Membrane Separation Processes for Post-Combustion Carbon Dioxide Capture: State of the Art and Critical Overview
Bouchra Belaissaoui, Eric Favre

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PART 2

Post Combustion CO₂ Capture

Capteur CO₂ en postcombustion

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Membrane Separation Processes for Post-Combustion Carbon Dioxide Capture: State of the Art and Critical Overview

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Abstract — Membrane Separation Processes for Post-Combustion Carbon Dioxide Capture: State of the Art and Critical Overview — Membrane processes have been initially seldom considered within a post-combustion carbon dioxide capture framework. More traditional processes, particularly gas-liquid absorption in chemical solvents, are often considered as the most appropriate solution for the first generation of technologies. In this paper, a critical state of the art of gas separation membranes for CO₂ capture is proposed. In a first step, the key performances (selectivity, permeability) of different membrane materials such as polymers, inorganic membranes, hybrid matrices and liquid membranes, including recently reported results, are reviewed. In a second step, the process design characteristics of a single stage membrane unit are studied. Purity and energy constraints are analysed as a function of operating conditions and membrane materials performances. The interest of multistage and hybrid systems, two domains which have not sufficiently investigated up to now, are finally discussed. The importance of technico-economical analyses is highlighted in order to better estimate the optimal role of membranes for CCS applications.
NOMENCLATURE

\[ E \] Energy requirement (J.ton\(^{-1}\))

\[ p' \] Membrane module upstream side pressure (bar)

\[ p'' \] Membrane module downstream side pressure (bar)

\[ p \] Pressure (bar)

\[ P \] Permeability coefficient (Barrer or mol.m\(^{-1}\).s\(^{-1}\).Pa\(^{-1}\) [SI])

\[ Q \] Flow rate (mol.s\(^{-1}\))

\[ R \] Recovery ratio (-)

\[ R \] Perfect gas constant (8.314 J.mol\(^{-1}\).K\(^{-1}\))

\[ T \] Temperature (K)

\[ x_{in} \] Mole fraction in the feed mixture (-)

\[ y \] Mole fraction in the permeate side (-)

GREEK LETTERS

\[ \alpha^* \] Ideal separation factor (-)

\[ \gamma \] Adiabatic gas expansion coefficient (-)

\[ \psi \] Pressure ratio (-)

\[ \theta \] Stage cut (-)

\[ \eta_C \] Compressor isentropic efficiency (-)

\[ \eta_V \] Vacuum pump efficiency (-)

\[ \eta_E \] Expander efficiency (-)

INTRODUCTION

The design of an efficient, low energy intensive and low cost CO\(_2\) capture unit is of utmost importance for the deployment of Carbon Capture and Storage (CCS) technologies at industrial scale (Davidson and Metz, 2005). This statement particularly holds for the post-combustion framework. In that case, flue gases to be treated are under atmospheric pressure conditions and carbon dioxide is diluted in nitrogen with a volume fraction typically between 4\% (i.e. gas turbine) and 15\% (coal combustion power plant); these two specificities address a major engineering challenge, especially in terms of the energy requirement of the separation process (Steeneveldt et al., 2006; Herzog, 2001).

Numerous capture technologies are being developed and evaluated for this application such as absorption in liquids, adsorption on solids, cryogenic processes, membranes, solid sorbents (Figuerola et al., 2008). The overall framework is complex and the identification of the most efficient process is subject to controversial debates. Generally speaking, the ideal carbon capture process should offer the following characteristics:

- a high CO\(_2\) over N\(_2\) selectivity, in order to achieve a concentration up to 90 to 99\% CO\(_2\) (before the compression, transport and storage steps);
- a minimal energy requirement, ideally below 2 GJ per ton of CO\(_2\) (thermal basis), so that secondary CO\(_2\) emissions generated by the capture process remain limited;
- a maximal productivity (sometimes also named capacity), which is required in order to minimize the size (footprint) of the capture unit;
- a minimal, if not negligible impact on environment, which can be assessed through criteria such as the waste produced by the capture unit (solids, liquid or gaseous emissions) and water footprint;
- a minimal overall cost of the unit, usually expressed in $ (or €) per ton of CO\(_2\) captured and including the operating (OPEX) plus capital (CAPEX) expenses.

Taking into account the different items listed above, gas liquid absorption in a chemical solvent (such as an aqueous amine solution) is currently considered as the best available and mature technology (Steeneveldt et al., 2006). Amine solvents are used for decades for natural gas treatment. Amines effectively offer a very high selectivity towards CO\(_2\) through carbamate formation; as a consequence a high purity CO\(_2\) stream can be attained after drying the mixture generated in the stripping unit. The energy requirement essentially corresponds to the stripper boiling duty; this is achieved by low pressure steam (typically 2 bar), a largely available energy source in a power plant. The two major drawbacks of the amine absorption process correspond to a too high energy requirement (Steeneveldt et al., 2006) (typically 3 GJ/ton in thermal basis, excluding compression) and the production of wastes: solids (heat stable salts) and traces of gaseous compounds due to amine degradation. New solvents or alternative separation processes which would not show these disadvantages are intensively explored for the second generation of carbon capture technologies.

Compared to other capture processes, membrane separation shows a very high parametric sensitivity, especially in terms of CO\(_2\) content in the feed stream (Belaissaoui et al., 2012a). Additionally, it has been shown that strong limitations occur when 90\% capture ratio and 90\% CO\(_2\) purity are both imposed for a single stage process. More specifically, given the performance of existing membrane materials, the target purity and capture ratio of 90\% can be attained in a single stage membrane module only providing that a flue gas stream with relatively concentrated CO\(_2\) (typically above
20% CO₂) is to be treated (Favre, 2007). Besides, a very large majority of CO₂ sources has a lower CO₂ content, such as 15% for coal power plants or 5% for natural gas turbine flue gases. In these cases, multistage membrane units or hybrid process design are needed, unless a concerted effort is first made to increase the CO₂ concentration through a mixture of Flue Gas Recirculation (FGR), supplementary firing, and combustion using Oxygen Enhanced Air (OEA) (Belaissaoui et al., 2012b).

Multistage membrane units and hybrid process design are discussed in more details in Section 4.

Generally speaking, membrane separation processes have been initially discarded for carbon capture (Davidson and Metz, 2005; Figueroa et al., 2008). Arguments such as a too high energy requirement, the impossibility to develop highly selective materials, too expensive capital costs characteristics (i.e. no economy of scale due to a scale up strategy based on numbering up) have been alternatively proposed in order to justify their inadequacy (Favre, 2007). Nevertheless, the situation evolved these last years with numerous membrane material developments, which suggest the potenti-alities of membrane processes to be reconsidered. A limited number of process design studies has been how-ever reported (Favre, 2007). It is worth to note that promising results have been recently reported on a pilot demonstration unit based on tailor made membrane modules, which has been launched in 2010 for the treatment of real flue gases in a coal power plant (Merkel et al., 2010).

In this context, a critical analysis of the pros and cons of membrane separation processes in a post-combustion capture framework, in light of new material developments, is of interest and is proposed hereafter.

Additionally, different types of design possibilities for membrane processes (single stage, multistage, hybrid process) shown in Figure 1 will be discussed in the next sections. However, the emphasis will be put on a single stage membrane process design.

1 MEMBRANES AND POST-COMBUSTION CARBON CAPTURE: FRAMEWORK

Schematically, the analysis of gas separation membrane processes for CO₂ capture depends on two different types of factors, which will be detailed in the next sections:

– a materials science aspect which addresses the evaluation of the intrinsic separation performances of the membrane. This topic typically requires the structure and/or the mass transfer mechanism(s) taking place in the membrane to be determined;

– a process engineering science aspect, which aims at the determination of the best design and operating conditions for a given membrane material. Classical tools of chemical engineering and separation science are used for this purpose.

Combining materials and process engineering considerations, the energy requirement and the size of the installation, the two key data of technico-economic studies, can be finally obtained (Zolandz and Fleming, 1995).

Generally speaking, the interplay between material and process engineering is a key feature of membrane science. Unfortunately, for CO₂ capture, these two domains are often disconnected: numerous publications report only on material structure and performances, while process design studies sometimes ignore the specificities of membrane materials. The two main objectives of the carbon capture step are the CO₂ purity level (noted y hereafter) and the capture ratio (noted R and usually fixed at 0.9 for post-combustion studies, Deschamps and Pilavachi, 2004). The capture ratio of a single stage membrane unit (Fig. 1a) is obtained through:

\[ R = \theta \cdot \frac{y}{x_{in}} \]  

with the stage-cut (\( \theta \)), corresponding to the fraction of the feed flow (\( Q_{in} \)) which effectively permeates through the membrane on the permeate side (\( Q_p \)), expressed as:

\[ \theta = \frac{Q_p}{Q_{in}} \]  

Any membrane separation process requires a driving force to operate, classically expressed through the pressure ratio, which corresponds to the ratio of the downstream over upstream pressure:

\[ \psi = \frac{P_p}{P} \]  

CO₂ purity (\( y \)) logically depends on the feed composition and the material performances, namely the membrane selectivity \( \alpha \); \( \alpha \) is determined from to the ratio of the CO₂ permeability over N₂ permeability under steady state conditions. In a large majority of cases, only the experimental permeabilities of the pure compounds are known, leading to the so-called ideal selectivity of the material:

\[ \alpha^* = \frac{P_1}{P_j} \]  

It is important to note that the effective purity produced by a membrane unit does not depend only on the
membrane selectivity, but is also affected by the module operating conditions, especially the pressure ratio ($\Psi$). Additionally, the capture ratio ($R$) also affects the purity level, due to mass balance constraints. This qualitative analysis illustrates how material and process variables are closely interconnected, as discussed hereafter.

2 MEMBRANE MATERIALS FOR CO$_2$ CAPTURE: STATE OF THE ART

A very large number of membrane materials has been investigated for gas separation applications for decades. In terms of industrial applications however, a very limited number of materials, almost exclusively polymers, is used. For CO$_2$ capture, a great number of studies have been reported on different materials such as polymeric, inorganic, hybrid organic-inorganic and facilitated transport membranes (Ebner and Ritter, 2009; Luis et al., 2012). Depending on the material characteristics, different transport mechanisms for the gaseous compounds (i.e. CO$_2$ and N$_2$ for post-combustion capture studies) summarized in Table 1, can take place (Koros, 1993; Baker, 2004).

For a given separation (such as CO$_2$/N$_2$ in post-combustion), a membrane material should ideally combine two target performances (Zolandz and Fleming, 1995):

- a high selectivity (i.e. ratio of the two compounds steady state permeability, Eq. 4), in order to achieve the purity target, ideally in a single stage process;
- a high permeability, in order to minimize the surface area required for a given application. Permeability of a dense polymeric material is usually expressed in Barrer, the most common unit, and the ratio of the permeability to the membrane thickness, usually named permeance is the most relevant characteristics for engineering calculation. For membrane gas
<table>
<thead>
<tr>
<th>Material type</th>
<th>Example</th>
<th>Separation mechanism</th>
</tr>
</thead>
<tbody>
<tr>
<td>Microporous inorganic membrane (silica, carbon, etc.) with pore diameter ranging between permeants kinetic diameter and mean free path</td>
<td><img src="image1" alt="Microporous inorganic membrane" /></td>
<td>Knudsen flow&lt;br&gt;Very low selectivity for CO₂/N₂</td>
</tr>
<tr>
<td>Dense skin polymeric membrane</td>
<td><img src="image2" alt="Dense skin polymeric membrane" /></td>
<td>Solution-diffusion, upper performances predictable by trade-off analysis. Industrial membrane gas separations make use of this type of materials</td>
</tr>
<tr>
<td>Mixed Matrix Membranes (continuous polymeric phase with dispersed inorganic adsorbing particles)</td>
<td><img src="image3" alt="Mixed Matrix Membranes" /></td>
<td>Hybrid mechanism: solution-diffusion + surface flow, possibility to overpass trade-off limitations</td>
</tr>
<tr>
<td>Zeolite or molecular sieve membrane (pore diameter close to or smaller than molecules kinetic diameter)</td>
<td><img src="image4" alt="Zeolite or molecular sieve membrane" /></td>
<td>Molecular sieving, very high selectivity theoretically achievable providing that defect free materials can be produced at a large scale</td>
</tr>
<tr>
<td>Liquid Membranes (LM), Fixed Site Carrier Membranes (FSCM): liquid impregnated in a porous membrane or dissolved in a dense polymer</td>
<td><img src="image5" alt="Liquid Membranes" /></td>
<td>Reaction with a mobile or fixed carrier molecule + diffusion in a liquid. High selectivity achievable. Usually requires water</td>
</tr>
</tbody>
</table>
separations, permeance is often expressed in GPU (Gas Permeation Unit).

An overview of the main different materials investigated for CO₂/N₂ separation and their performances are summarized in Tables 2-5. Selectivity and permeance are often antagonistic and an inverse relationship is obtained when they are simultaneously reported on a graph (Robeson Lloyd, 2008). Figure 2 presents the CO₂/N₂ selectivity (α) data for different membrane materials versus CO₂ permeance. The trade-off curve, obtained through theoretical calculation and specific to dense polymers only, has been added (Freeman, 1999). The most relevant advances in each class of membrane materials: polymeric, inorganic, hybrid organic-inorganic and facilitated transport membranes are detailed hereafter.

2.1 Polymeric Membranes
Commercial gas separation membranes include mainly: polyimide, polysulfones, polycarbonates, polyphenyl

<table>
<thead>
<tr>
<th>TABLE 2</th>
<th>Selected high performing membrane materials-data for dense polymeric membranes</th>
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</thead>
<tbody>
<tr>
<td>Membrane type</td>
<td>CO₂/N₂ selectivity (α)</td>
</tr>
<tr>
<td>Polymers of intrinsic microporosity (PIMs)</td>
<td>25</td>
</tr>
<tr>
<td>Polymers of intrinsic microporosity (PIMs)</td>
<td>40</td>
</tr>
<tr>
<td>Composite</td>
<td>50</td>
</tr>
<tr>
<td>Block copolymers</td>
<td>75</td>
</tr>
<tr>
<td>Block copolymers</td>
<td>60</td>
</tr>
<tr>
<td>Block copolymers</td>
<td>58</td>
</tr>
</tbody>
</table>

* Indicates that CO₂ permeance is estimated from membrane permeability, assuming a 1 μm dense layer thickness.

<table>
<thead>
<tr>
<th>TABLE 3</th>
<th>Selected high performing membrane materials – data for inorganic and hybrid organic-inorganic membranes</th>
</tr>
</thead>
<tbody>
<tr>
<td>Membrane type</td>
<td>CO₂/N₂ selectivity (α)</td>
</tr>
<tr>
<td>Zeolites</td>
<td>&gt;500</td>
</tr>
<tr>
<td>Zeolites</td>
<td>200</td>
</tr>
<tr>
<td>Zeolites</td>
<td>69</td>
</tr>
<tr>
<td>Silicas</td>
<td>50</td>
</tr>
</tbody>
</table>
oxides, cellulose derivatives and poly(ethylene oxide) (Ebner and Ritter, 2009; Luis et al., 2012). Numerous studies are reported on polymeric membrane materials attempting to improve their performances for CO$_2$/N$_2$ separation, which logically almost systematically fall below the trade-off curve (Fig. 2). For dense polymers, a composite membrane based on a segmented block copolymer (Polaris$^\text{TM}$), a type of poly(ethylene oxide) co-polymer, has been specifically designed for carbon capture applications up to the industrial scale (spiral wound modules) (Merkel et al., 2010). This material shows a CO$_2$ permeance of 1 000 GPU with a CO$_2$/N$_2$ selectivity of 50. Another example of block copolymers is poly(ethylene oxide) poly(butylenes terephtalate) block copolymers with a selectivity of 75 and a CO$_2$ permeance of 1 000 GPU (Metz et al., 2004). Up to day, these

<table>
<thead>
<tr>
<th>Membrane type</th>
<th>CO$_2$/N$_2$ selectivity ($\alpha$)</th>
<th>Membrane CO$_2$ permeance (GPU)</th>
<th>Membrane</th>
<th>Reference</th>
</tr>
</thead>
<tbody>
<tr>
<td>MOFs</td>
<td>30</td>
<td>&gt;30 000</td>
<td>MgMOF-74, theoretical value, $\varepsilon = 50$ µm</td>
<td>Krishna and van Baten (2012)</td>
</tr>
<tr>
<td>Mixed matrix membrane</td>
<td>72</td>
<td>154</td>
<td>(PEBAX)-Silica (90:10)</td>
<td>Kim and Lee (2001)</td>
</tr>
<tr>
<td>Mixed matrix membrane</td>
<td>118</td>
<td>205</td>
<td>PEBAX-Silica (81:19)</td>
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<th>Membrane CO$_2$ permeance (GPU)</th>
<th>Membrane</th>
<th>Reference</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fixed Site Carrier Membranes (FSCM)</td>
<td>150</td>
<td>&gt;300</td>
<td>Polyvinyl amine (PVAm) and polyvinyl alcohol (PVA) blend membrane. Selective layer thickness of 0.3-0.7µm</td>
<td>Deng et al. (2009); Deng and Hägg (2010); Matsuyama et al. (1999)</td>
</tr>
<tr>
<td>Fixed Site Carrier Membranes (FSCM)</td>
<td>120</td>
<td>320</td>
<td>Fixed carrier in a solvent-swollen membrane Water-swollen chitosan (poly(D glucosamine) membrane</td>
<td>El-Azzami and Grulke (2008)</td>
</tr>
<tr>
<td>Fixed Site Carrier Membranes (FSCM)</td>
<td>500</td>
<td>&gt;6 000 B</td>
<td>Poly (allylamine) as a fixed carrier 2-aminoisoobutyric acid-potassium salt as a mobile carrier in a crosslinked PVA matrix</td>
<td>Huang et al. (2008)</td>
</tr>
<tr>
<td>Fixed Site Carrier Membranes (FSCM)</td>
<td>500</td>
<td>840</td>
<td>Composite membranes (coating PPO and PSf with high molecular PVAm)</td>
<td>Sandru et al. (2010)</td>
</tr>
<tr>
<td>Fixed Site Carrier Membranes (FSCM)</td>
<td>&gt;500</td>
<td>&gt;1 800</td>
<td>PVAm composite membranes with PH control</td>
<td>Kim et al. (2013)</td>
</tr>
<tr>
<td>Fixed Site Carrier Membranes (FSCM)</td>
<td>800</td>
<td>300</td>
<td>Amine-modified mesoporous silica membranes</td>
<td>Barillas et al. (2011)</td>
</tr>
<tr>
<td>Liquid membrane</td>
<td>200</td>
<td>100</td>
<td>DEA supported on poly (vinyl alcohol) membranes</td>
<td>Francisco et al. (2010)</td>
</tr>
<tr>
<td>Liquid membrane</td>
<td>820</td>
<td>52</td>
<td>Enzyme supported on polymer</td>
<td>Zhang et al. (2010)</td>
</tr>
</tbody>
</table>
performances can be considered as the upper limit for dense polymeric membranes based on a physical separation mechanism.

2.2 Inorganic Membranes

Inorganic membranes are very attractive because they can potentially show better performances and resistance against high temperature and pressure conditions in comparison to polymeric membranes. Classical materials include carbon, alumina, zeolites and silicas (Ebner and Ritter, 2009). Two outstanding results have been recently reported for CO\textsubscript{2}/N\textsubscript{2} separation on tailor made materials at lab scale. A selectivity of 200 and permeance of 200 000 GPU has been obtained based on molecular sieve separation mechanism with a NAY zeolite membrane (Krishna and van Baten, 2011). A higher selectivity of 500 and a CO\textsubscript{2} permeance of 300 GPU have been reported with another type of zeolite membrane (White et al., 2010). It can be seen in Figure 2 that these performances stand far above the trade-off curve of dense polymers.

2.3 Hybrid Organic/Inorganic Membranes

These membranes consist of mixtures of organic and inorganic phases. Classically, inorganic particles (adsorbents) are dispersed into a dense polymeric matrix (Tab. 4). The corresponding material is often named Mixed Matrix Membrane (MMM). This strategy offers the possibility to overcome the permeability/selectivity trade-off of polymeric membranes, as shown in Figure 2. The dispersed inorganic phase, which may consist of various types of materials (silicas, zeolite), may act as a molecular sieve or as a Selective Surface Flow (SSF) material. In other cases, the two phases interactions can open inter-chain distance improving both selectivity and permeability. The inorganic phase can also improve both the mechanical and thermal resistance of the polymeric phase. New organic-inorganic hybrid materials based on metal cation sites in porous crystalline phase have also been developed. This class of adsorbent particles is called Metal Organic Frameworks (MOF). Molecular simulations have recently been reported with the possibility to get MgMOF-74 membranes with a permeance greater than 30 000 and a selectivity of about 30 (Krishna and van Baten, 2012) higher selectivity of about 118 is reported for a mixed matrix membrane based on PEBAX-Silica (81:19) with a CO\textsubscript{2} permeance of 205 GPU (Kim and Lee, 2001).

2.4 Facilitated Transport Membranes (FTM)

Facilitated Transport Membranes (FTM) are based on a selective reversible reaction between the incorporated carrier agents and the target gas component. It should be stressed that for these membranes the transport is different from the solution-diffusion mechanism which takes place in dense polymeric membranes. These membranes are known to show high selectivities together with high fluxes (Ebner and Ritter, 2009; Luis et al., 2012; Ramasubramanian and Winston Ho, 2011; Xomeritakis et al., 2005). They can be divided into two categories: Fixed Site Carrier Membranes (FSCM) where the carriers are chemically bonded to the polymer and mobile site carrier membranes, usually named Liquid Membranes (LM) where the carrier can diffuse in the membrane. The two membrane categories have shown attracting performances for CO\textsubscript{2}/N\textsubscript{2} mixture. Recently, a permeability above 6 000 Barrer and a selectivity of about 500 has been reported for a membrane based on poly (allylamine) as a fixed carrier and 2-aminoisobutyric acid-potassium salt as a mobile carrier in a crosslinked PVA matrix (Huang et al., 2008). One peculiarity of FSCM and LM is to require water for the selective reaction towards CO\textsubscript{2} to take place; this constraint does not hold for the other membrane materials. Consequently, the separation performances of reactive membranes strongly depend on humidity and a certain level of hydration has to be maintained on the two sides of the membrane for correct operation. More precisely, a classical operation with vacuum pumping on the downstream side is the more favorable possibly
with 1.5-3 bar feed pressure, and 200-300 mbar vacuum. The carriers in FTM will become saturated quicker if compression strategy is used on feed side. This specificity logically has a strong impact on process design features.

The overview of membrane characteristics and performances described above shows that impressive improvements have been recently achieved for the CO₂/N₂ separation challenge (Ramasubramanian and Winston Ho, 2011). It should be stressed however that most studies concern sample materials tested at the lab scale, with pure gases, leading to ideal selectivity data; experiments with gas mixtures and/or humid feeds, in order to better mimic real flue gas conditions, are scarce. Up to now only one tailor made material, the composite Polaris membrane based on a dense polymeric skin, has been tested, including membrane module development, up to the pilot scale on real flue gases (Merkel et al., 2010). The evaluation of the interest of more selective and/or permeable materials is obviously a key issue for materials development purposes; this target is best evaluated through simulation studies. Process design characteristics of membrane separation will now be detailed.

It is important to note that the simulation framework used in this study does not hold for FTM membranes because of the specificity of their transport mechanism as mentioned in Section 2.4 (permeability and selectivity constancy).

3 SINGLE STAGE MEMBRANE PROCESS DESIGN FOR CO₂ CAPTURE APPLICATIONS

The simulation of membrane based gas separations has been investigated for decades and the possibility to correctly predict the separation performances of a unit through a series of key equations and associated assumptions is abundantly documented (Bounaceur et al., 2006; Kaldis et al., 2000; Chowdhury et al., 2005; Coker et al., 1999; Matson, 1983; Zanderighi, 1996). Similarly to numerous studies in this field, a cross-plug flow model has been used for the simulations shown in this section. Models assumptions and mathematical model have been presented in details in our previous work (Favre, 2007).

The performances of the separation can be simulated based on numerical resolution using the DASSL (Differential/Algebraic System Solver) method (Bounaceur et al., 2006; Petzold, 1983).

Like any process design study, a series of variables has first to be defined; in a second step, a fixed number of unknowns can be identified through analytical or numerical resolution. For carbon capture studies, two key constraints, detailed before, are imposed: the CO₂ capture ratio \( R = \theta \frac{y}{x_{in}} \) corresponding to the fraction of inlet CO₂ recovered in the permeate stream and the CO₂ purity produced at the permeate side \( y \).

Compared to other carbon capture processes, membrane separations show a high parametric sensitivity and the following variables will play a key role:

- feed conditions through the inlet CO₂ content \( x_{in} \);
- driving force, expressed through the pressure ratio across the membrane \( P' \) / \( P'' \);
- material properties: the membrane selectivity \( \frac{x}{y} = \frac{P_{CO₂}}{P_{N₂}} \) and CO₂ permeance, discussed in the first part.

For post-combustion applications, the permeate flux of each compound is assumed to result from its own driving, flux coupling effects being negligible (Scholes et al., 2010; Hughes and Jiang, 1995; Dhingra and Marand, 1998). The influence of these different variables will be discussed hereafter, through a series of simulations, in link with the materials performances detailed in the previous section. For all simulations, the feed gas stream of the membrane module has been considered to be a binary dry CO₂/N₂ mixture at 30°C and under atmospheric pressure.

3.1. Interplay Between Pressure Ratio and Membrane Selectivity for a Target CO₂ Purity

In a first step, the interplay between the membrane material performances and process operating conditions is illustrated. More specifically, the influence of pressure ratio and membrane selectivity is investigated through an illustrative example for a single stage module. Results are shown in Figure 3. The permeate purity \( y \) is shown as a function of the pressure ratio \( \psi \) for different inlet CO₂ content \( x_{in} = 0.05 \) typical of a gas turbine, \( x_{in} = 0.15 \) corresponding to coal combustion, and \( x_{in} = 0.3 \) for oxygen enriched air combustion (Lie et al., 2007; Hendriks et al., 2009); CO₂ recovery ratio is fixed at 90%. The results are given for different membrane selectivity of \( x = 50, 100 \) and 200, in order to cover the range of materials selectivity performances detailed in the previous section.

First, one can remark that the required pressure ratio decreases as the desired permeate CO₂ purity increases. This induces an increased energy requirement when a higher purity target is fixed. Besides, high membrane selectivity logically permits to attain high permeate purity, for a given pressure ratio.

Also, it can be seen that at low pressure ratios (high driving force), the pressure ratio limitation is reduced and the role of membrane selectivity becomes important.
(Merkel et al., 2010; Bounaceur et al., 2006). Generally, as the pressure ratio decreases, the benefit from higher membrane selectivity increases.

Moreover, it can be seen that for a given permeate purity, the pressure ratio decreases with increasing the inlet CO₂ mole fraction ($x_{in}$). Consequently, when low inlet CO₂ concentrations are to be treated and when high permeate CO₂ purity is desired, the membrane selectivity will play a key role and energy requirement will strongly increase (Belaissaoui et al., 2012a).

More concretely, for a classical 0.9 purity target and 0.15 CO₂ inlet content (the most frequent flue gas composition), it can be noticed that a high selectivity level (around 200) is needed. Coming back to Tables 2-5 and Figure 2, one can notice that dense polymeric materials will not enable this target to be attained. As a consequence, multistage processes are needed in that case (Belaissaoui et al., 2012a). At this stage, it can be concluded from the materials point of view that the possibility to reach high selectivity levels through non polymeric membranes (Tab. 3-5) could be attractive because it offers the possibility to theoretically use a single stage unit for coal power plants.

### 3.2. Compression Strategy and Energy Requirement

It is important to note that, for a given pressure ratio such as shown in Figure 3, several possibilities are offered in terms of operating conditions. The choice between these options can be made, among others, with respect to the energy requirement, the main process selection constraint of carbon capture challenge. The driving force strategy adopted to achieve the pressure ratio across the membrane can indeed be generated by feed compression and/or permeate vacuum pumping and/or gas sweeping. Figure 4 shows a schematic diagram of these configurations.

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**Figure 3**
Permeate CO₂ mole fraction as a function of membrane pressure ratio for different inlet CO₂ concentrations ($x_{in} = 0.05-0.15-0.3$) and different membrane selectivity ($\alpha = 50-100-200$). $R = 0.9$.

**Figure 4**
Schematic representation of the different compression strategies of a single stage membrane unit for post-combustion CO₂ capture.
Generally, three major options can be proposed:

- a direct feed compression strategy is most often selected for gas separations but, for carbon capture, it corresponds to the largest energy requirement because the total feed flowrate has to be compressed (Merkel et al., 2010; Bounaceur et al., 2006);
- a vacuum pumping strategy is recommended when a minimal energy requirement is necessary. In that case, a smaller flowrate, namely the permeate flowrate, has to be pumped. For a stage cut of 0.1, the energy requirement will be in principle ten times lower than that required for vacuum pumping strategy. This option shows however several drawbacks: vacuum pumps energy efficiency can be much lower than compressors. Furthermore, the surface area will be much larger than that required for direct compression because of the lower driving force generated;
- a third possibility, consists to use an Energy Recovery System (ERS) at the retentate exit (high pressure side) of a direct compression unit. In that case, energy requirement can be close to vacuum pumping and surface requirement will correspond to direct compression. This interesting trade-off demands however an increased capital cost (i.e. ERS).

The calculation of the energy requirement of each two last strategies is detailed in Table 6.

Figure 5 illustrates the above considerations; the aggregate membrane surface area and the energy needed for vacuum pumping and compression with an ERS, such as turbo-expander are compared. The results are presented for \( x_{in} = 0.3 \), a membrane selectivity of 100, \( \text{CO}_2 \) permeance of 1 000 GPU and \( \text{CO}_2 \) permeate purity of 0.9. Energy requirement, first obtained from the electrical power of the compressor or vacuum pump, is converted in GJ per tonne of \( \text{CO}_2 \) captured, on a thermal basis (with a 2.7 work/heat factor), in order to enable a comparison with amine absorption processes. One may note that the two strategies lead to values which are similar, the vacuum pumping energy requirement being slightly lower for lower values of the recovery ratio. Nevertheless, it should be stressed that the feed compression strategy leads to a significantly lower membrane surface area than that required for the vacuum pumping strategy.

This analysis illustrates the compromise existing between energy requirement and membrane surface. It suggests the existence of an optimal recovery ratio for each strategy with a minimum total cost that could be determined with a techno-economical analysis.

### 3.3. Synthesis: Membrane Selectivity, Permeate Purity and Energy Requirement

The strong interaction between membrane selectivity, inlet \( \text{CO}_2 \) concentration, \( \text{CO}_2 \) purity and energy requirement has been discussed above. The combined relationship between these key variables is shown in Figure 6. The energy requirement (GJ/ton thermal basis) of a single stage membrane unit with feed compression + ERS is shown as a function of the permeate purity for different inlet \( \text{CO}_2 \) content of \( x_{in} = 0.05, 0.15 \) and 0.3 and \( \text{CO}_2 \) recovery ratio of 90\%.

In this figure, the benefit of higher membrane selectivity in reducing the energy requirement of the single stage membrane process is highlighted. Nevertheless, taking into account 2 GJ/ton as a key target for energy...
requirement, it can be concluded that a single stage membrane unit does not offer attractive performances for gas turbines ($x_{in} = 0.05$), whatever the material selectivity level. Clearly, alternative approaches such as hybrid or integrated systems are thus required for membrane implementation on gas turbines.

For coal combustion ($x_{in} = 0.15$), energy requirement limitations restrict permeate purity to be limited to 0.65 with a membrane selectivity of 50 (a typical performance of dense polymeric membranes). This limitation again suggest multistage or hybrid systems to be applied. With higher selectivities (such as 100 or 200, potentially achievable by fixed site carrier membranes or inorganic membranes), permeate purity can reach 0.85. This level is often considered to be too low for compression and transportation. A second membrane stage or polishing process remains probably necessary.

Finally, a single stage membrane unit offers attractive performances when inlet CO$_2$ content is around 0.3. From the energy requirement point of view, there is no need for highly selective membranes in that case. This situation is however rather uncommon in post-combustion, expect special emissions (biogas, cement, etc.) (Sheinbaum and Ozawa (1998)) or oxygen enriched air combustion.

It is worth to note that the simulated reported in this section are based on dry binary feed gas mixture CO$_2$/N$_2$. The influence of the other gas component in the flue gas e.g. should be taken into account. Indeed, more complex process design and/or lower efficiency can result from the presence of O$_2$ and H$_2$O (Zha and Rubin, 2013; Franz et al., 2013; Low et al., 2013).

4 SYNTHESIS AND CONCLUDING REMARKS

The materials and process design issues of membrane for post-combustion carbon capture have been presented and discussed above. In summary, it can be stated that despite impressive materials improvements in membrane selectivity and/or permeance, it is impossible to attain carbon capture specifications with a single stage membrane unit for the large majority of flue gases (gas turbines and coal power plants). Consequently, while numerous design studies focus on the single stage situation (Fig. 1a), there is a crucial need to tackle more complex systems such as multistage (Fig. 1b) (Merkel et al., 2010) or hybrid processes (Fig. 1c) (Favre et al., 2009; Belaissaoui et al., 2012c). Compared to the single stage case, these systems show a large number of variables, combined to numerous architecture possibilities (such as recycling loops). It is important to notice that a simple two stage membrane process without recycling loops cannot compete to amine absorption in terms of energy requirement, even with a CO$_2$/N$_2$ selectivity of 100 (Skorek-Osikowska et al., 2013). This observation shows that more sophisticated designs have to be explored for multistage membrane units. Unfortunately, very few studies addressed these problems up to now with the carbon capture constraints (Bhown and Freeman, 2011), as indicated by the limited
number of references in Table 7. A summary of the publications recently published for hybrid systems including one membrane separation step, with some key characteristics, is provided in Table 8. For each of these studies, a certain number of variables such as material performances or operating conditions has necessarily to be fixed among the numerous variables which have to be taken into account; consequently an holistic analysis of the situation is far to be achieved and a critical comparison and synthesis is extremely difficult. Similarly to other carbon capture process studies, the overall energy requirement of the system usually logically remains the objective variable number one. Nevertheless, it should be stressed that taking the energy requirement as the only objective variable can generate erroneous conclusions. A capture unit has also to attain the purity and recovery targets, and, more importantly, it has to be economically competitive.

Consequently, the ultimate and most relevant objective function of design studies should ideally be the effective cost of the capture unit, to be compared to other technologies. Unfortunately, the number of technico-economic analyses on membranes for post-combustion capture is extremely limited (Merkel et al., 2010; Herzog, 1991; Van der Sluijs et al., 1992; Zhao et al., 2008;
REFERENCES


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