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Economic analysis of microaerobic removal of H_2S from biogas in full-scale sludge digesters

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HIGHLIGHTS

- The cost-effectiveness of the microaerobic removal of H_2S from biogas was studied.
- Operating costs between 0.0019 and 0.0039 €/m^3 of biogas were estimated.
- The supply of concentrated O_2 was the most profitable treatment alternative.

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ABSTRACT

The application of microaerobic conditions during sludge digestion has been proven to be an efficient method for H_2S removal from biogas. In this study, three microaerobic treatments were considered as an alternative to the technique of biogas desulfurization applied ($FeCl_3$ dosing to the digesters) in a WWTP comprising three full-scale anaerobic reactors treating sewage sludge, depending on the reactant: pure O_2 from cryogenic tanks, concentrated O_2 from PSA generators, and air. These alternatives were compared in terms of net present value (*NPV*) with a fourth scenario consisting in the utilization of iron-sponge-bed filter inoculated with thiobacteria. The analysis revealed that the most profitable alternative to $FeCl_3$ addition was the injection of concentrated O_2 (0.0019 €/m^3 biogas), and this scenario presented the highest robustness towards variations in the price of $FeCl_3$, electricity, and in the H_2S concentration.

1. Introduction

Anaerobic digestion (AD) is widely employed in wastewater treatment plants (WWTPs) in order to reduce excess sludge production from the activated sludge process, and generate an energy-rich gas called biogas. Among other applications, biogas can be burnt in combined heat and power (CHP) generation plants for the production of heat and electricity by combustion engines. Therefore, AD can improve the economic balance of WWTPs (Appels et al., 2008).

Besides CH_4 and CO_2, which are its main components, biogas also contains several trace compounds that must be removed before combustion. H_2S is the most relevant of these pollutants, which is produced during sludge digestion as a result of reduction and fermentation of inorganic and organic S-containing compounds (respectively) present in both primary and secondary sludge (Lens and Hulshoff Pol, 2000). It is a toxic malodorous compound, highly corrosive to many types of steel, and considerably shortens the lifetime of the installations for biogas utilization. Therefore, for trouble-free operation of CHP generation plants, H_2S concentrations of less than 100–500 mg/N m^3 must be maintained (Deublein and Steinhauser, 2008).

Several techniques can be applied in order to achieve the required H_2S concentration in biogas, including physical, chemical, or biological, or combinations of these three alternative methods. Despite their high chemical requirements, energy and disposal cost, physicochemical technologies are widely implemented at industrial scale due to extensive experience in design and operation. Biological processes have the potential to overcome some or all the disadvantages of physicochemical technologies (Syed et al., 2006). In fact, there are several technical solutions offered commercially worldwide for biogas desulfurization involving biological removal mechanisms, such as the combination of chemical scrubbing with a bioreactor (http://en.paques.nl/products/featured/thiopaq), or the iron-sponge-bed filter inoculated with

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thiobacteria, which combines chemical and biological oxidation (http://odorfilter.com/hydrogen-sulfide-removal/technologies/h2splus-systems.html).

In the particular case of sludge digestion, the addition of ferric salts directly to the digester in order to precipitate insoluble FeS is a widespread method for biogas desulfurization. This is due to its simplicity, since the H_2S removal takes place directly in the anaerobic reactor, while the biogas recirculation employed for sludge mixing favors the mechanism of biogas desulfurization itself (namely, the displacement of the H_2S equilibrium from the gas to the liquid phase). Therefore, only a vessel for the storage of the salt solution and a pump are required. However, this technique presents several drawbacks, such as accumulation of FeS in the digester, increased amount of iron in the treated sludge, possible re-oxidation of FeS if aerobic conditions are reached, and high cost due to the price of the reactant and the existence of parallel reactions (Devai and Delaune, 2002; Gutierrez et al., 2010; Speece, 2008). The application of other physicochemical methods in industrial sludge digesters is limited due to the scale (for instance, iron-sponges are only suitable for small/medium productions) or the high chemical consumption and the consequent operating cost (as in the case of scrubbing) (Abatzoglou and Boivin, 2009).

An alternative for the desulfurization of biogas directly in the reactor consists in injecting limited amounts of O_2, namely, imposing microaerobic conditions during digestion. Thus, most of the H_2S is chemically and biologically oxidized to S^0 (Kleinjan, 2005). This is possible because sulfide-oxidizing bacteria (SOB) are present in the sludge and proliferate in the digester (Weiland, 2010).

Fdz-Polanco et al. (2009) achieved efficiencies of H_2S removal higher than 99% in sewage sludge digesters (200 L) by injecting limited amounts of pure O_2, thereby maintaining H_2S concentrations of 50–500 mg/N m^3, while the biogas production, CH_4 yield and volatile solids (VS) removal remained stable. Moreover, they reported dissolved sulfide removal when biogas recirculation was implemented, which was indeed consistent with Díaz et al. (2011). Under equivalent oxygenation rates, Díaz et al. (2010) reached similar efficiencies of H_2S removal with both air and pure O_2. However, they reported a significant decrease in the biogas CH_4 content with micro-aeration, due to the presence of N_2. This dilution effect was also reported by Jenicek et al. (2008, 2010), who achieved removal efficiencies higher than 99% with limited air supply to full-scale digesters (1500 and 2100 m^3) treating sewage sludge, while the efficiency of VSS degradation was improved, and the soluble COD, ammonia and phosphate concentration in the digested sludge was reduced.

During their lab-pilot-scale investigation, Díaz et al. (2011) found that, under microaerobic conditions, S^0 was deposited all over the gas space of the sewage sludge digester, where different SOB developed, independently of the dosing point of the oxidative reactant (headspace or liquid phase) and the mixing method (sludge or biogas recirculation). Accordingly, the main drawback of the microaerobic treatment for gaseous sulfide control is the accumulation of S^0 in the gas space, which reduces the volume of this area and could cause clogging problems, thereby shortening the cleaning interval of digesters. Conversely, after 240 d with injection of concentrated O_2, Ramos et al. (2014) observed that scarce amount of S^0 was deposited in the headspace of an industrial-pilot digester treating sewage sludge, despite this S-compound was found to be the main by-product of sulfide oxidation. Nonetheless, photos of the gas space of full-scale agricultural reactors suggest that the accumulation of S^0 in this area could effectively be a problem during microaerobic digestion at industrial scale (Pérez et al., 2012).

Despite its promising nature, microaerobic digestion has not been evaluated in economic terms, which still hinders the widespread application of this technology for H_2S control. The present study analyzes the economic viability of the microaerobic control of gaseous sulfide in a WWTP comprising three industrial anaerobic reactors (5000 m^3 each), where the existing method of biogas desulfurization consisted in $FeCl_3$ addition. Three different microaerobic scenarios are proposed as an alternative to ferric salt dosing, consisting in the injection of pure O_2 from cryogenic tanks, concentrated O_2 from pressure swing adsorption (PSA) generators, and air. A fourth scenario consisting in the utilization of iron-sponge-bed filter inoculated with thiobacteria is also considered. Furthermore, a sensitivity analysis is carried out in order to evaluate the robustness of the different scenarios towards variations in four of the most influential economic and operational parameters (price of $FeCl_3$ and electricity, H_2S concentration, and cleaning interval of digesters).

2. Methods

2.1. Plant description

The municipal WWTP object of this study was located at Villalonquéjar (Burgos, Spain) and comprised three anaerobic digesters with a working volume of 5000 m^3 treating sewage sludge (750 m^3/d). The reactors were operated at 35 °C and 20 d of hydraulic retention time (HRT), and mixed by both sludge and biogas recirculation. The biogas produced in the plant usually presented a H_2S concentration of approximately 2300 mg/N m^3, (which was controlled by $FeCl_3$ addition to the raw sludge), and was fed to a combustion engine for electricity generation. Data contained in Table 1 (specifically, VS influent and removal, and biogas production) were estimated from the results of Ramos et al. (2014), which were obtained during the digestion of the sewage sludge treated in the WWTP object of the present study.

2.2. Scenarios

As shown in Table 1, four scenarios were considered as an alternative to that existing in the WWTP (the FSA scenario). In all of them, the objective was to achieve a H_2S concentration of 100–500 mg/N m^3. Three different microaerobic scenarios were proposed, depending on the oxidative reactant: O_2 (MA_{O2}), concentrated O_2 ($MA_{95\%O2}$), and air (MA_{air}). In order to make a comparison with another novel technology for biogas desulfurization, a fourth scenario involving the utilization of iron-sponge-bed filter inoculated with thiobacteria was considered (ISF).

2.2.1. FSA scenario

In the FSA scenario, the dosage of $FeCl_3$ (230 €/t) was set at 1.8 mol Fe supplied/mol S removed, which equal the double of the minimal molar ratio of Fe to S required to achieve efficient sulfide removal in sewer systems (Firer et al., 2008). It was calculated by considering that the biogas sulfide content (2300 mg/N m^3) decreased to 150 mg/N m^3, and assuming that dissolved sulfide was reduced from 50 mg/L (Ramos et al., 2014) to 0.2 mg/L, which is the lowest level usually achieved by this technique (Haaning Nielsen et al., 2005; Speece, 2008). The dosage represented 760 g $FeCl_3$/m^3 raw sludge, or equivalently, 0.017 kg $FeCl_3$/kg total solids influent.

2.2.2. MA scenarios

MA_{O2} involved the utilization of cryogenic tanks, which were leased. Since the O_2 consumption was higher than 1000 N m^3/month, this alternative implied lower cost of reactant (0.32 €/kg) than the utilization of gas cylinders. $MA_{95\%O2}$ involved the utilization of concentrated O_2 (air with an O_2 concentration

Table 1
Technique of H_2S removal and characteristics of AD and in the different scenarios for a WWTP comprising three anaerobic digesters (5000 m^3 each) treating sewage sludge at 20 d of HRT.

	FSA	ISF	MA_{O2}	$MA_{95\%O2}$	MA_{air}
Method of H_2S removal	$FeCl_3$ addition to the raw sludge	Iron-sponge-bed filter inoculated with thiobacteria	Supply of pure O_2 to the reactors	Supply of concentrated O_2 to the reactors	Air supply to the reactors
Reactive	$FeCl_3$	Fe_2O_3 and air	O_2	O_2 (>95%v.)	Air
Rate of reactive per N m^3 of biogas	44 g	N/A	5.7 NL[a]	6.0 NL[a]	27.2 NL[a]
VS influent (kg/m^3)[a]	33	33	33	33	33
VS removal (%)[a]	47	47	47	47	47
Biogas production (N m^3/d)[b]	13,100	13,100	13,100	13,100	13,400
CH_4 concentration (%v.)[a]	62	62	62	62	60
Engine efficiency (%)	35.00	35.00	35.00	35.00	34.84[c]

[a] Data obtained from Ramos et al. (2014).
[b] Estimated from Ramos et al. (2014).
[c] Estimated from Porpatham et al. (2008).

higher than 95%v.) produced from air by an industrial PSA generator (1.2 kW h/m^3 O_2). Thus, negligible amounts of N_2 were introduced in the digester (Ramos et al., 2014). MA_{air} implied the injection of air into the digesters by compressors (1.4 kW h). As a result, the CH_4 concentration in biogas was 2%v. lower than in MA_{O2} and $MA_{95\%O2}$ (Table 1), as reported by Díaz et al. (2010). The consequent loss of efficiency of the engine (0.16%) was estimated according to Porpatham et al. (2008), who reported a decrease of 0.9% in the efficiency of the spark-ignition engine when the CH_4 concentration was reduced from 70 to 59%v. (within this range, a linear behavior was assumed).

In all the MA scenarios, microaerobic conditions were applied by injecting the oxidative reactant into the headspace of the digesters. The results of Díaz et al. (2011) and Ramos et al. (2014) indicate that this is the optimal dosing point. The dose of every reactant was obtained from Ramos et al. (2014) and increased by 20% (Table 1), assuming that less efficient mass transfer conditions were maintained at industrial scale. In order to avoid that the residual oxygen concentration reaches the limiting oxygen concentration (LOC), 12%v. for CH_4 and 7.5%v. for H_2S (Crowl, 2012), that may lead to a fire or explosion, an online biogas monitoring system and automated valve actuators were included in all the microaerobic scenarios (safety section) so that residual oxygen concentration remains below 1%v.

The implementation of microaerobic conditions in the WWTP involved legal cost, which was assumed to be 3% of the total equipment cost, according to the recommendations of the technical manager of the plant. It also implied building cost, since each digester remained empty for 20 d while the required modifications were carried out. Once one digester was emptied, its content was thickened with centrifuges (3.5 €/m^3) and inertized with N_2 (0.32 €/kg), and, for this purpose, the addition of polyelectrolyte (2.5 €/t TS) was required. Final disposal cost of sludge was 10 €/t TS. After 20 d, the content of one of the operative reactors was transferred to the empty one. Hence, building required a total of 60 d, during which digestion took place at 2/3 of the conventional HRT. Under these conditions, VS removal and CH_4 production was calculated with the following equations (Eqs. 1–3) (Metcalf and Eddy, 2003):

$$VS_{removal} = 13.7X \ln(HRT) + 18.9 \quad (1)$$

$$P_x = YES_0/(1 + k_d\theta_c) \quad (2)$$

$$V = 0.35\ m^3/kg\ \{[ES_0] - 1.42(Px)\} \quad (3)$$

where $VS_{removal}$ is the VS destruction (%), P_x the net mass of the cells produced (kg/d), Y the yield coefficient (g/g) (for municipal sludge, a value of 0.1 mg VS/mg BOD was assumed), E the efficiency of waste utilization (0.6), S_0 the ultimate BOD of the influent sludge (kg/d), k_d the endogenous coefficient (0.04 d^{-1}), Θ_c the mean cell residence time (d) equal to the HRT, V the volume of CH_4 produced (m^3/d), 0.35 the theoretical conversion factor for the amount of CH_4 produced from the conversion of 1 kg of BOD, and 1.42 the conversion factor for cellular material into BOD.

A technical stoppage and reactors emptying for maintenance (cleaning) was carried out in all the MA alternatives after 12 yr of operation. The cleaning interval in real WWTP is of 10–12 yr, independently of the technique of H_2S control. Nevertheless, when microaerobic scenarios are applied, S^0 partly accumulates in the digesters and disposal (26 €/t) was considered.

2.2.3. ISF scenario

Fixed cost of this alternative was estimated by Eq. (4) (Green and Perry, 2008), considering that the desulfurization of 1700 m^3/h of biogas in an iron-sponge-bed filter inoculated with thiobacteria involved a capital cost of 450,000 $ (Abatzoglou and Boivin, 2009):

$$C_2 = C_1(q_2/q_1)^k \quad (4)$$

where C_1 is the capital cost for a plant with a capacity of treatment q_1, C_2 the cost for a plant with a capacity q_2, and k is the scale factor, which was assumed to be 0.7, since this is a typical value for WWTPs in many European countries (Maurer, 2009).

The operating cost included only the expense on bed material, which was calculated on the web of the company selling this technology (http://odorfilter.com/resources/h2splus-calculator.html). It should be highlighted that, in contrast to the rest of the scenarios, maintenance in ISF (periodic replacement of the media) did not interfere with the digestion process.

2.3. Profitability analysis

The profitability analysis was based on cash flows, and carried out in terms of net present value (*NPV*) and internal return rate (*IRR*), according to Chauvel et al. (2003) (Eq. (5)):

$$NPV = A_{CF_0} + \sum_{1}^{n} \frac{A_{CF_n}}{(1+i)^n (1+i_i)^n} \quad (5)$$

where A_{CF0} is the initial investment, A_{CFn} the annual cash flow for the n year, i the discount rate, which was assumed to be 10%, and i_i the inflation rate (3%). *NPV* was evaluated for a period of 20 yr, considering a loan of the capital cost with an interest of 10% (4 yr). *IRR* was calculated as the required value of i to obtain *NPV* = 0 for 5 yr.

For the FSA and MA scenarios, cash flows were calculated by estimating the operational and fixed costs. Market prices of equipment, reactants and services were compiled from Spanish suppliers (January 2015). Maintenance cost was assumed to be 10% of the cost of the equipment (Ulrich and Vaasudevan, 2004). Cash flows in ISF were estimated from the data reported by Abatzoglou and Boivin (2009).

2.4. Sensitivity analysis

Four parameters were considered for the sensitivity analysis: price of $FeCl_3$ (1) and electricity (2), H_2S concentration in biogas (3) and cleaning interval (4). The FSA, MA, and ISF scenarios were compared in terms of *NPV* at 5 and 20 yr (*NPV*5 and *NPV*20, respectively) taking into account a decrease of 25% in the $FeCl_3$ price (from 230 to 173 €/t) (a), an increase of 25% in the electricity price (from 0.12 to 0.15 €/kW h) (b), and an increase of 100% in the H_2S concentration in biogas (from 2300 to 4612 mg/N m^3) (c). Variation (a) was selected in order to take into account the volatility of the price of $FeCl_3$, which strongly depends on the amount of available iron scrap habitually employed for the production of this chemical. Variation (b) was proposed in order to consider the geographical variability in the price of this commodity, and (c) was included due to the fact that Ramos et al. (2014) observed fluctuations in the H_2S concentration in biogas of approximately 100%. Finally, the *NPV* of the most profitable MA scenario was evaluated for a period of 20 yr, assuming that maintenance was performed every 6 and 8 yr. 40% of the S^0 production was considered to accumulate in the digesters (Díaz et al., 2011). Thus, the effect of the cleaning interval on the profitability of this alternative was studied.

3. Results and discussion

3.1. Fixed cost

A detailed list of all the fixed costs (including those related to equipment, building and legal aspects) arising from the implementation of each technique proposed as an alternative to the solution of H_2S control applied in the WWTP object of this study (the FSA scenario) is shown in Table 2. The total fixed cost in the ISF alternative (85,000 €) was lower than in any of the MA scenarios (111,497–132,931 €). This was due to the cost resulting from the digesters emptying, installation and modifications required during the building period in the MA alternatives, which was of 94,876 € (75–91% of the total fixed cost). The main expense during the building period was that resulting from the loss of CH_4 generation (40,464 €), which represented 43% of the building cost. Regarding the legal cost arising from the implementation of microaerobic conditions, since it was assumed to be 3% of the equipment cost, it represented less than 1% of the total fixed cost (484–1108 €).

Table 2 lists the costs related to the equipment, including that resulting from the acquisition of the source of the specific oxidative reactant. Among the MA alternatives, MA_{O2} implied the lowest outlay for equipment (13,136 €), due to the fact that the source of O_2 (cryogenic tanks) was leased. As a result, since the building cost was the same, and the legal cost was negligible in all the MA scenarios, MA_{O2} involved the lowest total fixed cost (111,497 €). Conversely, $MA_{95\%O2}$ required the highest outlay for equipment (36,946 €), since the purchase of the PSA generators involved an expense of 25,476 €. As a result, among the MA alternatives, $MA_{95\%O2}$ implied the highest total fixed cost (132,931 €). Regarding MA_{air}, it required an investment of 11,180 € in gas compressors, and a total outlay for equipment of 25,472 €, which resulted in a medium fixed cost (121,113 €).

Table 2
Fixed cost (€) in the different scenarios.

	FSA	ISF	MA_{O2}	$MA_{95\%O2}$	MA_{air}
Equipment	N/A	N/A			
Reactant source			0	25,476	11,180
Flow control			4816	0	1920
Safety			10,520	10,670	11,572
Valves			300	300	300
Piping			500	500	500
Subtotal			13,136	36,946	25,472
Building	N/A	N/A			
Digester emptying					
Centrifuges (including labor)			21,850	21,850	21,850
Inertization (including labor)			6000	6000	6000
Polyelectrolyte			4375	4375	4375
Sludge disposal			11,500	11,500	11,500
Electricity[a]			7188	7188	7188
CH_4 loss			40,464	40,464	40,464
Installation and modifications			3500	3500	3500
Subtotal			94,876	94,876	94,876
Legal	N/A	N/A	484	1108	764
Total fixed cost	**0**	**85,000**[b]	**111,497**	**132,931**	**121,113**

[a] 0.12 €/kW h.
[b] Scaled with data from Abatzoglou and Boivin (2009).

From the data reported by Abatzoglou and Boivin (2009) and the application of Eq. (4), the investment cost in vessels for the SulfaTreat® and the Media-G2® processes (both based on pure chemical removal in iron oxide absorbents), and for H_2S removal with impregnated activated carbon was estimated at about 50,000, 82,000 and 50,000 €, respectively. Obviously, any other specific equipment required to implement the aforementioned technologies would increase the investment cost. Kazemi et al. (2014) performed a simulation in order to calculate the cost arising from the implementation of four different commercial alternatives for H_2S removal in a plant treating sour in Iran: the mixed amine, Sulfinol-M, Chelated iron (LO CAT), and Shell Paques processes. By applying Eq. (4), the investment cost for the application of these technologies in the WWTP object of this study was estimated to exceed 56,000 €. Nevertheless, it must be taken into account that the capital outlay for the implementation of any of the aforementioned processes in Spain would be higher than in Iran. Therefore, independently of the oxidative reactant used, the microaerobic treatment would outcompete any other commercial alternative for H_2S removal from biogas in terms of fixed cost in a new WWTP, where building cost would not exist (Table 2).

3.2. Operating cost

The expenditures arising during the operation of the different technologies considered in this study (including those on reactant, installations, electricity and maintenance, and the loss of power generation in the combustion engine) are listed in Table 3. The method of biogas desulfurization applied in the WWTP involved the highest operating cost, requiring an annual outlay of 48,111 €. It must be mentioned that this figure resulted from the utilization of $FeCl_3$ mainly, since this chemical involved a cost of 47,196 €/yr. Therefore, the expense involved in the desulfurization of the biogas produced in the WWTP would be 2–5 times lower if the existing method of H_2S removal (the FSA scenario) were replaced by any of the alternatives proposed. Specifically, the treatment cost in the ISF and MA scenarios ranged between 0.0049 and 0.0019 €/m^3 of biogas treated, against 0.010 €/m^3 in FSA.

As shown in Table 3, all the MA alternatives presented lower operating cost that the ISF scenario. $MA_{95\%O2}$ was the most

Table 3
Operating cost (€/yr) in the different scenarios.

	FSA	ISF	MA_{O2}	$MA_{95\%O2}$	MA_{air}
Reactive	47,196	N/A	12,856	0	0
Installations (rent)	0	N/A	3600	0	0
Electricity[a]	415	N/A	207	5406	5276
Maintenance	500	N/A	1614	3695	2547
Engine loss	0	N/A	0	0	5083
Total operating cost	**48,111**	**23,322**[b]	**18,277**	**9101**	**12,906**

[a] 0.12 €/kW h.
[b] Data from http://odorfilter.com/resources/h2splus-calculator.html.

economical alternative, since concentrated O_2 was obtained from a costless substance (air). Although air was also used as the source of O_2 in MA_{air}, this scenario implied higher operating cost than $MA_{95\%O2}$ (12,906 against 9101 €/yr, or equivalently, 0.0027 against 0.0019 €/m^3 biogas) as a result of the loss of power generation during the combustion of biogas. This loss was due to the fact that air was introduced directly to the digesters, thereby reducing the energy content (CH_4 concentration) of the biogas. The operating cost in MA_{O2} was the highest of the MA alternatives, as a result of the utilization of pure O_2, which implied an annual outlay of 12,856 €, and the fact that the cryogenic tanks were leased (involving a cost of 3600 €/yr). Nevertheless, and although the operating cost in ISF included only the expense on the iron-sponge, the operating cost of MA_{O2} was lower than that of the ISF scenario (18,277 against 23,322 €/yr, or equivalently, 0.0039 against 0.0049 €/m^3 biogas).

The operating cost of the different alternatives for H_2S removal considered by Kazemi et al. (2014) and Abatzoglou and Boivin (2009) was estimated with Eq. (4). It ranged between 14,000 (for the LO-CAT technology) and 94,000 € (utilization of impregnated activated carbon). Therefore, in the WWTP object of this study, the operating cost in the $MA_{95\%O2}$ and MA_{air} scenarios would be lower than that involved in the biogas desulfurization by the implementation of the SulfaTreat®, Media-G2®, mixed amine, Sulfinol-M, LO CAT, and Shell Paques processes, as well as by the utilization of impregnated activated carbon (Table 3).

3.3. Profitability analysis

In Fig. 1, the cost-effectiveness of the different scenarios proposed as an alternative to the method of H_2S control applied in the WWTP object of the present study is compared in terms of *NPV*. As shown in Fig. 1a, the short-term profitability of $MA_{95\%O2}$ and MA_{air} was really similar (*NPV*5 ≈ 10,000 €) and slightly higher than that of ISF, which presented a *NPV*5 of 7000 € while *NPV*5 of MA_{O2} was negative. $MA_{95\%O2}$, MA_{air} and ISF scenarios also presented similar *IRRs* around 25% (Fig. 1b). However, the *NPV*5 of $MA_{95\%O2}$ was higher than that of MA_{air} and ISF at low discount rates. Therefore, in the short term, the microaerobic treatment consisting in supplying concentrated O_2 to the reactors was proved to be the most profitable alternative to $FeCl_3$ addition in the WWTP.

Noticeable differences in terms of *NPV*20 were observed between all the scenarios as a result of the dissimilarities in the operating costs (see Section 3.2), as illustrated in Fig. 1a. $MA_{95\%O2}$ and ISF presented the highest and the lowest *NPV*20, respectively. Nevertheless, it should be highlighted that the replacement of the technique of biogas desulfurization implemented in the WWTP with that proposed in the ISF scenario would also bring economical benefits in the long term. Hence, $MA_{95\%O2}$ proved to be the most profitable alternative to $FeCl_3$ dosing in both the short and long term, and should be the preferred investment in the WWTP.

Despite the highest initial outlay required to acquire the PSA generators in $MA_{95\%O2}$ (Table 2), the profitability of this scenario was higher than that of MA_{air} as a result of the loss of power generation due to the dilution of biogas in MA_{air} (Table 3). As noted in Section 2.2.2, this loss (0.16%) was estimated from the experimental data reported by Porpatham et al. (2008). However, if the loss of efficiency of the combustion engine were less than 0.10%, the injection of air into the digesters would be the most profitable alternative solution to replace $FeCl_3$ addition in the WWTP.

Obviously, any other fixed cost not considered in the present analysis, such as freight, site preparation, start-up or contingencies, would reduce the profitability of the scenarios in the short term (namely, the *NPV*5). However, in the long term (*NPV*20), the introduction of air and concentrated O_2 to the reactors would outcompete the addition of $FeCl_3$ due to the high operating cost of this last technique, which, as shown in Table 3, results from the price of the salt.

3.4. Sensitivity analysis

3.4.1. Sensitivity of $FeCl_3$ and electricity prices, and H_2S concentration on the NPV5

The variations in the economic and operational parameters evaluated (price of $FeCl_3$ and electricity, and H_2S concentration) severely impacted the profitability of the scenarios in the short term, as shown in Fig. 2a. Specifically, when a 25% discount on the price of $FeCl_3$ was considered, all the scenarios showed negative *NPV*5. Due to the fact that the outlay for $FeCl_3$ implied approximately 98% of the operating cost in the baseline FSA scenario (Table 3), the operating cost of this alternative was reduced by almost 25%. Therefore, in the event of a reduction in the price of $FeCl_3$, the implementation of any of the others methods of biogas desulfurization proposed (microaerobic treatment or the iron-sponge) would involve a longer payback period than in the baseline conditions.

The $MA_{95\%O2}$ and MA_{air} scenarios showed the best behavior when the sensitivity of an increase of 25% and 100% in the electricity price and the concentration of H_2S in biogas (respectively) was evaluated, which is illustrated in Fig. 2a. When the aforementioned variations were considered, $MA_{95\%O2}$ and MA_{air} presented a payback period shorter than 5 yr (as in the baseline conditions), whereas MA_{O2} showed negative *NPV*5. In fact, according to Fig. 2a, MA_{O2} was the most sensitive scenario, since its short-term response to all the variations considered in this analysis was always a negative *NPV*5. Regarding ISF, in the event of an increase of 25% in the electricity price, this alternative would still show positive *NPV*5. Conversely, in case of an increase of 100% in the H_2S concentration, the ISF scenario would be an unprofitable alternative in the short term.

3.4.2. Sensitivity of $FeCl_3$ and electricity prices, and H_2S concentration on the NPV20

The profitability of the scenarios was also severely impacted in the long term by the 25% discount and raise in the price of $FeCl_3$ and electricity (respectively), and the increase of 100% in the H_2S concentration. However, in contrast to the short-term profitability (Fig. 2a), the *NPV*20 of all the scenarios taking into account the aforementioned variations was positive, with the only exception of that of ISF when the increase in concentration of H_2S in biogas was considered (Fig. 2b).

As shown in Fig. 2b, the raise in the electricity price reduced slightly the *NPV*20 of all the scenarios. Specifically, this parameter caused the lowest influence on the long-term profitability of all the treatment alternatives considered. Conversely, the reduction in the $FeCl_3$ price implied a decrease of 57–88% in the *NPV*20. $MA_{95\%O2}$ and MA_{air} proved to be the most robust scenarios against this variation, despite their *NPV*20 was reduced by more than 50%. MA_{O2} and ISF presented an *NPV*20 around 85% lower than under the baseline conditions.

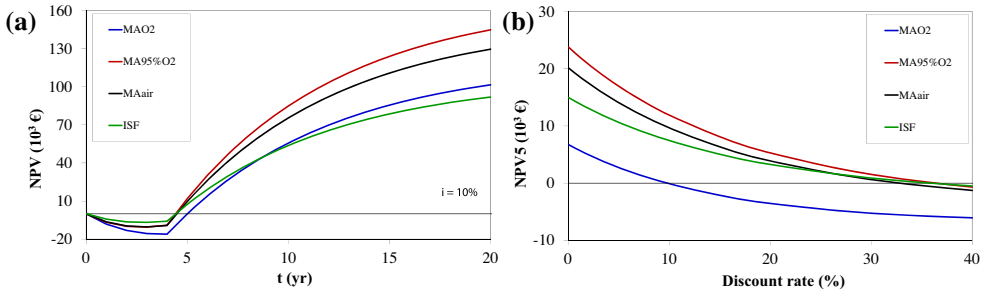

Fig. 1. Comparison of the profitability of the different alternatives proposed in this study for the replacement of the existing scenario (FSA) in terms of (a) *NPV*, considering a discount rate of 10% for a period of 20 years, and (b) *NPV*5, for different discount rates.

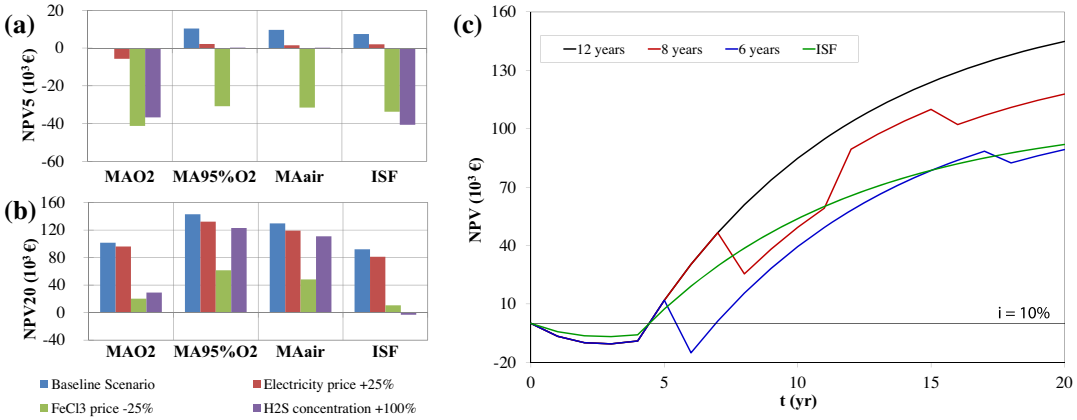

Fig. 2. Sensitivity of an increase of 25% and 100% in the electricity price and the H_2S concentration in biogas (respectively), and a decrease of 25% in the $FeCl_3$ price on the (a) *NPV*5 and (b) *NPV*20 of the different scenarios proposed in the study. (c) Sensitivity of the cleaning interval on the *NPV* of the $MA_{95\%O2}$ scenario, and comparison with the ISF alternative.

When an increase of 100% in the H_2S concentration in biogas was considered, the *NPV*20 of the MA_{O2} scenario decreased drastically (specifically, it was 71% lower than in the baseline scenario), while that of ISF became negative (Fig. 2b). This was the result of the higher operating cost arising from an increased demand of reactant (in the MA_{O2} alternative) and a lower periodicity of replacement of the bed material (in ISF). Conversely, the *NPV*20 of $MA_{95\%O2}$ and MA_{air} was only reduced by approximately 15%, since no media replacement was required, and the utilization of air as the source of O_2 avoided any additional expense on reactant.

3.4.3. Sensitivity of the interval of headspace cleaning

Obviously, the potential accumulation of S^0 in the headspace of the digesters would affect the profitability of the microaerobic treatment, independently of the oxidative reactant used. As noted in Section 2.2.2, in all the MA scenarios, the reactors were stopped and emptied for cleaning every 12 yr. Therefore, shorter periods of maintenance were considered for $MA_{95\%O2}$, which proved to be the most profitable MA alternative (see Section 3.3). Fig. 2c presents the *NPV* of ISF and $MA_{95\%O2}$ for a period of 20 yr, considering a cleaning interval of 12 (the baseline), 8 and 6 yr in the aforementioned MA scenario.

S^0 accumulation does not alter significantly the costs of maintenance if performed every 12 years. The estimation of S^0 accumulation during 12 years is around 90*t* and 2600 € for disposal, less than 3% of the total maintenance costs in the most adverse scenario (longer cleaning interval and larger accumulation). Besides, S^0 is a valuable product for agriculture or wastes composting (Roig et al., 2004) and disposal might be avoided since recovered sulfur from sludge digesters presented a high purity (86%w.) in pilot-plant studies (Ramos et al., 2014).

The reduction of the cleaning interval from 12 to 8 yr reduced considerably the profit resulting from the replacement of the technique of H_2S control implemented in the WWTP with the $MA_{95\%O2}$ alternative, as demonstrated in Fig. 2c. However, the payback period in $MA_{95\%O2}$ with a periodicity of maintenance of 8 yr was similar than in the baseline $MA_{95\%O2}$ scenario. Moreover, the *NPV*5 and *NPV*20 in the $MA_{95\%O2}$ alternative with a cleaning interval of 8 yr were both higher than in ISF. Hence, the injection of concentrated O_2 into the digesters would be the most profitable alternative solution to $FeCl_3$ addition if maintenance was required every 8 yr (or longer).

As expected, a cleaning interval of 6 yr would imply a longer amortization period of the $MA_{95\%O2}$ alternative. As shown in Fig. 2c, the *NPV* at 10 and 20 yr in $MA_{95\%O2}$ with a periodicity of maintenance of 6 yr and in ISF were both similar (respectively). Therefore, the cleaning interval in the $MA_{95\%O2}$ scenario should be higher than 6 yr in order to compete effectively with the ISF scenario. In this regard, it should be noted that Pérez et al. (2012) reported that no operation or maintenance issue arose from the application of micro-aeration in full-scale reactors treating crops after up to 7 yr. This observation is indeed consistent with the results of Ramos et al. (2014). Accordingly, the introduction of concentrated O_2 in the digesters could effectively be more profitable than the utilization of the iron-sponge-bed filter in order to replace the method of H_2S control implemented in the WWTP.

4. Conclusions

The utilization of concentrated O_2 (>95%v.) produced from air by PSA generators ($MA_{95\%O2}$) was the most profitable alternative to addition of ferric salt for the removal of H_2S from the biogas

produced in the WWTP. The sensitivity analysis revealed that, among all the scenarios proposed, $MA_{95\%O2}$ was the most robust against the variations in the price of $FeCl_3$ and electricity, and the H_2S concentration, in both the short and long term. Additionally, $MA_{95\%O2}$ would still be the preferred alternative if a cleaning interval of 8 years were required.

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