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Techno-economic analysis of bio-methane production from agriculture and food industry waste

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Abstract

Bio-methane production via anaerobic digestion is a promising technology for the decarbonization of the energy system. Bio-gas obtained from anaerobic digestion of farm and food industry waste is largely composed of 60% CH₄ and 40% CO₂. For injection of bio-methane into the gas distribution network it is necessary to remove CO₂ from the biogas so that a richer CH₄ stream is injected to satisfy gas network requirements. Chemical separation processes using solvents that react with CO₂ or physically processes using adsorbents or membranes in which CO₂ is retained are currently under investigation to reduce associated energy consumption whilst maximizing CO₂ removal. In the case of sorbent based processes, research is mainly focused on the optimal design of pressure swing adsorption (PSA) cycles. In this work, a comparative techno-economic study of bio-gas upgrading for bio-methane production using solvent based processes and pressure swing adsorption cycles is presented. The results show that, pressure swing adsorption cycles exhibit 37% lower capital costs and 10% lower average life-time costs compared to solvent based technologies.

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Keywords: bio-gas, anaerobic digestion, pressure swing adsorption, solvent based units

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Nomenclature

AD	Anaerobic Digestion
ALTCBM	Average life-time cost for bio-methane production
CIC	Capital investment cost
D	Diameter
DEA	Di-ethanol amine
L	Column length
LHV	Low heating value
MEA	Mono-ethanol amine
NCOB	Normalised cost of bio-methane
Nm ³	Normal m ³ of gas-refers to 0 °C and 1 atm
O&M	Operation and maintenance costs
P	Pressure
P/F	Purge and feed ratio
PSA	Pressure swing adsorption
PVSA	Pressure vacuum swing adsorption
r	Discount rate
TEA	Tri-ethanol amine
Subscripts	
<i>Feed</i>	Feed
<i>Biogas</i>	Biogas
<i>C</i>	Column
<i>Removed</i>	Removed CO ₂

1. Introduction

Anaerobic digestion (AD) is a proven technology to treat farm and food industry wastes [1]. Due to microbiological activities, macromolecules such as lipids, proteins and carbohydrates are converted into small chain acids and ultimately into a gaseous mixture (bio-gas) of CH₄ and CO₂ [2]. Farm originated bio-gas presents a CH₄ mole fraction close to 60% while CO₂ largely accounts for the remainder. Bio-gas can be upgraded to bio-methane, for injection to the gas network for use in domestic, commercial and industrial heating or used as fuel in road transport.

Different gas separation technologies can be used for bio-gas upgrading to bio-methane such as chemical and physical driven processes [3]. In chemical driven processes, the CO₂ and other acid gases like H₂S react with a solvent in an absorption column resulting in a bio-methane stream free from CO₂ and H₂S. Solvent is then thermally regenerated and recirculated to the absorption column. In physically driven separations, CO₂ is retained in the surface or in the pores of the separation agent that can be a membrane or an adsorbent [4]. The separation agent is then regenerated using pressure difference and /or vacuum processes [5].

Solvent based processes use mainly aqueous solutions of amines (primary, secondary or tertiary) such as mono-ethanol (MEA), di-ethanol (DEA) or tri-ethanol amine (TEA) as separation agents. Acid gas removal is a widely known and employed technology for natural gas sweetening however the CO₂ mole fractions for natural gas are far lower than the analogous for bio-gas (4-5% CO₂ mole fraction for untreated natural gas and 40-50% CO₂ mole fraction for bio-gas). Significant differences in terms of operating pressures also exist (around 20 bar for the natural gas sweetening process and between 1 and 1.5 bar for the bio-gas upgrading) [6]. The performance of amine based separation units for bio-gas upgrading was studied in previous works [7-8]. Gamba et al [7] reported energy consumption close to 3.52kWh/ Nm³ of treated bio-gas or 0.205 kWh/mol of removed CO₂ when a 15% w/w MEA aqueous solution is used as solvent. Larger MEA weight percentage in the solvent would enable a larger CO₂ load thus lower solvent flowrates and lower energy consumption in the reboiler of the stripper. Solutions of 30% w/w have

been suggested in the literature for CO₂ capture applications, allowing energy consumptions around 0.041 kWh/mol of removed CO₂ [9].

Pressure swing adsorption (PSA) cycles are the most commercially employed sorbent based technology applied to bio-gas upgrading for bio-methane production [10]. In PSA cycles, gas mixtures are separated due to adsorption and desorption steps driven by pressure swings. CO₂ is retained in the pores of the adsorbent during the adsorption step while a nearly CO₂ free methane stream is produced. The sorbent is regenerated during the desorption step, which may occur at atmospheric pressure or vacuum. The separation performance and the associated energy consumption are a consequence of the cycle step design, the adsorption and desorption pressure and the purge and feed ratio. Several works have previously assessed the techno-economic feasibility of the use of pressure swing adsorption cycles in bio-gas to bio-methane upgrading [11-13]. They mainly focused on analyzing high pressure adsorption (up to 8 bar) and atmospheric pressurized purge step, considering as well the expansion of the product gas as a way to produce part of the energy required by the PSA cycles. The use of nearly atmospheric pressurized adsorption and vacuum operated desorption steps (pressure vacuum swing adsorption cycles, PVSA) would enable a better economic performance; requiring lower duty blowers, less expensive columns and a lower energy consumption without the need to expand the product gas.

In this paper, a techno-economic comparison is made between absorption based (using MEA 30%) and pressure vacuum swing adsorption cycles for bio-gas upgrading to bio-methane in the context of anaerobic digestion processes for a 125 herd size cow dairy farms. Economic assessment relies on highly detailed process models for the separation units and on an extensive literature review of different capital investment costs. Results will be valuable for the analysis and deployment of the studied technologies in the farming and the food industry sectors.

2. Plant islands under study

2.1. AD reactor and biogas yield

For the study, the biogas composition and flow-rate were estimated using the Bunswell Boyle equation [14] for a 125 dairy cow herd. It was assumed that each cow produced a daily amount of slurry close to 64 kg [15]. Chemical and physical properties for the slurry were based on the information presented by [16]. Bio-gas composition and volumetric inlet flow to the gas upgrading units are shown in Table 1. Capital investment cost for AD reactor was estimated using the correlation presented in [17] while yearly operating costs were assumed to be 7% of the life-time capital investment cost [18,19].

Table 1. Biogas flow and composition

Biogas flow	
Biogas flow-rate (kg/kg VS)	1.75
Biogas flow-rate (kg/s)	0.015
Composition (mole fraction)	
CH ₄	57.3
CO ₂	40.9
NH ₃	1.5
H ₂ S	0.2

In this study, it was decided to employ a two stage 4 step Skarstrom PVSA cycle adsorption based separation process. In the basic cycle configuration, shown in Fig. 1, each bed undergoes four different steps: adsorption, blowdown, purge and pressurization. During the adsorption step, the most strongly adsorbing specie is retained in the solid while a nearly free CO₂ stream is obtained. It was assumed that the adsorption step took place at 1.5 bar. During the blowdown step, the most strongly adsorbing specie is concentrated in the column due to the pressure decrease that causes the least adsorbing component to leave the column. The sorbent is regenerated at the purge step using a fraction

of the adsorption product gas while the column reaches adsorption pressure due to bio-gas inlet during the pressurization step. Energy is consumed for pressurizing the feed inlet and to produce vacuum. The use of two stages is commonly suggested in order to meet the required separation target with a lower energy consumption [20-22].

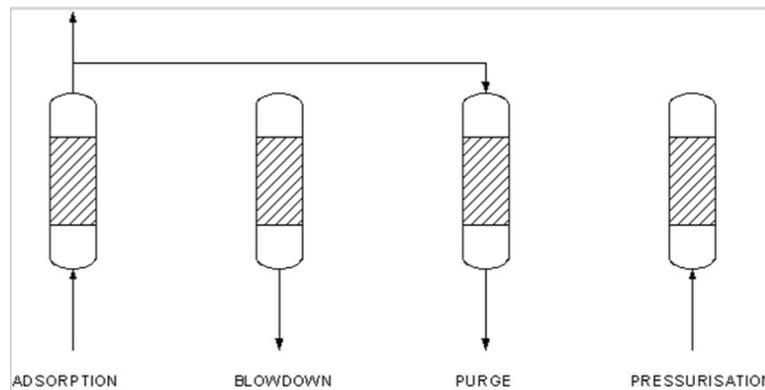


Fig. 1: Steps in Skarstrom PSA cycles

Zeolite 13 X was assumed as a separation agent for the adsorption units. Equilibrium functions for CO_2 and CH_4 and sorbent physical properties were published by [23]. Simulations for the PVSA unit were conducted by using gPROMS software [24] having implemented the model presented in [25], which considers different mechanisms for mass and heat transfer between the phases and the column wall. The simulation of PSA cycles is a computationally challenging problem, due to the non-linear relationships and interdependency among the variables, which leads to a complex system of non-linear partial differential equations. Considering the multi-variable problem and the difficulty of identifying a set of optimal solutions using genetic algorithms, a manual optimization was carried out by effectuating a sensitivity analysis of the different variables. Initial guess for column residence times and cycle configurations were based on the cycle design presented in [26]. Purge pressure and purge to feed ratio were selected in the different stages so that aimed CH_4 purity was reached, maximizing CH_4 recovery but with the lowest specific energy consumption. Table 2 displays operating conditions for the two stages as well as column dimension and overall performance parameters. It was assumed that the electrical power for the operation of the PSA cycles was supplied from the national electricity grid. This electricity accounted for 55% of the electricity consumption of the farm.

The cost for the PSA columns was estimated by quantifying the vessel thickness using ASME codes [27]. Capital investment cost for the vessels was determined using cost tools available online [28]. Zeolite cost was obtained from the vendors. Capital investment cost for the required blower and vacuum pumps was determined from manufacturers and cost estimator tools.

2.3 Solvent based unit

The conventional MEA process, shown in Fig. 2, is one of the most widespread gas sweetening technologies [7]. Bio-gas is fed to the absorber in which CO_2 reacts with the solvent through chemical absorption and a mixture of CH_4 and steam is produced at the top. The rich gaseous CH_4 stream is cooled down and steam is condensed so that a 95.9% CH_4 mole fraction stream is produced. The rich CO_2 solvent stream is processed in a stripper column in which the CO_2 is desorbed by coming into contact with steam produced in the reboiler, and a rich CO_2 gaseous stream is obtained. The lean CO_2 solvent is pumped to the absorption column in combination with solvent and water make up to compensate possible losses in the process.

Table 2. Technical performance and equipment sizing for sorbent based units

Parameter	Value	Parameter	Value
	First stage		Second stage
P_{feed} (bar)	1.5	P_{feed} (bar)	1.5
P_{vacuum} (bar)	0.2	P_{vacuum} (bar)	0.1
P/F ratio (%)	3.0	P/F ratio	1.0
L_c (m)	3.0	L_c (m)	2.6
D_c (m)	0.3	D_c (m)	0.2
Feed compression power (kW)	0.46	Feed compression power (kW)	--
Vacuum power (kW)	1.02	Vacuum power (kW)	0.9
	Overall unit		
CH ₄ recovery (%)			84.1
CH ₄ purity (%)			96.2
Specific energy consumption (kJ/Nm ³ biogas)			200.6
Specific energy consumption (kJ/CO ₂ removed)			10.1

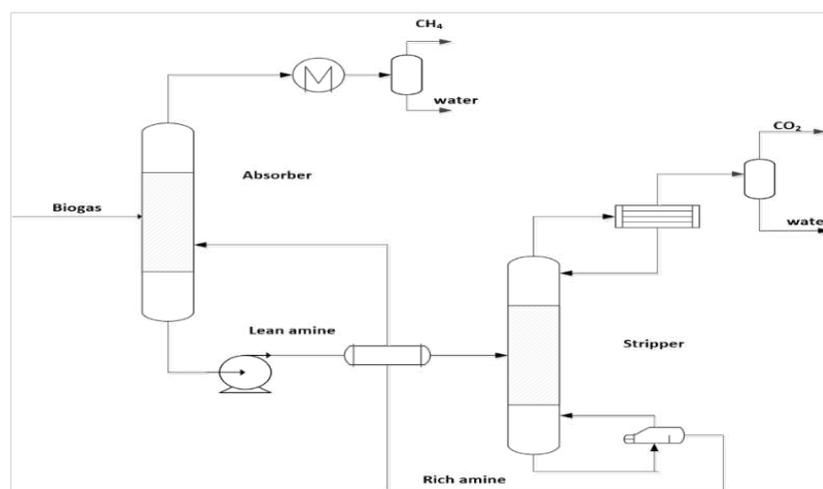


Fig. 2. Conventional MEA process applied to bio-gas upgrading for bio-methane production

Table 3 displays the main operating conditions for the absorption and stripping columns, selected packing, number of theoretical trays, column dimensions and thermal and electrical duty in heat and flow transfer devices. Capital investment cost for the absorption and stripping columns were estimated from information for costs and equivalent height in packed columns presented by [29]. Capital investment cost for biogas blower, stripper condenser and reboiler were quantified using the information from quotes and correlation present in online cost estimator tools [20].

3. Comparative assessment

To enable a good comparison between both gas upgrading technologies, two indicators were analyzed: the net bio-methane thermal flow per kg of volatile solids (NBM), Eq. 1, and the average life-time cost per thermal unit of bio-methane, Eq. 2. Thermal flow was estimated by accounting for the mass of bio-methane produced and the bio-methane low heating value (LHV). The specific thermal flow per kg of volatile solids (VS) was quantified by considering the

ratio between the thermal bio-methane flow to the mass of volatile solids in the slurry, which was assumed to be equal to the mass of bio-degradable waste.

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Table 3. Technical performance and equipment sizing for sorbent based units

Absorber		Stripper	
Operating pressure (bar)	1.3	Operating pressure (bar)	2.0
Operating temperature (C)	43.6	Operating temperature (C)	100
Packing	FLEXIPAC 250Y	Packing	IMTP#40
Number of trays	25	Number of trays	30
L_c (m)	12.5	L_c (m)	15.0
D_c (m)	0.12	D_c (m)	0.08
Feed compression power (kW)	0.58	Reboiler (kW)	19.0
Pump rich amine (kW)	0.005	Condenser (kW)	0.9
Heat rich and lean amine (kW)	13.6		
Overall performance			
CH ₄ recovery considering biogas burned for reboiler operation (%)		81.7	
CH ₄ purity (%)		95.9	
Specific heat consumption (kJ/Nm ³ biogas)		1607.1	
Specific heat consumption (MJ/kg CO ₂ removed)		3.85	

$$NBM_{kgVS} = BMFR / \left(MSL \times \frac{VS(\%)}{100} \right) \quad (1)$$

$$ALTCBM = \left(\frac{\sum_n CIC + O \& M}{(1+r)^n} \right) / \left(\frac{\sum_n BMFR}{(1+r)^n} \right) \quad (2)$$

The average life-time cost of bio-methane is defined as the discounted life-time cost of ownership and use of the plant per unit of bio-methane energy produced in pence/kWh [30]. The average life-time cost of bio-methane was estimated as the ratio of the total cost of the plant (including both capital and operating cost), to the total amount of bio-methane energy flow expected to be produced over the plant's lifetime (20 years).

Higher values for the bio-methane thermal flow per unit of bio-degradable mass unit of waste were obtained for the AD plant with PVSA cycles. This was due to the fact that the energy consumption for the bio-gas upgrading unit was supplied by using electricity from the grid instead of burning part of the produced bio-gas in a CHP system to generate the electricity. It must be noted as well that for the MEA process, the duty of the reboiler of the stripper is 229 % larger than the heat requirement for the farm and the AD reactor while the power consumption for the PSA cycles account for 55% of the electricity consumption of the farm and the AD reactor.

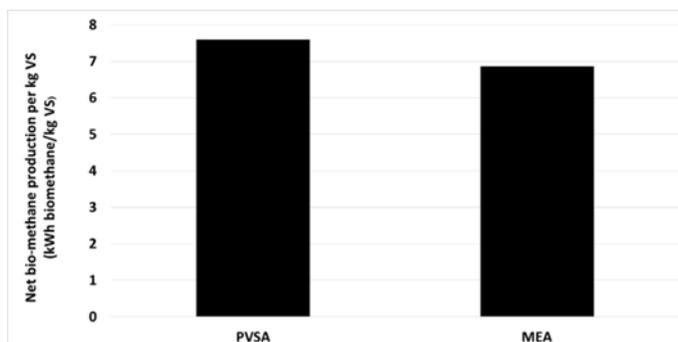


Fig. 3. Net bio-methane production per kg of volatile solid in the slurry

The use of technical indicators to compare separation units with thermal and electrical inputs is not straight forward thus the average life-time cost of methane (ALTCBM) becomes a useful tool to compare the performance of both processes under discussion in this article. As it can be seen from Fig. 4, lower values of ALTCBM for the plant with PVSA cycles were obtained. This was mainly due to a larger bio-methane production rate and to lower capital investment costs in comparison with the AD plant with installed MEA process. Capital investment costs for AD plants with PVSA cycles accounted for approx. £127300 whilst it was close to £190000 (including bio-gas boiler used for steam generation for the operation of the reboiler) for the AD plant with solvent based bio-gas upgrading units. Larger capital investment costs and lower bio-methane production led to larger values for the numerator and lower values for the denominator of Eq. 2 when applied to the AD plant with MEA process for CO₂ removal. Considering that the required heat for the farm is provided by the installed boilers (to produce steam in the MEA unit and to burn the off-gas of the PVSA units), a decrease of propane consumption was reported leading to savings to the farms (or negative cost contributions in Fig.4).

Currently paid incentives in the UK for bio-methane production at this plant scale are around 3.89 p/kWh of bio-methane consequently the ALTCBM for the plant with PVSA cycles represents approximately 35% of the paid tariff. The ALTCBM for the plant with amines is around 39% of the values for the revenues. It shows that for both kind of separation processes, there is a quite significant profit margin for the farms that opt for the deployment of bio-methane production.

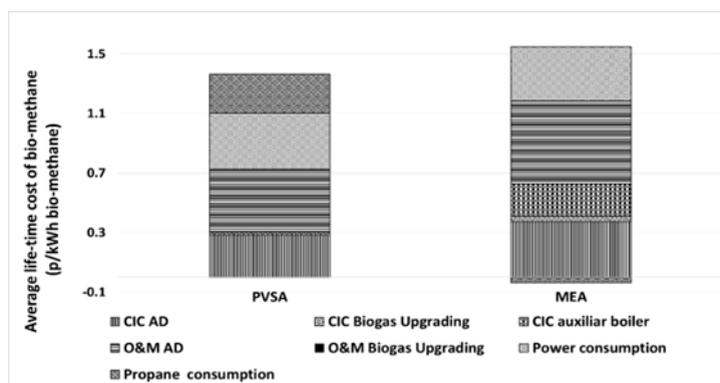


Fig. 4. Average life-time cost of bio-methane for AD plants with PVSA cycles and MEA

Conclusion

In this paper, a techno-economic analysis of solvent based (MEA) and sorbent based (PVSA cycles) bio-gas upgrading technologies for bio-methane production is presented. Highly detailed process simulations were used for the selection of operating conditions that enabled the fulfillment of the separation targets (CH₄ purity over 95% and

CH₄ recovery close to 85%) with the lowest energy consumption. Based on the results for the sizing of the separation units and the duties for the heat and fluid flow equipment, capital investment and maintenance costs were quantified by using cost estimation correlations and quotations from existing systems. It was shown that the average life-time cost of bio-methane for the plant with PVSA cycles was 10% lower than the average life-time cost of bio-methane for the plant with MEA based units while capital investment cost was 37% lower. For both technologies, it was found that the average annualized life-cycle cost of bio-methane production was lower than the incentives paid by the government in the UK. This demonstrates that the production of bio-methane from farm waste can offer a good commercial opportunity to the farm sector in the UK.

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