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COST ESTIMATION OF COMBINED MEMBRANE SEPARATION AND DIFFERENT ADVANCED OXIDATION PROCESSES PRE-TREATMENT OF OILY WASTEWATER

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Abstract: The current study deals with comparison of the cost of oily wastewater treatment by membrane filtration associated with different pretreatment methods. The cost evaluation was completed for model oily wastewaters containing 100 ppm crude oil in saline or distilled water, and in some cases for real produced water, aiming at elimination of oil content. Three different pre-treatment methods – ozonation, Fenton and photo-Fenton pretreatments – and the modification of the used membrane surfaces with photocatalytic nanoparticles were used for the mitigation of fouling and enhancing oil elimination efficiency during the process. The cost estimation was evaluated on the basis of flux decline and the cost of pre-treatments. Results showed that pre-oxidation enhances the flux, thus potentially decreasing the cost of combined treatment. The lower cost was achieved by the application of modified membrane surfaces, which shows that the development of antifouling surfaces may further decrease costs of membrane filtration.

Keywords: *Oily wastewaters, membrane filtration, waste reduction, titanium dioxide, carbon nanotubes, cost evaluation*

1. INTRODUCTION

Oily wastewaters originate from several industries, especially the oil industry, oil mining, oil refining, oil storage transportation, and petrochemical industries. Besides them, other industries like machinery or food industry also produce a considerable amount of oily wastewaters. The composition and characteristics of these waters strongly depend on the releasing industry; however many of them contaminated by crude oil, with nondescript composition. As crude oil is a mixture of different organic and inorganic compounds, which are often toxic, discharged oily wastewaters pollute surface and underground water, endangering aquatic resources and human health [1], [2].With the continuous improvement in environmental standards, the existing

methods have been unable to meet the requirements, thus more efficient technologies should be developed [1].

In order to meet environmental standards and to achieve reusable and recyclable water from oily wastewaters, several investigations have been focused on developing water treatment technologies for oily wastewaters [3]. Conventional water treatment technologies involve physical, chemical and biological methods. Although the biological treatments are cost-effective, these industries produce large volumes of wastewater contain recalcitrant compounds, possibly toxic organic pollutants which are difficult to treat by biological methods [4].

In oily wastewater emulsions, the typical oil concentrations vary from 50 up to 1,000 ppm of oil [5]. Several techniques have been used for oil-in-water emulsion purification, such as air flotation, heating, ozonation, coagulation-flocculation, and membrane filtration [6]. The physical and chemical methods that can be applied have high capital costs, and the cost of chemicals for chemical treatment is high [3]. Moreover, the current methods cannot remove the dispersed, suspended oil content.

Among the physical processes, membrane-based separation is a promising technology for 21st century since there is no necessity for chemical additives and the energy costs are relatively low [6]. Ultrafiltration (UF) is the most effective treatment for oily wastewater, possessing high oil-removal efficiency however, membrane fouling is a considerable limiting factor.

Membrane separation processes combined with pretreatments may enhance the elimination efficiency and reduce the filtration resistance. Earlier studies showed that pre-oxidation of oily wastewaters may be an advantageous solution, as it degrades the organic pollutants in oily wastewaters, and in parallel, it may improve the flocculation efficiency and particle removal during the filtration step [7], [8]. These advanced oxidation processes may include ozone, UV/hydrogen peroxide, Fenton or photo-Fenton reaction, and heterogeneous photocatalysis. Advanced oxidation processes and ozone treatment generate free radicals, which are able to react with the contaminants directly and indirectly, and finally decompose to oxygen. In these reactions two typical pathways have been observed, influencing membrane filtration parameters: (1) the micro-flocculating effect producing associated colloidal particles, and (2) degradation of organic materials *(Figure 1)*. The latter decreases the retention of pollutants and may increase the pore fouling, thus short-term oxidation pretreatment may lead to micro-flocculation and result in large floccules, thereby reducing membrane fouling [9], [10]. This means that for combined techniques the pre-oxidation processes should be optimized in terms of filtration parameters.

Another approach to membrane fouling mitigation is to modify membranes with photocatalytic nanoparticles, therefore combining the advantages of membrane filtration (physical separation) and the advantage of photocatalysis (non-selective organic matter degradation) [12]. TiO_2 is one of the most commonly used photocatalysts due to its good physical and chemical properties, availability, high photocatalytic activity, and desirable hydrophilic properties [13–17], but recently the application of new nanoparticles is being widely investigated [18].

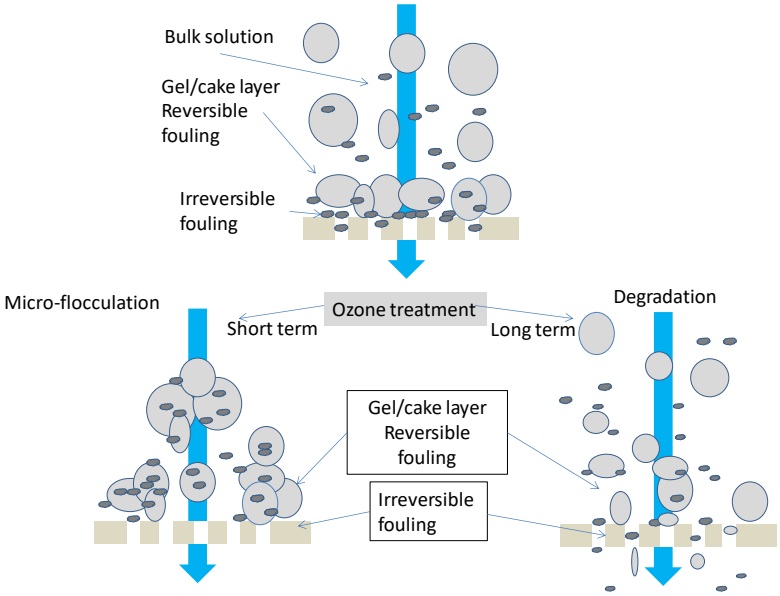

Figure 1
Possible effects of ozone pretreatment on membrane filtration [11]

Several investigations have focused on the optimal operational conditions, and only a few works contain cost evaluation, which is essential to choose an appropriate method for water treatment technology. The cost of water treatment depends on both the technological aspects and the regulation requirements.

The economics of membrane filtration are determined by the initial investment cost of the membrane modules, equipment, and facilities. The key parameter is the permeate flux, which is determined by the driving force (typically the transmembrane pressure) and the hydrodynamic resistance of the membrane and the particle layer accumulates on the membrane surface, namely the fouling. Membrane fouling affects both capital and operating costs by affecting the membrane area and energy requirements [19].

For estimating the cost of technology to be chosen, preliminary design and economic potential estimation should be attained [20]. Preliminary experimental data provide a basis for the operational parameters and technical requirements, while economic evaluations give an estimate of capital and operational costs. The costs of advanced oxidation processes are highly dependent on the quality of the source water to be treated and effluent treatment goals [21]. Although very few studies aim at investigation of advanced oxidation processes (AOPs) involving cost estimation analysis, summarizing them, they conclude that the main factors influencing the prices are the removal efficiency and flow rate; capital, operational and management costs increased with higher removal efficiency [22].

Although AOPs are known to be relatively expensive processes, by applying them as pre-treatment, without aiming for total pollutant elimination during oxide-

tion, their cost may become competitive to membrane filtration. In this work a comparison of the cost of oily wastewater treatment by membrane filtration associated with different pre-treatment methods was performed. The cost evaluation was completed for model oily wastewaters containing 100 ppm crude oil in saline or distilled water, and in some cases for real produced water, aiming elimination of oil content. In the present study the four different promising methods of pre-ozonation [23], Fenton or photo-Fenton reaction, and photocatalytic nanoparticle modified membranes [18], [12] were used for the mitigation of fouling during membrane filtration. The cost estimation was evaluated on the basis of flux decline and the cost of pre-treatments.

2. EXPERIMENTAL

2.1. Materials and methods

Model wastewater composition: oil in water (o/w) emulsions (c_{oil} = 100 mg L^{-1}; $d_{droplets}$ < 1.5 μm) were prepared in 2 steps, using crude oil (provided by MOL Zrt.; Hungary), and ultrapure water (PureLab Pulse, ELGA Labwater, UK). Intensive stirring (35,000 rpm, 1 min) of crude oil and water was followed by 10 min ultrasonic homogenization (Hielscher UP200S, Germany) at 25 °C (using maximal amplitude and cycle). For the photocatalytic experiments 20 ppm oil concentration was set by the dilution of the emulsion. In case of saline water, a model of real groundwater located in south Hungary was prepared, which contained the following salts: 2.26 g L^{-1} $NaHCO_3$; 53.4 mg L^{-1} NH_4Cl; 19.1 mg L^{-1} $CaCl_2$; 20.9 mg L^{-1} KCl; 93.5 mg L^{-1} NaCl; 4.5 mg L^{-1} $FeCl_3$ and 35.1 mg L^{-1} $MgSO_4$ (Sigma Aldrich; analytical grade).

The investigated real wastewater was produced from the southern part of Hungary, containing crude oil. (Extractable oil content 28±2 ppm, turbidity 44.2±0.5 NTU, pH 7.95±0.05, COD 927±10 mg L^{-1}.)

Pre-ozonation was carried out in a glass batch reactor, the ozone was generated from clean oxygen (Messer; 3.5) by a flow-type ozone generator (BMT 802N, Germany) and it was bubbled through a diffuser into a batch reactor containing 400 mL of the given oil-in-water emulsion, equipped with a magnetic stirrer (Fig. 2a). The applied flow rate was 1 L min^{-1} and ozone concentration of inlet and outlet was measured using a WPA Biowave II type UV spectrophotometer, with the wavelength set at λ = 254 nm) to determine the absorbed volume of ozone. If pre-ozonation was applied, its duration was only 5 minutes in all cases, since our previous studies [23], [24] proved that longer pre-ozonation can result in pore blocking (due to the fragmentation of the oil droplets) and also results in lower purification efficiency because of the generated water-soluble organic oxidation by-products. The applied 5 min long pre-ozonation resulted in 30 ± 5 mg L^{-1} of absorbed ozone dose. The remaining dissolved ozone was purged by oxygen after the treatment to avoid damage of the used membrane. In the case of real wastewater, the duration of ozone treatment was 2 min, with 28±2 mg L^{-1} absorbed ozone.

Fenton-type reactions were carried out at pH = 4, the concentration of H_2O_2 (VWR Hungary) was 300 ppm, ratio of n_{Fe}:n_{H2O2} = 1 : 25 *(Figure 2b)*, and before membrane filtration the pH was set to be 7, and (possibly) remaining H_2O_2 was

decomposed enzymatically with catalase enzyme. Photo-Fenton reactions were carried out by the irradiation of the emulsion with an immersed compact fluorescent UV-tube ($\lambda_{max}\approx254$ nm, 10 W, Lighttech Ltd, Hungary). For determination of the residual amount of hydrogen peroxide, COD measurements were performed before and after the addition of catalase enzyme.

For the membrane surface modification, commercial titanium dioxide (TiO_2; Aeroxide P25, Germany, d = 25–39 nm, a^S_{BET} = 50.6 m^2 g^{-1}) and carbon nanotubes (CNT; Nanothinx NTX1 multi-walled carbon nanotube, Greece, $l \geq 10$ μm; d = 15-35 nm) were applied. Nanomaterials (TiO_2 and CNT) by themselves or in composites (TiO_2 with 1 wt% CNT) were suspended in 2-propanol (c = 400 mg L^{-1}) by 1 min ultrasonic homogenization (Hielscher UP200S, Germany) at 25 °C (maximal amplitude and cycle were applied). 40 mg of the given nanomaterial (suspended in 100 mL of 2-propanol) was immobilized on a polyvinylidene fluoride (PVDF) membrane (New Logic Research INC, USA, 100 kDa; ~1.0 mg cm^{-2} catalyst coverage) by physical deposition method: the suspension was filtered through the membrane, applying 0.3 MPa transmembrane pressure in a batch-stirred membrane reactor (Millipore, XFUF07601, USA), followed by drying in air at room temperature *(Figure 2c)*.

Membrane filtration experiments were carried out in a Millipore XFUF07601 batch-stirred membrane reactor, which was equipped with commercial polyethersulfone (PES) or polyvinylidene fluoride ultrafilter membranes (PVDF, 100 kDa), or in the case of Fenton-type reactions microfiltration PES membranes (pore size 0.2μm) (New Logic Research INC, USA) or with photocatalytic nanomaterial covered membranes. Filtration was carried out using 0.1 MPa transmembrane pressure and 5.83 s^{-1} stirring speed (350 rpm). In all filtration experiments, 250 mL emulsion was filled into the reactor and filtered until the production of 200 mL permeate (volume reduction ratio: VRR = 5).

The purification efficiencies were determined by measuring the chemical oxygen demand (COD) and the extractable oil content (TOG/TPH) of the feed and the permeate. COD was measured by a standard potassium-dichromate oxidation based method, using standard test tubes (Lovibond). The digestions were carried out in a COD digester (Lovibond, ET 108) for 2 h at 150 °C and the COD values were measured with a COD photometer (Lovibond PC-CheckIt). Extractable oil content was measured by a Wilks InfraCal TOG/TPH type analyzer, using hexane as extracting solvent. The purification efficiency (R) was calculated as:

$$R = \left(1 - \frac{c}{c_0}\right) \cdot \mathbf{100\%}\,, \tag{1}$$

where c_0 is the COD or the TOG/TPH value of the feed and c indicates the values of the permeate.

To compare the performance of different AOPs, Oxygen-equivalent Chemical-oxidation Capacity (OCC, kg O_2/m^{-3}) was used to quantify the oxidants used in the

ozone treatment and Fenton process, and was determined on the basis of stoichiometric calculations as [22]:

$$OCC=1.000[O_3] =0.471[H_2O_2], \quad (2)$$

where $[O_3]$ is the demanded ozone concentration (kg O_3/m^3) and $[H_2O_2]$ is the demanded hydrogen peroxide concentration (kg H_2O_2 / m^3).

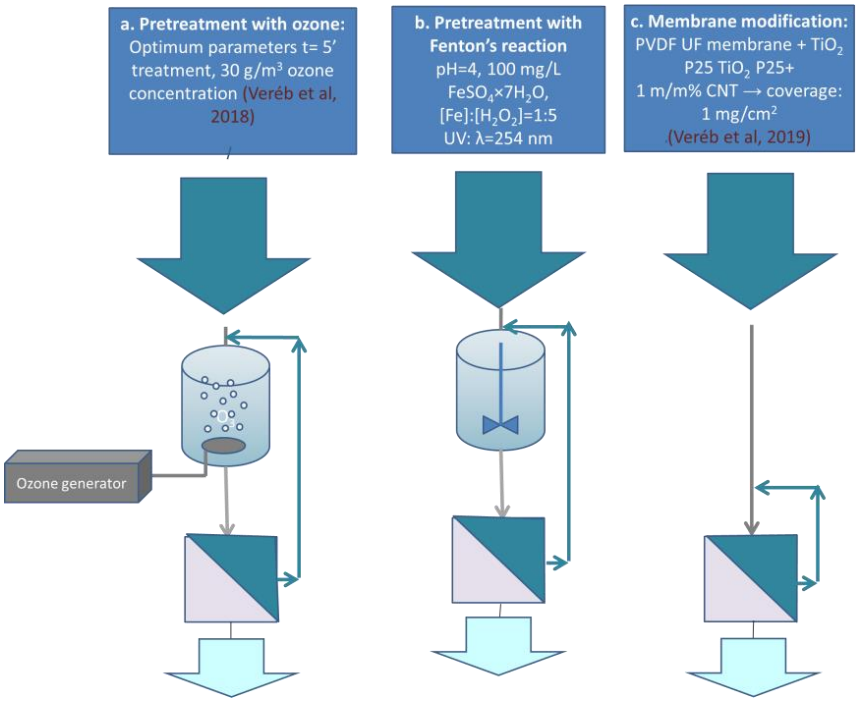

Figure 2

Experimental design of pre-treatments and membrane filtration experiments [24], [18]

2.2. Cost estimation methodology

The cost estimation of combined processes was based on the costs of membrane filtration and pretreatments. During calculations 1,000 m^3/day feed volume was assumed *(Figure 2)*. Amortization was assumed to be 30 years. Taking into consideration the capital investments and operational costs, the total cost of 1 m^3 of wastewater purification was calculated. The calculation of costs of membrane filtration was carried out by the work published by [20]. The calculation of total capital investment for the membrane unit and plant includes fixed capital and working capital (operational cost) investments, while fixed capital investment comprises both direct and indirect costs. A similar method was applied for the cost calculation of the pretreatments [21].

3. RESULTS AND DISCUSSION

3.1. Effect of different pre-treatments on pollutant elimination efficiencies

In order to make the oxidation pre-treatments comparable, the Oxygen-equivalent Chemical-oxidation Capacity (OCC) was calculated during the pretreatments. Due to the negative long-term effects of pre-oxidation on filtration efficiency, short time (5 min, and in case of real wastewater 2 min) pretreatments were performed. In case of ozone treatments, the OCC was 30±5 g m^{-3}, in case of real wastewaters 28±2 g m^{-3}. In case of Fenton reaction, the OCC was 11.28, and in case of photo-Fenton 33 g m^{-3}. In case of catalyst-modified membrane surfaces, in one set of experiments the wastewaters were not pretreated before filtration; in these cases the photocatalytic effect was used for cleaning the membranes [24], while in another set of experiments ozone-pretreatments were performed (OCC=28±2 g m^{-3}).

The pretreatments alone caused only a slight, if any, decrease in COD and extractable oil content. After the filtration, the overall extractable oil elimination was very high, while COD elimination efficiency was a little bit lower, but generally above 97% *(Table 1)*. The salinity of the water generally decreased the elimination efficiency.

Table 1

Chemical oxygen demand (COD) and extractable oil elimination efficiency after different pre-treatments and membrane filtration in case of model wastewaters (DW in distilled water, SW in saline water).

	MF (DW)	MF (SW)	UF	UF+TiO_2	UF+ TiO_2/CNT	Ozone +MF DW	Ozone +MF SW	Fenton + MF	Photo-Fenton +MF
Extractable oil elimination (%)	>99	>99	>99	97	97	99	99	98	98
COD elimination (%)	98	98.3	99	97	97	97.5	93	98	98

In case of real wastewater, the short term ozone treatment alone resulted in 10.6% COD elimination, and 30% elimination of extractable oil content. After filtration (Table 2), the extractable oil elimination efficiencies were lower than in the case of model wastewaters, presumably due to the saline water matrix. The pre-ozonation worsened the oil retention, presumably due to decomposition of oil to smaller molecules. The COD elimination efficiency was very low (around only 20%), which means that in this case the chemical oxygen demand mainly related to small organic and inorganic pollutants, not to oil.

Table 2

COD and extractable oil elimination efficiency after different pre-treatments and membrane filtration in case of real produced water

	UF	**UF+TiO_2**	**UF+TiO_2+CNT**	**Ozone+ UF**	**Ozone+ UF+TiO_2**	**Ozone+ UF+TiO_2+CNT**
Extracable oil elimination (%)	83	98	99	87	95	98
COD elimination (%)	22	25.5	24	24	24.1	23

3.2. Effect of different pre-treatments on permeate flux

Separation performance of the selected membranes was studied in this and earlier studies [23], [24], [18]. A preliminary design can be developed on the basis of flux data by calculating the membrane area. During preliminary experiments the membrane filtration was done until VRR = 5. The measured flux at VRR = 5 (*Tables 3* and *4*) was considered as steady-state flux and used for calculations. Assuming 1,000 $m^3\ d^{-1}$ feed volume, the membrane area was calculated by the following equation:

$$A = \frac{Y \cdot V_F}{J} \tag{3}$$

where A is the filtration area (m^2), Y is the yield of the process, V_F is feed volume ($L\ h^{-1}$), and J is the flux ($L\ m^{-2}\ h^{-1}$) at VRR = 5.

It was found that pretreatments significantly increased the flux. Although the ozone pretreatment significantly increased the flux, both Fenton pretreatment, and membrane surface modification were found to be more effective *(Table 3)*.

Table 3

Steady-state flux and calculated membrane area for filtration of pre-treated model wastewaters

	MF (DW)	**MF (SW)**	**UF**	**UF+TiO_2**	**UF+ TiO_2/CNT**	**Ozone +MF DW**	**Ozone +MF SW**	**Fenton +MF**	**Photo-Fenton +MF**
Flux ($L\ m^{-2} h^{-1}$)	54	46.0	45	150	294	173	85	278	244
Membrane area (m^2)	616	723.0	740	222	113	192	392	119	136

Similar results were obtained in case of real produced water; while ozone treatment slightly decreased the flux, the effect of membrane modification was more expressed. However, in this case – unlike in model wastewaters – the presence of

carbon nanotubes did not enhance the flux, and this effect was observed after pre-ozonation, too. The explanation of this finding requires more experiments, but the salinity of the water may play an important role in the fouling behavior of modified membrane surfaces.

Table 4

Steady-state flux and calculated membrane area for filtration of pre-treated real produced waters

	UF	**UF+TiO_2**	**UF+TiO_2+CNT**	**Ozone+ UF**	**Ozone+ UF+TiO_2**	**Ozone+ UF+TiO_2+CNT**
Flux (L/m^2h)	121.00	326.0	301.00	155.00	362.00	339.00
Membrane area (m^2)	275.04	102.1	110.56	214.71	91.93	98.17.

3.3. Cost evaluation

For the estimation of AOPs cost, Kommineni's method was followed [21]. The costs were divided into categories direct investment costs, indirect costs and operating costs *(Table 5).* Amortization was assumed to be 30 years, 7% discount rate. The volume of ozonation reaction vessel was estimated by the ozone demand (30 g m^{-3}), and the contact time (5 min). On the basis of the feed volume, the volume of the tank is:

$$V=V_F/t = 1000\ m^3\ d^{-1}/5\ min = 3.47 \sim 4\ m^3, \quad (4)$$

The amount of absorbed ozone is 1,440 g/4 m^3/h, which can be provided by 18 × 80 g/h ozone generator. The price of a 80 g/h ozone generator starts from 700 EUR, but it strongly depends on the manufacturer and other specifications; thus the costs calculated are minimum costs. Energy cost calculations were based on the power of ozone generators (680 W/item), 350 working days/year was assumed.

For calculation of Fenton and photo-Fenton treatment costs 120 min contact time was assumed, thus a 3 × 30 m^3 flocculator tank was considered (approx. 10,000 EUR each). The chemical demand was calculated as 100 g m^{-3} $FeSO_4$ and 0,2 Lm^{-3} H_2O_2. For the photo-Fenton process, lamps with 1,300 W power were considered in each reactor, and the UV energy demand was 0.096 kWh m^{-3}.

According to the method previously described by Salehi et al. [20] the total capital investment for the membrane unit includes can be divided into fixed capital and working capital investments. Fixed capital investment contains both direct and indirect costs, according to *Table 5*. The results of calculations are summarized in *Tables 6* and *7*.

Table 5

Cost estimation methodology

	Costs of Advanced Oxidation Processes [21]	**Membrane separation costs [20]**
Direct investment costs	a) Advanced oxidation unit b) Piping, valves, electrical (30% of (a)) c) Site work (10% of (a))	a) Main operating system (membrane systems). b) Installation of main systems (15% of (a)) c) Instrumentation and controls (6% of (a)) d) Electrical (10% of (a)) e) Installation (30% of (a)) f) Buildings, yard and auxiliary (15% of (a)) g) Land (6% of (a))
Indirect costs	Contractor's fees (15% of direct costs Engineering (15% of Direct costs +contractor) Contingency (20% of subtotal)	Engineering and supervision (30% of (a)) Contractor's fees (5% of direct cost) Construction expenses (10% of direct costs) Contingency (8% of fixed capitals)
Operating costs (annual)	Replacement Parts (1.5% of capital cost) Labor (3% of fixed capital) Analytical costs (2% of fixed capital) Chemical costs (based on experiments) Power (0.08 EUR/kWh)	Energy consumption (4% of fixed capital) Maintenance (4% of fixed capital) Operation and performance (2% of fixed capital) Labor (3% of fixed capital) Cleaning (3% of fixed capital)
Total annual cost	Amortization + annual operating costs Amortization: (based on 30-year period, 7% discount rate)	Amortization: lifetime of the polymeric membranes is 2 years. Amortization: (1/30) maintenance cost + (1/15) engineering and supervision + (1/2) membrane system

Table 6

Cost estimation results of combined pretreatment + membrane filtration processes in model wastewaters

	MF (DW)	**MF (SW)**	**UF**	**UF+ TiO_2**	**UF+ TiO_2/ CNT**	**Ozone +MF DW**	**Ozone +MF SW**	**Fenton + MF**	**Photo -Fen-ton+ MF**
Direct costs (investm.) (EUR)	140207	216410	168249	50475	25752	83314	128623	49285	60080
Indirect costs (EUR/year)	21031	24689	25237	7571	3863	29780	36577	17029	21707
Operating costs (EUR/year)	29496	34626	35395	10619	5418	22426	31958	71332	75541
Total annual cost (EUR)	69801	81940	83761	25128	12821	40028	62584	81961	88149
Water treatment cost (EUR/m^3)	**0.199**	**0.223**	**0.239**	**0.072**	**0.037**	**0.114**	**0.179**	**0.234**	**0.252**

The results show that the salinity of the water slightly increases the costs, in spite of this technology aiming only at oil removal (salinity removal requires more complex and expensive technology). The low costs of ozone pretreatment (assuming the cheapest ozone generators in the market) may be surprising, but it can be noted that the price is mainly determined by the cost of filtration, not by the ozone pretreatment, due to the short contact time. As the Fenton processes were considered with longer contact time (120 min), its investment costs were considerably higher than those of the ozone treatment unit. Together with its relatively high operating costs this resulted in higher total costs. The membrane surface modification presents the lowest investment cost with low operational costs, resulting in the most beneficial cost/m^3 wastewater. Although this technology would be very beneficial, it should be noted that these types of modified membranes are not available on the market at this time. Nevertheless, our results show that due to their good antifouling properties, the development of modified membrane surfaces may lead to a decrease in membrane filtration costs. Accordingly, the development of these kinds of commercial membranes is of great interest, and their widespread application is expected in several industrial activities.

In case of real wastewater *(Table 7)*, due to the lower oil content of the produced water, the cost of oil elimination decreased further. Although in terms of flux, the ozone treatment before filtration through modified membranes was beneficial – preozonation enhanced the flux – in terms of costs this cannot be stated. The investment and operational costs of ozone treatment unit cannot be compensated by gains in smaller membrane area.

Table 7

Cost estimation results of combined pre-treatment + membrane filtration processes in real wastewaters

	UF	**UF+TiO_2**	**UF+TiO_2+CNT**	**Ozone+ UF**	**Ozone+ UF+TiO_2**	**Ozone+ UF+TiO_2+CNT**
Direct costs (investments) (EUR)	62572	23225	25153	88396	60465	62575
Indirect costs (EUR/year)	9386	3484	3773	7327	26353	23622
Operating costs (without amortization) (EUR/year)	13163	4886	5292	23495	17619	17989
Total annual cost (EUR)	31151	11562	12522	42558	28653	29518
Water treatment cost (EUR/m^3)	0.089	0.033	0.036	0.122	0.082	0.084

4. CONCLUSIONS

The current study deals with a comparison of the cost of oily wastewater treatment by membrane filtration associated with different pretreatment methods. The cost

evaluation was completed for model oily wastewaters containing 100 ppm crude oil in saline or distilled water, and in some cases for real produced water, aiming at elimination of oil content. Three different pre-treatment methods (ozonation, Fenton and photo-Fenton pretreatments) and the modification of the used membrane surfaces with photocatalytic nanoparticles were used for the mitigation of fouling and enhancing the oil elimination efficiency during the process. The cost estimation was evaluated on the basis of flux decline and the cost of pretreatments. Results showed that pre-oxidation enhances the flux, and thus may decrease the cost of combined treatment.

Comparison of model and real wastewater's costs allows us to conclude that the cost depends on pollutant concentration – a lower oil concentration decreases the cost.

Analyzing the costs of AOPs and membrane separation, it appears that the capital cost of the ozone treatment unit may be reasonable, as the price is mainly determined by the cost of filtration rather than by the ozone pretreatment, due to the short contact time.

The membrane surface modification showed the lowest investment cost with low operational costs, resulting in the most beneficial cost/m^3 treated wastewater; due to their good antifouling properties, the development of modified membrane surfaces may lead to decreasing of membrane filtration costs, so there will be great interest in the development and the widespread application of these kind of commercial membranes in the near future, due to the continuously tightening emission limits.

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