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Technical-economic assessment of different biogas upgrading routes from vinasse anaerobic digestion in the Brazilian bioethanol industry

Rodrigo Marcelo Leme[*], Joaquim E.A. Seabra

Faculdade de Engenharia Mecânica, Universidade Estadual de Campinas (Unicamp), Rua Mendeleyev, 200, CEP 13083-860, Cidade Universitária "Zeferino Vaz" Barão Geraldo - Campinas, SP, Brazil

A B S T R A C T

This paper presents a technical-economic assessment of biomethane production from vinasse in the Brazilian bioethanol industry, considering five technological routes of biogas upgrading. The technologies assessed were pressure water scrubbing, organic-physical scrubbing, amine scrubbing, membrane separation and pressure swing adsorption. The biomethane costs of the five technological routes overlapped in the range between R\$30/GJ$_{HHV}$ (US\$13/GJ$_{HHV}$) and R\$34/GJ$_{HHV}$ (US\$14/GJ$_{HHV}$), which indicates a certain equivalence of the options. Those costs were comparable to the prices of potential alternative fuels at 2014 prices, such as the Bolivian natural gas, priced at R\$20/GJ (US\$8/GJ); imported LNG, at R\$31/GJ (US\$13/GJ); and diesel, at R\$42/GJ (US\$18/GJ). The effects of scale on biomethane cost were also assessed and pointed out that the likely minimum scale for vinasse-to-biomethane projects aiming at diesel substitution would be at sugarcane mills with a minimum capacity of producing 87 million liters of ethanol per season, whereas if the aim were to target natural gas markets, 174 million liters of ethanol per season would likely be the minimum capacity for competitive prices.

1. Introduction

The sugarcane industry is one of the most relevant in the Brazilian economy. The industry boasts more than 350 sugarcane mills, mostly located in the South of the country, that processed 630 million tonnes of sugarcane in the 2014/2015 harvesting season, placing the country amongst the top producers of fuel ethanol in the world [1]. In Brazil, ethanol is produced through the fermentation of sugars obtained from the sugarcane. As in other fermentative routes, a main production waste, known as vinasse or stillage, results. Vinasse is a liquid of dark color and strong smell, with high organic content and high polluting potential, produced at a ratio of 12 L of vinasse for each liter of ethanol. An estimated 336 billion liters of vinasse must have been produced during the 2014/2015 harvesting season.

The vast majority of the vinasse is disposed directly in the sugarcane field, without prior treatment, in an operation referred to as "fertirrigation" (a combination of fertilization and irrigation). Fertirrigation associated to the use of other residues from sugarcane processing are important elements in the recycling of nutrients for sugarcane production in Brazil [2,3]. Nevertheless, there is still debate as to whether further treatment and control of fertirrigation should be required. Notwithstanding the alleged benefits of vinasse fertirrigation and the lack of evidence of environmental impacts, there is little scientific research available in the literature that effectively investigates the potential detrimental effects of fertirrigation in the environment in the long term [4,5]. Based on extensive literature research, Fuess e Garcia [5] state that the main adverse effects of vinasse fertirrigation on the soil that should be further investigated are salinization; aquifer contamination by salts; organic overloading; reduction of soil oxygen; contamination by chemical species, such as nitrates, chlorides, lead, copper and zinc; and acidification.

Given the benefits of fertirrigation on nutrients recycling and its potential of environmental damage due to overuse, it is reasonable that further options be investigated for vinasse pre-treatment and use. Harnessing its energetic potential, which can be tapped by anaerobic digestion, is one of them.

Anaerobic digestion results in a double-dividend as it not only reduces the polluting potential of vinasse, due to organic matter removal, but also produces biogas. Tapping the energetic potential of vinasse would represent gains between 0.83 MJ and 2.33 MJ per liter of ethanol, which represents improvements between 4% and 10% in the energy yield of a bioethanol distillery [6]. The energetic

potential of vinasse remains mostly unexplored though, as reported by Ref. [7]. There is only a handful of sugarcane mills in Brazil that produce biogas from vinasse and many of those tried it on a pilot scale. Therefore, it is worth to further the research on options to promote for the development of vinasse anaerobic digestion and the biogas industry in the ethanol distilleries.

Most of the literature published so far on the topic focused mainly on its use for electricity generation such as in Refs. [8—11]. There is a lack of studies exploring the use of vinasse for biomethane production. For instance, Moraes et al. [12] assess different applications of biogas, including electricity generation, and concludes that the use of biogas for diesel replacement would be the most economically attractive alternative. The authors conclude that biogas could replace up to 40% of the annual diesel demand in the agricultural operations of the sugarcane mill. Budzianowski et al. [13] report that the high subsidies to biogas-based electricity in Europe have promoted the development of biogas technology towards combined heat and power systems, in spite of the greater economic benefits that biomethane would offer in many cases.

Given the lack of in-depth studies of vinasse-to-biomethane projects in the literature that cover the whole, integrated process from both technical and economic standpoints, the present paper assesses the unit cost of biomethane of five different technological routes of biogas upgrading in the Brazilian bioethanol industry. The paper presents a detailed technical assessment and estimates the unit cost of biomethane considering the whole, integrated process, including vinasse collection in the sugarcane mill, its anaerobic digestion, raw biogas H_2S removal, biogas upgrading and biomethane dispatch to the natural gas grid. The technologies assessed were pressure water scrubbing, organic-physical scrubbing, amine scrubbing, membrane separation and pressure swing adsorption.

2. Materials and methods

The technical-economic assessment was undertaken at an example vinasse-to-biomethane project. The project was chosen at a distillery with the capacity of producing 174 million liters of bioethanol per season. That figure translates into a sugarcane mill with the capacity of processing 4 million tonnes of sugarcane per season, if the distillery is annexed to a sugar factory (assuming that about 50% of the fermentable sugars are destined for sugar production); or a mill with the capacity of processing 2 million tonnes of sugarcane per season, if the distillery is autonomous. Those are relatively typical installations amongst the modern mills in Brazil. That project size results in a nominal biomethane production close to 3000 Nm^3/h.

It has been assumed that all the vinasse produced by the sugarcane mill was processed for biomethane production, as per the quality standards defined by the Brazilian legislation, for injection in a nearby natural gas grid at a pressure of 20 bar(g). The assessment considered the five technological routes for biogas upgrading as presented before, namely, amine scrubbing, organic-physical scrubbing, pressure water scrubbing, membrane separation and pressure swing adsorption.

The technical-economic assessment was undertaken by firstly defining a conceptual process flow diagram for each biogas upgrading route; then estimating the corresponding mass and energy balances; and finally calculating the unit cost of biomethane, for the five biogas upgrading technologies.

The input data used in the assessment was obtained from interviews with technology providers, sugarcane mill operators, biogas upgrading plant operators, and from the literature. Whenever information was lacking, the expert judgement of the authors was used to fill data gaps. Generic process flow diagrams; performance parameters; and mass and energy balances obtained during the interviews were used for the technical assessment and process design. Process integration was made by the authors and resulted in the overall mass and energy balances that generated the information on investment and consumables that were later used for the economic assessment. Prices quotations were obtained also from the interviews for the main process equipment and consumables. The economic assessment was accomplished by calculating the equivalent uniform annual cost (EUAC) of the plant, which was then used to determine the unit cost of biomethane in energy terms.

3. Technical assessment and process design

The conceptual process flow used for each of the biogas upgrading routes is slightly different due to their specific requirements, as it will be explained along the following sections. However, the generic process flow diagram presented in Fig. 1
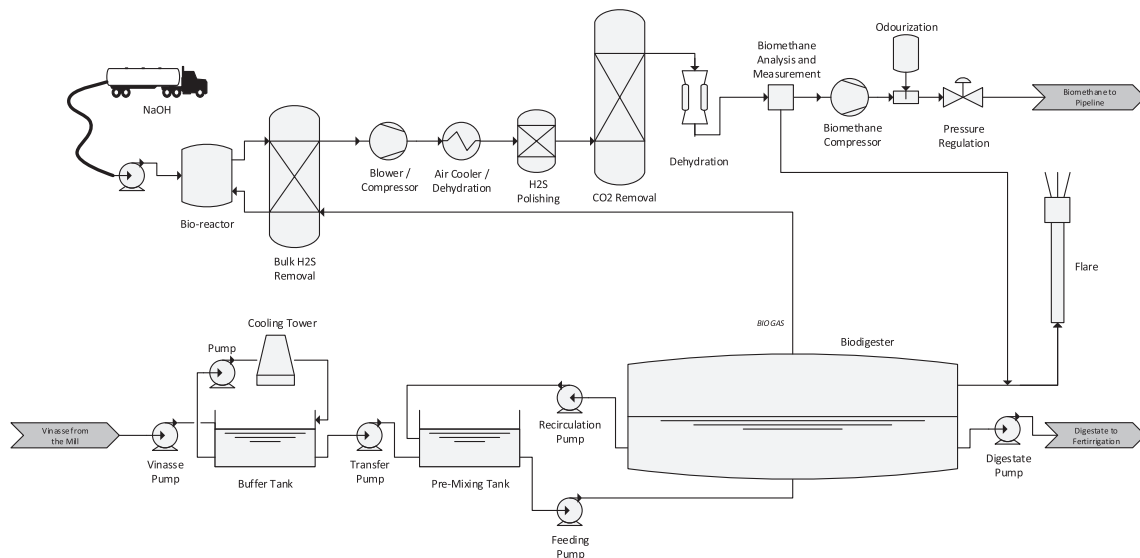

Fig. 1. Generic process flow diagram.

illustrates the main features of the five options.

3.1. Sugarcane mill and vinasse yield

There are three types of ethanol distilleries in Brazil, according to the sources of fermentable sugars [4]. In autonomous distilleries, that produce ethanol only, the fermentable sugars are obtained exclusively from fresh sugarcane juice. In annexed distilleries, that operate next to sugar factories, the fermentable sugars can be obtained from molasses (which is a byproduct from sugar production), originating what is referred to as molasses must; or from a mixture of molasses and fresh juice, producing the so-called mixed must. The source of the fermentable sugars (juice, molasses or a combination of both) used in ethanol production is one of the main factors that determine the characteristics of vinasse, as described in Table 1. In any case the fermentable sugars are fermented by yeast that converts sugar into ethanol, producing what is commonly referred to as wine [14]. The wine is then separated into hydrated ethanol and vinasses through distillation. Further dehydration of ethanol is achieved by means of another distillation step or pressure swing adsorption.

The production rate of vinasse in the sugarcane industry in Brazil is well known and reported in the literature. Vinasse is produced at rates that range from 7 to 16 L of vinasse per liter of hydrated ethanol [6]. A good average value for the Brazilian industry is 12 L of vinasse per liter of ethanol [4]. Variations are mainly due to different process setpoints for alcoholic content in the wine and due to steam recovery from the distillation column.

Table 1 shows significant variation in the values of COD (Chemical Oxygen Demand). In part, that is due to process variations amongst sugarcane mills and the source of fermentable sugars. Another reason is the sampling point of vinasse that can include other inert liquid effluents from the sugarcane mill, which may be mixed with vinasse and then destined for fertirrigation. From interviews with plant operators and the literature [15], it has been identified that the most likely range for vinasse COD lies within 20,000 to 40,000 mg/L, with values between 25,000 and 30,000 mg/L being the most usual.

Vinasse is always acidic, but without detrimental effects on anaerobic digestion, as it is shown by the anaerobic digestion trials and projects in the industry.

The table shows vinasse temperatures as it leaves the distillation columns. For anaerobic digestion, however, vinasse at lower temperatures can be obtained from storage ponds, after vinasse cooling, either by heat exchange with other plant operations or by mixing with other effluents from the plant (which is a relatively common situation in many mills). In fact, the best source of vinasse for anaerobic digestion is probably the vinasse stored in ponds and tanks, that act as homogenization tanks, due to its lower temperatures and relatively homogeneous characteristics, as opposed to fresh vinasse obtained directly from the distillation columns.

The main source of sulphate in vinasse is sulphuric acid (H_2SO_4). It is used in many distilleries for the correction of pH prior to fermentation [5] and in sugar factories it is used in white sugar production. Thus when molasses from sugar production are used in the production of ethanol, an additional load of sulphates may be present. That becomes evident from the much higher sulphate content in vinasse from molasses. Nevertheless, from interviews with plant operators and the literature [15], it has been identified that the most common values of sulphate content in vinasse lie in the range between 500 and 1500 mg/L.

In the example project the following data was used: according to data from the three most recent crushing seasons the average ethanol yield of the bioethanol sector was calculated as 43 L of hydrated plus anhydrous ethanol (expressed as hydrated ethanol) per tonne of sugarcane processed for annexed distilleries [1]; and as 85 L of total ethanol per tonne of sugarcane for autonomous distilleries [14]. The vinasse yield was chosen as 12 L/L of total ethanol (expressed as hydrated ethanol) [4]. The average season in the sugarcane industry was chosen as 200 days per year, operating 24 h per day, with an uptime of around 85%, based in interviews with mill operators. Vinasse COD (Q_{vin}) was chosen as 27.5 kg/m³ and sulphur content (S_{vin}) as 0.8 kg/m³ of sulphate. Considering those parameters, and a production of 174 million liters of ethanol per season as in the example project, the average vinasse production rate was determined as 510 m³/hour with an average organic load of 14,021 kg/h of COD and 3.3 tonne/day of sulphur.

It was assumed that vinasse was available at a storage pond, nearby the project site at 60 °C. Vinasse was transported to the project site through piping and centrifugal pumps and at the project site it was stored in an open pond that served as buffer and homogenization tank. Additional cooling by means of forced ventilation, open cooling towers was considered in order to ensure temperature control inside the anaerobic digester.

3.2. Anaerobic digestion and methane yield

Due to its high organic content, vinasse has a significant energetic potential that can be tapped by anaerobic digestion. Anaerobic digestion can convert into biogas between 70% and 90% of the biodegradable organic matter contained in substrates. Only a small fraction of the organic matter is converted into microbial biomass (from 5% to 15%), which constitutes the excess sludge produced in the system, and the remaining of the organic matter exits the system as non-degraded organic matter [16]. Extensive literature reviews [15] [6] list several examples of successful applications of anaerobic digestion of vinasse. Table 2 presents the main types of anaerobic digesters that could be used for vinasse digestion and their characteristics, namely lagoon digesters, CSTR (Continuously Stirred Tank Reactors) and UASB (Upflow Anaerobic Sludge

Table 1
Selected physicochemical parameters of vinasse.

Parameter	Unit	Range[a]	Molasses[b]	Juice[b]	Mixed[b]
pH	–	3.5–4.9	4.2–5.0	3.7–4.6	4.4–4.6
Temperature	°C	65–110	80–100	80–100	80–100
COD	mg/L	9200–97,400	65,000	15,000–33,000	45,000
Sulphate (SO_4^-)	mg/L	92.3–3363	6400	600–760	3700–3730
Total solids	mg/L	10,780–56,780	81,500	23,700	52,700
Volatile solids	mg/L	628–45,225	60,000	20,000	40,000
Nitrogen (N)	mg/L	81.2–1214	450–1610	150–700	480–710
Phosporus (P_2O_5)	mg/L	<10–188	100–290	10–210	9–200
Potassium (K_2O)	mg/L	814–7611	3740–7830	1200–2100	3340

[a] Source: *apud* [4].
[b] Source: *apud* [4].

Table 2
Main characteristics of selected anaerobic reactors.

	Lagoons	CSTR	UASB
Construction	Lagoons excavated directly in the soil, lined and covered by plastic membranes	Cylindrical tanks made of either concrete or steel	Cylindrical or rectangular tanks made of either concrete or steel
Number of stages	Usually single stage	Arrangements involving two stages are very common	Single stage
Temperature	Mesophilic	Mesophilic or Thermophilic	Mesophilic
Mixing	Internal mechanical mixers and/or high velocity of injection of the substrate ensure proper mixing	Internal mechanical mixers and agitators	Based on the inlet velocity of the substrate and formation of suspended sludge blanket
Solids	Low (<3%)	Medium (<11%)	Medium (<10%)
Organic loading rate	Low (2 kgCOD/m³.day)	Medium (5 kgCODm³.day)	High (10 kgCOD/m³.day)
Biomass retention	In general a settling zone with sludge recirculation is included in the final stage of the lagoon	High hydraulic retention time ensures biomass retention	Based on the formation of sedimentable sludge blanket
Hydraulic retention time	15—20 days	>30 days	<5 days
Footprint	Normally requires the largest area of the three	Moderate as compared to lagoons	The smallest of the three for the same organic load
General comments	More robust due to large buffer volume, higher response times and simple construction	Most versatile, covering a wider range of substrates, but more expensive for low solids substrates	Save space and promote high levels of COD removal, with much smaller retention times and faster response

Source: based on interviews with technology providers.

Blanket). The three of them have been used in Brazil for digestion of vinasse, as supported by interviews with plant owners and technology providers.

Based on literature reviews on the performance of several applications of anaerobic digestion of bioethanol vinasse, average values for COD removal efficiencies in different types of reactors can be within the range of 76.9 ± 11% [6] and 71.20 ± 9.33% [15]. Those ranges of efficiency were also confirmed by interviews with technology providers.

Irrespective of the anaerobic reactor chosen, and assuming that the process conditions are met for good microbiological activity to develop, the production of biomethane can be estimated from the organic load of the substrate and the efficiency of the anaerobic process through the COD balance shown in Fig. 2 [16].

From Fig. 2, the inlet COD can be divided in two types: biodegradable and non-biodegradable. Part of the biodegradable COD will be converted into microbial biomass (microbial COD) and will not generate any biogas. From the remaining biodegradable COD (the acidified COD), part will eventually be converted into biogas (methane COD) and part will not be degraded (fatty acids COD). Methane generation can be determined from that balance according to Equation (1) (as derived in Appendix A):

$$V_{CH4} = 0.35 \cdot (\eta - \beta) \cdot (Q_{vin} - 64/96 \cdot S_{vin}) \tag{1}$$

where, V_{CH4} is the volume of methane [Nm³]; η is the COD removal efficiency [ratio]; β is the fraction of COD that is converted into microbial biomass [ratio]; Q_{vin} is the COD of the incoming from vinasse [kg]; S_{vin} is the sulphate content of vinasse [kg].

Another relevant parameter was the production of sulphidric gas (H_2S), which was conservatively estimated as per Equation (2) (derived in Appendix B):

$$V_{H_2S} = 22.4/96 \cdot S_{vin} \tag{2}$$

where V_{H2S} is the volume of sulphidric gas [Nm³]; and S_{vin} is the sulphate content of vinasse [kg/m³].

In the example project, the COD removal efficiency (η) was chosen as 0.72 based on interviews with plant operators; and a value of 0.1 was chosen for the COD converted into microbial biomass (β), which is the average of the typical range for this parameter between 5% e 15% of inlet COD [16]. Assuming that biogas contains 60% of methane (volume, dry-basis) [17], and substituting values in Equation (1), the total biogas yield resulted in 5020 Nm³/h, or 9.84 Nm³/m³ of vinasse (dry basis, 0 °C, 1 atm).

With the aid of Equation (2) and the result for biogas generation, the concentration of H_2S in the biogas resulted in 1.9% (volume, dry basis). For the sake of simplification, it was assumed that the remaining gas, on a dry basis, was comprised of carbon dioxide only. That resulted in CO_2 content equal to 38.1%. Other gaseous components that might be present in biogas, such as NH_3, H_2, N_2 and O_2, were disregarded. The water content in the biogas was estimated assuming that the biogas would be water saturated at 35 °C and 1 atm. At those conditions, the water vapour pressure is
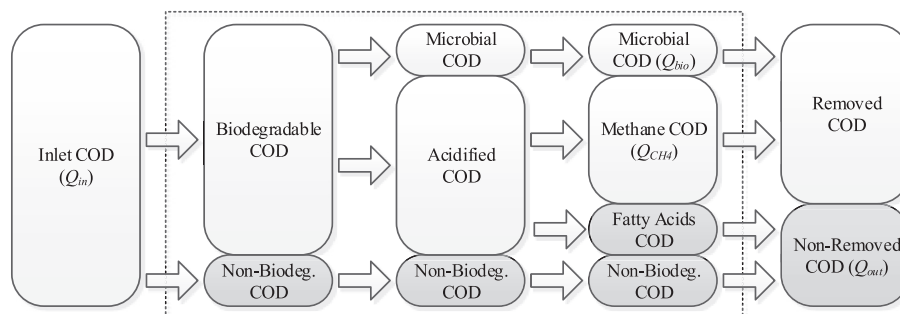
Fig. 2. COD balance of anaerobic digestion.

0.0555 atm, which means that the water content in biogas would be 5.55%.

In all scenarios assessed, the choice was for a lagoon reactor. Based on interviews with technology providers, that type of reactor can achieve the efficiency of COD removal considered in the example project. An alleged advantage of that technology is that further pre-treatment and conditioning of the vinasse would not be required except for pre-cooling. Also, due to the thermal inertia of the reactor and the feeding temperature of vinasse, temperature control of the reactor would not be required, according to technology providers. Prior to the digester itself it has been considered that a pre-mixing tank is used to mix the fresh incoming substrate with partially digested substrate taken from inside of the reactor.

3.3. Biogas upgrading

The biogas produced by anaerobic digestion is comprised of a mixture of gases that varies according to the substrate and the conditions of anaerobic digestion. In general, however, it can be assumed that the biogas composition, on dry-basis, includes between 60% and 70% by volume of methane and between 30% and 40% of carbon dioxide [17,18]. It also contains trace components such as hydrogen, ammonia, hydrogen sulphide, nitrogen and oxygen, depending on a number of factors.

Resolution 8/2015 [19] of the Brazilian oil and gas agency establishes the quality standards of biomethane so that there is perfect interchangeability between biomethane and natural gas. It defines biomethane as a gaseous biofuel comprised of mainly methane, produced from upgrading biogas; whereas biogas is defined as the raw gas obtained from the biological degradation of organic matter. The biogas produced from the anaerobic digestion of vinasse, and the biomethane produced from it, clearly comply with the definitions above. Table 3 lists the components of biogas and compares them to the biomethane standards established by Resolution 8/2015.

From Table 3 it can be concluded that the main components to be removed from biogas, in order for it to attain biomethane status as per Resolution 8/2015, are carbon dioxide (CO_2), sulphidric gas (H_2S) and water (H_2O).

3.3.1. H_2S removal

H_2S is a highly toxic and corrosive gas and its removal is strongly recommended since the first stages of the process, not only to improve process safety, but also to facilitate the design of the next steps of the process. This is best accomplished following a two-step removal process. In a first step, the bulk removal is accomplished using a technology that is robust and economical, that normally involves regenerable medias, but that will still leave some H_2S in the gaseous stream. In a second step, commonly referred to as polishing step, the remainder of the H_2S can be removed by a more efficient method that is normally also more expensive because it involves non-regenerable medias.

There are various such methods, each one with its application niche [17]. In general terms, H_2S removal technologies can be divided according to gas flow and H_2S concentration (which combined result in the daily load of sulphur − S-load) into three zones [20]:

- Sulphur Load < 50 kgS/day − for such cases, absorption or adsorption in non-regenerable liquid or solid medias is recommended. Although these technologies tend to be very efficient in removing H_2S, delivering a virtually sulphur-free gaseous stream, and have a relatively simple plant design, their disadvantage is that the media is consumed in the process, which can represent a high operational cost;
- 50 kgS/day < Sulphur Load < 50 tonS/day − Biological systems are a good option for these levels of sulphur load. In this case the operation of the system is based on the removal of sulphur through the aerobic metabolism of bacteria that convert sulphides into elemental sulphur or sulphates. These methods tend to have affordable investment costs, associated with relatively lower operational costs;
- Sulphur Load > 50 tonS/day − For higher loads, the use of regenerable solid or liquid media is the best option. In this case the principle of operation is similar to the first case, however the media is regenerated at a later stage so that it is not consumed in the process. This reduces significantly the operational cost of these technologies. However they tend to require more complex installations and a higher investment cost. This is mainly because of the steps required for media regeneration, which commonly involves heat and/or pressure.

Given the H_2S concentration and biogas flow, the sulphur load of the example project is 3.3 t/day of sulphur. For such a sulphur load, the sulphur removal technology would be in the range of biological scrubbers. Therefore, this type of technology was selected for H_2S removal in this study. The technology chosen involves a combination of a physicochemical sodium hydroxide (NaOH) scrubber with a biological reactor. The scrubber is responsible for H_2S removal from the biogas flow and the biological reactor, for the final removal of sulphur from the system and regeneration of alcalinity, so that the consumption of sodium hydroxide is reduced. Sulphur is removed from the system mostly in the form of elemental sulphur. That sulphur can be further processed for sulphur recovery or it can be used *in natura* as a fertilizer, according to technology providers. The target H_2S concentration downstream of the H_2S scrubber was set at 25 ppmv.

Besides the biological scrubber, a final polishing step was also

Table 3
Biomethane quality standards and biogas composition.

Parameter	Units[a,b]	Biomethane (ANP 8/2015)	Vinasse biogas[d]
CH_4	% mol	>96,5	55−65
O_2	% mol	<0,5	~0
CO_2	% mol	<3	45−35
$CO_2 + O_2 + N_2$	% mol	<3,5	45−35
Total sulphur[c]	mg/m³	<70	9500−42,500
H_2S	mg/m³	<10[e]	10,000−45,000
H_2O (dew point at 1 atm)	°C	<−45	Saturated
Other	−	Not mentioned	Traces

[a] Reference conditions are 20 °C and 1 atm.
[b] Composition is dry-basis.
[c] It is assumed that all sulphur in biogas is in H_2S.
[d] Based on interviews with plant owners and technology providers.
[e] This is equivalent to approximately 7 ppmv.

included, when necessary, in order to bring H_2S concentrations further down to the level required by Resolution 8/2015. In that case, an iron-oxide-based media was chosen with a target output of less than 3 ppmv of H_2S in the biogas.

A note on the interaction between CO_2 and H_2S removal is due at this point because some of the CO_2 removal technologies can be used for H_2S removal as well. The pressure water scrubber also removes H_2S due to the high solubility of H_2S in water. In this paper, however, the water scrubber is not used for bulk H_2S removal. It is only used for H_2S polishing. Two reasons explain this choice. Firstly, the H_2S content in the biogas is well above what commercially available water scrubbers would be able to handle. Furthermore the H_2S absorbed in water must be separated and discarded through the vent stream of the plant so that water can be re-used. And this vent stream would have to be treated for H_2S anyway. For the same reasons, amine plants that could also be used for H_2S removal are not used for that purpose.

In the case of organic physical scrubbing, which could also be used to remove H_2S, it has been ascertained with technology providers that this is not the practice in the biogas industry. Usually if such a plant is chosen, H_2S is removed upstream of the plant due to the increased complexity of the upgrading unit, should both CO_2 and H_2S be removed on a single step in the organic-physical scrubber. In the case of membrane separation and pressure swing adsorption H_2S must be removed upstream of the plant because H_2S is a contaminant that can hamper the performance of CO_2 removal.

For those reasons the option in the example project was to remove H_2S upstream of the CO_2 removal step, so that a virtually sulphur free gaseous stream was available for further processing. Furthermore, in the case of organic-physical scrubbing, membrane separation and pressure swing adsorption an additional step of H_2S polishing was added after bulk H_2S removal.

3.3.2. H_2O removal

Water removal can be accomplished by different methods [21]. In biogas applications three processes are most relevant and, in general, are combined in biogas upgrading plants.

The first one is incidental and depends on the level of gas compression and cooling, upstream of the CO_2 removal step. In general, the higher the pressure of the biogas, the more water is removed during compression after successive stages of compression, cooling and condensate separation. Nevertheless, in most cases, additional dehydration steps are required.

The second method commonly used for water removal in biogas upgrading is through cooling-effect dehydrators, which are normally combined with the compression stage of biogas. The principle is based on the condensation and separation of water from a pressurized gaseous stream.

The third relevant technology is the so-called temperature swing adsorption (TSA). This technology is similar to pressure swing adsorption and is based on the adsorption of water on solid media. It is particularly recommended for applications where water in gaseous phase need to be removed from a mixture of gases that may contain some H_2S and/or when the target water concentration in the gaseous stream is very low, as in biogas applications (the quality standard for water in biomethane is a dew point of -45 °C). The main consumable in this process is the heat required to regenerate the media. In biogas applications, it is not uncommon that a simple electrical heater is used for that purpose.

Water removal is dependent on the CO_2 removal technology. In the case of membrane separation and pressure swing adsorption water must be removed upstream of CO_2 removal, because water is a contaminant. Cooling-effect dehydrators combined with the compression step of biogas were chosen.

In the case of organic-physical scrubbers, due their operating pressure, the level or water removal achieved during biogas compression is enough, thus no additional water removal step was required. And for pressure water scrubbers and amine scrubbers, water must be removed downstream of CO_2 removal, because CO_2 removal itself saturates the gaseous flow with water. In those cases, temperature swing adsorption (TSA) dehydrators were selected. The heat for the TSA was obtained from an electrical heater. In all cases, the target dew point of water in biomethane at 1 atm was set at less than -45 °C, as per ANP standards.

3.3.3. CO_2 removal

There are a number of different technologies that can remove CO_2 from biogas [22–31]. They range from well established technologies that have been in use in the oil and gas industry, such as amine scrubbing; to technologies that are widespread in the industrial gases production, such as pressure swing adsorption and membrane separation; as well as specific ones, like pressure water scrubbing, that seem to have its main application niche in the biogas industry. From the literature review and from interviews with technology providers and plant owners, five technologies were identified as the most relevant for CO_2 removal and those will be assessed in this study: pressure water scrubbing, organic-physical scrubbing, amine scrubbing, pressure swing adsorption and membrane separation. Table 4 briefly addresses each one of these technologies in the context of vinasse-to-biomethane projects.

In the example project, the aim was to produce biomethane at methane concentrations higher than 96.5%mol and CO_2 less than 3%mol, on dry-basis, in compliance with Resolution 8/2015.

In the case of pressure water scrubbing, the operating pressure was chosen as 8 bar(g) and a conventional electrical glycol chiller was selected to ensure that process water was cooled down to 6–7 °C. Dehydration downstream of CO_2 removal was required and accomplished using a TSA, as explained before. After the process, biogas is available at around 8 bar(g) and methane losses were identified as being close to 2% of inlet methane.

For the organic-physical scrubber, an operating pressure of 30 bar(g) was indicated by technology providers. The plant involved two steps, in the low temperature step an electrical glycol chiller was chosen to provide the required cooling effect for the solvent. In the high temperature step, heat removed in the interstage cooling of biogas compression was used to provide the required thermal energy for the solvent. The upgraded biogas is available at around 30 bar(g) downstream of the plant, so no further compression was required. Methane losses were identified as being close to 3% of inlet methane.

The amine scrubber was chosen to operate at 100 mbar(g), which was achieved by centrifugal blowers. The main utility used was heat for solvent regeneration. In this case it was assumed that heat could be obtained from the cogeneration system operating in the sugarcane mill. In fact, given the heat demand indicated by the technology provider of the amine plant ($\sim$700 kcal/Nm3 of biogas at 125–150 °C), it was reasonable to assume that the sugarcane mill could supply the heat required by the amine plant in the form of saturated steam. After CO_2 removal, the biogas achieves very high quality standards with methane concentrations above 99% mol (dry-basis). Water must be removed and a dehydration step was required downstream of the plant. The biogas is available at very low pressure downstream of the plant and must be compressed to reach pipeline pressure. Methane losses are very low, below 0.1% of inlet methane.

Pressure swing adsorption was chosen to operate at 8 bar(g). Because water was removed upstream of the plant, no further dehydration was required. The biogas is available at around 7 bar(g)

Table 4
CO_2 removal technologies.

	Description	Pre-/Post-treatment	Consumables	Operating pressure
Pressure water scrubbing	Based on the solubility of CO_2 in water. Water is regenerated through flashing and stripping operations.	Robust as far as other gas contaminants are concerned, including H_2S. Downstream dehydration is required.	Electricity, process water and chilled water.	8 bar(g)
Organic-physical scrubbing	Based on the solubility of CO_2 in organic solvents, e.g. polyethylene glycol. The solvent is regenerated by flashing and stripping.	H_2S need to be removed upstream of the plant. There is no need for further dehydration downstream of the plant.	Electricity, solvent, heat and chilled water.	30 bar(g)
Amine scrubbing	Based on the removal of CO_2 by means of a weak chemical reaction between CO_2 and amines. Amine is regenerated through heating.	Oxygen is an important contaminant and must be removed if present in biogas. Downstream dehydration is required.	Electricity, amine and heat.	<1 bar(g)
Pressure swing adsorption	Based on the adsorption of CO_2 in specific solid media. The plant can have several vessels working in parallel so that while one vessel adsorbs others are regenerated, through pressure decrease.	H_2S, water and any other trace contaminants must be removed upstream.	Electricity and adsorbent media.	8 bar(g)
Membranes	Based on the separation of gas components due to their different permeability through a set of polymeric membranes. The plant can have several stages, with increasing methane purity in the outlet.	Similarly to the PSA technology, a good pre-treatment of the gaseous stream is required.	Electricity and membranes.	15 bar(g)

Source: based on literature review [22–31] and interviews with technology providers.

and methane losses are very high, above 8% of inlet methane.

The membrane separation plant can be offered with different number of stages, depending on the targeted quality of product gas. For this study, a plant with three stages was considered so that the quality standards for biomethane could be met. The plant operating pressure was chosen as 16 bar(g) and downstream of the plant the biogas is available at around 14 bar(g). No dehydration is required as water was removed upstream of the plant. Methane losses are low, in the order of 1% of inlet methane.

3.4. Compression, analysis and odorisation

For all biogas-upgrading routes, some level of biogas compression was required. Apart from amine scrubbing, which required very low pressure to operate, biogas compression was accomplished in two steps. For amine scrubbing, the plant was fed directly from the first step. The first step of compression involved the use of centrifugal blowers that provided the required vacuum to pull the biogas from the digester through the H_2S removal facility and deliver it at a slightly positive pressure to the next step of compression. The second step depended on the operating pressure of the different CO_2 removal technologies. Conventional reciprocating or screw compressors were chosen to bring the biogas up to the operating pressure. In fact, depending on the technology provider, the CO_2 removal plant includes the biogas compressor.

In addition to biogas compression, biomethane compression was required except for organic-physical scrubbing, assuming that the pipeline pressure was 20 bar(g) in the example project. In the case of organic-physical scrubbing the operating pressure was above 20 bar(g) and no further compression was required. For biomethane compression, an electrical reciprocating compressor was chosen. Besides biomethane compression, the example project also included gas analysis by gas chromatography, in compliance with Resolution 8/2015, which was then followed by gas measurement and odorisation. Odorisation was accomplished using conventional equipment that injects controlled amounts of odorizer (e.g. mercaptans) directly into the gaseous stream based on the gaseous stream flow rate.

Apart from the main process flow described above, the example project also considered all ancillary equipment and installations required for plant operation, such as, emergency flare, fire and gas protection, plant integration and plant automation. The supply of electrical power was assumed to be by the sugarcane mill, considering that enough surplus capacity would be available from the mill's bagasse cogeneration system, which is a reasonable assumption for most modern sugar mills. The digested vinasse effluent of the anaerobic digester was sent back to the sugarcane mill for further use in the usual fertirrigation system.

3.5. Technical assessment and process design summary and results

Tables 5 and 6 summarize the process parameters for the example project, as well as the results of the technical assessment. These parameters and results were used in the economic assessment of the following section. Table 5 lists the process parameters for biogas production and bulk H_2S removal. Table 6 lists the parameters for biogas compression and upgrading, including biogas compression, dehydration, CO_2 removal and final biomethane compression and conditioning. The energy base used to report the parameters below is based on the higher heating value of biogas in Table 5 and the higher heating value of biomethane in Table 6, both calculated in compliance with Resolution ANP16/2008 [32].

4. Economic assessment

4.1. Economic assumptions and parameters

The economic assessment of the different options was accomplished through the evaluation of the unit cost of biomethane, determined from the equivalent uniform annual cost (EUAC). The

Table 5
Process parameters for biogas production and bulk H_2S removal.

Parameter	Units	Value
Sugarcane mill and vinasse production		
Distillery capacity[a]	million L/season	174
Ethanol production — Annexed distillery (50% mix)[b]	L/tonne	43
Ethanol production — Autonomous distillery[c]	L/tonne	85
Vinasse production[d]	L/L	12
Season[e]	days	200
Uptime[e]	—	85%
Vinasse		
Flow[f]	m^3/h	510
Vinasse COD (Q_{vin})[g]	kg/m^3	27.5
Vinasse sulphur (S_{vin})[g]	$kgSO_4/m^3$	0.8
Vinasse transportation and cooling		
Electricity[f]	kWh/GJ	0.898
Anaerobic digestion		
Volume of the anaerobic reactor[p]	m^3	140,000
COD removal (η)[h]	fraction	0.72
COD converted into microbial biomass (β)[i]	fraction	0.10
Electricity[f]	kWh/GJ	1.89
Biogas		
Flow[j]	Nm^3/h (dry basis)	5020
P[k]	bar(g)	0
T[k]	°C	35
CH_4[l]	% mol (dry basis)	60
CO_2[m]	% mol (dry basis)	38.1
H_2S[n]	% mol (dry basis)	1.9
H_2O[f]	% mol	5.5
H_2S removal		
Load[f]	kgS/day	3.3
Outlet H_2S[o]	ppmv	<25
Consumables - Electricity[o]	kWh/GJ	1.86
Consumables - Water[o]	L/GJ	9.69
Consumables - Nutrients[o]	L/GJ	0.03
Consumables - NaOH[o]	L/GJ	2.69
Infrastructure and utilities		
Electricity[f]	kWh/GJ	1.25

[a] Chosen by the authors based on interviews with plant operators. The size is consistent with modern distilleries currently operating in the country.

[b] Obtained from interviews with plant operators and cross-checked with estimates calculated from Ref. [1] which resulted in 42.8 L/tonne.

[c] Obtained from interviews with plant operators and cross-checked with the value presented in Ref. [14].

[d] Obtained from interviews with plant operators and cross-checked with literature references, such as [4] that presents 12 L/L; and [6] that presents a range from 7 to 16 L/L.

[e] Season duration and uptime are highly variable depending on the mill, season, sugarcane yield and weather. The values are representative of a large number of mills and were obtained from interviews with plant operators.

[f] Calculated by the authors from mass and energy balances. The values were cross-checked and showed consistency with data from plant operators and technology providers.

[g] Obtained from interviews with plant operators and cross-checked with the literature. The values were consistent with the ranges presented in Refs. [4] and [15].

[h] Obtained from interviews with plant operators and cross-checked with the literature. The values were consistente with the ranges presented in Ref. [6], 0.77 ± 0.11%; and [15], 0.71 ± 0.09%.

[i] Obtained from interviews with plant operators and cross-checked with the literature. The values are consistente with the ranges presented in Ref. [16], range 5%–15%.

[j] Calculated by the authors from mass balance and equation (1). The result is consistent with estimates provided by technology providers and plant operators.

[k] Obtained from interviews with technology providers and plant operators. Values may vary but within a range that cause little impact on project performance.

[l] Chosen by the authors based on [17] and data from plant operators. Consistent with estimates provided by technology providers.

[m] Calculated by authors from mass balance. Consistent with [17] and data from plant operators.

[n] Calculated by the authors based on mass balance and equation (2) Consistent with data from technology providers and plant operators.

[o] Obtained from interviews with technology providers and consistent with [20].

[p] This is a preliminary estimate by technology providers base on a covered lagoon reactor, considering proper mixing and adequate management of sludge.

analysis was based on prices and quotations obtained from interviews with technology providers and market players for the year 2014. Whenever the commercial conditions were ex-works or free-on-board (FOB, as per the Incoterms), the costs for importation including packaging, freight, insurance, import duties and customs clearance was estimated as an additional 40% of the ex-works or FOB price, which is consistent with estimates from technology providers. The effects of inflation were not taken into account, thus real interest rates were used. No salvage value was considered in the analysis.

Table 7 lists the financial and economic parameters used in the assessment. The opportunity cost of vinasse was deemed as zero because no alternative uses for vinasse, apart from fertirrigation, are relevant. It was considered that vinasse would return to the sugarcane mill with similar agronomic properties, for further application in the sugarcane field after anaerobic digestion.

The cost structure was divided in two parts: investment costs and operational costs. Within the investment costs, three

Table 6
Process parameters for biogas compression and upgrading.

Parameter	Units	Pressure water scrubbing	Organic-physical scrubbing	Amine scrubbing	Pressure swing adsorption	Membrane separation
Operating pressure at CO_2 removal[a]	bar(g)	8	30	0.1	8	16
Downstream compression[a]	–	Required	Not required	Required	Required	Required
Dehydration[a]	–	Downstream	Not required	Downstream	Upstream	Upstream
H_2S polishing[a]	–	Not required	Required	Not required	Required	Required
H_2S polishing media replacement[a]	g/GJ	–	17.97	–	19.37	17.61
Electricity[a]	kWh/GJ	11.74	10.36	5.82	11.20	8.61
CO_2 removal media replacement[a]	per GJ	11.8 L (water)	5.31 g (solvent)	7.75 g (amine)	14.1 g	77×10^{-6} units
	[R$/GJ]	[0.02]	[0.13]	2.90 L (water) [0.11]	(adsorbent) [1.09]	(membrane) [1.80]
Heat (150 °C)[a]	GJ/GJ	–	–	0.117	–	–
Chilled water[a,b]	kWh/GJ	2.43	0.70	–	–	–
Methane losses[a]	% of inlet methane	2%	3%	0.05%	8%	1%
Biomethane yield[c]	GJ$_{HHV}$/season	479,055	474,167	488,584	439,948	483,941
	[Nm^3/h at %CH_4]	[3043 at 97%]	[3012 at 97%]	[3041 at 99%]	[2795 at 97%]	[3027 at 98.5%]

[a] This includes electricity used for compression, CO_2 removal, H_2O removal and other remaining steps of the downstream process. It excludes electricity used in chillers (see line "Chilled water"). All data was obtained from interviews with technology providers and plant operators. For amine scrubbing, pressure swing adsorption and membrane separation, at least two technology providers were interviewed for each technology. In the case of pressure water scrubbing and organic-physical scrubbing one technology provider and one plant operator was interviewed for each technology. In general the data is consistent with references in the literature such as [17] and [20–31].
[b] Expressed in terms of electricity consumption assuming chilled water is produced in conventional, vapour compression, electrical, water-glycol chillers.
[c] Calculated by the authors based on mass balance.

Table 7
Parameters used in the economic assessment.

	Unit	Value
Parameter		
Investment term[a]	years	20
Discount rate[a]	%	10
Dollar[b]	R$	2.35
Euro[b]	R$	3.12
Importation costs (added to FOB prices)[a]	%	40
Consumables and utilities costs		
Electricity[c]	R$/MWh	200
Water[d]	R$/m3	2.00
Nutrients[e]	R$/L	3.15
NaOH[e]	R$/L	2.21
H_2S adsorbent (iron oxides)[e]	R$/kg	4.35
Amine[e]	R$/kg	13
Membranes[e]	R$/unit	23,316
CO_2 adsorbent[e]	R$/kg	77
Solvent[e]	R$/kg	25
Heat[f]	R$/GJ	23

[a] Chosen/estimated by the authors based on interview with plant operators and technology providers.
[b] Average exchange rate during 2014 obtained from the Central Bank of Brazil.
[c] Chosen by the authors based on prices of electricity in public auctions.
[d] Highly variable, depending on the location of the project. Chosen by the authors as a conservative estimate based on water prices for industrial users.
[e] Price obtained from supplier.
[f] Estimated by the authors based on the energy balance of the sugarcane mill with a thermal efficiency of heat generation equal to 90% and an opportunity cost of fuel (bagasse) equal to R$160/tonne.

categories were considered: (1) engineering, procurement and construction management (EPCM), which accounted for all engineering and management costs associated with project implementation; (2) equipment, which accounted for the investment in process equipment; and (3) infrastructure and construction, which accounted for the civil works, electromechanical assembly and ancillary installations, such as electricity supply and distribution, water supply and distribution, offices, etc.

Operational costs were comprised of: (1) personnel, which accounted for operators, technicians, engineers and administrative staff required for the operation of the plant, assuming that some synergy exists between the biomethane plant and the sugarcane mill; (2) consumables, which accounted for electricity, water, media replacement and other items used directly in the process; and (3)

maintenance, which accounted for materials (such as lubricating oil, gaskets, seals, etc.) and specialized labour for routine maintenance of the equipment. In particular, maintenance costs were estimated as 1.5% of investment costs, except for pressure swing adsorption and membrane separation, for which 75% of that value was used considering that maintenance can be significantly easier in those cases due to the nature of those processes which does not involve solvents and liquids.

In order to account for uncertainties in the costs above, fixed levels of allowances were considered on the reference case, so that a range of most likely values was obtained for the biomethane cost. These allowances were chosen for each cost category based on the perceived uncertainty involved in the cost estimates obtained from the interviews and the personal judgement of the authors, as indicated in Table 8. The allowances were added to the costs of the reference case so that best and worst case scenarios resulted for the biomethane cost. Table 8 also lists the investment and operational costs of each technological route.

4.2. Biomethane unit cost

The results for the unit cost of biomethane are shown in Fig. 3, which also shows some price benchmarks for most likely alternative fuels. The lowest cost was that of organic-physical scrubbing, at R$29/GJ$_{HHV}$ (US$12/GJ$_{HHV}$), with pressure swing adsorption resulting in the highest value, at R$33/GJ$_{HHV}$ (US$14/GJ$_{HHV}$). Although the range of variation was quite narrow, close to 14%. The inclusion of allowances resulted in a range between R$30/GJ$_{HHV}$ (US$13/GJ$_{HHV}$) and R$34/GJ$_{HHV}$ (US$14/GJ$_{HHV}$) where the five technological routes overlap.

It is worth noting that the value estimated above is very specific to the case proposed and certainly bears no comparison with biomethane costs calculated in different contexts (i.e. industry type, year, country, etc.). Furthermore, the present paper considers the whole, integrated process – including vinasse collection, anaerobic digestion, H_2S removal, biogas upgrading and biomethane dispatch.

Further details are presented on Figs. 4 and 5 that show the breakdown of the unit cost of biomethane. Fig. 4 presents the composition of cost based on production of biogas, after bulk H_2S removal, and biogas upgrading. The cost of biogas production associated with bulk H_2S removal (i.e. down to 25 ppmv of H_2S) represents more than 50% of the unit cost of biomethane, for all five

Table 8
Investment and operational costs, from vinasse collection to biomethane dispatch – reference case.

Parameter	Allowances[a]	Pressure water scrubbing	Organic-physical scrubbing	Amine scrubbing	Pressure swing adsorption	Membrane separation
Investment costs (million R$)						
EPCM[b]	−8/+30%	5.0	5.0	5.0	5.0	5.0
Equipment[c]	−8/+8%	44.4	41.8	45.7	46.2	45.9
Infrastructure & construction[d]	−8/+30%	20.3	20.3	20.2	20.3	20.2
Operational costs (million R$)						
Personnel[e]	−8/+15%	0.65	0.65	0.65	0.65	0.65
Consumables[f]	−15/+15%	4.9	4.7	5.5	5.0	5.3
Maintenance[g]	−15/+15%	0.66	0.63	0.68	0.61	0.60

[a] Allowances were chosen based on the expert judgement of the authors and the uncertainties involved in each cost category. They are consistent with the levels of allowances practiced in the industry for this level of assessment.

[b] EPCM costs were obtained from engineering companies for projects of similar characteristics.

[c] Equipment costs were obtained from technology providers.

[d] Infrastructure and construction costs were estimated by the authors based on market prices for projects of similar characteristics.

[e] Costs with personnel were estimated by the authors considering a minimum amount of people needed to operate and maintain the plant and current salary and wage rates in the country.

[f] Costs with consumables were calculated by the authors based on information presented in Tables 5, 6 and 7.

[g] Maintenance costs were estimated by the authors as being 1.5% of equipment price (except for pressure swing adsorption and membrane separation, for which 75% of that value was used), based on interviews with technology providers and the author's expert judgement.
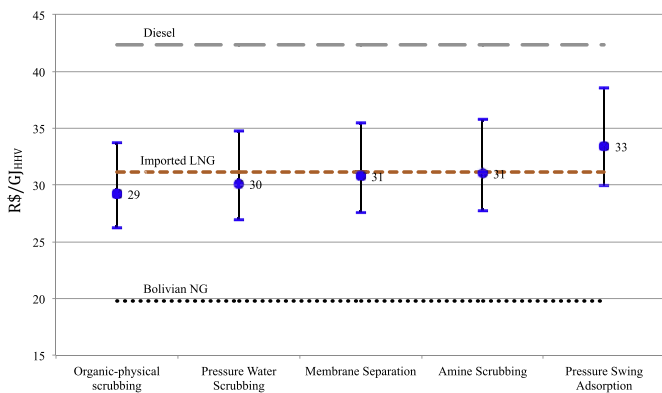
Fig. 3. Unit cost of biomethane.
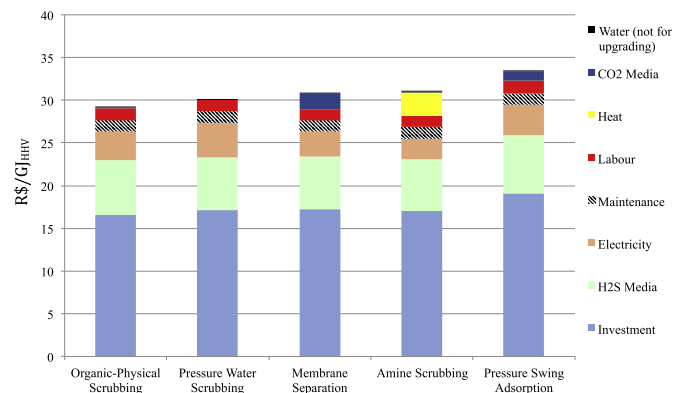
Fig. 5. Unit cost breakdown.
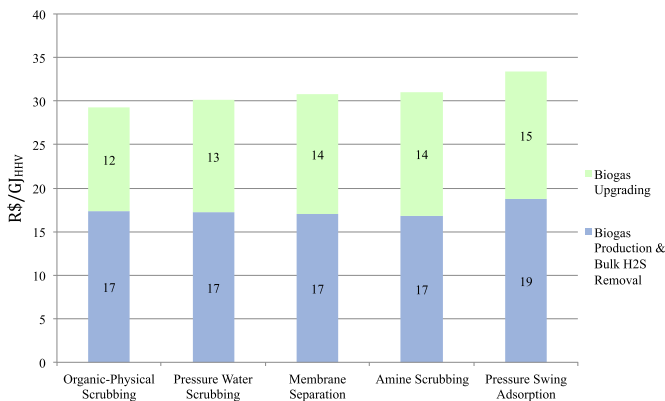
Fig. 4. Impact of biogas production and upgrading on the unit cost.

technological routes. Therefore biogas production and bulk H_2S removal explains a significant part of the cost of biomethane production.

Fig. 5 shows the cost breakdown in terms of the main cost categories. Two cost categories appear as the most relevant for all cases, namely, investment costs and operational costs associated with H_2S removal. The third most important category, except for amine scrubbing, is the operational cost associated with electricity consumption. For amine scrubbing, the third most important cost

category is the cost heat. Maintenance, labour and other consumables resulted in a smaller share of the total cost.

Investment costs on biogas production are mainly affected by the cost of the anaerobic digester and the bulk H_2S removal equipment. Both of them can be procured and built locally, although the number of local suppliers is limited and both are likely to include royalties due to technology transfer. On the other hand, the investment costs on biogas upgrading are significantly influenced by exchange rates and importation costs as a major part of the biogas upgrading equipment is imported. An additional challenge is the lack of technology providers with proven experience in such type of projects with permanent base in Brazil. Importation costs alone would represent a reduction of 3–6% in the cost, according to estimates of the authors. Perhaps some incentive on the nationalization of equipment, development of national suppliers or abatement on import duties would help improve the costs of biomethane and accelerate the learning curve of vinasse-to-biomethane projects in the country.

The second most important cost item is the operational costs associated with H_2S removal. In fact this is one of the significant sources of uncertainty in vinasse-to-biomethane projects because the references on H_2S generation in anaerobic digestion of vinasse are very limited. In particular, there is a lack of information on the relationship between vinasse characteristics and resulting H_2S in the biogas. Perhaps, from the technical point of view, this is one of the most relevant topics for further investigation. There certainly is

potential for cost reduction, not only by better upfront estimation of H_2S generation in the process, but also in terms of H_2S removal options for such high H_2S concentrations in the biogas. There are also changes in the sugarcane mill that could be investigated in order to reduce sulphate content in the vinasse and consequently, reduce H_2S generation.

As regards the impact of electricity use on the unit cost of biomethane, there is probably little that can be done as far as technological improvements are concerned. Most of the electricity is actually used for biogas compression, which is not different from the compression of natural gas, with little room for significant technological improvements. Cost reduction is more likely to happen in this case from a reduction in electricity prices, which was assumed as the opportunity cost for the mill. An assessment of future prices of electricity was beyond the scope of this paper. Particularly in the case of amine, the cost of heat has an important impact and may vary significantly depending on the energy balance of the sugarcane mill. The case presented here is likely to be a worst-case scenario with the highest cost for heat. On a case-by-case basis lower heat costs may be achieved.

Fig. 4 also shows some price benchmarks for biomethane. Two relevant benchmarks are pipeline natural gas and diesel oil used in transportation. Pipeline natural gas is the most obvious target market as biomethane and natural gas are perfectly interchangeable. Two important sources of natural gas in Brazil in 2013 were the natural gas imported from Bolivia and liquified natural gas (LNG), imported from diverse countries. These two sources accounted for about 50% of the total natural gas supplied in the Brazilian market in 2013 [33]. The average price of the Bolivian gas in that same year was US$8.88/MMbtu, and that of imported LNG was US$13.98/MMbtu [33]. Those prices are equivalent to R$19.78/GJ and R$31.14/GJ, respectively, at the average dollar exchange rate of 2014. At a cost in the range of R$31/GJ, biomethane would be competitive with imported LNG at 2013 prices, but not so much with Bolivian natural gas, with a price gap of about 50%.

In the case of diesel oil, the sugarcane mill itself is an important potential end-user due to the large volumes of diesel used in its operations. The average consumer price of diesel, including taxes, was US$1.067/L in 2014, according to the Brazilian energy balance [34]. That corresponds to R$70.59/GJ. Taxes can be estimated at around 40% of the final price, thus the price excluding taxes would be R$42.35/GJ. That shows a great potential for biomethane, to compete with diesel oil in the consumer market. Especially in the case of sugarcane mills, which are consumers of diesel oil and pay the full final price for it. There is significant potential of economy if biomethane is used in the sugarcane mills, in substitution for diesel oil, by using diesel-gas engines in trucks and agricultural machinery.

4.3. Economies of scale

The impact of scale on the cost of biomethane was also considered. To that end, the unit cost of biomethane of three other project scales were calculated. The plant sizes considered were 43, 87 and 260 million liters of ethanol per season, corresponding to 750, 1500 and 4500 Nm^3/h of biomethane, respectively.

The cost scaling was performed using different criteria depending on the cost item under consideration. Engineering costs were considered fixed given the small difference in plant sizes and resulting engineering hours. For the anaerobic reactor, price quotations for different sizes of plants were used to derive a scaling factor equal to 0.95. In the case of the CO_2 removal equipment, price quotations were used for each different plant size. The H_2S removal units were scaled based upon the sulphur removal load using a scaling factor equal to 0.6. Blowers, compressors and pumps were scaled based on the nominal power of the equipment using a scaling factor equal to 0.6. Labour costs were considered constant, given that the difference in plant size does not justify additional work force. Consumables and maintenance costs were scaled linearly, using a factor equal to one, according to plant size. These criteria were based on the expert judgement of the authors and interviews with plant operators and equipment suppliers.

The resulting unit costs are presented in Fig. 6 and clearly show increasing economies of scale at those project capacities. The results also point out that the likely minimum scale for vinasse-to-biomethane projects aiming at diesel substitution is at distilleries with a capacity of at 87 million liters of ethanol per season. That translates into sugarcane mills of 2 million tonnes of sugarcane per season for an annexed distillery and 1 million tonne per season for an autonomous distillery. If the aim is to target natural gas markets, 174 million liters per season seems like the minimum capacity for competitive prices (assuming that the full potential for biomethane production is exploited), i.e. sugarcane mills of 4 million tonnes of sugarcane per season for an annexed distillery and 2 million tonnes per season for an autonomous distillery.

5. Analysis and conclusions

The difference between the lowest and highest unit costs is in the order of 14% for the reference case. If the allowances are included, all five technological routes for biogas upgrading overlap in the cost range between R$30/GJ$_{HHV}$ and R$34/GJ$_{HHV}$, which shows an equivalence of the options, as far as cost is concerned, with a positive outlook for organic-physical scrubbing and a negative outlook for pressure swing adsorption. In fact, pressure swing adsorption is the option that lies farther from the average of the remaining options. Nevertheless, no option should be ruled out based on costs and this study alone, given the uncertainties involved.

In that regard, it should be noted that the investment costs on biogas production, the operational costs associated with personnel and maintenance, the operational costs associated with water consumption (in the dessulphurization unit), and the operational costs of H_2S removal seem not to have a significant impact on the differences amongst the five options as they are all about the same, without including the effect of biomethane yield. Conversely, the investment costs on biogas upgrading, the operational costs associated with electricity and heat consumption, the operational costs of CO_2 media replacement, and the methane losses (i.e. biomethane yield) have a significant impact on the costs of the different options.

Investment costs associated with biogas upgrading were very similar for amine scrubbing, membrane separation and pressure
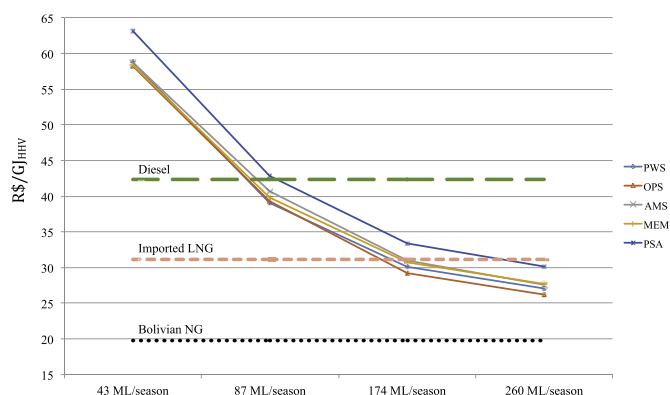

Fig. 6. Impact of scale on the unit cost of biomethane.

swing adsorption. However, pressure water scrubbing and, in particular, organic-physical scrubbing presented relatively lower investment costs, mainly due to lower equipment prices. It is uncertain whether this difference is statistically significant, given the small sample of technology providers interviewed, but this is one of the sources of cost differentiation in this study.

Electricity consumption is a function of mainly two operations: biogas compression and solvent chilling and pressurization. When it comes to gas compression, the technologies that operate the CO_2 removal step at a lower pressure (such as the case of amine scrubbing) have an advantage over the others because electricity is not wasted in the compression of CO_2. As regards solvent pressurization and cooling, organic-physical scrubbing and pressure water scrubbing are in disadvantage. In particular, pressure water scrubbing faces a significantly higher electricity use for solvent pressurization and cooling due to the very large volumes of water that need to be processed in the plant, as compared to relatively much smaller volumes of organic solvent. Amine scrubbing, pressure swing adsorption and membrane technologies do not require solvent pressurization and cooling, with amine scrubbing resulting in the lowest electricity cost of all technologies due to its very low operating pressure. Nevertheless, the operational costs associated with heat consumption overshadow the lower costs of electricity use offered by the amine technology, at least at the relative prices considered in this study.

Another source of difference between the costs is the operational costs associated with CO_2 media replacement. The impact of that cost is most significant in the case of membrane separation and pressure swing adsorption, whereas for the other options, media replacement seems not to be an important cost category. In particular, those costs weigh heavily on membrane separation. If it were not for the high costs involved in membrane replacement, membrane separation could potentially move to first or second place in the rank. It has been reported by technology providers that due to the novelty of membrane separation in the biogas industry, in particular in large scale projects, those costs may be overestimated at present. Perhaps after the learning phase goes by, those costs can be reassessed.

The other component that explains the differences in costs amongst the technologies is the biomethane yield, which is a function of methane losses during CO_2 removal. That impact is more evident in the case of pressure swing adsorption, due to its much higher methane losses (8% as compared to values below 3% for the other options). In spite of a relatively small difference in absolute costs between pressure swing adsorption and the other technological routes (~6%), pressure swing adsorption appears at a much worse position when the unit cost is considered, relative to biomethane yield, with a difference of ~14%. An improvement in its methane losses would certainly place pressure swing adsorption at a much better position as compared to its alternatives.

Overall, the decision between one alternative and another is likely to be made based on other factors than cost alone, especially given the close range of unit costs. Aspects such as availability of local suppliers, maintenance and operational efforts involved in the technology, safety and environmental risks, and personal preferences are all factors that would influence the decision. Perhaps, in this regard, technologies such as pressure swing adsorption and membranes may have a slight advantage due to the inexistence of liquid media involved in the upgrading process. And pressure water scrubbing, which is based on water, may be more attractive than amine and organic-physical scrubbing, which involve the use of industrial chemical solvents.

Irrespective of the biogas upgrading technology, the technical-economic assessment undertaken for the example vinasse-to-biomethane project indicates that the anaerobic digestion of vinasse is a promising option for harnessing the energetic potential of vinasse and adding value to the sugarcane industry production chain. The unit cost of biomethane resulted in the same price range as that of some alternative fuels, such as natural gas and diesel oil. Biomethane is definitely competitive with diesel oil, at a cost base, and may also be competitive with natural gas if marginal, more expensive sources of natural gas, such as imported LNG, are considered.

The results also showed that the most likely scenario is the implementation of vinasse-to-biomethane projects at larger sugarcane mills, those with capacities above 2 million tonnes of sugarcane for annexed distilleries and above 1 million tonnes, for autonomous distilleries, if diesel is the target fuel for displacement. Larger mills would be required if natural gas is the target market.

Given the risks associated with the large-scale development of such a novel activity in the industry, it is worth considering options for further cost reductions. A better assessment of H_2S generation and removal techniques would certainly be a boon in that regard, as H_2S removal contributes significantly to costs and, amongst the process steps, it is probably the least studied. It remains to be seen whether improvements on the costs of membrane replacement and a reduction in methane losses will further improve the costs of membrane separation and pressure swing adsorption, respectively. As regards amine scrubbing, the cost of heat, which is very sensitive to the specific conditions of each mill, deserve better attention on a case-by-case basis due to its impact on the final cost.

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Appendix A. Derivation of Equation (1) for CH_4 generation

The COD (Chemical Oxygen Demand) balance in Fig. 1 results in the following equation $Q_{CH4} = Q_{in} - Q_{out} - Q_{bio}$. The volume of methane at standard temperature and pressure (V_{CH4}) can be determined from the amount of COD converted into methane (Q_{CH4}) using the chemical equation $CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$, which shows that the COD of methane is 64 g (2 mol) of oxygen per mol of methane. Therefore, each mol of methane (i.e. 22.4 L at standard temperature and pressure) is equivalent to the removal of 64 g of COD. Thus, V_{CH4} and Q_{CH4} are such that $Q_{CH4} = 64/22.4 \cdot V_{CH4}$. The COD converted into microbial biomass (Q_{bio}) is usually represented as a fraction (β) of input COD (Q_{in}) so that $Q_{bio} = \beta \cdot Q_{in}$. Furthermore, let the COD removal efficiency be defined as $\eta = (Q_{in} - Q_{out})/Q_{in}$. It, then, results that $V_{CH4} = 0.35 \cdot (\eta - \beta) \cdot Q_{in}$. Sulphates reduction can impair the production of methane by competing with methanogenesis for intermediate anaerobic digestion compounds. A first estimate of the impact of sulphate reduction in methane generation can be derived from the following chemical equation: $2H^+ + SO_4^- \rightarrow H_2S + 2O_2$, which shows that the reduction of 1 mol (96 g) of sulphate generates 2 mol of oxygen that will potentially remove an equivalent amount of COD. In other words, each mol of H_2S, or 22.4 L at standard temperature and pressure, is equivalent to the removal of 64 g (2 mol) of the total COD available in the vinasse (Q_{vin}). Therefore, a penalty factor equal to $64/22.4 \cdot V_{H2S}$ (where V_{H2S} is the volume of H_2S) was be deduced from the COD available in the vinasse Q_{vin}, which resulted in $Q_{in} = Q_{vin} - 64/22.4 \cdot V_{H2S}$. Substituting in the previous equation results $V_{CH4} = 0.35 \cdot (\eta - \beta) \cdot (Q_{vin} - 64/22.4 \cdot V_{H_2S})$. Substituting V_{H2S} from the equation in Appendix B, results Equation (1):

$$V_{CH4} = 0.35 \cdot (\eta - \beta) \cdot (Q_{vin} - 64/96 \cdot S_{vin}).$$

Appendix B. Derivation of Equation (2) for H$_2$S generation

The volume of sulphidric gas produced during the anaerobic digestion of vinasse (V_{H2S}) depends on a number of factors and is hard to be predicted. A conservative (higher-end) figure can be estimated based on the simplifying assumptions that all sulphur in the substrate will be reduced to sulphide; no other sources of sulphur are present other than the sulphur in the substrate; and that all sulphide will end up as sulphidric gas. Based on that, each mol of sulphate (96 g) will produce one mol of sulphidric gas (22.4 L), so that the volume of sulphidric gas at standard temperature and pressure (V_{H2S}) can be estimated from the sulphate content in vinasse (S_{vin}) as $V_{H2S} = 22.4/96 \cdot S_{vin}$.

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