**Optimal technology selection for the biogas upgrading into biomethane**

**Edgar Martín-Hernández, Lidia S. Guerras, Mariano Martín[[1]](#footnote-1)**

Department of Chemical Engineering. University of Salamanca. Plz. Caídos 1-5, 37008. Salamanca (Spain)

S1 Literature review

The literature available referred to CO2 capture is extensive, not only for combustion processes also for the particular case of biogas upgrading, there exists a lack of literature regarding the selection of the most appropriate biogas upgrading technology. It should be noted that the goal of this work it is not limited to the optimization the biogas production and upgrading processes, which it is implicitly done in the work, but the aim pursued is to determine the optimal biogas upgrading technology among the different feasible processes.

This lack in the literature was detected after a literature review, carried out with special emphasis in specific studies about biogas upgrading. On one hand, as a result of the literature review made, it could be concluded that the literature about reviews of biogas upgrading processes is extensive, even when only recent works are considered, as it is shown in Table 1. On the other hand, just a few recent works are approaching to carry out a systematic comparison of the processes, such as the work of Collet et al. (2017), where a comparison of several CO2 capture technologies using experimental data from other studies is presented, but without integrating and optimizing the biogas production and upgrading processes, or the study of Vo et al. (2018), where simulations of biogas upgrading processes limited to amine scrubbing and biological methanation are carried out, not including some essential upgrading technologies such as membranes or PSA. Other works, including but not limited to, Capra et al., 2018; Curto and Martin, 2019; and Gilassi et al,. 2019, determine the optimal biogas upgrading process but analyzing only one technology in each case (amines scrubbing, biogas methanation, and membrane separation). In the work presented 5 technologies have been evaluated considering both heuristic and mathematical modelling stages (biogas methanation, water scrubbing, pressure swing adsorption systems, amines scrubbing, and membranes separation systems).

Table 1: Relevant literature for biogas upgrading

|  |  |
| --- | --- |
| **Author** | **Study developed** |
| Angelidaki et al., 2018 | Review of biogas upgrading technologies |
| Awe et al., 2017 | Review of biogas upgrading technologies |
| Bauer et al., 2013 | Review of biogas upgrading technologies |
| Khan et al., 2017 | Review of biogas upgrading technologies |
| Miltner et al., 2017 | Review of biogas upgrading technologies |
| Sun et al., 2015 | Review of biogas upgrading technologies |
| Zhou et al., 2017 | Review of biogas upgrading technologies |
| Collet et al., 2017 | Techno-economic and life cycle assessment of biogas upgrading |
| Ferella et al., 2019. | Techno-economic assesment of strategic plans for biogas upgrading plants |
| Toledo-Cervantes et al., 2017 | Comparison of photosynthetic and physico-chemical biogas upgrading processes |
| Vo et al., 2018 | Simulation of amine scrubbing and biological methanation |
| Capra et al., 2018. | Optimal selection of amine scrubbing process for biogas upgrading |
| Curto and Martin, 2019 | Optimal selection of renewable biogas methanation processes |
| Filipetto et al., 2019 | Optimal selection of membrane separation process for biogas upgrading |
| Gilassi et al,. 2019 | Optimal selection of membrane separation process for biogas upgrading |
| Morero et al., 2017 | Optimal selection of amine scrubbing process for biogas upgrading |

S2 Modelling approach

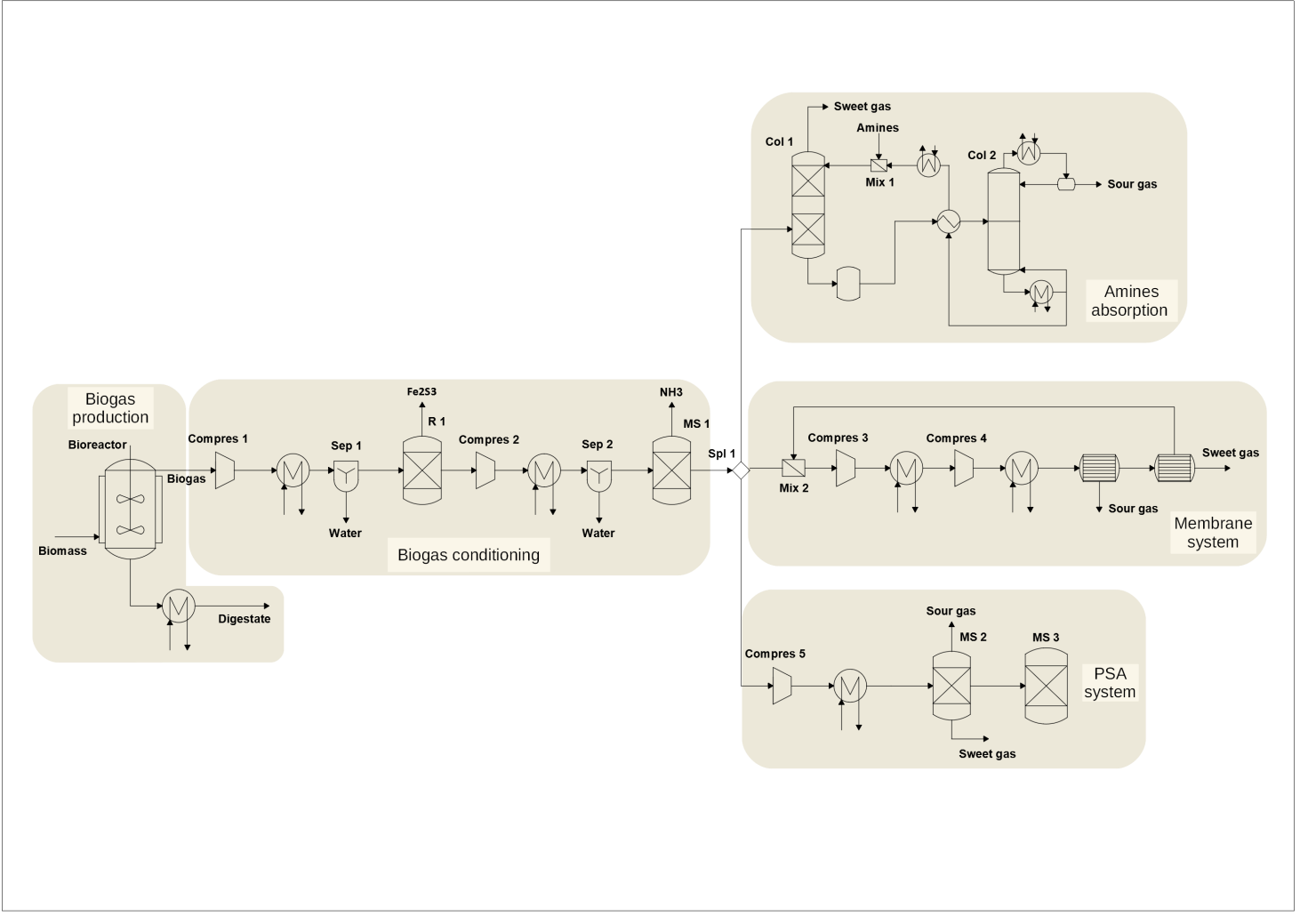


Figure 1: Scheme of the proposed superstructure for biogas upgrading into biomethane

S2.1 Amines

The CO2 absorption systems using amines typically operate at low temperatures, around 25-30ºC, and partial pressures above 0.05 bar, reaching removal yields of 90%-95% (Zhang and Chen, 2013). In contrast to post-combustion gases, which contains large amounts of nitrogen from air, biogas is composed mainly by methane and CO2, resulting in higher carbon dioxide partial pressures and in the need for lower operating pressures. CO2 partial pressures above 0.1 bar have been assumed to secure high removal yields (Zhang and Cheng 2013), resulting in the need to operate at total pressures around 1-1.5 bar to secure the appropriate CO2 partial pressures (Movagharnejad and Akbari, 2011; Xue et al, 2017).

The total amount of solution of amines needed to absorb the CO2 from the gas stream is calculated as a function of the amount of sour gases eliminated, Eq. 1. According to GPSA (2012), the concentration of solution and the correction factor (GPSA) depend on the amine used, as is shown in Table 3. The flow of the amine solutions depends on their solubility. Thus, for a fixed removal ratio, the amines solution flow required changes from one to another. It is considered there is no methane absorption in the amines flow, therefore all biomethane entering in the unit leaves the column and it is sent to storage.

 (1)

The solution of amine used in column 1 comes from two sources, as it shown in Eq. 2: the amines from the regeneration column (column 2 in Figure 1), and some make-up solution to replace the amine losses with the gas outlet stream in the regeneration column. Additionally, to mix the two streams of amines at the same temperature, a heat exchanger is used to adjust the temperature of the amine flow stream which leaves the regeneration column to 25 ºC,

 (2)

The mixing of the two amines streams is assumed to be adiabatic. The energy balance to the heat exchanger is as follows, Eq. 3.

 (3)

Where *Famine* is referred to the total mass flow of the amines stream, and *qamine* the heat flow ratio based on the rules of thumb reported by GPSA (2012). The values are collected in Table 4.

The assumed CO2 removal efficiency in the absorption column is 0.95, based on literature data (Zhang and Chen, 2013). The sour gas is absorbed by the amines in the absorption column, being withdrawn from the gas phase. The absorption is an exothermic process. Therefore, this energy is to be refrigerated, and the operation of the column is isothermal, Eq. 4. The heats of reaction are shown in Table 3.

 (4)

The biomethane leaves the column and it is sent to storage. On the other hand, the amine with the CO2 absorbed is sent to the regeneration column (column 2). The solution is heated up before being fed to column 2 through the heat transfer from the amines stream leaving the reboiler of the regeneration column with the aim of improving the desorption process. Rules of thumb reported in the literature are used to compute the energy involved in the heat exchanger using Eq. 6, considering the corresponding values of *qamine* collected in Table 4.

The operation of column 2 is also based on rules of thumb (GPSA, 2004; GPSA, 2012) including the estimation of the energy consumption in the reboiler and the cooler refrigeration requirements. According to the literature, the inlet temperature to column 2, *TCol2*, is equal to 93 ºC, whereas the temperature of the outlet amines stream, *Tbottom*,is equal to 125 ºC, while at the condenser, the temperature, *Ttop*,is equal to 54 ºC. Thus, the energy balances of the stream entering column 2 are described in Eq. 5.

 (5)

From the reboiler, the regenerated amine is cooled down heating up the feed to the column, Eq. 6.

 (6)

The gas leaving the regeneration column is saturated with the amines aqueous solution, producing the losses of amines which have to be replaced before be recirculated to the absorption column. In order to calculate these losses, humidification models are used. First, the saturation pressure of the amine solution is calculated, Eq. 7. Then, the specific humidity of the gaseous stream is computed to determine the amount of amines solution accompanying the CO2 gaseous stream which leaves the regeneration column, Eq. 8, where the operating pressure of the regeneration column, *PCol2*, is equal to 1.7 bar (GPSA, 2004). As an approximation, it is assumed that the amine solution behaves as water. The amount of amine lost with the sour gases is the one to be fed as fresh amine.

 (7)

 (8)

Three different amines are selected aiming a high selectivity, MEA, DEA, and MDEA. Table 2 shows the parameters used in the amines absorption modelling for each amine considered (GPSA, 2004&2012).

Table 2.- Amine properties for CO2 capture GPSA (2004 & 2012)

|  |  |  |  |
| --- | --- | --- | --- |
|  | MEA | DEA | MDEA |
| Gas pickup mol/mol amine | 0.35 | 0.35-0.65 | 0.2-0.55 |
| Solution concentration (wt %) | 20 | 35 | 45 |
| Heat of reaction /BTU/lb CO2) | 620-700 | 580-650 | 570-600 |
| GPSA | 0.35 | 0.5 | 0.375 |
| Density | 1.01 | 1.05 | 1.045 |
| Cost (€/kg)a | 1.3 | 1.32 | 3.09 |
| Molecular Weight | 61 | 105 | 119 |

Table 3.- Amine regeneration heat loads (GPSA, 2004)

|  |  |
| --- | --- |
|  | Duty (BTU/hr) |
| Reboiler | 72000 GPM |
| Condenser | 30000 GPM |
| Amine feed to distillation | 45000 GPM |
| Amine cooler | 15000·GPM |

The biomethane production model through amines scrubbing includes the units described in sections 3.2.1 and S1.1. The NLP problem consists of 288 equations and 953 variables per amine evaluated and is solved using a multistart initialization approach with CONOPT as the preferred solver where the main decision variable are the pressures temperatures and flow rates.

S.2.2 PSA

The stream of gases passes through the bed of zeolites and the carbon dioxide is captured by adsorption. The system consists of the compression train and the zeolite beds.

The adsorption capacity of the zeolites is directly related to the partial pressure of the CO2. Therefore, a system of compressors with intermediate cooling is implemented to determine the optimal operating pressure. As it is described previously, the compressors are modelled assuming polytropic behavior, with a polytropic coefficient *z* of 1.4 and an efficiency of the compression stages of 0.85, Eq. (9).

 (9)

The removal ratio is given by the breakthrough curve of the adsorbent bed. Based on experimental data (Hauchhum and Mahanta, 2014) for adsorption stages below 20 min, the CO2 removal yield at the PSA,  is assumed to be 98%, containing below 2% CO2 at the outlet stream, (Ferella et al 2017). The mass balance at the bed is as shown in Eq. 10.

 (10)

To compute the mass of bed the adsorption capacity of the bed is evaluated. Based on experimental data, the Langmuir isotherm is the adsorption model considered for this process, Eq. 11, since this is the one that best fits the performance of the zeolite 13X - CO2 system (Hauchhum and Mahanta, 2014).

 (11)

Where the parameters *qm* and *K* depend on the adsorbent material. Considering zeolite 13X, the effect of the operating temperature, within the range of 25ºC-60ºC, for both parameters can be correlated using data available in the literature, Eq. 12a (Hauchhum and Mahanta, 2014).

 (12a)

As result of the breakthrough curve for the zeolite 13X – carbon dioxide system, the operating time must be below 20 min so that the exit gas (methane) contains only traces of CO2. Similarly, correlations are developed for zeolite 4A, Eq. 12b. The operating time of zeolite 4A must be below 20 min (Hauchhum and Mahanta, 2014). Note that the expression for the computing of the constant K in the Langmuir correlation is the same for both adsorbents evaluated.

 (12b)

However, the adsorption capacity decays cycle after cycle until it stabilizes around 65% of the initial capacity computed by Eq. 13 (Hauchhum and Mahanta (2014)). Therefore, a corrected value for q is applied to compute the amount of zeolite used in the PSA system, as it can be shown in Eq. 14, where the CO2 removal yield at the PSA,  is assumed to be 98%, containing below 2% CO2 at the outlet stream, (Ferella et al 2017), and  is equal to 20 min, based on the results of (Hauchhum and Mahanta, 2014). Furthermore, a lifetime of the zeolites bed of 5 years has been considered based on data reported by Xiao et al. (2013). Thus for the cost, we considered that over the 20 years life time of the plant, 5 beds will be used.

 (13)

In the case of the PSA system, the objective function is shown in Eq. 22. To estimate the cost of the PSA system, Eq. 23, it is assumed that the zeolite bed loses efficiency over time, resulting in a lifetime of 5 years before it needs to be replaced (Xiao et al., 2013). As the plant life is considered 20 years, the zeolites bed must be replaced 4 times during the plant life, *NCycle*.

 (22)

 (23)

The cost of the zeolites considered is 5$/kg for both zeolite 13 X and zeolite 4A (Xiao et al., 2013).

The production of biomethane through pressure swing adsorption includes the units described in sections 3.2.1 and S1.2 consisting of an NLP problem of 283 equations and 828 variables that is solved similarly as in the case of the selection of amines where the main decision variables are the operating pressure of the adsorption tower and the size of the bed.

S2.3 Membranes

The membrane system considered is dual-stage membrane systems with single compression stage before the membrane system and no recompression stage between membrane units, have been deemed as the most economic under a wide range of feed compositions (Kim et al., 2017). The compressor is modelled as presented above, assuming polytropic compression of the gas, Eq. 9. Each membrane module is modelled using mass balances, considering the permeate and retentate streams, Eqs. 14 and 15, and the flux of the gases through the membrane, that is a function of the concentration gradient between both sides of the membrane, Eq. 17 (Fernandes Rodrigues, 2009). The flux is the parameter which allows computing the area of the membrane, as it is shown in Eq. 16, based on the permeability of the membrane, Eq. 19. As the driving force in the membrane separation process is the concentration gradient, the removal of CO2 results in a change in the composition of the stream along the membrane, leading to a change in the driving force which controls the process. Therefore, an average molar fraction between the feed and the retentate composition is used to compute the separation driving force, Eq. 18.

 (14)

 (15)

 (16)

 (17) (18) (19)

Where  is the membrane thickness, and Permi the permeability of the component *i*. The usual membrane thickness for industrial units is equal to 30 nm. Three different membrane materials are selected aiming a large CO2 permeability, low methane permeability, and therefore, high selectivity; cellulose acetate, polyamide, and polycarbonate. Table 4 shows the permeabilities for each membrane material considered in the model at 25 ºC (Vrbová and Ciahotný, 2017). The solution of the optimization problem will yield intermediate conditions to assure natural gas composition of the biomethane.

Table 4.-Gases permeability (Vrbová and Ciahotný, 2017)

|  |  |  |
| --- | --- | --- |
|  | Permeability (Barrer) | |
| Polymer | CH4 | CO2 |
| Cellulose acetate | 0.21 | 6.30 |
| Polycarbonate | 0.13 | 4.23 |
| Polyimide | 0.25 | 10.7 |

Finally, the simplified profit objective function for the membrane separation system, Eq. 20, considers cost of the gas compression and the amortization of the investment costs of the membranes, Eq. 21.

 (20)

 (21)

A value of 50 $/m2 will be used based on the literature (Kim et al., 2017). Considering the plant life equal to 20 years, the membranes with a typical lifetime of 4 years must be replaced 5 times during the plant life, *NMembranes* (Scholz et al., 2015).

The biogas production and upgrading considering a membranes separation system is formulated as an NLP problem considering the units described in sections 3.2.1 and S1.3 consisting of 299 equations and 869 variables for each material evaluated where the major decision variables are the flows, operating pressures, temperatures and the area required by the membrane units.

S3 Economic evaluation.

The evaluation of the cost for biomethane production is also estimated from different wastes. The production and investment costs are estimated. The CAPEX, or investment cost, is estimated based on Sinnot and Towler (2009) that presents a factorial method. This method estimates the CAPEX usinf as reference the cost of all the major units of the flowsheet.

The sizing of all the units involved in the flowsheet, see Figure 2, is carried out following the method presented in Martín and Grossmann (2011), and updated by Almena and Martin (2015). However, the following considerations for some specific units have been assumed:

-For the digester cost, assumed to be 365 €/m3 (Taifouris and Martín, 2018).

-The cost estimation of the columns for the amines is carried out using the rules reported in GPSA (2012). The diameter for the amine contactor can be estimated as given in Eq. 22.

 (22)

While the dimeter of the regenerator column can be estimated as given in Eq. 23.

 (23)

We use the same factors as presented in Davis and Martin (2014) to estimate the investment, CAPEX, of the facility for further comparison with other renewable based methane production processes. The costs for piping, isolation, instrumentation, and utilities infrastructure are estimated with respect to the equipment cost as 20%, 15%, 20% and 10% of its value respectively. Land and buildings costs are estimated to be 8 M€. These items add up to the fixed cost. The fees represent 1% of the fixed cost, other administrative expenses and overheads, and the plant layout represent 10% of the direct costs (fees plus fixed capital) and 5% of the fixed cost, respectively. Finally, the plant start-up cost represents 5% of the investment.

Furthermore, the production costs of biomethane are estimated using also the factorial method with the coefficients presented and validated in Davis and Martín (2014). For the average annual cost, the labour costs (0.5% of investment), equipment maintenance and raw materials (1% of fix costs), amortization (linear with time in 20 years), taxes (0.5% of investment), overheads (1% of investment), and administration (5% of labour, equipment maintenance, amortization, taxes and overheads) are considered. The utilities, particularly cooling water, power, and steam, are taken from the mass and energy balances performed in the superstructure model. Heat integration is carried out to reduce the utilities consumption. The cost of electricity is 0.06€/kWh. The hot compressed gas is used to evaporate the water and ammonia to prepare the digestate to be sold as fertilizer. However, the credit for the digestate to be used as fertilizer is difficult to be estimated. It depends on the market that typically is saturated. Although the cost is around 0.48 €/kg, only a fraction is typically obtained. To be conservative, a value of one third of this one is assumed to calculate the benefits received by the sold of the digestate produced (León and Martín, 2016). In addition, the cost of the digestate for the methane to be competitive with other resources, targeting 5 €/MMBTU (0.17€/Nm3), is computed.

S4 Scaling-up study

The scaling-up study, based on the previous work of Sánchez and Martín (2018), is performed in the following stages. Firstly, the capital cost of the different units is correlated as a function of a design variable, such as the membrane area for membranes systems or the column sizes in function of the flow processed. This variable will be denoted as scaling variable. The scaling variable of each equipment is directly related to the processing capacity of the facility studied through the mass or energy flows involved in that particular unit. The limitations in the size of equipment are considered in the scaling-up study. When an equipment excess the size defined by the rules of thumb, that unit is duplicated, affecting to the cost estimation. In a second stage, after defined the scaling variables for all units and the relation between these units, the mass and energy flows and the processing capacity of the facility, different capacities are evaluated, calculating the mas and energy balances, and estimating the sizes of the equipment, and the number of units in case some equipment needs to be duplicated. In a third stage, the capital and operating costs are estimated considering the equipment sizes and number of unit of each equipment using the factorial method presented in Sinnot and Towler (2009) and the estimation of the unit costs presented in Almena and Martín (2015), as described in the section S3 of the Supplementary Material.

**References:**

Almena, A., Martín, M., 2016. Technoeconomic analysis of the production of epichlorohydrin from glycerol. Ind. Eng. Chem. Res. 55, 3226–3238. https://doi.org/ 10.1021/acs.iecr.5b02555

Angelidaki, I., Treu, L., Tsapekos, P., Luo, G., Campanaro, S., Wenzel, H., Kougias, P.G., 2018. Biogas upgrading and utilization: Current status and perspectives. Biotechnology Advances 36, 452-466.

Awe, O.W., Zhao, Y., Nzihou, A., Minh, D.P., Lyczko, N., 2017. A Review of Biogas Utilisation, Purification and Upgrading Technologies. Waste Biomass Valor 8, 267–283.

Bauer, F, Hulteberg, C., Persson, T., Tamm, D., 2013. Biogas upgrading - Review of commercial technologies. (SGC Rapport; Vol. 270). Svenskt Gastekniskt Center AB.

Capra, F., Fettarappa, F., Magli, F., Gatti, M., Martelli, E., 2018. Biogas upgrading by amine scrubbing: solvent comparison between MDEA and MDEA/MEA blend. Energy Procedia 148, 970–977.

Collet, P., Flottes, E., Favre, A., Raynal, L., Pierre, H., Capela, S., Peregrina, C., 2017. Techno-economic and Life Cycle Assessment of methane production via biogas upgrading and power to gas technology. Applied Energy 192, 282-295.

Curto, D., Martín, M., 2019. Renewable based biogas upgrading. Journal of Cleaner Production 224, 50-59.

Davis, W., Martín, M., 2014a. Optimal year-round operation for methane production from CO2 and water using wind energy. Energy 69, 497-505. <https://doi.org/10.1016/j.energy.2014.03.043>

Ferella, F., Cucchiella, F., D'Adamo, I., Gallucci, K., 2019. A techno-economic assessment of biogas upgrading in a developed market. J. Clean. Prod., 210, 945-957.

Ferella, F., Puca, A., taglieri, G., Rossi, L., Gallucci, K., 2017. Separation of carbon dioxide for biogas upgrading to biomethane. J. Clean. Prod. 164, 1205-1218. <https://doi.org/10.1016/j.jclepro.2017.07.037>

Fernandes Rodrigues, D., 2009. Model development of a membrane gas permeation unit for the separation of Hydrogen and Carbon Dioxide. MSc Thesis. Instituto Superior Technico, Lisbon.

Filipetto, D., Capra, F., Magli, F., Gatti, M., Martelli, E., 2019. Optimization of semi-permeable membrane systems for biogas upgrading. Computer Aided Chemical Engineering 46, 1693-1698.

Gilassi, G., Taghavi, S.M., Rodrigue, D., Kaliaguine, S., 2019. Optimizing membrane module for biogas separation. International Journal of Greenhouse Gas Control 83, 195-207.

# GPSA Engineering Data Book FPS version 21-10, 2004.

# GPSA Engineering Data Book FPS version 21-10, 2012.

Hauchhum, L., Mahanta, P,. 2014. Carbon dioxide adsorption on zeolites and activated carbon by pressure swing adsorption in a fixed bed. Int. J. Energy Environ. Eng. 5, 349–356. https://doi.org/10.1007/s40095-014-0131-3

Khan, I.U., Othman, M.H.D., Hashim, H, Matsuura, T., Ismail, A.F., Rezaei-DashtArzhandi, M., Azelee, I.W., 2017. Biogas as a renewable energy fuel – A review of biogas upgrading, utilisation and storage. Energy Conversion and Management 150, 277-294.

Kim, M., Kim, S., Kim, J., 2017. Optimization-based approach for design and integration of carbon dioxide separation processes using membrane technology Energ Proc. 136 336–341. <https://doi.org/10.1016/j.egypro.2017.10.284>

León, E., Martín, M., 2016. Optimal production of power in a combined cycle from manure based biogas Energ. Conv. Manag. 114 89–99. <https://doi.org/10.1016/j.enconman.2016.02.002>

Miltner, M., Makaruk, A., Harasek, M., 2017 Review on available biogas upgrading technologies and innovations towards advanced solutions J Cleaner Prod. 161, 1329-1337

Morero, B., Groppelli, E.S., Campanella, E.S., 2017. Evaluation of biogas upgrading technologies using a response surface methodology for process simulation. Journal of Cleaner Production 141, 978-988.

Movagharnejad, K., Akbari, M., 2011. Simulation of CO2 Capture Process Int. J. Chem. Mol Eng. 5 (10), 843-847.

Sánchez, A., Martín, M., 2018. Scale up and Scale down issues of renewable ammonia plants: towards modular design. Sust. Prod. Consump. 16, 176-192. https://doi.org/10.1016/j.spc.2018.08.001

Scholz, M., Alders, M., Lohaus, T., Wessling, M., 2015. Structural optimization of membrane-based biogas upgrading processes. J. Memb. Sci. 474, 1-10. <https://doi.org/10.1016/j.memsci.2014.08.032>

Sinnott, R., Towler, G. Chemical Engineering Design. Oxford: Elsevier Ltd. 2009.

Sun, Q., Li, H., Yan, J., Liu, L., Yu, Z., Yu, X., 2015. Selection of appropriate biogas upgrading technology-a review of biogas cleaning, upgrading and utilization. Renewable and Sustainable Energy Reviews 51, 521-532.

Taifouris M., Martín M., 2018. Multiscale scheme for the optimal use of residues for the production of biogas across Castile and Leon. J. Cleaner Prod. 185, 239-251. <https://doi.org/10.1016/j.jclepro.2018.03.018>

Toledo-Cervantes, A., Estrada, J.M., Lebrero, R., Muñoz, R., 2017. A comparative analysis of biogas upgrading technologies: Photosynthetic vs physical/chemical processes. Algal Research 25, 237-243.

Vo, T.T.Q., Wall, D.M., Ring, D, Rajendran, K., Murphy, J.D., 2018. Techno-economic analysis of biogas upgrading via amine scrubber, carbon capture and ex-situ methanation. Applied Energy 212, 1191-1202.

[Vrbová](https://pubs.acs.org/author/Vrbov%C3%A1%2C+Veronika), V., [Ciahotný](https://pubs.acs.org/author/Ciahotn%C3%BD%2C+Karel), K., 2017. Upgrading Biogas to Biomethane Using Membrane Separation. [Energy Fuel 31, 9, 9393-9401](file:///D:\Mariano\Process%20System%20Engineering\Renew\WASTE\151%20Upgrading\Energy%20Fuel%2031, 9, 9393-9401). http://doi.org/10.1021/acs.energyfuels.7b00120

Xiao, G., Webley, P., Hoadley, A., Ho, M., Wiley, D., 2013. Low Cost Hybrid Capture Technology Development Final Report ANLEC R&D, Ref No: 3-0510-0046

Xue, B., Yu, Y., Chen, J., Luo, X., Wang, M., 2017. A comparative study of MEA and DEA for post-combustion CO2 capture with different process configurations. Int. J. Coal. Sci. Technol. 4 (1):15–24. <https://doi.org/10.1007/s40789-016-0149-7>

Zhang, Y., Chen, C.C., 2013. Modelling CO2 absorption and desorption by aqueous monoethanolamine solution with Aspen rate based models. Energ. Proc. 37, 1584-1586. <https://doi.org/10.1016/j.egypro.2013.06.034>

Zhou, K., Chaemchuen, S., Verpoort, F., 2017. Alternative materials in technologies for Biogas upgrading via CO2 capture. Renewable and Sustainable Energy Reviews 79, 1414-1441.

1. Corresponding author. Tel.: +34 923294479

   Email address: mariano.m3@usal.es [↑](#footnote-ref-1)