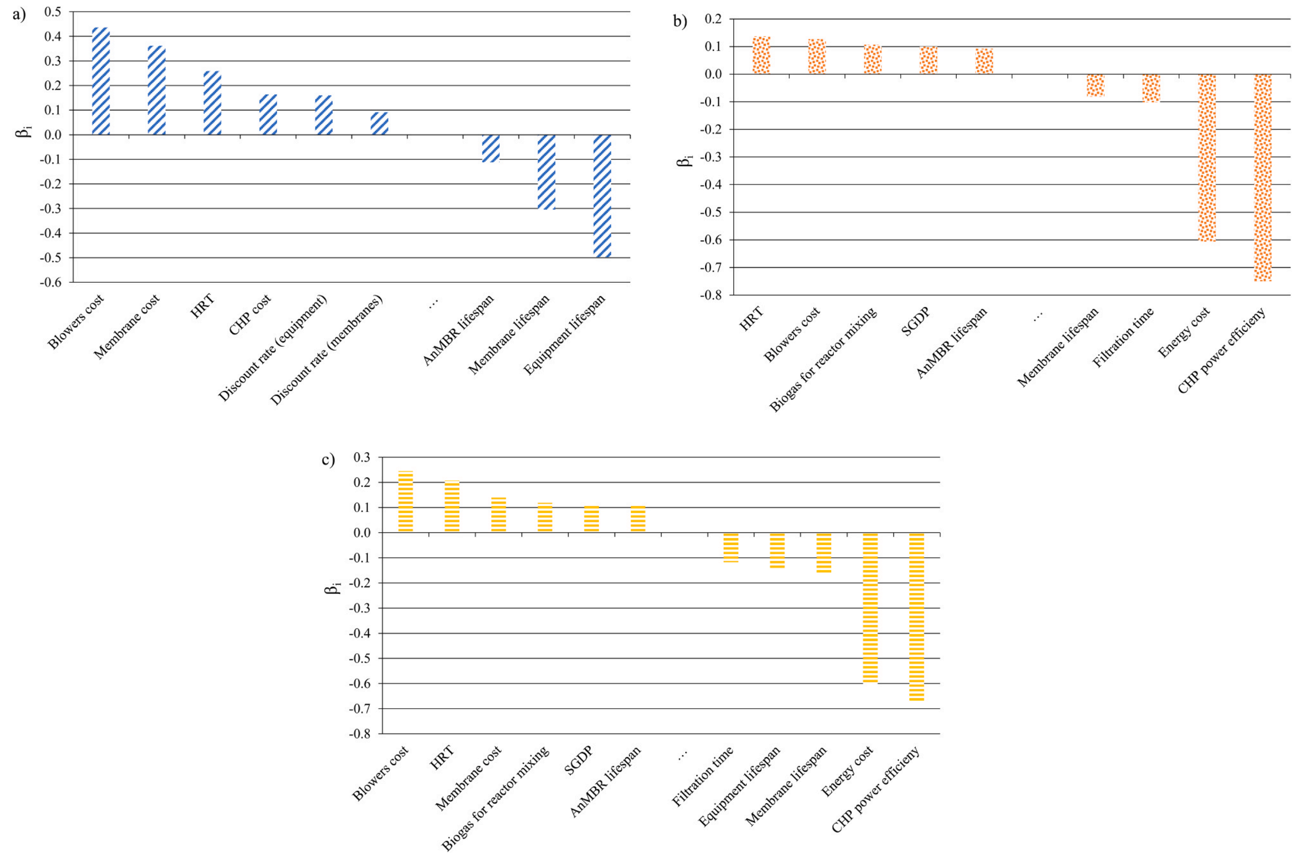


Fig. 5. Sensitivity analysis results for Scenario I. Main influential factors on model outputs: a) CAPEX, b) OPEX, and c) TAC.

period without downtime for physical cleaning of the membrane) would reduce the OPEX related to membrane replacement and filtration productivity. The largest effect would be associated with energy costs and CHP efficiency. Since Scenario I showed high energy production and performed as a net energy producer, favoring energy recovery through improved CHP power efficiency, and increasing the energy price for its sales would produce additional revenues and reduce OPEX. It is important to highlight that the effect of energy price would be the opposite in a net energy consumer scenario.

3.3.2. CAPEX output

The parameters with the greatest influence on CAPEX are shown in Fig. 5b. The SA results indicated that the costs associated with the blowers, membranes, size of the reactors (through the HRT) and capacity of the CHP (and its consequent cost) had a significant direct influence on CAPEX, as did the discount rates applied to the acquisition of membranes and equipment.

Therefore, reducing the gas sparging requirements, improving the filtration productivity (and therefore the necessary membrane surface), and optimizing the HRT to reduce the volume of the reactors, would be useful strategies to reduce CAPEX. Moreover, an accurate selection of CHP capacity would also be important to avoid over dimension since its influence on CAPEX was also significant.

Finally, CAPEX could also be reduced by increasing the useful lifespan of the plant, membranes and equipment. This necessarily involves adequate maintenance and optimization of the filtration operation, thus reducing the need for chemical cleaning of the membranes, which is mainly responsible for their deterioration.

3.3.3. TAC output

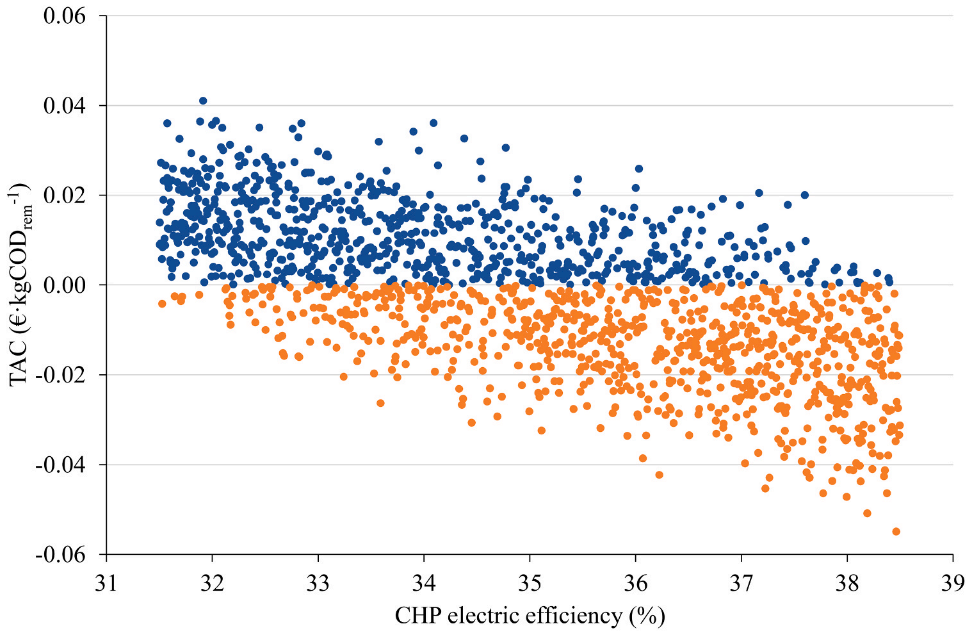
Fig. 5c shows the input parameters with a strong influence on TAC output, all of which were identified as influential parameters on CAPEX and OPEX outputs. The elements associated with gas sparging (blowers, SGD_P and reactor mixing) jointly with the equipment replacement needs derived from extending the AnMBR lifespan implied an increase in TAC. On the other hand, increasing the membrane and equipment lifespan, improving filtration productivity, and obtaining better conditions for energy recovery (e.g., through higher profits from the sale of electric power, or an improvement in the efficiency of CHP systems) would improve the overall economic assessment represented by the TAC.

Finally, by way of example, Fig. 6 shows the Monte Carlo simulation results for TAC with regard to CHP power efficiency (Fig. 6a) and energy tariff variation (Fig. 6b) in Scenario I. Fig. 6a shows that CHP power efficiencies over 38.5% would lead to a TAC below €0 $kgCOD_{rem}^{-1}$ in all simulations, which represents revenues for treating wastewater. On the contrary, cogeneration energy efficiencies lower than 31.5% obtained a positive TAC in all simulations, which indicates an economic outlay for treating the water. In those situations, with efficiencies between 31.5% and 38.5%, the TAC may be positive or negative depending on the rest of the variables, but the probability of a negative TAC increases as the efficiency of the CHP increases.

Moreover, a 10% increase in the price of the electricity sold would led to a negative TAC (see Fig. 6b), while a similar reduction of in the energy price would imply a TAC greater than €0 $kgCOD_{rem}^{-1}$. As already mentioned in Section 3.3.1, the influence of the electricity tariff on the results is marked by the fact that the scenario considered performed as a net producer of energy. Otherwise, an increase in the energy tariff would increase costs. These results highlight the importance of considering energy in AnMBR feasibility studies.

3.4. Uncertainty analysis

Fig. 7 gives the empirical cumulative distribution function (eCDF) and confidence interval for the OPEX, CAPEX and TAC outputs in



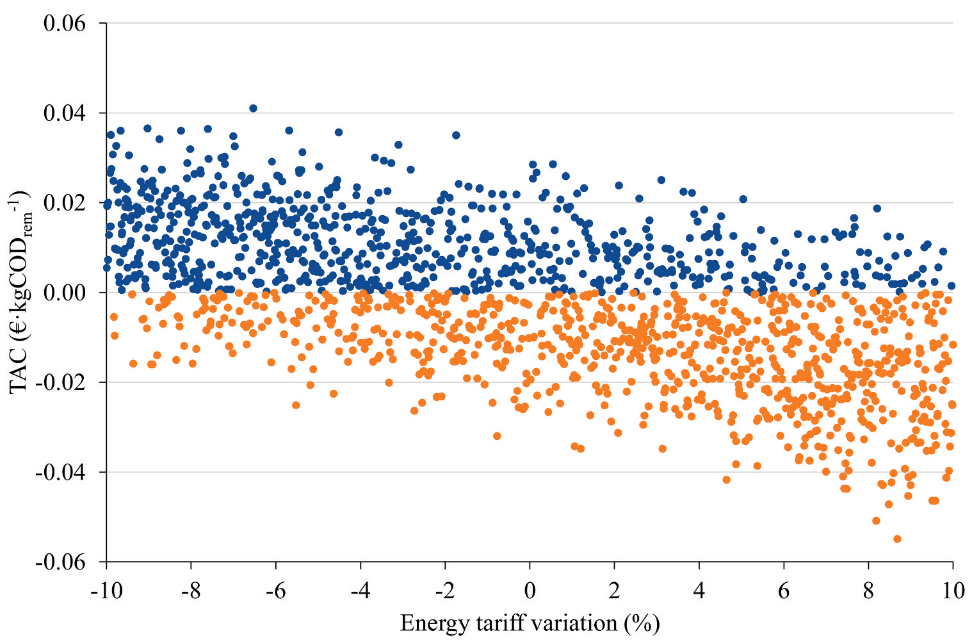


Fig. 6. Monte Carlo simulation results for TAC regarding a) CHP power efficiency (%) and b) energy tariff variation (%). Red dots indicate simulation outputs with TAC < 0 and blue dots represent simulation outputs with TAC > 0.

Scenario I. The eCDF followed a standard normal distribution in all cases and the narrow confidence bands indicated low variability of each value. Fig. 7 also shows that CAPEX remained at positive values, which is coherent with investment requirements, while OPEX sign mainly depends on energy recovery. Finally, Fig. 7c shows that the TAC could change from negative to positive results, going from a benefit scenario for water treatment to one of treatment expenditure.

Regarding the percentiles, the 5th-95th range for OPEX, CAPEX, and TAC resulted in €0.014, €0.051, and €0.052 per $kgCOD_{rem}$, respectively (see Table 3)(Table 4). These results show that the uncertainty in CAPEX is higher than in OPEX, being the former the main contributor to TAC uncertainty. Therefore, in order to reduce sources of uncertainty, efforts should be made to obtain reliable and robust investment data, while a careful design of AnMBR systems should be conducted to ensure an optimal performance. According to Section 3.3, improving data regarding UF membranes, CHP and blower cost would reduce CAPEX uncertainty. In the case of OPEX, accurate data related to reactor mixing and membrane sparging requirements and revenues through energy recovery (CHP power efficiency and energy tariff) would also improve overall UA.

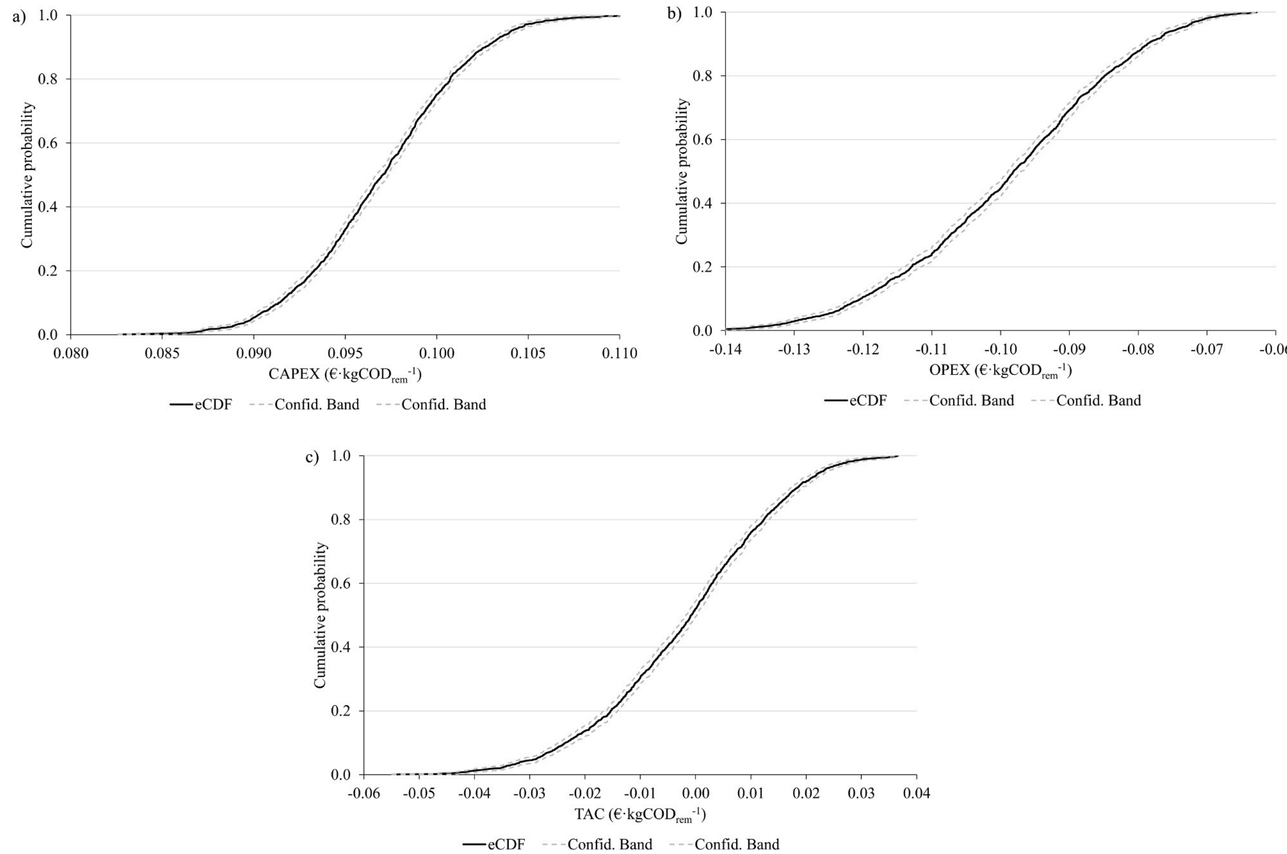


Fig. 7. Uncertainty analysis results for Scenario I. Empirical cumulative distribution function (eCDF) for: a) CAPEX, b) OPEX, and c) TAC.

Table 3
Power energy requirements, total energy recovery (TER), total energy consumption (TEC), net energy demand (NED) and % of TEC for each scenario.

Scenario	Rotofilter	Feeding pumping	Permeate pumping	Membrane scouring	Sludge wasting	Sludge recycling	Sludge dewatering	Reactor mixing	Dissolved methane vacuum blower	TER (biogas + dissolved methane capture)	TER (only biogas)	TEC	NED
						kWh·kgDQO$^{-1}_{rem}$							
I	0.001	0.002	0.010	0.089	≈ 0.000	0.006	0.016	0.099	0.001	-0.870	-0.827	0.225	-0.645
II	0.001	0.003	0.010	0.080	≈ 0.000	0.006	0.009	0.056	0.001	-0.553	-0.513	0.166	-0.387
III	0.002	0.005	0.022	0.120	≈ 0.000	0.010	0.016	0.097	0.001	-0.503	-0.429	0.273	-0.230
IV	0.002	0.006	0.005	0.108	≈ 0.000	0.011	0.018	0.112	0.001	-0.472	-0.386	0.262	-0.210
V	0.001	0.003	0.014	0.048	≈ 0.000	0.008	0.019	0.124	0.001	-0.643	-0.581	0.218	-0.425
						TEC (%)							
I	0.5	1.0	4.4	39.8	0.1	2.8	6.9	44.3	0.3				
II	0.6	1.7	6.3	48.2	0.1	3.6	5.4	33.8	0.3				
III	0.6	1.8	8.1	44.1	0.1	3.6	5.7	35.6	0.5				
IV	0.7	2.1	1.9	41.0	0.1	4.3	6.8	42.6	0.4				
V	0.6	1.3	6.3	21.9	0.1	3.6	8.9	56.8	0.5				

Table 4
Uncertainty results for Scenario I: 5th-95th percentile and 5th-95th range.

	€·kgCOD$^{-1}_{rem}$		
	CAPEX	OPEX	TAC
5th percentile	0.090	-0.125	-0.029
95th percentile	0.104	-0.074	0.023
5th-95th range	0.014	0.051	0.052

4. Conclusions

Favorable energy balances were achieved in all the scenarios evaluated. NED varied between − 0.210 kWh·kgCOD$_{rem}$ $^{-1}$ and − 0.645 kWh·kgCOD$^{-1}_{rem}$. Reactor mixing (0.056–0.124 kWh·kgCOD$^{-1}_{rem}$) and membrane scouring (0.048–0.120 kWh·kgCOD$^{-1}_{rem}$) showed the highest energy consumption.

The major contributors to CAPEX were UF membranes (€0.029–0.073 kgCOD$^{-1}_{rem}$; 31–49%), reciprocating engines acquisition (€0.012–0.023 kgCOD$_{rem}$ $^{-1}$; 8–24%), and reactor building (€0.007–0.014 kgCOD$^{-1}_{rem}$; 8–13%). OPEX was mainly influenced by equipment energy needs (€0.042–0.069 kgCOD$^{-1}_{rem}$; 41–46%), UF membrane replacement (€0.011–0.028 kgCOD$^{-1}_{rem}$; 9–17%) and discharge fee

(€0.010–0.020 $kgCOD_{rem}^{-1}$; 9–12%). CHP maintenance and replacement was also a significant operating cost (€0.009–0.017 $kgCOD_{rem}^{-1}$; 6–14%), although it finally provided revenues through energy sale (€0.120–0.222 $kgCOD_{rem}$ $^{-1}$). The TAC varied between -€0.003 and €0.188 $kgCOD_{rem}^{-1}$, with CAPEX as the main cost.

The results obtained from LCC assessment combined with SA and UA showed that AnMBR represents a potential alternative to conventional aerobic treatments. Its definitive impulse will go through, among other aspects, reinforcing its economic sustainability, for which it is necessary to optimize operational aspects (e.g., HRT, transmembrane flux, reactor mixing and membrane sparging) and technological developments that allow increasing membrane lifespan and improve CHP power efficiency. Finally, other aspects that would also result in improving its economic performance have to do with nutrient recovery present in its effluent and favorable pricing for the sale of electricity produced from the methane generated.

CRediT authorship contribution statement

A. Jiménez-Benítez: Writing – original draft, Investigation, Methodology, Formal analysis, Writing – review & editing. **A. Ruiz-Martínez** Investigation, Methodology, Resources, Writing – review & editing. **J. Ferrer:** Investigation, Validation, Writing – review & editing, Supervision, Management and coordination responsibility for the research activity planning and execution. **J. Ribes:** Resources, Investigation, Management and coordination responsibility for the research activity planning and execution, Validation, Writing – review & editing. **F. Rogalla:** Investigation, Methodology, Resources. **Á. Robles:** Definition, Methodology, Formal analysis, Investigation, Validation, Writing – review & editing, Supervision, Management and coordination responsibility for the research activity planning and execution.

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. Supporting information

Supplementary data associated with this article can be found in the online version at doi:10.1016/j.jece.2023.110267.

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